ECM Evaluation Study Report

Prepared For

Alberta Innovates – Energy and Environment Solutions

November 2013

JACOBS® Consultancy

JACOBS Consultancy

205 Quarry Park Boulevard SE
Calgary, Alberta T2C 3E7 Canada
+403.258.6896 (phone)
+403.385.1543 (fax)

ECM Evaluation Study Report

Prepared For

Alberta Innovates – Energy and Environment Solutions

For Jacobs Consultancy

Rich Hill **Laura Weaver**

Contributing Authors:

Stephen Scott
Dieter Lamprecht
Dave Butler

November 2013

This report was prepared based in part on information not within the control of the consultant, Jacobs Consultancy Canada Inc. Jacobs Consultancy has not made an analysis, verified, or rendered an independent judgment of the validity of the information provided by others. While it is believed that the information contained herein will be reliable under the conditions and subject to the limitations set forth herein, Jacobs Consultancy does not guarantee the accuracy thereof. Use of this report or any information contained therein shall constitute a release and contract to defend and indemnify Jacobs Consultancy from and against any liability (including but not limited to liability for special, indirect or consequential damages) in connection with such use. Such release from and indemnification against liability shall apply in contract, tort (including negligence of such party, whether active, passive, joint or concurrent), strict liability or other theory of legal liability, provided, however, such release limitation and indemnity provisions shall be effective to, and only to, the maximum extent, scope, or amount allowed by law.

This document, and the opinions, analysis, evaluations, or recommendations contained herein are for the sole use and benefit of the contracting parties. There are no intended third party beneficiaries, and Jacobs Consultancy shall have no liability whatsoever to third parties for any defect, deficiency, error, omission in any statement contained in or in any way related to this document or the services provided.

Table of Contents

Section **Page**

Table of Contents

Section **Page**

Appendices:

Section A.

Executive Summary

Background

Alberta Innovates – Energy and Environment Solutions ("AI-EES") and an industry consortium of six companies commissioned Jacobs Consultancy Canada Inc. ("Jacobs Consultancy") and our partner, David Butler and Associates Ltd. ("Butler"), to perform a high-level commercial-scale technical and economic evaluation of electrochemical membranes ("ECM") for CO_2 capture and solid oxide fuel cells ("SOFC") for combined heat and power in a Steam Assisted Gravity Drainage ("SAGD") production application. Under AI-EES leadership, a Steering Committee was developed among the following parties to fund and direct the Study:

- AI-EES (Sponsor and Steering Committee Chair)
- BP Canada
- Cenovus Energy
- MEG Energy
- Shell Canada
- Suncor Energy (Suncor)

Dr. Viola Birss, from the University of Calgary, also participated in the Study as a technology advisor to the Steering Committee.

This report documents our methodology, basis, findings and recommendations for the application of ECM for CO_2 capture and SOFC for the generation of combined heat and power in a SAGD production application.

Problem Statement

Inherent in any SAGD complex is the need to produce heat to generate steam for the production of bitumen. The amount of steam required can vary between two barrels of cold water equivalent for every barrel of oil to four or more, depending on the reservoir. Two primary methods of generating the heat are used today:

- The first is straightforward steam generation using a boiler (usually once-through steam generator) with natural gas and produced gas as the fuel.
- The other method is generically referred to as *combined heat and power* ("CHP") and typically involves the use of a gas turbine to generate power and the use of waste heat from the turbine plus supplemental duct firing to generate steam. Power generation is

usually more than required by the SAGD plant, so the excess is sold as an export to the "grid" or another power consumer.

One of the key environmental issues plaguing both types of steam production is the generation of CO_2. Commercially available technologies to capture CO_2 are expensive, have high operating costs, and incur parasitic losses that generate additional CO_2. The result is that the avoided cost of capture can be more than $150/tonne of CO_2.

Therefore, there is interest in developing new technologies that reduce the cost of capture or generate steam more efficiently. For this Study, it is hoped that ECM provides a cost effective and technically viable means of reducing the cost of capture and that SOFC provides a better and more efficient means of CHP, while making the CO_2 generated more amenable to capture.

Study Highlights

The following list summarizes the highlights of the Study:

- Fuel cells are commercial or near commercial and are in the "Early Adopter" phase of product development
- There are a limited number of vendors for fuel cells
- Most fuel cell development has been focused on the production of power
- Solid oxide fuel cells focused on power generation are poor fits for low-power, high-heat-demand operations such as SAGD
- ECM fuel cells that concentrate CO_2, and incidentally, produce power appear to offer significant economic advantages over conventional capture technologies for SAGD complexes
- ECM is estimated to offer carbon capture cost in the range of $40/t as opposed to conventional technologies with costs well over $100/t. On an avoided CO_2 basis the benefit for ECM is even larger. For the avoided cost basis, the cost of capture is divided by the amount of CO_2 avoided. Avoided CO_2 is the difference in the CO2 emitted in the base case (without CO_2 capture) minus the CO_2 emitted in the capture case. The amount of CO_2 avoided is almost always less than the amount captured because most CO_2 capture technologies have parasitic losses which increase the amount of unrecoverable CO_2 (especially indirect CO_2 as a result of purchased power) and therefore increase, numerically, the cost of capture. ECM, on the other hand, produces power, and, therefore, it's avoided CO_2 is more than the captured CO_2 which reduces its cost of capture on an avoided basis.

- Any relatively "clean" dilute CO_2 flue gas streams appear to be candidates for capture through the use of ECM fuel cells

Study Basis

The evaluation of ECM and SOFC in this Study required significant thought in the creation of the comparison cases to provide an objective and transparent means of evaluation. Some of the questions discussed in the kick-off meeting included:

1. What is the appropriate capacity of the SAGD facility?
2. Should the evaluation include both the surface and subsurface facilities?
3. How should the co-production of heat and power contribute to the economics of either steam or power production?
4. What sets the size of the conventional cogeneration options?
5. What sets the size and capacity of ECM and SOFC?
6. What is the impact of electrical power exports regarding value and CO_2 credits?
7. Should the cases include indirect CO_2 from the generation of power, chemicals consumed and wastes generated?

After much discussion the cases were established as summarized in Table A-1:

Table A-1.
Cases

Category	Case	Description	Electricity	Steam Generation	Steam Gen (BPD. CWE)	CO_2 Capture	Amount of CO_2 Capture	CO_2 credits/ debits for Power Export / Import (MT/MWh)
OTSG Cases	**1a**	OTSG + Import Power	Imported	OTSG	99,000	---	None	Sensitivity up to .88
	1b	OTSG + Import Power + PCC	Imported	OTSG	99,000	PCC	90% of direct (incl. capture)	Sensitivity up to .88
	1c	OTSG + ECM	Imported + ECM Generated	OTSG	99,000	ECM	90% of direct (incl. capture)	Sensitivity up to .88
CHP Cases	**2a**	Cogen	Cogen	Cogen	99,000	---	None	Sensitivity up to .88

Category	Case	Description	Electricity	Steam Generation	Steam Gen (BPD. CWE)	CO_2 Capture	Amount of CO_2 Capture	CO_2 credits/ debits for Power Export / Import (MT/MWh)
	2b	SOFC	SOFC	SOFC	99,000	---	None	Sensitivity up to .88
CHP + CO_2 Capture Cases	**2c**	Cogen + PCC	Cogen	Cogen	99,000	PCC	90% of direct (incl. capture)	Sensitivity up to .88
	2d	SOFC + CC	SOFC	SOFC	99,000	by FCE	90%+ of direct (incl. capture)	Sensitivity up to .88

Regarding the questions above, we agreed on the following basis:

1. The Central Processing Facility ("CPF") is designed to produce 99,000 BPCD of steam.
2. Only the CPF is considered except for power requirements related to mechanical lift at the wells.
3. The CPF is sized to produce 99,000 BPCD of steam via main steam generation technology. Excess power is exported with a credit for sales to the grid.
4. Cogeneration is sized to produce 100% of the steam using duct firing of the Heat Recovery Steam Generators (HRSGs). For this study, two Frame 7 gas turbines and HRSGs were required.
5. The ECM in Case 1c is sized to capture 90% of the CO_2, and excess power, if any, is exported. The SOFC in Cases 2b and 2d is sized to produce all the steam necessary for the facility; excess power is exported with a credit for sales to the grid.
6. The value of export power is the same as imported power adjusted for transmission charges. The economics of the cases were analyzed under various power prices. The CO_2 credits associated with power exports are based on sensitivities up to 0.88 Mt/MWh.
7. Electrical consumption at the site contributes to indirect CO_2 emission and will be included with direct CO_2 emissions for this study.

Methodology

At a high level, Jacobs Consultancy and Butler performed the following main activities:

1. Estimated the technical information and all the balances for each of the cases as follows:
    a. Based on in-house information, developed the technical data and balances for the conventional SAGD CPFs for Cases 1a, 1b, 1c, 2a, and 2c.
    b. Using publicly available information, developed the technical data and balances for the conventional amine-based post-combustion capture for Cases 2c and 2d.
    c. Entered into a non-disclosure agreement with Fuel Cell Energy, Inc. ("FCE") to obtain information for ECM in Option 1c, SOFC in Option 2b, and SOFC and carbon capture in Option 2d.
    d. Transmitted design parameters for the ECM and SOFC to FCE.
    e. Sense-checked the data and balances from FCE and documented in the technical summary.
2. Designed a technical summary sheet to compare each of the cases and provided the necessary inputs for the economic calculation.
3. Set up Butler's proprietary economic model to calculate the metrics established by the Steering Committee.
4. Calculated the technical and economic metrics.
5. Assessed the ECM and SOFC technologies for technical readiness.
6. Documented findings and provided recommendations for the Steering Committee.

New Technology Readiness Assessment

Our technology readiness assessment is based on a 9-step Technology Readiness Level ("TRL") category system developed by NASA and modified for process plant applications where a TRL 1 is a concept and TRL 10 is commercial. Table A-2 below gives the TRL levels for the ECM and SOFC technologies as considered, respectively, by Cases 1a, Cases 2b and 2d.

Table A-2.
Technology Readiness

Technology	TRL	Description
ECM	5	Bench-Scale Validation Underway
Cryogenic CO_2 purification	7	System in operation at or near full commercial scale
SOFC	7	System in operation at or near full commercial scale. Most functions available for demonstration and test.
SOFC with carbon capture (oxyburn portion only)	7	System prototyping underway. Components in commercial operation.

Results

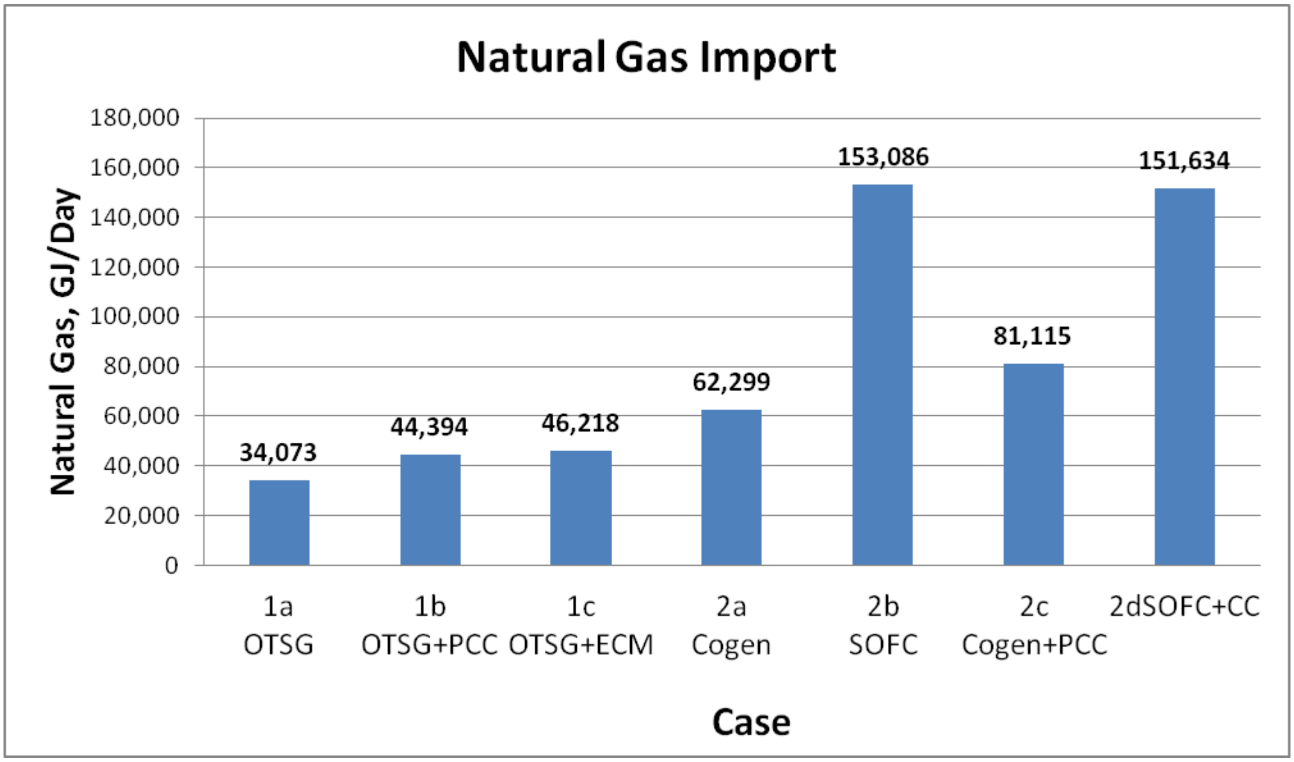
Regarding the objectives identified above, the results of our Study are as follows:

Technical Results

A few key results emerge from the technical information, as shown in Figures A-1 through A-4.

- As indicated by Figures A-1 and A-2, SAGD facilities have a large natural gas demand but are not power intensive.
- Not surprisingly, Case 1a—which is the standard OTSG case—has the lowest natural gas use.
- Case 2a, the Cogen case, consumes about 80% more natural gas than the OTSG case.
- CO_2 capture requires more natural gas than the cases without capture, except for Case 2d versus Case 2b.
- CO_2 capture with Case 2d requires less natural gas than Case 2b due to energy recovered from the oxygen combustion of the SOFC offgas.

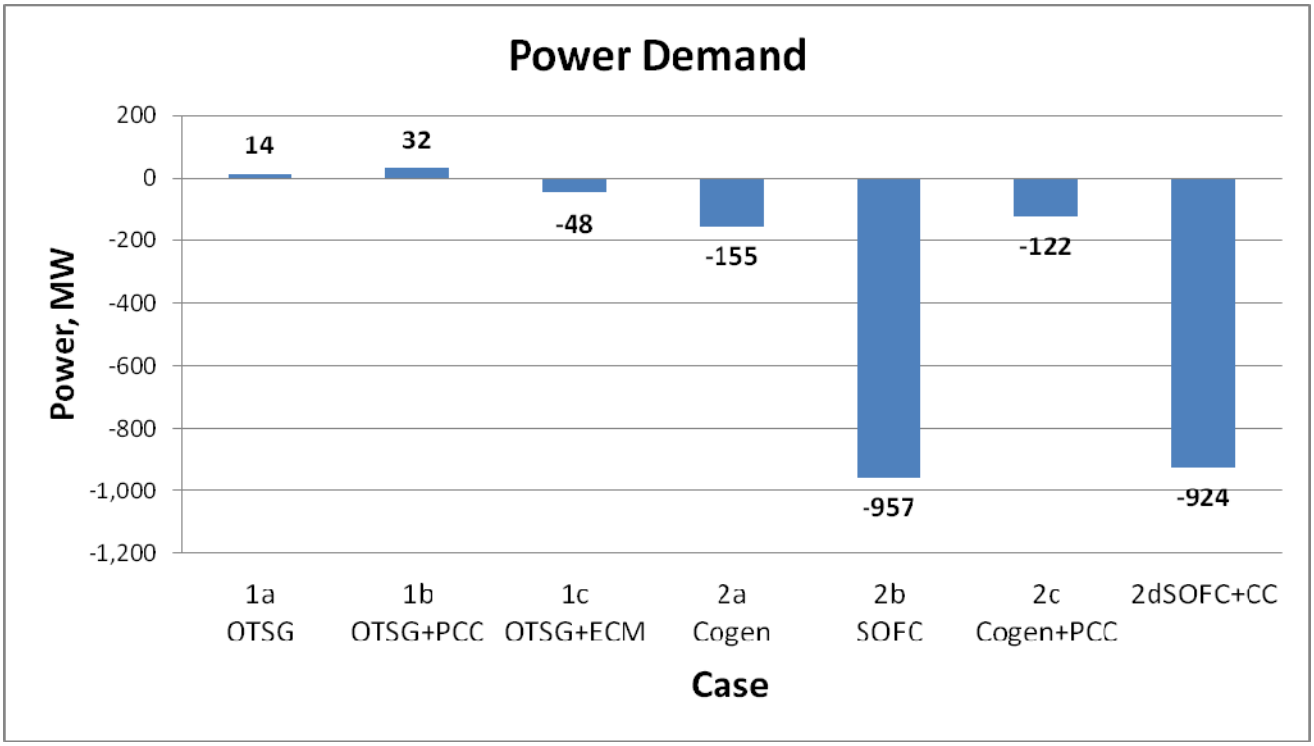
Figure A-1.
Natural Gas Import



From a power perspective, key points as shown in Figure A-2 are as follows:

- Due to relatively low internal demand for power, all cases except two (i.e., the OTSG cases, Case 1a and 1b) export power.
- Case 1c, with ECM, is unique in that the CO_2 captured in the fuel cell produces power and even provides sufficient quantity for export.
- The Cogen cases have power exports in the range of 100 to 175 MW.
- The SOFC cases have power exports in excess of 900 MW.

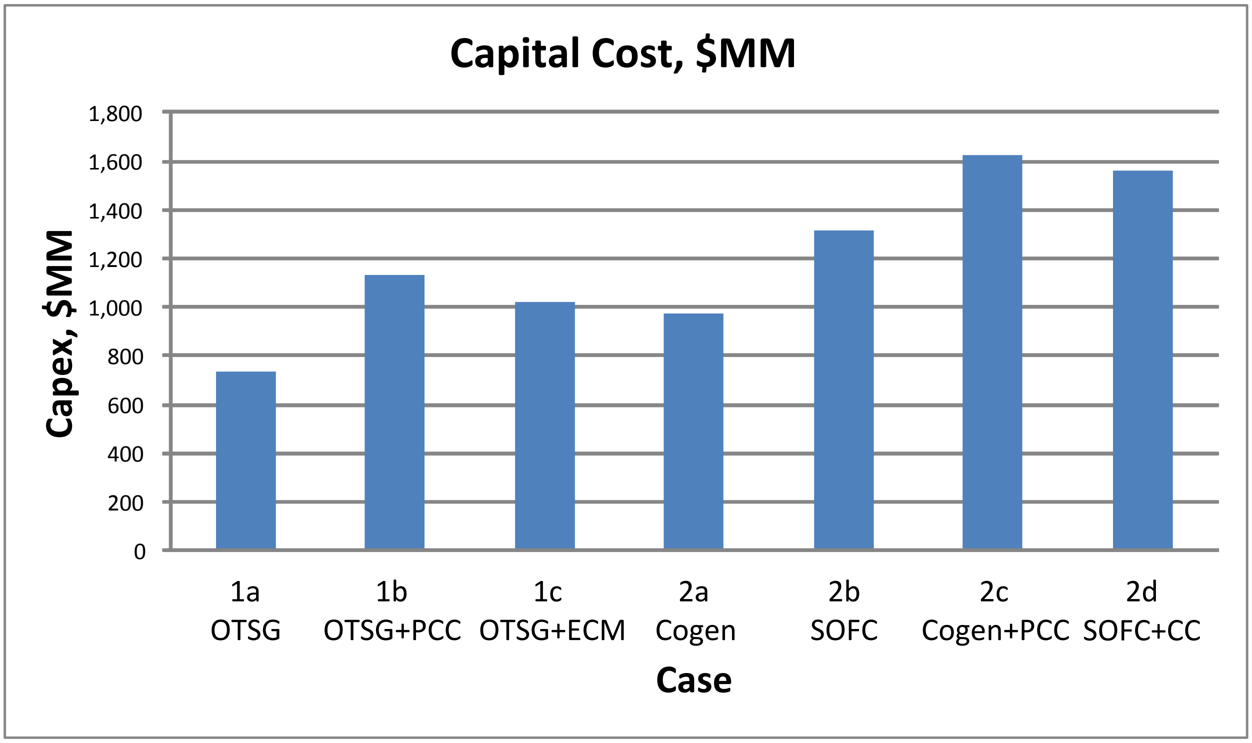
Figure A-2.
Net Power



As shown in Figure A-3, key points from a capital cost perspective are as follows:

- Not surprisingly, carbon capture universally increases the capital cost of the facilities.
- ECM, as shown in Case 1c, is less expensive than Case 1b with equivalent carbon capture.
- Cogen with post-combustion capture is more expensive than the SOFC technology cases.
- Conventional post-combustion capture is costly, adding nearly 50% to the capital costs for OTSG and Cogen.

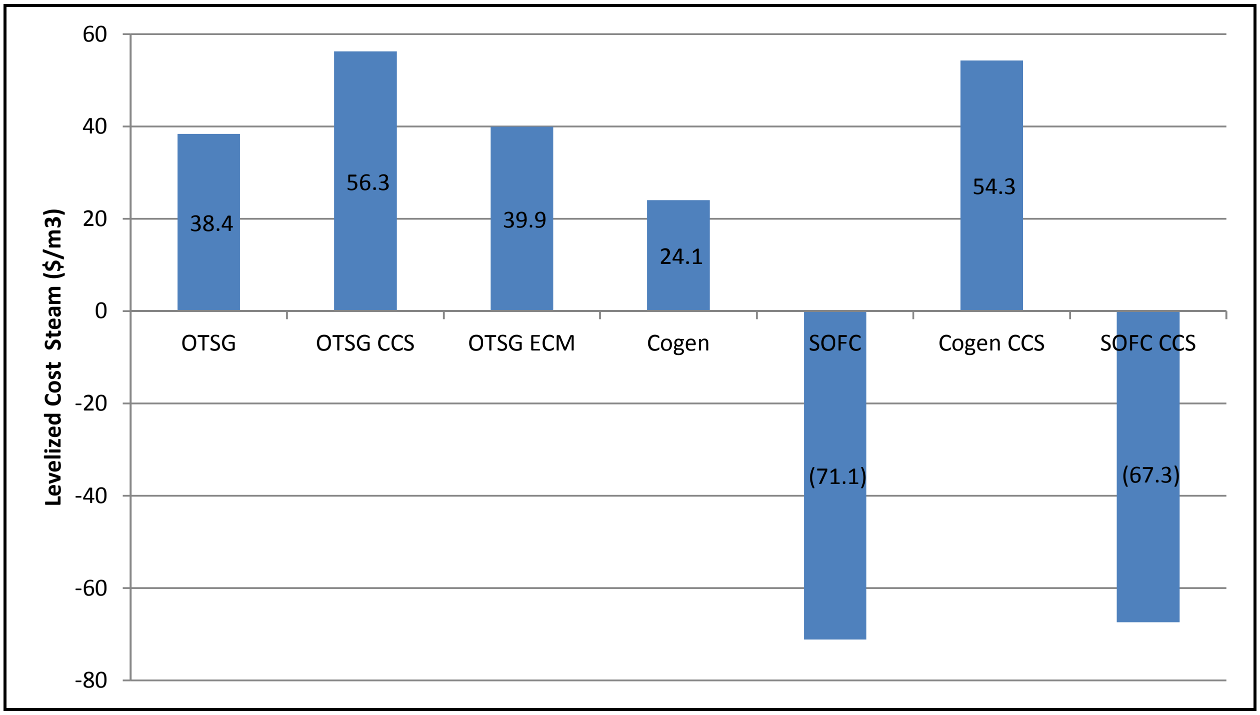
Figure A-3.
Estimated Capital Costs



Economic Results

The levelized cost of steam is the primary economic metric for comparing the various technologies. Figure A-4 below shows the net levelized cost of steam produced for the various cases. Generally one would expect that because capturing CO_2 requires additional energy and equipment cost, the levelized cost of steam for cases with lower CO_2 emissions should be higher than their base case without CCS. This is clearly seen for Cogen cases and OTSG. However, due to carbon credits (at 0.88 tonne/MWh) for the export of power, the OTSG ECM case has a levelized cost of steam comparable to an OTSG without CO_2 capture and has a much lower GHG emission intensity. Cogen without CCS offers the lowest levelized cost of steam. The SOFC case without CCS has the lowest levelized cost of steam. Adding carbon capture to the SOFC case slightly increases its levelized cost of steam. Both SOFC cases provide negative costs for steam production because of the carbon credits applied to the high amount of power production in those cases.

Figure A-4.
Net Levelized Cost of Steam



Another important economic metric is the cost of capture. Figure A-5 shows the cost to capture for the four CO_2 capture cases considered. These values were derived by assessing the cost of CO_2 capture on the same base option without CCS. The SOFC CCS case has the lowest cost of capture followed by the OTSG ECM case. The SOFC CCS case has a low capture cost because it does not cost much to purify the CO_2 produced by the fuel cell. The OTSG ECM case has a low capture cost because the incremental power produced helps offset the cost to capture CO_2.

Figure A-5.
Net CO_2 Capture Costs

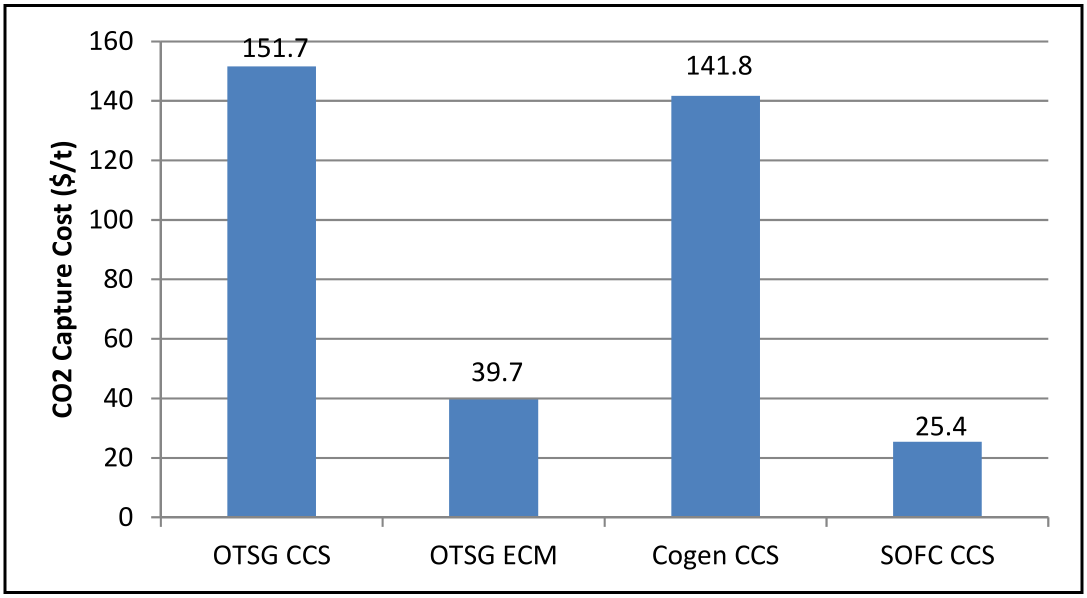
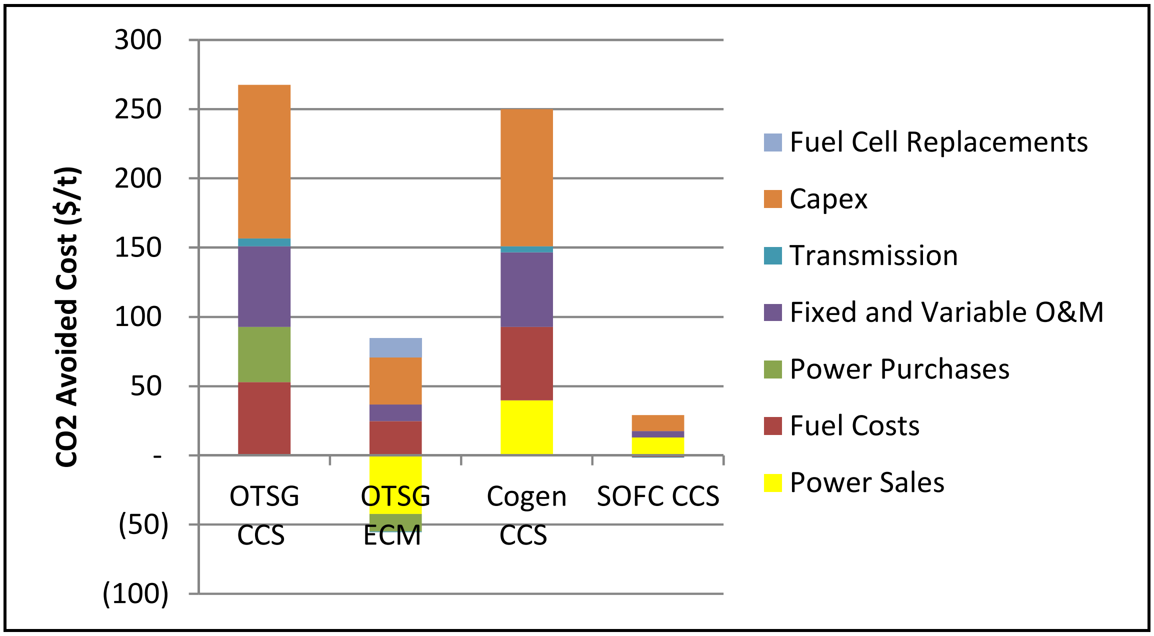


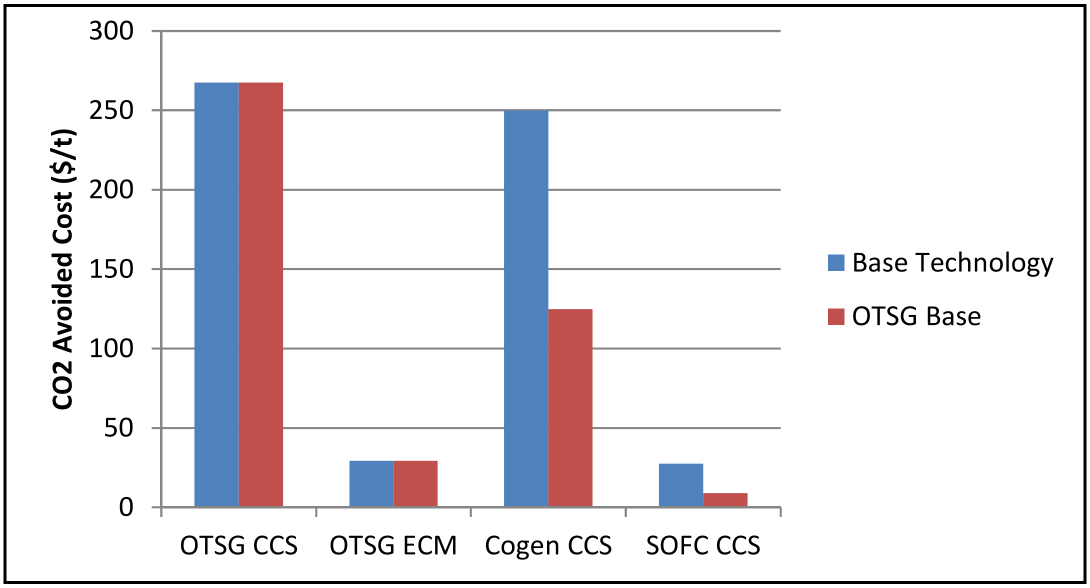
Figure A-6 shows the avoided cost of CO_2 for the four CO_2 capture options. All four options are compared against their own base technology without CCS. Therefore the OTSG CCS (using post combustion capture) and OTSG ECM is compared to the OTSG without CCS, while the Cogen CCS case is compared to the Cogen without CCS. The OTSG ECM case has a low capital and fuel cost required to capture CO_2 which is offset by the value of the incremental power produced. The SOFC CCS case incurs little cost to purify the CO_2 produced by the fuel cell.

Figure A-6.
Avoided CO_2 Costs



Avoided cost calculations are based on comparing a case with lower GHG emissions to a reference case. It could be argued that one would simply build an OTSG rather than install CCS technology. Therefore, an OTSG without CCS could be used as the reference case. The blue bars in Figure A-7 show the avoided cost where the base technology without CCS is used as the reference case. For example, an SOFC is used as the reference case for the SOFC CCS avoided costs. The red bars show the avoided costs where an OTSG without CCS is used as the reference case.

Figure A-7.
Avoided CO_2 Costs with Two Base Cases



Findings and Recommendations

The following are Jacobs Consultancy's findings regarding the use of fuel cell technologies for SAGD:

- The Solid Oxide Fuel cell is a power producer with heat as a by-product and is not well matched to thermal in-situ which has a large heat load and small power load. A typical thermal in-situ plant has a demand of about 10 MW of heat for every MW (thermal equivalent) of power. An SOFC, on the other hand, produces about one MW of high-temperature heat for every 7 MW (thermal equivalent) of power. This results in a mismatch of 70 to one as compared to the requirements for bitumen production. Any efforts to reduce the power to heat ratio on an SOFC result in bypassing fuel around the anode side and combusting it with air—which, in effect, becomes like having an OTSG to produce steam with a small SOFC to produce the power for the site. Additionally, there may be other issues such as land use and capital cost uncertainty that may negatively impact SOFC as a CHP technology for SAGD.
- However, combined heat and power ("CHP") can be attractive for a thermal in-situ site and provides a better match between heat and power. For example:
    - Conventional cogen produces steam with a cost about 40% lower than WLS+OTSG and with a manageable amount of power sold to grid.
- CO_2 capture and compression increase the cost of producing steam.

- However, fuel cells are relatively cost effective for CO_2 recovery assuming power produced can be sold to grid
  - ECM on conventional OTSG captures and compresses CO_2 at a cost that is roughly one-third of conventional PCC with manageable power sales.
  - On paper, due to **very high power sales**, SOFC has capture and compression costs that are roughly one-fourth of conventional CHP with PCC (but again, there is a mismatch of heat and power for SAGD).
- ECM is a potential "bolt-on" carbon capture technology for existing or new facilities, including:
  - OTSG / Cogen-based thermal in-situ
  - Steam Methane Reforming ("SMR")
  - Petcoke combustion
  - Other fired heaters

Our recommendations are summarized as follows:

- Focus efforts on SOFC as a power producer, not as a CHP technology for thermal in-situ plants.
- ECM within the context of Combined Electric Power and Carbon Dioxide Separation ("CEPACS") may have merit for the oil sands industry, however key risks remain, including:
  - Capital cost
  - Reliability, operability and maintenance costs
- Therefore, we recommend in the near term:
  - Feasibility studies on a commercial-scale plant to confirm capital costs and commercial viability for:
    - Cogen and OTSG-based thermal in-situ, SMR and Fired Heaters.
    - High-level screening evaluation for ECM on petcoke combustion.
    - A design study to determine location and capital cost of an ECM demonstration facility at the megawatt scale to confirm reliability, operability, and commercial readiness, and to highlight other development issues.

Section B.

Background, Scope, and Assumptions

Background

Alberta Innovates – Energy and Environment Solutions ("AI-EES") and an industry consortium of six companies commissioned Jacobs Consultancy Canada Inc. ("Jacobs Consultancy") and our partner, David Butler and Associates Ltd. ("Butler"), to perform a high-level commercial-scale technical and economic evaluation of electrochemical membranes ("ECM") for CO_2 capture and solid oxide fuel cells ("SOFC") for combined heat and power in a Steam Assisted Gravity Drainage ("SAGD") production application. Under AI-EES leadership, a Steering Committee was developed among the following parties to fund and direct the Study:

- AI-EES (Sponsor and Steering Committee Chair)
- BP Canada
- Cenovus Energy
- MEG Energy
- Shell Canada
- Suncor Energy (Suncor)

Dr. Viola Birss, from the University of Calgary, also participated in the Study as a technology advisor to the Steering Committee.

This report documents our methodology, basis, findings and recommendations for the application of ECM for CO_2 capture and SOFC for the generation of combined heat and power in a SAGD production application.

Inherent in any SAGD complex is the need to produce heat to generate steam for the production of bitumen. The amount of steam required can vary between two barrels of cold water equivalent for every barrel of oil to four or more, depending on the reservoir. Two primary methods of generating the heat are used today:

- The first is straightforward steam generation using a boiler (usually once-through steam generator) with natural gas and produced gas as the fuel.
- The other method is generically referred to as *combined heat and power* ("CHP") and typically involves the use of a gas turbine to generate power and the use of waste heat from the turbine plus supplemental duct firing to generate steam. Power generation is usually more than required by the SAGD plant, so the excess is sold as an export to the "grid" or another power consumer.

One of the key environmental issues plaguing both types of steam production is the generation of CO_2. Commercially available technologies to capture CO_2 are expensive, have high operating costs, and incur parasitic losses that generate additional CO_2. The result is that the avoided cost of capture can be more than $150/tonne of CO_2.

Therefore, there is interest in developing new technologies that reduce the cost of capture or generate steam more efficiently. For this Study, it is hoped that ECM provides a cost effective and technically viable means of reducing the cost of capture and that SOFC provides a better and more efficient means of CHP, while making the CO_2 generated more amenable to capture.

In a recent workshop on "Membrane and Ionic Liquids for Carbon Capture" organized by AI-EES, FuelCell Energy, Inc.'s ("FCE") ECM technology was identified as having a potential for low-cost carbon capture from thermal in-situ recovery processes. The ECM device has the same core component as FCE's molten carbonate fuel cell technology and has been commercialized through their DirectFuel Cell® (DFC®) product line for on-site power generation. FCE is now developing a similar process using ECM for recovery of CO_2 from flue gas streams.

In addition to FCE's ECM system, SOFCs have also been considered as a potentially advantageous technology for generating CHP. Alberta has been one of the global leaders in SOFC development. The application of SOFC in thermal in-situ processes has also been contemplated by some SOFC researchers and developers.

Scope of Study

The objectives of the Study were as follows:

- Evaluate FCE's ECM technology for CO_2 capture associated with thermal in-situ recovery.
- Evaluate SOFC technology for CHP associated with thermal in-situ recovery.
- Prepare a technology readiness assessment of FCE's two technologies.
- Compare ECM to a commercial post-combustion capture ("PCC") technology for a conventional OTSG-based production scheme.
- Compare SOFC to commercial CHP processes from both a technical and economic perspective.

The deliverables of the Study are as follows:

- Facilitate a kick-off meeting to finalize the cases for evaluation.
- Estimate and provide technical data needed for evaluation of all cases, either through in-house data, publicly available information, and/or licensor involvement.
- Prepare a technology readiness assessment of FCE's two technologies.
- Facilitate an interim review meeting to discuss preliminary results and refine plans for the rest of the Study.
- Prepare and deliver a final presentation and report.
- Facilitate a site visit for Steering Committee members to visit an operating fuel cell site.

SAGD Facility

For the purposes of the Study, each case was designed in concept to support SAGD production consistent with the following parameters:

- All SAGD units are sized for 33,000 BPCD of oil production.
- The SAGD steam-to-oil ratio ("SOR") is assumed to be 3 (based on barrels of water consumed per barrel of oil produced). This means that in all cases the CPF will produce 99,000 BPCD of dry steam (CWE basis).
- Steam is to be supplied at 100 bar and assumed to be 77% quality steam from the generator but before the high-pressure separator.
- Electric submersible pumps are assumed for lifting the bitumen from the reservoir.
- Imported electricity is supplied from the Alberta grid.
- Natural gas is supplied by pipeline.
- CO_2 produced will meet Kinder Morgan pipeline specifications at battery limits.
- The life of the plant is 30 years.
- At least 90% CO_2 capture is required for carbon capture cases.

CO_2 Calculation Assumptions

Both direct and indirect GHG emissions were considered:

- Direct GHG emission consisted of the CO_2 produced from the burning of produced gas and natural gas for steam generation.
- Indirect GHG emissions resulting from:
    - Imported power generation
    - Trucking of wastes
    - CO_2 generated in the production of imported chemicals

Comparative and Economic Assumptions

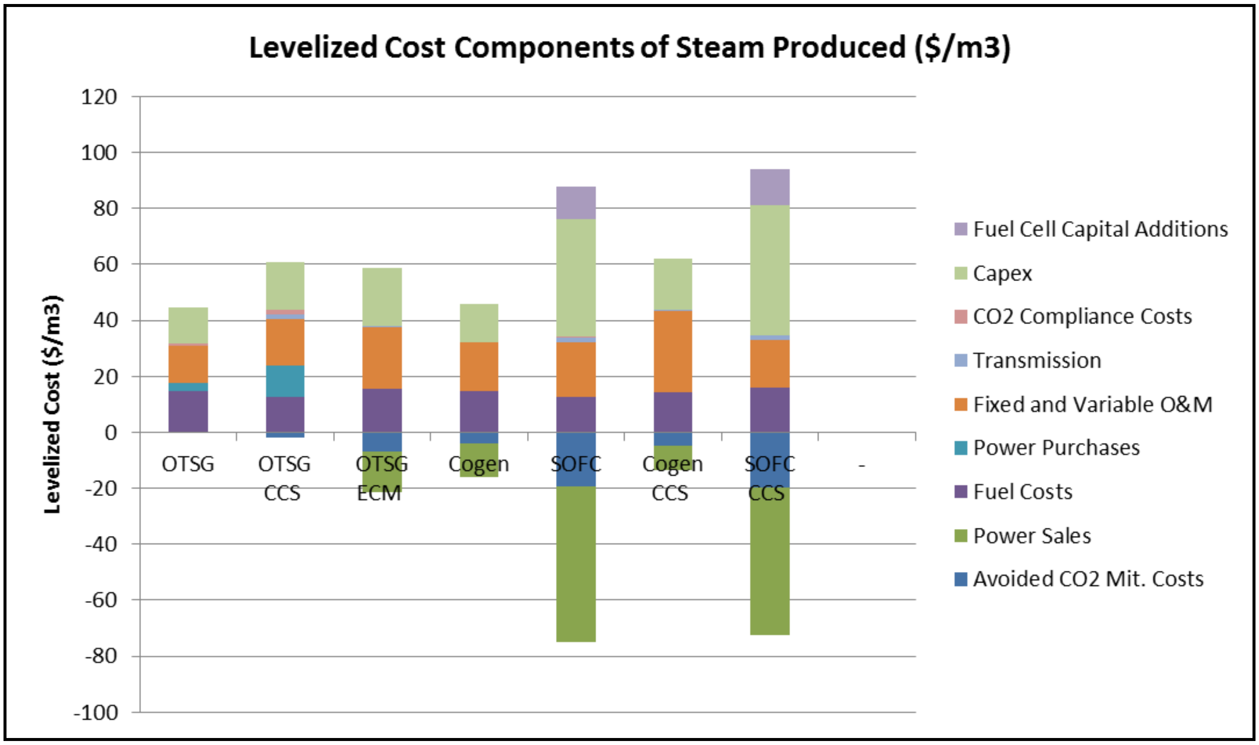
The following are the key comparative and economic assumptions for the Study as agreed to by the Steering Committee.

As the comparison of the cases was a ranking assessment as opposed to a full cost assessment, the following simplifications were made:

1. Taxes (including property taxes) or royalties are not included.
2. Un-levered analysis using a cost of capital of 10 percent.
3. The Steering Committee has agreed to use the flat levelized approach. All the underlying costs and other revenues have been escalated by 2% in a given year. The same price for steam for all years is derived by setting the NPV of a given case to zero with a discount rate of 10 percent.
4. The cost of capture will only include the capital and operating cost necessary to get the CO_2 to the plant gate at Kinder Morgan specifications.
5. Power transmission charges are 3% of net sales to AESO grid and 15% for purchases from grid.
6. Abandonment or reclamation charges are zero.

The technologies are ranked by the Levelized Cost of Steam.

Figure B-1.
Example of Levelized Cost of Steam Approach



The following commodity prices and other operating costs were used for the Study:

1. Power Price Forecast: $90/MWh escalating 2% per year
2. Gas Price Forecast: $5.00/GJ escalating 2% per year
3. Inflation and Escalation: 2% per year
4. Water has no charge once permitted for use
5. Water treating chemicals cost roughly $0.77/m^3 of water treated
6. MEA costs (for PCC): $2.40/kg
7. Corrosion inhibitor (for PCC): $7.20/kg
8. Mole sieve: $4.80/kg
9. WLS sludge disposal: $150/tonne
10. MEA solid waste disposal: $1,400/tonne
11. Other fixed and variable operating costs are 3.5% of capital cost

The primary focus of the Study was to understand how the new fuel cell technologies perform relative to commercially available processes on the cost of carbon capture. Therefore, the following conventions and sensitivities were included in the analysis:

1. **CO_2 Credits**—If a case has no carbon capture then it will be required to mitigate 12% of its CO_2 emissions at $15/t. The cases with carbon capture should be able to sell the volume of CO_2 captured less 12% of the plant CO_2 production as a CO_2 credit. (It can also be used internally to avoid having to purchase credits elsewhere.)
2. **Avoided CO_2 Costs**—OTSG without capture will be the reference case for the OTSG cases (1b and 1c) and the base technology without CCS for the other cases. Avoided costs were also completed using the OTSG without capture as the reference case. The GHG emissions associated with this additional energy required to capture CO_2 are netted out of the mass of CO_2 captured to determine the mass of CO_2 avoided. Therefore:

   CO_2 Avoided = CO_2 Captured – CO_2 for power from the grid used to capture CO_2 + CO_2 credits for power produced - CO_2 for incremental fuel.

   The avoided cost is determined by taking the incremental cost to produce steam in a given year, as derived above and dividing it by the mass of CO_2 avoided. Some entities are more familiar with the avoided mass of CO_2 expressed as:

   CO_2 Avoided = CO_2 emitted in reference case without CCS – CO_2 emissions with CCS – CO_2 from power used from grid + CO_2 credits for power produced.

   CO_2 emissions with CCS would include CO_2 emissions produced to provide steam used to capture CO_2.

To understand the sensitivity of the case rankings and economics to various factors, the following sensitivities were included in the economic results:

1. Natural gas prices: $2 to $9/GJ, escalating each year by inflation
2. Technology life of fuel cells: 3 to 10 years
3. Capital cost: -15, 0, +50, +100 of base capital cost of the CO_2 capture technology
4. Power price effect, power prices at $50 to $110/MWh escalating by inflation
5. CO_2 credit value/compliance cost: $0/te, $15/te, $30/te, $40/te, and $80/te
6. Show impact of GHG intensity for power ranging from 0 to 0.88 t/MWh
7. Increase in the capital cost of SOFC cases by up to 5 times
8. 100% increase in the capital cost for carbon capture

Capital Cost Estimate Basis

Cost Curve and Unit Rate Basis

The curve costs used in some of the cases were developed from Jacobs Engineering Group's ("Jacobs") in-house cost estimating work and take into account:

- Historical construction indirect factors
- Historical home office factors
- All-in unit rates (infrastructure, gathering lines and pipelines)
    - All-in unit rates include:
        - Direct labour and material costs
        - Contractor's construction indirect costs
        - Construction management

Jacobs' Unit Cost Curves ("UCCs") were used in the preparation of the process unit costs. The UCCs are based on the following key information from past and current SAGD projects:

- Sized equipment lists
- Engineering Design Study quality (30% engineering complete) material take-offs
- Firm equipment quotes
- Current market pricing for all major commodities (second quarter 2011)
- Bulk material unit installation man-hours (actual)
- Labour costs (adjusted for specific project requirements)
    - $65.00 per direct field labour hour
    - $130.00 per module hour

Jacobs' cost estimates are based on a variety of sources as follows:

- For Cases 1a and 2a, we used UCCs for SAGD CPFs using both OTSGs and Cogen for steam generation. The cost curves are intended to be Total Installed Costs ("TIC") for various operating blocks of the plant and are based largely on capacity. We have TIC cost curves with capacity drivers for the following:
    - Oil Treating—emulsion flow rate

  - De-oiling—produced water rate
  - Water Treating—Boiler Feed Water rate and Lime Softening feed
  - Steam Generation—Dry steam rate
  - Hydrocarbon Tankage—hydrocarbon hold-up volume
  - Excavation—approximate CPF land area
  - Utilities and Offsites—factored from dil-bit rate

  Other costs, home office, and engineering for the CPF are typically factored from the sum of the TIC of the operating blocks.

- For Cases 1b and 2c, we prepared our estimate from the cost curve method for Cases 1a and 2a and then added data for post-combustion capture using publicly available data on Fluor's Econamine process. For convenience, we adjusted all costs for PCC to a USGC basis and then used a location factor of 1.5. The publicly available sources for PCC performance and costs include:
  - David and Herzog—2012 (the cost of carbon capture)
  - Chapel and Mariz—NETL 1999
  - Global CCS Institute (Economic Assessment of Carbon Capture and Storage)—2011 update
  - DOE/NETL—401/110907
  - DOE/NETL—401/110509
  - DOE/NETL—402/102309
  - Simmonds, Hurst, Wilkinson, Reddy, Khambaty—May 2003
  - ROAD | Maasvlakte CCS Project C.V.—November 2011
  - US EIA—November 2011—Updated Capital Cost Estimates for Electricity Generation Plants
- For Case 1c, we prepared our capital cost estimate from the cost curve method for Case 1a and then added a TIC for the ECM, which was prepared based on an equipment cost from FCE and our installation costs for modular equipment.
- For Cases 2b and 2d, the estimate was prepared based on FCE's future expected equipment costs (assuming substantial increases in the number of fuel cells sold) and our installation costs.

Alberta Labour Market

The estimate is based on the Alberta labour market using the key assumptions shown in Table B-1:

Table B-1.
Labour Assumptions

Description	All Cases
Pricing Basis	1Q 2013
Workforce	CLAC
Construction Schedule (Construction to Mechanical Completion)	22 months
Work Schedule	10/4
All-in Field Rate Build-Up:	
(1) Direct Labour Wage Rate	65.00$
(2) Contractor Indirects (Incl. Fee)	92.10$
(3) Camp (Per Direct Hour)	33.44$
(4) CM	22.50$
All-in Rate (1+2+3+4):	***213.04$***
Scaffolding (% OF AG Hours)	26%
Productivity	1.35
Modularization %	40%
Modularization All In Rate	130.00$

Inclusions / Exclusions

The costs for each case include the major processing blocks mentioned above as well as the following cost components:

- Mechanical equipment
- Electrical equipment
- Bulk materials
- Direct field labour costs
- Contractor construction indirects (including contractor fee, overhead, and minor craft attraction incentives)
- Construction management costs
- Material-related costs (freight, module transportation, vendor assistance)

- Home office engineering (FEED, detail design, home office construction support and third party engineering)
- Contingency

The following key items are **excluded** from the estimate:

- Owner’s costs
- Pre-FEED costs
- Incurred costs
- Field facilities (well pads, gathering lines)
- LACT (lease automatic custody transfer)
- Main Substation
- HV incoming power cables and associated HV tie-ins
- Access roads
- Off-sites (pipelines, salt cavern)
- Disposal facilities/pond
- Commissioning and start-up costs
- Drilling and completions
- Any process unit not explicitly listed (Cogen, DRU)
- Demolition and disposal
- Escalation
- Foreign labour
- Construction workforce fly/in and fly/out

Contingency (Engineering, Procurement, Construction Risk)

For most of cases contingency was fixed at 20%. However, we did perform some sensitivity analysis of the economics at other capex points for the new technology options. Contingency for this type of estimate accounts for the following:

- Minor process unit capacity changes
- Minor construction work-hours (productivity) and labour pricing fluctuations
- Engineering work-hours and services pricing fluctuations

- Minor schedule delays
- Minor estimating errors and omissions
- Minor impacts to costs due to client-specific design requirements not accounted for in the Jacobs cost curves and historical unit rates
- Minor impacts to costs due to project-/client-specific execution requirements not accounted for in the Jacobs cost curves and historical unit rates
- Minor pricing fluctuations due to supply and demand (outside of typical escalation)
- Minor risks associated with site soil conditions (e.g., site grading, muskeg quantities)

Not Covered by Contingency

- Technical cost risk associated with new and emerging technologies
- Project location changes
- Major process unit capacity changes
- Product specifications
- Major material price fluctuations (extreme change in market conditions)
- Major schedule delays
- Scope changes / additional scope including:
    - Plant capacity changes
    - SOR changes
    - Technology changes
    - Increases in number of pipelines or pipeline lengths, sizes and thickness
    - Increases in gathering line lengths, sizes and thickness
    - Additional gathering lines (only emulsion / steam lines have been included in estimate)
    - Changes in number of well pairs
    - Costs associated with ROW sharing or crossing of other Owner organizations
- Major estimating errors and omissions
- Major impacts to costs due to client-specific design requirements not accounted for in the Jacobs cost curves and historical unit rates
- Major impacts to costs due to project-/client-specific execution requirements not accounted for in the Jacobs cost curves and historical unit rates
- Major risks associated with site soil conditions (e.g., site grading, muskeg quantities)

Section C.

Case Descriptions

Case Development

Two groups of cases were agreed to at the kickoff meeting to compare the FCE technologies in both a steam only SAGD CPF as well as a CHP CPF. Between the two groups, a total of seven cases were developed:

- **Group 1—Heat/Steam Only Cases**
    - Case 1a—Warm Lime Softening ("WLS") + OTSG without carbon capture ("CC")
    - Case 1b—WLS+OTSG with commercially available (amine based) post-combustion capture ("PCC") to capture at least 90% of the direct CO_2
    - Case 1c—WLS+ OTSG with ECM for CC to capture at least 90% of the direct CO_2 (capture percentages in excess of 90% may be necessitated by what is practical for ECM)
- **Group 2—Combined Heat and Power ("CHP") Cases (sized to produce 100% of the steam required)**
    - Case 2a—WLS + gas turbine ("GT") + heat recovery steam generator ("HRSG") without CC
    - Case 2b—WLS + SOFC without CC
    - Case 2c—WLS + GT + HRSG with PCC to capture at least 90% of direct CO_2
    - Case 2d—WLS + SOFC with CC to capture at least 90% of direct CO_2

The cases are summarized in Table C-1 below:

Table C-1.
Cases

Category	Case	Description	Electricity	Steam Generation	Steam Gen (BPD. CWE)	CO_2 Capture	Amount of CO_2 Capture	CO_2 credits / debits for Power Export / Import (MT/MWh)
OTSG Cases	1a	OTSG + Import Power	Imported	OTSG	99,000	---	None	Sensitivity up to .88
	1b	OTSG + Import Power + PCC	Imported	OTSG	99,000	PCC	90% of direct (incl. capture)	Sensitivity up to .88
	1c	OTSG + ECM	Imported + ECM Generated	OTSG	99,000	ECM	90% of direct (incl. capture)	Sensitivity up to .88

Category	Case	Description	Electricity	Steam Generation	Steam Gen (BPD. CWE)	CO_2 Capture	Amount of CO_2 Capture	CO_2 credits / debits for Power Export / Import (MT/MWh)
CHP Cases	2a	Cogen	Cogen	Cogen	99,000	---	None	Sensitivity up to .88
	2b	SOFC	SOFC	SOFC	99,000	---	None	Sensitivity up to .88
CHP + CO_2 Capture Cases	2c	Cogen + PCC	Cogen	Cogen	99,000	PCC	90% of direct (incl. capture)	Sensitivity up to .88
	2d	SOFC + CC	SOFC	SOFC	99,000	by FCE	90%+ of direct (incl. capture)	Sensitivity up to .88

The following conceptual flow diagrams provide a high-level flow-scheme for each case, illustrating the flow of each major water and energy streams.

Figure C-1.
Case 1a

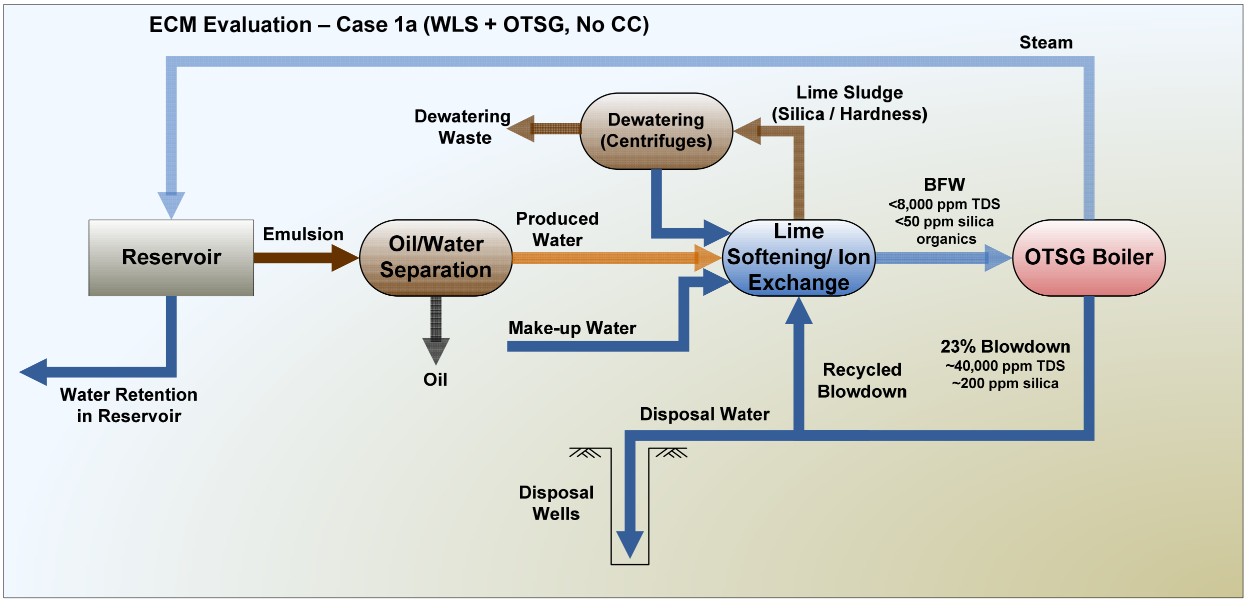


Figure C-2.
Case 1b

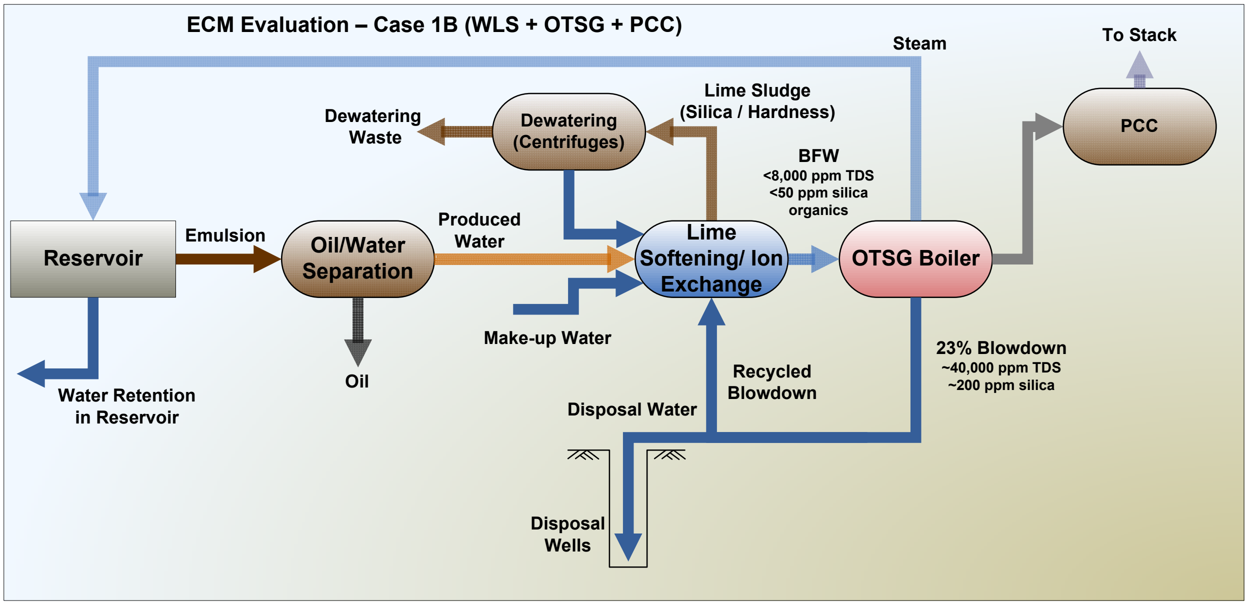


Figure C-3.
Case 1c

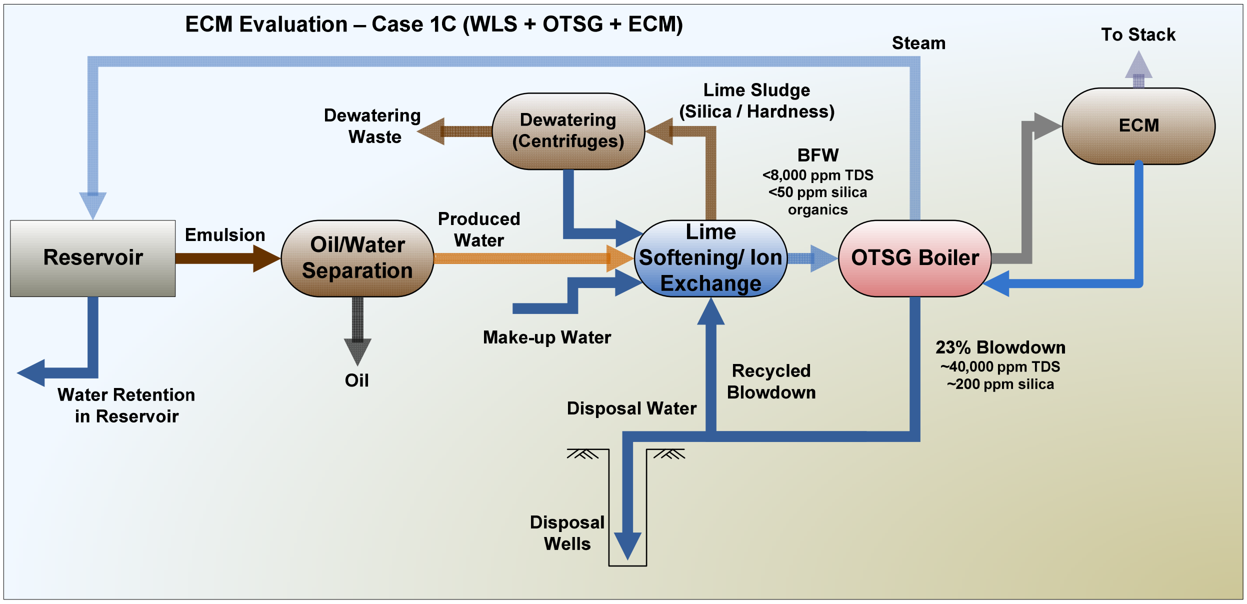


Figure C-4.
Case 2a

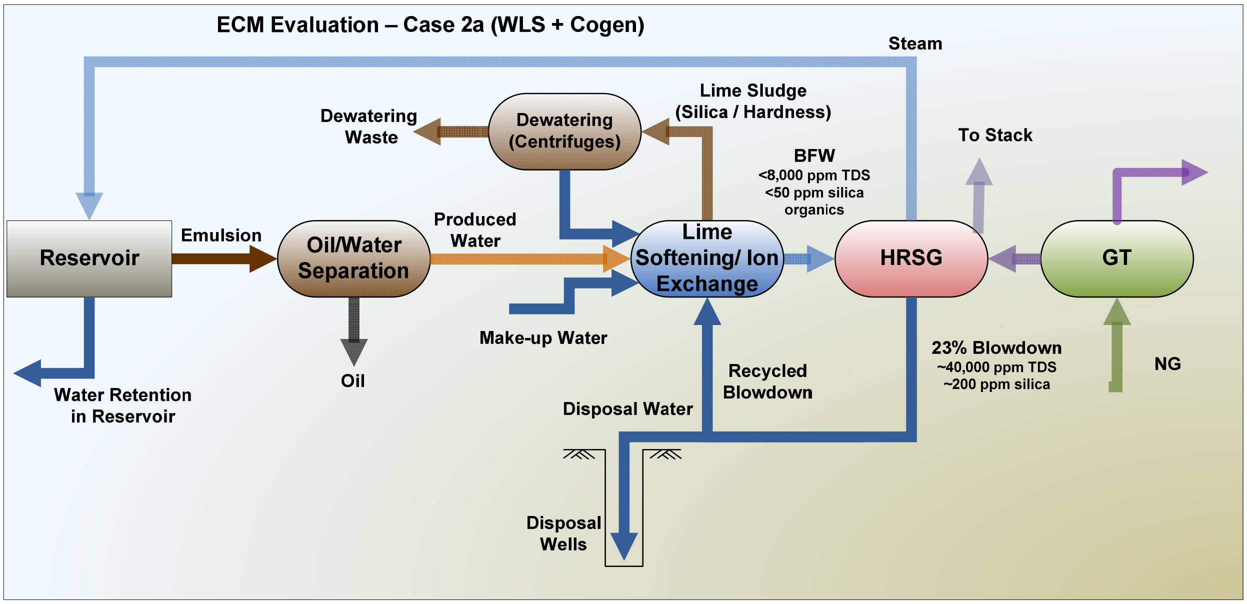


Figure C-5.
Case 2b

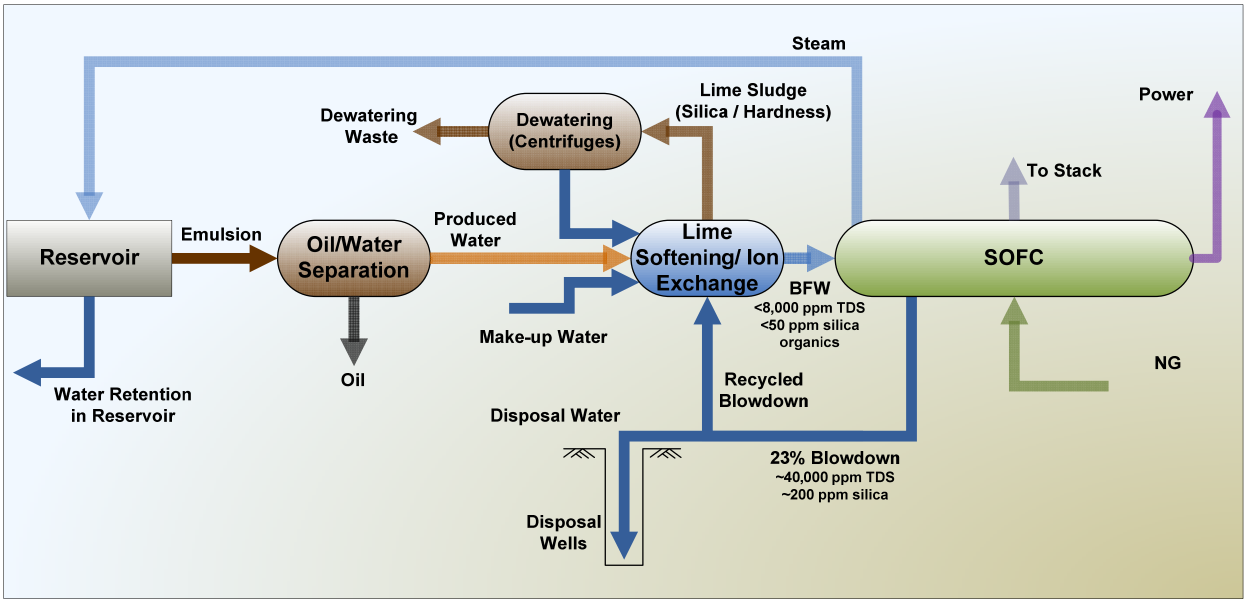


Figure C-6.
Case 2c

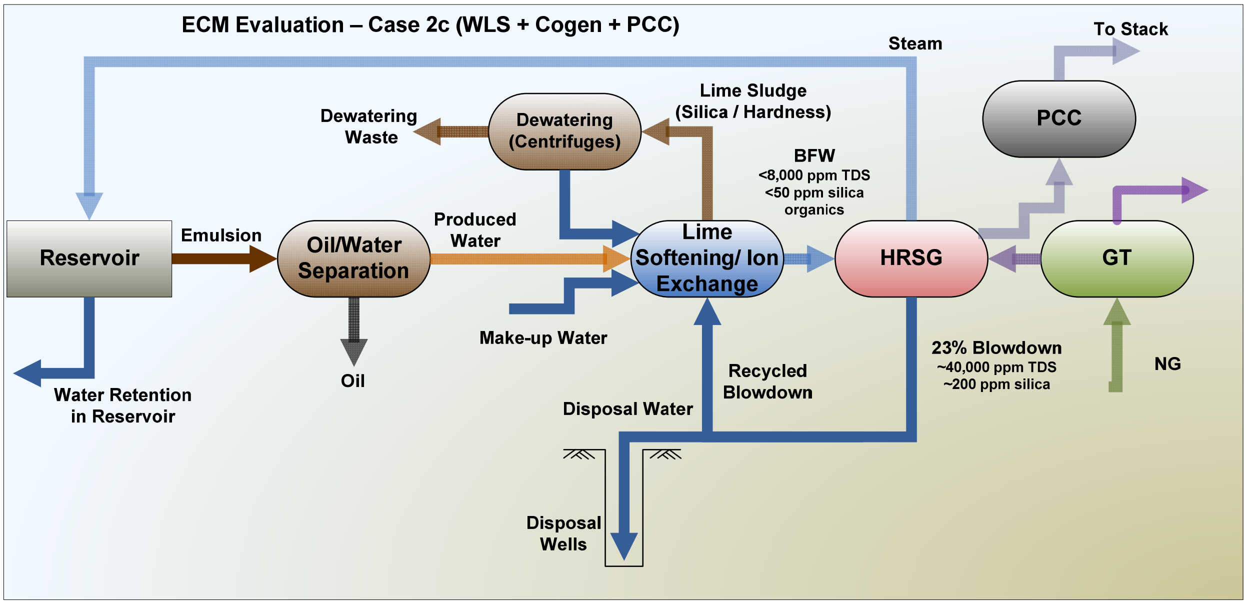
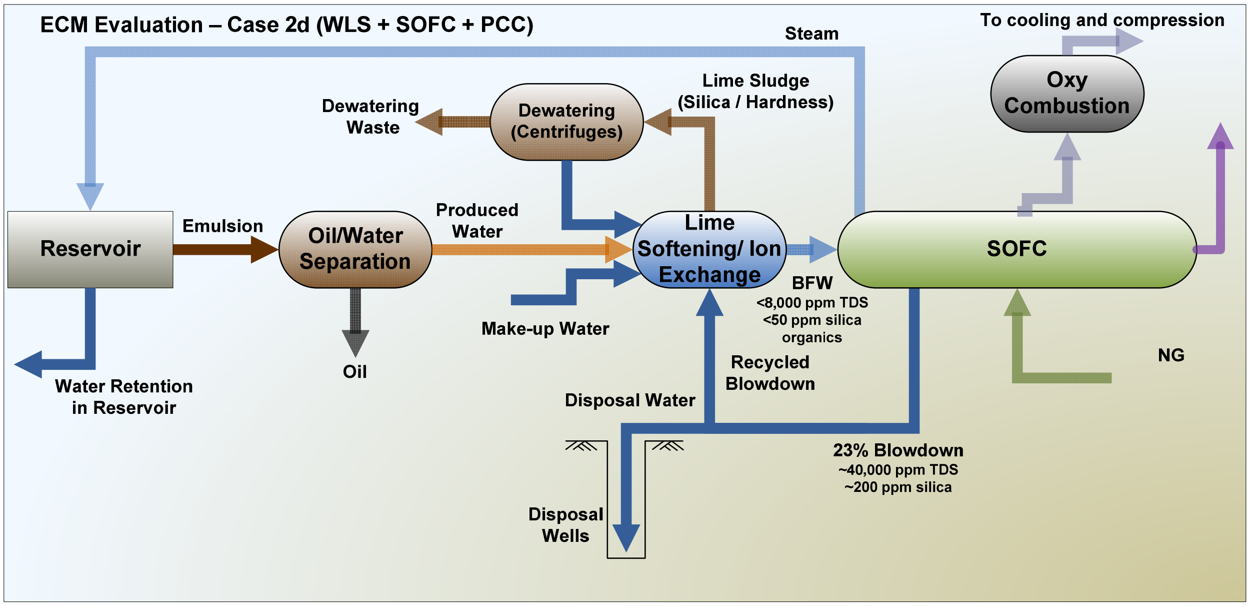


Figure C-7.
Case 2d



Section D.

Technology Description and Readiness Assessments

Combined Electric Power and Carbon Dioxide Separation (CEPACS)

Technology Background

The Combined Electric Power and Carbon Dioxide Separation (CEPACS) unit was specifically designed to produce power and concentrate CO_2 from dilute flue gases. It consists of two main sections:

- In the first section, the flue gas stream is fed to the ECM which produces a raw CO_2 containing stream. The CO_2 depleted flue gas steam is vented to atmosphere.
- In the second stage, the CO_2 containing stream is further purified and pressurized to produce pipeline quality CO_2 by compression and cooling until the CO_2 in the stream liquefies. The liquid CO_2 is separated from the residual permanent gases from the anode side of the ECM and is pumped to pipeline pressure and heated to ambient temperature by exchanging heat with the incoming CO_2 rich stream.

The ECM part of the CEPACS unit is based on FCE's commercially available Molten Carbonate Fuel Cell ("MCFC") technology.

The principle of the commercially available MCFC is that a carbon-containing fuel (usually natural gas) is supplied to the anode, and air is supplied to the cathode. On the anode side, the fuel is reacted with steam over a catalyst to produce CO_2 and H_2 (syngas). At the cathode, CO_2 in the air is converted to carbonate ions (CO_3^{2-}) according to the following equation:

$$CO_2 + \frac{1}{2}O_2 + 2e^- \rightarrow CO_3^{2-}$$

At the anode the hydrogen in the syngas reacts with the carbonate ions which pass through an electrolyte producing CO_2, water and free electrons according to the equation:

$$CO_3^{2-} + H_2 \rightarrow CO_2 + H_2O + 2e^-$$

Carbonate ions transfer electricity across the electrolyte releasing CO_2 into the anode offgas steam. The offgas from the anode contains unreacted fuel, hydrogen, and CO_2. This offgas is combusted with the air going to the cathode side, thereby increasing the concentration of CO_2 in the cathode feed stream. Excess CO_2 is vented with the cathode offgas. The MCFC acts in effect as a CO_2 concentrator in the anode stream.

The principle of the ECM is that flue gas, rather than anode offgas, is the source of the CO_2. Flue gas from a furnace, boiler, or engine is fed to the cathode side of the ECM. Additional air may also be required to maintain the required O_2 content. The action of the cathode is to remove the CO_2 from the flue gas stream and transfer it to the anode side (NOx and SOx are also removed by a similar process). The flue gas depleted in CO_2 passes to heat recovery before being vented to atmosphere. As with the MCFC, the carbonate ions migrate across the electrolyte and are converted to CO_2 and H_2O on the anode releasing the electrons in the process. Furthermore, fuel, usually natural gas, is fed to the anode where it is reacted with steam over a catalyst to produce CO_2 and H_2 (syngas). The hydrogen in the syngas reacts with the carbonate ions to produce water and CO_2, releasing the electrons and CO_2 in the carbonate ions in the process. The concentration of CO_2 in the anode offgas can be as high as about 80 percent. The rest of the anode offgas is unreformed methane and unconsumed fuel, H_2, and CO.

In the ECM, the anode offgas is not combusted in the cathode air stream as is the case with the MCFC. However, the offgas stream from the anode still contains a significant quantity of uncombusted fuel (CO, H_2, and CH_4); therefore, it is not ‘pure’ enough for CO_2 sequestration, or EOR. Further purification of CO_2 in the anode offgas is required.

First the anode flue gas is cooled by preheating the incoming flue gas and then fed to the CO_2 purification unit. The CO_2 is purified by cooling and compression of the anode offgas until the CO_2 in the stream liquefies. The liquid CO_2 is separated from the fuel gases—primarily CH_4, CO, and H_2 in a liquid vapour separator. The cold in the streams is recovered by heat exchange with the incoming feed, and the pressure of the vent stream gas is let down to the anode inlet pressure. The remaining fuel components are recycled back to the anode.

Process Description

In Case 1c, ECM is used for CO_2 capture in thermal in-situ production. In this case, flue gas from the OTSG recovers heat from the ECM offgas and is fed to the cathode side of the ECM together with additional air via independent blowers. The CO_2 reacts with the oxygen over the cathode to generate the carbonate ion and consume two electrons. Sensible heat is recovered from the CO_2 depleted flue gas which is then vented to atmosphere.

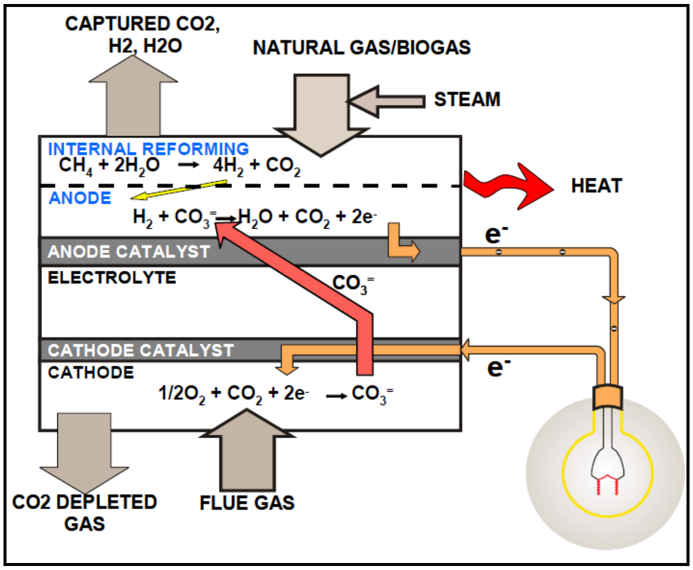
As in the MCFC, natural gas, steam, and residual fuel gas (recycled offgas from the CO_2 purification step) are preheated and fed to the anode. The natural gas is reacted (reformed) with steam over a catalyst to produce syngas, a mixture of H_2, CO, and CO_2 together with unreacted methane and steam, according to the steam methane reforming and water gas shift reactions (below):

$$CH_4 + H_2O \Leftrightarrow 3H_2 + CO$$

$$CO + H_2O \Leftrightarrow H_2 + CO_2$$

The hydrogen produced reacts with the carbonate ions over the anode catalyst to produce water, CO_2, and electrons which flow around the external circuit producing electricity. The CO_2 exhausts with the anode side together with water and unreacted CO, H_2, and methane. Due to their higher activity, any higher hydrocarbons in the feed natural gas are completely reformed to syngas. (See Figure D-1 below.)

Figure D-1.
ECM Flow Sketch



The anode offgas is cooled to ambient temperature; the water is knocked out and recycled back to the ECM. The impurities in the CO_2 in the anode offgas are higher than are permitted for sequestration, or EOR, and also contain valuable fuel. The anode offgas steam is dried, compressed, and cooled until the CO_2 in the stream liquefies. The liquid CO_2 is separated from the fuel gases—primarily CH_4, CO and H_2 in a liquid vapour separator—and pumped to pipeline pressure. The cold in the product streams is recovered by heat exchange with the incoming feed and the vent stream gas is let down to the anode inlet pressure. The remaining fuel components are recycled back to the anode. It should be noted that the excess heat on the anode side is used to generate the steam used on the steam reforming.

The principle of the CO_2 purification system is well known, and there are many patents and much know-how in the field of cryogenic CO_2 separation. The precise details of the CO_2 purification system developed by FCE were described to Jacobs for validation, but are covered under a confidentiality agreement. Therefore, Jacobs cannot release the details to the Steering Committee.

History and Development

The ECM technology is a variant of the MCFC technology, which has been in commercial operation (MW scale) for many years. The cryogenic CO_2 purification is a new application of existing commercial technologies. However, the ECM fuel cell has only been tested at a bench-top scale. Development of the ECM technology is currently underway and is supported by the US DoE and US EPA involving FCE, Pacific Northwest National Labs, and URS.

Readiness of the Technology

There are two distinct parts to the technology: the ECM and the cryogenic CO_2 purification system. They operate independently of each other and each can be substituted by alternative technologies. For example, ECM could be replaced by amine or membrane CO_2 removal technologies, and the CO_2 purification system could be replaced by amine or CO_2 solidification separation technologies. Therefore, the two technologies' readiness levels will be considered separately. The ECM is closely tied to MCFC technology; this will be considered also.

Using a version of the 9-step Technology Readiness Level ("TRL") categories developed by NASA (Appendix 1) and modified for process plant applications, the MCFC technology is at a TRL of 8. Commercial plants of over 1 MW are in operation with the Early Adopter cadre of customers. They are not yet in mainstream commercial operation as are, for example, gas turbines, steam turbines, and reciprocating engines. This is primarily due to the much higher capital cost per unit of electricity generated by MCFC compared to other forms of electricity generation from fossil fuels, rather than any inherent issue with the technology.

The ECM itself is at a TRL of 5, because bench-top scale testing is complete. However, much of the manufacturing, scale-up, support and auxiliary components and processes (e.g., stack, enclosure, integral reformer and anode systems, cathode, fuel supply, and power rectification and step-up) have already been solved for the MCFC technology upon which it is based. This means that the technology should move through Stages 6 and 7—prototype demonstration and system prototype—more rapidly than a technology that is radically different from anything developed before.

The one issue that may present operating problems is impurities in the fuel. While the pipeline natural gas fired in the OTSGs is clean and relatively free from impurities, the offgas from bitumen production may contain sulphur compounds, trace metals, ash, and particulates which can foul the cathode side electrolyte. A development program is underway at the Pacific Northwest National Labs investigating the impact of impurities, typically found in flue gases, primarily aimed at coal-fired rather than gas-fired applications, which may have applicability for bitumen production.

Determining a TRL for the cryogenic CO_2 purification system is more complicated. The individual items of equipment are all commercially available—no one equipment item needs development, or extension of current operating service. The equipment can be bought from multiple competent vendors. The issue is that the configuration of the components is unique. The production of food-grade CO_2 is the closest technology comparison to FCE's CO_2 purification unit. For food-grade CO_2, the raw CO_2 is recovered from an amine unit or similar. This stream is then compressed and cooled, so that the CO_2 is liquefied. The liquid CO_2 is purified in a distillation column to meet food-grade quality, and the impurities are released to atmosphere. These food-grade CO_2 purification units are typically sized to produce a few hundred tonnes per day of CO_2. CO_2 for EOR, or sequestration, does not need to achieve food-grade quality and FCE does not claim that the cryogenic CO_2 purification system technology being developed achieves that level of purity; therefore the reliability and operability of such technology should be higher than those of commercial food-grade CO_2 production units.

Jacobs Consultancy's evaluation is that the cryogenic CO_2 purification system is at a TRL of 7. The system is in operation at or near commercial scale.

Solid Oxide Fuel Cell (SOFC)

Background

The SOFC design considered for SAGD operation is derived from the conventional SOFC designed for power generation with heat recovery for steam generation. The principal of the SOFC is that a carbon-containing fuel is supplied to the anode, and air is supplied to the cathode. Oxygen ions transfer electricity across the electrolyte 'combusting' the fuel and releasing CO_2 and water into the anode offgas steam. The SOFC operates at high temperatures (500 to 1000°C), and waste heat can be recovered to generate steam for the SAGD operation.[1]

Process Description

For use in thermal in-situ bitumen production as in Cases 2b and 2d, the SOFC commercial configuration remains unchanged. The only change is that excess heat is used to generate steam for bitumen production rather than being rejected to the atmosphere. Air is preheated by heat exchange with the anode offgas combustion stream and fed to the cathode in the SOFC. Oxygen ions diffuse across the electrolyte to the anode where they 'react' with the fuel and

---

[1] For more information visit http://www.csa.com/discoveryguides/fuecel/overview.php

release electrons. The electrons flow through the external circuit back to the cathode to generate more oxygen ions from the air on the cathode side.

The sensible heat is recovered from the cathode offgas which is then vented to atmosphere. The offgas from the anode contains some unconverted fuel, which is catalytically combusted with air; the heat recovered is used to preheat the incoming air and generate steam for the SAGD operations.

In carbon capture mode, the offgas from the anode is combusted catalytically with high-purity oxygen (oxy-combustion). In this case, the ‘flue gas’ is predominantly CO_2 and steam. Heat is recovered from this stream for preheating the incoming air and fuel and for generating steam. It is then cooled to ambient temperature to knock out the water. The stream is dried and compressed to pipeline pressure.

History and Development

The SOFC technology has been under development for power generation for 50 years. It has been deployed commercially on the hundreds of kW scale.

The commercial use of pure oxygen to promote combustion dates back over a century, and there are many commercial examples of using oxygen to catalytically combust fuel gas components to generate an inert gas stream. Using oxygen and catalytic combustion has not been practiced on this scale nor has it been used to capture CO_2 from SOFC flue gases on a commercial scale.

SOFCs for power generation are being actively marketed by several vendors at sizes up to several hundred kWe.

Oxy-combustion—the practice of burning fuels in pure oxygen to capture CO_2—is under development by a number of commercial organizations, research centres and government agencies around the world. However, these development programs have not reached the scale required by a 1 GW SOFC plant. There is no active development or commercialization program underway for using pure oxygen and catalytic combustion to capture CO_2 from SOFC.

Readiness of the Technology

The SOFC and oxy-combustion represent two distinct parts of the technology and are considered separately for technical readiness.

SOFC for power generation is at a TRL of 8. Commercial plants of a few hundred kW are in operation with the Early Adopter cadre of customers. They are not yet in mainstream commercial operation as are, for example, gas turbines, steam turbines, and reciprocating engines. This is primarily due to the much higher capital cost per unit of electricity generated by SOFC than for other forms of electricity generation from fossil fuels, rather than any inherent issue with the technology.

Configuring the technology for power and steam production results in a change to the balance of plant rather than the fuel cell stack itself, and as such reduces the TRL to 7 because although electricity generation from SOFC is commercial, high-pressure steam generation with SOFC has not been done commercially.

Using pure oxygen to catalytically combust the residual fuel in the anode offgas ready for CO_2 capture is at a TRL of 7. High-purity air separation plants of this scale are in widespread operation, and most of that oxygen is used in either combustion or catalytic processes. However, using combustion with pure oxygen to capture CO_2 is only at the prototype stage and integration with SOFC has only been modelled.

Section E.

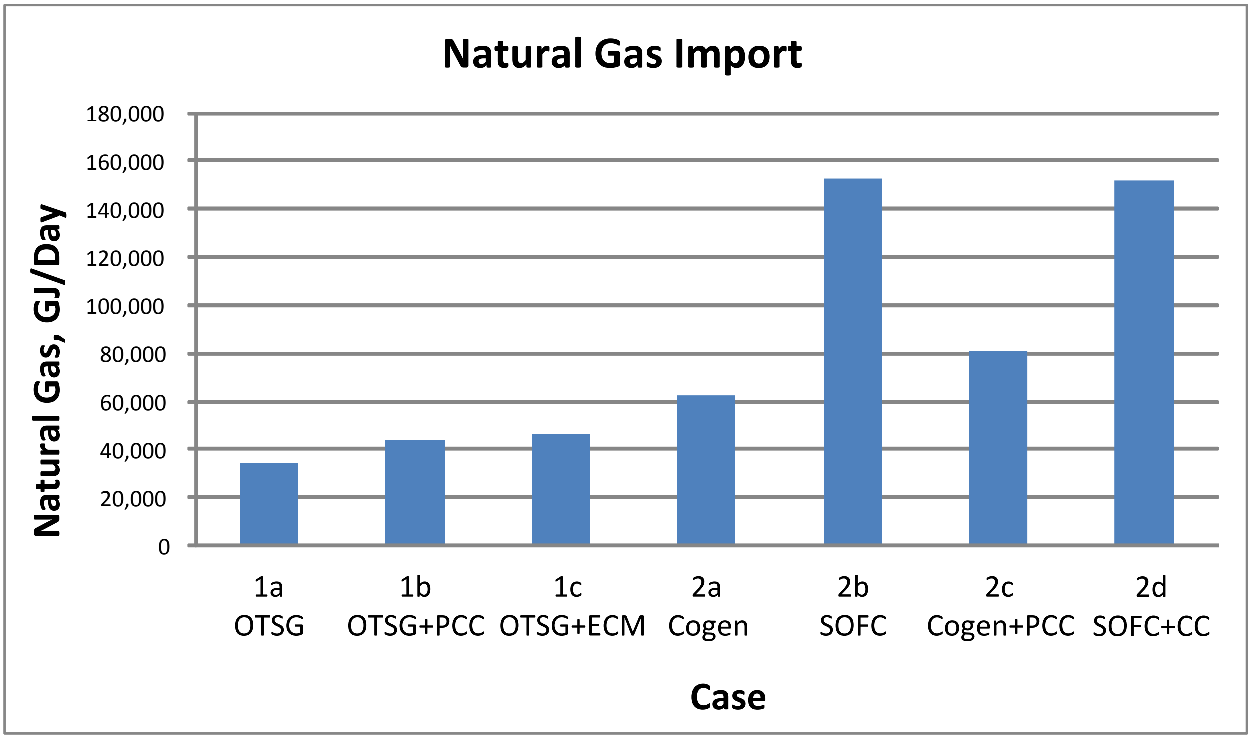
Technical Results

A summary of the technical information for the cases is included in Table E-1. As indicated in Section B of this report, our technical analyses are based on in-house information for Cases 1a and 2a. The post-combustion capture costs for Cases 1b and 2c are based on publicly available information for the Fluor Economine process. The technical information for Cases 1c, 2b and 2d are based on simulations and results provided by FCE.

A few interesting results emerge from the technical information, as shown in Figures E-1 through E-3:

- As indicated by Figures E-1 and E-2, SAGD facilities have a large natural gas demand but are not power intensive.
- Not surprisingly, Case 1a, which is the standard OTSG case, has the lowest natural gas use.
- Case 2a, the Cogen case, consumes about 80% more natural gas than the OTSG case.
- CO_2 capture requires more natural gas than the cases without capture, except for Case 2d versus Case 2b.
- CO_2 capture with Case 2d requires less natural gas than Case 2b, due to energy recovered from the oxygen combustion of the SOFC offgas.

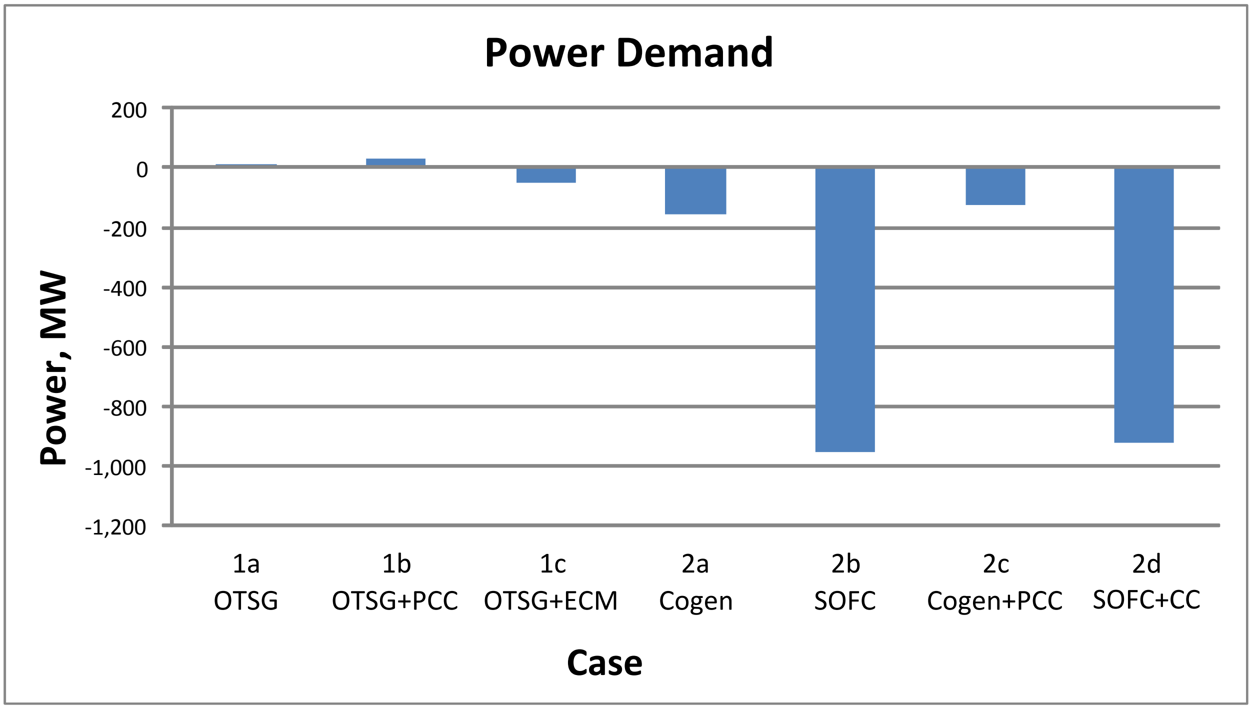
Figure E-1.
Natural Gas Import



From a power perspective:

- Due to relatively low internal demand for power, all cases except two (i.e., the OTSG cases, Case 1a and 1b) export power.
- Case 1c, with ECM, is unique in that the CO_2 capture with the fuel cell produces power and even provides sufficient quantity for export.
- The Cogen cases have modest power exports.
- The SOFC cases have large power exports.

Figure E-2.
Net Power



From a capital cost perspective:

- Not surprisingly, carbon capture universally increases the capital cost of the facilities.
- ECM, as shown in Case 1c, is less expensive than Case 1b with equivalent carbon capture.
- Cogen with post-combustion capture is less expensive than the SOFC technologies.
- Conventional post-combustion capture is costly, adding nearly 50% to the capital costs for OTSG and Cogen.

Figure E-3.
Estimated Capital Costs

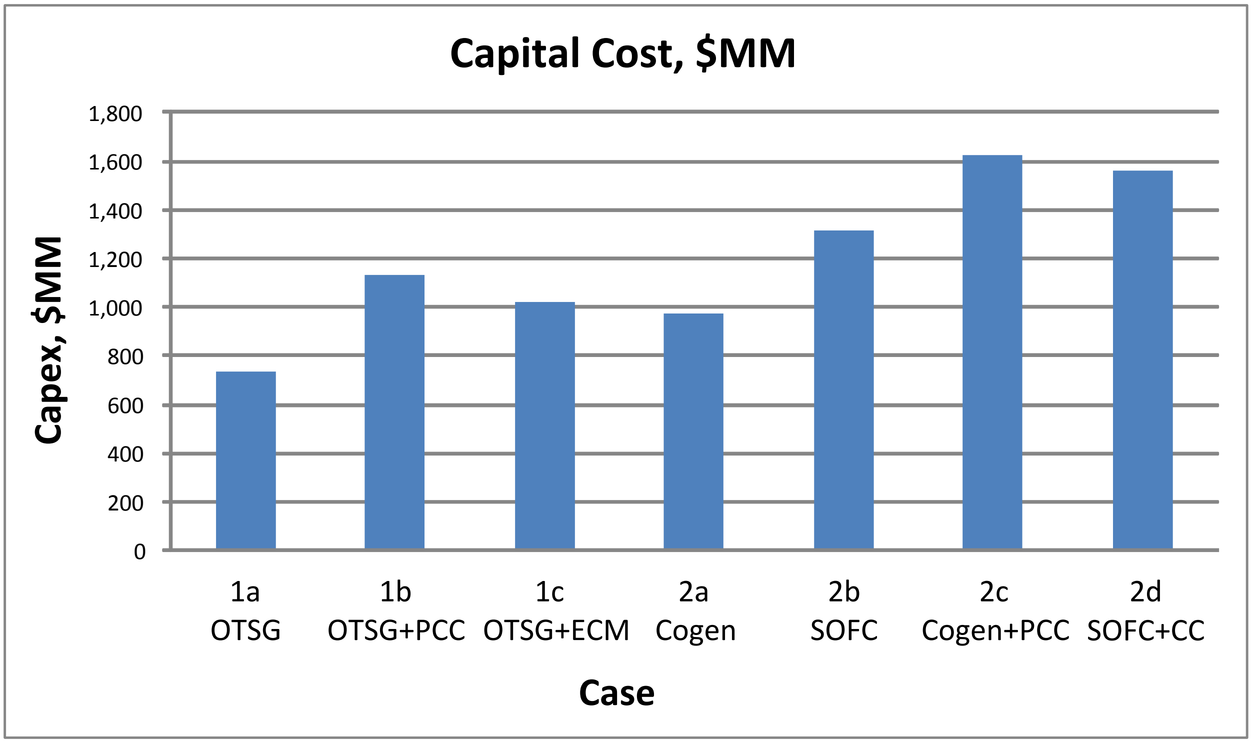


Table E-1.
Technical Data Summary

Calendar Day Rates by Case (Rev 13)		1a OTSG	1b OTSG+PCC	1c OTSG+ECM	2a Cogen	2b SOFC	2c Cogen+PCC	2d SOFC+CC
Bitumen Production	**BPD**	33,000	33,000	33,000	33,000	33,000	33,000	33,000
Steam Production for SAGD	**BPD CWE**	99,000	99,000	99,000	99,000	99,000	99,000	99,000
Steam Production for Carbon Capture	**BPD CWE**		21,425				39,061	
Sulphur Production	**tonnes/day**	0.03	0.03	0.03	0.03	0.03	0.03	0.03
Disposal Water								
Steam Production for SAGD	**m^3/day CWE**	1,693	1,693	1,693	1,693	1,693	1,693	4,391
PCC Boiler System	**m^3/day CWE**	0	68	0	0	0	125	0
Total	**m^3/day CWE**	1,693	1,762	1,693	1,693	1,693	1,818	4,391
WLS Sludge	**tonnes/day**	20	20	20	20	20	20	10
WLS Sludge	**CAD/day**	3,053	3,053	3,053	3,053	3,053	3,053	1,500
Makeup Water Rate								
Steam Production for SAGD	**m^3/day**	3,375	3,375	3,375	3,375	3,375	3,375	3,375
PCC	**m^3/day**		68	0			125	0
ECM/SOFC Water Produced	**m^3/day**			-1,488				-6,072
Total	**m^3/day**	3,375	3,443	1,887	3,375	0	3,499	0
Water Treating Chemicals								
Lime	**tonnes/day**	6.9	6.9	6.3	6.9	6.9	6.9	5.0
MagOx	**tonnes/day**	2.6	2.6	2.6	2.6	2.6	2.6	2.3
Soda Ash	**tonnes/day**	0.8	0.8	0.7	0.8	0.8	0.8	0.5
Total Water Treating Chemicals	**CAD/day**	15,841	15,841	15,049	15,841	15,841	15,841	11,881

Calendar Day Rates by Case (Rev 13)		1a OTSG	1b OTSG+PCC	1c OTSG+ECM	2a Cogen	2b SOFC	2c Cogen+PCC	2d SOFC+CC
Amine Carbon Capture Chemicals								
MEA	**MMCAD/yr**		1.1				2.0	0
Corrosion Inhibitor	**MMCAD/yr**		0.2				0.4	0
Mole Sieve	**MMCAD/yr**		0.05				0.09	0
Total	**MMCAD/yr**		1.3				2.4	0
Amine Sludge Disposal	**MMCAD/yr**		0.8				1.4	0
Produced Gas Consumed								
Bitumen Production (LHV)	**GJ/day**	898	898	898	898	898	898	898
Bitumen Production (HHV)	**GJ/day**	1,002	1,002	1,002	1,002	1,002	1,002	1,002
Natural Gas Consumed								
Bitumen Production (LHV)	**GJ/day**	30,712	30,712	30,712				
Bitumen Production (HHV)	**GJ/day**	34,073	34,073	34,073				
Cogen Steam & Power Production (LHV)	**GJ/day**				56,153		56,153	
Cogen Steam & Power Production (HHV)	**GJ/day**				62,299		62,299	
Carbon Capture (LHV)	**GJ/day**		9,302				16,960	
Carbon Capture (HHV)	**GJ/day**		10,320				18,816	
ECM (LHV)	**GJ/day**			10,946				
ECM (HHV)	**GJ/day**			12,145				
SOFC (LHV)	**GJ/day**					137,090		135,782
SOFC (HHV)	**GJ/day**					152,084		150,632
Total (LHV)	**GJ/day**	30,712	40,014	41,658	56,153	137,988	73,113	136,680
Total (HHV)	**GJ/day**	34,073	44,394	46,218	62,299	153,086	81,115	151,634

Calendar Day Rates by Case (Rev 13)		1a OTSG	1b OTSG+PCC	1c OTSG+ECM	2a Cogen	2b SOFC	2c Cogen+PCC	2d SOFC+CC
Power Consumed								
Bitumen Production	**MW**	14	14	14	11	11	11	11
Carbon Capture	**MW**		8				14	15
CO_2 Compression	**MW**	0	10		0		19	39
ECM	**MW**			11				
SOFC						61		45
Inverter losses				3		43		43
Total Power Consumed	**MW**	14	32	28	11	115	44	153
Power Generated	**MW**	0	0	76	166	1072	166	1077
Power Imported	**MW**	14	32	0	0		0	
Power Exported	**MW**	0	0	48	155	957	122	924
CO_2 Direct								
Steam Production for SAGD	**tonnes/day**	1,765	1,765	1,765				
Cogen Steam & Power Production Total	**tonnes/day**				3,217		3,217	
Carbon Capture Steam Generation	**tonnes/day**		505				921	
ECM Operation	**tonnes/day**			608				
SOFC	**tonnes/day**					7,755		7,617
CO2 Direct Total	**tonnes/day**	1,765	2,270	2,373	3,217	7,755	4,138	7,617
CO_2 Indirect	**tonnes/day**							
Without Power	**tonnes/day**	5	5	5	5	5	5	4
CO_2 Captured	**tonnes/day**	0	2,043	2,205	0	0	3,724	7,615
CO_2 Generated (Direct)	**tonnes/day**	1,765	227	168	3,217	7,755	414	2

Calendar Day Rates by Case (Rev 13)		1a OTSG	1b OTSG+PCC	1c OTSG+ECM	2a Cogen	2b SOFC	2c Cogen+PCC	2d SOFC+CC
% CO_2 Capture (Direct Only)		0%	90%	93%	0%	0%	90%	100%
Total Costs								
Total Chemicals & Disposal Costs	**MM CAD/yr**	7	11	7	7	7	13	7
Fixed O&M Costs	**MM CAD/yr**	26	40	36	34	46	57	55
Capital Cost (3Q2012, Ft McMurray)	**MM CAD**	735	1,132	1,023	975	1,311	1,623	1,563
Capital Spend Profile								
Start up - 5yr		0%	0%	0%	0%	0%	0%	0%
Start up - 4yr		0%	0%	0%	0%	0%	0%	0%
Start up - 3yr		15%	15%	30%	15%	30%	15%	30%
Start up - 2yr		55%	55%	50%	55%	50%	55%	50%
Start up - 1yr		30%	30%	20%	30%	20%	30%	20%
Steam Generation								
Availability Factor		94%	94%	94%	94%	94%	94%	94%
Wet Steam Quality		77%	77%	77%	77%	77%	77%	77%
Fuel Cell Replacement costs	**MMCAD/yr**			18		80		80
Steam Generation								
Availability Factor		94%	94%	94%	94%	94%	94%	94%
Wet Steam Quality		77%	77%	77%	77%	77%	77%	77%

Calendar Day Rates by Case (Rev 13)		1a OTSG	1b OTSG+PCC	1c OTSG+ECM	2a Cogen	2b SOFC	2c Cogen+PCC	2d SOFC+CC
CO$_2$ capture and compression costs								
Capex	**MM CAD**		397	288		336	648	252
Annual Capex charge (12%)	**MM CAD/yr**		48	53	0	120	78	110
Base Utility costs	**MM CAD/yr**							
Chemicals and Catalyst costs	**MM CAD/yr**		3	0	0	0	6	0
Fixed O&M Costs	**MM CAD/yr**		14	10	0	12	23	9
Total CO$_2$ capture and compression costs	**MM CAD/yr**	**0**	**65**	**63**	**0**	**132**	**107**	**119**

Section F.

Economic Results

Introduction

One of the key objectives of this Study was to identify technologies that may have lower GHG emission intensities while providing steam at a reasonable cost. Given that reducing GHG emissions generally requires energy and additional capital and operating cost, it is expected that technologies with lower GHG emissions will have higher costs to provide steam. Unique to the fuel cell cases evaluated, however, is that they produced significant amounts of additional power. The value of this power contributed significantly to the economics of these cases, providing some surprising results. To compare the various options several metrics were used to assess the cost of providing steam.

The following section describes the assumptions and methodologies used, the key results, and the sensitivities associated with the key variables.

Levelized Cost of Steam

Figure F-1 shows the major cost and revenue components that make up the levelized cost of steam produced. The yellow negative values are related to the benefits associated with avoiding paying $15/t to mitigate CO_2 emissions or selling CO_2 credits. The other blue negative bar is the CO_2 benefit associated with selling additional green power to the grid. The top light purple value is associated with the cost of replacing fuel cells as they degrade over time. For most cases CO_2 compliance costs are a very small component of steam costs. The OTSG and OTSG CCS cases are the only cases that purchase power from the grid. CO_2 compliance costs are $15/t for the OTSG and Cogen cases to mitigate 12% of emissions.

Figure F-1.
Levelized Cost of Steam ($/m^3)

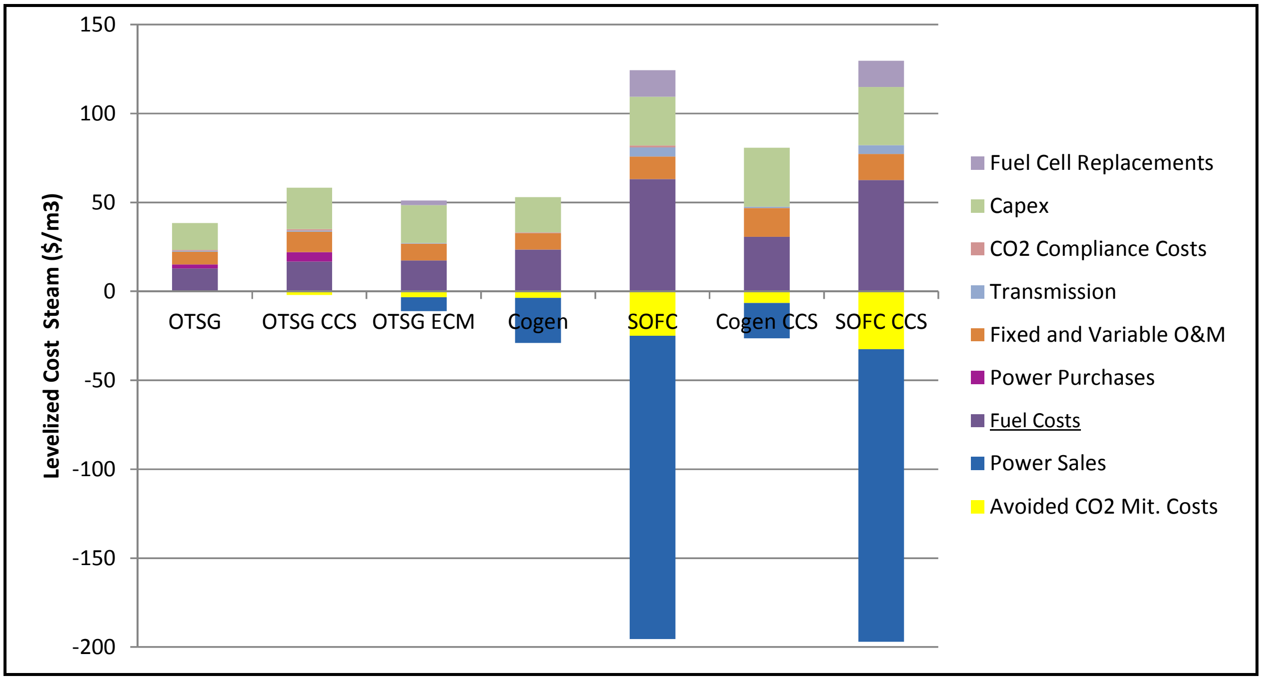


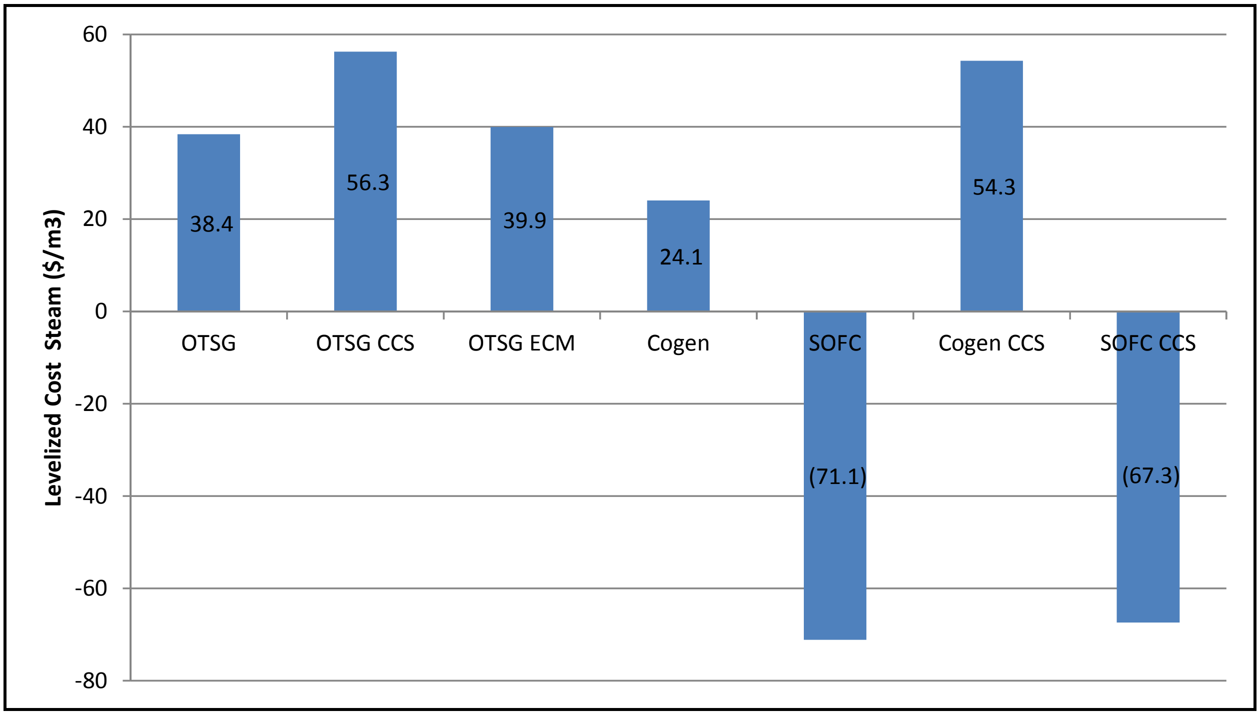
Table F-1 shows the numerical values for the components in Figure F-1 above. For the OTSG CCS case, -2.0/m^3 is related to CO_2 being captured and the value of GHG credits associated with doing so. There is also a 0.8 CO_2 Compliance Cost for the OTSG CCS case because it was assumed that the user of power would be mitigating the emissions associated with power consumption. There is an incremental amount of power required to capture CO_2 for this case compared to the OTSG case. The OTSG ECM case eliminates power purchases and also produces more credits related to a larger net power sale compared to the OTSG ECM case.

Table F-1.
Levelized Cost of Steam

	OTSG	OTSG CCS	OTSG ECM	Cogen	SOFC	Cogen CCS	SOFC CCS
Avoided CO2 Mit. Costs	0.0	-2.0	-3.3	-3.7	-25.0	-6.6	-32.4
Power Sales	0.0	0.0	-7.8	-25.2	-170.4	-19.9	-164.6
Fuel Costs	12.9	16.8	17.4	23.5	63.1	30.6	62.5
Power Purchases	2.3	5.3	0.0	0.0	0.0	0.0	0.0
Fixed and Variable O&M	7.2	11.5	9.4	9.1	12.7	16.3	14.7
Transmission	0.3	0.8	0.2	0.0	5.1	0.6	4.9
CO2 Compliance Costs	0.6	0.8	0.0	0.4	1.2	0.0	0.0
Capex	15.1	23.2	21.4	20.0	27.4	33.3	32.7
Fuel Cell Replacements	0.0	0.0	2.6	0.0	14.8	0.0	14.8
Total Cost of Steam	38.4	56.3	39.9	24.1	-71.1	54.3	-67.3

Figure F-2 below shows the net levelized cost of steam produced. Generally one would expect that because capturing CO_2 requires additional energy and equipment cost, the levelized cost of steam for cases with lower CO_2 emissions should be higher than their base case without CCS. This is clearly seen for Cogens and OTSG. However, the OTSG ECM case has a levelized cost of steam comparable to an OTSG without CO_2 capture and has a much lower GHG emission intensity. The SOFC case without CCS has the lowest levelized cost of steam. Adding carbon capture to the SOFC case slightly increases its levelized cost of steam. Both SOFC cases provide negative costs for steam production because of the very high amount of power production provided.

Figure F-2.
Net Levelized Cost of Steam



Cost of CO_2 Capture

Figure F-3 shows the cost components that make up the cost to capture CO_2. The cost of capture is based on the incremental cost of CO_2 capture on the same steam production technology. The cost to capture CO_2 includes the cost to supply CO_2 at high pressure to a pipeline at the plant gate. The OTSG ECM case has both a benefit associated with selling 47.8 MW of power produced (dark pink bar) and not having to purchase the 14.2 MW of power (dark green bar) required by the OTSG case without CCS. The sum of these two values represents the economic benefit associated with producing 62 MW of power, which was not available in the base OTSG case. The fuel cell replacement cost is not shown for the SOFC cases because it is a cost born by both the SOFC and the SOFC CCS case, and therefore it is netted out and is not incremental to CO_2 capture. Recall we have defined the cost of capture based on the incremental cost of completing capture on the base technology—in this case a SOFC—without CCS. For cogen, 32.7 MW of power has been used for carbon capture and is no longer available for sale (grey bar). For SOFC, 33.3 MW of power has been used for carbon capture and is not longer available for sale (grey bar). The dark blue bars on top of the X-axis for the Cogen and SOFC case indicate that less power is available for sale because a significant amount of it is required to capture CO_2.

The values in the graph below were derived by taking the difference in the cost of steam for each component found in Figure F-2 for a CCS case and its reference case, multiplying by 5,744,742 m^3/yr of steam, and then dividing by the mass of CO_2 captured as found in Table F-2.

Figure F-3.
Cost of CO_2 Capture

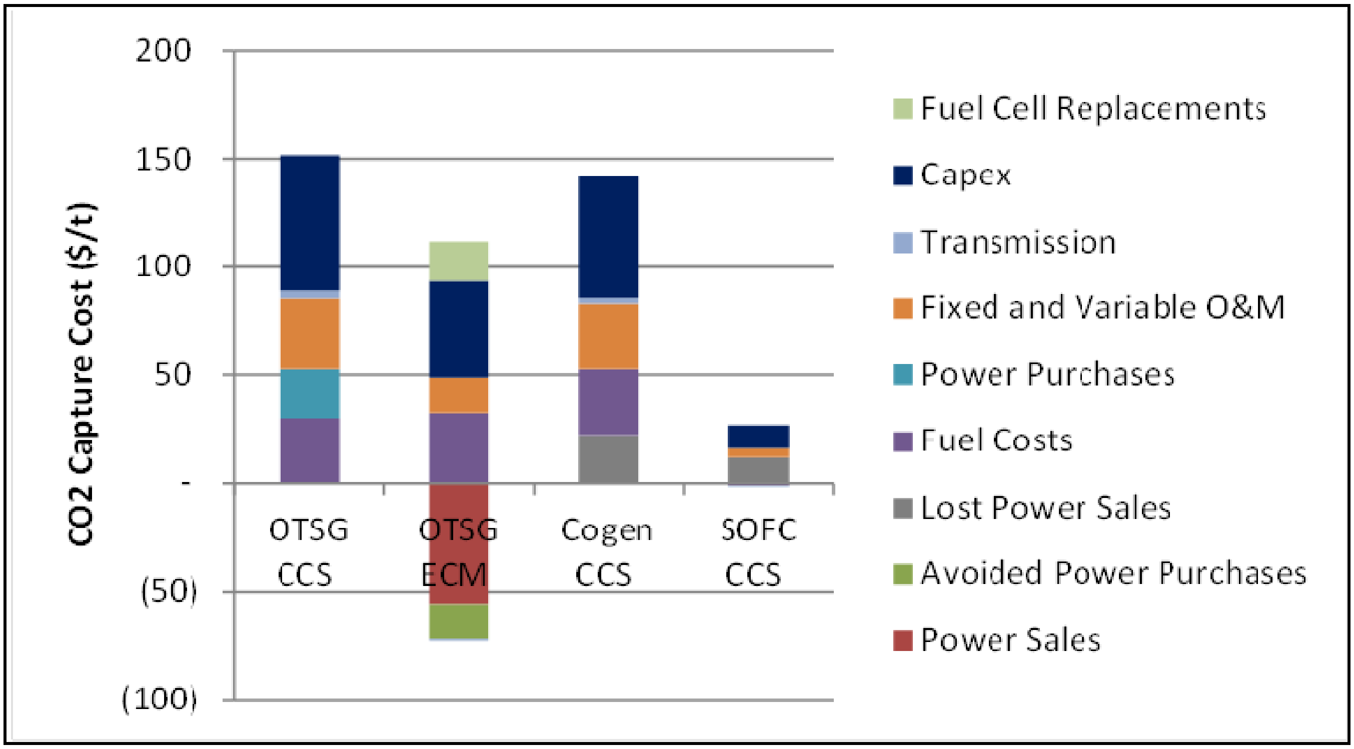


Table F-2 shows the components of costs and revenues which make up the cost of capturing CO2 for all of the cases with CO2 capture. The values in this table use a reference case of the underlying technology without CCS. For instance a SOFC without CCS was used as the reference case for the SOFC CCS case.

Table F-2.
CO_2 Capture Cost Components

	OTSG CCS	OTSG ECM	Cogen CCS	SOFC CCS
Power Sales	0.0	-55.6	22.5	12.0
Fuel Costs	30.0	32.7	30.0	-1.1
Power Purchases	22.6	-16.6	0.0	0.0
Fixed and Variable O&M	32.9	15.8	30.5	4.2
Transmission	3.4	-0.8	2.5	-0.4
Capex	62.7	45.0	56.2	10.9
Fuel Cell Replacements	0.0	18.3	0.0	0.0
Taxes	0.0	0.0	0.0	0.0
Total Capture Cost ($/t)	151.7	38.8	141.7	25.5

Figure F-4 shows the cost to capture CO_2 for the four CO_2 capture cases considered. These values were derived by assessing the cost of CO_2 capture on the same base option without CCS. The SOFC CCS case has the lowest cost of capture followed by the OTSG ECM case. The SOFC CCS case has a low capture cost because it does not cost much to purify the CO_2 produced by the fuel cell, while the OTSG ECM case has a low capture cost because the incremental power produced helps offset the cost to capture CO_2. Table F-3 shows that for the same amount of steam, the SOFC cases capture several times more CO_2 than the OTSG cases. The reason is that more fuel is required to produce the power generated in the SOFC cases than in the OTSG cases. Therefore, as shown in Table F-3, the fuel cost component cost of steam with and without CO_2 capture, for the SOFC cases, is greater than all the other cases.

Figure F-4.
Net CO_2 Capture Costs

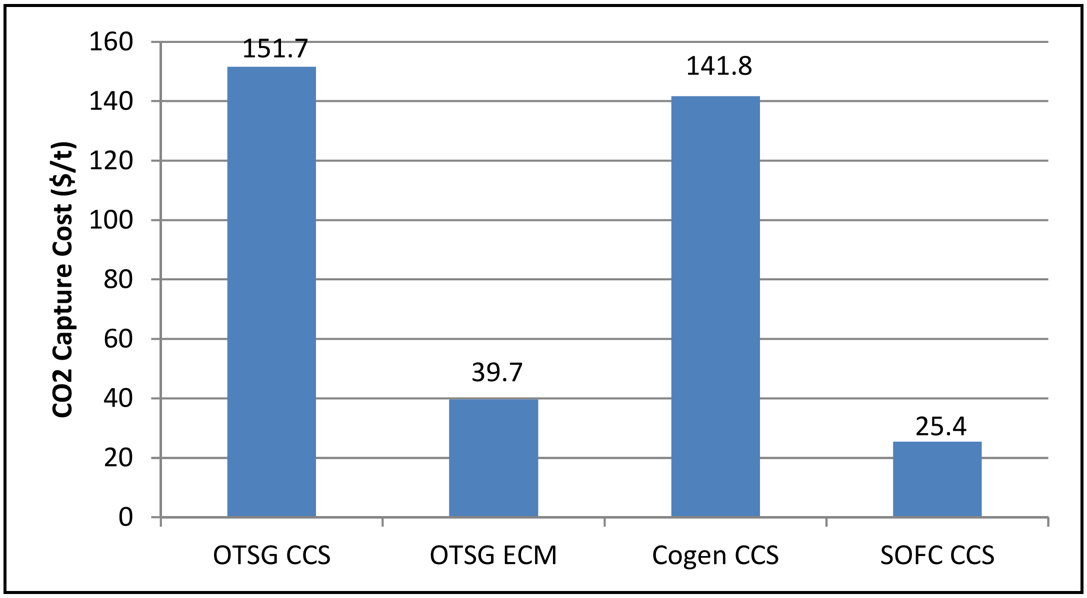


Table F-3 shows the values used to derive the mass of CO_2 avoided. For this table the reference case is the base technology without CCS. Avoided cost is the cost incurred to install and operate equipment to avoid these emissions.

As shown in Table F-3, CO_2 avoided is equal to CO_2 Captured, less CO_2 associated with power used to capture CO_2, plus credits associated with additional power generated less extra fuel used to produce steam used to capture CO_2. For the OTSG cases, 109,773 t/yr of CO_2 is emitted. Power consumption increased for the OTSG CCS case, causing 248,223 t/yr of emissions for a net increase of 138,450 t/yr. For the OTSG ECM case the plant produces enough power to back out power purchased from the grid. Therefore, -109,773 t/yr is avoided. For the ECM case there is a net sale of power to the grid compared to the OTSG case. For the SOFC cases, less power is sold to the grid because some is used to capture CO_2. Incremental fuel use to produce steam is determined by comparing the emissions associated with fuel use in the base case without CCS to the case with CCS.

Generally the reference case used to calculate the cost of capture is what one would otherwise build. For instance, one would not build an IGCC without CCS because the cost is too high. Therefore, generally a supercritical coal plant without CCS is used as the reference case. If it is assumed that one would otherwise build an OTSG, then the capture cost for the Cogen becomes $97/t and -$150/t for the SOFC.

Table F-3.
Components of CO_2 Avoided

	OTSG CCS	OTSG ECM	Cogen CCS	SOFC CCS
CO2 Captured	745,695	804,825	1,359,260	2,779,475
OTSG GHG for Power Purchases	109,773	109,773		
Power Purchased with CCS (t/yr)	248,223	0		
Less: CO2 from Power for CCS	138,450	-109,773		
Add: Credits for Power Gen. (t/yr)	0	368,481	-252,078	-256,703
Base Technology Fuel Use wo CCS	644,225	644,225	1,174,205	2,830,575
Fuel Use with CCS	828,550	865,780	1,510,370	2,780,205
Less: Incremental Fuel Use	184,325	221,555	336,165	-50,370
CO2 Avoided	422,920	1,061,524	771,017	2,573,142

Figure F-5 shows the components contributing to the avoided cost of CO_2. The values in the Figure F-5 below were derived by taking the difference in the cost of steam for each component found in Table F-1 for the CCS and the reference case and multiplying it by 5,744,742 m^3/yr of steam and then dividing by the mass of CO_2 avoided, as found in Table F-3.

Figure F-5.
Avoided CO_2 Costs

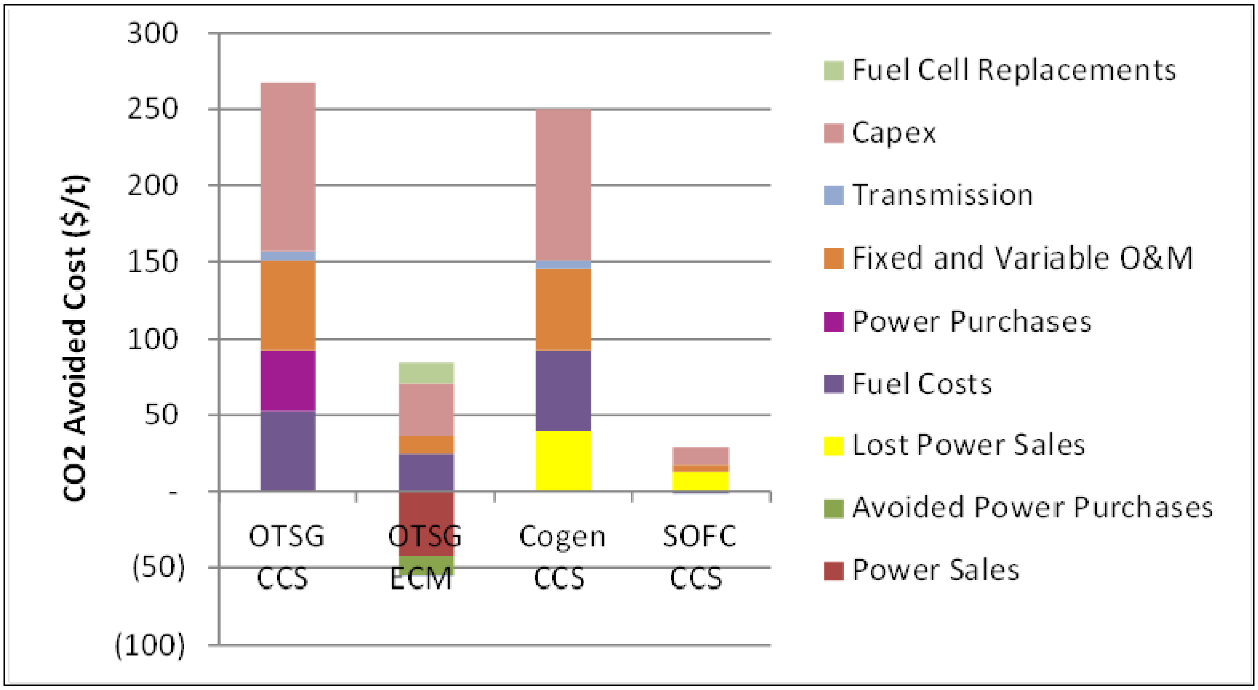


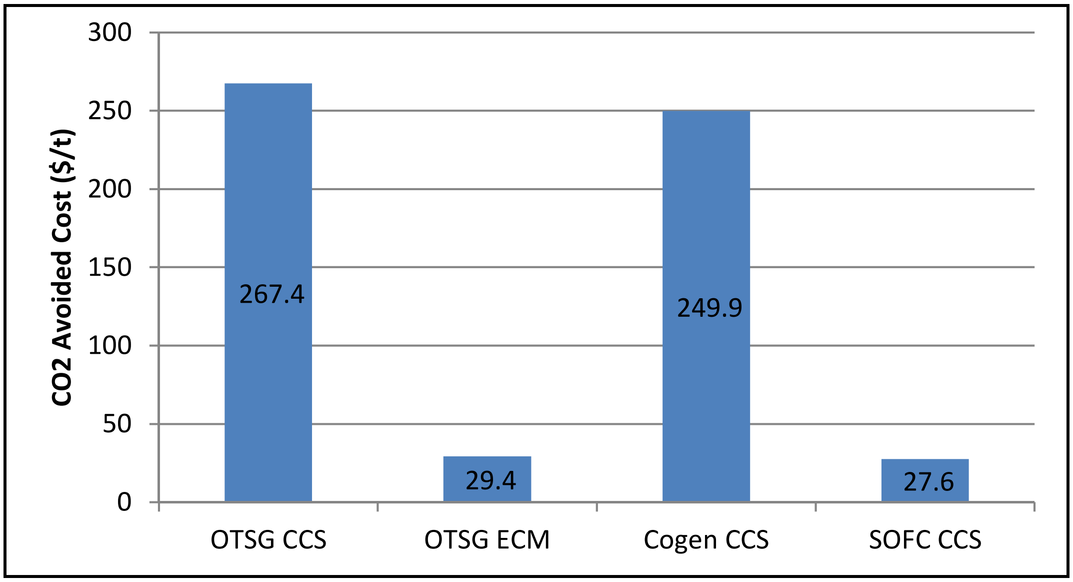
Table F-4 shows the components of costs and revenues which make up the avoided cost of CO_2 for all of the cases with CO_2 capture. The values in this table use a reference case of the underlying technology without CCS. For instance a SOFC without CCS was used as the reference case for the SOFC CCS case.

Table F-4.
Avoided Cost Components

	OTSG CCS	OTSG ECM	Cogen CCS	SOFC CCS
Power Sales	0.0	-42.2	39.7	12.9
Fuel Costs	52.9	24.8	52.9	-1.2
Power Purchases	39.8	-12.6	0.0	0.0
Fixed and Variable O&M	58.1	12.0	53.7	4.5
Transmission	6.0	-0.6	4.4	-0.4
Capex	110.6	34.1	99.0	11.8
Fuel Cell Replacements	0.0	13.9	0.0	0.0
Total Avoided Cost ($/t)	267.4	29.4	249.9	27.6

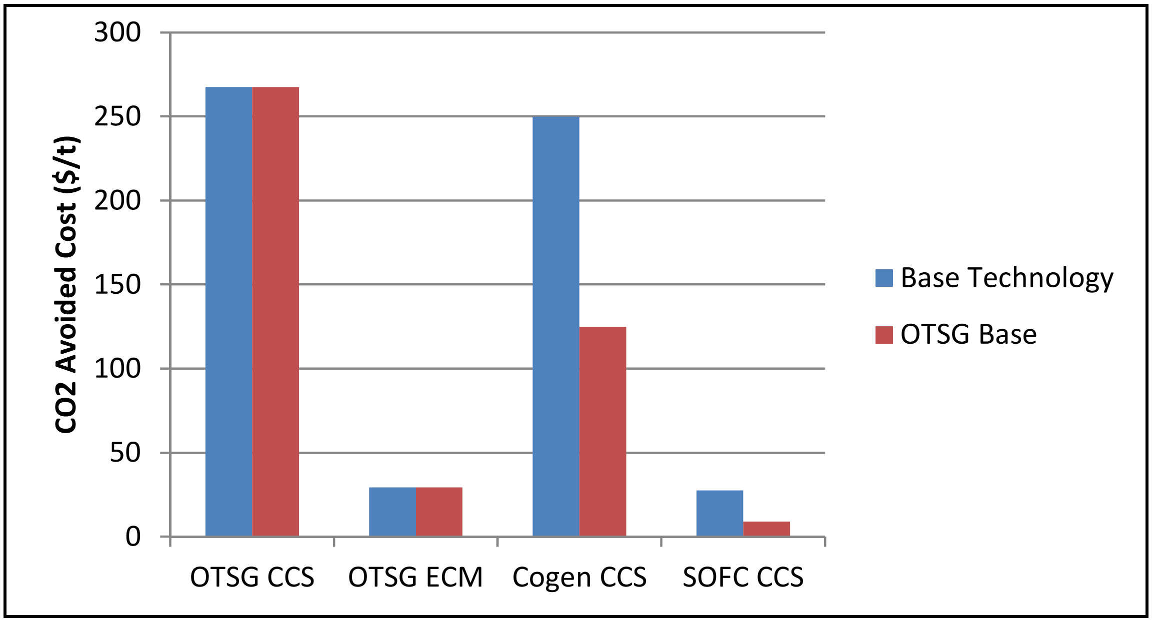
Figure F-6 shows the avoided cost of CO_2 for the four CO_2 capture options. The OTSG ECM case has a low capital and fuel cost required to capture CO_2 which is offset by the value of the incremental power produced. The SOFC CCS case incurs little cost to purify the CO_2 produced by the fuel cell.

Figure F-6.
Avoided CO_2 Costs



Avoided cost calculations are based on comparing a case with lower GHG emissions to a reference case. It could be argued that one would simply build an OTSG rather than install CCS technology. Therefore, an OTSG without CCS could be used as the reference case. The blue bars in Figure F-7 show the avoided cost where the base technology without CCS is used as the reference case. For example, an SOFC is used as the reference case for the SOFC CCS avoided costs. The red bars show the avoided costs where an OTSG without CCS is used as the reference case.

Figure F-7.
Avoided CO_2 Costs with Two Base Cases



Economic Sensitivities

Introduction

This Study was based on, in some cases, fairly uncertain costs given that some of the technologies are emerging or early in commercialization. In addition, the operational characteristics over the long term are not well known. The economics of many of these cases are highly dependent upon the price forecasts for CO_2 credits, natural gas, and power. Therefore, several sensitivities were constructed to compare the economic results for a range of assumptions.

Natural Gas Prices

The SOFC options have a very high fuel cost component compared to the other cases. It is expected that the cost of steam for these cases would be very sensitive to changes in natural gas prices.

Figure F-8 shows how the cost of steam is expected to change as the price of natural gas changes. The base natural gas price used in the economic modeling is $5.00/GJ HHV. Clearly the SOFC cases are most sensitive to changes in natural gas prices. However, while the cost of steam for all cases increases as gas price increases, only the Cogen CCS case becomes more expensive than another case, the OTSG CCS case. Aside from this, the relative ranking of the options remains the same at all gas prices considered. Figure F-8 suggests that the price of natural gas would have to exceed $10.00/GJ before the SOFC cases would have a cost of steam greater than zero.

Figure F-8.
Natural Gas Price Sensitivity

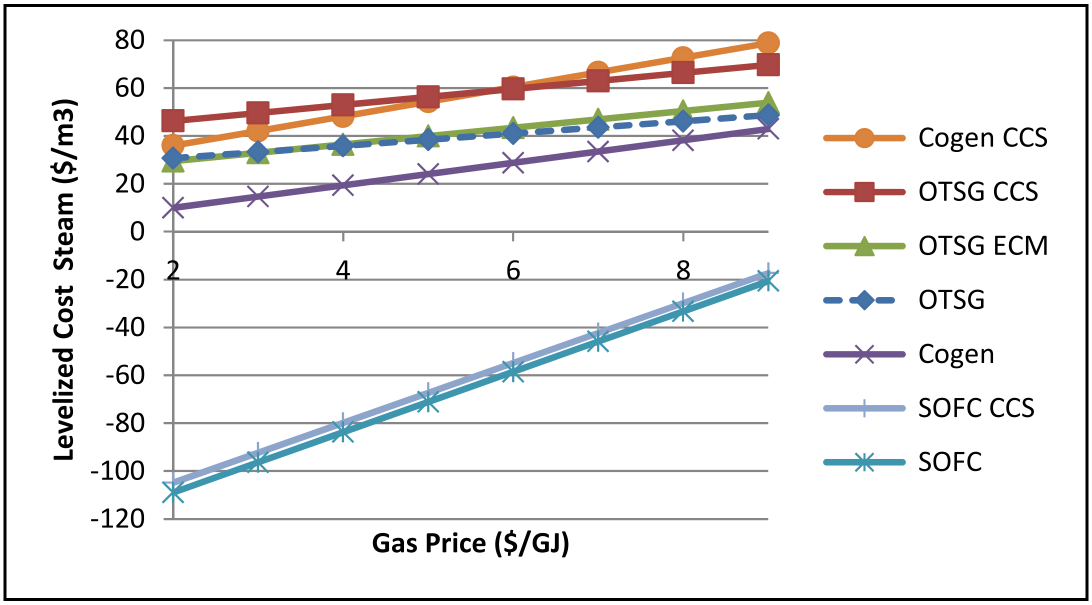
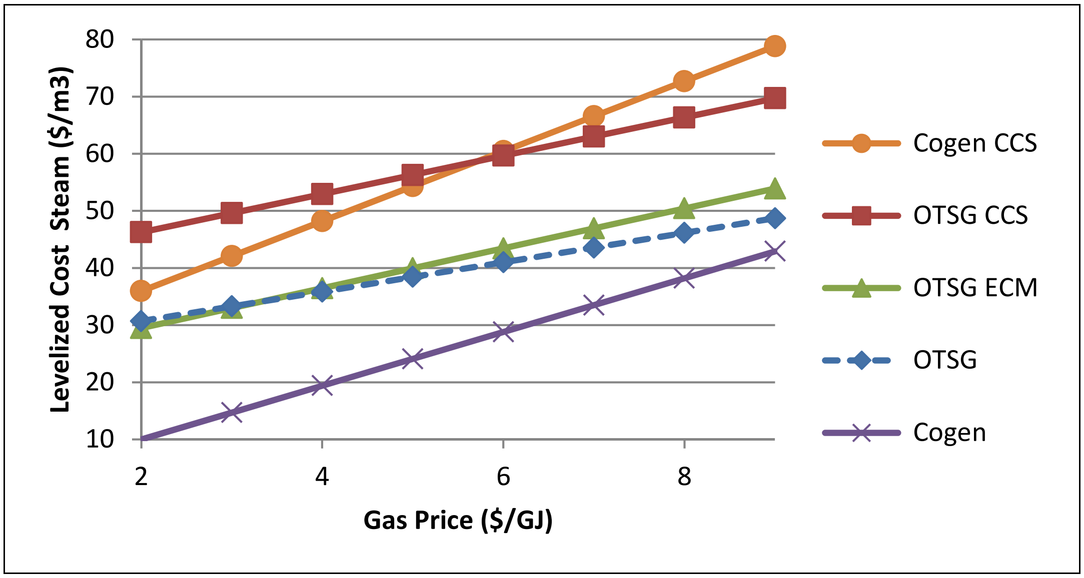


Figure F-9.
Natural Gas Price Sensitivity – without SOFC



Power Price Forecast

Power consumption in Alberta is expected to continue to grow by about 300 MW per year. Over the next 15 years, several thousand MW of coal plants may be decommissioned. The SOFC cases considered would bring over 900 MW onto the system over a short period of time. Generally, if more power is brought onto the system compared to load growth, the price of power will decrease. Bringing large amounts of SOFC power onto the grid along with other new power projects could suppress the power price. If the price of power is suppressed, then the economics of the SOFC cases could be less attractive.

The SOFC cases have a very large component related to power sales. Therefore, the economics of these cases should be sensitive to changes in power price. Some of the options near the top of the graph in Figure F-10 become more attractive or less attractive than other options as the power price changes. The price of power would have to collapse to $30/MWh before the SOFC CCS cases have a cost of steam similar to that of an OTSG—that is, if the cost of steam is zero or no steam is made available, the cost of power is $50/MWh for the SOFC cases. The OTSG ECM has the same cost as the OTSG at about $110/MWh. On a standalone basis, without capturing CO_2 from the OTSG or supplying steam to the OTSG, the cost of power from the ECM is about $120/MWh. It could still capture the CO_2 produced by the natural gas fuel leading to a low power GHG intensity.

Cogen units generally price steam similar to that for an OTSG and then determine the required selling price of power to yield a reasonable return. The Cogen case would have to sell its power for about $50/MWh to have a similar cost of steam as an OTSG. This is low compared to other assessments of the required selling price of power for cogens. However, if steam is priced to provide a 20% discount to that of an OTSG at about $40/t steam, then the required selling price of power for the Cogen rises to about $80/MWh. Cogen units are one of the lowest-cost forms of new generation. They may set a floor price on the cost of power over the long term.

Many companies in the oil industry have decided not to build Cogens to supply steam even though they may have better economics than OTSGs. Oil and gas companies may not be willing to invest the additional capital required to support a Cogen, preferring to allocate capital to other opportunities. The SOFC cases considered require three to four times the capital for the same amount of steam production as the OTSG case.

In addition, oil and gas companies may not be interested in selling large quantities of power or working with power utilities to do so. For these reasons oil and gas companies may not want to allocate capital to sell large quantities of power from SOFC. In addition, the owners of a large SOFC project will assume significant exposure to the volatile power market in Alberta. Historically Cogens have faced the situation where off-peak prices were insufficient to recover fuel costs. Figure F-10 shows that the fuel chargeable to power heat rate of the SOFC cases is relatively low compared to the other cases, helping to mitigate this issue.

The proportion of power produced relative to steam can however be reduced in a number of ways. First, power could be used in electric boilers to produce steam. This would reduce the overall size of the project, reducing power production even further. However, an OTSG requires about 2.16 GJ to produce one m^3 of steam. The cost to produce one m^3 of steam with natural gas is about 2.16 GJ/m3 X $5.00/GJ = $10.80/m^3. If we assume an electric boiler is 100% efficient and the OTSG is only 85% efficient, then an electric boiler will require 1.84 GJ to produce one m^3 of steam. There are 3.6 GJ/MWh; therefore, 1.84 GJ/m^3 / 3.6 GJ/MWh = .51 MWh/m^3. Assuming the power price is $90/MWh, then the .51 MWh/m^3 X $90/MWh = $46/m^3. Thus, using power to generate steam will cost about four times more than using gas. The price of power would have to drop to about $20/MWh before using power to create steam would make economic sense. However, there are new developing technologies where electricity can be used to heat oil under the ground.

In addition, as a technology, the SOFC is designed to produce as much power as possible for a given amount of fuel and to reduce heat not used for other purposes. There may be other SOFC designs or configurations that will produce more heat for steam production and less power, but they were not analyzed as part of the scope of this Study..

Figure F-10.
Power Price Sensitivity

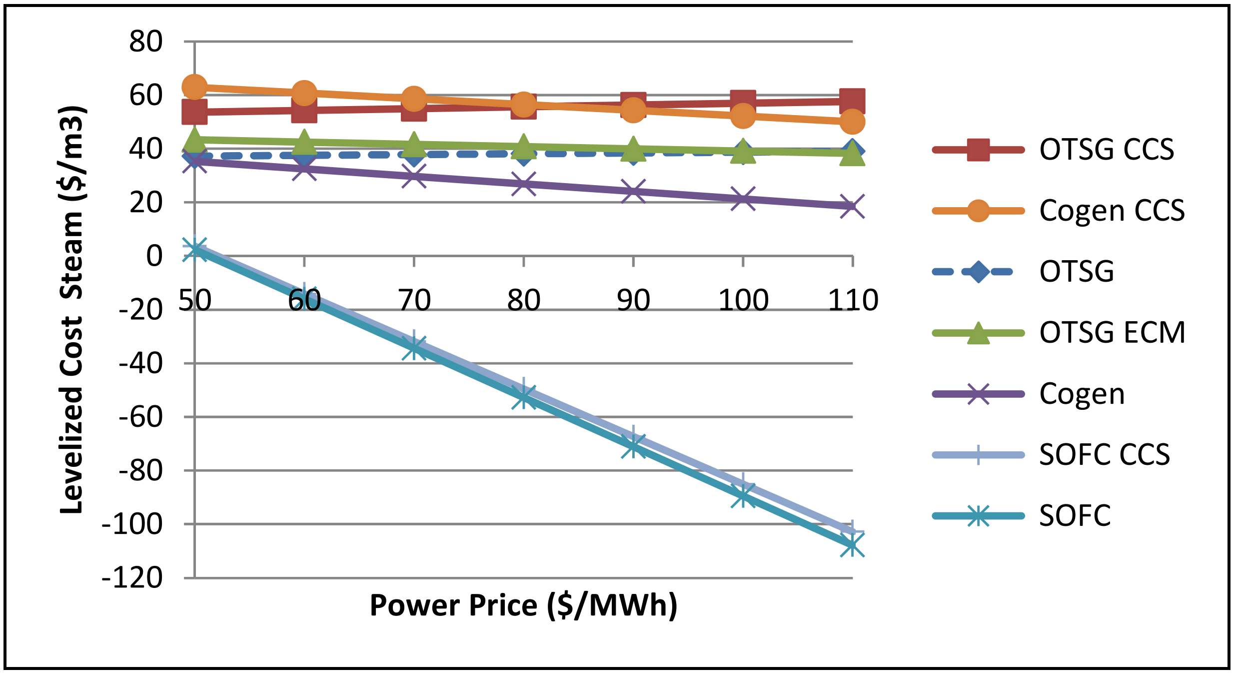
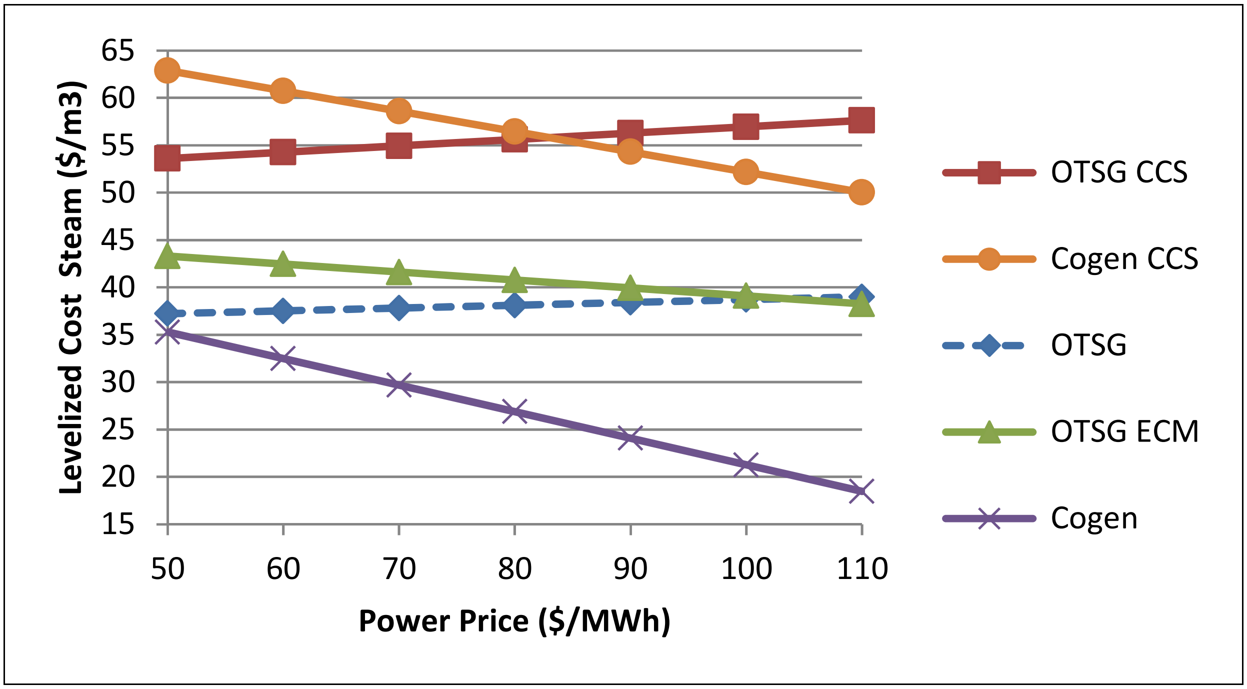


Figure F-11.
Power Price Sensitivity – without SOFC



Capital Cost

Given that some of the technologies evaluated have not been mass produced on the scale required of the options considered in this Study, there is some uncertainty about the capital costs of these technologies. Figure F-12 shows how the cost of steam changes as the capital cost changes.

As shown above, SOFC CCS had the largest component cost associated with capital, followed by Cogen and SOFC. It is expected that the cost of steam for the SOFC CCS case should change most with changes in capital cost. However, as shown in Figure F-12, the relative ranking of all the cases is generally preserved over a wide range of changes in capital cost. The SOFC cases have the lowest cost of steam even with a 100% increase in the capital costs. This graph is based on the assumption that both the underlying steam production equipment and the carbon capture equipment both increase in costs.

Figure F-12.
Capex Sensitivity

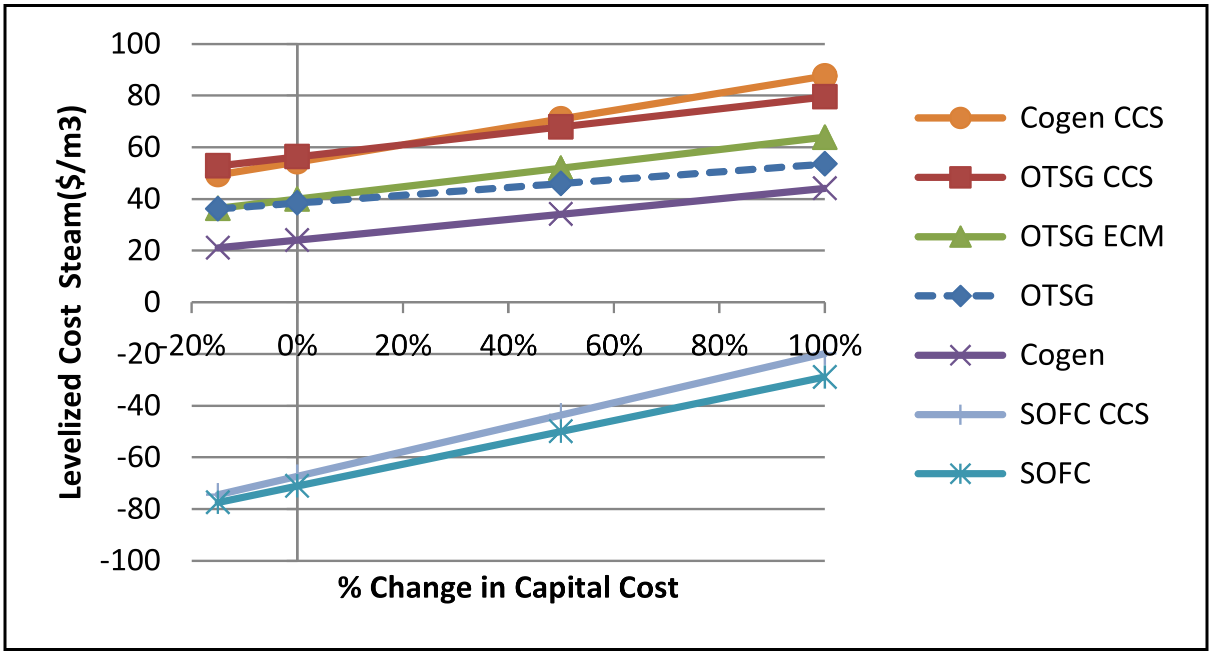
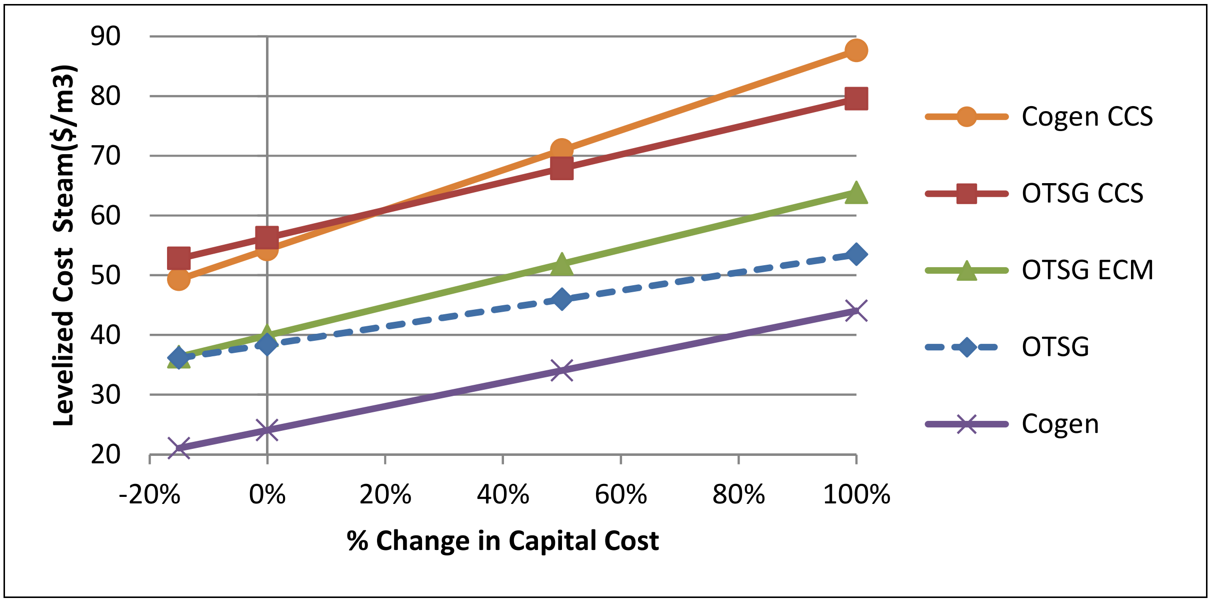


Figure F-13.
Capex Sensitivity – without SOFC



The current cost for SOFC is high compared to other forms of power generation. The expectation is that once SOFCs are mass produced their capital cost should decrease substantially. The capital cost of SOFC provided by the vendor was $900/kW; this is substantially lower than current prices. This value was escalated to $1,500/kW to account for costs in Alberta and U&O. Figure F-14 shows how the cost of steam increases as the cost of SOFC increases. Recall current costs of steam with an OTSG are close to $40/t. The cost of the SOFC would have to increase by a factor of about 3.5 or to about $5,300/kW to exceed the cost of steam from an OTSG. The cost of the SOFC would have to increase by a factor of 4 or to about $6,000 to exceed the cost of steam from an OTSG with CCS.

Figure F-14.
Steam Costs Relative to Capex Sensitivity

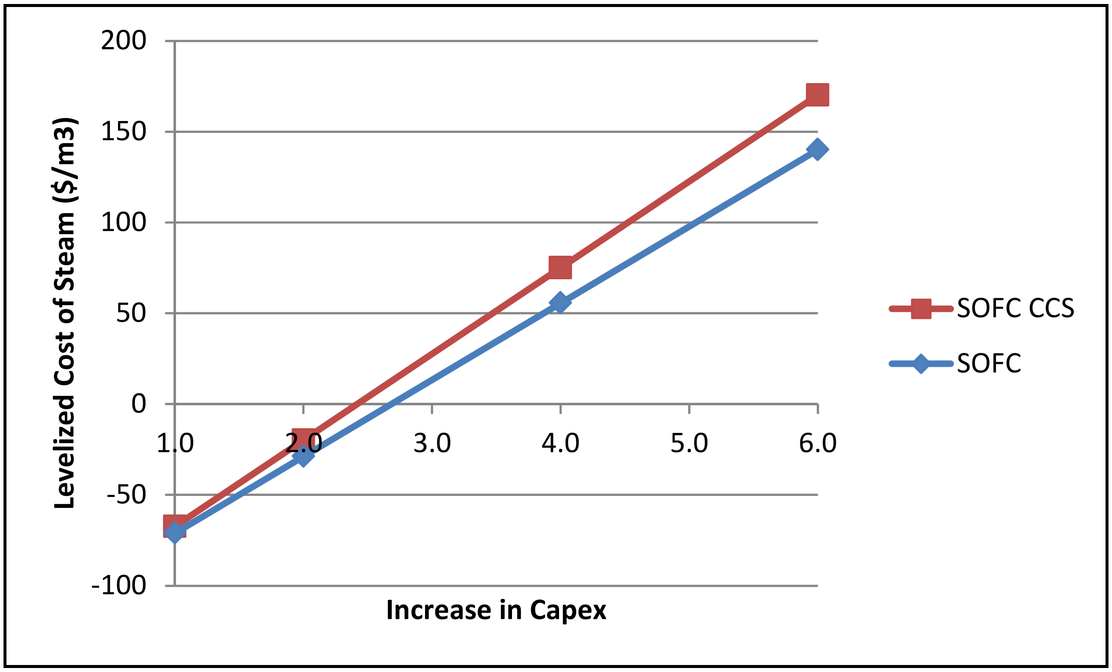


Figure F-14 shows how the cost of steam increases for increases in the whole cost of the plant, including the steam production equipment. Table F-5 shows the increase in steam cost for just a 100% increase in the capex for the CO_2 capture equipment. The first three rows show the base steam cost, the portion related to capital costs for CO_2 capture equipment, and the steam cost with the additional 100% increase in the capital cost for the CO_2 capture equipment. Even with a 100% increase in the capex for CO_2 capture equipment, the OTSG ECM case is only $10/m^3 higher than the OTSG case.

The middle part of the table shows the base cost of CO_2 capture, plus the capture cost component related just to capture capital followed by the cost of CO_2 capture with a 100% increase in the cost component related to CO_2 capture capital. The OTSG ECM case with a 100% increase in capture capital cost is still lower than the base CO_2 capture cost for the other two technologies.

The final part of this table shows the same analysis as for the avoided cost of CO_2 capture.

Table F-5.
Capex Sensitivity

	OTSG CCS	OTSG ECM	Cogen CCS
Base Steam Cost ($/m3)	56.3	39.9	54.3
Capex ($/m3)	8.1	8.9	13.3
Steam Cost +100% Inc. in Capex ($/m3)	64.4	48.8	67.6
Base CC ($/t)	152	39	142
Capex CC ($/t)	63	45	56
Capture Cost +100% Inc. in Capex ($/t)	214	84	198
Base Avoided Cost ($/t)	267	29	250
Capex Avoided ($/t)	111	34	99
Avoided Cost +100% Inc. in Capex ($/t)	378	63	349

CO_2 Compliance Costs

There is a significant amount of uncertainty regarding the cost to mitigate GHG emissions. Currently industry must mitigate about 12% of its GHG emissions and can do so by paying the CCEMC $15/tonne. Figure F-15 shows the impact of changes in the carbon price on this 12% of emissions to be mitigated. As explained above, the SOFC cases had a very large benefit associated with the value of credits sold from green power. Therefore, as shown in the graph below, the cost of steam for the SOFC cases changes most for the SOFC cases.

An increase in the CO_2 compliance cost materially decreases the cost of steam for the SOFC cases. Even over a wide range of the compliance costs the relative ranking does not change for most cases, except for the OTSG case. As expected, increasing compliance costs increase the cost of steam for the OTSG case but decrease it for all other cases because they sell credits. The compliance cost for a 40% reduction in emissions and a $40/t carbon price would be $133/t on the following graph. The OTSG is relatively unaffected by changes in CO_2 compliance cost because it is only mitigating 12% of its emissions. The other cases producing CO_2 credits related to CO_2 capture or for power sales have much larger volumes of CO_2 multiplied by $15/t. Therefore, they are more sensitive to CO_2 compliances costs. Figure F-15 shows that the mass of credits available is in some cases an order of magnitude greater than the direct compliance requirement.

Figure F-15.
CO_2 Compliance Cost Sensitivity

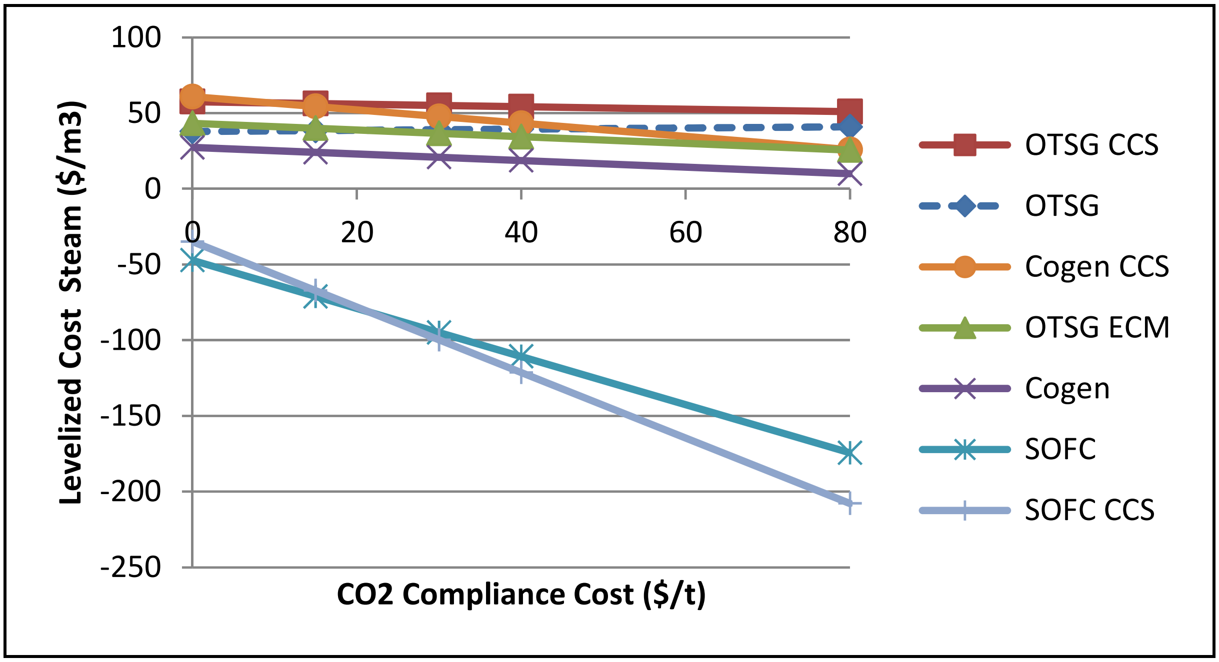
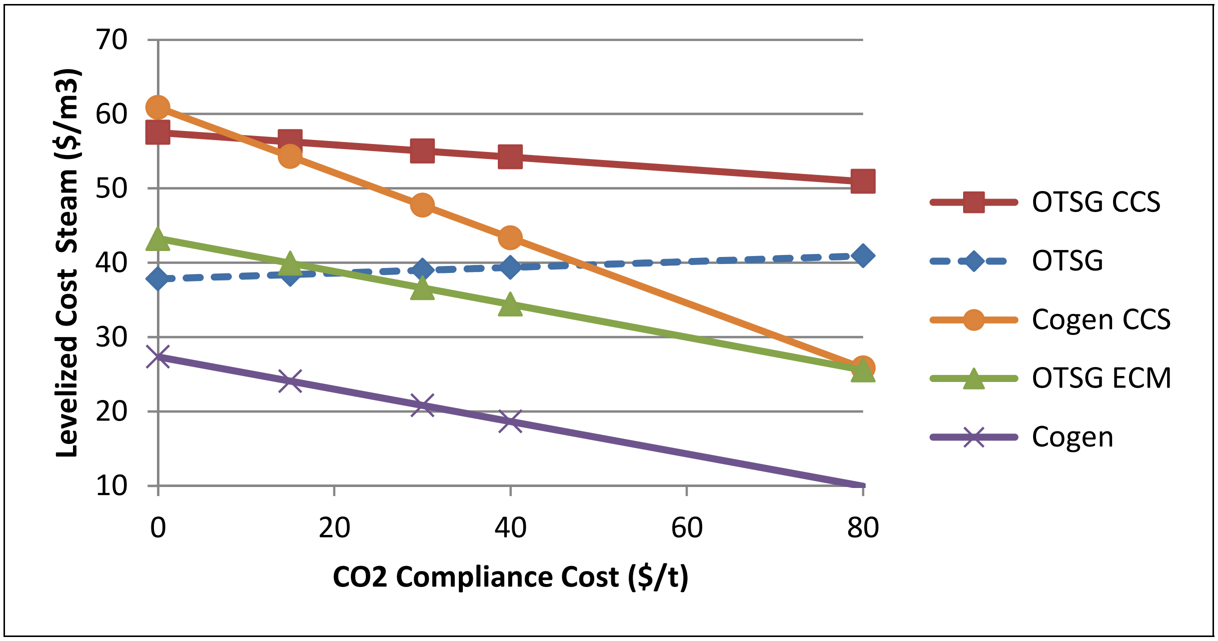


Figure F-16.
CO_2 Compliance Cost Sensitivity – without SOFC



Emission Intensity for Power

Currently in Alberta the grid emission intensity is about 0.88 t/MWh due to the high proportion of power generated by coal plants with high emission intensities. Power plants selling green power are provided GHG credits based on 0.65 t/MWh. Figure F-17 shows what would happen if the GHG emission intensity used to quantify GHG credits for green power changed. As expected, the SOFC cases are very sensitive to changes in the emission intensity. Even if power is not provided any CO_2 credits, the SOFC cases have the lowest steam production costs.

Figure F-17.
Impact of CO_2 Credits for Power Export Sensitivity

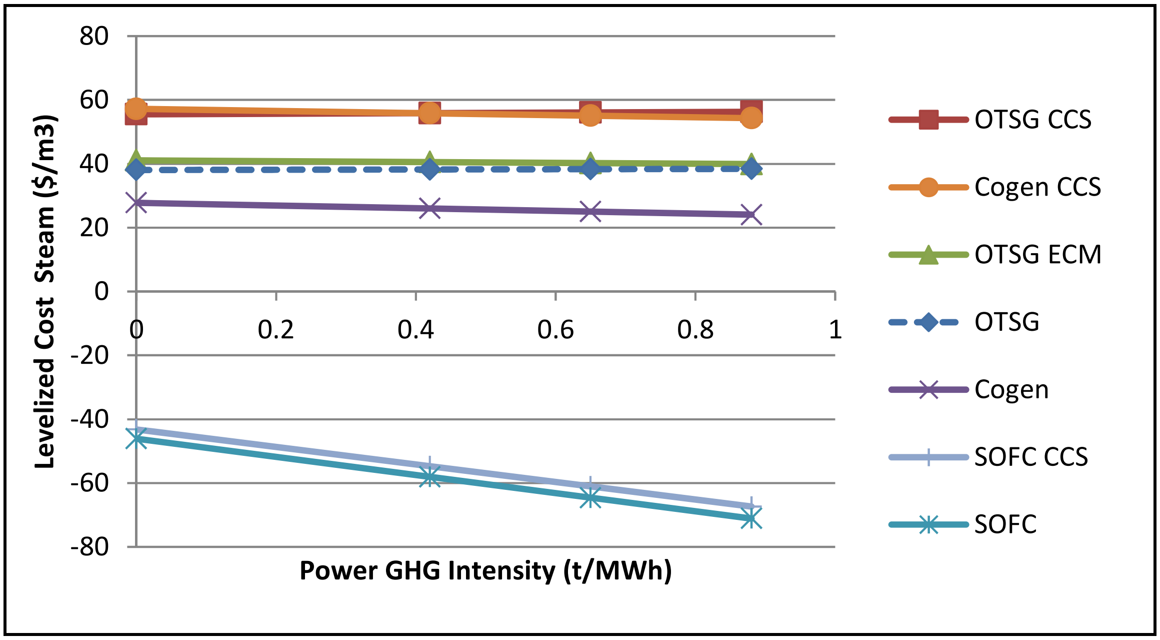
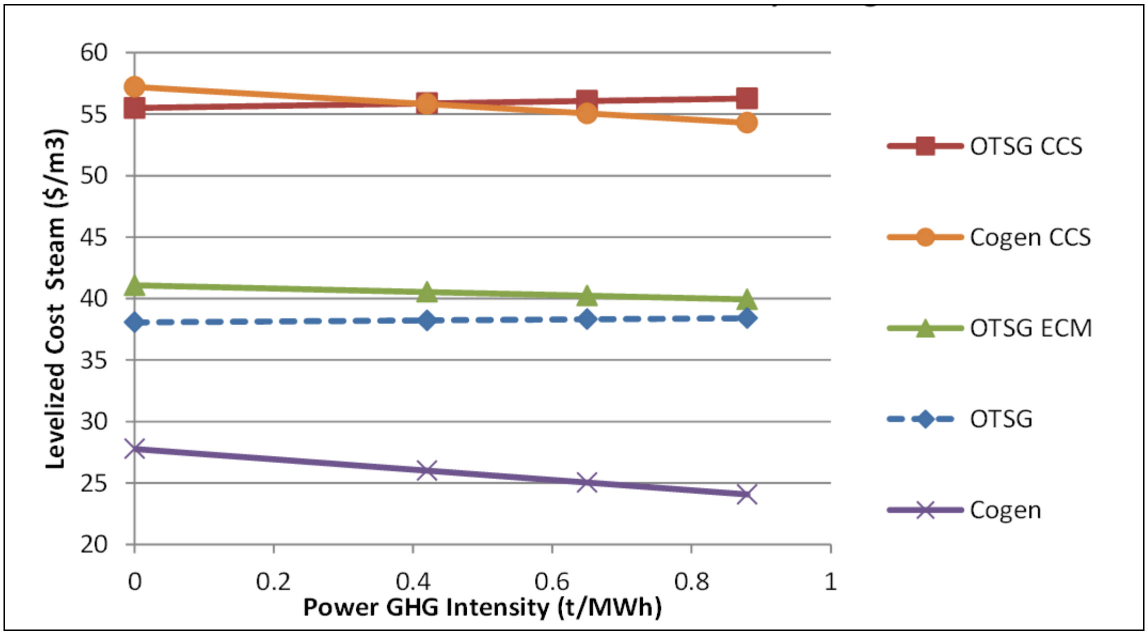
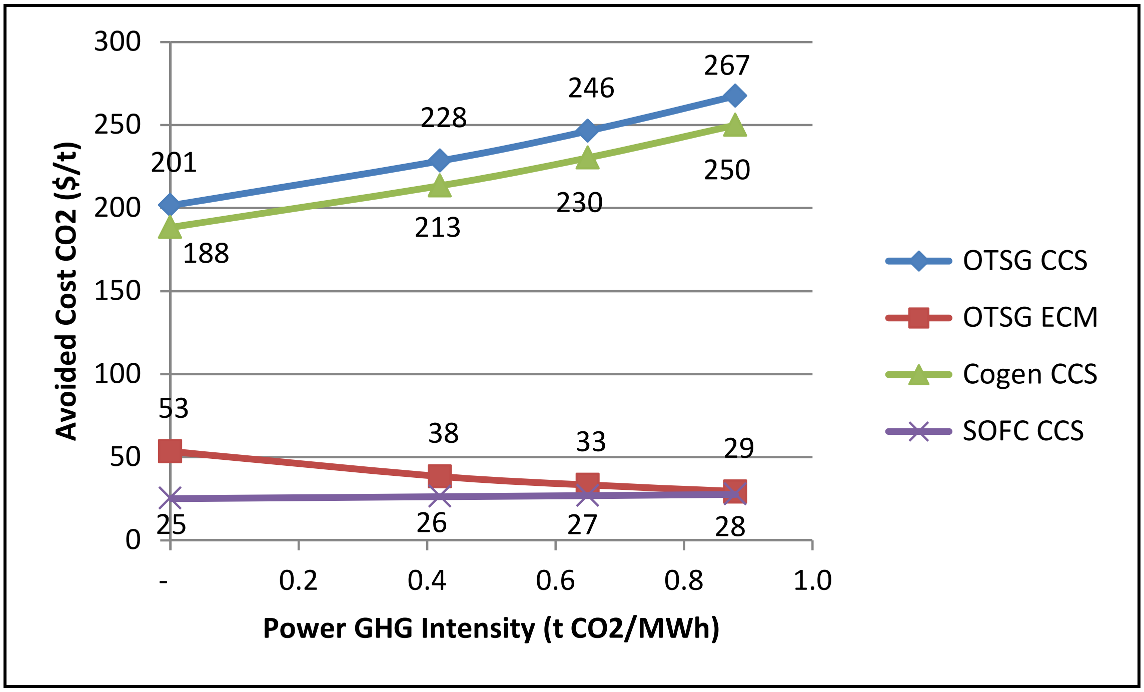


Figure F-18.
Impact of CO_2 Credits for Power Export Sensitivity – without SOFC



The GHG intensity of power will also have a substantial impact on avoided cost. Figure F-19 shows how the avoided cost of CO_2 changes as the GHG emission intensity of power ranges from 0 to 0.88 t/MWh.

Figure F-19.
Sensitivity to Power Generation CO_2 Intensity

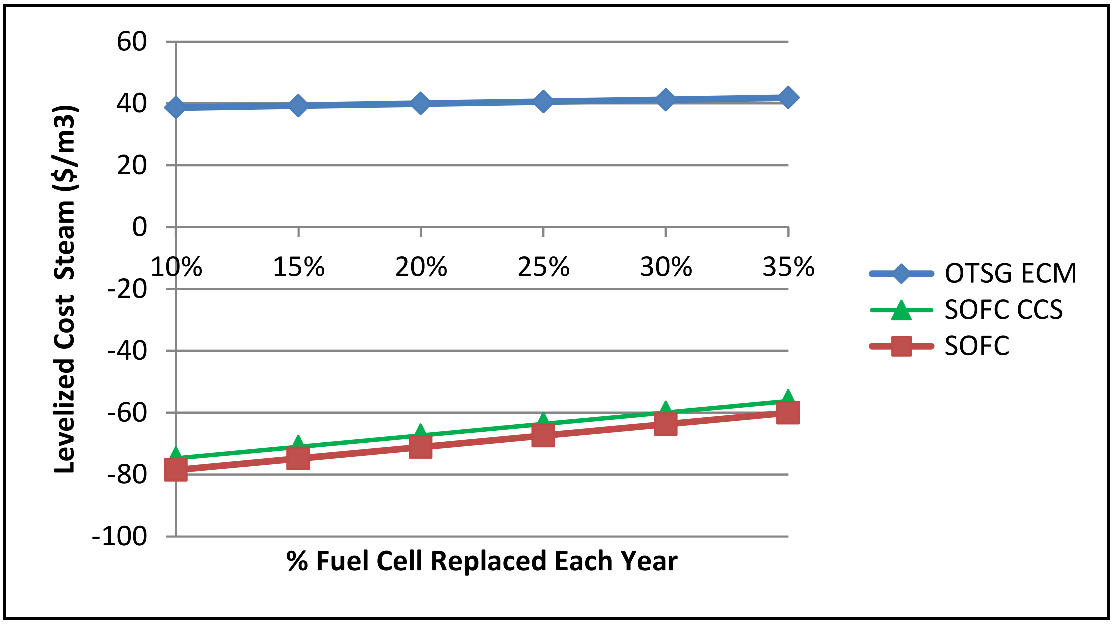


Fuel Cell Replacement

SOFCs are installed in equal portions over a 3-year period prior to January 1, 2016. When the first third of capacity is installed in 2013, it is assumed to produce power in 2014. When the next third of capacity is installed in 2014, two-thirds of the ultimate installed capacity is assumed to be producing power in early 2015. However, it is unlikely that the required amount of fuel cells can be installed in 2013. Therefore, the in-service date for the SOFC cases should likely be moved into the future beyond January 1, 2016. The fuel cells for the OTSG ECM case are unlikely to produce additional power prior to 2016 given the OTSG portion of the plant is not operating for a greenfield OTSG case. Thus, this option is assumed to begin producing power beginning in 2016, and fuel cells begin being replaced in 2021 for the OTSC ECM case.

However, SOFCs are only expected to operate for about five years, at which point the power production will decrease by about 10 percent. As such, beginning in 2018, 20% of the fuel cell capacity is replaced each year. The manufacturer believed the expected life can be increased to ten years in the future. Therefore, a sensitivity on fuel cell life is shown in Figure F-20. The replacement of fuel cells for the OTSG ECM case has little impact on the cost of steam. However, the replacement cost of fuel cells for the SOFC cases is more sensitive to changes in the pace of replacement.

Figure F-20.
Fuel Cell Replacement Cost Sensitivity



Economic Conclusions

The Cogen case offers a lower cost of steam than the OTSG case and may have a lower emission intensity than the OTSG case, depending upon how power credits are allocated. The Cogen CCS and OTSG CCS cases have a similar cost of steam production; however, the Cogen CCS case may have a lower GHG emission intensity than the OTSG CCS case, depending upon how the credits for power are treated. Cogens, though, produce a significant amount of power. It is conceivable that most of this power will need to be sold to the grid, putting the steam producer into the power business. Cogens with CCS also require a capital investment of 50% more than an OTSG with CCS.

The SOFC cases offer the greatest reduction in emission intensity largely due to the credit for power export. However, the emissions associated with steam and power production can be allocated in different ways, so other allocation options may yield lower emission intensities for steam. The cost of CO_2 capture for the SOFC cases is very low. The SOFC cases also have a very low cost of steam production due to the large amount of power produced.

One of the issues associated with the SOFC cases is the large amount of power produced and the large capital outlay required. The OTSG ECM case offers an emission intensity without producing a large amount of power and with a capital outlay less than an OTSG with CCS. It also has the virtue of having a low cost of CO_2 capture and a cost of steam production similar to that for an OTSG.

In Table F-6, the term "EI" represents emission intensity. Direct emission intensities assume all GHG emissions from the plant are allocated to steam. The D&I row assumes that in addition to direct emissions, GHG emissions associated with power purchases are added and GHG emissions associated with power sales are deducted in the calculation. (Negative intensities may not be allowed.)

Table F-6.
Economic Summary Data

	OTSG	OTSG CCS	OTSG ECM	Cogen	SOFC	Cogen CCS	SOFC CCS
Direct EI (t CO_2/bbl)	.053	.007	.005	.097	.235	.013	.000
D&I EI (t CO_2/bbl)	.063	.027	-.026	-.002	-.378	-.066	-.591
COS ($/$m^3$)	38.4	56.3	39.9	24.1	-71.1	54.3	-67.3
Capture Cost ($/t)		151.7	38.8			141.7	25.5
Capex $000's	795,337	1,224,919	1,112,396	1,054,536	1,426,060	1,755,757	1,699,917
Power (MW)	-14	-32	48	155	957	122	924

Other Metrics

Energy Required for Capture

One of the key metrics for assessing the cost of post combustion capture technologies is the amount of energy required to capture one tonne of CO_2. Table F-7 shows this energy requirement. The first three cases are for solvent-based CO_2 capture systems. The SOFC CCS case is based on a system using oxygen to burn out the unused fuel left in the CO_2. These values are based on the difference in natural gas consumption between the CCS case and the relevant base technology without CCS divided by the mass of CO_2 captured in a given year.

Table F-7.
Energy Required for Capture

	OTSG CCS	OTSG ECM	Cogen CCS	SOFC CCS
Energy (GJ HHV / t CO_2)	5.1	5.5 5.1		(.2)

Fuel Chargeable to Power Heat Rates

The fuel chargeable to power heat rate is often used to compare technologies that produce both steam and power. The fuel chargeable to power heat rate is derived by subtracting the fuel energy required to produce a certain quantity of steam from the fuel used in the case and dividing this term by the electric energy produced. The fuel energy required to produce steam is assumed to be the fuel used in the OTSG case without CCS. Table F-8 shows the fuel chargeable to power heat rate.

Table F-8.
Fuel Chargeable to Power Heat Rates

	OTSG ECM	Cogen	SOFC	Cogen CCS	SOFC CCS
FCP HR (GJ/MWh)	**10.8**	**7.6**	**5.1**	**16.1**	**5.2**

GHG Emission Intensities

One of the key metrics used to assess the GHG emissions associated with oil production is the GHG emission intensity. The emission intensities shown in Table F-9 relate just to the GHG emission associated with the production of steam used to recover oil. For the cases with CO_2 capture, the emissions are those that occur because the CO_2 capture system was unable to capture emissions related to steam production used to recover oil and steam produced to capture CO_2. Emission intensity is the mass of CO_2 emitted divided by the barrels of oil produced in a year.

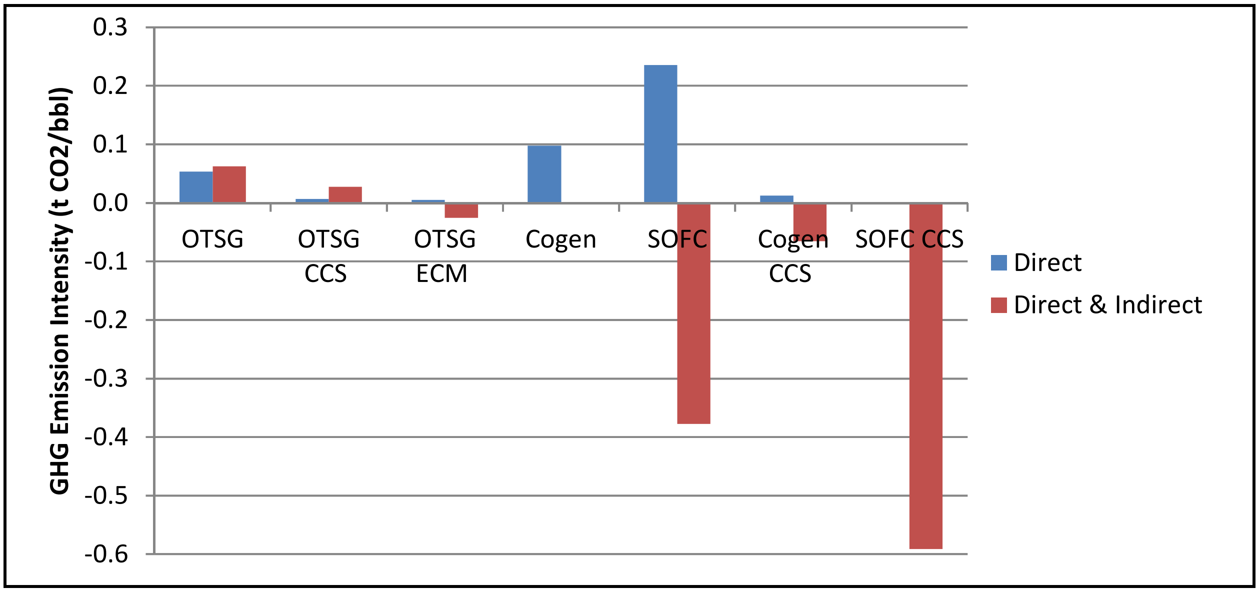
There are two emission intensity values reported in Table F-9. The first is related to the direct CO_2 emissions for an option. Direct emission intensities are based on the mass of CO_2 emitted. The Direct and Indirect intensities include the direct CO_2 emissions from the plant, plus indirect CO_2 emissions associated with power used to capture CO_2 or credits from additional power produced by the capture option at a rate of 0.88t/MWh. Some of the cases have negative GHG emission intensities because these cases produce a significant amount of additional power and GHG credits. However, generating a negative emission intensity for steam production may not be permitted. The first two OTSG cases see an increase in the Direct & Indirect emission intensities because they account for the emissions associated with using power from the grid to capture CO_2. The SOFC case has a high direct emission intensity largely because a significant amount of the natural gas used in the fuel cell is used to produce power. All of the CO_2 capture cases offer a significant reduction in the emission intensity compared to an OTSG. Both of the non-CO_2 capture cases have higher direct emission intensities than an OTSG.

Table F-9.
GHG Emission Intensities in tonnes of CO_2/bbl of Oil

	OTSG	OTSG CCS	OTSG ECM	Cogen	SOFC	Cogen CCS	SOFC CCS
Direct	.053	.007	.005	.097	.235	.013	.000
Difference on OTSG		.047	.048	-.044	-.182	.041	.053
Direct & Indirect	.063	.027	-.026	-.002	-.378	-.066	-.591

Emission Intensity in Figure F-21 below shows the emission intensity of steam used to recover oil. The blue values are the direct CO_2 emissions. The red bars include the indirect emissions associated with power purchases, and sales are included based on 0.88 t/MWh.

Figure F-21.
Direct and Indirect Emission Intensity of Steam



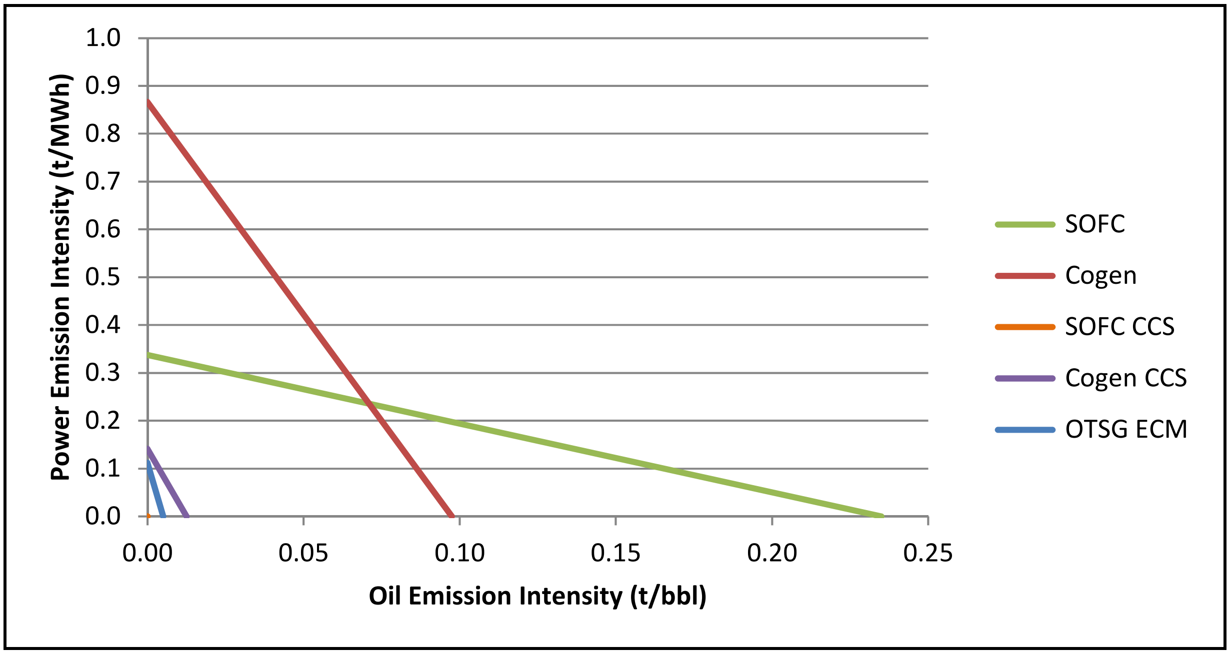
All of the CO_2 emissions from the plant are attributed to the production of steam and none to power in Table F-8 above. For this reason, any power produced will have a GHG emission intensity of zero and will generate GHG credits in the power market. However, if some of the GHG emissions are attributed to power, then the emission intensity of steam production will fall. The point is that there are two legitimate commodities being produced. There is no reason all the GHG emissions need to be allocated to steam production; in fact, doing so may overestimate the GHG emission intensity of the steam produced. The five right most Direct & Indirect values in Table F-9 are based on credits generated based on 0.88 t/MWh of power sold or to power purchases avoided. Table F-10 shows the GHG emission intensities for power assuming no GHG emissions are allocated to steam produced and all of the GHG emissions are allocated to power produced. To be clear, the GHG emission intensity of the steam produced would be zero. All cases but the Cogen case would likely still generate credits in the power market.

Table F-10.
GHG Emission Intensity Assuming All Emissions Allocated to Power

	OTSG ECM	Cogen	SOFC	Cogen CCS	SOFC CCS
Intensity (t/MWh)	.11	.87	.34	.14	.00

However, emission intensities will likely have to be greater than or equal to zero. Therefore, one could use a value between 0.0 t/MWh and 0.88 t/MWh and back out the GHG emission intensity for steam. Figure F-22 shows the range in emission intensity for power sales and steam for oil production. The values on the X axis assume all direct GHG emissions are allocated to steam and none to power. All the values on the Y axis assume all direct GHG emissions are allocated to power production and none to steam. The value for the SOFC CCS case is zero because close to 100% of the CO_2 generated is captured. The emission intensity for the OTSG with CCS case is fixed at 0.007 t/bbl because this case does not produce power. The base OTSG emission intensity is 0.053 t/MWh.

Figure F-22.
Range of Emission Intensity



For Cogens, credits are created based on deducting reasonable emission values associated with the CO_2 that would have otherwise been emitted to produce the steam supplied. We know that the OTSG case emits 644,225 tonnes CO_2/yr. If an emission intensity of 0.053 t/bbl of steam is applied to a Cogen, then from the graph above power would have an emission intensity of about 0.39 t/MWh. The power emission intensity of 0.39t/MWh for the Cogen case is similar to values derived for other projects. The SOFC emission intensity for power at 0.053t/bbl is 0.26 t/MWh. There may be an economic optimum allocation of GHG emissions between oil and power.

GHG Credits Generated

As discussed previously, there are two sources of CO_2 credits. The first is related to the reduction in GHG emissions associated with CO_2 capture, and the other is related to the production of incremental power. Emissions related to CO_2 capture are predicated on the notion that one would have to mitigate 12% of one's emissions if no CO_2 capture occurred. An entity cannot sell credits for the mass of CO_2 it is obligated to mitigate. It can only sell credits for the mass of CO_2 it captures over and above its obligations to mitigate. Therefore, it is assumed that the baseline for this calculation would be the emissions that would have occurred if no CO_2 capture had been completed on the underlying technology without CCS. Given that all GHG emissions are allocated to the production of steam, all power sold to the grid is assumed to have a GHG emission intensity of zero and is therefore allocated 0.88 t CO_2/MWh of credits.

Table F-11 shows the values used to estimate the amount of GHG obligations and credits. Two sources of CO_2 emissions were considered in this assessment: direct CO_2 emissions and emissions associated with power used from the grid. The Direct Compliance requirement in the fifth row is 12% of the first row CO_2 Produced. It was also assumed that the entity producing steam will be obligated to purchase credits to mitigate GHG emissions associated with power usage from the grid. Generally speaking, this is the obligation of the power producer. The compliance requirement is the sum of the 12% emission reduction requirement and the GHG mitigation requirements associated with power purchases. CO_2 credits available for sale are based on the mass of CO_2 captured less 12% of the CO_2 generated plus credits associated with power sales.

The OTSG ECM case generates credits that exceed the mass of CO_2 produced in the base OTSG case. Some of the credits produced are related to the extra power produced and sold to the grid. However, the direct reductions include CO_2 captured from the base OTSG plus emissions associated with the extra fuel required. This value is greater than the base OTSG emissions as well. Taking credit for the emissions associated with the extra fuel required may not be permitted under the rules. Table F-11 also shows that if Cogens without CCS received 0.88t/MWh related to CO_2 credits in the power market, they could effectively offset all of the CO_2 emissions from the plant.

Table F-11.
Values Used to Derive GHG Obligations and Credits

GHG Production	OTSG	OTSG CCS	OTSG ECM	Cogen	SOFC	Cogen CCS	SOFC CCS
CO2 Produced (t/yr)	**644,225**	**828,550**	**865,780**	**1,174,205**	**2,830,575**	**1,510,370**	**2,780,205**
CO2 Captured (t/yr)	-	**745,695**	**804,825**	-	-	**1,359,260**	**2,779,475**
CO2 Emissions (t/yr)	644,225	82,855	60,955	1,174,205	2,830,575	151,110	730
% Comp. Requirement (t/yr)	**12%**	12%	12%	12%	12%	12%	12%
Compliance Requirement (t/yr)	77,307	-	-	140,905	339,669	-	-
GHG Compliance							
Direct Compliance Req. (t/yr)	77,307	-	-	140,905	339,669	-	-
GHG From Power Purchases (t/yr)	109,773	248,223	-	-	-	-	-
Compliance Requirement (t/yr)	187,080	248,223	-	140,905	339,669	-	-
GHG Credits							
Direct Reductions in CO2 (t/yr)	-	646,269	700,931	-	-	1,178,016	2,445,850
Credits for Clean Power Sales (t/yr)	-	-	368,481	1,192,551	7,379,634	940,474	7,122,931
Credits Available for Sale (t/yr)	-	646,269	1,069,412	1,192,551	7,379,634	2,118,489	9,568,782

Capture Cost Assuming Standalone ECM as Part of Base Case

Another base case option is to have an OTSG with an independent ECM producing 47.8 MW the net output for use as the reference case for the OTSG ECM case. The cost of the ECM can be determined by taking the cost components for the OTSG ECM case and subtracting them from the OTSG. We also have to add back the value of the power that could have been sold if not used by the OTSG. Furthermore, we cannot take credit for CO_2 captured by the ECM on the OTSG. (There will still be CO_2 captured from the ECM and CO_2 emitted to produce the oxygen now required.) That is 39.93 - (38.40 - 3.32X (1 -.344) + 2.32) = \$1.39 / m^3 steam. For a 47.8 MW case we need to multiply this by 47.8 / (47.8 + 14.2) = \$1.07/m^3. When we calculate the capture cost we add back the CO_2 mitigation costs and any CO_2 credits generated. For the ECM standalone this credit is 3.32*.334*47.8/(47.8+14.2) =\$ .88/m^3.

The revised cost of capture is:

$$((39.9 + 1.07 + .88) - (38.4 - .58)) \times 5,744,742 / 808,825 = \$28.6 /t\ CO_2$$

The cost of power for the ECM is about:

$$(39.93 - (-3.32 \times (1-.344) - 38.40 + 7.8) \times 5,744,742 / ((47.8 + 14.2) \times 8,760) = \$122/MWh$$

Section G.

Findings and Recommendations

Findings

Based on the Study, Jacobs Consultancy's findings are summarized as follows:

- The Solid Oxide Fuel Cell ("SOFC") is a power producer with heat as a by-product and is a mismatch for thermal in-situ, which has a large heat load and small power load. A typical thermal in-situ plant has a demand of about 10 MW of heat for every MW (thermal equivalent) of power. An SOFC, on the other hand, produces about one MW of heat for every 7 MW (thermal equivalent) of power. This results in a mismatch of 70 to one as compared to the requirements for bitumen production. Any efforts to reduce the power-to-heat ratio on an SOFC just result in bypassing fuel around the anode side and combusting it with air which, in effect, is like having an OTSG to produce steam with a small SOFC to produce the power for the site.
- However, combined heat and power ("CHP") can be attractive for a thermal in-situ site provided a better match between heat and power. For example:
  - Conventional cogen produces steam with a cost about 40% lower than WLS+OTSG with a manageable amount of power sold to grid.
- CO_2 capture and compression increase the cost of producing steam.
- However, fuel cells are relatively cost effective for CO_2 recovery assuming power produced can be sold to grid
  - ECM on conventional OTSG captures and compresses CO_2 at a cost that is roughly one-third of conventional PCC with manageable power sales.
  - On paper, due to **very high power sales**, SOFC has capture and compression costs that are roughly one-fourth of conventional CHP with PCC (but again, there is a mismatch of heat and power for SAGD).
- ECM is a potential "bolt-on" carbon capture technology for existing or new facilities including:
  - OTSG / Cogen-based thermal in-situ
  - SMR
  - Petcoke combustion
  - Other fired heaters

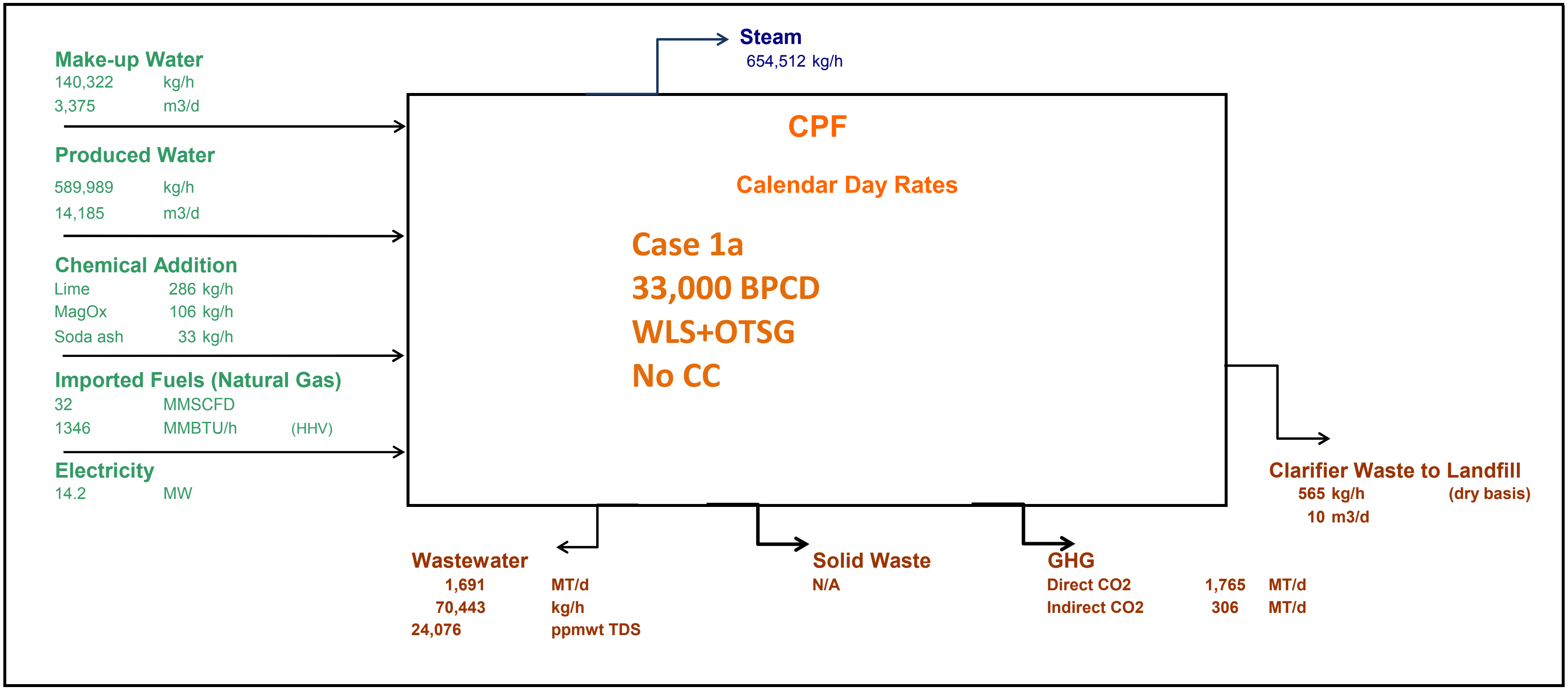
Recommendations

Our recommendations are summarized below:

- Focus efforts on SOFC as a power producer, not as a CHP technology for thermal in-situ plants
- ECM within the context of CEPACS may have merit for the oil sands industry; however, key risks remain, including:
    - Capital cost
    - Reliability, operability and maintenance costs
- Therefore, we recommend in the near term:
    - Feasibility studies on a commercial-scale plant to confirm capital costs and commercial viability for:
        - Cogen and OTSG-based thermal in-situ, SMR and Fired Heaters
        - High-level screening evaluation for ECM on petcoke combustion
        - A demonstration facility at the megawatt scale will confirm reliability, operability, and commercial readiness, and highlight other issues

Appendix 1

High Level Material Balances



CPF

Calendar Day Rates

Case 1a
33,000 BPCD
WLS+OTSG
No CC

Make-up Water

140,322	kg/h
3,375	m3/d

Produced Water

589,989	kg/h
14,185	m3/d

Chemical Addition

Lime	286 kg/h
MagOx	106 kg/h
Soda ash	33 kg/h

Imported Fuels (Natural Gas)

32	MMSCFD	
1346	MMBTU/h	(HHV)

Electricity

14.2	MW

Steam

654,512 kg/h

Clarifier Waste to Landfill

565 kg/h (dry basis)
10 m3/d

Wastewater

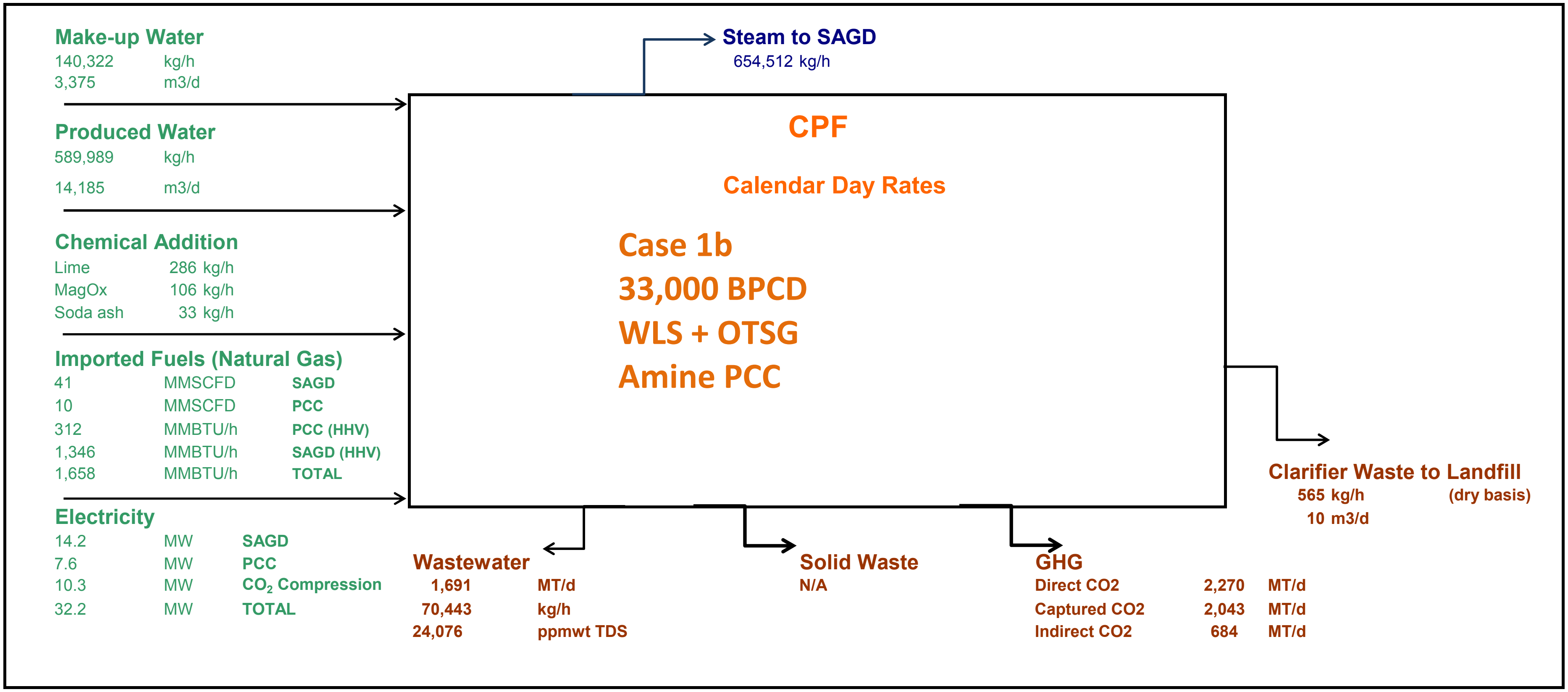
1,691	MT/d
70,443	kg/h
24,076	ppmwt TDS

Solid Waste

N/A

GHG

Direct CO2	1,765	MT/d
Indirect CO2	306	MT/d



Make-up Water

140,322	kg/h
3,375	m3/d

Produced Water

589,989	kg/h
14,185	m3/d

Chemical Addition

Lime	286	kg/h
MagOx	106	kg/h
Soda ash	33	kg/h

Imported Fuels (Natural Gas)

41	MMSCFD	SAGD
10	MMSCFD	PCC
312	MMBTU/h	PCC (HHV)
1,346	MMBTU/h	SAGD (HHV)
1,658	MMBTU/h	TOTAL

Electricity

14.2	MW	SAGD
7.6	MW	PCC
10.3	MW	CO_2 Compression
32.2	MW	TOTAL

Steam to SAGD

654,512 kg/h

CPF

Calendar Day Rates

Case 1b

33,000 BPCD

WLS + OTSG

Amine PCC

Clarifier Waste to Landfill

565 kg/h (dry basis)

10 m3/d

Wastewater

1,691	MT/d
70,443	kg/h
24,076	ppmwt TDS

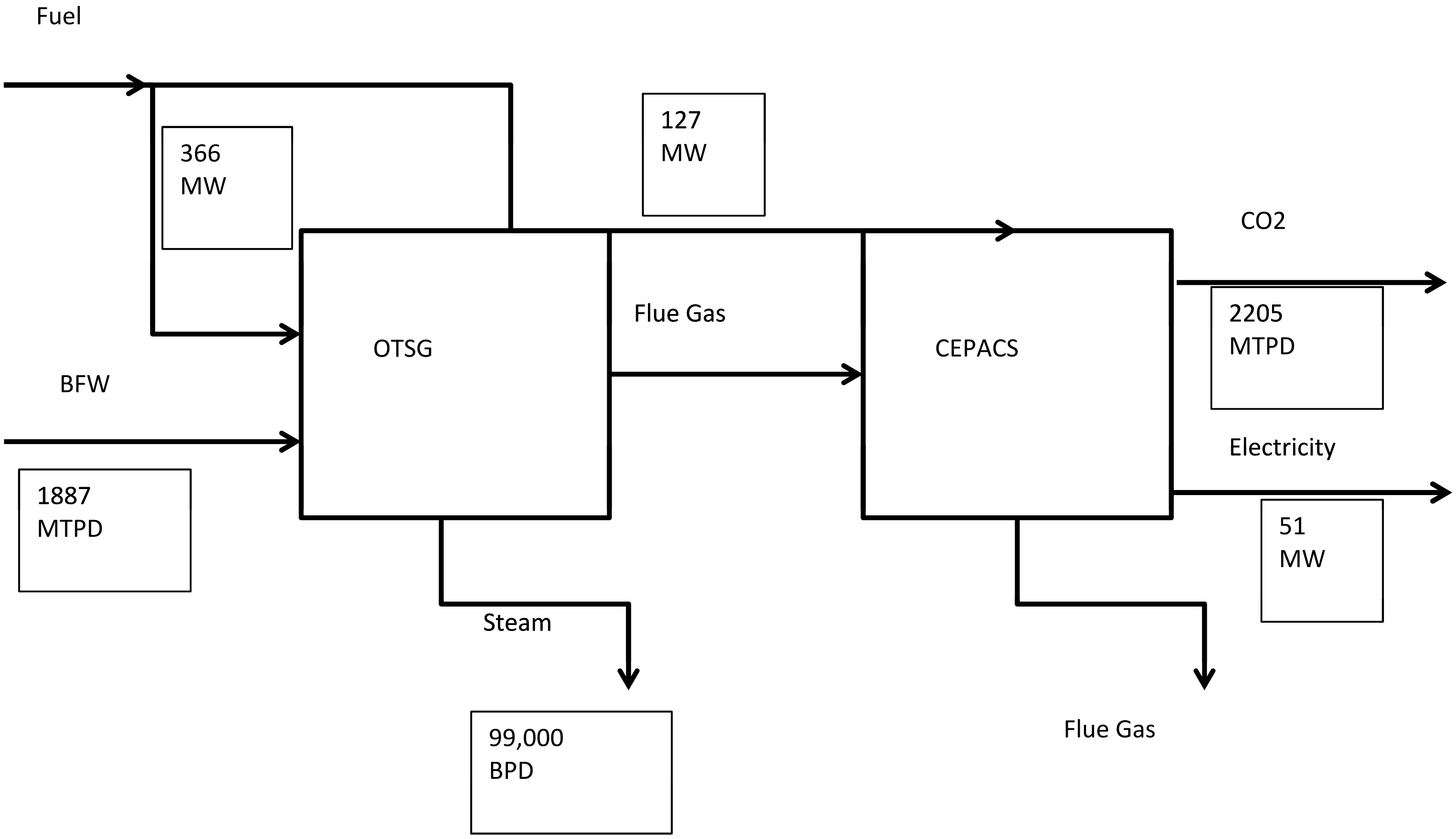
Solid Waste

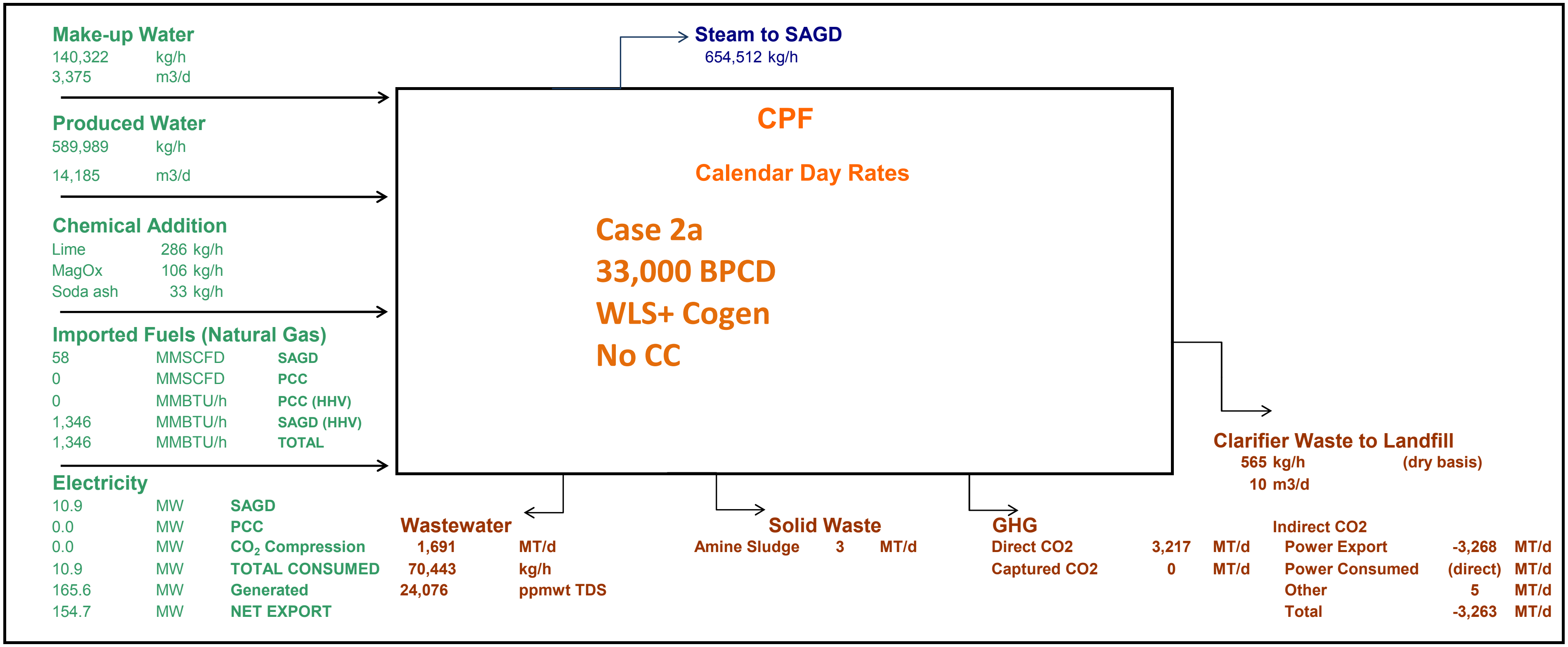
N/A

GHG

Direct CO2	2,270	MT/d
Captured CO2	2,043	MT/d
Indirect CO2	684	MT/d

Case 1C Balances





Make-up Water

140,322	kg/h
3,375	m3/d

Produced Water

589,989	kg/h
14,185	m3/d

Chemical Addition

Lime	286	kg/h
MagOx	106	kg/h
Soda ash	33	kg/h

Imported Fuels (Natural Gas)

58	MMSCFD	SAGD
0	MMSCFD	PCC
0	MMBTU/h	PCC (HHV)
1,346	MMBTU/h	SAGD (HHV)
1,346	MMBTU/h	TOTAL

Electricity

10.9	MW	SAGD
0.0	MW	PCC
0.0	MW	CO_2 Compression
10.9	MW	TOTAL CONSUMED
165.6	MW	Generated
154.7	MW	NET EXPORT

Steam to SAGD

654,512 kg/h

CPF

Calendar Day Rates

Case 2a
33,000 BPCD
WLS+ Cogen
No CC

Clarifier Waste to Landfill

565	kg/h	(dry basis)
10	m3/d	

Wastewater

1,691	MT/d
70,443	kg/h
24,076	ppmwt TDS

Solid Waste

Amine Sludge	3	MT/d

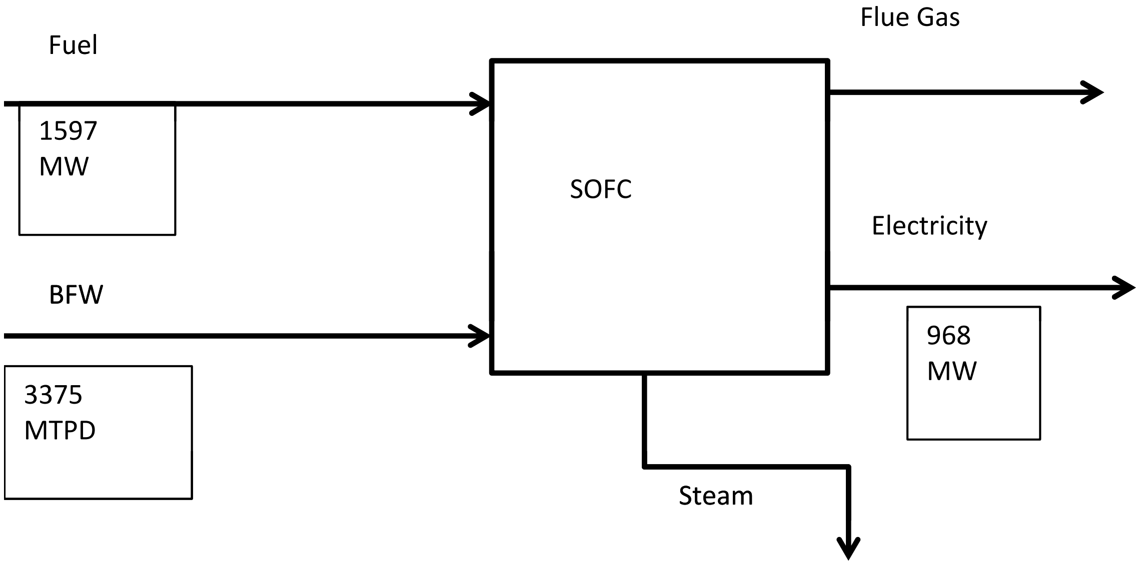
GHG

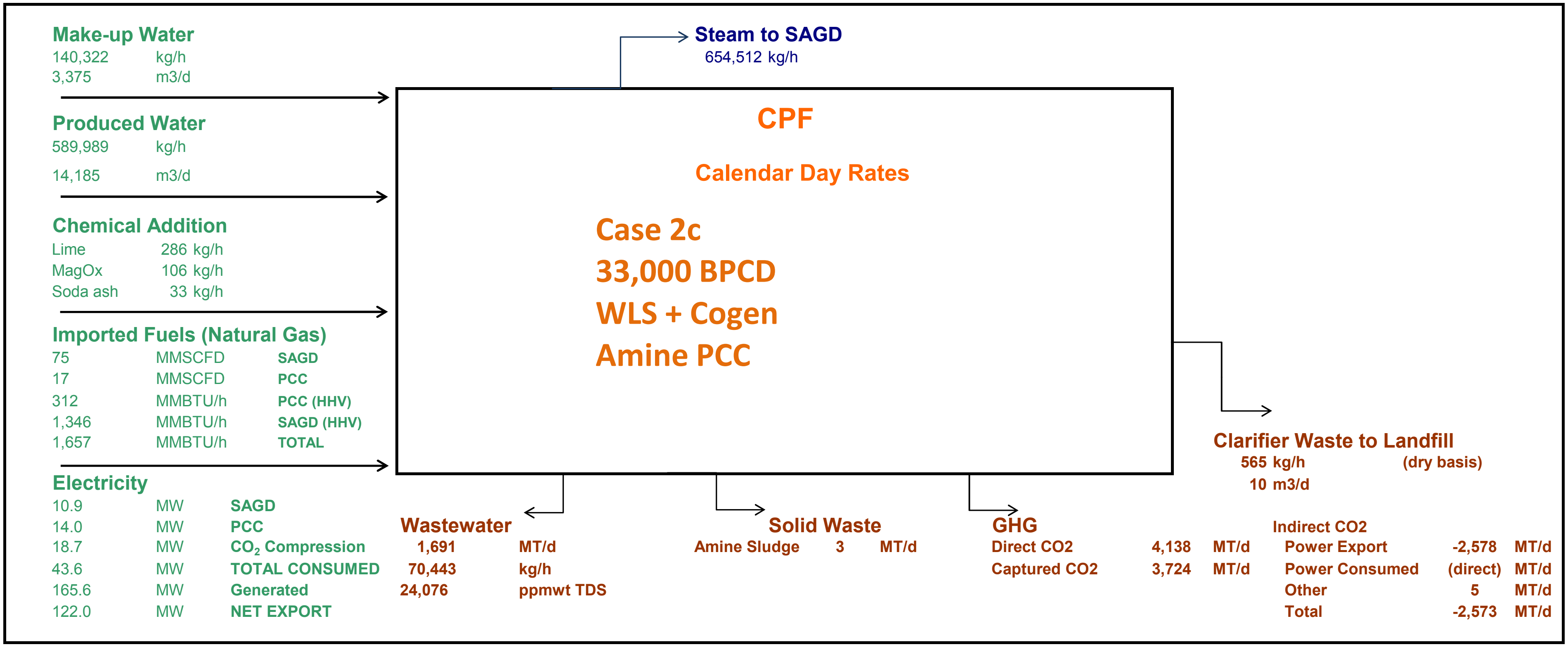
Direct CO2	3,217	MT/d
Captured CO2	0	MT/d

Indirect CO2

Power Export	-3,268	MT/d
Power Consumed	(direct)	MT/d
Other	5	MT/d
Total	-3,263	MT/d

Case 2b SOC


Fuel
1597 MW
BFW
3375 MTPD
SOFC
Flue Gas
Electricity
968 MW
Steam



Make-up Water
140,322 kg/h
3,375 m3/d

Produced Water
589,989 kg/h
14,185 m3/d

Chemical Addition

Lime	286 kg/h
MagOx	106 kg/h
Soda ash	33 kg/h

Imported Fuels (Natural Gas)

75	MMSCFD	**SAGD**
17	MMSCFD	**PCC**
312	MMBTU/h	**PCC (HHV)**
1,346	MMBTU/h	**SAGD (HHV)**
1,657	MMBTU/h	**TOTAL**

Electricity

10.9	MW	**SAGD**
14.0	MW	**PCC**
18.7	MW	**CO_2 Compression**
43.6	MW	**TOTAL CONSUMED**
165.6	MW	**Generated**
122.0	MW	**NET EXPORT**

Steam to SAGD
654,512 kg/h

CPF

Calendar Day Rates

Case 2c
33,000 BPCD
WLS + Cogen
Amine PCC

Clarifier Waste to Landfill
565 kg/h (dry basis)
10 m3/d

Wastewater
1,691 MT/d
70,443 kg/h
24,076 ppmwt TDS

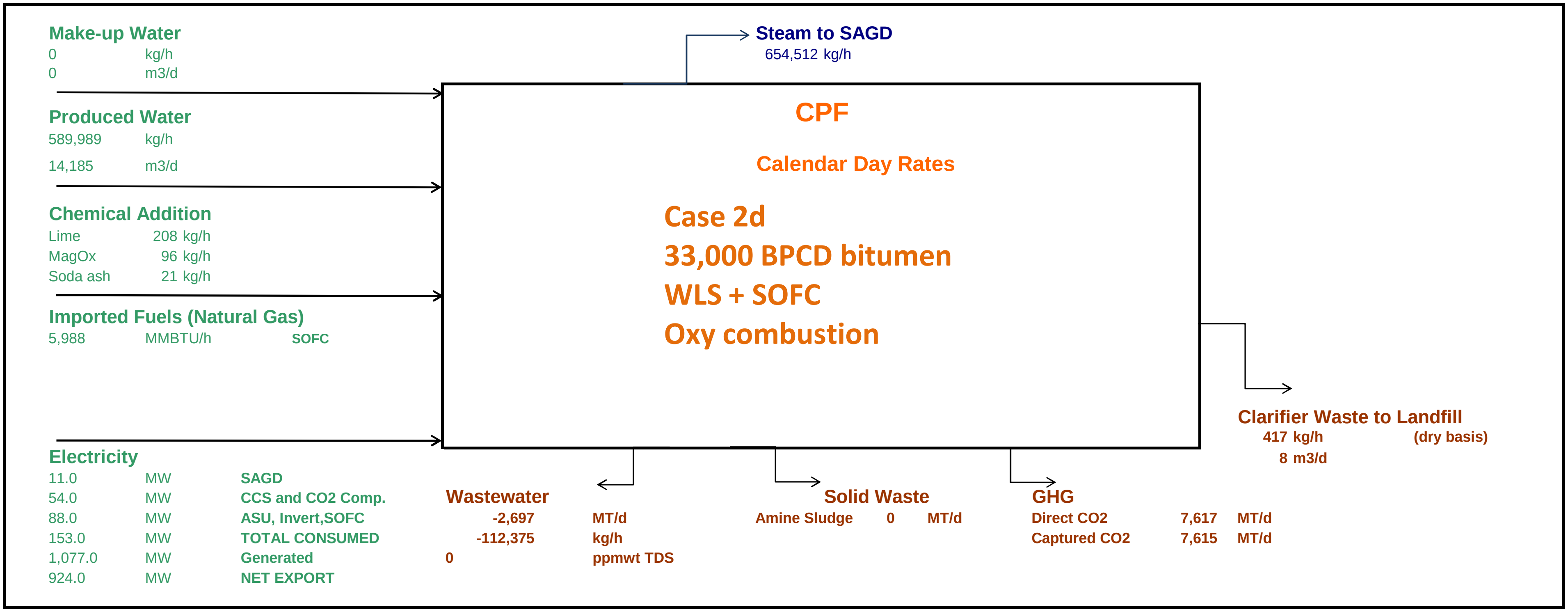
Solid Waste
Amine Sludge 3 MT/d

GHG

Direct CO2	**4,138**	**MT/d**
Captured CO2	**3,724**	**MT/d**

Indirect CO2

Power Export	**-2,578**	**MT/d**
Power Consumed	**(direct)**	**MT/d**
Other	**5**	**MT/d**
Total	**-2,573**	**MT/d**



Make-up Water

0	kg/h
0	m3/d

Produced Water

589,989	kg/h
14,185	m3/d

Chemical Addition

Lime	208	kg/h
MagOx	96	kg/h
Soda ash	21	kg/h

Imported Fuels (Natural Gas)

5,988	MMBTU/h	SOFC

Electricity

11.0	MW	SAGD
54.0	MW	CCS and CO2 Comp.
88.0	MW	ASU, Invert,SOFC
153.0	MW	TOTAL CONSUMED
1,077.0	MW	Generated
924.0	MW	NET EXPORT

Steam to SAGD

654,512 kg/h

CPF

Calendar Day Rates

Case 2d

33,000 BPCD bitumen

WLS + SOFC

Oxy combustion

Clarifier Waste to Landfill

417	kg/h	(dry basis)
8	m3/d	

Wastewater

-2,697	MT/d
-112,375	kg/h
0	ppmwt TDS

Solid Waste

Amine Sludge	0	MT/d

GHG

Direct CO2	7,617	MT/d
Captured CO2	7,615	MT/d

Appendix 2

Technical Data Summary

Calendar Day Rates by Case (Rev 13)		1a OTSG	1b OTSG+PCC	1c OTSG+ECM	2a Cogen	2b SOFC	2c Cogen+PCC
Bitumen Production	BPD	33,000	33,000	33,000	33,000	33,000	33,000
Steam Production for SAGD	BPD CWE	99,000	99,000	99,000	99,000	99,000	99,000
Steam Production for Carbon Capture	BPD CWE		21,425				39,061
Sulphur Production	tonnes/day	0.03	0.03	0.03	0.03	0.03	0.03
Disposal Water							
Steam Production for SAGD	m3/day CWE	1,693	1,693	1,693	1,693	1,693	1,693
PCC Boiler System	m3/day CWE	0	68	0	0	0	125
Total	m3/day CWE	1,693	1,762	1,693	1,693	1,693	1,818
WLS Sludge	tonnes/day	20	20	20	20	20	20
WLS Sludge	CAD/day	3,053	3,053	3,053	3,053	3,053	3,053
Makeup Water Rate							
Steam Production for SAGD	m3/day	3,375	3,375	3,375	3,375	3,375	3,375
PCC	m3/day		68	0			125
ECM/SOFC Water Produced	m3/day			-1,488			
Total	m3/day	3,375	3,443	1,887	3,375	0	3,499
Water Treating Chemicals							
Lime	tonnes/day	6.9	6.9	6.3	6.9	6.9	6.9
MagOx	tonnes/day	2.6	2.6	2.6	2.6	2.6	2.6
Soda Ash	tonnes/day	0.8	0.8	0.7	0.8	0.8	0.8
Total Water Treating Chemicals	CAD/day	15,841	15,841	15,049	15,841	15,841	15,841
Amine Carbon Capture Chemicals							
MEA	MMCAD/yr		1.1				2.0
Corrosion Inhibitor	MMCAD/yr		0.2				0.4
Mole Sieve	MMCAD/yr		0.05				0.09
Total	MMCAD/yr		1.3				2.4
Amine Sludge Disposal	MMCAD/yr		0.8				1.4
Produced Gas Consumed							

Calendar Day Rates by Case (Rev 13)		1a OTSG	1b OTSG+PCC	1c OTSG+ECM	2a Cogen	2b SOFC	2c Cogen+PCC
Bitumen Production (LHV)	GJ/day	898	898	898	898	898	898
Bitumen Production (HHV)	GJ/day	1,002	1,002	1,002	1,002	1,002	1,002
Natural Gas Consumed							
Bitumen Production (LHV)	GJ/day	30,712	30,712	30,712			
Bitumen Production (HHV)	GJ/day	34,073	34,073	34,073			
Cogen Steam & Power Production (LHV)	GJ/day				56,153		56,153
Cogen Steam & Power Production (HHV)	GJ/day				62,299		62,299
Carbon Capture (LHV)	GJ/day		9,302				16,960
Carbon Capture (HHV)	GJ/day		10,320				18,816
ECM (LHV)	GJ/day			10,946			
ECM (HHV)	GJ/day			12,145			
SOFC (LHV)	GJ/day					137,090	
SOFC (HHV)	GJ/day					152,084	
Total (LHV)	GJ/day	30,712	40,014	41,658	56,153	137,988	73,113
Total (HHV)	GJ/day	34,073	44,394	46,218	62,299	153,086	81,115
Power Consumed							
Bitumen Production	MW	14	14	14	11	11	11
Carbon Capture	MW		8				14
CO2 Compression	MW	0	10		0		19
ECM	MW			11			
SOFC						61	
Inverter losses				3		43	
Total Power Consumed	MW	14	32	28	11	115	44
Power Generated	MW	0	0	76	166	1072	166
Power Imported	MW	14	32	0	0		0
Power Exported	MW	0	0	48	155	957	122
CO2 Direct							
Steam Production for SAGD	tonnes/day	1,765	1,765	1,765			
Cogen Steam & Power Production Total	tonnes/day				3,217		3,217

Calendar Day Rates by Case (Rev 13)		1a OTSG	1b OTSG+PCC	1c OTSG+ECM	2a Cogen	2b SOFC	2c Cogen+PCC
Carbon Capture Steam Generation	tonnes/day		505				921
ECM Operation	tonnes/day			608			
SOFC	tonnes/day					7,755	
CO2 Direct Total	tonnes/day	1,765	2,270	2,373	3,217	7,755	4,138
CO2 Indirect	tonnes/day						
Without Power	tonnes/day	5	5	5	5	5	5
CO2 Captured	tonnes/day	0	2,043	2,205	0	0	3,724
CO2 Generated (Direct)	tonnes/day	1,765	227	168	3,217	7,755	414
% CO2 Capture (Direct Only)		0%	90%	93%	0%	0%	90%
Total Costs							
Total Chemicals & Disposal Costs	MM CAD/yr	7	11	7	7	7	13
Fixed O&M Costs	MM CAD/yr	26	40	36	34	46	57
Capital Cost (3Q2012, Ft McMurray)	MM CAD	735	1,132	1,023	975	1,311	1,623
Capital Spend Profile							
Start up - 5yr		0%	0%	0%	0%	0%	0%
Start up - 4yr		0%	0%	0%	0%	0%	0%
Start up - 3yr		15%	15%	30%	15%	30%	15%
Start up - 2yr		55%	55%	50%	55%	50%	55%
Start up - 1yr		30%	30%	20%	30%	20%	30%
Steam Generation							
Availability Factor		94%	94%	94%	94%	94%	94%
Wet Steam Quality		77%	77%	77%	77%	77%	77%
Fuel Cell Replacement costs	MMCAD/yr			18		80	

Appendix 3

Publicly Available Demo

Second Annual Conference on Carbon Sequestration – May 2003

Amine Based CO_2 Capture from Gas Turbines

M. Simmonds [1], P. Hurst [1], M.B. Wilkinson [1], S. Reddy [2] & S. Khambaty [2]

[1] BP plc, UK Sunbury on Thames, Middlesex, TW16 7LN, UK
E-mail: **simmonm@bp.com**, Tel: +44 (0)1932 775641
[2] Fluor Enterprises Inc, USA 1 Fluor Daniel Drive, Aliso Viejo, California 92698, USA
E-mail: **satish.reddy@fluor.com**, Tel: +001 (949) 349 4959
E-mail: **shakir.khambaty@fluor.com**, Tel: +001 (949) 349 2252

Abstract

The CO_2 Capture Project (CCP) is a joint project of eight major energy companies, developing a wide range of CO_2 capture technologies to significantly reduce the cost of capturing and storing CO_2.

A feasibility study has been carried out to examine post combustion CO_2 capture from eleven simple cycle gas turbines using best available technology. The paper presents the findings of the study and examines the complex issues of retrofitting an amine-based capture facility onto existing gas processing plant located on the Alaskan North Slope.

The facility, designed to capture 1.78 million tonnes per year of CO_2, produces a high pressure, high purity product stream suitable for delivery to an enhanced oil recovery project. The paper highlights the benefits and the key challenges to be addressed by anyone considering post combustion capture of CO_2 at industrial scale and in an environmentally challenging location.

Introduction

The CO_2 Capture Project (CCP) is a joint project of eight major energy companies who are working to reduce significantly the cost of capturing and storing CO_2. The CCP is developing a wide range of CO_2 capture technologies, classified within the generic groupings of post-combustion, oxyfuel and pre-combustion decarbonisation.

A feasibility study has been carried out to examine post combustion CO_2 capture from eleven simple cycle gas turbines using today's best available technology. The study will provide one of four baselines, against which new and improved technologies for CO_2 capture will subsequently be assessed.

The CCP selected amine scrubbing as the best available technology for post combustion CO_2 capture and they requested Fluor to use their proprietary Econamine FG^{SM} process to produce a baseline process design and cost estimate. The Econamine FG^{SM} technology uses MEA scrubbing with chemical inhibitors to counter the effects of corrosion caused by oxygen in the flue gas. The process is well developed and has been widely used at a relatively smaller scale to produce high purity CO_2 for the food industry and feedstock for urea and methanol plants. However the process has never been implemented at a scale that is envisaged for the current feasibility study.

CO_2 Capture Scenario

The CO_2 capture scenario described in this paper is based upon BP's Central Gas Facility (CGF) at Prudhoe Bay, located on the North Slope in Alaska. The CGF employs a range of simple cycle gas turbines providing various mechanical shaft power duties including, for example, re-injection compression and process refrigeration. The objective of the study is to provide a detailed process design and cost estimate for a retrofit CO_2 capture facility, designed to capture the majority of the CO_2 currently emitted in the flue gases from eleven of these turbines, using a commercially proven amine based process (Fluor's Econamine FG^{SM}).

The resulting capture plant facility is designed to deliver 1.78 million tonnes per year of CO_2 (equivalent to around 5,200 tonnes per day), representing around 85% of the total CO_2 emitted by the selected turbines on an annual basis. The anticipated sink for the CO_2 is a potential Enhanced Oil Recovery (EOR) project on the North Slope, requiring the CO_2 from the turbine sources to be separated from other constituents in the flue gas prior to compression, to produce a high pressure, high purity CO_2 product stream.

The number and type of gas turbines selected for CO_2 capture in the study are listed below in Table 1.

Table 1: Number and Type of Gas Turbines selected for CO_2 Capture

Turbine Type	Number
General Electric Frame 6-1B	4
General Electric Frame 5-2B	3
Rolls Royce RB-211C	4

Process Overview

The CO_2 capture plant consists of the process equipment and supporting utility systems required to recover CO_2 from turbine flue gases. Figure 1 shows the main component blocks of the process schematically. The turbines utilise associated natural gas as a fuel, resulting in a flue gas that contains only dilute levels of CO_2 (around 3.3 mol% amongst other combustion products). The CO_2 is removed from the flue gases using Fluor's proprietary Econamine FG^{SM} solvent. It is subsequently regenerated from the solvent, dehydrated and compressed. The product is dry and of high purity and pressure (>99.9 vol. % CO_2 and 50 ppmv H_2O at 220 barg), suitable for use as a miscible injectant in a potential Enhanced Oil Recovery project.

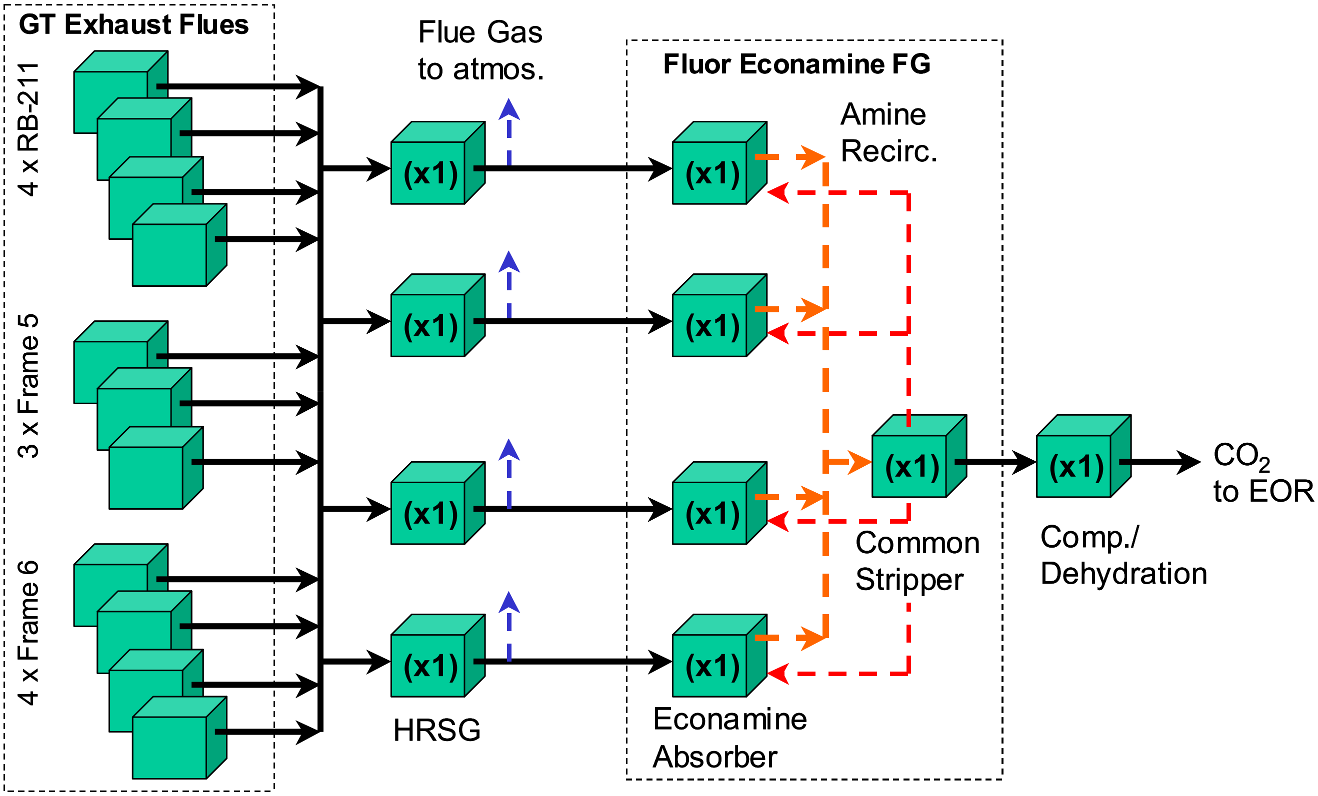
At the start of the study, it was quickly determined that the most cost effective configuration for the CO_2 capture plant involved maximising the size of the absorption trains. It was found that the size of each train was limited by the largest commercially available Heat Recovery Steam Generator (HRSG) and by the largest Absorber diameter that can be built with confidence. On this basis, the flue gases from the gas turbines are combined and fed to four identical absorption trains. Each absorption train contains a HRSG, a Direct Contact Cooler (DCC), a blower and an Econamine FG^{SM} absorber, complete with its associated heat exchangers, filters and pumps.

Regeneration of the Econamine FG^{SM} solvent takes place in a single stripping system, common to all four absorption trains. The stripping system includes a common stripper, a reclaimer, an amine filtration package, the associated heat exchangers and pumps and a solvent storage facility.

In its present configuration, the facility only utilises a small proportion of the heat energy available from simple cycle gas turbine exhaust (available at almost 500°C), by recovering heat through a single waste heat recovery unit connected to one of the Frame 5 turbines. The specification of new heat recovery steam generators to each proposed absorption train can satisfy two key process requirements, allowing essential pre-cooling of the turbine flue gases before they enter the absorption system whilst meeting the full energy demands of the capture plant.

Given the large scale of the proposed capture plant, the cost of providing virtually all of the other supporting utility systems have been included in the study, as existing infrastructure would be unable to accommodate the additional requirements of the new facilities.

Figure 1: Block diagram of the Gas Gathering, Capture and Product Treatment System



The facility is essentially self-sufficient in terms of energy and all utilities except treated seawater (used to generate boiler feed water for the HRSGs and general process water). During normal operation, the specification of a steam turbine generator allows the CO_2 plant to produce enough electricity to satisfy its own power demands, whilst exporting excess power to the local grid. It is envisaged that this excess power could be used to displace power generation elsewhere on the facility.

Process Description

The flue gases from several turbines are commingled before entering the HRSG, where the flue gases are cooled to a temperature that has a safe margin over its dewpoint. The heat available from the flue gas is used to raise three levels of steam, with each HRSG designed to recover around 140 MW of heat from the incoming flue gas. High pressure steam is used to generate electricity via a steam turbine power generator before subsequently being used as motive steam for the CO_2 product compressors. Intermediate pressure steam is used to provide heat to the stripper reboiler and the amine reclaiming system. Low pressure steam is used to provide deaeration for the boiler feed water. In addition to raising steam, a heating coil in the HRSG is used to further recover energy, for space heating of both new and existing modules at the CGF.

The partly cooled flue gases then flow to the DCC, where they are quenched. The DCC circulating water is cooled and filtered, providing a means of removing particulates that may be present in the flue gas stream. After leaving the DCC, the cooled flue gas is passed into a blower to maintain the required pressure in the inlet flue gas ducts and to ensure proper distribution of flue gases between the four absorption trains.

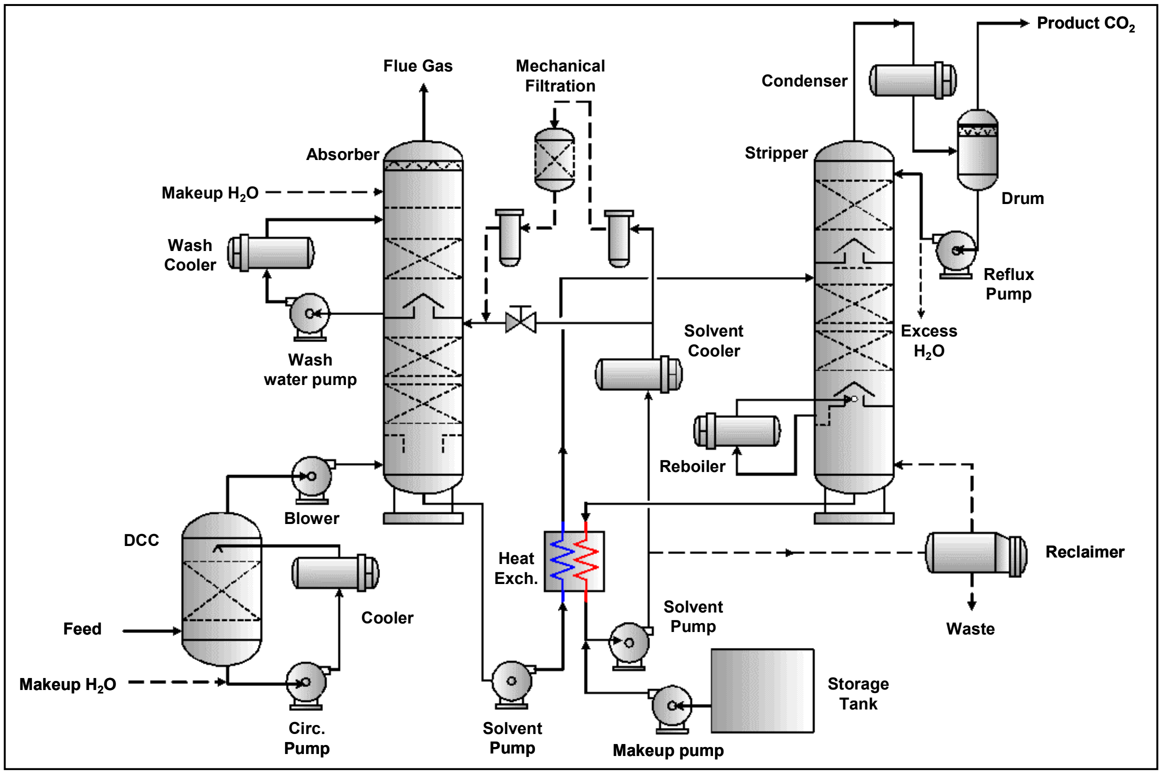
On exiting the blower, the flue gas enters the absorption column where it is counter-currently contacted with the Econamine FG^{SM} solution. CO_2 is absorbed from the flue gas stream as it passes up the column, before passing out through the top section of the absorber, which provides a wash for the gases before venting to atmosphere, ensuring that less than 1 ppmv of entrained solvent is emitted to atmosphere.

CO_2 is absorbed into the lean solvent, with the solvent circulation rate controlled by measuring the amount of CO_2 in the lean solvent feed to the absorber. The CO_2 rich solvent leaves the absorber and is sent for generation in the stripping system, where heat is provided to reversibly release the CO_2 from solution using four kettle type reboilers. A wash section at the top of the stripper ensures that a minimal amount of entrained and vaporized solvent leaves the column in the product CO_2 stream.

To maintain an efficient operation, a reclaimer is operated in parallel to the reboilers (as an intermittent batch process) to limit the build up of heat stable salts in the lean solvent, thereby reducing solvent losses over time. Additionally, a small percentage of the lean solvent returning to the absorbers is continuously filtered via a carbon bed to remove solids and other degradation products.

After leaving the top of the stripper, the CO_2 stream is compressed in a five stage centrifugal compressor, powered by a condensing steam turbine drive. Between the third and fourth stage, the CO_2 is dehydrated using a proprietary dehydration system, to ensure that the moisture content of the stream at the final export pressure of 220 bar(g) is sufficiently low for transportation via the export pipeline. Figure 2 shows the Econamine FG^{SM} process schematically.

Figure 2: Schematic Layout of the Econamine FG$^{\text{SM}}$ Process



Site Layout Considerations

The gas turbines at the Central Gas Facility are physically arranged in reasonably close proximity to one another, occupying an existing plot of around 400 m length by 200 m width. The harsh climatic environment of the Alaskan North Slope required a decision to be made early on in the study regarding the modularisation of the CO_2 capture equipment. Modularisation of the new plant provides three key benefits:

- An opportunity to minimise the high cost of installation labour in the field;
- The relative ease of transportation of process equipment to the site;
- A means for enclosing and winterising the process equipment for the ease of operation.

The construction strategy for the study is based on prefabricating the process and utility modules at Anchorage, Alaska and then transporting these modules to the Prudhoe Bay site via two sea-lifts. The four absorption trains and all supporting process equipment (stripping system, product compression, dehydration and utility plant) required in the capture facility are therefore strategically arranged in seven large sized modules with interconnecting pipe racks. The key limiting factors for each process module are:

- its overall size, dictated by both the available plot space and the maximum dimensions of the sea-lift barge;
- its weight, set by the sea-lift and road based transportation requirements.

Table 2 summarises the allocation of different process units to each module.

Table 2: Process Unit Allocation to Modules

Process Unit	Module Number
Heat Recovery Steam Generator (HRSG), Direct Contact Cooler (DCC), Blower, Absorber (Train 1)	1
HRSG, DCC, Blower, Absorber (Train 2)	2
HRSG, DCC, Blower, Absorber (Train 3)	3
HRSG, DCC, Blower, Absorber (Train 4)	4
Power Generation – Steam Turbine, CO_2 Compression & Dehydration, Plant Air, Instrument Air & Nitrogen System	5
Solvent Stripping & Reclaiming	6
Solvent Storage & Make-up, Glycol Circulation & Storage, Seawater Treatment & Waste Storage	7

In addition to the process modules, approximately eighteen pipe-way modules and sixty ductwork modules are required to connect the gas turbines and the process and utility systems together. The decision to modularise the capture plant, together with limitations imposed by the availability of transport routes to the Alaskan North Slope has a significant impact on the project schedule, estimated during the study to take a total of 57 months, with the first two trains being available for start-up after 45 months and an additional 12 months for the remaining 2 trains.

Utility Plants

Retrofitting a post combustion capture process of this scale requires the provision of significant utility systems to meet the needs of the CO_2 capture plant. Table 3 summarises the design capacities of the various utility systems required.

Table 3: Utility Plant Design Capacities

Utility	Design Capacity	Comments
HP Steam	602 Te/hr	Motive force for CO_2 compressor; Steam turbine for power generation
IP Steam	119 Te/hr	Stripper reboiler (combined with IP steam from Steam Turbine)
LP Steam	28 Te/hr	-
Cooling Medium	32,300 m^3/hr	-
Heating medium	2,310 m^3/hr	-
Sea water supply	125 m^3/hr	Compensating for steam blow down, solvent water balance and reverse osmosis unit waste
Demineralised Water	43 m^3/hr	-
Plant Air	643 Nm^3/hr	-
Instrument Air	965 Nm^3/hr	-
Nitrogen	80 Nm^3/hr	-

The steam turbine produces 69 MW(e) net of electrical power. After internal distribution of power to various process users on the CO_2 capture plant, a balance of 18 MW(e) is available for export to the local grid.

Environmental Emissions

There are a series of waste streams associated with the operation of the plant. Table 4 summarises these emissions.

The most notable emission from the CO_2 capture plant (apart from the balance of the treated flue gas components) is the reclaimer waste stream. The reclaimer produces around 5,000 tonnes per year of material, which is anticipated to require disposal. The reclaimer waste will contain a mixture of organic and inorganic compounds, typically including higher molecular weight nitrogen compounds, sodium salts and other metal salts.

The overall purpose of the capture plant is to reduce CO_2 emissions to atmosphere by capturing the CO_2 for subsequent storage. Whilst the amount of CO_2 targeted for capture equates to a total of 1.78 million tonnes of CO_2.per annum, the quantity of CO_2 emissions avoided is slightly higher than this, as the 18 MW(e) of electricity exported to the local grid from the steam turbine could be used to offset gas turbine generated power produced elsewhere on the facility.

If it is assumed that the exported power can displace essentially one Frame 5 gas turbine, a further 98,000 tonnes per annum of CO_2 emissions can be credited to the CO_2 capture plant. Hence, the total quantity of CO_2 avoided on an annual basis rises to 1.88 million tonnes per annum.

Table 4: Capture Plant Emissions

Type	Type of Emission	Average Rate	Frequency
Gas	Flue Gas from Absorbers (per train)	1,073,000 m^3/h	Continuous
Gas	Vent from Nitrogen generation package	322 Nm^3/h	Intermittent
Gas	Steam vent from intermittent blow down drum	Normally no flow	
Gas	Moisture vent from dehydration package	Small	Continuous
Liquid	Boiler drum blow down	15 m^3/h	Continuous
Liquid	Excess water from stripper reflux	Normally no flow	Intermittent
Liquid	Reject water from water treatment system	55 m^3/h	Continuous
Liquid	Reclaimer waste	100 metric tonnes/ week	Intermittent
Liquid	Filter backwash	Normally no flow	Intermittent
Solid	Spent carbon from amine filter package	63,500 kg	Once every 6 months
Solid	Disposable filter cartridges	Infrequent	Intermittent

Cost of Carbon Dioxide Capture

Tables 5 and 6 provide a break down of the total capital cost and annual operating costs associated with the proposed CO_2 capture plant.

Table 5: Capital Costs

Capture Plant Cost	Millions US$
Direct Field Cost - Process Modules (Off site)	705
Direct Field Cost - Process Modules (at North Slope)	252
Indirects	116
Home Office Costs	161
Other (start up support, owner's costs etc.)	149
Contingency (at 20%)	277
Total:	**1,660**

Table 6: Annual Operating Costs

Capture Plant Cost	Millions US$/yr
Chemicals	12.4
Maintenance (Materials & Labour)	24.9
Labour	2.2
Overheads	21.5
Insurance & Taxes	16.6
Total:	**77.7**

An indicative assessment places the likely cost of CO_2 capture from the facility (expressed as US$/tonne of CO_2 captured) at around $137/tonne (or $130/tonne of CO_2 avoided). These figures include elements of annualised capital and operating costs. The authors believe that the figures accurately reflect the potential costs of retrofitting capture technology at a location with a very harsh working environment. Some of the reasons why the cost of capture on the North Slope is so high are listed below:

- An execution strategy to cope with a two to three month annual construction window;
- A prolonged schedule of 57 months due to limited sea-lifts;
- A very high labor field cost;

- Construction strategy based on super modules weighing approximately 10,000 Te each;
- A very dilute feed gas with an average of concentration of only 3.3% CO_2 by volume;
- Multiple sources of low density gas (at 550°C) led to large collection ducts;
- A design for severe cold weather conditions mandated a costly glycol cooling system;
- Lack of fresh water resulted in an expensive water supply system that included an RO unit.

Had a less climatically challenging location been selected as the basis for this study, these costs would almost certainly be significantly reduced.

Conclusions

The present study has demonstrated that, despite the unprecedented scale, the post combustion capture of 5,200 Te/d of carbon dioxide from several gas turbines operating at BP's Central Gas Facility on the Alaskan North Slope is technically feasible, using today's best available technology.

The size of process equipment and the associated infrastructure required to support the capture plant would have a significant impact on the existing complex. The study has both highlighted and gone on to assess the cost impacts of a wide range of issues, some of which are generic to any post combustion capture scenario, but several of which are specific to the harsh environmental conditions presented by the Alaskan climate.

The capture plant produces a high purity, high pressure stream of 5,200 tonnes per day CO_2 (equivalent to 1.78 million tonnes per annum), suitable for use with an Enhanced Oil Recovery project or as an injectant stream for storage of CO_2 in a depleted reservoir or deep saline formation and in doing so, allows credit to be taken for an additional 98,000 tonnes per annum of CO_2 as a result of local electricity export.

Costs have been derived which, together with similar studies on other CO_2 capture scenarios (not reported here), will form the basis for comparison with future technology developments and will allow the participant companies of the CO_2 Capture Project to assess the cost reduction opportunities available from such developments.

NATIONAL ENERGY TECHNOLOGY LABORATORY

Assessment of Power Plants That Meet Proposed Greenhouse Gas Emission Performance Standards

November 5, 2009

DOE/NETL-401/110509

U.S. DEPARTMENT OF ENERGY

Disclaimer

ASSESSMENT OF POWER PLANTS THAT MEET PROPOSED GREENHOUSE GAS EMISSION PERFORMANCE STANDARDS

DOE/NETL-401/110509

FINAL REPORT

November 5, 2009

NETL Contact:

Eric Grol
General Engineer
Office of Systems, Analyses and Planning

National Energy Technology Laboratory

www.netl.doe.gov

Table of Contents

List of Exhibits

Prepared by:

Research and Development Solutions, LLC (RDS)

John L. Haslbeck
Parsons Corporation

Mark C. Woods
Parsons Corporation

Norma J. Kuehn
Parsons Corporation

Robert D. Brasington
Parsons Corporation

DOE Contract #DE-AC26-04NT41817

Acknowledgments

This report was prepared by **Research and Development Solutions, LLC (RDS)** for the United States Department of Energy's National Energy Technology Laboratory. This work was completed under DOE NETL Contract Number DE-AC-26-04NT41817. This work was performed under RDS Subtask 41817.401.01.01.008.

The authors wish to acknowledge the excellent guidance, contributions, and cooperation of the NETL staff, particularly:

Jared Ciferno & Eric Grol, NETL Technical Monitors

LIST OF ACRONYMS AND ABBREVIATIONS

AACE	Association for the Advancement of Cost Engineering
AEO	Annual Energy Outlook
AGR	Acid gas removal
ANSI	American National Standards Institute
ASU	Air separation unit
BACT	Best available control technology
BART	Best available retrofit technology
BEC	Bare erected cost
BFD	Block flow diagram
Btu	British thermal unit
Btu/h	British thermal unit per hour
Btu/kWh	British thermal unit per kilowatt hour
Btu/lb	British thermal unit per pound
Btu/scf	British thermal unit per standard cubic foot
CAAA	Clean Air Act Amendments of 1990
CAIR	Clean Air Interstate Rule
CAMR	Clean Air Mercury Rule
CCF	Capital Charge Factor
CDR	Carbon Dioxide Recovery
CF	Capacity factor
CFM	Cubic feet per minute
CFR	Code of Federal Regulations
cm	Centimeter
CO_2	Carbon dioxide
COE	Cost of electricity
COR	Contracting Officer's Representative
COS	Carbonyl sulfide
CRT	Cathode ray tube
CS	Carbon steel
CT	Combustion turbine
CTG	Combustion Turbine-Generator
CWT	Cold water temperature
dB	Decibel
DCS	Distributed control system
DI	De-ionized
Dia.	Diameter
DLN	Dry low NOx
DOE	Department of Energy
EAF	Equivalent availability factor
EIA	Energy Information Administration
EPA	Environmental Protection Agency

EPC	Engineer/Procure/Construct
EPRI	Electric Power Research Institute
EPCM	Engineering/Procurement/Construction Management
FERC	Federal Energy Regulatory Commission
FOAK	First of a kind
FRP	Fiberglass-reinforced plastic
ft	Foot, Feet
ft, w.g.	Feet of water gauge
GADS	Generating Availability Data System
gal	Gallon
GDP	Gross domestic product
gpm	Gallons per minute
GT	Gas turbine
h	Hour
H_2	Hydrogen
Hg	Mercury
HDPE	High density polyethylene
HHV	Higher heating value
hp	Horsepower
HP	High pressure
HRSG	Heat recovery steam generator
HVAC	Heating, ventilating, and air conditioning
HWT	Hot water temperature
Hz	Hertz
ICR	Information Collection Request
IEEE	Institute of Electrical and Electronics Engineers
IGCC	Integrated gasification combined cycle
IGVs	Inlet guide vanes
In. H_2O	Inches water
In. Hga	Inches mercury (absolute pressure)
In. W.C.	Inches water column
IOU	Investor-owned utility
IP	Intermediate pressure
IPM	Integrated Planning Model
IPP	Independent power producer
ISO	International Standards Organization
kg/GJ	Kilogram per gigajoule
kg/h	Kilogram per hour
kJ	Kilojoules
kJ/h	Kilojoules per hour
kJ/kg	Kilojoules per kilogram
KO	Knockout
kPa	Kilopascal absolute

kV	Kilovolt
kW	Kilowatt
kWe	Kilowatts electric
kWh	Kilowatt-hour
LAER	Lowest Achievable Emission Rate
lb	Pound
lb/h	Pounds per hour
lb/ft^2	Pounds per square foot
lb/MMBtu	Pounds per million British thermal units
lb/MWh	Pounds per megawatt hour
lb/TBtu	Pounds per trillion British thermal units
LCOE	Levelized cost of electricity
LF_{Fn}	Levelization factor for category n fixed operating cost
LF_{Vn}	Levelization factor for category n variable operating cost
LHV	Lower heating value
LNB	Low NOx burner
LP	Low pressure
lpm	Liters per minute
m	Meters
m/min	Meters per minute
m^3/min	Cubic meter per minute
MAF	Moisture and Ash Free
MCR	Maximum continuous rate
MDEA	Methyldiethanolamine
MHz	Megahertz
MJ/Nm^3	Megajoule per normal cubic meter
MMBtu	Million British thermal units (also shown as 10^6 Btu)
MMBtu/h	Million British thermal units (also shown as 10^6 Btu) per hour
MMkJ	Million kilojoules (also shown as 10^6 kJ)
MMkJ/h	Million kilojoules (also shown as 10^6 kJ) per hour
MNQC	Multi Nozzle Quiet Combustor
MPa	Megapascals
MVA	Mega volt-amps
MWe	Megawatts electric
MWh	Megawatt-hour
MWt	Megawatts thermal
N/A	Not applicable
NAAQS	National Ambient Air Quality Standards
NEMA	National Electrical Manufacturers Association
NERC	North American Electric Reliability Council
NETL	National Energy Technology Laboratory
NFPA	National Fire Protection Association
Nm^3	Normal cubic meter

NOx	Oxides of nitrogen
NSPS	New Source Performance Standards
NSR	New Source Review
O&M	Operation and maintenance
OC_{Fn}	Category n fixed operating cost for the initial year of operation
OC_{Vnq}	Category n variable operating cost for the initial year of operation
OD	Outside diameter
OP/VWO	Over pressure/valve wide open
OSHA	Occupational Safety and Health Administration
PF	Power Factor
PM	Particulate matter
PM_{10}	Particulate matter measuring 10 µm or less
POTW	Publicly Owned Treatment Works
ppm	Parts per million
ppmv	Parts per million volume
ppmvd	Parts per million volume, dry
PRB	Powder River Basin coal region
PSA	Pressure Swing Adsorption
PSD	Prevention of Significant Deterioration
psia	Pounds per square inch absolute
psid	Pounds per square inch differential
psig	Pounds per square inch gage
PTFE	Teflon (Polytetrafluoroethylene)
Qty	Quantity
RDS	Research and Development Solutions, LLC
RH	Reheater
scfh	Standard cubic feet per hour
scfm	Standard cubic feet per minute
Sch.	Schedule
scmh	Standard cubic meter per hour
SCOT	Shell Claus Off-gas Treating
SG	Specific gravity
SGC	Synthesis gas cooler
SGS	Sour gas shift
SO_2	Sulfur dioxide
SOx	Oxides of sulfur
SRU	Sulfur recovery unit
SS	Stainless steel
STG	Steam turbine generator
TCR	Total capital requirement
TEWAC	Totally Enclosed Water-to-Air Cooled
TGTU	Tail gas treating unit
Tonne	Metric Ton (1000 kg)

TPC	Total plant cost
TPD	Tons per day
TPH	Tons per hour
TPI	Total plant investment
TS&M	Transport, storage and monitoring
V-L	Vapor Liquid portion of stream (excluding solids)
vol%	Volume percent
WB	Wet bulb
wg	Water gauge
wt%	Weight percent
$/MMBtu	Dollars per million British thermal units
$/MMkJ	Dollars per million kilojoule

This page intentionally left blank

EXECUTIVE SUMMARY

Revision 2 Updates

The technologies modeled in this study, namely integrated gasification combined cycle, subcritical pulverized coal and supercritical pulverized coal, are the subject of other ongoing systems analysis studies at the Department of Energy's National Energy Technology Laboratory. Vendor discussions that occurred as part of the other studies led to improved technology information that was incorporated into the Aspen models for this study. The updated models led to revised performance estimates, which were then used to update the cost estimates. The reference costs used for this study were also updated through efforts on other studies, and the most recent costs have been incorporated. In addition, owner's costs were added to the Total Plant Cost previously reported, and the capital component of levelized cost of electricity now includes owner's costs. The updated results are presented in the current revision (revision 2) of this report. Details of the modeling updates and cost methodology changes are included in the body of the report and are identified by italicized font.

Greenhouse gas (GHG) emissions continue to receive increased scrutiny because of their perceived relation to global warming. Numerous bills have been introduced in both the United States Senate and House of Representatives that would limit GHG emissions. The bills vary primarily in the economy sectors regulated, the extent of GHG reductions and the compliance year, but all represent reductions from the "business-as-usual" scenario. In June, 2009 the House of Representatives passed the American Clean Energy and Security Act of 2009 (H.R. 2454) which would limit GHG emissions starting as soon as 2012. The Senate has not yet acted on the House bill or any companion bill, but deliberations are ongoing. Adding to the legislative momentum for carbon regulation, in September, 2009 the Environmental Protection Agency proposed a rule that would limit future regulation of GHG emissions under the Clean Air Act to industrial facilities that emit 25,000 tons or more of carbon dioxide annually. The proposed rule would impact facilities such as power plants, refineries, and factories, which produce nearly 70 percent of domestic GHGs.

In addition to proposed Federal regulations, various states have proposed or enacted legislation to reduce GHG emissions. The most imminent regulations were enacted by the state of California and would limit GHG emissions from in-state energy producers or out-of-state producers supplying electricity to California to 1,100 lb CO_2/net-MWh [1]. A sampling of the legislation is provided in Section 1.1.

The objective of this report is to present the baseline cost and performance of greenfield integrated gasification combined cycle (IGCC) plants, greenfield supercritical (SC) pulverized coal (PC) plants, and retrofit subcritical PC plants that limit carbon dioxide (CO_2) emissions to the California standard of 1,100 lb CO_2/net-MWh and that achieve 90 percent CO_2 capture. For each plant type, three cases were modeled:

- Baseline performance with no CO_2 capture
- CO_2 emissions reduced to 1,100 lb CO_2/net-MWh

- CO_2 emissions reduced by 90 percent

The subcritical PC retrofit case was based on a generic plant site, but is representative of a western plant that could supply electricity to California and hence have to meet the 1,100 lb CO_2/net-MWh standard. The elevation used was 6,700 ft, which is the average elevation of Wyoming. For consistency between cases, this same elevation was used for all technologies. The fuel used in all nine cases was representative of a coal from the Powder River Basin (PRB) and has the same composition as the subbituminous coal used in an as yet unpublished NETL study entitled "Cost and Performance Baseline for Low-Rank Coal Fossil Energy Plants." The nine cases are summarized in Exhibit 1-1.

The cost and performance of the various fossil fuel-based technologies will most likely determine which combination of technologies will be utilized to meet the demands of the power market. Selection of new generation technologies will depend on many factors, including:

- Capital and operating costs
- Overall energy efficiency
- Fuel prices
- Cost of electricity (COE)
- Availability, reliability and environmental performance
- Current and potential regulation of air, water, and solid waste discharges from fossil-fueled power plants
- Market penetration of clean coal technologies that have matured and improved as a result of recent commercial-scale demonstrations under the Department of Energy's (DOE's) Clean Coal Programs

Nine power plant configurations were analyzed as listed in Exhibit 1-1. The list includes three IGCC cases utilizing Shell gasifiers each with and without CO_2 capture; six PC cases, three greenfield supercritical and three existing subcritical plants, each with and without CO_2 capture.

The methodology used information provided by the technology vendors (IGCC) and conventional models and existing plant information (PC) to perform steady-state simulations of the technology using the Aspen Plus (Aspen) modeling program. The resulting mass and energy balance results from the Aspen model were used to size major pieces of equipment. These equipment sizes formed the basis for cost estimating. Costs were scaled from estimates provided previously on similar technologies using PRB coal. The original estimates were developed through a combination of vendor quotes and scaled estimates from previous design/build projects. Performance and process limits were based upon published reports, information obtained from vendors and users of the technology, and cost and performance data from design/build utility projects. Baseline fuel costs for this analysis were determined using data from the Energy Information Administration's (EIA) Annual Energy Outlook (AEO) 2007. The first year (2015) cost used is \$0.57/GJ (\$0.61/MMBtu) for coal (Montana Rosebud Powder River Basin) on a higher heating value (HHV) basis and in 2007 U.S. dollars.

All plant configurations were evaluated based on installation at a greenfield site, with the exception of the existing subcritical PC plant. Typically, greenfield plants are state-of-the-art plants with higher efficiencies than the existing average power plant population. Consequently, these plants would be expected to be near the top of the dispatch list, and the study capacity factor is chosen to reflect the maximum availability demonstrated for the specific plant type, i.e. 80 percent for IGCC and 85 percent for PC. A capacity factor of 85 percent was also used for the subcritical PC case to be consistent with the SC PC greenfield plant.

PERFORMANCE

Plant Output

The performance results are presented in Exhibit ES-1 and Exhibit ES-2. The net power output varies between technologies because the combustion turbines in the IGCC cases are manufactured in discrete sizes, but the boilers and steam turbines in the greenfield PC cases are readily available in a wide range of capacities. The net output in the subcritical retrofit PC plant is limited by the capacity of the existing boiler and steam turbine. The result is that all of the greenfield supercritical PC cases have a net output of 550 MW, the subcritical retrofit cases have net outputs ranging from 532 to 359 MW, and the IGCC cases have net outputs ranging from 502 to 401 MW.

The range in IGCC net output is caused by the increased elevation, the much higher auxiliary load imposed in the CO_2 capture cases primarily due to CO_2 compression, and the need for extraction steam in the water-gas shift reactions, which reduces steam turbine output. Higher auxiliary load and extraction steam requirements can be accommodated in the greenfield supercritical PC cases (larger boiler and steam turbine) but not in the IGCC or subcritical retrofit PC cases. For the IGCC cases or subcritical retrofit PC cases, it is impossible to maintain a constant net output from the steam cycle given the fixed input (combustion turbine for IGCC and existing boiler capacity for subcritical retrofit cases). In addition, the combustion turbine output increases with increasing capture levels because of the higher flue gas moisture content due to the higher hydrogen content of the syngas, while the mass flow remains relatively the same.

Energy Efficiency

The definition of Energy Penalty used in this study to evaluate the impact of energy losses due to the addition of CO_2 capture controls is the difference in net power plant efficiency expressed in absolute percentage points as shown in the following equation.

$$\text{Energy Penalty} = (\text{Net Power Plant Efficiency})_{\textit{no capture}} - (\text{Net Power Plant Efficiency})_{\textit{with capture}}$$

The net plant efficiency (HHV basis) for all 9 cases is shown in Exhibit ES-3.

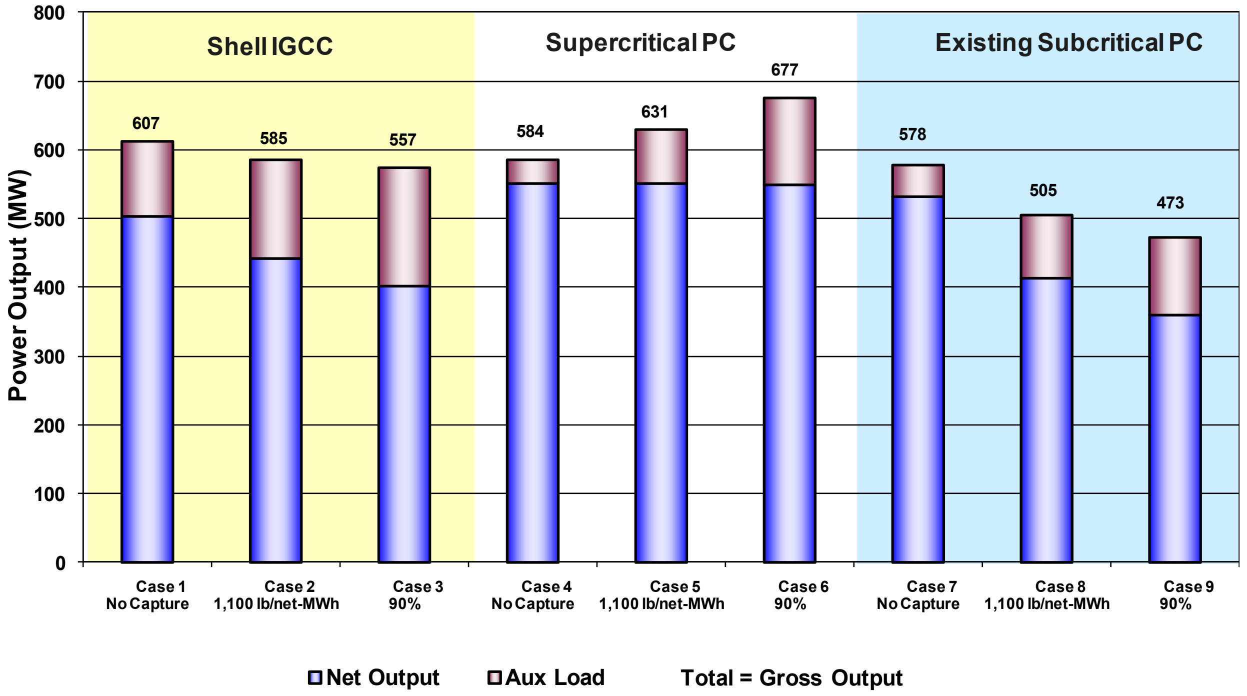
Exhibit ES-1 Performance Summary

		Shell IGCC			Supercritical PC			Existing Subcritical PC		
		Case 1 **No Capture**	**Case 2** **1,100 lb/MWh$_{net}$**	**Case 3** **90%**	**Case 4** **No Capture**	**Case 5** **1,100 lb/MWh$_{net}$**	**Case 6** **90%**	**Case 7** **No Capture**	**Case 8** **1,100 lb/MWh$_{net}$**	**Case 9** **90%**
Output										
Syngas Fuel Flow	(lbm/h)	709,921	425,936	216,381						
Combustion Turbine Generator	(kW)	372,500	377,000	380,600						
Main Steam Flow	(lbm/h)	1,290,732	919,549	942,596	3,640,595	4,366,633	5,136,290	3,924,635	3,924,075	3,925,353
Steam Turbine Generator	(kW)	240,400	208,000	192,900	585,300	629,800	675,500	577,800	476,800	432,000
Let Down Turbine Generator	(kW)					0	0		28,100	40,600
Total Gross Power	(kW)	612,900	585,000	573,500	585,300	629,800	675,500	577,800	504,900	472,600
Auxiliary Power Summary										
Base Plant Power	(kW)	28,360	28,760	30,040	32,750	40,920	49,880	43,010	48,590	51,120
Air Separation Unit	(kW)	81,290	90,830	98,160						
Flue Gas Cleanup	(kW)	880	9,520	18,200	2,540	14,510	26,550	2,760	15,570	21,310
CO_2 Compression	(kW)		13,130	25,960		24,340	49,170		28,200	40,780
Total Auxiliary Power	(kW)	110,530	142,240	172,360	35,290	79,770	125,600	45,770	92,360	113,210
Net Plant Output	(kW)	502,370	442,760	401,140	550,010	550,030	549,900	532,030	412,540	359,390
Boiler Efficiency (HHV)[1]	(fraction)				0.86	0.86	0.86	0.83	0.83	0.83
Coal Feed Rate	(lbm/h)	478,697	495,356	517,357	568,411	691,955	814,119	650,360	650,355	650,360
Coal Heat Input (HHV)	(10^6 Btu/h)	410	424	443	487	593	697	557	557	557
CO_2 Capture Efficiency	%	0	46	90	0	53	90	0	62	90
Net Plant Heat Rate (HHV)	(Btu/kWh)	8,160	9,581	11,045	8,851	10,774	12,679	10,469	13,501	15,498
Net Plant Thermal Efficeincy (HHV)	(%)	41.8	35.6	30.9	38.6	31.7	26.9	32.6	25.3	22.0
Energy Penalty[2]	(% Net Points)	-	6.2	10.9	-	6.9	11.7	-	7.3	10.6

[1] Boiler Heat Output/ ($C_{\text{oal-HHV}}$)

[2] Percentage points decrease in efficiency due to CO_2 capture

Exhibit ES-2 Power Output Summary

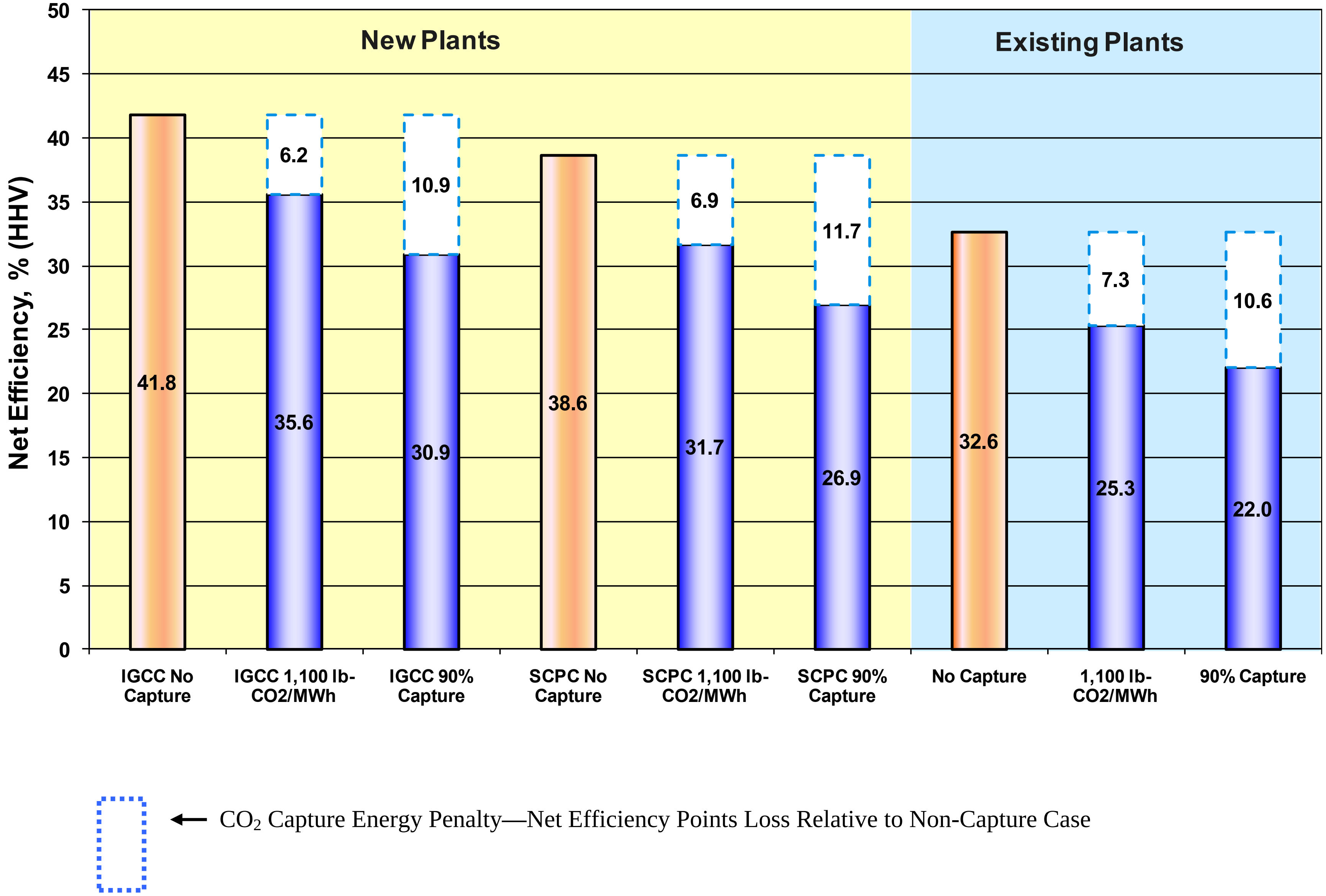


Performance Highlights:

- The IGCC no-capture case has the highest net efficiency of the technologies modeled in this study with an efficiency of 41.8 percent. The energy penalty for the 1,100 lb/net-MWh CO_2 emission level is smallest for IGCC and highest for the retrofit subcritical PC case.
- The new SCPC no-capture case has a lower net efficiency compared to the new IGCC no-capture case, and the CO_2 Capture Energy Penalty for the SCPC cases is higher at 6.9 and 11.7 net efficiency points lost relative to the SCPC no-capture case for the 1,100 lb CO_2/net-MWh case and 90 percent capture case, respectively, *resulting in lower net power plant efficiencies compared to the IGCC power plants equipped with CO_2 capture.*
- The estimated efficiency of the existing subcritical PC using Montana Rosebud PRB coal is 32.6 percent. There is a 7.3 percent penalty to achieve the 1,100 lb/net-MWh CO_2 emission limit and 10.6 percent penalty for the 90 percent capture case. The retrofit cases have the lowest efficiency for each of the three cases, but the smallest energy penalty for the 90 percent capture case.

Exhibit ES-3 Net Plant Efficiency (HHV Basis)

Subbituminous PRB Coal at 6,700 feet elevation



Water Use

Three water values are presented for each case in Exhibit ES-4 and Exhibit ES-5: raw water withdrawal, process water discharge and raw water consumption. Each of these values is normalized by net plant output for Exhibit ES-4.

Exhibit ES-5 shows absolute water withdrawal and consumption. Water demand represents the total amount of water required for a particular process. Some water is recovered within the process, primarily as coal moisture from the drying process and syngas condensate (IGCC cases) or flue gas condensate (PC cases), and that water is re-used as internal recycle. Raw water withdrawal is the difference between water demand and internal recycle. Some water is returned to the source, namely sour water stripper blowdown (IGCC cases) and cooling tower blowdown (IGCC and PC cases). The difference between raw water withdrawal and water returned to the source (process discharge) is raw water consumption, which represents the net impact on the water source.

The largest consumer of water in the plant is the makeup to the cooling system. The greenfield plants (IGCC and supercritical PC) use parallel wet (50 percent)/dry (50 percent) cooling. The existing subcritical PC uses only an evaporative cooling tower. The difference in cooling systems has a significant impact on water consumption.

Water Usage Highlights:

- In all cases the primary water consumer is cooling tower makeup, which ranges from 60 to 94 percent of the total raw water consumption.
- For the non-capture cases, IGCC has the least amount of raw water withdrawal and consumption, followed by the new SCPC and existing subcritical PC. The relative total raw water consumption is 4.0: 1.6: 1.0 normalized by net power output (gpm/MWnet) (subcritical PC: SCPC: IGCC). The relative results are as expected due to the cooling systems employed and the higher steam turbine output in the PC cases. These factors combined result in higher condenser duties, higher cooling water requirements and ultimately higher cooling water makeup.
- Among the CO_2 capture cases, raw water consumption increases (relative to non-capture cases) much more dramatically for the PC than for IGCC cases because of the large cooling water demand of the CO_2 scrubbing process which results in much greater cooling water makeup requirements. Comparing the 1,100 lb CO_2/net-MWh emission limit cases, the relative raw water consumption is 3.6 : 1.7 : 1.0 normalized by net power output (subcritical PC: SCPC: IGCC)—meaning that the SCPC plant has a net consumption that is 1.7 times greater than the IGCC plant at the same capture rate. The relative raw water consumption comparison for the 90 percent capture cases follows the same trend.
- CO_2 capture increases the absolute raw water consumption of the subcritical PC retrofit cases by the least amount of the technologies evaluated: a maximum of 30 percent at the 90 percent capture level. The primary reason for the small increase is the reduction in condenser duty, which mostly offsets the increase in duty due to the Econamine process. The consequence is a significant reduction in net plant power output. With the addition of CO_2 capture, supercritical PC raw water consumption increases by 208 percent and IGCC by 106 percent. The substantial increase in the SC PC case is driven by the high Econamine water requirement and the increase in size of the steam turbine to maintain a constant net output. Hence the steam turbine condenser duty remains high and the Econamine cooling load is simply additive.

Exhibit ES-4 Normalized Water Withdrawal and Consumption

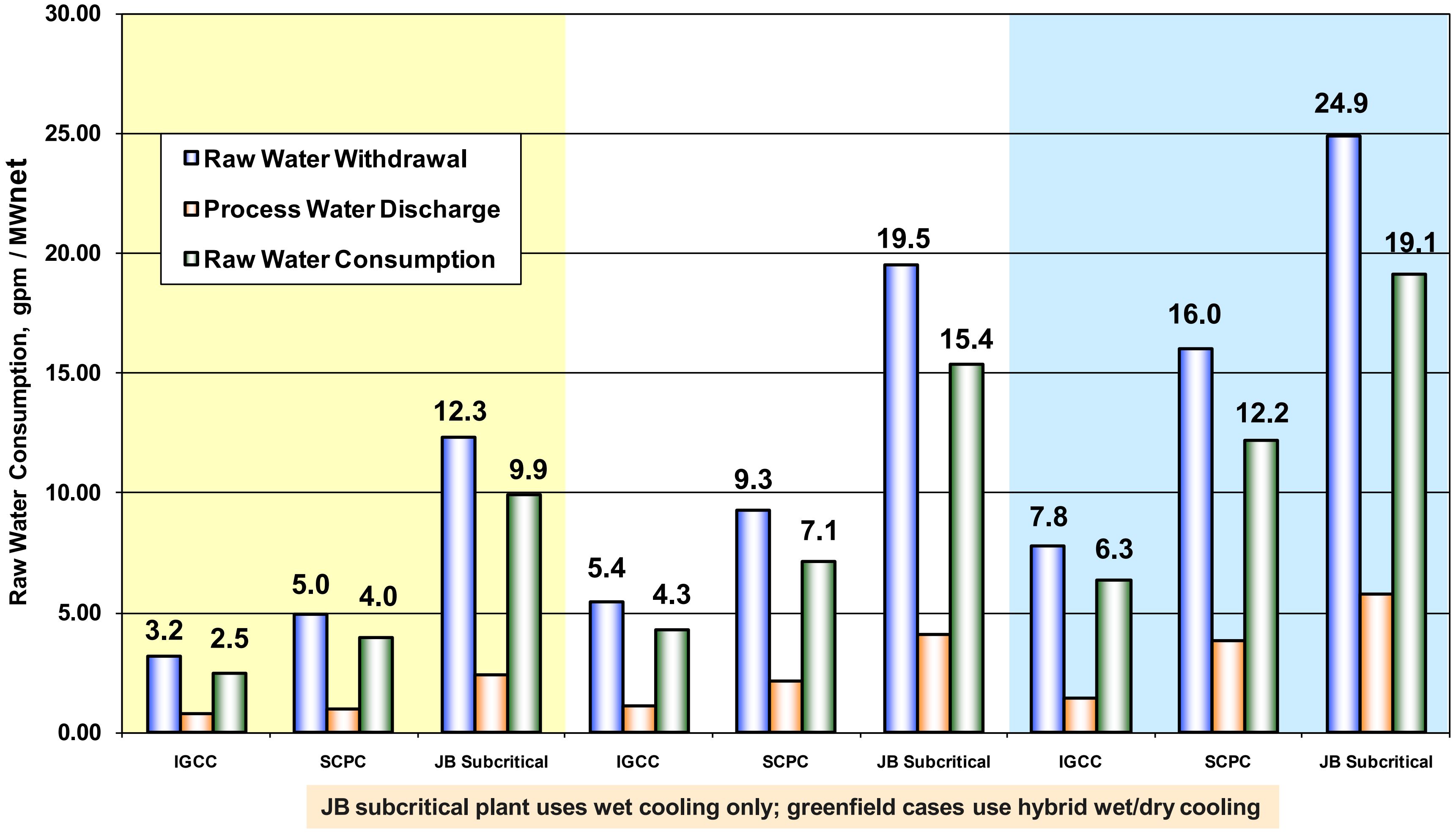
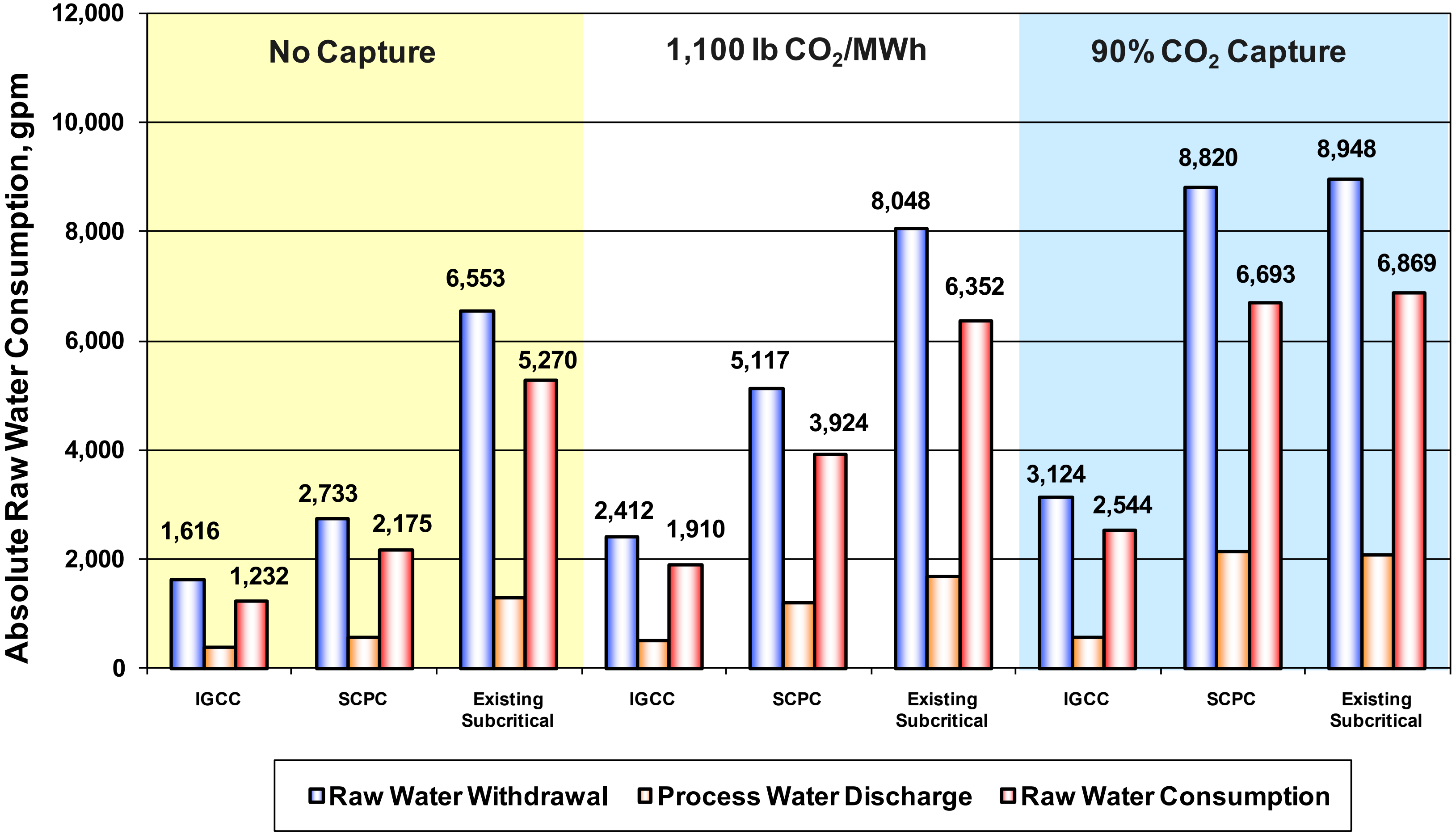


Exhibit ES-5 Absolute Water Demand and Usage



COST RESULTS

Total Plant Cost

The total plant cost (TPC) for each technology was determined through a combination of vendor quotes and scaled estimates from previous design/build projects. For the existing subcritical retrofit PC plant, the cost was determined through the use of a Best Available Retrofit Technology (BART) analysis and scaled estimates. TPC includes all equipment, materials, labor (direct and indirect), engineering and construction management, and contingencies (process and project). *Owner's costs, including preproduction costs, inventory capital, initial cost for catalyst and chemicals, land, financing costs and other owner's costs were added to TPC to generate total overnight cost (TOC). Property taxes and insurance were included in the fixed operating costs as an additional owner's cost. TOC was used to calculate the capital component of the levelized cost of electricity (LCOE). A factor was applied to TOC to convert to total as spent cost (TASC), which includes interest and escalation during the construction period. The inclusion of owner's costs increases the TPC by about 18 percent for each of the cases modeled.*

The cost estimates carry an accuracy of ±30 percent, consistent with the screening study level of design engineering applied to the various cases in this study. The value of the study lies not in the absolute accuracy of the individual case results but in the fact that all cases were evaluated under the same set of technical and economic assumptions. This consistency of approach allows meaningful comparisons among the cases evaluated.

Project contingencies were added to the Engineering/Procurement/Construction Management (EPCM) capital accounts to cover project uncertainty and the cost of any additional equipment that would result from a detailed design. The contingencies represent costs that are expected to occur. Each bare erected cost (BEC) account was evaluated against the level of estimate detail and field experience to determine project contingency. Process contingency was added to cost account items that were deemed to be first-of-a-kind or posed significant risk due to lack of operating experience. The cost accounts that received a process contingency include:

- Gasifiers and Syngas Coolers – 15 percent on all IGCC cases – next-generation commercial offering and integration with the power island.
- Two Stage Selexol – 20 percent on all IGCC capture cases – lack of operating experience at commercial scale in IGCC service.
- Mercury Removal – 5 percent on all IGCC cases – minimal commercial scale experience in IGCC applications.
- CO_2 Removal System – 20 percent on all PC capture cases - post-combustion process unproven at commercial scale for power plant applications.
- Combustion Turbine Generator – 5 percent on all IGCC non-capture cases – syngas firing and ASU integration; 10 percent on all IGCC capture cases – high hydrogen firing.
- Instrumentation and Controls – 5 percent on all IGCC accounts and 5 percent on the PC capture cases – integration issues.

The TPC, TOC and TASC for the nine power plant configurations are shown in Exhibit ES-6 in June 2007 dollars. The normalized TOC for each technology is shown in Exhibit ES-7.

Exhibit ES-6 Plant Costs

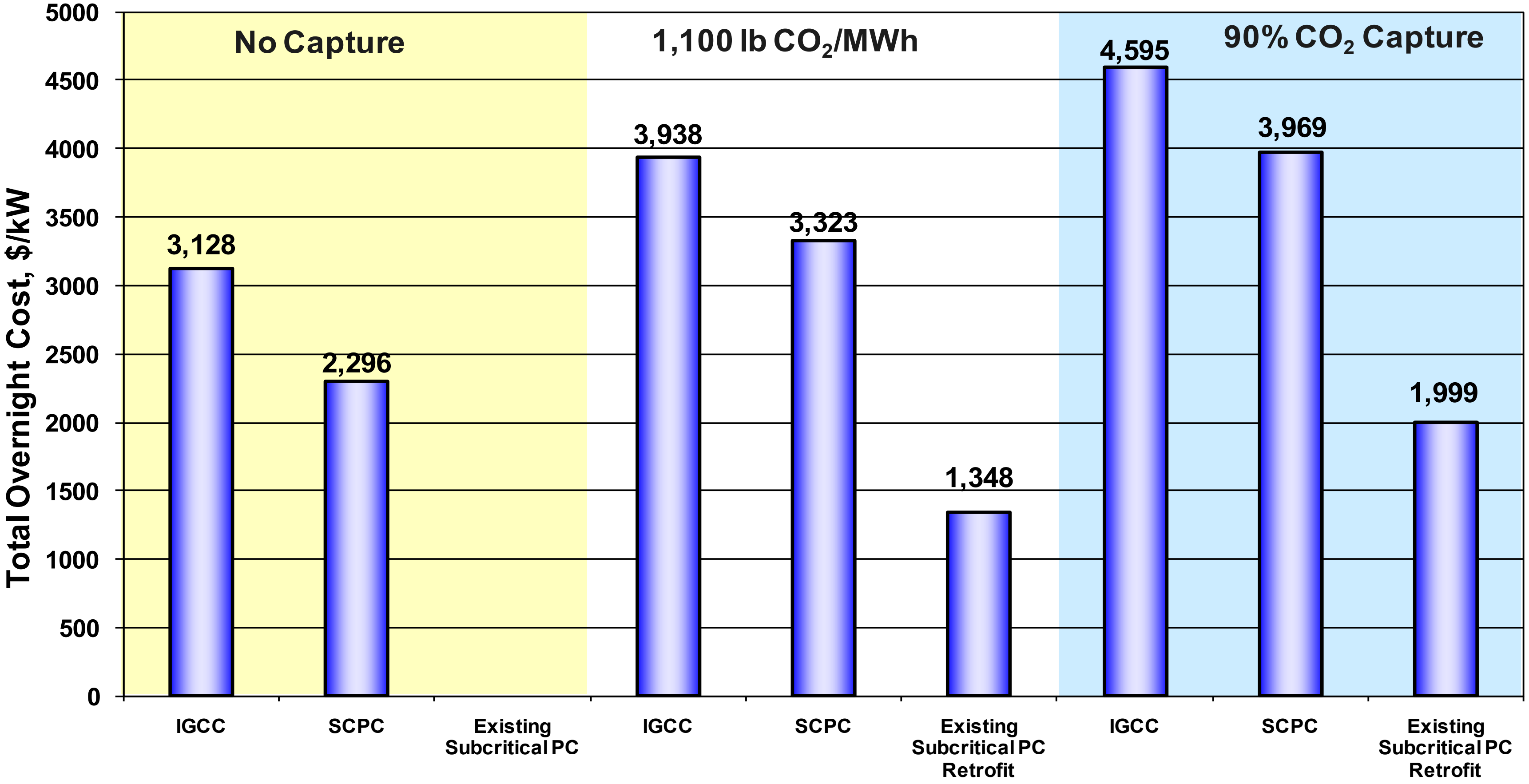
Study Case	Net Plant Output	Total Plant Cost (TPC)		Total Over Night Cost (TOC)		Total As-Spent Cost (TASC)[1]	
	kW	1000$	$\$/kW_{net}$	1000$	$\$/kW_{net}$	1000$	$\$/kW_{net}$
Case 1, IGCC w/o CO_2 Capture	502,370	1,290,415	2,569	1,571,409	3,128	1,791,407	3,566
Case 2, IGCC w/ 1,100 lb/MWh_{net} CO_2 Capture	442,760	1,432,055	3,234	1,743,413	3,938	1,987,490	4,489
Case 3, IGCC w/ 90% CO_2 Capture	401,140	1,513,013	3,772	1,843,305	4,595	2,101,368	5,238
Case 4, SCPC w/o CO_2 Capture	550,010	1,036,345	1,884	1,262,625	2,296	1,357,322	2,468
Case 5, SCPC w/ 1,100 lb/MWh_{net} CO_2 Capture	550,030	1,501,061	2,729	1,827,611	3,323	1,970,165	3,582
Case 6, SCPC w/ 90% CO_2 Capture	549,900	1,792,301	3,259	2,182,729	3,969	2,352,982	4,279
Case 7, Existing Subcritical PC w/o CO_2 Capture	532,030	0	0	0	0	0	0
Case 8, Existing Subcritical PC Retrofit w/ 1,100 lb/MWh_{net} CO_2 Capture	412,540	458,271	1,111	555,992	1,348	599,360	1,453
Case 9, Existing Subcritical PC Retrofit w/ 90% CO_2 Capture	359,390	591,983	1,647	718,587	1,999	774,637	2,155

[1] Construction duration for the greenfield cases is 5 years and for the retrofit cases is 3 years.

Capital Cost Highlights:

- Retrofitting the existing subcritical PC plant with CO_2 capture designed to meet a CO_2 emissions limit of 1,100 lb/net-MWh has a TOC of \$1,348/kW$_e$ and one designed for 90% CO_2 capture has a TOC of \$1,999/kW$_e$. The retrofit cost for an existing plant is 59 percent lower than a greenfield application at an emissions limit of 1,100 lb CO_2/net-MWh and 50 percent less at 90 percent CO_2 capture primarily because of the assumption that the subcritical PC plant is paid for and the cost of capital pertains only to the carbon capture and sequestration portion of the plant, while in the greenfield cases the cost of capital pertains to the total plant cost.
- If New Source Review (NSR) is triggered, the existing subcritical PC retrofit plant could require SCR. The retrofit with SCR designed to meet a CO_2 emissions limit of 1,100 lb/net-MWh and 90% CO_2 capture will have a TOC of \$1,717/ kW$_e$ and \$2,430/ kW$_e$, respectively. Even with SCR, the TOC of the retrofit cases is less than greenfield plants with CO_2 capture for equivalent emissions reduction primarily because of the assumption that the original subcritical PC plant capital cost has been paid.
- Comparing the greenfield IGCC and PC cases without CO_2 capture, the supercritical PC has the lowest TOC at \$2,296/kW$_e$ followed by IGCC with a cost of \$3,128/kW$_e$. The IGCC cost is 36 percent greater than the supercritical PC cost.
- Comparing the greenfield IGCC and PC cases with CO_2 capture, the supercritical PC has the lowest TOC at \$3,323/kW$_e$ at an emissions level of 1,100 lb CO_2/net-MWh and \$3,969/kW$_e$ at a capture level of 90 percent. The corresponding IGCC cases have a TOC of \$3,938/kW$_e$ and \$4,595/kW$_e$.
- A new IGCC power plant built with CO_2 capture designed to meet a CO_2 emissions limit of 1,100 lb/net-MWh will add \$810/kW$_e$ in incremental capital cost. The same IGCC power plant designed for 90% CO_2 capture will have an incremental cost of \$1,467/kW$_e$.
- A new SCPC power plant built with CO_2 capture designed to meet a CO_2 emissions limit of 1,100 lb/net-MWh will add \$1,027/kW$_e$ in incremental capital cost and one designed for 90% CO_2 capture will have an incremental cost of \$1,673/kW$_e$.

Exhibit ES-7 Total Overnight Cost, 2007 Dollars



Note: Compared to the Greenfield IGCC and SCPC power plants, the capital cost for the subcritical PC no capture case is zero (assumes the original plant costs are paid). The capital costs for the new IGCC and PC power plants are total (base plant + capture); whereas the capital cost for the subcritical PC retrofit is for the CO_2 capture process only.

Levelized Cost of Electricity (LCOE)

The current dollar, 30-year LCOE was calculated for each case using the economic parameters shown in Exhibit ES-8. The cases were divided into three categories, all undertaken at an investor owned utility: high risk projects with a five year construction duration (all IGCC cases and greenfield SC PC capture cases); low risk projects with a five year construction duration (greenfield SC PC non-capture case); and high risk projects with a three year duration (retrofit PC capture cases). High risk projects are those in which commercial scale operating experience is limited. The IGCC cases (with and without CO_2 capture) and the PC cases with CO_2 capture were considered to be high risk. The non-capture PC case was considered to be low risk.

Exhibit ES-8 Economic Parameters Used to Calculate LCOE

	High Risk (5 year construction period)	Low Risk (5 year construction period)	High Risk (3 year construction period)
Capital Charge Factor	0.1773	0.1691	0.1567
General Levelization Factor	1.443	1.4299	1.4101

The LCOE results are shown in Exhibit ES-9 with the capital cost fixed operating cost, variable operating cost, fuel cost and TS&M cost shown separately. When carbon capture is implemented, the net power output of the subcritical PC plant decreases to 413 MW and 359 MW for the 1,100 lb/net-MWh and 90 percent capture cases, respectively. The current electricity cost for the subcritical PC plant was estimated to be $19/MWh using the Energy Velocity Database. An estimated cost for the plant property taxes and insurance was added to the fixed O&M costs for a total current cost of electricity of $26.29/MWh. The current cost was levelized using the same factor applied to the retrofit cases yielding an LCOE of $33.78/MWh.

LCOE Cost Highlights:

- By virtue of having the intitial plant capital investment paid off, the subcritical PC retrofit case has the lowest LCOE of all cases.
- Comparing the greenfield IGCC and PC cases without CO_2 capture, the LCOE of the PC case is 32 percent lower than the IGCC case.
- Comparing the greenfield IGCC and PC cases with CO_2 capture, the supercritical PC has the lowest LCOE at $120.01/MWh at a capture level to meet a CO_2 emissions level of 1,100 lb/net-MWh and $143.89/MWh at a capture level of 90 percent. The LCOEs of the SCPC cases are 20 percent and 18 percent lower than the corresponding IGCC cases. For reference, the Energy Information Administration reports the average residential retail price of electricity in 2007 was $106.5/MWh.

Exhibit ES-9 Levelized Cost of Electricity for Power Plants

Study Case	**Levelized Cost of Electricity ($/MWh)**						**Incremental COE[a] (¢/kWh)**	**Increase COE[a] (%)[a]**
	Capital	Fixed O&M	Variable O&M	Fuel	TS&M	**Total**		
Case 1, IGCC w/o CO_2 Capture	79.13	20.17	11.41	7.13	0.00	117.84	-	-
Case 2, IGCC w/ 1,100 lb/MWh_{net} CO_2 Capture	99.62	24.57	13.33	8.37	3.49	149.38	31.54	26.8%
Case 3, IGCC w/ 90% CO_2 Capture	116.25	28.39	15.85	9.65	4.72	174.86	57.02	48.4%
Case 4, SCPC w/o CO_2 Capture	52.13	12.86	7.21	7.66	0.00	79.86	-	-
Case 5, SCPC w/ 1,100 lb/MWh_{net} CO_2 Capture	79.12	17.96	10.38	9.41	3.13	120.01	40.15	50.3%
Case 6, SCPC w/ 90% CO_2 Capture	94.52	20.84	13.09	11.07	4.37	143.89	64.02	80.2%
Case 7, Existing Subcritical PC w/o CO_2 Capture	0	13.16	1.48	19.14	0	33.78	-	-
Case 8, Existing Subcritical PC Retrofit w/ 1,100 lb/MWh_{net} CO_2 Capture	28.36	22.44	5.53	24.69	3.79	84.81	51.03	151%
Case 9, Existing Subcritical PC Retrofit w/ 90% CO_2 Capture	42.08	28.09	7.89	28.34	5.24	111.64	77.86	230%

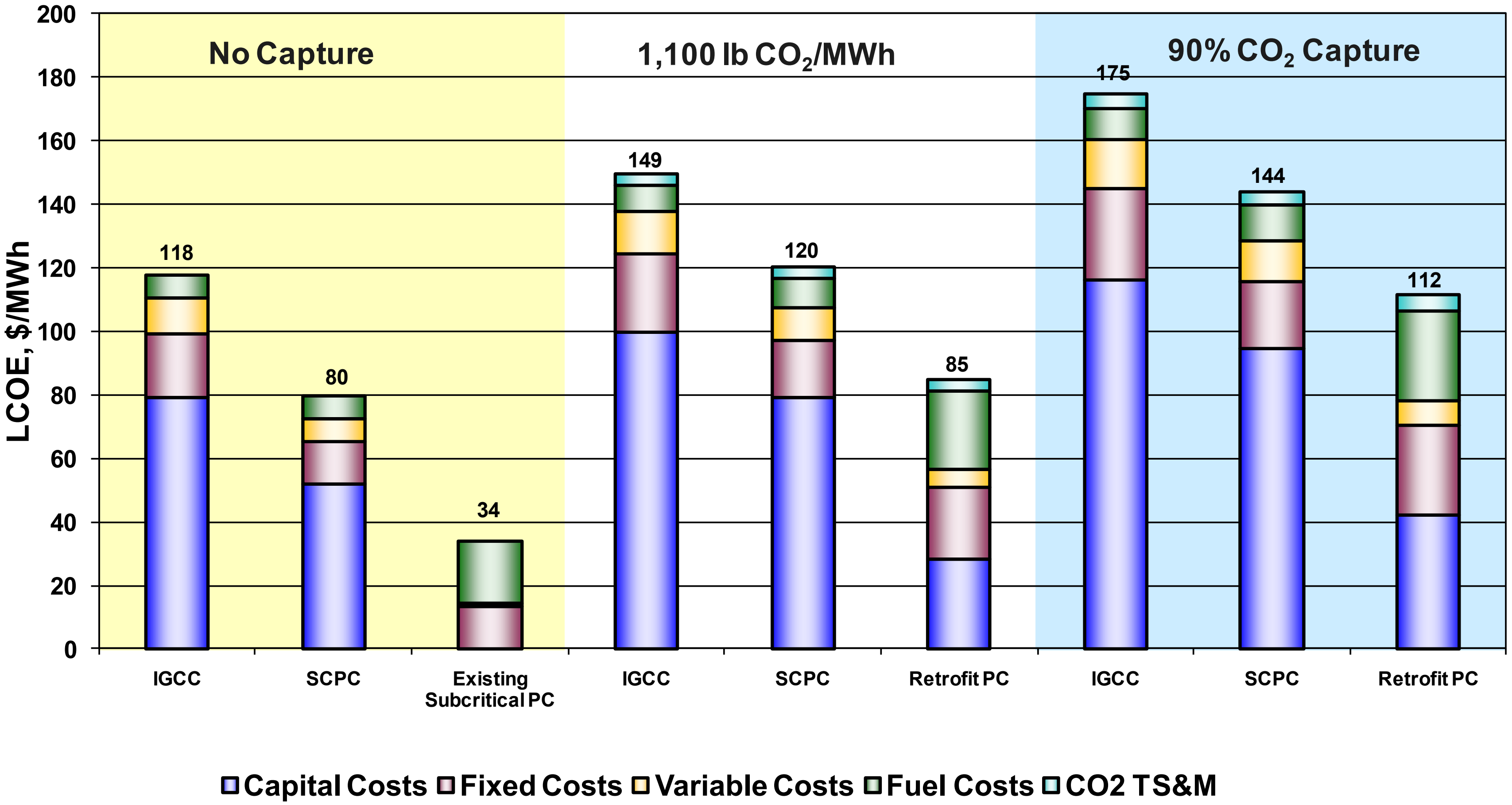
[a] Relative to non-capture case for each respective technology

The TS&M in the costs assume the CO_2 is transported 50 miles via pipeline to a geological sequestration field, injected into a saline formation at a depth of 4,055 ft and monitored for 30 years during plant life and 50 years following for a total of 80 years. These values are shown in Exhibit ES-10. The Total Levelized Costs of Electricity including TS&M are shown in Exhibit ES-11.

Exhibit ES-10 Levelized Cost of Electricity for CO_2 Transport, Storage, and Monitoring

	30 yr Levelized Costs ($/MWh)			
Study Case	CO_2 Transport	CO_2 Storage	CO_2 Monitoring	TS&M Total
Case 2, IGCC w/ 1,100 lb/MWh_{net} CO_2 Capture	2.62	0.42	0.40	3.45
Case 3, IGCC w/ 90% CO_2 Capture	3.34	0.47	0.91	4.72
Case 5, SCPC w/ 1,100 lb/MWh_{net} CO_2 Capture	2.29	0.32	0.52	3.13
Case 6, SCPC w/ 90% CO_2 Capture	2.90	0.41	1.05	4.37
Case 8, Existing Subcritical PC Retrofit w/ 1,100 lb/MWh_{net} CO_2 Capture	2.72	0.39	0.68	3.79
Case 9, Existing Subcritical PC Retrofit w/ 90% CO_2 Capture	3.53	0.57	1.14	5.24

Exhibit ES-11 LCOE By Cost Component



CO_2 Mitigation Cost

The cost of CO_2 capture was calculated in two ways, the cost of CO_2 removed and the cost of CO_2 avoided, as illustrated in Equations 1 and 2, respectively.

$$Removal\ Cost = \frac{\{LCOE_{with\ removal} - LCOE_{w/o\ removal}\}\ \$/MWh}{\{CO_2\ removed\}\ tons/MWh} \qquad \text{(Equation-1)}$$

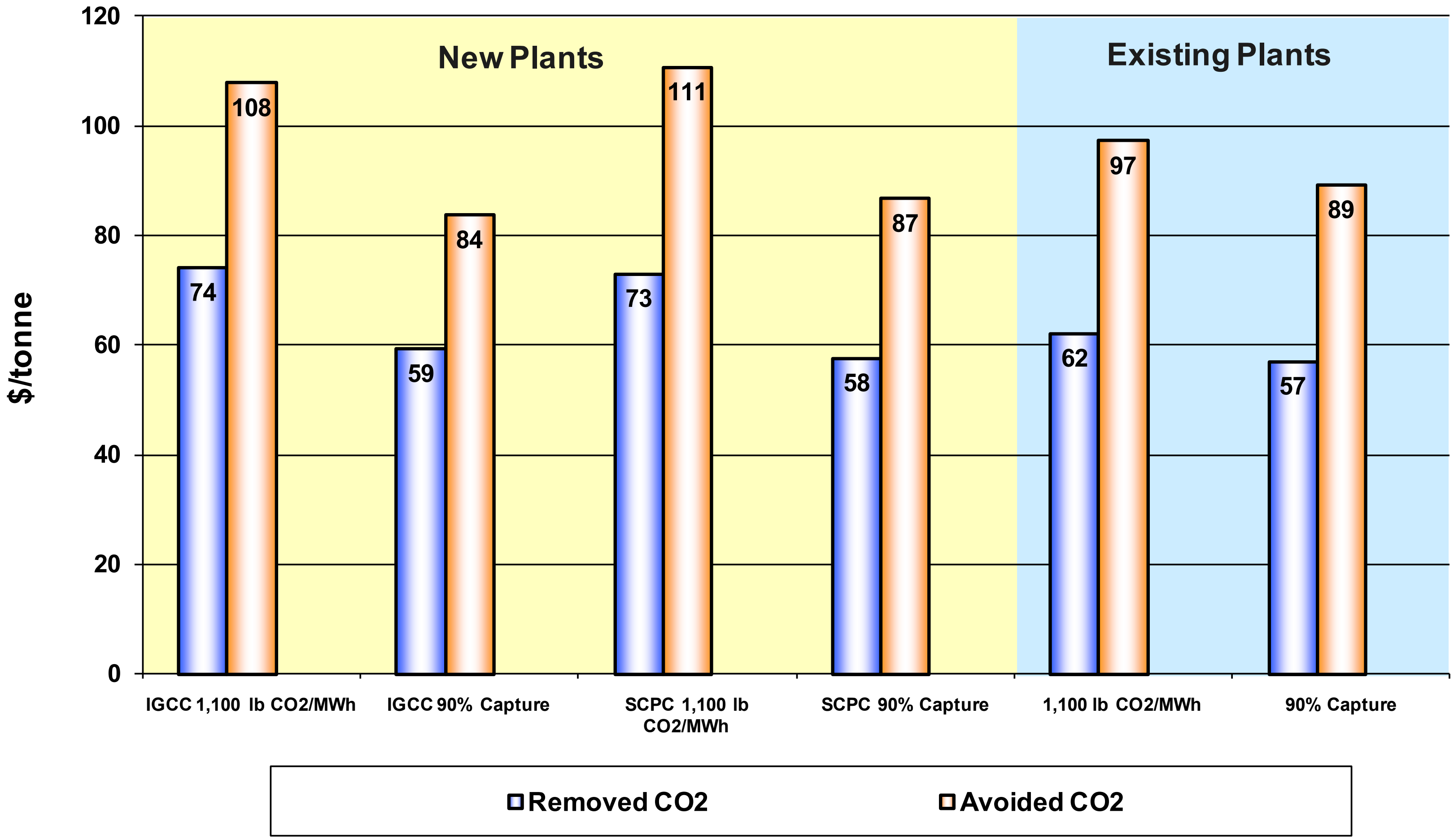
$$Avoided\ Cost = \frac{\{LCOE_{with\ removal} - LCOE_{w/o\ removal}\}\ \$/MWh}{\{Emissions_{w/o\ removal} - Emissions_{with\ removal}\}\ tons/MWh} \qquad \text{(Equation-2)}$$

The LCOE with CO_2 removal includes the costs of capture and compression as well as TS&M costs. The resulting removal and avoided costs are shown in Exhibit ES-12 for each of the six capture technologies modeled.

CO_2 Mitigation Cost Highlights:

- The total cost of CO_2 avoided meeting a 1,100 lb/net-MWh CO_2 emissions limit was estimated to be $108/tonne ($98/ton) for a new IGCC, $111/tonne ($100/ton) for a new SCPC, and $97/tonne ($88/ton) to retrofit the existing subcritical PC plant. At the 90 percent carbon capture level, the avoided cost was estimated to be $84/tonne ($76/ton) for a new IGCC, $87/tonne ($79/ton) for a new SCPC, and $89/tonne ($81/ton) to retrofit the subcritical PC plant.
- The CO_2 avoided costs are lowest for the existing subcritical PC plant at an emission level of 1,100 lb CO_2/net-MWh primarily because there is no capital cost associated with the power plant itself (CO_2 capture only) and because of the shorter construction duration (3 years versus 5 years for the greenfield plants). At 90 percent capture the reduction in net output offsets the capital cost savings and the existing PC plant has a slightly higher CO_2 avoided cost compared to greenfield SCPC and IGCC.
- The cost of CO_2 removed is 16 percent lower for the subcritical PC retrofit case compared to the greenfield PC and IGCC cases at a CO_2 emission limit of 1,100 lb/net-MWh.
- The cost of CO_2 removed is fairly constant for all technologies at the 90 percent capture level, ranging from $57-$59/tonne ($52-$54/ton).
- The normalized cost of CO_2 removed and avoided is lower for 90 percent capture cases compared to 1,100 lb/net-MWh cases mainly because of economies of scale. The size of CO_2 capture equipment increases in proportion to the amount of CO_2 captured, but the costs increase less than proportionately.

Exhibit ES-12 CO_2 Mitigation Cost



ENVIRONMENTAL PERFORMANCE

The environmental targets for each power plant scenario are summarized in Exhibit ES-13 and emission rates of SO_2, NOx, PM, Hg and CO_2 are shown graphically in Exhibit ES-14 through Exhibit ES-16. Targets were chosen on the basis of the environmental regulations that would most likely apply to plants built in 2015.

Exhibit ES-13 Study Environmental Targets

Pollutant	IGCC	Supercritical PC	Existing Subcritical Plant	Subcritical Retrofit Plant
SO_2	0.0128 lb/MMBtu	0.085 lb/MMBtu	0.256 lb/MMBtu	0.054 lb/MMBtu
NOx	15 ppmv (dry) @ 15% O_2	0.070 lb/MMBtu	0.45 lb/MMBtu	0.24 lb/MMBtu
PM (Filterable)	0.0071 lb/MMBtu	0.013 lb/MMBtu	0.027 lb/MMBtu	0.027 lb/MMBtu
Hg	>90% capture	0.70 lb/TBtu	6.00 lb/TBtu	6.00 lb/TBtu
CO_2	Three CO_2 emission levels were evaluated in this analysis for each case. Baseline—no CO_2 controls California Standard—1,100 $lbCO_2$/MWh-net Maximum Study Capture—90% CO_2 Capture			

Environmental targets were established for each of the scenarios as follows:

- IGCC cases use the EPRI targets established in their CoalFleet for Tomorrow work as documented in the CoalFleet User Design Basis Specification for Coal-Based Integrated Gasification Combined Cycle (IGCC) Power Plants: Version 4.
- Supercritical PC cases are based on best available control technology.
- The Existing Subcritical Plant environmental targets are based on typical subcritical PC emissions and the Subcritical Retrofit Plant is based on upgrades required to accommodate CO_2 capture.

Exhibit ES-14 Criteria Pollutant Emissions

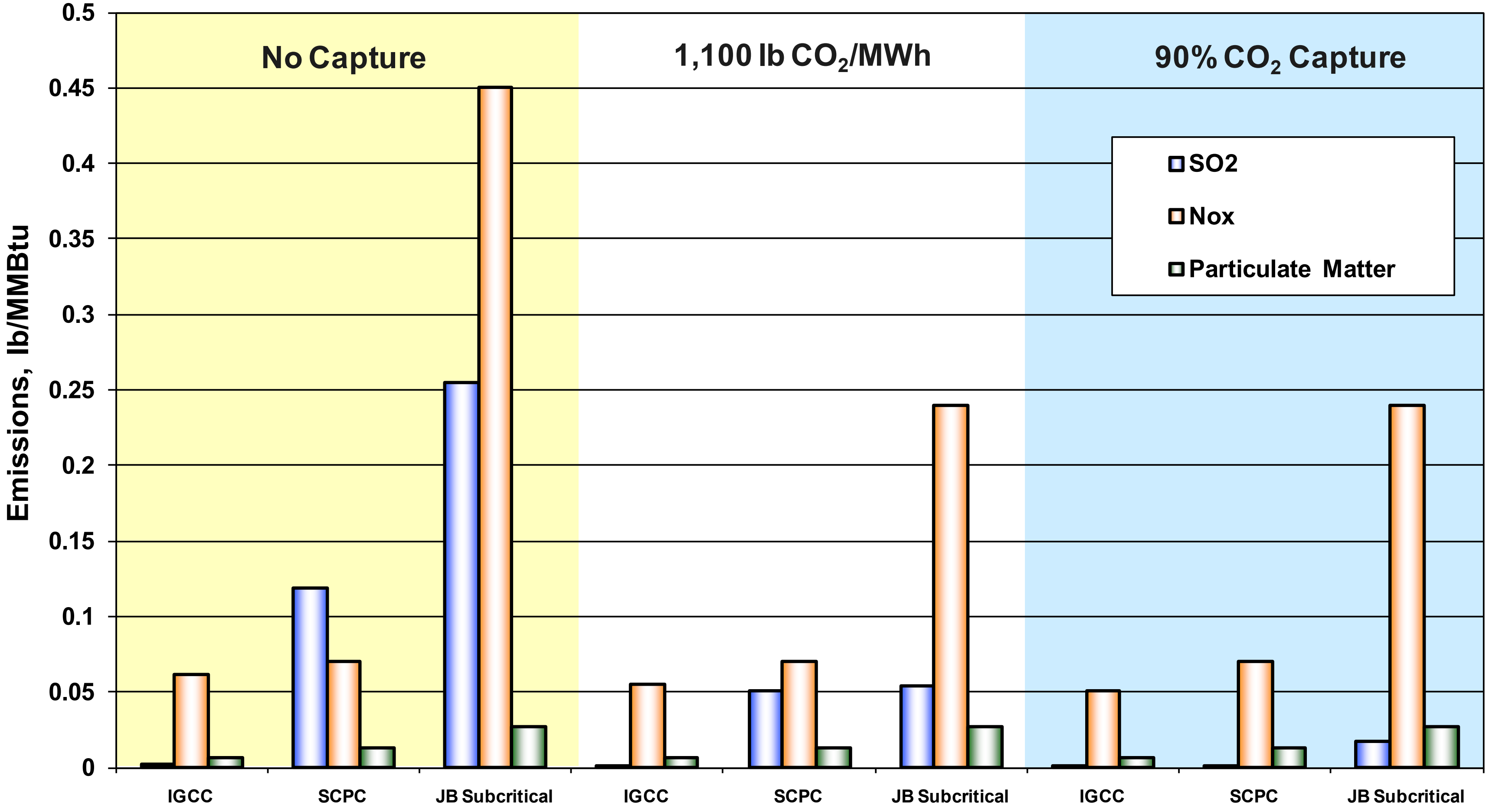


Exhibit ES-15 Mercury Emissions Rates

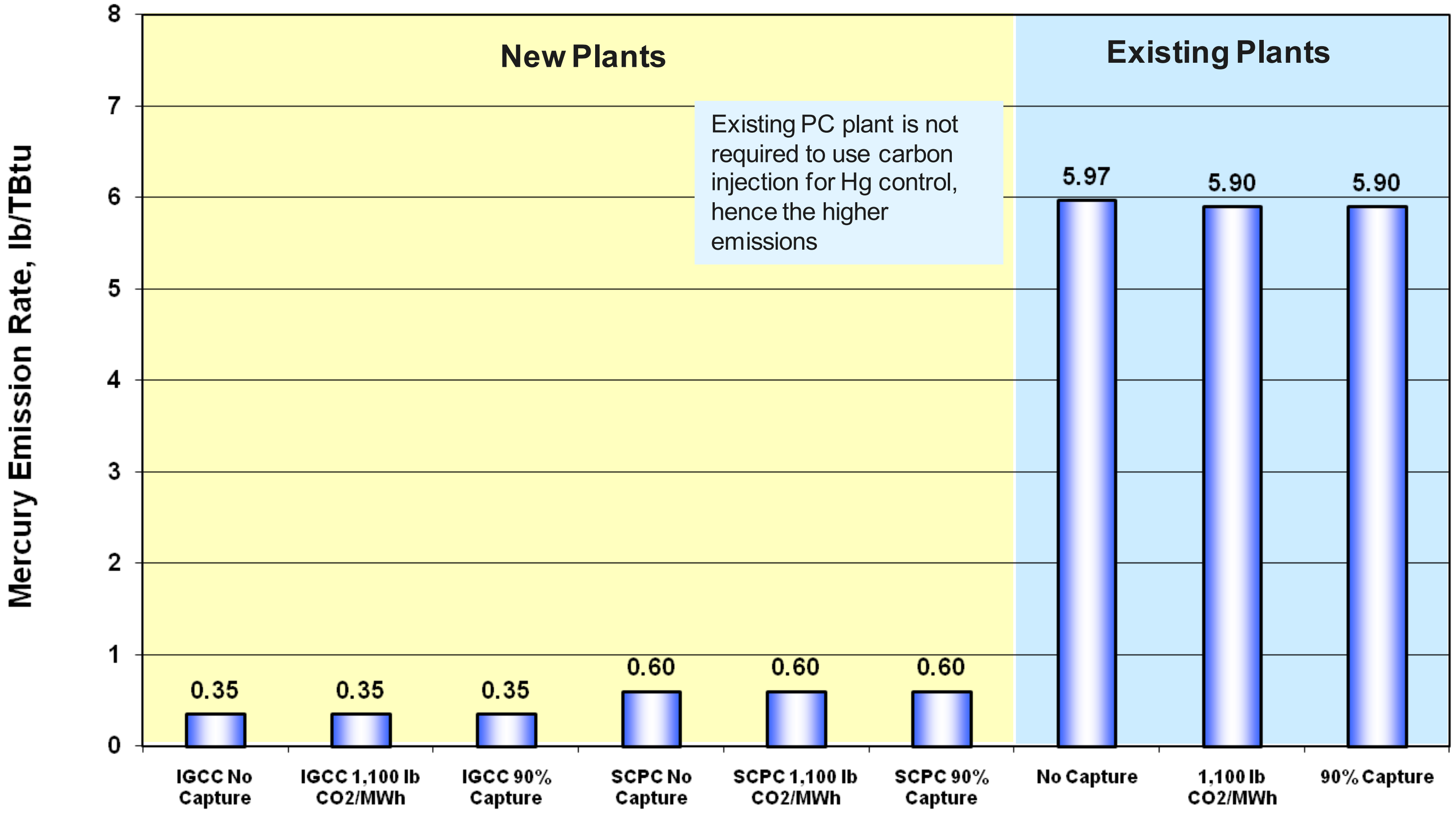
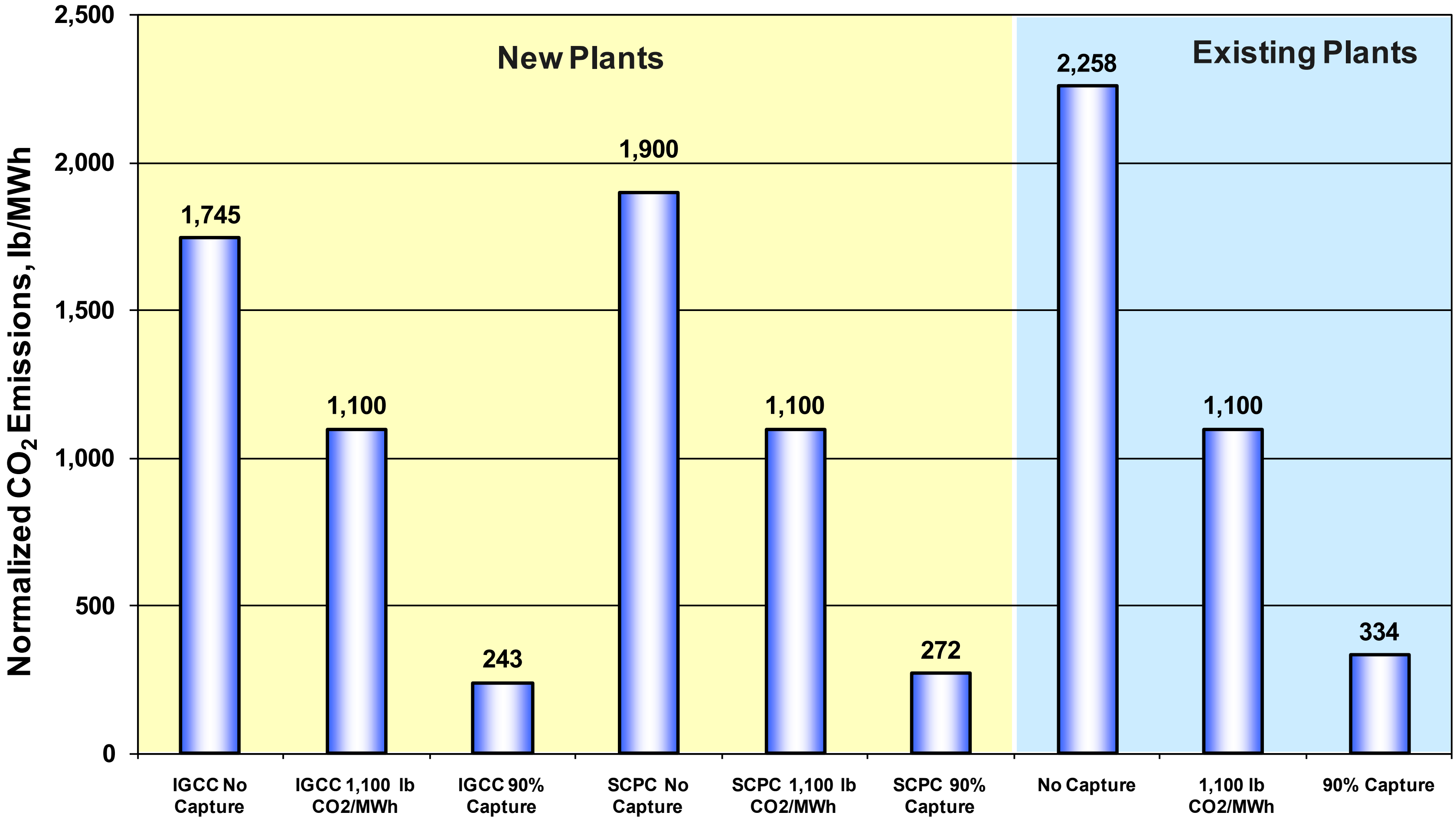


Exhibit ES-16 CO_2 Emissions Normalized By Net Power Output



Emission Key Findings:

- The IGCC baseline plant generates the lowest criteria pollutant emissions (NO_x, SO_x, PM, and mercury), followed by supercritical PC and then existing subcritical PC.
- In cases with no carbon capture, IGCC emits 8.2 percent less CO_2 than supercritical PC and 23 percent less CO_2 than existing subcritical PC per unit of net output. The relative emissions are indicative of the net efficiencies of each technology.
- In the IGCC cases the nominal 90 percent CO_2 reduction was accomplished by adding two sour gas shift (SGS) reactors to convert CO to CO_2 and using a two-stage Selexol process with a second stage CO_2 removal efficiency of up to 95 percent, which resulted in 90 percent reduction of CO_2 in the syngas. This number was supported by vendor quotes. In the 1,100 lb CO_2/net-MWh capture case, the 2 gasifier trains each have 1 SGS reactor with a bypass to achieve the emission limit, which resulted in 46 percent carbon capture.

This page intentionally left blank

1. INTRODUCTION

Revision 2 Updates

The technologies modeled in this study, namely integrated gasification combined cycle, subcritical pulverized coal and supercritical pulverized coal, are the subject of other ongoing systems analysis studies at the Department of Energy's National Energy Technology Laboratory. Vendor discussions that occurred as part of the other studies led to improved technology information that was incorporated into the Aspen models for this study. The updated models led to revised performance estimates, which were then used to update the cost estimates. The reference costs used for this study were also updated through efforts on other studies, and the most recent costs have been incorporated. In addition, owner's costs were added to the Total Plant Cost previously reported, and the capital component of levelized cost of electricity now includes owner's costs. The updated results are presented in the current revision (revision 2) of this report. Details of the modeling updates and cost methodology changes are identified throughout the report by the use of italicized font.

Greenhouse gas (GHG) emissions continue to receive increased scrutiny because of their perceived relation to global warming. Over the past several years, numerous bills have been introduced in both the United States Senate and House of Representatives that would limit GHG emissions. The bills vary primarily in the economy sectors regulated, the extent of GHG reductions and the compliance year, but all represent reductions from the "business-as-usual" scenario. *In June, 2009, the American Clean Energy and Security Act of 2009 (H.R. 2454) was passed by the House of Representatives. The bill requires that GHG emissions from regulated sources be reduced to 97 percent of 2005 levels by 2012; to 83 percent by 2020; to 58 percent by 2030; and to 17 percent by 2050 [2]. While the bill has not yet come to a vote in the Senate, it continues to be debated and similar bills are also under consideration. Adding to the legislative momentum for carbon regulation, in September, 2009 the Environmental Protection Agency proposed a rule that would limit future regulation of GHG emissions under the Clean Air Act to industrial facilities that emit 25,000 tons or more of carbon dioxide annually. The proposed rule would impact facilities such as power plants, refineries, and factories, which produce nearly 70 percent of domestic GHGs.* In addition to proposed Federal regulations, various states have proposed or enacted legislation to reduce GHG emissions. A sampling of the legislation is provided in Section 1.1.

The objective of this report is to present the baseline cost and performance of greenfield integrated gasification combined cycle (IGCC) plants, greenfield supercritical (SC) pulverized coal (PC) plants, and retrofit subcritical PC plants that limit carbon dioxide (CO_2) emissions to various levels. For each plant type, three cases were modeled:

- Baseline performance with no CO_2 capture
- CO_2 emissions reduced to 1,100 lb/net-MWh
- CO_2 emissions reduced by 90 percent

The intermediate value of 1,100 lb/net-MWh was chosen to match the recent interim California standard established in January 2007 [1].

The fuel used in all nine cases was representative of a coal from the Powder River Basin (PRB). The nine cases are summarized in Exhibit 1-1.

Generating Unit Configurations

The three IGCC cases are each comprised of two advanced F-class turbines and a single steam turbine. The advanced F-class turbine comes in a standard size of 232 MW when operated on syngas at ISO conditions. Because these cases are operated at elevations greater than sea level (2,042 m [6,700 ft]), the output is reduced from the turbine's ISO condition potential. In the combined cycle, a heat recovery steam generator extracts heat from the combustion turbine exhaust to power a steam turbine. However, the carbon capture cases consume more extraction steam than the non-capture cases, thus reducing the steam turbine output. In addition, the capture cases have a higher auxiliary load requirement than non-capture cases, which serves to further reduce net plant output. Although the gross combustion turbine output increases with increasing levels of carbon capture, the net plant output decreases as CO_2 capture increases because of the higher auxiliary loads and the increased extraction steam requirement for the water-gas shift reactions. The gross combustion turbine output increases because the syngas has a higher heat of combustion (Btu/lb) with increase capture levels, while maintaining a relatively similar mass flow. The resulting net output ranges from 504 MW for Case 1 (no capture) to 396 MW for Case 3 (90% CO_2 capture).

The three greenfield SC PC cases are all modeled to maintain a nominal net output of 550 MW. The increased auxiliary loads due to CO_2 capture plus the extraction steam required for amine regeneration result in higher gross outputs for the capture cases, ranging from 584 MW for Case 4 (no capture) to 677 MW in Case 6 (90% CO_2 capture).

Since the boiler heat input (coal flowrate and boiler size), and therefore the amount of steam generated, is a fixed quantity in the existing subcritical power plant, the gross and net output are both decreased when retrofitted for CO_2 capture. The extraction steam required to regenerate the amine solvent in the capture cases causes a de-rating to the existing steam turbine and the higher auxiliary loads (i.e. CO_2 compression) further reduces the net output. The baseline case or 'current state" is Case 7 and has a net power output equal to 532 MW. When the plant is retrofitted with 90 percent CO_2 capture (Case 9) the net power output is reduced to 359 MW (a 33% reduction or 173 MW loss of power to the grid).

1.1 GHG EMISSION STANDARDS

The United States continues to progress towards restrictions of greenhouse gas (GHG) emissions, including carbon dioxide (CO_2). In 1992, the United States ratified the United Nations' Framework Convention on Climate Change (UNFCCC), which called on industrialized countries to take the lead in reducing greenhouse gas emissions. The UNFCCC did not include mandatory reductions, but set an ultimate objective of stabilizing greenhouse gas concentrations "at a level that would prevent dangerous anthropogenic interference with the climate system." The convention requires precise and regularly updated inventories of greenhouse gas emissions from industrialized countries. The "base year" for tabulating greenhouse gas emissions was set as 1990.

The first addition to the treaty, the Kyoto Protocol, was adopted in 1997. The United States chose not to ratify the Kyoto Protocol, which called for legally binding commitments by developed countries to reduce their greenhouse gas emissions. Instead, the domestic initiatives have focused on voluntary reductions in the growth of GHG emissions per unit of gross domestic product (GDP) and the development of advanced technologies to improve energy efficiency and control GHG emissions reliably and cost-effectively.

Many bills were introduced in the 109th and 110th Congresses with the goal of reducing emissions of GHG. During the 111th Congress, the House of Representatives passed H.R. 2454 which includes provisions for reducing GHG emissions from regulated sources. While the Senate has yet to act on H.R. 2454 or any companion bills, momentum continues to gather for climate change legislation. A sampling of bills that were previously proposed along with a summary of H.R. 2454 are shown below to illustrate the format that climate change legislation may ultimately take [2,3,4,5,6].

- *H.R. 2454 (introduced by Representatives Waxman and Markey and passed by the House of Representatives in June 2009) would require reduction of greenhouse gases from regulated sources to 97 percent of 2005 levels by 2012; to 83 percent by 2020; to 58 percent by 2030; and to 17 percent by 2050.*
- S. 342 (introduced by Senators McCain and Lieberman) and H.R. 759 (introduced by Representatives Gilchrest and Oliver) would cap emissions of CO_2 in the year 2010 based on the affected facilities' 2000 emissions (for any entity that emits from a single facility more than 10,000 metric tons of GHG annually [CO_2 equivalent]). The legislation would be implemented through an expansive allowance trading program that would allow cross-sector trading, increases in carbon sequestration and limited acquisition of allowances from foreign sources.
- S. 150 (introduced by Senator Jeffords) would require electric generating facilities producing 15 MW or greater to meet an aggregate CO_2 emissions cap in the year 2010. The national emissions cap would be set at 1990 emissions levels for electric generating facilities and would be implemented through a tradable allowance program.
- S. 730 (introduced by Senator Leahy) is the same as S. 150 except that trading is restricted to within a single facility, not across different sites.
- H.R. 1451 (introduced by Representative Waxman) is essentially the same as S. 150 with regard to CO_2 emissions.
- H.R. 1873 (introduced by Representative Bass) would set a CO_2 cap in 2010 at the projected 2006 levels, declining to actual 2001 levels in 2015.
- H.R. 1590 (introduced by Representative Waxman and also called The Safe Climate Act of 2007) freezes domestic GHG emissions in 2010 at 2009 levels. Beginning in 2011, it would cut emissions by roughly 2 percent per year, reaching 1990 emission levels by 2020. After 2020, it would cut emissions by roughly 5 percent per year. By 2050, emissions would be 80 percent lower than in 1990.
- S. 280 (introduced by Senator Lieberman) would cap emissions of the six greenhouse gases specified in the United Nations Framework Convention on Climate Change, at reduced levels, form the electric generation, transportation, industrial, and commercial sectors. The initial cap would restrict 2012 emissions from the affected facility to their 2004 emission levels. The cap steadily declines until it is equal to one-third of the facilities' 2004 level.
- S. 317 (introduced by Senator Feinstein) would cap GHG emissions from electric generators over 25 MW. Beginning in 2011, affected generators would be capped at their

- 2006 levels, declining to 2001 levels by 2015. After that, the emission cap would decline 1% annually until 2020, when the rate of decline would increase to 1.5%.
- The Low Carbon Economy Act of 2007 (introduced by Senators Bingaman and Specter) would require economy-wide CO_2 emissions to be reduced to 2006 levels by 2020. Following that, the bill proposes to achieve reductions to 1990 levels by 2030 and at least 60 percent below current levels by 2050. The bill also contains a “safety valve” provision, which allows entities to purchase climate change credits at a relatively low cost in early years, to allow time for carbon sequestration technologies to be developed.

While GHG limits continue to be debated at a national level, many state and local governments have already passed climate change legislation. Twenty-five states have passed Renewable Portfolio Standards (RPS) and two other states have voluntary standards. Seventeen states have established GHG emission targets [7].

In 2006, the California Legislature passed Assembly Bill 32 (the Global Warming Solutions Act of 2006), which mandates that California must reduce its GHG emissions to 2000 levels by 2010 and to 1990 levels by 2020. Senate Bill No. 1368 further required that, “On or before February 1, 2007, the commission, through a rulemaking proceeding, and in consultation with the Energy Commission and the State Air Resources Board, shall establish a greenhouse gases emission performance standard for all baseload generation of load-serving entities, at a rate of emissions of greenhouse gases that is no higher than the rate of emissions of greenhouse gases for combined-cycle natural gas baseload generation.”

In response to Senate Bill No. 1368, on January 25, 2007 the California Public Utilities Commission adopted an interim GHG Emissions Performance Standard. The level established is 1,100 lb CO_2/net-MWh for all new long-term commitments for baseload generation that serves California consumers. “New long-term commitment” refers to new plant investments (new construction), new or renewal contracts with a term of five years or more, or major investments by the utility in its existing baseload power plants. The clause “baseload generation that serves California consumers” also makes it applicable to imported power supplies.

The California GHG Emissions Performance Standard provided the impetus for this study. The results provide an overview of the cost and performance impacts that such a standard will have on new and existing coal-fired power plants.

The balance of this report is organized as follows:

- Chapter 2 provides the basis for technical, environmental and cost evaluations.
- Chapter 3 describes the systems of all three greenfield Shell IGCC cases.
- Chapter 4 provides the results of the three greenfield Shell IGCC cases.
- Chapter 5 describes the systems common to all six PC cases.
- Chapter 6 describes the results of the supercritical PC cases.

- Chapter 7 describes the results of the existing subcritical PC retrofit.
- Chapter 8 provides the conclusions that summarize the major differences for all nine cases.
- Chapter 9 contains the reference list.

Exhibit 1-1 Summary of Cases

Case	Power Plant Type	Steam Cycle, psig/°F/°F	Oxidant	Sulfur Removal	PM Control	NOx Control	CO_2 Capture	Capture Level	CO_2 Storage[c]
1	**NEW IGCC** Shell Gasifier 2 x Advanced F Class Combustion Turbines	1800/1048/1048	95 mol% O_2	Sulfinol with Claus Plant	Cyclone, barrier filter and scrubber	N_2 dilution	Case 1—IGCC No CO_2 Capture		
2		1800/1020/1020[a]		Selexol with Claus Plant			Selexol 2nd Stage added	1,100 lb/net-MWh	Off-Site
3		1800/996/996[a]						90%	
4	**NEW** **PULVERIZED COAL** Supercritical	3500/1100/1100	Air	Spray Dryer Absorber	Baghouse	LNB w/OFA and SCR	Case 4—PC No CO_2 Capture		
5							Amine Scrubbing	1,100 lb/net-MWh	Off-Site
6								90%	
7	**EXISTING** **PULVERIZED COAL** Subcritical	2400/1000/1000	Air	Existing Wet FGD/ Sodium based	ESP	OFA and 'retro' LNB	Case 7—Existing PC No CO_2 Capture		
8				Upgrade[b] Existing Wet FGD/ Sodium based		OFA and Reconfigure LNB	Amine Scrubbing	1,100 lb/net-MWh	Off-Site
9								90%	

ESP = Electrostatic Precipitator, OFA = Overfired air, LNB = Low NOx Burners

[a] For the IGCC w/CO_2 capture cases, the steam conditions are lowered due to a lower temperature flue gas exiting the combustion turbine.
[b] Upgrade existing Wet FGD by removing the internal bypass and modifying the stack to handle wet operation. This increases efficiency from 85% to 92% to decrease the sulfur concentration entering the CO_2 scrubbing process.
[c] Transported 50 miles via pipeline to a geologic sequestration field for injection into a saline formation

2. GENERAL EVALUATION BASIS

For each of the plant configurations in this study an AspenPlus model was developed and used to generate material and energy balances, which in turn were used to provide a design basis for items in the major equipment list. The equipment list and material balances were used as the basis for creating factored capital and operating cost estimates. Cost estimates were generated for the greenfield Shell IGCC and SCPC cases (on PRB coal) from a previous study and those costs were used as the scaling basis. The original cost estimates were based on simulation results and through a combination of vendor quotes, scaled estimates from previous design/build projects, or a combination of the two. Ultimately, a constant dollar, 30-year levelized cost of electricity (LCOE) was calculated for each of the cases and is reported as the revenue requirement figure-of-merit.

Performance and process limits were based upon published reports, information obtained from vendors and users of the technology, performance data from design/build utility projects, and/or best engineering judgment. The performance of the subcritical PC plant was based on a suite of publicly available data, including a report published by the Energy Information Administration (EIA) and a Best Available Retrofit Technology (BART) report [8].

The balance of this chapter documents the design basis, environmental targets and cost assumptions used in the study.

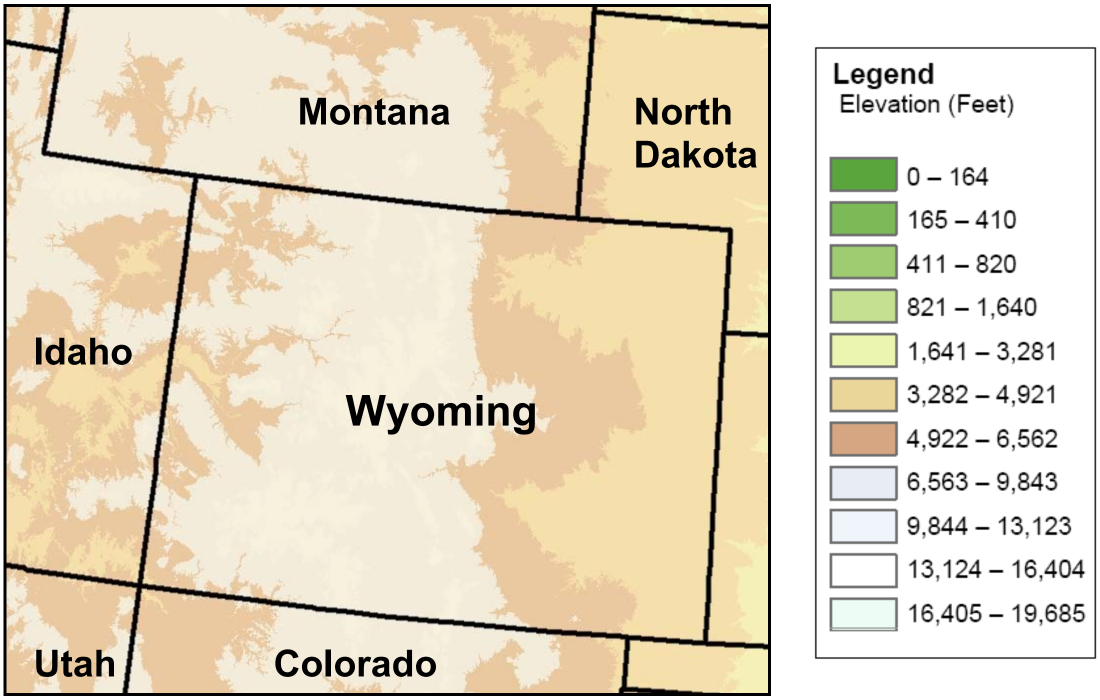
2.1 SITE CHARACTERISTICS

The plants are located in Wyoming, U.S. The ambient conditions for the site are shown in Exhibit 2-1. The average elevation of the state of Wyoming is 6,700 ft and was the basis for the elevation chosen for this study. A topographical map of the state of Wyoming is shown in Exhibit 2-2.

Exhibit 2-1 Site Ambient Conditions for All Cases

Elevation, m (ft)	2,042 (6,700)
Barometric Pressure, MPa (psia)	0.08 (11.4)
Design Ambient Temperature, Dry Bulb, °C (°F)	5.6 (42)
Design Ambient Temperature, Wet Bulb, °C (°F)	2.8 (37)
Design Ambient Relative Humidity, %	62

Exhibit 2-2 Topographical Map of Wyoming



The assumed site characteristics are shown in Exhibit 2-3.

Exhibit 2-3 Site Characteristics

Location	Wyoming, USA
Topography	Level
Size, acres	300
Ash/Slag Disposal	Off Site
Water Source	Municipal (50%) / Groundwater (50%) for Cases 1-6 Green River for Cases 7-9
CO_2 Storage	Compressed to 15.3 MPa (2,215 psia), transported 80 kilometers (50 miles) and sequestered in a saline formation at a depth of 1,239 meters (4,055 feet)

The land area for all greenfield cases (PC and IGCC) assumes 30 acres are required for the plant proper and the balance provides a buffer of approximately 0.25 miles to the fence line. Sufficient land area for additional controls, including CO_2 capture and compression, is assumed available in the retrofit PC cases.

In all cases it was assumed that the steam turbine is enclosed in a turbine building and the boiler in the PC cases is also enclosed, but the gasifier in the IGCC cases is not enclosed.

The following design parameters are considered site-specific, and are not quantified for this study. Allowances for normal conditions and construction are included in the cost estimates.

- Flood plain considerations
- Existing soil/site conditions
- Water discharges and reuse
- Rainfall/snowfall criteria
- Seismic design
- Buildings/enclosures
- Fire protection
- Local code height requirements
- Noise regulations – Impact on site and surrounding area

2.2 COAL CHARACTERISTICS

The design coal is a subbituminous PRB coal from Montana. The coals properties are from NETL's Coal Quality Guidelines and are shown in Exhibit 2-4 [9].

The first year cost of coal used in this study is \$0.57/GJ (\$0.61/MMBtu). The first year coal cost is the EIA projected cost of Montana Rosebud PRB coal for 2015 in 2005 dollars. This cost is then scaled to 2007 dollars. The projected 2015 coal cost was used to correspond with the start-up date for the greenfield IGCC and SCPC cases, but is applied to all cases to enable comparison. The costs were determined using the following information from the Energy Information Administration's (EIA) 2007 Annual Energy Outlook (AEO):

- The 2015 minemouth cost of PRB coal in 2005 dollars, \$10.85/tonne (\$9.84/ton), was obtained from Supplemental Table 113 of the EIA's 2007 AEO for western Montana medium-sulfur subbituminous coal.
- The plants are assumed to be minemouth so transportation costs are zero.
- The 2015 cost of PRB coal was escalated to 2007 dollars using the gross domestic product (GDP) chain-type price index from AEO 2007, resulting in a price of \$11.43/tonne (\$10.37/ton) or \$0.57/GJ (\$0.61/MMBtu) [10]. (Note: The PRB coal cost conversion of \$10.37/ton to dollars per million Btu results in \$0.6053/MMBtu which was used in calculations, but only two decimal places are shown in the report.)

2.3 ENVIRONMENTAL TARGETS

The current federal regulation governing new fossil-fuel fired electric utility steam generating units is the New Source Performance Standards (NSPS) as amended in February 2006 and shown in Exhibit 2-5, which represents the minimum level of control that would be required for a new fossil energy plant.

Exhibit 2-4 Montana Rosebud PRB, Area D, Western Energy Co. Mine, Subbituminous Design Coal Analysis

Proximate Analysis	**Dry Basis, %**	**As Received, %**
Moisture	0.0	25.77
Ash	11.04	8.19
Volatile Matter	40.87	30.34
Fixed Carbon	48.09	35.70
Total	100.0	100.0
Ultimate Analysis	**Dry Basis, %**	**As Received, %**
Carbon	67.45	50.07
Hydrogen	4.56	3.38
Nitrogen	0.96	0.71
Sulfur	0.98	0.73
Chlorine	0.01	0.01
Ash	11.03	8.19
Moisture	0.00	25.77
Oxygen (Note A)	15.01	11.14
Total	100.0	100.0
Heating Value	**Dry Basis, (Dulong Calc.)**	**As Received, %**
HHV, kJ/kg	26,787	19,920
HHV, Btu/lb	11,516	8,564
LHV, kJ/kg	25,810	19,195
LHV, Btu/lb	11,096	8,252
Hardgrove Grindability Index	57	
Ash Mineral Analysis		**%**
Silica	SiO_2	38.09
Aluminum Oxide	Al_2O_3	16.73
Iron Oxide	Fe_2O_3	6.46
Titanium Dioxide	TiO_2	0.72
Calcium Oxide	CaO	16.56
Magnesium Oxide	MgO	4.25
Sodium Oxide	Na_2O	0.54
Potassium Oxide	K_2O	0.38
Sulfur Trioxide	SO_3	15.08
Phosphorous Pentoxide	P_2O_5	0.35
Barium Oxide	Ba_2O	0.00
Strontium Oxide	SrO	0.00
Unknown	---	0.84
	Total	100.0
Trace Components		**ppmd**
Mercury (Note B)	Hg	0.081

Notes: A. By Difference

B. Mercury value is the mean plus one standard deviation using EPA's ICR data

Exhibit 2-5 Standards of Performance for Electric Utility Steam Generating Units Built, Reconstructed, or Modified After February 28, 2005

	New Units		Reconstructed Units		Modified Units	
	Emission Limit	% Reduction	Emission Limit (lb/MMBtu)	% Reduction	Emission Limit (lb/MMBtu)	% Reduction
PM	0.015 lb/MMBtu	99.9	0.015	99.9	0.015	99.8
SO_2	1.4 lb/MWh[1]	95	0.15	95	0.15	90
NOx	1.0 lb/MWh[1]	N/A	0.11	N/A	0.15	N/A

[1] Gross MWh output

The new NSPS standards apply to units with the capacity to generate greater than 73 MW of power by burning fossil fuels, as well as cogeneration units that sell more than 25 MW of power and more than one-third of their potential output capacity to any utility power distribution system. The rule also applies to combined cycle, including IGCC plants, and combined heat and power combustion turbines that burn 75 percent or more synthetic-coal gas. In cases where both an emission limit and a percent reduction are presented, the unit has the option of meeting one or the other. All limits with the unit lb/MWh are based on gross power output.

Other regulations that could affect emissions limits from a new plant include the New Source Review (NSR) permitting process and Prevention of Significant Deterioration (PSD). The NSR process requires installation of emission control technology meeting either Best Available Control Technology (BACT) determinations for new sources being located in areas meeting ambient air quality standards (attainment areas), or Lowest Achievable Emission Rate (LAER) technology for sources being located in areas not meeting ambient air quality standards (non-attainment areas). Environmental area designation varies by county and can be established only for a specific site location. Based on the Environmental Protection Agency (EPA) Green Book Non-attainment Area Map relatively few areas in the Western U.S. are classified as "non-attainment" so the greenfield plant site for this study was assumed to be in an attainment area [11].

In addition to federal regulations, state and local jurisdictions can impose even more stringent regulations on a new facility. However, since each new plant has unique environmental requirements, it was necessary to apply some judgment in setting the environmental targets for this study.

The Clean Air Mercury Rule (CAMR) established NSPS limits for Hg emissions. While CAMR is no longer legally binding, it is used as a reference until new regulations are established. The

IGCC limits are independent of coal type and the PC limits are dependent on the type of coal used. The applicable limit for IGCC cases in this study is 20×10^{-6} lb/MWh. The applicable limit for the PC cases is 97×10^{-6} lb/MWh. The NSPS limits, based on gross output, are shown in Exhibit 2-6.

Exhibit 2-6 NSPS Mercury Emission Limits

Coal Type / Technology	Hg Emission Limit
Bituminous/ PC	20×10^{-6} lb/MWh
Subbituminous (wet units)/ PC	66×10^{-6} lb/MWh
Subbituminous (dry units)/ PC	**97×10^{-6} lb/MWh**
Lignite/ PC	175×10^{-6} lb/MWh
Coal refuse/ PC	16×10^{-6} lb/MWh
All coals/ IGCC	**20×10^{-6} lb/MWh**

The coal mercury concentration used for this study was determined from the Environmental Protection Agency's (EPA) Information Collection Request (ICR) database. The ICR database has 137 records of Montana Rosebud subbituminous coal with an average Hg concentration of 0.056 ppm (dry) and a standard deviation of 0.025 ppm. The mercury value in Exhibit 2-4 is the mean plus one standard deviation, or 0.081 ppm (dry) [12]. It was further assumed that all of the coal Hg enters the gas phase and none leaves with the bottom ash or slag.

2.3.1 IGCC Environmental Targets

The IGCC environmental targets were chosen to match the Electric Power Research Institute's (EPRI) design basis for their CoalFleet for Tomorrow Initiative and are shown in Exhibit 2-7 [13]. The design targets were established specifically for bituminous coal, but are applied to subbituminous as well. Because of the lower coal sulfur content in the Montana Rosebud PRB coal, actual SO_2 emissions in this study are substantially lower than the environmental target. EPRI notes that these are design targets and are not to be used for permitting values.

Exhibit 2-7 IGCC Environmental Targets

Pollutant	Environmental Target	NSPS Limit[1]	Control Technology
NOx	15 ppmv (dry) @ 15% O_2	1.0 lb/MWh (0.117 lb/MMBtu)	Low NOx burners and syngas nitrogen dilution
SO_2	0.0128 lb/MMBtu	1.4 lb/MWh (0.163 lb/MMBtu)	Sulfinol—non capture cases Selexol—capture cases
Particulate Matter (Filterable)	0.0071 lb/MMBtu	0.015 lb/MMBtu	Full quench (capture cases), water scrubber, and cyclones
Mercury	> 90% capture	20×10^{-6} lb/MWh (2.3 lb/TBtu)	Carbon bed

[1] The NSPS value in parentheses is calculated based on an average heat rate of 8,570 Btu/kWh from the two non-CO_2 capture gasifier cases.

IGCC Emissions Design Assumptions

1. **NOx:** Based on published vendor literature, it was assumed that low NOx burners (LNB) and nitrogen dilution can achieve 15 ppmv (dry) at 15 percent O_2, and that value was used for all IGCC cases [14, 15].
2. **SO_2:** To achieve an environmental target of 0.0128 lb/MMBtu of SO_2 (see Exhibit 2-7) requires approximately 28 ppmv sulfur in the sweet syngas. The acid gas removal (AGR) process requires a sulfur capture efficiency of about 99.7 percent to reach the environmental target using bituminous coal with a sulfur content of 2.51 percent. Vendor data on the AGR processes used in this study indicate that this level of sulfur removal is possible, resulting in substantially lower SO_2 emissions because of the lower coal sulfur content. In the CO_2 capture cases, the two-stage Selexol process was designed for just over 90 percent CO_2 removal, which results in a sulfur capture of greater than 99.7 percent due to co-sequestration of some sulfur containing compounds.
3. **PM:** Most of the coal ash is removed from the gasifier as slag. The ash that remains entrained in the syngas is captured in the downstream equipment, including the syngas scrubber and a cyclone and either ceramic or metallic candle filters. The environmental target of 0.0071 lb/MMBtu filterable particulates can be achieved with this combination of particulate control devices so that it was assumed the environmental target was met exactly.
4. **Mercury:** The environmental target for mercury capture is greater than 90 percent. Based on experience at the Eastman Chemical plant, where syngas from a GEE gasifier is treated, the actual mercury removal efficiency used is 95 percent. Sulfur-impregnated activated carbon is used by Eastman as the adsorbent in the packed beds operated at 30°C (86°F) and 6.2 MPa (900 psig). Mercury removal between 90 and 95 percent has been reported with a bed life of 18 to 24 months. Removal efficiencies may be even higher, but at 95 percent the measurement precision limit was reached. Eastman has yet to experience any mercury contamination in its product [16]. Mercury removals of greater

than 99 percent can be achieved by the use of dual beds, i.e., two beds in series. However, this study assumes that the use of sulfur-impregnated carbon in a single carbon bed achieves 95 percent reduction of mercury emissions which meets the environmental target and NSPS limits in all cases.

2.3.2 Pulverized Coal Environmental Targets

Best available control technology (BACT) was applied to the greenfield supercritical PC cases, and the resulting emissions were compared to NSPS limits and recent permit averages. Since the BACT results met or exceeded the NSPS requirements and the average of recent permits, they were used as the environmental targets. The average of recent permits is comprised of 8 units at 5 locations. The 5 plants include Elm Road Generating Station, Longview Power, Prairie State, Thoroughbred and Cross.

The existing subcritical PC plant used in this study does not, and is not required to, meet the NSPS limits. However, Best Available Retrofit Technology (BART) for SO_2 and NOx control was applied to the subcritical PC cases retrofitted for CO_2 capture because of the amine-based system limits on those pollutants [8]. In addition, it was assumed that the retrofit modifications for the existing subcritical PC plant would not trigger New Source Review (NSR) environmental standards because additional capacity and subsequent emissions rates are not increased. Per U.S.C. §7411(a), NSR is only applicable when “any physical change in, or change in the method of operation of, a stationary source which increases the amount of any air pollutant emitted by such source which results in the emissions of any air pollutant not previously emitted.” However, in the event that NSR is triggered and selective catalytic reduction (SCR) needs to be implemented for NOx control, a sensitivity case was added to determine the impact on costs. The environmental targets for the greenfield supercritical PC plant and the existing subcritical PC plant are shown in Exhibit 2-8.

Pulverized Coal Emission Control Design Assumptions

1. **NOx:** In the new SCPC cases, the NOx emissions exiting the boiler equipped with low NOx burners and overfire air would be 0.20 lb/MMBtu . Adding an SCR unit would further reduce the NOx by 65 percent, resulting in the emission of 0.070 lb/MMBtu.

   The current subcritical PC plant NOx emissions are 0.45 lb/MMBtu. With the implementation of new LNBs and improved OFA the emissions would be reduced to 0.24 lb/MMBtu according to the CH2MHill Bart Analysis [8]. This level of control was assumed to meet the amine-based CO_2 capture NO_2 limit.

   With the addition of SCR to the subcritical PC retrofit, the NOx emissions would be further reduced to 0.070 lb/MMBtu.

2. **SO_2:** The lime-based spray dry absorber utilized in the new SCPC cases was assumed to be 93 percent efficient which results in SO_2 emissions of 0.119 lb/MMBtu for the non-capture case (Case 4). Current technology allows flue gas desulfurization (FGD) removal efficiencies in excess of 99 percent, but based on NSPS requirements and recent permit averages, such high removal efficiency is not necessary.

Exhibit 2-8 Environmental Targets for Pulverized Coal Cases

Pollutant	Environmental Target	NSPS Limit	Average of Recent Permits	Control Technology
NOx				
New SCPC	0.07 lb/MMBtu	1.0 lb/MWh (0.111 lb/MMBtu)	0.08 lb/MMBtu	LNB, OFA, SCR
Existing Plant	0.45 lb/MMBtu			OFA and ‘retro’ LNB
Existing Plant CO$_2$ Retrofit[1]	0.24 lb/MMBtu			New LNB, Improved OFA
SO$_2$				
New SCPC	0.119 lb/MMBtu	1.4 lb/MWh (0.156 lb/MMBtu)	0.16 lb/MMBtu	Dry lime-based spray dry absorber
Existing Plant	0.255 lb/MMBtu			Wet soda ash scrubber
Existing Plant CO$_2$ Retrofit[1]	0.017 lb/MMBtu			Upgraded wet soda ash scrubber
Particulate Matter				
New SCPC	0.0150 lb/MMBtu	0.0150 lb/MMBtu	0.017 lb/MMBtu	Fabric filter
Existing Plant	0.0270 lb/MMBtu			ESP
Existing Plant CO$_2$ Retrofit[1]	0.0270 lb/MMBtu			ESP
Mercury				
New SCPC	0.70 lb/TBtu	97 x 10^{-6} lb/MWh (11 lb/TBtu)	2.49 lb/TBtu	Co-benefit capture plus carbon injection
Existing Plant	6.00 lb/TBtu			Co-benefit capture
Existing Plant CO$_2$ Retrofit[1]	6.00 lb/TBtu			Co-benefit capture

[1] Both 1,100 lb/net-MWh and 90 percent CO$_2$ capture cases

LNB: Low NOx Burners
OFA: Over-fired Air
SCR: Selective Catalytic Reduction

The wet soda ash scrubber utilized in the existing subcritical retrofit PC case was assumed to be 85 percent efficient for (Case 7), which results in SO_2 emissions of 0.255 lb/MMBtu. SO_2 emissions for this technology are currently greater than NSPS limits. Should NSPS requirements become relevant, the wet soda ash scrubber would have to be modified to meet NSPS limits.

In the CO_2 capture cases, the Econamine system employs a polishing scrubber to reduce the flue gas SO_2 concentration to 10 ppmv entering the CO_2 absorber. This results in SO_2 emissions of 0.017 lb/MMBtu for the new SCPC and existing plant 90 percent CO_2 capture cases. In the 1,100 lb CO_2/net-MWh PC cases, the SO_2 emissions increase because a portion of the flue gas is bypassed around the Econamine system polishing scrubber. The SO_2 emissions at this capture level are 0.068 lb/MMBtu for the SCPC case (Case 5) and 0.054 lb/MMBtu for the subcritical retrofit PC case (Case 8).

3. **PM:** In new SCPC cases, a fabric filter will remove 99.97 percent of the entering particulate. In the existing subcritical PC cases, the ESP will remove 99.65 percent. There is an 80/20 split between fly ash and bottom ash in all PC cases. The result is the emission of 0.0150 lb/MMBtu for supercritical PC cases and 0.0270 lb/MMBtu for the existing subcritical PC cases. The SCPC technology meets NSPS and recent permit average requirements. PM emissions from the existing subcritical PC are currently greater than NSPS limits. Should NSPS requirements become relevant, the ESP would have to be replaced by a baghouse.

4. **Mercury:** EPA's documentation for their Integrated Planning Model (IPM) provides mercury emission modification factors (EMF) based on 190 combinations of boiler types and control technologies [17]. The EMF is simply one minus the removal efficiency. Based on the IPM estimates, mercury control was assumed to occur through 15 percent co-benefit capture for the fabric filter, dry FGD scrubber, and SCR in the new supercritical PC cases. Activated carbon injection provides an additional 90 percent reduction for a total Hg environmental target for the new SCPC of 0.70 lb/TBtu. In the subcritical PC plant, the co-benefit capture is assumed to be 16 percent with a wet FGD, a cold-side ESP, and no post-combustion NOx control. The estimated Hg emissions for the existing subcritical PC plant are 6.00 lb/TBtu. With the addition of SCR to the subcritical PC retrofit, the mercury emissions would be further reduced to 2.39 lb/TBtu because of the increased co-benefit capture.

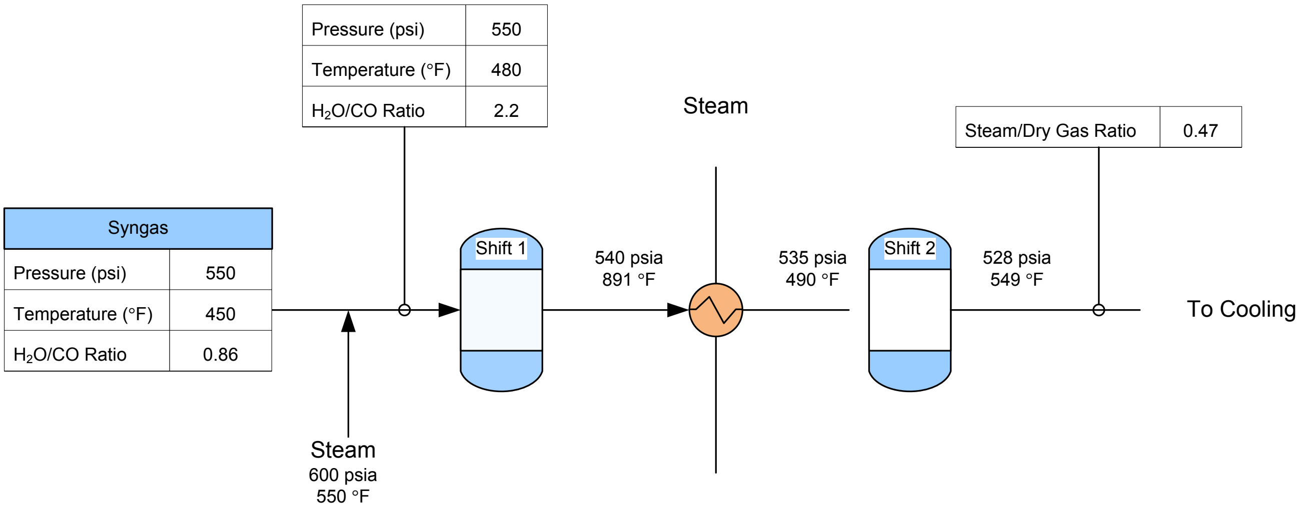
2.3.3 Carbon Dioxide

Carbon dioxide (CO_2) is not currently regulated nationally, but the California Public Utilities Commission adopted an interim GHG Emissions Performance Standard of 1,100 lb CO_2/net-MWh for carbon dioxide.

For the IGCC cases that have CO_2 capture, the emissions benchmarks are a nominal 90 percent overall carbon capture and an emissions limit equal to 1,100 lb CO_2/net-MWh. These are based on carbon input from the coal and excluding carbon that exits the gasifier with the slag. For the 90 percent capture case, two water gas shift (WGS) reactors were used with a Selexol CO_2 removal efficiency of 90.1 percent (based on a vendor quote and a given syngas CO_2

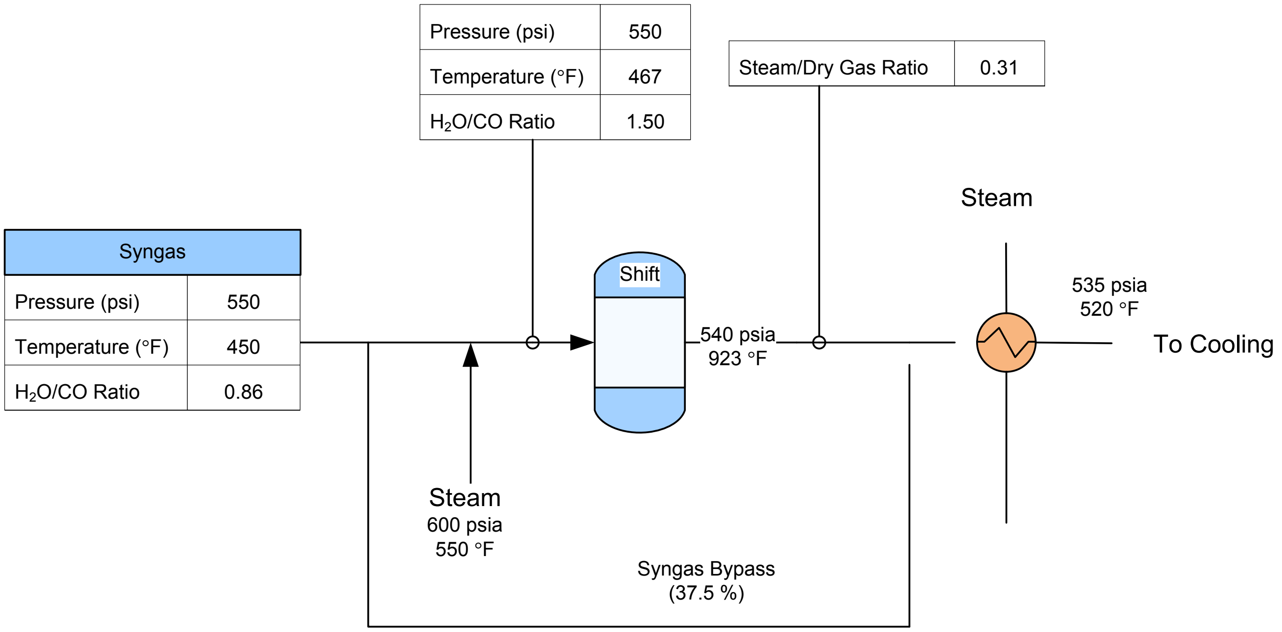
concentration). In addition, to achieve 90 percent CO_2 capture, shift steam had to be increased above the minimum value of 0.30 (steam: dry gas at shift outlet) to 0.47 in order to increase conversion to CO_2 as shown in Exhibit 2-9.

Exhibit 2-9 IGCC with 90 Percent CO_2 Capture WGS Process



For the IGCC case that meets the 1,100 lb CO_2/net-MWh emission standard, a partial flue gas bypass around a single water gas shift reactor was implemented and the shift steam was reduced to near the minimum value of 0.30:1 (steam: dry gas) as shown in Exhibit 2-10. To achieve the CO_2 emissions target of 1,100 lb CO_2/net-MWh, 50 percent removal was required.

Exhibit 2-10 IGCC with Partial WGS to Meet 1,100 lb/net-MWh CO_2 Emission Limit



For the SCPC cases that have CO_2 capture, it is assumed that all of the fuel carbon is converted to CO_2 in the flue gas. Ninety percent of the CO_2 entering the Econamine FG Plus unit from the FGD is subsequently captured. For the 1,100 lb CO_2/net-MWh cases, a partial flue gas bypass is implemented to reduce the amount of CO_2 entering the Econamine unit to achieve the desired emission limit.

The cost of CO_2 capture was calculated in two ways, the cost of CO_2 removed and the cost of CO_2 avoided, as illustrated in Equations 1 and 2, respectively. The cost of electricity in the CO_2 capture cases includes transport, storage and monitoring (TS&M) as well as capture and compression.

$$(1) \qquad Removal\ Cost = \frac{\{LCOE_{with\ removal} - LCOE_{w/o\ removal}\}\ \$/MWh}{\{CO_2\ removed\}\ tons/MWh}$$

$$(2) \qquad Avoided\ Cost = \frac{\{LCOE_{with\ removal} - LCOE_{w/o\ removal}\}\ \$/MWh}{\{Emissions_{w/o\ removal} - Emissions_{with\ removal}\}\ tons/MWh}$$

2.4 CO_2 TRANSPORT AND STORAGE

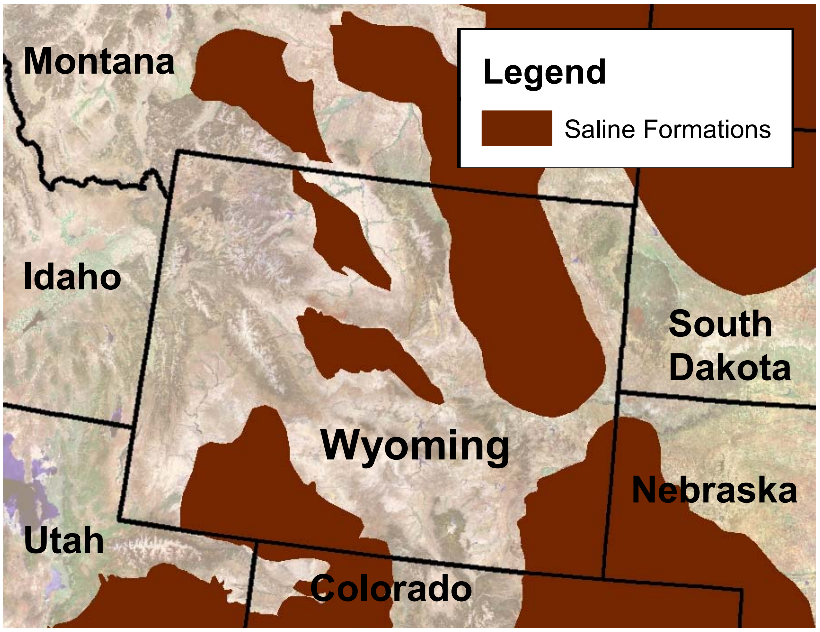
CO_2 is compressed to a pressure of 15.3 MPa (2,215 psia) in preparation for sequestration. The CO_2 product gas composition varies in the cases presented, but is expected to meet the specification described in Exhibit 2-11.

Exhibit 2-11 CO_2 Pipeline Specification

Parameter	Units	Parameter Value
Inlet Pressure	MPa (psia)	15.3 (2,215)
Outlet Pressure	MPa (psia)	10.4 (1,515)
Inlet Temperature	°C (°F)	26 (79)
N_2 Concentration	ppmv	< 300
O_2 Concentration	ppmv	< 40
Ar Concentration	ppmv	< 10

The CO_2 is transported 50 miles via pipeline to a geologic sequestration field for injection into a saline formation. Exhibit 2-12 shows the possible saline formations in Wyoming and the surrounding areas that could be used for CO_2 sequestration.

Exhibit 2-12 Saline Formations of Wyoming and Surrounding States



The CO_2 is transported and injected as a supercritical fluid in order to avoid two-phase flow and achieve maximum efficiency [18]. The pipeline is assumed to have an outlet pressure (above the supercritical pressure) of 10.4 MPa (1,515 psia) with no recompression along the way. Accordingly, CO_2 flow in the pipeline was modeled to determine the pipe diameter that results in a pressure drop of 4.8 MPa (700 psi) over a 50 mile pipeline length [19]. (Although not explored in this study, the use of boost compressors and a smaller pipeline diameter could possibly reduce capital costs for sufficiently long pipelines.) The diameter of the injection pipe will be of sufficient size that frictional losses during injection are minimal and no booster compression is required at the well-head in order to achieve an appropriate down-hole pressure.

The saline formation is at a depth of 4,055 ft and has a permeability of 22 millidarcy (a measure of permeability defined as roughly 10^{-12} Darcy) and formation pressure of 8.4 MPa (1,220 psig) [20]. This is considered an average storage site and requires roughly one injection well for each 10,320 short tons of CO_2 injected per day [20]. The assumed aquifer characteristics are tabulated in Exhibit 2-13.

Exhibit 2-13 Deep Saline Aquifer Specifications

Parameter	Units	Base Case
Pressure	MPa (psi)	8.4 (1,220)
Thickness	m (ft)	161 (530)
Depth	m (ft)	1,236 (4,055)
Permeability md		22

Parameter	Units	Base Case
Pipeline Distance	km (miles)	80 (50)
Injection Rate per Well	Tonne (ton) CO_2/day	9,360 (10,320)

2.5 CAPACITY FACTOR

This study assumes that each new plant would be dispatched any time it is available and would be capable of generating maximum capacity when online. Therefore, capacity factor and availability are assumed to be equal. The availability for PC cases was determined using the Generating Availability Data System (GADS) for the North American Electric Reliability Council [21]. Input from EPRI and their work on the CoalFleet for Tomorrow Initiative were used to set the IGCC case capacity factor.

NERC defines an equivalent availability factor (EAF), which is essentially a measure of plant capacity factor assuming there is always a demand for the output. The EAF accounts for planned and scheduled derated hours as well as seasonal derated hours. As such, the EAF matches this study’s definition of capacity factor.

The average EAF for pulverized coal-fired plants in the 400-599 MW size range was 84.9 percent in 2004 and averaged 83.9 percent from 2000-2004. Given that many plants of this size range are older, the EAF was rounded up to 85 percent and that value was used as the greenfield supercritical PC plant capacity factor. The BART analysis of the existing subcritical PC plant uses 90 percent for the capacity factor, but 85 percent was used in this study to be consistent with the greenfield supercritical PC cases.

EPRI examined the historical forced and scheduled outage times for IGCCs and concluded that the reliability factor (which looks at forced or unscheduled outage time only) for a single train IGCC (no spares) would be about 90 percent [22]. To get the availability factor, one has to deduct the scheduled outage time. In reality the scheduled outage time differs from gasifier technology-to-gasifier technology, but the differences are relatively small and would have minimal impact on the capacity factor, so for this study it was assumed to be constant at a 30-day planned outage per year (or two 15-day outages). The planned outage would amount to 8.2 percent of the year, so the availability factor would be (90 percent - 8.2 percent), or 81.2 percent.

There are four operating IGCC’s worldwide that use a solid feedstock and are primarily power producers (Polk, Wabash, Buggenum and Puertollano). A 2006 report by Higman et al. examined the reliability of these IGCC power generation units and concluded that typical annual on-stream times are around 80 percent [23]. The capacity factor would be somewhat less than the on-stream time since most plants operate at less than full load for some portion of the operating year. Given the results of the EPRI study and the Higman paper, a capacity factor of 80 percent was chosen for IGCC with no spare gasifier required.

The addition of CO_2 capture to each technology was assumed <u>not</u> to impact the capacity factor. This assumption was made to enable a comparison based on the impact of capital and variable operating costs only. Any reduction in assumed capacity factor would further increase the levelized cost of electricity (LCOE) for the CO_2 capture cases.

2.6 RAW WATER WITHDRAWAL AND CONSUMPTION

A water balance was performed for each case on the major water consumers in the process. The total water demand for each subsystem was determined. The internal recycle water available from various sources like boiler feedwater blowdown, moisture recovered from the coal in the drying process (IGCC cases only), and condensate from syngas was applied to offset the water demand. The difference between demand and recycle is raw water withdrawal.

In the greenfield cases, raw water makeup was assumed to be provided 50 percent by a publicly owned treatment works (POTW) and 50 percent from groundwater. In the existing subcritical PC cases, raw water is obtained from the Green River. Raw water withdrawal is defined as the water metered from a water source and used in the plant processes for any and all purposes, such as cooling tower makeup, boiler feedwater makeup, ash handling makeup, syngas humidification, and quench system makeup. Withdrawal represents the gross impact of the process on the water source.

Some water from the process can be treated and returned to the source, referred to as process discharge. The main source of process discharge is cooling tower blowdown with smaller amounts from the sour water stripper in the IGCC cases. It was assumed that 90 percent of the cooling tower blowdown could be returned to the source, and the remaining 10 percent would be sent to the ash ponds to evaporate. Similarly, 90 percent of the sour water stripper blowdown is recycled as process discharge and the balance is sent to the slag pile. The difference between raw water withdrawal and process discharge is raw water consumption and represents the net impact on the water source.

The largest consumer of raw water in all cases is cooling tower makeup. The IGCC and supercritical PC cases utilize a parallel cooling system with half of the turbine exhaust steam condensed in an air-cooled condenser and half in a water-cooled condenser. The subcritical PC retrofit cases utilize a water-cooled condenser only. The cooling water is provided by a mechanical draft, evaporative cooling tower, and all process blowdown streams were assumed to be treated and recycled to the cooling tower. The design ambient wet bulb temperature of 3°C (37°F) (Exhibit 2-1) was used to achieve a cooling water temperature of 9°C (48°F), using an approach of 6°C (11°F). The cooling water range was assumed to be 11°C (20°F). The cooling tower makeup rate was determined using the following [24]:

- Evaporative losses of 0.8 percent of the circulating water flow rate per 10°F of range
- Drift losses of 0.001 percent of the circulating water flow rate
- Blowdown losses were calculated as follows:
    - Blowdown Losses = Evaporative Losses / (Cycles of Concentration - 1)

Where cycles of concentration are a measure of water quality, and a mid-range value of 4 was chosen for this study.

The water balances presented in subsequent sections include the water demand of the major water consumers within the process, the amount provided by internal recycle, raw water withdrawal by difference, process discharge, and raw water consumption. The existing subcritical PC plant water balance was calculated using the same methodology as the greenfield cases.

2.7 COST ESTIMATING METHODOLOGY

The cost estimates for this project were derived from previous estimates on similar plant types and sizes. The original estimates were done for an ongoing DOE project by WorleyParsons Group Inc. (WorleyParsons) that included a Shell IGCC using PRB coal with and without CO_2 capture and a supercritical PC plant using PRB coal with and without CO_2 capture. WorleyParsons estimated the Total Plant Cost (TPC) and the Operation and Maintenance (O&M) costs for each technology. The estimates have an accuracy of ±30 percent.

The costing methodology used by WorleyParsons for the baseline estimates is described below. At the end of this section the methodology used to scale the WorleyParsons estimates is described.

WorleyParsons used an in-house database and conceptual estimating models for the capital cost and O&M cost estimates. Costs were further calibrated using a combination of adjusted vendor-furnished and actual cost data from recent design and design/build projects.

The capital costs for each cost account were reviewed by comparing individual accounts across each of the similar technologies to ensure an accurate representation of the relative cost differences between the cases and accounts. All capital costs are presented as “overnight costs” expressed in June 2007 dollars.

Capital costs are presented at the TPC level. TPC includes:

- Equipment (complete with initial chemical and catalyst loadings),
- Materials,
- Labor (direct and indirect),
- Engineering and construction management, and
- Contingencies (process and project).

Owner’s costs were subsequently calculated and added to the TPC, the result of which is Total Overnight Cost (TOC). Additionally, financing costs were estimated and added to TOC to provide Total As-Spent Cost (TASC). The levelized cost of electricity was calculated using TOC.

System Code-of-Accounts

The costs are grouped according to a process/system oriented code of accounts. This type of code-of-account structure has the advantage of grouping all reasonably allocable components of a system or process so they are included in the specific system account. (This would not be the case had a facility, area, or commodity account structure been chosen instead).

Non-CO_2 Capture Plant Maturity

The non-capture IGCC cases are based on commercial offerings; however, there have been very limited sales of these units so far. These non-CO_2-capture IGCC plant costs are less mature in the learning curve than PC plants, and the costs listed reflect the “next commercial offering” level of cost rather than mature nth-of-a-kind cost. Thus, each of these cases reflects the expected cost for the next commercial sale of each of these respective technologies.

CO_2 Removal Maturity

The pre-combustion CO_2 removal technology for the IGCC capture cases has a stronger commercial experience base than post-combustion technologies for PC plants. Pre-combustion CO_2 removal from syngas streams has been proven in chemical processes with similar conditions to that in IGCC plants, but has not been demonstrated in IGCC applications. While no commercial IGCC plant yet uses CO_2 removal technology in commercial service, there are currently IGCC plants with CO_2 capture well along in the planning stages.

While the post-combustion technology for the PC plants has been practiced at smaller scale, it has never been practiced at a scale equivalent to that required in this study. There are domestic amine-based CO_2 capture systems operating on coal-derived flue gas at scales ranging from 150-800 TPD [25]. Plants in this study will capture on average 11,500 TPD. Consequently the CO_2 capture cases are treated as first-of-a-kind (FOAK).

Contracting Strategy

The estimates are based on an Engineering/Procurement/Construction Management (EPCM) approach utilizing multiple subcontracts. This approach provides the Owner with greater control of the project, while minimizing, if not eliminating most of the risk premiums typically included in an Engineer/Procure/Construct (EPC) contract price.

In a traditional lump sum EPC contract, the Contractor assumes all risk for performance, schedule, and cost. However, as a result of current market conditions, EPC contractors appear more reluctant to assume that overall level of risk. Rather, the current trend appears to be a modified EPC approach where much of the risk remains with the Owner. Where Contractors are willing to accept the risk in EPC type lump-sum arrangements, it is reflected in the project cost. In today’s market, Contractor premiums for accepting these risks, particularly performance risk, can be substantial and increase the overall project costs dramatically.

The EPCM approach used as the basis for the estimates here is anticipated to be the most cost effective approach for the Owner. While the Owner retains the risks and absorbs higher project management costs, the risks become reduced with time, as there is better scope definition at the time of contract award(s).

Estimate Scope

The estimates represent a complete power plant facility on a generic site. Site-specific considerations such as unusual soil conditions, special seismic zone requirements, or unique local conditions such as accessibility, local regulatory requirements are not considered in the estimates.

The estimate boundary limit is defined as the total plant facility within the "fence line" including coal receiving and water supply system, but terminating at the high voltage side of the main power transformers. The single exception to the fence line limit is in the CO_2 capture cases where costs are included for TS&M of the CO_2.

Labor costs are based on Merit Shop (non-union), in a competitive bidding environment.

Capital Costs

WorleyParsons developed the capital cost estimates for each plant using the company's in-house database and conceptual estimating models for each of the specific technologies. This database and the respective models are maintained by WorleyParsons as part of a commercial power plant design base of experience for similar equipment in the company's range of power and process projects. A reference bottoms-up estimate for each major component provides the basis for the estimating models. This provides a basis for subsequent comparisons and easy modification when comparing between specific case-by-case variations.

Key equipment costs for each of the cases were calibrated to reflect recent quotations and/or purchase orders for other ongoing in-house power or process projects. These include, but are not limited to the following equipment:

- Pulverized Coal Boilers
- Combustion Turbine Generators
- Steam Turbine Generators
- Circulating Water Pumps and Drivers
- Cooling Towers
- Condensers
- Air Separation Units (partial)
- Main Transformers

Other key estimate considerations include the following:

- Labor costs are based on Midwest, Merit Shop using factors from PAS, Inc. [26]. PAS presents information for eight separate regions. Previous studies used a generic

Midwestern site, typical of Region 5 (IL, IN, MI, MN, OH, WI). The weighted average rate for Region 8 (CO, MT, ND, SD, UT, WY) is within less than one-half of one percent of that for Region 5. The difference is inconsequential so the same rates used in other NETL studies were maintained in this study.

- The estimates are based on a competitive bidding environment, with adequate skilled craft labor available locally.
- Labor is based on a 50-hour work-week (5-10s). No additional incentives such as per-diems or bonuses have been included to attract craft labor.
- While not included at this time, labor incentives may ultimately be required to attract and retain skilled labor depending on the amount of competing work in the region, and the availability of skilled craft in the area at the time the projects proceed to construction. Current indications are that regional craft shortages are likely over the next several years. The types and amounts of incentives will vary based on project location and timing relative to other work. The cost impact resulting from an inadequate local work force can be significant.
- The estimates are based on a greenfield site.
- The site is considered to be Seismic Zone 1, relatively level, and free from hazardous materials, archeological artifacts, or excessive rock. Soil conditions are considered adequate for spread footing foundations. The soil bearing capability is assumed adequate such that piling is not needed to support the foundation loads.
- Costs are limited to within the “fence line,” terminating at the high voltage side of the main power transformers with the exception of costs included for TS&M of CO_2 in all capture cases.
- Engineering and Construction Management were estimated as a percent of bare erected cost. These costs consist of all home office engineering and procurement services as well as field construction management costs. Site staffing generally includes a construction manager, resident engineer, scheduler, and personnel for project controls, document control, materials management, site safety and field inspection.
- All capital costs are presented as “Overnight Costs” in June 2007 dollars. Escalation to period-of-performance is specifically excluded.

Price Escalation

A significant change in power plant cost occurred in recent years due to the significant increases in the pricing of equipment and bulk materials. This estimate includes these increases. All vendor quotes used to develop these estimates were received within the last three years. The price escalation of vendor quotes incorporated a vendor survey of actual and projected pricing increases from 2004 through mid-2007 that WorleyParsons conducted for a recent project. The results of that survey were used to validate/recalibrate the corresponding escalation factors used in the conceptual estimating models.

Cross-comparisons

In all technology comparison studies, the relative differences in costs are often more significant than the absolute level of TPC. This requires cross-account comparison between technologies to review the consistency of the direction of the costs. As noted above, the capital costs were reviewed and compared across each of the similar technologies to ensure that a consistent representation of the relative cost differences is reflected in the estimates.

In performing such a comparison, it is important to reference the technical parameters for each specific item, as these are the basis for establishing the costs. Scope or assumption differences can quickly explain any apparent anomalies. There are a number of cases where differences in design philosophy occur. Some key examples are:

- The combustion turbines for the IGCC capture cases include an additional cost for firing a high hydrogen content fuel.
- The Shell gasifier syngas cooling configuration is different between the CO_2-capture and non-CO_2-capture cases, resulting in a significant differential in thermal duty between the syngas coolers for the two cases.

Exclusions

The capital cost estimate includes all anticipated costs for equipment and materials, installation labor, professional services (Engineering and Construction Management), and contingency. The following items are extremely project and site specific and are therefore excluded from the capital costs:

- Escalation to period-of-performance
- Owner's costs – these are accounted for separately and are described below.
- Site specific considerations – including but not limited to seismic zone, accessibility, local regulatory requirements, excessive rock, piles and laydown space
- Labor incentives in excess of a 5-day/10-hour work week
- Additional premiums associated with an EPC contracting approach

Contingency

Both the project contingency and process contingency costs represent costs that are expected to be spent in the development and execution of the project that are not yet fully reflected in the design. It is industry practice to include project contingency in the TPC to cover project uncertainty and the cost of any additional equipment that would result during detailed design. Likewise, the estimates include process contingency to cover the cost of any additional equipment that would be required as a result of continued technology development.

Project Contingency

Project contingencies were added to each of the capital accounts to cover project uncertainty and the cost of any additional equipment that could result from detailed design. The project contingencies represent costs that are expected to occur. Each bare erected cost account was evaluated against the level of estimate detail, field experience, and the basis for the equipment pricing to define project contingency.

The capital cost estimates associated with the plant designs in this study were derived from various sources which include prior conceptual designs and actual design and construction of both process and power plants.

The Association for the Advancement of Cost Engineering (AACE) International recognizes five classes of estimates. On the surface, the level of project definition of the cases evaluated in this study would appear to fall under an AACE International Class 5 Estimate, associated with less than 2 percent project definition, and based on preliminary design methodology. However, the study cases are actually more in line with the AACE International Class 4 Estimate, which is associated with equipment factoring, parametric modeling, historical relationship factors, and broad unit cost data.

Based on the AACE International contingency guidelines as presented in NETL’s "Quality Guidelines for Energy System Studies" it would appear that the overall project contingencies for the subject cases should be in the range of 30 to 40 percent [27]. However, such contingencies are believed to be too high when the basis for the cost numbers is considered. The costs have been extrapolated from an extensive data base of project costs (estimated, quoted, and actual), based on both conceptual and detailed designs for the various technologies. This information has been used to calibrate the costs in the current studies, thus improving the quality of the overall estimates. As such, the overall project contingencies should be more in the range of 15 to 20 percent with the capture cases being higher than the non-capture cases.

Process Contingency

Process contingency is intended to compensate for uncertainties arising as a result of the state of technology development. Process contingencies have been applied to the estimates as follows:

- Gasifiers and Syngas Coolers – 15 percent on all cases – next-generation commercial offering and integration with the power island
- Two Stage Selexol – 20 percent on all capture cases - unproven technology at commercial scale in IGCC service
- CO_2 Removal System – 20 percent on all PC capture cases – post-combustion process unproven at commercial scale for power plant applications
- Mercury Removal – 5 percent on all cases – minimal commercial scale experience in IGCC applications
- Combustion Turbine Generator – 5 percent on all non-capture cases – syngas firing and ASU integration; 10 percent on all capture cases – high hydrogen firing.
- Instrumentation and Controls – 5 percent on all accounts

AACE International provides standards for process contingency relative to technology status; from commercial technology at 0 to 5 percent to new technology with little or no test data at 40 percent. The process contingencies as applied in this study are consistent with the AACE International standards.

All contingencies included in the TPC, both project and process, represent costs that are expected to be spent in the development and execution of the project.

Operations and Maintenance (O&M)

The production costs or operating costs and related maintenance expenses (O&M) pertain to those charges associated with operating and maintaining the power plants over their expected life. These costs include:

- Operating labor
- Maintenance – material and labor
- Administrative and support labor
- Consumables
- Fuel
- Waste disposal
- Co-product or by-product credit (that is, a negative cost for any by-products sold)

There are two components of O&M costs; fixed O&M, which is independent of power generation, and variable O&M, which is proportional to power generation.

Operating Labor

Operating labor cost was determined based on the number of operators required for each specific case. The average base labor rate used to determine annual cost is $34.65/hr [26]. The associated labor burden is estimated at 30 percent of the base labor rate.

Maintenance Material and Labor

Maintenance cost was evaluated on the basis of relationships of maintenance cost to initial capital cost. This represents a weighted analysis in which the individual cost relationships were considered for each major plant component or section. The exception to this is the maintenance cost for the combustion turbines, which is calculated as a function of operating hours.

Administrative and Support Labor

Labor administration and overhead charges are assessed at rate of 25 percent of the burdened operation and maintenance labor.

Consumables

The cost of consumables, including fuel, was determined on the basis of individual rates of consumption, the unit cost of each specific consumable commodity, and the plant annual operating hours.

Quantities for major consumables such as fuel were taken from technology-specific heat and mass balance diagrams developed for each plant application. Other consumables were evaluated on the basis of the quantity required using reference data.

The quantities for initial fills and daily consumables were calculated on a 100 percent operating capacity basis. The annual cost for the daily consumables was then adjusted to incorporate the annual plant operating basis, or capacity factor.

Initial fills of the consumables, fuels and chemicals, are different from the initial chemical loadings, which are included with the equipment pricing in the capital cost.

Waste Disposal

Waste quantities and disposal costs were determined similarly to the consumables. The slag from the IGCC cases is considered a waste with a disposal cost of $16.23/ton. The carbon used for mercury control is considered a hazardous waste with disposal cost of $834/ton.

Co-Products and By-Products (Other than CO_2)

IGCC Cases

By-product quantities were also determined similarly to the consumables. However, due to the variable marketability of these by-products, specifically sulfur, no credit was taken for its potential saleable value. Nor were any of the cases penalized for their potential disposal cost. That is, for this evaluation, it is assumed that the by-product or co-product value simply offset disposal costs, for a net zero in operating costs. Similarly slag is a potential by-product in certain markets and would have potential marketability. However, slag is also considered a waste in this study with a concomitant disposal cost.

PC Cases

Due to the variable marketability of these by-products (bottom ash and fly ash co-mingled with flue gas desulfurization (FGD) products) no credit was taken for potential saleable value.

It should be noted that by-product credits and/or disposal costs could potentially be an additional determining factor in the choice of technology for some companies and in selecting some sites. A high local value of the product can establish whether or not added capital should be included in the plant costs to produce a particular co-product. Ash is a potential by-product in certain markets and would have potential marketability. However, since in these cases the fly ash contains mercury from carbon injection and FGD byproducts, it was assumed to be a waste material rather than a saleable byproduct. Similarly the bottom ash was considered a waste with both materials having a concomitant disposal cost of $17.89/tonne ($16.23/ton).

Owner's Costs

The owner's costs included in the TOC cost estimate are shown in Exhibit 2-14.

Exhibit 2-14 Owner's Costs Included in TOC

Owner's Cost	***Comprised of***
Preproduction Costs	• *6 months operating, maintenance, and administrative & support labor* • *1 month maintenance materials* • *1 month non-fuel consumables* • *1 month of waste disposal costs* • *25% of one month's fuel cost @ 100% capacity factor* • *2% of TPC*
Inventory Capital	• *60 day supply of fuel and consumables @100% capacity factor* • *0.5% of TPC (spare parts)*
Land	• *$3,000/acre (300 acres for greenfield IGCC and PC)*
Financing Costs	• *2.7% of TPC*
Other Owner's Costs	• *15% of TPC*
Initial Cost for Catalyst and Chemicals	• *All initial fills not included in BEC*
Prepaid Royalties	• *Not included in owner's costs (included with BEC)*
Taxes & Insurance	• *2% of TPC (Fixed O&M cost)*
AFUDC and Escalation	• *Varies based on levelization period and financing scenario* • *33-yr IOU high risk: TASC = TOC *1.078* • *33-yr IOU low risk: TASC = TOC * 1.075* • *35-yr IOU high risk: TASC = TOC * 1.140* • *35-yr IOU low risk: TASC = TOC * 1.134*

The category labeled "Other Owner's Costs" includes the following:

- *Preliminary feasibility studies, including a Front-End Engineering Design (FEED) study*
- *Economic development (costs for incentivizing local collaboration and support)*
- *Construction and/or improvement of roads and/or railroad spurs outside of site boundary.*
- *Legal fees*
- *Permitting costs*
- *Owner's engineering (staff paid by owner to give third-party advice and to help the owner oversee/evaluate the work of the EPC contractor and other contractors)*

- *Owner's contingency: sometimes called "management reserve", these are funds to cover costs relating to delayed startup, fluctuations in equipment costs, unplanned labor incentives in excess of a five-day/ten-hour-per-day work week*

Cost items excluded from "Other Owner's Costs" include:

- *EPC Risk Premiums: Costs estimates are based on an Engineering Procurement Construction Management (EPCM) approach utilizing multiple subcontracts, in which the owner assumes project risks for performance, schedule and cost. This approach provides the owner with greater control of the project, while minimizing, if not eliminating most of the risk premiums typically included in a lump-sum, "turnkey" Engineer/Procure/Construct (EPC) contract, under which the EPC contractor assumes some or all of the project risks. The EPCM approach used as the basis for the estimates here is anticipated to be the most cost effective approach for the owner.*
- *Transmission interconnection: the cost of interconnecting with power transmission infrastructure beyond the plant busbar.*
- *Taxes on capital costs: all capital costs are assumed to be exempt from state and local taxes.*
- *Unusual site improvements: normal costs associated with improvements to the plant site are included in the bare erected cost, assuming that the site is level and requires no environmental remediation. Unusual costs associated with the following design parameters are excluded: flood plain considerations, existing soil/site conditions, water discharges and reuse, rainfall/snowfall criteria, seismic design, buildings/enclosures, fire protection, local code height requirements, noise regulations.*

CO_2 Transport, Storage and Monitoring

For those cases that feature CO_2 capture, the capital and operating costs for CO_2 TS&M were independently estimated by NETL. Those costs were converted to a levelized cost of electricity (LCOE) and combined with the plant capital and operating costs to produce an overall LCOE.

The transport and storage (T&S) capital and operating costs were assessed using metrics published in a DOE sponsored report entitled *Economic Evaluation of CO_2 Storage and Sink Enhancement Options* [28]. These costs were escalated from the 1999-year dollars described in the report to June 2007-year dollars using cost indices appropriate to that cost type. Capital costs were escalated using the Chemical Engineering Plant Cost Index Report and operating costs were escalated using the U.S. Bureau of Labor Statistics (BLS) Producer Price Indices (PPI) for the oil and gas industry.

Capital costs were levelized over a 30-year period and include both a 30 percent process contingency factor and a 20 percent project contingency factor in accordance with NETL's Systems Analysis Guidelines [29].

T&S costs are also assessed in terms of removed or avoided emissions cost, which requires power plant specific information such as plant efficiency, capacity factor, and emission rates.

Monitoring costs were evaluated based on the methodology set forth in the IEA Greenhouse Gas R&D Programme's *Overview of Monitoring Projects for Geologic Storage Projects* report [30]. In this scenario, operational monitoring of the CO_2 plume occurs over thirty years and closure monitoring occurs for the following fifty years (for a total of eighty years). Operational and closure monitoring costs are assumed to be proportional to the plume size plus a fixed cost, with closure monitoring costs evaluated at half the value of the operational costs. The present value of the life-cycle costs is assessed at a 10 percent discount rate and a capital fund is set up to pay for these costs over the eighty year monitoring cycle.

High pressure (2,200 psig) CO_2 is provided at the power plant gate and is transported via pipeline to a geologic storage site where it can be safely sequestered. It is transported and injected as a supercritical fluid in order to avoid two-phase flow and achieve maximum efficiency [28]. A minimum pipeline outlet pressure of 1,500 psig is utilized in order to ensure the CO_2 exiting the pipeline is supercritical and the pipeline is sized such that no recompression stations are needed. Utilizing this large pressure drop also minimizes the pipeline diameter required, and therefore transport capital cost.

The storage site evaluated is a saline aquifer at a depth of 4,055 feet with a permeability of 22 md and down-hole pressure of 1,220 psig [28] as shown in Exhibit 2-13. This is considered an average storage site and requires roughly one injection well for each 10,300 tons of CO_2 injected per day [28].

Exhibit 2-15 and Exhibit 2-16 detail the T&S cost metrics for the deep, saline aquifer described above. Transport capital costs are directly dependent on both pipeline length and diameter and constitute a significant portion of the overall transport, storage, and monitoring costs. Specific costs will be site specific based on right-of-way, topography, and other issues, but in this study the basis costs outlined in this section will be used. Costs from the *Economic Evaluation of CO_2 Storage and Sink Enhancement Options* were escalated from $33,000/inch-Diameter/mile in 1999-year dollars to $47,175/inch-Diameter/mile in June-2007 dollars using the Chemical Engineering Plant Cost Index for piping, valves and fittings.

Exhibit 2-15 Transport (Pipeline) Costs

Cost Type	Units	Cost
Capital	$/inch-Diameter/mile	$47,175
Fixed O&M	$/mile/year	$8,350

The order of magnitude of this cost appears to be valid based on a recent testimony from Ronald T. Evans, Senior Vice President of Denbury Resources, Inc. to the U.S. Senate Committee on Energy and Natural Resources. In his testimony, Mr. Evans states that pipeline costs have dramatically increased in recent years and of the three CO_2 pipelines Denbury has constructed in recent years, the costs have ranged from $30,000//inch-Diameter/mile in 2006, $55,000/inch-Diameter/mile in 2007 and an approximate $100,000/inch-Diameter/mile for a planned pipeline

[31]. With regards to the latter \$100,000/inch-Diameter/mile pipeline, he states that issues such as route obstacles and terrain inflate the cost of that particular pipeline. However, it provides a data point that shows the \$47,175/inch-Diameter/mile figure used in this study is a reasonable cost metric [31].

The fixed O&M costs related to transport are inclusive of pipeline maintenance and monitoring and constitute a large portion of the combined transport and storage costs. These costs were escalated using the Support Activities for Oil and Gas Operations BLS PPI [32]. No variable O&M costs were assessed [28].

Storage costs include initial site assessment, injection wells, and associated injection well equipment. The site assessment cost is a fixed cost and was escalated using the Drilling Oil and Gas Wells BLS PPI [32].

Exhibit 2-16 Geological Storage Costs

Cost Type	Units	Cost
Capital		
Initial Site Assessment	\$	\$4,931,547
Injection Wells	\$/injection well $(see\ formula)^{1,2}$	$\$189{,}242 \times e^{0.0008 \times well-depth}$
Injection Equipment	\$/injection well $(see\ formula)^{2}$	$\$92{,}916 \times \left(\frac{7{,}389}{280 \times \#\ of\ injection\ wells} \right)^{0.5}$
O&M		
Normal Daily Expenses (Fixed O&M)	\$/injection well	\$11,086
Consumables (Variable O&M)	\$/injection well	\$29,619
Surface Maintenance (Fixed O&M)	*see formula*	$\$22{,}504 \times \left(\frac{7{,}389}{280 \times \#\ of\ injection\ wells} \right)^{0.5}$

Subsurface Maintenance **(Fixed O&M)**	\$/ft-depth/inject. well	\$2.07

[1] The units for the "well depth" term in the formula are meters of depth.
[2] The formulas at right describe the cost per injection well and in each case the number of injection wells should be multiplied the formula in order to determine the overall capital cost.

The injection well and well equipment costs are a function of the number of wells. The number of injection wells is largely determined by reservoir characteristics such as Permeability, Downhole Injection Pressure Differential, and Thickness which can result in significantly different storage costs. The Downhole Injection Pressure Differential is the difference between the reservoir pressure and the CO_2 pressure at the bottom of the well hole. The pressure differential can be changed by manipulating the injection pressure. These costs were evaluated based on what is considered to be an average storage site, as described in Table 1.

The storage fixed O&M costs consist of Normal Daily Expenses, Surface Maintenance, and Subsurface Maintenance costs, with Surface Maintenance comprising the largest portion of costs. Consumables represent the only variable O&M cost. All storage O&M Costs were escalated using the Support Activities for Oil and Gas Operations BLS [32].

Levelized Cost of Electricity

The revenue requirement method of performing an economic analysis of a prospective power plant has been widely used in the electric utility industry. This method permits the incorporation of the various dissimilar components for a potential new plant into a single value that can be compared to various alternatives. The revenue requirement figure-of-merit in this report is cost of electricity (COE) levelized over a 30 year period and expressed in \$/MWh (numerically equivalent to mills/kWh). The 30-year LCOE was calculated using a simplified model derived from the NETL Power Systems Financial Model [33].

The equation used to calculate LCOE is as follows:

$$LCOE_P = \frac{(CCF_P)(TPC) + (LF_P)[(OC_{F1}) + (OC_{F2}) + \ldots] + (CF)(LF_P)[(OC_{V1}) + (OC_{V2}) + \ldots]}{(CF)(MWh)}$$

where

$LCOE_P$ = levelized cost of electricity over P years, \$/MWh

P = levelization period (e.g., 10, 20 or 30 years)

CCF_P = capital charge factor for a levelization period of P years

TPC = total plant cost, \$

LF = levelization factor

OC_{Fn} = category n fixed operating cost for the initial year of operation (but expressed in "first-year-of-construction" year dollars)

CF = plant capacity factor

OC_{Vn} = category n variable operating cost at 100 percent capacity factor for the initial year of operation (but expressed in "first-year-of-construction" year dollars)

MWh = annual net megawatt-hours of power generated at 100 percent capacity factor

All costs are expressed in June 2007 dollars, and the resulting LCOE is also expressed in June 2007 year dollars.

In CO_2 capture cases, the LCOE for TS&M costs was added to the LCOE calculated using the above equation to generate a total cost including CO_2 capture, sequestration and subsequent monitoring.

Although their useful life is usually well in excess of thirty years, a thirty-year levelization period is typically used for large energy conversion plants and is the levelization period used in this study.

The technologies modeled in this study were categorized as investor owned utility (IOU) high risk except for the SC PC and subcritical PC non-capture cases, which were categorized as low risk. The resulting capital charge factor and levelization factors are shown in Exhibit 2-17. The levelization factors assume a nominal 3 percent escalation for all cost categories.

Exhibit 2-17 Economic Parameters for LCOE Calculation

	High Risk (5 year construction period)	***Low Risk (5 year construction period)***	***High Risk (3 year construction period)***
Capital Charge Factor	*0.1773*	*0.1691*	*0.1567*
General Levelization Factor	*1.443*	*1.4299*	*1.4101*

The economic assumptions used to derive the capital charge factors are shown in Exhibit 2-18. The difference between the high risk and low risk categories is manifested in the debt-to-equity ratio and the weighted cost of capital. The values used to generate the capital charge factors and levelization factors in this study are shown in Exhibit 2-19.

Exhibit 2-18 Parameter Assumptions for Capital Charge Factors

Parameter	*Value*
Income Tax Rate	*38% (Effective 34% Federal, 6% State)*
Repayment Term of Debt	*15 years*
Grace Period on Debt Repayment	*0 years*
Debt Reserve Fund	*None*
Capital Depreciation	*20 years, 150% declining balance*
Working Capital	*zero for all parameters*
Plant Economic Life	*30 years*
Investment Tax Credit	*0%*
Tax Holiday	*0 years*
All other additional capital costs ($)	*0*
Capital Cost Escalation During Construction (nominal annual rate)	*3.6%*[1]
Construction Duration	*5 years (greenfield) / 3 years (retrofit)*

[1] A nominal average annual rate of 3.6 percent is assumed for escalation of capital costs during construction. This rate is equivalent to the nominal average annual escalation rate for process plant construction costs between 1947 and 2008 according to the Chemical Engineering Plant Cost Index.

Exhibit 2-19 Financial Structure for Investor Owned Utility High and Low Risk Projects

Type of Security	***% of Total***	***Current (Nominal) Dollar Cost***	***Weighted Current (Nominal) Cost***	***After Tax Weighted Cost of Capital***
Low Risk				
Debt	*50*	*4.5%*	*2.25%*	
Equity	*50*	*12%*	*6%*	
Total			*8.25%*	*7.39%*
High Risk				
Debt	*45*	*5.5%*	*2.475%*	
Equity	*55*	*12%*	*6.6%*	
Total			*9.075%*	*8.13%*

Cost Scaling Procedures

The WorleyParsons estimates were scaled for this study as described below.

Total Plant Cost

Each cost subaccount was scaled using an appropriate process parameter and a scaling exponent derived from the WorleyParsons baseline estimates. For example, each Coal Handling subaccount was scaled based on coal feed rate using an exponent of 0.62 as follows:

Scaled Cost = Reference Cost x (Scaled coal feed rate / Reference coal feed rate)$^{0.62}$

In total, 25 process parameters were used to scale the IGCC costs, 15 parameters were used to scale the greenfield SC PC costs and 7 parameters were required to scale the subcritical PC plant retrofit costs. Additional cost data for the subcritical plant was derived from the recent CH2MHill BART analysis of Unit 4 [8].

The TPC for Case 7, the existing subcritical PC plant, was assumed to be zero. The TPC for the CO_2 retrofit cases (8 and 9) included the Econamine FG Plus process and ancillary components that were scaled based on incremental process requirements above existing plant capacity. For example, in Case 8 the circulating water flow rate requirement increased by 81,000 gpm over the current plant capacity. The cost accounts related to the circulating water flow (circulating water pumps, circulating water system auxiliaries, circulating water piping and component cooling water systems and circulating water systems and foundations) were scaled from the reference estimate based on the incremental flow requirement and the appropriate scaling exponent.

The CH2MHill BART analysis provided costs for the new low NOx burners and the required upgrades to the flue gas desulfurization system. These costs were used directly in subaccount 4.2 (LNB’s and OFA) and subaccount 5.1 (Absorber Vessel and Accessories) [8]. The SCR costs were also taken from the BART analysis in the sensitivity case that assumed NSR would be activated.

O & M Costs

The O&M costs for the greenfield IGCC and SC PC cases were calculated using the same staffing requirements, labor rates, labor burdens, overhead charges, waste disposal costs and commodity unit costs as in the reference cases estimated previously by WorleyParsons. The maintenance labor and material costs were calculated by maintaining the same percentage of bare erected cost as used in the reference estimates.

The existing subcritical PC retrofit plant O&M costs were obtained from Global Energy Decisions’ Energy Velocity Database [34]. The O&M costs represent the marginal cost of electricity exclusive of any capital charges. The database provided the fuel component of the O&M costs and the total O&M costs. By difference the total of the variable and fixed O&M costs was calculated. The magnitude of the fixed O&M costs indicated that property taxes and insurance were excluded. To be consistent with the greenfield cases, an estimate of taxes and insurance was made and applied to the fixed operating cost. The estimate was based on 2 percent of the TPC, and the TPC was estimated by multiplying the ratio of the gross power output to the 0.7 power using the corresponding greenfield PC case.

The existing subcritical PC retrofit O&M costs in the CO_2 retrofit cases include the baseline costs from the Energy Velocity Database plus the additional costs incurred from retrofit of the CO_2 capture technology. The additional O&M costs include the following:

- One additional skilled operator and 1.3 additional operators (represents the delta between capture and non-capture in the SC PC cases)
- Maintenance labor and maintenance materials calculated as a percentage of the bare erected cost of the CO_2 capture technology and ancillary equipment
- Additional raw water makeup at a cost of $1.22/1000 gallons (obtained from the BART analysis)
- Additional water treatment chemicals estimated at the same relative makeup rate as the SC PC cases (on the incremental makeup water only)
- Makeup chemicals required by the Econamine FG Plus system, including amine solvent, sodium hydroxide, sulfuric acid, activated carbon and corrosion inhibitor at the same unit costs as used in the SC PC cases
- Incremental soda ash required after the FGD upgrade at a cost of $80/ton (obtained from the BART analysis)
- Incremental FGD waste disposal at a cost of $24.33/ton (obtained from the BART analysis)

The addition of CO_2 capture to an existing plant results in a de-rating of the plant output because of extraction steam required to regenerate the solvent and because of the additional auxiliary load from the CO_2 capture and compression process. In this analysis it was assumed that the plant would simply operate with a reduced net output. Alternatively, the plant could purchase power to compensate for the de-rated capacity. However, that option was not investigated in this study.

The IGCC plants also experience a decrease in net power output with CO_2 capture because of the fixed combustion turbine output constraint. However, the greenfield IGCC plants could add an additional train of gasification and a third combustion turbine if additional output is required with minimal impact to the cost of electricity. Thus makeup electricity cost was also not considered for the IGCC plant cases.

3. IGCC POWER PLANTS

Three IGCC power plant configurations were evaluated and the results are presented in this section. Each design is based on a market-ready technology that is assumed to be commercially available to support startup in 2015.

The three cases are based on the Shell gasifier using Montana Rosebud PRB coal, with and without CO_2 capture. As discussed in Section 1, the net output for the three cases varies because of the constraint imposed by the fixed gas turbine output, the site elevation, and the high auxiliary loads imparted by the CO_2 capture process.

The combustion turbine is based on an advanced F-class design. The HRSG/steam turbine cycle is 12.4 MPa/564°C/564°C (1800 psig/1048°F/1048°F) for the non-CO_2 capture case; 12.4 MPa/549°C/549°C (1800 psig/1020°F/1020°F) for the partial CO_2 capture case; and 12.4 MPa/536°C/536°C (1800 psig/996°F/996°F) for the 90 percent CO_2 capture case. The capture cases have a lower main and reheat steam temperature primarily because the turbine inlet temperature is reduced to allow for a parts life equivalent to NGCC operation with a high-hydrogen content fuel, which results in a lower turbine exhaust temperature. The effect is more pronounced in the 90 percent capture case than the 1,100 lb CO_2/net-MWh capture case. The combustion turbine output is also de-rated from ISO conditions because of operating at altitude at the location used in this study.

The evaluation scope included developing heat and mass balances and estimating plant performance. Equipment lists were developed for each design to support plant capital and operating cost estimates. The evaluation basis details, including site ambient conditions, fuel composition and environmental targets, were provided in Section 2. Section 3.1 covers general information that is common to all IGCC cases, and case specific information is subsequently presented in Section 4.

3.1 COMMON PROCESS AREAS

The three Shell IGCC cases have process areas which are common to each plant configuration such as coal receiving and storage, coal drying, oxygen supply, gas cleanup, and power generation. As detailed descriptions of these process areas for each case would be burdensome and repetitious, they are presented in this section for general background information. Where there is case-specific performance information, the performance features are presented in the relevant case sections.

3.1.1 Coal Receiving and Storage

The function of the Coal Receiving and Storage system is to convey, prepare, and store the coal delivered to the plant. The scope of the system is from the minemouth up to and including the slide gate valves at the outlet of the coal storage silos. Coal receiving and storage is identical for all three IGCC cases.

Operation Description – Coal is delivered to the site by conveyors from the nearby minemouth. Two conveyors with an intermediate transfer tower are assumed to convey the coal to the coal stacker, which transfer the coal to either the long-term storage pile or to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron and then to the reclaim pile.

The reclaimer loads the coal into two vibratory feeders located in the reclaim hopper under the pile. The feeders transfer the coal onto a belt conveyor that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3 cm x 0 (1¼" x 0) by the crusher. A conveyor then transfers the coal to a transfer tower. In the transfer tower the coal is routed to the tripper, which loads the coal into one of three silos. Two sampling systems are supplied: the as-received sampling system and the as-fired sampling system. Data from the analyses are used to support the reliable and efficient operation of the plant.

3.1.2 Coal Drying

Reduction in coal moisture content improves the efficiency of dry-feed gasifiers, but is also required for materials handling reasons. Coal moisture consists of two components, surface moisture and inherent moisture. Low rank coals have higher inherent moisture content and total moisture content than bituminous and other high rank coals. It is necessary to reduce most, if not all, of the surface moisture for coal transport properties to be acceptable.

In a recent GTC paper, Shell examined the WTA process for drying low rank coals and considered two cases [35]:

1) Case 1: Lignite coal dried from 53 to 12 percent
2) Case 2: Subbituminous coal dried from 30 to 6 percent

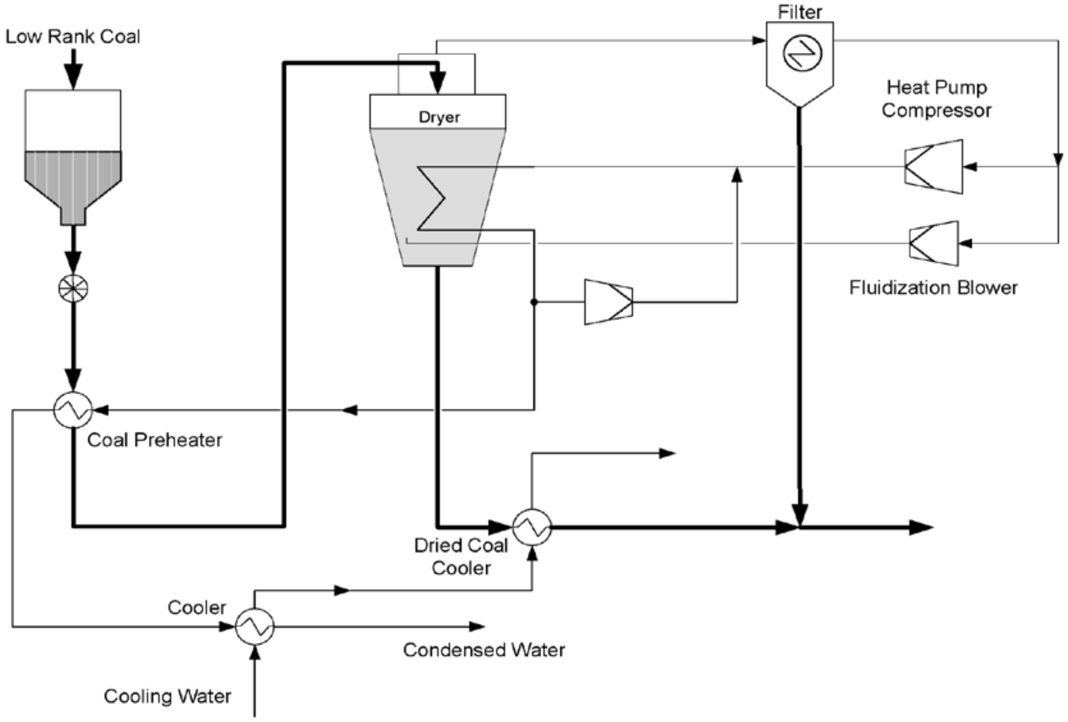
In personal correspondence with Shell, they indicated the moisture content of the coal after drying should be 3-14 percent depending on coal type [36]. EPRI and IEA recently performed studies that included the Shell gasifier using lignite coal that used a design moisture content of 5 percent entering the gasifier [37, 38].

For the Shell IGCC cases it is assumed that the subbituminous coal is dried to 6 percent moisture. This is consistent with the Shell GTC presentation and in the range suggested by the personal correspondence with Shell.

The WTA coal drying system was used in this study because of its ability to recover the water from the coal in liquid state for use in the process and the fact that syngas is not used to provide heat for drying. In conventional dryers, the water is mixed with the heating gas and discharged to atmosphere as vapor. Recovery of the coal moisture in a liquid state results in a sizable electric auxiliary load.

The ‘closed’ WTA process has been demonstrated at pilot scale. Plans for a commercial demonstration of an ‘open’ version of the process have been delayed. In spite of the uncertainty of the commercial demonstration, the potential benefit of the technology was viewed to be significant enough to use the ‘closed version’ of the process in this study. A process schematic is shown in Exhibit 3-1.

Exhibit 3-1 WTA Process Schematic



3.1.3 Air Separation Unit Choice and Integration

In order to economically and efficiently support IGCC projects, air separation equipment has been modified and improved in response to production requirements and the consistent need to increase single train output. “Elevated pressure” air separation designs have been implemented that result in distillation column operating pressures that are about twice as high as traditional plants. In this study, the main air compressor discharge pressure was set at 1.3 MPa (190 psia) compared to a traditional ASU plant operating pressure of about 0.7 MPa (105 psia) [39]. For IGCC designs, the elevated pressure ASU process minimizes power consumption and decreases the size of some of the equipment items. When the air supply to the ASU is integrated with the gas turbine, the ASU operates at or near the supply pressure from the gas turbine’s air compressor.

Residual Nitrogen Injection

The residual nitrogen that is available after gasifier oxygen and nitrogen requirements have been met is often compressed and sent to the gas turbine. Since all product streams are being compressed, the ASU air feed pressure is optimized to reduce the total power consumption and to provide a good match with available compressor frame sizes.

Increasing the diluent flow to the gas turbine by injecting residual nitrogen from the ASU can have a number of benefits, depending on the design of the gas turbine:

- Increased diluent increases mass flow through the turbine, thus increasing the power output of the gas turbine while maintaining optimum firing temperatures for syngas operation. This is particularly beneficial for locations where the ambient temperature and/or elevation are high and the gas turbine would normally operate at reduced output.
- By mixing with the syngas or by being injected directly into the combustor, the diluent nitrogen lowers the firing temperature (relative to natural gas) and reduces the formation of NOx.
- In this study, the ASU nitrogen product was used as the primary diluent with a design target of reducing the syngas lower heating value (LHV) to 4.3-4.8 MJ/Nm^3 (115-129 Btu/scf). If the amount of available nitrogen was not sufficient to meet this target, additional dilution was provided through syngas humidification, and if still more dilution was required, the third option was steam injection. For the three Shell IGCC cases, nitrogen dilution was sufficient in the capture cases and humidification was required for the non-capture case.

Air Integration

Integration between the ASU and the combustion turbine can be practiced by extracting some, or all, of the ASU's air requirement from the gas turbine. Medium Btu syngas streams result in a higher mass flow than natural gas to provide the same heat content to the gas turbine. Some gas turbine designs may need to extract air to maintain stable compressor or turbine operation in response to increased fuel flow rates. Other gas turbines may balance air extraction against injection of all of the available nitrogen from the ASU. The amount of air extracted can also be varied as the ambient temperature changes at a given site to optimize year-round performance.

An important aspect of air-integrated designs is the need to efficiently recover the heat of compression contained in the air extracted from the gas turbine. Extraction air temperature is normally in the range 399 - 454°C (750 - 850°F), and must be cooled to the last stage main air compressor discharge temperature prior to admission to the ASU. High-level recovery from the extracted air occurs by transferring heat to the nitrogen stream to be injected into the gas turbine with a gas-to-gas heat exchanger.

Elevated Pressure ASU Experience in Gasification

The Buggenum, Netherlands unit built for Demkolec was the first elevated-pressure, fully integrated ASU to be constructed. It was designed to produce up to 1,796 tonnes/day (1,980 TPD) of 95 percent purity oxygen for a Shell coal-based gasification unit that fuels a Siemens V94.2 gas turbine. In normal operation at the Buggenum plant the ASU receives all of its air supply from and sends all residual nitrogen to the gas turbine.

The Polk County, Florida ASU for the Tampa Electric IGCC is also an elevated-pressure, 95 percent purity oxygen design that provides 1,832 tonnes/day (2,020 TPD) of oxygen to a GEE coal-based gasification unit, which fuels a General Electric 7FA gas turbine. All of the nitrogen produced in the ASU is used in the gas turbine. The original design did not allow for air extraction from the combustion turbine. After a combustion turbine air compressor failure in January, 2005, a modification was made to allow air extraction which in turn eliminated a bottleneck in ASU capacity and increased overall power output [40].

ASU Basis

For this study, air integration is used for the non-carbon capture case only. In the carbon capture cases, once the syngas is diluted to the target heating value, all of the available combustion air is required to maintain mass flow through the turbine and hence maintain power output.

The amount of air extracted from the gas turbine in the non-capture case is determined through a process that includes the following constraints:

- The combustion turbine must be fully loaded; i.e., sufficient gas mass flow is supplied to maximize the turbine power output at the given elevation.
- The diluted syngas must meet heating value requirements specified by a combustion turbine vendor, which ranged from 4.3-4.8 MJ/Nm^3 (115-129 Btu/scf) (LHV).

The air extraction for the non-CO_2 capture case is shown in Exhibit 3-2. It was not a goal of this project to optimize the integration of the combustion turbine and the ASU, although several recent papers have shown that providing 25-30 percent of the ASU air from the turbine compressor provides the best balance between maximizing plant output and efficiency without compromising plant availability or reliability [41, 42].

Exhibit 3-2 Air Extracted from the Combustion Turbine and Supplied to the ASU in Non-Carbon Capture Cases

	Case 1
Air Extracted from Gas Turbine, %	5.7
Air Provided to ASU, % of ASU Total	22.5

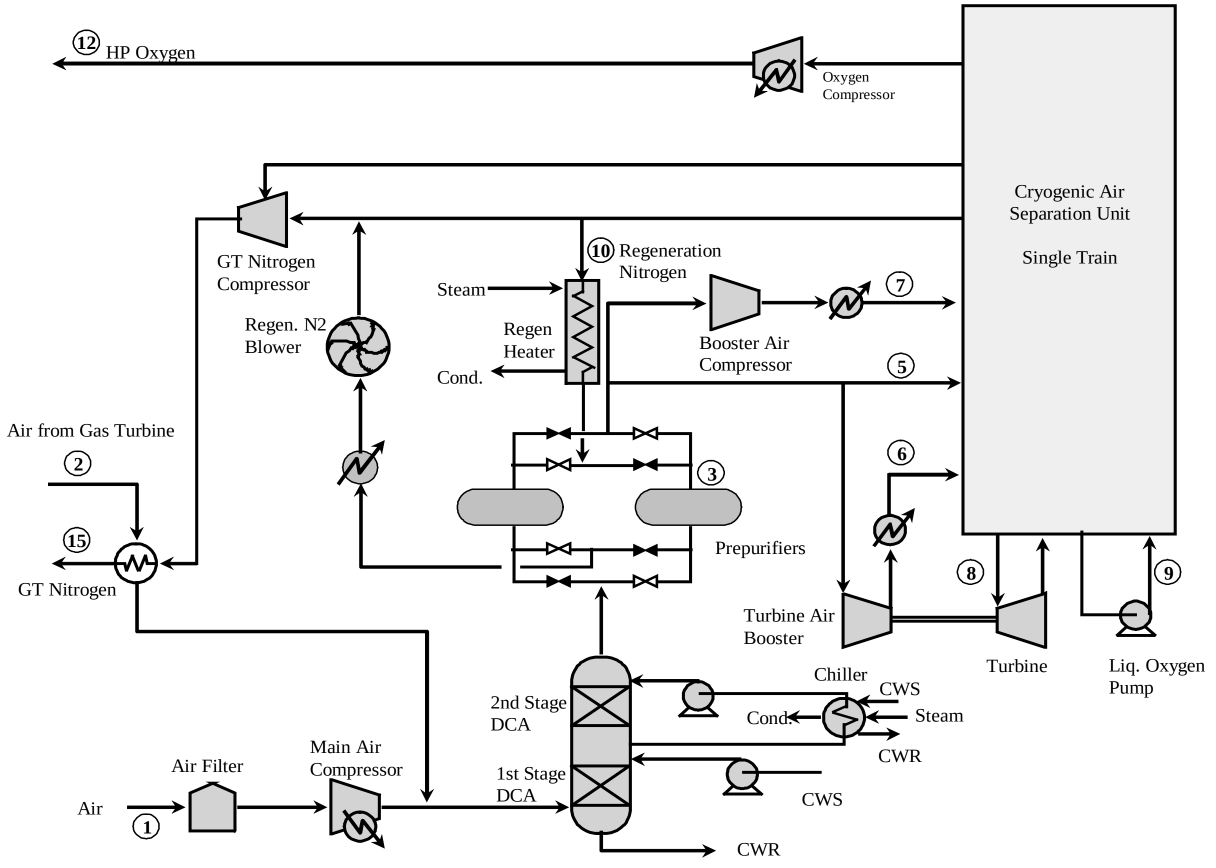
Air Separation Plant Process Description [43]

The air separation plant is designed to produce 95 mole percent O_2 for use in the gasifier. The plant is designed with two production trains, one for each gasifier. The air compressor is powered by an electric motor. Nitrogen is also recovered, compressed, and used as dilution in the gas turbine combustor. A process schematic of a typical ASU is shown in Exhibit 3-3.

The air feed to the ASU is supplied from two sources. A portion of the air is extracted from the compressor of the gas turbine (non-CO_2 capture cases only). The remaining air is supplied from a stand-alone compressor. Air to the stand-alone compressor is first filtered in a suction filter upstream of the compressor. This air filter removes particulate, which may tend to cause compressor wheel erosion and foul intercoolers. The filtered air is then compressed in the centrifugal compressor, with intercooling between each stage.

Air from the stand-alone compressor is combined with the extraction air, and the combined stream is cooled and fed to an adsorbent-based pre-purifier system. The adsorbent removes water, carbon dioxide, and C_4+ saturated hydrocarbons in the air. After passing through the adsorption beds, the air is filtered with a dust filter to remove any adsorbent fines that may be present. Downstream of the dust filter a small stream of air is withdrawn to supply the instrument air requirements of the ASU.

Exhibit 3-3 Typical ASU Process Schematic



Regeneration of the adsorbent in the pre-purifiers is accomplished by passing a hot nitrogen stream through the off-stream bed(s) in a direction countercurrent to the normal airflow. The nitrogen is heated against extraction steam (1.7 MPa [250 psia]) in a shell and tube heat exchanger. The regeneration nitrogen drives off the adsorbed contaminants. Following regeneration, the heated bed is cooled to near normal operating temperature by passing a cool nitrogen stream through the adsorbent beds. The bed is re-pressurized with air and placed on stream so that the current on-stream bed(s) can be regenerated.

The air from the pre-purifier is then split into three streams. About 70 percent of the air is fed directly to the cold box. About 25 percent of the air is compressed in an air booster compressor. This boosted air is then cooled in an aftercooler against cooling water in the first stage and against chilled water in the second stage before it is fed to the cold box. The chiller utilizes low pressure process steam at 0.45 MPa (65 psia). The remaining 5 percent of the air is fed to a turbine-driven, single-stage, centrifugal booster compressor. This stream is cooled in a shell and tube aftercooler against cooling water before it is fed to the cold box.

All three air feeds are cooled in the cold box to cryogenic temperatures against returning product oxygen and nitrogen streams in plate-and-fin heat exchangers. The large air stream is fed directly to the first distillation column to begin the separation process. The second largest air stream is liquefied against boiling liquid oxygen before it is fed to the distillation columns. The third, smallest air stream is fed to the cryogenic expander to produce refrigeration to sustain the cryogenic separation process.

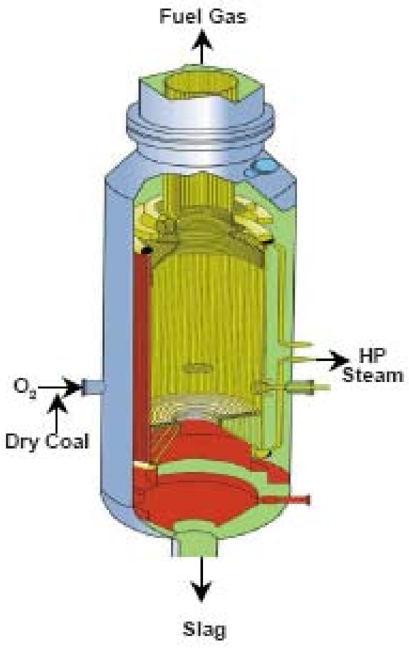
Inside the cold box the air is separated into oxygen and nitrogen products. The oxygen product is withdrawn from the distillation columns as a liquid and is pressurized by a cryogenic pump. The pressurized liquid oxygen is then vaporized against the high-pressure air feed before being warmed to ambient temperature. The gaseous oxygen exits the cold box and is fed to the centrifugal compressor with intercooling between each stage of compression. The compressed oxygen is then fed to the gasification unit.

Nitrogen is produced from the cold box at two pressure levels. Each stream is compressed to 2.63 MPa (381 psia) for use as combustion turbine diluent nitrogen. Some of the nitrogen stream is compressed further for use as transport gas in the lockhoppers.

3.1.4 Gasifier

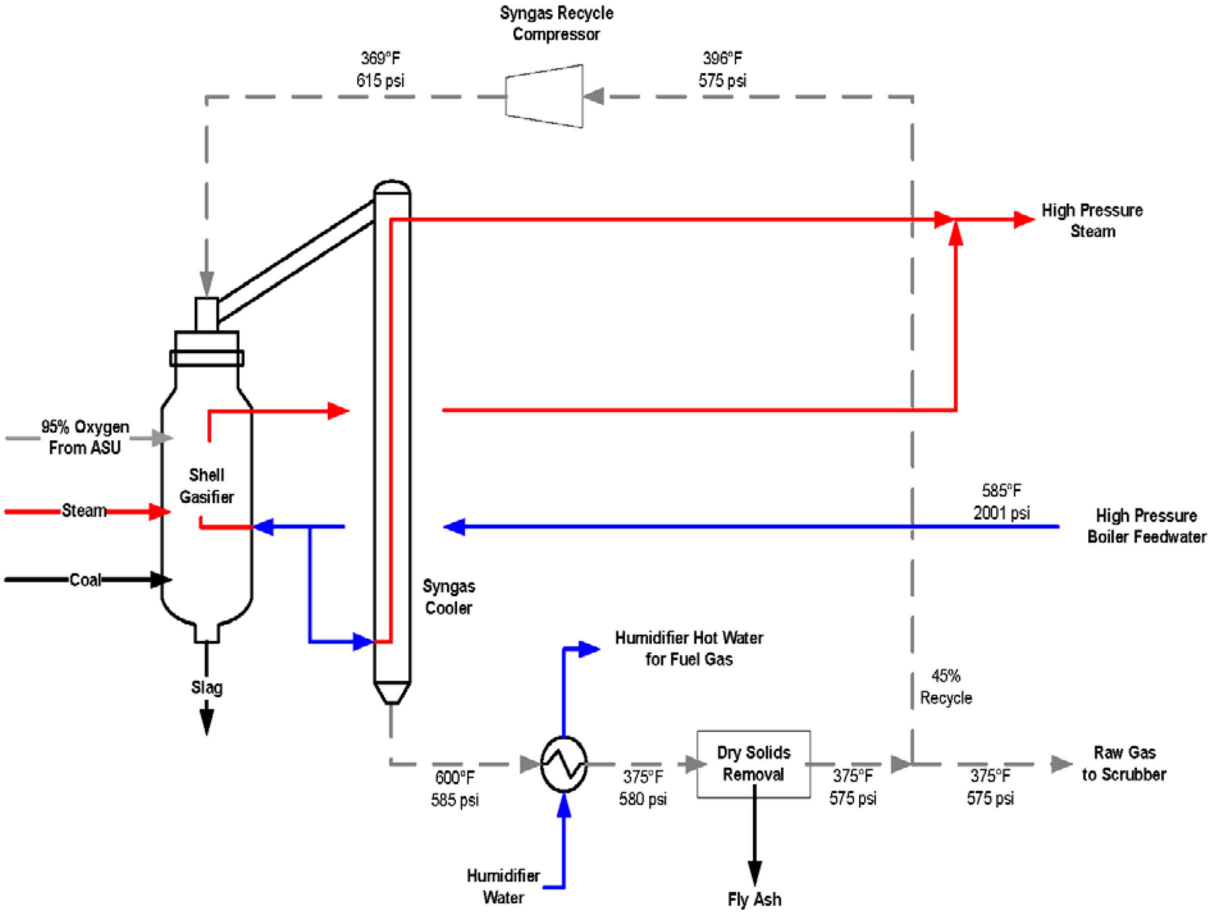
The Shell gasifier, which is a single-stage, entrained-flow, dry-feed gasifier, is modeled as an equilibrium reactor. A schematic of the Shell gasifier as a stand-alone unit is shown in Exhibit 3-4. Many literature references support this modeling strategy [37,44,45]. Steam injection is based on published data and the oxygen injection is controlled to maintain published heat losses for the gasifier. The predicted raw gas composition for the PRB coal is reasonable relative to published data.

Exhibit 3-4 Shell Gasifier



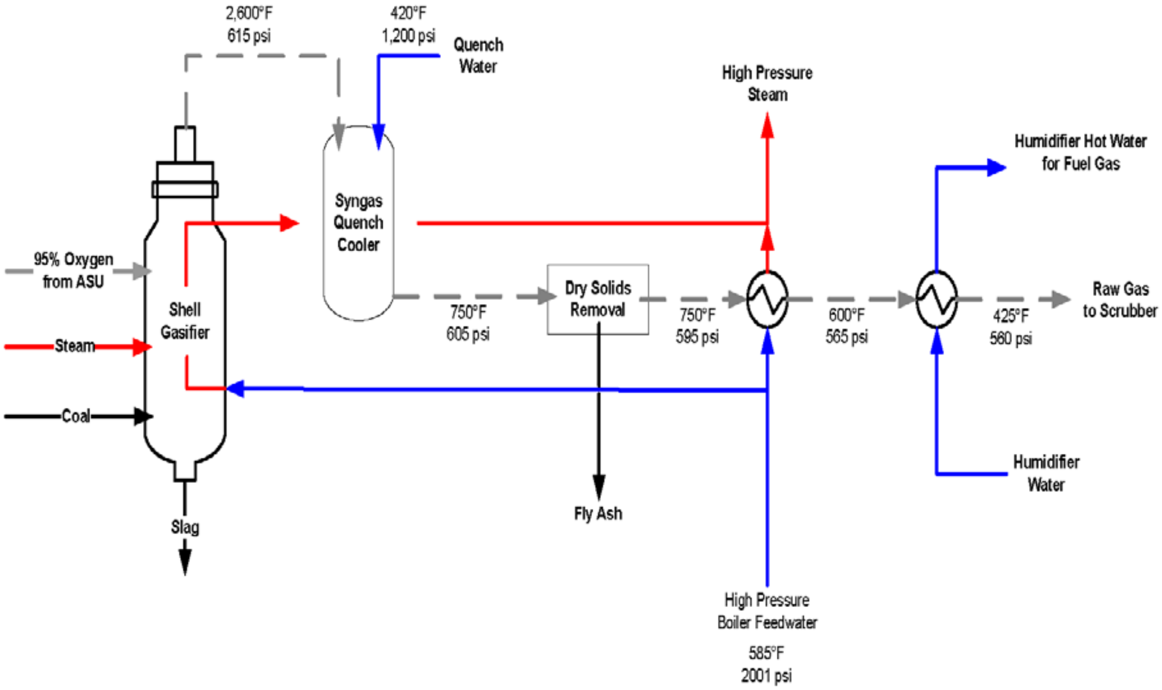
Two different raw gas cooling configurations were used in this study with the Shell gasifier. One configuration is a gasifier with a syngas cooler and the other is a gasifier with full quench. For the non-capture case (Case 1), a syngas cooler was implemented. The syngas cooler cools the raw gas from the gasifier to 600°F by creating high pressure steam. This configuration was utilized in the non-capture case for the ability of the syngas cooler to produce high pressure steam that supplements the steam produced in the heat recovery steam generator and subsequently is used for power generation in the steam cycle. A process schematic of the gasifier and syngas cooler is shown in Exhibit 3-5.

Exhibit 3-5 Shell Gasifier with Syngas Cooler



For the capture cases (Cases 2 and 3), a full quench design was implemented. This configuration is implemented because it reduces the amount of steam extracted from the steam cycle, which would be used for power generation, necessary for the Water Gas Shift (WGS) reactors to achieve the required levels of carbon capture. This is accomplished by using water to quench the raw gas from the gasifier to 750°F. The quench water is subsequently used in the WGS reactors to create the desired amount of shift of carbon monoxide (CO) to carbon dioxide (CO_2), which is later separated and captured to the specified levels of 1,100 lb CO_2/net-MWh or 90 percent capture. The full quench gasifier configuration is shown in Exhibit 3-6.

Exhibit 3-6 Shell Gasifier with Full Quench



With the syngas quench cooler configuration high pressure steam is still produced, but at a reduced quantity because the temperature available for steam production is now at 750°F instead of 2,600°F, as in the syngas cooler configuration. This causes a decrease in overall plant efficiency, but the quench design is still utilized to enhance the shift reaction necessary for carbon capture. For comparison, if full quench is implemented on the non-capture case, the net efficiency is reduced from 41.8 percent to 37.9 percent, a decrease of 3.9 percent.

3.1.5 Water Gas Shift Reactors

Selection of Technology - In the cases with CO_2 separation and capture, the gasifier product must be converted to hydrogen-rich syngas. The first step is to convert most of the syngas carbon monoxide (CO) to CO_2 by reacting the CO with water over a bed of catalyst. The H_2O:dry gas molar ratio at the exit of the final shift reactor is adjusted to a minimum of 0.3:1 by the addition of steam to the syngas stream thus promoting a high conversion of CO. The H_2O:dry gas molar ratio is adjusted as necessary (but maintaining a minimum 0.3:1) to achieve 90 percent overall CO_2 removal. In the cases without CO_2 separation and capture, CO shift convertors are not required.

Water Gas Shift: $CO + H_2O \leftrightarrow CO_2 + H_2$

The CO shift converter can be located either upstream of the acid gas removal step (sour gas shift) or immediately downstream (sweet gas shift). If the CO converter is located downstream of the acid gas removal, then the metallurgy of the unit is less stringent but additional equipment must be added to the process. Products from the gasifier are quenched with water and contain a portion of the water vapor necessary to meet the water-to-dry gas criterion at the reactor outlet. If the CO converter is located downstream of the acid gas removal, then the gasifier product would first have to be cooled and the free water separated and treated. Then additional steam would have to be generated and re-injected into the CO converter feed to meet the required water-to-dry gas ratio. If the CO converter is located upstream of the acid gas removal step, no additional equipment is required. This is because the CO converter promotes carbonyl sulfide (COS) hydrolysis without a separate catalyst bed. Therefore, for this study the CO converter was located upstream of the acid gas removal unit and is referred to as sour gas shift (SGS). In the 1,100 lb CO_2/net-MWh capture case, the partial bypass around the SGS reactor in each train causes an elevation in the sulfur content of the CO_2 product because not all of the COS gets converted to H_2S and consequently is removed to a much lesser extent in the AGR process.

Process Description - The SGS consists of two paths of parallel fixed-bed reactors arranged in series. Two reactors in series are used in each parallel path to achieve sufficient conversion to meet the 90 percent CO_2 capture target. Only one reactor in each train is necessary to achieve the emission limit of 1,100 lb CO_2/net-MWh. In addition for the 1,100 lb CO_2/net-MWh case, a bypass stream around the SGS is implemented to further reduce the conversion of CO to CO_2 to reach the required emissions limit.

In the 1,100 lb CO_2/net-MWh capture case, the 2 gasifier trains each have 1 SGS reactor with a bypass to achieve the emission limit, which resulted in 46 percent carbon capture. Since less than 50 percent carbon capture is required, 2 or 3 stages of SGS could be used in one train and none in the second train. This configuration would require a separate one-stage Selexol unit and a two-stage Selexol unit, which was deemed to not offer any particular advantage.

Cooling is provided between the series of reactors in the 90 percent case to control the exothermic temperature rise. The parallel set of reactors is required due to the high gas mass flow rate. In the 90 percent CO_2 capture case the heat exchanger after the first SGS reactor is used to superheat steam that is then used to adjust the syngas H_2O:dry gas ratio to greater than 0.3:1 on a molar basis. The heat exchanger after the second SGS reactor is a gas-gas exchanger used to preheat the syngas prior to the first SGS reactor to raise the syngas temperature above the dew point.

3.1.6 Mercury Removal

An IGCC power plant has the potential of removing mercury in a more simple and cost-effective manner than conventional PC plants. This is because mercury can be removed from the syngas at elevated pressure and prior to combustion so that syngas volumes are much smaller than flue gas volumes in comparable PC cases. A conceptual design for a carbon bed adsorption system was developed for mercury control in the IGCC plants being studied. Data on the performance of carbon bed systems were obtained from the Eastman Chemical Company, which uses carbon beds at its syngas facility in Kingsport, Tennessee [16]. The coal mercury content (0.081 ppm dry for PRB) and carbon bed removal efficiency (95 percent) were discussed previously in Section 2.3. IGCC-specific design considerations are discussed below.

Carbon Bed Location – The packed carbon bed vessels are located upstream of the acid gas removal (AGR) process and syngas enters at a temperature near 38°C (100°F). Consideration was given to locating the beds further upstream before the COS hydrolysis unit (in non-CO_2 capture cases) at a temperature near 204°C (400°F). However, while the mercury removal efficiency of carbon has been found to be relatively insensitive to pressure variations, temperature adversely affects the removal efficiency [46]. Eastman Chemical also operates their beds ahead of their sulfur recovery unit at a temperature of 30°C (86°F) [16].

Consideration was also given to locating the beds downstream of the AGR. However, it was felt that removing the mercury and other contaminants before the AGR unit would enhance the performance of both the AGR and sulfur recover unit (SRU) and increase the life of the various solvents.

Process Parameters – An empty vessel basis gas residence time of approximately 20 seconds was used based on Eastman Chemical's experience [16]. Allowable gas velocities are limited by considerations of particle entrainment, bed agitation, and pressure drop. One-foot-per-second superficial velocity is in the middle of the range normally encountered [46] and was selected for this application.

The bed density of 30 lb/ft^3 was based on the Calgon Carbon Corporation HGR-P sulfur-impregnated pelletized activated carbon [47]. These parameters determined the size of the vessels and the amount of carbon required. Each gasifier train has one mercury removal bed and there are two gasifier trains in each IGCC case, resulting in two carbon beds per case.

Carbon Replacement Time – Eastman Chemicals replaces its bed every 18 to 24 months [16]. However, bed replacement is not because of mercury loading, but for other reasons including:

- A buildup in pressure drop
- A buildup of water in the bed
- A buildup of other contaminants

For this study a 24 month carbon replacement cycle was assumed. Under these assumptions, the mercury loading in the bed would build up to 0.64 weight percent (wt%). Mercury capacity of sulfur-impregnated carbon can be as high as 20 wt% [48]. The mercury laden carbon is considered to be a hazardous waste, and the disposal cost estimate reflects this categorization.

3.1.7 Acid Gas Removal (AGR) Process Selection

Gasification of coal to generate power produces a syngas that must be treated prior to further utilization. A portion of the treatment consists of acid gas removal (AGR) and sulfur recovery. The environmental target for these IGCC cases, 0.0128 lb SO_2/MMBtu, is based on the EPRI CoalFleet values for bituminous coal [13] and requires that the total sulfur content of the syngas be reduced to less than 30 ppmv. This includes all sulfur species, but in particular the total of COS and H_2S, thereby resulting in stack gas emissions of less than 4 ppmv SO_2. Because the low rank western coals have substantially less sulfur than eastern bituminous coal, the resulting sulfur emissions are significantly below the environmental target.

COS Hydrolysis

The use of COS hydrolysis pretreatment in the feed to the acid gas removal process provides a means to reduce the COS concentration. This method was first commercially proven at the Buggenum plant, and was also used at both the Tampa Electric and Wabash River IGCC projects. Several catalyst manufacturers including Haldor Topsoe and Porocel offer a catalyst that promotes the COS hydrolysis reaction. The non-carbon capture COS hydrolysis reactor designs are based on information from Porocel. In cases with carbon capture, the SGS reactors reduce COS to H_2S as discussed in Section 3.1.4.

The COS hydrolysis reaction is equimolar with a slightly exothermic heat of reaction. The reaction is represented as follows.

$$COS + H_2O \leftrightarrow CO_2 + H_2S$$

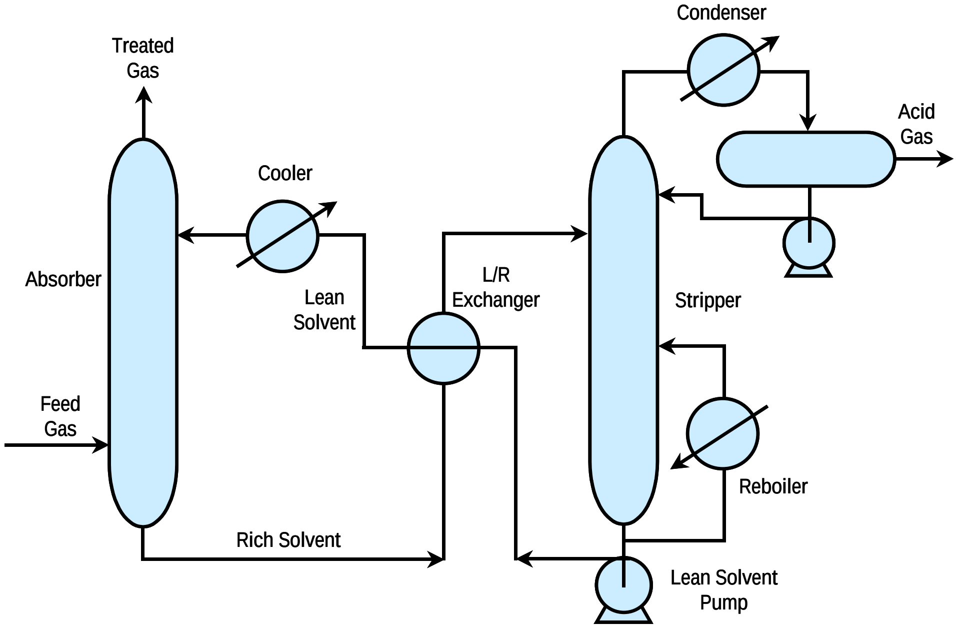
Since the reaction is exothermic, higher conversion is achieved at lower temperatures. However, at lower temperatures the reaction kinetics are slower. Since the exit gas COS concentration is critical to the amount of H_2S that must be removed with the AGR process, a retention time of 50-75 seconds was used to achieve 99.5 percent conversion of the COS. The Porocel activated alumina-based catalyst, designated as Hydrocel 640 catalyst, promotes the COS hydrolysis reaction without promoting reaction of H_2S and CO to form COS and H_2.

Although the reaction is exothermic, the heat of reaction is dissipated among the large amount of non-reacting components. Therefore, the reaction is essentially isothermal. The product gas, now containing less than 4 ppmv of COS, is cooled prior to entering the mercury removal process and the AGR.

Sulfur Removal

Hydrogen sulfide removal generally consists of absorption by a regenerable solvent. The most commonly used technique is based on countercurrent contact with the solvent. Acid-gas-rich solution from the absorber is stripped of its acid gas in a regenerator, usually by application of heat. The regenerated lean solution is then cooled and recirculated to the top of the absorber, completing the cycle. Exhibit 3-7 is a simplified diagram of the AGR process [49].

Exhibit 3-7 Flow Diagram for a Conventional AGR Unit



There are well over 30 AGR processes in common commercial use throughout the oil, chemical, and natural gas industries. However, in a 2002 report by SFA Pacific a list of 42 operating and planned gasifiers shows that only six AGR processes are represented: Rectisol, Sulfinol, methyldiethanolamine (MDEA), Selexol, aqueous di-isoproponal (ADIP) amine and FLEXSORB [50]. These processes can be separated into three general types: chemical reagents, physical solvents, and hybrid solvents. A summary of these common AGR processes is shown in Exhibit 3-8. The optimum technology choice for a particular IGCC plant depends on many factors such as gasifier operating pressure, availability of low/medium pressure steam, acid gas removal requirements, and capital cost.

Chemical Solvents

Frequently used for acid gas removal, chemical solvents are more suitable than physical or hybrid solvents for applications at lower operating pressures. The chemical nature of acid gas absorption makes solution loading and circulation less dependent on the acid gas partial pressure. Because the solution is aqueous, co-absorption of hydrocarbons is minimal.

In a conventional amine unit, the chemical solvent reacts exothermically with the acid gas constituents. They form a weak chemical bond that can be broken, releasing the acid gas and regenerating the solvent for reuse.

In recent years MDEA, a tertiary amine, has acquired a much larger share of the gas-treating market. Compared with primary and secondary amines, MDEA has superior capabilities for selectively removing H_2S in the presence of CO_2, is resistant to degradation by organic sulfur compounds, has a low tendency for corrosion, has a relatively low circulation rate, and consumes less energy. Commercially available are several MDEA-based solvents that are formulated for high H_2S selectivity.

Exhibit 3-8 Summary of Common AGR Processes

Solvent Type	Process	H_2S Selectivity	Solvent Circulation	Heat Input	Capital Cost	Pressure Sensitive	High Removal
Physical	Rectisol, Selexol	Good	High, decreases with increased pressure	Low	High	Yes	Yes, at high acid gas partial pressures
Mixed	Sulfinol, FLEXSORB	Good but more complicated to achieve	Intermediate	Intermediate	Intermediate	Yes, but to a lesser extent than physical solvents only	Yes, at optimum operating conditions
Chemical	Amines (MEA, DEA, MDEA)	Varies depending on amine selected, highest for MDEA	Low	High	Low	No	Yes, but with refrigeration

Chemical reagents are used to remove the acid gases by a reversible chemical reaction of the acid gases with an aqueous solution of various alkanolamines or alkaline salts in water. Exhibit 3-9 lists commonly used chemical reagents along with principal licensors that use them in their processes. The process consists of an absorber and regenerator, which are connected by a circulation of the chemical reagent aqueous solution. The absorber contacts the lean solution with the main gas stream (at pressure) to remove the acid gases by absorption/ reaction with the chemical solution. The acid-gas-rich solution is reduced to low pressure and heated in the stripper to reverse the reactions and strip the acid gas. The acid-gas-lean solution leaves the bottom of the regenerator stripper and is cooled, pumped to the required pressure and recirculated back to the absorber. For some amines, a filter and a separate reclaiming section (not shown) are needed to remove undesirable reaction byproducts.

Exhibit 3-9 Common Chemical Reagents Used in AGR Processes

Chemical Reagent	Designation	Process Licensors Using the Reagent
Monoethanolamine	MEA	Dow, Exxon, Lurgi, Union Carbide
Diethanolamine	DEA	Elf, Lurgi
Diglycolamine	DGA	Texaco, Fluor
Triethanolamine	TEA	AMOCO
Diisopropanolamine	DIPA	Shell
Methyldiethanolamine	MDEA	BASF, Dow, Elf, Snamprogetti, Shell, Union Carbide, Coastal Chemical
Hindered amine		Exxon
Potassium carbonate	“hot pot”	Eickmeyer, Exxon, Lurgi, Union Carbide

Typically, the absorber temperature is 27 to 49°C (80 to 120°F) for amine processes, and the regeneration temperature is the boiling point of the solutions, generally 104 to 127°C (220 to 260°F). The liquid circulation rates can vary widely, depending on the amount of acid gas being captured. However, the most suitable processes are those that will dissolve 2 to 10 scf acid gas per gallon of solution circulated. Steam consumption can vary widely also: 0.7 to 1.5 pounds per gallon of liquid is typical, with 0.8 to 0.9 being a typical “good” value.

The major advantage of these systems is the ability to remove acid gas to low levels at low to moderate H_2S partial pressures.

Physical Solvents

Physical solvents involve absorption of acid gases into certain organic solvents that have a high solubility for acid gases. As the name implies, physical solvents involve only the physical solution of acid gas – the acid gas loading in the solvent is proportional to the acid gas partial pressure (Henry’s Law). Physical solvent absorbers are usually operated at lower temperatures than is the case for chemical solvents. The solution step occurs at high pressure and at or below ambient temperature while the regeneration step (dissolution) occurs by pressure letdown and indirect stripping with low-pressure 0.45 MPa (65 psia) steam. It is generally accepted that physical solvents become increasingly economical, and eventually superior to amine capture, as the partial pressure of acid gas in the syngas increases.

The physical solvents are regenerated by multistage flashing to low pressures. Because the solubility of acid gases increases as the temperature decreases, absorption is generally carried out at lower temperatures, and refrigeration is often required.

Most physical solvents are capable of removing organic sulfur compounds. Exhibiting higher solubility of H_2S than CO_2, they can be designed for selective H_2S or total acid gas removal. In applications where CO_2 capture is desired the CO_2 is flashed off at various pressures, which reduces the compression work and parasitic power load associated with sequestration.

Physical solvents co-absorb heavy hydrocarbons from the feed stream. Since heavy hydrocarbons cannot be recovered by flash regeneration, they are stripped along with the acid gas during heated regeneration. These hydrocarbon losses result in a loss of valuable product and may lead to CO_2 contamination.

Several physical solvents that use anhydrous organic solvents have been commercialized. They include the Selexol process, which uses dimethyl ether of polyethylene glycol as a solvent; Rectisol, with methanol as the solvent; Purisol, which uses N-methyl-2-pyrrolidone (NMP) as a solvent; and the propylene-carbonate process.

Exhibit 3-10 is a simplified flow diagram for a physical reagent type acid gas removal process [49]. Common physical solvent processes, along with their licensors, are listed in Exhibit 3-11.

Exhibit 3-10 Physical Solvent AGR Process Simplified Flow Diagram

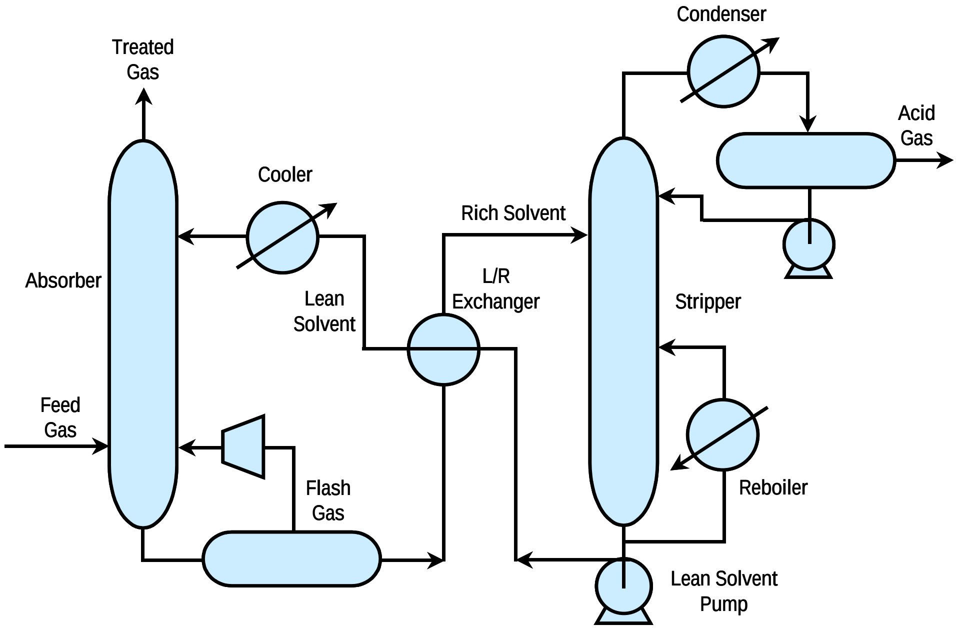


Exhibit 3-11 Common Physical Solvents Used in AGR Processes

Solvent	Solvent/Process Trade Name	Process Licensors
Dimethyl ether of polyethylene glycol	Selexol	UOP
Methanol	Rectisol	Linde AG and Lurgi
Methanol and toluene	Rectisol II	Linde AG
N—methyl pyrrolidone	Purisol	Lurgi
Polyethylene glycol and dialkyl ethers	Sepasolv MPE	BASF
Propylene carbonate	Fluor Solvent	Fluor
Tetrahydrothiophenedioxide	Sulfolane	Shell
Tributyl phosphate	Estasolvan	Uhde and IFP

Hybrid Solvents

Hybrid solvents combine the high treated-gas purity offered by chemical solvents with the flash regeneration and lower energy requirements of physical solvents. Some examples of hybrid solvents are Sulfinol, Flexsorb PS, and Ucarsol LE.

Sulfinol is a mixture of sulfolane (a physical solvent), diisopropanolamine (DIPA) or MDEA (chemical solvent), and water. DIPA is used when total acid gas removal is specified, while MDEA provides for selective removal of H_2S.

Flexsorb PS is a mixture of a hindered amine and an organic solvent. Physically similar to Sulfinol, Flexsorb PS is very stable and resistant to chemical degradation. High treated-gas purity, with less than 50 ppmv of CO_2 and 4 ppmv of H_2S, can be achieved. Both Ucarsol LE-701, for selective removal, and LE-702, for total acid gas removal, are formulated to remove mercaptans from feed gas.

Mixed chemical and physical solvents combine the features of both systems. The mixed solvent allows the solution to absorb an appreciable amount of gas at high pressure. The amine portion is effective as a reagent to remove the acid gas to low levels when high purity is desired.

Mixed solvent processes generally operate at absorber temperatures similar to those of the amine-type chemical solvents and do not require refrigeration. They also retain some advantages of the lower steam requirements typical of the physical solvents. Common mixed chemical and physical solvent processes, along with their licensors, are listed in Exhibit 3-12. The key advantage of mixed solvent processes is their apparent ability to remove H_2S and, in some cases, COS to meet very stringent purified gas specifications.

Exhibit 3-12 Common Mixed Solvents Used in AGR Processes

Solvent/Chemical Reagent	Solvent/Process Trade Name	Process Licensors
Methanol/MDEA or diethylamine	Amisol	Lurgi
Sulfolane/MDEA or DIPA	Sulfinol	Shell
Methanol and toluene	Selefining	Snamprogetti
(Unspecified) /MDEA	FLEXSORB PS	Exxon

Exhibit 3-13 shows reported equilibrium solubility data for H_2S and CO_2 in various representative solvents [49]. The solubility is expressed as standard cubic feet of gas per gallon liquid per atmosphere gas partial pressure.

The figure illustrates the relative solubilities of CO_2 and H_2S in different solvents and the effects of temperature. More importantly, it shows an order of magnitude higher solubility of H_2S over CO_2 at a given temperature, which gives rise to the selective absorption of H_2S in physical solvents. It also illustrates that the acid gas solubility in physical solvents increases with lower solvent temperatures.

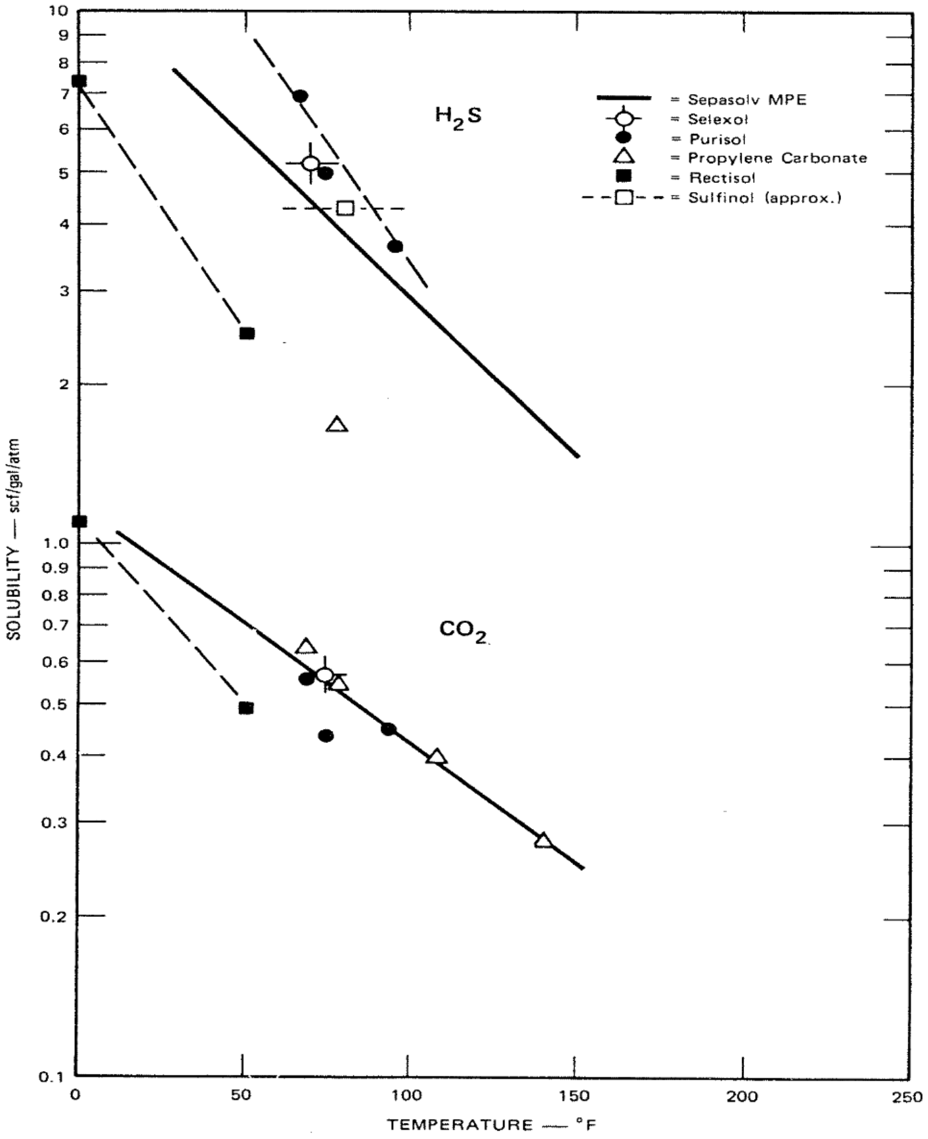
The ability of a process to selectively absorb H_2S may be further enhanced by the relative absorption rates of H_2S and CO_2. Thus, some processes, besides using equilibrium solubility differences, will use absorption rate differences between the two acid gases to achieve selectivity. This is particularly true of the amine processes where the CO_2 and H_2S absorption rates are very different.

CO_2 Capture

A two-stage Selexol process is used for both IGCC CO_2 capture cases in this report. A brief process description follows.

Untreated syngas enters the first of two absorbers where H_2S is preferentially removed using loaded solvent from the CO_2 absorber. The gas exiting the H_2S absorber passes through the second absorber where CO_2 is removed using first flash regenerated, chilled solvent followed by thermally regenerated solvent added near the top of the column. The treated gas exits the absorber and is sent either directly to the combustion turbine or is partially humidified prior to entering the combustion turbine.

Exhibit 3-13 Equilibrium Solubility Data on H_2S and CO_2 in Various Solvents



The amount of hydrogen remaining in the syngas stream is dependent on the Selexol process design conditions. In this study, hydrogen recovery is 99.4 percent. The minimal hydrogen slip to the CO_2 sequestration stream maximizes the overall plant efficiency. The Selexol plant cost estimates are based on a plant designed to recover this high percentage of hydrogen. The balance of the hydrogen is either co-sequestered with the CO_2, destroyed in the Claus plant burner, or recycled to the gasifier.

The CO_2 loaded solvent exits the CO_2 absorber and a portion is sent to the H_2S absorber, a portion is sent to a reabsorber and the remainder is sent to a series of flash drums for regeneration. The CO_2 product stream is obtained from the three flash drums, and after flash regeneration the solvent is chilled and returned to the CO_2 absorber.

The rich solvent exiting the H_2S absorber is combined with the rich solvent from the reabsorber and the combined stream is heated using the lean solvent from the stripper. The hot, rich solvent enters the H_2S concentrator and partially flashes. The remaining liquid contacts nitrogen from the ASU and a portion of the CO_2 along with lesser amounts of H_2S and COS are stripped from the rich solvent. The stripped gases from the H_2S concentrator are sent to the reabsorber where the H_2S and COS that were co-stripped in the concentrator are transferred to a stream of loaded solvent from the CO_2 absorber. The clean gas from the reabsorber is combined with the clean gas from the H_2S absorber and sent to the combustion turbine.

The solvent exiting the H_2S concentrator is sent to the stripper where the absorbed gases are liberated by hot gases flowing up the column from the steam heated reboiler. Water in the overhead vapor from the stripper is condensed and returned as reflux to the stripper or exported as necessary to maintain the proper water content of the lean solvent. The acid gas from the stripper is sent to the Claus plant for further processing. The lean solvent exiting the stripper is first cooled by providing heat to the rich solvent, then further cooled by exchange with the product gas and finally chilled in the lean chiller before returning to the top of the CO_2 absorber.

AGR/Gasifier Pairings

There are numerous commercial AGR processes that could meet the sulfur environmental target of this study. The most frequently used AGR systems (Selexol, Sulfinol, MDEA, and Rectisol) have all been used with the Shell gasifier in various applications. Since there is no compelling reason to select one AGR process over another, the Sulfinol-M process was chosen to be consistent with the Shell cases in previous studies. Previous vendor performance estimates for Sulfinol systems showed high removals for H_2S (99.77 percent) and CO_2 (97.5 percent). With the higher CO_2 and lower H_2S concentrations in the raw gas for the lower rank coals, it is necessary for the AGR to slip a significant amount of CO_2. The high slip is necessary to reduce the volume and increase the H_2S concentration of the acid gas stream to the Claus plant for adequate performance and minimum capital cost.

The literature indicates that Sulfinol systems with CO_2 slips of 60% have been designed. The Shell non-capture cases assume a 60 percent CO_2 slip.

The two-stage Selexol process is used in both cases that require carbon capture. According to the previously referenced SFA Pacific report, “For future IGCC with CO_2 removal for sequestration, a two-stage Selexol process presently appears to be the preferred AGR process – as indicated by ongoing engineering studies at EPRI and various engineering firms with IGCC interests.” [50]

3.1.8 Sulfur Recovery/Tail Gas Cleanup Process Selection

Currently, most of the world’s sulfur is produced from the acid gases coming from gas treating. The Claus process remains the mainstay for sulfur recovery. Conventional three-stage Claus plants, with indirect reheat and feeds with a high H_2S content, can approach 98 percent sulfur recovery efficiency. However, since environmental regulations have become more stringent, sulfur recovery plants are required to recover sulfur with over 99.8 percent efficiency. To meet these stricter regulations, the Claus process underwent various modifications and add-ons.

The Claus Process

The Claus process converts H_2S to elemental sulfur via the following reactions:

$$H_2S + 3/2\ O_2 \leftrightarrow H_2O + SO_2$$

$$2H_2S + SO_2 \leftrightarrow 2H_2O + 3S$$

The second reaction, the Claus reaction, is equilibrium limited. The overall reaction is:

$$3H_2S + 3/2\ O_2 \leftrightarrow 3H_2O + 3S$$

The sulfur in the vapor phase exists as S_2, S_6, and S_8 molecular species, with the S_2 predominant at higher temperatures, and S_8 predominant at lower temperatures.

A simplified process flow diagram of a typical three-stage Claus plant is shown in Exhibit 3-14 [50]. One-third of the H_2S is burned in the furnace with oxygen from the air to give sufficient SO_2 to react with the remaining H_2S. Since these reactions are highly exothermic, a waste heat boiler that recovers this heat to generate high-pressure steam usually follows the furnace. Sulfur is condensed in a condenser that follows the high-pressure steam recovery section. Low-pressure steam is raised in the condenser. The tail gas from the first condenser then goes to several catalytic conversion stages, usually 2 to 3, where the remaining sulfur is recovered via the Claus reaction. Each catalytic stage consists of gas preheat, a catalytic reactor, and a sulfur condenser. The liquid sulfur goes to the sulfur pit, while the tail gas proceeds to the incinerator or for further processing in a TGTU.

Claus Plant Sulfur Recovery Efficiency

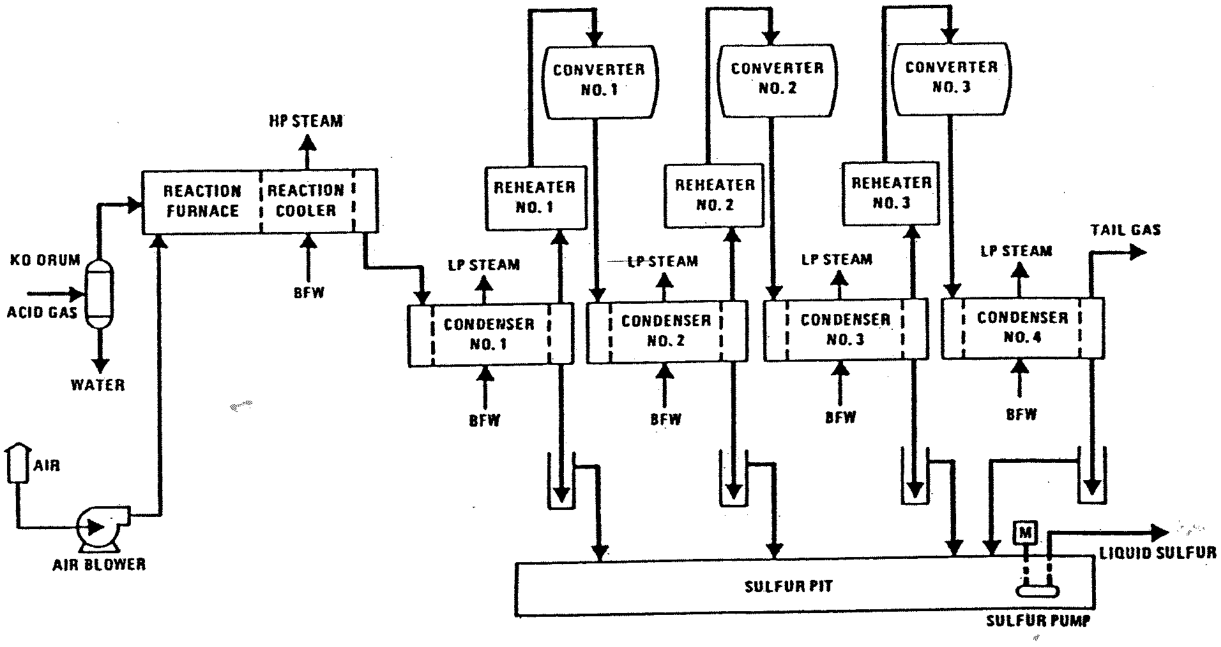
The Claus reaction is equilibrium limited, and sulfur conversion is sensitive to the reaction temperature. The highest sulfur conversion in the thermal zone is limited to about 75 percent. Typical furnace temperatures are in the range from 1093 to 1427°C (2000 to 2600°F), and as the temperature decreases, conversion increases dramatically.

Claus plant sulfur recovery efficiency depends on many factors:

- H_2S concentration of the feed gas
- Number of catalytic stages
- Gas reheat method

In order to keep Claus plant recovery efficiencies approaching 94 to 96 percent for feed gases that contain about 20 to 50 percent H_2S, a split-flow design is often used. In this version of the Claus plant, part of the feed gas is bypassed around the furnace to the first catalytic stage, while the rest of the gas is oxidized in the furnace to mostly SO_2. This results in a more stable temperature in the furnace.

Exhibit 3-14 Typical Three-Stage Claus Sulfur Plant



Oxygen-Blown Claus

Large diluent streams in the feed to the Claus plant, such as N_2 from combustion air, or a high CO_2 content in the feed gas, lead to higher cost Claus processes and any add-on or tail gas units. One way to reduce diluent flows through the Claus plant and to obtain stable temperatures in the furnace for dilute H_2S streams is the oxygen-blown Claus process.

The oxygen-blown Claus process was originally developed to increase capacity at existing conventional Claus plants and to increase flame temperatures of low H_2S content gases. The process has also been used to provide the capacity and operating flexibility for sulfur plants where the feed gas is variable in flow and composition such as often found in refineries. The application of the process has now been extended to grass roots installations, even for rich H_2S feed streams, to provide operating flexibility at lower costs than would be the case for conventional Claus units. At least four of the recently built gasification plants in Europe use oxygen enriched Claus units.

Oxygen enrichment results in higher temperatures in the front-end furnace, potentially reaching temperatures as high as 1593 to 1649°C (2900 to 3000°F) as the enrichment moves beyond 40 to 70 volume percent O_2 in the oxidant feed stream. Although oxygen enrichment has many benefits, its primary benefit for lean H_2S feeds is a stable furnace temperature. Sulfur recovery is not significantly enhanced by oxygen enrichment. Because the IGCC process already requires an ASU, the oxygen-blown Claus plant was chosen for all cases.

Tail Gas Treating

In many refinery and other conventional Claus applications, tail gas treating involves the removal of the remaining sulfur compounds from gases exiting the sulfur recovery unit. Tail gas from a typical Claus process, whether a conventional Claus or one of the extended versions of the process, usually contains small but varying quantities of COS, CS_2, H_2S, SO_2, and elemental sulfur vapors. In addition, there may be H_2, CO, and CO_2 in the tail gas. In order to remove the rest of the sulfur compounds from the tail gas, all of the sulfur-bearing species must first be converted to H_2S. Then, the resulting H_2S is absorbed into a solvent and the clean gas vented or recycled for further processing. The clean gas resulting from the hydrolysis step can undergo further cleanup in a dedicated absorption unit or be integrated with an upstream AGR unit. The latter option is particularly suitable with physical absorption solvents. The approach of treating the tail gas in a dedicated amine absorption unit and recycling the resulting acid gas to the Claus plant is the one used by the Shell Claus Off-gas Treating (SCOT) process. With tail gas treatment, Claus plants can achieve overall removal efficiencies in excess of 99.9 percent.

In the case of IGCC applications, the tail gas from the Claus plant can be catalytically hydrogenated and then recycled back into the system with the choice of location being technology dependent, or it can be treated with a SCOT-type process. The Shell cases in this report all use a catalytic hydrogenation step with tail gas recycle to the gasifier. The Shell Puertollano plant treats the tail gas in a similar manner, but the recycle endpoint is not specified [51].

Flare Stack

A self-supporting, refractory-lined, carbon steel flare stack is typically provided to combust and dispose of unreacted gas during startup, shutdown, and upset conditions. However, in the three IGCC cases a flare stack was provided for syngas dumping during startup and shutdown. This flare stack eliminates the need for a separate Claus plant flare.

3.1.9 Slag Handling

The slag handling system conveys, stores, and disposes of slag removed from the gasification process. Spent material drains from the gasifier bed into a water bath in the bottom of the gasifier vessel. A slag crusher receives slag from the water bath and grinds the material into pea-sized fragments. A slag/water slurry that is between 5 and 10 percent solids leaves the gasifier pressure boundary through the use of lockhoppers to a series of dewatering bins.

The slag is dewatered, the water is clarified and recycled and the dried slag is transferred to a storage area for disposal. The specifics of slag handling vary among different gasification technologies regarding how the water is separated and the end uses of the water recycle streams.

In this study the slag bins were sized for a nominal holdup capacity of 72 hours of full-load operation. At periodic intervals, a convoy of slag-hauling trucks will transit the unloading station underneath the hopper and remove a quantity of slag for disposal. While the slag is suitable for use as a component of road paving mixtures, it was assumed in this study that the slag would be landfilled at a specified cost.

3.1.10 Power Island

Combustion Turbine

The gas turbine generator selected for this application is representative of the advanced F Class turbines. This machine is an axial flow, single spool, and constant speed unit, with variable inlet guide vanes. The turbine includes advanced bucket cooling techniques, compressor aerodynamic design and advanced alloys, enabling a higher firing temperature than the previous generation machines. The standard production version of this machine is fired with natural gas and is also commercially offered for use with IGCC derived syngas, although only earlier versions of the turbine are currently operating on syngas. For the purposes of this study, it was assumed that the advanced F Class turbine will be commercially available to support a 2015 startup date on both conventional and high hydrogen content syngas representative of the cases with CO_2 capture. High H_2 fuel combustion issues like flame stability, flashback and NOx formation were assumed to be solved in the time frame needed to support deployment. However, because these are first-of-a-kind applications, process contingencies were included in the cost estimates as described in Section 2.7. Performance typical of an advanced F class turbine on natural gas at ISO conditions is presented in Exhibit 3-15.

Exhibit 3-15 Advanced F Class Combustion Turbine Performance Characteristics Using Natural Gas

	Advanced F Class
Firing Temperature Class, °C (°F)	1371+ (2500+)
Airflow, kg/s (lb/s)	431 (950)
Pressure Ratio	18.5
NOx Emissions, ppmv	25
Simple Cycle Output, MW	185
Combined cycle performance	
Net Output, MW	280
Net Efficiency (LHV), %	57.5
Net Heat Rate (LHV), kJ/kWh (Btu/kWh)	6,256 (5,934)

In this service, with syngas from an IGCC plant, the machine requires some modifications to the burner and turbine nozzles in order to properly combust the low-Btu gas and expand the combustion products in the turbine section of the machine.

The modifications to the machine include some redesign of the original can-annular combustors. A second modification involves increasing the nozzle areas of the turbine to accommodate the mass and volume flow of low-Btu fuel gas combustion products, which are increased relative to those produced when firing natural gas. Other modifications include rearranging the various auxiliary skids that support the machine to accommodate the spatial requirements of the plant general arrangement. The generator is a standard hydrogen-cooled machine with static exciter.

Combustion Turbine Package Scope of Supply

The combustion turbine (CT) is typically supplied in several fully shop-fabricated modules, complete with all mechanical, electrical and control systems as required for CT operation. Site CT installation involves module inter-connection, and linking CT modules to the plant systems. The CT package scope of supply for combined cycle application, while project specific, does not vary much from project-to-project. The typical scope of supply is presented in Exhibit 3-16.

Exhibit 3-16 Combustion Turbine Typical Scope of Supply

	System	System Scope
1.0	**ENGINE ASSEMBLY**	Coupling to Generator, Dry Chemical Exhaust Bearing Fire Protection System, Insulation Blankets, Platforms, Stairs and Ladders
1.1	Engine Assembly with Bedplate	Variable Inlet Guide, Vane System Compressor, Bleed System, Purge Air System, Bearing Seal Sir System, Combustors, Dual Fuel Nozzles Turbine Rotor Air Cooler
1.2	Walk-in acoustical enclosure	HVAC, Lighting, and Low Pressure CO_2 Fire Protection System
2.0	**MECHANICAL PACKAGE**	HVAC and Lighting, Air Compressor for Pneumatic System, Low Pressure CO_2 Fire Protection System
2.1 2.2	Lubricating Oil System and Control Oil System	Lube Oil Reservoir, Accumulators, 2x100% AC Driven Oil Pumps DC Emergency Oil Pump with Starter, 2x100% Oil Coolers, Duplex Oil Filter, Oil Temperature and Pressure Control Valves, Oil Vapor Exhaust Fans and Demister Oil Heaters Oil Interconnect Piping (SS and CS) Oil System Instrumentation Oil for Flushing and First Filling
3.0	**ELECTRICAL PACKAGE**	HVAC and Lighting, AC and DC Motor Control Centers, Generator Voltage Regulating Cabinet, Generator Protective Relay Cabinet, DC Distribution Panel, Battery Charger, Digital Control System with Local Control Panel (all control and monitoring functions as well as data logger and sequence of events recorder), Control System Valves and Instrumentation Communication link for interface with plant DCS Supervisory System, Bentley Nevada Vibration Monitoring System, Low Pressure CO_2 Fire Protection System, Cable Tray and Conduit Provisions for Performance Testing including Test Ports, Thermowells, Instrumentation and DCS interface cards

	System	System Scope
4.0	**INLET AND EXHAUST SYSTEMS**	Inlet Duct Trash Screens, Inlet Duct and Silencers, Self Cleaning Filters, Hoist System For Filter Maintenance, Evaporative Cooler System, Exhaust Duct Expansion Joint, Exhaust Silencers Inlet and Exhaust Flow, Pressure and Temperature Ports and Instrumentation
5.0	**FUEL SYSTEMS**	
5.1	Fuel Syngas System	Gas Valves Including Vent, Throttle and Trip Valves Gas Filter/Separator Gas Supply Instruments and Instrument Panel
5.2	Backup Fuel System	Specific to backup fuel type
6.0	**STARTING SYSTEM**	Enclosure, Starting Motor or Static Start System, Turning Gear and Clutch Assembly, Starting Clutch Torque Converter
7.0	**GENERATOR**	Static or Rotating Exciter (Excitation transformer to be included for a static system), Line Termination Enclosure with CTs, VTs, Surge Arrestors, and Surge Capacitors, Neutral Cubicle with CT, Neutral Tie Bus, Grounding Transformer, and Secondary Resistor, Generator Gas Dryer, Seal Oil System (including Defoaming Tank, Reservoir, Seal Oil Pump, Emergency Seal Oil Pump, Vapor Extractor, and Oil Mist Eliminator), Generator Auxiliaries Control Enclosure, Generator Breaker, Iso-Phase bus connecting generator and breaker, Grounding System Connectors
7.1	Generator Cooling	TEWAC System (including circulation system, interconnecting piping and controls), or Hydrogen Cooling System (including H_2 to Glycol and Glycol to Air heat exchangers, liquid level detector circulation system, interconnecting piping and controls)
8.0	**Miscellaneous**	Interconnecting Pipe, Wire, Tubing and Cable, Instrument Air System Including Air Dryer, On Line and Off Line Water Wash System, LP CO_2 Storage Tank, Drain System, Drain Tanks, Coupling, Coupling Cover and Associated Hardware

CT Firing Temperature Control Issue for Low Calorific Value Fuel

A gas turbine when fired on low calorific value syngas has the potential to increase power output due to the increase in flow rate through the turbine. The higher turbine flow and moisture content of the combustion products can contribute to overheating of turbine components, affect rating criteria for the parts lives, and require a reduction in syngas firing temperatures (compared to the natural gas firing) to maintain design metal temperature [52]. Uncontrolled syngas firing temperature could result in more than 50 percent life cycle reduction of stage 1 buckets. Control systems for syngas applications include provisions to compensate for these effects by maintaining virtually constant generation output for the range of the specified ambient conditions. Inlet guide vanes (IGV) and firing temperature are used to maintain the turbine output at the maximum torque rating, producing a flat rating up to the IGV full open position. Beyond the IGV full open position, flat output may be extended to higher ambient air temperatures by steam/nitrogen injection.

In the three Shell IGCC cases, the turbine inlet temperature ranged from 1,348°C (2,459°F) in the non-capture case to 1,322°C (2,412°F) in the 90 percent capture case. The outlet temperature was 592°C (1,098°F) in the non-capture case, 563°C (1,046°F) in the 90 percent capture case and 577°C (1,070°F) for the 1,100 lb CO_2/net-MWh case. The H_2O content of the combustion products is low, 5.2 volume percent (vol%), in the non-capture cases and increases up to 12 vol% in the 90 percent capture case.

Combustion Turbine Syngas Fuel Requirements

Typical fuel specifications and contaminant levels for successful combustion turbine operation are provided in reference [53] and presented for F Class machines in Exhibit 3-17 and Exhibit 3-18. The vast majority of published CT performance information is specific to natural gas operation. Turbine performance using syngas requires vendor input as was obtained for these cases.

Normal Operation

Inlet air is compressed in a single spool compressor to a pressure ratio of approximately 16:1. This pressure ratio was vendor specified and less than the 18.5:1 ratio used in natural gas applications. The majority of compressor discharge air remains on-board the machine and passes to the burner section to support combustion of the syngas. Compressed air is also used in burner, transition, and film cooling services. About 5.7 percent of the compressor air is extracted and integrated with the air supply of the ASU in non-carbon capture cases. It may be technically possible to integrate the CT and ASU in CO_2 capture cases as well; however, in this study integration was not recommended by the CT vendor and is considered only for non-carbon capture cases.

Pressurized syngas is combusted in several (14) parallel diffusion combustors and syngas dilution is used to limit NOx formation. As described in Section 3.1.2 nitrogen from the ASU is used as the primary diluent followed by syngas humidification and finally by steam dilution, if necessary, to achieve an LHV of 4.3-4.8 MJ/Nm^3 (115-129 Btu/scf). In the three IGCC cases discussed in this report, nitrogen dilution alone was sufficient for the capture cases, but humidification was necessary for the non-capture case. The advantages of using nitrogen as the primary diluent include:

- Nitrogen from the ASU is already partially compressed and using it for dilution eliminates wasting the compression energy.
- Limiting the water content reduces the need to de-rate firing temperature, particularly in the high-hydrogen (CO_2 capture) cases.

Exhibit 3-17 Typical Fuel Specification for F-Class Machines

	Max	Min
LHV, kJ/m^3 (Btu/scf)	None	3.0 (100)
Gas Fuel Pressure, MPa (psia)	3.1 (450)	
Gas Fuel Temperature, °C (°F)	(1)	Varies with gas pressure (2)
Flammability Limit Ratio, Rich-to-Lean, Volume Basis	(3)	2:2.1
Sulfur	(4)	

Notes:

1. The maximum fuel temperature is defined in reference [54]
2. To ensure that the fuel gas supply to the gas turbine is 100 percent free of liquids the minimum fuel gas temperature must meet the required superheat over the respective dew point. This requirement is independent of the hydrocarbon and moisture concentration. Superheat calculation shall be performed as described in GEI-4140G [53].
3. Maximum flammability ratio limit is not defined. Fuel with flammability ratio significantly larger than those of natural gas may require start-up fuel
4. The quantity of sulfur in syngas is not limited by specification. Experience has shown that fuel sulfur levels up to 1 percent by volume do not significantly affect oxidation/corrosion rates.

There are some disadvantages to using nitrogen as the primary diluent, and these include:

- There is a significant auxiliary power requirement to further compress the large nitrogen flow from the ASU pressure to two pressure levels at the CT (2.7 and 3.2 MPa [384 and 469 psia]).
- The low quality heat used in the syngas humidification process does not provide significant benefit to the process in other applications.
- Nitrogen is not as efficient as water in limiting NOx emissions

It is not clear that one dilution method provides a significant advantage over the other. However, in this study nitrogen was chosen as the primary diluent based on suggestions by turbine industry experts.

Hot combustion products are expanded in the three-stage turbine-expander. Given the assumed ambient conditions, back-end loss, and HRSG pressure drop, the CT exhaust temperature for the capture cases is nominally 563°C (1,046°F) for Case 3 and 577°C (1,070°F) for Case 2 and 592°C (1,098°F) for the non-capture Case 1.

Gross turbine power, as measured prior to the generator terminals, is 232 MW at ISO conditions. Turbine output was reduced for all cases due to the site elevation. The CT generator is a standard hydrogen-cooled machine with static exciter.

Exhibit 3-18 Allowable Gas Fuel Contaminant Level for F-Class Machines

	Turbine Inlet Limit, ppbw	Fuel Limit, ppmw		
		Turbine Inlet Flow/Fuel Flow		
		50	*12*	*4*
Lead	20	1.0	0.240	.080
Vanadium	10	0.5	0.120	0.040
Calcium	40	2.0	0.480	0.160
Magnesium	40	2.0	0.480	0.160
Sodium + Potassium				
Na/K = 28 (1)	20	1.0	0.240	0.080
Na/K = 3	10	0.5	0.120	0.40
Na/K ≤ 1	6	0.3	0.072	0.024
Particulates Total (2)	600	30	7.2	2.4
Above 10 microns	6	0.3	0.072	0.024

Notes:

1. Na/K=28 is nominal sea salt ratio
2. The fuel gas delivery system shall be designed to prevent generation or admittance of solid particulate to the gas turbine gas fuel system

The power output of the combustion turbine increases slightly with increased carbon capture primarily because of the increased hydrogen content of the syngas. The higher hydrogen concentration results in a higher water concentration after the combustor, which increases the specific heat of the flue gas stream. The higher specific heat more than offsets the small decrease in mass flow rate that occurs in the carbon capture cases and the net result is a 1.2 percent power output increase in the 1,100 lb CO_2/net-MWh emission rate case and a 2.2 percent increase in the 90 percent capture case relative to the non-capture case.

3.1.11 Steam Generation Island

Heat Recovery Steam Generator

The heat recovery steam generator (HRSG) is a horizontal gas flow, drum-type, multi-pressure design that is matched to the characteristics of the gas turbine exhaust when firing medium-Btu gas. High-temperature flue gas exiting the CT is conveyed through the HRSG to recover the large quantity of thermal energy that remains. Flue gas travels through the HRSG gas path and exits at 132°C (270°F) in all three Shell IGCC cases.

The high pressure (HP) drum produces steam at main steam pressure, while the intermediate pressure (IP) drum produces process steam and turbine dilution steam, if required. The HRSG drum pressures are nominally 12.4/2.9 MPa (1800/420 psia) for the HP/IP turbine sections, respectively. In addition to generating and superheating steam, the HRSG performs reheat duty for the cold/hot reheat steam for the steam turbine, provides condensate and feedwater heating, and also provides deaeration of the condensate.

Natural circulation of steam is accomplished in the HRSG by utilizing differences in densities due to temperature differences of the steam. The natural circulation HRSG provides the most cost-effective and reliable design.

The HRSG drums include moisture separators, internal baffles, and piping for feedwater/steam. All tubes, including economizers, superheaters, and headers and drums, are equipped with drains.

Safety relief valves are furnished in order to comply with appropriate codes and ensure a safe work place.

Superheater, boiler, and economizer sections are supported by shop-assembled structural steel. Inlet and outlet duct is provided to route the gases from the gas turbine outlet to the HRSG inlet and the HRSG outlet to the stack. A diverter valve is included in the inlet duct to bypass the gas when appropriate. Suitable expansion joints are also included.

Steam Turbine Generator and Auxiliaries

The steam turbine consists of an HP section, an IP section, and one double-flow low pressure (LP) section, all connected to the generator by a common shaft. The HP and IP sections are contained in a single-span, opposed-flow casing, with the double-flow LP section in a separate casing. The LP turbine has a last stage bucket length of 76 cm (30 in).

Main steam from the HRSG and gasifier island is combined in a header, and then passes through the stop valves and control valves and enters the turbine at either 12.4 MPa/566°C (1800 psig/1050°F) for the non-carbon capture cases, or 12.4 MPa/538°C (1800 psig/1000°F) for the carbon capture cases. The steam initially enters the turbine near the middle of the high-pressure span, flows through the turbine, and returns to the HRSG for reheating. The reheat steam flows through the reheat stop valves and intercept valves and enters the IP section at 3.2 MPa/566°C (467 psia/1050°F) for the non-carbon capture cases or 3.2 MPa/538°C (467 psia/1000°F) for the carbon capture cases. After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the LP section. The steam divides into two paths and flows through the LP sections, exhausting downward into the condenser.

Turbine bearings are lubricated by a closed-loop, water-cooled, pressurized oil system. The oil is contained in a reservoir located below the turbine floor. During startup or unit trip an emergency oil pump mounted on the reservoir pumps the oil. When the turbine reaches 95 percent of synchronous speed, the main pump mounted on the turbine shaft pumps oil. The oil flows through water-cooled heat exchangers prior to entering the bearings. The oil then flows through the bearings and returns by gravity to the lube oil reservoir.

Turbine shafts are sealed against air in-leakage or steam blowout using a modern positive pressure variable clearance shaft sealing design arrangement connected to a low-pressure steam seal system. During startup, seal steam is provided from the main steam line. As the unit increases load, HP turbine gland leakage provides the seal steam. Pressure-regulating valves control the gland header pressure and dump any excess steam to the condenser. A steam packing exhauster maintains a vacuum at the outer gland seals to prevent leakage of steam into the turbine room. Any steam collected is condensed in the packing exhauster and returned to the condensate system.

The generator is a hydrogen-cooled synchronous type, generating power at 24 kV. A static, transformer type exciter is provided. The generator is cooled with a hydrogen gas recirculation system using fans mounted on the generator rotor shaft. The heat absorbed by the gas is removed as it passes over finned tube gas coolers mounted in the stator frame. Gas is prevented from escaping at the rotor shafts by a closed-loop oil seal system. The oil seal system consists of storage tank, pumps, filters, and pressure controls, all skid-mounted.

The steam turbine generator is controlled by a triple-redundant, microprocessor-based electro-hydraulic control system. The system provides digital control of the unit in accordance with programmed control algorithms, color CRT operator interfacing, and datalink interfaces to the balance-of-plant DCS, and incorporates on-line repair capability.

Condensate System

The condensate system transfers condensate from the condenser hotwell to the deaerator, through the gland steam condenser, gasifier, and the low-temperature economizer section in the HRSG. The system consists of one main condenser; two 50 percent capacity, motor-driven, vertical condensate pumps; one gland steam condenser; and a low-temperature tube bundle in the HRSG. Condensate is delivered to a common discharge header through separate pump discharge lines, each with a check valve and a gate valve. A common minimum flow recirculation line discharging to the condenser is provided to maintain minimum flow requirements for the gland steam condenser and the condensate pumps.

Feedwater System

The function of the feedwater system is to pump the various feedwater streams from the deaerator storage tank in the HRSG to the respective steam drums. Two 50 percent capacity boiler feed pumps are provided for each of three pressure levels, HP, IP, and LP. Each pump is provided with inlet and outlet isolation valves, and outlet check valve. Minimum flow recirculation to prevent overheating and cavitation of the pumps during startup and low loads is provided by an automatic recirculation valve and associated piping that discharges back to the deaerator storage tank. Pneumatic flow control valves control the recirculation flow.

The feedwater pumps are supplied with instrumentation to monitor and alarm on low oil pressure, or high bearing temperature. Feedwater pump suction pressure and temperature are also monitored. In addition, the suction of each boiler feed pump is equipped with a startup strainer.

Main and Reheat Steam Systems

The function of the main steam system is to convey main steam generated in the synthesis gas cooler (SGC) and HRSG from the HRSG superheater outlet to the HP turbine stop valves. The function of the reheat system is to convey steam from the HP turbine exhaust to the HRSG reheater, and to the turbine reheat stop valves.

Main steam at approximately 12.4 MPa/566°C (1800 psig/1050°F) (non-carbon capture cases) or 12.4 MPa/538°C (1800 psig/1000°F) (carbon capture cases) exits the HRSG superheater through a motor-operated stop/check valve and a motor-operated gate valve, and is routed to the HP turbine. Cold reheat steam at approximately 3.45 MPa/352-376°C (500 psia/666-708°F) exits the HP turbine, flows through a motor-operated isolation gate valve, to the HRSG reheater. Hot reheat steam at approximately 3.2 MPa/566°C (467 psia/1050°F) for non-carbon capture cases and 3.2 MPa/538°C (467 psia/1000°F) for carbon capture cases exits the HRSG reheater through a motor-operated gate valve and is routed to the IP turbines.

Steam piping is sloped from the HRSG to the drip pots located near the steam turbine for removal of condensate from the steam lines. Condensate collected in the drip pots and in low-point drains is discharged to the condenser through the drain system.

Steam flow is measured by means of flow nozzles in the steam piping. The flow nozzles are located upstream of any branch connections on the main headers.

Safety valves are installed to comply with appropriate codes and to ensure the safety of personnel and equipment.

Circulating Water System

Exhaust steam from the steam turbine is split 50/50 to a surface condenser cooled with cooling water and to an air-cooled condenser used ambient air and forced convection. A decision to use a parallel wet/dry cooling system was based primarily on the plans for the Xcel Energy Comanche 3 PC plant currently under construction, and the desire to reduce the plant water requirement. Parallel cooling has less of a performance impact on combined cycle systems than on PC systems; and with the relatively low ambient temperature, the performance impact from the parallel cooling, as compared to wet cooling, is minor.

The major impact of parallel cooling is a significant reduction in water requirement when compared to a wet cooling system. This impact is included in the water balance presented later in this report.

With this cooling system and the specific ambient temperature, a condenser pressure of 0.005 MPa (0.698 psia) (condensing temperature of 32°C [90°F]) is used in the model.

The circulating water system is a closed-cycle cooling water system that supplies cooling water to the surface condenser to condense one-half of the main turbine exhaust steam. The system also supplies cooling water to the AGR plant as required, and to the auxiliary cooling system. The auxiliary cooling system is a closed-loop process that utilizes a higher quality water to remove heat from compressor intercoolers, oil coolers and other ancillary equipment and transfers that heat to the main circulating cooling water system in plate and frame heat exchangers. The heat transferred to the circulating water in the surface condenser and other applications is removed by a mechanical draft cooling tower.

The system consists of two 50 percent capacity vertical circulating water pumps, a mechanical draft evaporative cooling tower, and carbon steel cement-lined interconnecting piping. The pumps are single-stage vertical pumps. The piping system is equipped with butterfly isolation valves and all required expansion joints. The cooling tower is a multi-cell, counterflow mechanical draft cooling tower.

The surface condenser is a single-pass, horizontal type with divided water boxes. There are two separate circulating water circuits in each box. One-half of the condenser can be removed from service for cleaning or for plugging tubes. This can be done during normal operation at reduced load. The air-cooled condenser utilizes ambient air and forced convection across tube bundles to condense the balance of the turbine exhaust steam.

Both condensers are equipped with an air extraction system to evacuate the condenser steam space for removal of non-condensable gases during steam turbine operation and to rapidly reduce the condenser pressure from atmospheric pressure before unit startup and admission of steam to the condenser.

Raw Water, Fire Protection, and Cycle Makeup Water Systems

The raw water system supplies cooling tower makeup, cycle makeup, service water and potable water requirements. The water source is 50 percent from municipal sources and 50 percent from groundwater. Booster pumps within the plant boundary provide the necessary pressure.

The fire protection system provides water under pressure to the fire hydrants, hose stations, and fixed water suppression system within the buildings and structures. The system consists of pumps, underground and aboveground supply piping, distribution piping, hydrants, hose stations, spray systems, and deluge spray systems. One motor-operated booster pump is supplied on the intake structure of the cooling tower with a diesel engine backup pump installed on the water inlet line.

The cycle makeup water system provides high quality demineralized water for makeup to the HRSG cycle, for steam injection ahead of the water gas shift reactors in CO_2 capture cases, and for injection steam to the auxiliary boiler for control of NOx emissions, if required.

The cycle makeup system consists of two 100 percent trains, each with a full-capacity activated carbon filter, primary cation exchanger, primary anion exchanger, mixed bed exchanger, recycle pump, and regeneration equipment. The equipment is skid-mounted and includes a control panel and associated piping, valves, and instrumentation.

3.1.12 Accessory Electric Plant

The accessory electric plant consists of switchgear and control equipment, generator equipment, station service equipment, conduit and cable trays, and wire and cable. It also includes the main power transformer, all required foundations, and standby equipment.

3.1.13 Instrumentation and Control

An integrated plant-wide distributed control system (DCS) is provided. The DCS is a redundant microprocessor-based, functionally distributed control system. The control room houses an array of multiple video monitor (CRT) and keyboard units. The CRT/keyboard units are the primary interface between the generating process and operations personnel. The DCS incorporates plant monitoring and control functions for all the major plant equipment. The DCS is designed to be operational and accessible 99.5 percent of the time it is required (99.5 percent availability). The plant equipment and the DCS are designed for automatic response to load changes from minimum load to 100 percent. Startup and shutdown routines are manually implemented, with operator selection of modular automation routines available. The exception to this, and an important facet of the control system for gasification, is the critical controller system, which is a part of the license package from the gasifier supplier and is a dedicated and distinct hardware segment of the DCS.

This critical controller system is used to control the gasification process. The partial oxidation of the fuel feed and oxygen feed streams to form a syngas product is a stoichiometric, temperature- and pressure-dependent reaction. The critical controller utilizes a redundant microprocessor executing calculations and dynamic controls at 100- to 200-millisecond intervals. The enhanced execution speeds as well as evolved predictive controls allow the critical controller to mitigate process upsets and maintain the reactor operation within a stable set of operating parameters.

4. GREENFIELD IGCC CASES (CASES 1-3)

Revision 2 Updates

The modeling updates made to the IGCC cases consisted of the following:

- *Eliminated the ambient heat loss from the gasifier (previously was 2.7 percent of the heat input to the gasifier)*
- *Added syngas recycle to the CO_2 capture cases*
- *Added heat recovery to simulate the jacketed transfer duct between the gasifier and quench vessel (capture cases) or between the gasifier and the syngas cooler (non-capture case)*
- *Relocated the sulfur plant tail gas recycle stream to upstream of the AGR instead of to the gasifier*
- *Updated the combustion turbine model to be more predictive over the range of performance estimates*
- *Updated process heat integration to take advantage of the transfer duct heat recovery and account for the syngas recycle stream in the capture cases*

This section contains an evaluation of plant designs for Cases 1 through 3, which are based on the Shell Global Solutions (Shell) gasifier. These three cases are very similar in terms of process, equipment, scope and arrangement, except that Case 1 employs a syngas cooler as opposed to the full water quench in Cases 2 and 3. There are no provisions for CO_2 removal in Case 1.

Section 4.1 covers Case 1 (non-capture case) using Montana Rosebud PRB coal and Cases 2 and 3 (CO_2-capture cases) also using Montana Rosebud PRB coal. The cases are organized analogously as follows:

- Process and System Description provides an overview of the technology operation as applied to Cases 1 - 3.
- Key Assumptions is a summary of study and modeling assumptions relevant to Cases 1 - 3.
- Sparing Philosophy is provided for Cases 1 - 3.
- Performance Results provide the main modeling results from Cases 1 - 3, including the performance summary, environmental performance, carbon balance, sulfur balance, water balance, mass and energy balance diagrams and mass and energy balance tables.

- Equipment List provides an itemized list of major equipment for Cases 1 - 3 with account codes that correspond to the cost accounts in the Cost Estimates section.

- Cost Estimates provide a summary of capital and operating costs for Cases 1 - 3.

4.1 SHELL IGCC NON-CAPTURE CASE 1 AND CAPTURE CASES 2 AND 3

4.1.1 Process Description for Non-Capture Case 1

In this section the overall Shell gasification process for Case 1 is described. The system description follows the BFD in Exhibit 4-1 and stream numbers reference the same Exhibit. The stream tables provide process data in Exhibit 4-2 for the numbered streams in the BFD.

Coal Preparation and Feed Systems

Coal receiving and handling is common to all cases and was covered in Section 3.1.1. The receiving and handling subsystem ends at the coal silo. The Shell process uses a dry feed system, which is sensitive to the coal moisture content. Coal moisture consists of two parts, surface moisture and inherent moisture. For coal to flow smoothly through the lock hoppers, the surface moisture must be removed. The PRB coal used in this study contains 25.77 percent total moisture on an as-received basis (stream 9). It was assumed that the PRB coal must be dried to 6 percent moisture to allow for smooth flow through the dry feed system (stream 11).

The raw coal is crushed in the coal mill then delivered to a surge hopper with an approximate 2-hour capacity, which in turn delivers the coal to the coal pre-heater (stream 9). The WTA coal drying process includes a water-heated-, horizontal-, rotary-kiln coal pre-heater, a fluidized bed coal dryer and a water-cooled-, horizontal-, rotary-kiln coal cooler. The moisture driven from the coal in the fluid bed dryer passes through a baghouse for particulate removal and then is split into two streams. The smaller of the two streams is compressed and used as the fluidizing medium in the coal dryer. The bulk of the removed moisture is compressed to 0.66 MPa (96 psia) and the temperature is raised to about 413°C (776°F) in the process. The high temperature vapor passes through internal coils in the dryer to provide the heat to drive off the coal moisture and then exits the dryer as liquid water. The warm water is used in the coal pre-heater before being used as cooling tower makeup water (stream 10). The vapor compressor consumes the vast majority of the WTA process auxiliary load.

The coal is drawn from the surge hoppers and fed through a pressurization lock hopper system to a dense phase pneumatic conveyor, which uses nitrogen from the ASU (stream 6) to convey the coal to the gasifiers.

Gasifier

There are two Shell dry feed, pressurized, upflow, entrained, slagging gasifiers, operating at 4.2 MPa (615 psia) and processing a total of 5,211 tonnes/day (5,744 TPD) of as-received coal. Coal reacts with oxygen (stream 7) and steam (stream 8) at a temperature of 1,454°C (2,650°F) to produce principally hydrogen and carbon monoxide with little carbon dioxide formed (stream 13).

The gasifier includes a refractory-lined water wall that is also protected by molten slag that solidifies on the cooled walls.

Raw Gas Cooling/Particulate Removal

High-temperature heat recovery in each gasifier train is accomplished in three steps, including the gasifier jacket, which cools the syngas by maintaining the reaction temperature at 1454°C (2650°F). The product gas from the gasifier (stream 13) is cooled to 1,093°C (2,000°F) by adding cooled recycled fuel gas (stream 14) and then further cooled in a jacketed duct cooler to 899°C (1,650°F) to lower the temperature below the ash melting point. Syngas then goes through a raw gas cooler, which lowers the gas temperature from 899°C (1,650°F) to 335°C (635°F), and produces high-pressure steam for use in the steam cycle. The syngas is further cooled to 191°C (375°F) by vaporizing high-pressure water and subsequently low-pressure water.

After passing through the raw gas coolers, the syngas passes through a cyclone and a raw gas candle filter where a majority of the fine particles are removed and returned to the gasifier with the coal fuel. The filter consists of an array of ceramic candle elements in a pressure vessel. Fines produced by the gasification system are recirculated to extinction. The ash that is not carried out with the gas forms slag and runs down the interior walls, exiting the gasifier in liquid form. The slag is solidified in a quench tank for disposal (stream 12). Lockhoppers are used to reduce the pressure of the solids from 4.2 MPa (615 psia) to ambient. The syngas scrubber removes additional particulate matter further downstream.

Quench Gas Compressor

About 30 percent of the raw gas from the filter is recycled back to the gasifier as quench gas. A single-stage compressor is utilized to boost the pressure of a cooled fuel gas stream from 4.0 MPa (575 psia) to 4.2 MPa (615 psia) (stream 14) to provide quench gas to cool the gas stream from the gasifier.

Syngas Scrubber/Sour Water Stripper

The raw synthesis gas exiting the ceramic particulate filter at 191°C (375°F) (stream 15) then enters the scrubber for removal of chlorides and remaining particulate. The quench scrubber washes the syngas in a counter-current flow in two packed beds. The syngas leaves the scrubber saturated at a temperature of 98°C (208°F). The quench scrubber removes essentially all traces of entrained particles, principally unconverted carbon, slag, and metals. The bottoms from the scrubber are sent to the slag removal and handling system for processing.

The sour water stripper removes NH_3, SO_2, and other impurities from the waste stream of the scrubber. The sour gas stripper consists of a sour drum that accumulates sour water from the gas scrubber and condensate from synthesis gas coolers. Sour water from the drum flows to the sour stripper, which consists of a packed column with a steam-heated reboiler. Sour gas is stripped from the liquid and sent to the sulfur recovery unit. Remaining water is sent to wastewater treatment.

COS Hydrolysis, Mercury Removal and Acid Gas Removal

H_2S and COS are at significant concentrations, requiring removal for the power plant to achieve the low design level of SO_2 emissions. H_2S is removed in an acid gas removal process; however, because COS is not readily removed, it is first catalytically converted to H_2S in a COS hydrolysis unit.

Following the water scrubber, the gas is reheated to 177°C (350°F) and fed to the COS hydrolysis reactor. The COS in the sour gas is hydrolyzed with steam over a catalyst bed to H_2S, which is more easily removed by the AGR solvent. Before the raw fuel gas can be treated in the AGR process (stream 18), it must be cooled to about 35°C (95°F). During this cooling through a series of heat exchangers, part of the water vapor condenses. This water, which contains some NH_3, is sent to the sour water stripper. The cooled syngas (stream 17) then passes through a carbon bed to remove 95 percent of the Hg (Section 3.1.6).

The Sulfinol process, developed by Shell in the early 1960s, is a combination process that uses a mixture of amines and a physical solvent. The solvent consists of an aqueous amine and sulfolane. Sulfinol-D uses diisopropanolamine (DIPA), while Sulfinol-M uses MDEA. The mixed solvents allow for better solvent loadings at high acid gas partial pressures and higher solubility of COS and organic sulfur compounds than straight aqueous amines. Sulfinol-M was selected for the non-CO_2 capture applications.

The sour syngas is fed directly into an HP contactor. The HP contactor is an absorption column in which the H_2S, COS, CO_2, and small amounts of H_2 and CO are removed from the gas by the Sulfinol solvent. The overhead gas stream from the HP contactor is then washed with water in the sweet gas scrubber before leaving the unit as the feed gas to the sulfur polishing unit.

The rich solvent from the bottom of the HP contactor flows through a hydraulic turbine and is flashed in the rich solvent flash vessel. The flashed gas is then scrubbed in the LP contactor with lean solvent to remove H_2S and COS. The overhead from the LP contactor is flashed in the LP KO drum. This gas can be used as a utility fuel gas, consisting primarily of H_2 and CO, at 0.8 MPa (118 psia) and 38°C (101°F). The solvent from the bottom of the LP contactor is returned to the rich solvent flash vessel.

Hot, lean solvent in the lean/rich solvent exchanger then heats the flashed rich solvent before entering the stripper. The stripper strips the H_2S, COS, and CO_2 from the solvent at low pressure with heat supplied through the stripper reboiler. The acid gas stream to sulfur recovery/tail gas cleanup is recovered as the flash gas from the stripper accumulator. The lean solvent from the bottom of the stripper is cooled in the lean/rich solvent exchanger and the lean solvent cooler. Most of the lean solvent is pumped to the HP contactor. A small amount goes to the LP contactor.

The Sulfinol process removes about 40 percent of the CO_2 along with the H_2S and COS. The acid gas fed to the SRU contains 28 vol% H_2S and 41 vol% CO_2. The CO_2 passes through the SRU, the hydrogenation reactor and is recycled to the gasifier. Since the amount of CO_2 in the syngas is small initially, this does not have a significant effect on the mass flow reaching the gas turbine. However, the costs of the sulfur recovery/tail gas treatment are higher than for a sulfur removal process producing an acid gas stream with a higher sulfur concentration.

Claus Unit

The sulfur recovery unit is a Claus bypass type sulfur recovery unit utilizing oxygen (stream 4) instead of air. The Claus plant produces molten sulfur (stream 20) by reacting approximately one third of the H_2S in the feed to SO_2, then reacting the H_2S and SO_2 to sulfur and water. The combination of Claus technology and tail gas recycle to the gasifier results in an overall sulfur recovery exceeding 99 percent.

Utilizing oxygen instead of air in the Claus plant reduces the overall cost of the sulfur recovery plant. The sulfur plant produces approximately 38 tonnes/day (42 TPD) of elemental sulfur. Feed for each case consists of acid gas from both the acid gas cleanup unit (stream 22) and a vent stream from the sour water stripper in the gasifier section.

In the furnace waste heat boiler steam is generated. This steam is used to satisfy all Claus process preheating and reheating requirements as well as to provide some steam to the medium-pressure steam header. The sulfur condensers produce 0.34 MPa (50 psig) steam for the low-pressure steam header.

Power Block

Clean syngas exiting the Sulfinol absorber (stream 23) is humidified and reheated (stream 24), diluted with nitrogen from the ASU (stream 5), and enters the advanced F Class combustion turbine (CT) burner. The CT compressor provides combustion air to the burner and also 14 percent of the air requirements in the ASU (stream 2). The exhaust gas exits the CT at 591°C (1,095°F) (stream 26) and enters the HRSG where additional heat is recovered until the flue gas exits the HRSG at 132°C (270°F) (stream 27) and is discharged through the plant stack. The steam raised in the HRSG is used to power an advanced, commercially available steam turbine using a nominal 12.4 MPa/566°C/566°C (1800 psig/1050°F/1050°F) steam cycle.

Air Separation Unit (ASU)

The ASU is designed to produce a nominal output of 3,201 tonnes/day (3,529 TPD) of 95 mole percent O_2 for use in the gasifier (stream 7) and sulfur recovery unit (stream 4). The plant is designed with two production trains. The air compressor is powered by an electric motor. Approximately 10,220 tonnes/day (11,265 TPD) of nitrogen are also recovered, compressed, and used as dilution in the gas turbine combustor or as transport nitrogen. Approximately 3.25 percent of the gas turbine air is used to supply approximately 14 percent of the ASU air requirements.

Exhibit 4-1 Case 1: IGCC without CO_2 Capture - Block Flow Diagram

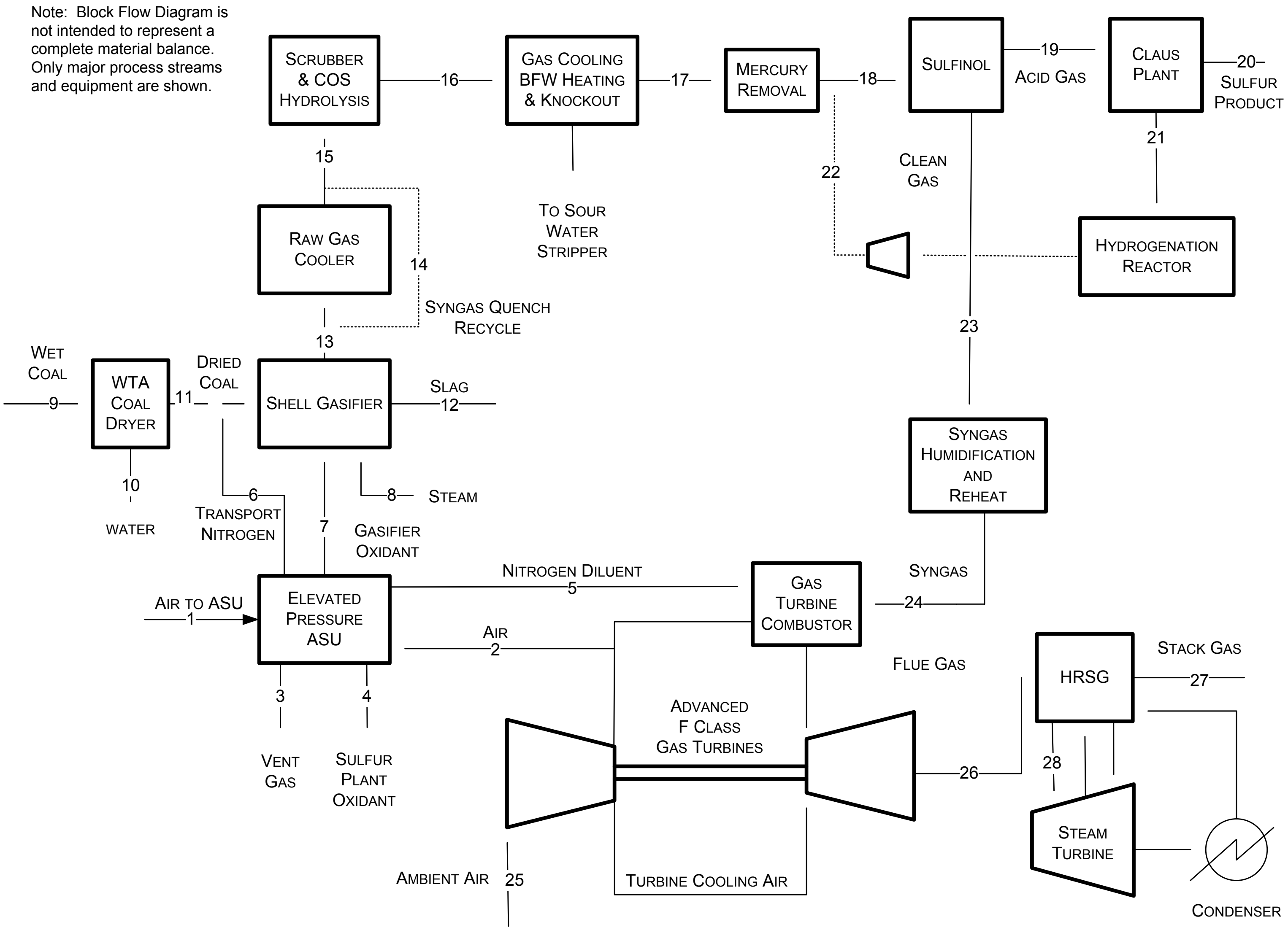


Exhibit 4-2 Case 1: IGCC without CO_2 Capture - Stream Table

	1	2	3	4	5	6	7	8	9	10	11	12	13	14
V-L Mole Fraction														
Ar	0.0093	0.0093	0.0240	0.0318	0.0023	0.0000	0.0318	0.0000	0.0000	0.0000	0.0000	0.0000	0.0104	0.0104
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.6018	0.6018
CO_2	0.0003	0.0003	0.0083	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0312	0.0312
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0005	0.0005
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2613	0.2613
H_2O	0.0071	0.0071	0.1723	0.0000	0.0003	0.0000	0.0000	0.0000	0.0000	1.0000	0.0000	0.0000	0.0289	0.0289
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0030	0.0030
N_2	0.7753	0.7753	0.5912	0.0178	0.9920	1.0000	0.0178	0.0000	0.0000	0.0000	0.0000	0.0000	0.0607	0.0607
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0022	0.0022
O_2	0.2080	0.2080	0.2042	0.9504	0.0054	0.0000	0.9504	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	1.0000	0.0000	0.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	17,257	2,860	796	46	14,411	765	4,099	0	0	2,535	0	0	14,216	6,237
V-L Flowrate (kg/hr)	498,497	82,611	21,912	1,466	404,377	21,433	131,921	0	0	45,667	0	0	306,499	134,463
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	217,133	0	171,466	18,322	0	0
Temperature (°C)	6	411	20	32	196	293	32	---	6	37	71	1,454	1,454	202
Pressure (MPa, abs)	0.08	1.25	0.11	0.86	2.65	5.62	0.86	---	0.08	0.49	0.08	4.24	4.24	4.24
Enthalpy (kJ/kg)^A	16.42	435.50	38.23	26.67	202.67	306.58	26.67	---	---	53.85	---	---	2,291.03	336.24
Density (kg/m^3)	1.0	6.3	1.5	11.0	18.9	32.8	11.0	---	---	981.9	---	---	6.3	22.8
V-L Molecular Weight	28.887	28.887	27.524	32.181	28.061	28.013	32.181	---	---	18.015	---	---	21.561	21.560
V-L Flowrate (lb_{mol}/hr)	38,045	6,305	1,755	100	31,771	1,687	9,038	0	0	5,589	0	0	31,340	13,749
V-L Flowrate (lb/hr)	1,098,999	182,127	48,308	3,232	891,498	47,252	290,836	0	0	100,679	0	0	675,714	296,440
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	478,697	0	378,018	40,392	0	0
Temperature (°F)	42	771	68	90	385	560	90	---	42	98	160	2,650	2,650	396
Pressure (psia)	11.4	181.8	16.4	125.0	385.0	815.0	125.0	---	11.4	71.1	11.1	614.7	614.7	615.0
Enthalpy (Btu/lb)^A	7.1	187.2	16.4	11.5	87.1	131.8	11.5	---	---	23.2	---	---	985.0	144.6
Density (lb/ft^3)	0.061	0.396	0.095	0.687	1.183	2.045	0.687	---	---	61.300	---	---	0.395	1.425
	A - Reference conditions are 32.02 F & 0.089 PSIA													

Exhibit 4-2 Case 1: IGCC without CO_2 Capture - Stream Table (continued)

	15	16	17	18	19	20	21	22	23	24	25	26	27	28
V-L Mole Fraction														
Ar	0.0104	0.0104	0.0107	0.0107	0.0008	0.0000	0.0074	0.0118	0.0108	0.0099	0.0093	0.0090	0.0090	0.0000
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.6018	0.6041	0.6199	0.6131	0.0311	0.0000	0.0337	0.0011	0.6208	0.5658	0.0000	0.0000	0.0000	0.0000
CO_2	0.0312	0.0318	0.0326	0.0380	0.4087	0.0000	0.2984	0.5259	0.0331	0.0302	0.0003	0.0832	0.0832	0.0000
COS	0.0005	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.2613	0.2623	0.2692	0.2669	0.0148	0.0000	0.0161	0.0576	0.2702	0.2462	0.0000	0.0000	0.0000	0.0000
H_2O	0.0289	0.0252	0.0016	0.0016	0.0122	0.0000	0.3918	0.0027	0.0014	0.0899	0.0071	0.0524	0.0524	1.0000
H_2S	0.0030	0.0035	0.0036	0.0036	0.2788	0.0000	0.0018	0.0101	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0607	0.0609	0.0625	0.0660	0.2536	0.0000	0.2464	0.3909	0.0636	0.0579	0.7753	0.7480	0.7480	0.0000
NH_3	0.0022	0.0017	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2080	0.1073	0.1073	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0042	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	14,216	14,161	13,801	13,953	182	0	241	152	13,771	15,110	87,994	108,520	108,520	32,498
V-L Flowrate (kg/hr)	306,499	305,520	299,060	304,386	6,494	0	6,910	5,326	297,892	322,015	2,541,887	3,185,667	3,185,667	585,466
Solids Flowrate (kg/hr)	0	0	0	0	0	1,577	0	0	0	0	0	0	0	0
Temperature (°C)	191	177	35	34	51	173	232	38	34	196	6	591	132	563
Pressure (MPa, abs)	3.96	3.79	3.69	3.62	0.41	0.4	0.335	5.106	3.618	3.216	0.079	0.082	0.079	12.512
Enthalpy (kJ/kg)^A	319.72	292.77	42.48	41.43	60.82	---	893.748	7.747	41.723	461.248	16.424	720.811	216.287	3,507.354
Density (kg/m^3)	21.9	21.6	31.2	30.9	5.6	5,289.3	2.3	77.5	30.6	17.5	1.0	0.3	0.7	35.0
V-L Molecular Weight	21.561	21.575	21.669	21.815	35.747	---	28.713	35.104	21.631	21.311	28.887	29.356	29.356	18.015
V-L Flowrate (lb_{mol}/hr)	31,340	31,220	30,426	30,761	400	0	531	334	30,360	33,312	193,993	239,245	239,245	71,647
V-L Flowrate (lb/hr)	675,714	673,557	659,314	671,055	14,317	0	15,234	11,742	656,739	709,921	5,603,902	7,023,194	7,023,194	1,290,732
Solids Flowrate (lb/hr)	0	0	0	0	0	3,477	0	0	0	0	0	0	0	0
Temperature (°F)	375	351	95	94	124	344	450	100	94	385	42	1,095	270	1,045
Pressure (psia)	574.7	549.7	534.7	524.7	60.0	53.6	48.6	740.5	524.7	466.4	11.4	11.9	11.4	1,814.7
Enthalpy (Btu/lb)^A	137.5	125.9	18.3	17.8	26.1	---	384.2	3.3	17.9	198.3	7.1	309.9	93.0	1,507.9
Density (lb/ft^3)	1.367	1.348	1.945	1.926	0.347	330	0.143	4.838	1.908	1.090	0.061	0.021	0.043	2.187

4.1.2 Process Description for Capture Cases 2 and 3

Cases 2 and 3 are configured to produce electric power with CO_2 capture. The plant configurations are similar to Case 1 with the major difference being the use of a two-stage Selexol AGR plant instead of Sulfinol and subsequent compression of the captured CO_2 stream. The gross power output is constrained by the capacity of the two combustion turbines, and since the CO_2 capture and compression processes increase the auxiliary load on the plant, the net output is significantly reduced relative to Case 1.

The process description for Case 2 and Case 3 is similar to Case 1 with several notable exceptions to accommodate CO_2 capture. A BFD for the 1,100 lb/net-MWh CO_2 capture Case 2 is shown in Exhibit 4-3 and stream tables for Case 2 are shown in Exhibit 4-4. The BFD for Case 3 is shown in Exhibit 4-5 and the associated stream tables are in Exhibit 4-6. Instead of repeating the entire process description, only differences from Case 1 are reported here.

Gasification

The gasification process is the same as Case 1 with the following exceptions:

- The syngas exiting the duct cooler is quenched to 399°C (750°F) with water rather than a syngas cooler to provide a portion of the water required for water gas shift
- Total coal feed (as-received) to the two gasifiers is 5,393 tonnes/day (5,944 TPD) in Case 2 and 5,632 tonnes/day (6,208 TPD) in Case 3 (stream 8)
- The ASU provides 3,283 tonnes/day (3,619 TPD) of 95 mole percent oxygen to the gasifier and Claus plant in the Case 2 and 3,444 tonnes/day (3,796 TPD) in the Case 3 (streams 6 and 3, respectively)

Raw Gas Cooling/Particulate Removal

Following the water quench and particulate removal the syngas is cooled to 260°C (500°F) prior to the syngas scrubber (stream 13) by vaporizing HP BFW and pre-heating IP BFW.

Syngas Scrubber/Sour Water Stripper

Syngas exits the scrubber at 189°C (373°F).

Sour Gas Shift (SGS)

The SGS process was described in Section 3.1.4. In Cases 2 and 3 the syngas after the scrubber is reheated to 232°C (450°F) and then steam (stream 16) is added to adjust the steam:dry gas molar ratio to a minimum of 0.3:1 at the exit of the shift reactor in Case 2 and to 0.47:1 in Case 3. The higher ratio is required in Case 3 to achieve sufficient CO conversion to achieve an overall capture of 90 percent. The hot syngas exiting the first stage of SGS is used to superheat the steam that is added in stream 16. One more stage of SGS (for a total of two) results in 97.4 percent overall conversion of the CO to CO_2 in Case 3. Case 2 uses a single SGS reactor with a bypass stream (stream 32) to achieve a conversion of 49.7 percent CO to CO_2 to reach the 1,100 lb CO_2/net-MWh emission limit. The warm syngas from the second stage of SGS is cooled to 275°C (527°F) by preheating the syngas prior to the first stage of SGS in Case 3. The SGS catalyst also serves to hydrolyze COS thus eliminating the need for a separate COS hydrolysis reactor. The bypass utilized in Case 2 prevents all COS from being hydrolyzed. Therefore, the CO_2 product contains more sulfur than in Case 3. Following the second stage (or first and only in Case 2) of SGS, the syngas is further cooled to 35°C (95°F) prior to the mercury removal beds.

Mercury Removal and Acid Gas Removal

Mercury removal is the same as in Case 1.

The AGR process in Cases 2 and 3 is a two-stage Selexol process where H_2S is removed in the first stage and CO_2 in the second stage of absorption. The process results in three product streams, the clean syngas, a CO_2-rich stream and an acid gas feed to the Claus plant. The feed to the Claus plant in Case 3 contains 16 percent H_2S, 66 percent CO_2, 13 percent H_2, and the balance primarily H_2O. In Case 2 the acid gas contains about 22 percent H_2S, 49 percent CO_2, 12 percent H_2, 11 percent CO, and the balance primarily H_2O. The higher concentration of CO in Case 2 relative to Case 3 is due to the bypass stream around the SGS reactor in each train. The CO_2-rich stream is discussed further in the CO_2 compression section.

CO_2 Compression and Dehydration

CO_2 from the AGR process is generated at two pressure levels. The LP stream is compressed from 0.12 MPa (17 psia) to 1.0 MPa (150 psia) and then combined with the HP stream. The combined stream is further compressed to a supercritical condition at 15.3 MPa (2215 psia) using a multiple-stage, intercooled compressor. During compression, the CO_2 stream is dehydrated to a dewpoint of -40ºC (-40°F). The raw CO_2 stream from the Selexol process contains at least 97.9 percent CO_2. The dehydrated CO_2 (stream 26 in Case 2 and stream 25 in Case 3) is transported to the plant fence line and is sequestration ready. CO_2 TS&M costs were estimated using the methodology described in Section 2.7.

Claus Unit

The Claus plant is the same as Case 1 with the following exception:

- 38 tonnes/day (42 TPD) of sulfur are produced in Case 2 and 41 tonnes/day (45 TPD) in Case 3.

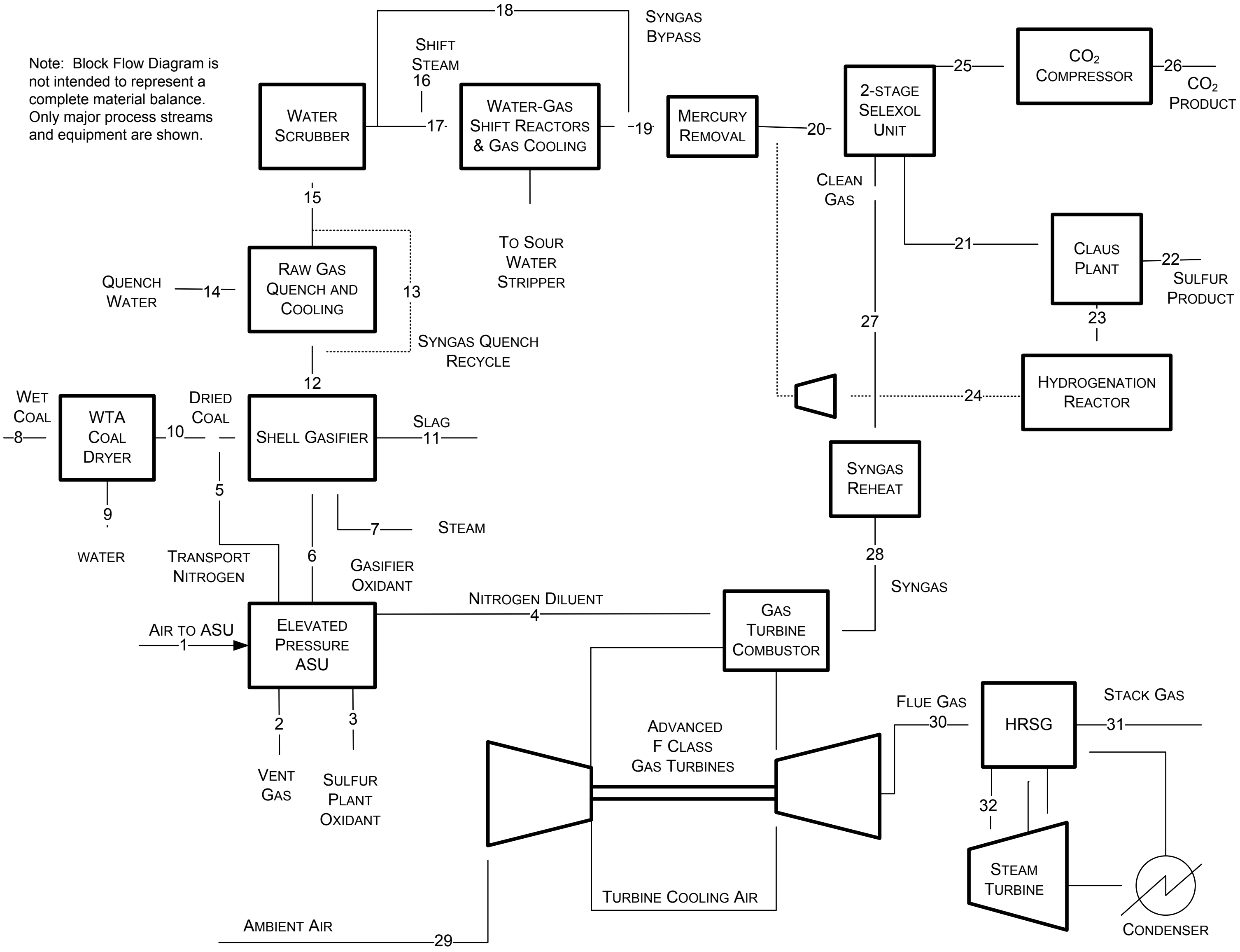
Power Block

Clean syngas from the AGR plant (stream 27 in Case 2 and stream 26 in Case 3) is reheated to 196°C (385°F) using HP boiler feedwater, diluted with nitrogen (stream 4), and then enters the CT burner. The exhaust gas (stream 30 in Case 2 and stream 29 in Case 3) exits the CT at 577°C (1,071°F) in Case 2 and 563°C (1,046°F) in Case 3 and enters the HRSG where additional heat is recovered. The flue gas exits the HRSG at 132°C (270°F) (stream 31 in Case 2 and stream 30 in Case 3) and is discharged through the plant stack. The steam raised in the HRSG is used to power an advanced commercially available steam turbine using a nominal 12.4 MPa/538°C/538°C (1800 psig/1000°F/1000°F) steam cycle. There is no integration between the CT and the ASU in either capture case.

Air Separation Unit

The same elevated pressure ASU is used as in Case 1 except the output is 3,283 tonnes/day (3,619 TPD) of 95 mole percent oxygen and 10,291 tonnes/day (11,344 TPD) of nitrogen in Case 2 and 3,444 tonne/day (3,796 tpd) of 95 mole percent oxygen and 11,068 tonne/day (12,201 tpd) of nitrogen in Case 3.

Exhibit 4-3 Case 2: IGCC with CO_2 Capture to an Emission Limit of 1,100 lb CO_2/net-MWh - Block Flow Diagram



Note: Block Flow Diagram is not intended to represent a complete material balance. Only major process streams and equipment are shown.

Exhibit 4-4 Case 2: IGCC with CO_2 Capture to an Emission Limit of 1,100 lb CO_2/net-MWh - Stream Table

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
V-L Mole Fraction																
Ar	0.0093	0.0185	0.0318	0.0023	0.0000	0.0318	0.0000	0.0000	0.0000	0.0000	0.0000	0.0103	0.0066	0.0000	0.0066	0.0000
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.6040	0.3872	0.0000	0.3872	0.0000
CO_2	0.0003	0.0061	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0295	0.0189	0.0000	0.0189	0.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0005	0.0003	0.0000	0.0003	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2634	0.1689	0.0000	0.1689	0.0000
H_2O	0.0071	0.1296	0.0000	0.0003	0.0000	0.0000	0.0000	0.0000	1.0000	0.0000	0.0000	0.0260	0.3755	1.0000	0.3755	1.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0030	0.0019	0.0000	0.0019	0.0000
N_2	0.7753	0.6930	0.0178	0.9920	1.0000	0.0178	0.0000	0.0000	0.0000	0.0000	0.0000	0.0603	0.0387	0.0000	0.0387	0.0000
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0029	0.0019	0.0000	0.0019	0.0000
O_2	0.2080	0.1528	0.9504	0.0054	0.0000	0.9504	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	1.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	20,634	1,102	45	14,490	792	4,206	0	0	2,623	0	0	14,699	5,913	8,226	22,925	3,570
V-L Flowrate (kg/hr)	596,059	30,474	1,451	406,596	22,179	135,358	0	0	47,257	0	0	316,011	119,741	148,194	464,205	64,317
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	224,690	0	177,433	18,959	0	0	0	0	0
Temperature (°C)	6	19	32	196	293	32	---	6	33	71	1,427	1,427	274	216	260	288
Pressure (MPa, abs)	0.08	0.11	0.86	2.65	5.62	0.86	---	0.08	0.46	0.08	4.24	4.24	4.24	8.27	3.93	4.14
Enthalpy (kJ/kg)^A	16.42	36.76	26.67	202.68	306.58	26.67	---	---	37.62	---	---	2,241.81	1,243.02	904.03	1,220.77	2,956.19
Density (kg/m^3)	1.0	1.5	11.0	18.9	32.8	11.0	---	---	985.4	---	---	6.4	19.1	782	18.2	18.2
V-L Molecular Weight	28.887	27.646	32.181	28.060	28.013	32.181	---	---	18.015	---	---	21.499	20.249	18.015	20.249	18.015
V-L Flowrate (lb_{mol}/hr)	45,490	2,430	99	31,945	1,745	9,273	0	0	5,783	0	0	32,405	13,037	18,135	50,540	7,871
V-L Flowrate (lb/hr)	1,314,085	67,183	3,200	896,392	48,897	298,414	0	0	104,183	0	0	696,686	263,983	326,712	1,023,397	141,795
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	495,356	0	391,173	41,798	0	0	0	0	0
Temperature (°F)	42	65	90	385	560	90	---	42	92	160	2,600	2,600	525	420	500	550
Pressure (psia)	11.4	16.4	125.0	384.0	815.0	125.0	---	11.4	66.8	11.1	614.7	614.7	615.0	1,200.0	569.7	600.0
Enthalpy (Btu/lb)^A	7.1	15.8	11.5	87.1	131.8	11.5	---	---	16.2	---	---	963.8	534.4	388.7	524.8	1,270.9
Density (lb/ft^3)	0.061	0.091	0.687	1.180	2.045	0.687	---	---	61.518	---	---	0.400	1.193	48.817	1.136	1.135

A - Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 4-4 Case 2: IGCC with CO_2 Capture to an Emission Limit of 1,100 lb CO_2/net-MWh - Stream Table (Continued)

	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32
V-L Mole Fraction																
Ar	0.0055	0.0069	0.0082	0.0082	0.0031	0.0000	0.0077	0.0093	0.0003	0.0003	0.0107	0.0107	0.0093	0.0091	0.0091	0.0000
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.3213	0.4050	0.2538	0.2509	0.1079	0.0000	0.1456	0.0075	0.0097	0.0097	0.3280	0.3280	0.0000	0.0000	0.0000	0.0000
CO_2	0.0157	0.0198	0.2500	0.2540	0.4912	0.0000	0.2864	0.5166	0.9790	0.9834	0.0256	0.0256	0.0003	0.0455	0.0455	0.0000
COS	0.0003	0.0003	0.0002	0.0002	0.0058	0.0000	0.0004	0.0000	0.0004	0.0004	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.1401	0.1767	0.4357	0.4340	0.1188	0.0000	0.0657	0.2486	0.0057	0.0058	0.5716	0.5716	0.0000	0.0000	0.0000	0.0000
H_2O	0.4819	0.3469	0.0016	0.0016	0.0444	0.0000	0.4287	0.1380	0.0045	0.0000	0.0001	0.0001	0.0071	0.0789	0.0789	1.0000
H_2S	0.0016	0.0020	0.0026	0.0026	0.2200	0.0000	0.0012	0.0021	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0321	0.0405	0.0480	0.0485	0.0087	0.0000	0.0642	0.0778	0.0004	0.0004	0.0640	0.0640	0.7753	0.7574	0.7574	0.0000
NH_3	0.0015	0.0019	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2080	0.1091	0.1091	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	17,269	8,219	18,491	18,709	218	0	305	252	4,376	4,357	14,114	14,114	87,994	110,250	110,250	23,153
V-L Flowrate (kg/hr)	343,112	167,277	384,368	390,946	7,385	0	8,160	7,198	190,359	190,008	193,201	193,201	2,541,887	3,141,685	3,141,685	417,101
Solids Flowrate (kg/hr)	0	0	0	0	0	1,571	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	242	232	35	35	48	178	232	49	16	54	31	196	6	577	132	549
Pressure (MPa, abs)	3.79	3.79	3.48	3.41	0.16	0.1	0.1	0.073	1.032	15.270	3.238	3.203	0.079	0.082	0.079	12.512
Enthalpy (kJ/kg)^A	1,453.12	1,106.37	43.59	43.04	106.80	---	1,024.4	275.092	6.977	-146.702	60.738	421.700	16.424	759.629	263.644	3,471.595
Density (kg/m^3)	18.1	18.6	28.6	28.2	2.1	5,280	0.5	0.8	20.0	592.7	17.4	11.1	1.0	0.3	0.7	35.8
V-L Molecular Weight	19.869	20.352	20.787	20.897	33.896	---	27	28.558	43.498	43.612	13.688	13.688	28.887	28.496	28.496	18.015
V-L Flowrate (lb_{mol}/hr)	38,072	18,121	40,766	41,245	480	0	673	556	9,648	9,605	31,117	31,117	193,993	243,059	243,059	51,043
V-L Flowrate (lb/hr)	756,433	368,783	847,387	861,888	16,281	0	17,991	15,869	419,671	418,895	425,936	425,936	5,603,902	6,926,230	6,926,230	919,549
Solids Flowrate (lb/hr)	0	0	0	0	0	3,464	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	467	450	95	94	119	352	450	120	60	130	87	385	42	1,071	270	1,021
Pressure (psia)	549.7	549.7	504.7	494.7	23.7	17.3	12.3	10.6	149.7	2,214.7	469.6	464.6	11.4	11.9	11.4	1,814.7
Enthalpy (Btu/lb)^A	624.7	475.7	18.7	18.5	45.9	---	440.4	118.3	3.0	-63.1	26.1	181.3	7.1	326.6	113.3	1,492.5
Density (lb/ft^3)	1.133	1.163	1.783	1.760	0.130	330	0	0.049	1.246	37.001	1.085	0.693	0.061	0.021	0.041	2.236

Exhibit 4-5 Case 3: IGCC with 90% CO_2 Capture - Block Flow Diagram

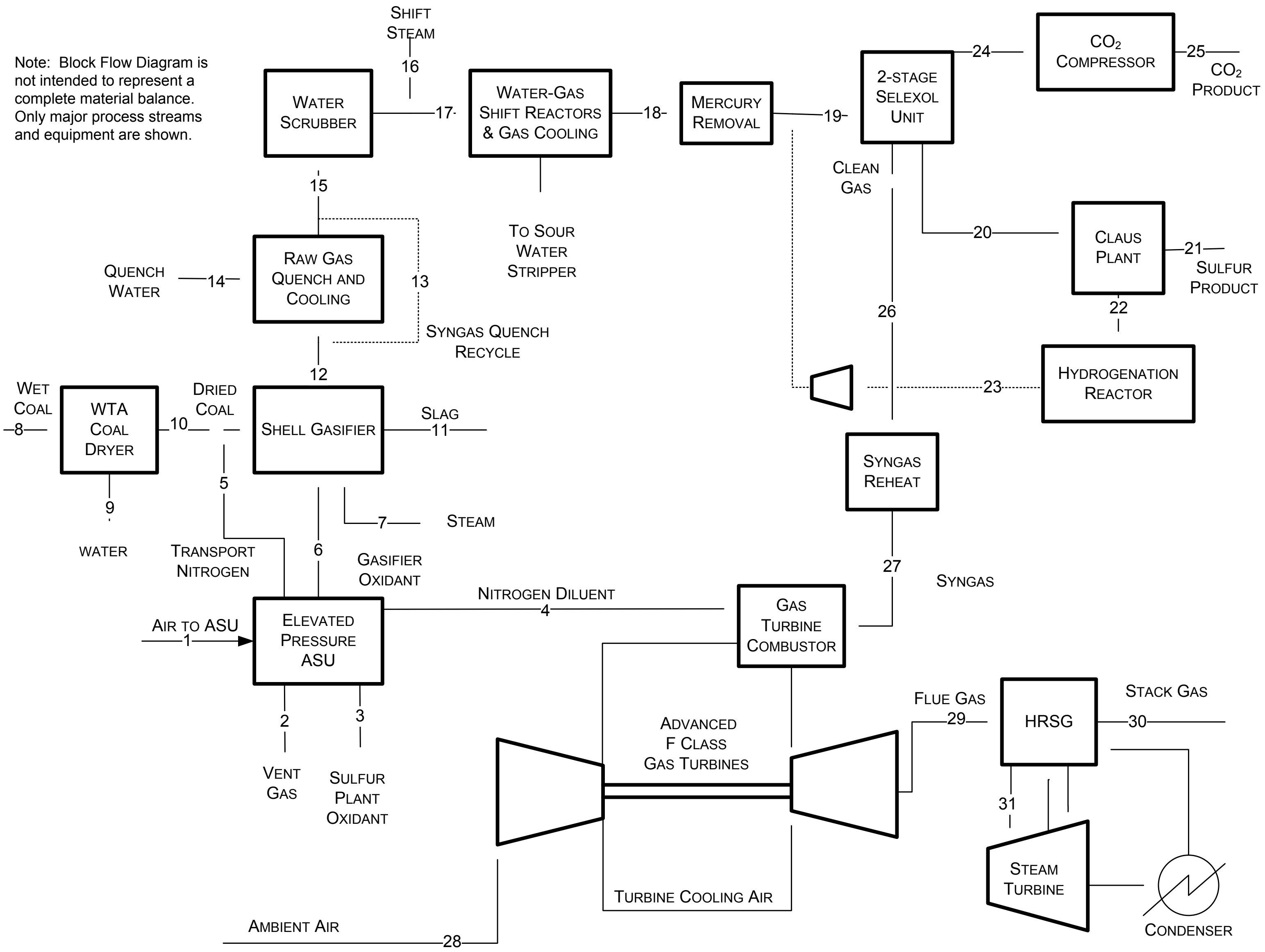


Exhibit 4-6 Case 3: IGCC with 90% CO_2 Capture - Stream Table

	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
V-L Mole Fraction																
Ar	0.0093	0.0272	0.0318	0.0023	0.0000	0.0318	0.0000	0.0000	0.0000	0.0000	0.0000	0.0103	0.0066	0.0000	0.0066	0.0000
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.6026	0.3860	0.0000	0.3860	0.0000
CO_2	0.0003	0.0095	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0308	0.0197	0.0000	0.0197	0.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0005	0.0003	0.0000	0.0003	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2624	0.1681	0.0000	0.1681	0.0000
H_2O	0.0071	0.2005	0.0000	0.0003	0.0000	0.0000	0.0000	0.0000	1.0000	0.0000	0.0000	0.0271	0.3768	1.0000	0.3768	1.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0030	0.0019	0.0000	0.0019	0.0000
N_2	0.7753	0.5294	0.0178	0.9920	1.0000	0.0178	0.0000	0.0000	0.0000	0.0000	0.0000	0.0604	0.0387	0.0000	0.0387	0.0000
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0029	0.0019	0.0000	0.0019	0.0000
O_2	0.2080	0.2334	0.9504	0.0054	0.0000	0.9504	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	1.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg$_{mol}$/hr)	21,640	746	47	15,610	827	4,412	0	0	2,740	0	0	15,353	6,133	8,614	23,967	12,382
V-L Flowrate (kg/hr)	625,129	20,460	1,497	438,022	23,164	141,985	0	0	49,355	0	0	330,662	124,331	155,192	485,853	223,066
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	234,669	0	185,314	19,801	0	0	0	0	0
Temperature (°C)	6	21	32	196	293	32	---	6	33	71	1,427	1,427	274	216	260	288
Pressure (MPa, abs)	0.08	0.11	0.86	2.65	5.62	0.86	---	0.08	0.46	0.08	4.24	4.24	4.24	8.27	3.93	4.14
Enthalpy (kJ/kg)^A	16.42	38.94	26.67	202.68	306.58	26.67	---	---	37.62	---	---	2,242.42	1,244.56	904.03	1,222.34	2,956.19
Density (kg/m^3)	1.0	1.6	11.0	18.9	32.8	11.0	---	---	985.4	---	---	6.4	19.1	782.0	18.2	18.2
V-L Molecular Weight	28.887	27.415	32.181	28.061	28.013	32.181	---	---	18.015	---	---	21.538	20.272	18.015	20.272	18.015
V-L Flowrate (lb$_{mol}$/hr)	47,709	1,645	103	34,414	1,823	9,727	0	0	6,040	0	0	33,847	13,522	18,992	52,839	27,298
V-L Flowrate (lb/hr)	1,378,173	45,107	3,300	965,674	51,068	313,023	0	0	108,810	0	0	728,984	274,103	342,139	1,071,123	491,776
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	517,357	0	408,547	43,654	0	0	0	0	0
Temperature (°F)	42	69	90	385	560	90	---	42	92	160	2,600	2,600	525	420	500	550
Pressure (psia)	11.4	16.4	125.0	384.0	815.0	125.0	---	11.4	66.8	11.1	614.7	614.7	615.0	1,200.0	569.7	600.0
Enthalpy (Btu/lb)^A	7.1	16.7	11.5	87.1	131.8	11.5	---	---	16.2	---	---	964.1	535.1	388.7	525.5	1,270.9
Density (lb/ft^3)	0.061	0.097	0.687	1.180	2.045	0.687	---	---	61.518	---	---	0.401	1.195	48.817	1.138	1.135

A - Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 4-6 Case 3: IGCC with 90% CO_2 Capture - Stream Table (continued)

	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31
V-L Mole Fraction															
Ar	0.0045	0.0066	0.0066	0.0022	0.0000	0.0057	0.0069	0.0002	0.0002	0.0105	0.0105	0.0093	0.0090	0.0090	0.0000
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.2622	0.0099	0.0098	0.0036	0.0000	0.0847	0.0060	0.0002	0.0002	0.0156	0.0156	0.0000	0.0000	0.0000	0.0000
CO_2	0.0134	0.3964	0.4004	0.6618	0.0000	0.4271	0.6199	0.9919	0.9948	0.0489	0.0489	0.0003	0.0090	0.0090	0.0000
COS	0.0002	0.0000	0.0000	0.0001	0.0000	0.0002	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.1141	0.5445	0.5401	0.1264	0.0000	0.0595	0.1695	0.0046	0.0046	0.8623	0.8623	0.0000	0.0000	0.0000	0.0000
H_2O	0.5768	0.0017	0.0017	0.0388	0.0000	0.3743	0.1381	0.0029	0.0000	0.0001	0.0001	0.0071	0.1218	0.1218	1.0000
H_2S	0.0013	0.0022	0.0022	0.1610	0.0000	0.0009	0.0016	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0263	0.0387	0.0390	0.0060	0.0000	0.0474	0.0580	0.0002	0.0002	0.0625	0.0625	0.7753	0.7553	0.7553	0.0000
NH_3	0.0012	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2080	0.1049	0.1049	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	35,289	23,954	24,258	331	0	430	352	8,829	8,803	15,098	15,098	87,994	112,074	112,074	23,733
V-L Flowrate (kg/hr)	689,820	485,653	496,128	11,876	0	12,750	11,342	386,104	385,637	98,149	98,149	2,541,887	3,078,059	3,078,059	427,554
Solids Flowrate (kg/hr)	0	0	0	0	1,703	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	249	35	35	48	175	232	49	16	51	31	196	6	563	132	535
Pressure (MPa, abs)	3.79	3.43	3.36	0.16	0.12	0.1	0.073	1.032	15.270	3.238	3.203	0.079	0.082	0.079	12.512
Enthalpy (kJ/kg)^A	1,705.35	43.90	42.97	94.95	---	851.3	245.007	5.268	-162.321	135.691	896.467	16.424	835.278	343.152	3,435.794
Density (kg/m^3)	17.8	27.7	27.5	2.2	5,284.8	0.6	0.9	20.1	642.0	8.2	5.3	1.0	0.3	0.6	36.7
V-L Molecular Weight	19.548	20.275	20.453	35.932	---	30	32.234	43.733	43.809	6.501	6.501	28.887	27.464	27.464	18.015
V-L Flowrate (lb_{mol}/hr)	77,799	52,809	53,479	729	0	948	776	19,464	19,407	33,286	33,286	193,993	247,081	247,081	52,322
V-L Flowrate (lb/hr)	1,520,792	1,070,681	1,093,776	26,181	0	28,108	25,004	851,213	850,185	216,381	216,381	5,603,902	6,785,957	6,785,957	942,596
Solids Flowrate (lb/hr)	0	0	0	0	3,754	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	480	95	94	119	348	450	120	60	124	87	385	42	1,046	270	996
Pressure (psia)	549.7	497.6	487.6	23.7	17.3	12.3	10.6	149.7	2,214.7	469.6	464.6	11.4	11.9	11.4	1,814.7
Enthalpy (Btu/lb)^A	733.2	18.9	18.5	40.8	---	366.0	105.3	2.3	-69.8	58.3	385.4	7.1	359.1	147.5	1,477.1
Density (lb/ft^3)	1.114	1.731	1.714	0.138	329.917	0	0.055	1.254	40.076	0.514	0.329	0.061	0.020	0.040	2.289

4.1.3 Key System Assumptions

System assumptions for Cases 1 through 3, Shell IGCC using Montana Rosebud PRB coal with and without CO_2 capture, are compiled in Exhibit 4-7.

Exhibit 4-7 Cases 1 - 3 IGCC Plant System Assumptions/ Configuration Matrix

	Case 1	Cases 2 and 3
Gasifier Pressure, MPa (psia)	4.2 (615)	4.2 (615)
O_2:Coal Ratio, kg O_2/kg dried coal	0.77	0.76
Carbon Conversion, %	99.5	99.5
Syngas HHV at Gasifier Outlet, kJ/Nm^3 (Btu/scf)	10,470 (281)	10,546 (283)
Nominal Steam Cycle, MPa/°C/°C (psig/°F/°F)	12.4/566/566 (1800/1050/1050)	12.4/538/538 (1800/1000/1000)
Condenser Pressure, mm Hg (in Hg)	36 (1.4)	36 (1.4)
Combustion Turbine	2x Advanced F Class (Nominal 232 MW output each, reduced by elevation considerations)	2x Advanced F Class (Nominal 232 MW output each, reduced by elevation considerations)
Gasifier Technology	Shell	Shell
Oxidant	95 vol% Oxygen	95 vol% Oxygen
Coal	Montana Rosebud PRB	Montana Rosebud PRB
Coal Feed Moisture Content, %	6	6
COS Hydrolysis	Yes	Yes (Part of WGS)
Water Gas Shift	No	Yes
H_2S Separation	Sulfinol-M	Selexol (1st Stage)
Sulfur Removal, %	99.9	96-99.8
CO_2 Separation	None	Selexol (2nd Stage)
CO_2 Removal	N/A	1,100 lb/net MWh / 90%
Sulfur Recovery	Claus Plant with Tail Gas Treatment / Elemental Sulfur	Claus Plant with Tail Gas Treatment / Elemental Sulfur
Particulate Control	Cyclone, Candle Filter, Scrubber, and AGR Absorber	Cyclone, Candle Filter, Scrubber, and AGR Absorber
Mercury Control	Carbon Bed	Carbon Bed
NOx Control	MNQC (LNB) and N_2 Dilution	MNQC (LNB) and N_2 Dilution

Balance of Plant – All Cases

The balance of plant assumptions are common to all cases and are presented in Exhibit 4-8. Items were also covered in Sections 3.1.10, 3.1.11, 3.1.12 and 3.1.13.

Exhibit 4-8 Balance of Plant Assumptions

Cooling water system	Recirculating Wet Cooling Tower
Fuel and Other storage	
Coal	30 days
Slag	30 days
Sulfur	30 days
Sorbent	30 days
Plant Distribution Voltage	
Motors below 1 hp	110/220 volt
Motors between 1 hp and 250 hp	480 volt
Motors between 250 hp and 5,000 hp	4,160 volt
Motors above 5,000 hp	13,800 volt
Steam and Gas Turbine Generators	24,000 volt
Grid Interconnection Voltage	345 kV
Water and Waste Water	
Makeup Water	The water supply is 50 percent from a local Publicly Owned Treatment Works and 50 percent from groundwater, and is assumed to be in sufficient quantities to meet plant makeup requirements. Makeup for potable, process, and de-ionized (DI) water is drawn from municipal sources
Process Wastewater	Water associated with gasification activity and storm water that contacts equipment surfaces is collected and treated for discharge through a permitted discharge.
Sanitary Waste Disposal	Design includes a packaged domestic sewage treatment plant with effluent discharged to the industrial wastewater treatment system. Sludge is hauled off site. Packaged plant was sized for 5.68 cubic meters per day (1,500 gallons per day)
Water Discharge	Most of the process wastewater is recycled to the cooling tower basin. Blowdown is treated for chloride and metals, and discharged.

4.1.4 Sparing Philosophy

The sparing philosophy for Cases 1 through 3 is provided below. Single trains are utilized throughout with exceptions where equipment capacity requires an additional train. There is no redundancy other than normal sparing of rotating equipment.

The plant design consists of the following major subsystems:

- Two air separation units (2 x 50%).
- Two trains of coal drying and dry feed systems (2 x 50%).
- Two trains of gasification, including gasifier, synthesis gas cooler, cyclone, and barrier filter (2 x 50%).
- Two trains of syngas clean-up process (2 x 50%).
- Two trains of Sulfinol-M acid gas removal in non-capture cases and two trains to two-stage Selexol in CO_2 capture cases (2 x 50%).
- One train of Claus-based sulfur recovery (1 x 100%).
- Two combustion turbine/HRSG tandems (2 x 50%).
- One steam turbine (1 x 100%).

4.1.5 Cases 1 - 3 Performance Results

The non-capture Shell IGCC plant using PRB coal produces a net output of 502 MWe at a net plant efficiency of 41.8 percent (HHV basis). The net output in the 1,100 lb CO_2/net-MWh case is 443 MWe at a net efficiency of 35.6 percent and 401 MWe at a net efficiency of 30.9 percent in the 90 percent capture case.

Overall performance for the three plants is summarized in Exhibit 4-9 which includes auxiliary power requirements. The ASU accounts for approximately 74 percent, 64 percent, and 57 percent of the total auxiliary load in Case 1, Case 2, and Case 3, respectively. The ASU auxiliary load is distributed between the main air compressor, the oxygen compressor, the nitrogen compressor, and ASU auxiliaries. The coal drying process accounts for 7.6 percent of the auxiliary load in the non-capture case and 5 to 6 percent in the capture cases. The cooling water system, including the circulating water pumps and cooling tower fan, and the air-cooled condenser account for about 3 to 5 percent of the auxiliary load in all three cases, and the BFW pumps account for an additional 3 percent in the non-capture case and less than 2 percent in the two capture cases. All other individual auxiliary loads are less than 3 percent of the total.

Exhibit 4-9 Cases 1 - 3 Plant Performance Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	**Case 1**	**Case 2**	**Case 3**
Gas Turbine Power	372,500	377,000	380,600
Steam Turbine Power	240,400	208,000	192,900
TOTAL POWER, kWe	**612,900**	**585,000**	**573,500**
AUXILIARY LOAD SUMMARY, kWe			
Coal Handling	460	470	480
Coal Milling	2,230	2,310	2,410
Slag Handling	480	490	520
WTA Coal Dryer Compressor	7,860	7,910	8,260
WTA Coal Dryer Auxiliaries	510	520	540
Air Separation Unit Auxiliaries	1,000	1,000	1,000
Air Separation Unit Main Air Compressor	46,830	56,000	58,730
Oxygen Compressor	7,220	7,410	7,770
Nitrogen Compressor	26,250	26,420	30,660
CO_2 Compressor	0	13,130	25,960
Boiler Feedwater Pumps	3,650	2,800	3,040
Condensate Pump	180	210	250
Quench Water Pump	0	490	510
Syngas Recycle Compressor	650	780	810
Circulating Water Pumps	1,660	2,150	2,470
Ground Water Pumps	150	220	280
Cooling Tower Fans	1,080	1,400	1,610
Air-Cooled Condenser Fans	2,670	2,740	2,540
Scrubber Pumps	500	310	330
Acid Gas Removal Auxiliaries	220	7,940	16,240
Gas Turbine Auxiliaries	1,000	1,000	1,000
Steam Turbine Auxiliaries	100	100	100
Claus Plant/TGTU Auxiliaries	250	250	250
Claus Plant Tail Gas Recycle Compressor	410	1,020	1,380
Miscellaneous Balance of Plant (Note 1)	3,000	3,000	3,000
Transformer Loss	2,160	2,170	2,220
TOTAL AUXILIARIES, kWe	**110,520**	**142,240**	**172,360**
NET POWER, kWe	**502,380**	**442,760**	**401,140**
Net Plant Efficiency, % (HHV)	41.8%	35.6%	30.9%
Net Plant Heat Rate, kJ/kWh (Btu/kWh)	8,610 (8,160)	10,109 (9,581)	11,653 (11,045)
CONDENSER COOLING DUTY GJ/h (10^6 Btu/h)	**1,192 (1,130)**	**1,213 (1,150)**	**1,129 (1,070)**
CONSUMABLES			
As-Received Coal Feed, kg/h (lb/h)	217,133 (478,697)	224,690 (495,356)	234,669 (517,357)
Thermal Input, kWt	1,201,463	1,243,274	1,298,493
Raw Water Withdrawal, m^3/min (gpm)	6.1 (1,616)	9.1 (2,412)	11.8 (3,124)
Raw Water Consumption, m^3/min (gpm)	4.7 (1,232)	7.2 (1,910)	9.6 (2,544)

Environmental Performance

The environmental targets for emissions of Hg, NOx, SO_2 and particulate matter were presented in Section 2.3. A summary of the plant air emissions for Cases 1 - 3 is presented in Exhibit 4-10.

Exhibit 4-10 Cases 1 - 3 Air Emissions

	Case 1	Case 2	Case 3
kg/GJ (lb/10^6 Btu)			
SO_2	0.001 (0.002)	0.0003 (0.0008)	0.0004 (0.0008)
NO_X	0.026 (0.062)	0.024 (0.056)	0.022 (0.051)
Particulates	0.003 (0.0071)	0.003 (0.0071)	0.003 (0.0071)
Hg	1.51E-7 (3.51E-7)	1.51E-7 (3.51E-7)	1.51E-7 (3.51E-7)
CO_2	92 (214)	49 (115)	9.4 (22)
Tonne/year (tons/year) 80% capacity			
SO_2	30 (33)	11 (12)	12 (13)
NO_X	802 (884)	749 (826)	718 (792)
Particulates	93 (102)	96 (106)	100 (110)
Hg	0.005 (0.005)	0.005 (0.005)	0.005 (0.005)
CO_2	2,786,239 (3,071,303)	1,548,138 (1,706,530)	309,368 (341,020)
kg/MWh (lb/gross-MWh)			
SO_2	0.007 (0.015)	0.003 (0.006)	0.003 (0.007)
NO_X	0.187 (.411)	0.183 (0.403)	0.179 (0.394)
Particulates	0.022 (.047)	0.023 (0.051)	0.025 (0.055)
Hg	1.07E-6 (2.35E-6)	1.15E-6 (2.55E-6)	1.23E-6 (2.71E-6)
CO_2	649 (1,430)	378 (833)	77 (170)
kg/MWh (lb/net-MWh)			
CO_2	791 (1,745)	499 (1,100)	110 (243)

The low level of SO_2 emissions is achieved by capture of the sulfur in the syngas by the Sulfinol-M AGR process in the non-capture case and a two-stage Selexol process in the capture cases. The AGR process removes over 99 percent of the sulfur compounds in the fuel gas down to a level of less than 3 ppmv in all three cases. This results in a concentration in the flue gas of less than 1 ppmv. The H_2S-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur. The Claus plant tail gas is hydrogenated and recycled to the inlet of the AGR process to capture most of the remaining sulfur. Because the environmental target was set based on higher sulfur bituminous coal, the resulting SO_2 emissions with lower sulfur western coals are substantially less than the environmental target.

NO_X emissions are limited by the use of nitrogen dilution to 15 ppmvd (as NO_2 @ 15 percent O_2). Ammonia in the syngas is removed with process condensate prior to the low-temperature AGR process and destroyed in the Claus plant burner. This helps lower NO_X levels as well.

Particulate discharge to the atmosphere is limited to extremely low values by the use of a cyclone and a barrier filter in addition to the syngas scrubber and the gas washing effect of the AGR absorber. The particulate emissions represent filterable particulate only.

Ninety five percent of the mercury is captured from the syngas by an activated carbon bed. CO_2 emissions represent the uncontrolled discharge from the process.

The carbon balance for all three IGCC cases is shown in Exhibit 4-11. The carbon input to the plant consists of carbon in the coal plus carbon in the air. Carbon leaves the plant as unburned carbon in the slag, CO_2 in the stack gas, CO_2 in the ASU vent, and CO_2 in the product gas in capture cases. Slag contains 2.97 percent carbon. The percent of total carbon sequestered for the capture cases is defined as the amount of carbon product produced (as sequestration-ready CO_2) divided by the carbon in the coal feedstock, less carbon contained in solid byproducts (slag).

Exhibit 4-11 Cases 1 - 3 Carbon Balance

	Case 1	Case 2	Case 3
Carbon In, kg/hr (lb/hr)			
Coal	108,715 (239,675)	112,498 (248,015)	117,494 (259,031)
Air (CO_2)	414 (913)	428 (942)	431 (951)
Total In	**109,129 (240,588)**	**112,925 (248,958)**	**117,926 (259,982)**
Carbon Out, kg/hr (lb/hr)			
Slag	544 (1,198)	562 (1,240)	587 (1,295)
Stack Gas	108,506 (239,215)	60,290 (132,917)	12,048 (26,561)
ASU Vent	79 (175)	81 (179)	85 (188)
CO_2 Product	0 (0)	51,992 (114,622)[1]	105,205 (231,938)[2]
Total Out	**109,129 (240,588)**	**112,925 (248,958)**	**117,926 (259,982)**

[1] Carbon capture is 46.4 percent to achieve an emission rate of 1,100 lb CO_2/net-MWh
[2] Carbon capture is 90 percent

Exhibit 4-12 shows the sulfur balance for all three IGCC cases. Sulfur input is the sulfur in the coal. Sulfur output is the sulfur recovered in the Claus plant, sulfur emitted in the stack gas, and sulfur sequestered with the CO_2 product in the capture cases. Sulfur in the slag and sulfur stripped from the wastewater streams are considered negligible.

Note that a significant amount of unconverted COS in Case 2 (because of the bypass around the SGS reactor) ends up as sulfur in the CO_2 product, thus reducing the capture fraction without increasing sulfur emissions.

Exhibit 4-12 Cases 1 – 3 Sulfur Balance

	Case 1	Case 2	Case 3
Sulfur In, kg/h (lb/hour)			
Coal	1,580 (3,482)	1,635 (3,603)	1,707 (3,764)
Total In	**1,580 (3,482)**	**1,635 (3,603)**	**1,707 (3,764)**
Sulfur Out, kg/h (lb/hour)			
Elemental Sulfur	1,577 (3,477)[1]	1,571 (3,464)[2]	1,703 (3,754)[3]
Stack Gas	2 (5)	1 (2)	1 (2)
CO_2 Product	0 (0)	62 (137)	3 (7)
Total Out	**1,580 (3,482)**	**1,635 (3,603)**	**1,707 (3,764)**

[1] Sulfur capture is 99.9 percent
[2] Sulfur capture is 96.1 percent
[3] Sulfur capture is 99.8 percent

Some water is returned to the source, namely cooling tower blowdown and sour water stripper blowdown. The difference between raw water withdrawal and water returned to the source (process discharge) is raw water consumption, which represents the net impact on the water source. Exhibit 4-13 shows the overall water balance for the plant. Raw water is obtained from groundwater (50 percent) and from municipal sources (50 percent). Water demand represents the total amount of water required for a particular process. Some water is recovered within the process, primarily as coal moisture from the drying process and syngas condensate, and that water is re-used as internal recycle. Raw water withdrawal is the difference between water demand and internal recycle. Some water is returned to the source, namely cooling tower blowdown and sour water stripper blowdown. The difference between raw water withdrawal and water returned to the source (process discharge) is raw water consumption, which represents the net impact on the water source.

Exhibit 4-13 Cases 1 – 3 Water Balance

	Case 1	**Case 2**	**Case 3**
Water Demand, m^3/min (gpm)			
Slag Handling	**0.40 (105)**	**0.41 (109)**	**0.43 (113)**
Quench/Wash	**0.00 (0)**	**2.5 (653)**	**2.6 (684)**
Humidification	**0.42 (110)**	**0.00 (0)**	**0.00 (0)**
Condenser Makeup	**0.14 (36)**	**1.2 (316)**	**3.9 (1,020)**
Shift Steam	0.00 (0)	1.1 (284)	3.7 (984)
BFW Makeup	0.14 (36)	0.12 (33)	0.14 (37)
Cooling Tower	**6.5 (1,705)**	**8.4 (2,207)**	**9.6 (2,541)**
Total	**7.4 (1,956)**	**12.4 (3,285)**	**16.5 (4,358)**
Internal Recycle, m^3/min (gpm)			
Slag Handling	**0.35 (93)**	**0.41 (109)**	**0.43 (113)**
Quench/Wash	**0.0 (0)**	**1.8 (466)**	**2.6 (684)**
Humidification	**0.0 (0)**	**0.00 (0)**	**0.00 (0)**
Condenser Makeup	**0.0 (0)**	**0.00 (0)**	**0.00 (0)**
Cooling Tower	**0.93 (247)**	**1.1 (299)**	**1.7 (436)**
Water from Coal Drying	0.76 (201)	0.79 (208)	0.82 (218)
BFW Blowdown	0.14 (36)	0.12 (33)	0.14 (37)
SWS Blowdown	0.04 (9)	0.22 (57)	0.34 (89)
SWS Excess	0.00 (0)	0.00 (0)	0.35 (93)
Total	**1.3 (340)**	**3.3 (874)**	**4.7 (1,234)**
Raw Water Withdrawal, m^3/min (gpm)			
Slag Handling	**0.05 (12)**	**0.00 (0)**	**0.00 (0)**
Quench/Wash	**0.00 (0)**	**0.71 (187)**	**0.00 (0)**
Humidification	**0.42 (110)**	**0.00 (0)**	**0.00 (0)**
Condenser Makeup	**0.14 (36)**	**1.2 (316)**	**3.9 (1,020)**
Shift Steam	0.00 (0)	1.1 (283)	3.7 (983)
BFW Makeup	0.14 (36)	0.12 (33)	0.14 (37)
Cooling Tower	**5.5 (1,458)**	**7.2 (1,908)**	**8.0 (2,104)**
Total	**6.1 (1,616)**	**9.1 (2,412)**	**11.8 (3,124)**
Process Water Discharge, m^3/min (gpm)			
SWS Blowdown	**0.00 (0.9)**	**0.02 (6)**	**0.03 (9)**
Cooling Tower Blowdown	**1.5 (383)**	**1.9 (496)**	**2.2 (571)**
Total	**1.5 (384)**	**1.9 (502)**	**2.2 (580)**

Raw Water Consumption, m^3/min (gpm)			
Slag Handling	0.05 (12)	0.00 (0)	0.00 (0)
Quench/Wash	0.00 (0)	0.71 (187)	0.00 (0)
Humidification	0.42 (110)	0.00 (0)	0.00 (0)
SWS Blowdown	-0.00 (-0.9)	-0.02 (-6)	-0.03 (-9)
Condenser Makeup	0.14 (36)	1.2 (316)	3.9 (1,020)
Cooling Tower	4.1 (1,074)	5.3 (1,412)	5.8 (1,533)
Total	4.7 (1,232)	7.2 (1,910)	9.6 (2,544)
Total, gpm/MWnet	2.5	4.3	6.3

Heat and Mass Balance Diagrams

Heat and mass balance diagrams are shown for all three IGCC cases for the following subsystems in Exhibit 4-14 through Exhibit 4-22.

- ASU and Gasifier Units
- Gas Cleanup System
- Power block

An overall plant energy balance is provided in tabular form in Exhibit 4-23 for the three cases.

Exhibit 4-14 Case 1: IGCC without CO_2 Capture - ASU and Gasification Heat and Mass Balance Schematic

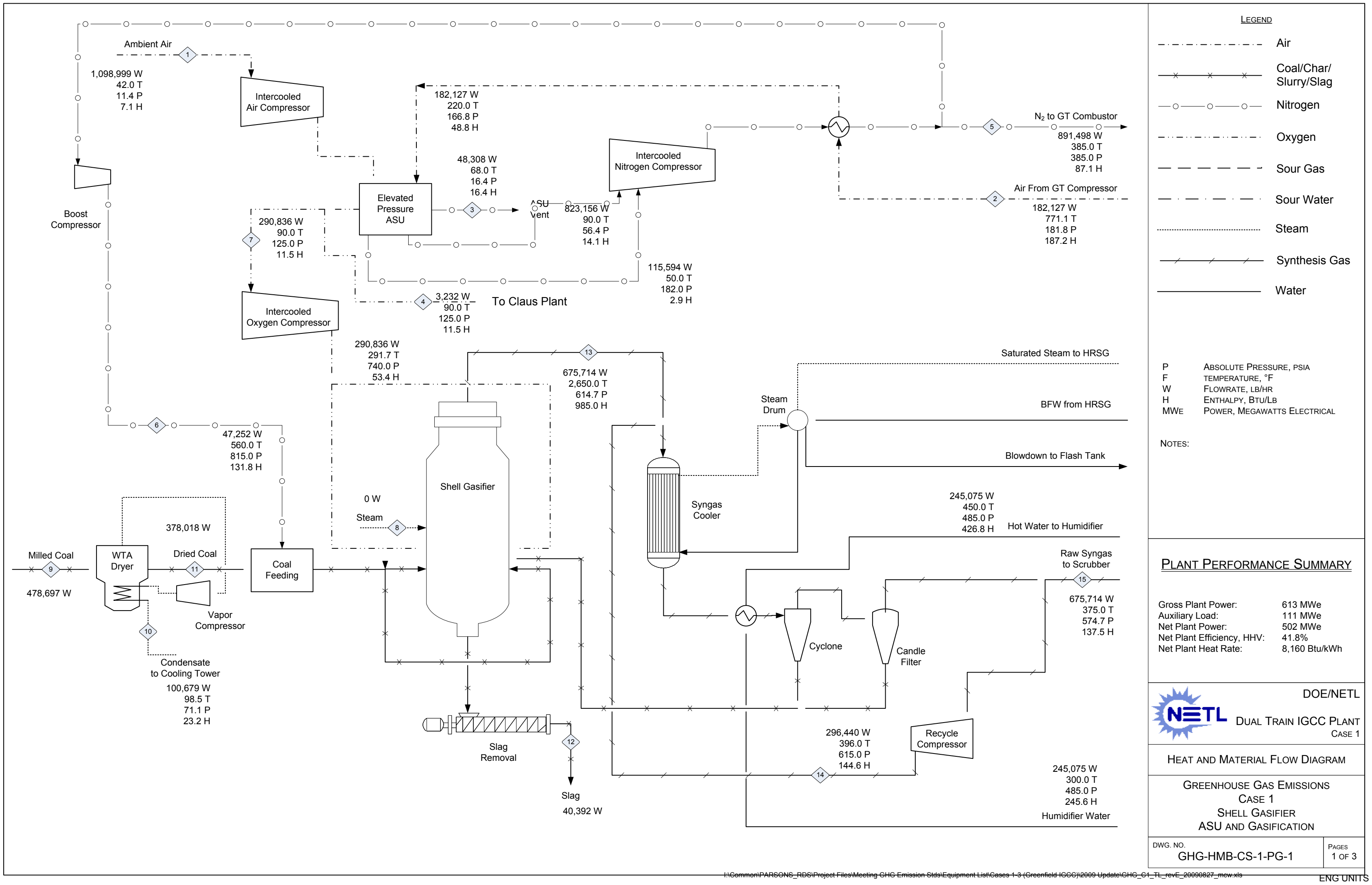


Exhibit 4-15 Case 1: IGCC without CO_2 Capture - Gas Cleanup System Heat and Mass Balance Schematic

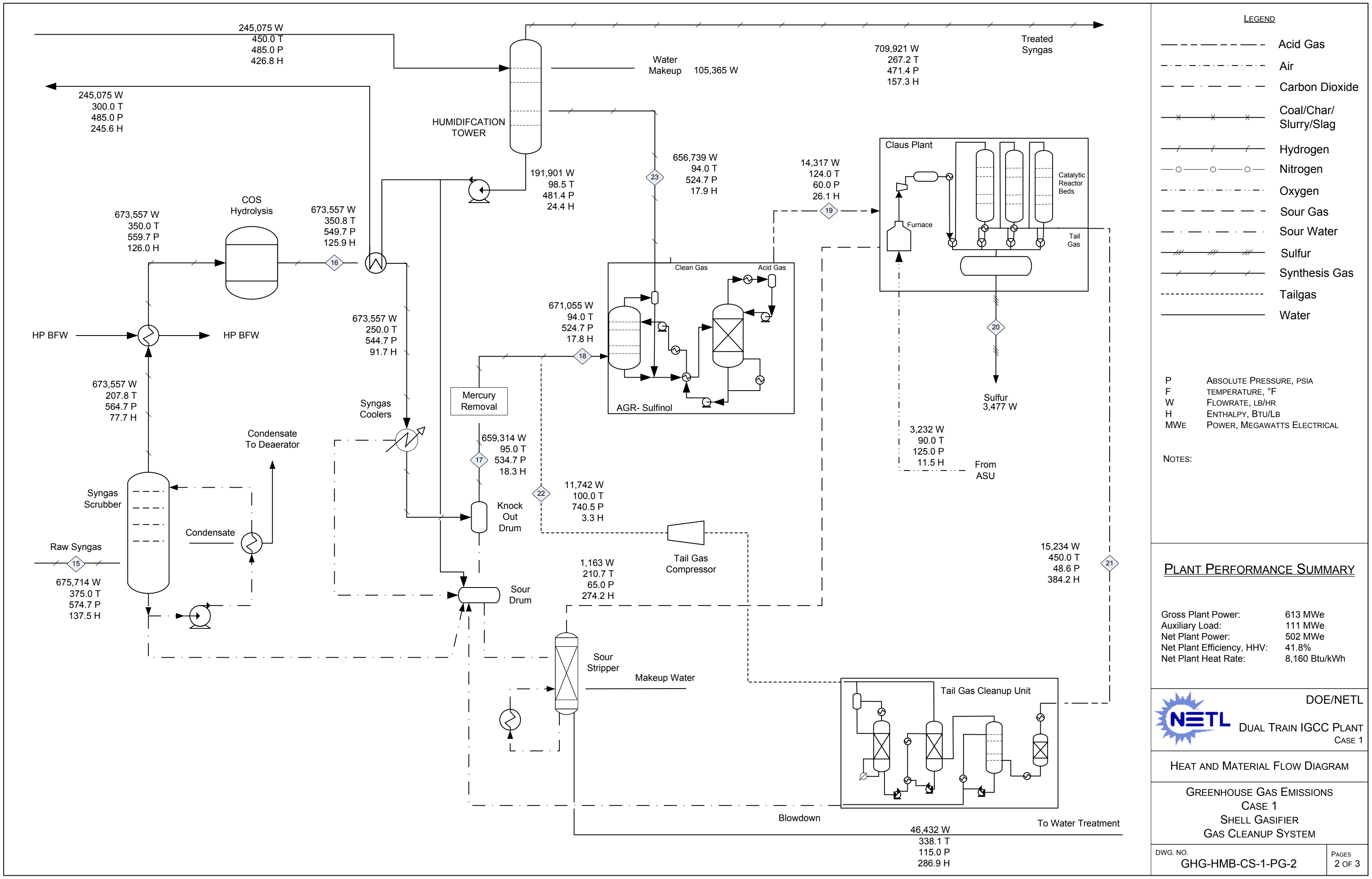


Exhibit 4-16 Case 1: IGCC without CO_2 Capture - Power Block System Heat and Mass Balance Schematic

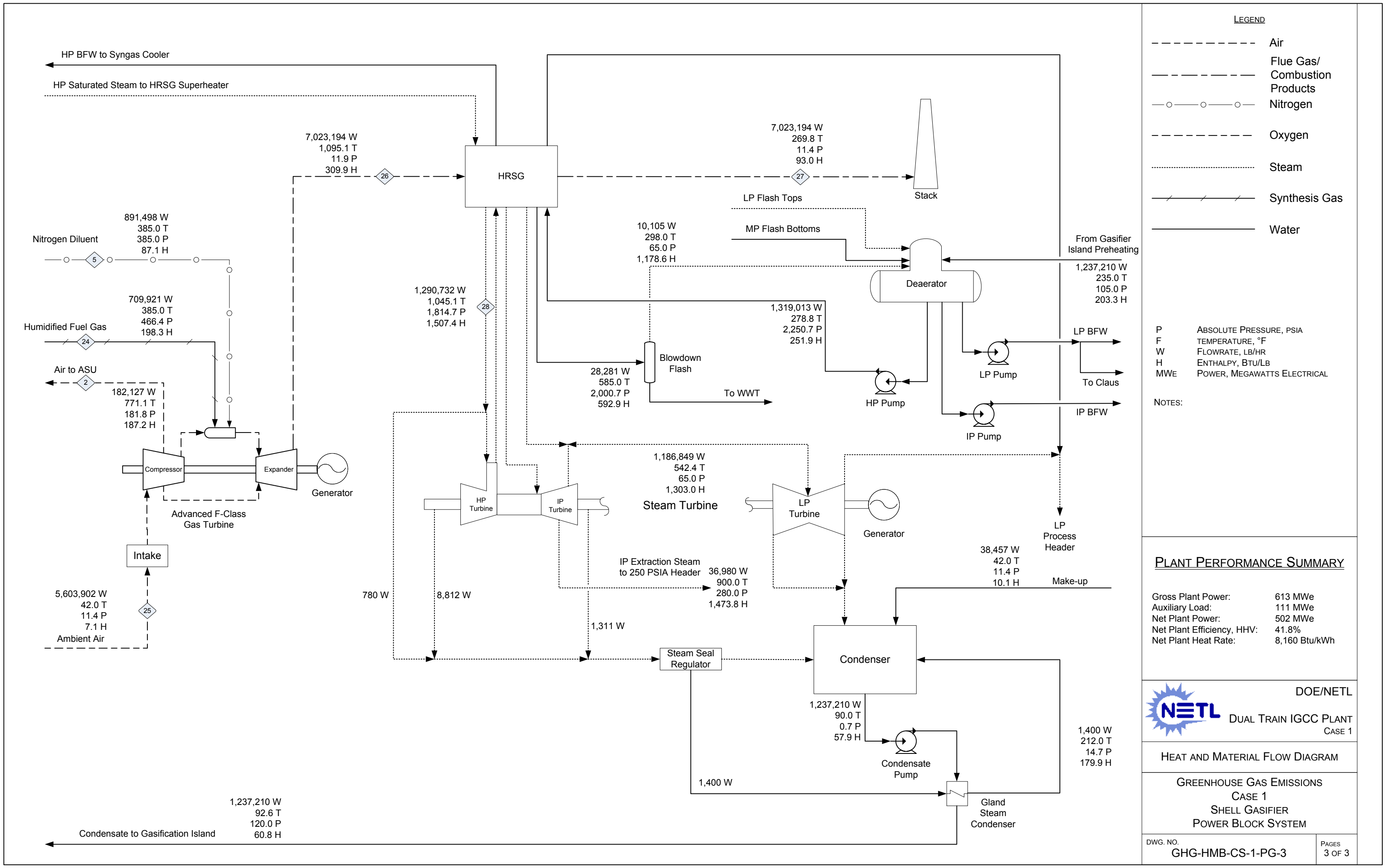


Exhibit 4-17 Case 2: IGCC with CO_2 Capture to an Emission Limit of 1,100 lb/net-MWh - ASU and Gasification Heat and Mass Balance Schematic

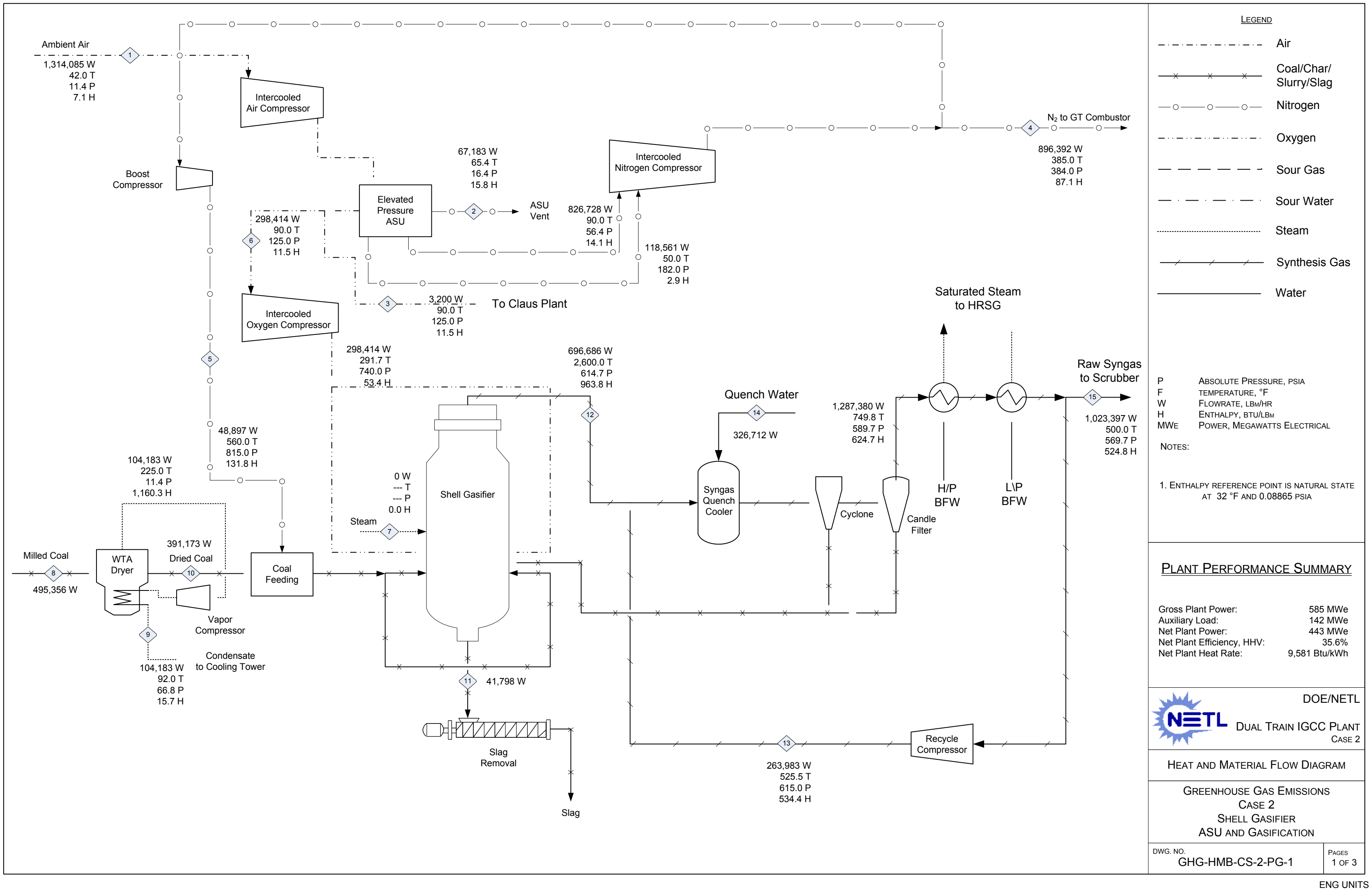


Exhibit 4-18 Case 2: IGCC with CO_2 Capture to an Emission Limit of 1,100 lb/net-MWh - Gas Cleanup Heat and Mass Balance Schematic

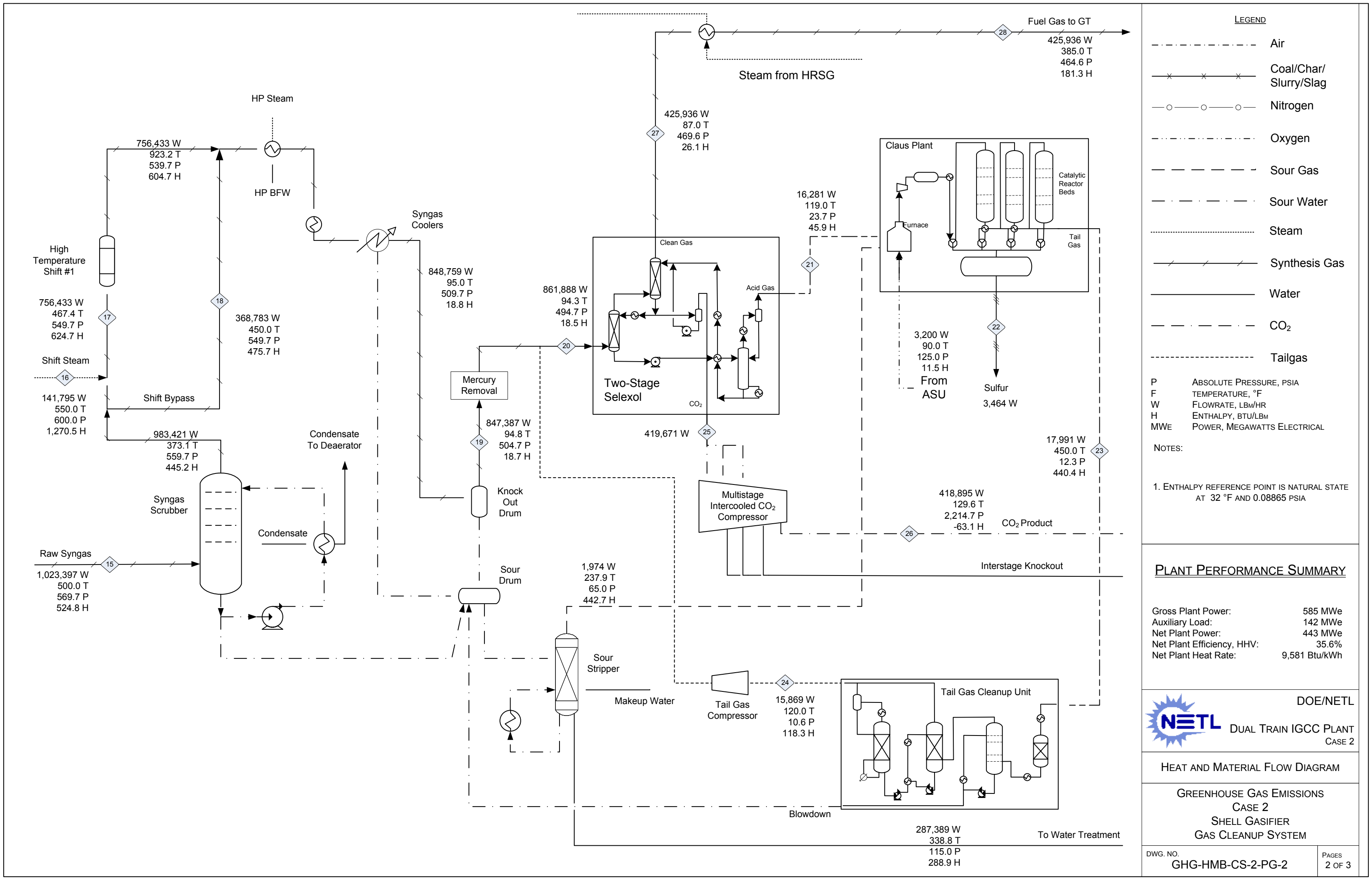


Exhibit 4-19 Case 2: IGCC with CO_2 Capture to an Emission Limit of 1,100 lb/net-MWh - Power Block System Heat and Mass Balance Schematic

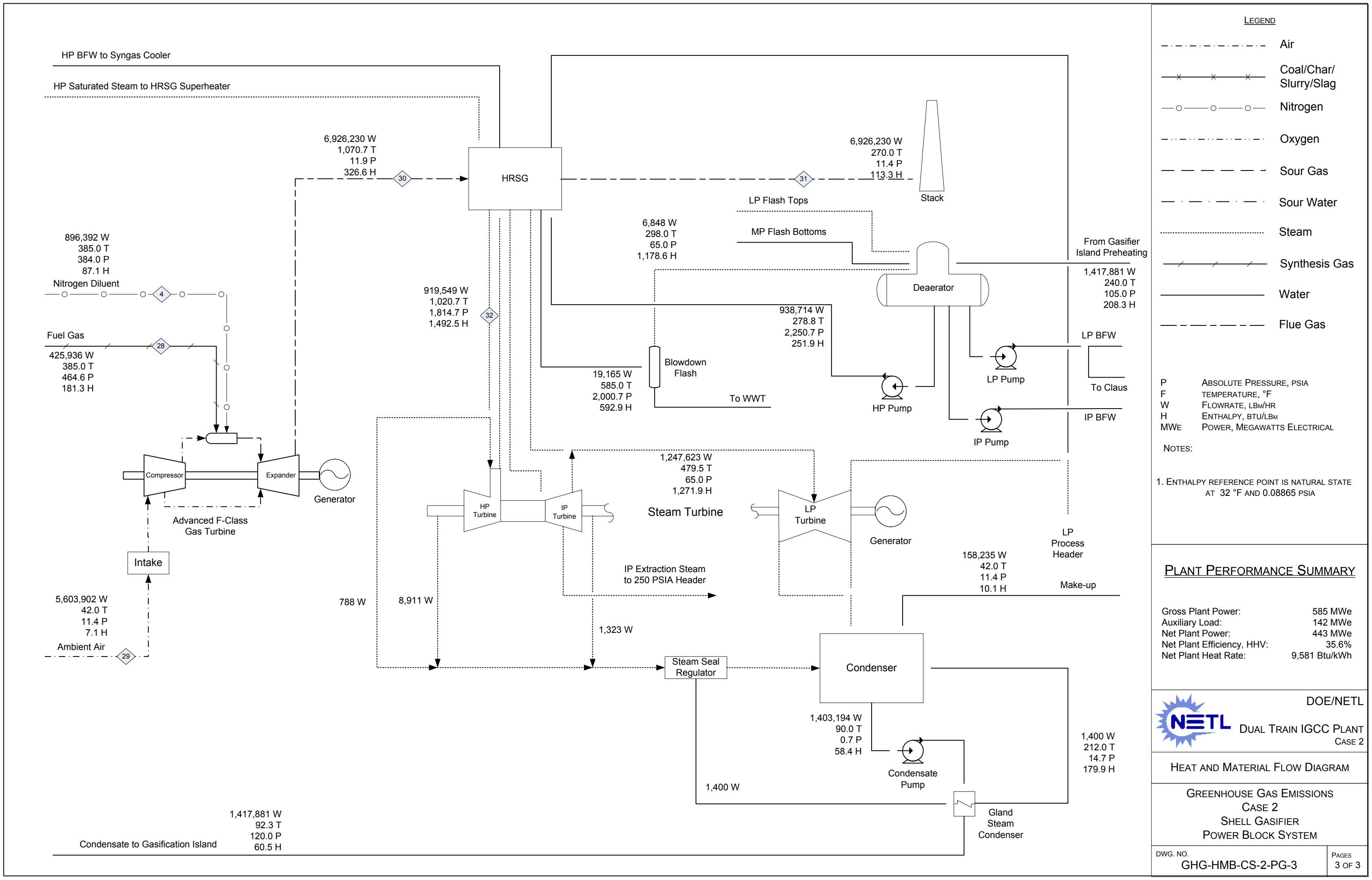


Exhibit 4-20 Case 3: IGCC with 90% CO_2 Capture - ASU and Gasification Heat and Mass Balance Schematic

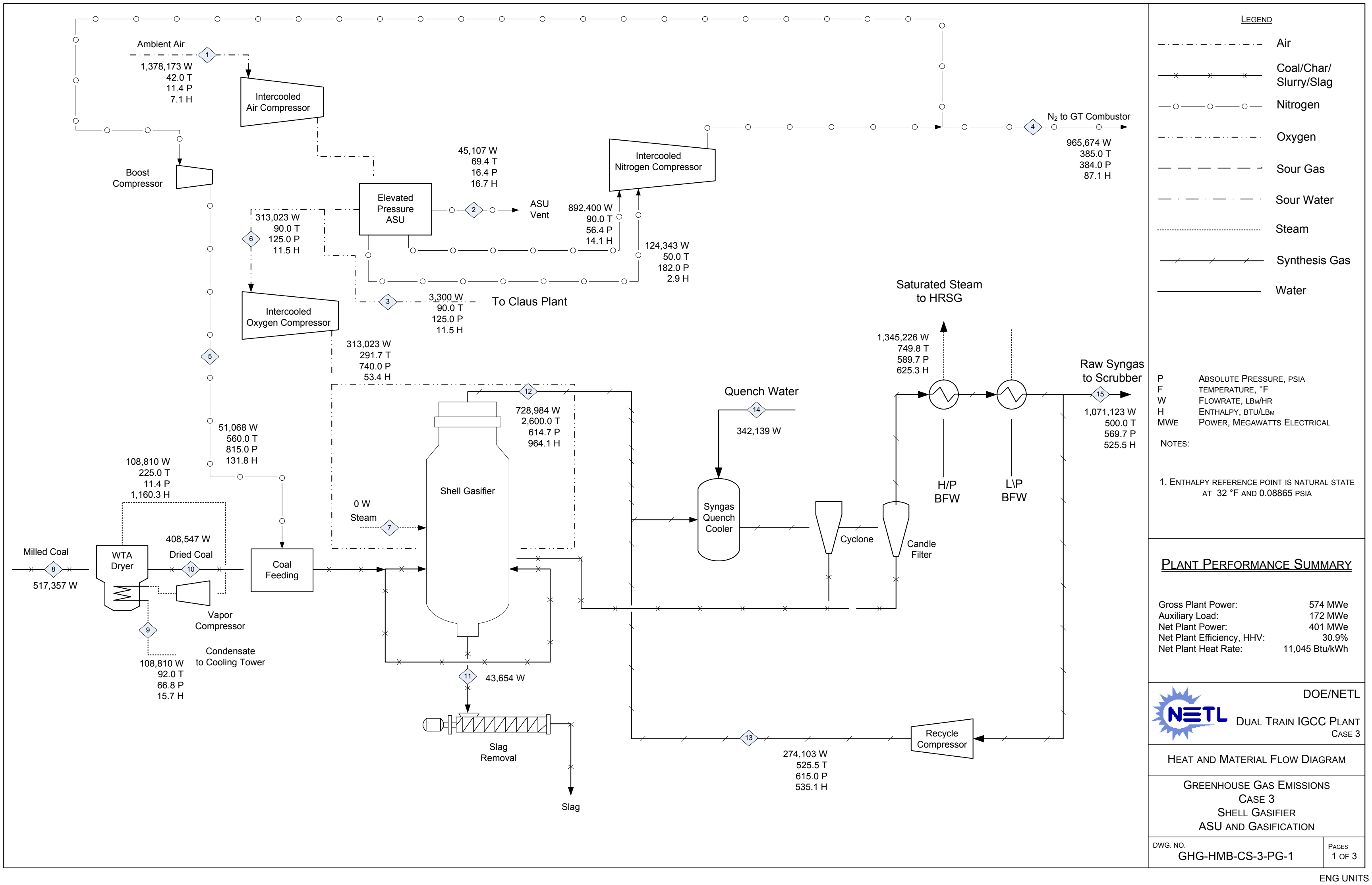


Exhibit 4-21 Case 3: IGCC with 90% CO_2 Capture - Gas Cleanup System Heat and Mass Balance Schematic

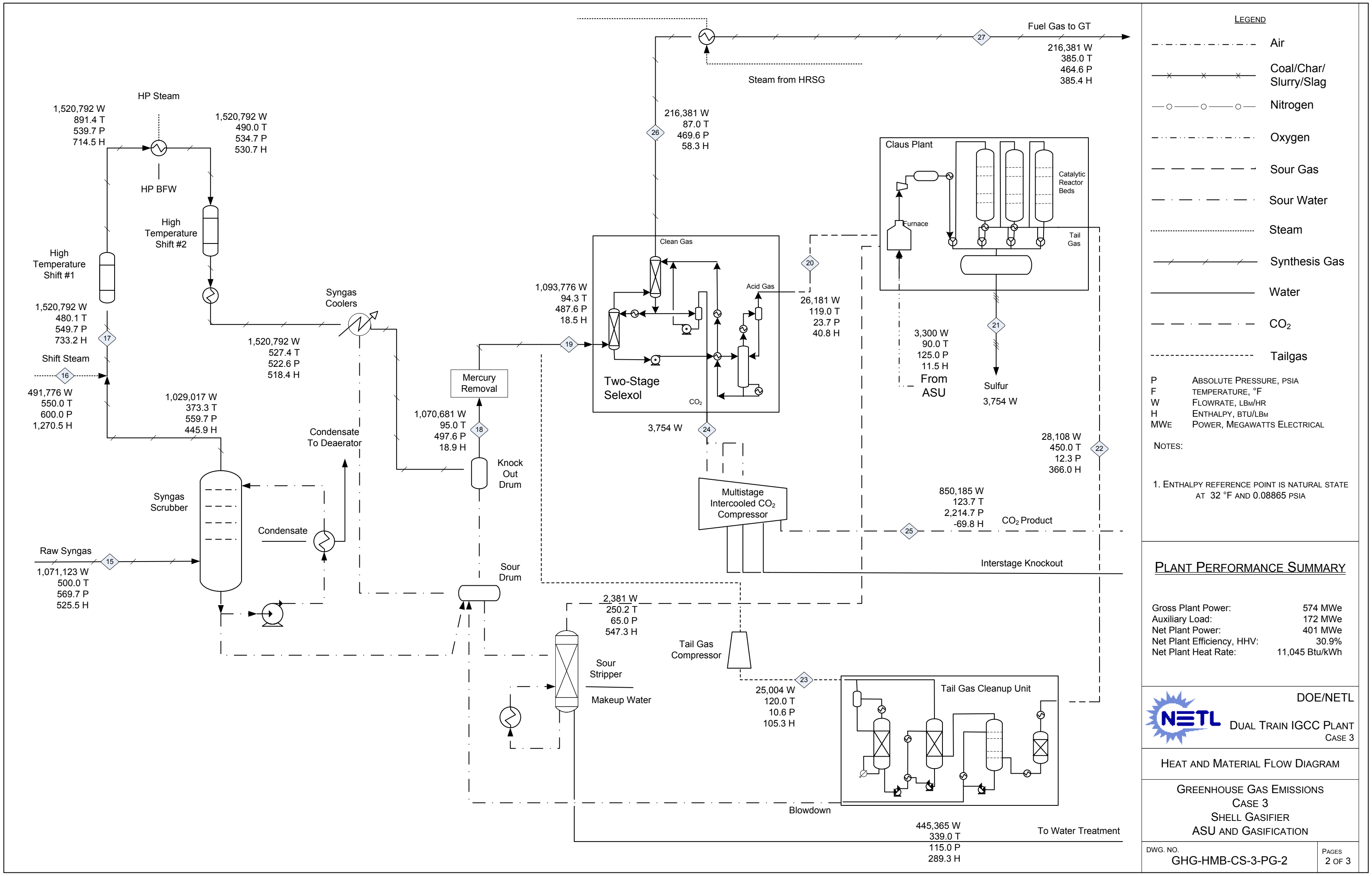


Exhibit 4-22 Case 3: IGCC with 90% CO_2 Capture - Power Block System Heat and Mass Balance Schematic

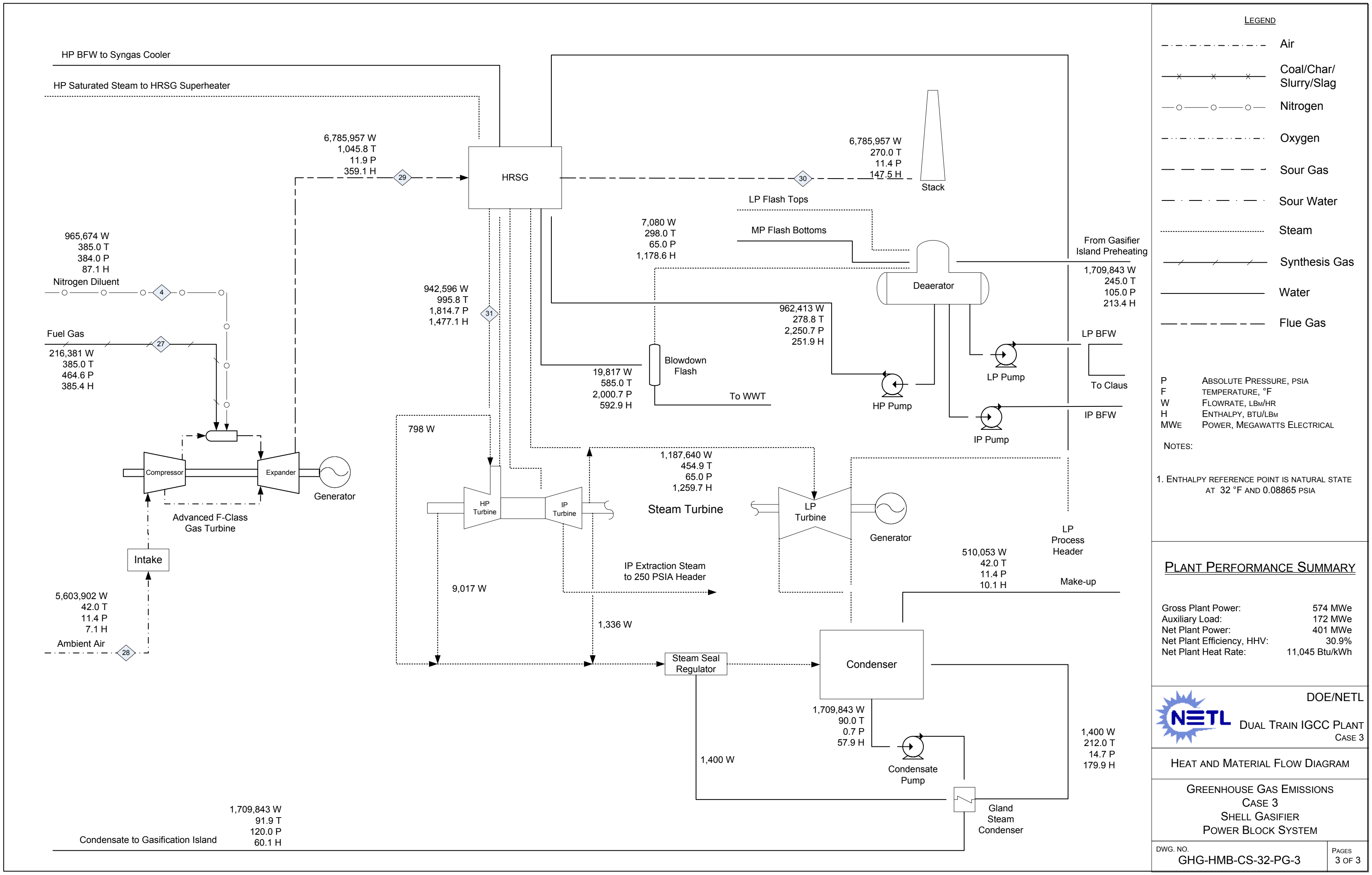


Exhibit 4-23 Cases 1 - 3 Energy Balance

	Case 1	**Case 2**	**Case 3**
Energy In, GJ/hr (MMBtu/hr)[1]			
Coal, HHV	**4,325 (4,100)**	**4,476 (4,242)**	**4,675 (4,431)**
Sensible + Latent			
Coal,	2.2 (2.1)	2.3 (2.2)	2.4 (2.3)
ASU Air	8.2 (7.8)	9.8 (9.3)	10.3 (9.7)
GT Air	41.7 (39.6)	41.7 (39.6)	41.7 (39.6)
Raw Water Makeup	8.5 (8.1)	12.7 (12.0)	16.5 (15.6)
Auxiliary Power	**398 (377)**	**512 (485)**	**620 (588)**
Total In	**4,784 (4,534)**	**5,054 (4,791)**	**5,366 (5,086)**
Energy Out, GJ/hr (MMBtu/hr)[1]			
Slag, HHV	**18 (17)**	**18 (17)**	**19 (18)**
Sulfur, HHV	**15 (14)**	**15 (14)**	**16 (15)**
Sensible + Latent			
ASU Intercoolers	169 (160)	190 (180)	201 (191)
ASU Vent	0.8 (0.8)	1.1 (1.1)	0.8 (0.8)
Slag	31.7 (30.1)	32.0 (30.3)	33.4 (31.7)
Sulfur	0.2 (0.2)	0.2 (0.2)	0.2 (0.2)
CO_2	0.0 (0.0)	-27.9 (-26.4)	-62.6 (-59.3)
CO_2 Compressor Intercoolers	0.0 (0.0)	73.4 (69.6)	152.6 (144.7)
Cooling Tower Blowdown	8.1 (7.7)	10.5 (9.9)	12.1 (11.4)
Gasifier Heat Loss	0.0 (0.0)	0.0 (0.0)	0.0 (0.0)
Combustion Turbine Heat Loss	63.3 (60.0)	63.3 (60.0)	63.3 (60.0)
HRSG Flue Gas	689 (653)	828 (785)	1,056 (1,001)
Condenser	1,188 (1,126)	1,218 (1,154)	1,132 (1,073)
Auxliary Cooling Load[2]	22 (21)	119 (113)	160 (152)
Process Losses[3]	*373 (353)*	*407 (386)*	*516 (489)*
Power	**2,206 (2,091)**	**2,106 (1,996)**	**2,065 (1,957)**
Total Out	**4,784 (4,534)**	**5,054 (4,791)**	**5,366 (5,086)**

[1] Enthalpy reference conditions are 0°C (32°F) and 614 Pa (0.089 psia)

[2] Auxiliary cooling load includes the sour water stripper condenser, syngas cooler (low level heat rejection) and the extraction air cooler (in extraction cases)

[3] Process losses are calculated by difference to close the energy balance

4.1.6 Cases 1 - 3 Equipment Lists

Major equipment items for all three IGCC cases are shown in the following tables. The accounts used in the equipment list correspond to the account numbers used in the cost estimates in Section 4.1.7. In general, the design conditions include a 10 percent contingency for flows and heat duties and a 21 percent contingency for heads on pumps and fans.

ACCOUNT 1 COAL HANDLING

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
1	Feeder	Belt	2	0	572 tonne/hr (630 tph)	572 tonne/hr (630 tph)	572 tonne/hr (630 tph)
2	Conveyor No. 1	Belt	1	0	1,134 tonne/hr (1,250 tph)	1,134 tonne/hr (1,250 tph)	1,134 tonne/hr (1,250 tph)
3	Transfer Tower No. 1	Enclosed	1	0	N/A	N/A	N/A
4	Conveyor No. 2	Belt	1	0	1,134 tonne/hr (1,250 tph)	1,134 tonne/hr (1,250 tph)	1,134 tonne/hr (1,250 tph)
5	As-Received Coal Sampling System	Two-stage	1	0	N/A	N/A	N/A
6	Stacker/Reclaimer	Traveling, linear	1	0	1,134 tonne/hr (1,250 tph)	1,134 tonne/hr (1,250 tph)	1,134 tonne/hr (1,250 tph)
7	Reclaim Hopper	N/A	2	1	45 tonne (50 ton)	45 tonne (50 ton)	45 tonne (50 ton)
8	Feeder	Vibratory	2	1	181 tonne/hr (200 tph)	181 tonne/hr (200 tph)	191 tonne/hr (210 tph)
9	Conveyor No. 3	Belt w/ tripper	1	0	354 tonne/hr (390 tph)	372 tonne/hr (410 tph)	390 tonne/hr (430 tph)
10	Crusher Tower	N/A	1	0	N/A	N/A	N/A
11	Coal Surge Bin w/ Vent Filter	Dual outlet	2	0	181 tonne (200 ton)	181 tonne (200 ton)	191 tonne (210 ton)
12	Crusher	Impactor reduction	2	0	8 cm x 0-3 cm x 0 (3" x 0-1-1/4" x 0)	8 cm x 0-3 cm x 0 (3" x 0-1-1/4" x 0)	8 cm x 0-3 cm x 0 (3" x 0-1-1/4" x 0)
13	As-Fired Coal Sampling System	Swing hammer	1	1	N/A	N/A	N/A
14	Conveyor No. 4	Belt w/tripper	1	0	354 tonne/hr (390 tph)	372 tonne/hr (410 tph)	390 tonne/hr (430 tph)
15	Transfer Tower No. 2	Enclosed	1	0	N/A	N/A	N/A
16	Conveyor No. 5	Belt w/ tripper	1	0	354 tonne/hr (390 tph)	372 tonne/hr (410 tph)	390 tonne/hr (430 tph)
17	Coal Silo w/ Vent Filter and Slide Gates	Field erected	3	0	816 tonne (900 ton)	816 tonne (900 ton)	816 tonne (900 ton)

ACCOUNT 2 COAL PREPARATION AND FEED

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
1	Feeder	Vibratory	3	0	82 tonne/hr (90 tph)	82 tonne/hr (90 tph)	82 tonne/hr (90 tph)
2	Conveyor No. 6	Belt w/tripper	1	0	236 tonne/hr (260 tph)	245 tonne/hr (270 tph)	254 tonne/hr (280 tph)
3	Roller Mill Feed Hopper	Dual Outlet	1	0	481 tonne (530 ton)	499 tonne (550 ton)	517 tonne (570 ton)
4	Weigh Feeder	Belt	2	0	118 tonne/hr (130 tph)	127 tonne/hr (140 tph)	127 tonne/hr (140 tph)
5	Pulverizer	Rotary	2	0	118 tonne/hr (130 tph)	127 tonne/hr (140 tph)	127 tonne/hr (140 tph)
6	Coal Dryer Feed Hopper	Vertical Hopper	2	0	236 tonne (260 ton)	245 tonne (270 ton)	254 tonne (280 ton)
7	Coal Preheater	Water Heated Horizontal Rotary Kiln	1	0	Coal feed: 236 tonne/hr (260 tph) Heat duty: 22.3 GJ/hr (21.2 MMBtu/hr)	Coal feed: 245 tonne/hr (270 tph) Heat duty: 23.1 GJ/hr (21.9 MMBtu/hr)	Coal feed: 254 tonne/hr (280 tph) Heat duty: 24.1 GJ/hr (22.9 MMBtu/hr)
8	Coal Dryer	Fluidized Bed with Internal Coils	2	0	Coal feed: 118 tonne/hr (130 tph) Heat duty: 69.2 GJ/hr (65.6 MMBtu/hr) Bed diameter: 11.3 m (37 ft)	Coal feed: 127 tonne/hr (140 tph) Heat duty: 71.6 GJ/hr (67.9 MMBtu/hr) Bed diameter: 11.3 m (37 ft)	Coal feed: 127 tonne/hr (140 tph) Heat duty: 74.8 GJ/hr (70.9 MMBtu/hr) Bed diameter: 11.6 m (38 ft)
9	Steam Compressor	Reciprocating, Multi-Stage	2	0	500 m^3/min (17,670 scfm) Suction - 0.08 MPa (11.4 psia) Discharge - 0.66 MPa (96 psia)	518 m^3/min (18,300 scfm) Suction - 0.08 MPa (11.4 psia) Discharge - 0.63 MPa (92 psia)	541 m^3/min (19,100 scfm) Suction - 0.08 MPa (11.4 psia) Discharge - 0.63 MPa (92 psia)
10	Dryer Exhaust Filter	Hot Baghouse	2	0	Steam - 25,129 kg/hr (55,400 lb/hr) Temperature - 107°C (225°F)	Steam - 25,991 kg/hr (57,300 lb/hr) Temperature - 107°C (225°F)	Steam - 27,125 kg/hr (59,800 lb/hr) Temperature - 107°C (225°F)
11	Dry Coal Cooler	Water Cooled Horizontal Rotary Kiln	1	0	189 tonne/hr (208 tph) Heat duty - 11 GJ/hr (10 MMBtu/hr)	195 tonne/hr (215 tph) Heat duty - 11 GJ/hr (11 MMBtu/hr)	204 tonne/hr (225 tph) Heat duty - 12 GJ/hr (11 MMBtu/hr)

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
1	Demineralized Water Storage Tank	Vertical, cylindrical, outdoor	2	0	526,172 liters (139,000 gal)	480,747 liters (127,000 gal)	537,528 liters (142,000 gal)
2	Condensate Pumps	Vertical canned	2	1	5,186 lpm @ 91 m H_2O (1,370 gpm @ 300 ft H_2O)	5,943 lpm @ 91 m H_2O (1,570 gpm @ 300 ft H_2O)	7,154 lpm @ 91 m H_2O (1,890 gpm @ 300 ft H_2O)
3	Deaerator (integral w/ HRSG)	Horizontal spray type	2	0	342,916 kg/hr (756,000 lb/hr)	409,140 kg/hr (902,000 lb/hr)	506,663 kg/hr (1,117,000 lb/hr)
4	Intermediate Pressure Feedwater Pump	Horizontal centrifugal, single stage	2	1	227 lpm @ 27 m H_2O (60 gpm @ 90 ft H_2O)	1,779 lpm @ 27 m H_2O (470 gpm @ 90 ft H_2O)	2,385 lpm @ 27 m H_2O (630 gpm @ 90 ft H_2O)
5	High Pressure Feedwater Pump No. 1	Barrel type, multi-stage, centrifugal	2	1	HP water: 5,716 lpm @ 1,890 m H_2O (1,510 gpm @ 6,200 ft H_2O)	HP water: 4,050 lpm @ 1,890 m H_2O (1,070 gpm @ 6,200 ft H_2O)	HP water: 4,164 lpm @ 1,890 m H_2O (1,100 gpm @ 6,200 ft H_2O)
6	High Pressure Feedwater Pump No. 2	Barrel type, multi-stage, centrifugal	2	1	IP water: 1,703 lpm @ 223 m H_2O (450 gpm @ 730 ft H_2O)	IP water: 1,098 lpm @ 223 m H_2O (290 gpm @ 730 ft H_2O)	IP water: 1,173 lpm @ 223 m H_2O (310 gpm @ 730 ft H_2O)
7	Auxiliary Boiler	Shop fabricated, water tube	1	0	18,144 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F)	18,144 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F)	18,144 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F)
8	Service Air Compressors	Flooded Screw	2	1	28 m^3/min @ 0.7 MPa (1,000 scfm @ 100 psig)	28 m^3/min @ 0.7 MPa (1,000 scfm @ 100 psig)	28 m^3/min @ 0.7 MPa (1,000 scfm @ 100 psig)
9	Instrument Air Dryers	Duplex, regenerative	2	1	28 m^3/min (1,000 scfm)	28 m^3/min (1,000 scfm)	28 m^3/min (1,000 scfm)
10	Closed Cycle Cooling Heat Exchangers	Plate and frame	2	0	120 GJ/hr (113 MMBtu/hr) each	225 GJ/hr (213 MMBtu/hr) each	297 GJ/hr (282 MMBtu/hr) each

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
11	Closed Cycle Cooling Water Pumps	Horizontal centrifugal	2	1	42,775 lpm @ 21 m H_2O (11,300 gpm @ 70 ft H_2O)	80,629 lpm @ 21 m H_2O (21,300 gpm @ 70 ft H_2O)	106,749 lpm @ 21 m H_2O (28,200 gpm @ 70 ft H_2O)
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	1	1	3,785 lpm @ 107 m H_2O (1,000 gpm @ 350 ft H_2O)	3,785 lpm @ 107 m H_2O (1,000 gpm @ 350 ft H_2O)	3,785 lpm @ 107 m H_2O (1,000 gpm @ 350 ft H_2O)
13	Fire Service Booster Pump	Two-stage horizontal centrifugal	1	1	2,650 lpm @ 76 m H_2O (700 gpm @ 250 ft H_2O)	2,650 lpm @ 76 m H_2O (700 gpm @ 250 ft H_2O)	2,650 lpm @ 76 m H_2O (700 gpm @ 250 ft H_2O)
14	Raw Water Pumps	Stainless steel, single suction	2	1	1,741 lpm @ 18 m H_2O (460 gpm @ 60 ft H_2O)	2,574 lpm @ 18 m H_2O (680 gpm @ 60 ft H_2O)	3,293 lpm @ 18 m H_2O (870 gpm @ 60 ft H_2O)
15	Ground Water Pumps	Stainless steel, single suction	1	1	3,445 lpm @ 268 m H_2O (910 gpm @ 880 ft H_2O)	2,574 lpm @ 268 m H_2O (680 gpm @ 880 ft H_2O)	3,293 lpm @ 268 m H_2O (870 gpm @ 880 ft H_2O)
16	Filtered Water Pumps	Stainless steel, single suction	2	1	606 lpm @ 49 m H_2O (160 gpm @ 160 ft H_2O)	2,347 lpm @ 49 m H_2O (620 gpm @ 160 ft H_2O)	3,861 lpm @ 49 m H_2O (1,020 gpm @ 160 ft H_2O)
17	Filtered Water Tank	Vertical, cylindrical	2	0	295,262 liter (78,000 gal)	1,124,267 liter (297,000 gal)	1,858,637 liter (491,000 gal)
18	Makeup Water Demineralizer	Anion, cation, and mixed bed	2	0	151 lpm (40 gpm)	757 lpm (200 gpm)	2,196 lpm (580 gpm)
19	Liquid Waste Treatment System		1	0	10 years, 24-hour storm	10 years, 24-hour storm	10 years, 24-hour storm

ACCOUNT 4 GASIFIER, ASU AND ACCESSORIES INCLUDING LOW TEMPERATURE HEAT RECOVERY

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
1	Gasifier	Pressurized dry-feed, entrained bed	2	0	2,903 tonne/day, 4.2 MPa (3,200 tpd, 614.696 psia)	2,994 tonne/day, 4.2 MPa (3,300 tpd, 614.696 psia)	3,084 tonne/day, 4.2 MPa (3,400 tpd, 614.696 psia)
2	Synthesis Gas Cooler	Convective spiral-wound tube boiler	2	0	242,672 kg/hr (535,000 lb/hr)	321,143 kg/hr (708,000 lb/hr)	335,658 kg/hr (740,000 lb/hr)
3	Synthesis Gas Cyclone	High efficiency	2	0	242,672 kg/hr (535,000 lb/hr) Design efficiency 90%	321,143 kg/hr (708,000 lb/hr) Design efficiency 90%	335,658 kg/hr (740,000 lb/hr) Design efficiency 90%
4	Candle Filter	Pressurized filter with pulse-jet cleaning	2	0	metallic filters	metallic filters	metallic filters
5	Syngas Scrubber Including Sour Water Stripper	Vertical upflow	2	0	168,736 kg/hr (372,000 lb/hr)	255,373 kg/hr (563,000 lb/hr)	267,166 kg/hr (589,000 lb/hr)
6	Raw Gas Coolers	Shell and tube with condensate drain	6	0	164,654 kg/hr (363,000 lb/hr)	240,858 kg/hr (531,000 lb/hr)	305,721 kg/hr (674,000 lb/hr)
7	Raw Gas Knockout Drum	Vertical with mist eliminator	2	0	164,654 kg/hr, 35°C, 3.6 MPa (363,000 lb/hr, 95°F, 525 psia)	211,828 kg/hr, 35°C, 3.5 MPa (467,000 lb/hr, 95°F, 510 psia)	267,619 kg/hr, 35°C, 3.5 MPa (590,000 lb/hr, 95°F, 503 psia)
8	Saturation Water Economizers	Shell and tube	2	0	164,654 kg/hr (363,000 lb/hr)	N/A	N/A
9	Fuel Gas Saturator	Vertical tray tower	2	0	176,901 kg/hr, 131°C, 3.3 MPa (390,000 lb/hr, 267°F, 480 psia)	N/A	N/A
10	Saturator Water Pump	Centrifugal	2	2	757 lpm @ 12 m H_2O (200 gpm @ 40 ft H_2O)	N/A	N/A

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
11	Synthesis Gas Reheater	Shell and tube	2	0	176,901 kg/hr (390,000 lb/hr)	106,141 kg/hr (234,000 lb/hr)	53,977 kg/hr (119,000 lb/hr)
12	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	2	0	168,736 kg/hr (372,000 lb/hr) syngas	255,373 kg/hr (563,000 lb/hr) syngas	267,166 kg/hr (589,000 lb/hr) syngas
13	ASU Main Air Compressor	Centrifugal, multi-stage	2	0	3,738 m^3/min @ 1.3 MPa (132,000 scfm @ 190 psia)	4,474 m^3/min @ 1.3 MPa (158,000 scfm @ 190 psia)	4,701 m^3/min @ 1.3 MPa (166,000 scfm @ 190 psia)
14	Cold Box	Vendor design	2	0	1,724 tonne/day (1,900 tpd) of 95% purity oxygen	1,814 tonne/day (2,000 tpd) of 95% purity oxygen	1,905 tonne/day (2,100 tpd) of 95% purity oxygen
15	Oxygen Compressor	Centrifugal, multi-stage	2	0	878 m^3/min (31,000 scfm) Suction - 0.9 MPa (130 psia) Discharge - 5.1 MPa (740 psia)	906 m^3/min (32,000 scfm) Suction - 0.9 MPa (130 psia) Discharge - 5.1 MPa (740 psia)	963 m^3/min (34,000 scfm) Suction - 0.9 MPa (130 psia) Discharge - 5.1 MPa (740 psia)
16	Primary Nitrogen Compressor	Centrifugal, multi-stage	2	0	2,888 m^3/min (102,000 scfm) Suction - 0.4 MPa (60 psia) Discharge - 2.7 MPa (390 psia)	2,888 m^3/min (102,000 scfm) Suction - 0.4 MPa (60 psia) Discharge - 2.7 MPa (390 psia)	3,143 m^3/min (111,000 scfm) Suction - 0.4 MPa (60 psia) Discharge - 2.7 MPa (390 psia)
17	Secondary Nitrogen Compressor	Centrifugal, single-stage	2	0	396 m^3/min (14,000 scfm) Suction - 2.7 MPa (390 psia) Discharge - 5.7 MPa (820 psia)	425 m^3/min (15,000 scfm) Suction - 2.6 MPa (380 psia) Discharge - 5.7 MPa (820 psia)	425 m^3/min (15,000 scfm) Suction - 2.6 MPa (380 psia) Discharge - 5.7 MPa (820 psia)

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
18	Extraction Air Heat Exchanger	Gas-to-gas, vendor design	2	0	45,359 kg/hr, 411°C, 1.3 MPa (100,000 lb/hr, 771°F, 182 psia)	N/A	N/A
19	Transport Nitrogen Boost Compressor	Centrifugal, single-stage	2	0	167 m^3/min (5,900 scfm) Suction - 2.7 MPa (389 psia) Discharge - 5.6 MPa (815 psia)	173 m^3/min (6,100 scfm) Suction - 2.6 MPa (384 psia) Discharge - 5.6 MPa (815 psia)	178 m^3/min (6,300 scfm) Suction - 2.6 MPa (384 psia) Discharge - 5.6 MPa (815 psia)
20	Syngas Dilution Nitrogen Boost Compressor	Centrifugal, single-stage	2	0	N/A	N/A	1,478 m^3/min (52,200 scfm) Suction - 2.6 MPa (384 psia) Discharge - 3.2 MPa (469 psia)

ACCOUNT 5 SYNGAS CLEANUP

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
1	Mercury Adsorber	Sulfated carbon bed	2	0	164,654 kg/hr (363,000 lb/hr) 35°C (95°F) 3.7 MPa (535 psia)	211,374 kg/hr (466,000 lb/hr) 35°C (95°F) 3.5 MPa (505 psia)	267,166 kg/hr (589,000 lb/hr) 35°C (95°F) 3.4 MPa (498 psia)
2	Sulfur Plant	Claus type	1	0	42 tonne/day (46 tpd)	41 tonne/day (46 tpd)	45 tonne/day (50 tpd)
3	COS/WGS Reactor(s)	Fixed bed, catalytic	Case 1 - 2 Case 2 - 2 Case 3 - 4	0	167,829 kg/hr (370,000 lb/hr) 177°C (350°F) 3.9 MPa (560 psia)	188,694 kg/hr (416,000 lb/hr) 243°C (470°F) 3.8 MPa (550 psia)	379,203 kg/hr (836,000 lb/hr) 249°C (480°F) 3.8 MPa (550 psia)

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
4	WGS Heat Exchangers	Shell and tube	Case 1 – 0 Case 2 – 2 Case 3 - 4	0	N/A	Exchanger 1: 73 GJ/hr (69 MMBtu/hr) Exchanger 2: -17 GJ/hr (-16 MMBtu/hr)	Exchanger 1: 162 GJ/hr (154 MMBtu/hr) Exchanger 2: 8 GJ/hr (8 MMBtu/hr)
5	Acid Gas Removal Plant	Sulfinol/ Selexol/ Selexol	2	0	167,376 kg/hr (369,000 lb/hr) 34°C (94°F) 3.6 MPa (525 psia)	215,003 kg/hr (474,000 lb/hr) 35°C (94°F) 3.4 MPa (495 psia)	273,063 kg/hr (602,000 lb/hr) 35°C (94°F) 3.4 MPa (488 psia)
6	Hydrogenation Reactor	Fixed bed, catalytic	1	0	6,910 kg/hr (15,234 lb/hr) 232°C (450°F) 0.3 MPa (48.6 psia)	8,976 kg/hr (19,790 lb/hr) 232°C (450°F) 0.1 MPa (12.3 psia)	14,025 kg/hr (30,919 lb/hr) 232°C (450°F) 0.1 MPa (12.3 psia)
7	Tail Gas Recycle Compressor	Centrifugal	1	0	5,326 kg/hr (11,742 lb/hr)	7,243 kg/hr (15,969 lb/hr)	11,534 kg/hr (25,428 lb/hr)

ACCOUNT 5B CO$_2$ COMPRESSION

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
	CO$_2$ Compressor	Integrally geared, multi-stage centrifugal	4	0	N/A	473 m^3/min @ 15.3 MPa (16,700 scfm @ 2,215 psia)	957 m^3/min @ 15.3 MPa (33,800 scfm @ 2,215 psia)

ACCOUNT 5C CO$_2$ TRANSPORT, STORAGE, AND MONITORING (not shown in Total Plant Cost Details)

Equipment No.	Description	Type	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
1	CO$_2$ Pipeline	Carbon Steel	N/A	50 miles @ 12 in diameter w/ inlet pressure of 2,200 psi and outlet pressure of 1,500 psi	50 miles @ 14 in diameter w/ inlet pressure of 2,200 psi and outlet pressure of 1,500 psi

Equipment No.	Description	Type	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
2	CO_2 Sequestration Source	Saline Formation	N/A	1 well with bottom hole pressure @ 1,220 psi, 530 ft thickness, 4,055 ft depth, 22 Md permeability	2 wells with bottom hole pressure @ 1,220 psi, 530 ft thickness, 4,055 ft depth, 22 Md permeability
3	CO_2 Monitoring	N/A	N/A	20 year monitoring during plant life / 80 years following / Total of 100 years	20 year monitoring during plant life / 80 years following / Total of 100 years

ACCOUNT 6 COMBUSTION TURBINE AND AUXILIARIES

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
1	Gas Turbine	Advanced F class	2	0	185 MW	190 MW	190 MW
2	Gas Turbine Generator	TEWAC	2	0	210 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	210 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	210 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase

ACCOUNT 7 HRSG, DUCTING AND STACK

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
1	Stack	CS plate, type 409SS liner	1	0	76 m (250 ft) high x 8.6 m (28 ft) diameter	76 m (250 ft) high x 8.7 m (28 ft) diameter	76 m (250 ft) high x 8.8 m (29 ft) diameter

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
2	Heat Recovery Steam Generator	Drum, multi-pressure with economizer section and integral deaerator	2	0	Main steam - 322,006 kg/hr, 12.4 MPa/563°C (709,902 lb/hr, 1,800 psig/1,045°F) Reheat steam - 310,531 kg/hr, 3.1 MPa/563°C (684,604 lb/hr, 452 psig/1,045°F)	Main steam - 229,405 kg/hr, 12.4 MPa/549°C (505,752 lb/hr, 1,800 psig/1,021°F) Reheat steam - 254,747 kg/hr, 3.1 MPa/549°C (561,621 lb/hr, 452 psig/1,021°F)	Main steam - 235,155 kg/hr, 12.4 MPa/535°C (518,428 lb/hr, 1,800 psig/996°F) Reheat steam - 228,561 kg/hr, 3.1 MPa/535°C (503,891 lb/hr, 452 psig/996°F)

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
1	Steam Turbine	Commercially available	1	0	253 MW 12.4 MPa/563°C/563°C (1800 psig/ 1,045°F/1,045°F)	219 MW 12.4 MPa/549°C/549°C (1800 psig/ 1,021°F/1,021°F)	203 MW 12.4 MPa/535°C/535°C (1800 psig/ 996°F/996°F)
2	Steam Turbine Generator	Hydrogen cooled, static excitation	1	0	280 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	240 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	230 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase
3	Surface Condenser	Single pass, divided waterbox including vacuum pumps	1	0	654 GJ/hr (620 MMBtu/hr), Condensing temperature 32°C (90°F), Inlet water temperature 9°C (48°F), Water temperature rise 11°C (20°F)	665 GJ/hr (630 MMBtu/hr), Condensing temperature 32°C (90°F), Inlet water temperature 9°C (48°F), Water temperature rise 11°C (20°F)	622 GJ/hr (590 MMBtu/hr), Condensing temperature 32°C (90°F), Inlet water temperature 9°C (48°F), Water temperature rise 11°C (20°F)

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
4	Air-cooled Condenser	---	1	0	654 GJ/hr (620 MMBtu/hr), Condensing temperature 32°C (90°F), Ambient temperature 6°C (42°F)	665 GJ/hr (630 MMBtu/hr), Condensing temperature 32°C (90°F), Ambient temperature 6°C (42°F)	622 GJ/hr (590 MMBtu/hr), Condensing temperature 32°C (90°F), Ambient temperature 6°C (42°F)

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
1	Circulating Water Pumps	Vertical, wet pit	2	1	166,558 lpm @ 30 m (44,000 gpm @ 100 ft)	215,768 lpm @ 30 m (57,000 gpm @ 100 ft)	246,052 lpm @ 30 m (65,000 gpm @ 100 ft)
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	1	0	3°C (37°F) wet bulb / 9°C (48°F) CWT / 20°C (68°F) HWT / 928 GJ/hr (880 MMBtu/hr) heat duty	3°C (37°F) wet bulb / 9°C (48°F) CWT / 20°C (68°F) HWT / 1,203 GJ/hr (1140 MMBtu/hr) heat duty	3°C (37°F) wet bulb / 9°C (48°F) CWT / 20°C (68°F) HWT / 1382 GJ/hr (1,310 MMBtu/hr) heat duty

ACCOUNT 10 SLAG RECOVERY AND HANDLING

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
1	Slag Quench Tank	Water bath	2	0	193,056 liters (51,000 gal)	200,627 liters (53,000 gal)	208,198 liters (55,000 gal)
2	Slag Crusher	Roll	2	0	10 tonne/hr (11 tph)	11 tonne/hr (12 tph)	11 tonne/hr (12 tph)
3	Slag Depressurizer	Lock Hopper	2	0	10 tonne/hr (11 tph)	11 tonne/hr (12 tph)	11 tonne/hr (12 tph)
4	Slag Receiving Tank	Horizontal, weir	2	0	117,348 liters (31,000 gal)	121,133 liters (32,000 gal)	124,919 liters (33,000 gal)
5	Black Water Overflow Tank	Shop fabricated	2		52,996 liters (14,000 gal)	52,996 liters (14,000 gal)	56,781 liters (15,000 gal)

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
6	Slag Conveyor	Drag chain	2	0	10 tonne/hr (11 tph)	11 tonne/hr (12 tph)	11 tonne/hr (12 tph)
7	Slag Separation Screen	Vibrating	2	0	10 tonne/hr (11 tph)	11 tonne/hr (12 tph)	11 tonne/hr (12 tph)
8	Coarse Slag Conveyor	Belt/bucket	2	0	10 tonne/hr (11 tph)	11 tonne/hr (12 tph)	11 tonne/hr (12 tph)
9	Fine Ash Settling Tank	Vertical, gravity	2	0	162,773 liters (43,000 gal)	170,344 liters (45,000 gal)	177,914 liters (47,000 gal)
10	Fine Ash Recycle Pumps	Horizontal centrifugal	2	2	38 lpm @ 14 m H_2O (10 gpm @ 46 ft H_2O)	38 lpm @ 14 m H_2O (10 gpm @ 46 ft H_2O)	38 lpm @ 14 m H_2O (10 gpm @ 46 ft H_2O)
11	Grey Water Storage Tank	Field erected	2	0	52,996 liters (14,000 gal)	52,996 liters (14,000 gal)	56,781 liters (15,000 gal)
12	Grey Water Pumps	Centrifugal	2	2	189 lpm @ 433 m H_2O (50 gpm @ 1,420 ft H_2O)	189 lpm @ 433 m H_2O (50 gpm @ 1,420 ft H_2O)	189 lpm @ 433 m H_2O (50 gpm @ 1,420 ft H_2O)
13	Slag Storage Bin	Vertical, field erected	2	0	726 tonne (800 tons)	726 tonne (800 tons)	816 tonne (900 tons)
14	Unloading Equipment	Telescoping chute	1	0	82 tonne/hr (90 tph)	91 tonne/hr (100 tph)	91 tonne/hr (100 tph)

ACCOUNT 11 ACCESSORY ELECTRIC PLANT

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
1	CTG Step-up Transformer	Oil-filled	2	0	24 kV/345 kV, 210 MVA, 3-ph, 60 Hz	24 kV/345 kV, 210 MVA, 3-ph, 60 Hz	24 kV/345 kV, 210 MVA, 3-ph, 60 Hz
2	STG Step-up Transformer	Oil-filled	1	0	24 kV/345 kV, 280 MVA, 3-ph, 60 Hz	24 kV/345 kV, 240 MVA, 3-ph, 60 Hz	24 kV/345 kV, 230 MVA, 3-ph, 60 Hz
3	High Voltage Auxiliary Transformer	Oil-filled	2	0	345 kV/13.8 kV, 49 MVA, 3-ph, 60 Hz	345 kV/13.8 kV, 62 MVA, 3-ph, 60 Hz	345 kV/13.8 kV, 73 MVA, 3-ph, 60 Hz
4	Medium Voltage Auxiliary Transformer	Oil-filled	1	1	24 kV/4.16 kV, 22 MVA, 3-ph, 60 Hz	24 kV/4.16 kV, 32 MVA, 3-ph, 60 Hz	24 kV/4.16 kV, 43 MVA, 3-ph, 60 Hz
5	Low Voltage Transformer	Dry ventilated	1	1	4.16 kV/480 V, 3 MVA, 3-ph, 60 Hz	4.16 kV/480 V, 5 MVA, 3-ph, 60 Hz	4.16 kV/480 V, 6 MVA, 3-ph, 60 Hz
6	CTG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	2	0	24 kV, 3-ph, 60 Hz	24 kV, 3-ph, 60 Hz	24 kV, 3-ph, 60 Hz
7	STG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	1	0	24 kV, 3-ph, 60 Hz	24 kV, 3-ph, 60 Hz	24 kV, 3-ph, 60 Hz
8	Medium Voltage Switchgear	Metal clad	1	1	4.16 kV, 3-ph, 60 Hz	4.16 kV, 3-ph, 60 Hz	4.16 kV, 3-ph, 60 Hz

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
9	Low Voltage Switchgear	Metal enclosed	1	1	480 V, 3-ph, 60 Hz	480 V, 3-ph, 60 Hz	480 V, 3-ph, 60 Hz
10	Emergency Diesel Generator	Sized for emergency shutdown	1	0	750 kW, 480 V, 3-ph, 60 Hz	750 kW, 480 V, 3-ph, 60 Hz	750 kW, 480 V, 3-ph, 60 Hz

ACCOUNT 12 INSTRUMENTATION AND CONTROLS

Equipment No.	Description	Type	Operating Qty.	Spares	Case 1 Design Condition	Case 2 Design Condition	Case 3 Design Condition
1	DCS - Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W)	1	0	Operator stations/printers and engineering stations/printers	Operator stations/printers and engineering stations/printers	Operator stations/printers and engineering stations/printers
2	DCS - Processor	Microprocessor with redundant input/output	1	0	N/A	N/A	N/A
3	DCS - Data Highway	Fiber optic	1	0	Fully redundant, 25% spare	Fully redundant, 25% spare	Fully redundant, 25% spare

4.1.7 Case 1 – Cost Estimating

The cost estimating methodology was described previously in Section 2.6. Exhibit 4-24 shows the TPC cost details organized by cost account as well as TOC and TASC. Exhibit 4-25 shows the initial and annual O&M costs.

The estimated TOC of the IGCC case with no CO_2 capture is $3,128/kW. Owner's costs represent 18 percent of the TOC. The current dollar, 30-year LCOE is $117.84/MWh.

Exhibit 4-24 Case 1 Total Plant Cost Details

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		Case 1 TOTAL PLANT COST	
				Direct	Indirect				Process	Project	$	$/kW
1	COAL & SORBENT HANDLING											
1.1	Coal Receive & Unload	$3,710	$0	$1,813	$0	$0	$5,523	$495	$0	$1,204	$7,222	$14
1.2	Coal Stackout & Reclaim	$4,794	$0	$1,162	$0	$0	$5,957	$522	$0	$1,296	$7,775	$15
1.3	Coal Conveyors	$4,458	$0	$1,150	$0	$0	$5,608	$492	$0	$1,220	$7,320	$15
1.4	Other Coal Handling	$1,166	$0	$266	$0	$0	$1,432	$125	$0	$312	$1,869	$4
1.5	Sorbent Receive & Unload	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.6	Sorbent Stackout & Reclaim	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.7	Sorbent Conveyors	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.8	Other Sorbent Handling	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.9	Coal & Sorbent Hnd.Foundations	$0	$2,625	$6,565	$0	$0	$9,190	$881	$0	$2,014	$12,086	$24
	SUBTOTAL 1.	**$14,128**	**$2,625**	**$10,957**	**$0**	**$0**	**$27,710**	**$2,515**	**$0**	**$6,045**	**$36,271**	**$72**
2	COAL & SORBENT PREP & FEED											
2.1	Coal Crushing & Drying	$42,557	$2,557	$6,201	$0	$0	$51,315	$4,428	$0	$11,149	$66,892	$133
2.2	Prepared Coal Storage & Feed	$2,016	$482	$316	$0	$0	$2,814	$241	$0	$611	$3,666	$7
2.3	Dry Coal Injection System	$66,338	$770	$6,161	$0	$0	$73,269	$6,311	$0	$15,916	$95,495	$190
2.4	Misc.Coal Prep & Feed	$1,108	$807	$2,418	$0	$0	$4,334	$398	$0	$946	$5,678	$11
2.5	Sorbent Prep Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.6	Sorbent Storage & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.7	Sorbent Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.8	Booster Air Supply System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.9	Coal & Sorbent Feed Foundation	$0	$4,309	$3,537	$0	$0	$7,846	$727	$0	$1,714	$10,287	$20
	SUBTOTAL 2.	**$112,019**	**$8,924**	**$18,634**	**$0**	**$0**	**$139,577**	**$12,104**	**$0**	**$30,336**	**$182,017**	**$362**
3	FEEDWATER & MISC. BOP SYSTEMS											
3.1	FeedwaterSystem	$2,718	$4,668	$2,464	$0	$0	$9,850	$913	$0	$2,153	$12,915	$26
3.2	Water Makeup & Pretreating	$289	$30	$162	$0	$0	$481	$46	$0	$158	$685	$1
3.3	Other Feedwater Subsystems	$1,487	$503	$452	$0	$0	$2,442	$219	$0	$532	$3,194	$6
3.4	Service Water Systems	$165	$341	$1,182	$0	$0	$1,688	$165	$0	$556	$2,409	$5
3.5	Other Boiler Plant Systems	$888	$344	$852	$0	$0	$2,084	$198	$0	$456	$2,738	$5
3.6	FO Supply Sys & Nat Gas	$292	$551	$514	$0	$0	$1,357	$131	$0	$298	$1,786	$4
3.7	Waste Treatment Equipment	$404	$0	$246	$0	$0	$650	$63	$0	$214	$928	$2
3.8	Misc. Equip.(cranes,AirComp.,Comm.)	$999	$134	$513	$0	$0	$1,646	$159	$0	$541	$2,346	$5
	SUBTOTAL 3.	**$7,242**	**$6,570**	**$6,386**	**$0**	**$0**	**$20,198**	**$1,893**	**$0**	**$4,908**	**$27,000**	**$54**
4	GASIFIER & ACCESSORIES											
4.1	Gasifier, Syngas Cooler & Auxiliaries	$137,274	$0	$59,482	$0	$0	$196,756	$17,552	$27,621	$36,980	$278,909	$555
4.2	Syngas Cooling (w/4.1)	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.3	ASU/Oxidant Compression	$142,499	$0	w/equip.	$0	$0	$142,499	$13,812	$0	$15,631	$171,943	$342
4.4	LT Heat Recovery & FG Saturation	$16,347	$0	$6,214	$0	$0	$22,561	$2,202	$0	$4,953	$29,716	$59
4.5	Misc. Gasification Equipment w/4.1 & 4.2	w/4.1&4.2	$0	w/4.1&4.2	$0	$0	$0	$0	$0	$0	$0	$0
4.6	Other Gasification Equipment	$0	$819	$333	$0	$0	$1,152	$110	$0	$253	$1,515	$3
4.8	Major Component Rigging	w/4.1&4.2	$0	w/4.1&4.2	$0	$0	$0	$0	$0	$0	$0	$0
4.9	Gasification Foundations	$0	$8,073	$4,606	$0	$0	$12,679	$1,161	$0	$3,460	$17,299	$34
	SUBTOTAL 4.	**$296,120**	**$8,892**	**$70,636**	**$0**	**$0**	**$375,648**	**$34,837**	**$27,621**	**$61,276**	**$499,383**	**$994**

Exhibit 4-24 Case 1 Total Plant Cost Details **(Continued)**

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		Case 1 TOTAL PLANT COST	
				Direct	Indirect				Process	Project	$	$/kW
5A	GAS CLEANUP & PIPING											
5A.1	Sulfinol/Selexol System	$33,314	$0	$15,556	$0	$0	$48,870	$4,693	$0	$10,713	$64,275	$128
5A.2	Elemental Sulfur Plant	$4,460	$889	$5,754	$0	$0	$11,103	$1,078	$0	$2,436	$14,617	$29
5A.3	Mercury Removal	$833	$0	$634	$0	$0	$1,466	$142	$73	$336	$2,018	$4
5A.4	COS Hydrolysis/WGS Reactors	$2,459	$0	$3,211	$0	$0	$5,671	$551	$0	$1,244	$7,466	$15
5A.5	Particulate Removal	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
5A.5	Blowback Gas Systems	$1,273	$214	$121	$0	$0	$1,608	$153	$0	$352	$2,113	$4
5A.6	Fuel Gas Piping	$0	$694	$486	$0	$0	$1,179	$109	$0	$258	$1,546	$3
5A.9	HGCU Foundations	$0	$635	$410	$0	$0	$1,045	$96	$0	$342	$1,483	$3
	SUBTOTAL 5A.	**$42,339**	**$2,432**	**$26,171**	**$0**	**$0**	**$70,942**	**$6,822**	**$73**	**$15,681**	**$93,518**	**$186**
5B	CO2 REMOVAL & COMPRESSION											
5B.1	CO2 Removal System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5B.2	CO2 Compression & Drying	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 5B.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Combustion Turbine Generator	$87,466	$0	$6,269	$0	$0	$93,736	$8,886	$4,687	$10,731	$118,040	$235
6.2	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.3	Compressed Air Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.9	Combustion Turbine Foundations	$0	$806	$892	$0	$0	$1,699	$159	$0	$557	$2,415	$5
	SUBTOTAL 6.	**$87,466**	**$806**	**$7,162**	**$0**	**$0**	**$95,434**	**$9,045**	**$4,687**	**$11,288**	**$120,454**	**$240**
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	$30,167	$0	$4,289	$0	$0	$34,457	$3,276	$0	$3,773	$41,506	$83
7.2	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.3	Ductwork	$0	$1,774	$1,298	$0	$0	$3,071	$270	$0	$668	$4,009	$8
7.4	Stack	$3,460	$0	$1,300	$0	$0	$4,760	$456	$0	$522	$5,738	$11
7.9	HRSG,Duct & Stack Foundations	$0	$693	$666	$0	$0	$1,359	$127	$0	$446	$1,931	$4
	SUBTOTAL 7.	**$33,628**	**$2,467**	**$7,553**	**$0**	**$0**	**$43,648**	**$4,128**	**$0**	**$5,409**	**$53,185**	**$106**
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	$26,335	$0	$4,527	$0	$0	$30,862	$2,961	$0	$3,382	$37,206	$74
8.2	Turbine Plant Auxiliaries	$182	$0	$418	$0	$0	$600	$59	$0	$66	$724	$1
8.3a	Condenser & Auxiliaries	$2,602	$0	$831	$0	$0	$3,433	$328	$0	$376	$4,137	$8
8.3b	Air Cooled Condenser	$23,849	$0	$4,781	$0	$0	$28,631	$2,863	$0	$6,299	$37,792	$75
8.4	Steam Piping	$4,958	$0	$3,488	$0	$0	$8,446	$726	$0	$2,293	$11,465	$23
8.9	TG Foundations	$0	$903	$1,527	$0	$0	$2,430	$230	$0	$798	$3,459	$7
	SUBTOTAL 8.	**$57,926**	**$903**	**$15,572**	**$0**	**$0**	**$74,402**	**$7,167**	**$0**	**$13,214**	**$94,783**	**$189**

Exhibit 4-24 Case 1 Total Plant Cost Details **(Continued)**

Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		Case 1 TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$	$/kW
9	COOLING WATER SYSTEM											
9.1	Cooling Towers	$3,494	$0	$636	$0	$0	$4,130	$393	$0	$678	$5,202	$10
9.2	Circulating Water Pumps	$909	$0	$74	$0	$0	$983	$83	$0	$160	$1,227	$2
9.3	Circ.Water System Auxiliaries	$84	$0	$12	$0	$0	$96	$9	$0	$16	$121	$0
9.4	Circ.Water Piping	$0	$3,551	$921	$0	$0	$4,472	$404	$0	$975	$5,851	$12
9.5	Make-up Water System	$187	$0	$268	$0	$0	$455	$44	$0	$100	$599	$1
9.6	Component Cooling Water Sys	$418	$500	$356	$0	$0	$1,274	$119	$0	$279	$1,672	$3
9.9	Circ.Water System Foundations& Structures	$0	$1,349	$2,293	$0	$0	$3,642	$345	$0	$1,196	$5,184	$10
	SUBTOTAL 9.	**$5,093**	**$5,400**	**$4,559**	**$0**	**$0**	**$15,053**	**$1,398**	**$0**	**$3,404**	**$19,855**	**$40**
10	ASH/SPENT SORBENT HANDLING SYS											
10.1	Slag Dewatering & Cooling	$14,744	$0	$7,271	$0	$0	$22,014	$2,115	$0	$2,413	$26,542	$53
10.2	Gasifier Ash Depressurization	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.3	Cleanup Ash Depressurization	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.4	High Temperature Ash Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.5	Other Ash Rrecovery Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.6	Ash Storage Silos	$503	$0	$548	$0	$0	$1,051	$102	$0	$173	$1,326	$3
10.7	Ash Transport & Feed Equipment	$675	$0	$163	$0	$0	$838	$78	$0	$137	$1,054	$2
10.8	Misc. Ash Handling Equipment	$1,043	$1,278	$382	$0	$0	$2,702	$257	$0	$444	$3,403	$7
10.9	Ash/Spent Sorbent Foundation	$0	$45	$56	$0	$0	$100	$9	$0	$33	$143	$0
	SUBTOTAL 10.	**$16,965**	**$1,322**	**$8,419**	**$0**	**$0**	**$26,706**	**$2,562**	**$0**	**$3,200**	**$32,468**	**$65**
11	ACCESSORY ELECTRIC PLANT											
11.1	Generator Equipment	$853	$0	$843	$0	$0	$1,696	$162	$0	$186	$2,044	$4
11.2	Station Service Equipment	$3,688	$0	$332	$0	$0	$4,020	$371	$0	$439	$4,830	$10
11.3	Switchgear & Motor Control	$6,817	$0	$1,240	$0	$0	$8,057	$747	$0	$1,321	$10,125	$20
11.4	Conduit & Cable Tray	$0	$3,167	$10,448	$0	$0	$13,615	$1,317	$0	$3,733	$18,664	$37
11.5	Wire & Cable	$0	$6,051	$3,976	$0	$0	$10,027	$728	$0	$2,689	$13,444	$27
11.6	Protective Equipment	$0	$655	$2,385	$0	$0	$3,041	$297	$0	$501	$3,838	$8
11.7	Standby Equipment	$214	$0	$209	$0	$0	$424	$40	$0	$70	$534	$1
11.8	Main Power Transformers	$13,713	$0	$127	$0	$0	$13,840	$1,047	$0	$2,233	$17,120	$34
11.9	Electrical Foundations	$0	$138	$362	$0	$0	$500	$48	$0	$164	$712	$1
	SUBTOTAL 11.	**$25,285**	**$10,011**	**$19,922**	**$0**	**$0**	**$55,218**	**$4,757**	**$0**	**$11,335**	**$71,310**	**$142**
12	INSTRUMENTATION & CONTROL											
12.1	IGCC Control Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.2	Combustion Turbine Control	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
12.3	Steam Turbine Control	w/8.1	$0	w/8.1	$0	$0	$0	$0	$0	$0	$0	$0
12.4	Other Major Component Control	$978	$0	$653	$0	$0	$1,632	$154	$82	$280	$2,148	$4
12.5	Signal Processing Equipment	W/12.7	$0	W/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.6	Control Boards,Panels & Racks	$225	$0	$144	$0	$0	$369	$35	$18	$84	$507	$1
12.7	Computer & Accessories	$5,220	$0	$167	$0	$0	$5,387	$494	$269	$615	$6,765	$13
12.8	Instrument Wiring & Tubing	$0	$1,823	$3,728	$0	$0	$5,551	$471	$278	$1,575	$7,874	$16
12.9	Other I & C Equipment	$3,489	$0	$1,694	$0	$0	$5,183	$488	$259	$890	$6,820	$14
	SUBTOTAL 12.	**$9,912**	**$1,823**	**$6,386**	**$0**	**$0**	**$18,122**	**$1,642**	**$906**	**$3,444**	**$24,114**	**$48**

Exhibit 4-24 Case 1 Total Plant Cost Details (Continued)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	Case 1 TOTAL PLANT COST $	Case 1 TOTAL PLANT COST $/kW
13	Improvements to Site											
13.1	Site Preparation	$0	$101	$2,164	$0	$0	$2,265	$225	$0	$747	$3,237	$6
13.2	Site Improvements	$0	$1,801	$2,393	$0	$0	$4,194	$414	$0	$1,382	$5,990	$12
13.3	Site Facilities	$3,227	$0	$3,405	$0	$0	$6,632	$654	$0	$2,186	$9,472	$19
	SUBTOTAL 13.	**$3,227**	**$1,902**	**$7,962**	**$0**	**$0**	**$13,092**	**$1,293**	**$0**	**$4,315**	**$18,699**	**$37**
14	Buildings & Structures											
14.1	Combustion Turbine Area	$0	$265	$150	$0	$0	$414	$36	$0	$90	$541	$1
14.2	Steam Turbine Building	$0	$2,429	$3,460	$0	$0	$5,889	$542	$0	$965	$7,395	$15
14.3	Administration Building	$0	$825	$598	$0	$0	$1,423	$127	$0	$232	$1,782	$4
14.4	Circulation Water Pumphouse	$0	$153	$81	$0	$0	$234	$21	$0	$38	$293	$1
14.5	Water Treatment Buildings	$0	$258	$252	$0	$0	$510	$46	$0	$83	$639	$1
14.6	Machine Shop	$0	$425	$291	$0	$0	$716	$64	$0	$117	$896	$2
14.7	Warehouse	$0	$686	$443	$0	$0	$1,129	$100	$0	$184	$1,413	$3
14.8	Other Buildings & Structures	$0	$409	$318	$0	$0	$727	$65	$0	$158	$951	$2
14.9	Waste Treating Building & Str.	$0	$903	$1,725	$0	$0	$2,627	$245	$0	$574	$3,447	$7
	SUBTOTAL 14.	**$0**	**$6,351**	**$7,318**	**$0**	**$0**	**$13,669**	**$1,245**	**$0**	**$2,443**	**$17,357**	**$35**
	TOTAL COST	**$711,351**	**$60,430**	**$217,637**	**$0**	**$0**	**$989,418**	**$91,410**	**$33,287**	**$176,300**	**$1,290,415**	**$2,569**
	Owner's Costs											
	Preproduction Costs											
	6 Months All Labor										$11,703	$23
	1 Month Maintenance Materials										$2,567	$5
	1 Month Non-fuel Consumables										$92	$0
	1 Month Waste Disposal										$240	$0
	25% of 1 Months Fuel Cost at 100% CF										$453	$1
	2% of TPC										$25,808	$51
	Total										**$40,863**	**$81**
	Inventory Capital											
	60 day supply of fuel and consumables at 100% CF										$3,808	$8
	0.5% of TPC (spare parts)										$6,452	$13
	Total										**$10,260**	**$20**
	Initial Cost for Catalyst and Chemicals										$568	$1
	Land										$900	$2
	Other Owner's Costs										$193,562	$385
	Financing Costs										$34,841	$69
	Total Overnight Costs (TOC)										**$1,571,409**	**$3,128**
	TASC Multiplier								(IOU, high-risk, 35 year)		1.140	
	Total As-Spent Cost (TASC)										**$1,791,407**	**$3,566**

Exhibit 4-25 Case 1 Initial and Annual Operating and Maintenance Costs

INITIAL & ANNUAL O&M EXPENSES					Cost Base (June)	2007
Case 1 - Shell IGCC w/o CO2					Heat Rate-net(Btu/kWh):	8,160
					MWe-net:	502
					Capacity Factor: (%):	80
OPERATING & MAINTENANCE LABOR						
Operating Labor						
Operating Labor Rate(base):	34.65	$/hour				
Operating Labor Burden:	30.00	% of base				
Labor O-H Charge Rate:	25.00	% of labor				
			Total			
Skilled Operator	2.0		2.0			
Operator	9.0		9.0			
Foreman	1.0		1.0			
Lab Tech's, etc.	3.0		3.0			
TOTAL-O.J.'s	15.0		15.0			
					Annual Cost	Annual Unit Cost
					$	$/kW-net
Annual Operating Labor Cost	Maintenance labor cost		% of BEC	1.2942	$5,918,913	$11.782
Maintenance Labor Cost	(Case S1A is reference)		BEC	$989,418	$12,805,347	$25.490
Administrative & Support Labor					$4,681,065	$9.318
Property Taxes & Insurance					$25,808,296	$51.373
TOTAL FIXED OPERATING COSTS					**$49,213,621.24**	**$97.963**
VARIABLE OPERATING COSTS						
						$/kWh-net
Maintenance Material Cost			% of BEC	2.4905	**$24,641,658**	**$0.00700**
Consumables	Consumption		Unit	Initial Fill		
	Initial Fill	/Day	Cost	Cost		
Water(/1000 gallons)	0	1,164	1.08	**$0**	**$367,497**	**$0.00010**
Chemicals		5.959				
MU & WT Chem.(lb)	0	6,934	0.17	$0	$350,395	$0.00010
Carbon (Mercury Removal) (lb)	56,520	77	1.05	$59,356	$23,742	$0.00001
COS Catalyst (m3)	212	0.15	2,397.36	$509,009	$101,732	$0.00003
Water Gas Shift Catalyst(ft3)	0	0	498.83	$0	$0	$0.00000
Selexol Solution (gal.)	0	0	13.40	$0	$0	$0.00000
MDEA Solution (gal)	0	0	8.70	$0	$0	$0.00000
Sulfinol Solution (gal)	w/equip.	7	10.05	$0	$19,366	$0.00001
SCR Catalyst (m3)	0	0	0.00	$0	$0	$0.00000
Aqueous Ammonia (ton)	0	0	0.00	$0	$0	$0.00000
Claus Catalyst(ft3)	w/equip.	0.63	131.27	$0	$24,232	$0.00001
Subtotal Chemicals				**$568,365**	**$519,468**	**$0.00015**
Other						
Supplemental Fuel(MBtu)	0	0	0.00	$0	$0	$0.00000
Gases,N2 etc.(/100scf)	0	0	0.00	$0	$0	$0.00000
L.P. Steam(/1000 pounds)	0	0	0.00	$0	$0	$0.00000
Subtotal Other				**$0**	**$0**	**$0.00000**
Waste Disposal						
Spent Mercury Catalyst (lb)	0	77	0.42	$0	$9,429	$0.00000
Flyash (ton)	0	0	0.00	$0	$0	$0.00000
Slag (ton)	0	485	16.23	$0	$2,296,856	$0.00065
Subtotal-Waste Disposal				**$0**	**$2,306,285**	**$0.00066**
By-products & Emissions						
Sulfur(tons)	0	42	0.00	$0	$0	$0.00000
Subtotal By-Products				**$0**	**$0**	**$0.00000**
TOTAL VARIABLE OPERATING COSTS				**$568,365**	**$27,834,907.70**	**$0.00791**
Fuel(ton)	0	5,744	10.367	**$0**	**$17,389,132**	**$0.00494**

4.1.8 Case 2 – Cost Estimating

Exhibit 4-26 shows the TPC cost details organized by cost account as well as TOC and TASC. Exhibit 4-27 shows the initial and annual O&M costs.

The estimated TOC of the IGCC case with an emission rate of 1,100 lb CO_2/net-MWh is $3,938/kW. Owner's costs represent 18 percent of the TOC. The current dollar, 30-year LCOE, including TS&M, is $149.33/MWh.

Exhibit 4-26 Case 2 Total Plant Cost Details

Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		Case 2 TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$	$/kW
1	COAL & SORBENT HANDLING											
1.1	Coal Receive & Unload	$3,790	$0	$1,852	$0	$0	$5,642	$505	$0	$1,229	$7,377	$17
1.2	Coal Stackout & Reclaim	$4,897	$0	$1,187	$0	$0	$6,084	$533	$0	$1,324	$7,941	$18
1.3	Coal Conveyors	$4,553	$0	$1,175	$0	$0	$5,728	$503	$0	$1,246	$7,477	$17
1.4	Other Coal Handling	$1,191	$0	$272	$0	$0	$1,463	$128	$0	$318	$1,909	$4
1.5	Sorbent Receive & Unload	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.6	Sorbent Stackout & Reclaim	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.7	Sorbent Conveyors	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.8	Other Sorbent Handling	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.9	Coal & Sorbent Hnd.Foundations	$0	$2,682	$6,706	$0	$0	$9,387	$900	$0	$2,057	$12,345	$28
	SUBTOTAL 1.	**$14,431**	**$2,682**	**$11,191**	**$0**	**$0**	**$28,304**	**$2,569**	**$0**	**$6,175**	**$37,048**	**$84**
2	COAL & SORBENT PREP & FEED											
2.1	Coal Crushing & Drying	$43,529	$2,615	$6,343	$0	$0	$52,487	$4,529	$0	$11,403	$68,419	$155
2.2	Prepared Coal Storage & Feed	$2,062	$493	$323	$0	$0	$2,878	$246	$0	$625	$3,749	$8
2.3	Dry Coal Injection System	$67,853	$788	$6,301	$0	$0	$74,942	$6,455	$0	$16,279	$97,676	$221
2.4	Misc.Coal Prep & Feed	$1,134	$825	$2,474	$0	$0	$4,433	$407	$0	$968	$5,808	$13
2.5	Sorbent Prep Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.6	Sorbent Storage & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.7	Sorbent Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.8	Booster Air Supply System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.9	Coal & Sorbent Feed Foundation	$0	$4,407	$3,618	$0	$0	$8,025	$743	$0	$1,754	$10,522	$24
	SUBTOTAL 2.	**$114,577**	**$9,128**	**$19,059**	**$0**	**$0**	**$142,764**	**$12,381**	**$0**	**$31,029**	**$186,174**	**$420**
3	FEEDWATER & MISC. BOP SYSTEMS											
3.1	FeedwaterSystem	$2,128	$3,654	$1,929	$0	$0	$7,711	$714	$0	$1,685	$10,111	$23
3.2	Water Makeup & Pretreating	$384	$40	$215	$0	$0	$639	$61	$0	$210	$910	$2
3.3	Other Feedwater Subsystems	$1,164	$393	$354	$0	$0	$1,912	$172	$0	$417	$2,500	$6
3.4	Service Water Systems	$220	$453	$1,571	$0	$0	$2,243	$219	$0	$739	$3,201	$7
3.5	Other Boiler Plant Systems	$1,180	$457	$1,133	$0	$0	$2,769	$263	$0	$606	$3,638	$8
3.6	FO Supply Sys & Nat Gas	$289	$546	$509	$0	$0	$1,344	$129	$0	$295	$1,768	$4
3.7	Waste Treatment Equipment	$537	$0	$328	$0	$0	$864	$84	$0	$285	$1,233	$3
3.8	Misc. Equip.(cranes,AirComp.,Comm.)	$993	$133	$510	$0	$0	$1,636	$158	$0	$538	$2,332	$5
	SUBTOTAL 3.	**$6,894**	**$5,676**	**$6,548**	**$0**	**$0**	**$19,118**	**$1,800**	**$0**	**$4,774**	**$25,692**	**$58**
4	GASIFIER & ACCESSORIES											
4.1	Gasifier, Syngas Cooler & Auxiliaries	$116,362	$0	$49,863	$0	$0	$166,225	$14,844	$22,983	$31,320	$235,372	$532
4.2	Syngas Cooling (w/4.1)	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.3	ASU/Oxidant Compression	$160,585	$0	w/equip.	$0	$0	$160,585	$15,565	$0	$17,615	$193,765	$438
4.4	LT Heat Recovery & FG Saturation	$26,641	$0	$10,128	$0	$0	$36,768	$3,588	$0	$8,071	$48,428	$109
4.5	Misc. Gasification Equipment w/4.1 & 4.2	w/4.1&4.2	$0	w/4.1&4.2	$0	$0	$0	$0	$0	$0	$0	$0
4.6	Other Gasification Equipment	$0	$1,646	$670	$0	$0	$2,316	$222	$0	$508	$3,046	$7
4.8	Major Component Rigging	w/4.1&4.2	$0	w/4.1&4.2	$0	$0	$0	$0	$0	$0	$0	$0
4.9	Gasification Foundations	$0	$8,212	$4,686	$0	$0	$12,898	$1,181	$0	$3,520	$17,598	$40
	SUBTOTAL 4.	**$303,588**	**$9,858**	**$65,346**	**$0**	**$0**	**$378,792**	**$35,401**	**$22,983**	**$61,033**	**$498,209**	**$1,125**

Exhibit 4-26 Case 2 Total Plant Cost Details (Continued)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		Case 2 TOTAL PLANT COST	
				Direct	Indirect				Process	Project	$	$/kW
5A	GAS CLEANUP & PIPING											
5A.1	Sulfinol/Selexol System	$56,111	$0	$48,095	$0	$0	$104,206	$10,079	$20,841	$27,025	$162,152	$366
5A.2	Elemental Sulfur Plant	$4,449	$887	$5,740	$0	$0	$11,075	$1,076	$0	$2,430	$14,581	$33
5A.3	Mercury Removal	$1,416	$0	$1,078	$0	$0	$2,494	$241	$125	$572	$3,432	$8
5A.4	COS Hydrolysis/WGS Reactors	$5,066	$0	$2,039	$0	$0	$7,105	$681	$0	$1,557	$9,343	$21
5A.5	Particulate Removal	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
5A.5	Blowback Gas Systems	$1,850	$311	$175	$0	$0	$2,337	$222	$0	$512	$3,070	$7
5A.6	Fuel Gas Piping	$0	$699	$490	$0	$0	$1,189	$110	$0	$260	$1,559	$4
5A.9	HGCU Foundations	$0	$687	$443	$0	$0	$1,131	$104	$0	$370	$1,605	$4
	SUBTOTAL 5A.	**$68,893**	**$2,585**	**$58,059**	**$0**	**$0**	**$129,537**	**$12,513**	**$20,966**	**$32,727**	**$195,742**	**$442**
5B	CO2 REMOVAL & COMPRESSION											
5B.1	CO2 Removal System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5B.2	CO2 Compression & Drying	$8,653	$0	$5,015	$0	$0	$13,668	$1,316	$0	$2,997	$17,980	$41
	SUBTOTAL 5B.	**$8,653**	**$0**	**$5,015**	**$0**	**$0**	**$13,668**	**$1,316**	**$0**	**$2,997**	**$17,980**	**$41**
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Combustion Turbine Generator	$93,866	$0	$6,583	$0	$0	$100,449	$9,522	$10,045	$12,002	$132,017	$298
6.2	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.3	Compressed Air Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.9	Combustion Turbine Foundations	$0	$806	$892	$0	$0	$1,699	$159	$0	$557	$2,415	$5
	SUBTOTAL 6.	**$93,866**	**$806**	**$7,475**	**$0**	**$0**	**$102,147**	**$9,681**	**$10,045**	**$12,559**	**$134,432**	**$304**
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	$29,527	$0	$4,198	$0	$0	$33,725	$3,207	$0	$3,693	$40,625	$92
7.2	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.3	Ductwork	$0	$1,794	$1,312	$0	$0	$3,106	$273	$0	$676	$4,055	$9
7.4	Stack	$3,499	$0	$1,315	$0	$0	$4,814	$461	$0	$528	$5,803	$13
7.9	HRSG,Duct & Stack Foundations	$0	$701	$673	$0	$0	$1,374	$128	$0	$451	$1,953	$4
	SUBTOTAL 7.	**$33,026**	**$2,495**	**$7,499**	**$0**	**$0**	**$43,020**	**$4,069**	**$0**	**$5,347**	**$52,435**	**$118**
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	$23,844	$0	$3,994	$0	$0	$27,838	$2,671	$0	$3,051	$33,560	$76
8.2	Turbine Plant Auxiliaries	$164	$0	$376	$0	$0	$540	$53	$0	$59	$652	$1
8.3a	Condenser & Auxiliaries	$2,631	$0	$840	$0	$0	$3,471	$332	$0	$380	$4,184	$9
8.3b	Air Cooled Condenser	$24,118	$0	$4,835	$0	$0	$28,953	$2,895	$0	$6,370	$38,218	$86
8.4	Steam Piping	$3,908	$0	$2,749	$0	$0	$6,658	$572	$0	$1,807	$9,037	$20
8.9	TG Foundations	$0	$813	$1,374	$0	$0	$2,187	$207	$0	$718	$3,113	$7
	SUBTOTAL 8.	**$54,665**	**$813**	**$14,169**	**$0**	**$0**	**$69,647**	**$6,730**	**$0**	**$12,386**	**$88,763**	**$200**

Exhibit 4-26 Case 2 Total Plant Cost Details (Continued)

Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		Case 2 TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$	$/kW
9	COOLING WATER SYSTEM											
9.1	Cooling Towers	$4,188	$0	$762	$0	$0	$4,950	$471	$0	$813	$6,235	$14
9.2	Circulating Water Pumps	$1,097	$0	$73	$0	$0	$1,170	$99	$0	$190	$1,459	$3
9.3	Circ.Water System Auxiliaries	$98	$0	$14	$0	$0	$112	$11	$0	$18	$141	$0
9.4	Circ.Water Piping	$0	$4,155	$1,077	$0	$0	$5,232	$473	$0	$1,141	$6,846	$15
9.5	Make-up Water System	$238	$0	$341	$0	$0	$579	$56	$0	$127	$762	$2
9.6	Component Cooling Water Sys	$488	$584	$415	$0	$0	$1,487	$139	$0	$325	$1,951	$4
9.9	Circ.Water System Foundations& Structures	$0	$1,578	$2,682	$0	$0	$4,260	$404	$0	$1,399	$6,063	$14
	SUBTOTAL 9.	**$6,110**	**$6,316**	**$5,364**	**$0**	**$0**	**$17,790**	**$1,652**	**$0**	**$4,014**	**$23,457**	**$53**
10	ASH/SPENT SORBENT HANDLING SYS											
10.1	Slag Dewatering & Cooling	$15,071	$0	$7,432	$0	$0	$22,503	$2,162	$0	$2,466	$27,131	$61
10.2	Gasifier Ash Depressurization	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.3	Cleanup Ash Depressurization	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.4	High Temperature Ash Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.5	Other Ash Rrecovery Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.6	Ash Storage Silos	$513	$0	$558	$0	$0	$1,071	$104	$0	$176	$1,352	$3
10.7	Ash Transport & Feed Equipment	$688	$0	$166	$0	$0	$854	$80	$0	$140	$1,074	$2
10.8	Misc. Ash Handling Equipment	$1,063	$1,302	$389	$0	$0	$2,754	$262	$0	$452	$3,469	$8
10.9	Ash/Spent Sorbent Foundation	$0	$45	$57	$0	$0	$102	$10	$0	$34	$146	$0
	SUBTOTAL 10.	**$17,335**	**$1,348**	**$8,602**	**$0**	**$0**	**$27,285**	**$2,617**	**$0**	**$3,269**	**$33,171**	**$75**
11	ACCESSORY ELECTRIC PLANT											
11.1	Generator Equipment	$830	$0	$821	$0	$0	$1,650	$158	$0	$181	$1,989	$4
11.2	Station Service Equipment	$4,109	$0	$370	$0	$0	$4,479	$413	$0	$489	$5,381	$12
11.3	Switchgear & Motor Control	$7,596	$0	$1,382	$0	$0	$8,978	$833	$0	$1,472	$11,282	$25
11.4	Conduit & Cable Tray	$0	$3,529	$11,641	$0	$0	$15,170	$1,467	$0	$4,159	$20,797	$47
11.5	Wire & Cable	$0	$6,742	$4,430	$0	$0	$11,172	$812	$0	$2,996	$14,980	$34
11.6	Protective Equipment	$0	$655	$2,385	$0	$0	$3,041	$297	$0	$501	$3,838	$9
11.7	Standby Equipment	$210	$0	$205	$0	$0	$414	$40	$0	$68	$522	$1
11.8	Main Power Transformers	$15,376	$0	$124	$0	$0	$15,500	$1,172	$0	$2,501	$19,173	$43
11.9	Electrical Foundations	$0	$133	$350	$0	$0	$483	$46	$0	$159	$689	$2
	SUBTOTAL 11.	**$28,121**	**$11,060**	**$21,708**	**$0**	**$0**	**$60,889**	**$5,237**	**$0**	**$12,526**	**$78,652**	**$178**
12	INSTRUMENTATION & CONTROL											
12.1	IGCC Control Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.2	Combustion Turbine Control	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
12.3	Steam Turbine Control	w/8.1	$0	w/8.1	$0	$0	$0	$0	$0	$0	$0	$0
12.4	Other Major Component Control	$1,013	$0	$677	$0	$0	$1,690	$160	$84	$290	$2,225	$5
12.5	Signal Processing Equipment	W/12.7	$0	W/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.6	Control Boards,Panels & Racks	$233	$0	$149	$0	$0	$382	$36	$19	$87	$525	$1
12.7	Computer & Accessories	$5,405	$0	$173	$0	$0	$5,578	$512	$279	$637	$7,006	$16
12.8	Instrument Wiring & Tubing	$0	$1,888	$3,860	$0	$0	$5,749	$488	$287	$1,631	$8,154	$18
12.9	Other I & C Equipment	$3,613	$0	$1,755	$0	$0	$5,368	$505	$268	$921	$7,062	$16
	SUBTOTAL 12.	**$10,265**	**$1,888**	**$6,614**	**$0**	**$0**	**$18,767**	**$1,701**	**$938**	**$3,567**	**$24,973**	**$56**

Exhibit 4-26 Case 2 Total Plant Cost Details (Continued)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	Case 2 TOTAL PLANT COST $	Case 2 TOTAL PLANT COST $/kW
13	Improvements to Site											
13.1	Site Preparation	$0	$102	$2,179	$0	$0	$2,281	$226	$0	$752	$3,259	$7
13.2	Site Improvements	$0	$1,813	$2,409	$0	$0	$4,223	$417	$0	$1,392	$6,031	$14
13.3	Site Facilities	$3,249	$0	$3,429	$0	$0	$6,678	$658	$0	$2,201	$9,537	$22
	SUBTOTAL 13.	**$3,249**	**$1,915**	**$8,017**	**$0**	**$0**	**$13,181**	**$1,301**	**$0**	**$4,345**	**$18,827**	**$43**
14	Buildings & Structures											
14.1	Combustion Turbine Area	$0	$265	$150	$0	$0	$414	$36	$0	$90	$541	$1
14.2	Steam Turbine Building	$0	$2,062	$2,938	$0	$0	$5,000	$460	$0	$819	$6,279	$14
14.3	Administration Building	$0	$833	$605	$0	$0	$1,438	$128	$0	$235	$1,801	$4
14.4	Circulation Water Pumphouse	$0	$140	$74	$0	$0	$215	$19	$0	$35	$269	$1
14.5	Water Treatment Buildings	$0	$343	$335	$0	$0	$678	$61	$0	$111	$850	$2
14.6	Machine Shop	$0	$431	$295	$0	$0	$726	$64	$0	$119	$909	$2
14.7	Warehouse	$0	$696	$449	$0	$0	$1,145	$101	$0	$187	$1,434	$3
14.8	Other Buildings & Structures	$0	$418	$325	$0	$0	$743	$66	$0	$162	$971	$2
14.9	Waste Treating Building & Str.	$0	$903	$1,725	$0	$0	$2,628	$245	$0	$575	$3,448	$8
	SUBTOTAL 14.	**$0**	**$6,091**	**$6,896**	**$0**	**$0**	**$12,987**	**$1,182**	**$0**	**$2,332**	**$16,501**	**$37**
	TOTAL COST	**$763,672**	**$62,661**	**$251,562**	**$0**	**$0**	**$1,077,896**	**$100,149**	**$54,932**	**$199,078**	**$1,432,055**	**$3,234**
	Owner's Costs											
	Preproduction Costs											
	6 Months All Labor										$12,092	$27
	1 Month Maintenance Materials										$2,573	$6
	1 Month Non-fuel Consumables										$164	$0
	1 Month Waste Disposal										$249	$1
	25% of 1 Months Fuel Cost at 100% CF										$469	$1
	2% of TPC										$28,641	$65
	Total										**$44,188**	**$100**
	Inventory Capital											
	60 day supply of fuel and consumables at 100% CF										$4,077	$9
	0.5% of TPC (spare parts)										$7,160	$16
	Total										**$11,238**	**$25**
	Initial Cost for Catalyst and Chemicals										$1,558	$4
	Land										$900	$2
	Other Owner's Costs										$214,808	$485
	Financing Costs										$38,665	$87
	Total Overnight Costs (TOC)										**$1,743,413**	**$3,937.6**
	TASC Multiplier								(IOU, high-risk, 35 year)		1.140	
	Total As-Spent Cost (TASC)										**$1,987,490**	**$4,489**

Exhibit 4-27 Case 2 Initial and Annual Operating and Maintenance Costs

INITIAL & ANNUAL O&M EXPENSES					Cost Base (June)	2007
Case 2 - Shell IGCC w/ CO2 capture (1,100 lb/net MWh)					Heat Rate-net(Btu/kWh):	9,581
					MWe-net:	443
					Capacity Factor: (%):	80
OPERATING & MAINTENANCE LABOR						
Operating Labor						
Operating Labor Rate(base):	34.65	$/hour				
Operating Labor Burden:	30.00	% of base				
Labor O-H Charge Rate:	25.00	% of labor				
			Total			
Skilled Operator	2.0		2.0			
Operator	10.0		10.0			
Foreman	1.0		1.0			
Lab Tech's, etc.	3.0		3.0			
TOTAL-O.J.'s	16.0		16.0			
					Annual Cost $	Annual Unit Cost $/kW-net
Annual Operating Labor Cost	Maintenance labor cost		% of BEC	1.2092	$6,313,507	$14.259
Maintenance Labor Cost	(Case S1B is reference)		BEC	$1,077,896	$13,033,374	$29.437
Administrative & Support Labor					$4,836,720	$10.924
Property Taxes & Insurance					$28,641,102	$64.688
TOTAL FIXED OPERATING COSTS					**$52,824,703.55**	**$119.308**
VARIABLE OPERATING COSTS						
						$/kWh-net
Maintenance Material Cost			% of BEC	2.2918	**$24,702,887**	**$0.00796**
Consumables	Consumption		Unit Cost	Initial Fill Cost		
	Initial Fill	/Day				
Water(/1000 gallons)	0	1,737	1.08	**$0**	**$548,516**	**$0.00018**
Chemicals		5.959				
MU & WT Chem.(lb)	0	10,349	0.17	$0	$522,991	$0.00017
Carbon (Mercury Removal) (lb)	79,200	108	1.05	$83,174	$33,269	$0.00001
COS Catalyst (m3)	0	0.00	2,397.36	$0	$0	$0.00000
Water Gas Shift Catalyst(ft3)	2,957	2.0	498.83	$1,475,043	$294,807	$0.00010
Selexol Solution (gal.)	w/equip.	39	13.40	$0	$153,362	$0.00005
MDEA Solution (gal)	0	0	8.70	$0	$0	$0.00000
Sulfinol Solution (gal)	0	0	10.05	$0	$0	$0.00000
SCR Catalyst (m3)	0	0	0.00	$0	$0	$0.00000
Aqueous Ammonia (ton)	0	0	0.00	$0	$0	$0.00000
Claus Catalyst(ft3)	w/equip.	0.63	131.27	$0	$24,142	$0.00001
Subtotal Chemicals				**$1,558,217**	**$1,028,570**	**$0.00033**
Other						
Supplemental Fuel(MBtu)	0	0	0.00	$0	$0	$0.00000
Gases,N2 etc.(/100scf)	0	0	0.00	$0	$0	$0.00000
L.P. Steam(/1000 pounds)	0	0	0.00	$0	$0	$0.00000
Subtotal Other				**$0**	**$0**	**$0.00000**
Waste Disposal						
Spent Mercury Catalyst (lb)	0	108	0.42	$0	$13,213	$0.00000
Flyash (ton)	0	0	0.00	$0	$0	$0.00000
Slag (ton)	0	502	16.23	$0	$2,376,450	$0.00077
Subtotal-Waste Disposal				**$0**	**$2,389,663**	**$0.00077**
By-products & Emissions						
Sulfur(tons)	0	42	0.00	$0	$0	$0.00000
Subtotal By-Products				**$0**	**$0**	**$0.00000**
TOTAL VARIABLE OPERATING COSTS				**$1,558,217**	**$28,669,636.71**	**$0.00924**
Fuel(ton)	0	5,944	10.37	**$0**	**$17,994,286**	**$0.00580**

4.1.9 Case 3 – Cost Estimating

Exhibit 4-28 shows the TPC cost details organized by cost account as well as TOC and TASC. Exhibit 4-29 shows the initial and annual O&M costs.

The estimated TOC of the IGCC case with 90 percent carbon capture is $4,595/kW. Owner’s costs represent 18 percent of the TOC. The current dollar, 30-year LCOE, including TS&M, is $174.86/MWh.

Exhibit 4-28 Case 3 Total Plant Cost Details

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		Case 3 TOTAL PLANT COST	
				Direct	Indirect				Process	Project	$	$/kW
1	COAL & SORBENT HANDLING											
1.1	Coal Receive & Unload	$3,893	$0	$1,903	$0	$0	$5,796	$519	$0	$1,263	$7,578	$19
1.2	Coal Stackout & Reclaim	$5,031	$0	$1,220	$0	$0	$6,251	$548	$0	$1,360	$8,158	$20
1.3	Coal Conveyors	$4,677	$0	$1,207	$0	$0	$5,884	$517	$0	$1,280	$7,681	$19
1.4	Other Coal Handling	$1,224	$0	$279	$0	$0	$1,503	$132	$0	$327	$1,961	$5
1.5	Sorbent Receive & Unload	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.6	Sorbent Stackout & Reclaim	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.7	Sorbent Conveyors	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.8	Other Sorbent Handling	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.9	Coal & Sorbent Hnd.Foundations	$0	$2,755	$6,889	$0	$0	$9,644	$924	$0	$2,114	$12,682	$32
	SUBTOTAL 1.	**$14,825**	**$2,755**	**$11,497**	**$0**	**$0**	**$29,077**	**$2,639**	**$0**	**$6,343**	**$38,060**	**$95**
2	COAL & SORBENT PREP & FEED											
2.1	Coal Crushing & Drying	$44,796	$2,691	$6,527	$0	$0	$54,014	$4,661	$0	$11,735	$70,410	$176
2.2	Prepared Coal Storage & Feed	$2,122	$508	$333	$0	$0	$2,962	$253	$0	$643	$3,859	$10
2.3	Dry Coal Injection System	$69,827	$810	$6,485	$0	$0	$77,122	$6,642	$0	$16,753	$100,518	$251
2.4	Misc.Coal Prep & Feed	$1,167	$849	$2,546	$0	$0	$4,561	$419	$0	$996	$5,977	$15
2.5	Sorbent Prep Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.6	Sorbent Storage & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.7	Sorbent Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.8	Booster Air Supply System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.9	Coal & Sorbent Feed Foundation	$0	$4,535	$3,723	$0	$0	$8,258	$765	$0	$1,805	$10,828	$27
	SUBTOTAL 2.	**$117,911**	**$9,393**	**$19,614**	**$0**	**$0**	**$146,918**	**$12,741**	**$0**	**$31,932**	**$191,591**	**$478**
3	FEEDWATER & MISC. BOP SYSTEMS											
3.1	FeedwaterSystem	$2,165	$3,719	$1,963	$0	$0	$7,847	$727	$0	$1,715	$10,289	$26
3.2	Water Makeup & Pretreating	$462	$48	$258	$0	$0	$768	$73	$0	$252	$1,093	$3
3.3	Other Feedwater Subsystems	$1,185	$400	$360	$0	$0	$1,945	$175	$0	$424	$2,544	$6
3.4	Service Water Systems	$264	$544	$1,888	$0	$0	$2,695	$263	$0	$888	$3,846	$10
3.5	Other Boiler Plant Systems	$1,417	$549	$1,361	$0	$0	$3,328	$316	$0	$729	$4,372	$11
3.6	FO Supply Sys & Nat Gas	$292	$551	$514	$0	$0	$1,357	$131	$0	$298	$1,785	$4
3.7	Waste Treatment Equipment	$645	$0	$394	$0	$0	$1,039	$101	$0	$342	$1,482	$4
3.8	Misc. Equip.(cranes,AirComp.,Comm.)	$1,003	$134	$515	$0	$0	$1,652	$160	$0	$543	$2,355	$6
	SUBTOTAL 3.	**$7,433**	**$5,946**	**$7,252**	**$0**	**$0**	**$20,631**	**$1,945**	**$0**	**$5,190**	**$27,766**	**$69**
4	GASIFIER & ACCESSORIES											
4.1	Gasifier, Syngas Cooler & Auxiliaries	$117,054	$0	$50,159	$0	$0	$167,213	$14,932	$23,119	$31,506	$236,770	$590
4.2	Syngas Cooling (w/4.1)	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.3	ASU/Oxidant Compression	$166,026	$0	w/equip.	$0	$0	$166,026	$16,093	$0	$18,212	$200,331	$499
4.4	LT Heat Recovery & FG Saturation	$26,799	$0	$10,188	$0	$0	$36,987	$3,610	$0	$8,119	$48,716	$121
4.5	Misc. Gasification Equipment w/4.1 & 4.2	w/4.1&4.2	$0	w/4.1&4.2	$0	$0	$0	$0	$0	$0	$0	$0
4.6	Other Gasification Equipment	$0	$1,029	$419	$0	$0	$1,447	$139	$0	$317	$1,903	$5
4.8	Major Component Rigging	w/4.1&4.2	$0	w/4.1&4.2	$0	$0	$0	$0	$0	$0	$0	$0
4.9	Gasification Foundations	$0	$8,392	$4,789	$0	$0	$13,181	$1,207	$0	$3,597	$17,984	$45
	SUBTOTAL 4.	**$309,879**	**$9,421**	**$65,554**	**$0**	**$0**	**$384,854**	**$35,980**	**$23,119**	**$61,751**	**$505,705**	**$1,261**

Exhibit 4-28 Case 3 Total Plant Cost Details (Continued)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		Case 3 TOTAL PLANT COST	
				Direct	Indirect				Process	Project	$	$/kW
5A	GAS CLEANUP & PIPING											
5A.1	Sulfinol/Selexol System	$69,149	$0	$59,269	$0	$0	$128,418	$12,421	$25,684	$33,304	$199,826	$498
5A.2	Elemental Sulfur Plant	$4,695	$936	$6,057	$0	$0	$11,688	$1,135	$0	$2,565	$15,388	$38
5A.3	Mercury Removal	$2,145	$0	$1,633	$0	$0	$3,778	$365	$189	$866	$5,198	$13
5A.4	COS Hydrolysis/WGS Reactors	$7,705	$0	$3,101	$0	$0	$10,806	$1,036	$0	$2,368	$14,210	$35
5A.5	Particulate Removal	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
5A.5	Blowback Gas Systems	$1,874	$315	$178	$0	$0	$2,367	$225	$0	$518	$3,110	$8
5A.6	Fuel Gas Piping	$0	$735	$515	$0	$0	$1,249	$116	$0	$273	$1,638	$4
5A.9	HGCU Foundations	$0	$732	$472	$0	$0	$1,205	$111	$0	$395	$1,710	$4
	SUBTOTAL 5A.	**$85,568**	**$2,718**	**$71,224**	**$0**	**$0**	**$159,511**	**$15,408**	**$25,872**	**$40,290**	**$241,081**	**$601**
5B	CO2 REMOVAL & COMPRESSION											
5B.1	CO2 Removal System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5B.2	CO2 Compression & Drying	$16,187	$0	$9,381	$0	$0	$25,568	$2,461	$0	$5,606	$33,635	$84
	SUBTOTAL 5B.	**$16,187**	**$0**	**$9,381**	**$0**	**$0**	**$25,568**	**$2,461**	**$0**	**$5,606**	**$33,635**	**$84**
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Combustion Turbine Generator	$93,866	$0	$6,583	$0	$0	$100,449	$9,522	$10,045	$12,002	$132,017	$329
6.2	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.3	Compressed Air Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.9	Combustion Turbine Foundations	$0	$806	$892	$0	$0	$1,699	$159	$0	$557	$2,415	$6
	SUBTOTAL 6.	**$93,866**	**$806**	**$7,475**	**$0**	**$0**	**$102,147**	**$9,681**	**$10,045**	**$12,559**	**$134,432**	**$335**
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	$28,950	$0	$4,116	$0	$0	$33,067	$3,144	$0	$3,621	$39,832	$99
7.2	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.3	Ductwork	$0	$1,814	$1,327	$0	$0	$3,142	$276	$0	$684	$4,101	$10
7.4	Stack	$3,540	$0	$1,330	$0	$0	$4,869	$467	$0	$534	$5,869	$15
7.9	HRSG,Duct & Stack Foundations	$0	$709	$681	$0	$0	$1,390	$129	$0	$456	$1,976	$5
	SUBTOTAL 7.	**$32,490**	**$2,524**	**$7,455**	**$0**	**$0**	**$42,468**	**$4,016**	**$0**	**$5,294**	**$51,778**	**$129**
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	$22,586	$0	$3,783	$0	$0	$26,369	$2,530	$0	$2,890	$31,789	$79
8.2	Turbine Plant Auxiliaries	$155	$0	$355	$0	$0	$511	$50	$0	$56	$617	$2
8.3a	Condenser & Auxiliaries	$2,513	$0	$803	$0	$0	$3,315	$317	$0	$363	$3,996	$10
8.3b	Air Cooled Condenser	$23,035	$0	$4,618	$0	$0	$27,654	$2,765	$0	$6,084	$36,503	$91
8.4	Steam Piping	$3,975	$0	$2,796	$0	$0	$6,772	$582	$0	$1,838	$9,192	$23
8.9	TG Foundations	$0	$769	$1,300	$0	$0	$2,069	$196	$0	$680	$2,945	$7
	SUBTOTAL 8.	**$52,264**	**$769**	**$13,656**	**$0**	**$0**	**$66,690**	**$6,440**	**$0**	**$11,911**	**$85,041**	**$212**

Exhibit 4-28 Case 3 Total Plant Cost Details (Continued)

Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		Case 3 TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$	$/kW
9	COOLING WATER SYSTEM											
9.1	Cooling Towers	$4,616	$0	$840	$0	$0	$5,456	$520	$0	$896	$6,872	$17
9.2	Circulating Water Pumps	$1,206	$0	$80	$0	$0	$1,286	$108	$0	$209	$1,603	$4
9.3	Circ.Water System Auxiliaries	$107	$0	$15	$0	$0	$123	$12	$0	$20	$154	$0
9.4	Circ.Water Piping	$0	$4,527	$1,174	$0	$0	$5,701	$515	$0	$1,243	$7,459	$19
9.5	Make-up Water System	$278	$0	$398	$0	$0	$676	$65	$0	$148	$890	$2
9.6	Component Cooling Water Sys	$534	$639	$455	$0	$0	$1,628	$152	$0	$356	$2,137	$5
9.9	Circ.Water System Foundations& Structures	$0	$1,714	$2,913	$0	$0	$4,627	$439	$0	$1,520	$6,586	$16
	SUBTOTAL 9.	**$6,743**	**$6,880**	**$5,875**	**$0**	**$0**	**$19,497**	**$1,811**	**$0**	**$4,393**	**$25,701**	**$64**
10	ASH/SPENT SORBENT HANDLING SYS											
10.1	Slag Dewatering & Cooling	$15,485	$0	$7,636	$0	$0	$23,122	$2,222	$0	$2,534	$27,877	$69
10.2	Gasifier Ash Depressurization	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.3	Cleanup Ash Depressurization	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.4	High Temperature Ash Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.5	Other Ash Rrecovery Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.6	Ash Storage Silos	$525	$0	$572	$0	$0	$1,097	$106	$0	$180	$1,384	$3
10.7	Ash Transport & Feed Equipment	$705	$0	$170	$0	$0	$875	$82	$0	$143	$1,099	$3
10.8	Misc. Ash Handling Equipment	$1,088	$1,333	$398	$0	$0	$2,820	$268	$0	$463	$3,551	$9
10.9	Ash/Spent Sorbent Foundation	$0	$46	$58	$0	$0	$105	$10	$0	$34	$149	$0
	SUBTOTAL 10.	**$17,803**	**$1,380**	**$8,835**	**$0**	**$0**	**$28,017**	**$2,688**	**$0**	**$3,356**	**$34,061**	**$85**
11	ACCESSORY ELECTRIC PLANT											
11.1	Generator Equipment	$820	$0	$811	$0	$0	$1,631	$156	$0	$179	$1,966	$5
11.2	Station Service Equipment	$4,462	$0	$402	$0	$0	$4,864	$448	$0	$531	$5,844	$15
11.3	Switchgear & Motor Control	$8,249	$0	$1,500	$0	$0	$9,749	$904	$0	$1,598	$12,251	$31
11.4	Conduit & Cable Tray	$0	$3,832	$12,641	$0	$0	$16,473	$1,593	$0	$4,517	$22,583	$56
11.5	Wire & Cable	$0	$7,321	$4,811	$0	$0	$12,132	$881	$0	$3,253	$16,267	$41
11.6	Protective Equipment	$0	$655	$2,385	$0	$0	$3,041	$297	$0	$501	$3,838	$10
11.7	Standby Equipment	$208	$0	$203	$0	$0	$410	$39	$0	$67	$517	$1
11.8	Main Power Transformers	$15,134	$0	$122	$0	$0	$15,256	$1,154	$0	$2,461	$18,871	$47
11.9	Electrical Foundations	$0	$132	$345	$0	$0	$477	$46	$0	$157	$679	$2
	SUBTOTAL 11.	**$28,872**	**$11,940**	**$23,220**	**$0**	**$0**	**$64,033**	**$5,518**	**$0**	**$13,264**	**$82,816**	**$206**
12	INSTRUMENTATION & CONTROL											
12.1	IGCC Control Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.2	Combustion Turbine Control	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
12.3	Steam Turbine Control	w/8.1	$0	w/8.1	$0	$0	$0	$0	$0	$0	$0	$0
12.4	Other Major Component Control	$1,039	$0	$694	$0	$0	$1,733	$164	$87	$297	$2,281	$6
12.5	Signal Processing Equipment	W/12.7	$0	W/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.6	Control Boards,Panels & Racks	$239	$0	$153	$0	$0	$392	$37	$20	$90	$538	$1
12.7	Computer & Accessories	$5,542	$0	$177	$0	$0	$5,719	$525	$286	$653	$7,183	$18
12.8	Instrument Wiring & Tubing	$0	$1,936	$3,958	$0	$0	$5,894	$500	$295	$1,672	$8,361	$21
12.9	Other I & C Equipment	$3,705	$0	$1,799	$0	$0	$5,503	$518	$275	$944	$7,241	$18
	SUBTOTAL 12.	**$10,524**	**$1,936**	**$6,781**	**$0**	**$0**	**$19,241**	**$1,744**	**$962**	**$3,657**	**$25,604**	**$64**

Exhibit 4-28 Case 3 Total Plant Cost Details (Continued)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		Case 3 TOTAL PLANT COST	
				Direct	Indirect				Process	Project	$	$/kW
13	Improvements to Site											
13.1	Site Preparation	$0	$102	$2,188	$0	$0	$2,290	$227	$0	$755	$3,273	$8
13.2	Site Improvements	$0	$1,821	$2,419	$0	$0	$4,240	$418	$0	$1,397	$6,056	$15
13.3	Site Facilities	$3,262	$0	$3,442	$0	$0	$6,705	$661	$0	$2,210	$9,576	$24
	SUBTOTAL 13.	**$3,262**	**$1,923**	**$8,049**	**$0**	**$0**	**$13,235**	**$1,307**	**$0**	**$4,362**	**$18,904**	**$47**
14	Buildings & Structures											
14.1	Combustion Turbine Area	$0	$265	$150	$0	$0	$414	$36	$0	$90	$541	$1
14.2	Steam Turbine Building	$0	$2,081	$2,964	$0	$0	$5,045	$464	$0	$826	$6,335	$16
14.3	Administration Building	$0	$838	$608	$0	$0	$1,446	$129	$0	$236	$1,811	$5
14.4	Circulation Water Pumphouse	$0	$149	$79	$0	$0	$228	$20	$0	$37	$285	$1
14.5	Water Treatment Buildings	$0	$412	$402	$0	$0	$814	$74	$0	$133	$1,021	$3
14.6	Machine Shop	$0	$432	$295	$0	$0	$727	$65	$0	$119	$910	$2
14.7	Warehouse	$0	$697	$450	$0	$0	$1,147	$102	$0	$187	$1,435	$4
14.8	Other Buildings & Structures	$0	$418	$326	$0	$0	$744	$66	$0	$162	$972	$2
14.9	Waste Treating Building & Str.	$0	$924	$1,766	$0	$0	$2,690	$251	$0	$588	$3,529	$9
	SUBTOTAL 14.	**$0**	**$6,215**	**$7,039**	**$0**	**$0**	**$13,254**	**$1,206**	**$0**	**$2,379**	**$16,839**	**$42**
	TOTAL COST	**$797,628**	**$64,606**	**$272,908**	**$0**	**$0**	**$1,135,142**	**$105,586**	**$59,999**	**$212,287**	**$1,513,013**	**$3,772**
	Owner's Costs											
	Preproduction Costs											
	6 Months All Labor										$12,524	$31
	1 Month Maintenance Materials										$2,710	$7
	1 Month Non-fuel Consumables										$247	$1
	1 Month Waste Disposal										$260	$1
	25% of 1 Months Fuel Cost at 100% CF										$489	$1
	2% of TPC										$30,260	$75
	Total										**$46,491**	**$116**
	Inventory Capital											
	60 day supply of fuel and consumables at 100% CF										$4,410	$11
	0.5% of TPC (spare parts)										$7,565	$19
	Total										**$11,975**	**$30**
	Initial Cost for Catalyst and Chemicals										$3,123	$8
	Land										$900	$2
	Other Owner's Costs										$226,952	$566
	Financing Costs										$40,851	$102
	Total Overnight Costs (TOC)										**$1,843,305**	**$4,595**
	TASC Multiplier								(IOU, high-risk, 35 year)		1.140	
	Total As-Spent Cost (TASC)										**$2,101,368**	**$5,238**

Exhibit 4-29 Case 3 Initial and Annual Operating and Maintenance Costs

INITIAL & ANNUAL O&M EXPENSES					Cost Base (June)	2007
Case 3 - Shell IGCC w/ 90% CO2 capture					Heat Rate-net(Btu/kWh):	11,045
					MWe-net:	401
					Capacity Factor: (%):	80
OPERATING & MAINTENANCE LABOR						
Operating Labor						
Operating Labor Rate(base):	34.65	$/hour				
Operating Labor Burden:	30.00	% of base				
Labor O-H Charge Rate:	25.00	% of labor				
			Total			
Skilled Operator	2.0		2.0			
Operator	10.0		10.0			
Foreman	1.0		1.0			
Lab Tech's, etc.	3.0		3.0			
TOTAL-O.J.'s	16.0		16.0			
					Annual Cost	Annual Unit Cost
					$	$/kW-net
Annual Operating Labor Cost	Maintenance labor cost		% of BEC	1.2092	$6,313,507	$15.739
Maintenance Labor Cost	(Case S1B is reference)		BEC	$1,135,142	$13,725,566	$34.216
Administrative & Support Labor					$5,009,768	$12.489
Property Taxes & Insurance					$30,260,266	$68.345
TOTAL FIXED OPERATING COSTS					**$55,309,108.00**	**$130.789**
VARIABLE OPERATING COSTS						
						$/kWh-net
Maintenance Material Cost			% of BEC	2.2918	**$26,014,838**	**$0.00925**
Consumables	Consumption		Unit	Initial Fill		
	Initial Fill	/Day	Cost	Cost		
Water(/1000 gallons)	0	2,249	1.08	**$0**	**$710,433**	**$0.00025**
Chemicals		5.959				
MU & WT Chem.(lb)	0	13,404	0.17	$0	$677,373	$0.00024
Carbon (Mercury Removal) (lb)	103,110	141	1.05	$108,283	$43,313	$0.00002
COS Catalyst (m3)	0	0.00	2,397.36	$0	$0	$0.00000
Water Gas Shift Catalyst(ft3)	6,043.0	4.1	498.83	$3,014,436	$602,474	$0.00021
Selexol Solution (gal.)	w/equip.	80	13.40	$0	$313,765	$0.00011
MDEA Solution (gal)	0	0	8.70	$0	$0	$0.00000
Sulfinol Solution (gal)	0	0	10.05	$0	$0	$0.00000
SCR Catalyst (m3)	0	0	0.00	$0	$0	$0.00000
Aqueous Ammonia (ton)	0	0	0.00	$0	$0	$0.00000
Claus Catalyst(ft3)	w/equip.	0.68	131.27	$0	$26,163	$0.00001
Subtotal Chemicals				**$3,122,719**	**$1,663,089**	**$0.00059**
Other						
Supplemental Fuel(MBtu)	0	0	0.00	$0	$0	$0.00000
Gases,N2 etc.(/100scf)	0	0	0.00	$0	$0	$0.00000
L.P. Steam(/1000 pounds)	0	0	0.00	$0	$0	$0.00000
Subtotal Other				**$0**	**$0**	**$0.00000**
Waste Disposal						
Spent Mercury Catalyst (lb)	0	141	0.42	$0	$17,202	$0.00001
Flyash (ton)	0	0	0.00	$0	$0	$0.00000
Slag (ton)	0	523	16.23	$0	$2,478,785	$0.00088
Subtotal-Waste Disposal				**$0**	**$2,495,987**	**$0.00089**
By-products & Emissions						
Sulfur(tons)	0	45	0.00	$0	$0	$0.00000
Subtotal By-Products				**$0**	**$0**	**$0.00000**
TOTAL VARIABLE OPERATING COSTS				**$3,122,719**	**$30,884,347.39**	**$0.01099**
Fuel(ton)	0	6,208	10.37	**$0**	**$18,793,494**	**$0.00669**

5. PULVERIZED COAL RANKINE CYCLE PLANTS

Six pulverized coal-fired Rankine cycle power plant configurations were evaluated and the results are presented in Sections 6 and 7 of this report. Cases 4 through 6 are based on greeenfield sites, and assume supercritical steam conditions. Case 7 is based on an existing subcritical PC unit, and Cases 8 and 9 are a retrofit of the existing subcritical PC plant.

The greenfield supercritical PC Cases 4 through 6 are evaluated with and without carbon capture on a common 550 MWe net basis. The designs that include carbon capture have a larger gross unit size to compensate for the higher auxiliary loads. The constant net output sizing basis is selected because it provides for a meaningful side-by-side comparison of the results. The boiler and steam turbine industry ability to match unit size to a custom specification has been commercially demonstrated enabling common net output comparison of the greenfield PC cases in this study. As discussed in Section 3, this was not possible in the IGCC cases because of the fixed output from the combustion turbine.

The subcritical PC retrofit Cases 7 through 9 are evaluated with and without carbon capture. Current performance parameters were taken from two sources, the NETL Coal Plant Database and a recent study performed by CH2MHill [8,55]. The initial Aspen model used the coal composition currently burned at Unit 4. Once performance parameters like coal feed rate, net plant heat rate, net stack output, and stack exit temperature were matched as closely as possible, the coal composition was changed to Montana Rosebud PRB coal and the results represent the baseline performance without CO_2 capture. This established a common 250,000 kg/hr (650,360 lb/hr) basis for coal feed rate for Cases 7 through 9.

Steam conditions for the Rankine cycle Cases 4 through 6 were selected to be consistent with supercritical steam conditions used in previous systems analysis studies [56]. For Cases 7 through 9 the steam cycle conditions were matched according to typical subcritical steam plant operation:

- For supercritical cases (4 - 6) – 24.1 MPa/593°C/593°C (3500 psig/1100°F/1100°F)
- For subcritical cycle cases (7 - 9) – 16.5 MPa/538°C/538°C (2400 psig/1000°F/1000°F)

The evaluation basis details, including site ambient conditions, fuel composition and the emissions control basis, are provided in Section 2 of this report.

5.1 PC COMMON PROCESS AREAS

The PC cases have process areas which are common to each plant configuration such as coal receiving and storage, emissions control technologies and power generation. As detailed descriptions of these process areas in each case section would be burdensome and repetitious, they are presented in this section for general background information. The performance features of these sections are then presented in the case-specific sections.

5.1.1 Coal and Sorbent Receiving and Storage

The function of the coal portion of the Coal and Sorbent Receiving and Storage system for PC plants is identical to the IGCC facilities. It is to provide the equipment required for conveying, preparing, and storing the fuel delivered to the plant. The scope of the system is from the minemouth up to the coal storage silos. The system is designed to support short-term operation at the 5 percent over pressure/valves wide open (OP/VWO) condition (16 hours) and long-term operation of 90 days or more at the maximum continuous rating (MCR).

The scope of the sorbent receiving and storage system includes truck roadways, turnarounds, unloading hoppers, conveyors and the day storage bin.

Operation Description - The coal is delivered to the site in the same manner as the IGCC cases. The 8 cm x 0 (3" x 0) coal from the minemouth is discharged onto a belt conveyor. Two conveyors with an intermediate transfer tower are assumed to convey the coal to the coal stacker, which transfer the coal to either the long-term storage pile or to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron and then to the reclaim pile.

Coal from the reclaim pile is fed by two vibratory feeders, located under the pile, onto a belt conveyor, which transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 2.5 cm x 0 (1" x 0) by the coal crushers. The coal is then transferred by conveyor to the transfer tower. In the transfer tower the coal is routed to the tripper that loads the coal into one of the six boiler silos.

Limestone is delivered to the site using 23 tonne (25 ton) trucks. The trucks empty into a below grade hopper where a feeder transfers the limestone to a conveyor for delivery to the storage pile. Limestone from the storage pile is transferred to a reclaim hopper and conveyed to a day bin.

5.1.2 Steam Generator and Ancillaries

The steam generator for the subcritical PC plants is a tangentially fired, totally enclosed dry bottom furnace, with superheater, reheater, economizer and air-heater.

The steam generator for the supercritical plants is a once-through, spiral-wound, Benson-boiler, wall-fired, balanced draft type unit with a water-cooled dry bottom furnace. It includes superheater, reheater, economizer, and air heater.

It is assumed for the purposes of this study that the greenfield power plants are designed to be operated as a base-loaded unit but with some consideration for daily or weekly cycling, as can be cost effectively included in the base design.

The combustion systems for both subcritical and supercritical steam conditions are equipped with LNBs and OFA. In the subcritical CO_2 capture cases, the existing subcritical PC LNBs are replaced with state-of-the-art LNBs to reduce NOx emissions below the current performance. It is assumed for the purposes of this study that the power plant is designed for operation as a base-load unit.

Scope

The steam generator comprises the following for both subcritical and supercritical PCs (this is standard equipment and assumed applicable to the existing subcritical PC plant.):

- Drum-type evaporator (subcritical only)
- Once-through type steam generator (supercritical only)
- Startup circuit, including integral separators (supercritical only)
- Water-cooled furnace, dry bottom
- Two-stage superheater
- Reheater
- Economizer
- Spray type desuperheater
- Soot blower system
- Air preheaters (Ljungstrom type)
- Coal feeders and pulverizers
- Low NOx Coal burners and light oil ignitors/ warmup system
- Overfire air system
- Forced draft (FD) fans
- Primary air (PA) fans
- Induced draft (ID) fans

The steam generator description for the subcritical case is for a generic unit, but it is assumed that the description would apply to the existing subcritical PC plant. The supercritical PC description is also for a generic greenfield application.

Feedwater and Steam

For the subcritical PC cases, feedwater enters the economizer, recovers heat from the combustion gases exiting the steam generator, and then passes to the boiler drum, from where it is distributed to the water wall circuits enclosing the furnace. After passing through the lower and upper furnace circuits and steam drum in sequence, the steam passes through the convection enclosure circuits to the primary superheater and then to the secondary superheater.

The steam then exits the steam generator en route to the HP turbine. Steam from the HP turbine returns to the steam generator as cold reheat and returns to the IP turbine as hot reheat.

For the supercritical PC cases, feedwater enters the bottom header of the economizer and passes upward through the economizer tube bank, through stringer tubes which support the primary superheater, and discharges to the economizer outlet headers. From the outlet headers, water flows to the furnace hopper inlet headers via external downcomers. Water then flows upward through the furnace hopper and furnace wall tubes. From the furnace, water flows to the steam water separator. During low load operation (operation below the Benson point), the water from the separator is returned to the economizer inlet with the boiler recirculating pump. Operation at loads above the Benson point is once through.

Steam flows from the separator through the furnace roof to the convection pass enclosure walls, primary superheater, through the first stage of water attemperation, to the furnace platens. From the platens, the steam flows through the second stage of attemperation and then to the intermediate superheater. The steam then flows to the final superheater and on to the outlet pipe terminal. Two stages of spray attemperation are used to provide tight temperature control in all high temperature sections during rapid load changes.

Steam returning from the turbine passes through the primary reheater surface, then through crossover piping containing inter-stage attemperation. The crossover piping feeds the steam to the final reheater banks and then out to the turbine. Inter-stage attemperation is used to provide outlet temperature control during load changes.

Air and Combustion Products

Combustion air from the FD fans is heated in Ljungstrom type air preheaters, recovering heat energy from the exhaust gases exiting the boiler. This air is distributed to the burner windbox as secondary air. Air for conveying pulverized coal to the burners is supplied by the PA fans. This air is heated in the Ljungstrom type air preheaters to permit drying of the pulverized coal, and a portion of the air from the PA fans bypasses the air preheaters to be used for regulating the outlet coal/air temperature leaving the mills.

The pulverized coal and air mixture flows to the coal nozzles at various elevations of the furnace. The hot combustion products rise to the top of the boiler and pass through the superheater and reheater sections. The gases then pass through the economizer and air preheater. The gases exit the steam generator at this point and flow to the SCR reactor (SC PC cases only), fabric filter (or ESP in the existing subcritical PC plant cases), ID fan, FGD system, and stack.

Fuel Feed

The crushed Montana Rosebud PRB coal is fed through feeders to each of the mills (pulverizers), where its size is reduced to approximately 72% passing 200 mesh and less than 0.5% remaining on 50 mesh [57]. The pulverized coal exits each mill via the coal piping and is distributed to the coal nozzles in the furnace walls using air supplied by the PA fans.

Ash Removal

The furnace bottom comprises several hoppers, with a clinker grinder under each hopper. The hoppers are of welded steel construction, lined with refractory. The hopper design incorporates a water filled seal trough around the upper periphery for cooling and sealing. Water and ash discharged from the hopper pass through the clinker grinder to an ash sluice system for conveyance to hydrobins, where the ash is dewatered before it is transferred to trucks for offsite disposal. The description of the balance of the bottom ash handling system is presented in Section 5.1.9. The steam generator incorporates fly ash hoppers under the economizer outlet and air heater outlet.

Burners

In the SC PC cases, a boiler of this capacity employs approximately 24 to 36 coal nozzles arranged at multiple elevations. Each burner is designed as a low-NOx configuration with staging of the coal combustion to minimize NOx formation. In addition, overfire air nozzles are provided to further stage combustion and thereby minimize NOx formation.

The existing subcritical PC plant is a tangentially fired unit with older-vintage LNBs.

Oil fired pilot torches are provided for each coal burner for ignition, warm-up and flame stabilization at startup and low loads.

Air Preheaters

Each steam generator is furnished with two vertical-shaft Ljungstrom regenerative type air preheaters. These units are driven by electric motors through gear reducers.

Soot Blowers

The soot-blowing system utilizes an array of 50 to 150 retractable nozzles and lances that clean the furnace walls and convection surfaces with jets of high-pressure steam. The blowers are sequenced to provide an effective cleaning cycle depending on the coal quality and design of the furnace and convection surfaces. Electric motors drive the soot blowers through their cycles.

5.1.3 NO_x Control System

NOx Operation Performance (Greenfield SC PC)

The plant is designed to achieve the environmental target of 0.07 lb NOx/MMBtu. Two measures are taken to reduce the NOx. The first is a combination of low-NOx burners and the introduction of staged overfire air in the boiler. The low-NOx burners and overfire air reduce the emissions to about 0.2 lb/MMBtu.

The second measure taken to reduce the NOx emissions is the installation of an SCR system prior to the air heater. SCR uses ammonia and a catalyst to reduce NOx to N_2 and H_2O. The SCR system consists of three subsystems: reactor vessel, ammonia storage and injection, and gas flow control. The SCR system is designed for 65 percent reduction with 2 ppmv ammonia slip at the end of the catalyst life. This, along with the low-NOx burners, achieves the emission limit of 0.07 lb/MMBtu.

The SCR capital costs are included with the boiler costs, as is the cost for the initial load of catalyst.

SCR Operation Description

The reactor vessel is designed to allow proper retention time for the ammonia to contact the NOx in the boiler exhaust gas. Ammonia is injected into the gas immediately prior to entering the reactor vessel. The catalyst contained in the reactor vessel enhances the reaction between the ammonia and the NOx in the gas. Catalysts consist of various active materials such as titanium dioxide, vanadium pentoxide, and tungsten trioxide. The operating range for vanadium/titanium-based catalysts is 260°C (500°F) to 455°C (850°F). The boiler is equipped with economizer bypass to provide flue gas to the reactors at the desired temperature during periods of low flow rate, such as low load operation. Also included with the reactor vessel is soot-blowing equipment used for cleaning the catalyst.

The ammonia storage and injection system consists of the unloading facilities, bulk storage tank, vaporizers, dilution air skid, and injection grid.

The flue gas flow control consists of ductwork, dampers, and flow straightening devices required to route the boiler exhaust to the SCR reactor and then to the air heater. The economizer bypass and associated dampers for low load temperature control are also included.

NOx Operation Performance (Existing Subcritical PC)

The existing subcritical PC plant uses overfire air with a non-optimal configuration of low NOx burners for NOx control with emissions of 0.45 lb/MMBtu. This is the assumed performance for Case 7. Because the Econamine process requires low concentrations of NO_2 as well as SO_2, Cases 8 and 9 include new, reconfigured low NOx burners in addition to the overfire air to reduce NOx emissions to 0.24 lb/MMBtu.

In the event that NSR standards become applicable, an economic sensitivity case was performed with SCR retrofitted downstream of the LNBs in Cases 8 and 9. The projected NOx emissions with SCR are 0.07 lb/MMBtu.

5.1.4 Particulate Control

Greenfield SC PC

The fabric filter (or baghouse), for supercritical Cases 4 through 6, consists of two separate single-stage, in-line, multi-compartment units. Each unit is of high (0.9-1.5 m/min [3-5 ft/min]) air-to-cloth ratio design with a pulse-jet on-line cleaning system. The ash is collected on the outside of the bags, which are supported by steel cages. The dust cake is removed by a pulse of compressed air. The bag material is polyphenylensulfide (PPS) with intrinsic Teflon (PTFE) coating [58]. The bags are rated for a continuous temperature of 180°C (356°F) and a peak temperature of 210°C (410°F). Each compartment contains a number of gas passages with filter bags, and heated ash hoppers supported by a rigid steel casing. The fabric filter is provided with necessary control devices, inlet gas distribution devices, insulators, inlet and outlet nozzles, expansion joints, and other items as required.

Existing Subcritical PC Plant

The electrostatic precipitator (ESP), for subcritical Cases 7 through 9, consists of a hopper-bottomed, fully enclosed casing containing rows of vertical plates forming passages through which the flue gas flows horizontally. Centrally located in each passage are emitting electrodes energized with high-voltage, negative-polarity direct current. The applied voltage is of sufficient strength to ionize gas molecules close to the electrodes, resulting in a visible corona. When passing the flue gas, the charged ions collide with, and attach themselves to, fly ash particles suspended in the gas. The electric field forces the charged particles out of the gas stream toward the grounded plates, and here they collect and layer. The plates are periodically cleaned by a rapping system to release the layer into ash hoppers as an agglomerated mass. The ESP is located after the air heater and is referred to as a cold-side ESP.

5.1.5 Mercury Removal

Mercury removal is based on a coal Hg content of 0.081 ppmd. The basis for the coal Hg concentration was discussed in Section 2.2. The combination of pollution control technologies used in the PC plants, SCR, fabric filters, ESP, and FGD result in some co-benefit capture of mercury. The SCR promotes the oxidation of elemental mercury, which in turn enhances the mercury removal capability of the fabric filter and FGD unit. The mercury co-benefit capture for SC PC Cases 4 through 6 is assumed to be 15 percent for this combination of control technologies. Activated carbon injection is used to remove an additional 90 percent of the Hg at a carbon injection rate of 1 lb/MMscf. For Cases 7 through 9 mercury co-benefit capture is assumed to be 16 percent with wet FGD and a cold-side ESP.

5.1.6 Flue Gas Desulfurization

Greenfield SC PC

The FGD process uses a lime-based spray dryer system. The function of the FGD system is to scrub the boiler exhaust gases to remove the SO_2 prior to release to the environment, or prior to entering the Carbon Dioxide Removal (CDR) facility. Sulfur removal efficiency is 93 percent in the FGD unit for all cases. The CDR unit includes a polishing scrubber to reduce the flue gas SO_2 concentration from about 55 ppmv at the FGD exit to the required 10 ppmv prior to the CDR absorber. The scope of the FGD system is from the outlet of the combustion air preheater to the ID fan.

A lime-based spray dryer absorber is a dry scrubbing process that is generally used for low-sulfur coal [59]. Flue gas is treated in an absorber by mixing the gas stream concurrently with atomized lime slurry droplets. The lime slurry is atomized through rotary cup spray atomizers or through dual fluid nozzles. Water in the spray droplets evaporates, cooling the gas from the inlet temperature of 300°F or higher to 160°F to 180°F. The final temperature is maintained at approximately 30ºF above the flue gas saturation temperature by regulating the quantity of the slurry water. The droplets absorb SO_2 from the gas and react the SO_2 with the lime in the slurry. The desulfurized flue gas, along with reaction products, unreacted lime, and the fly ash pass out of the dry scrubber to the baghouse. Sorbent utilization is increased by about 40 percent by slurrying and recycling a portion of the solid effluent collected in the baghouse into the absorber with the fresh lime slurry.

The system description is divided into three sections:

- Lime Handling and Reagent Preparation
- SO_2 Removal
- Baghouse

Reagent Handling and Preparation

Lime is received by truck and conveyed to storage. Lime is stored in a 14-day capacity bulk storage lime silo. The lime is pneumatically conveyed to a 16-hour capacity day bin. The lime day bin and a gravimetric feeder supply the lime to a 150 percent slaking system. This will allow two shift operations for the unit operating continuously at 100 percent load. A conventional lime slaker with high-efficiency grit removal and lime recovery system is used. Two 100 percent capacity slurry transfer pumps are used to provide high reliability to transfer the slurry to the slurry tank. The process makeup water is added to the slaker to produce 20 percent solids slurry. The slurry is diluted on line, if required, prior to injection into an absorber. The slurry is fed to the absorber by a dedicated reagent feed pump (100 percent spare capacity provided).

SO_2 Removal

Two absorbers, each treating 50 percent of the flue gas, are provided to achieve 93 percent SO_2 removal efficiency in the absorber and baghouse. The absorber is a vertical, open chamber with concurrent contact between the flue gas and lime slurry. The slurry is injected into the tower at the top using a rotary atomizer. The hopper in the bottom of the carbon steel absorber also removes large particles that may drop in the absorber. The absorber will be operated at 30°F adiabatic approach to saturation temperature. In the past, a lower approach had been proposed. However, over the years, operational problems associated with the lower adiabatic approach to saturation temperature, due to wetting of the walls and large deposits in the absorber, were alleviated by designs with 30°F adiabatic approach to saturation temperature.

Existing Subcritical PC Plant

The current FGD system configuration is a wet sodium carbonate-based forced oxidation positive pressure absorber with a bypass used to reheat the flue gas. The retrofit cases (Cases 8 and 9) will have no bypass and a modification to the stack to handle wet operation. The function of the FGD system is to scrub the boiler exhaust gases to remove the SO_2 prior to release to the environment, or entering into the Carbon Dioxide Removal (CDR) facility. SO_2 removal efficiency is 85 percent in the existing plant (Case 7) and 93 percent for the retrofit cases with a modifidied wet FGD (Cases 8 and 9). For Cases 8 and 9 with CO_2 capture, the SO_2 content of the scrubbed gases must be further reduced to approximately 10 ppmv to minimize formation of amine heat stable salts during the CO_2 absorption process. The CDR unit includes a polishing scrubber to reduce the flue gas SO_2 concentration from about 38 ppmv at the FGD exit to the required 10 ppmv prior to the CDR absorber. The scope of the FGD system is from the outlet of the ID fans to the stack inlet (Case 7) or to the CDR process inlet (Cases 8 and 9).

Sodium sulfate is produced by the injection of oxygen into the sodium carbonate in the absorber tower sump. The bleed from the absorber contains approximately 20 wt% sodium sulfate. The absorber slurry is pumped by an absorber bleed pump to a primary dewatering hydrocyclone cluster. The primary hydrocyclone performs two process functions. The first function is to dewater the slurry from 20 to 50 wt% solids. The second function of the primary hydrocyclone is to perform a $NaCO_3$ and $NaSO_4•2H_2O$ separation. This process ensures a sodium carbonate stoichiometry in the absorber vessel of 1.10 and an overall limestone stoichiometry of 1.05. This system reduces the overall operating cost of the FGD system. The underflow from the hydrocyclone flows into the filter feed tank, from which it is pumped to a horizontal belt vacuum filter. Two 100 percent filter systems are provided for redundant capacity.

5.1.7 Carbon Dioxide Recovery Facility

A Carbon Dioxide Recovery (CDR) facility is used in Cases 5, 6, 8 and 9 to remove the specified amount of the CO_2 in the flue gas exiting the FGD unit, purify it, and compress it to a supercritical condition. In Cases 8 and 9 the flue gas exiting the FGD unit contains about 1 percent more CO_2 than the raw flue gas because of the CO_2 liberated by the sodium carbonate in the FGD absorber vessel. The CDR is comprised of the flue gas supply, a bypass system, SO_2 polishing, CO_2 absorption, solvent stripping and reclaiming, and CO_2 compression and drying.

The CO_2 absorption/stripping/solvent reclaim process for Cases 5, 6, 8 and 9 is based on the Fluor Econamine FG Plus technology [60]. A typical flowsheet is shown in Exhibit 5-1. The Econamine FG Plus process uses a formulation of monoethanolamine (MEA) and a proprietary inhibitor to recover CO_2 from the flue gas. This process is designed to recover high-purity CO_2 from low-pressure streams that contain oxygen, such as flue gas from coal-fired power plants, gas turbine exhaust gas, and other waste gases. The Econamine process used in this study differs from previous studies, including the 2004 IEA study [60], in the following ways:

- The complexity of the control and operation of the plant is significantly decreased
- Solvent consumption is decreased
- Hard to dispose waste from the plant is eliminated

The above are achieved at the expense of a slightly higher steam requirement in the stripper (3,556 kJ/kg [1,530 Btu/lb] versus 3,242 kJ/kg [1,395 Btu/lb] used in the IEA study) [61].

<u>SO_2 Polishing and Flue Gas Cooling and Supply</u>

To prevent the accumulation of heat stable salts, the incoming flue gas must have an SO_2 concentration of 10 ppmv or less. The gas exiting the FGD system passes through an SO_2 polishing step to achieve this objective. The polishing step consists of a non-plugging, low-differential-pressure, spray-baffle-type scrubber using a 20 wt% solution of sodium hydroxide (NaOH). A removal efficiency of about 82 percent (Cases 5 and 6) or 74 percent (Cases 8 and 9) is necessary to reduce SO_2 emissions from the FGD outlet to 10 ppmv as required by the Econamine process. The polishing scrubber proposed for this application has been demonstrated in numerous industrial applications throughout the world and can achieve removal efficiencies of over 95 percent if necessary.

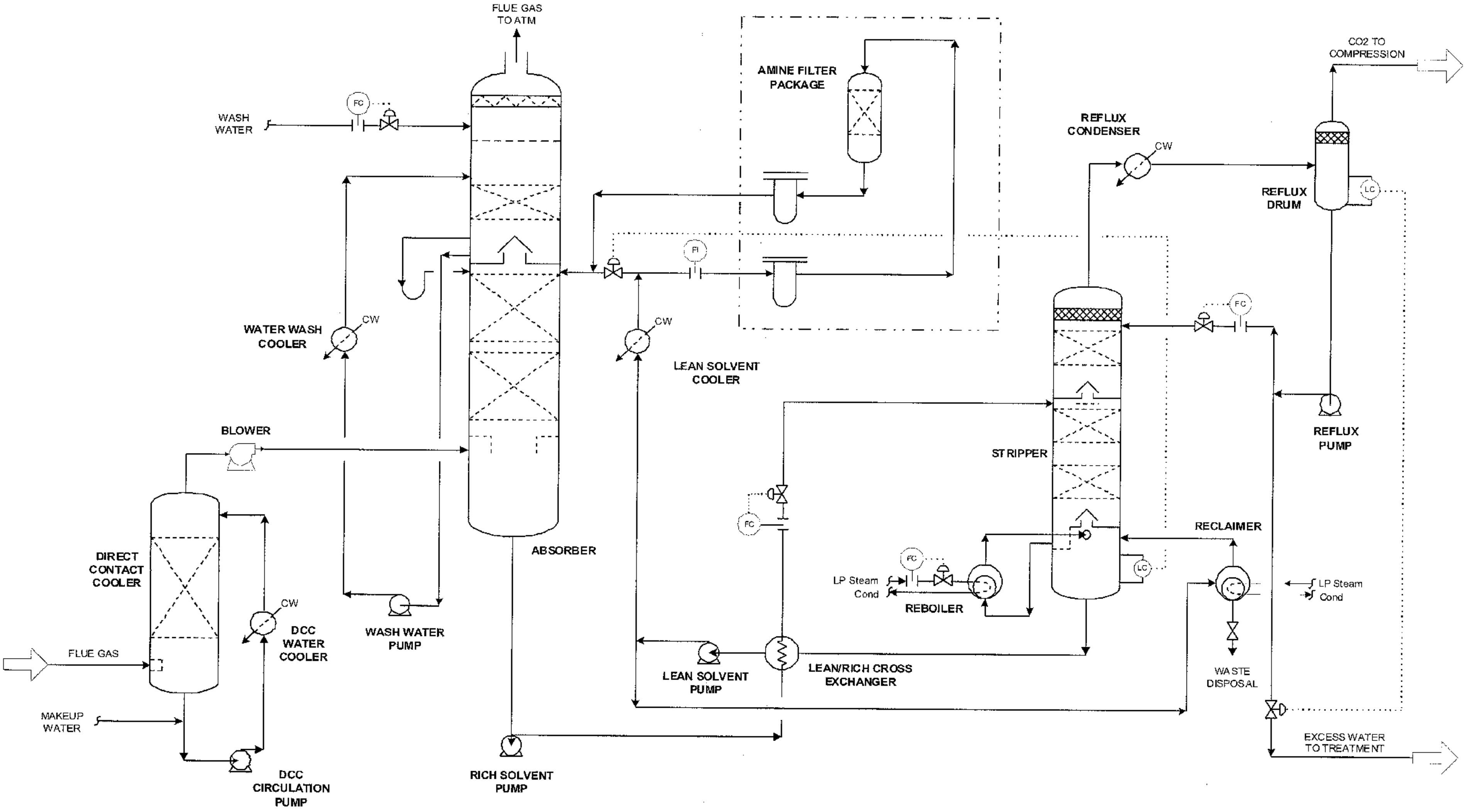
The polishing scrubber also serves as the flue gas cooling system. Cooling water from the PC plant is used to reduce the temperature and hence moisture content of the saturated flue gas exiting the FGD system. Flue gas is cooled beyond the CO_2 absorption process requirements to 32°C (90°F) to account for the subsequent flue gas temperature increase of about 17°C (30°F) in the flue gas blower. Downstream from the Polishing Scrubber flue gas pressure is boosted in the Flue Gas Blowers by approximately 0.014 MPa (2 psi) to overcome pressure drop in the CO_2 absorber tower.

Circulating Water System

Cooling water is provided from the PC plant circulating water system and returned to the PC plant cooling tower. The CDR facility requires a significant amount of cooling water for flue gas cooling, water wash cooling, absorber intercooling, reflux condenser duty, reclaimer cooling, the lean solvent cooler, and CO_2 compression interstage cooling. The cooling water requirements for the plants with a CDR facility in the four PC capture cases range from 946,361-1,705,500 lpm (250,000-450,000 gpm), which exceeds the PC plant cooling water requirement of 340,690-681,380 lpm (90,000-180,000 gpm)

.

Exhibit 5-1 Fluor Econamine FG Plus Typical Flow Diagram



CO_2 Absorption

The cooled flue gas enters the bottom of the CO_2 Absorber and flows up through the tower countercurrent to a stream of lean MEA-based solvent (Econamine FG Plus). Approximately 90 percent of the CO_2 in the feed gas is absorbed into the lean solvent, and the rest leaves the top of the absorber section and flows into the water wash section of the tower. The lean solvent enters the top of the absorber, absorbs the CO_2 from the flue gases and leaves the bottom of the absorber with the absorbed CO_2.

Water Wash Section

The purpose of the Water Wash section is to minimize solvent losses due to mechanical entrainment and evaporation. The flue gas from the top of the CO_2 Absorption section is contacted with a re-circulating stream of water for the removal of most of the lean solvent. The scrubbed gases, along with unrecovered solvent, exit the top of the wash section for discharge to the atmosphere via the vent stack. The water stream from the bottom of the wash section is collected on a chimney tray. A portion of the water collected on the chimney tray spills over to the absorber section as water makeup for the amine with the remainder pumped via the Wash Water Pump and cooled by the Wash Water Cooler, and recirculated to the top of the CO_2 Absorber. The wash water level is maintained by water makeup from the Wash Water Makeup Pump.

Rich/Lean Amine Heat Exchange System

The rich solvent from the bottom of the CO_2 Absorber is preheated by the lean solvent from the Solvent Stripper in the Rich/Lean Solvent Exchanger. The heated rich solvent is routed to the Solvent Stripper for removal of the absorbed CO_2. The stripped solvent from the bottom of the Solvent Stripper is pumped via the Hot Lean Solvent Pumps through the Rich Lean Exchanger to the Solvent Surge Tank. Prior to entering the Solvent Surge Tank, a slipstream of the lean solvent is pumped via the Solvent Filter Feed Pump through the Solvent Filter Package to prevent buildup of contaminants in the solution. From the Solvent Surge Tank the lean solvent is pumped via the Warm Lean Solvent Pumps to the Lean Solvent Cooler for further cooling, after which the cooled lean solvent is returned to the CO_2 Absorber, completing the circulating solvent circuit.

Solvent Stripper

The purpose of the Solvent Stripper is to separate the CO_2 from the rich solvent feed exiting the bottom of the CO_2 Absorber. The rich solvent is collected on a chimney tray below the bottom packed section of the Solvent Stripper and routed to the Solvent Stripper Reboilers where the rich solvent is heated by steam, stripping the CO_2 from the solution. It was assumed that the steam turbine extraction point pressure could be selected to match the reboiler requirements in the greenfield cases, but would be fixed in the retrofit cases. The steam is extracted from the LP turbine at a pressure of 73 psia in the supercritical PC cases and requires only to be de-superheated prior to use in the stripper reboiler. The steam is extracted from the LP turbine at a pressure of 168 psia in the subcritical retrofit cases. The extracted steam in the subcritical PC cases is sent to a Let-Down Turbine to reduce the pressure to 71 psia and generate power from the extracted steam before being de-superheated. The hot wet vapor from the top of the stripper containing CO_2, steam, and solvent vapor, is partially condensed in the Solvent Stripper Condenser by cross exchanging the hot wet vapor with cooling water. The partially condensed stream then flows to the Solvent Stripper Reflux Drum where the vapor and liquid are separated. A portion of the condensate is combined with the vapor stream from the Let-Down Turbine to saturate the superheated vapor before entering the solvent stripper. The uncondensed CO_2-rich gas is then delivered to the CO_2 product compressor. The condensed liquid from the Solvent Stripper Reflux Drum is pumped via the Solvent Stripper Reflux Pumps where a portion of condensed overhead liquid is used as make-up water for the Water Wash section of the CO_2 Absorber. The rest of the pumped liquid is routed back to the Solvent Stripper as reflux, which aids in limiting the amount of solvent vapors entering the stripper overhead system.

Solvent Stripper Reclaimer

A small slipstream of the lean solvent from the Solvent Stripper bottoms is fed to the Solvent Stripper Reclaimer for the removal of high-boiling nonvolatile impurities (heat stable salts - HSS), volatile acids and iron products from the circulating solvent solution. The solvent bound in the HSS is recovered by reaction with caustic and heating with steam. The solvent reclaimer system reduces corrosion, foaming and fouling in the solvent system. The reclaimed solvent is returned to the Solvent Stripper and the spent solvent is pumped via the Solvent Reclaimer Drain Pump to the Solvent Reclaimer Drain Tank.

Steam Condensate

Steam condensate from the Solvent Stripper Reclaimer accumulates in the Solvent Reclaimer Condensate Drum and is level controlled to the Solvent Reboiler Condensate Drum. A portion of the condensate is vaporized to de-superheat the steam entering the stripping section. Steam condensate from the Solvent Stripper Reboilers is also collected in the Solvent Reboiler Condensate Drum and returned to the steam cycle between boiler feedwater heaters 4 and 5 via the Solvent Reboiler Condensate Pumps.

Corrosion Inhibitor System

A proprietary corrosion inhibitor is continuously injected into the CO_2 Absorber rich solvent bottoms outlet line, the Solvent Stripper bottoms outlet line and the Solvent Stripper top tray. This constant injection is to help control the rate of corrosion throughout the CO_2 recovery plant system.

Gas Compression and Drying System

In the compression section, the CO_2 is compressed to 15.3 MPa (2,215 psia) by a six-stage centrifugal compressor. The discharge pressures of the stages were balanced to give reasonable power distribution and discharge temperatures across the various stages as shown in Exhibit 5-2.

Power consumption for this large compressor was estimated assuming an isentropic efficiency of 84 percent. During compression to 15.3 MPa (2,215 psia) in the multiple-stage, intercooled compressor, the CO_2 stream is dehydrated to a dewpoint of -40ºC (-40°F) with triethylene glycol. The virtually moisture-free supercritical CO_2 stream is delivered to the plant battery limit as sequestration ready. CO_2 TS&M costs were estimated and included in LCOE using the methodology described in Section 2.7.

Exhibit 5-2 CO_2 Compressor Interstage Pressures

Stage	Outlet Pressure, MPa (psia)
1	0.35 (51)
2	0.77 (112)
3	1.69 (245)
4	3.71 (538)
5	8.16 (1,184)
6	15.3 (2,215)

Power consumption for this large compressor was estimated assuming an isentropic efficiency of 84 percent. During compression to 15.3 MPa (2,215 psia) in the multiple-stage, intercooled compressor, the CO_2 stream is dehydrated to a dewpoint of -40ºC (-40°F) with triethylene glycol. The virtually moisture-free supercritical CO_2 stream is delivered to the plant battery limit as sequestration ready. CO_2 TS&M costs were estimated and included in LCOE using the methodology described in Section 2.7.

5.1.8 Power Generation

The steam turbine is designed for long-term operation (90 days or more) at MCR with throttle control valves 95 percent open. It is also capable of a short-term 5 percent OP/VWO condition (16 hours).

For the subcritical cases, the steam turbine is assumed to be a tandem compound type, consisting of HP-IP-two LP (double flow) sections enclosed in three casings, designed for condensing single reheat operation, and equipped with non-automatic extractions and four-flow exhaust. The turbine drives a hydrogen cooled generator. The turbine has DC motor-operated lube oil pumps, and main lube oil pumps, which are driven off the turbine shaft [62]. The exhaust pressure is 50.8 cm (20 in) Hg in the single pressure condenser. There are seven extraction points. The condenser is two-shell, transverse, single pressure with divided waterbox for each shell.

The steam-turbine generator systems for the supercritical plants are similar in design to the subcritical systems. The differences include steam cycle conditions and eight extractions points versus seven for the subcritical design.

Turbine bearings are lubricated by a closed-loop, water-cooled pressurized oil system. Turbine shafts are sealed against air in-leakage or steam blowout using a labyrinth gland arrangement connected to a low-pressure steam seal system. The generator stator is cooled with a closed-loop water system consisting of circulating pumps, shell and tube or plate and frame type heat exchangers, filters, and deionizers, all skid-mounted. The generator rotor is cooled with a hydrogen gas recirculation system using fans mounted on the generator rotor shaft.

Operation Description - The turbine stop valves, control valves, reheat stop valves, and intercept valves are controlled by an electro-hydraulic control system. Main steam from the boiler passes through the stop valves and control valves and enters the turbine at 16.5 MPa/ 538°C (2400 psig/1000ºF) for the subcritical cases and 24.1MPa /593°C (3500 psig/1100°F) for the supercritical cases. The steam initially enters the turbine near the middle of the high-pressure span, flows through the turbine, and returns to the boiler for reheating. The reheat steam flows through the reheat stop valves and intercept valves and enters the IP section at 528°C (1000ºF) in the subcritical cases and 593°C (1100°F) in the supercritical cases. After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the two LP sections. The steam divides into four paths and flows through the LP sections exhausting downward into the condenser.

The turbine is designed to operate at constant inlet steam pressure over the entire load range.

5.1.9 Balance of Plant

The balance of plant components consist of the condensate, feedwater, main and reheat steam, extraction steam, ash handling, ducting and stack, waste treatment and miscellaneous systems as described below.

Condensate

The function of the condensate system is to pump condensate from the condenser hotwell to the deaerator, through the gland steam condenser and the LP feedwater heaters. Each system consists of one main condenser; two variable speed electric motor-driven vertical condensate pumps each sized for 50 percent capacity; one gland steam condenser; four LP heaters; and one deaerator with storage tank.

Condensate is delivered to a common discharge header through two separate pump discharge lines, each with a check valve and a gate valve. A common minimum flow recirculation line discharging to the condenser is provided downstream of the gland steam condenser to maintain minimum flow requirements for the gland steam condenser and the condensate pumps.

LP feedwater heaters 1 through 4 are 50 percent capacity, parallel flow, and are located in the condenser neck. All remaining feedwater heaters are 100 percent capacity shell and U-tube heat exchangers. Each LP feedwater heater is provided with inlet/outlet isolation valves and a full capacity bypass. LP feedwater heater drains cascade down to the next lowest extraction pressure heater and finally discharge into the condenser. Pneumatic level control valves control normal drain levels in the heaters. High heater level dump lines discharging to the condenser are provided for each heater for turbine water induction protection. Pneumatic level control valves control dump line flow.

Feedwater

The function of the feedwater system is to pump the feedwater from the deaerator storage tank through the HP feedwater heaters to the economizer. One turbine-driven boiler feedwater pump sized at 100 percent capacity is provided to pump feedwater through the HP feedwater heaters. One 25 percent motor-driven boiler feedwater pump is provided for startup. The pumps are provided with inlet and outlet isolation valves, and individual minimum flow recirculation lines discharging back to the deaerator storage tank. The recirculation flow is controlled by automatic recirculation valves, which are a combination check valve in the main line and in the bypass, bypass control valve, and flow sensing element. The suction of the boiler feed pump is equipped with startup strainers, which are utilized during initial startup and following major outages or system maintenance.

Each HP feedwater heater is provided with inlet/outlet isolation valves and a full capacity bypass. Feedwater heater drains cascade down to the next lowest extraction pressure heater and finally discharge into the deaerator. Pneumatic level control valves control normal drain level in the heaters. High heater level dump lines discharging to the condenser are provided for each heater for turbine water induction protection. Dump line flow is controlled by pneumatic level control valves.

The deaerator is a horizontal, spray tray type with internal direct contact stainless steel vent condenser and storage tank. The boiler feed pump turbine is driven by main steam up to 60 percent plant load. Above 60 percent load, extraction from the IP turbine exhaust (1.16 MPa/367°C [168 psia/693°F] for subcritical PC and 0.50 MPa/292°C [73 psia/557°F] for SC PC) provides steam to the boiler feed pump steam turbine.

Main and Reheat Steam

The function of the main steam system is to convey main steam from the boiler superheater outlet to the HP turbine stop valves. The function of the reheat system is to convey steam from the HP turbine exhaust to the boiler reheater and from the boiler reheater outlet to the IP turbine stop valves.

Main steam exits the boiler superheater through a motor-operated stop/check valve and a motor-operated gate valve and is routed in a single line feeding the HP turbine. A branch line off the IP turbine exhaust feeds the boiler feed water pump turbine during unit operation starting at approximately 60 percent load.

Cold reheat steam exits the HP turbine, flows through a motor-operated isolation gate valve and a flow control valve, and enters the boiler reheater. Hot reheat steam exits the boiler reheater through a motor-operated gate valve and is routed to the IP turbine. A branch connection from the cold reheat piping supplies steam to feedwater heater 7.

Extraction Steam

The function of the extraction steam system is to convey steam from turbine extraction points to end use points as follows:

Greenfield SC PC Cases

- From HP turbine exhaust (cold reheat) to heater 7 and 8
- From IP turbine extraction to heater 6 and the deaerator (heater 5)
- From LP turbine extraction to heaters 1, 2, 3, and 4
- From the crossover pipe to the CDR facility via the Let-Down Turbine (capture cases only)

Existing Subcritical PC Plant Cases

- From HP turbine exhaust (cold reheat) to heater 7
- From IP turbine extraction to heater 6 and the deaerator (heater 5)
- From LP turbine extraction to heaters 1, 2, 3, and 4
- From the crossover pipe to the CDR facility via the Let-Down Turbine (capture cases only)

The turbine is protected from overspeed on turbine trip, from flash steam reverse flow from the heaters through the extraction piping to the turbine. This protection is provided by positive closing, balanced disc non-return valves located in all extraction lines except the lines to the LP feedwater heaters in the condenser neck. The extraction non-return valves are located only in horizontal runs of piping and as close to the turbine as possible.

The turbine trip signal automatically trips the non-return valves through relay dumps. The remote manual control for each heater level control system is used to release the non-return valves to normal check valve service when required to restart the system.

Circulating Water System

In the SC PC cases, it is assumed that the plant is serviced by a public water facility and has access to groundwater for use as makeup cooling water with minimal pretreatment. In the existing subcritical PC plant cases, makeup water comes from a nearby river. All filtration and treatment of the circulating water are conducted on site. A mechanical draft, counter-flow cooling tower is provided for the circulating water heat sink. Two 50 percent circulating water pumps are provided. The circulating water system provides cooling water to the condenser, the auxiliary cooling water system, and the CDR facility in capture cases.

The auxiliary cooling water system is a closed-loop system. Plate and frame heat exchangers with circulating water as the cooling medium are provided. This system provides cooling water to equipment such as the lube oil coolers, turbine generator, and boiler feed pumps. All pumps, vacuum breakers, air release valves, instruments and controls are included for a complete operable system.

The CDR system in Cases 5, 6, 8, and 9 requires a substantial amount of cooling water that is provided by the PC plant circulating water system. The additional cooling load imposed by the CDR is reflected in the significantly larger circulating water pumps and cooling tower in those cases.

Ash Handling System

The function of the ash handling system is to provide the equipment required for conveying, preparing, storing, and disposing of the fly ash and bottom ash produced on a daily basis by the boiler. The scope of the system is from the baghouse hoppers (SC PC cases) or the ESP hoppers (existing subcritical PC plant cases), air heater and economizer hopper collectors, and bottom ash hoppers to the hydrobins (for bottom ash) and truck filling stations (for fly ash). The system is designed to support short-term operation at the 5 percent OP/VWO condition (16 hours) and long-term operation at the 100 percent guarantee point (90 days or more).

The fly ash collected in the baghouse (Cases 4 – 6) or ESP (Cases 7 – 9) and the air heaters is conveyed to the fly ash storage silo. A pneumatic transport system using low-pressure air from a blower provides the transport mechanism for the fly ash. Fly ash is discharged through a wet unloader, which conditions the fly ash and conveys it through a telescopic unloading chute into a truck for disposal.

The bottom ash from the boiler is fed into a clinker grinder. The clinker grinder is provided to break up any clinkers that may form. From the clinker grinders the bottom ash is sluiced to hydrobins for dewatering and offsite removal by truck.

Ash from the economizer hoppers and pyrites (rejected from the coal pulverizers) is conveyed using water to the economizer/pyrites transfer tank. This material is then sluiced on a periodic basis to the hydrobins.

Ducting and Stack

One stack is provided with a single fiberglass-reinforced plastic (FRP) liner. The stack is constructed of reinforced concrete. The stack is 152 m (500 ft) high for adequate particulate dispersion in all PC cases. The existing subcritical PC plant used for a baseline (Case 7) does not have an existing stack liner and a stack liner is added per the retrofit analysis for Cases 8 and 9.

Waste Treatment/Miscellaneous Systems

An onsite water treatment facility treats all runoff, cleaning wastes, blowdown, and backwash to within the U.S. EPA standards for suspended solids, oil and grease, pH, and miscellaneous metals. Waste treatment equipment is housed in a separate building. The waste treatment system consists of a water collection basin, three raw waste pumps, an acid neutralization system, an oxidation system, flocculation, clarification/thickening, and sludge dewatering. The water collection basin is a synthetic-membrane-lined earthen basin, which collects rainfall runoff, maintenance cleaning wastes, and backwash flows.

The raw waste is pumped to the treatment system at a controlled rate by the raw waste pumps. The neutralization system neutralizes the acidic wastewater with hydrated lime in a two-stage system, consisting of a lime storage silo/lime slurry makeup system, dry lime feeder, lime slurry tank, slurry tank mixer, and lime slurry feed pumps.

The oxidation system consists of an air compressor, which injects air through a sparger pipe into the second-stage neutralization tank. The flocculation tank is fiberglass with a variable speed agitator. A polymer dilution and feed system is also provided for flocculation. The clarifier is a plate-type, with the sludge pumped to the dewatering system. The sludge is dewatered in filter presses and disposed offsite. Trucking and disposal costs are included in the cost estimate. The filtrate from the sludge dewatering is returned to the raw waste sump.

Miscellaneous systems consisting of fuel oil, service air, instrument air, and service water are provided. A storage tank provides a supply of No. 2 fuel oil used for startup and for a small auxiliary boiler. Fuel oil is delivered by truck. All truck roadways and unloading stations inside the fence area are provided.

Buildings and Structures

Foundations are provided for the support structures, pumps, tanks, and other plant components. The following buildings are included in the design basis:

- Steam turbine building
- Boiler building
- Administration and service building
- Makeup water and pretreatment building
- Fuel oil pump house
- Coal crusher building
- Continuous emissions monitoring building
- Pump house and electrical equipment building
- Guard house
- Runoff water pump house
- Industrial waste treatment building
- FGD system buildings

5.1.10 Accessory Electric Plant

The accessory electric plant consists of switchgear and control equipment, generator equipment, station service equipment, conduit and cable trays, and wire and cable. It also includes the main power transformer, required foundations, and standby equipment.

5.1.11 Instrumentation and Control

An integrated plant-wide control and monitoring DCS is provided. The DCS is a redundant microprocessor-based, functionally distributed system. The control room houses an array of multiple video monitor and keyboard units. The monitor/keyboard units are the primary interface between the generating process and operations personnel. The DCS incorporates plant monitoring and control functions for all the major plant equipment. The DCS is designed to provide 99.5 percent availability. The plant equipment and the DCS are designed for automatic response to load changes from minimum load to 100 percent. Startup and shutdown routines are implemented as supervised manual, with operator selection of modular automation routines available.

6. GREENFIELD SUPERCRITICAL PC CASES (CASES 4 – 6)

Revision 2 Updates

- *Changed the IP turbine outlet pressure to match the requirements of the Econamine system and eliminated the let-down turbine from the system*
- *Changed the flue gas exit temperature from the combustion air preheater from 166°C (330°F) to 149°C (300°F) to take advantage of the lower sulfur content of the design coal*
- *Updated the steam turbine stage efficiencies and exhaust losses to more closely match existing supercritical steam turbine energy balances*
- *Changed the primary/secondary air split from 23.5 percent primary air to 40 percent primary air*
- *Incorporated air pre-heater leakage into the models*
- *Updated CO_2 compression stage efficiencies based on vendor input*

This section contains an evaluation of plant designs for Cases 4 through 6 which are based on a supercritical PC plant with a nominal net output of 550 MWe. The plants use a single reheat 24.1 MPa/593°C/593°C (3500 psig/1100°F/1100°F) steam cycle. The only difference between the plants is that Case 6 includes 90 percent CO_2 capture and Case 5 is based on an emission rate of 1,100 lb CO_2/net-MWh. Case 4 does not include CO_2 capture.

The balance of Section 6 is organized in an analogous manner to the IGCC section:

- Process and System Description for Cases 4 - 6
- Key Assumptions for Cases 4 - 6
- Sparing Philosophy for Cases 4 - 6
- Comparison of Performance Results for Cases 4 - 6
- Equipment List for Cases 4 -6
- Cost Estimates for Cases 4 – 6

6.1 SC PC NON-CAPTURE CASE 4 AND CAPTURE CASES 5 AND 6

6.1.1 Process Description for Non-Capture Case 4

In this section the supercritical PC process without CO_2 capture is described. The description follows the BFD in Exhibit 6-1 and stream numbers reference the same Exhibit. The tables in Exhibit 6-2 provide process data for the numbered streams in the BFD.

Coal (stream 8) and primary air (stream 5) are introduced into the boiler through the wall-fired burners. Additional combustion air, including the overfire air, is provided by the forced draft fans (stream 2). The boiler operates at a slight negative pressure so air leaks into the boiler, and the infiltration air is accounted for in stream 7. Air leakage also occurs in the combustion air preheater and is accounted for in streams 3 and 6.

Flue gas exits the boiler through the SCR reactor (stream 10) and is cooled to 149°C (300°F) in the combustion air preheater (not shown) before passing to the spray-dryer absorbers. The gases from the absorbers are sent to the baghouse to collect the waste products and the fly ash. Activated carbon is injected for additional mercury removal prior to the baghouse (stream 13). Flue gas exits the baghouse and enters the ID fan suction (stream 15). The clean flue gas passes to the plant stack and is discharged to the atmosphere.

Exhibit 6-1 Case 4: SC PC without CO_2 Capture - Block Flow Diagram

Note: Block Flow Diagram is not intended to represent a complete material balance. Only major process streams and equipment are shown. For example, extraction steam used in the BFW heaters is not shown and accounts for the higher steam flow rate (stream 17) compared to the BFW flow rate (stream 22).

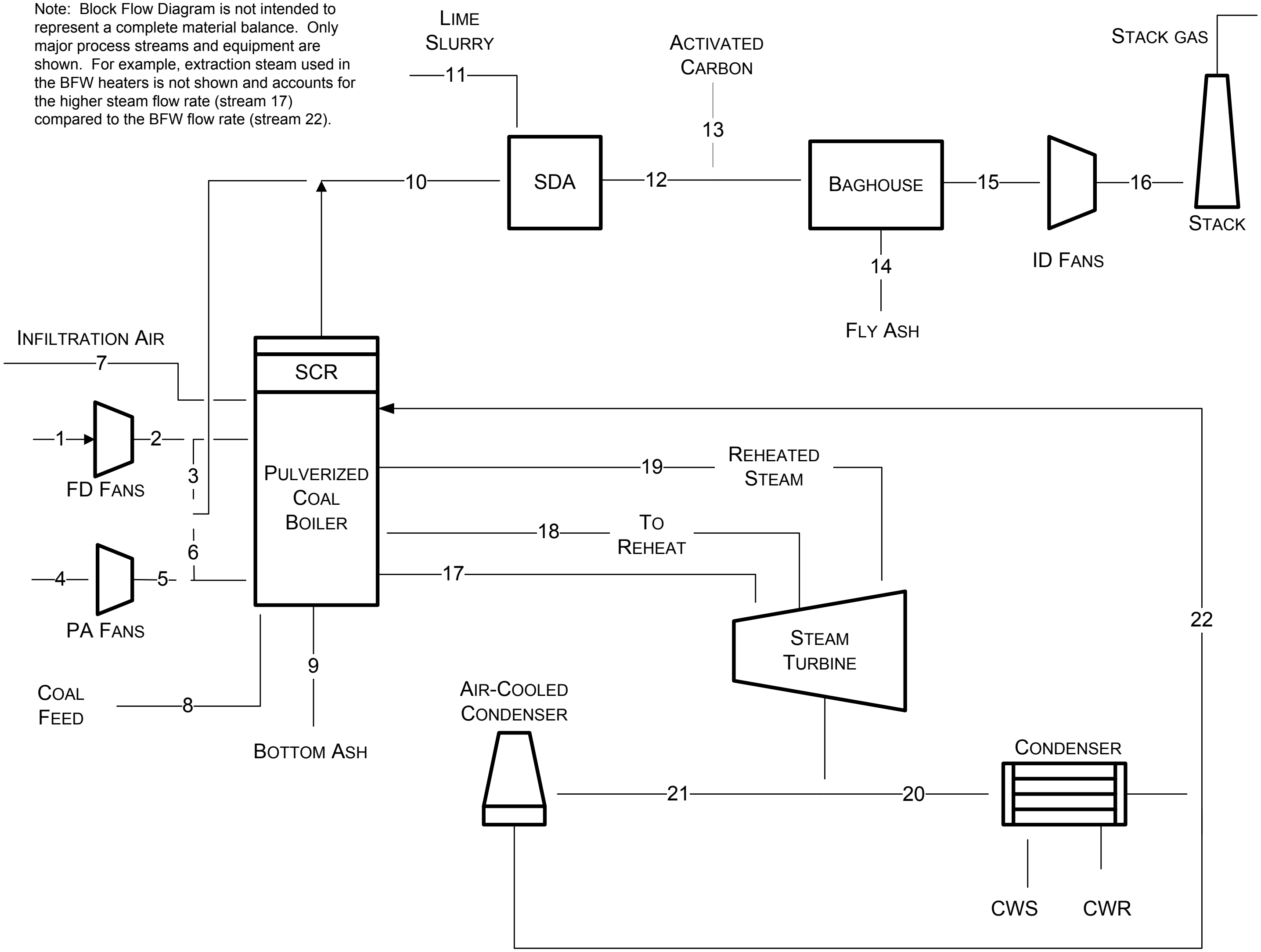


Exhibit 6-2 Case 4: SC PC without CO_2 Capture - Stream Table

	1	2	3	4	5	6	7	8	9	10	11
V-L Mole Fraction											
Ar	0.0093	0.0093	0.0093	0.0093	0.0093	0.0093	0.0093	0.0000	0.0000	0.0084	0.0000
CO_2	0.0003	0.0003	0.0003	0.0003	0.0003	0.0003	0.0003	0.0000	0.0000	0.1470	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0071	0.0071	0.0071	0.0071	0.0071	0.0071	0.0071	0.0000	0.0000	0.1159	1.0000
N_2	0.7753	0.7753	0.7753	0.7753	0.7753	0.7753	0.7753	0.0000	0.0000	0.7041	0.0000
O_2	0.2080	0.2080	0.2080	0.2080	0.2080	0.2080	0.2080	0.0000	0.0000	0.0239	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	39,185	39,185	1,480	26,123	26,123	2,112	1,156	0	0	73,280	3,134
V-L Flowrate (kg/hr)	1,131,934	1,131,934	42,745	754,623	754,623	61,015	33,381	0	0	2,156,656	56,460
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	257,827	4,222	16,888	3,925
Temperature (°C)	6	11	11	6	19	19	6	6	143	143	6
Pressure (MPa, abs)	0.08	0.08	0.08	0.08	0.09	0.09	0.08	0.08	0.08	0.08	0.08
Enthalpy (kJ/kg)^A	16.93	22.06	22.06	16.93	29.97	29.97	16.93	---	---	359.05	313.67
Density (kg/m^3)	1.0	1.0	1.0	1.0	1.1	1.1	1.0	---	---	0.7	1,012.1
V-L Molecular Weight	28.887	28.887	28.887	28.887	28.887	28.887	28.887	---	---	29.430	18.015
V-L Flowrate (lb_{mol}/hr)	86,387	86,387	3,262	57,592	57,592	4,657	2,548	0	0	161,555	6,909
V-L Flowrate (lb/hr)	2,495,488	2,495,488	94,237	1,663,659	1,663,659	134,516	73,593	0	0	4,754,612	124,473
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	568,411	9,308	37,231	8,653
Temperature (°F)	42	51	51	42	65	65	42	42	289	289	42
Pressure (psia)	11.4	12.0	12.0	11.4	12.8	12.8	11.4	11.4	11.1	11.1	11.4
Enthalpy (Btu/lb)^A	7.3	9.5	9.5	7.3	12.9	12.9	7.3	---	---	154.4	134.9
Density (lb/ft^3)	0.061	0.063	0.063	0.061	0.066	0.066	0.061	---	---	0.041	63.182

A - Reference conditions are 32.02 F & 0.089 PSIA

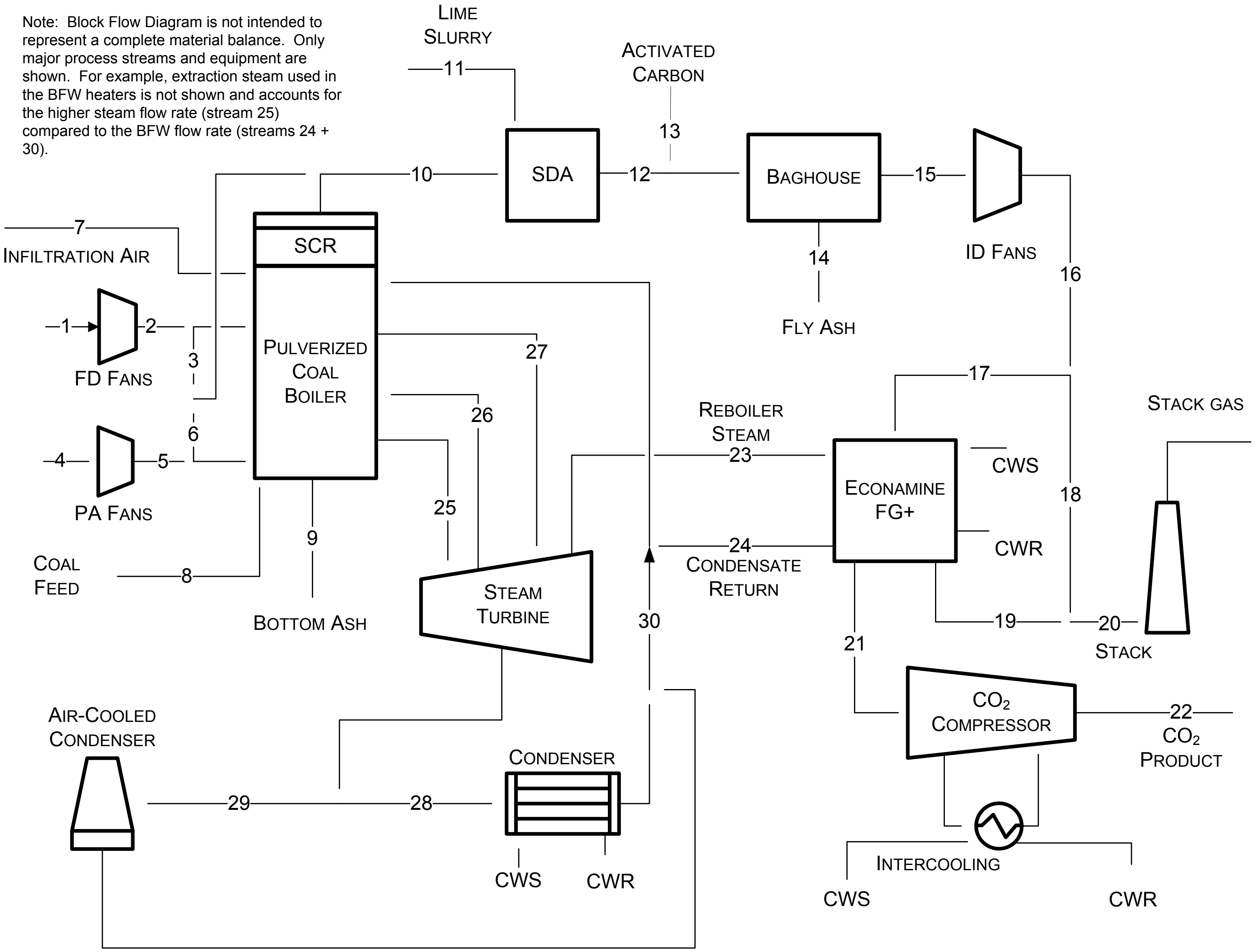
Exhibit 6-2 Case 4: SC PC without CO_2 Capture - Stream Table (Continued)

	12	13	14	15	16	17	18	19	20	21	22
V-L Mole Fraction											
Ar	0.0081	0.0000	0.0000	0.0081	0.0081	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.1411	0.0000	0.0000	0.1411	0.1411	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.1519	0.0000	0.0000	0.1519	0.1519	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
N_2	0.6759	0.0000	0.0000	0.6759	0.6759	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0229	0.0000	0.0000	0.0229	0.0229	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0001	0.0000	0.0000	0.0001	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	0.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	76,332	0	0	76,332	76,332	91,664	75,983	75,983	34,671	34,671	69,342
V-L Flowrate (kg/hr)	2,209,141	0	0	2,209,141	2,209,141	1,651,346	1,368,854	1,368,854	624,605	624,605	1,249,210
Solids Flowrate (kg/hr)	24,788	49	24,837	0	0	0	0	0	0	0	0
Temperature (°C)	82	6	82	82	93	593	354	593	32	32	32
Pressure (MPa, abs)	0.07	0.10	0.07	0.07	0.08	24.23	4.90	4.52	0.00	0.00	1.72
Enthalpy (kJ/kg)^A	333.49	---	---	335.14	347.33	3,476.62	3,082.92	3,652.22	1,932.01	1,932.01	136.94
Density (kg/m^3)	0.7	---	---	0.7	0.8	69.2	18.7	11.6	0.0	0.0	995.7
V-L Molecular Weight	28.941	---	---	28.941	28.941	18.015	18.015	18.015	18.015	18.015	18.015
V-L Flowrate (lb_{mol}/hr)	168,284	0	0	168,284	168,284	202,084	167,514	167,514	76,436	76,436	152,872
V-L Flowrate (lb/hr)	4,870,322	0	0	4,870,322	4,870,322	3,640,595	3,017,806	3,017,806	1,377,018	1,377,018	2,754,037
Solids Flowrate (lb/hr)	54,648	108	54,756	0	0	0	0	0	0	0	0
Temperature (°F)	180	42	180	180	200	1,100	669	1,100	90	90	90
Pressure (psia)	10.7	14.4	10.5	10.5	11.5	3,514.7	710.8	655.8	0.7	0.7	250.0
Enthalpy (Btu/lb)^A	143.4	---	---	144.1	149.3	1,494.7	1,325.4	1,570.2	830.6	830.6	58.9
Density (lb/ft^3)	0.045	---	---	0.044	0.047	4.319	1.164	0.722	0.003	0.003	62.162

6.1.2 Process Description for Capture Cases 5 and 6

Cases 5 and 6 are configured to produce electric power with CO_2 capture. Case 5 has an emission rate of 1,100 lb CO_2/net-MWh. This is achieved by bypassing a portion of the flue gas around the Econamine unit, leaving a portion untreated. Case 6 is designed to include a carbon capture rate of 90 percent. The plant configurations for Cases 5 and 6 are similar to Case 4, with the major difference being the use of an Econamine FG Plus system for CO_2 capture and subsequent compression of the captured CO_2 stream. Since the CO_2 capture and compression process increases the auxiliary load on the plant, the coal feed rate is increased and the overall efficiency is subsequently reduced relative to Case 4. Block flow diagrams for Cases 5 and 6 are shown in Exhibit 6-3 and in Exhibit 6-5, respectively. Stream tables for the BFD’s are presented in Exhibit 6-4 (Case 5) and Exhibit 6-6 (Case 6). The CO_2 removal system was described previously in Section 5.1.7.

Exhibit 6-3 Case 5: SC PC with CO_2 Capture to an Emission Limit of 1,100 lb CO_2/net-MWh - Block Flow Diagram



Note: Block Flow Diagram is not intended to represent a complete material balance. Only major process streams and equipment are shown. For example, extraction steam used in the BFW heaters is not shown and accounts for the higher steam flow rate (stream 25) compared to the BFW flow rate (streams 24 + 30).

Exhibit 6-4 Case 5: SC PC with CO_2 Capture to an Emission Limit of 1,100 lb CO_2/net-MWh - Stream Table

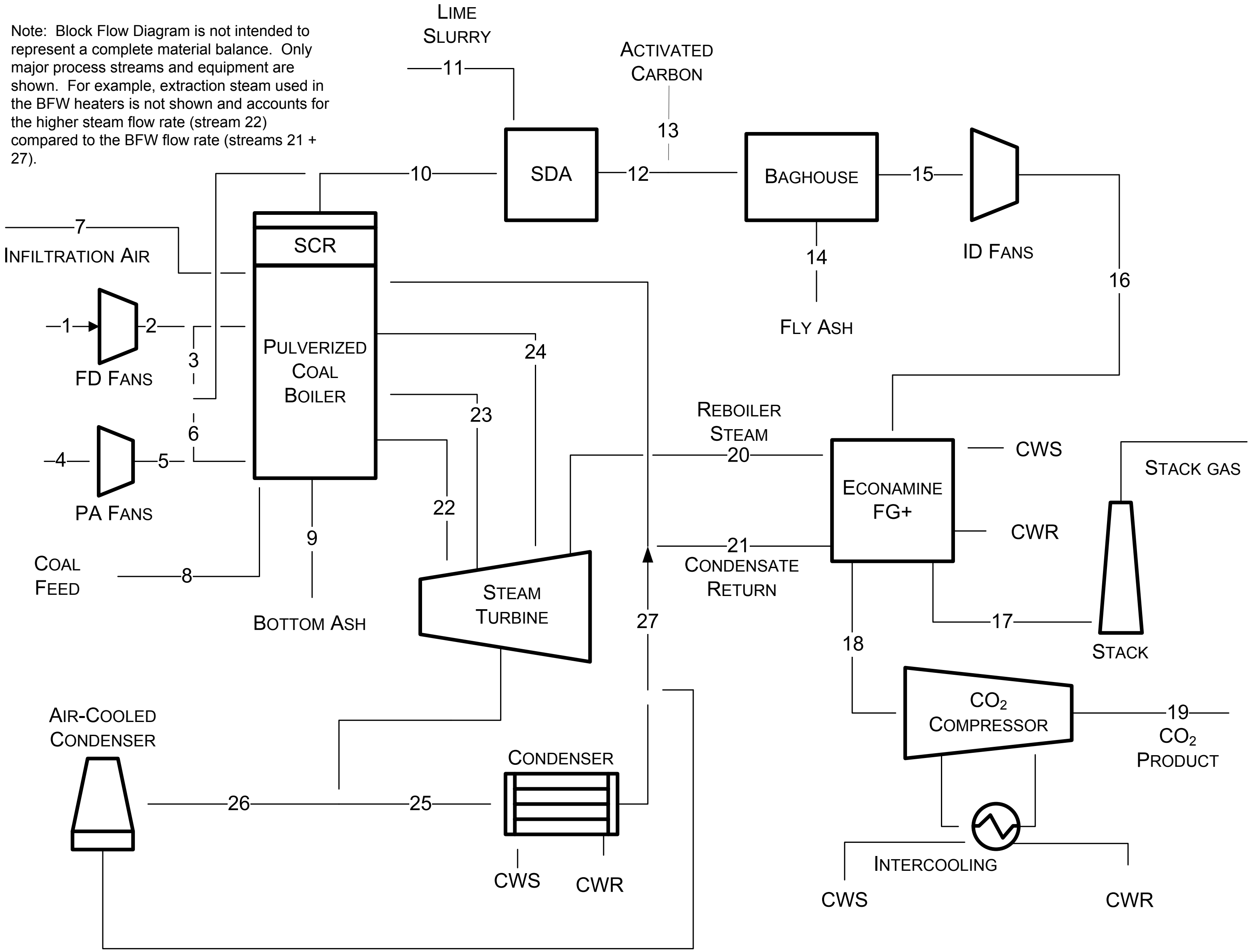
	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
V-L Mole Fraction															
Ar	0.0093	0.0093	0.0093	0.0093	0.0093	0.0093	0.0093	0.0000	0.0000	0.0084	0.0000	0.0081	0.0000	0.0000	0.0081
CO_2	0.0003	0.0003	0.0003	0.0003	0.0003	0.0003	0.0003	0.0000	0.0000	0.1468	0.0000	0.1410	0.0000	0.0000	0.1410
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0071	0.0071	0.0071	0.0071	0.0071	0.0071	0.0071	0.0000	0.0000	0.1158	1.0000	0.1518	0.0000	0.0000	0.1518
N_2	0.7753	0.7753	0.7753	0.7753	0.7753	0.7753	0.7753	0.0000	0.0000	0.7041	0.0000	0.6760	0.0000	0.0000	0.6760
O_2	0.2080	0.2080	0.2080	0.2080	0.2080	0.2080	0.2080	0.0000	0.0000	0.0240	0.0000	0.0231	0.0000	0.0000	0.0231
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0000	0.0001	0.0000	0.0000	0.0001
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	1.0000	1.0000	1.0000	0.0000	0.0000	
V-L Flowrate (kg_{mol}/hr)	47,743	47,743	1,803	31,828	31,828	2,574	1,407	0	0	89,276	3,818	92,995	0	0	92,995
V-L Flowrate (kg/hr)	1,379,153	1,379,153	52,081	919,435	919,435	74,341	40,637	0	0	2,627,392	68,777	2,691,330	0	0	2,691,330
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	313,865	5,140	20,558	4,707	30,104	60	30,164	0
Temperature (°C)	6	11	11	6	19	19	6	6	143	143	6	82	6	82	82
Pressure (MPa, abs)	0.08	0.08	0.08	0.08	0.09	0.09	0.08	0.08	0.08	0.08	0.09	0.07	0.10	0.07	0.07
Enthalpy (kJ/kg)^A	16.93	22.06	22.06	16.93	29.97	29.97	16.93	---	---	358.88	309.48	333.55	---	---	335.00
Density (kg/m^3)	1.0	1.0	1.0	1.0	1.1	1.1	1.0	---	---	0.7	1,012.1	0.7	---	---	0.7
V-L Molecular Weight	28.887	28.887	28.887	28.887	28.887	28.887	28.887	---	---	29.430	18.015	28.941	---	---	28.941
V-L Flowrate (lb_{mol}/hr)	105,255	105,255	3,975	70,170	70,170	5,674	3,101	0	0	196,821	8,417	205,018	0	0	205,018
V-L Flowrate (lb/hr)	3,040,511	3,040,511	114,819	2,027,007	2,027,007	163,895	89,589	0	0	5,792,408	151,627	5,933,367	0	0	5,933,367
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	691,955	11,331	45,323	10,377	66,369	131	66,500	0
Temperature (°F)	42	51	51	42	65	65	42	42	289	289	42	180	42	180	180
Pressure (psia)	11.4	12.0	12.0	11.4	12.8	12.8	11.4	11.4	11.1	11.1	13.0	10.7	14.4	10.5	10.5
Enthalpy (Btu/lb)^A	7.3	9.5	9.5	7.3	12.9	12.9	7.3	---	---	154.3	133.1	143.4	---	---	144.0
Density (lb/ft^3)	0.061	0.063	0.063	0.061	0.066	0.066	0.061	---	---	0.041	63.182	0.045	---	---	0.044

A - Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 6-4 Case 5: SC PC with CO_2 Capture to an Emission Limit of 1,100 lb CO_2/net-MWh - Stream Table (continued)

	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
V-L Mole Fraction															
Ar	0.0081	0.0081	0.0081	0.0107	0.0094	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.1410	0.1410	0.1410	0.0187	0.0783	0.9950	0.9998	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.1518	0.1518	0.1518	0.0431	0.0961	0.0050	0.0002	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
N_2	0.6760	0.6760	0.6760	0.8969	0.7892	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0231	0.0231	0.0231	0.0306	0.0269	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0001	0.0001	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000		1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	92,995	54,169	38,825	40,827	79,652	6,907	6,874	28,276	28,276	109,944	92,757	92,757	29,346	29,346	58,693
V-L Flowrate (kg/hr)	2,691,330	1,567,700	1,123,630	1,148,537	2,272,168	303,093	302,496	509,406	509,406	1,980,671	1,671,041	1,671,041	528,685	528,685	1,057,370
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	93	93	93	32	63	21	35	152	151	593	354	593	32	32	32
Pressure (MPa, abs)	0.08	0.08	0.08	0.09	0.08	0.16	15.27	0.51	0.49	24.23	4.90	4.52	0.00	0.00	1.72
Enthalpy (kJ/kg)^A	347.20	347.20	347.20	104.77	224.65	20.81	-212.29	2,746.79	635.72	3,476.62	3,082.16	3,652.22	1,989.69	1,989.69	136.94
Density (kg/m^3)	0.8	0.8	0.8	1.0	0.8	2.9	794.5	2.7	915.8	69.2	18.7	11.6	0.0	0.0	995.7
V-L Molecular Weight	28.941	28.941	28.941	28.132	28.526	43.881	44.006	18.015	18.015	18.015	18.015	18.015	18.015	18.015	18.015
V-L Flowrate (lb_{mol}/hr)	205,018	119,423	85,595	90,008	175,603	15,228	15,155	62,339	62,339	242,385	204,494	204,494	64,698	64,698	129,396
V-L Flowrate (lb/hr)	5,933,367	3,456,186	2,477,181	2,532,091	5,009,272	668,206	666,889	1,123,048	1,123,048	4,366,633	3,684,014	3,684,014	1,165,551	1,165,551	2,331,103
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	200	200	200	89	145	69	95	306	304	1,100	669	1,100	90	90	90
Pressure (psia)	11.5	11.5	11.5	13.1	11.5	23.2	2,215.0	73.5	71.0	3,514.7	710.8	655.8	0.7	0.7	250.0
Enthalpy (Btu/lb)^A	149.3	149.3	149.3	45.0	96.6	8.9	-91.3	1,180.9	273.3	1,494.7	1,325.1	1,570.2	855.4	855.4	58.9
Density (lb/ft^3)	0.047	0.047	0.047	0.063	0.050	0.181	49.600	0.169	57.172	4.319	1.165	0.722	0.003	0.003	62.162

Exhibit 6-5 Case 6: SC PC with 90% CO_2 Capture - Block Flow Diagram



Note: Block Flow Diagram is not intended to represent a complete material balance. Only major process streams and equipment are shown. For example, extraction steam used in the BFW heaters is not shown and accounts for the higher steam flow rate (stream 22) compared to the BFW flow rate (streams 21 + 27).

Exhibit 6-6 Case 6: SC PC with 90% CO_2 Capture - Stream Table

	1	2	3	4	5	6	7	8	9	10	11	12	13	14
V-L Mole Fraction														
Ar	0.0093	0.0093	0.0093	0.0093	0.0093	0.0093	0.0093	0.0000	0.0000	0.0084	0.0000	0.0081	0.0000	0.0000
CO_2	0.0003	0.0003	0.0003	0.0003	0.0003	0.0003	0.0003	0.0000	0.0000	0.1472	0.0000	0.1413	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0071	0.0071	0.0071	0.0071	0.0071	0.0071	0.0071	0.0000	0.0000	0.1160	1.0000	0.1521	0.0000	0.0000
N_2	0.7753	0.7753	0.7753	0.7753	0.7753	0.7753	0.7753	0.0000	0.0000	0.7040	0.0000	0.6758	0.0000	0.0000
O_2	0.2080	0.2080	0.2080	0.2080	0.2080	0.2080	0.2080	0.0000	0.0000	0.0236	0.0000	0.0227	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0000	0.0001	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	1.0000	1.0000	1.0000	0.0000	0.0000
V-L Flowrate (kg_{mol}/hr)	56,029	56,029	2,116	37,352	37,352	3,020	1,655	0	0	104,799	4,483	109,166	0	0
V-L Flowrate (kg/hr)	1,618,507	1,618,507	61,120	1,079,004	1,079,004	87,243	47,811	0	0	3,084,366	80,768	3,159,440	0	0
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	369,278	6,047	24,188	5,548	35,429	70	35,499
Temperature (°C)	6	11	11	6	19	19	6	6	143	143	6	82	6	82
Pressure (MPa, abs)	0.08	0.08	0.08	0.08	0.09	0.09	0.08	0.08	0.08	0.08	0.09	0.07	0.10	0.07
Enthalpy (kJ/kg)^A	16.93	22.06	22.06	16.93	29.97	29.97	16.93	---	---	359.38	310.45	334.00	---	---
Density (kg/m^3)	1.0	1.0	1.0	1.0	1.1	1.1	1.0	---	---	0.7	1,012.1	0.7	---	---
V-L Molecular Weight	28.887	28.887	28.887	28.887	28.887	28.887	28.887	---	---	29.431	18.015	28.942	---	---
V-L Flowrate (lb_{mol}/hr)	123,522	123,522	4,665	82,348	82,348	6,658	3,649	0	0	231,043	9,884	240,669	0	0
V-L Flowrate (lb/hr)	3,568,197	3,568,197	134,746	2,378,798	2,378,798	192,339	105,406	0	0	6,799,862	178,062	6,965,373	0	0
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	814,119	13,331	53,325	12,231	78,107	154	78,262
Temperature (°F)	42	51	51	42	65	65	42	42	289	289	42	180	42	180
Pressure (psia)	11.4	12.0	12.0	11.4	12.8	12.8	11.4	11.4	11.1	11.1	13.0	10.7	14.4	10.5
Enthalpy (Btu/lb)^A	7.3	9.5	9.5	7.3	12.9	12.9	7.3	---	---	154.5	133.5	143.6	---	---
Density (lb/ft^3)	0.061	0.063	0.063	0.061	0.066	0.066	0.061	---	---	0.041	63.182	0.045	---	---
	A - Referer	A - Reference conditions are 32.02 F & 0.089 PSIA												

Exhibit 6-6 Case 6: SC PC with 90% CO_2 Capture - Stream Table (continued)

	15	16	17	18	19	20	21	22	23	24	25	26	27
V-L Mole Fraction													
Ar	0.0081	0.0081	0.0107	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.1413	0.1413	0.0188	0.9951	0.9998	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.1521	0.1521	0.0431	0.0049	0.0002	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
N_2	0.6758	0.6758	0.8974	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0227	0.0227	0.0301	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0001	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	109,166	109,166	82,216	13,951	13,884	57,113	57,113	129,323	109,243	109,243	24,397	24,397	48,793
V-L Flowrate (kg/hr)	3,159,440	3,159,440	2,312,797	612,190	610,986	1,028,906	1,028,906	2,329,782	1,968,043	1,968,043	439,510	439,510	879,020
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	82	93	32	21	35	152	151	593	354	593	32	32	32
Pressure (MPa, abs)	0.07	0.08	0.09	0.16	15.27	0.51	0.49	24.23	4.90	4.52	0.00	0.00	1.72
Enthalpy (kJ/kg)^A	335.41	347.60	104.77	20.80	-212.29	2,861.67	635.72	3,476.62	3,081.64	3,652.22	2,002.88	2,002.88	136.94
Density (kg/m^3)	0.7	0.8	1.0	2.9	794.5	2.7	915.8	69.2	18.7	11.6	0.0	0.0	995.7
V-L Molecular Weight	28.942	28.942	28.131	43.881	44.006	18.015	18.015	18.015	18.015	18.015	18.015	18.015	18.015
V-L Flowrate (lb_{mol}/hr)	240,669	240,669	181,254	30,757	30,609	125,913	125,913	285,107	240,840	240,840	53,785	53,785	107,570
V-L Flowrate (lb/hr)	6,965,373	6,965,373	5,098,844	1,349,648	1,346,995	2,268,350	2,268,350	5,136,290	4,338,793	4,338,793	968,953	968,953	1,937,907
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	180	200	89	69	95	306	304	1,100	668	1,100	90	90	90
Pressure (psia)	10.5	11.5	13.1	23.2	2,215.0	73.5	71.0	3,514.7	710.8	655.8	0.7	0.7	250.0
Enthalpy (Btu/lb)^A	144.2	149.4	45.0	8.9	-91.3	1,230.3	273.3	1,494.7	1,324.9	1,570.2	861.1	861.1	58.9
Density (lb/ft^3)	0.044	0.047	0.063	0.181	49.600	0.169	57.172	4.319	1.166	0.722	0.003	0.003	62.162

6.1.3 Key System Assumptions

System assumptions for Cases 4 through 6, supercritical PC with and without CO_2 capture, are compiled in Exhibit 6-7.

Exhibit 6-7 Supercritical PC Plant Study Configuration Matrix

	Case 4 w/o CO_2 Capture	**Case 5 w/CO_2 Capture**	**Case 6 w/CO_2 Capture**
Steam Cycle, MPa/°C/°C (psig/°F/°F)	24.1/593/593 (3500/1100/1100)	24.1/593/593 (3500/1100/1100)	24.1/593/593 (3500/1100/1100)
Coal	Rosebud PRB	Rosebud PRB	Rosebud PRB
Condenser pressure, mm Hg (in Hg)	35.6 (1.4)	35.6 (1.4)	35.6 (1.4)
Boiler Efficiency, %	86	86	86
Cooling water to condenser, °C (ºF)	8.9 (48)	8.9 (48)	8.9 (48)
Cooling water from condenser, °C (ºF)	20 (68)	20 (68)	20 (68)
Stack temperature, °C (°F)	93 (200)	63 (145)	32 (89)
SO_2 Control	Dry Limestone FGD	Dry Limestone FGD (B)	Dry Limestone FGD (B)
FGD Efficiency, % (A)	93	93	93
NOx Control	LNB w/OFA and SCR	LNB w/OFA and SCR	LNB w/OFA and SCR
SCR Efficiency, % (A)	65	65	65
Ammonia Slip (end of catalyst life), ppmv	2	2	2
Particulate Control	Fabric Filter	Fabric Filter	Fabric Filter
Fabric Filter efficiency, % (A)	99.97	99.97	99.97
Ash Distribution, Fly/Bottom	80% / 20%	80% / 20%	80% / 20%
Mercury Control	Co-benefit Capture and Activated Carbon Injection	Co-benefit Capture and Activated Carbon Injection	Co-benefit Capture and Activated Carbon Injection
Mercury removal efficiency, % (A)	15% co-benefit capture and additional 90% with activated carbon injection	15% co-benefit capture and additional 90% with activated carbon injection	15% co-benefit capture and additional 90% with activated carbon injection
CO_2 Control	N/A	Econamine FG Plus	Econamine FG Plus
CO_2 Capture (A)	N/A	1,100 lb/net-MWh	90% (A)
CO_2 Sequestration	N/A	Off-site Saline Formation	Off-site Saline Formation

A. Removal efficiencies are based on the flue gas content

B. An SO_2 polishing step is included to meet more stringent SOx content limits in the flue gas (< 10 ppmv) to reduce formation of amine heat stable salts during the CO_2 absorption process

Balance of Plant – Cases 4 - 6

The balance of plant assumptions are common to all cases and are presented in Exhibit 6-8.

Exhibit 6-8 Balance of Plant Assumptions

Cooling system	Recirculating Wet Cooling Tower
Fuel and Other storage	
Coal	30 days
Ash	30 days
Lime	30 days
Plant Distribution Voltage	
Motors below 1 hp	110/220 volt
Motors between 1 hp and 250 hp	480 volt
Motors between 250 hp and 5,000 hp	4,160 volt
Motors above 5,000 hp	13,800 volt
Steam and Gas Turbine generators	24,000 volt
Grid Interconnection voltage	345 kV
Water and Waste Water	
Makeup Water	The water supply is 50 percent from a local Publicly Owned Treatment Works (POTW) and 50 percent from groundwater, and is assumed to be in sufficient quantities to meet plant makeup requirements. Makeup for potable, process, and de-ionized (DI) water is drawn from municipal sources.
Process Wastewater	Storm water that contacts equipment surfaces is collected and treated for discharge through a permitted discharge.
Sanitary Waste Disposal	Design includes a packaged domestic sewage treatment plant with effluent discharged to the industrial wastewater treatment system. Sludge is hauled off site. Packaged plant is sized for 5.68 cubic meters per day (1,500 gallons per day)
Water Discharge	Most of the process wastewater is recycled to the cooling tower basin. Blowdown will be treated for chloride and metals, and discharged.

6.1.4 Sparing Philosophy

Single trains are used throughout the design with exceptions where equipment capacity requires an additional train. There is no redundancy other than normal sparing of rotating equipment. The plant design consists of the following major subsystems:

- One dry-bottom, wall-fired PC supercritical boiler (1 x 100%)
- Two single-stage, in-line, multi-compartment fabric filters (2 x 50%)
- One lime-based spray dryer absorber (1 x 100%)
- One steam turbine (1 x 100%)
- For Case 6 only, two parallel Econamine FG Plus CO_2 absorption systems, with each system consisting of two absorbers, strippers and ancillary equipment (2 x 50%). Case 5 consists of a single train only.

6.1.5 Case 4 - 6 Performance Results

The plants produce a net output of 550 MWe at a net plant efficiency of 38.6, 31.7, and 26.9 percent (HHV basis) for Cases 4 through 6, respectively.

Overall performance for the plant is summarized in Exhibit 6-9, which includes auxiliary power requirements. The CDR facility, including CO_2 compression, accounts for over 45 and 57 percent of the auxiliary plant load for Cases 5 and 6, respectively. The CDR facility loads include a flue gas booster fan to overcome the absorber pressure drop and pumps to circulate the amine solvent. The circulating water system (circulating water pumps and cooling tower fan) accounts for over 11 percent the auxiliary load, largely due to the high cooling water demand of the CDR facility.

Exhibit 6-9 Cases 4 - 6 Plant Performance Summary

Power Output, kWe	Case 4	Case 5	Case 6
Steam Turbine Power	**585,300**	**629,800**	**675,500**
Gross Power	**585,300**	**629,800**	**675,500**
Auxiliary Load, kWe			
Coal Handling and Conveying	510	570	630
Pulverizers	3,870	4,710	5,540
Sorbent Handling & Reagent Preparation	170	210	250
Ash Handling	860	1,040	1,230
Primary Air Fans	2,830	3,450	4,050
Forced Draft Fans	1,670	2,040	2,390
Induced Draft Fans	7,750	9,450	11,030
SCR	10	20	20
Baghouse	120	150	170
Spray Dryer FGD	2,240	2,730	3,210
Econamine FG Plus Auxiliaries	N/A	11,400	22,900
CO_2 Compression	N/A	24,340	49,170
Miscellaneous Balance of Plant[1,2]	2,000	2,000	2,000
Steam Turbine Auxiliaries	400	400	400
Condensate Pumps	790	670	560
Circulating Water Pumps	2,410	5,160	9,190
Ground Water Pumps	250	460	800
Cooling Tower Fans	1,800	3,840	6,000
Air Cooled Condenser Fans	5,760	5,030	3,690
Transformer Losses	1,850	2,100	2,370
Total	35,290	79,770	125,600
Plant Performance			
Net Plant Power	550,010	550,030	549,900
Net Plant Efficiency (HHV)	38.6%	31.7%	26.9%
Net Plant Heat Rate (HHV)	9,338 (8,851)	11,367 (10,774)	13,377 (12,679)
Coal Feed Flowrate (kg/hr (lb/hr))	257,827 (568,411)	313,865 (691,955)	369,278 (814,119)
Thermal Input (kW$_{th}$)	1,426,632	1,736,710	2,043,325
Condenser Duty (GJ/hr (MMBtu/hr))	2,245 (2,128)	1,961 (1,859)	1,642 (1,556)
Raw Water Withdrawal (m^3/min (gpm))	10.3 (2,733)	19.4 (5,117)	33.4 (8,820)
Raw Water Consumption (m^3/min (gpm))	8.2 (2,175)	14.9 (3,924)	25.3 (6,693)
Other Consumables			
Activated Carbon (kg/day (lb/day))	1,174 (2,588)	1,430 (3,153)	1,679 (3,701)
SCR Catalyst (m^3 (ft^3))	379 (13,390)	462 (16,313)	542 (19,150)
FGD Sorbent (tonne/day (ton/day))	3.93 (4.33)	4.71 (5.19)	5.55 (6.12)
Ammonia (19% Solution) (tonne/day (ton/day))	20.5 (22.6)	25.0 (27.6)	29.4 (32.4)
Econamine Consumables			
MEA (tonne/day (ton/day))	N/A	0.73 (0.80)	1.47 (1.62)
Activated Carbon (kg/day (lb/day))	N/A	435 (960)	880 (1,939)
Sodium Hydroxide (NaOH) (tonne/day (ton/day))	N/A	5.21 (5.74)	10.53 (11.61)
Sulfuric Acid (H_2SO_4) (tonne/day (ton/day))	N/A	3.47 (3.82)	7.00 (7.72)
Corrosion Inhibitor ($/yr)	N/A	3,733	7,541

1 - Boiler feed pumps are turbine driven

2 - Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

Environmental Performance

The environmental targets for emissions of Hg, NOx, SO_2 and particulate matter were presented in Section 2.4. A summary of the plant air emissions for Cases 4 through 6 is presented in Exhibit 6-10.

Exhibit 6-10 Cases 4 - 6 Air Emissions

	Case 4	**Case 5**	**Case 6**
kg/GJ (lb/10^6 Btu)			
SO_2	0.051 (0.119)	0.022 (0.051)	0.001 (0.002)
NO_X	0.030 (0.070)	0.030 (0.070)	0.030 (0.070)
Particulates	0.006 (0.013)	0.006 (0.013)	0.006 (0.013)
Hg	0.257E-6 (0.597E-6)	0.257E-6 (0.597E-6)	0.257E-6 (0.597E-6)
CO_2	92 (215)	44 (102)	9.2 (21)
Tonne/year (tons/year) 85% capacity			
SO_2	1,953 (2,153)	1,013 (1,116)	37 (40)
NO_X	1,151 (1,269)	1,401 (1,544)	1,648 (1,817)
Particulates	214 (236)	260 (287)	306 (337)
Hg	0.010 (0.011)	0.012 (0.013)	0.014 (0.015)
CO_2	3,529,083 (3,890,148)	2,043,885 (2,252,998)	505,458 (557,172)
kg/MWh (lb/gross-MWh)			
SO_2	0.448 (0.988)	0.216 (0.476)	0.007 (0.016)
NO_X	0.264 (0.582)	0.299 (0.659)	0.328 (0.722)
Particulates	0.049 (0.108)	0.055 (0.122)	0.061 (0.134)
Hg	2.25E-6 (4.96E-6)	2.55E-6 (5.62E-6)	2.79E-6 (6.16E-6)
CO_2	810 (1,785)	436 (961)	100 (222)
kg/MWh (lb/net-MWh)			
CO_2	862 (1,900)	499 (1,100)	123 (272)

SO_2 emissions are controlled using a lime-based spray dryer absorber that achieves a removal efficiency of 93 percent. The saturated flue gas exiting the scrubber is vented through the plant stack (Case 4) or sent to the Econamine unit (Cases 5 and 6).

NOx emissions are controlled to about 0.20 lb/10^6 Btu through the use of LNBs and OFA. An SCR unit then further reduces the NOx concentration by 65 percent to 0.07 lb/10^6 Btu.

Particulate emissions are controlled using a pulse jet fabric filter which operates at an efficiency of 99.97 percent.

Co-benefit capture results in a 15 percent reduction of mercury emissions. Activated carbon injection provides an additional 90 percent reduction of mercury emissions. CO_2 emissions represent the discharge from the respective processes.

Exhibit 6-11 shows the overall water balance for the plant. Raw water is obtained from groundwater (50 percent) and from municipal sources (50 percent). Water demand represents the total amount of water required for a particular process. Some water is recovered within the process, primarily as flue gas condensate in CO_2 capture cases, and that water is re-used as internal recycle. Raw water withdrawal is the difference between water demand and internal recycle. Some water is returned to the source, namely cooling tower blowdown. The difference between raw water withdrawal and water returned to the source (process discharge) is raw water consumption, which represents the net impact on the water source.

Exhibit 6-11 Cases 4 - 6 Water Balance

	Case 4	**Case 5**	**Case 6**
Water Demand, m^3/min (gpm)			
Econamine	N/A	0.07 (20)	0.15 (40)
FGD Makeup	0.94 (249)	1.1 (303)	1.3 (356)
Cooling Tower	9.4 (2,484)	20.1 (5,305)	35.8 (9,456)
Total	**10.3 (2,733)**	**21.3 (5,628)**	**37.3 (9,852)**
Internal Recycle, m^3/min (gpm)			
Econamine	N/A	0.0 (0)	0.0 (0)
FGD Makeup	0.00 (0)	0.0 (0)	0.0 (0)
Cooling Tower	0.00 (0)	1.9 (511)	3.9 (1,032)
Total	**0.00 (0)**	**1.9 (511)**	**3.9 (1,032)**
Raw Water Withdrawal, m^3/min (gpm)			
Econamine	N/A	0.07 (20)	0.15 (40)
FGD Makeup	0.94 (249)	1.1 (303)	1.3 (356)
Cooling Tower	9.4 (2,484)	18.1 (4,794)	31.9 (8,424)
Total	**10.3 (2,733)**	**19.4 (5,117)**	**33.4 (8,820)**
Process Water Discharge, m^3/min (gpm)			
Cooling Tower	2.1 (559)	4.5 (1,193)	8.0 (2,127)
Total	**2.1 (559)**	**4.5 (1,193)**	**8.0 (2,127)**
Raw Water Consumption, m^3/min (gpm)			
Econamine	N/A	0.07 (20)	0.15 (40)
FGD Makeup	0.94 (249)	1.1 (303)	1.3 (356)
Cooling Tower	7.3 (1,926)	13.6 (3,601)	23.8 (6,297)
Total	**8.2 (2,175)**	**14.9 (3,924)**	**25.3 (6,693)**
Total, gpm/MWnet	**4.0**	**7.1**	**12.2**

The carbon balance for the plant is shown in Exhibit 6-12. The carbon input to the plant consists of carbon in the coal and carbon in the air. Carbon leaves the plant as carbon in the CO_2 in the stack gas and CO_2 product. The percent of total carbon sequestered for the capture cases is defined as the amount of carbon product produced (as sequestration-ready CO_2) divided by the carbon in the coal feedstock, less carbon contained in solid byproducts (ash), expressed as a percentage.

Exhibit 6-12 Cases 4 – 6 Carbon Balance

	Case 4	**Case 5**	**Case 6**
Carbon In, kg/hr (lb/hr)			
Coal	129,089 (284,593)	157,147 (346,449)	184,891 (407,614)
Air (CO_2)	262 (577)	319 (703)	374 (825)
Activated Carbon	49 (108)	60 (131)	70 (154)
Total In	**129,400 (285,277)**	**157,525 (347,283)**	**185,335 (408,593)**
Carbon Out, kg/hr (lb/hr)			
Ash	49 (108)	60 (131)	70 (154)
Stack Gas	129,351 (285,169)	74,914 (165,157)	18,526 (40,844)
CO_2 Product	N/A	82,551 (181,994)[1]	166,738 (367,594)[2]
Total Out	**129,400 (285,277)**	**157,525 (347,283)**	**185,335 (408,593)**

[1] Carbon capture is 52.5 percent to achieve an emission rate of 1,100 lb CO_2/net-MWh
[2] Carbon capture is 90 percent

The sulfur balance for the plant is shown in Exhibit 6-13. Sulfur input is the sulfur in the coal. Sulfur output is the sulfur combined with lime in the ash and the sulfur emitted in the stack gas.

Exhibit 6-13 Cases 4 - 6 Sulfur Balance

	Case 4	**Case 5**	**Case 6**
Sulfur In, kg/h (lb/hour)			
Coal	1,876 (4,135)	2,283 (5,034)	2,686 (5,922)
Total In	**1,876 (4,135)**	**2,283 (5,034)**	**2,686 (5,922)**
Sulfur Out, kg/h (lb/hour)			
Ash	1,744 (3,845)[1]	2,123 (4,681)[1]	2,498 (5,508)[1]
Stack Gas	131 (289)	68 (150)	2 (5)
Polishing Scrubber	N/A	92 (202)	186 (409)
Total Out	**1,876 (4,135)**	**2,283 (5,034)**	**2,686 (5,922)**

[1] Sulfur capture is 93 percent

Heat and Mass Balance Diagrams

Heat and mass balance diagrams are shown for all three supercritical PC cases, including the boiler, gas cleanup, and the power block system in Exhibit 6-14 through Exhibit 6-19.

An overall plant energy balance is provided in tabular form in Exhibit 6-20. The power out is the steam turbine power prior to generator losses.

Exhibit 6-14 Case 4 SC PC without CO_2 Boiler and Gas Cleanup Systems Heat and Mass Balance Schematic

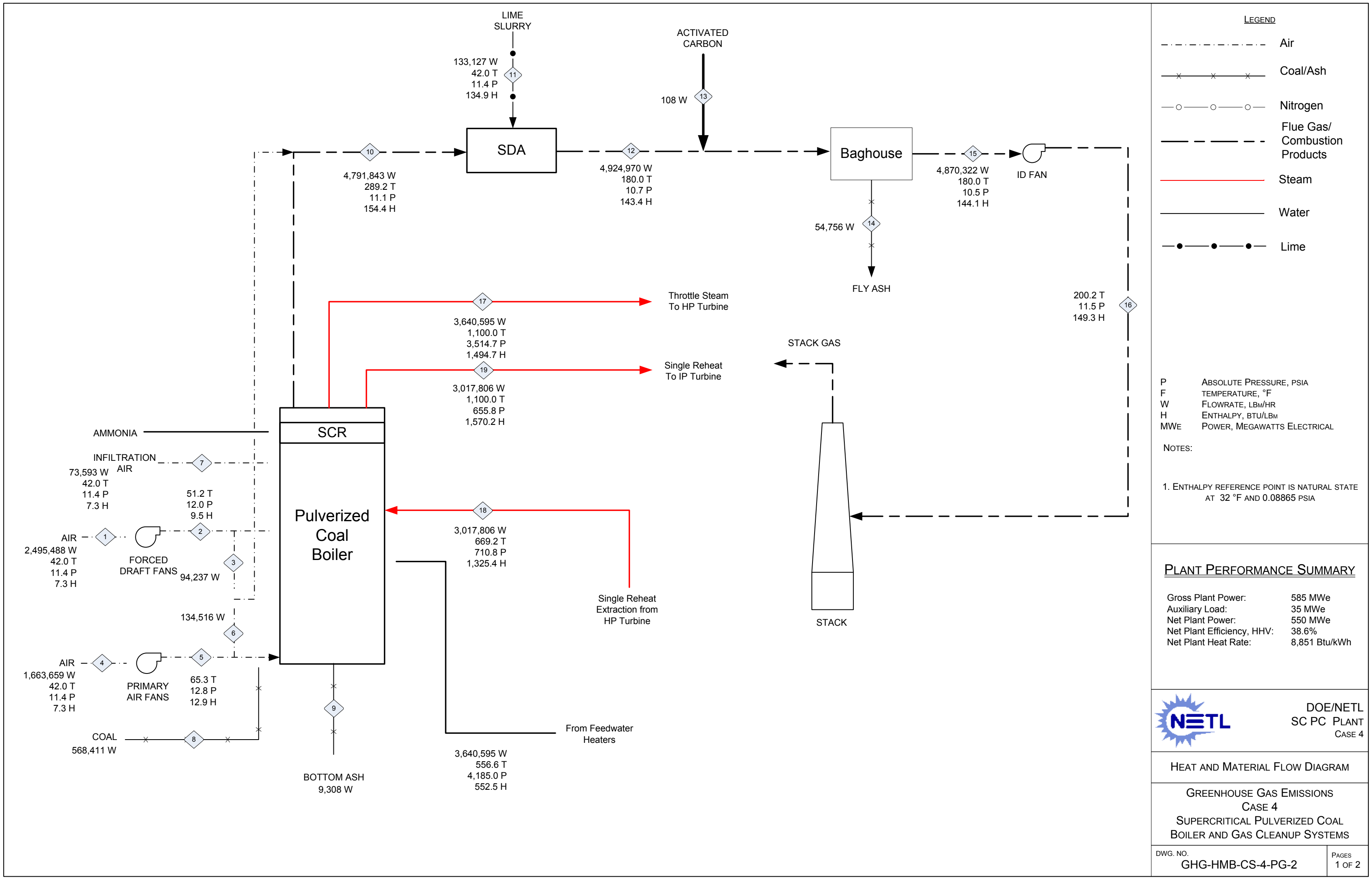


Exhibit 6-15 Case 4 SC PC without CO_2 Power Block Systems Heat and Mass Balance Schematic

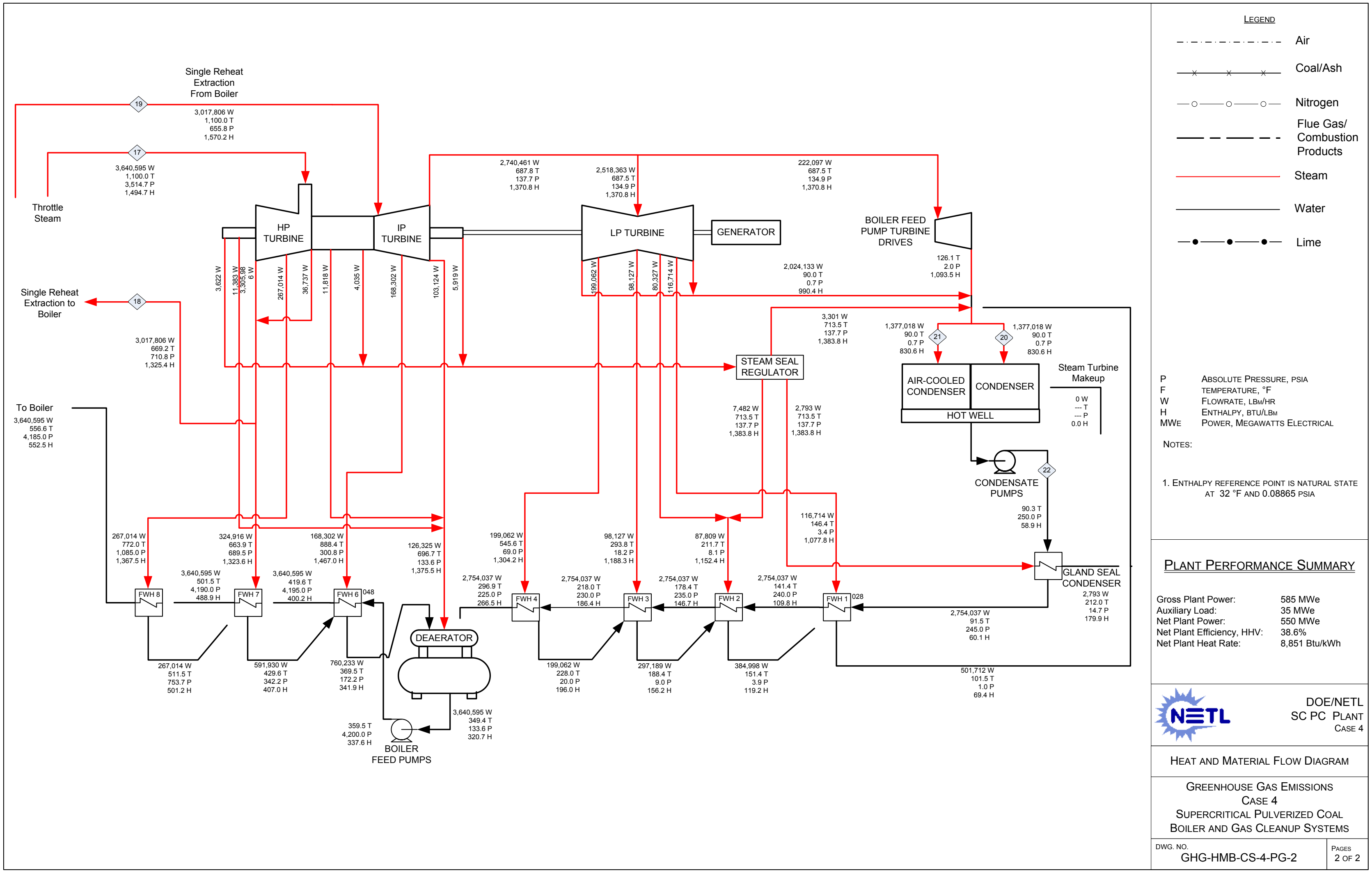


Exhibit 6-16 Case 5 SC PC with CO_2 Capture to an Emissions Limit of 1,100 lb/net-MWh Boiler and Gas Cleanup Heat and Mass Balance Schematic

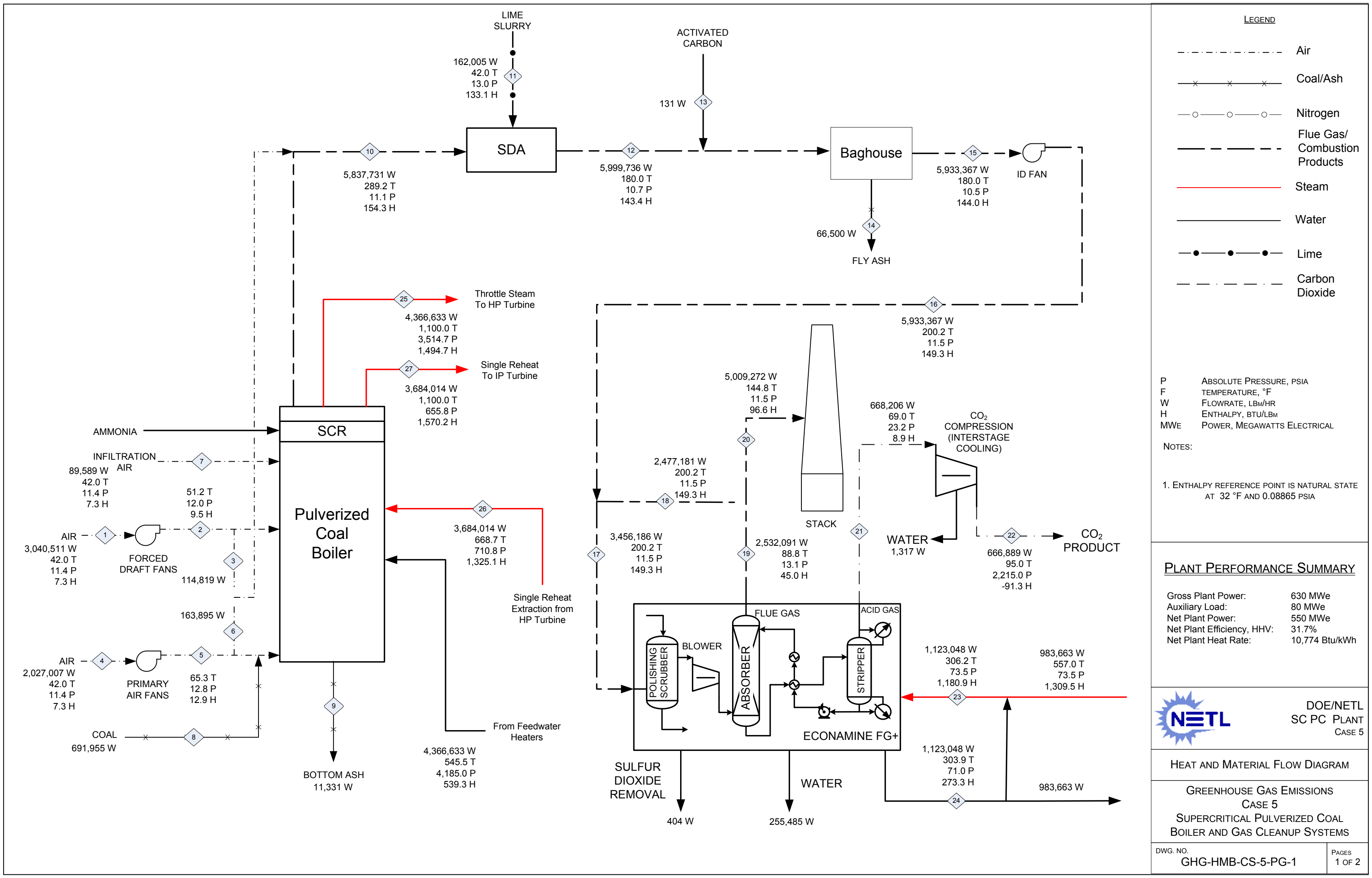


Exhibit 6-17 Case 5 SC PC with CO_2 Capture to an Emissions Limit of 1,100 lb/net-MWh Power Block Systems Heat and Mass Balance Schematic

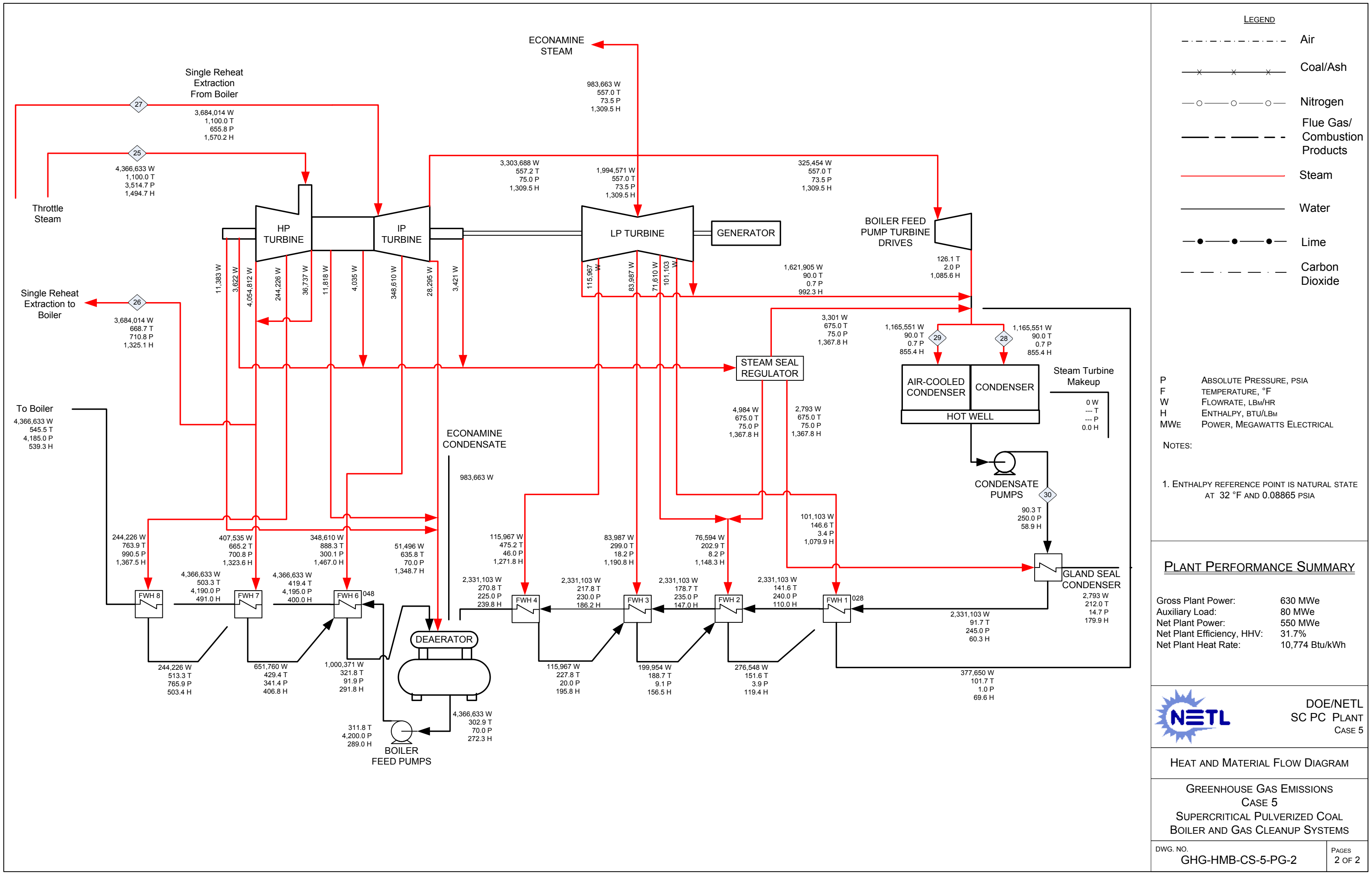


Exhibit 6-18 Case 6 SC PC with 90% CO_2 Capture Boiler and Gas Cleanup System Heat and Mass Balance Schematic

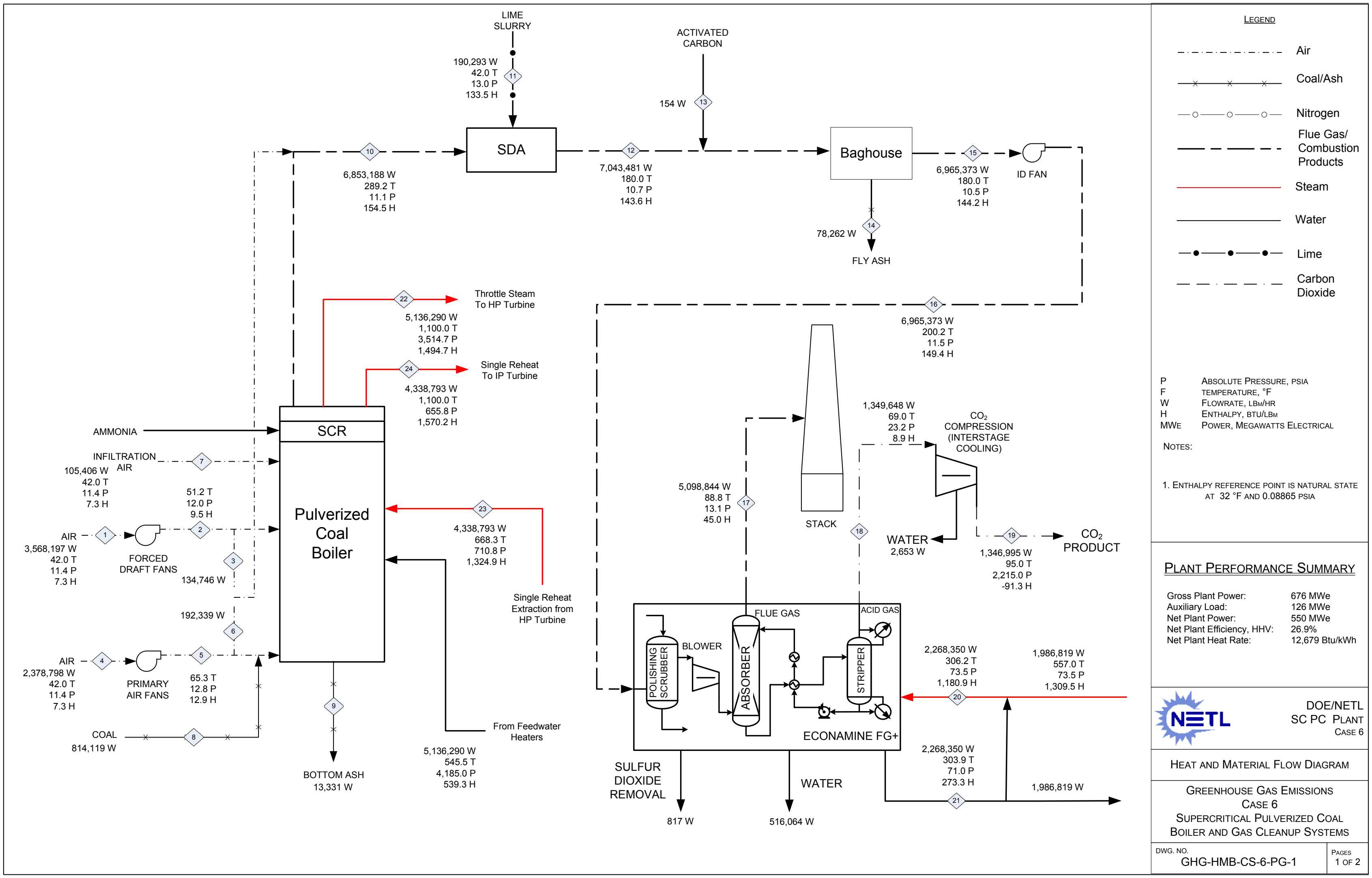


Exhibit 6-19 Case 6 SC PC with 90% CO_2 Capture Power Block System Heat and Mass Balance Schematic

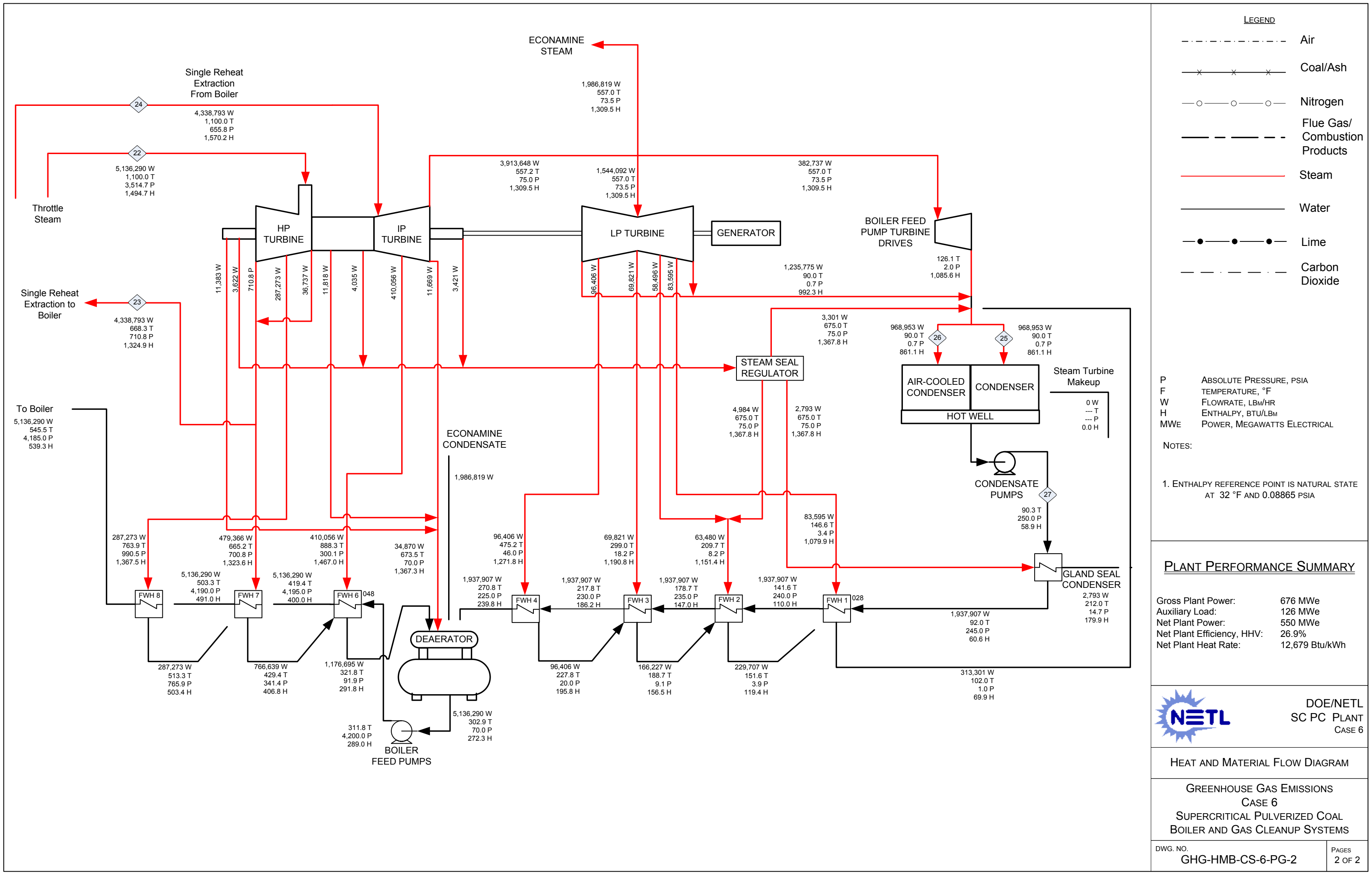


Exhibit 6-20 Cases 4 - 6 Overall Energy Balance

	Case 4	Case 5	Case 6
Energy In, GJ/hr (MMBtu/hr)[1]			
Coal, HHV	5,136 (4,868)	6,252 (5,926)	7,356 (6,972)
Sensible + Latent			
Coal	2.6 (2.5)	3.2 (3.0)	3.8 (3.6)
Air	32.5 (30.8)	39.6 (37.5)	46.5 (44.0)
Raw Water Makeup	14.4 (13.7)	27.0 (25.6)	46.5 (44.1)
Lime	0.02 (0.02)	0.02 (0.02)	0.02 (0.02)
Auxiliary Power	127 (120)	287 (272)	452 (429)
Total In	**5,312 (5,035)**	**6,609 (6,264)**	**7,905 (7,492)**
Energy Out, GJ/hr (MMBtu/hr)[1]			
Sensible + Latent			
Bottom Ash	0.5 (0.4)	0.6 (0.5)	0.7 (0.6)
Fly Ash + FGD Ash	1.6 (1.6)	2.0 (1.9)	2.4 (2.2)
Flue Gas	767 (727)	510 (484)	242 (230)
Condenser	2,245 (2,128)	1,961 (1,859)	1,642 (1,556)
CO_2	N/A	-64 (-61)	-130 (-123)
Cooling Tower Blowdown	11.8 (11.2)	25.2 (23.9)	44.9 (42.5)
Econamine Losses	N/A	1,536 (1,456)	3,298 (3,126)
Process Losses[2]	*179 (170)*	*370 (351)*	*373 (353)*
Power	2,107 (1,997)	2,267 (2,149)	2,432 (2,305)
Total Out	**5,312 (5,035)**	**6,609 (6,264)**	**7,905 (7,492)**

[1] Enthalpy reference conditions are 0°C (32°F) and 614 Pa (0.089 psia)

[2] Process losses are calculated by difference to close the energy balance

6.1.6 Case 4 - 6 – Major Equipment List

Major equipment items for the supercritical PC plant with and without CO_2 capture are shown in the following tables. The accounts used in the equipment list correspond to the account numbers used in the cost estimates in Section 6.1.7. In general, the design conditions include a 10 percent contingency for flows and heat duties and a 21 percent contingency for heads on pumps and fans.

ACCOUNT 1 COAL HANDLING

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
1	Feeder	Belt	2	0	572 tonne/hr (630 tph)	572 tonne/hr (630 tph)	572 tonne/hr (630 tph)
2	Conveyor No. 1	Belt	1	0	1,134 tonne/hr (1,250 tph)	1,134 tonne/hr (1,250 tph)	1,134 tonne/hr (1,250 tph)
3	Transfer Tower No. 1	Enclosed	1	0	N/A	N/A	N/A
4	Conveyor No. 2	Belt	1	0	1,134 tonne/hr (1,250 tph)	1,134 tonne/hr (1,250 tph)	1,134 tonne/hr (1,250 tph)
5	As-Received Coal Sampling System	Two-stage	1	0	N/A	N/A	N/A
6	Stacker/Reclaimer	Traveling, linear	1	0	1,134 tonne/hr (1,250 tph)	1,134 tonne/hr (1,250 tph)	1,134 tonne/hr (1,250 tph)
7	Reclaim Hopper	N/A	2	1	54 tonne (60 ton)	64 tonne (70 ton)	73 tonne (80 ton)
8	Feeder	Vibratory	2	1	209 tonne/hr (230 tph)	263 tonne/hr (290 tph)	308 tonne/hr (340 tph)
9	Conveyor No. 3	Belt w/ tripper	1	0	426 tonne/hr (470 tph)	517 tonne/hr (570 tph)	608 tonne/hr (670 tph)
10	Crusher Tower	N/A	1	0	N/A	N/A	N/A
11	Coal Surge Bin w/ Vent Filter	Dual outlet	2	0	209 tonne (230 ton)	263 tonne (290 ton)	308 tonne (340 ton)
12	Crusher	Impactor reduction	2	0	8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0)	8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0)	8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0)

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
13	As-Fired Coal Sampling System	Swing hammer	1	1	N/A	N/A	N/A
14	Conveyor No. 4	Belt w/tripper	1	0	426 tonne/hr (470 tph)	517 tonne/hr (570 tph)	608 tonne/hr (670 tph)
15	Transfer Tower No. 2	Enclosed	1	0	N/A	N/A	N/A
16	Conveyor No. 5	Belt w/ tripper	1	0	426 tonne/hr (470 tph)	517 tonne/hr (570 tph)	608 tonne/hr (670 tph)
17	Coal Silo w/ Vent Filter and Slide Gates	Field erected	3	0	907 tonne (1,000 ton)	1,179 tonne (1,300 ton)	1,361 tonne (1,500 ton)
18	Lime Truck Unloading System	N/A	1	0	18 tonne/hr (20 tph)	18 tonne/hr (20 tph)	27 tonne/hr (30 tph)
19	Lime Bulk Storage Silo w/Vent Filter	Field erected	3	0	454 tonne (500 ton)	544 tonne (600 ton)	726 tonne (800 ton)
20	Lime Live Storage Transport	Pneumatic	1	0	6 tonne/hr (7 tph)	8 tonne/hr (9 tph)	9 tonne/hr (10 tph)
21	Lime Day Bin	w/ actuator	2	0	54 tonne (60 ton)	64 tonne (70 ton)	73 tonne (80 ton)
22	Activated Carbon Storage Silo and Feeder System with Vent Filter	Shop assembled	1	0	Silo - 36 tonne (40 ton) Feeder - 54 kg/hr (120 lb/hr)	Silo - 45 tonne (50 ton) Feeder - 64 kg/hr (140 lb/hr)	Silo - 54 tonne (60 ton) Feeder - 77 kg/hr (170 lb/hr)

ACCOUNT 2 COAL PREPARATION AND FEED

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
1	Coal Feeder	Gravimetric	6	0	45 tonne/hr (50 tph)	54 tonne/h (60 tph)	64 tonne/hr (70 tph)
2	Coal Pulverizer	Ball type or equivalent	6	0	45 tonne/hr (50 tph)	54 tonne/h (60 tph)	64 tonne/hr (70 tph)
3	Lime Slaker	N/A	1	1	5 tonne/hr (6 tph)	7 tonne/h (8 tph)	8 tonne/hr (9 tph)
4	Lime Slurry Tank	Field Erected	1	1	276,337 liters (73,000 gal)	333,119 liters (88,000 gal)	389,900 liters (103,000 gal)
5	Lime Slurry Feed Pumps	Horizontal centrifugal	1	1	303 lpm @ 9m H_2O (80 gpm @ 30 ft H_2O)	341 lpm @ 9m H_2O (90 gpm @ 30 ft H_2O)	416 lpm @ 9m H_2O (110 gpm @ 30 ft H_2O)

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
1	Condensate Pumps	Vertical canned	1	1	23,091 lpm @ 213 m H_2O (6,100 gpm @ 700 ft H_2O)	19,684 lpm @ 213 m H_2O (5,200 gpm @ 700 ft H_2O)	16,277 lpm @ 213 m H_2O (4,300 gpm @ 700 ft H_2O)
2	Deaerator and Storage Tank	Horizontal spray type	1	0	1,816,637 kg/hr (4,005,000 lb/hr), 5 min. tank	2,178,604 kg/hr (4,803,000 lb/hr), 5 min. tank	2,562,797 kg/hr (5,650,000 lb/hr), 5 min. tank
3	Boiler Feed Pump/Turbine	Barrel type, multi-stage, centrifugal	1	1	30,662 lpm @ 3,444 m H_2O (8,100 gpm @	36,718 lpm @ 3,505 m H_2O (9,700 gpm @	43,154 lpm @ 3,505 m H_2O (11,400 gpm @

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
					11,300 ft H_2O)	11,500 ft H_2O)	11,500 ft H_2O)
4	Startup Boiler Feed Pump, Electric Motor Driven	Barrel type, multi-stage, centrifugal	1	0	9,085 lpm @ 3,444 m H_2O (2,400 gpm @ 11,300 ft H_2O)	10,978 lpm @ 3,505 m H_2O (2,900 gpm @ 11,500 ft H_2O)	12,870 lpm @ 3,505 m H_2O (3,400 gpm @ 11,500 ft H_2O)
5	LP Feedwater Heater 1A/1B	Horizontal U-tube	2	0	684,924 kg/hr (1,510,000 lb/hr)	580,598 kg/hr (1,280,000 lb/hr)	485,344 kg/hr (1,070,000 lb/hr)
6	LP Feedwater Heater 2A/2B	Horizontal U-tube	2	0	684,924 kg/hr (1,510,000 lb/hr)	580,598 kg/hr (1,280,000 lb/hr)	485,344 kg/hr (1,070,000 lb/hr)
7	LP Feedwater Heater 3A/3B	Horizontal U-tube	2	0	684,924 kg/hr (1,510,000 lb/hr)	580,598 kg/hr (1,280,000 lb/hr)	485,344 kg/hr (1,070,000 lb/hr)
8	LP Feedwater Heater 4A/4B	Horizontal U-tube	2	0	684,924 kg/hr (1,510,000 lb/hr)	580,598 kg/hr (1,280,000 lb/hr)	485,344 kg/hr (1,070,000 lb/hr)
9	HP Feedwater Heater 6	Horizontal U-tube	1	0	1,814,369 kg/hr (4,000,000 lb/hr)	2,177,243 kg/hr (4,800,000 lb/hr)	2,562,797 kg/hr (5,650,000 lb/hr)
10	HP Feedwater Heater 7	Horizontal U-tube	1	0	1,814,369 kg/hr (4,000,000 lb/hr)	2,177,243 kg/hr (4,800,000 lb/hr)	2,562,797 kg/hr (5,650,000 lb/hr)
11	HP Feedwater heater 8	Horizontal U-tube	1	0	1,814,369 kg/hr (4,000,000 lb/hr)	2,177,243 kg/hr (4,800,000 lb/hr)	2,562,797 kg/hr (5,650,000 lb/hr)
12	Auxiliary Boiler	Shop fabricated, water tube	1	0	18,144 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F)	18,144 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F)	18,144 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F)
13	Fuel Oil System	No. 2 fuel oil for light off	1	0	1,135,624 liter (300,000 gal)	1,135,624 liter (300,000 gal)	1,135,624 liter (300,000 gal)
14	Service Air Compressors	Flooded Screw	2	1	28 m^3/min @ 0.7 MPa (1,000 scfm	28 m^3/min @ 0.7 MPa (1,000 scfm	28 m^3/min @ 0.7 MPa (1,000 scfm

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
					@ 100 psig)	@ 100 psig)	@ 100 psig)
15	Instrument Air Dryers	Duplex, regenerative	2	1	28 m^3/min (1,000 scfm)	28 m^3/min (1,000 scfm)	28 m^3/min (1,000 scfm)
16	Closed Cycle Cooling Heat Exchangers	Shell and tube	2	0	53 GJ/hr (50 MMBtu/hr) each	53 GJ/hr (50 MMBtu/hr) each	53 GJ/hr (50 MMBtu/hr) each
17	Closed Cycle Cooling Water Pumps	Horizontal centrifugal	2	1	20,820 lpm @ 30 m H_2O (5,500 gpm @ 100 ft H_2O)	20,820 lpm @ 30 m H_2O (5,500 gpm @ 100 ft H_2O)	20,820 lpm @ 30 m H_2O (5,500 gpm @ 100 ft H_2O)
18	Engine-Driven Fire Pump	Vertical turbine, diesel engine	1	1	3,785 lpm @ 88 m H_2O (1,000 gpm @ 290 ft H_2O)	3,785 lpm @ 88 m H_2O (1,000 gpm @ 290 ft H_2O)	3,785 lpm @ 88 m H_2O (1,000 gpm @ 290 ft H_2O)
19	Fire Service Booster Pump	Two-stage horizontal centrifugal	1	1	2,650 lpm @ 64 m H_2O (700 gpm @ 210 ft H_2O)	2,650 lpm @ 64 m H_2O (700 gpm @ 210 ft H_2O)	2,650 lpm @ 64 m H_2O (700 gpm @ 210 ft H_2O)
20	Raw Water Pumps	Stainless steel, single suction	2	1	2,688 lpm @ 43 m H_2O (710 gpm @ 140 ft H_2O)	5,565 lpm @ 43 m H_2O (1,470 gpm @ 140 ft H_2O)	8,896 lpm @ 43 m H_2O (2,350 gpm @ 140 ft H_2O)
21	Ground Water Pumps	Stainless steel, single suction	2	1	2,688 lpm @ 268 m H_2O (710 gpm @ 880 ft H_2O)	2,801 lpm @ 268 m H_2O (740 gpm @ 880 ft H_2O)	2,953 lpm @ 268 m H_2O (780 gpm @ 880 ft H_2O)
22	Filtered Water Pumps	Stainless steel, single suction	2	1	151 lpm @ 49 m H_2O (40 gpm @ 160 ft H_2O)	227 lpm @ 49 m H_2O (60 gpm @ 160 ft H_2O)	265 lpm @ 49 m H_2O (70 gpm @ 160 ft H_2O)
23	Filtered Water Tank	Vertical, cylindrical	1	0	158,987 liter (42,000 gal)	200,627 liter (53,000 gal)	238,481 liter (63,000 gal)

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
24	Makeup Water Demineralizer	Multi-media filter, cartridge filter, RO membrane assembly, electrode ionization unit	1	1	341 lpm (90 gpm)	416 lpm (110 gpm)	492 lpm (130 gpm)
25	Liquid Waste Treatment System	--	1	0	10 years, 24-hour storm	10 years, 24-hour storm	10 years, 24-hour storm

ACCOUNT 4 BOILER AND ACCESSORIES

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
1	Boiler	Supercritical, drum, wall-fired, low NOx burners, overfire air	1	0	1,814,369 kg/hr steam @ 25.5 MPa/602°C/602°C (4,000,000 lb/hr steam @ 3,700 psig/1,115°F/1,115°F)	2,177,243 kg/hr steam @ 25.5 MPa/602°C/602°C (4,800,000 lb/hr steam @ 3,700 psig/1,115°F/1,115°F)	2,562,797 kg/hr steam @ 25.5 MPa/602°C/602°C (5,650,000 lb/hr steam @ 3,700 psig/1,115°F/1,115°F)
2	Primary Air Fan	Centrifugal	2	0	415,037 kg/hr, 7,054 m^3/min @ 123 cm WG (915,000 lb/hr, 249,100 acfm @ 48 in. WG)	505,755 kg/hr, 8,597 m^3/min @ 123 cm WG (1,115,000 lb/hr, 303,600 acfm @ 48 in. WG)	593,299 kg/hr, 10,086 m^3/min @ 123 cm WG (1,308,000 lb/hr, 356,200 acfm @ 48 in. WG)
3	Forced Draft Fan	Centrifugal	2	0	622,782 kg/hr, 10,582 m^3/min @ 47 cm WG (1,373,000 lb/hr, 373,700 acfm @ 19 in. WG)	758,406 kg/hr, 12,893 m^3/min @ 47 cm WG (1,672,000 lb/hr, 455,300 acfm @ 19 in. WG)	890,402 kg/hr, 15,133 m^3/min @ 47 cm WG (1,963,000 lb/hr, 534,400 acfm @ 19 in. WG)

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
4	Induced Draft Fan	Centrifugal 2		0	1,215,174 kg/hr, 28,524 m^3/min @ 82 cm WG (2,679,000 lb/hr, 1,007,300 acfm @ 32 in. WG)	1,480,072 kg/hr, 34,750 m^3/min @ 82 cm WG (3,263,000 lb/hr, 1,227,200 acfm @ 32 in. WG)	1,737,712 kg/hr, 40,793 m^3/min @ 82 cm WG (3,831,000 lb/hr, 1,440,600 acfm @ 32 in. WG)
5	SCR Reactor Vessel	Space for spare layer	2 0		2,431,255 kg/hr (5,360,000 lb/hr)	2,961,958 kg/hr (6,530,000 lb/hr)	3,474,518 kg/hr (7,660,000 lb/hr)
6 SCR	Catalyst	--	3	0	-- -- --		
7	Dilution Air Blower	Centrifugal 2		1	40 m^3/min @ 108 cm WG (1,400 acfm @ 42 in. WG)	51 m^3/min @ 108 cm WG (1,800 acfm @ 42 in. WG)	59 m^3/min @ 108 cm WG (2,100 acfm @ 42 in. WG)
8	Ammonia Storage	Horizontal tank	5 0		45,425 liter (12,000 gal)	52,996 liter (14,000 gal)	64,352 liter (17,000 gal)
9	Ammonia Feed Pump	Centrifugal 2		1	9 lpm @ 91 m H_2O 2 gpm @ 300 ft H_2O)	10 lpm @ 91 m H_2O (3 gpm @ 300 ft H_2O)	12 lpm @ 91 m H_2O (3 gpm @ 300 ft H_2O)

ACCOUNT 5 FLUE GAS CLEANUP

Equipment No.	**Description**	**Type**	**Operating Qty.**	**Spares**	**Case 4 Design Condition**	**Case 5 Design Condition**	**Case 6 Design Condition**
1	Fabric Filter	Single stage, high-ratio with pulse-jet online cleaning system, air-to-cloth ratio - 3.5 ft/min	2 0		1,215,174 kg/hr (2,679,000 lb/hr) 99.9% efficiency	1,480,072 kg/hr (3,263,000 lb/hr) 99.9% efficiency	1,737,712 kg/hr (3,831,000 lb/hr) 99.9% efficiency
2	Spray Dryer	Co-current open spray	2	0	30,356 m^3/min (1,072,000 acfm)	36,982 m^3/min (1,306,000 acfm)	43,410 m^3/min (1,533,000 acfm)

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
3	Atomizer	Rotary	2	1	151 lpm @ 64 m H_2O (40 gpm @ 210 ft H_2O)	189 lpm @ 64 m H_2O (50 gpm @ 210 ft H_2O)	227 lpm @ 64 m H_2O (60 gpm @ 210 ft H_2O)
4	Spray Dryer Solids Conveying	---	2	0	---	---	---
5	Carbon Injectors	---	1	0	54 kg/hr (120 lb/hr)	64 kg/hr (140 lb/hr)	77 kg/hr (170 lb/hr)

ACCOUNT 5B CO$_2$ COMPRESSION

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
1	Econamine FG Plus	Amine-based CO_2 capture technology	2	0	N/A	862,279 kg/h (1,901,000 lb/h) 21.4 wt % CO_2 concentration	1,737,712 kg/h (3,831,000 lb/h) 21.5 wt % CO_2 concentration
2	Econamine Condensate Pump	Centrifugal	1	1	N/A	9,350 lpm @ 52 m H_2O (2,470 gpm @ 170 ft H_2O)	18,889 lpm @ 52 m H_2O (4,990 gpm @ 170 ft H_2O)
3	CO2 Compressor	Reciprocating	2	0	N/A	166,363 kg/h @ 15.3 MPa (366,767 lb/h @ 2,215 psia)	336,022 kg/h @ 15.3 MPa (740,801 lb/h @ 2,215 psia)

ACCOUNT 5C CO$_2$ TRANSPORT, STORAGE, AND MONITORING (not shown in Total Plant Cost Details)

Equipment No.	Description	Type	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
1	CO$_2$ Pipeline	Carbon Steel	N/A	50 miles @ 14 in diameter w/ inlet pressure of 2,200 psi and outlet pressure of 1,500 psi	50 miles @ 18 in diameter w/ inlet pressure of 2,200 psi and outlet pressure of 1,500 psi
2	CO$_2$ Sequestration Source	Saline Formation	N/A	1 well with bottom hole pressure @ 1,220 psi, 530 ft thickness, 4,055 ft depth, 22 Md permeability	2 wells with bottom hole pressure @ 1,220 psi, 530 ft thickness, 4,055 ft depth, 22 Md permeability
3	CO$_2$ Monitoring	N/A	N/A	20 year monitoring life during plant life / 80 years following / Total of 100 years	20 year monitoring life during plant life / 80 years following / Total of 100 years

ACCOUNT 6 COMBUSTION TURBINE/ACCESSORIES

N/A

ACCOUNT 7 HRSG, DUCTING & STACK

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
1	Stack	Reinforced concrete with FRP liner	1	0	152 m (500 ft) high x 6.8 m (22 ft) diameter	152 m (500 ft) high x 6.7 m (22 ft) diameter	152 m (500 ft) high x 6.1 m (20 ft) diameter

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
1	Steam Turbine	Commercially available advanced steam turbine	1	0	616 MW 24.1 MPa/593°C/593°C (3500 psig/ 1100°F/1100°F)	663 MW 24.1 MPa/593°C/593°C (3500 psig/ 1100°F/1100°F)	711 MW 24.1 MPa/593°C/593°C (3500 psig/ 1100°F/1100°F)
2	Steam Turbine Generator	Hydrogen cooled, static excitation	1	0	680 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	740 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	790 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase
3	Surface Condenser	Single pass, divided waterbox including vacuum pumps	1	0	1,230 GJ/hr (1,170 MMBtu/hr), Condensing temperature 32°C (90°F), Inlet water temperature 9°C (48°F), Water temperature rise 11°C (20°F)	1,080 GJ/hr (1,020 MMBtu/hr), Condensing temperature 32°C (90°F), Inlet water temperature 9°C (48°F), Water temperature rise 11°C (20°F)	910 GJ/hr (860 MMBtu/hr), Condensing temperature 32°C (90°F), Inlet water temperature 9°C (48°F), Water temperature rise 11°C (20°F)
4	Air Cooled Condenser	Ambient air to steam	1	0	1,230 GJ/hr (1,170 MMBtu/hr), Condensing temperature 32°C (90°F), Ambient temperature 6°C (42°F)	1,080 GJ/hr (1,020 MMBtu/hr), Condensing temperature 32°C (90°F), Ambient temperature 6°C (42°F)	910 GJ/hr (860 MMBtu/hr), Condensing temperature 32°C (90°F), Ambient temperature 6°C (42°F)

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
1	Circulating Water Pumps	Vertical, wet pit	2	1	242,300 lpm @ 30 m (64,000 gpm @ 100 ft)	518,600 lpm @ 30 m (137,000 gpm @ 100 ft)	923,600 lpm @ 30 m (244,000 gpm @ 100 ft)
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	1	0	3°C (37°F) wet bulb / 9°C (48°F) CWT / 20°C (68°F) HWT / 1,350 GJ/hr (1,280 MMBtu/hr) heat duty	3°C (37°F) wet bulb / 9°C (48°F) CWT / 20°C (68°F) HWT / 2,880 GJ/hr (2,730 MMBtu/hr) heat duty	3°C (37°F) wet bulb / 9°C (48°F) CWT / 20°C (68°F) HWT / 5,138 GJ/hr (4,870 MMBtu/hr) heat duty

ACCOUNT 10 ASH/SPENT SORBENT RECOVERY AND HANDLING

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
1	Economizer Hopper (part of boiler scope of supply)	--	4	0	--	--	--
2	Bottom Ash Hopper (part of boiler scope of supply)	--	2	0	--	--	--
3	Clinker Grinder	--	1	1	4.5 tonne/hr (5 tph)	5.4 tonne/hr (6 tph)	6.4 tonne/hr (7 tph)
4	Pyrites Hopper (part of pulverizer scope of supply included with boiler)	--	6	0	--	--	--
5	Hydroejectors	--	12		--	--	--

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
6	Economizer /Pyrites Transfer Tank	--	1	0	--	--	--
7	Ash Sluice Pumps	Vertical, wet pit	1	1	189 lpm @ 17 m H_2O (50 gpm @ 56 ft H_2O)	227 lpm @ 17 m H_2O (60 gpm @ 56 ft H_2O)	265 lpm @ 17 m H2O (70 gpm @ 56 ft H2O)
8	Ash Seal Water Pumps	Vertical, wet pit	1	1	7,571 lpm @ 9 m H_2O (2000 gpm @ 28 ft H_2O)	7,571 lpm @ 9 m H_2O (2000 gpm @ 28 ft H_2O)	7,571 lpm @ 9 m H_2O (2000 gpm @ 28 ft H_2O)
9	Hydrobins	--	1	1	189 lpm (50 gpm)	227 lpm (60 gpm)	265 lpm (70 gpm)
10	Baghouse Hopper (part of baghouse scope of supply)	--	24	0	--	--	--
11	Air Heater Hopper (part of boiler scope of supply)	--	10	0	--	--	--
12	Air Blower	--	1	1	25 m^3/min @ 0.2 MPa (880 scfm @ 24 psi)	30 m^3/min @ 0.2 MPa (1070 scfm @ 24 psi)	36 m^3/min @ 0.2 MPa (1260 scfm @ 24 psi)
13	Fly Ash Silo	Reinforced concrete	2	0	1,630 tonne (1,800 ton)	2,000 tonne (2,200 ton)	2,360 tonne (2,600 ton)
14	Slide Gate Valves	--	2	0	--	--	--
15	Unloader	--	1	0	--	--	--
16	Telescoping Unloading Chute	--	1	0	154 tonne/hr (170 tph)	181 tonne/hr (200 tph)	218 tonne/hr (240 tph)
17	Recycle Waste Storage Silo	Reinforced concrete	2	0	272 tonne (300 ton)	363 tonne (400 ton)	454 tonne (500 ton)
18	Recycle Waste Conveyor	--	1	0	36 tonne/hr (40 tph)	45 tonne/hr (50 tph)	54 tonne/hr (60 tph)

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
19	Recycle Slurry Mixer	--	1	1	984 lpm (260 gpm)	1,211 lpm (320 gpm)	1,438 lpm (380 gpm)
20	Recycle Waste Slurry Tank	--	1	0	60,570 liters (16,000 gal)	71,920 liters (19,000 gal)	87,060 liters (23,000 gal)
21	Recycle Waste Pump	--	1	1	984 lpm (260 gpm)	1,211 lpm (320 gpm)	1,438 lpm (380 gpm)

ACCOUNT 11 ACCESSORY ELECTRIC PLANT

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
1	STG Transformer	Oil-filled	1	0	24 kV/345 kV, 650 MVA, 3-ph, 60 Hz	24 kV/345 kV, 650 MVA, 3-ph, 60 Hz	24 kV/345 kV, 650 MVA, 3-ph, 60 Hz
2	Auxiliary Transformer	Oil-filled	1	1	24 kV/4.16 kV, 37 MVA, 3-ph, 60 Hz	24 kV/4.16 kV, 86 MVA, 3-ph, 60 Hz	24 kV/4.16 kV, 137 MVA, 3-ph, 60 Hz
3	Low Voltage Transformer	Dry ventilated	1	1	4.16 kV/480 V, 6 MVA, 3-ph, 60 Hz	4.16 kV/480 V, 13 MVA, 3-ph, 60 Hz	4.16 kV/480 V, 21 MVA, 3-ph, 60 Hz
4	STG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	1	0	24 kV, 3-ph, 60 Hz	24 kV, 3-ph, 60 Hz	24 kV, 3-ph, 60 Hz
5	Medium Voltage Switchgear	Metal clad	1	1	4.16 kV, 3-ph, 60 Hz	4.16 kV, 3-ph, 60 Hz	4.16 kV, 3-ph, 60 Hz
6	Low Voltage Switchgear	Metal enclosed	1	1	480 V, 3-ph, 60 Hz	480 V, 3-ph, 60 Hz	480 V, 3-ph, 60 Hz
7	Emergency Diesel Generator	Sized for emergency shutdown	1	0	750 kW, 480 V, 3-ph, 60 Hz	750 kW, 480 V, 3-ph, 60 Hz	750 kW, 480 V, 3-ph, 60 Hz

ACCOUNT 12 INSTRUMENTATION AND CONTROL

Equipment No.	Description	Type	Operating Qty.	Spares	Case 4 Design Condition	Case 5 Design Condition	Case 6 Design Condition
1	DCS - Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W)	1	0	Operator stations/printers and engineering stations/printers	Operator stations/printers and engineering stations/printers	Operator stations/printers and engineering stations/printers
2	DCS - Processor	Microprocessor with redundant input/output	1	0	N/A	N/A	N/A
3	DCS - Data Highway	Fiber optic	1	0	Fully redundant, 25% spare	Fully redundant, 25% spare	Fully redundant, 25% spare

6.1.7 Case 4 – Cost Estimating

The cost estimating methodology was described previously in Section 2.6. Exhibit 6-21 shows the total plant capital cost details organized by cost. Exhibit 6-22 shows the initial and annual O&M costs.

The estimated TOC of the supercritical PC case with no CO_2 capture is \$2,296/kW. Owner's costs represent 18 percent of the TOC. The current dollar, 30-year LCOE is \$79.86/MWh.

Exhibit 6-21 Case 4 Total Plant Cost Details

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		Case 4 TOTAL PLANT COST	
				Direct	Indirect				Process	Project	$	$/kW
1	COAL & SORBENT HANDLING											
1.1	Coal Receive & Unload	$4,127	$0	$1,885	$0	$0	$6,012	$537	$0	$982	$7,531	$14
1.2	Coal Stackout & Reclaim	$5,333	$0	$1,208	$0	$0	$6,542	$572	$0	$1,067	$8,181	$15
1.3	Coal Conveyors	$4,959	$0	$1,196	$0	$0	$6,154	$539	$0	$1,004	$7,698	$14
1.4	Other Coal Handling	$1,297	$0	$277	$0	$0	$1,574	$138	$0	$257	$1,968	$4
1.5	Sorbent Receive & Unload	$50	$0	$15	$0	$0	$65	$6	$0	$11	$82	$0
1.6	Sorbent Stackout & Reclaim	$811	$0	$149	$0	$0	$960	$84	$0	$157	$1,200	$2
1.7	Sorbent Conveyors	$290	$63	$71	$0	$0	$423	$37	$0	$69	$529	$1
1.8	Other Sorbent Handling	$175	$41	$92	$0	$0	$308	$27	$0	$50	$385	$1
1.9	Coal & Sorbent Hnd.Foundations	$0	$5,069	$6,394	$0	$0	$11,463	$1,077	$0	$1,881	$14,420	$26
	SUBTOTAL 1.	**$17,042**	**$5,172**	**$11,286**	**$0**	**$0**	**$33,501**	**$3,016**	**$0**	**$5,478**	**$41,994**	**$76**
2	COAL & SORBENT PREP & FEED											
2.1	Coal Crushing & Drying	$2,395	$0	$467	$0	$0	$2,862	$249	$0	$467	$3,578	$7
2.2	Coal Conveyor to Storage	$6,132	$0	$1,338	$0	$0	$7,470	$653	$0	$1,218	$9,342	$17
2.3	Coal Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.4	Misc.Coal Prep & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.5	Sorbent Prep Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.6	Sorbent Storage & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.7	Sorbent Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.8	Booster Air Supply System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.9	Coal & Sorbent Feed Foundation	$0	$685	$575	$0	$0	$1,260	$117	$0	$206	$1,583	$3
	SUBTOTAL 2.	**$8,527**	**$685**	**$2,380**	**$0**	**$0**	**$11,592**	**$1,019**	**$0**	**$1,892**	**$14,502**	**$26**
3	FEEDWATER & MISC. BOP SYSTEMS											
3.1	FeedwaterSystem	$18,522	$0	$5,983	$0	$0	$24,506	$2,141	$0	$3,997	$30,644	$56
3.2	Water Makeup & Pretreating	$2,296	$0	$739	$0	$0	$3,035	$287	$0	$665	$3,987	$7
3.3	Other Feedwater Subsystems	$5,671	$0	$2,396	$0	$0	$8,067	$723	$0	$1,318	$10,108	$18
3.4	Service Water Systems	$450	$0	$245	$0	$0	$695	$65	$0	$152	$913	$2
3.5	Other Boiler Plant Systems	$7,472	$0	$7,377	$0	$0	$14,848	$1,410	$0	$2,439	$18,697	$34
3.6	FO Supply Sys & Nat Gas	$256	$0	$320	$0	$0	$576	$54	$0	$95	$725	$1
3.7	Waste Treatment Equipment	$1,557	$0	$887	$0	$0	$2,444	$238	$0	$536	$3,219	$6
3.8	Misc. Equip.(cranes,AirComp.,Comm.)	$2,717	$0	$830	$0	$0	$3,547	$341	$0	$778	$4,666	$8
	SUBTOTAL 3.	**$38,941**	**$0**	**$18,778**	**$0**	**$0**	**$57,719**	**$5,260**	**$0**	**$9,980**	**$72,959**	**$133**
4	PC BOILER											
4.1	PC Boiler & Accessories	$186,248	$0	$91,031	$0	$0	$277,279	$26,958	$0	$30,424	$334,660	$608
4.2	SCR (w/4.1)	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.3	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.4	Boiler BoP (w/ ID Fans)	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.5	Primary Air System	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.6	Secondary Air System	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.8	Major Component Rigging	$0	w/4.1	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.9	Boiler Foundations	$0	w/14.1	w/14.1	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 4.	**$186,248**	**$0**	**$91,031**	**$0**	**$0**	**$277,279**	**$26,958**	**$0**	**$30,424**	**$334,660**	**$608**

Exhibit 6-21 Case 4 Total Plant Cost Details (Continued)

Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		Case 4 TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$	$/kW
5	FLUE GAS CLEANUP											
5.1	Absorber Vessels & Accessories	$81,057	$0	$17,563	$0	$0	$98,620	$9,403	$0	$10,802	$118,825	$216
5.2	Other FGD	$947	$0	$488	$0	$0	$1,435	$138	$0	$157	$1,730	$3
5.3	Bag House & Accessories	w/5.1	$0	w/5.1	$0	$0	$0	$0	$0	$0	$0	$0
5.4	Other Particulate Removal Materials	$20,218	$0	$10,972	$0	$0	$31,190	$3,000	$0	$3,419	$37,609	$68
5.5	Gypsum Dewatering System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.6	Mercury Removal System	w/5.1	$0	w/5.1	$0	$0	$0	$0	$0	$0	$0	$0
5.9	Open											
	SUBTOTAL 5.	**$102,221**	**$0**	**$29,024**	**$0**	**$0**	**$131,245**	**$12,541**	**$0**	**$14,379**	**$158,164**	**$288**
5B	CO_2 REMOVAL & COMPRESSION											
5B.1	CO_2 Removal System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5B.2	CO_2 Compression & Drying	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5B.3	CO_2 Removal System Let Down Turbine	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 5B.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Combustion Turbine Generator	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
6.2	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.3	Compressed Air Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.9	Combustion Turbine Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 6.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
7.2	HRSG Accessories	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.3	Ductwork	$10,547	$0	$6,777	$0	$0	$17,324	$1,511	$0	$2,825	$21,660	$39
7.4	Stack	$11,542	$0	$6,753	$0	$0	$18,295	$1,761	$0	$2,006	$22,062	$40
7.9	Duct & Stack Foundations	$0	$1,348	$1,532	$0	$0	$2,880	$270	$0	$630	$3,779	$7
	SUBTOTAL 7.	**$22,089**	**$1,348**	**$15,062**	**$0**	**$0**	**$38,499**	**$3,541**	**$0**	**$5,461**	**$47,501**	**$86**
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	$52,000	$0	$6,918	$0	$0	$58,919	$5,647	$0	$6,457	$71,022	$129
8.2	Turbine Plant Auxiliaries	$349	$0	$748	$0	$0	$1,098	$107	$0	$121	$1,326	$2
8.3a	Condenser & Auxiliaries	$4,114	$0	$2,305	$0	$0	$6,418	$618	$0	$704	$7,740	$14
8.3b	Air Cooled Condenser	$37,199	$0	$7,458	$0	$0	$44,657	$4,466	$0	$9,824	$58,947	$107
8.4	Steam Piping	$20,436	$0	$10,076	$0	$0	$30,513	$2,564	$0	$4,961	$38,038	$69
8.9	TG Foundations	$0	$1,096	$1,732	$0	$0	$2,828	$268	$0	$619	$3,715	$7
	SUBTOTAL 8.	**$114,099**	**$1,096**	**$29,237**	**$0**	**$0**	**$144,432**	**$13,668**	**$0**	**$22,686**	**$180,786**	**$329**

Exhibit 6-21 Case 4 Total Plant Cost Details (Continued)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		Case 4 TOTAL PLANT COST	
				Direct	Indirect				Process	Project	$	$/kW
9	COOLING WATER SYSTEM											
9.1	Cooling Towers	$5,750	$0	$1,791	$0	$0	$7,540	$721	$0	$826	$9,088	$17
9.2	Circulating Water Pumps	$1,190	$0	$116	$0	$0	$1,307	$111	$0	$142	$1,559	$3
9.3	Circ.Water System Auxiliaries	$363	$0	$48	$0	$0	$411	$39	$0	$45	$495	$1
9.4	Circ.Water Piping	$0	$2,877	$2,788	$0	$0	$5,665	$530	$0	$929	$7,125	$13
9.5	Make-up Water System	$278	$0	$371	$0	$0	$649	$62	$0	$107	$818	$1
9.6	Component Cooling Water Sys	$287	$0	$228	$0	$0	$516	$49	$0	$85	$649	$1
9.9	Circ.Water System Foundations& Structures	$0	$1,712	$2,721	$0	$0	$4,433	$419	$0	$971	$5,823	$11
	SUBTOTAL 9.	**$7,868**	**$4,590**	**$8,064**	**$0**	**$0**	**$20,522**	**$1,932**	**$0**	**$3,104**	**$25,558**	**$46**
10	ASH/SPENT SORBENT HANDLING SYS											
10.1	Ash Coolers	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.2	Cyclone Ash Letdown	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.3	HGCU Ash Letdown	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.4	High Temperature Ash Piping	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.5	Other Ash Recovery Equipment	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.6	Ash Storage Silos	$770	$0	$2,371	$0	$0	$3,141	$308	$0	$345	$3,794	$7
10.7	Ash Transport & Feed Equipment	$4,982	$0	$5,103	$0	$0	$10,085	$964	$0	$1,105	$12,154	$22
10.8	Misc. Ash Handling Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.9	Ash/Spent Sorbent Foundation	$0	$183	$215	$0	$0	$398	$37	$0	$87	$523	$1
	SUBTOTAL 10.	**$5,751**	**$183**	**$7,690**	**$0**	**$0**	**$13,624**	**$1,310**	**$0**	**$1,537**	**$16,471**	**$30**
11	ACCESSORY ELECTRIC PLANT											
11.1	Generator Equipment	$1,607	$0	$261	$0	$0	$1,867	$173	$0	$153	$2,194	$4
11.2	Station Service Equipment	$3,007	$0	$988	$0	$0	$3,995	$373	$0	$328	$4,696	$9
11.3	Switchgear & Motor Control	$3,457	$0	$588	$0	$0	$4,045	$375	$0	$442	$4,861	$9
11.4	Conduit & Cable Tray	$0	$2,167	$7,494	$0	$0	$9,662	$935	$0	$1,590	$12,187	$22
11.5	Wire & Cable	$0	$4,090	$7,895	$0	$0	$11,985	$1,010	$0	$1,949	$14,944	$27
11.6	Protective Equipment	$271	$0	$923	$0	$0	$1,195	$117	$0	$131	$1,443	$3
11.7	Standby Equipment	$1,282	$0	$29	$0	$0	$1,312	$120	$0	$143	$1,575	$3
11.8	Main Power Transformers	$6,183	$0	$175	$0	$0	$6,358	$484	$0	$684	$7,526	$14
11.9	Electrical Foundations	$0	$313	$768	$0	$0	$1,081	$103	$0	$237	$1,421	$3
	SUBTOTAL 11.	**$15,808**	**$6,570**	**$19,121**	**$0**	**$0**	**$41,499**	**$3,691**	**$0**	**$5,657**	**$50,846**	**$92**
12	INSTRUMENTATION & CONTROL											
12.1	PC Control Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.2	Combustion Turbine Control	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
12.3	Steam Turbine Control	w/8.1	$0	w/8.1	$0	$0	$0	$0	$0	$0	$0	$0
12.4	Other Major Component Control	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
12.5	Signal Processing Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.6	Control Boards,Panels & Racks	$454	$0	$272	$0	$0	$726	$69	$0	$119	$914	$2
12.7	Distributed Control System Equipment	$4,586	$0	$802	$0	$0	$5,388	$499	$0	$589	$6,476	$12
12.8	Instrument Wiring & Tubing	$2,486	$0	$4,932	$0	$0	$7,418	$632	$0	$1,208	$9,258	$17
12.9	Other I & C Equipment	$1,296	$0	$2,941	$0	$0	$4,237	$411	$0	$465	$5,112	$9
	SUBTOTAL 12.	**$8,823**	**$0**	**$8,946**	**$0**	**$0**	**$17,769**	**$1,611**	**$0**	**$2,380**	**$21,760**	**$40**

Exhibit 6-21 Case 4 Total Plant Cost Details (Continued)

											Case 4	
Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$	$/kW
13	IMPROVEMENTS TO SITE											
13.1	Site Preparation	$0	$50	$1,004	$0	$0	$1,055	$105	$0	$232	$1,391	$3
13.2	Site Improvements	$0	$1,667	$2,071	$0	$0	$3,738	$369	$0	$821	$4,928	$9
13.3	Site Facilities	$2,988	$0	$2,947	$0	$0	$5,934	$585	$0	$1,304	$7,823	$14
	SUBTOTAL 13.	**$2,988**	**$1,718**	**$6,022**	**$0**	**$0**	**$10,727**	**$1,058**	**$0**	**$2,357**	**$14,143**	**$26**
14	BUILDINGS & STRUCTURES											
14.1	Boiler Building	$0	$9,149	$8,046	$0	$0	$17,196	$1,546	$0	$2,811	$21,552	$39
14.2	Turbine Building	$0	$11,890	$11,081	$0	$0	$22,971	$2,070	$0	$3,756	$28,797	$52
14.3	Administration Building	$0	$588	$622	$0	$0	$1,210	$110	$0	$198	$1,518	$3
14.4	Circulation Water Pumphouse	$0	$169	$134	$0	$0	$303	$27	$0	$49	$379	$1
14.5	Water Treatment Buildings	$0	$298	$246	$0	$0	$544	$49	$0	$89	$682	$1
14.6	Machine Shop	$0	$393	$264	$0	$0	$658	$58	$0	$107	$824	$1
14.7	Warehouse	$0	$267	$267	$0	$0	$534	$48	$0	$87	$670	$1
14.8	Other Buildings & Structures	$0	$218	$185	$0	$0	$403	$36	$0	$66	$505	$1
14.9	Waste Treating Building & Str.	$0	$408	$1,238	$0	$0	$1,646	$156	$0	$270	$2,073	$4
	SUBTOTAL 14.	**$0**	**$23,380**	**$22,084**	**$0**	**$0**	**$45,465**	**$4,101**	**$0**	**$7,435**	**$57,000**	**$104**
	TOTAL COST	**$530,405**	**$44,742**	**$268,724**	**$0**	**$0**	**$843,871**	**$79,706**	**$0**	**$112,767**	**$1,036,345**	**$1,884**
	Owner's Costs											
	Preproduction Costs											
	6 Months All Labor										$8,054	$15
	1 Month Maintenance Materials										$1,083	$2
	1 Month Non-fuel Consumables										$561	$1
	1 Month Waste Disposal										$379	$1
	25% of 1 Months Fuel Cost at 100% CF										$538	$1
	2% of TPC										$20,727	$38
	Total										**$31,341**	**$57**
	Inventory Capital											
	60 day supply of fuel and consumables at 100% CF										$5,424	$10
	0.5% of TPC (spare parts)										$5,182	$9
	Total										**$10,606**	**$19**
	Initial Cost for Catalyst and Chemicals										$0	$0
	Land										$900	$2
	Other Owner's Costs										$155,452	$283
	Financing Costs										$27,981	$51
	Total Overnight Costs (TOC)										**$1,262,625**	**$2,296**
	TASC Multiplier								(IOU, low risk, 35 year)		1.134	
	Total As-Spent Cost (TASC)										**$1,431,817**	**$2,603**

Exhibit 6-22 Case 4 Initial and Annual Operating and Maintenance Costs

INITIAL & ANNUAL O&M EXPENSES					Cost Base (June)	2007
Case 4 - Supercritical PC w/o CO2					Heat Rate-net(Btu/kWh):	8,851
					MWe-net:	550
					Capacity Factor: (%):	85
OPERATING & MAINTENANCE LABOR						
Operating Labor						
Operating Labor Rate(base):	34.65	$/hour				
Operating Labor Burden:	30.00	% of base				
Labor O-H Charge Rate:	25.00	% of labor				
			Total			
Skilled Operator	2.0		2.0			
Operator	9.0		9.0			
Foreman	1.0		1.0			
Lab Tech's, etc.	2.0		2.0			
TOTAL-O.J.'s	14.0		14.0			
					Annual Cost $	Annual Unit Cost $/kW-net
Annual Operating Labor Cost	Maintenance labor cost		% of BEC	0.8723	$5,524,319	$10.044
Maintenance Labor Cost	(Case S12A is reference)		BEC	$843,871	$7,361,412	$13.384
Administrative & Support Labor					$3,221,433	$5.857
Property Taxes & Insurance					$20,726,893	$37.685
TOTAL FIXED OPERATING COSTS					**$36,834,056**	**$66.970**
VARIABLE OPERATING COSTS						
						$/kWh-net
Maintenance Material Cost			% of BEC	1.3085	**$11,042,118**	**$0.00270**
Consumables	Consumption Initial Fill	/Day	Unit Cost	Initial Fill Cost		
Water(/1000 gallons)	0	1,566	1.08	**$0**	**$525,534**	**$0.00013**
Chemicals		4.841				
MU & WT Chem.(lb)	0	7,580	0.17	$0	$407,027	$0.00010
Lime (ton)	0	104	75.00	$0	$2,416,134	$0.00059
Carbon (Mercury Removal) (lb)	0	2,588	1.05	$0	$843,210	$0.00021
MEA Solvent (ton)	0	0	2,249.89	$0	$0	$0.00000
NaOH (tons)	0	0	433.68	$0	$0	$0.00000
H2SO4 (tons)	0	0	138.78	$0	$0	$0.00000
Corrosion Inhibitor	0	0	0.00	$0	$0	$0.00000
Activated Carbon(lb)	0	0	1.05	$0	$0	$0.00000
Ammonia (28% NH3) ton	0	23	129.80	$0	$912,133	$0.00022
Subtotal Chemicals				**$0**	**$4,578,504**	**$0.00112**
Other						
Supplemental Fuel(MBtu)	0	0	0.00	$0	$0	$0.00000
SCR Catalyst(m3)	w/equip.	0.346	5,775.94	$0	$620,518	$0.00015
Emission Penalties	0	0	0.00	$0	$0	$0.00000
Subtotal Other				**$0**	**$620,518**	**$0.00015**
Waste Disposal						
Flyash (ton)	0	657	16.23	$0	$3,307,580	$0.00081
Bottom Ash(ton)	0	112	16.23	$0	$562,230	$0.00014
Subtotal-Waste Disposal				**$0**	**$3,869,811**	**$0.00094**
By-products & Emissions						
Gypsum (tons)	0	0	0.00	$0	$0	$0.00000
Subtotal By-Products				**$0**	**$0**	**$0.00000**
TOTAL VARIABLE OPERATING COSTS				**$0**	**$20,636,484**	**$0.00504**
Fuel(ton)	0	6,821	10.37	**$0**	**$21,938,585**	**$0.00536**

6.1.8 Case 5 – Cost Estimating

Exhibit 6-23 shows the total plant capital cost details organized by cost account. Exhibit 6-24 shows the initial and annual O&M costs.

The estimated TOC of the supercritical PC case with a CO_2 emission rate of 1,100 lb CO_2/net-MWh is \$3,323/kW. Owner's costs represent 18 percent of the TOC. The current dollar, 30-year LCOE, including TS&M, is \$120.01/MWh.

Exhibit 6-23 Case 5 Total Plant Cost Details

Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		Case 5 TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$	$/kW
1	COAL & SORBENT HANDLING											
1.1	Coal Receive & Unload	$4,662	$0	$2,130	$0	$0	$6,792	$607	$0	$1,110	$8,508	$15
1.2	Coal Stackout & Reclaim	$6,025	$0	$1,365	$0	$0	$7,390	$647	$0	$1,206	$9,243	$17
1.3	Coal Conveyors	$5,602	$0	$1,351	$0	$0	$6,953	$609	$0	$1,134	$8,696	$16
1.4	Other Coal Handling	$1,466	$0	$313	$0	$0	$1,778	$155	$0	$290	$2,224	$4
1.5	Sorbent Receive & Unload	$56	$0	$17	$0	$0	$73	$6	$0	$12	$92	$0
1.6	Sorbent Stackout & Reclaim	$910	$0	$167	$0	$0	$1,076	$94	$0	$176	$1,346	$2
1.7	Sorbent Conveyors	$325	$70	$80	$0	$0	$474	$41	$0	$77	$593	$1
1.8	Other Sorbent Handling	$196	$46	$103	$0	$0	$345	$30	$0	$56	$432	$1
1.9	Coal & Sorbent Hnd.Foundations	$0	$5,726	$7,223	$0	$0	$12,949	$1,216	$0	$2,125	$16,290	$30
	SUBTOTAL 1.	**$19,242**	**$5,842**	**$12,747**	**$0**	**$0**	**$37,831**	**$3,406**	**$0**	**$6,186**	**$47,422**	**$86**
2	COAL & SORBENT PREP & FEED											
2.1	Coal Crushing & Drying	$2,727	$0	$531	$0	$0	$3,258	$284	$0	$531	$4,074	$7
2.2	Coal Conveyor to Storage	$6,982	$0	$1,524	$0	$0	$8,506	$744	$0	$1,387	$10,637	$19
2.3	Coal Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.4	Misc.Coal Prep & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.5	Sorbent Prep Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.6	Sorbent Storage & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.7	Sorbent Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.8	Booster Air Supply System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.9	Coal & Sorbent Feed Foundation	$0	$780	$655	$0	$0	$1,434	$133	$0	$235	$1,802	$3
	SUBTOTAL 2.	**$9,709**	**$780**	**$2,710**	**$0**	**$0**	**$13,199**	**$1,161**	**$0**	**$2,154**	**$16,513**	**$30**
3	FEEDWATER & MISC. BOP SYSTEMS											
3.1	FeedwaterSystem	$20,866	$0	$6,740	$0	$0	$27,606	$2,412	$0	$4,503	$34,522	$63
3.2	Water Makeup & Pretreating	$3,672	$0	$1,182	$0	$0	$4,854	$459	$0	$1,063	$6,375	$12
3.3	Other Feedwater Subsystems	$6,388	$0	$2,700	$0	$0	$9,088	$814	$0	$1,485	$11,387	$21
3.4	Service Water Systems	$720	$0	$392	$0	$0	$1,111	$105	$0	$243	$1,459	$3
3.5	Other Boiler Plant Systems	$8,635	$0	$8,525	$0	$0	$17,160	$1,630	$0	$2,819	$21,609	$39
3.6	FO Supply Sys & Nat Gas	$268	$0	$335	$0	$0	$603	$57	$0	$99	$759	$1
3.7	Waste Treatment Equipment	$2,489	$0	$1,419	$0	$0	$3,909	$380	$0	$858	$5,147	$9
3.8	Misc. Equip.(cranes,AirComp.,Comm.)	$2,847	$0	$870	$0	$0	$3,716	$357	$0	$815	$4,889	$9
	SUBTOTAL 3.	**$45,885**	**$0**	**$22,163**	**$0**	**$0**	**$68,048**	**$6,215**	**$0**	**$11,884**	**$86,146**	**$157**
4	PC BOILER											
4.1	PC Boiler & Accessories	$213,607	$0	$104,237	$0	$0	$317,844	$30,901	$0	$34,874	$383,619	$697
4.2	SCR (w/4.1)	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.3	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.4	Boiler BoP (w/ ID Fans)	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.5	Primary Air System	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.6	Secondary Air System	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.8	Major Component Rigging	$0	w/4.1	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.9	Boiler Foundations	$0	w/14.1	w/14.1	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 4.	**$213,607**	**$0**	**$104,237**	**$0**	**$0**	**$317,844**	**$30,901**	**$0**	**$34,874**	**$383,619**	**$697**

Exhibit 6-23 Case 5 Total Plant Cost Details (Continued)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		Case 5 TOTAL PLANT COST	
				Direct	Indirect				Process	Project	$	$/kW
5	FLUE GAS CLEANUP											
5.1	Absorber Vessels & Accessories	$94,056	$0	$15,822	$0	$0	$109,878	$10,458	$0	$12,034	$132,369	$241
5.2	Other FGD	$1,120	$0	$448	$0	$0	$1,569	$150	$0	$172	$1,891	$3
5.3	Bag House & Accessories	w/5.1	$0	w/5.1	$0	$0	$0	$0	$0	$0	$0	$0
5.4	Other Particulate Removal Materials	$23,740	$0	$10,002	$0	$0	$33,742	$3,236	$0	$3,698	$40,676	$74
5.5	Gypsum Dewatering System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.6	Mercury Removal System	w/5.1	$0	w/5.1	$0	$0	$0	$0	$0	$0	$0	$0
5.9	Open											
	SUBTOTAL 5.	**$118,917**	**$0**	**$26,272**	**$0**	**$0**	**$145,188**	**$13,845**	**$0**	**$15,903**	**$174,936**	**$318**
5B	CO_2 REMOVAL & COMPRESSION											
5B.1	CO_2 Removal System	$141,982	$0	$43,078	$0	$0	$185,060	$17,693	$37,012	$47,953	$287,719	$523
5B.2	CO_2 Compression & Drying	$19,150	$0	$6,008	$0	$0	$25,157	$2,406	$0	$5,513	$33,076	$60
5B.3	CO_2 Removal System Let Down Turbine	$9,400	$0	$1,248	$0	$0	$10,648	$1,021	$0	$1,167	$12,836	$23
	SUBTOTAL 5B.	**$170,532**	**$0**	**$50,334**	**$0**	**$0**	**$220,866**	**$21,120**	**$37,012**	**$54,633**	**$333,631**	**$607**
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Combustion Turbine Generator	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
6.2	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.3	Compressed Air Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.9	Combustion Turbine Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 6.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
7.2	HRSG Accessories	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.3	Ductwork	$10,553	$0	$6,780	$0	$0	$17,334	$1,511	$0	$2,827	$21,672	$39
7.4	Stack	$11,124	$0	$6,509	$0	$0	$17,633	$1,698	$0	$1,933	$21,264	$39
7.9	Duct & Stack Foundations	$0	$1,295	$1,471	$0	$0	$2,766	$259	$0	$605	$3,630	$7
	SUBTOTAL 7.	**$21,677**	**$1,295**	**$14,761**	**$0**	**$0**	**$37,733**	**$3,468**	**$0**	**$5,365**	**$46,566**	**$85**
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	$54,735	$0	$7,269	$0	$0	$62,004	$5,942	$0	$6,795	$74,741	$136
8.2	Turbine Plant Auxiliaries	$369	$0	$789	$0	$0	$1,158	$113	$0	$127	$1,398	$3
8.3a	Condenser & Auxiliaries	$3,734	$0	$2,253	$0	$0	$5,987	$577	$0	$656	$7,220	$13
8.3b	Air Cooled Condenser	$33,792	$0	$6,775	$0	$0	$40,567	$4,057	$0	$8,925	$53,548	$97
8.4	Steam Piping	$23,522	$0	$11,598	$0	$0	$35,120	$2,951	$0	$5,711	$43,781	$80
8.9	TG Foundations	$0	$1,158	$1,830	$0	$0	$2,988	$283	$0	$654	$3,925	$7
	SUBTOTAL 8.	**$116,152**	**$1,158**	**$30,514**	**$0**	**$0**	**$147,824**	**$13,922**	**$0**	**$22,868**	**$184,614**	**$336**

Exhibit 6-23 Case 5 Total Plant Cost Details (Continued)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		Case 5 TOTAL PLANT COST	
				Direct	Indirect				Process	Project	$	$/kW
9	COOLING WATER SYSTEM											
9.1	Cooling Towers	$9,940	$0	$3,095	$0	$0	$13,035	$1,247	$0	$1,428	$15,710	$29
9.2	Circulating Water Pumps	$2,125	$0	$203	$0	$0	$2,328	$197	$0	$253	$2,778	$5
9.3	Circ.Water System Auxiliaries	$582	$0	$78	$0	$0	$660	$63	$0	$72	$795	$1
9.4	Circ.Water Piping	$0	$4,614	$4,472	$0	$0	$9,086	$851	$0	$1,490	$11,427	$21
9.5	Make-up Water System	$415	$0	$554	$0	$0	$969	$93	$0	$159	$1,221	$2
9.6	Component Cooling Water Sys	$461	$0	$366	$0	$0	$827	$79	$0	$136	$1,041	$2
9.9	Circ.Water System Foundations& Structures	$0	$2,739	$4,352	$0	$0	$7,091	$671	$0	$1,552	$9,314	$17
	SUBTOTAL 9.	**$13,523**	**$7,353**	**$13,120**	**$0**	**$0**	**$33,996**	**$3,199**	**$0**	**$5,091**	**$42,286**	**$77**
10	ASH/SPENT SORBENT HANDLING SYS											
10.1	Ash Coolers	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.2	Cyclone Ash Letdown	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.3	HGCU Ash Letdown	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.4	High Temperature Ash Piping	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.5	Other Ash Recovery Equipment	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.6	Ash Storage Silos	$857	$0	$2,642	$0	$0	$3,499	$343	$0	$384	$4,227	$8
10.7	Ash Transport & Feed Equipment	$5,550	$0	$5,685	$0	$0	$11,235	$1,074	$0	$1,231	$13,540	$25
10.8	Misc. Ash Handling Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.9	Ash/Spent Sorbent Foundation	$0	$204	$240	$0	$0	$444	$42	$0	$97	$582	$1
	SUBTOTAL 10.	**$6,407**	**$204**	**$8,566**	**$0**	**$0**	**$15,177**	**$1,459**	**$0**	**$1,712**	**$18,349**	**$33**
11	ACCESSORY ELECTRIC PLANT											
11.1	Generator Equipment	$1,674	$0	$272	$0	$0	$1,946	$180	$0	$159	$2,286	$4
11.2	Station Service Equipment	$4,277	$0	$1,405	$0	$0	$5,683	$531	$0	$466	$6,680	$12
11.3	Switchgear & Motor Control	$4,917	$0	$836	$0	$0	$5,753	$533	$0	$629	$6,915	$13
11.4	Conduit & Cable Tray	$0	$3,083	$10,660	$0	$0	$13,743	$1,331	$0	$2,261	$17,335	$32
11.5	Wire & Cable	$0	$5,817	$11,230	$0	$0	$17,048	$1,436	$0	$2,773	$21,256	$39
11.6	Protective Equipment	$270	$0	$920	$0	$0	$1,190	$116	$0	$131	$1,437	$3
11.7	Standby Equipment	$1,326	$0	$30	$0	$0	$1,356	$124	$0	$148	$1,629	$3
11.8	Main Power Transformers	$6,222	$0	$194	$0	$0	$6,416	$489	$0	$690	$7,595	$14
11.9	Electrical Foundations	$0	$329	$807	$0	$0	$1,136	$109	$0	$249	$1,493	$3
	SUBTOTAL 11.	**$18,688**	**$9,229**	**$26,354**	**$0**	**$0**	**$54,271**	**$4,849**	**$0**	**$7,506**	**$66,626**	**$121**
12	INSTRUMENTATION & CONTROL											
12.1	PC Control Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.2	Combustion Turbine Control	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
12.3	Steam Turbine Control	w/8.1	$0	w/8.1	$0	$0	$0	$0	$0	$0	$0	$0
12.4	Other Major Component Control	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
12.5	Signal Processing Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.6	Control Boards,Panels & Racks	$485	$0	$291	$0	$0	$776	$73	$39	$133	$1,021	$2
12.7	Distributed Control System Equipment	$4,898	$0	$856	$0	$0	$5,753	$533	$288	$657	$7,232	$13
12.8	Instrument Wiring & Tubing	$2,655	$0	$5,266	$0	$0	$7,921	$675	$396	$1,349	$10,341	$19
12.9	Other I & C Equipment	$1,384	$0	$3,141	$0	$0	$4,525	$439	$226	$519	$5,709	$10
	SUBTOTAL 12.	**$9,421**	**$0**	**$9,553**	**$0**	**$0**	**$18,975**	**$1,720**	**$949**	**$2,658**	**$24,302**	**$44**

Exhibit 6-23 Case 5 Total Plant Cost Details (Continued)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		Case 5 TOTAL PLANT COST	
				Direct	Indirect				Process	Project	$	$/kW
13	IMPROVEMENTS TO SITE											
13.1	Site Preparation	$0	$54	$1,086	$0	$0	$1,141	$113	$0	$251	$1,505	$3
13.2	Site Improvements	$0	$1,803	$2,240	$0	$0	$4,043	$399	$0	$888	$5,330	$10
13.3	Site Facilities	$3,232	$0	$3,187	$0	$0	$6,419	$633	$0	$1,410	$8,462	$15
	SUBTOTAL 13.	**$3,232**	**$1,858**	**$6,513**	**$0**	**$0**	**$11,603**	**$1,145**	**$0**	**$2,549**	**$15,297**	**$28**
14	BUILDINGS & STRUCTURES											
14.1	Boiler Building	$0	$9,603	$8,445	$0	$0	$18,048	$1,622	$0	$2,951	$22,621	$41
14.2	Turbine Building	$0	$12,599	$11,742	$0	$0	$24,341	$2,194	$0	$3,980	$30,515	$55
14.3	Administration Building	$0	$637	$673	$0	$0	$1,310	$119	$0	$214	$1,644	$3
14.4	Circulation Water Pumphouse	$0	$233	$185	$0	$0	$418	$37	$0	$68	$523	$1
14.5	Water Treatment Buildings	$0	$481	$396	$0	$0	$877	$79	$0	$143	$1,099	$2
14.6	Machine Shop	$0	$426	$286	$0	$0	$712	$63	$0	$116	$892	$2
14.7	Warehouse	$0	$289	$290	$0	$0	$578	$52	$0	$95	$725	$1
14.8	Other Buildings & Structures	$0	$236	$201	$0	$0	$437	$39	$0	$71	$547	$1
14.9	Waste Treating Building & Str.	$0	$431	$1,307	$0	$0	$1,738	$165	$0	$285	$2,188	$4
	SUBTOTAL 14.	**$0**	**$24,934**	**$23,525**	**$0**	**$0**	**$48,459**	**$4,371**	**$0**	**$7,924**	**$60,754**	**$110**
	TOTAL COST	**$766,991**	**$52,653**	**$351,368**	**$0**	**$0**	**$1,171,013**	**$110,780**	**$37,961**	**$181,307**	**$1,501,061**	**$2,729**
	Owner's Costs											
	Preproduction Costs											
	6 Months All Labor										$10,480	$19
	1 Month Maintenance Materials										$1,518	$3
	1 Month Non-fuel Consumables										$911	$2
	1 Month Waste Disposal										$461	$1
	25% of 1 Months Fuel Cost at 100% CF										$655	$1
	2% of TPC										$30,021	$55
	Total										**$44,045**	**$80**
	Inventory Capital											
	60 day supply of fuel and consumables at 100% CF										$7,058	$13
	0.5% of TPC (spare parts)										$7,505	$14
	Total										**$14,563**	**$26**
	Initial Cost for Catalyst and Chemicals										$1,354	$2
	Land										$900	$2
	Other Owner's Costs										$225,159	$409
	Financing Costs										$40,529	$74
	Total Overnight Costs (TOC)										**$1,827,611**	**$3,323**
	TASC Multiplier								(IOU, high risk, 35 year)		1.140	
	Total As-Spent Cost (TASC)										**$2,083,477**	**$3,788**

Exhibit 6-24 Case 5 Initial and Annual Operating and Maintenance Costs

INITIAL & ANNUAL O&M EXPENSES					Cost Base (June)	2007
Case 5 - Supercritical PC w/ CO2 capture (1,100 lb/net MWh)					Heat Rate-net(Btu/kWh):	10,774
					MWe-net:	550
					Capacity Factor: (%):	85
	OPERATING & MAINTENANCE LABOR					
Operating Labor						
Operating Labor Rate(base):	34.65	$/hour				
Operating Labor Burden:	30.00	% of base				
Labor O-H Charge Rate:	25.00	% of labor				
			Total			
Skilled Operator	2.0		2.0			
Operator	11.3		11.3			
Foreman	1.0		1.0			
Lab Tech's, etc.	2.0		2.0			
TOTAL-O.J.'s	16.3		16.3			
					Annual Cost	Annual Unit Cost
					$	$/kW-net
Annual Operating Labor Cost	Maintenance labor cost		% of BEC	0.8815	$6,444,907	$11.717
Maintenance Labor Cost	(Case S12B is reference)		BEC	$1,171,013	$10,322,636	$18.767
Administrative & Support Labor					$4,191,886	$7.621
Property Taxes & Insurance					$30,021,222	$54.583
TOTAL FIXED OPERATING COSTS					**$50,980,650**	**$92.689**
VARIABLE OPERATING COSTS						
						$/kWh-net
Maintenance Material Cost			% of BEC	1.3223	**$15,483,953**	**$0.00378**
Consumables	Consumption		Unit	Initial Fill		
	Initial Fill	/Day	Cost	Cost		
Water(/1000 gallons)	0	2,825	1.08	**$0**	**$948,135**	**$0.00023**
Chemicals		4.841				
MU & WT Chem.(lb)	0	13,676	0.17	$0	$734,333	$0.00018
Lime (ton)	0	125	75.00	$0	$2,897,518	$0.00071
Carbon (Mercury Removal) (lb)	0	3,153	1.05	$0	$1,027,296	$0.00025
MEA Solvent (ton)	567	0.80	2,249.89	$1,275,685	$560,934	$0.00014
NaOH (tons)	0	5.74	433.68	$0	$772,821	$0.00019
H2SO4 (tons)	0	3.82	138.78	$0	$164,508	$0.00004
Corrosion Inhibitor	0	0	0.00	$78,395	$3,733	$0.00000
Activated Carbon(lb)	0	960	1.05	$0	$312,780	$0.00008
Ammonia (28% NH3) ton	0	28	129.80	$0	$1,110,265	$0.00027
Subtotal Chemicals				**$1,354,079**	**$7,584,188**	**$0.00185**
Other						
Supplemental Fuel(MBtu)	0	0	0.00	$0	$0	$0.00000
SCR Catalyst(m3)	w/equip.	0.422	5,775.94	$0	$755,949	$0.00018
Emission Penalties	0	0	0.00	$0	$0	$0.00000
Subtotal Other				**$0**	**$755,949**	**$0.00018**
Waste Disposal						
Flyash (ton)	0	798	16.23	$0	$4,017,011	$0.00098
Bottom Ash(ton)	0	136	16.23	$0	$684,452	$0.00017
Subtotal-Waste Disposal				**$0**	**$4,701,463**	**$0.00115**
By-products & Emissions						
Gypsum (tons)	0	0	0.00	$0	$0	$0.00000
Subtotal By-Products				**$0**	**$0**	**$0.00000**
TOTAL VARIABLE OPERATING COSTS				**$1,354,079**	**$29,473,688**	**$0.00720**
Fuel(ton)	0	8,303	10.37	**$0**	**$26,706,931**	**$0.00652**

6.1.9 Case 6 – Cost Estimating

Exhibit 6-25 shows the total plant capital cost details organized by cost account. Exhibit 6-26 shows the initial and annual O&M costs.

The estimated TOC of the supercritical PC case with 90 percent carbon capture is $3,969/kW. Owner’s costs represent 18 percent of the TOC. The current dollar, 30-year LCOE, including TS&M costs, is $143.89/MWh.

Exhibit 6-25 Case 6 Total Plant Cost Details

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		Case 6 TOTAL PLANT COST	
				Direct	Indirect				Process	Project	$	$/kW
1	COAL & SORBENT HANDLING											
1.1	Coal Receive & Unload	$5,157	$0	$2,355	$0	$0	$7,512	$671	$0	$1,228	$9,411	$17
1.2	Coal Stackout & Reclaim	$6,665	$0	$1,510	$0	$0	$8,175	$715	$0	$1,333	$10,223	$19
1.3	Coal Conveyors	$6,196	$0	$1,494	$0	$0	$7,690	$674	$0	$1,255	$9,619	$17
1.4	Other Coal Handling	$1,621	$0	$346	$0	$0	$1,967	$172	$0	$321	$2,459	$4
1.5	Sorbent Receive & Unload	$62	$0	$19	$0	$0	$81	$7	$0	$13	$102	$0
1.6	Sorbent Stackout & Reclaim	$1,009	$0	$185	$0	$0	$1,194	$104	$0	$195	$1,492	$3
1.7	Sorbent Conveyors	$360	$78	$88	$0	$0	$526	$45	$0	$86	$657	$1
1.8	Other Sorbent Handling	$217	$51	$114	$0	$0	$382	$34	$0	$62	$479	$1
1.9	Coal & Sorbent Hnd.Foundations	$0	$6,334	$7,990	$0	$0	$14,323	$1,345	$0	$2,350	$18,019	$33
	SUBTOTAL 1.	**$21,288**	**$6,462**	**$14,101**	**$0**	**$0**	**$41,851**	**$3,768**	**$0**	**$6,843**	**$52,462**	**$95**
2	COAL & SORBENT PREP & FEED											
2.1	Coal Crushing & Drying	$3,036	$0	$592	$0	$0	$3,628	$316	$0	$592	$4,535	$8
2.2	Coal Conveyor to Storage	$7,773	$0	$1,697	$0	$0	$9,470	$828	$0	$1,545	$11,842	$22
2.3	Coal Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.4	Misc.Coal Prep & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.5	Sorbent Prep Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.6	Sorbent Storage & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.7	Sorbent Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.8	Booster Air Supply System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.9	Coal & Sorbent Feed Foundation	$0	$868	$729	$0	$0	$1,597	$148	$0	$262	$2,007	$4
	SUBTOTAL 2.	**$10,809**	**$868**	**$3,017**	**$0**	**$0**	**$14,694**	**$1,292**	**$0**	**$2,398**	**$18,384**	**$33**
3	FEEDWATER & MISC. BOP SYSTEMS											
3.1	FeedwaterSystem	$23,208	$0	$7,497	$0	$0	$30,705	$2,683	$0	$5,008	$38,396	$70
3.2	Water Makeup & Pretreating	$5,615	$0	$1,807	$0	$0	$7,422	$702	$0	$1,625	$9,749	$18
3.3	Other Feedwater Subsystems	$7,105	$0	$3,003	$0	$0	$10,108	$905	$0	$1,652	$12,665	$23
3.4	Service Water Systems	$1,101	$0	$599	$0	$0	$1,700	$160	$0	$372	$2,231	$4
3.5	Other Boiler Plant Systems	$9,826	$0	$9,701	$0	$0	$19,527	$1,855	$0	$3,207	$24,589	$45
3.6	FO Supply Sys & Nat Gas	$279	$0	$348	$0	$0	$627	$59	$0	$103	$789	$1
3.7	Waste Treatment Equipment	$3,807	$0	$2,170	$0	$0	$5,977	$582	$0	$1,312	$7,870	$14
3.8	Misc. Equip.(cranes,AirComp.,Comm.)	$2,958	$0	$904	$0	$0	$3,862	$371	$0	$847	$5,080	$9
	SUBTOTAL 3.	**$53,898**	**$0**	**$26,029**	**$0**	**$0**	**$79,927**	**$7,317**	**$0**	**$14,125**	**$101,370**	**$184**
4	PC BOILER											
4.1	PC Boiler & Accessories	$239,129	$0	$116,692	$0	$0	$355,821	$34,593	$0	$39,041	$429,456	$781
4.2	SCR (w/4.1)	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.3	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.4	Boiler BoP (w/ ID Fans)	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.5	Primary Air System	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.6	Secondary Air System	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.8	Major Component Rigging	$0	w/4.1	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.9	Boiler Foundations	$0	w/14.1	w/14.1	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 4.	**$239,129**	**$0**	**$116,692**	**$0**	**$0**	**$355,821**	**$34,593**	**$0**	**$39,041**	**$429,456**	**$781**

Exhibit 6-25 Case 6 Total Plant Cost Details (Continued)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		Case 6 TOTAL PLANT COST	
				Direct	Indirect				Process	Project	$	$/kW
5	FLUE GAS CLEANUP											
5.1	Absorber Vessels & Accessories	$102,672	$0	$17,271	$0	$0	$119,943	$11,416	$0	$13,136	$144,495	$263
5.2	Other FGD	$1,214	$0	$486	$0	$0	$1,700	$163	$0	$186	$2,049	$4
5.3	Bag House & Accessories	w/5.1	$0	w/5.1	$0	$0	$0	$0	$0	$0	$0	$0
5.4	Other Particulate Removal Materials	$25,300	$0	$10,659	$0	$0	$35,959	$3,449	$0	$3,941	$43,349	$79
5.5	Gypsum Dewatering System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.6	Mercury Removal System	w/5.1	$0	w/5.1	$0	$0	$0	$0	$0	$0	$0	$0
5.9	Open											
	SUBTOTAL 5.	**$129,186**	**$0**	**$28,416**	**$0**	**$0**	**$157,602**	**$15,028**	**$0**	**$17,263**	**$189,893**	**$345**
5B	CO$_2$ REMOVAL & COMPRESSION											
5B.1	CO$_2$ Removal System	$217,227	$0	$65,907	$0	$0	$283,134	$27,070	$56,627	$73,366	$440,197	$801
5B.2	CO$_2$ Compression & Drying	$29,198	$0	$9,160	$0	$0	$38,358	$3,669	$0	$8,405	$50,432	$92
5B.3	CO$_2$ Removal System Let Down Turbine	$10,400	$0	$1,381	$0	$0	$11,781	$1,129	$0	$1,291	$14,201	$26
	SUBTOTAL 5B.	**$256,825**	**$0**	**$76,448**	**$0**	**$0**	**$333,273**	**$31,867**	**$56,627**	**$83,063**	**$504,830**	**$918**
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Combustion Turbine Generator	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
6.2	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.3	Compressed Air Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.9	Combustion Turbine Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 6.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
7.2	HRSG Accessories	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.3	Ductwork	$10,558	$0	$6,783	$0	$0	$17,342	$1,512	$0	$2,828	$21,682	$39
7.4	Stack	$9,475	$0	$5,544	$0	$0	$15,019	$1,446	$0	$1,646	$18,111	$33
7.9	Duct & Stack Foundations	$0	$1,087	$1,235	$0	$0	$2,323	$217	$0	$508	$3,048	$6
	SUBTOTAL 7.	**$20,033**	**$1,087**	**$13,563**	**$0**	**$0**	**$34,683**	**$3,175**	**$0**	**$4,983**	**$42,841**	**$78**
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	$57,457	$0	$7,631	$0	$0	$65,087	$6,238	$0	$7,133	$78,458	$143
8.2	Turbine Plant Auxiliaries	$388	$0	$830	$0	$0	$1,218	$119	$0	$134	$1,471	$3
8.3a	Condenser & Auxiliaries	$3,425	$0	$2,066	$0	$0	$5,491	$529	$0	$602	$6,622	$12
8.3b	Air Cooled Condenser	$29,987	$0	$6,012	$0	$0	$36,000	$3,600	$0	$7,920	$47,519	$86
8.4	Steam Piping	$26,668	$0	$13,149	$0	$0	$39,817	$3,345	$0	$6,474	$49,636	$90
8.9	TG Foundations	$0	$1,216	$1,922	$0	$0	$3,138	$297	$0	$687	$4,122	$7
	SUBTOTAL 8.	**$117,925**	**$1,216**	**$31,610**	**$0**	**$0**	**$150,751**	**$14,128**	**$0**	**$22,949**	**$187,828**	**$342**

Exhibit 6-25 Case 6 Total Plant Cost Details (Continued)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		Case 6 TOTAL PLANT COST	
				Direct	Indirect				Process	Project	$	$/kW
9	COOLING WATER SYSTEM											
9.1	Cooling Towers	$15,102	$0	$4,703	$0	$0	$19,805	$1,894	$0	$2,170	$23,869	$43
9.2	Circulating Water Pumps	$3,149	$0	$300	$0	$0	$3,449	$292	$0	$374	$4,115	$7
9.3	Circ.Water System Auxiliaries	$834	$0	$111	$0	$0	$945	$90	$0	$104	$1,139	$2
9.4	Circ.Water Piping	$0	$6,612	$6,408	$0	$0	$13,021	$1,219	$0	$2,136	$16,375	$30
9.5	Make-up Water System	$596	$0	$796	$0	$0	$1,392	$133	$0	$229	$1,754	$3
9.6	Component Cooling Water Sys	$660	$0	$525	$0	$0	$1,184	$112	$0	$195	$1,492	$3
9.9	Circ.Water System Foundations& Structures	$0	$3,917	$6,224	$0	$0	$10,141	$959	$0	$2,220	$13,321	$24
	SUBTOTAL 9.	**$20,341**	**$10,530**	**$19,067**	**$0**	**$0**	**$49,938**	**$4,700**	**$0**	**$7,427**	**$62,065**	**$113**
10	ASH/SPENT SORBENT HANDLING SYS											
10.1	Ash Coolers	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.2	Cyclone Ash Letdown	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.3	HGCU Ash Letdown	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.4	High Temperature Ash Piping	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.5	Other Ash Recovery Equipment	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.6	Ash Storage Silos	$938	$0	$2,891	$0	$0	$3,829	$376	$0	$421	$4,626	$8
10.7	Ash Transport & Feed Equipment	$6,074	$0	$6,222	$0	$0	$12,295	$1,176	$0	$1,347	$14,818	$27
10.8	Misc. Ash Handling Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.9	Ash/Spent Sorbent Foundation	$0	$223	$263	$0	$0	$486	$46	$0	$106	$637	$1
	SUBTOTAL 10.	**$7,012**	**$223**	**$9,375**	**$0**	**$0**	**$16,610**	**$1,597**	**$0**	**$1,874**	**$20,081**	**$37**
11	ACCESSORY ELECTRIC PLANT											
11.1	Generator Equipment	$1,746	$0	$284	$0	$0	$2,030	$188	$0	$166	$2,384	$4
11.2	Station Service Equipment	$5,204	$0	$1,710	$0	$0	$6,914	$646	$0	$567	$8,127	$15
11.3	Switchgear & Motor Control	$5,983	$0	$1,017	$0	$0	$7,000	$649	$0	$765	$8,414	$15
11.4	Conduit & Cable Tray	$0	$3,751	$12,970	$0	$0	$16,721	$1,619	$0	$2,751	$21,091	$38
11.5	Wire & Cable	$0	$7,078	$13,664	$0	$0	$20,742	$1,747	$0	$3,373	$25,863	$47
11.6	Protective Equipment	$270	$0	$920	$0	$0	$1,190	$116	$0	$131	$1,437	$3
11.7	Standby Equipment	$1,372	$0	$31	$0	$0	$1,403	$129	$0	$153	$1,685	$3
11.8	Main Power Transformers	$6,266	$0	$195	$0	$0	$6,461	$492	$0	$695	$7,649	$14
11.9	Electrical Foundations	$0	$346	$848	$0	$0	$1,194	$114	$0	$262	$1,570	$3
	SUBTOTAL 11.	**$20,842**	**$11,175**	**$31,638**	**$0**	**$0**	**$63,656**	**$5,701**	**$0**	**$8,863**	**$78,220**	**$142**
12	INSTRUMENTATION & CONTROL											
12.1	PC Control Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.2	Combustion Turbine Control	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
12.3	Steam Turbine Control	w/8.1	$0	w/8.1	$0	$0	$0	$0	$0	$0	$0	$0
12.4	Other Major Component Control	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
12.5	Signal Processing Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.6	Control Boards,Panels & Racks	$516	$0	$309	$0	$0	$825	$78	$41	$142	$1,086	$2
12.7	Distributed Control System Equipment	$5,208	$0	$910	$0	$0	$6,118	$567	$306	$699	$7,691	$14
12.8	Instrument Wiring & Tubing	$2,823	$0	$5,601	$0	$0	$8,424	$718	$421	$1,434	$10,997	$20
12.9	Other I & C Equipment	$1,472	$0	$3,340	$0	$0	$4,811	$467	$241	$552	$6,070	$11
	SUBTOTAL 12.	**$10,019**	**$0**	**$10,159**	**$0**	**$0**	**$20,178**	**$1,830**	**$1,009**	**$2,827**	**$25,844**	**$47**

Exhibit 6-25 Case 6 Total Plant Cost Details (Continued)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	Case 6 TOTAL PLANT COST $	Case 6 TOTAL PLANT COST $/kW
13	IMPROVEMENTS TO SITE											
13.1	Site Preparation	$0	$57	$1,130	$0	$0	$1,186	$118	$0	$261	$1,565	$3
13.2	Site Improvements	$0	$1,875	$2,329	$0	$0	$4,205	$415	$0	$924	$5,543	$10
13.3	Site Facilities	$3,361	$0	$3,314	$0	$0	$6,675	$658	$0	$1,467	$8,800	$16
	SUBTOTAL 13.	**$3,361**	**$1,932**	**$6,773**	**$0**	**$0**	**$12,066**	**$1,190**	**$0**	**$2,651**	**$15,908**	**$29**
14	BUILDINGS & STRUCTURES											
14.1	Boiler Building	$0	$9,838	$8,652	$0	$0	$18,490	$1,662	$0	$3,023	$23,174	$42
14.2	Turbine Building	$0	$12,968	$12,086	$0	$0	$25,055	$2,258	$0	$4,097	$31,410	$57
14.3	Administration Building	$0	$648	$685	$0	$0	$1,333	$121	$0	$218	$1,672	$3
14.4	Circulation Water Pumphouse	$0	$297	$236	$0	$0	$534	$48	$0	$87	$669	$1
14.5	Water Treatment Buildings	$0	$741	$611	$0	$0	$1,352	$121	$0	$221	$1,694	$3
14.6	Machine Shop	$0	$433	$291	$0	$0	$724	$64	$0	$118	$907	$2
14.7	Warehouse	$0	$294	$295	$0	$0	$588	$53	$0	$96	$738	$1
14.8	Other Buildings & Structures	$0	$240	$204	$0	$0	$444	$40	$0	$73	$557	$1
14.9	Waste Treating Building & Str.	$0	$452	$1,373	$0	$0	$1,825	$173	$0	$300	$2,298	$4
	SUBTOTAL 14.	**$0**	**$25,912**	**$24,433**	**$0**	**$0**	**$50,346**	**$4,541**	**$0**	**$8,233**	**$63,119**	**$115**
	TOTAL COST	**$910,668**	**$59,407**	**$411,322**	**$0**	**$0**	**$1,381,397**	**$130,728**	**$57,636**	**$222,541**	**$1,792,301**	**$3,259**
	Owner's Costs											
	Preproduction Costs											
	6 Months All Labor										$11,639	$21
	1 Month Maintenance Materials										$1,791	$3
	1 Month Non-fuel Consumables										$1,309	$2
	1 Month Waste Disposal										$461	$1
	25% of 1 Months Fuel Cost at 100% CF										$770	$1
	2% of TPC										$35,846	$65
	Total										**$51,816**	**$94**
	Inventory Capital											
	60 day supply of fuel and consumables at 100% CF										$8,779	$16
	0.5% of TPC (spare parts)										$8,962	$16
	Total										**$17,741**	**$32**
	Initial Cost for Catalyst and Chemicals										$2,734	$5
	Land										$900	$2
	Other Owner's Costs										$268,845	$489
	Financing Costs										$48,392	$88
	Total Overnight Costs (TOC)										**$2,182,729**	**$3,969**
	TASC Multiplier								(IOU, high risk, 35 year)		1.140	
	Total As-Spent Cost (TASC)										**$2,488,311**	**$4,525**

Exhibit 6-26 Case 6 Initial and Annual Operating and Maintenance Costs

INITIAL & ANNUAL O&M EXPENSES					Cost Base (June)	2007
Case 6 - Supercritical PC w/ 90% CO2 capture					Heat Rate-net(Btu/kWh):	12,679
					MWe-net:	550
					Capacity Factor: (%):	85
OPERATING & MAINTENANCE LABOR						
Operating Labor						
Operating Labor Rate(base):	34.65	$/hour				
Operating Labor Burden:	30.00	% of base				
Labor O-H Charge Rate:	25.00	% of labor				
			Total			
Skilled Operator	2.0		2.0			
Operator	11.3		11.3			
Foreman	1.0		1.0			
Lab Tech's, etc.	2.0		2.0			
TOTAL-O.J.'s	16.3		16.3			
					Annual Cost $	Annual Unit Cost $/kW-net
Annual Operating Labor Cost	Maintenance labor cost		% of BEC	0.8815	$6,444,907	$11.720
Maintenance Labor Cost	(Case S12B is reference)		BEC	$1,381,397	$12,177,202	$22.144
Administrative & Support Labor					$4,655,527	$8.466
Property Taxes & Insurance					$35,846,019	$65.186
TOTAL FIXED OPERATING COSTS					**$59,123,655**	**$107.517**
VARIABLE OPERATING COSTS						
						$/kWh-net
Maintenance Material Cost			% of BEC	1.3223	**$18,265,802**	**$0.00446**
Consumables	Consumption Initial	 /Day	Unit Cost	Initial Cost		
Water(/1000 gallons)	0	4,819	1.08	**$0**	**$1,617,194**	**$0.00039**
Chemicals		4.841				
MU & WT Chem.(lb)	0	23,327	0.17	$0	$1,252,521	$0.00031
Lime (ton)	0	147	75.00	$0	$3,415,201	$0.00083
Carbon (Mercury Removal) (lb)	0	3,701	1.05	$0	$1,205,947	$0.00029
MEA Solvent (ton)	1,145	1.62	2,249.89	$2,576,119	$1,133,037	$0.00028
NaOH (tons)	0	11.61	433.68	$0	$1,561,782	$0.00038
H2SO4 (tons)	0	7.72	138.78	$0	$332,276	$0.00008
Corrosion Inhibitor	0	0	0.00	$158,350	$7,540	$0.00000
Activated Carbon(lb)	0	1,939	1.05	$0	$631,756	$0.00015
Ammonia (28% NH3) ton	0	32.4	129.80	$0	$1,306,344	$0.00032
Subtotal Chemicals				**$2,734,469**	**$10,846,404**	**$0.00265**
Other						
Supplemental Fuel(MBtu)	0	0	0.00	$0	$0	$0.00000
SCR Catalyst(m3)	w/equip.	0.495	5,775.94	$0	$887,391	$0.00022
Emission Penalties	0	0	0.00	$0	$0	$0.00000
Subtotal Other				**$0**	**$887,391**	**$0.00022**
Waste Disposal						
Flyash (ton)	0	939	16.23	$0	$4,727,488	$0.00115
Bottom Ash(ton)	0	160	16.23	$0	$805,315	$0.00020
Subtotal-Waste Disposal				**$0**	**$5,532,803**	**$0.00135**
By-products & Emissions						
Gypsum (tons)	0	0	0.00	$0	$0	$0.00000
Subtotal By-Products				**$0**	**$0**	**$0.00000**
TOTAL VARIABLE OPERATING COSTS				**$2,734,469**	**$37,149,594**	**$0.00907**
Fuel(ton)	0	9,769	10.37	**$0**	**$31,422,015**	**$0.00767**

7. SUBCRITICAL PC CASES

This section contains an evaluation of plant designs for Cases 7 through 9 (which are based on typical subcritical PC plant operation), with a coal feed rate of 250,000 kg/hr (650,360 lb/hr), which is fixed due to the current size of the Unit 4 boiler. Once baseline performance parameters such as coal feed rate, net plant heat rate, net stack output and stack exit temperature were established, the coal composition was changed to Montana Rosebud PRB coal. All three cases use the same steam conditions, a single reheat 16.5 MPa/538°C/538°C (2400 psig/1,000°F/1,000°F) cycle. The more detailed modeling parameters are described in Section 7.1.3. Cases 7 through 9 are intended to represent a generic existing subcritical PC plant.

The balance of Section 7 is organized as follows:

- Process and System Description for Cases 7 - 9
- Key Assumptions for Cases 7 - 9
- Sparing Philosophy for Cases 7 - 9
- Comparison of Performance Results for Cases 7 - 9
- Equipment List for Cases 7 - 9
- Cost Estimates for Cases 7 – 9

7.1 SUBCRITICAL PC NON-CAPTURE CASE 7 AND CAPTURE CASES 8 AND 9

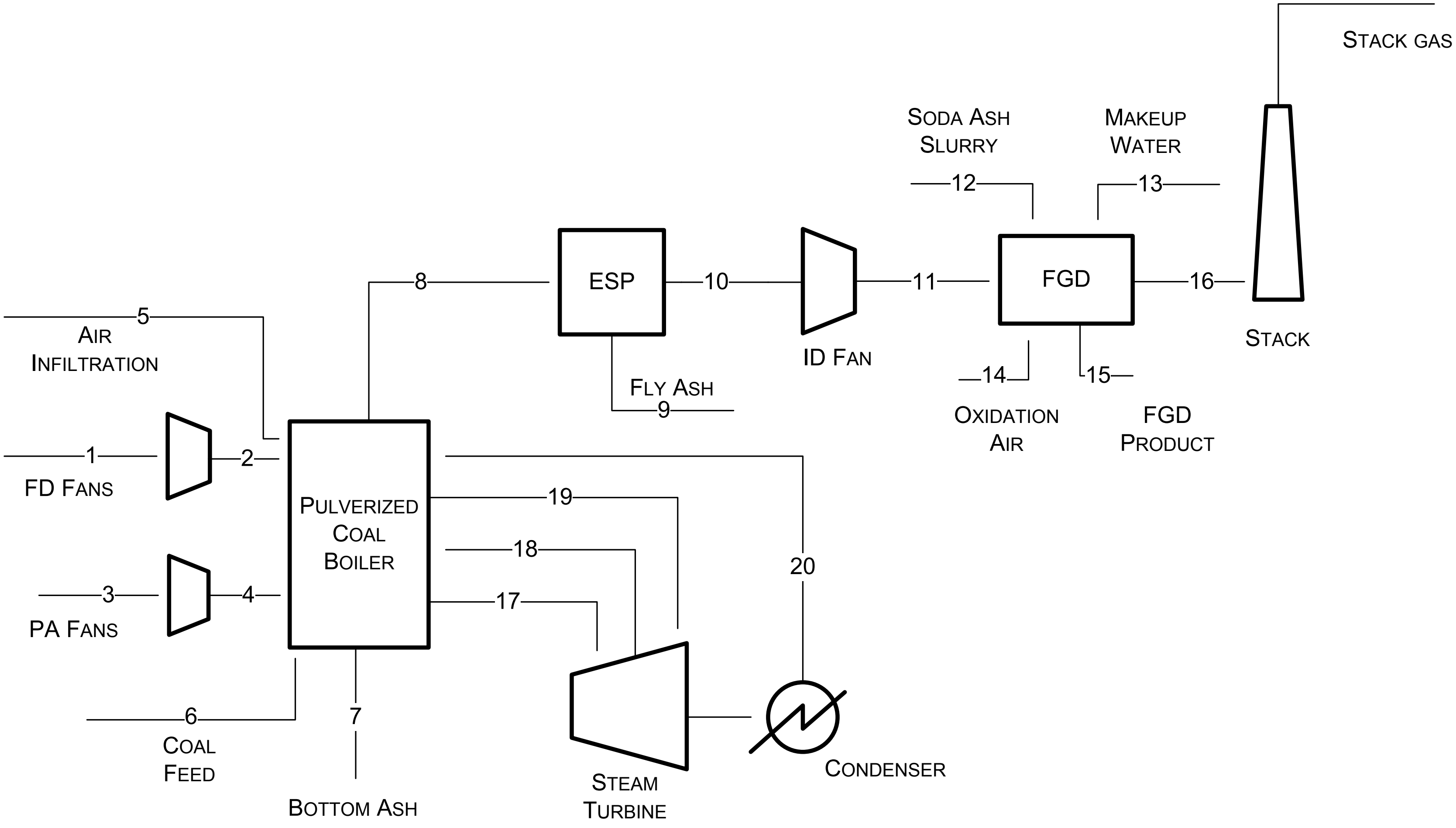
7.1.1 Case 7 Process Description

The system description is nearly identical to the supercritical PC case without CO_2 capture but is repeated here for completeness. The system description follows the block flow diagram (BFD) in Exhibit 7-1 and stream numbers reference the same Exhibit. The tables in Exhibit 7-2 provide process data for the numbered streams in the BFD.

Coal (stream 6) and primary air (stream 4) are introduced into the boiler through the tangentially fired burners. Additional combustion air, including the overfire air, is provided by the forced draft fans (stream 2). The boiler operates at a slight negative pressure so air leakage is into the boiler, and the infiltration air is accounted for in stream 5.

Flue gas exits the boiler (stream 8) and is cooled to 182°C (360°F) in the combustion air preheater (not shown) before passing through the ESP for particulate removal (stream 9). An ID fan increases the flue gas temperature to 199°C (390°F) and provides the motive force for the flue gas (stream 11) to pass through the FGD unit. FGD inputs and outputs include makeup water (stream 13), oxidation air (stream 14), soda ash slurry (stream 12) and FGD product (stream 15). The clean, saturated flue gas exiting the FGD unit (stream 16) passes to the plant stack and is discharged to atmosphere.

Exhibit 7-1 Case 7: Existing Subcritical PC - Block Flow Diagram



Note: Block Flow Diagram is not intended to represent a complete material balance. Only major process streams and equipment are shown.

Exhibit 7-2 Case 7: Existing Subcritical PC - Stream Table

	1	2	3	4	5	6	7	8	9	10
V-L Mole Fraction										
Ar	0.0093	0.0093	0.0093	0.0093	0.0093	0.0000	0.0000	0.0084	0.0000	0.0084
CO_2	0.0003	0.0003	0.0003	0.0003	0.0003	0.0000	0.0000	0.1434	0.0000	0.1434
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0071	0.0071	0.0071	0.0071	0.0071	0.0000	0.0000	0.1132	0.0000	0.1132
N_2	0.7753	0.7753	0.7753	0.7753	0.7753	0.0000	0.0000	0.7058	0.0000	0.7058
O_2	0.2080	0.2080	0.2080	0.2080	0.2080	0.0000	0.0000	0.0283	0.0000	0.0283
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0000	0.0008
Total	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	1.0000	0.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	58,751	58,751	18,048	18,048	1,322	0	0	85,921	0	85,921
V-L Flowrate (kg/hr)	1,697,155	1,697,155	521,348	521,348	38,194	0	0	2,527,543	0	2,527,543
Solids Flowrate (kg/hr)	0	0	0	0	0	294,998	4,831	19,323	19,323	0
Temperature (°C)	6	11	6	16	6	6	182	182	182	182
Pressure (MPa, abs)	0.08	0.08	0.08	0.09	0.08	0.08	0.08	0.08	0.08	0.08
Enthalpy (kJ/kg)	16.93	22.47	16.93	27.83	16.93	---	---	396.25	---	374.90
Density (kg/m^3)	1.0	1.0	1.0	1.0	1.0	---	---	0.6	---	0.6
V-L Molecular Weight	28.887	28.887	28.887	28.887	28.887	---	---	29.417	---	29.417
V-L Flowrate (lb_{mol}/hr)	129,524	129,524	39,788	39,788	2,915	0	0	189,422	0	189,422
V-L Flowrate (lb/hr)	3,741,587	3,741,587	1,149,376	1,149,376	84,204	0	0	5,572,278	0	5,572,278
Solids Flowrate (lb/hr)	0	0	0	0	0	650,360	10,650	42,599	42,599	0
Temperature (°F)	42	52	42	61	42	42	360	360	360	360
Pressure (psia)	11.4	12.0	11.4	12.6	11.4	11.4	11.2	11.2	11.0	11.0
Enthalpy (Btu/lb)	7.3	9.7	7.3	12.0	7.3	---	---	170.4	---	161.2
Density (lb/ft^3)	0.061	0.063	0.061	0.065	0.061	---	---	0.037	---	0.037
	A - Reference conditions are 32.02 F & 0.089 PSIA									

Exhibit 7-2 Case 7: Existing Subcritical PC - Stream Table (continued)

	11	12	13	14	15	16	17	18	19	20
V-L Mole Fraction										
Ar	0.0084	0.0000	0.0000	0.0093	0.0000	0.0077	0.0000	0.0000	0.0000	0.0000
CO_2	0.1434	0.0000	0.0000	0.0003	0.0000	0.1308	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.1132	1.0000	1.0000	0.0071	1.0000	0.1916	1.0000	1.0000	1.0000	1.0000
N_2	0.7058	0.0000	0.0000	0.7753	0.0000	0.6436	0.0000	0.0000	0.0000	0.0000
O_2	0.0283	0.0000	0.0000	0.2080	0.0000	0.0262	0.0000	0.0000	0.0000	0.0000
SO_2	0.0008	0.0000	0.0000	0.0000	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	85,921	1,025	9,620	369	2,237	94,670	98,815	90,709	90,709	81,920
V-L Flowrate (kg/hr)	2,527,543	18,461	173,311	10,668	40,294	2,687,638	1,780,185	1,634,149	1,634,149	1,475,804
Solids Flowrate (kg/hr)	0	8,022	0	0	10,073	0	0	0	0	0
Temperature (°C)	199	6	6	6	58	58	538	340	538	47
Pressure (MPa, abs)	0.08	0.08	0.08	0.08	0.08	0.08	16.65	4.28	3.90	2.76
Enthalpy (kJ/kg)	392.99	1,520.57	25.25	16.93	1,371.69	380.66	3,395.97	3,062.35	3,530.88	197.81
Density (kg/m^3)	0.6	1,012.1	1,012.1	1.0	961.0	0.9	50.2	16.6	10.7	990.7
V-L Molecular Weight	29.417	18.015	18.015	28.887	18.015	28.390	18.015	18.015	18.015	18.015
V-L Flowrate (lb_{mol}/hr)	189,422	2,259	21,209	814	4,931	208,711	217,850	199,979	199,979	180,602
V-L Flowrate (lb/hr)	5,572,278	40,699	382,086	23,518	88,832	5,925,226	3,924,635	3,602,682	3,602,682	3,253,590
Solids Flowrate (lb/hr)	0	17,686	0	0	22,208	0	0	0	0	0
Temperature (°F)	390	42	42	42	136	136	1,000	645	1,000	116
Pressure (psia)	12.2	11.4	12.0	11.4	12.0	12.0	2,415.0	620.5	565.5	400.0
Enthalpy (Btu/lb)	169.0	653.7	10.9	7.3	589.7	163.7	1,460.0	1,316.6	1,518.0	85.0
Density (lb/ft^3)	0.039	63.182	63.182	0.061	59.992	0.053	3.134	1.034	0.668	61.847

7.1.2 Cases 8 and 9 Process Description

Cases 8 and 9 are configured to produce electric power with CO_2 capture. Case 8 has an emission rate of 1,100 lb CO_2/net-MWh. This is achieved by bypassing a portion of the flue gas around the Econamine unit, therefore only treating a portion of the gas stream. Case 9 has a carbon capture rate of 90 percent. The plant configurations are similar to Case 7 with the major difference being the use of an Econamine FG Plus system for CO_2 capture and subsequent compression of the captured CO_2 stream. Low pressure steam (71 psi at approximately 305°F) is required for the Econamine system. For Case 8 (emission rate of 1,100 lb CO_2/net-MWh), approximately 35 percent of the total steam is extracted from the crossover pipe, sent to the let-down turbine and de-superheated before entering the Econamine system. For Case 9 (which includes 90 percent carbon capture), approximately 50 percent of the steam is extracted. For this analysis and based on results of the NETL/Alstom study, the existing steam turbine would be capable of turndown due to steam extraction. No other steam turbine retrofit is necessary other than the piping extraction from the IP/LP crossover [63]. Since the CO_2 capture and compression process increases the auxiliary load on the plant, the overall efficiency is significantly reduced relative to Case 7. A process block flow diagram for Cases 8 and 9 is shown in Exhibit 7-3 and Exhibit 7-5, respectively. Stream tables for Cases 8 and 9 are shown in Exhibit 7-4 and Exhibit 7-6, respectively. The CO_2 removal system is described in Section 5.1.7.

The boiler in the existing subcritical PC plant retrofit cases has a fixed heat duty and coal feed rate that is limited by its size and configuration. Therefore, the net power output decreases in the capture cases because of the extraction steam required in the CDR facility and the higher auxiliary loads.

Also, LNBs are upgraded for Cases 8 and 9 to reduce NOx emissions so that NO_2 is less than 20 ppmv as required by the Econamine process. The FGD system is upgraded to increase the efficiency from 85 to 92 percent to reduce the load on the polishing scrubber of the Econamine system. In the event that NSR would become relevant because of the CO_2 capture project, a cost sensitivity case was included with SCR retrofit to reduce NOx emissions to BACT limits.

Exhibit 7-3 Case 8: Subcritical PC Plant Retrofitted with Carbon Capture to an Emission Rate of 1,100 lb CO_2/net-MWh – Block Flow Diagram

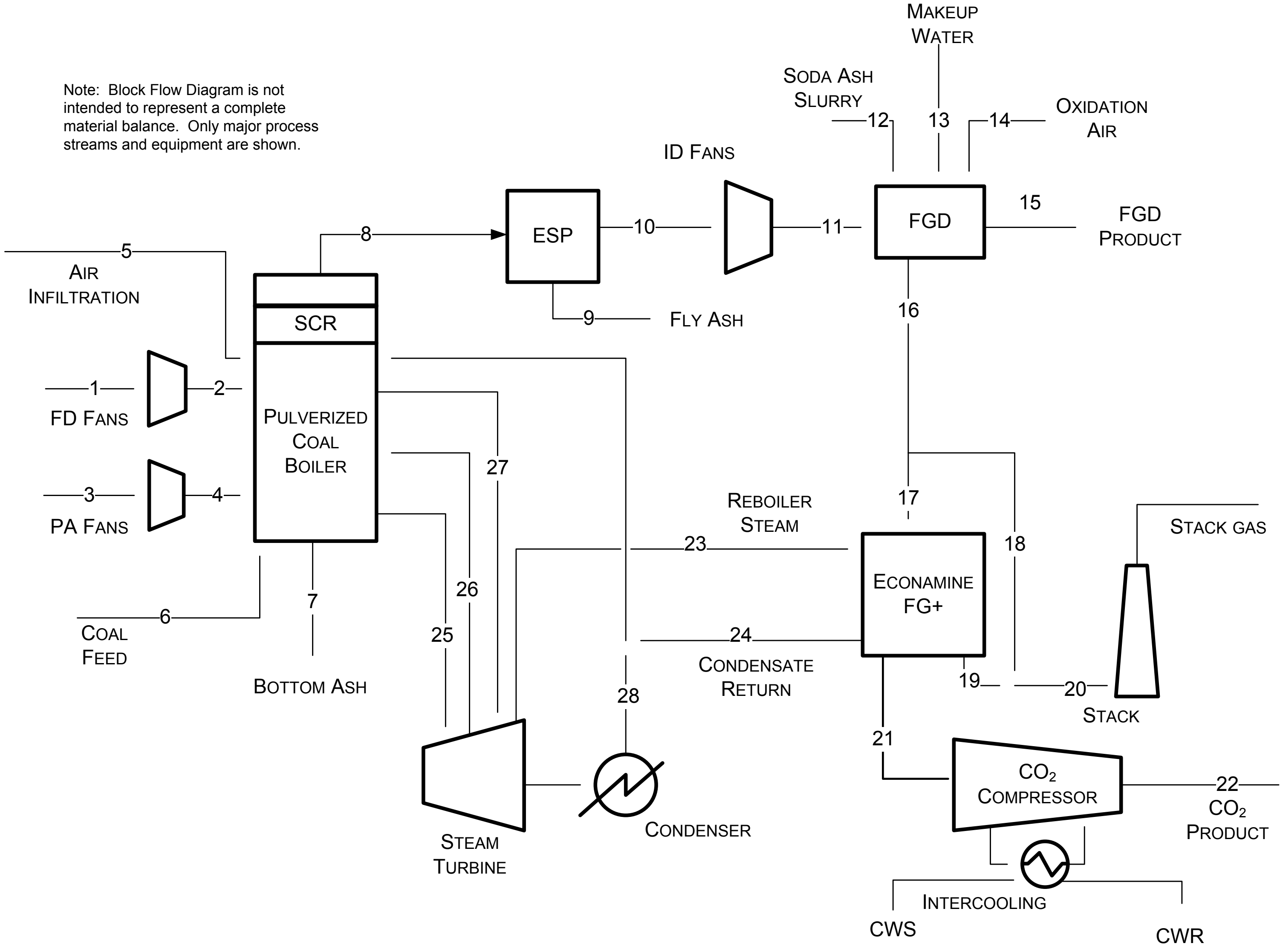


Exhibit 7-4 Case 8: Subcritical PC Plant Retrofitted with Carbon Capture to an Emission Rate of 1,100 lb CO_2/net-MWh - Stream Table

	1	2	3	4	5	6	7	8	9	10	11	12	13	14
V-L Mole Fraction														
Ar	0.0093	0.0093	0.0093	0.0093	0.0093	0.0000	0.0000	0.0084	0.0000	0.0084	0.0084	0.0000	0.0000	0.0093
CO_2	0.0003	0.0003	0.0003	0.0003	0.0003	0.0000	0.0000	0.1432	0.0000	0.1432	0.1432	0.0000	0.0000	0.0003
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0071	0.0071	0.0071	0.0071	0.0071	0.0000	0.0000	0.1131	0.0000	0.1131	0.1131	1.0000	1.0000	0.0071
N_2	0.7753	0.7753	0.7753	0.7753	0.7753	0.0000	0.0000	0.7059	0.0000	0.7059	0.7059	0.0000	0.0000	0.7753
O_2	0.2080	0.2080	0.2080	0.2080	0.2080	0.0000	0.0000	0.0286	0.0000	0.0286	0.0286	0.0000	0.0000	0.2080
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0000	0.0008	0.0008	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg$_{mol}$/hr)	58,852	58,852	18,079	18,079	1,322	0	0	86,052	0	86,052	86,052	1,113	9,692	374
V-L Flowrate (kg/hr)	1,700,066	1,700,066	522,242	522,242	38,194	0	0	2,531,345	0	2,531,345	2,531,345	20,057	174,600	10,818
Solids Flowrate (kg/hr)	0	0	0	0	0	294,996	4,831	19,322	19,322	0	0	8,436	0	0
Temperature (°C)	6	11	6	16	6	6	182	182	182	182	203	6	6	6
Pressure (MPa, abs)	0.08	0.08	0.08	0.09	0.08	0.08	0.08	0.08	0.07	0.07	0.08	0.08	0.08	0.08
Enthalpy (kJ/kg)	16.93	22.47	16.93	27.83	16.93	---	---	395.92	---	374.64	396.88	1,486.77	25.25	16.93
Density (kg/m^3)	1.0	1.0	1.0	1.0	1.0	---	---	0.6	---	0.6	0.6	1,012.1	1,012.1	1.0
V-L Molecular Weight	28.887	28.887	28.887	28.887	28.887	---	---	29.416	---	29.416	29.416	18.015	18.015	28.887
V-L Flowrate (lb$_{mol}$/hr)	129,746	129,746	39,857	39,857	2,915	0	0	189,713	0	189,713	189,713	2,455	21,367	826
V-L Flowrate (lb/hr)	3,748,004	3,748,004	1,151,348	1,151,348	84,203	0	0	5,580,661	0	5,580,661	5,580,661	44,219	384,927	23,850
Solids Flowrate (lb/hr)	0	0	0	0	0	650,355	10,650	42,599	42,599	0	0	18,599	0	0
Temperature (°F)	42	52	42	61	42	42	360	360	360	360	397	42	42	42
Pressure (psia)	11.4	12.0	11.4	12.6	11.4	11.4	10.9	10.9	10.7	10.7	12.2	11.4	12.0	11.4
Enthalpy (Btu/lb)	7.3	9.7	7.3	12.0	7.3	---	---	170.2	---	161.1	170.6	639.2	10.9	7.3
Density (lb/ft^3)	0.061	0.063	0.061	0.065	0.061	---	---	0.037	---	0.036	0.039	63.182	63.182	0.061

A - Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 7-4 Case 8: Subcritical PC Plant Retrofitted with Carbon Capture to an Emission Rate of 1,100 lb CO_2/net-MWh - Stream Table (Continued)

	15	16	17	18	19	20	21	22	23	24	25	26	27	28
V-L Mole Fraction														
Ar	0.0000	0.0077	0.0077	0.0077	0.0106	0.0095	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.0000	0.1306	0.1306	0.1306	0.0180	0.0609	0.9840	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	1.0000	0.1916	0.1916	0.1916	0.0469	0.1020	0.0160	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
N_2	0.0000	0.6436	0.6436	0.6436	0.8879	0.7948	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0265	0.0265	0.0265	0.0365	0.0327	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0001	0.0001	0.0001	0.0001	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000			
V-L Flowrate (kg_{mol}/hr)	2,366	94,835	65,569	29,266	47,527	76,794	7,832	7,706	31,690	31,690	98,801	90,696	90,696	53,946
V-L Flowrate (kg/hr)	42,626	2,691,975	1,861,232	830,744	1,335,889	2,166,632	341,414	339,150	570,895	570,895	1,779,931	1,633,913	1,633,913	971,852
Solids Flowrate (kg/hr)	10,656	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	58	58	58	58	32	45	21	35	152	151	538	340	538	47
Pressure (MPa, abs)	0.08	0.08	0.08	0.08	0.08	0.08	0.16	15.27	0.49	0.49	16.65	4.28	3.90	2.76
Enthalpy (kJ/kg)	1,375.35	380.70	380.70	380.70	111.30	214.60	28.91	-212.36	2,747.80	635.72	3,395.97	3,062.35	3,530.88	197.81
Density (kg/m^3)	958.8	0.9	0.9	0.9	0.9	0.9	2.9	794.4	2.6	915.8	50.2	16.6	10.7	990.7
V-L Molecular Weight	18.015	28.386	28.386	28.386	28.108	28.214	43.593	44.010	18.015	18.015	18.015	18.015	18.015	18.015
V-L Flowrate (lb_{mol}/hr)	5,216	209,075	144,555	64,521	104,780	169,301	17,266	16,989	69,863	69,863	217,819	199,950	199,950	118,931
V-L Flowrate (lb/hr)	93,973	5,934,789	4,103,313	1,831,476	2,945,131	4,776,607	752,688	747,699	1,258,609	1,258,609	3,924,075	3,602,161	3,602,161	2,142,566
Solids Flowrate (lb/hr)	23,493	0	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	136	136	136	136	89	113	69	95	306	304	1,000	645	1,000	116
Pressure (psia)	12.0	12.0	12.0	12.0	12.0	12.0	23.5	2,215.0	71.0	71.0	2,415.0	620.5	565.5	400.0
Enthalpy (Btu/lb)	591.3	163.7	163.7	163.7	47.9	92.3	12.4	-91.3	1,181.3	273.3	1,460.0	1,316.6	1,518.0	85.0
Density (lb/ft^3)	59.857	0.053	0.053	0.053	0.057	0.055	0.183	49.590	0.163	57.172	3.134	1.034	0.668	61.847

Exhibit 7-5 Case 9: Existing Subcritical PC Retrofitted with 90% CO_2 Capture - Block Flow Diagram

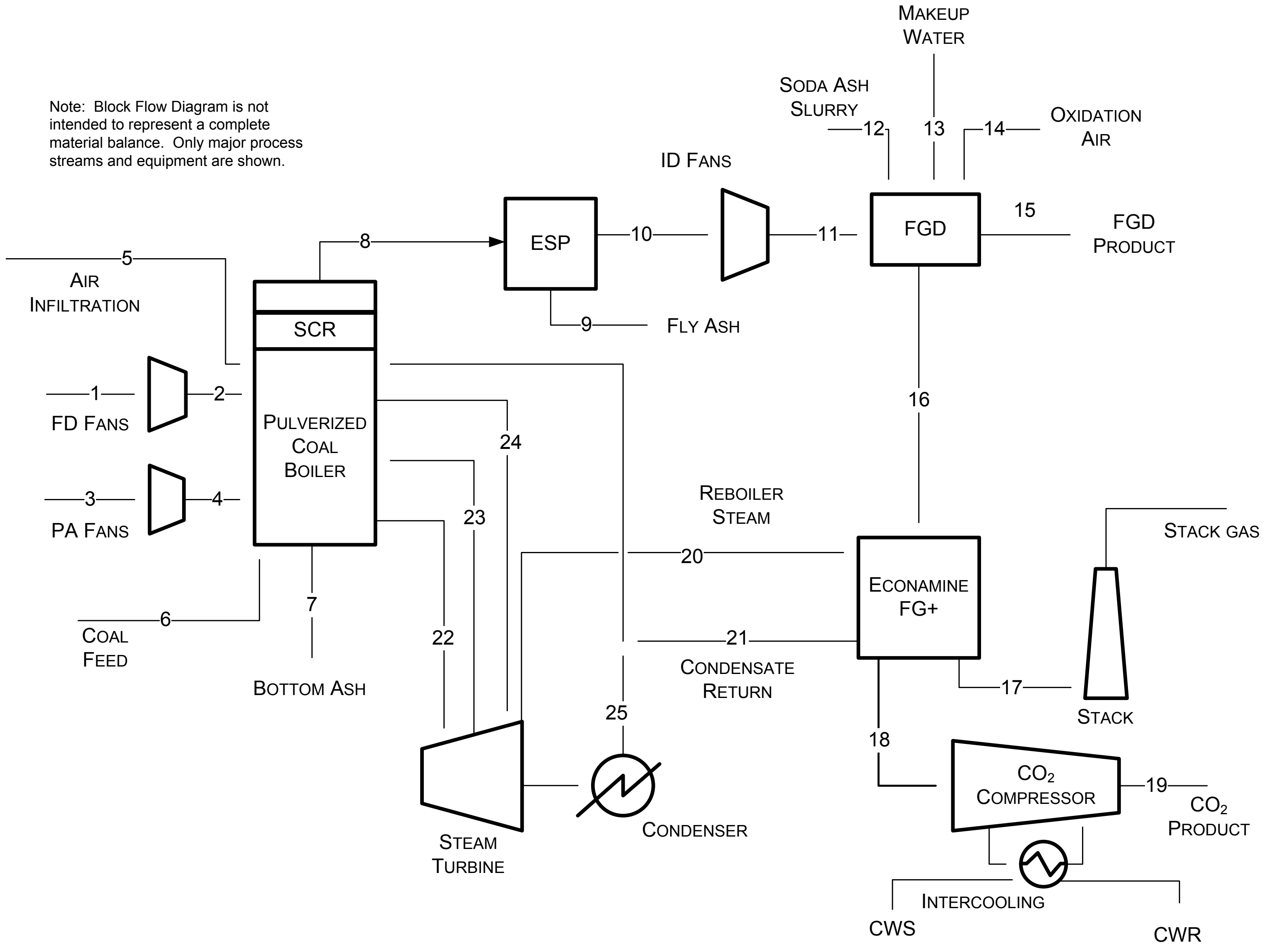


Exhibit 7-6 Case 9: Existing Subcritical PC Retrofitted with 90% CO_2 Capture - Stream Table

	1	2	3	4	5	6	7	8	9	10	11	12	13
V-L Mole Fraction													
Ar	0.0093	0.0093	0.0093	0.0093	0.0093	0.0000	0.0000	0.0084	0.0000	0.0084	0.0084	0.0000	0.0000
CO_2	0.0003	0.0003	0.0003	0.0003	0.0003	0.0000	0.0000	0.1437	0.0000	0.1437	0.1437	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0071	0.0071	0.0071	0.0071	0.0071	0.0000	0.0000	0.1135	0.0000	0.1135	0.1135	1.0000	1.0000
N_2	0.7753	0.7753	0.7753	0.7753	0.7753	0.0000	0.0000	0.7057	0.0000	0.7057	0.7057	0.0000	0.0000
O_2	0.2080	0.2080	0.2080	0.2080	0.2080	0.0000	0.0000	0.0279	0.0000	0.0279	0.0279	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0000	0.0008	0.0008	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	58,610	58,610	18,004	18,004	1,322	0	0	85,736	0	85,736	85,736	1,075	9,657
V-L Flowrate (kg/hr)	1,693,082	1,693,082	520,097	520,097	38,194	0	0	2,522,218	0	2,522,218	2,522,218	19,358	173,975
Solids Flowrate (kg/hr)	0	0	0	0	0	294,998	4,831	19,323	19,323	0	0	8,437	0
Temperature (°C)	6	11	6	16	6	6	182	182	182	182	203	6	6
Pressure (MPa, abs)	0.08	0.08	0.08	0.09	0.08	0.08	0.08	0.08	0.07	0.07	0.08	0.08	0.08
Enthalpy (kJ/kg)	16.93	22.47	16.93	27.83	16.93	---	---	396.72	---	375.28	397.52	1,523.61	25.25
Density (kg/m^3)	1.0	1.0	1.0	1.0	1.0	1,415.1	14.4	0.6	14.4	0.6	0.6	1,237.7	1,012.1
V-L Molecular Weight	28.887	28.887	28.887	28.887	28.887	---	---	29.418	---	29.418	29.418	18.015	18.015
V-L Flowrate (lb_{mol}/hr)	129,213	129,213	39,693	39,693	2,915	0	0	189,016	0	189,016	189,016	2,369	21,290
V-L Flowrate (lb/hr)	3,732,607	3,732,607	1,146,618	1,146,618	84,204	0	0	5,560,540	0	5,560,540	5,560,540	42,677	383,549
Solids Flowrate (lb/hr)	0	0	0	0	0	650,360	10,650	42,599	42,599	0	0	18,600	0
Temperature (°F)	42	52	42	61	42	42	360	360	360	360	397	42	42
Pressure (psia)	11.4	12.0	11.4	12.6	11.4	11.4	10.9	10.9	10.7	10.7	12.2	11.4	12.0
Enthalpy (Btu/lb)	7.3	9.7	7.3	12.0	7.3	---	---	170.6	---	161.3	170.9	655.0	10.9
Density (lb/ft^3)	0.061	0.063	0.061	0.065	0.061	88.340	0.899	0.037	0.899	0.036	0.039	77.265	63.182
	A - Reference conditions are 32.02 F & 0.089 PSIA												

Exhibit 7-6 Case 9: Existing Subcritical PC Retrofitted with 90% CO_2 Capture - Stream Table (Continued)

	14	15	16	17	18	19	20	21	22	23	24	25
V-L Mole Fraction												
Ar	0.0093	0.0000	0.0077	0.0106	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.0003	0.0000	0.1311	0.0181	0.9840	1.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0071	1.0000	0.1916	0.0469	0.0160	0.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
N_2	0.7753	0.0000	0.6437	0.8886	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.2080	0.0000	0.0259	0.0357	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0001	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000			
V-L Flowrate (kg_{mol}/hr)	368	2,366	94,439	68,405	11,327	11,146	45,862	45,862	98,833	90,726	90,726	41,492
V-L Flowrate (kg/hr)	10,635	42,628	2,681,339	1,922,597	493,787	490,527	826,217	826,217	1,780,510	1,634,452	1,634,452	747,492
Solids Flowrate (kg/hr)	0	10,657	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	6	58	58	32	21	35	152	151	538	340	538	47
Pressure (MPa, abs)	0.08	0.08	0.08	0.08	0.16	15.27	0.49	0.49	16.65	4.28	3.90	2.76
Enthalpy (kJ/kg)	16.93	1,375.34	380.63	111.31	28.91	-212.36	2,746.50	635.72	3,395.97	3,062.35	3,530.88	197.81
Density (kg/m^3)	1.0	1,099.0	0.9	0.9	2.9	794.4	2.6	915.8	50.2	16.6	10.7	990.7
V-L Molecular Weight	28.887	18.016	28.392	28.106	43.595	44.010	18.015	18.015	18.015	18.015	18.015	18.015
V-L Flowrate (lb_{mol}/hr)	812	5,217	208,203	150,807	24,971	24,572	101,108	101,108	217,890	200,016	200,016	91,474
V-L Flowrate (lb/hr)	23,447	93,978	5,911,340	4,238,600	1,088,614	1,081,427	1,821,496	1,821,496	3,925,353	3,603,349	3,603,349	1,647,938
Solids Flowrate (lb/hr)	0	23,494	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	42	136	136	89	69	95	305	304	1,000	645	1,000	116
Pressure (psia)	11.4	12.0	12.0	12.0	23.5	2,215.0	71.0	71.0	2,415.0	620.5	565.5	400.0
Enthalpy (Btu/lb)	7.3	591.3	163.6	47.9	12.4	-91.3	1,180.8	273.3	1,460.0	1,316.6	1,518.0	85.0
Density (lb/ft^3)	0.061	68.607	0.053	0.057	0.183	49.590	0.163	57.172	3.134	1.034	0.668	61.847

7.1.3 Key System Assumptions

System assumptions for Cases 7, 8 and 9, subcritical PC with and without CO_2 capture, are compiled in Exhibit 7-7.

Exhibit 7-7 Subcritical PC Plant Study Configuration Matrix

	Case 7 w/o CO_2 Capture	Case 8 w/CO_2 Capture	Case 9 w/CO_2 Capture
Steam Cycle, MPa/°C/°C (psig/°F/°F)	16.5/538/538 (2400/1000/1000)	16.5/538/538 (2400/1000/1000)	16.5/538/538 (2400/1000/1000)
Coal	Montana Rosebud PRB	Montana Rosebud PRB	Montana Rosebud PRB
Condenser pressure, mm Hg (in Hg)	35.6 (1.4)	35.6 (1.4)	35.6 (1.4)
Boiler Efficiency, %	83	83	83
Cooling water to condenser, °C (ºF)	8.9 (48)	8.9 (48)	8.9 (48)
Cooling water from condenser, °C (ºF)	20 (68)	20 (68)	20 (68)
Stack temperature, °C (°F)	58 (136)	45 (113)	32 (89)
SO_2 Control	Soda Ash-Based Wet Scrubber	Soda Ash-Based Wet Scrubber	Soda Ash-Based Wet Scrubber
FGD Efficiency, % (A)	85	92 (B)	92 (B)
NOx Control	LNB w/OFA	Advanced LNB w/OFA	Advanced LNB w/OFA
Particulate Control	ESP	ESP	ESP
ESP efficiency, % (A)	99.65	99.65	99.65
Ash Distribution, Fly/Bottom	80% / 20%	80% / 20%	80% / 20%
Mercury Control	Co-benefit Capture	Co-benefit Capture	Co-benefit Capture
Mercury removal efficiency, % (A)	16	16	16
CO_2 Control	N/A	Econamine FG Plus	Econamine FG Plus
CO_2 Capture (A)	N/A	1,100 lb/net-MWh	90%
CO_2 Sequestration	N/A	Off-site Saline Formation	Off-site Saline Formation

A. Removal efficiencies are based on the flue gas content

B. An SO_2 polishing step is included to meet more stringent SOx content limits in the flue gas (< 10 ppmv) to reduce formation of amine heat stable salts during the CO_2 absorption process

Balance of Plant – Cases 7 - 9

The balance of plant assumptions are common to all cases and are presented in Exhibit 7-8.

Exhibit 7-8 Balance of Plant Assumptions

Cooling system	Recirculating Wet Cooling Tower
Fuel and Other storage Coal	30 days
Ash	30 days
Soda Ash	30 days
Plant Distribution Voltage Motors below 1 hp	110/220 volt
Motors between 1 hp and 250 hp	480 volt
Motors between 250 hp and 5,000 hp	4,160 volt
Motors above 5,000 hp	13,800 volt
Steam and Gas Turbine generators	24,000 volt
Grid Interconnection voltage	345 kV
Water and Waste Water Makeup Water	The water supply 100 percent from the Green River. No municipal water sources are utilized.
Process Wastewater	Storm water that contacts equipment surfaces is collected and treated for discharge through a permitted discharge.
Sanitary Waste Disposal	Design includes a packaged domestic sewage treatment plant with effluent discharged to the industrial wastewater treatment system. Sludge is hauled off site. Packaged plant is sized for 5.68 cubic meters per day (1,500 gallons per day)
Water Discharge	Most of the process wastewater is recycled to the cooling tower basin. Blowdown will be treated for chloride and metals, and discharged.

7.1.4 Sparing Philosophy

Single trains are used throughout the design with exceptions where equipment capacity requires an additional train. There is no redundancy other than normal sparing of rotating equipment. The plant design consists of the following major subsystems:

- One dry-bottom, tangentially-fired PC subcritical boiler (1 x 100%)
- Two cold-side ESPs (2 x 50%)
- One soda ash-based wet forced oxidation positive pressure absorber (1 x 100%)
- One steam turbine (1 x 100%)
- For Case 9, two parallel Econamine FG Plus CO_2 absorption systems, with each system consisting of two absorbers, strippers and ancillary equipment (2 x 50%). Case 8 consists of a single train only.

7.1.5 Case 7 - 9 Performance Results

Cases 7 through 9 are based on a coal feed rate of 295,000 kg/hr (650,360 lb/hr). Overall performance for the plant is summarized in Exhibit 7-9 which includes auxiliary power requirements.

Exhibit 7-9 Cases 7 - 9 Plant Performance Summary

Power Output, kWe	**Case 7**	**Case 8**	**Case 9**
Steam Turbine Power	**577,800**	**476,800**	**432,000**
Econamine Let Down Turbine Power	**N/A**	**28,100**	**40,600**
Gross Power	**577,800**	**504,900**	**472,600**
Auxiliary Load, kWe			
Coal Handling and Conveying	550	550	550
Pulverizers	4,420	4,420	4,420
Sorbent Handling & Reagent Preparation	390	410	410
Ash Handling	710	710	710
Primary Air Fans	1,640	1,640	1,630
Forced Draft Fans	2,700	2,710	2,700
Induced Draft Fans	13,090	16,120	16,060
ESP	1,000	1,000	1,000
FGD Pumps and Agitators	1,370	1,370	1,370
Econamine FG Plus Auxiliaries	N/A	12,700	18,400
Econamine Condensate Return Pump	N/A	90	130
CO_2 Compression	N/A	28,200	40,780
Miscellaneous Balance of Plant[1,2]	6,500	6,500	6,500
Steam Turbine Auxiliaries	400	400	400
Condensate Pumps	1,470	990	760
Circulating Water Pumps	5,550	7,330	8,980
Cooling Tower Fans	4,130	5,460	6,690
Air Cooled Condenser Fans	0	0	0
Transformer Losses	1,850	1,760	1,720
Total	45,770	92,360	113,210
Plant Performance			
Net Plant Power	532,030	412,540	359,390
Net Plant Efficiency (HHV)	32.6%	25.3%	22.0%
Net Plant Heat Rate (HHV)	11,045 (10,469)	14,244 (13,501)	16,351 (15,498)
Coal Feed Flowrate (kg/hr (lb/hr))	294,998 (650,360)	294,996 (650,355)	294,998 (650,360)
Thermal Input (kW$_{th}$)	1,632,313	1,632,299	1,632,314
Condenser Duty (GJ/hr (MMBtu/hr))	2,716 (2,574)	1,779 (1,686)	1,361 (1,290)
Raw Water Withdrawal (m^3/min (gpm))	24.8 (6,553)	30.5 (8,048)	33.9 (8,948)
Raw Water Consumption (m^3/min (gpm))	19.9 (5,270)	24.0 (6,352)	26.0 (6,869)
Other Consumables			
SCR Catalyst (m^3 (ft^3))	N/A	N/A	N/A
FGD Sorbent (tonne/day (ton/day))	8.02 (8.84)	8.44 (9.30)	8.44 (9.30)
Ammonia (19% Solution) (tonne/day (ton/day))	N/A	N/A	N/A
Econamine Consumables			
MEA (tonne/day (ton/day))	N/A	0.82 (0.90)	1.18 (1.30)
Activated Carbon (kg/day (lb/day))	N/A	488 (1,076)	706 (1,557)
Sodium Hydroxide (NaOH) (tonne/day (ton/day))	N/A	6.72 (7.41)	9.73 (10.72)
Sulfuric Acid (H2SO4) (tonne/day (ton/day))	N/A	3.89 (4.28)	5.62 (6.20)
Corrosion Inhibitor ($/yr)	N/A	4,186	6,054

1 - Boiler feed pumps are turbine driven

2 - Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads. Miscellaneous loads were estimated to match the reported efficiency for the existing subcritical PC plant.

Environmental Performance

The environmental targets for emissions of Hg, NOx, SO_2 and particulate matter were presented in Section 2.4. A summary of the plant air emissions for Cases 7 through 9 is presented in Exhibit 7-10.

Exhibit 7-10 Cases 7 - 9 Air Emissions

	Case 7	Case 8	Case 9
kg/GJ (lb/10^6 Btu)			
SO_2	0.109 (0.255)	0.023 (0.054)	0.007 (0.017)
NO_X	0.193 (0.450)	0.103 (0.240)	0.103 (0.240)
Particulates	0.012 (0.0270)	0.012 (0.0270)	0.012 (0.027)
Hg	2.57E-6 (5.97E-6)	2.54E-6 (5.90E-6)	2.54E-6 (5.90E-6)
CO_2	93 (216)	35 (81)	9.3 (22)
Tonne/year (tons/year) 85% capacity			
SO_2	5,070 (5,589)	1,074 (1,184)	346 (381)
NO_X	8,963 (9,880)	4,780 (5,269)	4,780 (5,269)
Particulates	538 (593)	538 (593)	538 (593)
Hg	0.119 (0.131)	0.117 (0.129)	0.117 (0.129)
CO_2	4,295,414 (4,734,883)	1,623,154 (1,789,221)	429,701 (473,664)
kg/MWh (lb/gross-MWh)			
SO_2	1.11 (2.45)	0.270 (0.595)	0.093 (0.204)
NO_X	1.97 (4.34)	1.20 (2.65)	1.28 (2.83)
Particulates	0.118 (0.260)	0.135 (0.298)	0.144 (0.318)
Hg	2.61E-5 (5.75E-5)	2.95E-5 (6.51E-5)	3.15E-5 (6.95E-5)
CO_2	943 (2,079)	408 (899)	115 (254)
kg/MWh (lb/net-MWh)			
CO_2	1,024 (2,258)	499 (1,100)	152 (334)

SO_2 emissions are controlled using a wet soda ash-based forced oxidation scrubber that achieves a removal efficiency of 85, 92, and 92 percent for Cases 7 through 9, respectively. The byproduct sodium sulfate is dewatered and disposed of in a landfill. The flue gas exiting the scrubber is vented through the plant stack (Case 7) or sent to the Econamine unit (Cases 8 and 9).

NOx emissions are controlled to about 0.45 lb/10^6 Btu for Case 7 and 0.24 lb/10^6 Btu for Cases 8 and 9 through the use of LNBs and OFA. Particulate emissions are controlled using an electrostatic precipitator (ESP) which operates at an efficiency of 99.65 percent.

Co-benefit capture results in a 16 percent reduction of mercury emissions. CO_2 emissions represent the uncontrolled discharge from the process in Case 7. In Case 8 the CO_2 emission are limited to 1,100 lb/net-MWh, and in Case 9 there is a nominal 90 percent carbon capture.

Exhibit 7-11 shows the overall water balance for the plant. Raw water is obtained from the Green River.

Exhibit 7-11 Cases 7 - 9 Water Balance

	Case 7	**Case 8**	**Case 9**
Water Demand, m^3/min (gpm)			
Econamine	N/A	0.08 (22)	0.12 (32)
FGD Makeup	3.12 (846)	3.2 (858)	3.2 (852)
BFW Makeup	0.30 (78)	0.30 (78)	0.30 (79)
Cooling Tower	21.6 (5,708)	28.6 (7,543)	35.0 (9,246)
Total	**25.1 (6,632)**	**32.2 (8,502)**	**38.6 (10,209)**
Internal Recycle, m^3/min (gpm)			
Econamine	N/A	0.0 (0)	0.0 (0)
FGD Makeup	0.0 (0)	0.0 (0)	0.0 (0)
BFW Makeup	0.0 (0)	0.0 (0)	0.0 (0)
Cooling Tower	0.30 (78)	1.7 (454)	4.8 (1,260)
Total	**0.30 (78)**	**1.7 (454)**	**4.8 (1,260)**
Raw Water Withdrawal, m^3/min (gpm)			
Econamine	N/A	0.08 (22)	0.12 (32)
FGD Makeup	3.2 (846)	3.2 (858)	3.2 (852)
BFW Makeup	0.30 (78)	0.30 (78)	0.30 (79)
Cooling Tower	21.3 (5,629)	26.8 (7,089)	30.2 (7,985)
Total	**24.8 (6,553)**	**30.5 (8,048)**	**33.9 (8,948)**
Process Water Discharge, m^3/min (gpm)			
Cooling Tower	4.9 (1,284)	6.4 (1,696)	7.9 (2,079)
Total	**4.9 (1,284)**	**6.4 (1,696)**	**7.9 (2,079)**
Raw Water Consumption, m^3/min (gpm)			
Econamine	N/A	0.08 (22)	0.12 (32)
FGD Makeup	3.2 (846)	3.2 (858)	3.2 (852)
BFW Makeup	0.30 (78)	0.30 (78)	0.30 (79)
Cooling Tower	16.4 (4,346)	20.4 (5,393)	22.4 (5,906)
Total	**19.9 (5,270)**	**24.0 (6,352)**	**26.0 (6,869)**
Total, gpm/MWnet	**9.9**	**15.4**	**19.1**

Water demand represents the total amount of water required for a particular process. Some water is recovered within the process, primarily as flue gas condensate in CO_2 capture cases, and that water is re-used as internal recycle. Raw water withdrawal is the difference between water demand and internal recycle. Some water is returned to the source, namely cooling tower blowdown. The difference between raw water withdrawal and water returned to the source (process discharge) is raw water consumption, which represents the net impact on the water source.

The carbon balance for the plant is shown in Exhibit 7-12. The carbon input to the plant consists of carbon in the coal, carbon in the air, and carbon in the FGD reagent. Carbon leaves the plant as carbon in the FGD product, CO_2 in the stack gas, and CO_2 product. The percent of total carbon sequestered for the capture cases is defined as the amount of carbon product produced (as sequestration-ready CO_2) divided by the carbon in the coal feedstock, less carbon contained in solid byproducts (ash).

Exhibit 7-12
Cases 7 - 9 Carbon Balance

	Case 7	Case 8	Case 9
Carbon In, kg/hr (lb/hr)			
Coal	147,700 (325,623)	147,699 (325,620)	147,700 (325,623)
Air (CO_2)	309 (681)	309 (682)	308 (679)
FGD Reagent	818 (1,804)	860 (1,897)	860 (1,897)
Total In	**148,827 (328,108)**	**148,869 (328,200)**	**148,869 (328,200)**
Carbon Out, kg/hr (lb/hr)			
Ash	0 (0)	0 (0)	0 (0)
Stack Gas	148,692 (327,810)	56,188 (123,873)	14,875 (32,793)
FGD Product	135 (298)	121 (267)	121 (267)
CO_2 Product	N/A	92,560 (204,059)[1]	133,873 (295,139)[2]
Total Out	**148,827 (328,108)**	**148,869 (328,200)**	**148,869 (328,200)**

[1] Carbon capture is 62.3 percent to achieve an emission rate of 1,100 lb CO_2/net-MWh
[2] Carbon capture is 90 percent

Exhibit 7-13 shows the sulfur balance for the plant. Sulfur input is the sulfur in the coal. Sulfur output is the sulfur combined with lime in the ash, and the sulfur emitted in the stack gas.

Exhibit 7-13 Cases 4 - 6 Sulfur Balance

	Case 7	Case 8	Case 9
Sulfur In, kg/h (lb/hour)			
Coal	2,146 (4,731)	2,146 (4,731)	2,146 (4,731)
Total In	**2,146 (4,731)**	**2,146 (4,731)**	**2,146 (4,731)**
Sulfur Out, kg/h (lb/hour)			
FGD Product	1,824 (4,021)[1]	1,974 (4,353)[2]	1,974 (4,353)[2]
Stack Gas	322 (710)	172 (378)	172 (378)
CO_2 Product	N/A	0 (0)	0 (0)
Total Out	**2,146 (4,731)**	**2,146 (4,731)**	**2,146 (4,731)**

[1] Sulfur capture is 85 percent

[2] Sulfur capture is 92 percent

Heat and Mass Balance Diagrams

Heat and mass balance diagrams are shown Exhibit 7-14 through Exhibit 7-19 for the three existing subcritical PC plant configurations

An overall plant energy balance is provided in tabular form in Exhibit 7-20.

Exhibit 7-14 Case 7: Existing Subcritical PC Boiler – Boiler and Gas Cleanup Systems Heat and Mass Balance Schematic

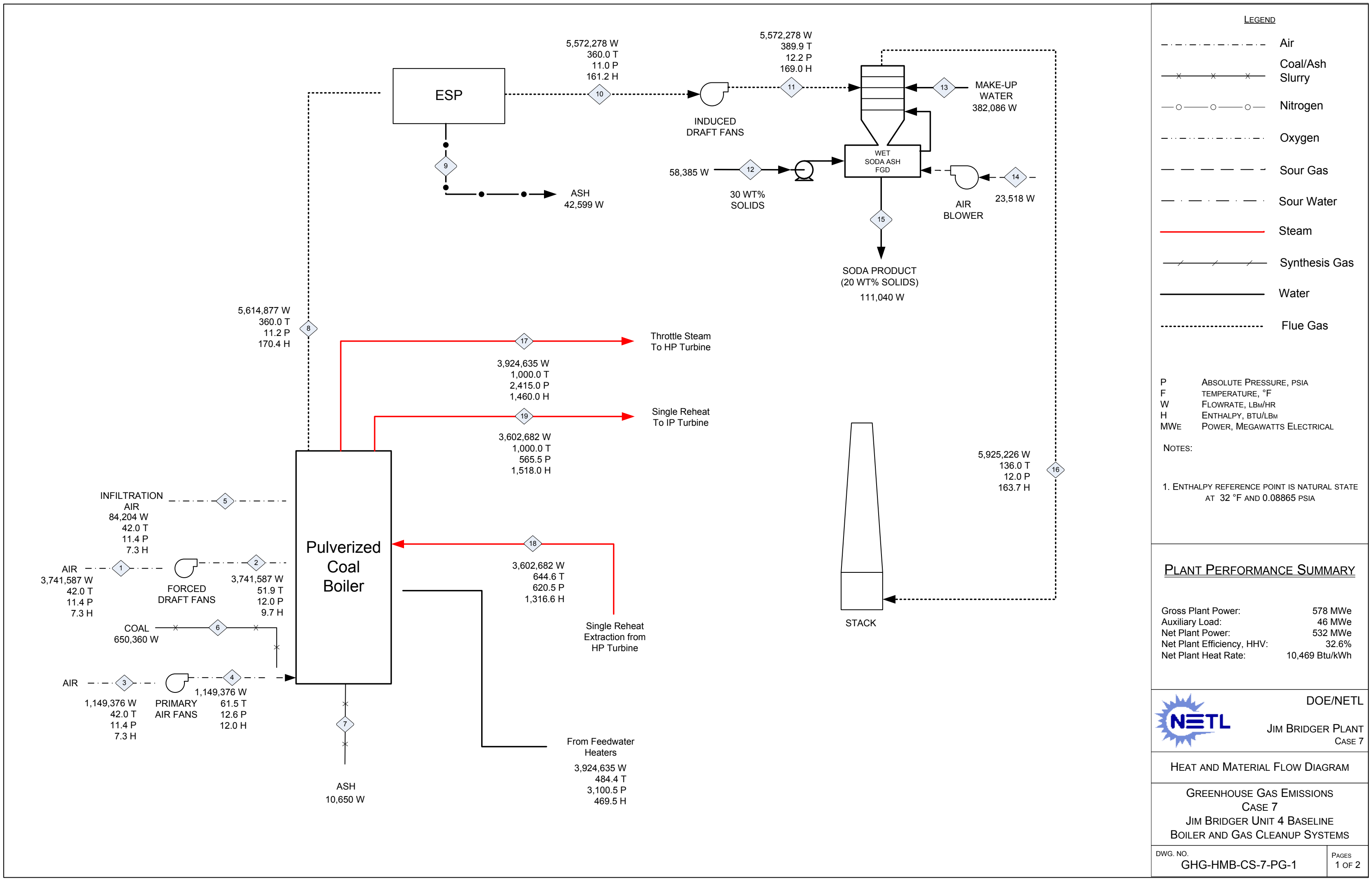
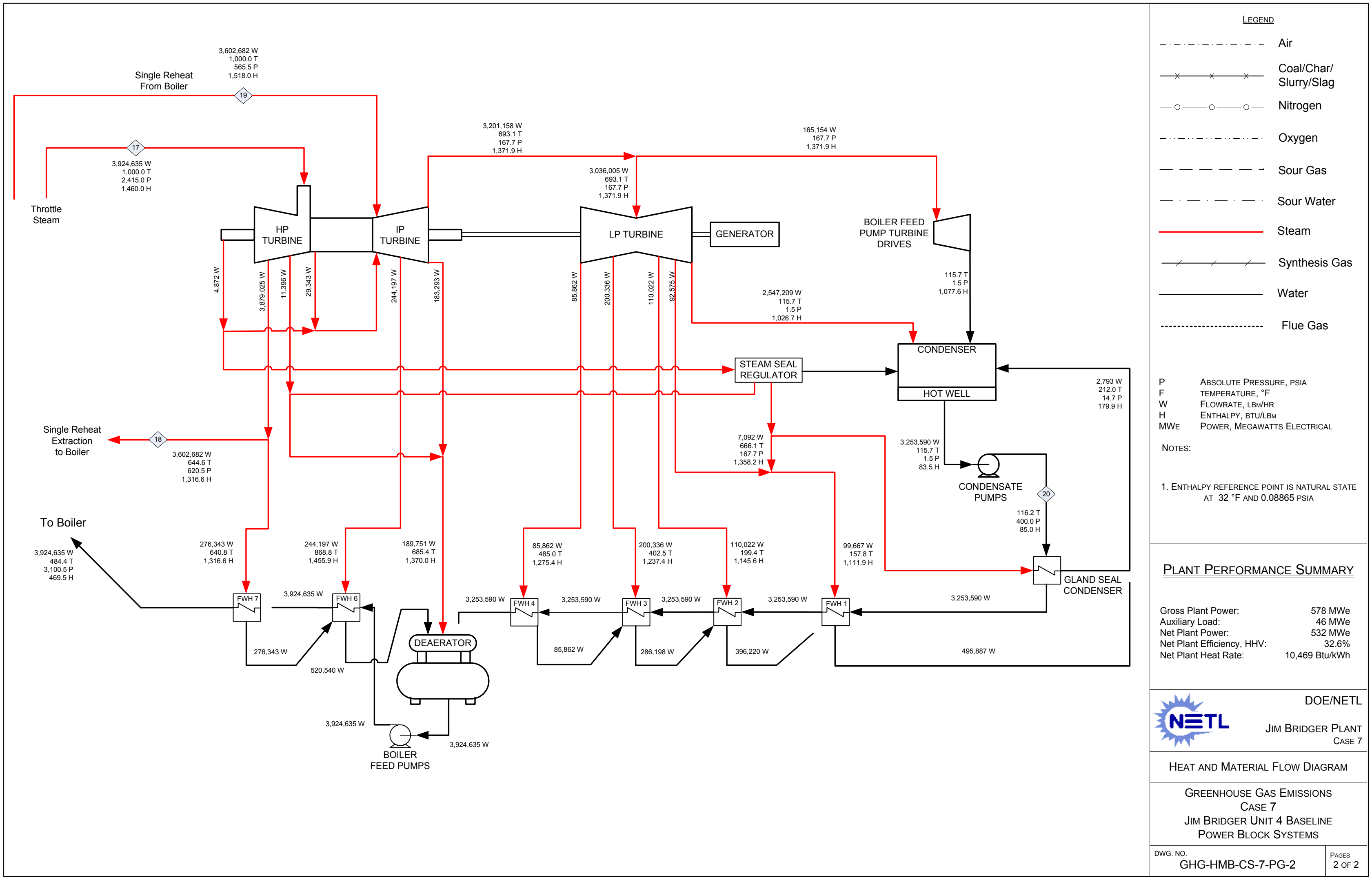


Exhibit 7-15 Case 7: Existing Subcritical PC - Power Block System Heat and Mass Balance Schematic



P	ABSOLUTE PRESSURE, PSIA
F	TEMPERATURE, °F
W	FLOWRATE, LBM/HR
H	ENTHALPY, BTU/LBM
MWE	POWER, MEGAWATTS ELECTRICAL

NOTES:

1. ENTHALPY REFERENCE POINT IS NATURAL STATE AT 32 °F AND 0.08865 PSIA

PLANT PERFORMANCE SUMMARY

Gross Plant Power:	578 MWe
Auxiliary Load:	46 MWe
Net Plant Power:	532 MWe
Net Plant Efficiency, HHV:	32.6%
Net Plant Heat Rate:	10,469 Btu/kWh

Exhibit 7-16 Case 8: Subcritical PC Plant Retrofitted with Carbon Capture to an Emission Rate of 1,100 lb CO_2/net-MWh - Boiler and Gas Cleanup Systems Heat and Mass Balance Schematic

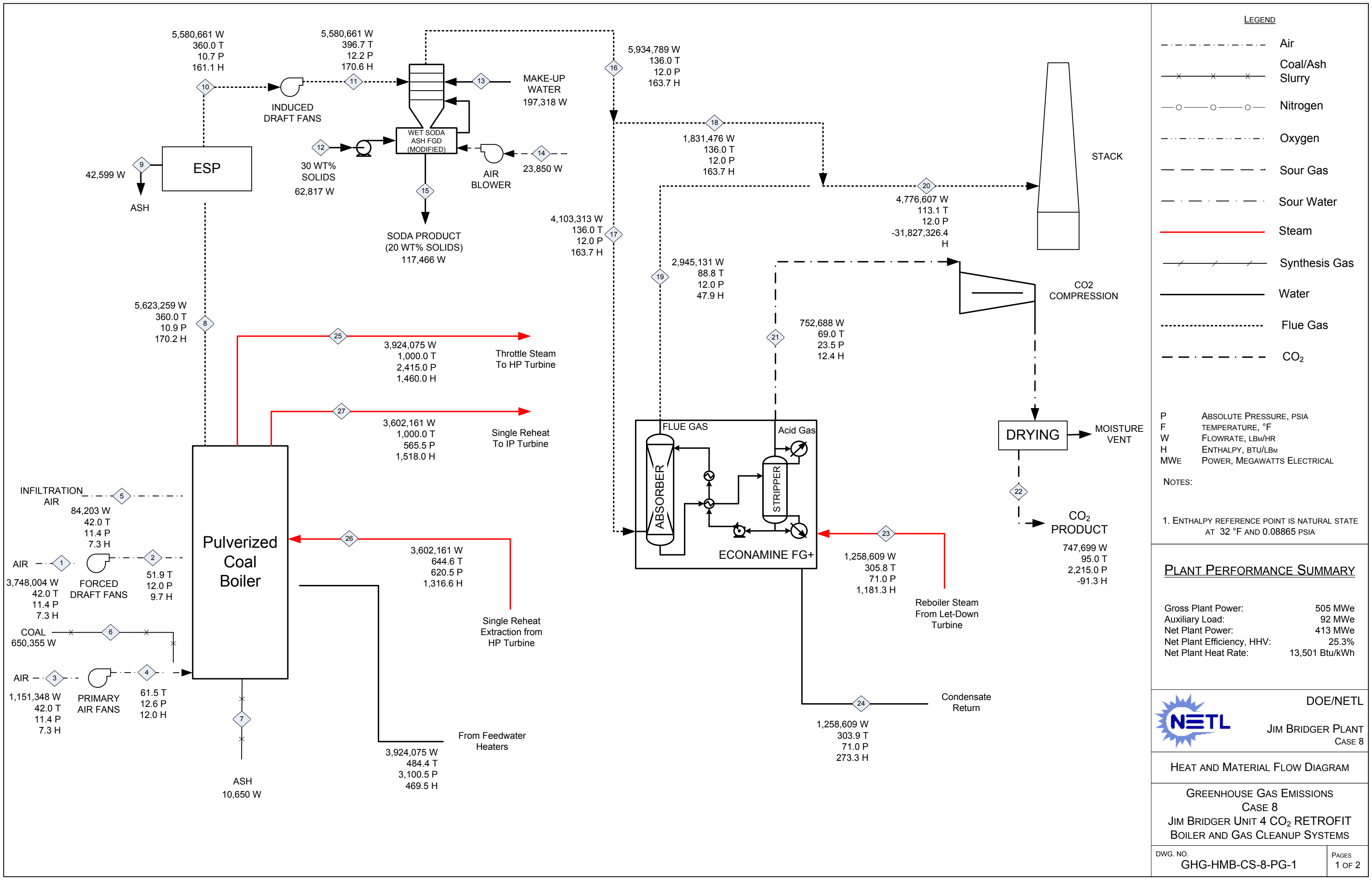


Exhibit 7-17 Case 8: Subcritical PC Plant Retrofitted with Carbon Capture to an Emission Rate of 1,100 lb CO_2/net-MWh - Power Block Systems Heat and Mass Balance Schematic

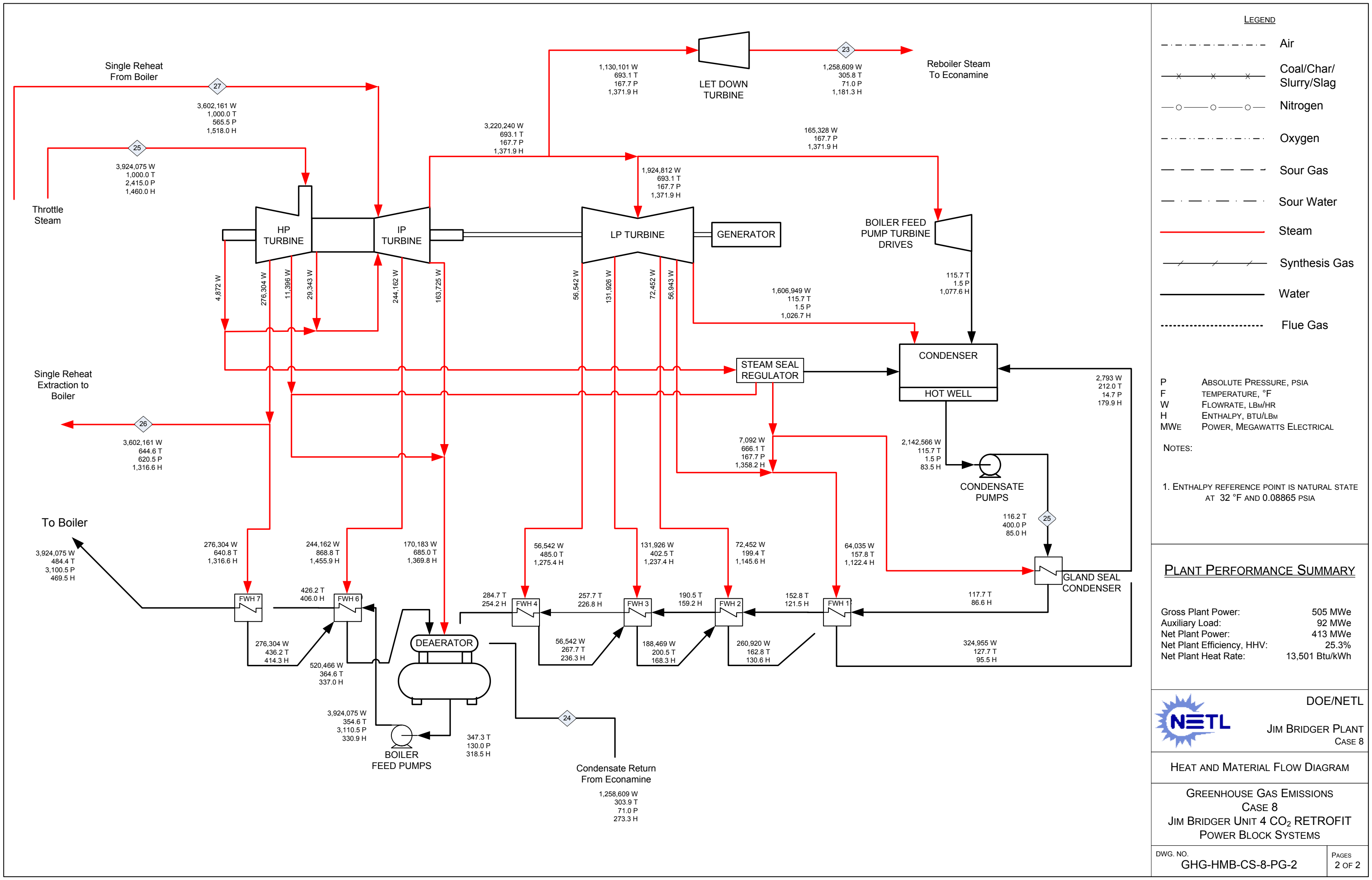


Exhibit 7-18 Case 9: Subcritical PC Plant Retrofitted with 90% CO_2 Capture - Boiler and Gas Cleanup Systems Heat and Mass Balance Schematic

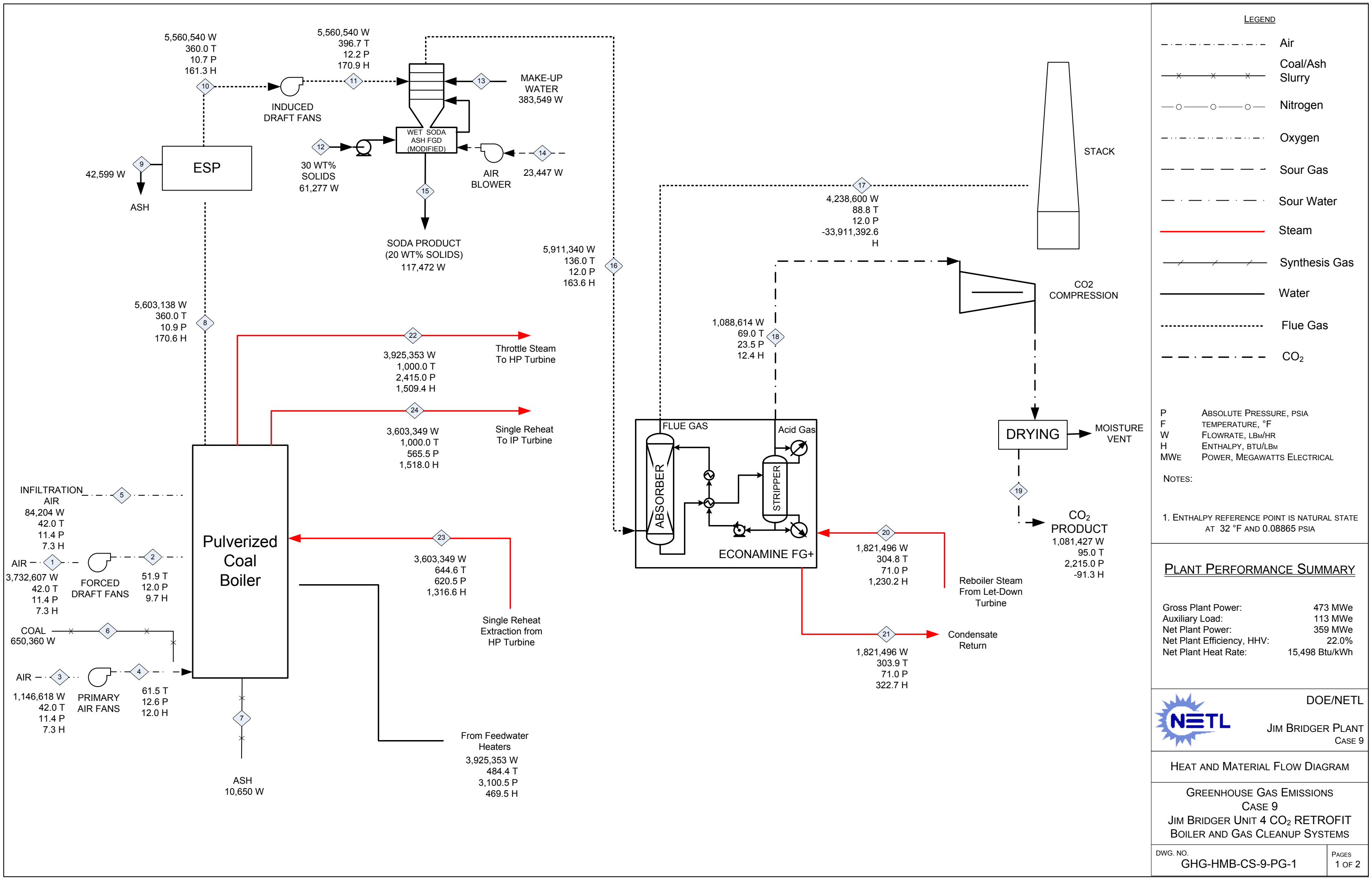


Exhibit 7-19 Case 9: Subcritical PC Plant Retrofitted with 90% CO_2 Capture - Power Block System Heat and Mass Balance Schematic

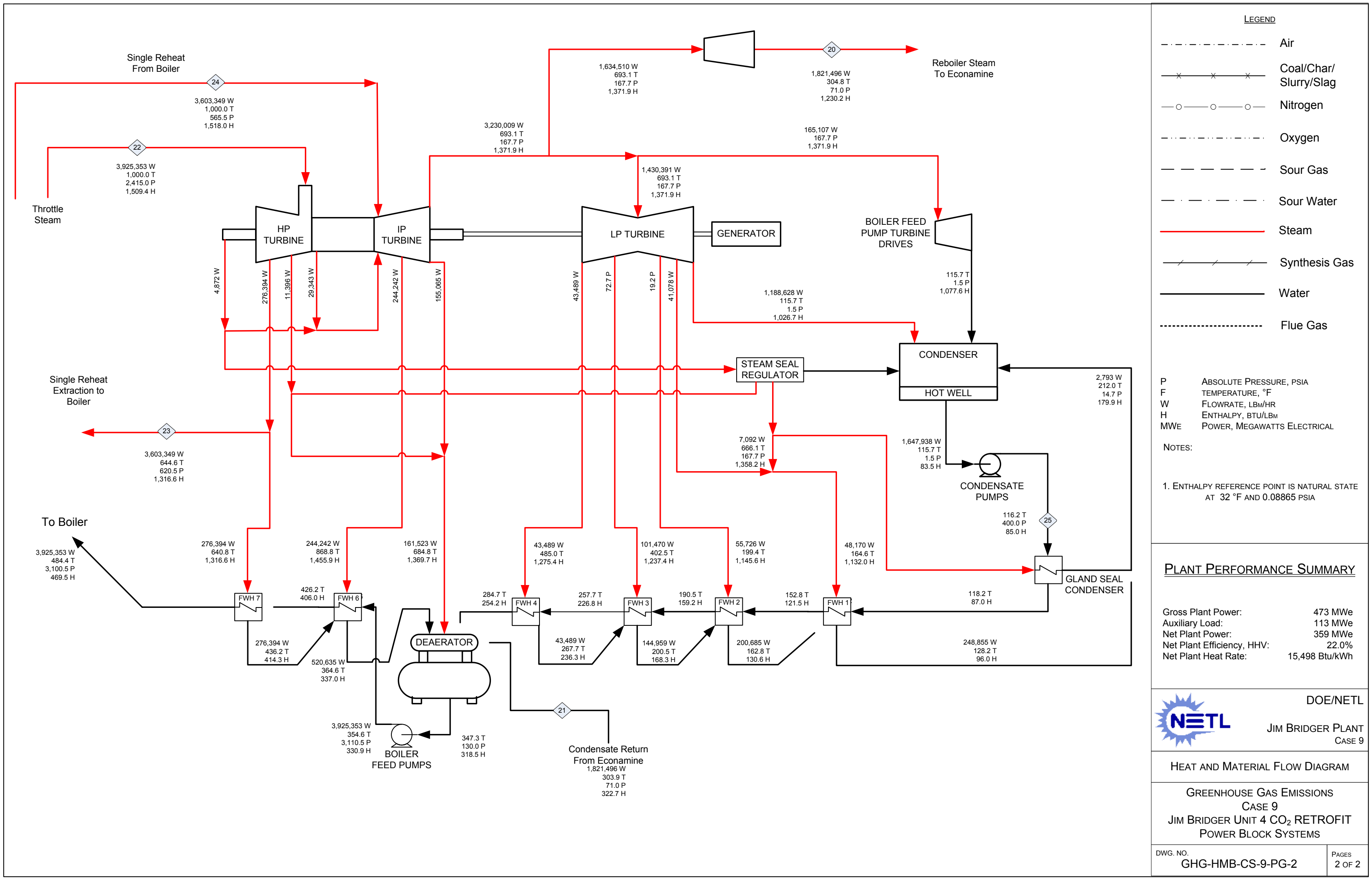


Exhibit 7-20 Cases 7 - 9 Overall Energy Balance

	Case 7	Case 8	Case 9
Energy In, GJ/hr (MMBtu/hr)[1]			
Coal, HHV	5,876 (5,570)	5,876 (5,570)	5,876 (5,570)
Sensible + Latent			
Coal	3.0 (2.9)	3.0 (2.9)	3.0 (2.9)
Air	38 (36)	38 (36)	38 (36)
Raw Water Makeup	35 (33)	42 (40)	47 (45)
Soda Ash	0.03 (0.03)	0.03 (0.03)	0.03 (0.03)
Auxiliary Power	165 (156)	332 (315)	408 (386)
Total In	**6,117 (5,798)**	**6,293 (5,964)**	**6,372 (6,040)**
Energy Out, GJ/hr (MMBtu/hr)[1]			
Sensible + Latent			
Bottom Ash	0.7 (0.7)	0.7 (0.7)	0.7 (0.7)
Fly Ash + FGD Ash	2.8 (2.6)	2.8 (2.6)	2.8 (2.6)
Flue Gas	1,023 (970)	465 (441)	214 (203)
Condenser	2,716 (2,574)	1,779 (1,686)	1,361 (1,290)
CO_2	N/A	-72 (-68)	-104 (-99)
Cooling Tower Blowdown	27 (26)	36 (34)	44 (42)
Econamine Losses	N/A	1,844 (1,748)	3,104 (2,942)
Process Losses[2]	*267 (253)*	*420 (398)*	*49 (46)*
Power	2,080 (1,972)	1,818 (1,723)	1,701 (1,613)
Total Out	**6,117 (5,798)**	**6,293 (5,964)**	**6,372 (6,040)**

[1] Enthalpy reference conditions are 0°C (32°F) and 614 Pa (0.089 psia)

[2] Process losses are calculated by difference to close the energy balance

7.1.6 Cases 7 - 9 Major Equipment List

Major equipment items for the existing subcritical PC plant with and without CO_2 capture are shown in the following tables. The Case 7 equipment list is an estimate of the existing subcritical PC plant using the same methodology as was used for the supercritical PC cases. Case 7 establishes a baseline equipment list that is used to determine the necessary plant equipment modifications. Cases 8 and 9 show the incremental increases due to the addition of equipment necessary for CO_2 capture. The accounts used in the equipment list correspond to the account numbers used in the cost estimates in Section 7.1.7. In general, the design conditions include a 10 percent contingency for flows and heat duties and a 21 percent contingency for heads on pumps and fans.

ACCOUNT 1 COAL AND SORBENT HANDLING

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
1	Feeder	Belt	2	0	572 tonne/h (630 tph)	No change	No change
2	Conveyor No. 1	Belt	1	0	1,134 tonne/h (1,250 tph)	No change	No change
3	Transfer Tower No. 1	Enclosed	1	0	N/A	N/A	N/A
4	Conveyor No. 2	Belt	1	0	1,134 tonne/h (1,250 tph)	No change	No change
5	As-Received Coal Sampling System	Two-stage	1	0	N/A	N/A	N/A
6	Stacker/Reclaimer	Traveling, linear	1	0	1,134 tonne/h (1,250 tph)	No change	No change
7	Reclaim Hopper	N/A	2	1	64 tonne (70 ton)	No change	No change
8	Feeder	Vibratory	2	1	245 tonne/h (270 tph)	No change	No change
9	Conveyor No. 3	Belt w/ tripper	1	0	490 tonne/h (540 tph)	No change	No change
10	Crusher Tower	N/A	1	0	N/A	N/A	N/A
11	Coal Surge Bin w/ Vent Filter	Dual outlet	2	0	245 tonne (270 ton)	No change	No change
12	Crusher	Impactor reduction	2	0	8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0)	No change	No change
13	As-Fired Coal Sampling System	Swing hammer	1	1	N/A	N/A	N/A
14	Conveyor No. 4	Belt w/tripper	1	0	490 tonne/h (540 tph)	No change	No change
15	Transfer Tower No. 2	Enclosed	1	0	N/A	N/A	N/A

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
16	Conveyor No. 5	Belt w/ tripper	1	0	490 tonne/h (540 tph)	No change	No change
17	Coal Silo w/ Vent Filter and Slide Gates	Field erected	3	0	1,089 tonne (1,200 ton)	No change	No change
18	Soda Ash Truck Unloading Hopper	N/A	1	0	36 tonne (40 ton)	No change	No change
19	Soda Ash Feeder	Belt	3	0	36 tonne/h (40 tph)	No change	No change
20	Soda Conveyor No. L1	Belt	1	0	36 tonne/h (40 tph)	No change	No change
21	Soda Reclaim Hopper	N/A	2	0	9 tonne (10 ton)	No change	No change
22	Soda Reclaim Feeder	Belt	1	0	27 tonne/h (30 tph)	No change	No change
23	Soda Conveyor No. L2	Belt	1	0	27 tonne/h (30 tph)	No change	No change
24	Soda Day Bin	w/ actuator	2	0	109 tonne (120 ton)	No change	No change

ACCOUNT 2 COAL AND SORBENT PREPARATION AND FEED

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
1	Coal Feeder	Gravimetric	6	0	54 tonne/h (60 tph)	No change	No change
2	Coal Pulverizer	Ball type or equivalent	6	0	54 tonne/h (60 tph)	No change	No change
3	Soda Weigh Feeder	Gravimetric	1	1	9 tonne/h (10 tph)	No change	No change
4	Soda Ball Mill	Rotary	1	1	9 tonne/h (10 tph)	No change	No change
5	Soda Mill Slurry Tank with Agitator	N/A	1	1	37,854 liters (10,000 gal)	No change	No change

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
6	Soda Mill Recycle Pumps	Horizontal centrifugal	1	1	606 lpm @ 12m H_2O (160 gpm @ 40 ft H_2O)	No change	No change
7	Hydroclone Classifier	4 active cyclones in a 5 cyclone bank	1	1	151 lpm (40 gpm) per cyclone	No change	No change
8	Distribution Box	2-way	1	1	N/A	N/A	N/A
9	Soda Slurry Storage Tank with Agitator	Field erected	1	1	189,271 liters (50,000 gal)	No change	No change
10	Soda Slurry Feed Pumps	Horizontal centrifugal	1	1	416 lpm @ 9 m H_2O (110 gpm @ 30 ft H_2O)	No change	No change

ACCOUNT 3 FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
1	Demineralized Water Storage Tank	Vertical, cylindrical, outdoor	2	0	1,177,263 liters (311,000 gal)	No change	No change
2	Condensate Pumps	Vertical canned	1	1	27,255 lpm @ 335 m H_2O (7,200 gpm @ 1,100 ft H_2O)	No change	No change
3	Deaerator and Storage Tank	Horizontal spray type	1	0	1,958,158 kg/h (4,317,000 lb/h), 5 min. tank	No change	No change

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
4	Boiler Feed Pump/Turbine	Barrel type, multi-stage, centrifugal	1	1	32,933 lpm @ 2,530 m H_2O (8,700 gpm @ 8,300 ft H_2O)	No change	No change
5	Startup Boiler Feed Pump, Electric Motor Driven	Barrel type, multi-stage, centrifugal	1	0	9,842 lpm @ 2,530 m H_2O (2,600 gpm @ 8,300 ft H_2O)	No change	No change
6	LP Feedwater Heater 1A/1B	Horizontal U-tube	2	0	811,930 kg/h (1,790,000 lb/h)	No change	No change
7	LP Feedwater Heater 2A/2B	Horizontal U-tube	2	0	811,930 kg/h (1,790,000 lb/h)	No change	No change
8	LP Feedwater Heater 3A/3B	Horizontal U-tube	2	0	811,930 kg/h (1,790,000 lb/h)	No change	No change
9	LP Feedwater Heater 4A/4B	Horizontal U-tube	2	0	811,930 kg/h (1,790,000 lb/h)	No change	No change
10	HP Feedwater Heater 6	Horizontal U-tube	1	0	1,959,519 kg/h (4,320,000 lb/h)	No change	No change
11	HP Feedwater Heater 7	Horizontal U-tube	1	0	1,959,519 kg/h (4,320,000 lb/h)	No change	No change
12	Auxiliary Boiler	Shop fabricated, water tube	1	0	18,144 kg/h, 2.8 MPa, 343°C (40,000 lb/h, 400 psig, 650°F)	No change	No change
13	Fuel Oil System	No. 2 fuel oil for light off	1	0	1,135,624 liter (300,000 gal)	No change	No change

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
14	Service Air Compressors	Flooded Screw	2	1	28 m^3/min @ 0.7 MPa (1,000 scfm @ 100 psig)	No change	No change
15	Instrument Air Dryers	Duplex, regenerative	2	1	28 m^3/min (1,000 scfm)	No change	No change
16	Closed Cycle Cooling Heat Exchangers	Shell and tube	2	0	53 GJ/h (50 MMBtu/h) each	No change	No change
17	Closed Cycle Cooling Water Pumps	Horizontal centrifugal	2	1	20,820 lpm @ 30 m H_2O (5,500 gpm @ 100 ft H_2O)	No change	No change
18	Engine-Driven Fire Pump	Vertical turbine, diesel engine	1	1	3,785 lpm @ 88 m H_2O (1,000 gpm @ 290 ft H_2O)	No change	No change
19	Fire Service Booster Pump	Two-stage horizontal centrifugal	1	1	2,650 lpm @ 64 m H_2O (700 gpm @ 210 ft H_2O)	No change	No change
20	Raw Water Pumps	Stainless steel, single suction	2	1	13,627 lpm @ 43 m H_2O (3,600 gpm @ 140 ft H_2O)	Δ3,634 lpm @ 43 m H_2O (Δ1,020 gpm @ 140 ft H_2O)	Δ7,344 lpm @ 43 m H_2O (Δ1,950 gpm @ 140 ft H_2O)
21	Filtered Water Pumps	Stainless steel, single suction	2	1	1,779 lpm @ 49 m H_2O (470 gpm @ 160 ft H_2O)	No change	No change
22	Filtered Water Tank	Vertical, cylindrical	1	0	1,688,294 liter (446,000 gal)	No change	No change

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
23	Makeup Water Demineralizer	Multi-media filter, cartridge filter, RO membrane assembly, electrodeionization unit	1	1	644 lpm (170 gpm)	Δ 113 lpm (Δ 30 gpm)	Δ 151 lpm (Δ 40 gpm)
24	Liquid Waste Treatment System	--	1	0	10 years, 24-hour storm	No change	No change

ACCOUNT 4 BOILER AND ACCESSORIES

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
1	Boiler	Subcritical, drum, wall-fired, low NOx burners, overfire air	1	0	1,959,519 kg/h steam @ 16.5 MPa/538°C/538°C (4,320,000 lb/h steam @ 2,400 psig /1,000°F /1,000°F)	No change	No change
2	Primary Air Fan	Centrifugal	2	0	286,670 kg/h, 4,873 m^3/min @ 102 cm WG (632,000 lb/h, 172,100 acfm @ 40 in. WG)	No change	No change
3	Forced Draft Fan	Centrifugal	2	0	933,493 kg/h, 15,866 m^3/min @ 51 cm WG (2,058,000 lb/h, 560,300 acfm @ 20 in. WG)	No change	No change

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
4	Induced Draft Fan	Centrifugal	2	0	1,390,261 kg/h, 39,312 m^3/min @ 128 cm WG (3,065,000 lb/h, 1,388,300 acfm @ 50 in. WG)	No change	No change
5	SCR Reactor Vessel	Space for spare layer	2	0	N/A	N/A	N/A
6	SCR Catalyst	--	3	0	--	N/A	N/A
7	Dilution Air Blower	Centrifugal	2	1	N/A	N/A	N/A
8	Ammonia Storage	Horizontal tank	5	0	N/A	N/A	N/A
9	Ammonia Feed Pump	Centrifugal	2	1	N/A	N/A	N/A

ACCOUNT 5 FLUE GAS CLEANUP

Equipment No.	**Description**	**Type**	**Operating Qty.**	**Spares**	**Case 7 Design Condition**	**Case 8 Design Condition**	**Case 9 Design Condition**
1	Electrostatic Precipitator	Cold-side	2	0	1,390,261 kg/h (3,065,000 lb/h) 99.65% efficiency	No change	No change
2	Absorber Module	Counter-current open spray	1	0	57,625 m^3/min (2,035,000 acfm)	Upgraded	Upgraded

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
3	Recirculation Pumps	Horizontal centrifugal	5	1	200,627 lpm @ 64 m H_2O (53,000 gpm @ 210 ft H_2O)	No change	No change
4	Bleed Pumps	Horizontal centrifugal	2	1	1,401 lpm (370 gpm) at 20 wt% solids	No change	No change
5	Oxidation Air Blowers	Centrifugal	2	1	100 m^3/min @ 0.3 MPa (3,520 acfm @ 42 psia)	No change	No change
6	Agitators	Side entering	5	1	50 hp	No change	No change
7	Dewatering Cyclones	Radial assembly, 5 units each	2	0	341 lpm (90 gpm) per cyclone	No change	No change
8	Vacuum Filter Belt	Horizontal belt	2	1	11 tonne/h (12 tph) of 50 wt % slurry	No change	No change
9	Filtrate Water Return Pumps	Horizontal centrifugal	1	1	227 lpm @ 12 m H_2O (60 gpm @ 40 ft H_2O)	No change	No change
10	Filtrate Water Return Storage Tank	Vertical, lined	1	0	151,416 lpm (40,000 gal)	No change	No change

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
11	Process Makeup Water Pumps	Horizontal centrifugal	1	1	341 lpm @ 21 m H_2O (90 gpm @ 70 ft H_2O)	No change	No change

ACCOUNT 5B CO$_2$ REMOVAL AND COMPRESSION

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
1	Econamine FG Plus	Amine-based CO_2 capture technology	2	0	N/A	1,023,758 kg/h (2,257,000 lb/h) 20.2 wt % CO_2 concentration	1,474,629 kg/h (3,251,000 lb/h) 20.3 wt % CO_2 concentration
2	Let-Down Turbine	Commercially available	1	0	N/A	30 MW 1.2 MPa/367°C (168 psig/ 693°F)	43 MW 1.2 MPa/367°C (168 psig/ 693°F)
3	Let-Down Turbine Generator		1	0	N/A	30 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	50 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase
4	Econamine Condensate Pump	Centrifugal	1	1	N/A	10,486 lpm @ 49 m H_2O (2,770 gpm @ 160 ft H_2O)	15,180 lpm @ 49 m H_2O (4,010 gpm @ 160 ft H_2O)
5	CO2 Compressor	Reciprocating	2	0	N/A	186,533 kg/h @ 15.3 MPa (411,234 lb/h @ 2,215 psia)	269,790 kg/h @ 15.3 MPa (594,785 lb/h @ 2,215 psia)

ACCOUNT 5C CO$_2$ TRANSPORT, STORAGE, AND MONITORING (not shown in Total Plant Cost Details)

Equipment No.	Description	Type	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
1	CO$_2$ Pipeline	Carbon Steel	N/A	50 miles @ 12 in diameter w/ inlet pressure of 2,200 psi and outlet pressure of 1,200 psi	50 miles @ 14 in diameter w/ inlet pressure of 2,200 psi and outlet pressure of 1,200 psi
2	CO$_2$ Sequestration Source	Saline Formation	N/A	1 well with bottom hole pressure @ 1,220 psi, 530 ft thickness, 4,055 ft depth, 22 Md permeability	2 wells with bottom hole pressure @ 1,220 psi, 530 ft thickness, 4,055 ft depth, 22 Md permeability
3	CO$_2$ Monitoring	N/A	N/A	20 year monitoring life during plant life / 80 years following / Total of 100 years	20 year monitoring life during plant life / 80 years following / Total of 100 years

ACCOUNT 6 COMBUSTION TURBINE/ACCESSORIES

N/A

ACCOUNT 7 HRSG, DUCTING & STACK

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
1	Stack	Reinforced concrete	1	0	152 m (500 ft) high x 7.1 m (23 ft) diameter	New stack liner for wet operation	New stack liner for wet operation

ACCOUNT 8 STEAM TURBINE GENERATOR AND AUXILIARIES

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
1	Steam Turbine	Commercially available advanced steam turbine	1	0	608 MW 16.5 MPa/538°C/538°C (2400 psig/ 1000°F/1000°F)	No change	No change
2	Steam Turbine Generator	Hydrogen cooled, static excitiation	1	0	680 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	No change	No change
3	Surface Condenser	Single pass, divided waterbox including vacuum pumps	1	0	2,986 GJ/h (2,830 MMBtu/h), Inlet water temperature 09°C (48°F), Water temperature rise 11°C (20°F)	No change	No change

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
1	Circulating Water Pumps	Vertical, wet pit	2	1	556,456 lpm @ 30 m (147,000 gpm @ 100 ft)	Δ355,828 lpm @ 30 m (Δ94,000 gpm @ 100 ft)	Δ688,944 lpm @ 30 m (Δ182,000gpm @ 100 ft)
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	1	0	3°C (37°F) wet bulb / 9°C (48°F) CWT / 20°C (68°F) HWT / 3102 GJ/h (2,940 MMBtu/h) heat duty	Δ1,002 GJ/h (Δ950 MMBtu/h) heat duty	Δ1,920 GJ/h (Δ1,820 MMBtu/h) heat duty

ACCOUNT 10 ASH/SPENT SORBENT RECOVERY AND HANDLING

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
1	Economizer Hopper (part of boiler scope of supply)	--	4	0	--	--	--
2	Bottom Ash Hopper (part of boiler scope of supply)	--	2	0	--	--	--
3	Clinker Grinder	--	1	1	5.4 tonne/h (6 tph)	No change	No change
4	Pyrites Hopper (part of pulverizer scope of supply included with boiler)	--	6	0	--	--	--

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
5	Hydroejectors	--	12		--	--	--
6	Economizer /Pyrites Transfer Tank	--	1	0	--	--	--
7	Ash Sluice Pumps	Vertical, wet pit	1	1	227 lpm @ 17 m H_2O (60 gpm @ 56 ft H_2O)	No change	No change
8	Ash Seal Water Pumps	Vertical, wet pit	1	1	7,571 lpm @ 9 m H_2O (2,000 gpm @ 28 ft H_2O)	No change	No change
9	Hydrobins	--	1	1	227 lpm (60 gpm)	No change	No change
10	Baghouse Hopper (part of baghouse scope of supply)	--	24	0	--	--	--
11	Air Heater Hopper (part of boiler scope of supply)	--	10	0	--	--	--
12	Air Blower	--	1	1	20 m^3/min @ 0.2 MPa (690 scfm @ 24 psi)	No change	No change
13	Fly Ash Silo	Reinforced concrete	2	0	635 tonne (1,400 ton)	No change	No change
14	Slide Gate Valves	--	2	0	--	--	--
15	Unloader	--	1	0	--	--	--
16	Telescoping Unloading Chute	--	1	0	118 tonne/h (130 tph)	No change	No change

ACCOUNT 11 ACCESSORY ELECTRIC PLANT

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
1	STG Transformer	Oil-filled	1	0	24 kV/345 kV, 620 MVA, 3-ph, 60 Hz	No change	No change
2	Auxiliary Transformer	Oil-filled	1	1	24 kV/4.16 kV, 49 MVA, 3-ph, 60 Hz	Δ52 MVA	Δ75 MVA
3	Low Voltage Transformer	Dry ventilated	1	1	4.16 kV/480 V, 7 MVA, 3-ph, 60 Hz	Δ8 MVA	Δ12 MVA
4	STG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	1	0	24 kV, 3-ph, 60 Hz	No change	No change
5	Medium Voltage Switchgear	Metal clad	1	1	4.16 kV, 3-ph, 60 Hz	As required	As required
6	Low Voltage Switchgear	Metal enclosed	1	1	480 V, 3-ph, 60 Hz	As required	As required
7	Emergency Diesel Generator	Sized for emergency shutdown	1	0	750 kW, 480 V, 3-ph, 60 Hz	No change	No change

ACCOUNT 12 INSTRUMENTATION AND CONTROL

Equipment No.	Description	Type	Operating Qty.	Spares	Case 7 Design Condition	Case 8 Design Condition	Case 9 Design Condition
1	DCS - Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W)	1	0	Operator stations/printers and engineering stations/printers	As required	As required
2	DCS - Processor	Microprocessor with redundant input/output	1	0	N/A	As required	As required
3	DCS - Data Highway	Fiber optic	1	0	Fully redundant, 25% spare	As required	As required

7.1.7 Case 7 – Cost Estimating

Case 7 represents the existing subcritical PC plant. The Energy Velocity Database reports the current cost of electricity for a typical subcritical PC plant as $19.10/MWh [34]. An estimated value for insurances and taxes was added to the fixed costs from the Energy Velocity Database to obtain the total current COE. The original plant capital costs are assumed to be paid for in this report and the COE was multiplied by the appropriate levelization factor to obtain the LCOE. The estimated O&M costs for the subcritical PC plant were distributed as follows (as explained in Section 2.7) and then levelized:

- Variable operating costs: $1.48/MWh
- Fixed operating costs: $13.16/MWh (includes $7.19/MWh for taxes and insurance)
- Fuel operating costs: $19.14/MWh
- Total LCOE: $33.78/MWh

7.1.8 Case 8 – Cost Estimating

The cost estimating methodology was described previously in Section 2.6. Exhibit 7-21 shows the total plant capital cost details organized by cost account as well as TOC and TASC. The costs represent the TOC for retrofitting the Econamine process and ancillary equipment. Exhibit 7-22 shows the initial and annual O&M costs.

The estimated TOC for adding carbon capture and sequestration to the existing subcritical PC plant with a CO_2 emission rate of 1,100 lb/net-MWh is $1,348/kW. Owner's costs represent 18 percent of the TOC. In the event that NSR is triggered and the addition of an SCR unit is necessary to achieve the NOx emission rate of 0.07 lb/MMBtu, the TOC for Case 8 including carbon capture and sequestration as well as the SCR unit would be $1,717/kW, which represents an increase of 27.4 percent over the no SCR case ($369/kW increase). The estimated cost for the SCR retrofit is taken from the referenced BART analysis and appears high (approximately $127 million, or $308/kW) [8]. The current dollar, 30-year LCOE, without SCR but including CO_2 TS&M costs, is $84.81/MWh. The net plant output is decreased by 22 percent because of the carbon capture and compression equipment. The current dollar, 30-year LCOE including SCR and TS&M costs is $94.01/MWh.

Exhibit 7-21 Case 8 Total Plant Cost Details

											Case 8	
Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$	$/kW
1	COAL & SORBENT HANDLING											
1.1	Coal Receive & Unload	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.2	Coal Stackout & Reclaim	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.3	Coal Conveyors	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.4	Other Coal Handling	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.5	Sorbent Receive & Unload	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.6	Sorbent Stackout & Reclaim	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.7	Sorbent Conveyors	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.8	Other Sorbent Handling	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.9	Coal & Sorbent Hnd.Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 1.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
2	COAL & SORBENT PREP & FEED											
2.1	Coal Crushing & Drying	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.2	Coal Conveyor to Storage	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.3	Coal Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.4	Misc.Coal Prep & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.5	Sorbent Prep Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.6	Sorbent Storage & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.7	Sorbent Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.8	Booster Air Supply System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.9	Coal & Sorbent Feed Foundation	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 2.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
3	FEEDWATER & MISC. BOP SYSTEMS											
3.1	FeedwaterSystem	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
3.2	Water Makeup & Pretreating	$1,793	$0	$577	$0	$0	$2,370	$224	$0	$519	$3,114	$8
3.3	Other Feedwater Subsystems	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
3.4	Service Water Systems	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
3.5	Other Boiler Plant Systems	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
3.6	FO Supply Sys & Nat Gas	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
3.7	Waste Treatment Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
3.8	Misc. Equip.(cranes,AirComp.,Comm.)	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 3.	**$1,793**	**$0**	**$577**	**$0**	**$0**	**$2,370**	**$224**	**$0**	**$519**	**$3,114**	**$8**
4	PC BOILER											
4.1	PC Boiler & Accessories	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.2	LNB's and OFA	$2,982	$0	$2,544	$0	$0	$5,526	$0	$0	$382	$5,907	$14
4.3	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.4	Boiler BoP (w/ ID Fans)	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.5	Primary Air System	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.6	Secondary Air System	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.8	Major Component Rigging	$0	w/4.1	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.9	Boiler Foundations	$0	w/14.1	w/14.1	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 4.	**$2,982**	**$0**	**$2,544**	**$0**	**$0**	**$5,526**	**$0**	**$0**	**$382**	**$5,907**	**$14**

Exhibit 7-21 Case 8 Total Plant Cost Details (Continued)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		Case 8 TOTAL PLANT COST	
				Direct	Indirect				Process	Project	$	$/kW
5	FLUE GAS CLEANUP											
5.1	Absorber Vessels & Accessories	$3,549	$0	$199	$0	$0	$3,748	$298	$0	$809	$4,855	$12
5.2	Other FGD	$100	$0	$0	$0	$0	$100	$0	$0	$0	$100	$0
5.3	Bag House & Accessories	w/5.1	$0	w/5.1	$0	$0	$0	$0	$0	$0	$0	$0
5.4	Other Particulate Removal Materials	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.5	Gypsum Dewatering System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.6	Mercury Removal System	w/5.1	$0	w/5.1	$0	$0	$0	$0	$0	$0	$0	$0
5.9	Open											
	SUBTOTAL 5.	**$3,649**	**$0**	**$199**	**$0**	**$0**	**$3,848**	**$298**	**$0**	**$809**	**$4,955**	**$12**
5B	CO_2 REMOVAL & COMPRESSION											
5B.1	CO_2 Removal System	$170,551	$0	$51,748	$0	$0	$222,299	$21,254	$44,460	$57,602	$345,614	$838
5B.2	CO_2 Compression & Drying	$20,574	$0	$6,454	$0	$0	$27,028	$2,585	$0	$5,923	$35,536	$86
5B.3	CO_2 Removal System Let Down Turbine	$9,900	$0	$1,315	$0	$0	$11,215	$1,075	$0	$1,229	$13,519	$33
	SUBTOTAL 5B.	**$201,025**	**$0**	**$59,517**	**$0**	**$0**	**$260,542**	**$24,913**	**$44,460**	**$64,754**	**$394,669**	**$957**
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Combustion Turbine Generator	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
6.2	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.3	Compressed Air Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.9	Combustion Turbine Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 6.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
7.2	HRSG Accessories	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.3	Ductwork	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.4	Stack	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.9	Duct & Stack Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 7.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
8.2	Turbine Plant Auxiliaries	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
8.3a	Condenser & Auxiliaries	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
8.3b	Air Cooled Condenser	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
8.4	Steam Piping	$2,353	$0	$1,160	$0	$0	$3,514	$295	$0	$571	$4,380	$11
8.9	TG Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 8.	**$2,353**	**$0**	**$1,160**	**$0**	**$0**	**$3,514**	**$295**	**$0**	**$571**	**$4,380**	**$11**

Exhibit 7-21 Case 8 Total Plant Cost Details (Continued)

Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		Case 8 TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$	$/kW
9	COOLING WATER SYSTEM											
9.1	Cooling Towers	$4,602	$0	$1,433	$0	$0	$6,035	$577	$0	$661	$7,273	$18
9.2	Circulating Water Pumps	$989	$0	$94	$0	$0	$1,083	$92	$0	$117	$1,292	$3
9.3	Circ.Water System Auxiliaries	$299	$0	$40	$0	$0	$339	$32	$0	$37	$408	$1
9.4	Circ.Water Piping	$0	$2,371	$2,298	$0	$0	$4,668	$437	$0	$766	$5,871	$14
9.5	Make-up Water System	$219	$0	$293	$0	$0	$512	$49	$0	$84	$645	$2
9.6	Component Cooling Water Sys	$237	$0	$188	$0	$0	$425	$40	$0	$70	$535	$1
9.9	Circ.Water System Foundations& Structures	$0	$1,413	$2,244	$0	$0	$3,657	$346	$0	$801	$4,803	$12
	SUBTOTAL 9.	**$6,345**	**$3,783**	**$6,590**	**$0**	**$0**	**$16,719**	**$1,574**	**$0**	**$2,536**	**$20,828**	**$50**
10	ASH/SPENT SORBENT HANDLING SYS											
10.1	Ash Coolers	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.2	Cyclone Ash Letdown	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.3	HGCU Ash Letdown	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.4	High Temperature Ash Piping	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.5	Other Ash Recovery Equipment	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.6	Ash Storage Silos	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.7	Ash Transport & Feed Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.8	Misc. Ash Handling Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.9	Ash/Spent Sorbent Foundation	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 10.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
11	ACCESSORY ELECTRIC PLANT											
11.1	Generator Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
11.2	Station Service Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
11.3	Switchgear & Motor Control	$2,091	$0	$355	$0	$0	$2,446	$227	$0	$267	$2,940	$7
11.4	Conduit & Cable Tray	$0	$1,311	$4,532	$0	$0	$5,843	$566	$0	$961	$7,370	$18
11.5	Wire & Cable	$0	$2,473	$4,775	$0	$0	$7,248	$611	$0	$1,179	$9,038	$22
11.6	Protective Equipment	$7	$0	$22	$0	$0	$29	$3	$0	$3	$35	$0
11.7	Standby Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
11.8	Main Power Transformers	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
11.9	Electrical Foundations	$0	$24	$58	$0	$0	$81	$8	$0	$18	$107	$0
	SUBTOTAL 11.	**$2,097**	**$3,808**	**$9,743**	**$0**	**$0**	**$15,648**	**$1,414**	**$0**	**$2,428**	**$19,490**	**$47**
12	INSTRUMENTATION & CONTROL											
12.1	PC Control Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.2	Combustion Turbine Control	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
12.3	Steam Turbine Control	w/8.1	$0	w/8.1	$0	$0	$0	$0	$0	$0	$0	$0
12.4	Other Major Component Control	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
12.5	Signal Processing Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.6	Control Boards,Panels & Racks	$59	$0	$35	$0	$0	$95	$9	$5	$16	$125	$0
12.7	Distributed Control System Equipment	$598	$0	$104	$0	$0	$702	$65	$35	$80	$883	$2
12.8	Instrument Wiring & Tubing	$324	$0	$643	$0	$0	$967	$82	$48	$165	$1,262	$3
12.9	Other I & C Equipment	$169	$0	$383	$0	$0	$552	$54	$28	$63	$697	$2
	SUBTOTAL 12.	**$1,150**	**$0**	**$1,166**	**$0**	**$0**	**$2,316**	**$210**	**$116**	**$324**	**$2,966**	**$7**

Exhibit 7-21 Case 8 Total Plant Cost Details (Continued)

Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		Case 8 TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$	$/kW
13	IMPROVEMENTS TO SITE											
13.1	Site Preparation	$0	$40	$802	$0	$0	$843	$84	$0	$185	$1,112	$3
13.2	Site Improvements	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
13.3	Site Facilities	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 13.	**$0**	**$40**	**$802**	**$0**	**$0**	**$843**	**$84**	**$0**	**$185**	**$1,112**	**$3**
14	BUILDINGS & STRUCTURES											
14.1	Boiler Building	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
14.2	Turbine Building	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
14.3	Administration Building	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
14.4	Circulation Water Pumphouse	$0	$148	$118	$0	$0	$266	$24	$0	$43	$333	$1
14.5	Water Treatment Buildings	$0	$226	$186	$0	$0	$412	$37	$0	$67	$516	$1
14.6	Machine Shop	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
14.7	Warehouse	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
14.8	Other Buildings & Structures	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
14.9	Waste Treating Building & Str.	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 14.	**$0**	**$374**	**$304**	**$0**	**$0**	**$678**	**$61**	**$0**	**$111**	**$849**	**$2**
	TOTAL COST	**$221,395**	**$8,005**	**$82,602**	**$0**	**$0**	**$312,002**	**$29,072**	**$44,576**	**$72,620**	**$458,271**	**$1,110.9**
	Owner's Costs											
	Preproduction Costs											
	6 Months All Labor										$2,285	$6
	1 Month Maintenance Materials										$401	$1
	1 Month Non-fuel Consumables										$367	$1
	1 Month Waste Disposal										$6	$0
	25% of 1 Months Fuel Cost at 100% CF										$0	$0
	2% of TPC										$9,165	$22
	Total										**$12,224**	**$30**
	Inventory Capital											
	60 day supply of consumables at 100% CF										$574	$1
	0.5% of TPC (spare parts)										$2,291	$6
	Total										**$2,865**	**$7**
	Initial Cost for Catalyst and Chemicals										$1,518	$4
	Land										$0	$0
	Other Owner's Costs										$68,741	$167
	Financing Costs										$12,373	$30
	Total Overnight Costs (TOC)										**$555,992**	**$1,348**
	TASC Multiplier								(IOU, high risk, 33 year)		1.078	
	Total As-Spent Cost (TASC)										**$599,360**	**$1,453**

Exhibit 7-22 Case 8 Initial and Annual Operating and Maintenance Costs

INITIAL & ANNUAL O&M EXPENSES					Cost Base (June)	2007
Case 8 - Subcritical PC w/ CO2 capture (1,100 lb/net MWh) - Retrofit					Heat Rate-net(Btu/kWh):	13,495
					MWe-net:	413
					Capacity Factor: (%):	85
OPERATING & MAINTENANCE LABOR						
Operating Labor						
Operating Labor Rate(base):	34.65	$/hour				
Operating Labor Burden:	30.00	% of base				
Labor O-H Charge Rate:	25.00	% of labor				
			Total			
Skilled Operator	1.0		1.0			
Operator	1.3		1.3			
Foreman	0.0		0.0			
Lab Tech's, etc.	0.0		0.0			
TOTAL-O.J.'s	2.3		2.3			
					Annual Cost	Annual Unit Cost
					$	$/kW-net
Annual Operating Labor Cost	Maintenance labor cost		% of BEC	0.8889	$907,567	$2.200
Maintenance Labor Cost	(Case S12B is reference)		BEC	$302,629	$2,749,062	$6.664
Administrative & Support Labor					$914,157	$2.216
Property Taxes & Insurance					$27,542,493	$66.763
TOTAL FIXED OPERATING COSTS					**$32,113,279**	**$77.843**
VARIABLE OPERATING COSTS						
						$/kWh-net
Maintenance Material Cost			% of BEC	1.3333	**$4,088,593**	**$0.00133**
Consumables	Consumption		Unit	Initial Fill		
	Initial Fill	/Day	Cost	Cost		
Water(/1000 gallons)	0	2,153	1.22	**$0**	**$814,846**	**$0.00027**
Chemicals		4.841				
MU & WT Chem.(lb)	0	10,421	0.17	$0	$559,545	$0.00018
Soda Ash (ton)	0	8	80.00	$0	$201,936	$0.00007
Carbon (Mercury Removal) (lb)	0	0	1.05	$0	$0	$0.00000
MEA Solvent (ton)	636	0.90	2,249.89	$1,430,574	$628,224	$0.00020
NaOH (tons)	0	7.41	433.68	$0	$997,006	$0.00032
H2SO4 (tons)	0	4.28	138.78	$0	$184,282	$0.00006
Corrosion Inhibitor	0	0	0.00	$87,799	$4,181	$0.00000
Activated Carbon(lb)	0	1,076	1.05	$0	$350,577	$0.00011
Ammonia (19% NH3) ton	0	0	129.80	$0	$0	$0.00000
Subtotal Chemicals				**$1,518,373**	**$2,925,751**	**$0.00095**
Other						
Supplemental Fuel(MBtu)	0	0	0.00	$0	$0	$0.00000
SCR Catalyst(m3)	w/equip.	0.000	5,775.94	$0	$0	$0.00000
Emission Penalties	0	0	0.00	$0	$0	$0.00000
Subtotal Other				**$0**	**$0**	**$0.00000**
Waste Disposal						
Flyash (ton)	0	8	24.33	$0	$61,414	$0.00002
Bottom Ash(ton)	0	0	16.23	$0	$0	$0.00000
Subtotal-Waste Disposal				**$0**	**$61,414**	**$0.00002**
By-products & Emissions						
Gypsum (tons)	0	0	0.00	$0	$0	$0.00000
Subtotal By-Products				**$0**	**$0**	**$0.00000**
TOTAL VARIABLE OPERATING COSTS				**$1,518,373**	**$7,890,604**	**$0.00257**
Fuel(ton)	0	0	10.37	**$0**	**$0**	**$0.00000**

7.1.9 Case 9 – Cost Estimating

Exhibit 7-23 shows the TPC cost details organized by cost account along with TOC and TASC. Exhibit 7-24 shows the initial and annual O&M costs.

The estimated TOC for adding CC&S to the existing subcritical PC boiler with a CO_2 capture level of 90 percent is $1,999/kW. In the event that NSR is triggered and the addition of an SCR unit is necessary to achieve the NOx emission rate of 0.07 lb/MMBtu, the TOC for Case 9 increases to $2,430/kW, which represents an increase of 21.5 percent ($431/kW increase). Owner's costs represent 18 percent of the TOC. The current dollar 30-year LCOE, excluding SCR but including CO_2 TS&M, is $111.65/MWh. The net plant output is decreased by 32.4 percent in the 90 percent capture case. The current dollar, 30-year LCOE, including SCR and CO_2 TS&M, is $122.46/MWh.

Exhibit 7-23 Case 9 Total Plant Cost Details

											Case 9	
Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$	$/kW
1	COAL & SORBENT HANDLING											
1.1	Coal Receive & Unload	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.2	Coal Stackout & Reclaim	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.3	Coal Conveyors	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.4	Other Coal Handling	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.5	Sorbent Receive & Unload	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.6	Sorbent Stackout & Reclaim	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.7	Sorbent Conveyors	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.8	Other Sorbent Handling	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.9	Coal & Sorbent Hnd.Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 1.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
2	COAL & SORBENT PREP & FEED											
2.1	Coal Crushing & Drying	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.2	Coal Conveyor to Storage	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.3	Coal Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.4	Misc.Coal Prep & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.5	Sorbent Prep Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.6	Sorbent Storage & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.7	Sorbent Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.8	Booster Air Supply System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.9	Coal & Sorbent Feed Foundation	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 2.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
3	FEEDWATER & MISC. BOP SYSTEMS											
3.1	FeedwaterSystem	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
3.2	Water Makeup & Pretreating	$2,506	$0	$807	$0	$0	$3,312	$313	$0	$725	$4,351	$12
3.3	Other Feedwater Subsystems	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
3.4	Service Water Systems	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
3.5	Other Boiler Plant Systems	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
3.6	FO Supply Sys & Nat Gas	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
3.7	Waste Treatment Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
3.8	Misc. Equip.(cranes,AirComp.,Comm.)	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 3.	**$2,506**	**$0**	**$807**	**$0**	**$0**	**$3,312**	**$313**	**$0**	**$725**	**$4,351**	**$12**
4	PC BOILER											
4.1	PC Boiler & Accessories	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.2	LNB's and OFA	$2,982	$0	$2,544	$0	$0	$5,526	$0	$0	$382	$5,907	$16
4.3	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.4	Boiler BoP (w/ ID Fans)	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.5	Primary Air System	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.6	Secondary Air System	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.8	Major Component Rigging	$0	w/4.1	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.9	Boiler Foundations	$0	w/14.1	w/14.1	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 4.	**$2,982**	**$0**	**$2,544**	**$0**	**$0**	**$5,526**	**$0**	**$0**	**$382**	**$5,907**	**$16**

Exhibit 7-23 Case 9 Total Plant Cost Details (Continued)

Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		Case 9 TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$	$/kW
5	FLUE GAS CLEANUP											
5.1	Absorber Vessels & Accessories	$3,549	$0	$199	$0	$0	$3,748	$298	$0	$809	$4,855	$14
5.2	Other FGD	$100	$0	$0	$0	$0	$100	$0	$0	$0	$100	$0
5.3	Bag House & Accessories	w/5.1	$0	w/5.1	$0	$0	$0	$0	$0	$0	$0	$0
5.4	Other Particulate Removal Materials	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.5	Gypsum Dewatering System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.6	Mercury Removal System	w/5.1	$0	w/5.1	$0	$0	$0	$0	$0	$0	$0	$0
5.9	Open											
	SUBTOTAL 5.	**$3,649**	**$0**	**$199**	**$0**	**$0**	**$3,848**	**$298**	**$0**	**$809**	**$4,955**	**$14**
5B	CO_2 REMOVAL & COMPRESSION											
5B.1	CO_2 Removal System	$223,265	$0	$67,742	$0	$0	$291,006	$27,823	$58,201	$75,406	$452,436	$1,259
5B.2	CO_2 Compression & Drying	$25,763	$0	$8,082	$0	$0	$33,845	$3,237	$0	$7,416	$44,498	$124
5B.3	CO_2 Removal System Let Down Turbine	$10,450	$0	$1,389	$0	$0	$11,839	$1,135	$0	$1,297	$14,270	$40
	SUBTOTAL 5B.	**$259,478**	**$0**	**$77,212**	**$0**	**$0**	**$336,690**	**$32,194**	**$58,201**	**$84,120**	**$511,205**	**$1,422**
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Combustion Turbine Generator	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
6.2	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.3	Compressed Air Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.9	Combustion Turbine Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 6.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
7.2	HRSG Accessories	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.3	Ductwork	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.4	Stack	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.9	Duct & Stack Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 7.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
8.2	Turbine Plant Auxiliaries	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
8.3a	Condenser & Auxiliaries	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
8.3b	Air Cooled Condenser	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
8.4	Steam Piping	$3,045	$0	$1,501	$0	$0	$4,546	$382	$0	$739	$5,667	$16
8.9	TG Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 8.	**$3,045**	**$0**	**$1,501**	**$0**	**$0**	**$4,546**	**$382**	**$0**	**$739**	**$5,667**	**$16**

Exhibit 7-23 Case 9 Total Plant Cost Details (Continued)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		Case 9 TOTAL PLANT COST	
				Direct	Indirect				Process	Project	$	$/kW
5	FLUE GAS CLEANUP											
5.1	Absorber Vessels & Accessories	$3,549	$0	$199	$0	$0	$3,748	$298	$0	$809	$4,855	$14
5.2	Other FGD	$100	$0	$0	$0	$0	$100	$0	$0	$0	$100	$0
5.3	Bag House & Accessories	w/5.1	$0	w/5.1	$0	$0	$0	$0	$0	$0	$0	$0
5.4	Other Particulate Removal Materials	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.5	Gypsum Dewatering System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.6	Mercury Removal System	w/5.1	$0	w/5.1	$0	$0	$0	$0	$0	$0	$0	$0
5.9	Open											
9	COOLING WATER SYSTEM											
9.1	Cooling Towers	$7,254	$0	$2,259	$0	$0	$9,513	$910	$0	$1,042	$11,465	$32
9.2	Circulating Water Pumps	$1,545	$0	$147	$0	$0	$1,692	$143	$0	$184	$2,019	$6
9.3	Circ.Water System Auxiliaries	$444	$0	$59	$0	$0	$503	$48	$0	$55	$606	$2
9.4	Circ.Water Piping	$0	$3,521	$3,412	$0	$0	$6,933	$649	$0	$1,137	$8,720	$24
9.5	Make-up Water System	$291	$0	$388	$0	$0	$679	$65	$0	$112	$856	$2
9.6	Component Cooling Water Sys	$351	$0	$280	$0	$0	$631	$60	$0	$104	$795	$2
9.9	Circ.Water System Foundations& Structures	$0	$2,093	$3,326	$0	$0	$5,419	$513	$0	$1,186	$7,118	$20
	SUBTOTAL 9.	**$9,885**	**$5,614**	**$9,871**	**$0**	**$0**	**$25,371**	**$2,388**	**$0**	**$3,820**	**$31,579**	**$88**
10	ASH/SPENT SORBENT HANDLING SYS											
10.1	Ash Coolers	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.2	Cyclone Ash Letdown	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.3	HGCU Ash Letdown	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.4	High Temperature Ash Piping	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.5	Other Ash Recovery Equipment	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.6	Ash Storage Silos	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.7	Ash Transport & Feed Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.8	Misc. Ash Handling Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.9	Ash/Spent Sorbent Foundation	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 10.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
11	ACCESSORY ELECTRIC PLANT											
11.1	Generator Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
11.2	Station Service Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
11.3	Switchgear & Motor Control	$2,453	$0	$417	$0	$0	$2,870	$266	$0	$314	$3,450	$10
11.4	Conduit & Cable Tray	$0	$1,538	$5,318	$0	$0	$6,856	$664	$0	$1,128	$8,647	$24
11.5	Wire & Cable	$0	$2,902	$5,602	$0	$0	$8,504	$716	$0	$1,383	$10,604	$30
11.6	Protective Equipment	$7	$0	$22	$0	$0	$29	$3	$0	$3	$35	$0
11.7	Standby Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
11.8	Main Power Transformers	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
11.9	Electrical Foundations	$0	$30	$74	$0	$0	$105	$10	$0	$23	$138	$0
	SUBTOTAL 11.	**$2,460**	**$4,470**	**$11,433**	**$0**	**$0**	**$18,363**	**$1,659**	**$0**	**$2,851**	**$22,873**	**$64**
12	INSTRUMENTATION & CONTROL											
12.1	PC Control Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.2	Combustion Turbine Control	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
12.3	Steam Turbine Control	w/8.1	$0	w/8.1	$0	$0	$0	$0	$0	$0	$0	$0
12.4	Other Major Component Control	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
12.5	Signal Processing Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.6	Control Boards,Panels & Racks	$62	$0	$37	$0	$0	$99	$9	$5	$17	$131	$0
12.7	Distributed Control System Equipment	$627	$0	$110	$0	$0	$736	$68	$37	$84	$925	$3
12.8	Instrument Wiring & Tubing	$340	$0	$674	$0	$0	$1,014	$86	$51	$173	$1,323	$4
12.9	Other I & C Equipment	$177	$0	$402	$0	$0	$579	$56	$29	$66	$731	$2
	SUBTOTAL 12.	**$1,206**	**$0**	**$1,223**	**$0**	**$0**	**$2,428**	**$220**	**$121**	**$340**	**$3,110**	**$9**

Exhibit 7-23 Case 9 Total Plant Cost Details (Continued)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		Case 9 TOTAL PLANT COST	
				Direct	Indirect				Process	Project	$	$/kW
5	FLUE GAS CLEANUP											
5.1	Absorber Vessels & Accessories	$3,549	$0	$199	$0	$0	$3,748	$298	$0	$809	$4,855	$14
5.2	Other FGD	$100	$0	$0	$0	$0	$100	$0	$0	$0	$100	$0
5.3	Bag House & Accessories	w/5.1	$0	w/5.1	$0	$0	$0	$0	$0	$0	$0	$0
5.4	Other Particulate Removal Materials	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.5	Gypsum Dewatering System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.6	Mercury Removal System	w/5.1	$0	w/5.1	$0	$0	$0	$0	$0	$0	$0	$0
5.9	Open											
13	IMPROVEMENTS TO SITE											
13.1	Site Preparation	$0	$43	$852	$0	$0	$894	$89	$0	$197	$1,179	$3
13.2	Site Improvements	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
13.3	Site Facilities	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 13.	**$0**	**$43**	**$852**	**$0**	**$0**	**$894**	**$89**	**$0**	**$197**	**$1,179**	**$3**
14	BUILDINGS & STRUCTURES											
14.1	Boiler Building	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
14.2	Turbine Building	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
14.3	Administration Building	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
14.4	Circulation Water Pumphouse	$0	$194	$154	$0	$0	$347	$31	$0	$57	$435	$1
14.5	Water Treatment Buildings	$0	$316	$260	$0	$0	$576	$52	$0	$94	$722	$2
14.6	Machine Shop	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
14.7	Warehouse	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
14.8	Other Buildings & Structures	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
14.9	Waste Treating Building & Str.	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 14.	**$0**	**$509**	**$414**	**$0**	**$0**	**$923**	**$83**	**$0**	**$151**	**$1,157**	**$3**
	TOTAL COST	**$285,210**	**$10,637**	**$106,055**	**$0**	**$0**	**$401,902**	**$37,626**	**$58,323**	**$94,133**	**$591,983**	**$1,647**
	Owner's Costs											
	Preproduction Costs											
	6 Months All Labor										$2,937	$8
	1 Month Maintenance Materials										$513	$1
	1 Month Non-fuel Consumables										$543	$2
	1 Month Waste Disposal										$6	$0
	25% of 1 Months Fuel Cost at 100% CF										$0	$0
	2% of TPC										$11,840	$33
	Total										**$15,838**	**$44**
	Inventory Capital											
	60 day supply of consumables at 100% CF										$829	$2
	0.5% of TPC (spare parts)										$2,960	$8
	Total										**$3,789**	**$11**
	Initial Cost for Catalyst and Chemicals										$2,196	$6
	Land										$0	$0
	Other Owner's Costs										$88,798	$247
	Financing Costs										$15,984	$44
	Total Overnight Costs (TOC)										**$718,587**	**$1,999**
	TASC Multiplier								(IOU, high risk, 33 year)		1.078	
	Total As-Spent Cost (TASC)										**$ 774,637**	**$2,155**

Exhibit 7-24 Case 9 Initial and Annual Operating and Maintenance Costs

INITIAL & ANNUAL O&M EXPENSES						Cost Base (June)	2007
Case 9 - Subcritical PC w/ 90% CO2 capture - Retrofit						Heat Rate-net(Btu/kWh):	15,496
						MWe-net:	359
						Capacity Factor: (%):	85
		OPERATING & MAINTENANCE LABOR					
Operating Labor							
Operating Labor Rate(base):		34.65	$/hour				
Operating Labor Burden:		30.00	% of base				
Labor O-H Charge Rate:		25.00	% of labor				
				Total			
Skilled Operator		1.0		1.0			
Operator		1.3		1.3			
Foreman		0.0		0.0			
Lab Tech's, etc.		0.0		0.0			
TOTAL-O.J.'s		2.3		2.3			
						Annual Cost	Annual Unit Cost
						$	$/kW-net
Annual Operating Labor Cost		Maintenance labor cost		% of BEC	0.8889	$907,567	$2.525
Maintenance Labor Cost		(Case S12B is reference)		BEC	$392,528	$3,791,163	$10.549
Administrative & Support Labor						$1,174,682	$3.269
Property Taxes & Insurance						$30,661,480	$85.315
TOTAL FIXED OPERATING COSTS						**$36,534,892**	**$101.658**
VARIABLE OPERATING COSTS							
							$/kWh-net
Maintenance Material Cost				% of BEC	1.3333	**$5,233,695**	**$0.00196**
Consumables		Consumption		Unit	Initial		
		Initial	/Day	Cost	Cost		
Water(/1000 gallons)		0	3,449	1.22	**$0**	**$1,305,388**	**$0.00049**
Chemicals			4.841				
MU & WT Chem.(lb)		0	16,694	0.17	$0	$896,395	$0.00033
Soda Ash (ton)		0	8	80.00	$0	$201,936	$0.00008
Carbon (Mercury Removal) (lb)		0	0	1.05	$0	$0	$0.00000
MEA Solvent (ton)		920	1.30	2,249.89	$2,069,097	$907,435	$0.00034
NaOH (tons)		0	10.72	433.68	$0	$1,442,362	$0.00054
H2SO4 (tons)		0	6.20	138.78	$0	$266,950	$0.00010
Corrosion Inhibitor		0	0	0.00	$126,820	$6,039	$0.00000
Activated Carbon(lb)		0	1,557	1.05	$0	$507,295	$0.00019
Ammonia (19% NH3) ton		0	0	129.80	$0	$0	$0.00000
Subtotal Chemicals					**$2,195,917**	**$4,228,412**	**$0.00158**
Other							
Supplemental Fuel(MBtu)		0	0	0.00	$0	$0	$0.00000
SCR Catalyst(m3)		w/equip.	0.000	5,775.94	$0	$0	$0.00000
Emission Penalties		0	0	0.00	$0	$0	$0.00000
Subtotal Other					**$0**	**$0**	**$0.00000**
Waste Disposal							
Flyash (ton)		0	8	24.33	$0	$61,414	$0.00002
Bottom Ash(ton)		0	0	16.23	$0	$0	$0.00000
Subtotal-Waste Disposal					**$0**	**$61,414**	**$0.00002**
By-products & Emissions							
Gypsum (tons)		0	0	0.00	$0	$0	$0.00000
Subtotal By-Products					**$0**	**$0**	**$0.00000**
TOTAL VARIABLE OPERATING COSTS					**$2,195,917**	**$10,828,908**	**$0.00405**
Fuel(ton)		0	0	10.37	**$0**	**$0**	**$0.00000**

8. CONCLUSIONS

The objective of this report was to present the baseline cost and performance of greenfield integrated gasification combined cycle (IGCC) plants, greenfield supercritical (SC) pulverized coal (PC) plants, and retrofit subcritical PC plants that limit carbon dioxide (CO_2) emissions to various levels. For each plant type, three cases were modeled:

- Baseline performance with no CO_2 capture
- CO_2 emissions reduced to 1,100 lb/net-MWh
- CO_2 emissions reduced by 90 percent

The intermediate value of 1,100 lb/net-MWh was chosen to match the recent interim California standard established in January 2007. The results show that the cost and performance of the technologies analyzed at this emission limit fall approximately half way between the non-capture cases and the 90 percent capture cases. While the cost and performance penalties incurred at the 1,100 lb CO_2/net-MWh emission rate are less than for 90 percent capture, they are still substantial.

The performance and cost results of the nine cases modeled in this study are summarized in Exhibit 8-1. The primary conclusions that can be drawn are:

- The lowest LCOE for all cases is the subcritical PC and subcritical PC with retrofit, mainly due to the assumption that the original plant debt has been retired. The non-capture LCOE for the subcritical PC is 71 percent less than the non-capture IGCC case and 58 percent less than the non-capture SC PC case. The 90 percent CO_2 capture LCOE for the subcritical PC is 36 percent less than the corresponding IGCC case and 22 percent less than the SC PC case.
- The IGCC cases have the lowest percent change in LCOE from the non-capture case ($117.84/MWh) to the capture cases, with 27 percent for the 1,100 lb CO_2/net-MWh case and 48 percent for the 90 percent capture case. However, the absolute LCOE is highest for IGCC cases relative to the SC PC and subcritical PC cases.
- The existing subcritical PC plant with SCR has the highest percentage change in LCOE from the non-capture case ($33.78/MWh) to the capture cases with 178 percent and 263 percent for the 1,100 lb CO_2/net-MWh case and the 90 percent capture case, respectively. This is somewhat misleading because the LCOE of the existing non-capture subcritical PC plant does not have a capital cost component (plant is assumed to be paid for).
- For the 1,100 lb CO_2/net-MWh cases, the IGCC plant has the smallest energy penalty relative to the non-capture case at 6.2 absolute percentage points. SC PC is next with an energy penalty of 6.9 percentage points and the subcritical PC retrofit has the largest energy penalty at 7.3 percentage points.
- For the 90 percent capture cases, the subcritical PC retrofit plant has the smallest energy penalty relative to the non-capture case at 10.6 absolute percentage points. IGCC is next with an energy penalty of 10.9 percentage points and the SC PC plant has the largest energy penalty at 11.7 percentage points.

- The greenfield supercritical PC plant has the highest change in normalized TOC at 45 percent for the 1,100 lb/net-MWh case and 73 percent for the 90 percent capture case. While the net power for the SC PC capture and non-capture cases remained the same, the gross power output increased. This caused the increase in the capital costs to be greater than the greenfield IGCC cases and the existing subcritical PC retrofit, in which net power was derated to accommodate CO_2 capture.
- The costs of CO_2 captured and avoided were nearly identical for the SC PC and IGCC cases. The existing subcritical PC case has the lowest CO_2 removal costs at both CO_2 emissions levels and also the lowest CO_2 avoided cost at the 1,100 lb CO_2/net-MWh capture level. The CO_2 avoided cost at 90 percent capture level is slightly higher than the IGCC and SC PC cases.

Exhibit 8-1 Cost and Performance Summary of Cases 1 – 9

	Non-Capture	1,100 lb/net-MWh		90% Capture	
	Absolute	Absolute	% Change Relative to Non-Capture	Absolute	% Change Relative to Non-Capture
IGCC					
Net Plant Power (MWe)	502	443	-12%	401	-20%
Net Plant Efficiency, % (HHV)	41.8	35.6	-15%	30.9	-26%
TOC ($/kW)	3,128	3,938	26%	4,595	47%
LCOE ($/MWh)	117.84	149.38	27%	174.86	48%
CO_2 Removal Cost ($/tonne)	N/A	74	N/A	59	-20% (A)
CO_2 Avoided Cost ($/tonne)	N/A	108	N/A	84	-22% (A)
Supercritical PC					
Net Plant Power (MWe)	550	550	N/A	550	N/A
Net Plant Efficiency, % (HHV)	38.6	31.7	-18%	26.9	-30%
TOC ($/kW)	2,296	3,323	46%	3,969	73%
LCOE ($/MWh)	79.86	120.01	50%	143.89	80%
CO_2 Removal Cost ($/tonne)	N/A	73	N/A	58	-21% (A)
CO_2 Avoided Cost ($/tonne)	N/A	111	N/A	87	-22% (A)
Existing Subcritical PC Retrofit Plant					
Net Plant Power (MWe)	532	413	-22%	359	-33%
Net Plant Efficiency, % (HHV)	32.6	25.3	-22%	22.0	-33%
TOC ($/kW)	N/A	1,348	N/A	1,999	48% (A)
LCOE ($/MWh)	33.78	84.81	151%	111.64	230%
CO_2 Removal Cost ($/tonne)	N/A	62	N/A	57	-8% (A)
CO_2 Avoided Cost ($/tonne)	N/A	97	N/A	89	-8% (A)
Existing Subcritical PC Retrofit Plant w/ SCR					
Net Plant Power (MWe)	532	409	-23%	356	-33%
Net Plant Efficiency, % (HHV)	32.6	25.3	-22%	22.0	-33%
TOC ($/kW)	N/A	1,717	N/A	2,430	42% (A)
LCOE ($/MWh)	33.78	94.01	178%	122.46	263%
CO_2 Removal Cost ($/tonne)	N/A	73	N/A	64	-12% (A)
CO_2 Avoided Cost ($/tonne)	N/A	116	N/A	102	-12% (A)

(A) - Values relative to 1,100 lb/net-MWh case

9. REFERENCES

1 “California Public Utilities Commission Sets Greenhouse Gas Emissions Performance Standard: Frequently Asked Questions,” California Public Utilities Commission, January 25, 2007

2 H.R. 2454: American Clean Energy and Security Act of 2009, 111th Congress, 2009-2010

3 Yacobucci, Brent D. and Parker, Larry, “Climate Change: Federal Laws and Policies Related to Greenhouse Gas Reductions,” CRS Report for Congress, RL31931, Updated February 22, 2006

4 Parker, Larry and Yacobucci, Brent D., “Climate Change: Greenhouse Gas Reduction Bills in the 110th Congress,” CRS Report for Congress, RL 33486, Updated April 24, 2007

5 http://www.house.gov/waxman/safeclimate/index.htm, The Safe Climate Act of 2007 (H.R. 1590), Summary of the Bill

6 “Low Carbon Economy Act of 2007,” http://energy.senate.gov/public/_files/END07842_xml1.pdf

7 Pew Center on Global Climate Change, State Action Maps, http://www.pewclimate.org/what_s_being_done/in_the_states/state_action_maps.cfm

8 “BART Analysis for Jim Bridger Unit 4,” Final Report, CH2MHill, January 12, 2007

9 Coal Specifications for Quality Guidelines, Revision 0, May 31, 2005

10 “Annual Energy Outlook 2007 with Projections to 2030,” AEO Reference Case Table 19, EIA, February 2007

11 US EPA Green Book Nonattainment Area Map, http://www.epa.gov/oar/oaqps/greenbk/mapnpoll.html

12 1999 Environmental Protection Agency Information Collection Request Database with Coal Compositions for Over 37,000 Samples

13 CoalFleet User Design Basis Specification for Coal-Based Integrated Gasification Combined Cycle (IGCC) Power Plants, EPRI, Palo Alto, CA: 2009. 1017501

14 Thomas, Greg, “Delivering the IGCC Solution”, GE Energy January 31, 2006

15 Morehead, Harry and Hannemann, Frank, “Siemens Technology Improvements Enhance IGCC Plant Economics,” Gasification Technologies 2005, San Francisco, CA, October 2005

16 Telephone communication with David Denton, Eastman Chemical Co.

17 Standalone Documentation for EPA Base Case 2004 (V.2.1.9) Using the Integrated Planning Model, U.S. Environmental Protection Agency, EPA 430-R-05-011, September 2005

18 Smith, Lawrence A., Gupta, Neeraj, Sass, Bruce M. and Bubenik, Thomas A., "Engineering and Economic Assessment of Carbon Dioxide Sequestration in Saline Formations," Battelle Memorial Institute, 2001

19 Ciferno, Jared P. and McIlvried, Howard, "CO_2 Flow Modeling and Pipe Diameter Determination," February, 2003

20 Economic Evaluation of CO2 Storage and Sink Enhancement Options, Tennessee Valley Authority, NETL and EPRI, December 2002

21 North American Electric Reliability Council website, http://www.nerc.com/~gads

22 "Evaluation of Alternative IGCC Plant Designs for High Availability and Near Zero Emissions: RAM Analysis and Impact of SCR," EPRI, Palo Alto, CA, 2005, 1010461

23 Higman, Christopher, DellaVilla, Sal, and Steele, Bob, "The Reliability of Integrated Gasification Combined Cycle (IGCC) Power Generation Units, Achema 2006, May 2006

24 Cooling Tower Fundamentals, ed. John C. Hensley, 2nd Edition, The Marley Cooling Tower Company, Mission, Kansas, 1985

25 Herzog, Howard J., "The Economics of CO_2 Separation and Capture," MIT Energy Laboratory

26 "Merit Shop Wage & Benefit Survey," PAS, Inc., http://www.pas1.com/

27 Quality Guidelines for Energy System Studies, prepared by DOE NETL Office of Systems and Policy Support, June 4, 2003

28 Economic Evaluation of CO_2 Storage and Sink Enhancement Options, Tennessee Valley Authority, NETL, EPRI, December 2002.

29 *Quality Guidelines for Energy System Studies*, Office of Systems and Policy Support, January 30, 2005.

30 *Overview of Monitoring Requirements for Geologic Storage Projects.* IEA Greenhouse Gas R&D Programme, Report Number PH4/29, November 2004.

31 Statement of Ronald T. Evans of Denbury Resources, Inc., before the U.S. Senate Committee on Energy and Natural Resources. Hearing on the Policy Aspects of Carbon Capture, Transportation, and Sequestration and Related Bills, S.2323 and S.2144, January 31, 2008.
http://energy.senate.gov/public/index.cfm?FuseAction=Hearings.Testimony&Hearing_ID=1672&Witness_ID=4847

32 U.S. Bureau of Labor Statistics Producer Price Index, Support activities for oil and gas operations, Series Id: PCU213112213112, February 2008.

33 Power Systems Financial Model Version 5.0, September 2006

34 Global Energy Decisions. Energy Velocity Database

35 Zuideveld, P.L., “Shell Coal Gasification Process Using Low Rank Coal,” Gasification Technologies Conference, San Francisco, CA, October 2005

36 Private Communication with Greg E. Johnson of Shell Development Corporation, February 5, 2007

37 “Feasibility Study for an Integrated Gasification Combined Cycle Facility at a Texas Site,” EPRI, 1014510, October 2006

38 “CO_2 Capture in Low Rank Coal Power Plants,” Technical Study 2006/1, IEA Greenhouse Gas R&D Programme, January 2006

39 Universal Industrial Gases, Inc., http://www.uigi.com/cryodist.html

40 Jenkins, Steve, “Real-World Experience with IGCC – Lessons Learned in Design, Permitting, Operations and Maintenance,” URS Corporation, August 17, 2005

41 Geosits, Robert F. and Schmoe, Lee A., “IGCC – The Challenges of Integration, Proceedings of GT2005 ASME Turbo Expo 2005: Power for Land, Sea, and Air, Reno, NV, June 2005

42 Jaeger, Harry, “Plant Design Net Rated 644 MW and 38% HHV on Low Rank Coal,” Gas Turbine World, March-April 2006

43 Personal communication, Ray Drnevich, Praxair, Inc., ray_drnevich@praxair.com, November 2002

44 Maurstad, Ola et al., “Impact of Coal Quality and Gasifier Technology on IGCC Performance”

45 Larson, Eric D., et al, “Synthetic Fuel Production by Indirect Coal Liquifaction,” Energy for Sustainable Development, Volume VII, No. 4, December 2003

46 http://www.calgoncarbon.com/bulletins/Mercury_Removal_Gas_Streams.htmn

47 http://www.calgoncarbon.com/bulletins/HGR-P.htm

48 http://www.calgoncarbon.com/bulletins/TYPE_HGR.htm

49 “Process Screening Study of Alternative Gas Treating and Sulfur Removal Systems for IGCC Power Plant Applications,” EPRI AP-5505, December 1987

50 “Process Screening Analysis of Alternative Gas Treating and Sulfur Removal for Gasification,” Revised Final Report, December 2002, SFA Pacific for NETL

51 IGCC Puertollano: A clean coal Gasification Plant, ELCOGAS, 2001

52 Brdar, R. Daniel, and Jones, Robert M. “GE IGCC Technology and Experience with Advanced Gas Turbines,” GER-4207, GE Power Systems, Schenectady, NY, 10/2000.

53 GEI-41040G, “Process Specification for Fuel Gases for Combustion in Heavy-Duty Turbines,” January 2002

54 GEK-4189, “Design Considerations for Heated Gas Fuel”

55 NETL Coal Plant Database, http://www.netl.doe.gov/energy-analyses/pubs/database/index.html

56 Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity, Final Report, DOE/NETL-2007/1281, Revision 1, August 2007

57 Richard F. Storm, Optimizing Combustion in Boilers with Low NOx Burners, Storm Engineering & Associates, Power Magazine, October 1993

58 Peter Wieslanger, Fabric Filters for Power Applications, ALSTOM Power Environmental Systems AB, 2000, www.power.alstom.com

59 Sargent & Lundy LLC. “Dry Flue Gas Desulfurization Technology Evaluation” prepared for National Lime Association, September 2002

60 Improvement in Power Generation with Post-Combustion Capture of CO_2, IEA Greenhouse Gas R&D Programme, Report PH4/33, November 2004

61 Updated Coal Power Systems Study, Pulverized Coal Boiler, Fluor Subcontract Report for Carbon Dioxide Capture Unit, Fluor, December 2005

62 W.G. Gorman, J.K. Reinker, Steam Turbines for Utility Applications, GER-3646C, GE Power Generation, 1994

63 Carbon Dioxide Capture from Existing Coal-Fired Power Plants. November 2007

Carbon Dioxide Capture from Existing Coal-Fired Power Plants

DOE/NETL-401/110907

Final Report (Original Issue Date, December 2006)

Revision Date, November 2007

Disclaimer

Carbon Dioxide Capture from Existing Coal-Fired Power Plants

DOE/NETL-401/110907

Final Report (Original Issue Date, December 2006)

Revision Date, November 2007

NETL Contact:

Jared P. Ciferno
Office of Systems Analysis and Planning Division

Prepared by:

Massood Ramezan, Project Manager
Timothy J. Skone, Project Engineer
Research and Development Solutions, LLC

Nsakala ya Nsakala, and Gregory N. Liljedahl, Project Co-Leaders
Alstom Power Inc.

With Support From:

Loren E. Gearhart, Principal Investigator
ABB Lummus Global, Inc.

Rolf Hestermann, Principal Investigator
Alstom Power Generation AG

Barry Rederstorff, Host Site Advisor
American Electric Power

National Energy Technology Laboratory
www.netl.doe.gov

TABLE OF CONTENTS

LIST OF TABLES

LIST OF FIGURES

ACRONYMS AND ABBREVIATIONS

ABB	ABB Lummus Global Inc.
AEP	American Electric Power
ANSI	American National Standards Institute
APC	Air Pollution Control System
AST	Adiabatic Saturation Temperature
bara	Bar absolute
barg	Bar gauge
BI	Boiler Island
B.L.	Boundary Limit
BOP	Balance of Plant
Btu	British Thermal Unit
cm H_2O	Centimeters of water
CO_2	Carbon Dioxide
COE	Cost of Electricity
DCC	Direct Contact Cooler
DOE/NETL	Department of Energy/National Energy Technology Laboratory
EOR	Enhanced Oil Recovery
EPC	Engineered, Procured, and Constructed
ESP	Electrostatic Precipitator
FD	Forced Draft
FGD	Flue Gas Desulfurization
FOM	Fixed Operation & Maintenance
GHG	Greenhouse Gases
gpm	Gallons per Minute
GPS	Gas Processing System
g	Grams
HHV	Higher Heating Value
HP	High Pressure
hr	Hour
HSS	Heat Stable Salts
ID	Induced Draft
in. H_2O	Inches of Water
in. Hga	Inches of Mercury, Absolute
IP	Intermediate Pressure
IRI	Industrial Risk Insurers
ISO	International Standards Organization
J	Joules
kg	Kilograms
kWe	Kilowatts electric
kWh	Kilowatt-hour
LAM	Lean MEA solution

lbm	Pound mass
LDT	Let Down Turbine
LHV	Lower Heating Value
LP	Low Pressure
LT	Low Temperature
MCC	Motor Control Center
MCR	Maximum Continuous Rating
MEA	Monoethanolamine
MJ	Mega joules
MMBtu	Million of British Thermal Units
MWe	Megawatt Electric
MUPC	Make-up Power Cost
NGCC	Natural Gas Combined Cycle
NPSH	Net Positive Suction Head
N_2	Nitrogen Gas
OCDO	Ohio Coal Development Office
OSBL	Outside Boundary Limits
O&M	Operation & Maintenance
PA	Primary Air
PC	Pulverized Coal
PFD	Process Flow Diagram
PFWH	Parallel Feedwater Heater
PHX	Primary Heat Exchanger
PLC	Programmable Logic Controller
ppm	Parts per million
psia	Pound per square inch, absolute
psig	Pound per square inch, gauge
RDS	Research and Development Solutions
s	Second
SA	Secondary Air
SCPC	Supercritical Pulverized Coal
TIC	Total Investment Cost
TPD	Ton Per Day
VOM	Variable Operation & Maintenance

ACKNOWLEDGMENTS

The authors appreciatively acknowledge the following people for their contributions to the successful performance of the work presented herein: Mark Borman and Tom Ross of AEP Conesville, Ohio, Plant for providing field unit information; Bill Anderson and Paul Milios of ABB Lummus Global for providing cost estimation and technical support respectively; Neil Canvin, Aku Rainio, Ray Chamberland, Glen Jukkola, and Larry Cellilli from Alstom for providing steam turbine modeling and analysis, economic analysis, contract administration, technical guidance, and project accounting respectively.

EXECUTIVE SUMMARY

There is growing concern that emissions of carbon dioxide (CO_2) and other greenhouse gases (GHG) to the atmosphere is resulting in climate change with undefined consequences. This has led to a comprehensive program to develop technologies to reduce CO_2 emissions from coal-fired power plants. New technologies, based upon both advanced combustion and gasification technologies hold promise for economically achieving CO_2 reductions through improved efficiencies. However, if the United States decides to embark on a CO_2 emissions control program employing these new and cleaner technologies in new plants only, it may not be sufficient. It may also be necessary to reduce emissions from the existing fleet of power plants.

This study was performed to evaluate the technical and economic feasibility of various levels of CO_2 capture (e.g., 90%, 70%, 50%, and 30%) for retrofitting an existing pulverized coal-fired power plant (Conesville #5 unit in Ohio) using advanced amine-based capture technology. Impacts on plant output, efficiency, and CO_2 emissions, resulting from addition of the CO_2 capture systems on an existing coal-fired power plant were considered. Cost estimates were developed for the systems required to produce, extract, clean, and compress CO_2, which could then be available for sequestration and/or other uses such as enhanced oil or gas recovery. Results are reported in terms of the incremental cost of electricity, levelized over 20 years, (LCOE) to retrofit and operate an existing pulverized coal-fired power plant at various levels of carbon capture. The cost of CO_2 mitigation is also reported for each level of carbon capture. Summary results are presented in Figure ES-1 and summarized in Table ES-1.

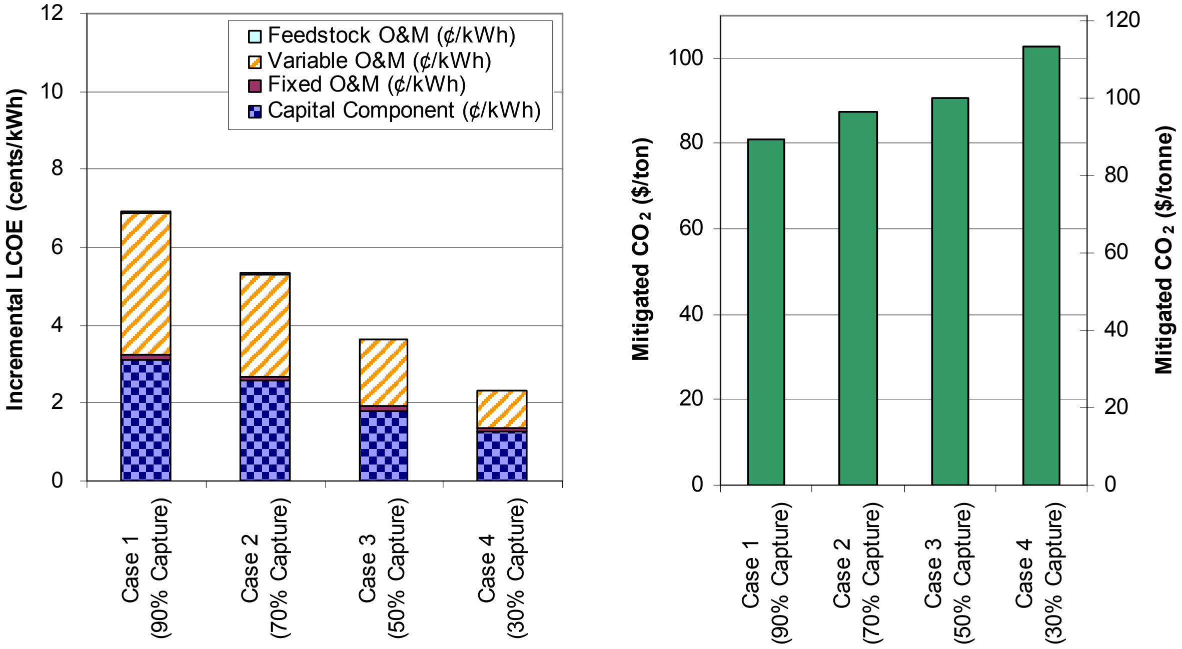
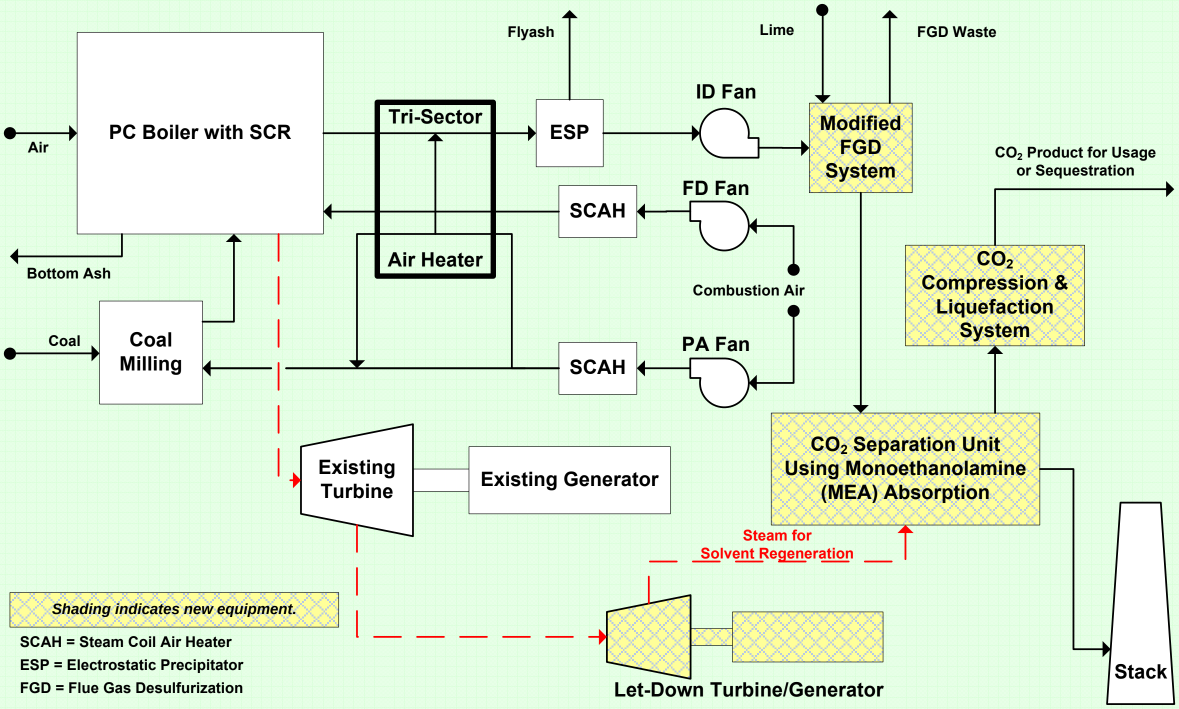


Figure ES-1: Incremental Levelized Cost of Electricity (LCOE) and CO_2 Mitigation Cost of Retrofitting a Pulverized Coal-fired Plant at Various Levels of Carbon Capture

The results demonstrate an almost linear relationship between the percent change in carbon capture and the incremental LCOE and CO_2 mitigation cost across the study range of 90% to 30% capture. A 10% reduction in the level of carbon capture equates to approximately an 11% reduction in the incremental LCOE and a 4% increase in the CO_2 mitigation cost.

Description of Plant Retrofit for Incorporating Carbon Capture Technology

A simplified process flow diagram for the study unit, modified with the addition of the post-combustion amine-based capture system, is shown in below. This simplified diagram is applicable to each of the CO_2 capture cases included in this study. The operation and performance of the existing boiler, air heater, and electrostatic precipitator (ESP) systems are identical to the Base Case for all the capture cases investigated and are not affected by the addition of the post-combustion amine-(MEA)-based CO_2 recovery systems.



Simplified Process Flow Diagram for Power Plant Modified with the Addition of an Advanced Amine Based CO_2 Capture System

The flue gas desulfurization (FGD) system is modified identically for each of the cases with the addition of a secondary absorber to reduce the SO_2 content of the flue gas entering the new amine system to below 10 ppmv. Recovery of less than 90% CO_2 (Cases 2, 3, and 4 with 70%, 50%, and 30% recovery respectively) is accomplished by bypassing a fraction of the total flue gas stream around the new CO_2 absorber. Flue gas bypass was determined to be the least costly way to obtain lower CO_2 recovery levels.

Table ES-1: Summary of Technical and Economic Performance for Retrofitting a Pulverized Coal-Fired Plant

Case	**Units**	**Base Case**	**Case 1 (90% Capture)**	**Case 2 (70% Capture)**	**Case 3 (50% Capture)**	**Case 4 (30% Capture)**
Boiler Parameters						
Main Steam Flow	lb/hr	3,131,619	3,131,651	3,131,651	3,131,651	3,131,651
Main Steam Pressure	psia	2,535	2,535	2,535	2,535	2,535
Main Steam Temp	Deg F	1,000	1,000	1,000	1,000	1,000
Reheat Steam Temp	Deg F	1,000	1,000	1,000	1,000	1,000
Boiler Efficiency	Percent	88.1	88.1	88.1	88.1	88.1
Coal Heat Input, HHV	10^6 Btu/hr	4,228.7	4,228.7	4,228.7	4,228.7	4,228.7
CO_2 Removal System Parameters						
Solvent Regeneration Energy	Btu/lbm-CO_2	n/a	1,550	1,550	1,550	1,550
Steam Pressure	psia	n/a	47	47	47	47
Steam Extraction Flow	lb/hr	n/a	1,210,043	940,825	671,949	403,170
Natural Gas Heat Input, HHV	10^6 Btu/hr	n/a	13.0	9.7	6.7	4.2
Steam Cycle Parameters						
Existing Steam Generator Output	kW	463,478	342,693	370,700	398,493	425,787
CO_2 Removal System Generator Output	kW	n/a	45,321	35,170	25,031	14,898
Total Turbine Generator Output	kW	463,478	388,014	405,870	423,524	440,685
Auxillary Power: Existing Plant	kW	29,700	29,758	29,928	30,113	30,306
Auxillary Power: CO_2 Removal System	kW	n/a	54,939	42,697	30,466	18,312
Net Plant Power	kW	433,778	303,317	333,245	362,945	392,067
Plant Performance Parameters						
Net Plant Heat Rate, HHV	Btu/kWh		13,984	12,728	11,686	10,818
Net Plant Efficiency, HHV	Percent	35.01	24.5	26.9	29.3	31.7
Energy Penalty	Percent	0	10.5	8.1	5.7	3.3
Capacity Factor	Percent	85	85	85	85	85
Plant CO_2 Profile						
CO_2 Produced	lb/hr	866,102	867,595	867,212	866,872	866,585
CO_2 Captured	lb/hr	0	779,775	607,048	433,606	260,164
CO_2 Emissions	lb/hr	866,102	87,820	260,164	433,266	606,421
Incremental Capital and O&M Costs						
Total Investment Cost	$1,000	n/a	400,094	365,070	280,655	211,835
Total Investment Cost	$/kW	n/a	1,319	1,095	773	540
Fixed O&M Costs	$1000/yr	n/a	2,494	2,284	2,079	1,869
Variable O&M Costs	$1000/yr	n/a	17,645	14,711	10,876	7,019
Levelized, Make-up Power Cost	$1000/yr	n/a	62,194	47,926	33,768	19,885
CO_2 By-product Revenue	$1000/yr	n/a	0	0	0	0
Feedstock (natural gas) O&M Costs	$1000/yr	n/a	653	488	337	211
Incremental LCOE Contributions						
Capital Component	¢/kWh	n/a	3.10	2.57	1.82	1.27
Fixed O&M	¢/kWh	n/a	0.13	0.11	0.09	0.07
Variable O&M	¢/kWh	n/a	3.66	2.62	1.72	0.96
Feedstock O&M	¢/kWh	n/a	0.03	0.02	0.01	0.01
Total	¢/kWh	n/a	6.92	5.32	3.64	2.31
CO_2 Mitigation Cost	$/ton	n/a	81	88	91	103
CO_2 Mitigation Cost	$/tonne	n/a	89	96	100	113
CO_2 Capture Cost	$/ton	n/a	54	58	61	70
CO_2 Capture Cost	$/tonne	n/a	59	64	67	77

Percent Change in Retrofit Investment Costs Show a Linear Correlation with CO_2 Capture Rate

The total investment required to retrofit an existing plant is also dependant on the level of carbon capture. Reductions in boiler modification costs and carbon capture equipment size are the primary factors. Figure ES-2 shows an almost linear relationship between percent CO_2 capture and total investment cost (TIC) based on the retrofitted plant net power output. As a result, this study shows a 10% reduction in CO_2 capture causes approximately a 10% reduction in the required retrofit investment across the study range of 90% to 30% capture. Table ES-1 summarizes the TIC on a $/kW-net and per $1000 dollar basis for each CO_2 capture rate.

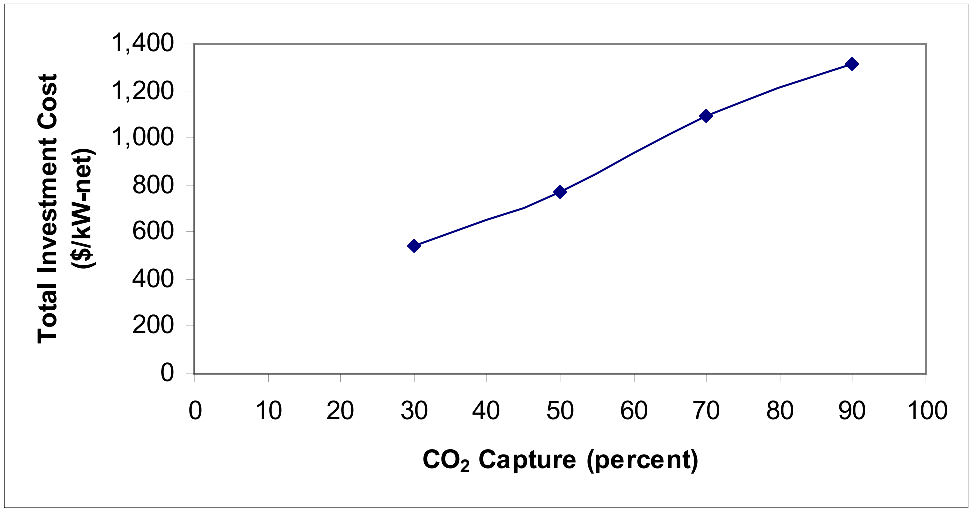


Figure ES-2: Affect of CO_2 Capture Rate on the Total Investment Cost for Retrofitting a Pulverized Coal-Fired Plant

Retrofit Investment and Operating Costs Included in the Study

The project capital cost estimates (July, 2006 cost date) include all required retrofit equipment such as the amine-based CO_2 scrubbing systems, the modified flue gas desulfurization (FGD) system, the CO_2 compression and liquefaction systems, and steam cycle modifications. Boiler island modifications other than for the FGD system are not required.

Operating and maintenance (O&M) costs were calculated for all systems. The O&M costs for the Base Case were provided by AEP. For the retrofit CO_2 capture system evaluations, additional O&M costs were calculated for the new equipment. The variable O&M (VOM) costs for the new equipment included such categories as chemicals and desiccants, waste handling, maintenance material and labor, and contracted services. A make-up power cost (MUPC) for the reduction in net power production is also included in the VOM costs. A levelized MUPC of 6.40 ¢/kWh-net, equivalent to a new subcritical pulverized coal (bituminous) power plant without carbon capture, was determined for each Case included within the study. The fixed O&M (FOM) costs for the new equipment include operating labor only.

Adding CO_2 Capture Technology Impacts Net Plant Output and Thermal Efficiency

Significant reductions in net plant output are incurred (10%-30% for Cases 1-4) as a result of the CO_2 capture system. For example, capturing 90% of the carbon reduces the net plant output from 433.8 MW to 303.3 MW. The capture system design includes a let down steam turbine generator that contributes 45.3 MW to the existing steam turbine generator output. Inclusion of the let down steam turbine improves the technical performance and lowers the incremental LCOE for retrofitting a pulverized coal-fired power plant with carbon capture technology.

Net plant thermal efficiency is also reduced from about 35.0% (HHV basis) for the Base Case to 24.4% - 31.6% for Cases 1-4. The efficiency reductions are due to reductions in the steam turbine output due to steam extraction for solvent regeneration and significant auxiliary power requirement increases as shown in Table ES-1. The auxiliary power increases are primarily due to the CO_2 compression and liquefaction system. The efficiency decrease is essentially a linear function of CO_2 recovery level over the range of CO_2 capture investigated.

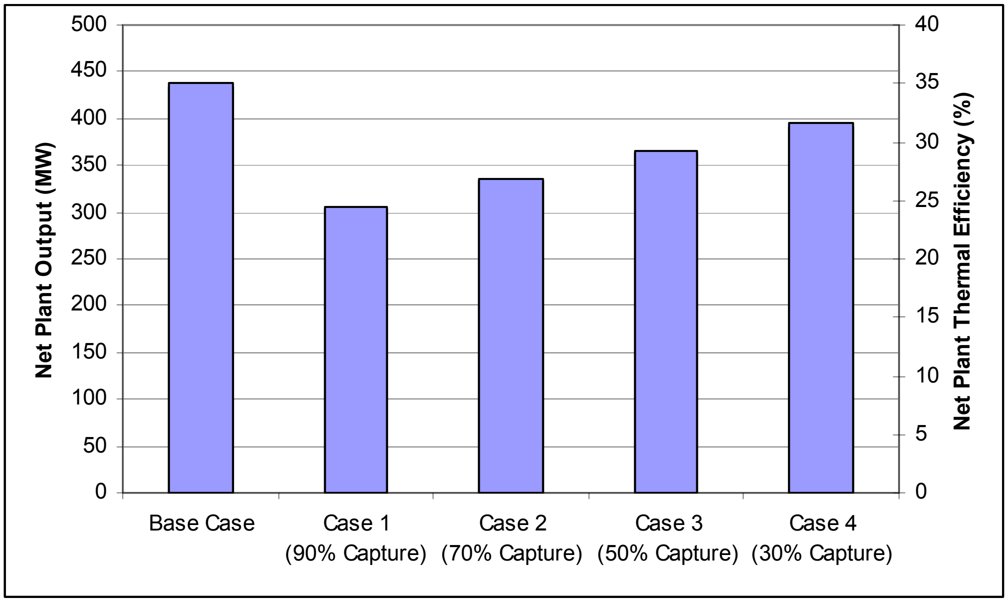


Figure ES-3: Plant Performance Impact of Retrofitting a Pulverized Coal-Fired Plant at Various Levels of Carbon Capture

Retrofitting Existing Coal-fired Plants Can Help Reduce U.S. GHG Emissions

Specific carbon dioxide emissions were reduced from about 908 g/kWh (2 lbm/kWh) for the Base Case to between 59-704 g/kWh (0.13-1.55 lbm/kWh) depending on CO_2 recovery level as shown in

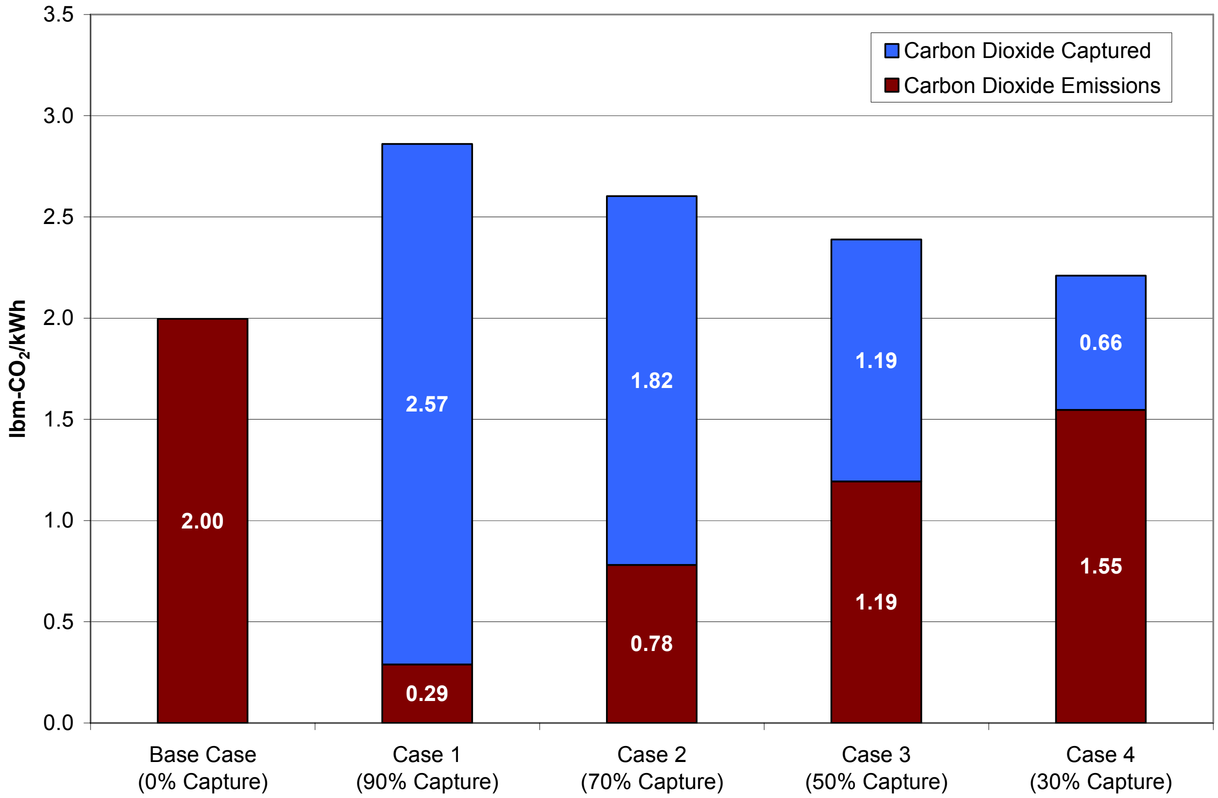


Figure ES-4. This corresponds to between 6.6% and 77.5% of the Base Case carbon dioxide emissions. The mass of carbon dioxide produced in each case is relatively the same[1], however the significant reduction in net power production in each of the retrofit cases (Cases 1-4) results in a higher CO_2 production rate per kilowatt-hour of power produced. Table ES-1 summarizes the mass of CO_2 produced, emitted, and captured for each case on a pound per hour basis. The mass of CO_2 emissions avoided is determined as the difference per kilowatt-hr in CO_2 emissions relative to the Base Case. For example, Case 1 (90% capture) emits 0.29 lbm-CO_2/kWh and the Base Case emits 2.00 lbm-CO_2/kWh. The difference is 1.71 lbm-CO_2/kWh. An 85.5% reduction in CO_2 released to the environment per kilowatt-hour of power produced.

---

[1] Coal feed rate is unchanged from the Base Case to each of the retrofit cases (Cases 1-4). A small amount of supplemental natural gas is utilized to regenerate the solvent media in the carbon capture system, therefore, adding to the total mass of CO_2 produced.

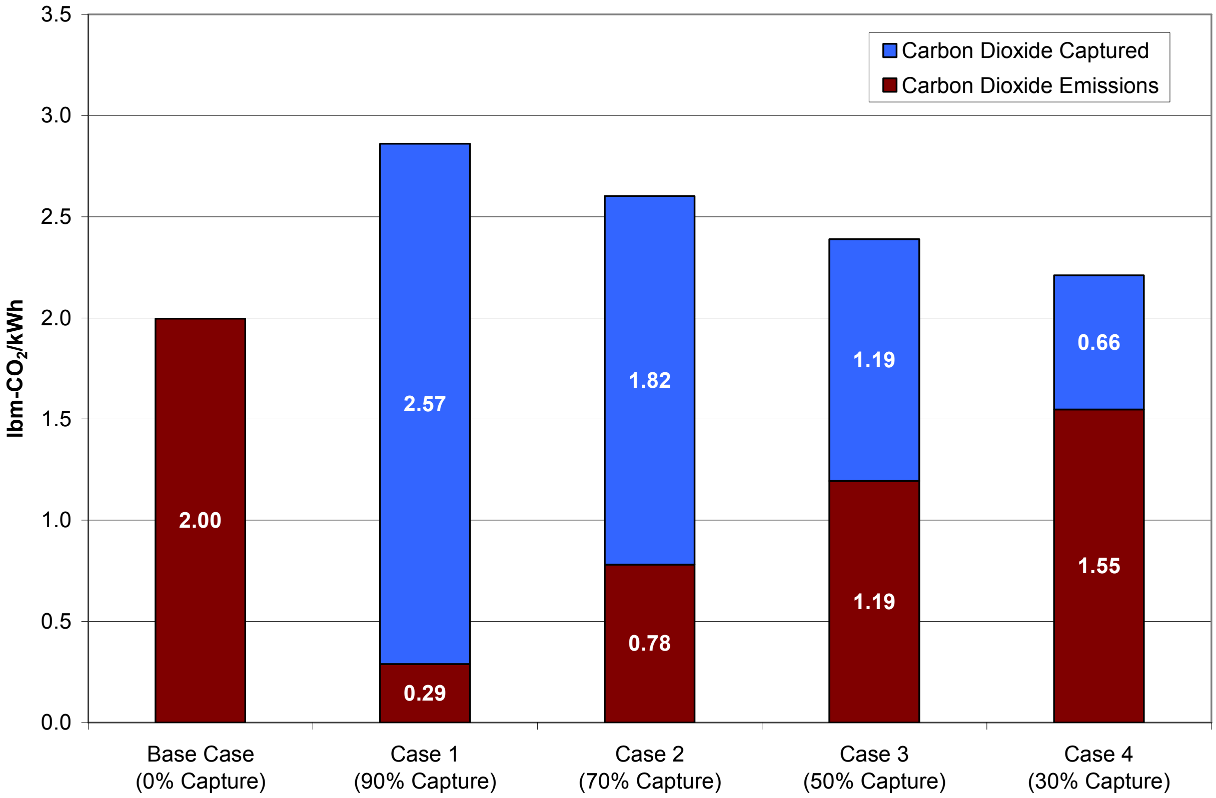


Figure ES-4: Reduction in Carbon Dioxide Emissions to the Environment

Sensitivity Analysis Results Demonstrate a Range of Scenarios for Evaluating the Benefits of Retrofitting Pulverized Coal-fired Plants with Carbon Capture Technology

Specific results from this study are limited to the retrofit of AEP's Conesville Unit #5. Therefore, a sensitivity analysis of the key economic variables was conducted to evaluate the applicability of retrofitting other pulverized coal-fired power plants with carbon capture technology. The economic sensitivity analysis was done by varying a number of parameters (Capacity Factor, Total Investment Cost, Make-up Power Cost, and CO_2 By-product Selling Price) that affect the economic results. These sensitivity parameters were chosen since the base values used for these parameters are site specific to this study.

The objective of this analysis was to determine the relative impacts of the sensitivity parameters and CO_2 capture level on incremental cost of electricity and CO_2 mitigation cost. The sensitivity analysis was conducted for each case analyzed within this study. The economic sensitivity results obtained from Case 1 (90% capture) are briefly discussed below.

Results for the Case 1 sensitivity study are shown in Figure ES-5. This figure shows the sensitivity of incremental LCOE to capacity factor, total investment cost, make-up power cost, and CO_2 by-product selling price. The base parameter values represent the point in Figure ES-5 where all the sensitivity curves intersect (point 0.0, 0.0). The incremental LCOE ranges from a low of -0.50 ¢/kWh to a high of 7.96 ¢/kWh for the Case 1 sensitivity analysis. The order of sensitivity (most sensitive to least sensitive) of these parameters to incremental LCOE is: CO_2

by-product selling price (levelized) > capacity factor > total investment cost > make-up power cost (levelized). For Cases 2 thru 5, the total investment cost becomes more significant than the make-up power cost, but they are approximately equivalent in Case 1.

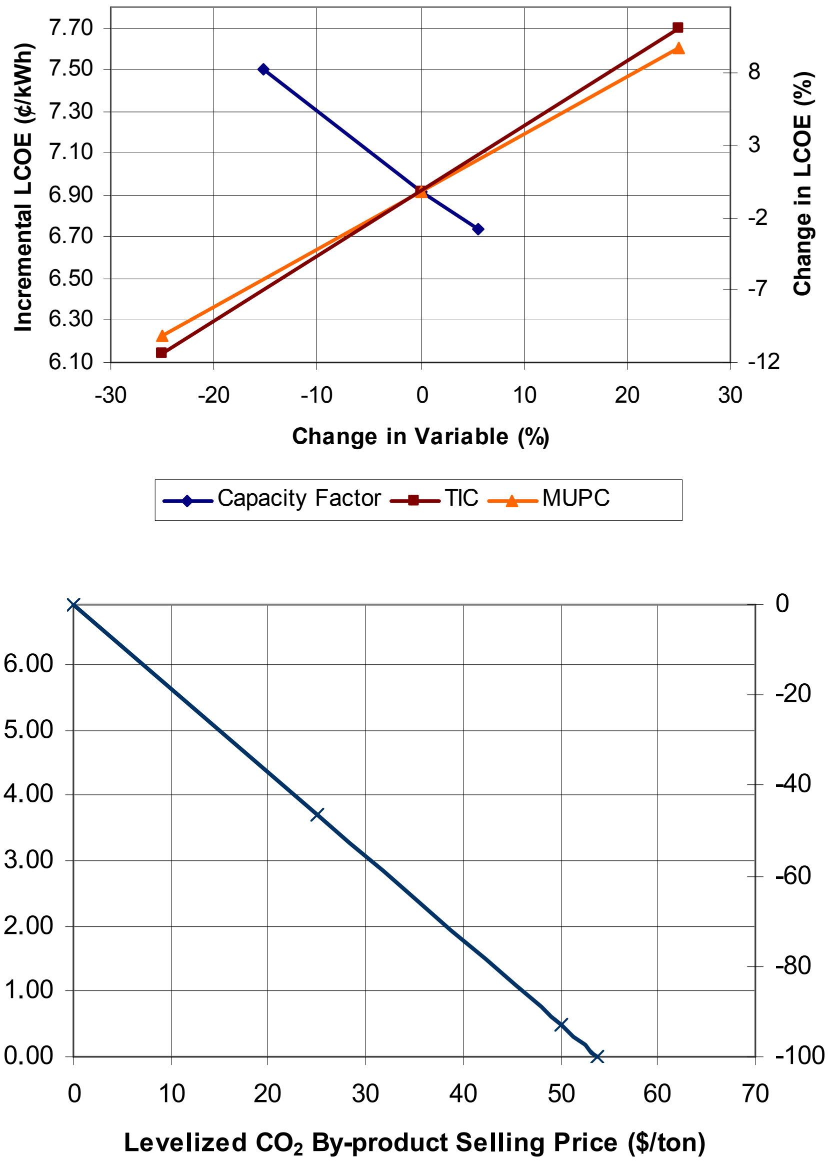


Figure ES-5: Economic Sensitivity Results (Case 1, 90% CO_2 Capture)

Reductions in Solvent Regeneration Energy Prove Key to Future Reductions in the Cost of Amine-Based Carbon Capture

Improvements in technical and economic performance resulting from reduction in solvent regeneration energy at the 90% carbon capture level were also evaluated as part of this study and compared to previous work conducted by NETL and Alstom (Bozzuto et al., 2001). The solvent regeneration energy used in this study is based on present day technology, and is 34% less than in the prior study (2,350 to 1,550 Btu/lbm-CO_2). The result is an improvement in plant thermal efficiency of 4.2 percentage points (from 20.3% to 24.5%). Additionally, retrofit specific investment costs ($/kWe) were reduced by 52% and incremental LCOE was reduced by 43%.

Because of this significant improvement in amine system performance (in particular, solvent regeneration energy) in the past several years, technical and economic performance of a near-future solvent regeneration level of 1,200 Btu/lbm-CO_2 was compared in a simplified manner to the current technology level of 1,550 Btu/lbm-CO_2. The results demonstrated a potential future improvement in plant thermal efficiency loss as low as 9.3 percentage points. Correspondingly, the retrofit specific investment costs ($/kWe) and incremental LCOE were further reduced by 3% and 9% respectively.

Table ES-2 shows the primary impacts of solvent regeneration energy level on plant performance and economics. The results show a significant improvement in plant efficiency relative to the earlier 2001 study (Case 5).

Table ES- 2: Plant Performance and Economics vs. Solvent Regeneration Energy

Case	Base Case	Case 5 (2001)	Case 1 (2006)	Case 1a (Near-future)
Solvent Regeneration Energy (Btu/lbm-CO_2)	---	2,350	1,550	1,200
Net Plant Efficiency (% HHV)	35.0	20.2	24.4 25.7	
Efficiency Loss (% Points)	---	14.8	10.6	9.3
Incremental LCOE (¢/kWh)	---	12.54	6.92	6.32
CO_2 Mitigation Cost ($/tonne)	---	134	81	73
CO_2 Mitigation Cost ($/tonne)	---	148	89	81
CO_2 Capture Cost ($/tonne)	---	76	54	52
CO_2 Capture Cost ($/tonne)	---	83	59	57

Conclusions

No major technical barriers exist for retrofitting AEP's Conesville Unit #5 to capture CO_2 with post-combustion amine-based capture systems. Nominally, four acres of new equipment space is needed for the amine-based capture and compression system and can be located in three primary locations on the existing 200-acre power plant site, which accommodates a total of six units (2,080 MWe). Slightly less acreage is needed as the capture level is reduced. However, if all six units on this site were converted to CO_2 capture, it may be difficult to accommodate all the new CO_2 capture equipment on the existing site and some additional land would probably need to be purchased.

Plant technical performance and the incremental cost of adding carbon capture technology was evaluated at 90%, 70%, 50%, and 30% capture levels with a solvent regeneration energy level of 1,550 Btu/lbm-CO_2, which represents the "state of the art" at the time of this study (ca. 2006). Lower levels of CO_2 capture can be achieved by simply bypassing some of the flue gas around the CO_2 capture system and only processing a fraction of the total flue gas in the amine based capture system, which can then be made smaller. Flue gas bypassing was determined to be the best approach, from a cost and economic standpoint, to obtain lower CO_2 recovery levels. Energy requirements and power consumption are high, resulting in significant decreases in overall power plant thermal efficiencies, which range from about 24.5% to 31.6% as the CO_2

capture level decreases from 90% to 30% for Cases 1-4 as compared to 35% for the Base Case (all HHV basis). The efficiency decrease is essentially a linear function of CO_2 recovery level.

Specific carbon dioxide emissions were reduced from about 908 g/kWh (2 lbm/kWh) for the Base Case to 132-704 g/kWh (0.29-1.55 lbm/kWh) as the CO_2 recovery level decreases from 90% to 30%. Case 2 (70% CO_2 capture) was found to yield approximately this same amount of CO_2 emissions, 362 g/kWh (0.781 lbm/kWh) as typical natural gas combined cycle (NGCC) plant without carbon capture.

Specific investment costs are $540 to $1,319/kWe-new as CO_2 capture level increases from 30% to 90%. The specific investment cost is a nearly linear function of CO_2 recovery level, although equipment selection and economy of scale effects make this relationship much less linear than efficiency.

All cases studied incur significant increases to the levelized cost of electricity (LCOE) as a result of CO_2 capture. The incremental LCOE, as compared to the Base case (air firing without CO_2 capture) increases from 2.31 to 6.92 ¢/kWh as CO_2 capture level increases from 30% to 90%. Conversely, CO_2 mitigation cost increases slightly from $89 to $113/tonne of CO_2 avoided as the CO_2 capture level decreases from 90% to 30%. The near linear decrease in incremental LCOE with reduced CO_2 capture indicates there is no optimum CO_2 recovery level.

For the ranges studied, the incremental LCOE is most impacted by the following parameters (in given order): CO_2 by-product selling price, CO_2 capture level, solvent regeneration energy, capacity factor, investment cost, and make-up power cost.

To examine the impact improvements in amine-based systems, a solvent regeneration energy sensitivity study was completed for the 90% capture level. Reduced solvent regeneration energy was found to have significant impacts on the plant performance and economics. Plant thermal efficiency was calculated to change by about 3.7 percentage points for a change in solvent regeneration energy of 1,000 Btu/lbm-CO_2. Similarly, incremental cost of electricity was determined to be sensitive to changes in solvent regeneration energy. The incremental LCOE was calculated to change by about 0.6 ¢/kWh (or about 10% relative to Case 1 at 6.92 ¢/kWh) for a change in solvent regeneration energy of 1,000 Btu/lbm-CO_2. Incremental LCOE reductions of about 49% were found, as compared to the original 2001 study.

Overall, the results demonstrate the technical and economical feasibility for retrofitting pulverized coal-fired power plants in the U.S. over a range of carbon capture levels. Research efforts continue to improve upon the technical and economic performance of amine-based carbon capture technology to ensure a potential option for existing U.S. power plants to contribute to reducing carbon emissions in the event the United States decides to embark on a CO_2 emissions control program.

1 INTRODUCTION

There is growing concern that emissions of CO_2 and other greenhouse gases (GHG) to the atmosphere is resulting in climate change with undefined consequences. This has led to a comprehensive program to develop technologies to reduce CO_2 emissions from coal-fired power plants. New technologies, such as advanced combustion systems and gasification technologies hold great promise for economically achieving CO_2 reductions. However, if the United States decides to embark on a CO_2 emissions control program employing these new and cleaner technologies in new plants only, it may not be sufficient. It will also be necessary to reduce emissions from the existing fleet of power plants. This study will build on the results of previous work to help determine better approaches to capturing CO_2 from existing coal-fired power plants.

This study significantly increases the information available on the impact of retrofitting CO_2 capture to existing PC-fired power plants. This study also provides input to potential electric utility actions concerning GHG emissions mitigation, should the U.S. decide to reduce CO_2 emissions. Such information is critical for deciding on the best path to follow for reduction of CO_2 emissions, should that become necessary. This study better informs the public as to the issues involved in reducing CO_2 emissions, provides regulators with information to assess the impact of potential regulations, and provides data to plant owners/operators concerning CO_2 capture technologies. If this is to be done in the most economic manner, it will be necessary to know what level of CO_2 recovery is most economical from the point of view of capital cost, cost of electricity (COE), and operability. All this will contribute to achieving necessary controls in the most economically feasible manner.

Although switching to natural gas as a fuel source is an option, a tight supply and rising costs may prevent this from being a universal solution. Also, fuel switching may not provide the desired CO_2 emission reductions; and, therefore, some form of CO_2 capture may be required. Captured CO_2 could be sold for enhanced oil or gas recovery or sequestered. The results of this CO_2 capture study will enhance the public's understanding of post-combustion control options and influence decisions and actions by government regulators and power plant operators relative to reducing GHG CO_2 emissions from power plants.

The objectives for this study are to evaluate the technical and economic impacts of removing CO_2 from a typical existing U.S. coal-fired electric power plant using advanced amine-based post-combustion CO_2 capture systems. By investigating various levels of CO_2 capture, potential exists for identifying an economically optimum CO_2 capture level as well as simply quantifying the effect of CO_2 capture level on typical measures of plant performance and economic merit. As a view of the future for amine-based CO_2 capture systems, a sensitivity analysis showing the effect of anticipated reductions in solvent regeneration energy was also investigated (Please refer to Section 4 for details). This sensitivity study was done at the 90% CO_2 capture level only with solvent regeneration energy values of 1,550 and 1,200 Btu/lbm-CO_2 (Cases 1 and 1a respectively). The 1,550 Btu/lbm-CO_2 level represents the state of the art at the time of this study (ca. 2006) (IEA, 2004), the 1,200 Btu/lbm-CO_2 level represents a near-future value which may be possible with improved solvents, as discussed in the literature. The primary impacts are quantified in terms of plant electrical output reduction, thermal efficiency reduction, CO_2

emissions reduction, retrofit investment costs, and the incremental cost of generating electricity resulting from the addition of the CO_2 capture systems to the selected study unit.

1.1 Background

In a report titled “Engineering Feasibility and Economics of CO_2 Capture on an Existing Coal-Fired Power Plant,” (Bozzuto et al., 2001), Alstom evaluated the impact of adding facilities to capture >90% of the CO_2 from AEP’s Conesville, Ohio, Unit #5. During the 1999-2001 time period of the study, Alstom teamed with American Electric Power (AEP), ABB Lummus Global Inc. (ABB), National Energy Technology Laboratory (NETL), and Ohio Coal Development Office (OCDO) and conducted a comprehensive study evaluating the technical and economic feasibility of three alternate CO_2 capture technologies applied to an existing U.S. coal-fired electric power plant. The power plant analyzed in this study was Conesville #5, a subcritical, pulverized-coal (PC) fired steam plant operated by AEP of Columbus, Ohio. Unit #5 is one of six coal-fired steam plants located on the Conesville site which has a total generating capacity of ~2,080 MWe. The Unit #5 steam generator is a nominal 450 MW, coal-fired, subcritical pressure, controlled circulation unit. The furnace is a single cell design that employs corner firing with tilting tangential burners. The fuel utilized is bituminous coal from the state of Ohio. The flue gas leaving the steam generator system is cleaned of particulate matter in an electrostatic precipitator (ESP) and of SO_2 in a lime-based flue gas desulfurization (FGD) system before being discharged to the atmosphere.

One of the CO_2 capture concepts investigated in this earlier study was a post-combustion system which consisted of an amine-based scrubber using monoethanolamine (MEA) as depicted in Figure 1-1. This system was referred to as **Concept A.** In Concept A, coal is burned conventionally in air as schematically depicted below. The flue gases leaving the modified FGD system (a secondary absorber is added to reduce the SO_2 concentration, as required by the MEA system) are cooled with a direct contact cooler and ducted to the MEA system where more than 96% of the CO_2 is removed, compressed, and liquefied for usage or sequestration. The MEA system uses the Kerr-McGee/ABB Lummus Global’s commercial MEA process. The remaining flue gases leaving the new MEA system (consisting of primarily oxygen, nitrogen, water vapor and a relatively small amount of sulfur dioxide and carbon dioxide) are discharged to the atmosphere. The CO_2 capture results were compared to a Base Case. The Base Case represents the “business as usual” operation scenario for the power plant without CO_2 capture.

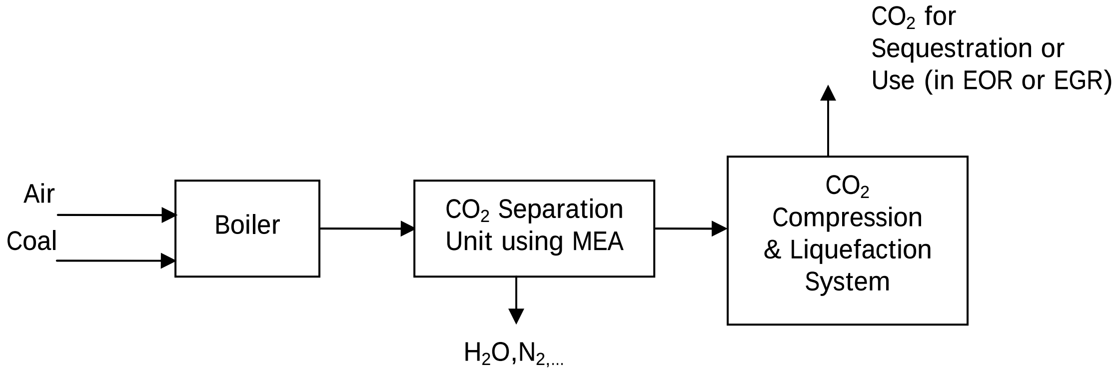


Figure 1-1: Post-Combustion Amine-Based CO_2 Capture Retrofit

Although boiler performance is identical to the Base Case in Concept A, there is a major impact to the steam cycle system where low-pressure steam is extracted to provide the energy for solvent regeneration. About 79% of the intermediate pressure (IP) turbine exhaust steam is extracted from the IP/LP crossover pipe. This steam is expanded from 200 psia to 65 psia through a new steam turbine/generator where electricity is produced. The exhaust steam leaving the new turbine provides the heat source for solvent regeneration in the reboilers of the CO_2 recovery system.

Solvent regeneration for this system requires about 5.46 MJ/Tonne CO_2 (2,350 Btu/lbm-CO_2). The condensate leaving the reboilers is pumped to the existing deaerator. The remaining 21% of the IP turbine exhaust steam is expanded in the existing low-pressure turbine before being exhausted to the existing condenser. The total electrical output from both the existing and new generators is 331,422 kW. This represents a gross output reduction of 132,056 kW (about 28%) as compared to the Base Case.

Investment costs required for adding the capture system to this existing unit were found to be very high (~$1,602/kWe-new: new refers to the new output level of 331,422 kW). The impact on the cost of electricity was found to be an increase of about 6.2¢/kWh (not including Make-up Power Costs, MUPC).[2]

Based on these results, further study was deemed necessary to find a better approach for capturing CO_2 from existing PC-fired power plants.

1.2 Current Study

In the current study NETL teamed with Alstom Power Inc., AEP, and ABB as well as with Science Applications International Corporation (SAIC)/Research and Development Solutions (RDS) to conduct a follow-up study. The follow up study again investigated post-combustion capture systems with amine scrubbing as applied to the Conesville #5 unit. The post-combustion CO_2 scrubbing system for the current study differs from the previous study in several ways:

- An advanced “state of the art” amine CO_2 scrubbing system is used for CO_2 removal from the flue gas stream. This advanced system requires significantly less energy for solvent regeneration. Solvent regeneration for this system, designed and selected in 2006, requires about 3.6 MJ/Tonne CO_2 (3.1×10^6 Btu/Ton CO_2) (~34% reduction as compared to the previous study, designed and selected in 2000). Additionally, the reboiler is operated at 3.1 bara (45 psia) as compared to 4.5 bara (65 psia) in the previous study.
- Several CO_2 capture levels are investigated in this study (90%, 70%, 50%, and 30%). These are referred to as **Cases 1, 2, 3, and 4** respectively in this study. In the previous study only one CO_2 recovery level (96%) was investigated. The costs and economic evaluation of this previous case (Case 5 in the current study) were updated.
- Alstom’s steam turbine retrofit group developed a detailed analysis of the modified existing steam turbine. Previously, a more simplified analysis was done for the existing steam turbine.

---

[2] Costs and economic evaluation were updated as part of the current study. Both the investment cost and incremental cost of electricity doubled as a result of the updated analysis; see Case 5 results.

- In the current study significant quantities of heat rejected from the CO_2 capture/compression system are integrated with the steam/water cycle. Previously, heat integration was not used because the CO_2 capture/compression system was located too far away from the steam/water system. The reboiler pressure for the current study was also lowered.
- It is recognized that solvent regeneration energy represents a key variable for amine-based post-combustion CO_2 capture systems in terms of the impact it ultimately has on the measures of power plant performance (thermal efficiency) and economic merit (cost of electricity). Knowing that the commercial implementation of these amine-based post-combustion capture systems will be several years in the future, and that research is continually improving the performance of these amines, a sensitivity analysis showing the effect of anticipated reductions in solvent regeneration energy was also investigated in this study. This sensitivity study was done at the 90% capture level only and the solvent regeneration energy levels investigated for this capture level were 1,550 and 1,200 Btu/lbm-CO_2. These cases are referred to as **Cases 1 and 1a** respectively in this study.

1.2.1 CO_2 Capture Level Sensitivity Study

The following list defines the five case included in the capture level sensitivity study presented in this report. The first four cases (Cases 1-4) use an advanced "state of the art" amine scrubbing system designed and cost estimated in 2006. The fifth case (Case 5) uses the Kerr-McGee/ABB Lummus Global's MEA scrubbing system, which was originally designed and cost-estimated in 2000.

- **Case 1:** 90% Capture
- **Case 2:** 70% Capture
- **Case 3:** 50% Capture
- **Case 4:** 30% Capture
- **Case 5:** 96% Capture "Concept A of 2001 study" using Kerr-McGee/ABB Lummus Global's commercial MEA-based process (cost and economic analysis update of previous study only).

To provide a frame of reference, each of the cases is evaluated against a Base Case from the standpoints of performance and impacts on power generation cost. The Base Case represents the "business as usual" operation scenario for the existing plant without CO_2 recovery. The Base Case which is used for the current study is identical to the Base Case used in the previous study from a plant performance standpoint. Fuel costs and other operating and maintenance costs for the Base Case have been updated based on AEP's current recommendations. All technical performance and cost results associated with these options are being evaluated in comparative manner.

Furthermore, in the current study, investment costs and economics are updated for "Concept A" from the original study in order to be directly comparable with the current study results. This is referred to as Case 5 in the current study. It should be pointed out that for Case 5 the process design and equipment selections were developed in 2000 and were not updated for the current study.

1.2.2 Solvent Regeneration Energy Sensitivity Study - A Look To The Future:

It is well known that commercial implementation of these amine-based post-combustion capture systems for power plant applications will not occur until several years in the future. This delay is because these systems need to be proven at large scale, CO_2 sequestration technology needs to be proven, and policies need to be implemented to make utilization of these systems economic. During this time period, numerous research and developmental efforts are ongoing to further advance post-combustion CO_2 capture technologies. These efforts seek to develop technologies that are focused on improving performance and reducing cost with post-combustion CO_2 capture.

One of the key parameters with post-combustion CO_2 capture systems that is an indicator of relative system performance is regeneration energy requirement (Btu/lbm-CO_2). When these post-combustion CO_2 capture systems are integrated with power plants, this variable is potentially quite sensitive with respect to the common measures of power plant performance (thermal efficiency) and economic merit (cost of electricity). Hence, as a look to the future, a simple sensitivity analysis for solvent regeneration energy and the impacts on power plant performance (thermal efficiency) and economics (cost of electricity) was carried out. This sensitivity study was done at the 90% capture level only and the solvent regeneration energy levels investigated were 1,550 and 1,200 Btu/lbm-CO_2. These cases are referred to as Cases 1 and 1a respectively.

- **Case 1 –** Existing power plant retrofit with an advanced “state of the art” amine system for 90% CO_2 capture (1,550 Btu/lbm-CO_2 solvent regeneration energy)
- **Case 1a –** Existing power plant retrofit with an advanced “near future” amine system for 90% CO_2 capture (1,200 Btu/lbm-CO_2 solvent regeneration energy)

The solvent regeneration energy level of 1,550 Btu/lbm-CO_2 represents the state of the art at the time of this study (ca. 2006) (IEA, 2004), the 1,200 Btu/lbm-CO_2 level represents a future value, which may be possible with improved solvents as discussed in the literature.

Alstom Power Inc. managed and performed the subject study from its U.S. Power Plant Laboratories office in Windsor, Connecticut. Alstom Steam Turbine Retrofit group performed the steam turbine analysis from its offices in Mannheim, Germany. ABB Lummus Global, from its offices in Houston, Texas, participated as a subcontractor. American Electric Power participated by offering their Conesville Unit #5 as the case study, and provided relevant technical and cost data. RDS is the prime contractor reporting to NETL for the project. AEP is one of the largest U.S. utilities and is the largest consumer of Ohio coal, and as such, brings considerable value to the project. Similarly, Alstom Power and ABB Lummus Global are well established as global leaders in the design and manufacture of power generation equipment, petrochemical and CO_2 separation technology. Alstom Environmental Business Unit is a world leader in providing equipment and services for power plant environmental control and provided their expertise to this project. The U.S. Department of Energy, National Energy Technology Laboratory through RDS provided consultation and funding.

The motivation for this study was to provide input to potential U.S. electric utility actions to meet Kyoto protocol targets. If the U.S. decides to reduce CO_2 emissions consistent with the Kyoto protocol, action would need to be taken to address the fleet of existing power plants. Although fuel switching from coal to gas is one likely scenario, it will not be a sufficient measure and some form of CO_2 capture for use or disposal may also be required. The output of this CO_2 capture study will enhance the public's understanding of CO_2 capture and influence decisions and actions by government, regulators, equipment suppliers, and power plant owners to reduce their greenhouse gas CO_2 emissions.

The primary objectives for this study are to evaluate the technical and economic impacts of removing CO_2 from this existing U.S. coal-fired electric power plant. By investigating various levels of capture, potential exists for identifying a "sweet spot," as well as simply quantifying the effect of this variable on typical measures of plant performance and economic merit. The impacts are quantified in terms of plant electrical output, thermal efficiency, CO_2 emissions, retrofit investment costs, and the incremental cost of generating electricity resulting from the addition of the CO_2 capture systems. All technical performance and cost results associated with these options are being evaluated in comparative manner. Technical and economic issues being evaluated include:

- Overall plant thermal efficiency
- Boiler efficiency
- Steam cycle thermal efficiency
- Steam cycle modifications
- Plant CO_2 emissions
- Plant SO_2 emissions
- Flue Gas Desulfurization system modifications and performance
- Plant systems integration and control
- Retrofit investment cost and cost of electricity (COE)
- Operating and Maintenance (O&M) costs
- CO_2 Mitigation Costs

Cost estimates were developed for all the systems required to extract, clean, compress and liquefy the CO_2, to a product quality acceptable for pipeline transport. The Dakota Gasification Company's CO_2 specification (Dakota 2005) for EOR, given in Table 1-1, was used as one of the bases for the design of the CO_2 capture system.

Table 1-1: Dakota Gasification Project's CO_2 Specification for EOR

Component	units	Value
CO_2	vol %	96
H_2S	vol %	1
CH_4	vol %	0.3
C_2 + HC's	vol %	2
CO	vol %	---
N_2	ppm by vol.	6000
H_2O	ppm by vol.	2
O_2	ppm by vol.	100
Mercaptans and other Sulfides	vol %	0.03

The CO_2 product could then be available for use in enhanced oil or gas recovery or for sequestration. Additionally, an economic evaluation, showing the impact of CO_2 capture on the incremental LCOE, was developed. Included in the economic evaluation was a sensitivity study showing the effects of plant capacity factor, CO_2 by-product selling price, investment cost, and make-up power cost, on the incremental LCOE (¢/kWh) and on the mitigation cost for the CO_2 ($/ton of mitigated CO_2).

This Page Intentionally Left Blank

2 STUDY UNIT DESCRIPTION AND BASE CASE PERFORMANCE

This section provides a brief description of the selected Conesville #5 study unit. The study unit is one of six existing coal-fired steam plants located on the site as shown in Figure 2-1. AEP owns and operates these units except for Unit #4, which is jointly owned by AEP, Cinergy, and Dayton Power and Light. The total electric generating capacity on this site is ~2,080 MWe, although two of the older units (Units 1 and 2, shown on the left) have been retired. The steam generated in Unit #5 is utilized in a subcritical steam cycle for electric power generation. The capacity of Conesville Unit #5 is ~430 MWe-net.

Figure 2-1: Conesville Power Station

The Base Case for this study is defined as the unmodified existing study unit firing coal at full load without capture of CO_2 from the flue gas. This represents the “business as usual” operating scenario and is used as the basis of comparison for the CO_2 removal options investigated in this study. The overall performance of the Base Case is presented in Section 2.2.

2.1 Study Unit Description

The power plant analyzed in this study is AEP’s Conesville Unit #5. This unit is a coal-fired steam plant which generates ~430 MWe-net using a subcritical pressure steam cycle. This plant has been in commercial operation since 1976. A general arrangement elevation drawing of the study unit steam generator is shown in Figure 2-2.

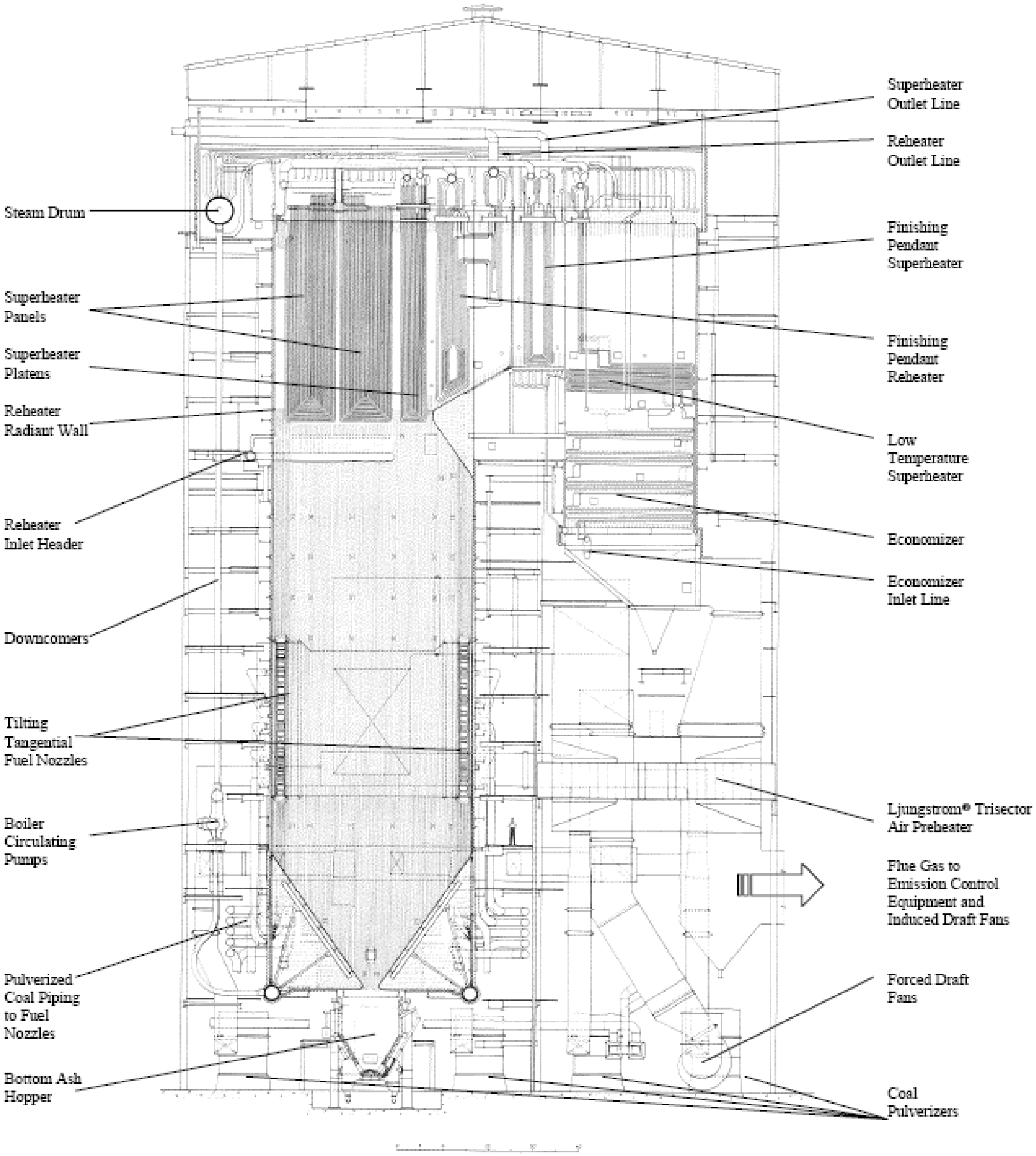


Figure 2-2: Study Unit Boiler (Existing Conesville Unit #5 Steam Generator)

The steam generator can be described as a tangentially coal-fired, subcritical pressure, controlled circulation, and radiant reheat wall unit. The furnace is a single cell design utilizing five elevations of tilting tangential coal burners. The furnace is about 15.75 m (51.67 ft) wide, 13.51 m (44.33 ft) deep and 52.33 m (171.67 ft) high. The unit fires mid-western bituminous coal. The coal is supplied to the five burner elevations with five RP-903 coal pulverizers. The unit is configured in a “Conventional Arch” type design and is representative in many ways of a large number of coal-fired units in use throughout the U.S. today. The unit is designed to generate about 391 kg/s (3.1 x 10^6 lbm/hr) of steam at nominal conditions of 175 bara (2,535 psia) and 538°C (1,000°F) with reheat steam also heated to 538°C (1,000°F). These represent the most common steam cycle operating conditions for the existing U.S. fleet of utility scale power generation systems. Outlet steam temperature control is provided with de-superheating spray and burner tilt.

The superheater is divided into four major sections. Saturated steam leaving the steam drum first cools the roof and walls of the rear pass before supplying the low-temperature superheater section. The low-temperature superheater section is located in the rear pass of the unit and is a horizontal section with the outlet tubes in a vertical orientation adjacent to the finishing superheater section. Steam leaving the low-temperature superheater section first flows through the de-superheater spray stations and then to the radiant superheat division panel section. The division panels are located in the upper furnace directly above the combustion zone of the lower furnace. Steam leaving the division panel section flows to the superheater platen section, which is a more closely spaced vertical section located between the panels and the finishing pendant reheater. Steam leaving the platens flows into the finishing superheater section which is also a pendant section located downstream of the pendant reheater, just before the gas turns downward to enter the low-temperature superheater section in the rear pass of the unit. Steam leaving the finishing superheater is piped to the high-pressure turbine where it is expanded to reheat pressure and then returned to the reheat de-superheating spray station.

The reheater is divided into two sections, a low-temperature radiant wall section followed by a spaced finishing pendent section. Steam is supplied to the reheater radiant wall from the de-superheating spray station, which is fed from the high-pressure turbine exhaust. The reheater radiant wall section is located in the upper furnace and covers the entire front wall and most of the two sidewalls of the upper furnace. The pendant finishing reheat section is located above the arch between the superheat platen and superheat finishing sections. Steam leaving the finishing reheater is returned to the intermediate pressure turbine where it continues its expansion through the intermediate and low-pressure turbines before being exhausted to the condenser.

The gases leaving the low-temperature superheater section are then further cooled in an economizer section. The economizer is comprised of four banks of spiral-finned tubes (0.79 fins/cm or 2 fins/inch), which heats high-pressure boiler feedwater before it is supplied to the steam drum. The feedwater supplying the economizer is supplied from the final extraction feedwater heater.

Flue gas leaving the economizer section then enters the Ljungstrom® trisector regenerative air heater, which is used to heat both the primary and secondary air streams prior to combustion in the lower furnace. Particulate matter is removed from the cooled flue gas leaving the air heater in an electrostatic precipitator (ESP) and sulfur dioxide is removed in a lime-based flue gas de-sulfurization (FGD) system. The induced draft fans are located in between the ESP and the FGD. The cleaned flue gas leaving the FGD system is then exhausted to the atmosphere through the stack, which also serves Unit #6. The induced draft and forced draft fans are controlled to operate the unit in a balanced draft mode with the furnace maintained at a slightly negative pressure (typically -0.5 in wg).

The high-pressure superheated steam leaving the finishing superheater is expanded through the high-pressure steam turbine, reheated in the two-stage reheater and returned to the intermediate pressure turbine. The steam continues its expansion through the low-pressure turbine sections where it expands to condenser pressure. The generator produces about 463 MW of electric power at Maximum Continuous Rating (MCR). The steam cycle utilizes six feedwater heaters (three low-pressure heaters, a deaerator, and two high-pressure heaters) where the feedwater is preheated to about 256°C (493°F) before entering the economizer of the steam generator unit. The boiler feed pump is steam turbine driven with steam provided from the intermediate pressure turbine exhaust and expanded to condenser pressure.

2.2 Base Case Performance Analysis

The Base Case can be described as the unmodified existing unit firing coal at full load and without capture of CO_2 from the flue gas. This represents the "business as usual" operating scenario and is used as the basis of comparison for the CO_2 removal options investigated in this study. The first step in the development of a Base Case was to set up a computer model of the boiler. Using test data from the existing unit, the computer model was then calibrated. The calibrated boiler model was then used for analysis of the Base Case and the CO_2 removal cases. The development of the Base Case was done as part of the original study (Bozzuto et al., 2001) and was not repeated for the current study. The Base Case of the original study was used as the Base Case for the current study description of the Base Case development (extracted from the original study report) is provided in this section.

2.2.1 Calibration of the Boiler Computer Model

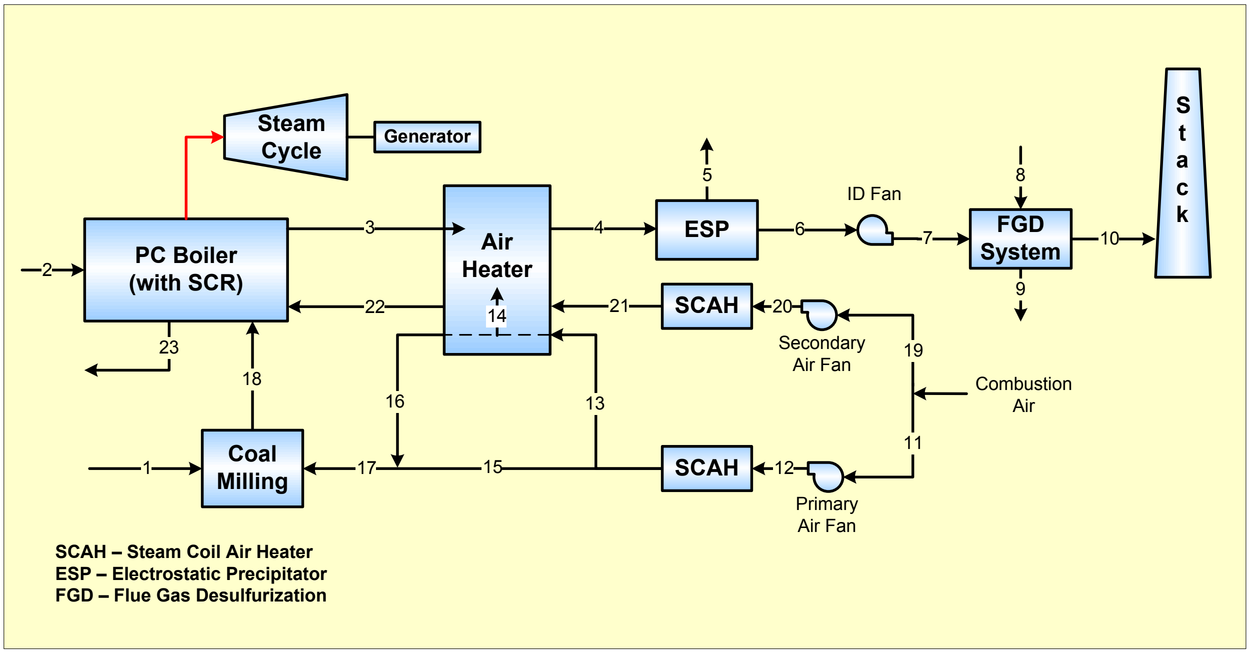
The first step in the calculation of a Base Case was to set up a steady state performance computer model of the Conesville #5 steam generator unit. This involves calculating or obtaining all the geometric information for the unit as required by the proprietary Reheat Boiler Program (RHBP). The RHBP provides an integrated, steady state performance model of the Boiler Island including, in addition to the steam generator unit, pulverizers, air heater, and steam temperature control logic. The RHBP is used to size components and/or predict performance of existing components. In this study, since the boiler island component sizes are known, the RHBP was used exclusively for calculating unit performance.

The next step in the heat transfer analysis of the Base Case was to calibrate the RHBP model of the unit. This involves obtaining test data (with air firing) for the existing unit and "adjusting" the performance model to match the test data. The required test data includes steam temperatures entering and leaving each major heat exchanger section in the unit, steam pressures, coal analysis, flue gas oxygen content, etc. The "adjustments" or "calibration factors" for the model are in the form of "surface effectiveness factors" and "fouling factors" for the various heat exchanger sections throughout the unit. Unfortunately, the test data used for calibration of this model was not totally complete and several assumptions were required in the calibration process. Although all the required data was not available, primarily due to existing instrumentation limitations, a satisfactory calibrated model was obtained.

Using the calibrated boiler model and providing it with new steam side inputs (mass flows, temperatures, and pressures) from the agreed upon MCR steam turbine material and energy balance, the model was run and performance was calculated for the Base Case. The performance for the overall power plant system is described in Section 2.2.2 with the boiler performance shown in Section 2.2.3 and the steam turbine performance in Section 2.2.4.

2.2.2 Overall System Description and Material and Energy Balance (Base Case)

The simplified gas side process flow diagram for the Base Case is shown in Figure 2-3 and the associated material and energy balance for this case is shown in Table 2-1. Overall plant performance is summarized in Table 2-2. This system is described previously in Section 2.2. Boiler efficiency is calculated to be 88.13%. The net plant heat rate is calculated to be 10,285 kJ/kWh (9,749 Btu/kWh) for this case as shown in Table 2-2. Auxiliary power is 29,700 kWe and the net plant output is 433,778 kWe. Carbon dioxide emissions are 109 kg/s (88,156 lbm/hr) or about 907 g/kWh (2.00 lbm/kWh).



Material Flow Stream Identification

1	Raw Coal to Pulverizers	9	FGD System Solids to Disposal	17	Mixed Primary Air to Pulverizers
2	Air Infiltration Stream	10	Flue Gas to Stack	18	Pulverized Coal and Air to Furnace
3	Flue Gas from Economizer to Air Heater	11	Air to Primary Air Fan	19	Secondary Air to Forced Draft Fan
4	Flue Gas Leaving Air Heater to ESP	12	Primary Air to Steam Coil Air Heater	20	Secondary Air to Steam Coil Air Heater
5	Flyash Leaving ESP	13	Primary Air to Air Heater	21	Secondary Air to Air Heater
6	Flue Gas Leaving ESP to Induced Draft Fan	14	Air Heater Leakage Air Stream	22	Heated Secondary Air to Furnace
7	Flue Gas to Flue Gas Desulfurization System	15	Tempering Air to Pulverizers	23	Bottom Ash from Furnace
8	Lime Feed to FGD System	16	Hot Primary Air to Pulverizers		

Figure 2-3: Simplified Gas Side Process Flow Diagram (Base Case)

Table 2-1: Gas Side Material and Energy Balance (Base Case)

Constituent	**(Units)**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**
O_2	(lbm/hr)	26586	42147	101097	144807		144817	144817	5335		144578	203237	203237	112918
N_2	"	4868	139626	2797385	2942220		2942220	2942220			2942220	673283	673283	374075
H_2O	"	37820	2357	228849	231294		231294	231294	250709	45979	436024	11365	11365	6314
CO_2	"			867210	867210		867210	867210			866156			
SO_2	"			20202	20202		20202	20202			1063			
H_2	"	16102												
Carbon	"	236665												
Sulfur	"	10110												
Ca	"								12452					
Mg	"								584					
MgO	"									484				
$MgSO_3$	"									1293				
$MgSO_4$	"									94				
$CaSO_3$	"									31579				
$CaSO_4$	"									2468				
$CaCO_3$	"									2398				
Ash/Inerts	"	42313		33851	33851	33851			968	968				
		Raw Coal	Leakage Air	Flue gas to AH	Flue gas to ESP	Flyash	Flue gas to ID Fan	Flue gas to FGD	Lime Slurry	FGD Disposal	Flue gas to CO_2 Sep	Pri Air to PA Fan	PA from PA Fan	Pri Air to AH
Total Gas	(lbm/hr)		184130	4014743	4205743		4205743	4205743			4390042	887885	887885	493308
Total Solids	"	374455		33851	33851	33851			14003	42884				
Total Flow	"	374455	184130	4048594	4239594	33851	4205743	4205743	270067	88863	4390042	887885	887885	493308
Temperature	(Deg F)	80	80	706	311	311	311	325	80	136	136	80	92	92
Pressure	(Psia)	14.7	14.7	14.6	14.3	14.7	14.2	15	14.7	14.7	14.7	14.7	15.6	15.6
$h_{sensible}$	(Btu/lbm)	0.000	0.000	161.831	57.924	57.750	57.924	61.384	0.000	14.116	14.116	0.000	2.899	2.899
Chemical	(10^6 Btu/hr)	4228.715												
Sensible	(10^6 Btu/hr)	0.000	0.000	655.007	245.567	1.955	243.612	258.166	0.000	3.314	63.916	0.000	2.574	1.430
Latent	(10^6 Btu/hr)	0.000	2.475	240.291	242.858	0.000	242.858	242.858	0.000	0	464.020	11.933	11.933	6.630
Total Energy$^{(1)}$	(10^6 Btu/hr)	4228.715	2.475	895.298	488.425	1.955	486.470	501.024	0.000	3.314	527.936	11.933	14.507	8.060

Constituent	**(Units)**	**14**	**15**	**16**	**17**	**18**	**19**	**20 21 22 23**						
O_2	(lbm/hr)	43720	90319	66680	156999	183585	641283	641283	641283	643801				
N_2	"	144835	299208	299208	520107	524975	2124443	2124443	2124443	3122785				
H_2O	"	2445	5051	3729	8779	46599	35860	35860	35860	36001				
CO_2	"													
SO_2	"													
H_2	"					16102								
Carbon	"					236655								
Sulfur	"					10110								
Ca	"													
Mg	"													
MgO	"													
$MgSO_3$	"													
$MgSO_4$	"													
$CaSO_3$	"													
$CaSO_4$	"													
$CaCO_3$	"													
Ash/Inerts	"					42313					8463			
		Air Htr Lkg Air	Tempering Air	Hot Pri Air	Mixed Pri Air	Coal-Pri Air Mix	Sec Air to FD	Sec air to SCAH	Sec Air to AH	Hot Sec Air	Bottom Ash			
Total Gas	(lbm/hr)	191000	394577	291308	685885		2801587	2801587	2801587	2812587				
Total Solids	"										8463			
Total Flow	"	191000	394577	291308	685885	1060340	2801587	2801587	2801587	2812587	8463			
Temperature	(Deg F)	92	92	666	339		80	86.4	86.4	616.1	2000			
Pressure	(Psia)	15.6	15.6	15.6	15.6	15.0	14.7	15.2 15.1	14.9 14.7					
$h_{sensible}$	(Btu/lbm)	2.899	2.899	145.249	63.358		0.000	1.549	1.549	132.582	480.000			
Chemical	(10^6 Btu/hr)					4228.715								

Sensible	(10^6 Btu/hr)	0.554	1.144	42.312	43.456		0.000	4.341	4.341	372.898	4.062			
Latent	(10^6 Btu/hr)	2.567	5.303	3.915	9.218		37.653	37.653	37.653	37.801	0.000			
Total Energy(1)	(10^6 Btu/hr)	3.121	6.447	46.227	52.674	4281.389	37.653	41.994	41.994	410.699	4.062			
Notes: (1) Energy Basis; Chemical Based on Higher Heating Value (HHV); Sensible energy above 80F; Latent based on 1050 Btu/lbm of water vapor														

Table 2-2: Overall Plant Performance Summary (Base Case)

	Units	Base Plant
Fuel Parameters		
Coal Heat Input (HHV)	10^6 Btu/hr	4228.7
Natural Gas Heat Input (HHV)	10^6 Btu/hr	---
Total Fuel Heat Input (HHV)	10^6 Btu/hr	4228.7
Steam Cycle Parameters		
Existing Steam Turbine Generator Output	kW	463478
CO_2 Removal System Turbine Generator Output	kW	0
Total Turbine Generator Output	kW	463478
Total Auxiliary Power	kW	29700
Net Plant Output	kW	433778
Overall Plant Performance Parameters		
Net Plant Efficiency (HHV)	fraction	0.3501
Net Plant Efficiency (LHV)	fraction	0.3666
Normalized Efficiency (HHV; Relative to Base Case)	fraction	1.0000
Net Plant Heat Rate (HHV)	Btu/kWhr	9749
Net Plant Heat Rate (LHV)	Btu/kWhr	9309
Overall Plant CO_2 Emissions		
Carbon Dioxide Emissions	lbm/hr	866102
Specific Carbon Dioxide Emissions	lbm/kWhr	1.997
Specific Carbon Dioxide Emissions	kg/kWhr	0.906
Normalized Specific CO_2 Emissions (Relative to Base Case)	fraction	1.000

2.2.3 Boiler Analysis Results (Base Case)

The main steam flow for this case and all other cases in this study is 395 kg/s (3,131,619 lbm/hr). The cold reheat flow leaving the high-pressure turbine for this case and all other cases in this study is 348 kg/s (2,765,058 lbm/hr). The hot reheat flow (including de-superheating spray) returning to the intermediate pressure turbine for this case is 359 kg/s (2,850,885 lbm/hr). The overall steam conditions produced by the existing Conesville #5 steam generator unit are shown in Table 2-3 below. To produce these conditions, the superheat circuit requires about 3.6% spray and the reheat circuit requires about 3.1% spray to maintain required steam outlet temperatures. The burner tilts are -10 degrees (the minimum value the customer uses). The boiler was fired with 15% excess air and the resulting boiler efficiency calculated for this case was 88.13% with an air heater exit gas temperature of 155°C (311°F).

Table 2-3: Boiler/Turbine Steam Flows and Conditions (Base Case)

		SHO	FWI	ECO	RHO	RHI
Mass Flow	(lbm/hr)	3131619	3131619	3017507	2850885	2850885
Pressure	(psia)	2535	3165	3070	590.8	656.5
Temperature	(Deg F)	1005	496.2	630	1005	607.7
Enthalpy	(Btu/lbm)	1459.7	483.2	652.8	1517.1	1290.4

Notes:
SHO = Superheater Outlet; FWI = Feedwater Inlet; ECO = Economizer Outlet; RHO = Reheater Outlet; RHI = Reheater Inlet

2.2.4 Steam Cycle Performance (Base Case)

The selected steam turbine energy and mass flow balance for Conesville #5, which provides the basis for developing the steam turbine performance calculations presented in this study is shown in Figure 2-4.

This turbine heat balance diagram, created by Black & Veatch, is a valves-wide-open, 5% over pressure case utilizing a condenser pressure of 6.35 cm Hga (2.5 in.-Hga) and a steam extraction for air heating of 6.3 kg/s (50,000 lbm/hr). Following general guidelines it is assumed that this diagram reflects the design maximum allowable flow conditions of the existing turbine.

In order to reflect the key performance parameters of the selected unit “as designed,” the Black & Veatch heat balance diagram was accurately re-modeled and the following adaptations to real mode operations were made:

- During normal operation no steam is required to feed the steam coil air heaters (6.3 kg/s or 50,000 lb/hr). Therefore, this extraction flow is set to zero.
- Reheat de-superheater spray water flow rate of 11 kg/s (85,827 lb/hr) is to be used as calculated in associated boiler performance computer simulation runs.

Keeping all other conditions constant, namely live steam (LS) pressure and temperature, reheat (RH) temperature and backpressure, the turbine base model reacts to the increase in RH spray (from zero to 11 kg/s or 85,827 lb/hr) and the switch-off of the extraction flow to the air pre-heaters (from 6.3 kg/s to 0 kg/s or from 50,000 lb/hr to 0 lb/hr) with a slight reduction in live steam flow due to the given swallowing capacity of the HP turbine (-0.26% in LS flow). In order to allow comparison with previous investigations the swallowing capacity was slightly re-adjusted to allow the nominal flow of 395 kg/s (3,131,619 lb/hr) at 5% overpressure.

The calculated power output applying this model showed some deficiency when compared to previous studies. This is partly due to the improved detailed modeling of the LP turbine performance, and to other differences between the previous and current models. Again, in order to allow comparison with previous investigations, the generator efficiency was adjusted in a way to allow easy comparison with previous results. Although the resulting generator efficiency may reach higher than typical values, this method allows easy comparison and simple adjustment between the two analyses, by just modifying the generator efficiency.

The final steam cycle for the Base Case is shown schematically in Figure 2-5. Figure 2-6 shows the associated Mollier diagram, which illustrates the process on enthalpy-entropy coordinates. The high-pressure turbine expands about 391 kg/s (3.1 x 10^6 lbm/hr) of steam at 175 bara (2,535 psia) and 538°C (1,000°F). Reheat steam is returned to the intermediate pressure turbine at 610 psia and 1,000°F. These conditions (temperature and pressure) represent the most common steam cycle operating conditions for existing utility-scale power generation systems in use today in the U.S. The condenser pressure used for the Base Case and all other cases in this study was 6.35 cm Hga (2.5 in. Hga). The steam turbine performance analysis results show the generator produces an output of 463,478 kWe and the steam turbine heat rate is about 8,200 kJ/kWh (7,773 Btu/kWh).

The key parameters describing the reference case are listed below:

Live steam pressure	2,535 / 175	psia / bara
Live steam temperature	1,000 / 538	°F / °C
Live steam flow	3,131,619 / 395	lbm/hr / kg/s
Steam for air pre-heating	0 / 0	lbm/hr / kg/s
RH de-superheating spray	85,827 / 11	lbm/hr / kg/s
Backpressure	2.5 / 6.35	In. Hg abs / cm Hg abs
Power output	463,478	kW

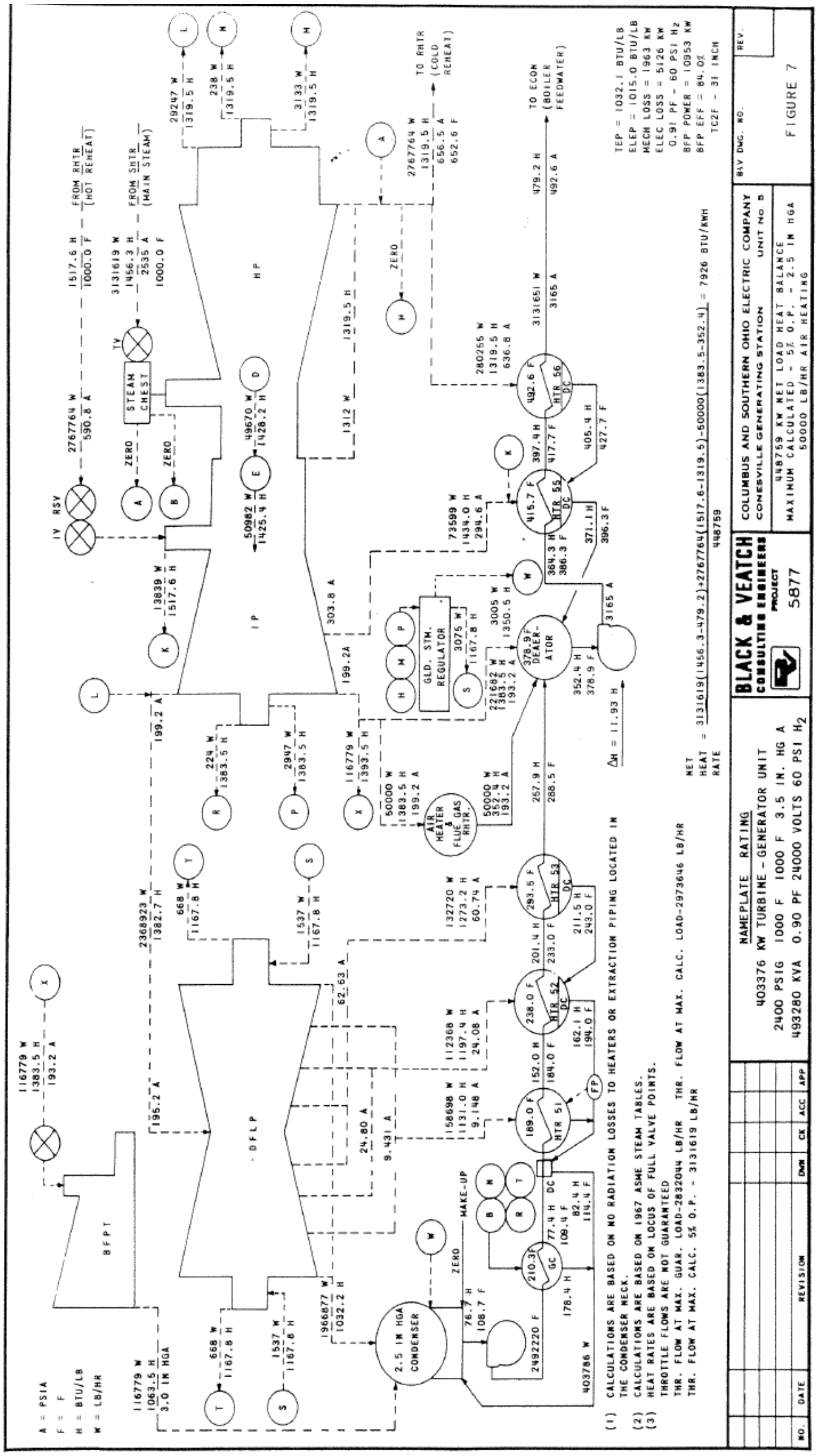


Figure 2-4: Selected Conesville #5 Turbine Heat Balance (basis for steam turbine modeling)

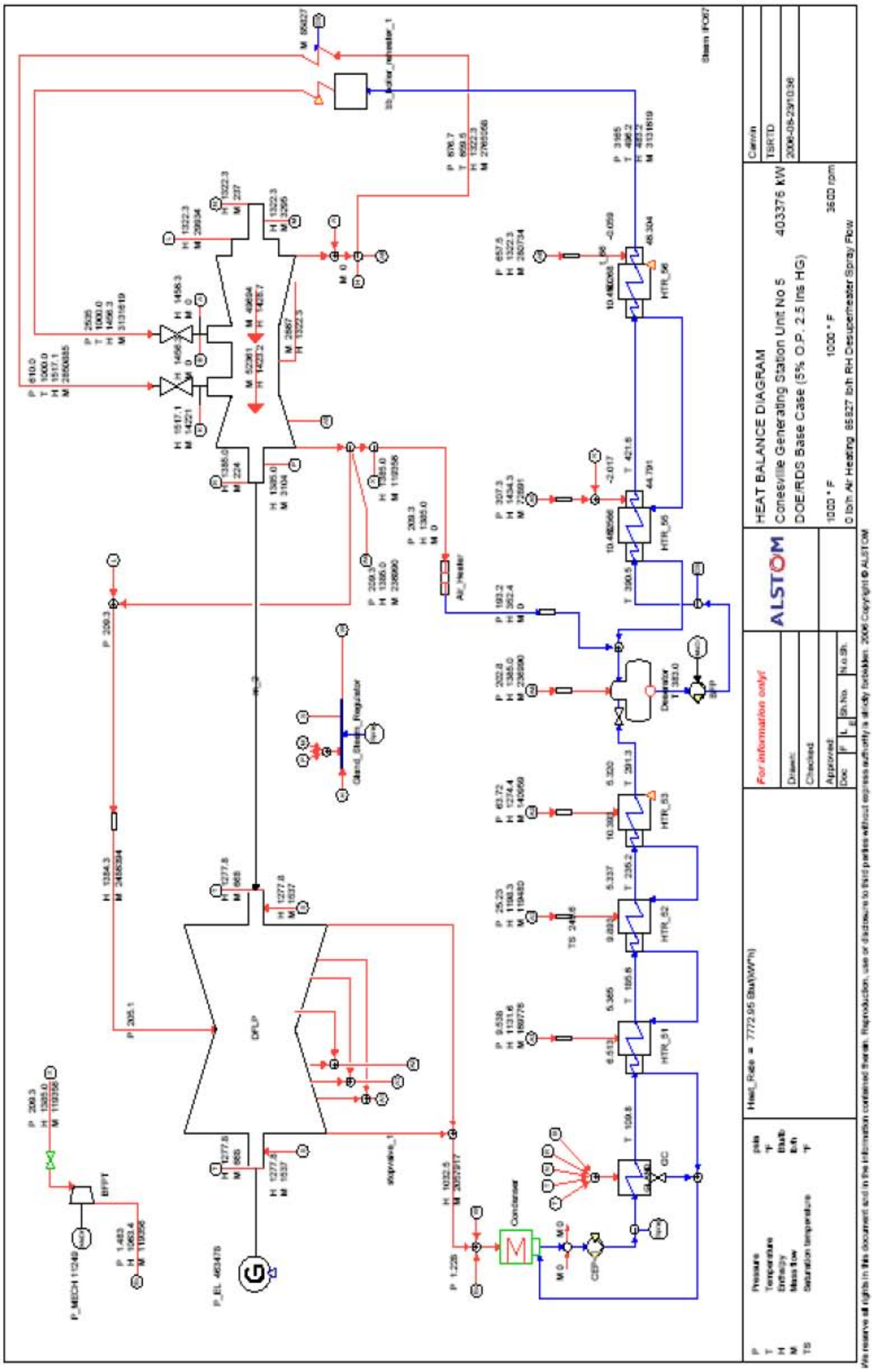


Figure 2-5: Steam Cycle Diagram and Performance (Base Case)

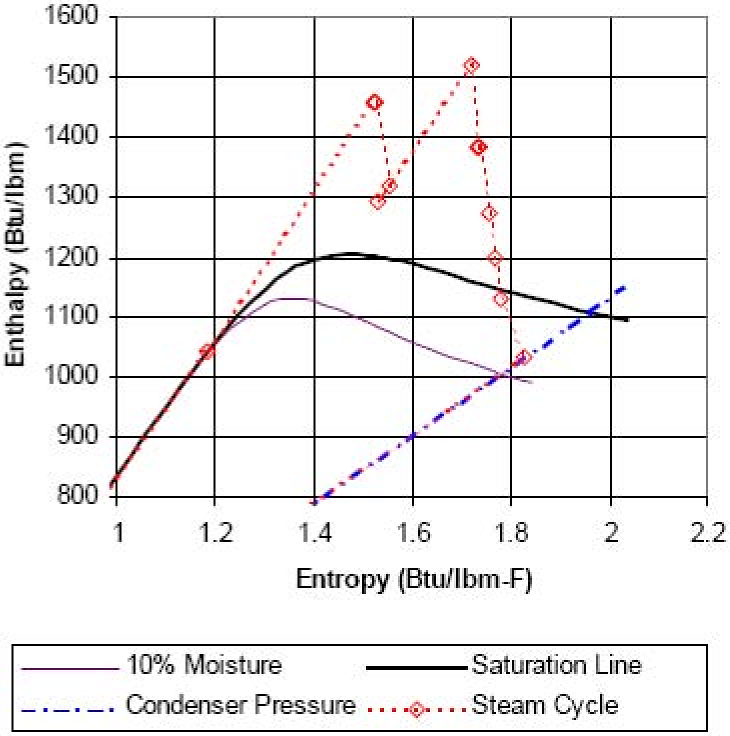


Figure 2-6: Steam Cycle Mollier Diagram (Base Case)

2.2.5 Flue Gas Desulfurization System Analysis (Base Case)

Figure 2-7 shows the process flow diagram for the existing Flue Gas Desulfurization System. The stream numbers in Figure 2-7 also correspond to stream numbers shown in Figure 2-3. The flue gas leaving the ID fan (Stream 7) is delivered to the absorber, which consists of a tray followed by a two-stage spray system. The incoming gas is saturated as it passes through the scrubbing slurry contained on the tray and through the two spray levels. The active component of the scrubbing slurry is calcium oxide (Stream 8a), which reacts with sulfur dioxide to form calcium bisulfite (Stream 9). The scrubbing slurry is circulated from the reagent feed tank that forms the base of the scrubber to the spray levels. The solids loading in the scrubbing slurry controls the blow down from the reaction tank to by-product disposal. The flue gas passes through chevron-type mist eliminators that remove entrained liquid before exiting the scrubber (Stream 10). The water utilized in spray washing the mist eliminators also serves as make-up (Stream 8b).

Table 2-4 identifies the assumptions that were made in predicting the FGD performance. Table 2-5 shows the gas constituents at the existing absorber inlet and outlet locations. Results show a CO_2/SO_2 mole ratio of 63 and an SO_2 removal efficiency of 94.9%, corresponding to a value of 104 ppmv at the outlet of the absorber.

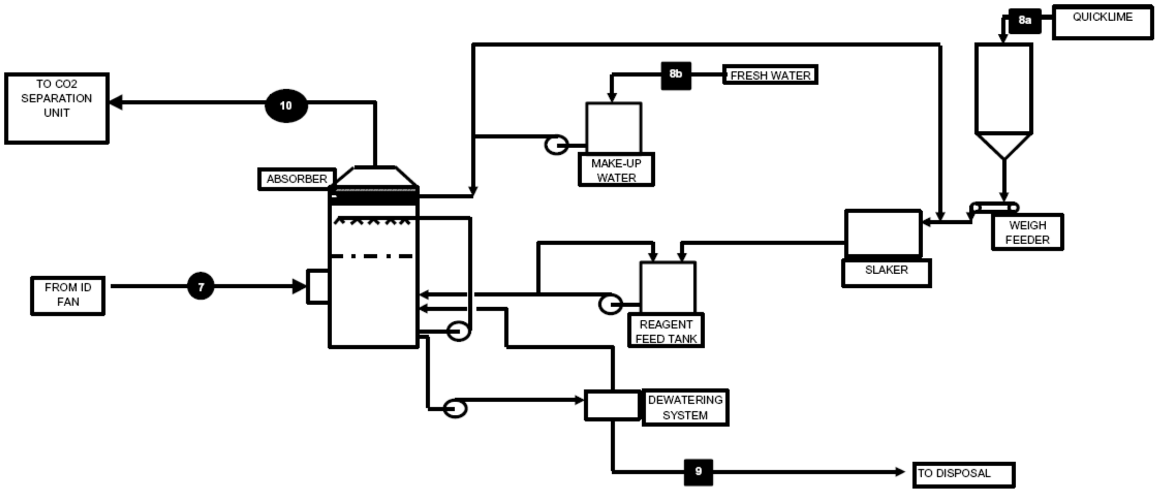


Figure 2-7: Existing Flue Gas Desulfurization System Process Flow Diagram

Table 2-4: FGD System Analysis Assumptions

Quantity	Unit	Existing Absorber
Ca/SO	Mol Ratio	1.04
Solids	Wt.%	20
CaO	Wt.%	90
MgO	Wt.%	5
Inerts	Wt.%	5
Bypass Leakage	Wt.%	2.5
Liquid/Gas (L/G) Ratio	gpm/1000 acfm	55
SO_2 Removal Efficiency		
APC	%	94.8
Absorber	%	97.2

Table 2-5: Existing FGD System Performance

Base Case						
	Existing Absorber Inlet			Existing Absorber Outlet		
Species	Mol/hr	Vol.%	Unit	Mol/hr	Vol.%	Unit
O_2	4,469	3.14	Vol.%	4,461	2.91	Vol.%
N_2	105,018	73.74	Vol.%	105,018	68.44	Vol.%
H_2O	12,863	9.03	Vol.%	24,228	15.79	Vol.%
CO_2	19,743	13.86	Vol.%	19,720	12.85	Vol.%
SO_2	315	2,212	ppmv	16	104	ppmv
SO_2 Removal Efficiency, %					94.9	
CO_2/SO_2 Mole Ratio		63				

3 THE SENSITIVITY OF PLANT PERFORMANCE AND ECONOMICS TO CO_2 CAPTURE LEVEL

This section describes the analysis of the impacts of CO_2 capture level. All CO_2 capture levels were done at the solvent regeneration energy level of 1,550 Btu/lbm-CO_2. As mentioned previously, the solvent regeneration energy level of 1,550 Btu/lbm-CO_2 represents the state of the art at the time of this study (ca. 2006). By investigating various levels of capture, the potential exists for identifying an economic optimum as well as simply quantifying the effect of this important variable on typical measures of plant performance and economic merit. Four CO_2 capture levels (90%, 70%, 50%, and 30%) are investigated in this study. These CO_2 capture levels are referred to as Cases 1, 2, 3, and 4, respectively and these four cases represent the primary case studies for this effort. Additionally, Concept A from the 2001 study was updated (costs and economics only) and is referred to as Case 5. This case used an amine system with ~96% CO_2 capture and solvent regeneration energy requirements of 2,350 Btu/lbm-CO_2.

The primary impacts are quantified in terms of plant electrical output reduction, thermal efficiency, CO_2 emissions, retrofit investment costs, and the incremental cost of generating electricity resulting from the addition of the CO_2 capture systems.

3.1 Study Unit Modifications and Definition of the Amine-Based CO_2 Capture Systems

This section provides most of the technical data for the retrofit cases comprising this study. It also discusses the complete retrofit to the power plant in terms of performance, equipment modifications and new equipment required. Each of the five study cases has equipment designed for the removal and recovery of CO_2 from the boiler flue gas using an amine scrubbing system. Plant material and energy balances are provided for the new and existing major systems and the equipment added or modified to complete the retrofit. The first subsection discusses the design basis used for the study. The second subsection (Section 3.1.2) discusses the boiler island and performance and equipment modifications. The third and fourth subsections discuss the amine-based CO_2 capture and compression systems. The advanced amine systems are discussed first (Section 3.1.3) followed by a review of the amine system from the previous study (Bozzuto et al., 2001) in Section 3.1.5. Finally a discussion of the steam/water cycle modifications and new equipment is presented in Section 3.1.6.

Cases 1-4 (90%, 70%, 50%, and 30% capture, respectively), which use the advanced amine systems, comprise the primary cases of the current study.

A fifth case **(Case 5)** is simply an update of “Concept A” from a previous study (Bozzuto et al., 2001). The update to this case consisted of simply escalating the investment and operating and maintenance costs from 2001 to 2006 $U.S. and re-calculating the economic analysis such that comparisons between the current study results and the previous results could be done on an equivalent basis. The process design and equipment selections for Case 5/Concept A were not updated.

The current study differs from the previous study in several ways, as listed below:

- First, an advanced amine CO_2 scrubbing system is used for CO_2 removal from the flue gas stream. This advanced system requires significantly less energy for solvent regeneration. Solvent regeneration for this system requires about 3.6 MJ/Tonne CO_2 ($3.1x10^6$ Btu/Ton CO_2) (~34% reduction). Additionally, the reboiler was operated at 3.1 bara (45 psia) as opposed to 4.5 bara (65 psia) in the previous study.
- Second, several CO_2 capture levels are investigated in this study (90%, 70%, 50%, and 30%). These are referred to as Cases 1, 2, 3, and 4 respectively in this study. Previously only one CO_2 capture level (96%) was investigated.
- Third, the current study differs from the previous study in that Alstom's steam turbine retrofit group developed a detailed analysis of the modified existing steam turbine. Previously, a more simplified analysis was used for the existing steam turbine.
- Another difference is that in the current study, significant quantities of heat rejected from the CO_2 capture/compression system are integrated with the steam/water cycle. Previously, heat integration was not used because the new CO_2 capture/compression system was located too far away (>1,500 ft) from the existing steam/water system.

3.1.1 Design Basis for CO_2 Capture Systems Retrofit Equipment and Performance Calculations (Cases 1-5)

This section describes many of the assumptions and data used for design of the equipment and in the calculation of process performance.

3.1.1.1 Site Data

Listed below is the summary of the site data used for equipment design:

- Plant is located in Conesville, Ohio, elevation 227 m (744 ft).
- Atmospheric pressure is 76 cm Hga (29.92 in. Hg).
- Dry bulb maximum temperature is 33°C (92°F) and minimum is -1°F.
- Wet bulb temperature for cooling tower design is 24°C (75°F).
- Average cooling tower water temperature is 27°C (80°F).
- Electric power is available from the existing facilities. Auxiliary power is provided through auxiliary transformers at 4,160-volt bus and is reduced down to 480 volts.
- 316L stainless steel is the preferred material of construction where the flue gas cooling systems contain halides and sulfur oxides.
- Pressure of product CO_2 is 139 bara (2,015 psia).
- For all plant performance calculations and material and energy balances the atmospheric conditions to be assumed are the standard conditions of 27°C /80°F, 1.014 bara/14.7 psia, 60% relative humidity).
- Condenser pressure used for all turbine heat balances is 2.5 in. Hga.

3.1.1.2 Fuel Analyses

Table 3-1 shows the coal analysis used for this study and Table 3-2 shows the natural gas analysis. Natural gas was used for desiccant regeneration in the CO_2 drying package.

Table 3-1: Coal Analysis

Proximate Analysis, Wt.%	
Moisture	10.1
Ash	11.3
Volatile Matter	32.7
Fixed Carbon	45.9
Total	**100.0**
Ultimate Analysis, Wt.%	
Moisture	10.1
Ash	11.3
H	4.3
C	63.2
S	2.7
N	1.3
O	7.1
Total	**100.0**
Higher Heating Value	
Btu/lbm	11,293
kJ/kg	26,266

Table 3-2: Natural Gas Analysis

Component		**Vol.%**
Methane		93.9
Ethane		3.2
Propane		0.7
n-butane		0.4
Carbon Dioxide		1.0
Nitrogen		0.8
Total		**100.0**
	LHV	**HHV**
kJ/kg	47805	53015
kJ/scm	35	39
Btu/lbm	20552	22792
Btu/scf	939	1040

Battery Limit Definition

Figure 3-1 shows a plot plan view of the existing Conesville Unit #5 with the major new equipment locations identified for Cases 1-4.

The new secondary SO_2 absorber for the modified FGD system is located just north and adjacent to the existing lime preparation and SO_2 scrubber equipment building in order to minimize the length of new ductwork and the associated draft losses.

The new amine plant absorbers are located ~30 m (100 feet) west of the Unit #5 stack to minimize the length of ductwork and the associated draft losses. The amine regenerators (Strippers) are located ~61 m (200 feet) south of Unit #5's steam turbine to minimize the length of low pressure steam piping and the associated pressure drops. The CO_2 compression, dehydration, and liquefaction facilities are located ~150 m (500 feet) south of the CO_2 strippers to minimize pressure drop in the connecting duct.

The CO_2 recovery and liquefaction equipment receives cooling water from the existing plant steam/water cycle (the existing plant cooling system). The availability of plant cooling water from the existing plant is the result of diverting steam that would have been used to generate power to the amine regeneration plant. This steam would have been condensed by water from the existing plant cooling tower but is now condensed by the amine regenerators.

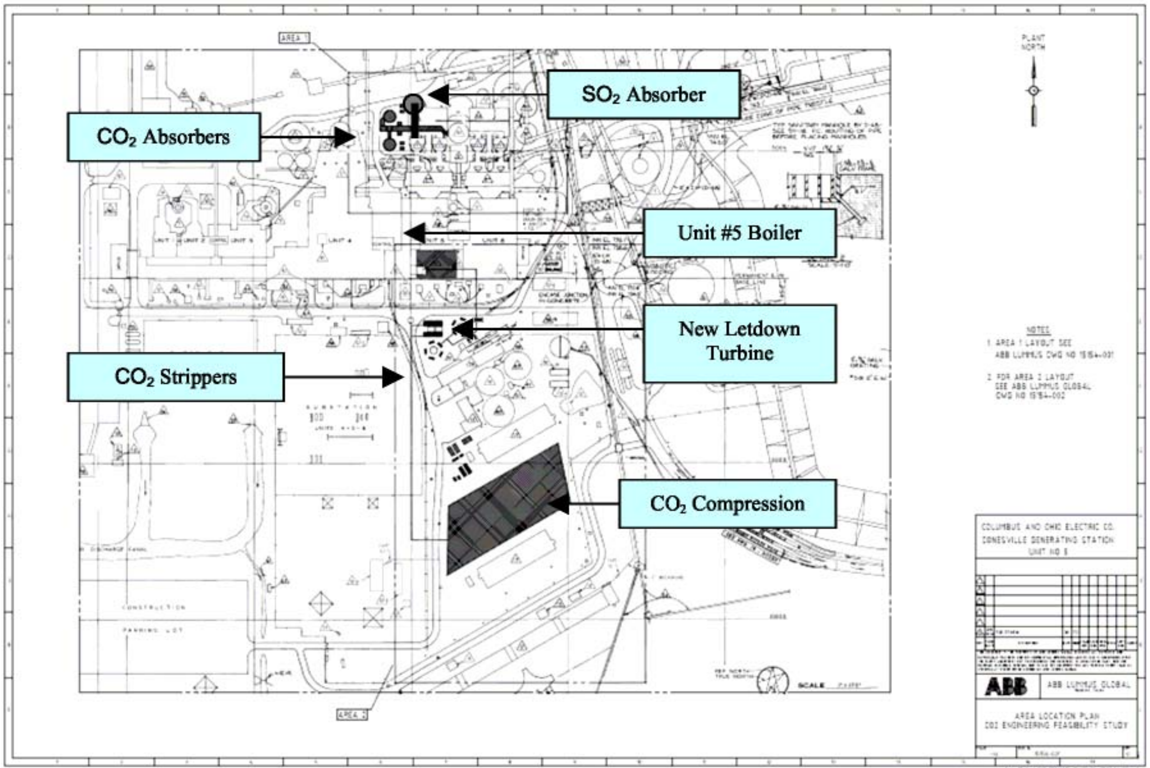


Figure 3-1: AEP Conesville, Ohio, Electric Power Generation Station Site and New Equipment Locations (Cases 1-4)

The CO_2 recovery and liquefaction sections have their own control room and MCC. In addition to the flue gas, which serves as the feed to the unit, it must also receive the required utilities and chemicals. Soda ash - if available from existing facilities - can be used to maintain levels in this facility's day tanks. Otherwise it can be off-loaded from trucks into the day tanks. Diatomaceous earth used in the amine filtration equipment will be off-loaded on skids. The spent diatomaceous earth leaves the plant in drums. Amine reclaimer effluent will be collected in a tank truck parked at one end of the unit. Potable water for eye washes and cooling tower make-up water for hose down will be routed along side the CO_2 gas duct. Corrosion inhibitor, to provide oxygen resistance to the amine, will be provided directly from drums into an injection package.

The CO_2 capture and liquefaction sections are based on the following flue gas analysis, which is taken after the modified Flue Gas Desulfurization system (FGD). See Table 3-3.

Table 3-3: Flue Gas Analysis Entering Amine System (Cases 1-5)

Component	Mole %
O_2	2.94
N_2	68.31
H_2O	15.95
CO_2	12.80
SO_2	<10 ppmv
MW	28.59
T (°F)	136
P(psia)	14.7

3.1.1.3 CO_2 Product Specification

The CO_2 product specification is shown in Table 3-4 below. This specification was taken from the Dakota Gasification Company product specification for EOR (Dakota, 2005). A CO_2 product pressure of 139 bara (2,015 psia) is used in all the cases that follow.

Table 3-4: CO_2 Product Specification

Component	Mole %
O_2	0.0100
N_2	0.6000
H_2O	0.0002
CO_2	96.000
H_2S	0.0001
Mercaptans	0.0300
CH_4	0.3000
C_2 + Hydrocarbons	2.0000

3.1.1.4 CO_2 Recovery Process Simulation Parameters

For Cases 1-4, which all use the advanced “state of the art” amine process, a commercial simulator called ProTreat® Version 3.3 was used to simulate the MEA process. Hysys® Version 2004.2 was used to simulate CO_2 compression and liquefaction systems.

The material balances for Case 5/Concept A were run on two process simulators: Hysim and Amsim. Amsim was used for the Absorption/Stripping systems while Hysim was used for the conventional systems as follows:

Flue Gas feed	Hysim
Absorber and Stripper	Amsim
Compression liquefaction	Hysim

The key process parameters used in the simulations are listed in Table 3-5 as well as data from a built and operating plant.

AES Corporation owns and operates a 200 STPD food grade CO_2 production plant in Oklahoma. This plant was designed and built by ABB Lummus Global as a part of the larger power station complex using coal-fired boilers. This plant was started up in 1990 and has been operating satisfactorily with lower than designed MEA losses. The key process parameters from the present designs for Cases 1-4, which use the advanced amine system, and Case 5/Concept A, which uses the Kerr/McGee ABB Lummus amine system, are compared with those from the built and operating AES plant (Barchas and Davis, 1992) in Table 3-5.

Table 3-5: Key Parameters for Process Simulation

Process Parameter	AEP Design Cases 1-4	AEP Design Case 5	AES Design
Plant Capacity, Ton/Day	9,350-3,120	9,888	200
CO_2 in Feed, mol %	12.8	13.9	14.7
O_2 in Feed, mol %	2.9	3.2	3.4
SO_2 in Feed, ppmv	10 (Max)	10 (Max)	10 (Max)
Solvent	MEA	MEA	MEA
Solvent Conc. Wt%	30	20	15 (Actual 17-18%Wt)
Lean Loading, mol CO_2/mol amine	0.19	0.21	0.10
Rich Loading, mol CO_2/mol amine	0.49	0.44	0.41
Stripper Feed Temp, °F	205	210	194
Stripper Bottom Temp, °F	247	250	245
Feed Temp To Absorber, °F	115	105	108
CO_2 Recovery, %	30-90	96	90 (Actual 96-97%)
Absorber Pressure Drop, psi	1	1	1.4
Stripper Pressure Drop, psi	0.7	0.6	4.35
Rich/Lean Exchanger Approach, °F	40	10	50
CO_2 Compressor 1st /Stage Temp, °F	125	105	115
Liquid CO_2 Temp, °F	82	82	-13
Steam Use, lbs Steam/ lb CO_2 captured	1.67	2.6	3.45
Liquid CO_2 Pressure, psia	2,015	2,015	247

3.1.1.5 Chemicals

This section provides data for the chemicals available on site and used by the CO_2 Recovery Unit. Conditions for liquid chemicals are specified at grade level.

Table 3-6: Soda Ash (Na_2CO_3) Requirements

Property	Pressure at B.L. Psia	Temperature °F
Normal	30	Ambient
Mechanical Design	65	125

- Available for reclaiming MEA
- The import and dilution facilities will be used to keep a day tank in the process area at desirable levels

3.1.1.6 Utilities

De-superheated steam at 3.2 bara (47 psia) is supplied to the amine regeneration system from a new low-pressure (LP) let down turbine that will operate in parallel with the existing LP turbine.

Steam for the new LP let down turbine comes from the existing intermediate pressure (IP) turbine outlet.

Steam:

Reboiler Source: Low-pressure steam from the new LP let down turbine outlet:
The steam leaving the let down turbine is used in the amine regeneration system reboilers for process heating.

Table 3-7: Process Steam Conditions (reboilers)

Property	Pressure at B.L. Psia	Temperature °F
Minimum (for process design)	43	272
Normal	45	274
Maximum	50	281
Mechanical Design	300	500

Reclaimer Source: Low-pressure steam from the existing IP turbine outlet:
The steam leaving the IP turbine is used in the amine system reclaimer for amine reclamation.

Table 3-8: Process Steam Conditions (reclaimer)

Property	Pressure at B.L. Psia	Temperature °F
Minimum (for process design)	85	316
Normal	90	320
Maximum	95	324
Mechanical Design	300	500

Water:

Cooling Water:

Source: Existing Cooling Towers

Table 3-9: Cooling Water Conditions

CW Supply	Pressure at B.L. (Psia)	Temperature °F
Minimum	60	70
Normal	65	80
Maximum	90	95
Mechanical Design	150	150

CW Return	Pressure at B.L. (Psia)	Temperature °F
Minimum		100
Normal	45	110
Maximum		135
Mechanical Design	150	175

Table 3-10: Surface Condensate (for amine make-up)

Property	Pressure at B.L. (Psia)	Temperature °F
Normal	135	110
Mechanical Design	175	200

Raw Water (Fresh Water):

Fresh water is distributed for general use at hose stations. The source of this water is the clarifier, which is used for cooling tower make-up. The capacity of the existing clarifier is sufficient for make up. Its quality is as follows:

Table 3-11: Raw Water (fresh water)

Components	Unit	Specifications
Si	ppm	22
Iron (as Fe)	ppm	0.18
Copper (as Cu)	ppm	0.05
Suspended Solids	ppm	15
Chlorine	ppm	100-180
Alkalinity	ppm	100
Na	ppm	100

Potable Water:

Potable water comes from public network for safety showers and eye washes and requirements are defined below:

Table 3-12: Potable Water

Property	Pressure at B.L. (Psia)	Temperature °F
Normal	115	Ambient
Mechanical Design	150	150

Air:

Plant air and instrument air requirements are defined below:

Table 3-13: Plant Air

Property	Pressure at B.L. Psia	Temperature °F
Normal	130	100
Mechanical Design	190	150

Dew point (at normal supply pressure - 40°C)

Table 3-14: Instrument Air

Property	Pressure at B.L. (Psia)	Temperature °F
Normal	130	100
Mechanical Design	190	150

Dew point (at normal supply pressure - 40°C)
Dust, oil and grease free

Fuel Gas:

Fuel gas (natural gas) requirements are defined below:

Table 3-15: LP Fuel Gas (natural gas)

Property	Pressure at OSBL (Psig)	Temperature °F
Normal	50	Ambient
Mechanical Design	100	150

Power Supply:

All of the required power (100%) for the CO_2 Recovery Unit will be provided by AEP either from the local supply or from the Ohio Grid.
Source: Conesville auxiliary power system at 4,160 volts or stepped down to 480 volts.

Table 3-16: Power Supply Requirements

Service	Voltage	Phase
Auxiliary plant power system	4160	3-phase
Large Motors	4160	3-phase
Small Motors	480	3-phase
Instruments, Lighting, etc.	480 / 230	3/1-phase

3.1.2 Boiler Island Modifications and Performance (Cases 1-5)

This section describes boiler island modifications and performance for the study unit. The modifications to the boiler island and the boiler island performance shown in this section are applicable to all five cases of this study.

3.1.2.1 Boiler Modifications

For this project the boiler scope is defined as everything on the gas side upstream of the FGD System. Therefore, it includes equipment such as the Conesville #5 steam generator, pulverizers, fans, ductwork, electrostatic precipitator (ESP), air heater, coal and ash handling systems, etc. Purposely not included in the boiler scope definition is the FGD system. The FGD system modifications are shown separately in Section 3.1.2.2.

For all the CO_2 capture options investigated in this study (Cases 1-5), Boiler Scope is not modified from the Base Case configuration.

3.1.2.2 Flue Gas Desulfurization System Modifications and Performance

The FGD system for all five cases is modified with the addition of a secondary absorber to reduce the SO_2 content to 10 ppmv or less as required by the amine system downstream.

Modified FGD System Process Description and Process Flow Diagram

The principle of operation of the FGD system is briefly described previously in Section 2.2.5 and is not repeated here. In the five capture cases, however, the entire flue gas stream leaving the existing FGD system absorber is supplied to the new secondary absorber and the flue gas stream leaving the secondary absorber provides the feed stream source for the new amine CO_2 absorption systems. Additional piping and ductwork is required as shown in Figure 3-2, which provides a simplified process flow diagram for the modified FGD system.

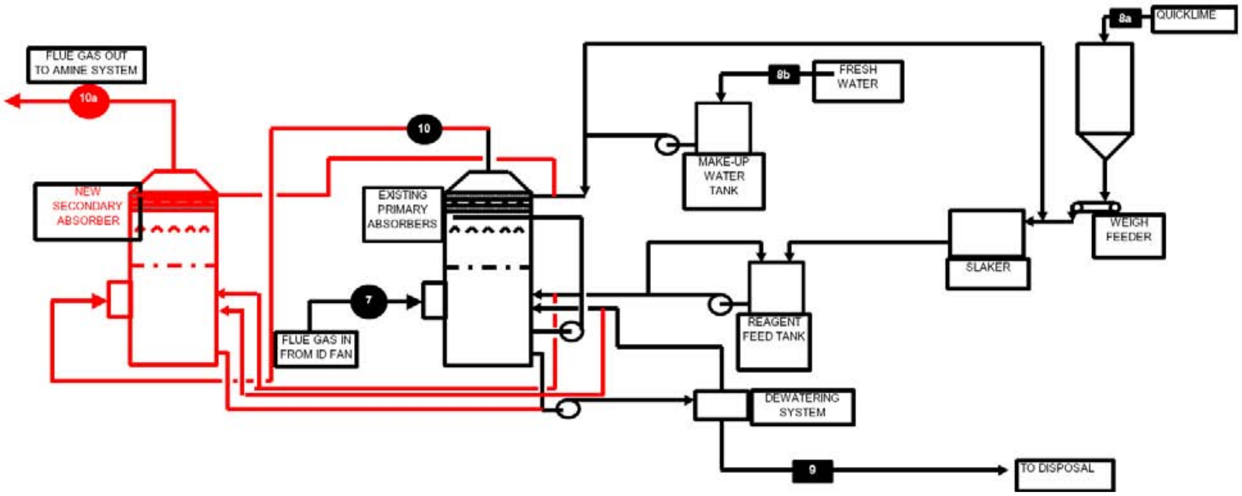


Figure 3-2: Modified FGD System Simplified Process Flow Diagram (Cases 1-5)

Modified FGD System Performance

Table 3-17 identifies the assumptions that were made in predicting the modified FGD system performance.

Table 3-17: Modified FGD System Assumptions (Cases 1-5)

Quantity	Unit	Existing Absorber	Secondary Absorber
Ca/S	Mol Ratio	1.04	1.04
Solids	Wt.%	20	20
CaO	Wt.%	90	90
MgO	Wt.%	5	5
Inerts	Wt.%	5	5
By-pass Leakage	Wt.%	2.5	0
Liquid/Gas (L/G) Ratio	gpm/1000 acfm	75	45
SO_2 Removal Efficiency			
APC	%	94.8	93.0
Absorber	%	97.2	93.0

Table 3-18 indicates the modified FGD system performance by identifying gas constituents at the existing absorber inlet and secondary absorber outlet. Results show a CO_2/SO_2 mole ratio of 63 and an overall SO_2 removal efficiency of 99.7%, corresponding to a value of 6.5 ppmv SO_2 at the outlet of the secondary absorbers.

Table 3-18: Modified FGD System Performance (Cases 1-5)

Constituent	Existing Absorber Inlet			Secondary Absorber Outlet		
	lbm/hr	Mol/hr	Vol %	lbm/hr	Mol/hr	Vol %
O_2	144817	4526	3.18	144566	4518	2.94
N_2	2942220	105019	73.75	2942220	105019	68.31
H_2O	231294	12838	9.02	441924	24530	15.95
CO_2	867210	19705	13.84	866102	19680	12.80
SO_2	20202	315	0.22	87	1	0.00
SO_2, ppmv			2215			8.8
Total	4205743	142403	100	4394900	153748	100
SO_2 Removal Efficiency, %		94.9			99.6	
CO_2/SO_2, Mole Ratio			62			

Modified FGD System Equipment Layout

Figure 3-3 shows the location of the new secondary SO_2 absorber. The new secondary absorber is a single vessel, which is 12.8 m (42 ft) in diameter, and is located just to the north and adjacent to the existing Conesville Unit #5 lime preparation and scrubber equipment building (i.e. label

#53 shown in green in the lower right part of Figure 3-3). This location minimizes the length of ductwork running from the existing FGD system to the new secondary SO_2 absorber and the ductwork length from the secondary SO_2 absorber to the new CO_2 absorbers. The blue lines indicate alterations, which must be made to the access roads located in this area.

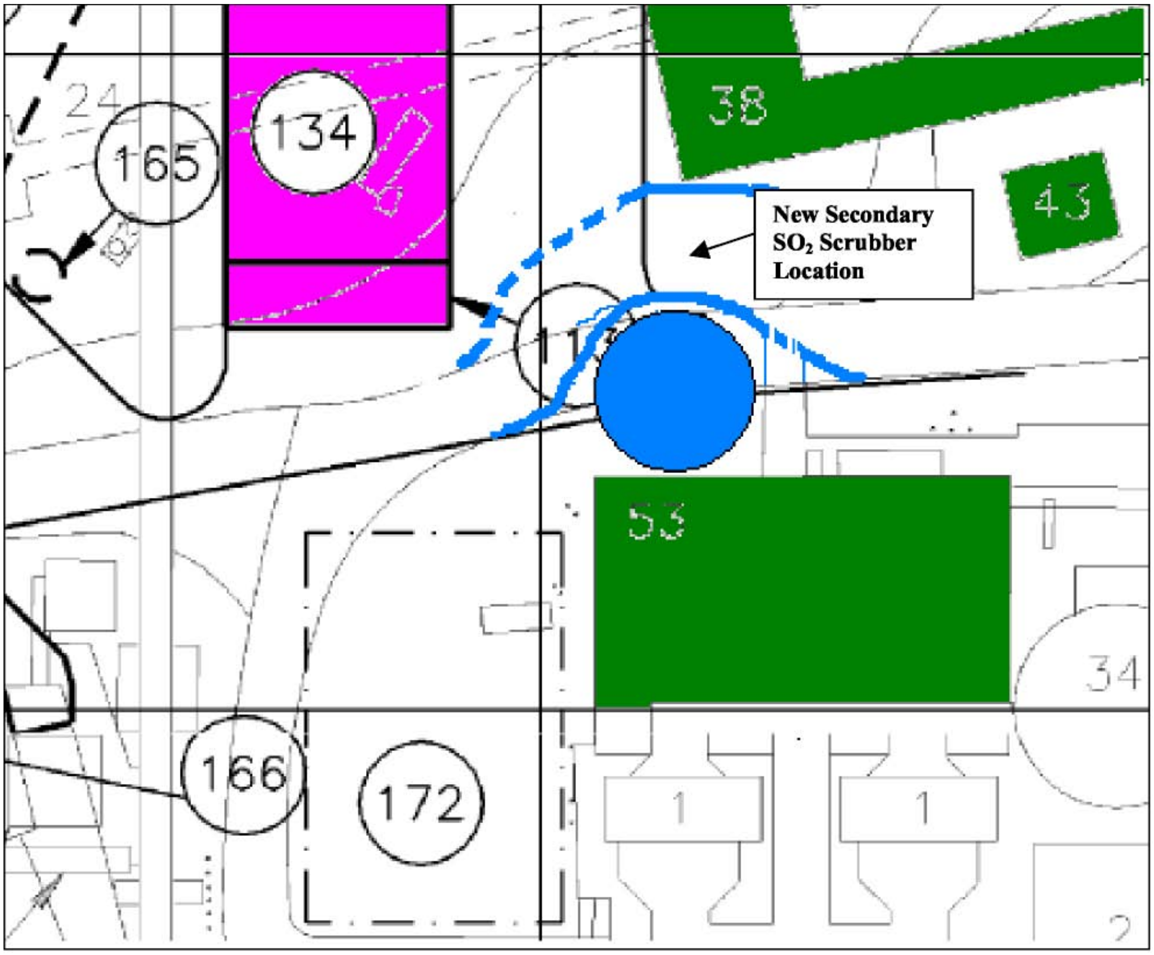


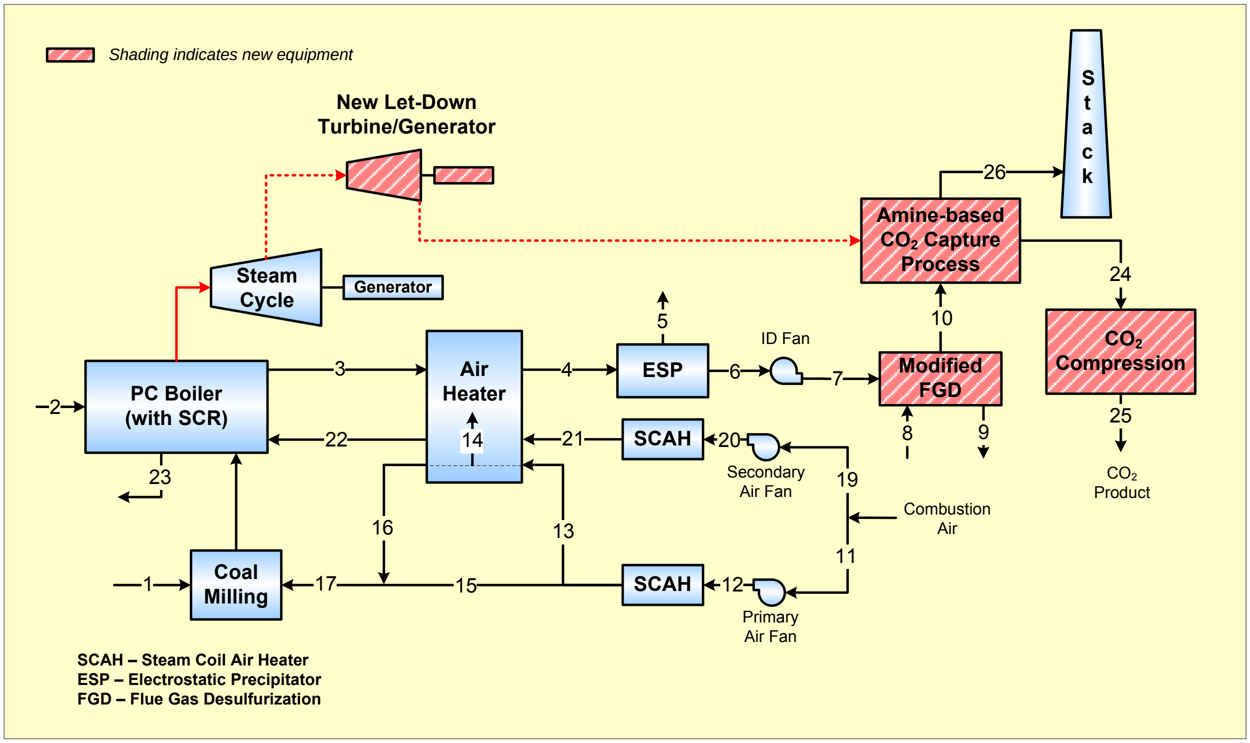
Figure 3-3: New Secondary SO_2 Scrubber Location (Cases 1-4)

Secondary FGD Absorber Effluent:

The existing plant uses lime in its FGD system. In the cost estimate of this plant, it has been assumed that the existing plant disposal facilities can include the relatively small additional load of the secondary regenerator.

3.1.2.3 Boiler Island Material and Energy Balance (Cases 1-5)

A simplified process flow diagram for the modified study unit boiler island is shown in Figure 3-4. This simplified diagram is applicable to each of the five cases included in this study. The operation and performance of the existing boiler and electrostatic precipitator (ESP) systems are identical to the Base Case for all five capture cases investigated and are not affected by the addition of the MEA-based CO_2 removal systems. The FGD system is modified for each of the five CO_2 removal cases with the addition of a secondary absorber to reduce the SO_2 content to less than 10 ppmv. The FGD system modification is described in Section 3.1.2.2.



Material Flow Stream Identification

1	Raw Coal to Pulverizers	9	FGD System Solids to Disposal	17	Mixed Primary Air to Pulverizers
2	Air Infiltration Stream	10	Flue Gas to Stack	18	Pulverized Coal and Air to Furnace
3	Flue Gas from Economizer to Air Heater	11	Air to Primary Air Fan	19	Secondary Air to Forced Draft Fan
4	Flue Gas Leaving Air Heater to ESP	12	Primary Air to Steam Coil Air Heater	20	Secondary Air to Steam Coil Air Heater
5	Flyash Leaving ESP	13	Primary Air to Air Heater	21	Secondary Air to Air Heater
6	Flue Gas Leaving ESP to Induced Draft Fan	14	Air Heater Leakage Air Stream	22	Heated Secondary Air to Furnace
7	Flue Gas to Flue Gas Desulfurization System	15	Tempering Air to Pulverizers	23	Bottom Ash from Furnace
8	Lime Feed to FGD System	16	Hot Primary Air to Pulverizers		

Figure 3-4: Simplified Boiler Island Gas Side Process Flow Diagram for CO_2 Separation by Monoethanolamine Absorption (Cases 1-5)

The overall material and energy balance for the boiler island system shown above in Figure 3-4 is provided in Table 3-19. The flue gases leaving the modified FGD system are ducted to the new MEA system where various levels (depending on the case in question) of the CO_2 is removed, compressed, and liquefied for usage or sequestration. The remaining flue gases leaving the new MEA system after removal of carbon dioxide (consisting of primarily oxygen, nitrogen, water vapor, and a relatively small amount of sulfur dioxide and carbon dioxide) are discharged to the atmosphere through the existing Unit 5/6 common stack.

Streams 24, 25, and 26 of Table 3-19 are purposely not filled in. These streams are dependent on the CO_2 recovery level and the attributes of these streams are defined in Section 3.1.4.1 for Cases 1-4 and Section 3.1.5.2 for Case 5.

Table 3-19: Gas Side Boiler Island Material and Material Energy Balance (Cases 1-5)

Constituent	(Units)	1	2	3	4	5 6		7	8	9	10	11	12	13
O_2	(lbm/hr)	26586	42147	101097	144817		144817	144817	5628		144566	203237	203237	112918
N_2	"	4868	139626	2797385	2942220		2942220	2942220			2942220	673283	673283	374075
H_2O	"	37820	2357	228849	231294		231294	231294	258954	48234	441924	11365	11365	6314
CO_2	"			867210	867210		867210	867210			8661102			
SO_2	"			20202	20202		20202	20202			87			
H_2	"	16102												
Carbon	"	236665												
Sulfur	"	10110												
Ca	"								13087					
Mg	"								613					
MgO	"									509				
$MgSO_3$	"									1251				
$MgSO_4$	"									76				
$CaSO_3$	"									34395				
$CaSO_4$	"									2051				
$CaCO_3$	"									2520				
Ash/Inerts	"	42313		33851	33851	33851			1017	1017				
		Raw Coal	Leakage Air	Flue Gas to AH	Flue Gas to ESP	Flyash	Flue Gas to ID Fan	Flue Gas to FGD	Lime Slurry	FGD Disposal	Flue Gas to CO_2 Sep	Pri Air to PA Fan	PA from PAA Fan	Pri Air to AH
Total Gas	(lbm/hr)		184130	4014743	4205743		4205743	4205743			4394900	887885	887885	493308
Total Solids	"	374455		33851	33851	33851			20346	41819				
Total Flow	"	374455	184130	4048594	4239594	33851	4205743	4205743	279300	90143	4394900	887885	887885	493308
Temperature	(Deg F)	80	80	706	311	311	311	325 80		136	136	80 92 9	2	
Pressure	(Psia)	14.7	14.7	14.6	14.3	14.7 14	2	15	14.7	14.7	14.7	14.7	15.6	15.6
h $_{sensible}$	(Btu/lbm)	0.000	0.000	161.831	57.924	57.750	57.924	61.384	0.000	14.116	14.543	0.000	2.899	2.899
Chemical	(10^6 Btu/hr)	4228.715												
Sensible	(10^6 Btu/hr)	0.000	0.000	655.007	245.567	1.955	243.612	258.166	0.000	3.314	63.916	0.000	2.574	1.430
Latent	(10^6 Btu/hr)	0.000	2.475	240.291	242.858	0.000	242.858	242.858	0.000	0.000	464.020	11.933	11.933	6.630
Total Energy[1]	(10^6 Btu/hr)	4228.715	2.475	895.298	488.425	1.955	486.470	501.024	0.000	3.314	527.936	11.933	14.507	8.060

Constituent	(Units)	14	15	16	17	18	19	20	21	22	23	24	25	26
O_2	(lbm/hr)	43720	90319	66680	156999	183585	641283	641283	641283	643801				
N_2	"	144835	299208	220899	520107	524975	2124443	2124443	2124443	2132785				
H_2O	"	2445	5051	3729	8779	46599	35860	35860	35860	36001				
CO_2	"													
SO_2	"													
H_2	"					16102								
Carbon	"					236655								
Sulfur	"					10110								
Ca	"													
Mg	"													
MgO	"													
$MgSO_3$	"													
$MgSO_4$	"													
$CaSO_3$	"													
$CaSO_4$	"													
$CaCO_3$	"													
Ash/Inerts	"					42313								
		Air Htr Lkg Air	Tempering Air	Hot Pri Air	Mixed Pri Air	Coal-Pri Air Mix	Sec Air to FD	Sec air to SCAH	Sec Air to AH	Hot Sec Air	Bottom Ash	CO_2 to Comp	CO_2 Product	Vent Stream
Total Gas	(lbm/hr)	191000	394577	291308	685885		2801587	280157	280157	281157				
Total Solids	"										8463			
Total Flow	"	191000	394577	291308	685885	1060340	2801587	280157	280157	2812587	8463			
Temperature	(Deg F)	92	92	666	339		80	86.4	86.4	616.1	2000			
Pressure	(Psia)	15.6	15.6	15.6	15.6	15.0	14.7	15.2	15.1	14.9	14.7			
h $_{sensible}$	(Btu/lbm)	2.899	2.899	145.249	63.358		0.000	1.549	1.549	132.582	480.000			
Chemical	(10^6 Btu/hr)					4228.715								
Sensible	(10^6 Btu/hr)	0.554	1.144	42.312	43.456		0.000	4.341	4.341	372.898	4.062			
Latent	(10^6 Btu/hr)	2.567	5.303	3.915	9.218		37.653	37.653	37.653	37.801	0.000			
Total Energy[1]	(10^6 Btu/hr)	3.121	6.447	46.227	52.674	4281.389	37.653	41.994	41.994	410.699	40.062			

(1) Energy Basis: Chemical based on Higher Heating Value (HHV); Sensible energy above 80°F; Latent based on 1050 Btu/lbm of water vapor.

3.1.3 Design and Performance of Advanced Amine CO_2 Removal Systems (Cases 1-4)

This section describes the advanced amine CO_2 Removal Systems used in this study. The amine technology used in this study is similar to existing advanced MEA amine processes. This process tolerates oxygen in the flue gas as well as a limited amount of sulfur dioxide. The process uses an oxygen-activated corrosion inhibitor, which also inhibits amine degradation. Low corrosion rates and minimal loss of the circulating solvent used to absorb CO_2 promotes economical and reliable operation. This study is based on the flue gases coming from the AEP's Conesville Unit #5 flue gas desulfurization system, shown later in this section.

ABB Lummus was responsible for the design, performance, and costs for the amine systems. The designs were based on information contained in the open literature (Bailey and Feron, 2005; Chapel and Mariz, 1999; Choi et al., 2005; Choi et al., 2004; Chinn et al., 2004; IEA, 2004) as well as their own proven experience (Barchas and Davis, 1992). The simulation tools used were ProTreat® Version 3.3 and Hysys® Version 2004.2. The resulting regeneration energy from this simulation was 1,550 Btu/lbm-CO_2.

There are four CO_2 capture cases using an advanced amine CO_2 removal systems investigated in this study. The four cases are described as follows:

- **Case 1:** 90% Capture
- **Case 2:** 70% Capture
- **Case 3:** 50% Capture
- **Case 4:** 30% Capture

An additional fifth case, also using the advanced amine system was originally planned to be evaluated in this study. This case was defined to be equivalent in CO_2 emissions to a NGCC plant without CO_2 capture, with CO_2 emissions of 362 g/kWh (0.799 lbm/kWh). Because Case 2 (70% CO_2 capture) of the current study was found to yield approximately this same amount of CO_2 emissions 354 g/kWh (0.78 1 lbm/kWh), the team decided not to evaluate this additional case.

The 90% recovery case (Case 1) processes the entire flue gas stream and adjusts the available process variables within the advanced MEA system to achieve 90% recovery in the absorber. The reduced recovery rates for Cases 2, 3, and 4 can be achieved by two methods. The 70%, 50%, and 30% recovery levels for Cases 2, 3, and 4 respectively are achieved by treating only part of the flue gas stream in the absorber and bypassing the remainder of the flue gas stream directly to the stack. The bypassing method allows the absorber and amine regeneration system to be smaller and less costly. The alternate method would involve treating the entire flue gas stream in the absorber and adjusting the available MEA process parameters to achieve a reduced recovery. This method was not chosen because it requires a larger absorber and a larger amine regeneration system, which was found to be significantly more costly than the selected flue gas bypass method.

3.1.4 Process Description - CO_2 Removal, Compression, and Liquefaction System (Cases 1-4)

The following process description applies to all the advanced amine cases in this study (i.e., Cases 1-4). The CO_2 Recovery Plant removes CO_2 from exhaust gas of the existing Conesville

#5 coal-fired steam boiler. The treated flue gas is returned to the existing stack. The captured CO_2 is compressed, dehydrated, and then liquefied in preparation for transport to a consumer.

Since the flue gas conditioning equipment flow scheme includes an existing blower, the pressure profile of the existing power generation equipment does not change from today's operation. To force the flue gas from the secondary flue gas desulfurizer (FGD) through the CO_2 Absorber, the pressure of the flue gas after the FGD is boosted ~0.1 bar (1.5 psi) by a motor driven fan. As the power consumption of the fan is considerable, the location of the absorbers is as close as possible to the new secondary FGD system and the existing stack, to minimize draft loss. The blower will run at constant speed. Each blower, provided as part of the boiler flue gas conditioning equipment, is equipped with its own suction and a discharge damper operated pneumatically. The suction damper controls the suction pressure to adjust for the flow variation resulting from the power plant performance. The suction pressure control will avoid any surges to blower. The discharge damper is an isolation damper.

Direct Contact Cooling

The following description refers to Figure 3-5. The direct contact cooler (DCC) Flue Gas Cooler is a packed column where hot 58°C (136°F) flue gas is brought into intimate contact with a recirculating stream of cool water. Physically the DCC and Absorber have been combined into a single compartmentalized tower. The lower compartment is designed to support the Absorber so that the top head of the DCC is the bottom head of the Absorber. Effectively, this dividing head acts as a chimney tray with a number of upward extending chimneys, which provide passages so the flue gas may flow directly from the DCC into the Absorber.

Theoretically, a direct contact cooler is capable of cooling the gas to a very close approach in a short bed. When the hot gas enters the DCC, it contains water but is highly superheated. At the bottom end of the bed, the gas quickly cools down to a temperature called the "Adiabatic Saturation Temperature" (AST). This is the temperature the gas reaches when some of its own heat content has been used to vaporize just the exact amount of water to saturate the gas.

Up to the point when the AST is reached, the mass flow of the gas stream increases due to evaporation of water. At the AST, water begins to condense as the gas is cooled further. As the gas travels up the column and is cooled further, more water is condensed. This internal refluxing increases the vapor/liquid (V/L) traffic at the bottom end of the bed significantly beyond the external flows and must be considered in the hydraulic design.

The water stream leaving the bottom of the DCC contains the water fed to the top as well as any water, which has condensed out of the flue gas. The condensed water may be somewhat corrosive due to sulfur and nitrogen oxides, which are present in the flue gas. Therefore, instead of using the condensate in the process, it will be blown down from the system. For the DCC to be effective, the temperature of the leaving water must always be lower than the AST.

The DCC Water Pump circulates most of the water leaving the bottom of the DCC back to the top of the direct contact cooler. However, before sending it back to the column, the water stream is first filtered in the DCC Water Filter and then cooled in DCC Water Cooler E-108. The temperature of the cooled water is controlled by a cascade loop, which maintains a constant flue gas exit temperature of 46°C (115°F).

Filtration is necessary to remove any particulate matter, which may enter the DCC in the flue gas. The blow down is taken out after the filter but before the cooler and mixed into the return water of cooler E-108. This way the cooler does not have to handle the extra duty, which would otherwise be imposed by the blow down.

Absorption

The following description refers to Figure 3-5.

CO_2 Absorber:

From the DCC the cooled flue gas enters the bottom of the CO_2 Absorber and flows up the tower counter-current through a stream of 30-wt% MEA solution. The lean MEA solution (LAM) enters the top of the column and heats up gradually as CO_2 is absorbed. By the time the stream leaves the bottom of the tower it has gained approximately 11°C (20°F). The tower has been designed to remove 90% of the CO_2 from the incoming gas. The CO_2 loading in LAM is approximately 0.19 mol CO_2/mol MEA, while the loading of the rich amine leaving the bottom is approximately 0.49 mol CO_2/mol MEA.

To maintain water balance in the process, the temperature of the LAM feed should be close to that of the feed gas stream. Thus, with feed gas temperature fixed at 46°C (115°F), the temperature of the LAM stream must also be close to 46°C (115°F), preferably within 5.5°C (10°F). If the feed gas comes in at a higher temperature than the LAM, it brings in excess moisture, which condenses in the Absorber and becomes excess water. Unless this water is purged from the system, the concentration of MEA will decrease and the performance of the system will suffer. If, on the other hand, the gas feed is colder than the LAM, it heats up in the tower and picks up extra moisture, which is then carried out of the system by the vent gas. The result is a water deficiency situation because more water is removed than comes into the system.

For the reasons explained above, it is essential that both the temperature of the flue gas and that of the LAM be accurately controlled. In fact, it is best to control one temperature and adjust the temperature of the other to maintain a fixed temperature difference.

The rich MEA solvent solution from the bottom of the absorber at 52°C (125°F) is heated to 96°C (205°F) by heat exchange with lean MEA solvent solution from the stripping column and then fed near the top of the stripping column. The lean MEA solvent solution is partially cooled by heat exchange with rich MEA and is further cooled to 4°C (105°F) by exchange with cooling water and fed back to the absorber to complete the circuit.

The CO_2 absorber contains two beds of structured packing and a “Wash Zone” at the very top of the column to reduce water and MEA losses. A liquid distributor is provided at the top of each bed of structured packing. There are several reasons for selecting structured packing for this service:

- Very low pressure drop which minimizes fan horsepower
- High contact efficiency / low packing height
- Good tolerance for mal-distribution in a large tower
- Smallest possible tower diameter
- Light weight

At the bottom of the tower, there is the equivalent of a chimney tray, which serves as the bottom sump for the absorber. Instead of being flat like a typical chimney tray, it is a standard dished head with chimneys. The hold-up volume of the bottom sump is sufficient to accept all the liquid held up in the packing both in the CO_2 absorber and in the Wash Zone. The Rich Solvent Pumps take suction from the chimney tray.

Absorber Wash Zone:

The purpose of the Wash Zone at the top of the tower is to minimize MEA losses both due to mechanical entrainment and also due to evaporation. This is achieved by recirculating wash water in this section to scrub most of the MEA from the lean gas exiting the Absorber. The key to minimizing MEA carryover is a mist separator pad between the wash section and the Absorber. The Wash Water Pump takes water from the bottom of the wash zone and circulates it back to the top of the wash zone.

The key to successful scrubbing is to maintain a low concentration of MEA in the circulating water. As MEA concentration is increased, the vapor pressure of MEA becomes higher and, consequently, the MEA losses are higher. Therefore, relatively clean water must be fed to the wash zone as make-up while an equal amount of MEA laden water is drawn out. A seal accomplishes this and maintains a level on the chimney tray at the bottom of the wash section. Overflow goes to the main absorber. Make-up water comes from the overhead system of the Solvent Stripper.

The lean flue gas leaving the wash zone is released to the existing flue gas stack at atmospheric pressure.

Rich/Lean Solvent Exchanger - E-100:

The Rich/Lean Solvent Exchanger is a plate type exchanger with rich MEA solution on one side and lean MEA solution on the other. The purpose of the exchanger is to recover as much heat as possible from the hot lean solvent at the bottom of the Solvent Stripper by heating the rich solvent feeding the Solvent Stripper. This reduces the duty of the Solvent Stripper Reboiler. This exchanger is the single most important item in the energy economy of the entire CO_2 Recovery Unit.

Lean Amine Cooler – E-104:

A plate frame water-cooled exchanger was added on the lean amine stream leaving the Rich/Lean Solvent Exchanger to reduce the plot space requirement and overall cost of the project. The lean amine cooler further cools the lean amine coming from the rich/lean exchanger E-100 from 66°C to 41°C (150°F to 105°F) with plant cooling water. Cooled amine from E-104 flows to the top of the absorber.

Stripping

Solvent Stripper:

The following description refers to Figure 3-5. The purpose of the Solvent Stripper is to separate CO_2 from the CO_2 rich solvent. The Solvent Stripper contains a top section with trays and a bottom section with structured packing. The top section of the stripper is a water wash zone designed to limit the amount of solvent (MEA) vapors entering the stripper overhead system.

The hot wet vapors from the top of the stripper contain the recovered CO_2, along with water vapor, and a limited amount of solvent vapor. The overhead vapors are cooled by water in the Solvent Stripper Condenser E-105, which is commonly called the reflux condenser, where most of the water and solvent vapors condense. The CO_2 does not condense. The condensed overhead liquid and CO_2 are separated in a reflux drum. CO_2 flows to the CO_2 Compression section on pressure control and the condensed liquid (called reflux) is returned to the top of the stripper. Rich solvent is fed to the stripper at the top of the packed section. As the solvent flows down over the packing to the bottom, hot vapor from the reboiler strips the CO_2 from the solution. The final stripping action occurs in the reboiler E-106.

Solvent Stripper Reboiler E-106:

The steam-heated reboiler consists of several plate frame thermo-siphon type exchangers arranged concentrically around the base of the Stripper. Circulating flow of the solvent through the reboiler is driven by gravity and density differences.

Solvent Reclaimer:

The Solvent Reclaimer is a horizontal heat exchanger. Certain acidic gases present in the flue gas feeding the CO_2 absorber form compounds with the MEA in the solvent solution, which cannot be regenerated by application of heat in the solvent stripper reboiler. These materials are referred to as “Heat Stable Salts” (HSS). A small slipstream of the lean solvent from the discharge of the Solvent Stripper Bottoms Pump is fed to the Solvent Reclaimer. The reclaimer restores the MEA usefulness by removing the high boiling and non-volatile impurities, such as HSS, suspended solids, acids, and iron products from the circulating solvent solution. Soda ash is added into the reclaimer to free MEA from its bond with sulfur oxides by its stronger basic attribute. This allows the MEA to be vaporized into the circulating mixture, minimizing MEA loss. This process is important in reducing corrosion, and fouling in the solvent system. The reclaimer bottoms are cooled intermittently with cooling tower water prior to being loaded on a tank truck.

Solvent Stripper Condenser E-105:

The solvent stripper condenser is a series water-cooled plate frame type heat exchangers. The purpose of the condenser is to completely condense all components contained in the overhead vapor stream leaving the stripper which are condensable under the operating conditions. Boiler feed water at 43°C (110°F) (integrated with the steam/water cycle) and 27°C (80°F) cooling tower water are used as the condensing medium. Components that do not condense include nitrogen, carbon dioxide, oxygen, nitrogen oxides and carbon monoxide. The water vapor and MEA solvent vapor will condense, and the condensed water will dissolve a small amount of carbon dioxide. This exchanger uses some of the cooling water capacity freed up due to the reduced load on the surface condensers of the existing Conesville #5 power plant.

Solvent Stripper Reflux Drum:

The reflux drum provides space and time for the separation of liquid and gases and provides liquid hold-up volume for suction to the reflux pumps.

Solvent Stripper Reflux Pump:

This pump takes suction from the reflux drum and discharges on flow control to the stripper top tray as reflux on flow control.

Semi-Lean Flash Drum:

Rich amine is pumped from the bottom of the absorber and is split into two streams. The first stream is heated in cross exchangers E-102 and E-100 with hot stripper bottoms and the preheated rich amine flowing to the stripper. The other part of the stream is flashed to produce steam, which is used in the stripping column. The Semi-Lean Flash Drum reduces the amount of steam needed in the reboiler. The rich amine prior to being flashed is heated in a pair of exchangers. The first is the semi-lean cooler E-101, where it is cross-exchanged with hot flashed semi-lean amine from the flash drum. The second is the flash preheater E-102, which is heated by hot stripper bottoms on its way to the amine cross exchanger.

Solvent Filtration Package:

The pre-coat filter is no ordinary filter; it is a small system. The main component is a pressure vessel, which has a number of so called “leaves” through which MEA flows. The leaves have a thin (~0.3 cm or 1/8 inch) coating of silica powder, which acts to filter any solids. For the purposes of such application the power is called “filter aid.”

To cover the leaves with the filter aid, the filter must be “pre-coated” before putting it into service. This is accomplished by mixing filter aid in water in a predetermined ratio (typically 10 wt%) to prepare slurry. This takes place in an agitated tank. A pump, which takes its suction from this tank, is then operated to pump the slurry into the filter. Provided the flow rate is high enough, the filter aid is deposited on the leaves while water passes through and can be recycled back to the tank. This is continued until the water in the tank becomes clear, indicating that all the filter aid has been transferred.

The volume of a single batch in the tank is typically 125% of the filter volume because there must be enough to fill the vessel and have some excess left over so the level in the tank is maintained and circulation can continue. In this design, water from the Stripper overhead is used as make-up water to fill the tank. This way the water balance of the plant is not affected.

During normal operation, it is often beneficial to add so called “body” which is the same material as the pre-coat but may be of different particle size. The body is also slurried in water but is continually added to the filter during operation. This keeps the filter coating porous and prevents rapid plugging and loss of capacity. As the description suggests, an agitated tank is needed to prepare the batch. A metering pump is then used to add the body at preset rate to the filter.

When the filter is exhausted (as indicated by pressure drop), it is taken off line so the dirty filter aid can be removed and replaced with fresh material. To accomplish this, the filter must be drained. This is done by pressurizing the filter vessel with nitrogen and pushing the MEA solution out of the filter. After this, the filter is depressurized. Then, a motor is started to rotate the leaves so a set of scrapers will wipe the filter cake off the leaves. The loosened cake then falls off and into a conveyor trough in the bottom of the vessel. This motor-operated conveyor then pushes the used cake out of the vessel and into a disposal container. The rejected cake has the consistency of toothpaste. This design is called “dry cake” filter and minimizes the amount of waste produced.

For this application, about 2% of the circulating MEA will be forced to flow through the filter. A Filter Circulating Pump draws the liquid through the filter. The advantage of placing the pump on the outlet side of the filter is reduced design pressure of the filter vessel and associated piping.

In spite of the restriction on its suction side, ample NPSH is still available for the pump. Flow is controlled downstream of the pump.

The MEA is also passed through a bed of activated carbon to reduce residual hydrocarbons. The presence of hydrocarbons in the amine can cause foaming problems. This study assumes that the bed is changed four times per year.

CO_2 Compression, Dehydration, and Liquefaction

The following description refers to Figure 3-6. CO_2 from the solvent stripper reflux drum, saturated with water, is compressed in a three-stage centrifugal compressor using 43°C (110°F) boiler feed water for interstage and after compression cooling. The heated boiler feedwater is returned to the existing feedwater system of the steam/water cycle, and this heat integration helps improve overall plant efficiency. The interstage coolers for first and second stage are designed to supply 52°C (125°F) CO_2 to the compressor suction.

Most of the water in the wet CO_2 stream is knocked out during compression and is removed from intermediate suction drums. A CO_2 dryer is located after the third stage to meet the water specifications in the CO_2 product. The water-free CO_2 is liquefied after the third stage of compression at about 13 bara (194 psia) by the use of a propane refrigeration system and is further pumped with a CO_2 pump to the required battery limit pressure of 139 bara (2,015 psia).

The propane refrigeration system requires centrifugal compressors, condensers, economizers, and evaporators to produce the required cold. The centrifugal compressor is driven by an electric motor and is used to raise the condensing temperature of the propane refrigerant above the temperature of the available cooling medium, which in this study is 110°F boiler feed water. The condenser is used to cool and condense the discharged propane vapor from the compressor back to its original liquid form. The economizer, which improves the refrigerant cycle efficiency, is designed to lower the temperature of the liquid propane by flashing or heat exchange. The evaporator liquefies the CO_2 vapor by transferring heat from the CO_2 vapor stream to the boiling propane refrigerant.

CO_2 Dryer

The following description refers to Figure 3-6. The purpose of the CO_2 dryer is to reduce the moisture content of the CO_2 product to a value less than pipeline transport specifications. The dryer package includes four dryer vessels loaded with Type 3A molecular sieve, three of which are in service while one is being regenerated or is on standby. The package also includes a natural gas fired regeneration heater and an air-cooled regeneration gas cooler. A water knockout, downstream from the gas cooler, removes the condensed water. The dryers are based on a 12-hour cycle.

The dryer is located on the discharge side of the third stage of the CO_2 Compressor. The temperature of the CO_2 stream entering the dryer is 125°F.

Once a bed is exhausted, it is taken off line and a slipstream of effluent from the on line beds is directed into this dryer after being boosted in pressure by a compressor. Before the slipstream enters the bed, which is to be regenerated, it is heated to a high temperature. Under this high temperature, moisture is released from the bed and carried away in the CO_2 stream. The regeneration gas is then cooled to the feed gas temperature to condense any excess moisture. After this, the regeneration gas stream is mixed with the feed gas upstream of the third stage knockout drum.

All the regeneration operations are controlled by a programmable logic controller (PLC), which switches the position of several valves to direct the flow to the proper dryer. It also controls the regeneration compressor, heater, and cooler.

Corrosion Inhibitor

Corrosion inhibitor chemical is injected into the process to help control the rate of corrosion throughout the CO_2 recovery plant system. The inhibitor is stored in a tank and is injected into the system via an injection pump (not shown in Figure 3-6). The pump is a diaphragm-metering type pump.

Process Flow Diagrams

The process flow diagram for the CO_2 recovery section is shown in Figure 3-5 and for the CO_2 compression, dehydration and liquefaction process is shown in Figure 3-6.

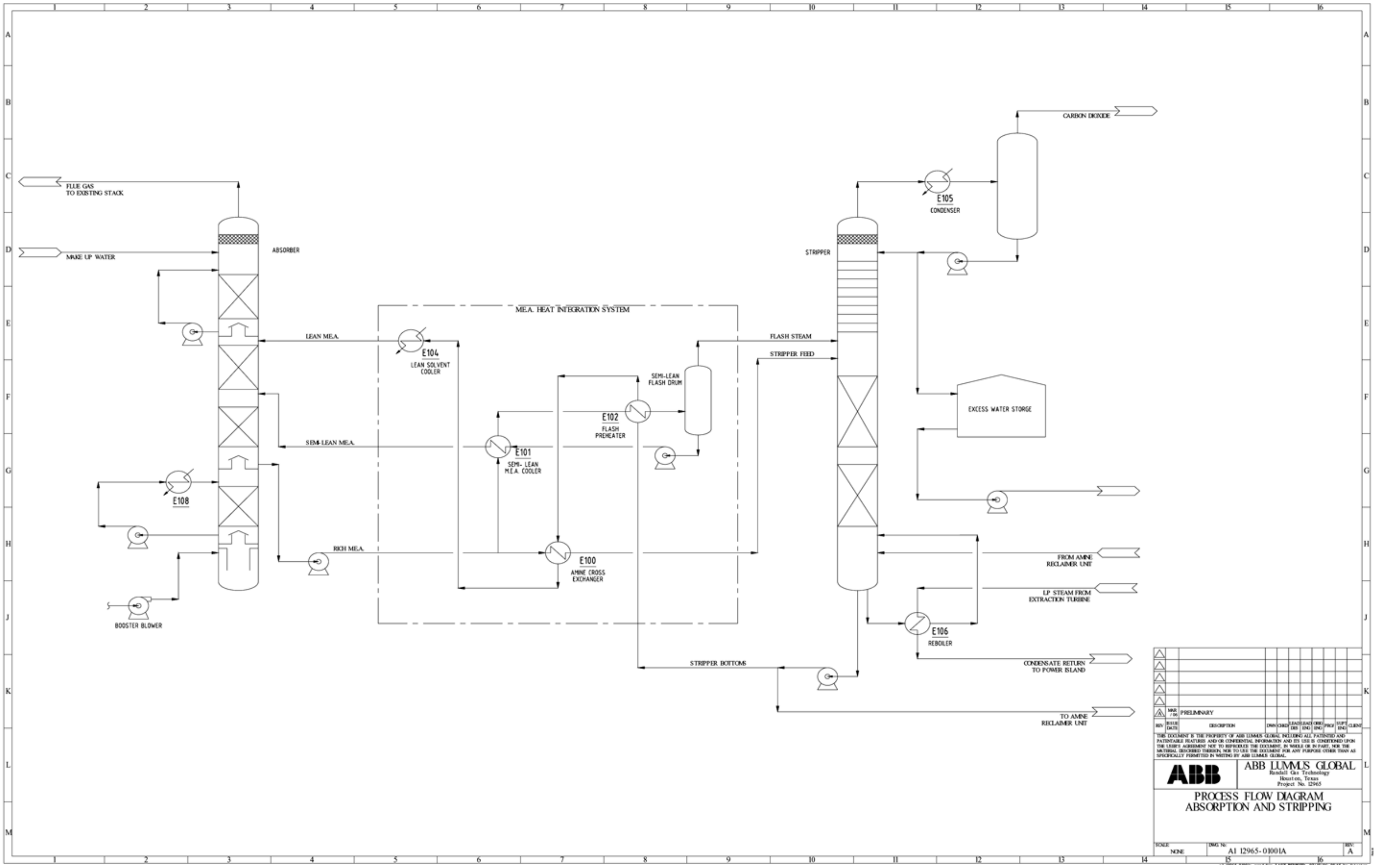


Figure 3-5: Advanced MEA Process Flow Diagram (Cases 1-4)

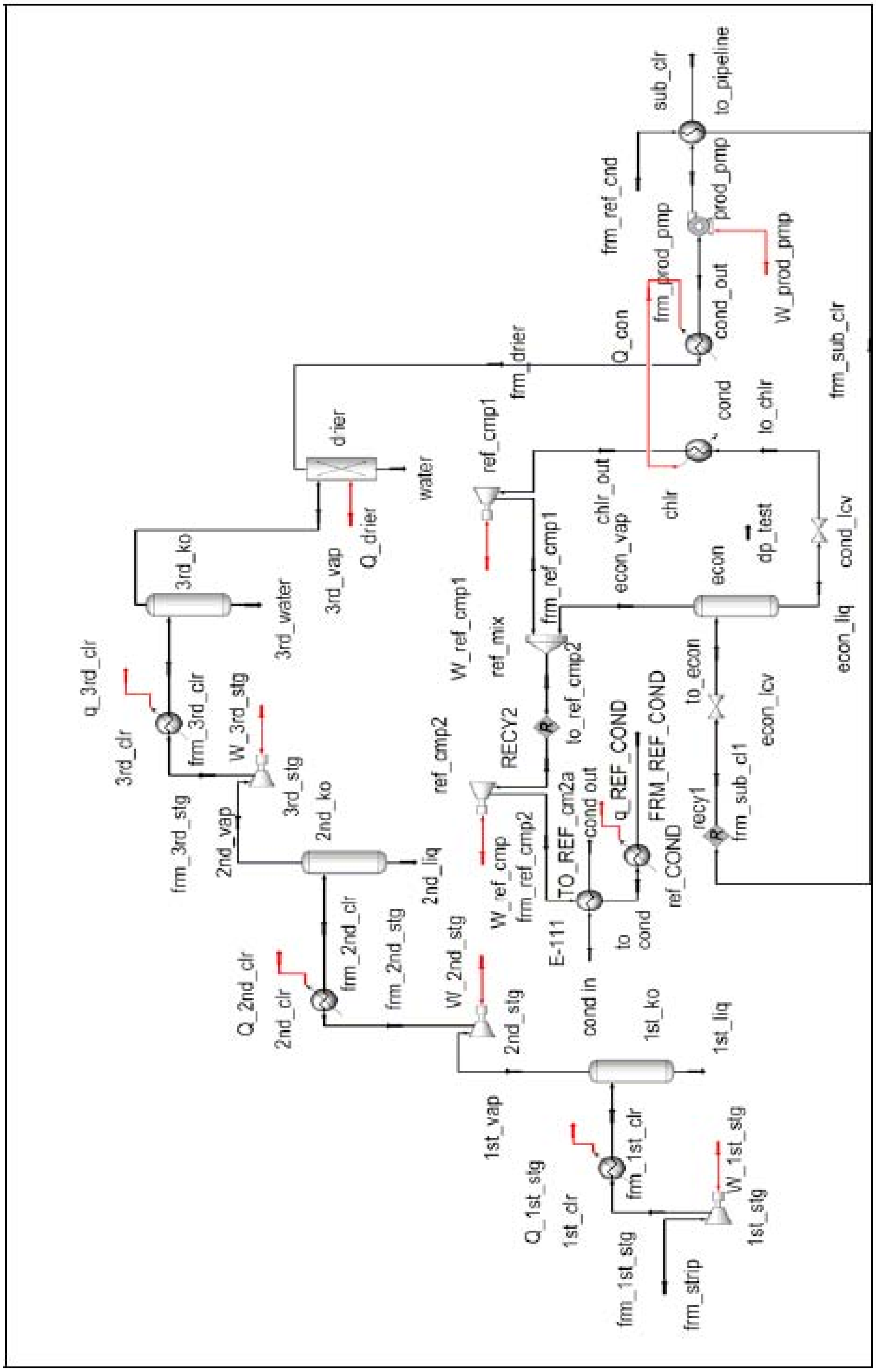


Figure 3-6: CO_2 Compression

3.1.4.1 Overall Material and Energy Balance - CO_2 Removal, Compression, and Liquefaction System (Cases 1-4)

This section provides material and energy balances for the CO_2 Removal and Compression Systems for Cases 1-4. Additionally, various other common parameters of comparison are provided for these systems.

Advanced Amine Plant Performance

Table 3-20 and Table 3-21 compare the amine plant material balance and energy demands, respectively, for each recovery case. The material balance shown in Table 3-20 is for the complete amine plant, as is Table 3-21. The CO_2 recovery cases below 90% (Cases 2, 3, and 4) are accomplished by combining the flue gas stream that bypasses the absorber, with the flue gas stream treated by the absorber, as shown in Figure 3-7. Even though the absorber and stripper recovery efficiencies are the same for each case, the net CO_2 recovery is lower due to the bypass.

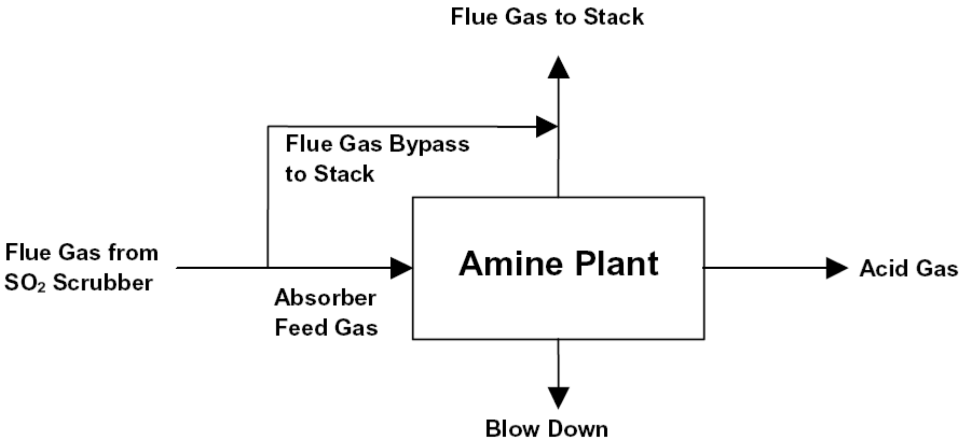


Figure 3-7: Flue Gas Bypass System used for 70%, 50%, and 30% CO_2 Absorption Cases (Cases 2, 3, and 4)

Table 3-20: Overall Material Balance for Amine Plants (Cases 1-4; 90%-30% CO_2 Capture)

Amine Plant	Case 1 (90% Capture)	Case 2 (70% Capture)	Case 3 (50% Capture)	Case 4 (30% Capture)
Feed to Absorber	moles/hr	moles/hr moles/hr moles/hr		
CO_2	19680	15306	10934	6560
H_2O	24530	19078	13628	8176
N_2	105020	81682	58344	35006
O_2	4518	3514	2510	1506
Total	153746	119582	85416	51248
From Top of Absorber	moles/hr	moles/hr moles/hr moles/hr		
CO_2	1962	1552	1102	650
H_2O	36460	28354	20252	12150
N_2	105016	81678	58342	35004
O_2	4518	3514	2510	1506
Total	147954	115098	82204	49312
Absorber Bypass*	moles/hr	moles/hr moles/hr moles/hr		
CO_2	0	4374	8746	13120
H_2O	0	5452	10902	16354
N_2	0	23338	46676	70014
O_2	0	1004	2008	3012
Total	0	34166	68330	102498
To Stack	moles/hr	moles/hr moles/hr moles/hr		
CO_2	1962	5924	9846	13770
H_2O	36460	33806	31154	28504
N_2	105016	105016	105018	105018
O_2	4518	4518	4518	4518
Total	147954	149264	150536	151810
Acid Gas	moles/hr	moles/hr moles/hr moles/hr		
CO_2	17720	13766	9822	5906
H_2O	1042	810	578 348	
N_2	0	0 0 0		
O_2	0	0 0 0		
Total	18762	14576	10400	6252
	moles/hr	moles/hr moles/hr moles/hr		
H_2O Blow Down	10714	8284	5860	3468

Note: "Bypass" method used to capture <90% CO_2

Table 3-21: Energy and Process Demands (Cases 1-4; 90%-30% CO_2 Capture)

Total Plant	Case 1 (90% Capture)	Case 2 (70% Capture)	Case 3 (50% Capture)	Case 4 (30% Capture)
CO_2 Captured, Metric TPD	8,481	6,595	4,706	2,829
CO_2 Captured, Short TPD	9,349	7,270	5,187	3,119
CO_2 captured, 10^6-scfd	161.2	125.4	89.5	53.8
H_2O Makeup to Amine Plant, gpm	427	331	235	140
H_2O Makeup to Cooling Tower, gpm	2,091	1,627	1,161	690
MEA Concentration, wt%	30.0%	30.0%	30.0%	30.0%
CO_2 Absorbed in the Absorber, %	90.0%	89.9%	89.8%	90.0%
Stripper Energy, Btu/lbm-CO_2 Absorbed	1,548	1,548	1,551	1,549
Solvent requirement, Gal MEA/lbm CO_2 Absorbed	2.042	2.044	2.047	2.042
Steam requirement, lbm/lbm CO_2 Absorbed	1.667	1.669	1.669	1.667
Lean Load, Mole CO_2/Mole MEA	0.188	0.190	0.190	0.186
Absorber Diameter, Ft	34.1	30.0	25.4	27.8
Stripper Diameter, Ft	22.0	19.3	16.3	17.9
Steam to Stripper, 10^3-lbm/h	1,300	1,010	722	433
Cooling Water (CW), gpm	69,694	54,217	38,693	22,991
Auxiliary power, Total kW Demand	54,939	42,697	30,466	18,247
Auxiliary power, kW w/o CO_2 Compression	11,802	9,169	6,549	3,866
Auxiliary power, kWh/Short Ton (ST) CO_2	141	141	141	140
Auxiliary power, kWh/ST CO_2 w/o CO_2 Compression	30	30	30	30
Cooling Water, Gallons/ST CO_2	10,735	10,739	10,742	10,615
Cooling Water, Cubic Meters/Metric Ton CO_2	46	46	46	45

CO_2 Compression and Liquefaction Plant Performance

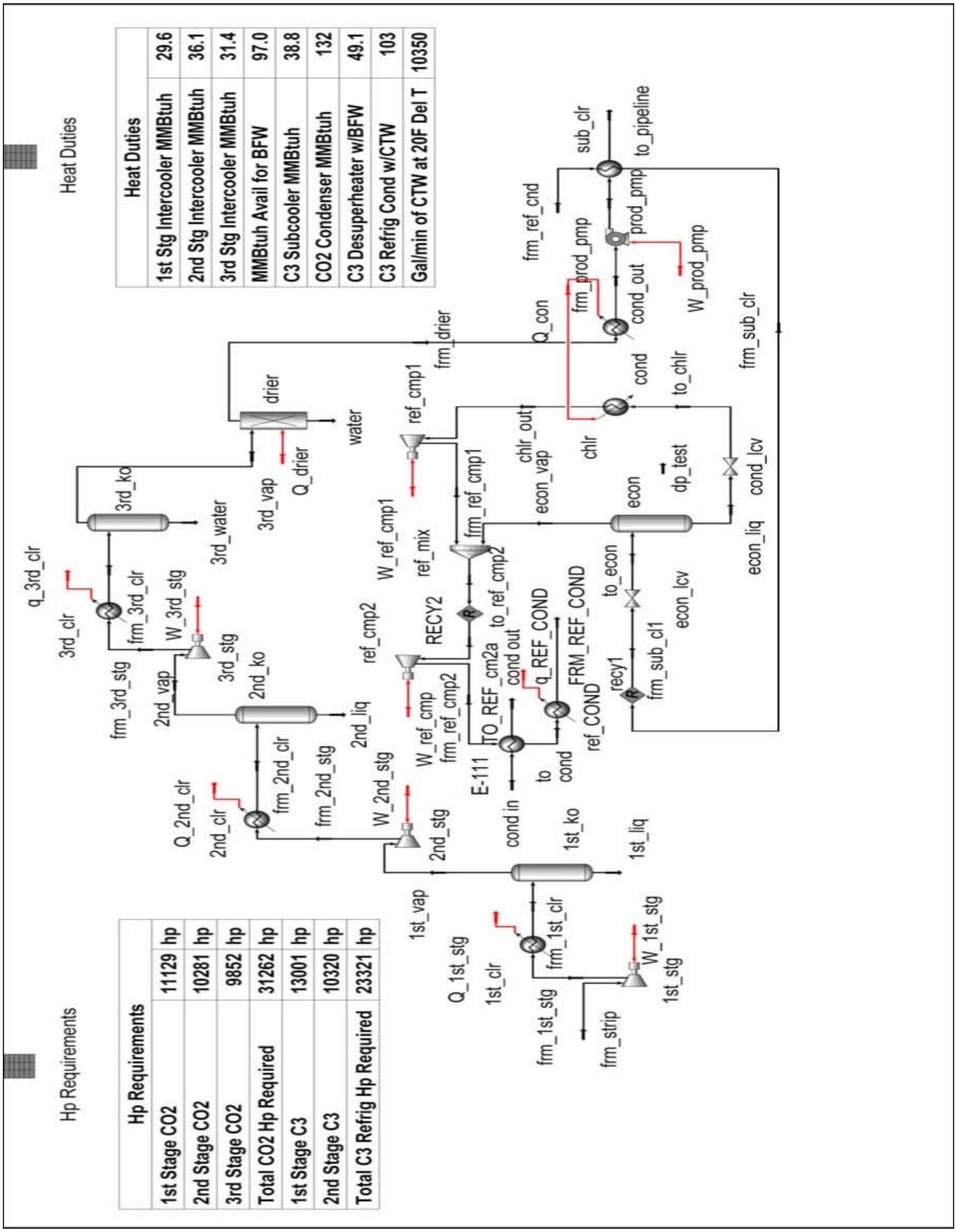
This section provides system schematics, material and energy balances, as well as heat duties and power requirements for the Compression and Liquefaction systems for Cases 1-4.

Table 3-22 shows the CO_2 compression and liquefaction system material and energy balance for Case 1 with 90% CO_2 recovery. Figure 3-8 shows the compression and liquefaction system schematic with heat duties and power requirements indicated.

Table 3-23 shows the CO_2 compression and liquefaction system material and energy balance for Case 2 with 70% CO_2 recovery. Figure 3-9 shows the compression and liquefaction system schematic with heat duties and power requirements indicated.

Table 3-24 shows the CO_2 compression and liquefaction system material and energy balance for Case 3 with 50% CO_2 recovery. Figure 3-10 shows the compression and liquefaction system schematic with heat duties and power requirements indicated.

Table 3-25 shows the CO_2 compression and liquefaction system material and energy balance for Case 4 with 30% CO_2 recovery. Figure 3-11 shows the compression and liquefaction system schematic with heat duties and power requirements indicated.



Heat Duties	
1st Stg Intercooler MMBtuh	29.6
2nd Stg Intercooler MMBtuh	36.1
3rd Stg Intercooler MMBtuh	31.4
MMBtuh Avail for BFW	97.0
C3 Subcooler MMBtuh	38.8
CO2 Condenser MMBtuh	132
C3 Desuperheater w/BFW	49.1
C3 Refrig Cond w/CTW	103
Gal/min of CTW at 20F Del T	10350

Hp Requirements		
1st Stage CO2	11129	hp
2nd Stage CO2	10281	hp
3rd Stage CO2	9852	hp
Total CO2 Hp Required	31262	hp
1st Stage C3	13001	hp
2nd Stage C3	10320	hp
Total C3 Refrig Hp Required	23321	hp

Figure 3-8: Case 1 CO_2 Compression, Dehydration, and Liquefaction Schematic (90% CO_2 Recovery)

Table 3-22: Case 1 Material & Energy Balance for CO_2 Compression, Dehydration, and Liquefaction (90% CO_2 Recovery)

STREAM NAME		Total Acid Gas from Strippers	First Stage Discharge	To Second Stage	First Stage Water KO	From Second Stage	2nd Stage Discharge	To 3rd Stage	2nd Stage Water KO	From 3rd Stage	From 3rd Stage Cooler	3rd Stage Water KO	To Drier	Water From Drier	From Drier To Condenser	Condensed CO2 Product	Frm C3 Desuperheater E111	Refrig Compressor Discharge	Section of 2nd Refrig Compressor	Discharge from 1st Refrig Compr	Refrig from CO2 Condenser
PFD STREAM NO.		frm_strip	frm_1st_clr	1st_vap	1st_liq	rm_2nd_stg	frm_2nd_clr	2nd_vap	2nd_liq	rm_3rd_stg	frm_3rd_clr	3rd_water	3rd_vap	water	frm_drier	cond out	to cond	m_ref_cmp	o_ref_cmp2	m_ref_cmp	chlr_out
VAPOR FRACTION	Molar	1.000	0.993	1.000	0.000	1.000	0.974	1.000	0.000	1.000	0.988	0.000	1.000	0.000	1.000	0.000	1.000	1.000	1.000	1.000	1.000
TEMPERATURE	°F	115.0	125	125	125	275	125	125	125	275	125	125	125	125	125	240	125	264	174	174	56
PRESSURE	PSIA	19.0	41	41	41	95	89	89	89	206	200	200	200	195	195	95	199	234	85	85	20
MOLAR FLOW RATE	lbmol/hr	18,762	18,762	18,625	136.50	18,625	18,625	18,142.03	483	18,142	18,142.03	218.92	17,923.11	205.61	17,717.51	20,119.70	16,100.00	16,100.00	16,100.00	16,100.00	16,100.00
MASS FLOW RATE	lb/hr	798,595.3	798,595.3	796,133.6	2,461.7	796,133.6	796,133.6	787,411.1	8,722.4	787,411.1	787,411.1	3,965.2	783,445.9	3,704.0	779,741.9	362,458.5	709,961.7	709,961.7	709,961.7	709,961.7	709,961.7
ENERGY	Btu/hr	-3.10E+09	-3.10E+09	-3.09E+09	-1.66E+07	-3.06E+09	-3.10E+09	-3.04E+09	-5.89E+07	-3.01E+09	-3.04E+09	-2.67E+07	-3.02E+09	-2.50E+07	-3.00E+09	-2.41E+09	-7.16E+08	-6.67E+08	-6.94E+08	-6.94E+08	-7.27E+08
COMPOSITON	Mol %																				
CO2		94.44%	94.44%	95.14%	0.08%	95.14%	95.14%	97.66%	0.17%	97.66%	97.66%	0.38%	98.85%	0.00%	100.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
H2O		5.56%	5.56%	4.86%	99.92%	4.86%	4.86%	2.34%	99.83%	2.34%	2.34%	99.62%	1.15%	100.00%	0.00%	100.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Nitrogen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	98.00%	98.00%	98.00%	98.00%	98.00%
Oxygen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Ethane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	1.00%	1.00%	1.00%	1.00%	1.00%
i-Butane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.50%	0.50%	0.50%	0.50%	0.50%
n-Butane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.50%	0.50%	0.50%	0.50%	0.50%
VAPOR																					
MOLAR FLOW RATE	lbmol/hr	18,762	18,625	18,625	0	18,625	18,142	18,142	0	18,142	17,923.1	-	17,923.1	-	17,717.5	-	16,100.0	16,100.0	16,100.0	16,100.0	16,100.0
MASS FLOW RATE	lb/hr	798,595.3	796,133.6	796,133.6	-	796,133.6	787,411.1	787,411.1	-	787,411.1	783,445.9	-	783,445.9	-	779,741.9	-	709,961.7	709,961.7	709,961.7	709,961.7	709,961.7
STD VOL. FLOW	MMSCFD	170.88	169.64	169.64	-	169.64	165.24	165.24	-	165.24	163.24	-	163.24	-	161.37	-	146.64	146.64	146.64	146.64	146.64
ACTUAL VOL. FLOW	ACFM	100,846	46,889	46,889	0	25,393	20,739	20,739	0	11,239	8,805.74	-	8,805.74	-	8,954.83	-	6,724.32	7,859.53	20,119.63	20,119.63	72,387.86
MOLECULAR WEIGHT	MW	42.57	42.75	42.75	-	42.75	43.40	43.40	-	43.40	43.71	-	43.71	-	44.01	-	44.10	44.10	44.10	44.10	44.10
DENSITY	lb/ft³	0.13	0.28	0.28	-	0.52	0.63	0.63	-	1.17	1.48	-	1.48	-	1.45	-	1.76	1.51	0.59	0.59	0.16
VISCOSITY	cP	0.0151	0.0155	0.0155	-	0.0201	0.0160	0.0160	-	0.0207	0.0164	-	0.0164	-	0.0165	-	0.0097	0.0118	0.0099	0.0099	0.0079
LIGHT LIQUID																					
MOLAR FLOW RATE	lbmol/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
DENSITY	lb/ft³	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MOLECULAR WEIGHT	MW	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
HEAVY LIQUID																					
MOLAR FLOW RATE	lbmol/hr	-	136.50	-	136.50	-	483.00	-	483.00	-	218.92	218.92	-	205.61	-	20,119.70	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	2,461.7	-	2,461.7	-	8,722.4	-	8,722.4	-	3,965.2	3,965.2	-	3,704.0	-	362,458.5	-	-	-	-	-
STD VOL. FLOW	BPD	-	169	-	169	-	599	-	599	-	273	273	-	254	-	24,869	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	4.98	-	4.98	-	17.64	-	17.64	-	8.01	8.01	-	7.49	-	774.31	-	-	-	-	-
DENSITY	lb/ft³	-	61.63	-	61.63	-	61.66	-	61.66	-	61.72	61.72	-	61.64	-	58.36	-	-	-	-	-
VISCOSITY	cP	-	0.5291	-	0.5291	-	0.5651	-	0.5651	-	0.5621	0.5621	-	0.5291	-	0.2394	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	67.39	-	67.39	-	67.33	-	67.33	-	67.19	67.19	-	67.44	-	55.61	-	-	-	-	-

90	7/17/2006	LEG	
No.	**Date**	**By**	**REVISION**

NOTES:

Alstom Power
AEP Unit 5, Conesville, OH
90% CO2 Recovery
Heat & Material Balance
90%_7T_R2CTW80F
JOB NO: LR12965 REV. A

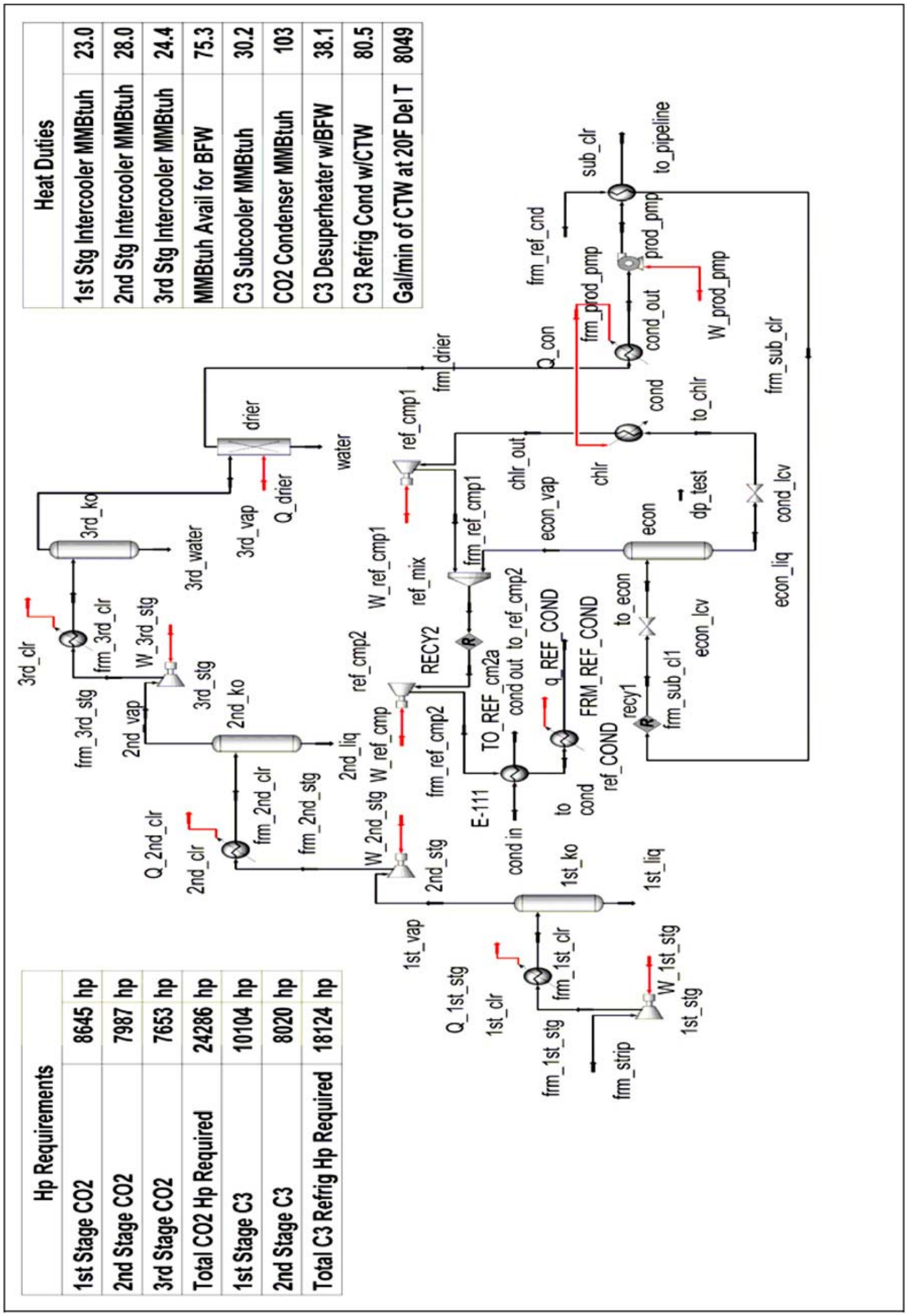
Table 3-22: Case 1 Material & Energy Balance for CO_2 Compression, Dehydration, and Liquefaction (90% CO_2 Recovery), continued

STREAM NAME		Vapor from Economizer	Refrig to CO2 Condenser	Economizer Liquid	To Economizer	From Subcooler	From Refrig Condenser	From Product Pump	CO2 To Pipeline
PFD STREAM NO.		econ_vap	to_chlr	econ_liq	to_econ	frm_sub_cl	frm_ref_cnd	frm_prod_pm	to_pipeline
VAPOR FRACTION	Molar	#DIV/0!	0.148	0.000	0.000	0.000	0.000	0.000	0.000
TEMPERATURE	°F	15	-32	15	15	15	100	-10	82
PRESSURE	PSIA	85	20	85	85	189	192	2,018	2,015
MOLAR FLOW RATE	lbmol/hr	-	16,100.00	16,100.00	16,100.00	16,100.00	16,100.00	17,717.51	17,717.51
MASS FLOW RATE	lb/hr	-	709,961.7	709,961.7	709,961.7	709,961.7	709,961.7	779,741.9	779,741.9
ENERGY	Btu/hr	0.00E+00	-8.59E+08	-8.59E+08	-8.59E+08	-8.59E+08	-8.20E+08	-3.12E+09	-3.08E+09
COMPOSITON	Mol %								
CO2		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	100.00%	100.00%
H2O		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Nitrogen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		95.58%	98.00%	98.00%	98.00%	98.00%	98.00%	0.00%	0.00%
Oxygen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Ethane		4.12%	1.00%	1.00%	1.00%	1.00%	1.00%	0.00%	0.00%
i-Butane		0.18%	0.50%	0.50%	0.50%	0.50%	0.50%	0.00%	0.00%
n-Butane		0.12%	0.50%	0.50%	0.50%	0.50%	0.50%	0.00%	0.00%
VAPOR									
MOLAR FLOW RATE	lbmol/hr	-	2,387.2	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	104,213.8	-	-	-	-	-	-
STD VOL. FLOW	MMSCFD	-	21.74	-	-	-	-	-	-
ACTUAL VOL. FLOW	ACFM	-	8,754.51	-	-	-	-	-	-
MOLECULAR WEIGHT	MW	43.56	43.66	-	-	-	-	-	-
DENSITY	lb/ft³	0.85	0.20	-	-	-	-	-	-
VISCOSITY	cP	0.0075	0.0065	-	-	-	-	-	-
LIGHT LIQUID									
MOLAR FLOW RATE	lbmol/hr	-	13,712.84	16,100.00	16,100.00	16,100.00	16,100.00	17,717.51	17,717.51
MASS FLOW RATE	lb/hr	-	605,747.9	709,961.7	709,961.7	709,961.7	709,961.7	779,741.9	779,741.9
STD VOL. FLOW	BPD	-	81,859	96,077	96,077	96,077	96,077	64,690	64,690
ACTUAL VOL. FLOW	GPM	-	2,107.66	2,617.99	2,617.99	2,609.19	3,007.65	1,416.98	1,914.60
DENSITY	lb/ft³	-	35.83	33.81	33.81	33.92	29.43	68.61	50.78
MOLECULAR WEIGHT	MW	-	44.17	44.10	44.10	44.10	44.10	44.01	44.01
VISCOSITY	cP	-	0.1841	0.1396	0.1396	0.1400	0.0881	0.1593	0.0622
SURFACE TENSION	Dyne/Cm	-	14.56	11.08	11.08	11.10	5.42	13.90	0.86
HEAVY LIQUID									
MOLAR FLOW RATE	lbmol/hr	-	-	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	-	-	-
DENSITY	lb/ft³	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	-	-	-

No.	Date	By	REVISION
90	7/17/2006	LEG	

ABB

Alstom Power
AEP Unit 5, Conesville, OH
90% CO2 Recovery
Heat & Material Balance
90%_7T_R2CTW80F
JOB NO: LR12965
REV. A



Heat Duties	
1st Stg Intercooler MMBtuh	23.0
2nd Stg Intercooler MMBtuh	28.0
3rd Stg Intercooler MMBtuh	24.4
MMBtuh Avail for BFW	75.3
C3 Subcooler MMBtuh	30.2
CO2 Condenser MMBtuh	103
C3 Desuperheater w/BFW	38.1
C3 Refrig Cond w/CTW	80.5
Gal/min of CTW at 20F Del T	8049

Hp Requirements	
1st Stage CO2	8645 hp
2nd Stage CO2	7987 hp
3rd Stage CO2	7653 hp
Total CO2 Hp Required	24286 hp
1st Stage C3	10104 hp
2nd Stage C3	8020 hp
Total C3 Refrig Hp Required	18124 hp

Figure 3-9: Case 2 CO_2 Compression, Dehydration, and Liquefaction Schematic
(70% CO_2 Recovery)

Table 3-23: Case 2 Material and Energy Balance for CO_2 Compression, Dehydration, and Liquefaction (70% CO_2 Recovery)

STREAM NAME		Total Acid Gas from Strippers	First Stage Discharge	To Second Stage	First Stage Water KO	From Second Stage	2nd Stage Discharge	To 3rd Stage	2nd Stage Water KO	From 3rd Stage	From 3rd Stage Cooler
PFD STREAM NO.		frm_strip	frm_1st_clr	1st_vap	1st_liq	rm_2nd_stg	frm_2nd_clr	2nd_vap	2nd_liq	frm_3rd_stg	frm_3rd_clr
VAPOR FRACTION	Molar	1.000	0.993	1.000	0.000	1.000	0.974	1.000	0.000	1.000	0.988
TEMPERATURE	°F	115.0	125	125	125	275	125	125	125	275	125
PRESSURE	PSIA	19.0	41	41	41	95	89	89	89	206	200
MOLAR FLOW RATE	lbmol/hr	14,575	14,575	14,469	106.04	14,469	14,469	14,093.74	375	14,094	14,093.74
MASS FLOW RATE	lb/hr	620,393.5	620,393.5	618,481.1	1,912.4	618,481.1	618,481.1	611,705.0	6,776.1	611,705.0	611,705.0
ENERGY	Btu/hr	-2.41E+09	-2.41E+09	-2.40E+09	-1.29E+07	-2.38E+09	-2.41E+09	-2.36E+09	-4.57E+07	-2.34E+09	-2.36E+09
COMPOSITON	Mol %										
CO2		94.44%	94.44%	95.14%	0.08%	95.14%	95.14%	97.66%	0.17%	97.66%	97.66%
H2O		5.56%	5.56%	4.86%	99.92%	4.86%	4.86%	2.34%	99.83%	2.34%	2.34%
Nitrogen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Oxygen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Ethane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
i-Butane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
n-Butane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
VAPOR											
MOLAR FLOW RATE	lbmol/hr	14,575	14,469	14,469	0	14,469	14,094	14,094	0	14,094	13,923.7
MASS FLOW RATE	lb/hr	620,393.5	618,481.1	618,481.1	-	618,481.1	611,705.0	611,705.0	-	611,705.0	608,624.6
STD VOL. FLOW	MMSCFD	132.75	131.78	131.78	-	131.78	128.37	128.37	-	128.37	126.82
ACTUAL VOL. FLOW	ACFM	78,342	36,426	36,426	0	19,726	16,112	16,112	0	8,731	6,840.79
MOLECULAR WEIGHT	MW	42.57	42.75	42.75	-	42.75	43.40	43.40	-	43.40	43.71
DENSITY	lb/ft^3	0.13	0.28	0.28	-	0.52	0.63	0.63	-	1.17	1.48
VISCOSITY	cP	0.0151	0.0155	0.0155	-	0.0201	0.0160	0.0160	-	0.0207	0.0164
LIGHT LIQUID											
MOLAR FLOW RATE	lbmol/hr	-	-	-	-	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	-	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	-	-	-	-	-
DENSITY	lb/ft^3	-	-	-	-	-	-	-	-	-	-
MOLECULAR WEIGHT	MW	-	-	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	-	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	-	-	-	-	-
HEAVY LIQUID											
MOLAR FLOW RATE	lbmol/hr	-	106.04	-	106.04	-	375.22	-	375.22	-	170.07
MASS FLOW RATE	lb/hr	-	1,912.4	-	1,912.4	-	6,776.1	-	6,776.1	-	3,080.4
STD VOL. FLOW	BPD	-	131	-	131	-	465	-	465	-	212
ACTUAL VOL. FLOW	GPM	-	3.87	-	3.87	-	13.70	-	13.70	-	6.22
DENSITY	lb/ft^3	-	61.63	-	61.63	-	61.66	-	61.66	-	61.72
VISCOSITY	cP	-	0.5291	-	0.5291	-	0.5651	-	0.5651	-	0.5621
SURFACE TENSION	Dyne/Cm	-	67.39	-	67.39	-	67.33	-	67.33	-	67.19

STREAM NAME		3rd Stage Water KO	To Drier	Water From Drier	From Drier To Condenser	Condensed CO2 Product	Frm C3 Desuperheater E111	Refrig Compressor Discharge	Section of 2nd Refrig Compressor	Discharge from 1st Refrig Compr	Refrig from CO2 Condenser
PFD STREAM NO.		3rd_water	3rd_vap	water	frm_drier	cond out	to cond	rm_ref_cmp	o_ref_cmp2	m_ref_cmp	chlr_out
VAPOR FRACTION	Molar	0.000	1.000	0.000	1.000	0.000	1.000	1.000	1.000	1.000	1.000
TEMPERATURE	°F	125	125	125	125	205	125	264	173	173	56
PRESSURE	PSIA	200	200	195	195	95	199	234	85	85	20
MOLAR FLOW RATE	lbmol/hr	170.07	13,923.68	159.73	13,763.95	21,526.53	12,522.00	12,522.00	12,522.00	12,522.00	12,522.00
MASS FLOW RATE	lb/hr	3,080.4	608,624.6	2,877.5	605,747.1	387,802.7	552,182.6	552,182.6	552,182.6	552,182.6	552,182.6
ENERGY	Btu/hr	-2.07E+07	-2.34E+09	-1.95E+07	-2.33E+09	-2.59E+09	-5.57E+08	-5.19E+08	-5.40E+08	-5.40E+08	-5.65E+08
COMPOSITON	Mol %										
CO2		0.38%	98.85%	0.00%	100.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
H2O		99.62%	1.15%	100.00%	0.00%	100.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Nitrogen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%	0.00%	0.00%	98.00%	98.00%	98.00%	98.00%	98.00%
Oxygen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Ethane		0.00%	0.00%	0.00%	0.00%	0.00%	1.00%	1.00%	1.00%	1.00%	1.00%
i-Butane		0.00%	0.00%	0.00%	0.00%	0.00%	0.50%	0.50%	0.50%	0.50%	0.50%
n-Butane		0.00%	0.00%	0.00%	0.00%	0.00%	0.50%	0.50%	0.50%	0.50%	0.50%
VAPOR											
MOLAR FLOW RATE	lbmol/hr	-	13,923.7	-	13,764.0	-	12,522.0	12,522.0	12,522.0	12,522.0	12,522.0
MASS FLOW RATE	lb/hr	-	608,624.6	-	605,747.1	-	552,182.6	552,182.6	552,182.6	552,182.6	552,182.6
STD VOL. FLOW	MMSCFD	-	126.82	-	125.36	-	114.05	114.05	114.05	114.05	114.05
ACTUAL VOL. FLOW	ACFM	-	6,840.79	-	6,956.61	-	5,229.93	6,107.89	15,636.40	15,636.40	56,255.60
MOLECULAR WEIGHT	MW	-	43.71	-	44.01	-	44.10	44.10	44.10	44.10	44.10
DENSITY	lb/ft^3	-	1.48	-	1.45	-	1.76	1.51	0.59	0.59	0.16
VISCOSITY	cP	-	0.0164	-	0.0165	-	0.0097	0.0118	0.0099	0.0099	0.0079
LIGHT LIQUID											
MOLAR FLOW RATE	lbmol/hr	-	-	-	-	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	-	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	-	-	-	-	-
DENSITY	lb/ft^3	-	-	-	-	-	-	-	-	-	-
MOLECULAR WEIGHT	MW	-	-	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	-	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	-	-	-	-	-
HEAVY LIQUID											
MOLAR FLOW RATE	lbmol/hr	170.07	-	159.73	-	21,526.53	-	-	-	-	-
MASS FLOW RATE	lb/hr	3,080.4	-	2,877.5	-	387,802.7	-	-	-	-	-
STD VOL. FLOW	BPD	212	-	197	-	26,608	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	6.22	-	5.82	-	813.90	-	-	-	-	-
DENSITY	lb/ft^3	61.72	-	61.63	-	59.40	-	-	-	-	-
VISCOSITY	cP	0.5621	-	0.5282	-	0.2915	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	67.19	-	67.42	-	59.38	-	-	-	-	-

90	7/17/2006	LEG	
No.	**Date**	**By**	**REVISION**

NOTES:

ABB

Alstom Power

AEP Unit 5, Conesville, OH

70% CO2 Recovery

Heat & Material Balance

70%_5T_R2CTW80

JOB NO: LR12965

REV. A

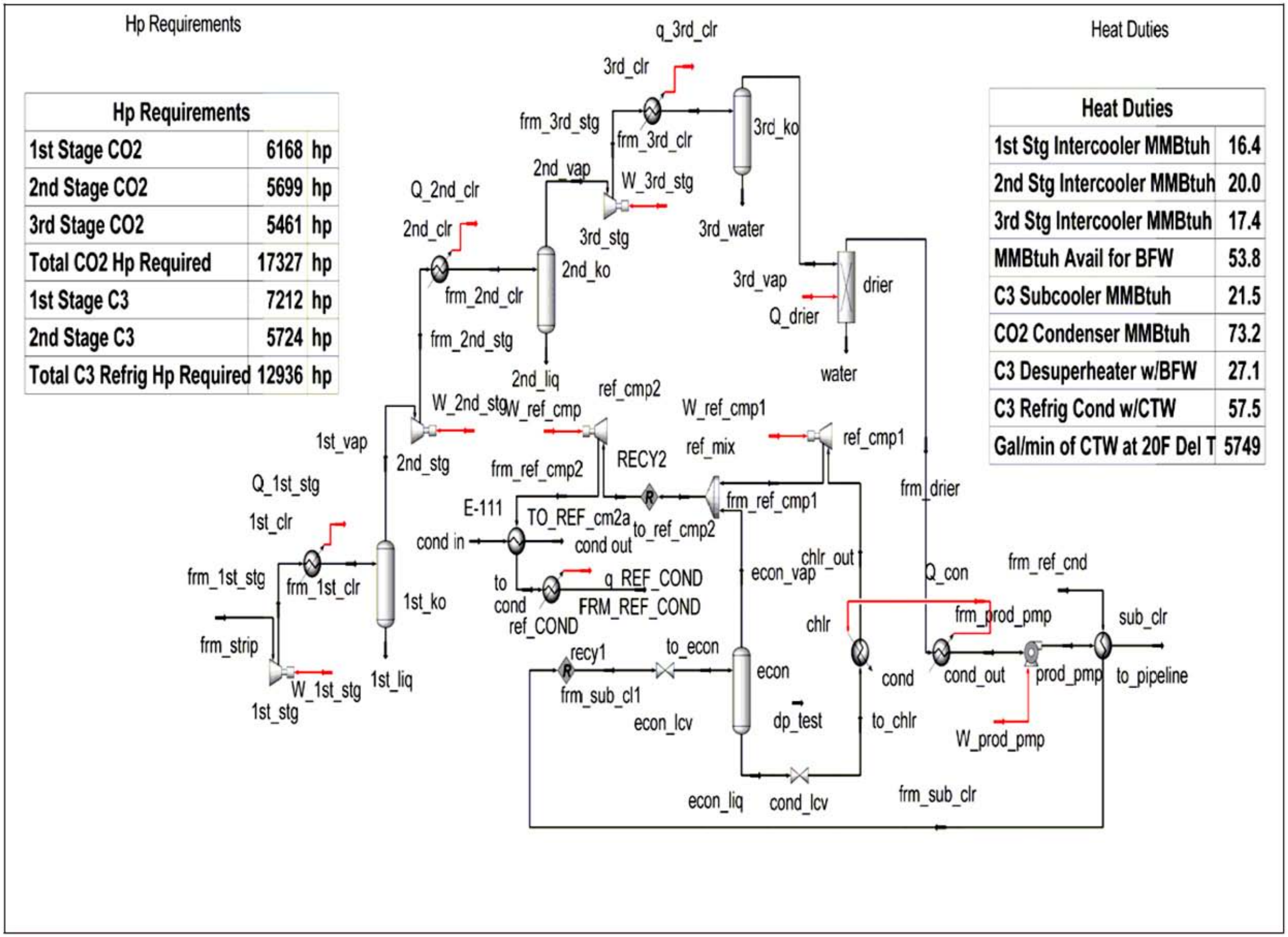
Table 3-23: Case 2 Material and Energy Balance for CO_2 Compression, Dehydration, and Liquefaction (70% CO_2 Recovery), continued

STREAM NAME		Vapor from Economizer	Refrig to CO2 Condenser	Economizer Liquid	To Economizer	From Subcooler	From Refrig Condenser	From Product Pump	CO2 To Pipeline
PFD STREAM NO.		econ_vap	to_chlr	econ_liq	to_econ	frm_sub_clr	frm_ref_cnd	frm_prod_pm	to_pipeline
VAPOR FRACTION	Molar	#DIV/0!	0.149	0.000	0.000	0.000	0.000	0.000	0.000
TEMPERATURE	°F	16	-32	16	16	15	100	-10	82
PRESSURE	PSIA	85	20	85	85	189	192	2,018	2,015
MOLAR FLOW RATE	lbmol/hr	-	12,522.00	12,522.00	12,522.00	12,522.00	12,522.00	13,763.95	13,763.95
MASS FLOW RATE	lb/hr	-	552,182.6	552,182.6	552,182.6	552,182.6	552,182.6	605,747.1	605,747.1
ENERGY	Btu/hr	0.00E+00	-6.68E+08	-6.68E+08	-6.68E+08	-6.68E+08	-6.38E+08	-2.43E+09	-2.40E+09
COMPOSITON	Mol %								
CO2		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	100.00%	100.00%
H2O		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Nitrogen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		95.58%	98.00%	98.00%	98.00%	98.00%	98.00%	0.00%	0.00%
Oxygen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Ethane		4.12%	1.00%	1.00%	1.00%	1.00%	1.00%	0.00%	0.00%
i-Butane		0.18%	0.50%	0.50%	0.50%	0.50%	0.50%	0.00%	0.00%
n-Butane		0.12%	0.50%	0.50%	0.50%	0.50%	0.50%	0.00%	0.00%
VAPOR									
MOLAR FLOW RATE	lbmol/hr	-	1,861.1	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	81,247.7	-	-	-	-	-	-
STD VOL. FLOW	MMSCFD	-	16.95	-	-	-	-	-	-
ACTUAL VOL. FLOW	ACFM	-	6,825.18	-	-	-	-	-	-
MOLECULAR WEIGHT	MW	43.56	43.66	-	-	-	-	-	-
DENSITY	lb/ft³	0.85	0.20	-	-	-	-	-	-
VISCOSITY	cP	0.0075	0.0065	-	-	-	-	-	-
LIGHT LIQUID									
MOLAR FLOW RATE	lbmol/hr	-	10,660.93	12,522.00	12,522.00	12,522.00	12,522.00	13,763.95	13,763.95
MASS FLOW RATE	lb/hr	-	470,934.9	552,182.6	552,182.6	552,182.6	552,182.6	605,747.1	605,747.1
STD VOL. FLOW	BPD	-	63,641	74,725	74,725	74,725	74,725	50,255	50,255
ACTUAL VOL. FLOW	GPM	-	1,638.59	2,036.48	2,036.48	2,029.63	2,339.24	1,100.79	1,487.37
DENSITY	lb/ft³	-	35.83	33.81	33.81	33.92	29.43	68.61	50.78
MOLECULAR WEIGHT	MW	-	44.17	44.10	44.10	44.10	44.10	44.01	44.01
VISCOSITY	cP	-	0.1841	0.1395	0.1395	0.1400	0.0881	0.1593	0.0622
SURFACE TENSION	Dyne/Cm	-	14.56	11.08	11.08	11.09	5.42	13.90	0.86
HEAVY LIQUID									
MOLAR FLOW RATE	lbmol/hr	-	0.00	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	-	-	-
DENSITY	lb/ft³	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	-	-	-

No.	Date	By	REVISION
90	7/17/2006	LEG	

ABB

Alstom Power
AEP Unit 5, Conesville, OH
70% CO2 Recovery
Heat & Material Balance
70%_5T_R2CTW80
JOB NO: LR12965 REV. A



Hp Requirements		
1st Stage CO2	6168	hp
2nd Stage CO2	5699	hp
3rd Stage CO2	5461	hp
Total CO2 Hp Required	17327	hp
1st Stage C3	7212	hp
2nd Stage C3	5724	hp
Total C3 Refrig Hp Required	12936	hp

Heat Duties	
1st Stg Intercooler MMBtuh	16.4
2nd Stg Intercooler MMBtuh	20.0
3rd Stg Intercooler MMBtuh	17.4
MMBtuh Avail for BFW	53.8
C3 Subcooler MMBtuh	21.5
CO2 Condenser MMBtuh	73.2
C3 Desuperheater w/BFW	27.1
C3 Refrig Cond w/CTW	57.5
Gal/min of CTW at 20F Del T	5749

Figure 3-10: Case 3 CO_2 Compression, Dehydration, and Liquefaction Schematic (50% CO_2 Recovery)

Table 3-24: Case 3 Material and Energy Balance for CO_2 Compression, Dehydration, and Liquefaction (50% CO_2 Recovery)

STREAM NAME		Total Acid Gas from Strippers	First Stage Discharge	To Second Stage	First Stage Water KO	From Second Stage	2nd Stage Discharge	To 3rd Stage	2nd Stage Water KO	From 3rd Stage	From 3rd Stage Cooler	3rd Stage Water KO	To Drier	Water From Drier	From Drier To Condenser	Condensed CO2 Product	Frm C3 Desuperheater E111	Refrig Compressor Discharge	Section of 2nd Refrig Compressor	Discharge from 1st Refrig Compr	Refrig from CO2 Condenser
PFD STREAM NO.		frm_strip	frm_1st_clr	1st_vap	1st_liq	rm_2nd_stg	frm_2nd_clr	2nd_vap	2nd_liq	frm_3rd_stg	frm_3rd_clr	3rd_water	3rd_vap	water	frm_drier	cond out	to cond	rm_ref_cmp	to_ref_cmp2	rm_ref_cmp	chlr_out
VAPOR FRACTION	Molar	1.000	0.993	1.000	0.000	1.000	0.974	1.000	0.000	1.000	0.988	0.000	1.000	0.000	1.000	0.000	1.000	1.000	1.000	1.000	1.000
TEMPERATURE	°F	115.0	125	125	125	275	125	125	125	275	125	125	125	125	125	205	125	263	173	173	56
PRESSURE	PSIA	19.0	41	41	41	95	89	89	89	206	200	200	200	195	195	95	199	234	85	85	20
MOLAR FLOW RATE	lbmol/hr	10,399	10,399	10,323	75.66	10,323	10,323	10,055.63	268	10,056	10,055.63	121.34	9,934.29	113.96	9,820.33	15,337.83	8,944.00	8,944.00	8,944.00	8,944.00	8,944.00
MASS FLOW RATE	lb/hr	442,639.6	442,639.6	441,275.1	1,364.5	441,275.1	441,275.1	436,440.5	4,834.6	436,440.5	436,440.5	2,197.8	434,242.7	2,053.0	432,189.7	276,312.6	394,403.6	394,403.6	394,403.6	394,403.6	394,403.6
ENERGY	Btu/hr	-1.72E+09	-1.72E+09	-1.71E+09	-9.22E+06	-1.70E+09	-1.72E+09	-1.68E+09	-3.26E+07	-1.67E+09	-1.69E+09	-1.48E+07	-1.67E+09	-1.39E+07	-1.66E+09	-1.85E+09	-3.98E+08	-3.71E+08	-3.85E+08	-3.85E+08	-4.04E+08
COMPOSITON	**Mol %**																				
CO2		94.44%	94.44%	95.14%	0.08%	95.14%	95.14%	97.66%	0.17%	97.66%	97.66%	0.38%	98.85%	0.00%	100.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
H2O		5.56%	5.56%	4.86%	99.92%	4.86%	4.86%	2.34%	99.83%	2.34%	2.34%	99.62%	1.15%	100.00%	0.00%	100.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Nitrogen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	98.00%	98.00%	98.00%	98.00%	98.00%
Oxygen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Ethane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	1.00%	1.00%	1.00%	1.00%	1.00%
i-Butane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.50%	0.50%	0.50%	0.50%	0.50%
n-Butane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.50%	0.50%	0.50%	0.50%	0.50%
VAPOR																					
MOLAR FLOW RATE	lbmol/hr	10,399	10,323	10,323	0	10,323	10,056	10,056	0	10,056	9,934.3	-	9,934.3	-	9,820.3	-	8,944.0	8,944.0	8,944.0	8,944.0	8,944.0
MASS FLOW RATE	lb/hr	442,639.6	441,275.1	441,275.1	-	441,275.1	436,440.5	436,440.5	-	436,440.5	434,242.7	-	434,242.7	-	432,189.7	-	394,403.6	394,403.6	394,403.6	394,403.6	394,403.6
STD VOL. FLOW	MMSCFD	94.71	94.03	94.03	-	94.03	91.59	91.59	-	91.59	90.48	-	90.48	-	89.44	-	81.46	81.46	81.46	81.46	81.46
ACTUAL VOL. FLOW	ACFM	55,896	25,989	25,989	0	14,074	11,495	11,495	0	6,230	4,880.78	-	4,880.78	-	4,963.42	-	3,735.54	4,359.32	11,160.53	11,160.53	40,151.15
MOLECULAR WEIGHT	MW	42.57	42.75	42.75	-	42.75	43.40	43.40	-	43.40	43.71	-	43.71	-	44.01	-	44.10	44.10	44.10	44.10	44.10
DENSITY	lb/ft³	0.13	0.28	0.28	-	0.52	0.63	0.63	-	1.17	1.48	-	1.48	-	1.45	-	1.76	1.51	0.59	0.59	0.16
VISCOSITY	cP	0.0151	0.0155	0.0155	-	0.0201	0.0160	0.0160	-	0.0207	0.0164	-	0.0164	-	0.0165	-	0.0097	0.0118	0.0099	0.0099	0.0079
LIGHT LIQUID																					
MOLAR FLOW RATE	lbmol/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
DENSITY	lb/ft³	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MOLECULAR WEIGHT	MW	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
HEAVY LIQUID																					
MOLAR FLOW RATE	lbmol/hr	-	75.66	-	75.66	-	267.71	-	267.71	-	121.34	121.34	-	113.96	-	15,337.83	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	1,364.5	-	1,364.5	-	4,834.6	-	4,834.6	-	2,197.8	2,197.8	-	2,053.0	-	276,312.6	-	-	-	-	-
STD VOL. FLOW	BPD	-	94	-	94	-	332	-	332	-	151	151	-	141	-	18,958	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	2.76	-	2.76	-	9.78	-	9.78	-	4.44	4.44	-	4.15	-	579.90	-	-	-	-	-
DENSITY	lb/ft³	-	61.63	-	61.63	-	61.66	-	61.66	-	61.72	61.72	-	61.65	-	59.41	-	-	-	-	-
VISCOSITY	cP	-	0.5291	-	0.5291	-	0.5651	-	0.5651	-	0.5621	0.5621	-	0.5311	-	0.2916	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	67.39	-	67.39	-	67.33	-	67.33	-	67.19	67.19	-	67.48	-	59.39	-	-	-	-	-

90	7/17/2006	LEG	
No.	**Date**	**By**	**REVISION**

NOTES:

Alstom Power
AEP Unit 5, Conesville, OH
50% CO2 Recovery
Heat & Material Balance
50%_4T_R2CTW80
JOB NO: LR12965 REV. A

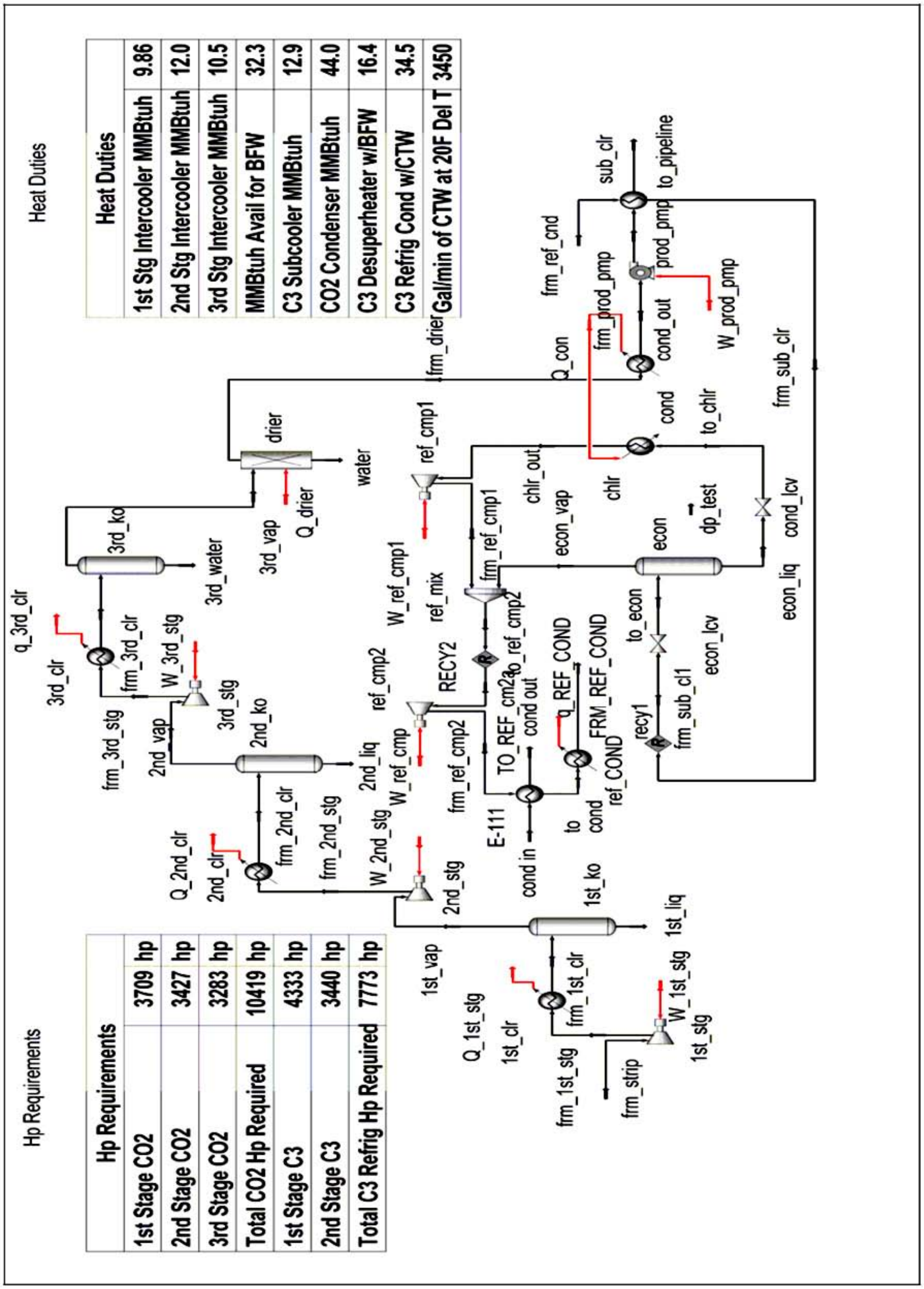
Table 3-24: Case 3 Material and Energy Balance for CO_2 Compression, Dehydration, and Liquefaction (50% CO_2 Recovery), continued

STREAM NAME		Vapor from Economizer	Refrig to CO2 Condenser	Economizer Liquid	To Economizer	From Subcooler	From Refrig Condenser	From Product Pump	CO2 To Pipeline
PFD STREAM NO.		econ_vap	to_chlr	econ_liq	to_econ	frm_sub_clr	frm_ref_cnd	m_prod_pm	to_pipeline
VAPOR FRACTION	Molar	#DIV/0!	0.149	0.000	0.000	0.000	0.000	0.000	0.000
TEMPERATURE	°F	16	-32	16	16	15	100	-10	82
PRESSURE	PSIA	85	20	85	85	189	192	2,018	2,015
MOLAR FLOW RATE	lbmol/hr	-	8,944.00	8,944.00	8,944.00	8,944.00	8,944.00	9,820.33	9,820.33
MASS FLOW RATE	lb/hr	-	394,403.6	394,403.6	394,403.6	394,403.6	394,403.6	432,189.7	432,189.7
ENERGY	Btu/hr	0.00E+00	-4.77E+08	-4.77E+08	-4.77E+08	-4.77E+08	-4.55E+08	-1.73E+09	-1.71E+09
COMPOSITON	Mol %								
CO2		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	100.00%	100.00%
H2O		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Nitrogen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		95.59%	98.00%	98.00%	98.00%	98.00%	98.00%	0.00%	0.00%
Oxygen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Ethane		4.12%	1.00%	1.00%	1.00%	1.00%	1.00%	0.00%	0.00%
i-Butane		0.18%	0.50%	0.50%	0.50%	0.50%	0.50%	0.00%	0.00%
n-Butane		0.12%	0.50%	0.50%	0.50%	0.50%	0.50%	0.00%	0.00%
VAPOR									
MOLAR FLOW RATE	lbmol/hr	-	1,332.2	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	58,161.8	-	-	-	-	-	-
STD VOL. FLOW	MMSCFD	-	12.13	-	-	-	-	-	-
ACTUAL VOL. FLOW	ACFM	-	4,885.82	-	-	-	-	-	-
MOLECULAR WEIGHT	MW	43.56	43.66	-	-	-	-	-	-
DENSITY	lb/ft^3	0.85	0.20	-	-	-	-	-	-
VISCOSITY	cP	0.0075	0.0065	-	-	-	-	-	-
LIGHT LIQUID									
MOLAR FLOW RATE	lbmol/hr	-	7,611.75	8,944.00	8,944.00	8,944.00	8,944.00	9,820.33	9,820.33
MASS FLOW RATE	lb/hr	-	336,241.8	394,403.6	394,403.6	394,403.6	394,403.6	432,189.7	432,189.7
STD VOL. FLOW	BPD	-	45,439	53,374	53,374	53,374	53,374	35,856	35,856
ACTUAL VOL. FLOW	GPM	-	1,169.93	1,454.78	1,454.78	1,449.89	1,670.83	785.40	1,061.21
DENSITY	lb/ft^3	-	35.83	33.80	33.80	33.91	29.43	68.61	50.78
MOLECULAR WEIGHT	MW	-	44.17	44.10	44.10	44.10	44.10	44.01	44.01
VISCOSITY	cP	-	0.1841	0.1394	0.1394	0.1399	0.0881	0.1593	0.0622
SURFACE TENSION	Dyne/Cm	-	14.56	11.07	11.07	11.09	5.42	13.90	0.86
HEAVY LIQUID									
MOLAR FLOW RATE	lbmol/hr	-	-	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	-	-	-
DENSITY	lb/ft^3	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	-	-	-

No.	Date	By	REVISION
90	7/17/2006	LEG	

ABB

Alstom Power
AEP Unit 5, Conesville, OH
50% CO2 Recovery
Heat & Material Balance
50%_4T_R2CTW80
JOB NO: LR12965 REV. A



Hp Requirements

Hp Requirements		
1st Stage CO2	3709	hp
2nd Stage CO2	3427	hp
3rd Stage CO2	3283	hp
Total CO2 Hp Required	10419	hp
1st Stage C3	4333	hp
2nd Stage C3	3440	hp
Total C3 Refrig Hp Required	7773	hp

Heat Duties

Heat Duties	
1st Stg Intercooler MMBtuh	9.86
2nd Stg Intercooler MMBtuh	12.0
3rd Stg Intercooler MMBtuh	10.5
MMBtuh Avail for BFW	32.3
C3 Subcooler MMBtuh	12.9
CO2 Condenser MMBtuh	44.0
C3 Desuperheater w/BFW	16.4
C3 Refrig Cond w/CTW	34.5
Gal/min of CTW at 20F Del T	3450

Figure 3-11: Case 4 CO_2 Compression, Dehydration, and Liquefaction Schematic (30% CO_2 Recovery)

Table 3-25: Case 4 Material and Energy Balance for CO_2 Compression, Dehydration, and Liquefaction (30% CO_2 Recovery)

STREAM NAME		Total Acid Gas from Strippers	First Stage Discharge	To Second Stage	First Stage Water KO	From Second Stage	2nd Stage Discharge	To 3rd Stage	2nd Stage Water KO	From 3rd Stage	From 3rd Stage Cooler	3rd Stage Water KO	To Drier	Water From Drier	From Drier To Condenser	Condensed CO2 Product	Frm C3 Desuperheater E111	Refrig Compressor Discharge	Section of 2nd Refrig Compressor	Discharge from 1st Refrig Compr	Refrig from CO2 Condenser
PFD STREAM NO.		frm_strip	frm_1st_clr	1st_vap	1st_liq	rm_2nd_stg	frm_2nd_clr	2nd_vap	2nd_liq	frm_3rd_stg	frm_3rd_clr	3rd_water	3rd_vap	water	frm_drier	cond out	to cond	rm_ref_cmp	to_ref_cmp2	rm_ref_cmp	chlr_out
VAPOR FRACTION	Molar	1.000	0.993	1.000	0.000	1.000	0.974	1.000	0.000	1.000	0.988	0.000	1.000	0.000	1.000	0.000	1.000	1.000	1.000	1.000	1.000
TEMPERATURE	°F	115.0	125	125	125	275	125	125	125	275	125	125	125	125	125	205	125	264	174	174	56
PRESSURE	PSIA	19.0	41	41	41	95	89	89	89	206	200	200	200	195	195	95	199	234	85	85	20
MOLAR FLOW RATE	lbmol/hr	6,253	6,253	6,208	45.49	6,208	6,208	6,046.53	161	6,047	6,046.53	72.96	5,973.57	68.53	5,905.04	9,247.81	5,367.00	5,367.00	5,367.00	5,367.00	5,367.00
MASS FLOW RATE	lb/hr	266,162.7	266,162.7	265,342.2	820.5	265,342.2	265,342.2	262,435.1	2,907.1	262,435.1	262,435.1	1,321.6	261,113.5	1,234.5	259,879.0	166,600.2	236,668.6	236,668.6	236,668.6	236,668.6	236,668.6
ENERGY	Btu/hr	-1.03E+09	-1.03E+09	-1.03E+09	-5.54E+06	-1.02E+09	-1.03E+09	-1.01E+09	-1.96E+07	-1.00E+09	-1.01E+09	-8.90E+06	-1.01E+09	-8.35E+06	-9.98E+08	-1.11E+09	-2.39E+08	-2.22E+08	-2.31E+08	-2.31E+08	-2.42E+08
COMPOSITON	**Mol %**																				
CO2		94.44%	94.44%	95.14%	0.08%	95.14%	95.14%	97.66%	0.17%	97.66%	97.66%	0.38%	98.85%	0.00%	100.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
H2O		5.56%	5.56%	4.86%	99.92%	4.86%	4.86%	2.34%	99.83%	2.34%	2.34%	99.62%	1.15%	100.00%	0.00%	100.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Nitrogen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	98.00%	98.00%	98.00%	98.00%	98.00%
Oxygen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Ethane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	1.00%	1.00%	1.00%	1.00%	1.00%
i-Butane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.50%	0.50%	0.50%	0.50%	0.50%
n-Butane		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.50%	0.50%	0.50%	0.50%	0.50%
VAPOR																					
MOLAR FLOW RATE	lbmol/hr	6,253	6,208	6,208	0	6,208	6,047	6,047	0	6,047	5,973.6	-	5,973.6	-	5,905.0	-	5,367.0	5,367.0	5,367.0	5,367.0	5,367.0
MASS FLOW RATE	lb/hr	266,162.7	265,342.2	265,342.2	-	265,342.2	262,435.1	262,435.1	-	262,435.1	261,113.5	-	261,113.5	-	259,879.0	-	236,668.6	236,668.6	236,668.6	236,668.6	236,668.6
STD VOL. FLOW	MMSCFD	56.95	56.54	56.54	-	56.54	55.07	55.07	-	55.07	54.41	-	54.41	-	53.78	-	48.88	48.88	48.88	48.88	48.88
ACTUAL VOL. FLOW	ACFM	33,611	15,628	15,628	0	8,463	6,912	6,912	0	3,746	2,934.85	-	2,934.85	-	2,984.54	-	2,241.58	2,619.65	6,706.10	6,706.10	24,127.54
MOLECULAR WEIGHT	MW	42.57	42.75	42.75	-	42.75	43.40	43.40	-	43.40	43.71	-	43.71	-	44.01	-	44.10	44.10	44.10	44.10	44.10
DENSITY	lb/ft³	0.13	0.28	0.28	-	0.52	0.63	0.63	-	1.17	1.48	-	1.48	-	1.45	-	1.76	1.51	0.59	0.59	0.16
VISCOSITY	cP	0.0151	0.0155	0.0155	-	0.0201	0.0160	0.0160	-	0.0207	0.0164	-	0.0164	-	0.0165	-	0.0097	0.0118	0.0099	0.0099	0.0079
LIGHT LIQUID																					
MOLAR FLOW RATE	lbmol/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
DENSITY	lb/ft³	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MOLECULAR WEIGHT	MW	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
HEAVY LIQUID																					
MOLAR FLOW RATE	lbmol/hr	-	45.49	-	45.49	-	160.98	-	160.98	-	72.96	72.96	-	68.53	-	9,247.81	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	820.5	-	820.5	-	2,907.1	-	2,907.1	-	1,321.6	1,321.6	-	1,234.5	-	166,600.2	-	-	-	-	-
STD VOL. FLOW	BPD	-	56	-	56	-	200	-	200	-	91	91	-	85	-	11,431	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	1.66	-	1.66	-	5.88	-	5.88	-	2.67	2.67	-	2.50	-	349.66	-	-	-	-	-
DENSITY	lb/ft³	-	61.63	-	61.63	-	61.66	-	61.66	-	61.72	61.72	-	61.63	-	59.40	-	-	-	-	-
VISCOSITY	cP	-	0.5291	-	0.5291	-	0.5651	-	0.5651	-	0.5621	0.5621	-	0.5272	-	0.2914	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	67.39	-	67.39	-	67.33	-	67.33	-	67.19	67.19	-	67.40	-	59.38	-	-	-	-	-

90	7/17/2006	LEG	
No.	Date	By	REVISION

NOTES:

ABB

Alstom Power
AEP Unit 5, Conesville, OH
30% CO2 Recovery
Heat & Material Balance
30%_3T_R2CTW80
JOB NO: LR12965 REV. A

Table 3-25: Case 4 Material and Energy Balance for CO_2 Compression, Dehydration, and Liquefaction (30% CO_2 Recovery), continued

STREAM NAME		Vapor from Economizer	Refrig to CO2 Condenser	Economizer Liquid	To Economizer	From Subcooler	From Refrig Condenser	From Product Pump	CO2 To Pipeline
PFD STREAM NO.		econ_vap	to_chlr	econ_liq	to_econ	frm_sub_clr	frm_ref_cnd	m_prod_pm	to_pipeline
VAPOR FRACTION	Molar	#DIV/0!	0.148	0.000	0.000	0.000	0.000	0.000	0.000
TEMPERATURE	°F	15	-32	15	15	15	100	-10	82
PRESSURE	PSIA	85	20	85	85	189	192	2,018	2,015
MOLAR FLOW RATE	lbmol/hr	-	5,367.00	5,367.00	5,367.00	5,367.00	5,367.00	5,905.04	5,905.04
MASS FLOW RATE	lb/hr	-	236,668.6	236,668.6	236,668.6	236,668.6	236,668.6	259,879.0	259,879.0
ENERGY	Btu/hr	0.00E+00	-2.86E+08	-2.86E+08	-2.86E+08	-2.86E+08	-2.73E+08	-1.04E+09	-1.03E+09
COMPOSITON	**Mol %**								
CO2		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	100.00%	100.00%
H2O		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Nitrogen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane		95.58%	98.00%	98.00%	98.00%	98.00%	98.00%	0.00%	0.00%
Oxygen		0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Ethane		4.12%	1.00%	1.00%	1.00%	1.00%	1.00%	0.00%	0.00%
i-Butane		0.18%	0.50%	0.50%	0.50%	0.50%	0.50%	0.00%	0.00%
n-Butane		0.12%	0.50%	0.50%	0.50%	0.50%	0.50%	0.00%	0.00%
VAPOR									
MOLAR FLOW RATE	lbmol/hr	-	796.1	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	34,754.1	-	-	-	-	-	-
STD VOL. FLOW	MMSCFD	-	7.25	-	-	-	-	-	-
ACTUAL VOL. FLOW	ACFM	-	2,919.52	-	-	-	-	-	-
MOLECULAR WEIGHT	MW	43.56	43.66	-	-	-	-	-	-
DENSITY	lb/ft³	0.85	0.20	-	-	-	-	-	-
VISCOSITY	cP	0.0075	0.0065	-	-	-	-	-	-
LIGHT LIQUID									
MOLAR FLOW RATE	lbmol/hr	-	4,570.91	5,367.00	5,367.00	5,367.00	5,367.00	5,905.04	5,905.04
MASS FLOW RATE	lb/hr	-	201,914.5	236,668.6	236,668.6	236,668.6	236,668.6	259,879.0	259,879.0
STD VOL. FLOW	BPD	-	27,286	32,028	32,028	32,028	32,028	21,561	21,561
ACTUAL VOL. FLOW	GPM	-	702.55	872.74	872.74	869.81	1,002.61	472.26	638.11
DENSITY	lb/ft³	-	35.83	33.81	33.81	33.92	29.43	68.61	50.78
MOLECULAR WEIGHT	MW	-	44.17	44.10	44.10	44.10	44.10	44.01	44.01
VISCOSITY	cP	-	0.1841	0.1396	0.1396	0.1400	0.0881	0.1593	0.0622
SURFACE TENSION	Dyne/Cm	-	14.56	11.08	11.08	11.10	5.42	13.90	0.86
HEAVY LIQUID									
MOLAR FLOW RATE	lbmol/hr	-	0.00	-	-	-	-	-	-
MASS FLOW RATE	lb/hr	-	-	-	-	-	-	-	-
STD VOL. FLOW	BPD	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW	GPM	-	-	-	-	-	-	-	-
DENSITY	lb/ft³	-	-	-	-	-	-	-	-
VISCOSITY	cP	-	-	-	-	-	-	-	-
SURFACE TENSION	Dyne/Cm	-	-	-	-	-	-	-	-

90	7/17/2006	LEG	
No.	**Date**	**By**	**REVISION**

ABB

Alstom Power
AEP Unit 5, Conesville, OH
30% CO2 Recovery
Heat & Material Balance
30% 3T R2CTW80
JOB NO: LR12965 REV. A

CO_2 Product Specification and Actual Composition (Cases 1-4)

The CO_2 product specification and actual composition are shown in Table 3-26. Note that no mercaptans nor methane and heavier hydrocarbons are shown in the flue gas analysis. Therefore, these components are shown as zero in Table 3-26. A CO_2 product pressure of 139 bara (2,015 psia) was used for all the cases.

Table 3-26: CO_2 Product Specification and Calculated Product Comparison (Cases 1-4)

Component	Specification	Calculated Results
	Mole %	Mole %
O_2	0.0100	<0.0050
N_2	0.6000	<0.0400
H_2O	0.0002	<0.0002
CO_2	96.000	>99.95
H_2S	0.0001	<0.0001
Mercaptans	0.0300	0.00
CH_4	0.3000	0.00
C_2 + Hydrocarbons	2.0000	0.00

3.1.4.2 Consumption of Chemicals and Desiccants - CO_2 Removal, Compression, and Liquefaction System (Cases 1-4)

The table below shows the daily chemical consumption for Cases 1-4 with 90%-30% CO_2 recovery respectively. These totals do not include chemicals provided by the cooling tower service people nor disposal of waste, which are handled as a component of operating costs referred to as contracted services and waste handling, respectively.

Table 3-27: Chemical and Desiccants Consumption (lbm/day) for Cases-1-4 (90%-30% CO_2 Recovery)

Chemical	Case 1 (90% Capture)	Case 2 (70% Capture)	Case 3 (50% Capture)	Case 4 (30% Capture)
Soda Ash	2,328	1,811 1,293		776
MEA	28,046	21,813	15,581	9,349
Corrosion inhibitor	1,028	800	571	343
Diatomaceous earth	458	356 254		153
Molecular sieve	257	200 143		86
Activated carbon	1,546	1,202	859	515

3.1.4.3 Equipment - CO_2 Removal, Compression, and Liquefaction System (Cases 1-4)

Complete equipment data summary sheets for Cases 1-4 are provided in Appendix II. These equipment lists have been presented in the so-called “short spec” format, which provides adequate data for developing a factored cost estimate. Table 3-28 shows a summary of the major equipment for the CO_2 Removal, Compression, and Liquefaction Systems. Three categories are shown in this table (Compressors, Towers/Internals, and Heat Exchangers). These three categories represent, in that order, the three most costly accounts in the cost estimates for these systems (See Section 3.3). These three accounts represent ~90% of the total equipment costs for these systems.

Table 3-28: Equipment Summary - CO_2 Removal, Compression, and Liquefaction System (Cases 1-4)

	Case 1 (90% Capture)		**Case 2** (70% Capture)		**Case 3** (50% Capture)		**Case 4** (30% Capture)	
Compressors	No.	HP each	No.	HP each	No.	HP each	No.	HP each
CO_2 Compressor	2	15,600	2	12,100	1	17,300	1	10,400
Propane Compressor	2	11,700	2	10,200	1	14,600	1	8,800
LP Let Down Turbine	1	60,800	1	47,200	1	33,600	1	20,000
Towers/Internals	No.	ID/Height (ft)	No.	ID/Height (ft)	No.	ID/Height (ft)	No.	ID/Height (ft)
Absorber/Cooler	2	34 / 126	2	30 / 126	2	25 / 126	1	28 / 126
Stripper	2	22 / 50	2	19 / 50	2	16 / 50	1	20 / 50
Heat Exchangers	No.	10^6-Btu/hr ea.	No.	10^6-Btu/hr ea.	No.	10^6-Btu/hr ea.	No.	10^6-Btu/hr ea.
Reboilers	10	120.0	8	120.0	6	120.0	4	120.0
Solvent Stripper CW Condenser	12	20.0	10	20.0	7	20.0	4	20.0
Other Heat Exchangers / Avg Duty	36	61.0	35	57.0	25	62.0	16	58.0
Total Heat Exchangers / Avg Duty	58	62.7	53	59.5	38	63.4	24	62.0

A review of this table shows how the number of compression trains is reduced from two trains for the 90% and 70% recovery cases to one train for the 50% and 30% recovery cases. Similarly the number of absorber/stripper trains is reduced from two trains for the 90%, 70% and 50% recovery cases to one train for the 30% recovery case. Additionally, the sizes of the vessels and power requirements for the compressors are also changing. The heat exchanger selections also show variation between the cases. Figure 3-12 is provided to help illustrate how the number of trains (compressor, absorber, and stripper), compressor power requirements, vessel sizes, and the number and heat duty of the heat exchangers in the system change as a function of the CO_2 recovery percentage.

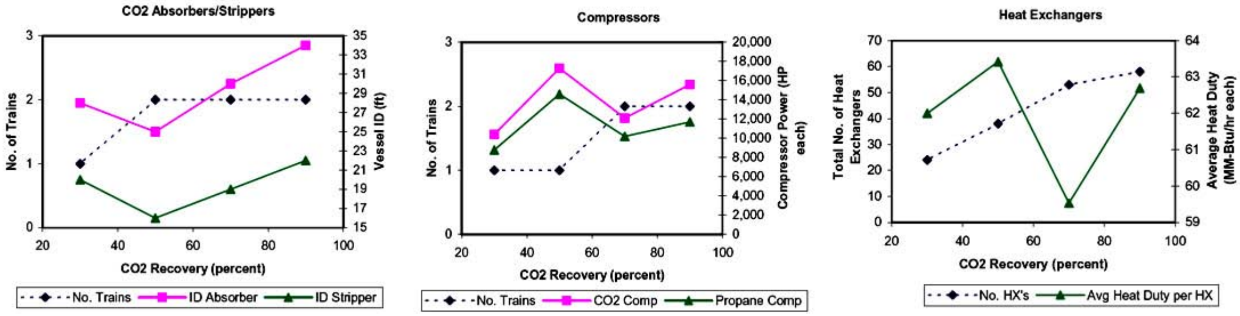


Figure 3-12: Equipment Variations – CO_2 Removal, Compression, and Liquefaction Systems (Cases 1-4)

3.1.4.4 Utilities Usage and Auxiliary Power Requirements - CO_2 Removal, Compression, and Liquefaction System (Cases 1-4)

Table 3-29 shows the CO_2 Removal and Compression System utilities usage for Cases 1-4. Table 3-30, Table 3-31, Table 3-32, and Table 3-33 show auxiliary power requirements for Cases 1-4 respectively (90%-30% CO_2 recovery).

Table 3-29: Consumption of Utilities for Cases 1-4 (90%-30% CO_2 Recovery)

Utility	Units	Case 1 (90% Capture)	Case 2 (70% Recovery)	Case 3 (50% Capture)	Case 4 (30% Capture)
Natural Gas for CO_2 Dryers	SCF/day	312,000	232,000	161,000	101,000
Saturated Steam at 45 psia	lbm/hr	1,300,000	1,010,000	722,000	433,333
80°F Cooling Tower Water	Gal/minute at 30°F rise	69,694	54,217	38,693	22,991

Table 3-30: Auxiliary Power Usage for Case 1 (90% CO_2 Recovery)

Number of Trains	Tag no.	Description	Number Operating per train	Power ea w/ 0.95 motor eff (kW)	Total all trains (kW)
2	Pump-2	Wash Water Pump	2	52	210
2	Pump-1	Direct Contact Cooler Water Pump	2	90	359
2	P-100	Rich Solvent Pump	2	430	1,719
2	P-102	Lean Solvent Pump	2	291	1,166
2	P-101	Semi-Lean Pump	2	130	519
2		Solvent Stripper Reflux Pump	1	11	22
2		Filter Circ. Pump	2	21	85
7		CO_2 Pipeline Pump	1	304	2,130
2		LP condensate booster pump	2	108	434
2		Soda ash metering pump	1	0	0
2		Flue Gas FD Fan	1	2,579	5,158
2		CO_2 Compressor (Motor driven)	1	12,270	24,539
2		Propane Refrigeration Compressors (2)	1	9,153	18,306
1		LP steam turbine/ generator	NA	NA	NA
2		CO_2 Dryer Package	1	146	292
		Total			54,939

Table 3-31: Auxiliary Power Usage for Case 2 (70% CO_2 Recovery)

Number of Trains	Tag no.	Description	Number Operating per train	Power ea w/ 0.95 motor eff (kW)	Total all trains (kW)
2	Pump-2	Wash Water Pump	2	41	163
2	Pump-1	Direct Contact Cooler Water Pump	2	69	277
2	P-100	Rich Solvent Pump	2	334	1,337
2	P-102	Lean Solvent Pump	2	228	912
2	P-101	Semi-Lean Pump	2	100	398
2		Solvent Stripper Reflux Pump	1	9	17
2		Filter Circ. Pump	2	17	66
5		CO_2 Pipeline Pump	1	330	1,650
2		LP condensate booster pump	2	84	337
2		Soda ash metering pump	1	0	0
2		Flue Gas FD Fan	1	2,006	4,012
2		CO_2 Compressor (Motor driven)	1	9,531	19,062
2		Propane Refrigeration Compressors (2)	1	7,113	14,226
1		LP steam turbine/ generator	NA	NA	NA
2		CO_2 Dryer Package	1	120	240
		Total			42,697

Table 3-32: Auxiliary Power Usage for Case 3 (50% CO_2 Recovery)

Number of Trains	Tag no.	Description	Number Operating per train	Power ea w/ 0.95 motor eff. (kW)	Total all trains (kW)
2	Pump-2	Wash Water Pump	2	29	117
2	Pump-1	Direct Contact Cooler Water Pump	2	49	196
2	P-100	Rich Solvent Pump	2	239	955
2	P-102	Lean Solvent Pump	2	163	651
2	P-101	Semi-Lean Pump	2	71	284
2		Solvent Stripper Reflux Pump	1	6	12
2		Filter Circ. Pump	2	12	47
4		CO_2 Pipeline Pump	1	295	1,180
2		LP condensate booster pump	2	60	241
2		Soda ash metering pump	1	0	0
2		Flue Gas FD Fan	1	1,433	2,866
1		CO_2 Compressor (Motor driven)	1	13,602	13,602
1		Propane Refrigeration Compressors (2)	1	10,154	10,154
1		LP steam turbine/ generator	NA	NA	NA
1		CO_2 Dryer Package	1	161	161
		Total			30,466

Table 3-33: Auxiliary Power Usage for Case 4 (30% CO_2 Recovery)

Number of Trains	Tag no.	Description	Number Operating per train	Power ea w/ 0.95 motor eff (kW)	Total all trains (kW)
1	Pump-2	Wash Water Pump	2	35	70
1	Pump-1	Direct Contact Cooler Water Pump	2	58	116
1	P-100	Rich Solvent Pump	2	287	574
1	P-102	Lean Solvent Pump	2	193	386
1	P-101	Semi-Lean Pump	2	88	176
1		Solvent Stripper Reflux Pump	1	8	8
1		Filter Circ. Pump	2	14	28
3		CO_2 Pipeline Pump	1	237	711
1		LP condensate booster pump	2	72	145
1		Soda ash metering pump	1	0	0
1		Flue Gas FD Fan	1	1,719	1,719
1		CO_2 Compressor (Motor driven)	1	8,178	8,178
1		Propane Refrigeration Compressors (2)	1	6,101	6,101
1		LP steam turbine/ generator	NA	NA	NA
1		CO_2 Dryer Package	1	101	101
		Total			18,312

3.1.4.5 Design Considerations and System Optimization - CO_2 Removal, Compression, and Liquefaction System (Cases 1-4)

A commercial simulator called ProTreat® Version 3.3 was used to simulate the advanced MEA process and Hysys® Version 2004.2 was used to simulate CO_2 compression and liquefaction system. The key process parameters used are listed in Table 3-34 below.

Table 3-34: Key Process Parameters for Simulation (Cases 1-4)

Process Parameter	Value
CO_2 in Feed, mol %	12.8
O_2 in Feed, mol %	2.9
SO_2 in Feed, ppmv	2
Solvent Type	MEA
Solvent Concentration, Wt%	30
Lean Loading, mol CO_2/mol amine	0.19
Rich Loading, mol CO_2/mol amine	0.49
Stripper Feed Temp, °F	205
Stripper Bottom Temp, °F	247
Feed Temp To Absorber, °F	115
CO_2 Recovery, %	30-90
Absorber Pressure Drop, psi	1
Stripper Pressure Drop, psi	0.7
Rich/Lean Exchanger Approach, °F	40
CO_2 Compressor 1st /Stage Temp, °F	125
Liquid CO_2 Temp, °F	82
Steam Use, lbs Steam/ lb CO_2 captured	1.67
Liquid CO_2 Pressure, psia	2,015

The following parameters were investigated with the objective of reducing the MEA plant energy requirements and ultimately the cost of electricity produced by the power plant.

Number of Absorber and Stripper Trains:

The number of absorbers and strippers is based on using a maximum diameter of 12.2 m (40 ft). The minimum diameter is achieved by bypassing available flue gas while keeping the percentage of CO_2 absorbed in the absorber at 90%.

Absorber Temperature:

Two temperatures were investigated: 58°C (136°F) and 46°C (115°F). A flue gas cooler was added upstream of the absorber to cool the flue gas from 58°C (136°F) to 46°C (115°F). At 58°C (136°F), 90% CO_2 recovery is not achievable due to equilibrium constraints.

Stripper Temperature / Reboiler Pressure:

A preliminary optimization study was done to define the best reboiler pressure for the design of this plant. This was done for the 90% capture case only (Case 1). In this study it was observed that a reduction in reboiler pressure (let down turbine exhaust pressure) would have the following primary impacts:

- Increased Let Down Turbine Output
- Increased Net Plant Output
- Higher Plant Thermal Efficiency
- Increased Let Down Turbine Cost
- Increased Reboiler Cost
- Higher Total Retrofit Costs

The results for the reboiler pressure optimization study are shown in Figure 3-13. The graph on the left shows how the plant thermal efficiency improves linearly and plant retrofit cost increases exponentially as let down turbine outlet pressure is reduced. The graph on the right shows how the combined effect of plant efficiency improvement and retrofit cost increase causes the incremental cost of electricity (COE) to be minimized at a let down turbine outlet pressure of about 2.8-3.4 bara (40-50 psia). A let down turbine outlet pressure of 3.2 bara (47 psia) was selected for this study. Allowing about 0.14 bar (2 psi) for pressure drop between the let down turbine exhaust and the reboiler yields a reboiler operating pressure of 3.1 bara (45 psia). The use of 3.1 bara (45 psia) pressure steam in the stripper reboiler causes no significant sacrifice in the CO_2 loading in the lean amine.

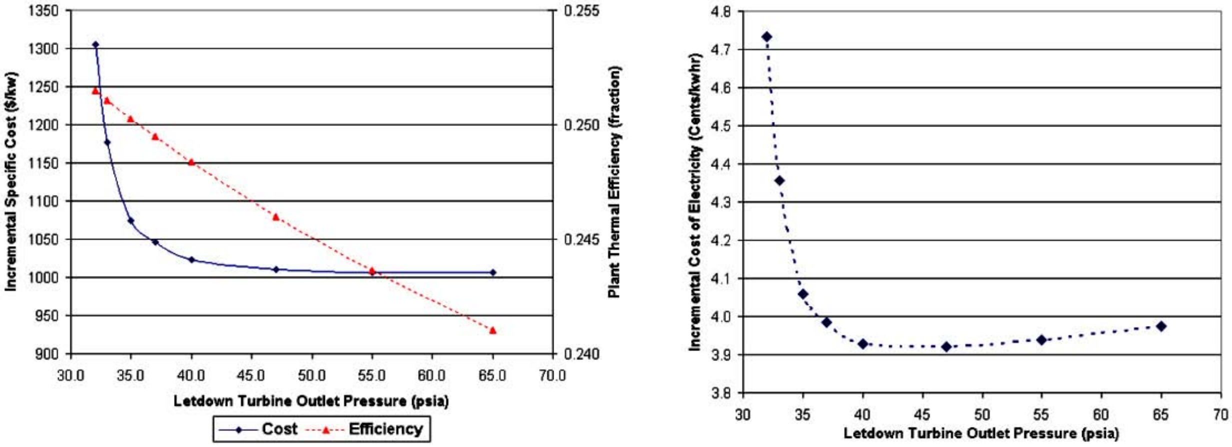


Figure 3-13: Reboiler Pressure Optimization Study Results (Case 1)

Absorber and Stripper Packing Type and Depth:

Eighty-five types of packing were investigated to optimize the absorber and stripper diameter. The packing depth in both the absorber and stripper was optimized until a 90% CO_2 recovery was achieved.

Location and Amount of the Semi-Lean Amine to the Absorber:

The entry location of the semi-lean amine stream to the absorber and the amount of semi-lean amine was varied to minimize energy consumption and maximize CO_2 recovery.

Heat Exchanger Types:

Plate Frame Heat Exchangers, Shell and Tube Exchangers, and Air-Cooled Exchangers were investigated. Plate frame type heat exchangers were used as much as possible to improve energy efficiency and reduce costs.

Number of CO_2 Compression Trains:

Two compression trains are specified to provide for plant turndown capability for the 90% and 70% CO_2 recovery cases. At lower recoveries (50% and 30%) just one train is provided.

3.1.4.6 Outside Boundary Limits (OSBL) Systems - CO_2 Removal, Compression, and Liquefaction System (Cases 1-4)

Reclaimer Bottoms:

The reclaimer bottoms are generated during the process of recovering MEA from heat stable salts (HSS). HSS are produced from the reaction of MEA with SO_2 and NO_2. The HSS accumulate in the reclaimer during the lean amine feed portion of the reclaiming cycle. The volume of reclaimer bottoms generated will depend on the quantity of SO_2 and NO_2 not removed in the Flue Gas Scrubber. A typical composition of the waste is presented below.

Table 3-35: Reclaimer Bottoms Composition (Cases 1-4)

MEA	9.5 wt.%
NH_3	0.02 wt.%
NaCl	0.6 wt.%
Na_2SO_4	6.6 wt.%
Na_2CO_3	1.7 wt.%
Insolubles	1.3 wt.%
Total Nitrogen	5.6 wt.%
Total Organic Carbon	15.6 wt.%
H_2O	59.08 wt.%
pH	10.7
Specific Gravity	1.14

Filter Residues:

A pressure leaf filter filters a slipstream of lean amine. Diatomaceous earth is used as a filter-aid for pre-coating the leaves and as a body feed. Filter cycles depend on the rate of flow through the filter, the amount of filter aid applied, and the quantity of contaminants in the solvent. A typical composition of the filter residue is provided in the table below. These will be disposed of by a contracted service hauling away the drums of spent cake.

Table 3-36: Filter Residue Composition (Cases 1-4)

MEA	2.5 wt.%
Total Organic Carbon	1.5 wt.%
SiO_2	43 wt.%
Iron Oxides	32 wt.%
Aluminum Oxides	15 wt.%
H_2O	6 wt.%
pH	10.0
Specific Gravity	2.6

Excess Solvent Stripper Reflux Water:

The CO_2 Recovery Facility has been designed to operate in a manner to avoid accumulation of water in the Absorber / Stripper system. By controlling the temperature of the scrubbed flue gas entering the absorber the MEA system can be kept in water balance. Excess water can accumulate in the Stripper Reflux Drum and can be reused once the system is corrected to operate in a balanced manner. Should water need to be discarded, contaminants will include small amounts of CO_2 and MEA.

Absorber Flue Gas Scrubber/Cooler:

The existing plant uses lime in its flue gas desulfurizer. In the cost estimate of this plant, it has been assumed that the existing plant disposal facilities can accommodate the additional water blow down load from the flue gas cooler located under the absorber.

Relief Requirements:

The relief valve discharges from the CO_2 Recovery Unit are discharged to atmosphere. No tie-ins to any flare header are necessary.

3.1.4.7 Plant Layout - CO_2 Removal, Compression, and Liquefaction System (Cases 1-4)

Please refer to Appendix I for the plant layout drawings for the modified Conesville #5 Unit. The plant layout for the CO_2 capture equipment has been designed in accordance with a spacing chart called “Oil and Chemical Plant Layout and Spacing” Section IM.2.5.2 issued by Industrial Risk Insurers (IRI).

The open-cup flash point of MEA is 93°C (200°F); and, therefore, it will not easily ignite. In addition to MEA, the corrosion inhibitor is the only other hydrocarbon liquid within the battery limits. The flash point of this material is higher than that of MEA and is handled in small quantities. Thus, no highly flammable materials are handled within the CO_2 Recovery Unit. As the chemicals used in the process present no fire hazard, there is an opportunity to reduce the minimum spacing between equipment from that normally considered acceptable in hydrocarbon handling plants. However, for the drawings that follow, standard spacing requirements, as suggested by IRI have been followed.

The relatively unoccupied plot areas available on the existing site in the immediate vicinity of Unit #5 for the installation of the desired equipment are small. Some equipment items are placed on structures to allow other pieces of equipment to be placed underneath them. This way, pumps and other equipment associated with the absorber can be located under the structure. Locating the pumps under the structure has been considered acceptable because the fluids being pumped are not flammable.

Discussions with vendors suggest that it will be possible to provide insulation on the flue gas fan casing to limit noise to acceptable level. Therefore, it has been assumed that no building needs to be provided for noise reasons.

The CO_2 absorbers are placed adjacent to the flue gas desulfurization (FGD) system scrubbers to minimize the length of the flue gas duct feeding the bottom of the absorbers. Figure 3-14 shows the existing FGD scrubbers (2 -50% units) located just left (west) of the common stack used for Units 5/6, which is shown on the far right side of Figure 3-14. The new CO_2 absorbers would be placed just to the left (west) of the existing FGD system scrubbers (far left side of Figure 3-14).

Figure 3-14: Conesville Unit #5 Existing Flue Gas Desulfurization System Scrubbers and Stack

The new strippers and the new let down turbine are placed ~30 m (100 ft) south of the existing Unit #5 intermediate pressure turbine just behind the existing turbine building shown in Figure 3-15. This location minimizes the length of the low-pressure steam line feeding the new LP let down turbine and the reboilers. The actual location for the new equipment would be just south of the road in the grassy area shown in the bottom part of Figure 3-15. The top of the Unit #5 boiler can be seen in the upper left side of Figure 3-15 and the duplicate Unit #6 boiler is on the upper right side.

Figure 3-15: Conesville Unit #5 Existing Turbine Building

The new low-pressure steam line runs from the IP/LP crossover pipe (shown in Figure 3-16) to the new let down low-pressure steam turbine, which is located near the strippers just beyond the outside wall shown in the background. The IP/LP crossover pipe will need to be modified with the addition of the steam extraction pipe to feed the let down turbine and the reboiler/reclaimer system. Additionally, a pressure control valve will need to be added downstream of the extraction point as described in Section 3.1.6.

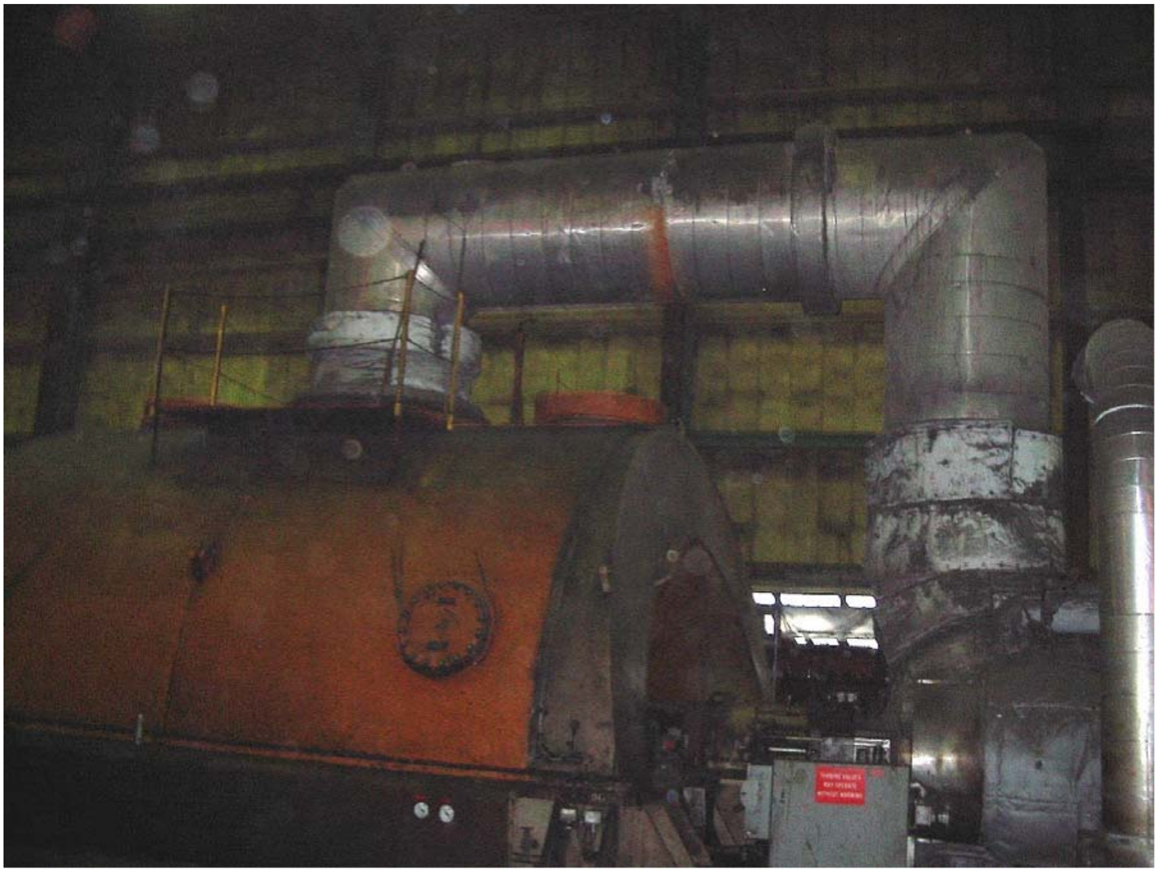


Figure 3-16: Conesville Unit #5 Existing LP Turbine and IP/LP Crossover Pipe

The new CO_2 compression and liquefaction system is located between two existing cooling tower banks as shown in Figure 3-17, ~150 m (500 ft) south of the new strippers. An abandoned warehouse must be removed to make room for the CO_2 Compression Facilities.

Figure 3-17: Existing Conesville Cooling Towers & CO_2 Compression/Liquefaction System Location

The corrosion inhibitor must be protected against freezing during winter. The soda ash solution will not freeze but will become very viscous when it gets cold. Therefore, a heated shed has been provided for housing the Corrosion Inhibitor and the soda ash injection packages.

3.1.5 Case 5/Concept A: Design and Performance of Kerr-McGee/ABB Lummus Amine CO_2 Removal System

Case 5 represents an update (costs and economics only) of a case (Concept A) from an earlier Alstom study (Bozzuto et al., 2001). The process design and equipment selection from the earlier study was not updated in this study. The information provided for Case 5/Concept A in this section and other sections in this report was copied or adapted from the earlier study. It should be noted that the design of Case 5 with ~96% CO_2 recovery (See Bozzuto et al., 2001) is not totally consistent with the design of Case 1 (90% CO_2 recovery) from the current study. Case 1 uses two absorbers, two strippers, and two compression trains. Whereas, Case 5, which was designed in 2000, used five absorbers, nine strippers, and seven compression trains. Additionally, Case 5 equipment, which occupies about twice as much land area, was all located about 1,500 feet from the Unit #5 stack whereas the Case 1 CO_2 Removal System equipment could be located much closer to the existing plant in three primary locations as explained previously.

Case 5/Concept A from this earlier study was a post-combustion system, which used an amine-based (MEA) scrubber for CO_2 recovery. In Concept A, coal is burned conventionally in air.

The flue gases leaving the modified FGD system (a secondary absorber is added to reduce the SO_2 concentration as required by the MEA system) are cooled with a direct contact cooler and ducted to the MEA system where more than 96% of the CO_2 is removed, compressed, and liquefied for usage or sequestration. The remaining flue gases leaving the new MEA system, consisting primarily of oxygen, nitrogen, water vapor and a relatively small amount of sulfur dioxide and carbon dioxide, are discharged to the atmosphere. The Kerr-McGee/ABB Lummus amine technology is used for the Case 5/Concept A CO_2 removal system.

The CO_2 Recovery Unit for Case 5/Concept A is comprised of the following sections:

- Flue Gas Pretreatment
- Absorption
- Stripping
- CO_2 Compression and Liquefaction
- CO_2 Drying

The flue gas pre-treatment section cools and conditions the flue gas, which is then fed to the CO_2 Absorber. In the Absorber, CO_2 is removed from the gas by contacting it, in counter current fashion, with MEA. The recovered CO_2 is then stripped off in the Stripper (or Regenerator) from where the lean solvent is recycled back to the Absorber. Solvent regeneration for Case 5/Concept A requires about 5.46 MJ/Tonne CO_2 (2,350 Btu/lbm-CO_2). The overhead vapor from the Stripper is cooled to condense most of the water vapor. The condensate is used as reflux in the Stripper, and the wet CO_2 stream is fed to the CO_2 Compression and Liquefaction System. Here the CO_2 product is compressed and dried so it can be pumped to its final destination. No specific destination has been chosen for the product pipeline. It has been assumed to end at the battery limit (outlet flange of the CO_2 pump) for costing purposes.

A brief description of the processing scheme for Case 5/Concept A is given in the following paragraphs. Description of the package units is indicative only and may vary for the chosen supplier of the package unit.

3.1.5.1 Case 5/Concept A Process Description - CO_2 Removal, Compression, and Liquefaction System

This section refers to the following process flow diagrams:

- Figure 3-18: Drawing D 09484-01001R-0: Flue Gas Cooling and CO_2 Absorption
- Figure 3-19: Drawing D 09484-01002R-0: Solvent Stripping
- Figure 3-20: Drawing D 09484-01003R-0: CO_2 Compression and Liquefaction

The designs include several process trains. Only one train is shown. The note section of the PFD tells how many trains are included in the complete system. To avoid confusion, suffixes have been used to indicate parallel equipment. These are mainly for spared pumps and drier vessels in parallel. Even if there are several trains, only one drawing (typical) has been prepared to represent all of the trains. On these drawings, flow splits to the other parallel trains have been shown. Similarly, flows coming from other parallel trains and converging to a single common stream have also been shown.

A note about stream numbering convention is also necessary. The stream numbers have not been tagged with "A," "B," etc. to indicate which train they belong to. Instead, the flow rate given in the material balance for each stream is the actual flow rate for the stream within the train. The combined flow from all of the trains leaving a process step shows the total flow going to the next process step. As an example, stream 8 (Drawing D 09484-01001R-0) is the rich amine stream leaving one train of the absorber process step, and comprises 1/5 of the total rich amine. Stream 9A is the total rich amine going to the solvent stripping process step. Stream 9A appears on both the absorber and solvent stripper PFDs. After the rich amine flow sheet continuation block, the stream splits nine ways for the nine stripping trains. Then, stream 9 continues for processing on the solvent stripper PFD (Drawing D 09484-01002R-0), with 1/9 of the flow entering the rich-lean solvent exchanger (EA-2205).

Flue Gas Pretreatment:

The pressure profile of the CO_2 capture equipment is contained in the material balance. Since the flue gas pre-treatment equipment flow scheme includes a blower, the pressure profile of the existing Conesville #5 power generation equipment does not change from current operation. To force the flue gas from the secondary FGD through the CO_2 Absorber, the pressure of the flue gas after sulfur removal is boosted to 0.1 barg (1.5 psig) by a motor driven fan. As the power consumption of the fan is considerable, the duct size must be chosen so as not to cause excessive pressure drop over the 460 m (1,500 ft) it takes to get to the absorbers. The blower will run at constant speed. Each blower, provided as part of the boiler flue gas conditioning equipment, is equipped with its own suction and a discharge damper operated pneumatically. The suction damper controls the suction pressure to adjust for the flow variation resulting from the power plant performance. The suction pressure control will avoid any surges to blower. The discharge damper is an isolation damper.

Direct Contact Cooling (Refer to Figure 3-18):

The Direct Contact flue gas Cooler (DCC) is a packed column where the hot flue gas flowing up is brought into intimate contact with cold water, which is fed to the top of the bed and flows down the tower. Physically, DA-2101 and DA-2102 have been combined into a single, albeit compartmentalized tower. DA-2101 is the lower compartment and is designed to support DA-2102 so that the top head of DA-2101 is the bottom head of DA-2102. Effectively, this dividing head acts as a chimney tray with a number of upward extending chimneys, which provide passages for the flue gas to flow directly from the DCC into the Absorber.

Theoretically, a direct contact cooler is capable of cooling the gas to a very close approach in a short bed. When the hot gas enters the DCC, the gas contains water but is highly superheated. At the bottom end of the bed, the gas is quickly cooled to a temperature known as the "Adiabatic Saturation Temperature" (AST). This is the temperature the gas reaches when some of its own heat content has been used to vaporize just the exact amount of water to saturate the gas.

Up to the point when the AST is reached, the mass flow of the gas stream increases due to evaporation of water. At the AST, water vapor contained in the gas begins to condense as the gas is further cooled. And, as the gas travels up the column and is cooled further, more water is condensed. This internal refluxing increases the V/L traffic at the bottom end of the bed significantly beyond the external flows and must be considered in the hydraulic design.

The water stream that leaves the bottom of the DCC contains the water fed to the top as well as any water that has condensed out of the flue gas. The condensed water may be somewhat corrosive due to sulfur and nitrogen oxides that may be present in the flue gas. Therefore, instead of using the condensate in the process, it will be blown down from the system. For the DCC to be effective, the temperature of the leaving water must always be lower than the AST.

DCC Water Pump GA-2102 A/B circulates most of the water leaving the bottom of the DCC back to the top of the direct contact cooler. However, before sending it back to the column the water stream is first filtered in DCC Water Filter FD-2101 and then cooled in DCC Water Cooler EA-2101 against the water from the new cooling tower. Temperature of the cooled water is controlled by a cascade loop, which maintains a constant flue gas exit temperature (Absorber feed temperature). Because of the relatively low cooling water temperature at the plant, the circulating water is cooled down to 35°C (95°F), which, in turn, easily cools the gas down to 46°C (115°F).

Filtration is necessary to remove any particulate matter that may enter the DCC in the flue gas. The blowdown is taken out after the filter but before the cooler and mixed into the return water of cooler EA-2101. This way the cooler does not have to handle the extra duty that would otherwise be imposed by the blowdown stream.

Absorption:

CO_2 Absorber DA-2102 (Refer to Figure 3-18):

From the DCC, the cooled flue gas enters the bottom of the CO_2 Absorber and flows up the tower counter current to a stream of 20 wt% MEA solution. The LAM enters the top of the column and heats up gradually as more and more CO_2 is absorbed. By the time the stream leaves the bottom of the tower, it has gained approximately 16°C (28°F). The tower has been designed to remove 96% of the CO_2 from the incoming gas. The CO_2 loading in LAM is 0.215 mol CO_2/mol MEA, while the loading of the rich amine leaving the bottom is 0.44 mol CO_2/mol MEA. These values are consistent with the values reported by Rochelle (2000).

To maintain water balance in the process, it is imperative that the temperature of the LAM feed be very close to that of the feed gas stream. Thus, with feed gas temperature fixed at 46°C (115°F), the temperature of the LAM stream must also be close to 46°C (115°F), preferably within 5.5°C (10°F). If the feed gas comes in at a higher temperature than the LAM, it brings in excess moisture, which condenses in the Absorber and becomes excess water. Unless this water is purged from the system, the concentration of MEA will decrease and the performance of the system will suffer. If on the other hand, the gas feed is colder than the LAM, it heats up in the tower and picks up extra moisture that is then carried out of the system by the vent gas. The result is a water deficiency situation because more water is removed than what comes into the system.

For the reasons explained above, it is essential that both the temperature of the flue gas and that of the LAM be accurately controlled. In fact, it is best to control one temperature and adjust the temperature of the other to maintain a fixed temperature difference. The design temperature difference is approximately 5.5°C (10°F). The LAM temperature was chosen to be the “master” and the gas temperature to be the “slave.”

The rich MEA solvent solution from the bottom of the absorber at 56°C (133°F) is heated to 95.5°C (204°F) by heat exchange with lean MEA solvent solution returning from the stripping column. The rich MEA solvent is then fed to the top of the stripping column. The lean MEA solvent solution, thus partially cooled to 62°C (143°F), is further cooled to 41°C (105°F) by exchange with cooling water and fed back to the absorber to complete the circuit.

CO_2 Absorber DA-2102 is a packed tower which contains two beds of structured packing and a third bed, the so-called “Wash Zone,” at the very top of the column. There is also a liquid distributor at the top of each bed. The distributors for the main beds are of high-quality design. There are several reasons for selecting structured packing for this service:

- Very low pressure drop which minimized fan horsepower
- High contact efficiency / low packing height
- Good tolerance for mal-distribution in a large tower
- Smallest possible tower diameter
- Light weight

At the bottom of the tower, there is the equivalent of a chimney tray, which serves as the bottom sump for the absorber. Instead of being flat like a typical chimney tray, it is a standard dished head with chimneys. The hold-up volume of the bottom sump is sufficient to accept all the liquid held up in the packing, both in the CO_2 absorber and in the Wash Zone. Rich Solvent Pump GA-2103 A/D takes suction from the chimney tray.

Absorber Wash Zone (Refer to Figure 3-18):

The purpose of the Wash Zone at the top of the tower is to minimize MEA losses, both due to mechanical entrainment and also due to evaporation. This is achieved by circulating wash water in this section to scrub most of the MEA from the lean gas exiting the Absorber. The key to minimizing MEA carryover is a mist separator pad between the wash section and the absorber. But, the demister cannot stop losses of gaseous MEA carried in the flue gas. This is accomplished by scrubbing the gas with counter current flow of water. Wash Water Pump GA-2101 takes water from the bottom of the wash zone and circulates it back to the top of the bed. The circulation rate has been chosen to irrigate the packing sufficiently for efficient operation.

The key to successful scrubbing is to maintain a low concentration of MEA in the circulating water. As the MEA concentration increases, the vapor pressure of MEA also increases and, consequently, higher MEA losses are incurred. Therefore, relatively clean water must be fed to the wash zone as make-up while an equal amount of MEA laden water is drawn out. A simple gooseneck seal accomplishes this and maintains a level in the chimney tray at the bottom of the wash section. Overflow goes to the main absorber. Make-up water comes from the overhead system of the Solvent Stripper.

The lean flue gas leaving the wash zone is released to atmosphere. The top of the tower has been designed as a stack, which is made high enough to ensure proper dispersion of the existing gas.

Rich/Lean Solvent Exchanger EA-2205 (Refer to Figure 3-19):

The Rich/Lean Solvent Exchange is a plate type exchanger with rich solution on one side and lean solution on the other. The purpose of the exchanger is to recover as much heat as possible from the hot lean solvent from the bottom of the solvent stripper by heating the rich solvent feeding the Solvent Stripper. This reduces the duty of the Solvent Stripper Reboiler. This exchanger is the single most important item in the energy economy of the entire CO_2 Recovery Unit. For this study, 5.5°C (10°F) approach was chosen to maximize the heat recovery. An air cooler (EC-2201) was added on the lean amine stream leaving the Solvent Stripper. This was to reduce the plot space requirement (compared to placing the air cooler downstream of the rich/lean exchanger) and overall cost of the project. A study was performed which determined that heat transfer via the plate frame lean/rich exchanger is relatively cheap, and thus justifies tight temperature approaches for the exchanger.

Stripping:

Solvent Stripper DA-2201 (Refer to Figure 3-19):

The solvent Stripper is a packed tower which contains two beds of structured packing and a third bed, also called “wash zone,” at the very top of the column. The purpose of the Solvent Stripper is to separate the CO_2 (contained in the rich solvent) from the bottom stream of the CO_2 Absorber that is feeding the stripper. As the solvent flows down, the bottom hot vapor from the reboiler continues to strip the CO_2 from the solution. The final stripping action occurs in the reboiler. The hot wet vapors from the top of the stripper contain the CO_2, along with water vapor and solvent vapor. Solvent Stripper CW Condenser (EA-2206) cools the overhead vapors, where most of the water and solvent vapors condense. The CO_2 does not condense. The condensed overhead liquid and gaseous CO_2 are separated in a reflux drum (FA-2201). CO_2 flows to the CO_2 purification section on pressure control and the liquid (called reflux) is returned via Solvent Stripper Reflux Pump (GA-2202 A/B) to the top bed in the stripper. The top bed of the stripper is a water wash zone designed to limit the amount of solvent (MEA) vapors entering the stripper overhead system.

Solvent Stripper Reboiler EA-2201 (Refer to Figure 3-19):

The steam-heated reboiler is a vertical shell-and-tube thermo-siphon type exchanger using inside coated high flux tubing proprietary of UOP. Circulation of the solvent solution through the reboiler is natural and is driven by gravity and density differences. The reboiler tube side handles the solvent solution and the shell side handles the steam. The energy requirement for the removal of CO_2 is about 2.36 tonnes of steam per tonne of CO_2 (2.6 tons of steam per ton of CO_2) for Case 5/Concept A.

Solvent Reclaimer EA-2203 (Refer to Figure 3-19):

The solvent Stripper Reclaimer is a horizontal heat exchanger. Certain acidic gases, present in the flue gas feeding the CO_2 absorber, form compounds with the MEA in the solvent solution that cannot be regenerated by application of heat in the solvent stripper reboiler. These materials are referred to as “Heat Stable Salts” (HSS). A small slipstream of the lean solvent from the discharge of the Solvent Stripper Bottoms Pump (GA-2201 A/B/C) is fed to the Solvent Reclaimer. The reclaimer restores the MEA usefulness by removing the high boiling and non-volatile impurities, such as HSS, suspended solids, acids, and iron products from the circulating solvent solution. Caustic is added into the reclaimer to free MEA up from its bond with sulfur oxides by its stronger basic attribute. This allows the MEA to be vaporized back into the circulating mixture, minimizing MEA loss. This process is important in reducing corrosion and fouling in the solvent system. The reclaimer bottoms are cooled (EA-2204) and are supplied to a tank truck without any interim storage.

Solvent Stripper Condenser EA-2206 (Refer to Figure 3-19):

EA-2206 is a water-cooled shell and tube exchanger. The purpose of the condenser is to completely condense all components contained in the overhead vapor stream that can condense under the operating conditions, with the use of cooling water as the condensing medium. Components that do not condense include nitrogen, carbon dioxide, oxygen, nitrogen oxides, and carbon monoxide. The water vapor and MEA solvent vapor will condense and the condensed water will dissolve some carbon dioxide. This exchanger uses cooling water capacity freed up due to the reduced load on the existing surface condensers of the power plant. The same is true for the lean solvent cooler (EA-2202).

Solvent Stripper Reflux Drum, FA-2201 (Refer to Figure 3-19):

The purpose of the reflux drum is to provide space and time for the separation of liquid and gases, provide liquid hold-up volume for suction to the reflux pumps, and provide surge for the pre-coat filter. The separation is not perfect, as a small amount of carbon dioxide is left in the liquid being returned to the stripper. The CO_2, saturated with water, is routed to the CO_2 compression and liquefaction system.

Solvent Stripper Reflux Pump, GA-2202 (Refer to Figure 3-19):

This pump takes suction from the reflux drum and discharges on flow control to the stripper top tray as reflux.

Solvent Filtration Package, PA-2251, (Refer to Figure 3-19):

Pre-coat Filter PA-2251 is no ordinary filter; it is a small system. The main component is a pressure vessel that has a number of so called “leaves” through which MEA flows. The leaves have a thin (1/8 inch) coating of silica powder, which acts to filer off any solids. For the purposes of such application the powder is called “filter aid.”

To cover the leaves with the filter aid, the filter must be “pre-coated” before putting it into service. This is accomplished by mixing filter aid in water at a predetermined ratio (typically 10 wt%) to prepare slurry. This takes place in an agitated tank. A pump, which takes it suction from this tank, is then operated to pump the slurry into the filter. Provided the flow rate is high enough, the filter aid is deposited on the leaves while water passes through and can be recycled back to the tank. This is continued until the water in the tank becomes clear, indicating that all the filter aid has been transferred.

The volume of a single batch in the tank is typically 125% of the filter volume because there must be enough to fill the vessel and have some excess left over so the level in the tank is maintained and circulation can continue. In this design, water from the Stripper overhead will be used as make-up water to fill the tank. This way, the water balance of the plant is not affected.

During normal operation, it is often beneficial to add so-called “body” which is the same material as the pre-coat but may be of different particle size. The body is also slurried in water but is continually added to the filter during operation. This keeps the filter coating porous and prevents rapid plugging and loss of capacity. As the description suggests, an agitated tank is needed to prepare the batch. A metering pump is then used to add the body at a prescribed rate to the filter.

When the filter is exhausted (as indicated by pressure drop), it is taken off line so the dirty filter aid can be removed and replaced with fresh material. To accomplish this, the filter must be drained. Pressurizing the filter vessel with nitrogen and pushing the MEA solution out of the filter accomplishes this. After this step, the filter is depressurized. Then, a motor is started to rotate the leaves so a set of scrapers will wipe the filter cake off the leaves. The loosened cake then falls off into a conveyor trough in the bottom of the vessel. This motor-operated conveyor then pushes the used cake out of the vessel and into a disposal container (oil drum or similar). The rejected cake has the consistency of toothpaste. This design is called “dry cake” filter and minimizes the amount of waste produced.

For this application, some 2% of the circulating MEA will be forced to flow through the filter. In fact, Filter Circulating Pump GA-2203 draws the liquid through the filter as it has been installed downstream of the filter. The advantage of placing the pump on the outlet side of the filter is reduced design pressure of the filter vessel and associated piping. In spite of the restriction on its suction side, ample NPSH is still available for the pump. Flow is controlled on the downstream side of the pump.

Corrosion Inhibitor (Refer to Figure 3-19):

Corrosion inhibitor chemical is injected into the process constantly to help control the rate of corrosion throughout the CO_2 recovery plant system. Since rates of corrosion increase with high MEA concentrations and elevated temperatures, the inhibitor is injected at appropriate points to minimize the corrosion potential. The inhibitor is stored in a tank (Part of the Package, not shown) and is injected into the system via injection pump (Part of the Package, not shown). The pump is a diaphragm-metering pump.

The selection of metallurgy in different parts of the plant is based on the performance feedback obtained from our similar commercial units in operation over a long period of time.

CO_2 Compression, Dehydration, and Liquefaction:

(Refer to Figure 30-20):

CO_2 from the solvent stripper reflux drum, GA-2201, is saturated with water, and is compressed in a three-stage centrifugal compressor using the air and cooling water from the new cooling tower for interstage and after-compression cooling. The interstage coolers for first and second stage are designed to supply 35°C (95°F) CO_2 to the compressor to minimize the compression power requirements.

Most of the water in the wet CO_2 stream is knocked out during compression and is removed from intermediate suction drums. A CO_2 drier is located after the third stage compressor to meet the water specifications for the CO_2 product. The water-free CO_2 is liquefied after the third stage of compression at about 13.4 barg (194 psig) pressure by transferring heat to propane refrigerant. CO_2 is then pumped (GA-2301) to the required battery limit pressure of 138 barg (2000 psig).

The propane refrigeration system requires centrifugal compressors, condensers, economizers, and evaporators to produce the required cold. The centrifugal compressor is driven by an electric motor and is used to raise the condensing temperature of the propane refrigerants above the temperature of the available cooling medium. The condenser is used to cool and condense the discharged propane vapor from the compressor back to liquid form. The economizer, which improves the refrigerant cycle efficiency, is designed to lower the temperature of the liquid propane by flashing or heat exchange. The evaporator liquefies the CO_2 vapor by transferring heat from the CO_2 vapor stream to the boiling propane refrigerant.

Drying:

CO_2 Drier, FF-2351 (Refer to Figure 3-20):

The purpose of the CO_2 drier is to reduce the moisture content of the CO_2 product to less than 20 ppmv to meet pipeline transport specifications. The drier package, FF-2351, includes four drier vessels, three of which are in service while one is being regenerated or is on standby. The package also includes a natural gas fired regeneration heater and a cooled regeneration cooler. The exchanger will have a knock out cooler downstream for separating the condensed water. The drier used as a basis for cost estimation is good for a 10-hour run length based on 3A molecular sieve.

The drier is located on the discharge side of the third stage of the CO_2 Compressor. Considering the cost of the vessel and the performance of the desiccant, this is the location favored by vendors. The temperature of the CO_2 stream entering the drier is 32°C (90°F).

Once a bed is exhausted, it is taken off line, and a slipstream of effluent from the online beds is directed into this drier after being boosted in pressure by a compressor. Before the slipstream enters the bed that is to be regenerated, it is heated to a high temperature. Under this high temperature, moisture is released from the bed and carried way in the CO_2 stream. The regeneration gas is then cooled to the feed gas temperature to condense any excess moisture. After this, the regeneration gas stream is mixed with the feed gas upstream of the third-stage knockout drum.

All the regeneration operations are controlled by a PLC that switches the position of several valves to direct the flow to the proper drier. It also controls the regeneration compressor, heater, and cooler. Because the regeneration gas has the same composition as the feed gas, it also contains some moisture. Thus, it is primarily the heat (“temperature swing”) that regenerates the bed.

Process Flow Diagrams:

The processes described above are illustrated in the following process flow diagrams:

- Figure 3-18: Drawing D 09484-01001R-0: Flue Gas Cooling and CO_2 Absorption
- Figure 3-19: Drawing D 09484-01002R-0: Solvent Stripping
- Figure 3-20: Drawing D 09484-01003R-0: CO_2 Compression and Liquefaction

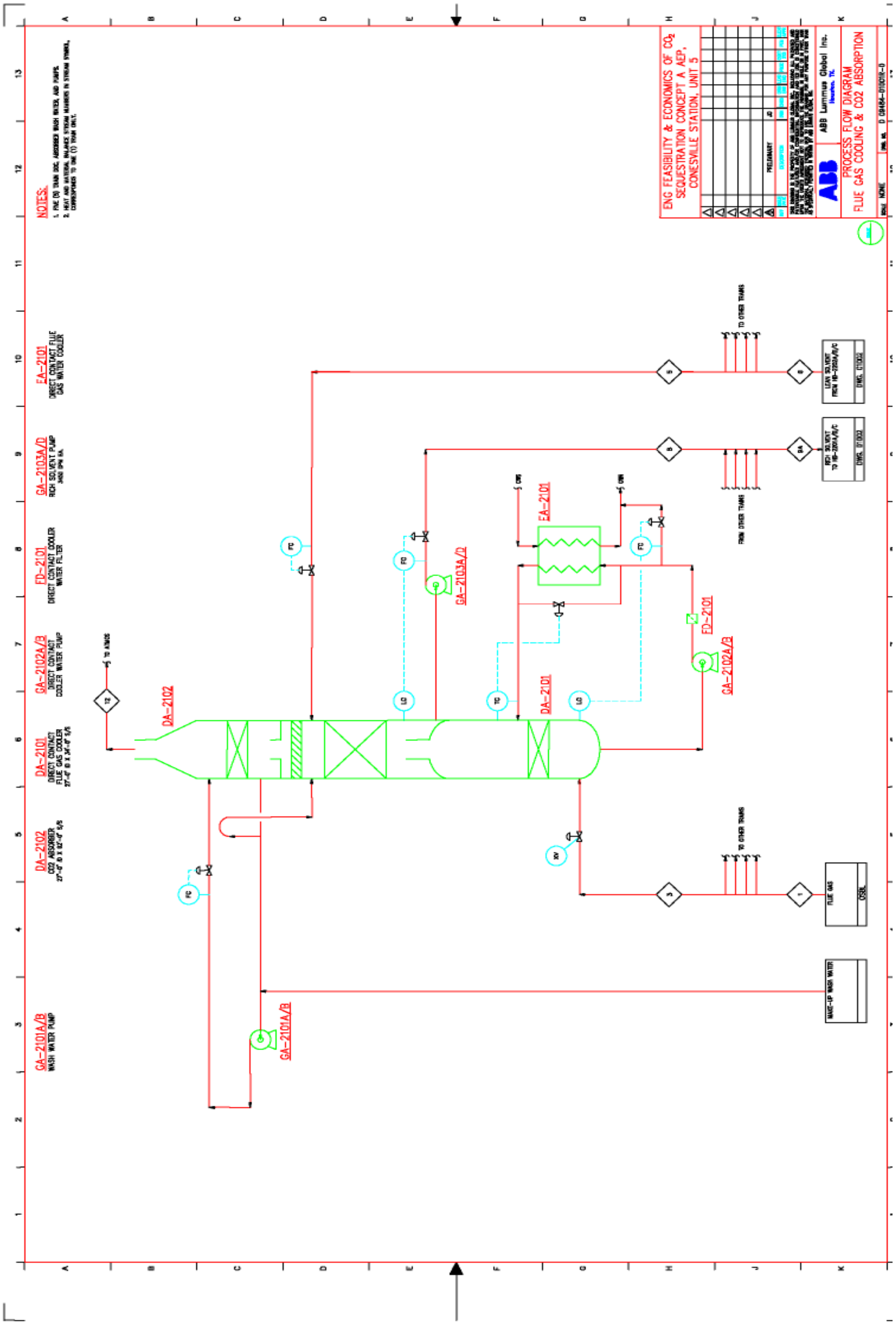


Figure 3-18: Process Flow Diagram for Case 5/Concept A: Flue Gas Cooling and CO_2 Absorption

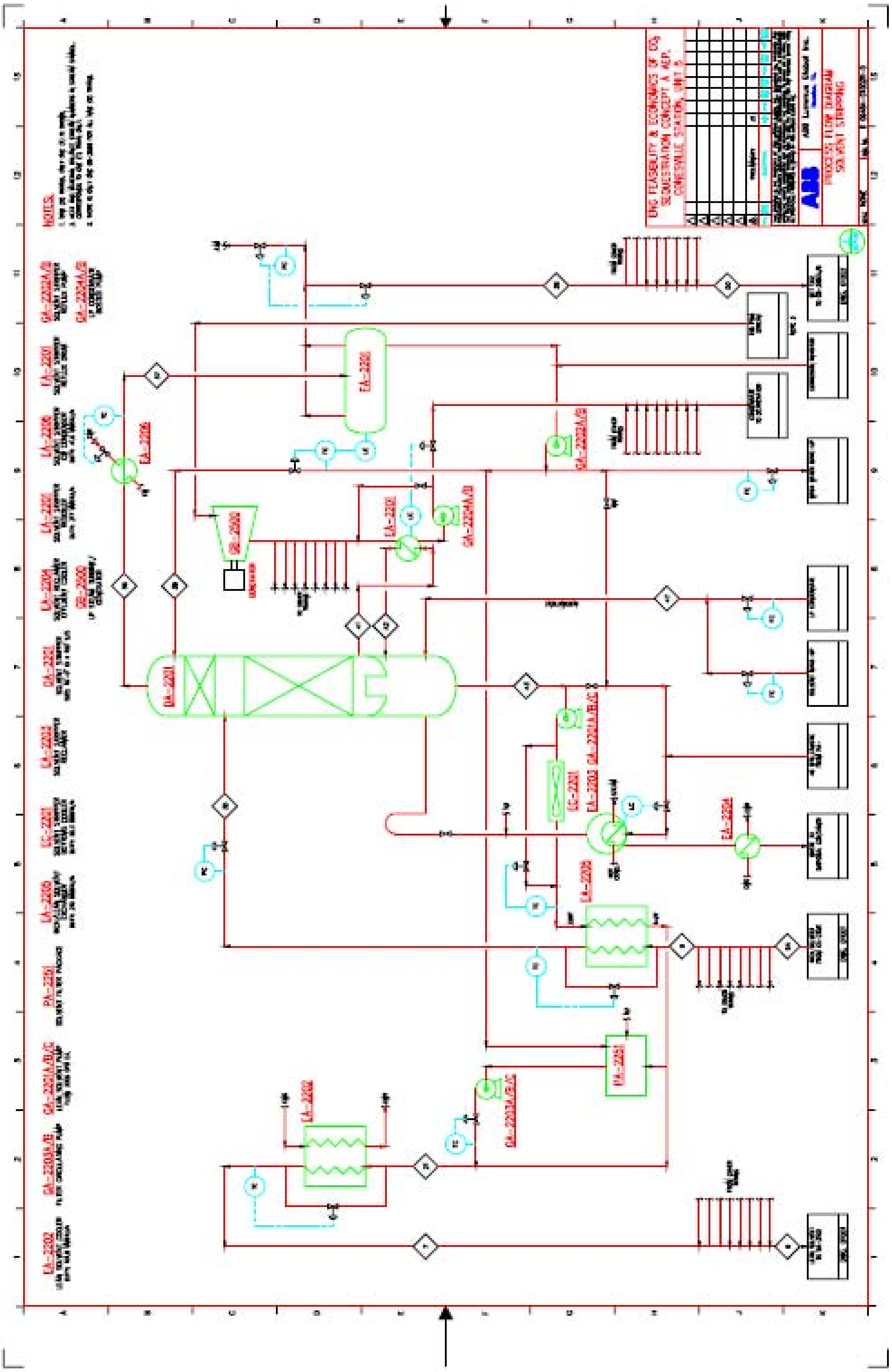


Figure 3-19: Process Flow Diagram for Case 5/Concept A: Solvent Stripping

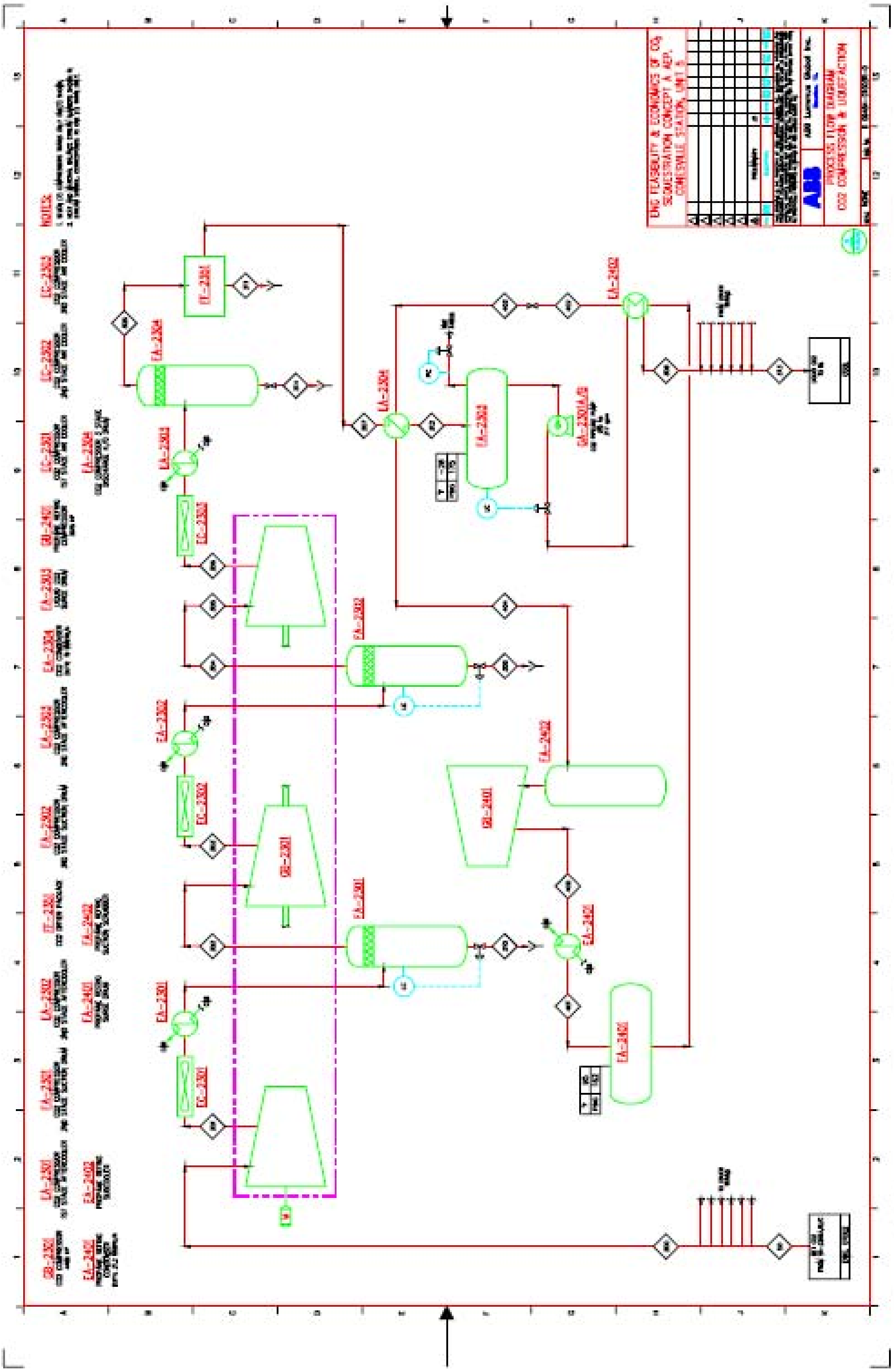


Figure 3-20: Process Flow Diagram for Case 5/Concept A: CO_2 Compression, Dehydration, and Liquefaction

3.1.5.2 Case 5/Concept A Overall Material and Energy Balance - CO_2 Removal, Compression, and Liquefaction System

The material balances (Table 3-37 and Table 3-38) were run on two process simulators: Hysim and Amsim. Amsim was used for the Absorption/Stripping systems while Hysim was used for the conventional systems as follows:

Flue Gas feed	Hysim
Absorber and Stripper	Amsim
Compression	Hysim

The two simulators use a different reference enthalpy. They also use slightly different calculation methods for determining water saturation quantities. There is no simple way to normalize the enthalpies to the same reference. Thus, the enthalpies given in the balance are the values copied directly from the simulation. This creates a discontinuity at the interface between Hysim and Amsim simulations - take for example the wet CO_2 flow to the CO_2 compressor. The stream comes from the Stripper overhead system, which was simulated with Amsim and enters the CO_2 compressor, which was simulated using Hysim. For this particular stream, the enthalpy value given in the balance comes from Hysim. Lastly, convergence algorithms allow the programs to slightly alter input streams. Thus, some leniency and care should be exercised when using such interface streams for heat balance checks. This section contains heat and material balances for Case 5/Concept A.

Table 3-37: Material and Energy Balance for Case 5/Concept A Amine System

TEMPERATURE F	150	115	115	115	115	105	133	106	133	133	133	204
PRESSURE PSIA	16.5	16.5	16.5	16.5	16.5	14.9	16.5	14.9	16.5	16.5	16.5	16.5
COMPONENTS												
CO2 (Carbon Dioxide) LbMol/HR	19,684.00	3,936.80	3,936.80	3,936.23	0.14	3,585.44	7,380.58	141.1	36,902.89	4,100.32	4,100.32	4,100.32
MEA LbMol/HR	0	0	0	0	0	16,765.89	16,763.07	2.82	83,815.36	9,312.82	9,312.82	9,312.82
H2O (Water) LbMol/HR	24,551.00	4,910.20	4,910.20	2,544.80	2,365.50	227,379.00	228,257.60	1,666.30	1,141,288.00	126,809.80	126,809.80	126,809.80
C1 (Methane) LbMol/HR	0	0	0	0	0	0	0	0	0	0	0	0
N2 (Nitrogen) LbMol/HR	105,079.00	21,015.80	21,015.80	21,016.14	0.02	0	1.75	21,014.40	8.76	0.97	0.97	0.97
O2 (Oxygen) LbMol/HR	4,518.00	903.6	903.6	903.61	0	0	0.14	903.47	0.7	0.08	0.08	0.08
Total Molar Flow Rate LbMol/HR	153,832.00	30,766.40	30,766.40	28,400.80	2,365.60	247,730.40	252,403.20	23,728.10	1,262,016.00	140,224.00	140,224.00	140,224.00
VAPOR												
MASS FLOW RATE LbMol/HR	446,600,625	3,572,805	3,572,805	3,397,068				2,438,328				
STD. VOL. FLOW RATE MMSCFD	1401.1	280.22	280.22	258.66				216.1				
ACTUAL VOL. FLOW RATE MMACFD	1378	275.6	275.6	254.5				231.72				
MOLECULAR WEIGHT MW	285.821	57.1642	57.1642	58.9234				55.1246				
STD. DENSITY Lb/Ft3	0.765	0.153	0.153	0.1576				0.1354				
GAS COMPRESSIBILITY	0	0	0	0				0				
VISCOSITY cP	0	0	0	0				0				
HEAT CAPACITY Btu/Lb-F	0	0	0	0				0				
THERMAL CONDUCTIVITY Btu/Hr-ft-F	127.958	25.5916	25.5916	27.7192				1.1892				
LIQUID												
MASS FLOW RATE Lb/Hr					85,263	10,557,848	10,923,302		273,082,551	3,371,390	3,371,390	3,371,390
STD. VOL. FLOW RATE GPM					85.26	10252.78	10352.54		51762.7	5751.41	5751.41	5751.41
ACTUAL VOL. FLOW RATE GPM					86.02	10308.54	10467.22		52336.1	5815.12	5815.12	5940.3
MOLECULAR WEIGHT MW					18.02	21.31	21.64		21.64	21.64	21.64	21.64
STD. DENSITY Lb/Ft3					62.34	64.19	65.77		65.77	65.77	65.77	65.77
VISCOSITY cP					0.6383	0.8608	0.6868		0.6868	0.6868	0.6868	0.3544
HEAT CAPACITY Btu/Lb-F					0.9948	0.9357	0.9221		0.9221	0.9221	0.9221	0.9325
THERMAL CONDUCTIVITY Btu/Hr-ft-F					0.3979	0.3557	0.3557		0.3557	0.3557	0.3557	0.3557

STREAM NAME	Rich Amine Feed to Regenerator	Regenerator Overhead Vapor	RegeneratorCon de nser Outlet	Acid Gas	RegeneratorReflux Liquid	Liquid to Regenerator Reboiler	RegeneratorRebo il er Vapor	Lean Amine from RegeneratorReboi l er	Lean Amine from Lean/Rich Heat Exchanger	Lean Amine to Cooler	Amine and Water Make-up	Total Acid Gas
STREAM NO.	35	36	37	38	39	41	42	43	21	21	47	24
LIQUID FRACTION	1	0	1	0	1	1	0	1	1	1	1	0
TEMPERATURE F	209	209	105	105	105	248	250	250	173	173	68	105
PRESSURE PSIA	28	26	23	23	23	29.8	30	30	30	30	30	23
COMPONENTS												
CO2 (Carbon Dioxide) LbMol/Hr	4,100.32	2,081.06	2,081.06	2,079.81	1.27	2,701.12	680.61	2,020.51	2,020.51	2,020.51	0	18,718.28
MEA	9,312.82	9.92	9.92	0.01	9.9	9,381.40	68.6	9,312.81	9,312.81	9,314.38	1.58	0.11
H2O (Water) LbMol/Hr	126,809.80	2,128.70	2,128.70	105.7	2,023.00	137,717.90	11,013.80	126,704.00	126,704.00	126,321.80	-382.3	951.3
C1 (Methane) LbMol/Hr	0	0	0	0	0	0	0	0	0	0	0	0
N2 (Nitrogen) LbMol/Hr	0.97	0.97	0.97	0.97	0	0	0	0	0	0	0	8.76
O2 (Oxygen)	0.08	0.08	0.08	0.08	0	0	0	0	0	0	0	0.7
Total Molar Flow Rate	140,224.00	4,220.70	4,220.70	2,186.60	2,034.10	149,800.30	11,763.00	138,037.30	138,037.30	137,656.70	-380.7	19,679.20
VAPOR												
MASS FLOW RATE Lb/Hr		221,688		166,131			429,305					121,109,333
STD. VOL. FLOW RATE MMSCFD		38.44		19.91			107.13					179.2
ACTUAL VOL. FLOW RATE MMACFD		27.73		13.72			70.62					123.5
MOLECULAR WEIGHT MW		34.37		47.5			21.97					427.46
STD. DENSITY Lb/Ft3		0.12		0.18			0.09					1.62
GAS COMPRESSIBILITY		0		0			0					0
VISCOSITY cP		0		0			0					0
HEAT CAPACITY Btu/Lb-F		0		0			0					0
THERMAL CONDUCTIVITY Btu/Hr-ft-F		54.78		105.69			6.43					951.17
LIQUID												
MASS FLOW RATE Lb/Hr	3,371,390		145,088		41,234	3,525,978		3,267,542	3,267,542	3,259,998	-7,547	
STD. VOL. FLOW RATE MMSCFD	5751.41		247.18		73.13	6116.13		5709.78	5709.78	5696.53	-13.59	
ACTUAL VOL. FLOW RATE MMACFD	5951.79		248.73		73.61	6434.23		6011.14	5839.38	5825.79	-13.6	
MOLECULAR WEIGHT MW	21.64		30.94		18.24	21.18		21.3	21.3	21.31	17.84	
STD. DENSITY Lb/Ft3	65.77		65.86		63.27	64.69		64.21	64.21	64.21	62.31	
VISCOSITY cP	0.3401		0.6888		0.6655	0.2592		0.2564	0.4548	0.4549	1.2839	
HEAT CAPACITY Btu/Lb-F	0.9324		0.4962		0.9902	0.9481		0.9491	0.9513	0.9513	0.9454	
THERMAL CONDUCTIVITY Btu/Hr-ft-F	0.3557		0.3945		0.3944	0.3583		0.3557	0.3557	0.3557	0.3664	

Table 3-38: Material and Energy Balance for Case 5/Concept A CO_2 Compression, Dehydration and Liquefaction System

STREAM NAME	Total Acid gas from strippers	To train A liquefaction	First stage discharge	To second stage	First stage water KO	2nd stage discharge	To 3rd stage	2nd stage water KO	From 3rd stage	To drier	3rd stage water KO
STREAM NO.	**300**	**300**	**301**	**302**	**310**	**303**	**304**	**309**	**306**	**305**	**314**
VAPOR FRACTION Molar	1.000	1.000	1.000	1.000	0.000	1.000	1.000	0.000	1.000	1.000	0.000
TEMPERATURE F	105	105	230	95	95	236	95	95	282	90	90
PRESSURE PSIG	4	4	25	19	19	62	56	56	191	185	185
MOLAR FLOW RATE LbMol/Hr	19,679.08	2 ,811.30	2,811.30	2,743.70	67.60	2,743.70	2,708.50	35.19	2,708.50	2,686.56	21.94
MASS FLOW RATE Lb/Hr	841,192	120,170	120,170	118,951	1,219	118,951	118,315	636	118,315	117,917	398
ENERGY Btu/Hr	8.79E+07	1.26E+07	1.58E+07	1.58E+07	-9.79E+05	1.56E+07	1.17E+07	-5.09E+05	1.64E+07	1.10E+07	-3.18E+05
COMPOSITON Mol %											
CO_2	95.12%	95.12%	95.12%	97.46%	0.09%	97.46%	98.72%	0.18%	98.72%	99.52%	0.54%
H_2O	4.83%	4.83%	4.83%	2.49%	99.91%	2.49%	1.23%	99.82%	1.23%	0.42%	99.46%
Nitrogen	0.04%	0.04%	0.04%	0.05%	0.00%	0.05%	0.05%	0.00%	0.05%	0.05%	0.00%
Ammonia	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Oxygen	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
VAPOR											
MOLAR FLOW RATE LbMol/Hr	19,679.10	2,811.30	2,811.30	2,743.70	--	2,743.70	2,708.50	--	2,708.50	2,686.60	--
MASS FLOW RATE Lb/Hr	841,192	120,170	120,170	118,951	--	118,951	118,315	--	118,315	117,917	--
STD VOL. FLOW MMSCFD	179.23	25.6	25.6	24.99	-	24.99	24.67	-	24.67	24.47	-
ACTUAL VOL. FLOW ACFM	103,907.68	14,843.95	8,749.53	8,063.83	-	4,417.63	3,728.32	-	1,698.44	1,224.03	-
MOLECULAR WEIGHT MW	42.75	42.75	42.75	43.35	-	43.35	43.68	-	43.68	43.89	-
DENSITY Lb/Ft3	0.13	0.13	0.23	0.25	--	0.45	0.53	--	1.16	1.61	--
VISCOSITY cP	0.0149	0.0149	0.0187	0.0149	--	0.0193	0.0152	--	0.0212	0.0154	--
HYDROCARBON LIQUID											
MOLAR FLOW RATE LbMol/Hr	--	--	--	--	--	--	--	--	--	--	--
MASS FLOW RATE Lb/Hr	--	--	--	--	--	--	--	--	--	--	--
STD VOL. FLOW MMSCFD	-	-	-	-	-	-	-	-	-	-	-
ACTUAL VOL. FLOW ACFM	-	-	-	-	-	-	-	-	-	-	-
MOLECULAR WEIGHT MW	-	-	-	-	-	-	-	-	-	-	-
DENSITY Lb/Ft3	-	-	-	-	-	-	-	-	-	-	-
VISCOSITY cP	-	-	-	-	-	-	-	-	-	-	-
SURFACE TENSION Dyne/Cm	-	-	-	-	-	-	-	-	-	-	-

STREAM NAME	From drier / To condenser	Water from drier	From condenser	From product pump	From Train A liquefaction	To pipeline	Refrig compressor discharge	From refrig condenser	From subcooler	Refrig to CO_2 condenser	Refrig from CO_2 condenser
STREAM NO.	**307**	**311**	**312**	**308**	**309**	**313**	**400**	**401**	**402**	**403**	**404**
VAPOR FRACTION Molar	1.000	0.726	0.000	0.000	0.000	0.000	1.000	0.000	0.000	0.173	0.996
TEMPERATURE F	90	380	-26	-12	82	82	149	95	24	-31	-31
PRESSURE PSIG	180	180	2,003	2,000	2,000	2,000	169	162	159	5	5
MOLAR FLOW RATE LbMol/Hr	2,675.15	11.41	2,675.15	2,675.15	2,675.15	18,726.05	2,928.57	2,928.57	2,928.57	2,928.57	2,928.57
MASS FLOW RATE Lb/Hr	117,711	206	117,711	117,711	117,711	823,979	129,141	129,141	129,141	129,141	129,141
ENERGY Btu/Hr	1.10E+07	2.51E+04	-8.07E+06	-7.29E+06	-1.36E+06	-9.50E+06	1.81E+07	7.63E+05	-5.17E+06	-5.17E+06	1.39E+07
COMPOSITON Mol %											
CO_2	99.95%	0.00%	99.95%	99.95%	99.95%	99.95%	0.00%	0.00%	0.00%	0.00%	0.00%
H_2O	0.00%	100.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Nitrogen	0.05%	0.00%	0.05%	0.05%	0.05%	0.05%	0.00%	0.00%	0.00%	0.00%	0.00%
Ammonia	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
Propane	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	100.00%	100.00%	100.00%	100.00%	100.00%
Oxygen	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%	0.00%
VAPOR											
MOLAR FLOW RATE LbMol/Hr	2 ,675.2	8 .3	--	--	--	--	2,928.6	--	--	506.5	2 ,915.8
MASS FLOW RATE Lb/Hr	1 17,711	1 49	--	--	--	--	129,141	--	--	22,334	1 28,577
STD VOL. FLOW MMSCFD	2 4.36	0 .08	--	--	--	--	2 6.67	--	--	4 .61	2 6.56
ACTUAL VOL. FLOW ACFM	1 ,253.44	5 .96	--	--	--	--	3,573.03	--	--	1,860.34	10,709.92
MOLECULAR WEIGHT MW	4 4.00	1 8.02	--	--	--	--	44.10	--	--	4 4.10	4 4.10
DENSITY Lb/Ft3	1.57	0 .42	--	--	--	--	0.6	--	--	0 .20	0.2
VISCOSITY cP	0 .0155	0 .0154	--	--	--	--	0.0082	--	--	0 .0065	0.0065
HYDROCARBON LIQUID											
MOLAR FLOW RATE LbMol/Hr	--	--	2,675.15	2,675.15	2,675.15	18,726.05	--	2,928.57	2,928.57	2,422.10	12.79
MASS FLOW RATE Lb/Hr			117,711.33	117,711.33	117,711.33	823,979.29	--	129,141.22	129,141.22	106,807.22	563.95
STD VOL. FLOW MMSCFD	-	-	9,766	9,766	9,766	68,360	--	17,452	17,452	14,434	76
ACTUAL VOL. FLOW ACFM	-	-	217.05	213.53	289.79	2,028.56	--	541.52	480.49	372.27	1.97
MOLECULAR WEIGHT MW	-	-	67.61	68.73	50.64	50.64	--	29.73	33.51	35.77	35.77
DENSITY Lb/Ft3	-	-	44	44	44	44	--	44.1	44.1	44.1	44.1
VISCOSITY cP	-	-	0.1752	0.1607	0.062	0.062	--	0.0906	0.1332	0.1823	0.1823
SURFACE TENSION Dyne/Cm	-	-	16.07	14.07	0.86	0.86	--	5.74	10.51	14.49	14.49

3.1.5.3 Case 5/Concept A Equipment List – CO_2 Removal, Compression, and Liquefaction

Complete equipment data summary sheets for Case 5/Concept A are provided in Appendix II. These equipment lists have been presented in the so-called “short spec” format, which provides adequate data for developing a factored cost estimate.

It should be noted that although Cases 1 and 5 both capture about the same amount of CO_2 (90% and 96% respectively), the design of Case 5 (See Bozzuto et al., 2001), which was developed in 2000, is not totally consistent with the design of Case 1 done in the current study. Table 3-39, which summarizes the major equipment categories for Case 1 and 5, shows that Case 1 uses two absorber trains, two stripper trains, and two compression trains. Case 5, which was designed in 2000, uses five absorber trains, nine stripper trains, and seven compression trains. Additionally, the total number of heat exchangers in the system for Case 1 is 58 whereas for Case 5 is 131. Because of these differences, Case 1 is able to take advantage of significant economy of scale effects for equipment cost with the larger equipment sizes in each train as compared to Case 5. Additionally, Case 5 equipment was all located about 457 m (1,500 ft) from the Unit #5 stack, which also increased the costs of Case 5 relative to Case 1.

Table 3-39: Equipment Summary CO_2 Removal, Compression, and Liquefaction System (Cases 1, 5)

	Case 1 (90% recovery)		**Case 5 (96% recovery)**	
Compressors	No.	HP each	No.	HP each
CO_2 Compressor	2	15,600	7	4,500
Propane Compressor	2	11,700	7	3,100
LP Let Down Turbine	1	60,800	1	82,300
Towers/Internals	No.	ID/Height (ft)	No.	ID/Height (ft)
Absorber/Cooler	2	34 / 126	5	27 / 126
Strippers	2	22 / 50	9	16 / 50
Heat Exchangers	No.	10^6-Btu/Hr ea.	No.	10^6-Btu/Hr ea.
Reboilers	10	120.0	9	217.0
Solvent Stripper CW Condenser	12	20.0	9	42.0
Other Heat Exchangers / Avg Duty	36	61.0	113	36.0
Total Heat Exchangers / Avg Duty	58	101.0	131	56.6

3.1.5.4 Case 5/Concept A Consumption of Utilities - CO_2 Removal, Compression, and Liquefaction System

The following utilities from OSBL are required in the CO_2 Recovery Unit.

- Steam
- High Pressure (HP) Steam
- Low Pressure (LP) Steam
- Water
- Demineralized Water
- Raw Water (Fresh Water, Cooling tower make-up)

- Potable Water (hoses, etc.)
- Air
- Plant Air (maintenance, etc.)
- Instrument Air
- Electric Power
- Natural Gas

Note: The CO_2 Recovery Plant includes cooling water pumps that supply all the cooling water required by this unit. Case 5/Concept A utility consumption is presented in Table 3-40 and the auxiliary power consumption is shown in Table 3-41.

Table 3-40: Utility Consumption for Case 5/Concept A

Utility	Amount Consumed	Units
Natural Gas	0.42	10^6 SCFD
Steam (180 psig)	1,950,000	Lb/hr
Cooling water	22,000	Gpm

Table 3-41: Auxiliary Power Usage for Case 5/Concept A

Number of Trains	Tag no.	Description	Number Operating per train	Power (ea) w/ 0.95 motor eff (kW)	Total all trains (kW)
5	GA-2101 A/B	Wash Water Pump	1	19	95
5	GA-2102 A/B	Direct Contact Cooler Water Pump	1	32	162
5	GA-2103 A/B/C/D	Rich Solvent Pump	3	146	729
9	GA-2201 A/B/C	Lean Solvent Pump	2	117	1,053
9	GA-2202 A/B	Solvent Stripper Reflux Pump	1	3	28
9	GA-2203 A/B	Filter Circ. Pump	1	12	107
7	GA-2301 A/B	CO_2 Pipeline Pump	1	184	1,288
9	GA-2204 A/B	LP condensate booster pump	1	74	667
3	GA-2501	Caustic metering pump	1	0	0
7	GB-2301	CO_2 Compressor (Motor driven)	1	3,557	24,901
7	GB-2401	Propane Refrigeration Compressor	1	2,395	16,765
1	GB-2500	LP steam turbine/ generator	NA	NA	NA
7	EC-2301	CO_2 Compressor 1st stage Air Cooler	1	9	66
7	EC-2302	CO_2 Compressor 2nd stage Air Cooler	1	10	69
7	EC-2303	CO_2 Compressor 3rd stage Air Cooler	1	15	103
9	EC-2201	Solvent Stripper Bottoms Cooler	1	256	2,305
7	PA-2351	CO_2 Drier Package	1	151	1054
1	PA-2551	Cooling Tower	1	962	962
		Total Power			**50,355**

3.1.5.5 Case 5/Concept A Consumption of Chemicals and Desiccants - CO_2 Removal, Compression, and Liquefaction System

The consumption of chemicals and desiccants for Case 5/Concept A are identified in Table 3-42.

Table 3-42: Chemicals and Desiccants Consumption for Case 5/Concept A

Chemical	Consumption per day (lbm)
Caustic (100%)	3600
MEA	14000
Corrosion inhibitor	1140
Diatomaceous earth	916
Molecular sieve	257
Sodium hypochlorite	3590
Sodium bisulfite	13.8

This total does not include chemicals provided by the cooling tower service people nor disposal of waste. These are handled as a component of operating costs referred to as contracted services and waste handling, respectively.

3.1.5.6 Case 5/Concept A Design Considerations - CO_2 Removal, Compression, and Liquefaction System

The following parameters were optimized for Case 5/Concept A with the objective of reducing the overall unit cost and energy requirements.

- Solvent Concentration
- Lean Amine Loading
- Rich Amine Loading
- Absorber Temperature
- Rich /Lean Exchanger approach
- CO_2 Compressor interstage temperatures
- CO_2 Refrigeration Pressure and Temperature

A minimum of 90% CO_2 recovery was targeted. The above parameters were adjusted to increase the recovery until a significant increase in equipment size and/or energy consumption was observed. AES Corporation owns and operates a 200 STPD food grade CO_2 production plant in Oklahoma. This plant was designed and built by ABB Lummus Global as a part of the larger power station complex using coal-fired boilers. This plant started up in 1990 and has been operating satisfactorily with lower than designed MEA losses. The key process parameters from the present design for Case 5/Concept A are compared with those from the AES plant (Barchas and Davis, 1992) in Table 3-43.

Table 3-43: Key Process Parameters Comparison for Case 5/Concept A

PROCESS PARAMETER	AEP DESIGN (Case 5/Concept A)	AES DESIGN
Plant Capacity (TPD)	9,888	200
CO_2 in Feed, (% mol)	13.9	14.7
O_2 in Feed, (% mol)	3.2	3.4
SO_2 in Feed, (ppmv)	10 (Max)	10 (Max)
Solvent	MEA	MEA
Solvent Conc. (wt%)	20	15 (Actual 17-18 wt%)
Lean Loading (mol CO_2 / mol MEA)	0.21	0.10
Rich Loading (mol CO_2 / mol MEA)	0.44	0.41
Stripper Feed Temperature, °F	210	194
Stripper Bottom Temperature, °F	250	245
Feed Temperature to Absorber, °F	105	108
CO_2 Recovery, %	96	90 (Actual 96%-97%)
Absorber Pressure Drop, psi	1	1.4
Stripper Pressure Drop, psi	0.6	4.35
R/L Exchanger Approach, °F	10	50
CO_2 Compressor I/STG Temperature, °F	105	115
Liquid CO_2 Temperature, °F	82	-13
Steam Consumption, lbm steam/ lbm CO_2	2.6	3.45
Liquid CO_2 Pressure (psia)	2,015	247

3.1.5.7 Case 5/Concept A OSBL Systems - CO_2 Removal, Compression, and Liquefaction System

Reclaimer Bottoms (Case 5/Concept A):

The reclaimer bottoms are generated during the process of recovering MEA from heat stable salts (HSS), which are produced from the reaction of MEA with SO_2 and NO_2. The HSS accumulate in the reclaimer during the lean amine feed portion of the reclaiming cycle. The volume of reclaimer bottoms generated will depend on the quantity of SO_2 and NO_2 that is not removed in the Flue Gas Scrubber. A typical composition of the waste is presented in Table 3-44.

Table 3-44: Reclaimer Bottoms Composition for Case 5/Concept A

MEA	9.5 wt%
NH_3	0.02 wt%
NaCl	0.6 wt%
Na_2SO_4	6.6 wt%
Na_2CO_3	1.7 wt%
Insolubles	1.3 wt%
Total Nitrogen	5.6 wt%
Total Organic Carbon	15.6 wt%
H_2O	59.08 wt%
pH	10.7
Specific Gravity	1.14

Filter Residues:

A pressure leaf filter filters a slipstream of lean amine. Diatomaceous earth is used as a filter-aid for pre-coating the leaves and as a body feed. Filter cycles depend on the rate of flow through the filter, the amount of filter aid applied, and the quantity of contaminants in the solvent. A typical composition of the filter residue is provided in Table 3-45. These will be disposed of by a contracted service which hawls away the drums of spent cake.

Table 3-45: Filter Residue Composition for Case 5/Concept A

MEA	2.5 wt%
Total Organic Carbon	1.5 wt%
SiO_2	43 wt%
Iron Oxides	32 wt%
Aluminum Oxides	15 wt%
H_2O	6 wt%
pH	10.0
Specific Gravity	1.0

Excess Solvent Stripper Reflux Water:

The CO_2 Recovery Facility has been designed to operate in a manner to avoid accumulation of water in the Absorber/Stripper system. Conversely, no continuous make-up stream of water is required, either. By controlling the temperature of the scrubbed flue gas to the absorber, the MEA system can be kept in water balance. Excess water can accumulate in the Stripper Reflux Drum and can be reused once the system is corrected to operate in a balanced manner. Should water need to be discarded, contaminants will include CO_2 and MEA.

Cooling Tower Blowdown:

The composition limits on cooling tower blowdown are shown in Table 3-46.

Table 3-46: Cooling Tower Blowdown Composition Limitations – Case 5/Concept A

Component	Specification
Suspended Solids	30 ppm average monthly, 100 ppm maximum daily
pH	6.9 to 9
Oil and Grease	15 ppm maximum monthly, 20 ppm maximum daily
Free Chlorine	0.035 ppm

There is a thermal limit specification for the entire river. However, the blowdown volume is too small to affect it significantly.

Relief Requirements:

The relief valve discharges from the CO_2 Recovery Unit to atmosphere. No tie-ins to any flare header are necessary.

3.1.5.8 Case 5/Concept A Plant Layout - CO_2 Removal, Compression, and Liquefaction System

The new equipment required for Case 5/Concept A covers ~7.8 acres of plot area. Plant layout drawings prepared for the Case 5/Concept A CO_2 Recovery System are as follows:

These drawings are shown in Appendix I.

- Plot Plan – Overall Site before CO_2 Unit Addition
- U01-D-0208 Plot Plan – Case 5/Concept A: Flue Gas Cooling & CO_2 Absorption
- U01-D-0214 Plot Plan – Case 5/Concept A: Solvent Stripping
- U01-D-0204 Plot Plan – Case 5/Concept A: CO_2 Compression & Liquefaction
- U01-D-0211 Plot Plan – Case 5/Concept A: Overall Layout Conceptual Plan
- U01-D-0200R Plot Plan – Case 5/Concept A: Modified Overall Site Plan

Plant layout has been designed in accordance with a spacing chart called “Oil and Chemical Plant Layout and Spacing” Section IM.2.5.2 issued by Industrial Risk Insurers (IRI).

When reviewing the layout, the first thing to observe is that no highly flammable materials are handled within the CO_2 Recovery Unit. The open cup flash point of MEA is 93°C (200°F) and, therefore, will not easily ignite. In addition to MEA, the corrosion inhibitor is the only other hydrocarbon liquid within the battery limits. The flash point of this material is higher than that of MEA and is handled in small quantities.

As the chemicals used in the process present no fire hazard, there is an opportunity to reduce the minimum spacing between equipment from that normally considered acceptable in hydrocarbon handling plants. Regardless, for the drawings that follow, standard spacing requirements - as imposed by IRI - have been followed.

The plot areas in the immediate vicinity of Unit #5 available for the installation of the desired equipment are small. Some equipment items are placed on structures to allow other pieces of equipment to be placed underneath them. This way pumps and other equipment associated with the Absorber can be located under the structure. Locating the pumps under the structure has been considered acceptable because the fluids being pumped are not flammable.

Noise is an issue with the flue gas fan as much as it is with compressors. Discussions with vendors suggest that it will be possible to provide insulation on the fan casing to limit noise to acceptable levels. Therefore, it has been assumed that no building needs to be provided for noise reasons.

Having economized on the required plot space as noted above, it was judged not to be practical to divide up the absorbers and strippers that are required into the relatively small plot areas initially offered for this purpose. Eventually, it was agreed that the units would be placed in an area about 460 m (1,500 ft) northeast of the Unit #5/6 common stack. By locating the units in a single location, the MEA piping between the absorber and stripper could be minimized, however, the flue gas duct length and steam piping with this location are quite long.

The corrosion inhibitor must be protected against freezing during winter. The caustic solution will not freeze but will become very viscous when it gets cold. Therefore, a heated shed has been provided for housing the Corrosion Inhibitor and the Caustic injection packages.

The plot plan shows a substation in the Stripper area, but none for the Absorber area. The assumption is that because the electrical consumption of the Absorber equipment is small (0.23 MW) compared to the Stripper equipment, the equipment can be run directly from the auxiliary power 480-volt power system.

For the Rich/Lean Solvent Exchanger, which is a plate and frame type exchanger, area estimates received from vendors based on similar conditions suggest that five units/train would be sufficient for the specified service.

3.1.6 Steam Cycle Modifications, Performance, and Integration with Amine Process (Cases 1-5)

This section presents the performance and modification requirements for the steam/water cycles for all five cases of this study.

3.1.6.1 Amine Process Integration

Figure 3-21 shows a simplified steam cycle schematic that highlights the basic modifications required to integrate the CO_2 capture process into the existing water-steam cycle. These modifications include:

- Addition of a new let down steam turbine generator (LSTG),
- Modification of the existing crossover piping (from existing IP turbine outlet to existing LP turbine inlet). Extracted steam will feed the new let down steam turbine generator and reclaim system of the amine CO_2 recovery system. The exhaust of the let down steam turbine generator (LSTG) ultimately provides the feed steam for the reboilers. This includes a new pressure control valve to maintain a required pressure level even at high extraction flow rates.

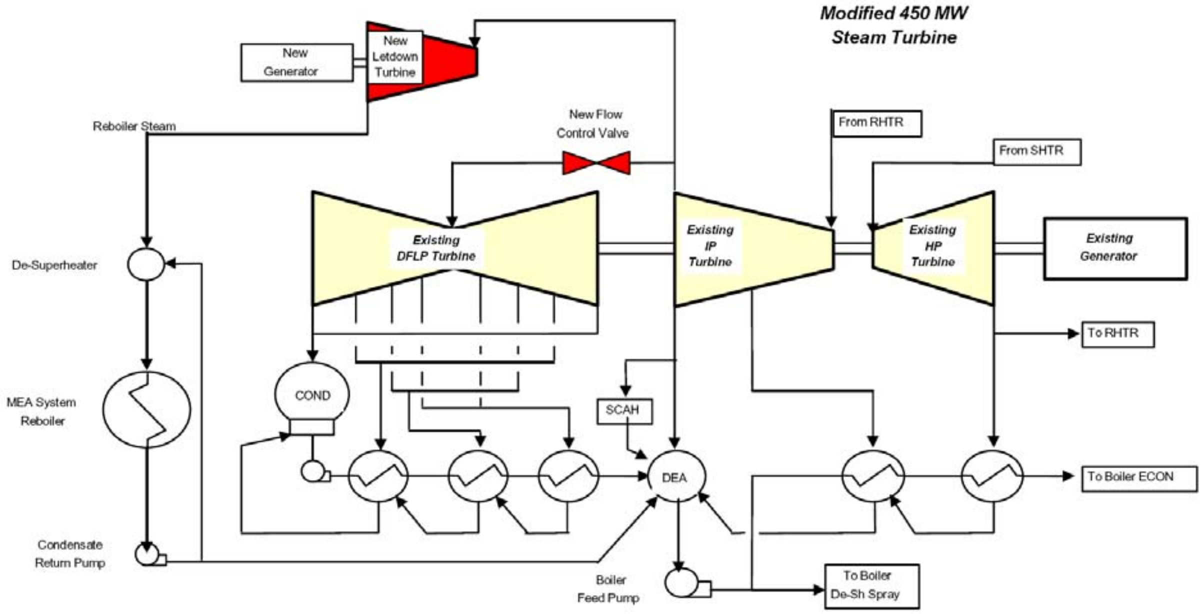


Figure 3-21: Modified Steam/Water Schematic (simplified)

Further modifications to the feedwater system, although not shown in Figure 3-21, are recommended in order to ensure optimum integration of the heat rejected within the CO_2 capture and compression system with the existing steam/water cycle.

For the efficient integration of the amine process into the existing water-steam cycle the locations where the steam needs to be extracted to feed the reboiler and the reclaimer need to be carefully matched. A thorough analysis of the overall process revealed that the amine system reboiler operation would be most economical at a steam pressure of 3.2 bara (47 psia) at the let down turbine exit (See Section 3.1.6.7). This pressure level also ensures that the amine will be protected from being destroyed by high temperatures. The amine system reclaimer needs steam at 6.2 bara (90 psia). When defining the locations of the extraction piping, it needs to be taken into account that these pressure levels need to be maintained also at loads differing from the MCR design load.

Another important assumption was made and is of crucial importance in determination of the potential modifications and, hence, performance of the unit with the MEA plant being in operation. It was assumed that the existing steam turbine/generator is required to continue operation at maximum load in case of a trip of the MEA plant. Additionally, all pressures should still be within a level that no steam will be blown off. This is of specific relevance for any turbine modifications, since changes in steam swallowing capacity of any turbine cylinder requires taking into account this requirement.

Four different scenarios were considered in the current study to assess the impact of various levels of CO_2 removal on the cost/benefit ratio. In the following paragraphs a description of the impact of the CO_2 removal system on water-steam cycle performance will be given. Five cases are discussed as defined below:

- Case 1 - 90% CO_2 removal with advanced amine system
- Case 2 - 70% CO_2 removal with advanced amine system
- Case 3 - 50% CO_2 removal with advanced amine system

- Case 4 - 30% CO_2 removal with advanced amine system
- Case 5 - 96% CO_2 removal with Kerr/McGee ABB Lummus amine system

For ease of performance comparison, the backpressure for each of the four cases was kept constant at 6.35 cm Hga (2.5 in. Hga).

The following subsections discuss the performance and modification requirements for the steam/water cycles for all five cases of this study.

3.1.6.2 Case 1: Steam Cycle for 90% CO_2 Recovery

In order to remove 90% of the CO_2 contained in the flue gas, the amine plant requires approximately 152.5 kg/s of steam (1.21×10^6 lbm/hr). This is approximately 50% of the steam that would enter the LP turbine cylinder in the absence of the amine plant. Out of this steam flow, roughly 4.5% supplies the reclaimer at a pressure of 6.2 bara (90 psia); whereas, the remaining larger portion is required for operation of the reboiler. Before entering the reboilers, steam is expanded through a new turbine, the so-called Let Down Turbine (LDT), to make the best use of the steam’s energy. Refer to Appendix IV for technical details regarding the Let Down Turbine.

Without any additional measures, the decrease in steam flow entering the existing LP turbine would result in a corresponding lower pressure at the LP turbine inlet (about 50% of the pressure level without extraction). Consequently, the pressure at the exhaust of the existing IP turbine would also be reduced to about this same value. Keeping the live steam conditions constant would then result in increased mechanical loading of the IP blades in excess of the permissible stress levels. For this reason, a pressure control valve needs to be added in the IP-LP crossover pipe to protect the IP turbine blading.

Due to the high amount of flow extracted from the IP-LP crossover and, consequently, the remaining low flow passing through the LP turbine, there is a potential risk for the LP blades being damaged. By comparing the load for the 90% CO_2 removal case with data given in the Conesville #5 instruction manual for “lower load limit,” it can be shown that the operation as shown in Figure 3-22 is well within the operational range of the existing LP turbine.

Care was taken to integrate the heat rejected within the amine process into the existing water-steam cycle in an efficient manner. The main sources of integrated heat are provided from three sources as listed below:

- CO_2 compressor intercoolers
- Stripper overhead cooler
- Refrigeration compressor cooler (de-superheating section)

Additionally, warm condensate is returned from the amine reboiler/reclaimer system to the existing deaerator. For the 90% CO_2 removal case, the most beneficial arrangement for heat integration is also shown in the lower part of Figure 3-22. It should be noted that with this arrangement the deaerator flow increases by approximately 26%. This may impact deaerator performance or require either modification of the deaerator or a change in the heat integration arrangement in order to reduce the duty of the deaerator. Although the cost for modification of the deaerator was not included in this study, given the relatively large costs required for the other plant modifications (new amine plant and CO_2 compression equipment), this omission should not impact the results of the study significantly.

In summary, the power output of the Conesville #5 Unit after modification to remove 90% of the CO_2 contained in the flue gas will decrease by approximately 16.3% (from 463.5 MWe to 388.0 MWe) when compared to the Base Case as shown in Section 2.2.4.

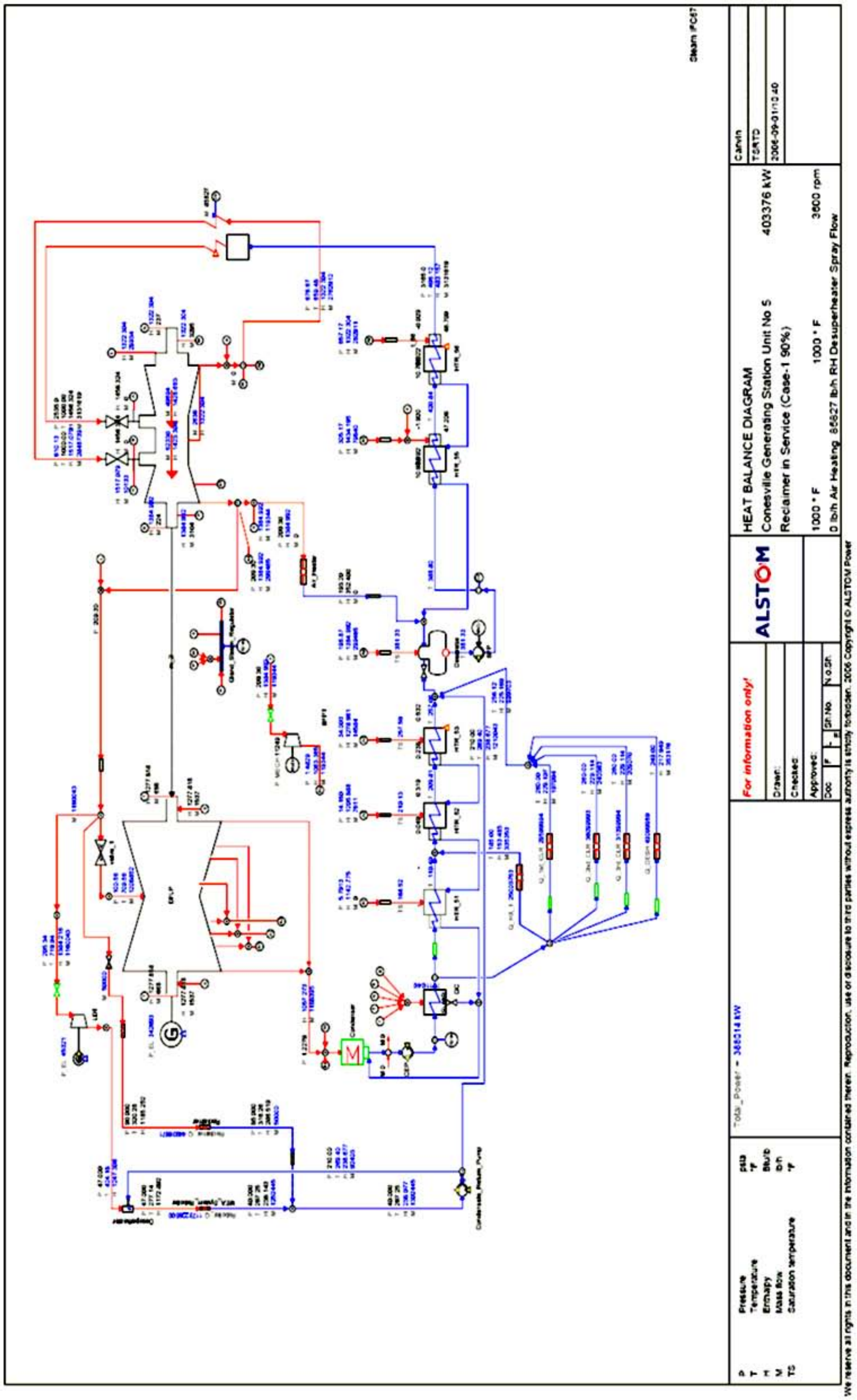


Figure 3-22: Case 1 – Modified Water-Steam Cycle for 90% CO_2 Removal

3.1.6.3 Case 2: Steam Cycle for 70% CO_2 Recovery

In the case of removal of 70% of the CO_2 contained in the flue gas, the steam required to operate the boiler/reclaimer of the amine process is approximately 118.5 kg/s (940.8 x 10^3 lbm/hr), equivalent to approximately 39% of the steam that would enter the LP turbine cylinder in the absence of the amine plant.

Similar to the 90% removal case, the lower steam flow entering the LP turbine would result in a correspondingly lower pressure at the LP turbine inlet (about 59% of the pressure without extraction). Consequently, the pressure at the exhaust of the IP turbine would also come down; therefore, a pressure control valve is required to protect the IP blading.

For this scenario of 70% CO_2 removal, a low load limitation within the LP is not expected to be an issue because even more steam remains within the LP turbine cylinder compared to the 90% removal case.

Heat integration is done in the same manner as for the 90% removal case and is shown in the lower part of Figure 3-23. The deaerator flow is somewhat less than in the 90% removal case, but still significantly higher than the flow as indicated for the reference case (approximately 24.5% larger). Again, this may impact performance of the deaerator or require either modification of the deaerator or a change in the heat integration arrangement in order to reduce the duty of the deaerator.

In summary, as illustrated in Figure 3-23, the power output of the Conesville #5 Unit after modification to remove 70% of the CO_2 contained in the flue gas will decrease by approximately 12.4% (from 463.5 MW to 405.9 MW) when compared to the Base Case (please refer Section 2.2.4).

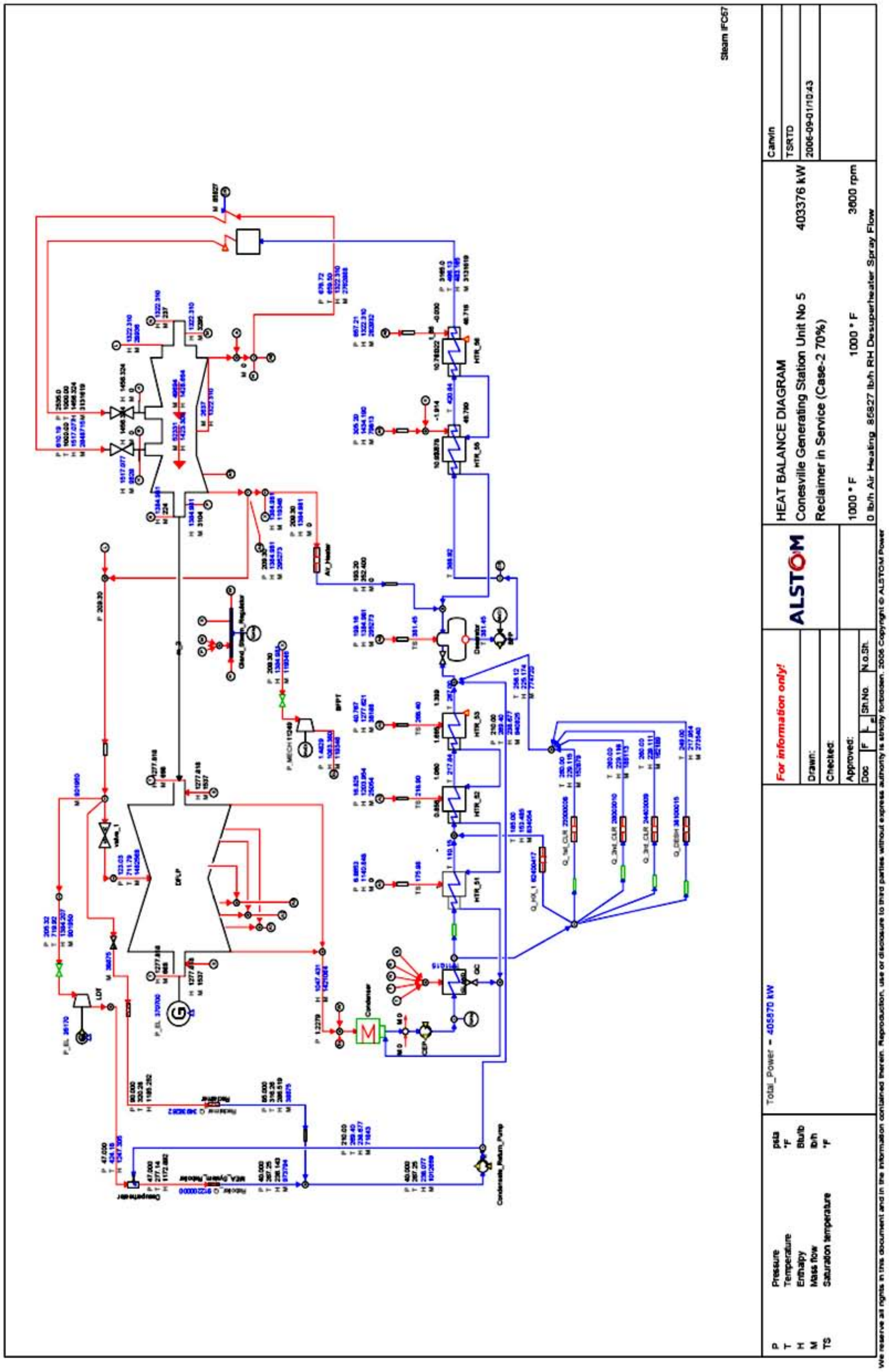


Figure 3-23: Case 2 – Modified Water-Steam Cycle for 70% CO_2 Removal

3.1.6.4 Case 3: Steam Cycle for 50% CO_2 Recovery

In the case of removal of 50% of the CO_2 contained in the flue gas, the steam required to operate the boiler/reclaimer of the amine process is approximately 84.7 kg/s (671.9 x 10^3 lbm/hr), equivalent to approximately 27.6% of the steam that would enter the LP turbine cylinder in the absence of the amine plant.

Again, the lower steam flow entering the LP turbine would result in a corresponding lower pressure at the LP turbine inlet (about 70% of the pressure without extraction) and, consequently, a lower pressure at IP exhaust. Therefore, a pressure control valve is required to protect the IP blading.

Operation close to low load limitation within the LP is not expected to be an issue.

Heat integration is done in the same manner as for the 90% removal case and is shown in Figure 3-24. The deaerator flow is somewhat less than in the 90% removal case, but still significantly higher than the flow as indicated for the reference case (approximately 20% higher). Again, this may impact performance of the deaerator or require either modification of the deaerator or a change in the heat integration arrangement in order to reduce the duty of the deaerator. Moving the location where the condensate from the amine plant is fed back into the turbine cycle up one feedwater heater, i.e., upstream of HTR #53 instead of downstream reduces the duty on the deaerator, but the power generated will be less by approximately 200 kW.

The modified water-steam cycle is shown in Figure 3-24. In summary, the power output of the Conesville #5 Unit after modification to remove 50% of the CO_2 will decrease by approximately 8.6% (from 463.5 MW to 423.5 MW) when compared to the Base Case (please refer to Section 2.2.4).

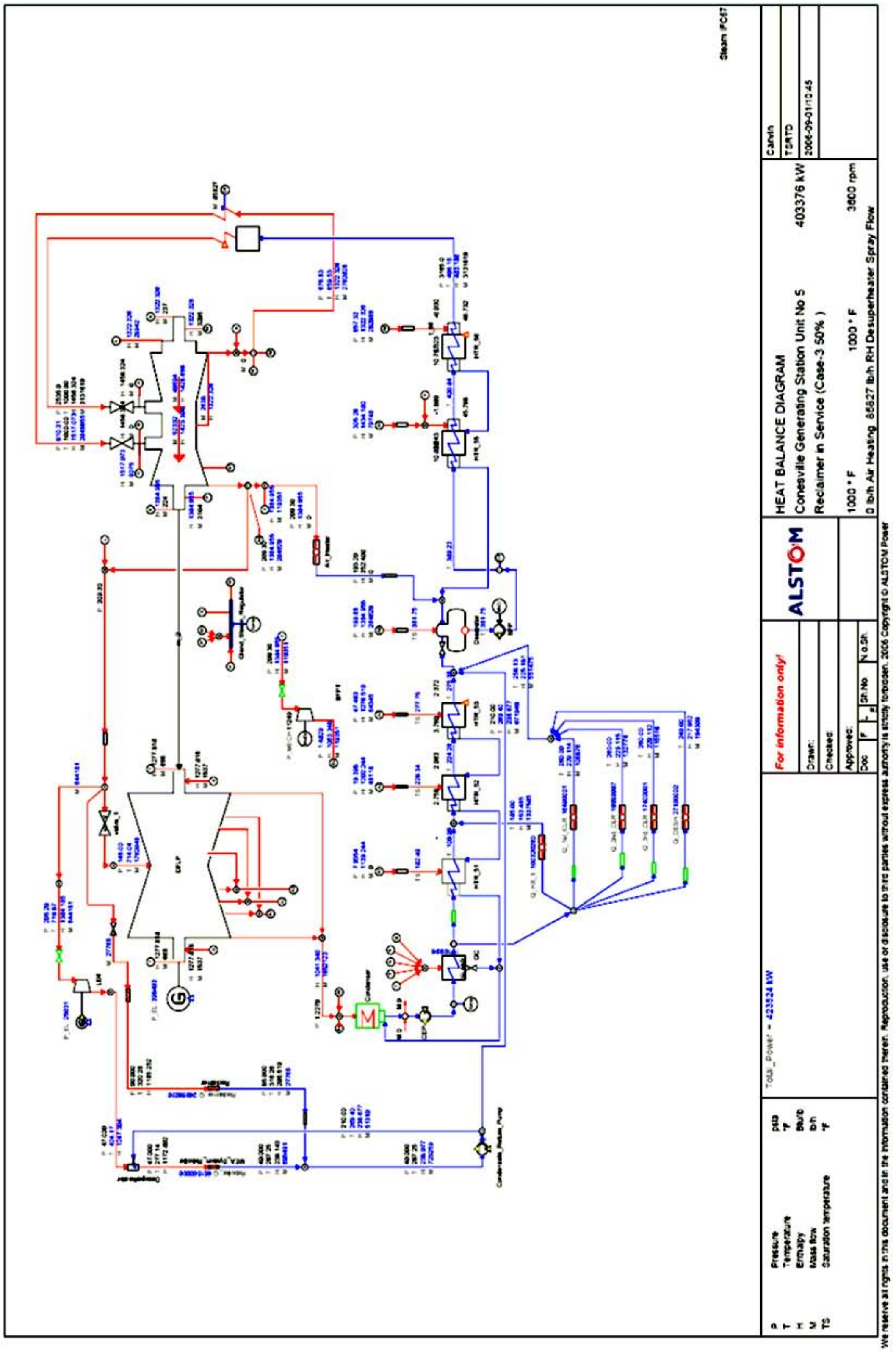


Figure 3-24: Case 3 - Modified Water-Steam Cycle for 50% CO_2 Removal

3.1.6.5 Case 4: Steam Cycle for 30% CO_2 Recovery

In the case of removal of 30% of the CO_2 contained in the flue gas, the steam required to operate the boiler/reclaimer of the amine process is approximately 50.8 kg/s (403.2 x 10^3 lbm/hr), equivalent to approximately 16.4% of the steam that would enter the LP turbine cylinder in the absence of the amine plant.

The lower steam flow entering the LP turbine results in a corresponding lower pressure at the LP turbine inlet (about 80.9% of the pressure without extraction). Consequently, the pressure at the exhaust of the IP turbine would also come down; therefore, a pressure control valve is required to protect the IP blading.

With the heat integration arrangement being the same as with the other cases, the deaerator flow still is approximately 13.4% greater than for the reference case. Again, this may impact performance of the deaerator, or require either modification of the deaerator, or a change in the heat integration arrangement in order to reduce the duty of the deaerator.

The modified water-steam cycle is shown in Figure 3-25. In summary, the power output of the Conesville #5 Unit after modification to remove 30% of the CO_2 will decrease by approximately 5% (from 463.5 MW to 440.7 MW) when compared to the reference case (please refer to Section 2.2.4).

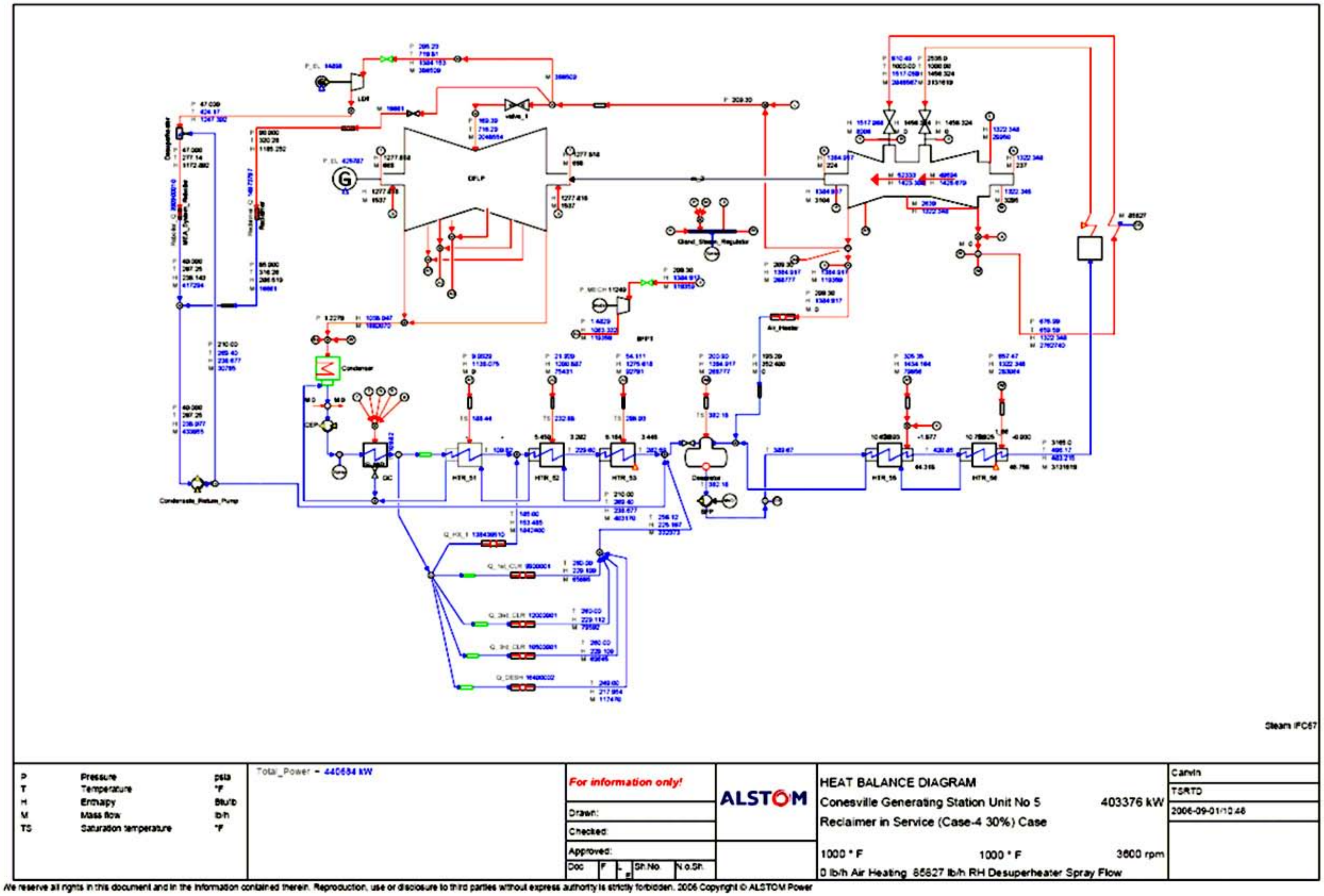
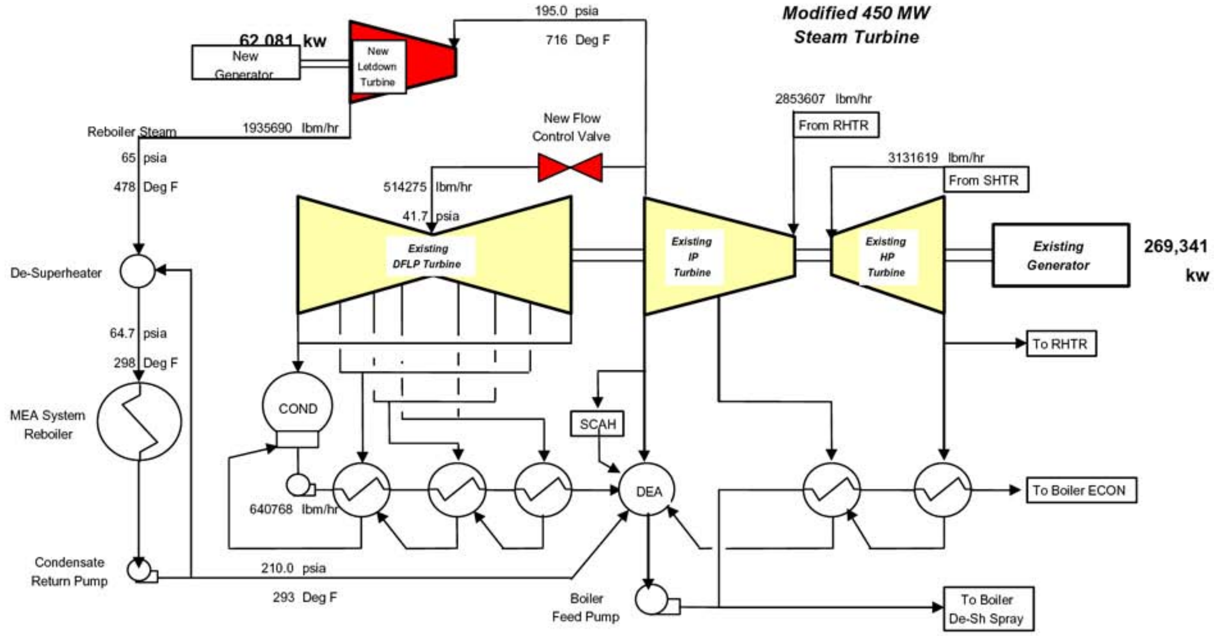


Figure 3-25: Case 4 – Modified Water-Steam Cycle for 30% CO_2 Removal

3.1.6.6 Case 5/Concept A: Steam Cycle for 96% CO_2 Recovery (from previous study)

The steam cycle system for Case 5/Concept A is modified as shown in Figure 3-26, while Figure 3-27 shows the associated Mollier diagram. It should be pointed out that the performance shown for the steam turbine in this case was developed in 2000 using a less detailed analysis than was used for Cases 1-4. About 79% of the IP turbine exhaust is extracted from the IP/LP crossover pipe. This steam is expanded to about 4.5 bara (65 psia) through a new let down steam turbine generating 62,081 kWe. The exhaust from the new turbine, at about 248°C (478°F), is de-superheated and then provides the energy requirement for the solvent regeneration done in the reboilers/stripper system of the MEA CO_2 removal process. The condensate from the reboilers is pumped to the existing deaerator. The remaining 21% of the IP turbine exhaust is expanded in the existing LP turbine. The current study confirmed that the existing LP turbine would be able to operate at this low flow condition. The modified existing steam cycle system produces 269,341 kWe. The total output from both generators is 331,422 kWe. This represents a gross output reduction of 132,056 kWe (about 28.5%) as compared to the Base Case.



Steam Cycle Energy Balance

Energy Outputs	(10^6 Btu/hr)	Energy Inputs	(10^6 Btu/hr)
Power Output (Existing and New Turbines)	1151	Boiler Heat Input	3707
Steam Coil Air Heater Output	0	Condensate Pump	0
MEA System Reboiler Duty Output	1953	**Total Energy Input**	3708
Condenser Loss	603		
Total Energy Output	3708	**In - Out**	0

Figure 3-26: Case 5/Concept A – Modified Water-Steam Cycle for 96% CO_2 Removal

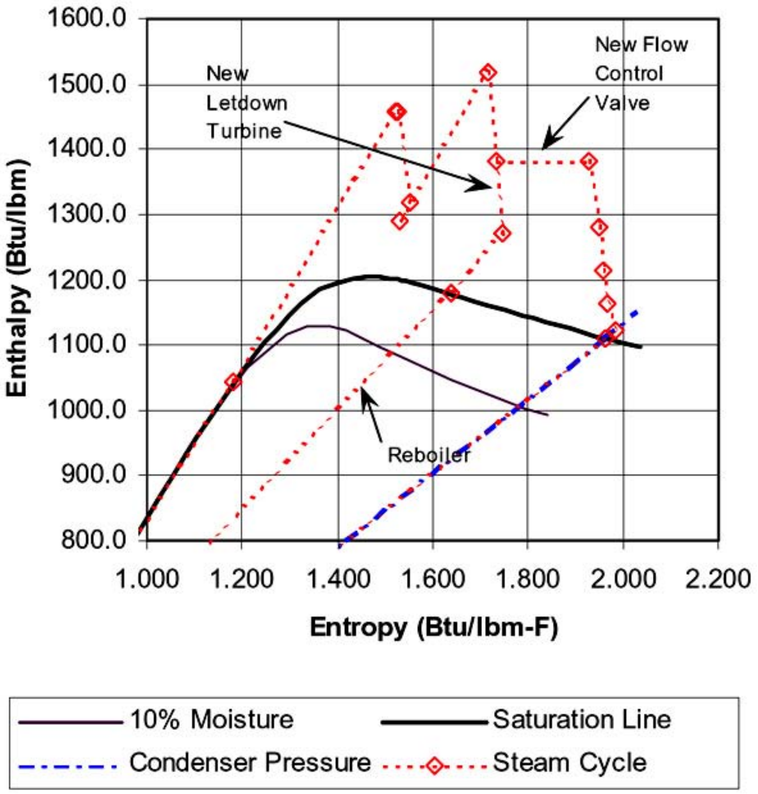


Figure 3-27: Case 5/Concept A - Modified Water-Steam Cycle Mollier Diagram for 96% CO_2 Removal

3.1.6.7 Discussion of Alternate Solutions for Steam Extraction

While this study focuses on the addition of a new LDT to the existing water-steam cycle to effectively use the energy contained in the steam while matching the requirements of the amine plant, the following paragraphs will give a brief overview of other available retrofit solutions as potential alternatives to the let down turbine approach. The common advantage of all the alternate retrofit scenarios under consideration is that there is no need for an additional turbine-generator with all the equipment and modifications that are linked to this (e.g., new foundations/foundation enforcements, additional transformer, piping, grid connection, etc).

As with all arrangements under consideration, retrofit scenarios have to take into account that the unit has to be able to run at maximum load both with and without the amine plant being in operation. It is this requirement that tremendously increases the mechanical design load acting on the turbine blades, since the pressure upstream of the location where the steam will be extracted drops approximately proportional to the relative amount of steam that will be extracted. This of course means that a scenario for 90% removal of CO_2, where approximately 50% of the steam entering the existing LP turbine cylinder (See Figure 3-28) will be extracted, puts the greatest load on the blading.

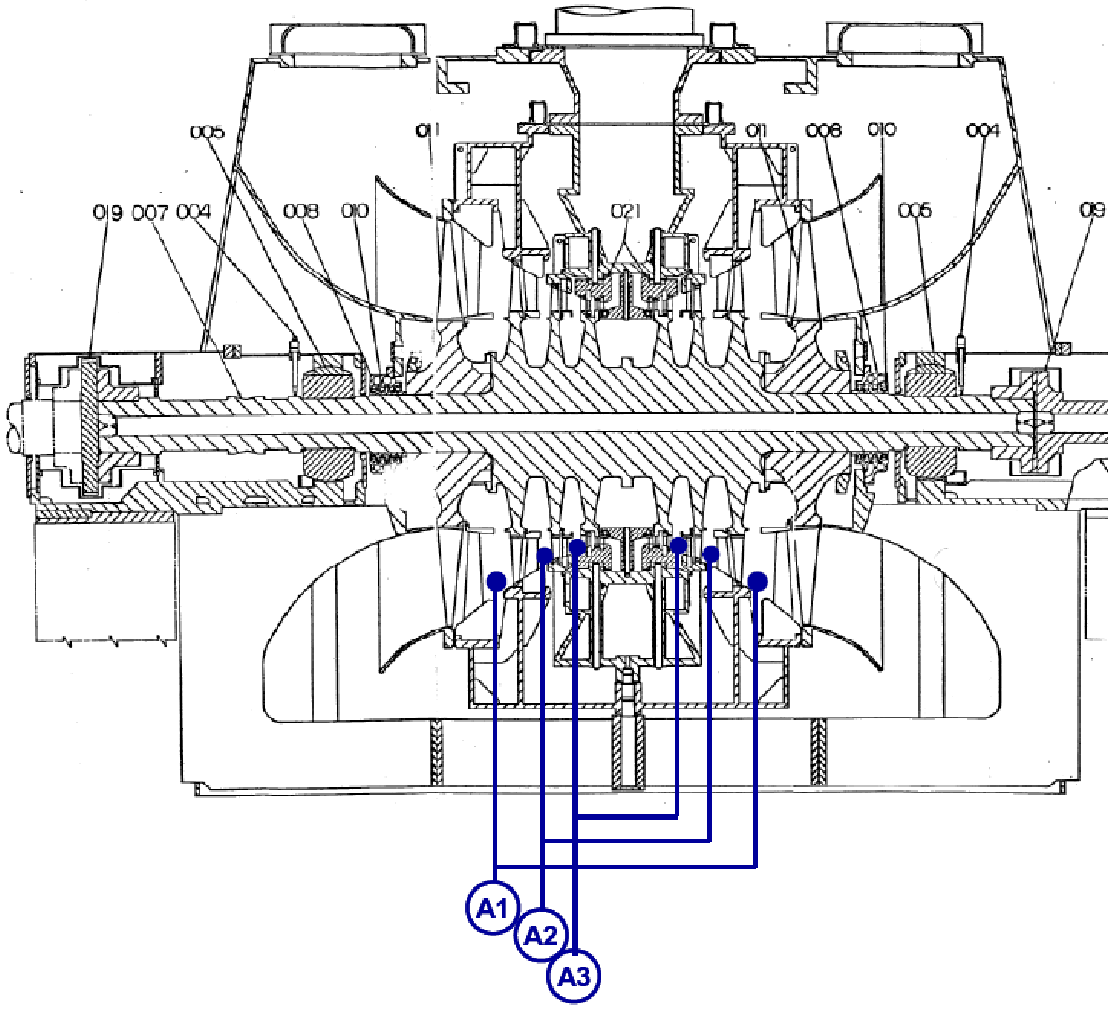


Figure 3-28: Existing LP Turbine at Conesville Unit #5

In Table 3-47 pressure data are given for a scenario with 30% CO_2 removal. The data in Row 2 of the table ("Reference Conditions") are for the 5% overpressure load condition without any modification. In Row 3 ("30% CO_2 removal"), the impact of steam extraction on the pressure distribution within the remaining LP turbine can be seen. Due to the given swallowing capacity of the existing LP turbine the pressure at the LP turbine, inlet drops down from ~14.1 bara (205 psia) with no steam extraction to ~11.7 bara (169 psia) with the amine plant being in operation [requiring ~51 kg/s (403,000 lbm/hr) of steam to remove 30% of the CO_2]. Without taking additional measures, about the same pressure would also act on the exhaust section of the IP turbine and the existing blading would not be able to withstand this increased mechanical loading.

Table 3-47: Expected Steam Conditions at Extraction Points for 30% CO_2 Removal

		A1	A2	A3	LP inlet	
Reference Conditions		9.5 psia 169.8 klb/hr	25.2 psia 119.5 klb/hr	63.7 psia 140.9 klb/hr	205.1 psia 2,486.4 klb/hr	No steam extraction
30% CO_2 removal	Existing turbine, pls. refer to Section "30% removal" above	9.0 psia 0 klb/hr	21.9 psia 75.4 klb/hr	54.1 psia 92.8 klb/hr	169.4 psia 2,048.6 klb/hr	Steam extraction in operation
Scenario "LP retrofit"	30% CO_2 removal, no LDT, retrofitted LP turbine	~9.0 psia; determined by turbine swallowing capacity & backpressure	47 psia to feed reboiler	90 psia to feed reclaimer	205.1 psia	Steam extraction in operation
Scenario "LP & HP/IP retrofit"	30% CO_2 removal, requirements for LP turbine retrofit	~9.0 psia; determined by turbine swallowing capacity & backpressure	~22 psia	~47 psia	~105 psia	Steam extraction in operation

A retrofit solution offers the potential to specifically address these issues. This can be done by designing the new blade path in such a way that the pressure levels required to feed the amine plant can be closely matched at the extraction points inside the LP turbine, thus minimizing the impact on the IP turbine. A preliminary engineering assessment revealed that a steam path could be designed to achieve a 6.2 bara (90 psia) pressure level at the first extraction point ("A3") to feed the reclaimer as well as a 3.2 bara (47 psia) pressure level at the second extraction point ("A2") to feed the reboilers. Since the steam flow to feed the reboiler with the 3.2 bara (47 psia) steam is significantly more than the flow that was originally extracted to feed the connected feedwater heater (48.7 kg/s vs. 15.1 kg/s or 386.5 x10^3 lbm/hr vs. 119.5 x10^3 lbm/hr) it is very likely that the piping requires modification, which in turn may mean that the LP turbine outer casing also needs to be modified in order to allow bigger pipe diameters to be connected. It also needs to be considered that the existing piping and the connected feedwater heater most likely will not be designed to allow operation at the higher pressure (3.2 bara vs. 1.7 bara or 47 psia vs. 25.2 psia). This could be overcome by either replacement of the existing piping and feedwater heater, or it needs to be checked whether the blade path and turbine casing could be modified to allow for an additional extraction point at approximately 1.7 bara (25 psia).

In principle, the comments above apply similarly to the 50%, 70%, and 90% CO_2 removal scenarios with the requirements for a proper steam path design getting more and more challenging as more steam is required for the amine plant, i.e., with increasing rate of CO_2 removal. At higher removal rates, in order to allow operation, both with and without the amine plant being in operation, it is likely that an HP/IP retrofit needs to be considered as well. This would allow not only reducing the mechanical load on the LP blading by reducing the pressure level at LP inlet, but also better matching of the extraction pressures to the new requirements while optimizing cycle efficiency.

In summary, alternative technically proven retrofit solutions are available that may offer attractive solutions that does not necessitate the addition of a new Let Down Turbine. For a typical LP turbine retrofit solution, please refer to Figure 3-29. It should be noted that all of the retrofit options (HP, IP, LP), in addition to the advantages indicated above, offer the potential advantage of improved heat rate and power output due to the application of state of the art blading technology, and therefore can mitigate, to some extent, the performance deterioration due to the addition of the post-combustion carbon capture equipment. To have a sound basis for comparison and evaluation, a detailed engineering assessment is required, taking into account unit specifics that go well beyond the intent and scope of this study.

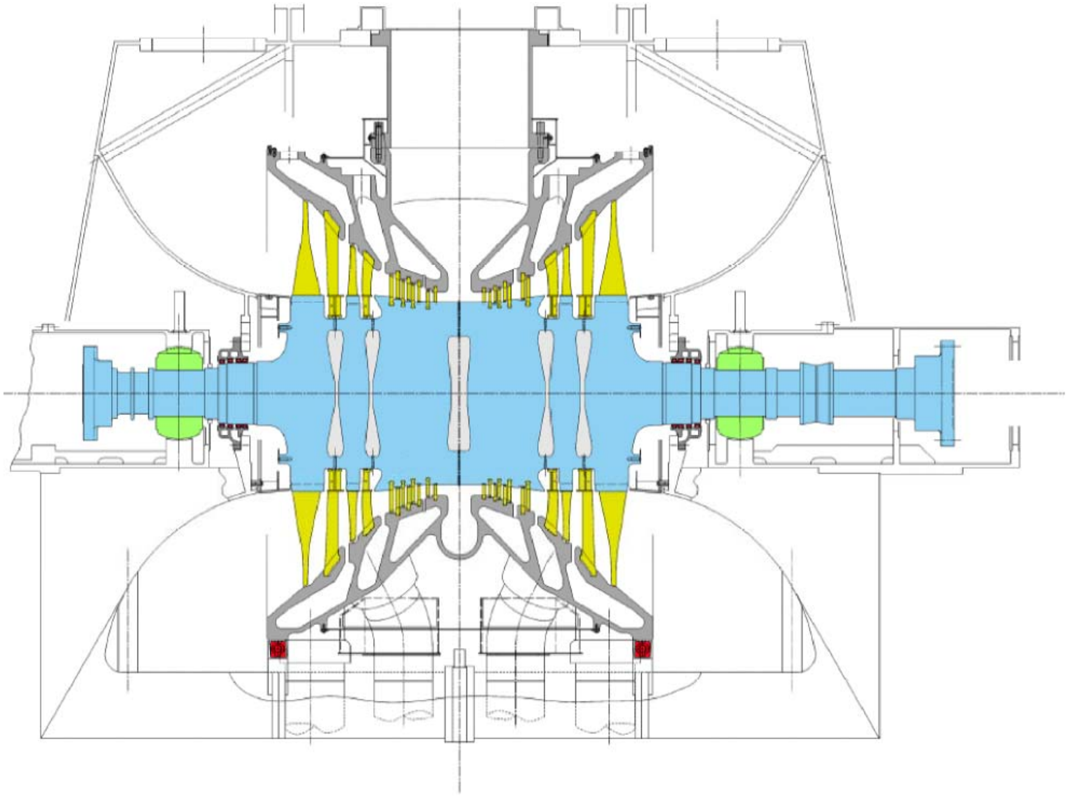


Figure 3-29: Typical Retrofit Solution for the Conesville Unit #5 LP Turbine Type

3.1.7 Project Construction Schedule (Cases 1-5)

Figure 3-30 shows the project construction schedule for the retrofit of Conesville Unit #5 to CO_2 capture, which is 36 months in duration. This schedule is assumed to apply to each of the five cases in this study (Cases 1-5). Engineering is completed in the first 15 months. Procurement occurs in months 9-23 and Construction takes place in months 14-34. Commissioning and startup are done in months 35 and 36.

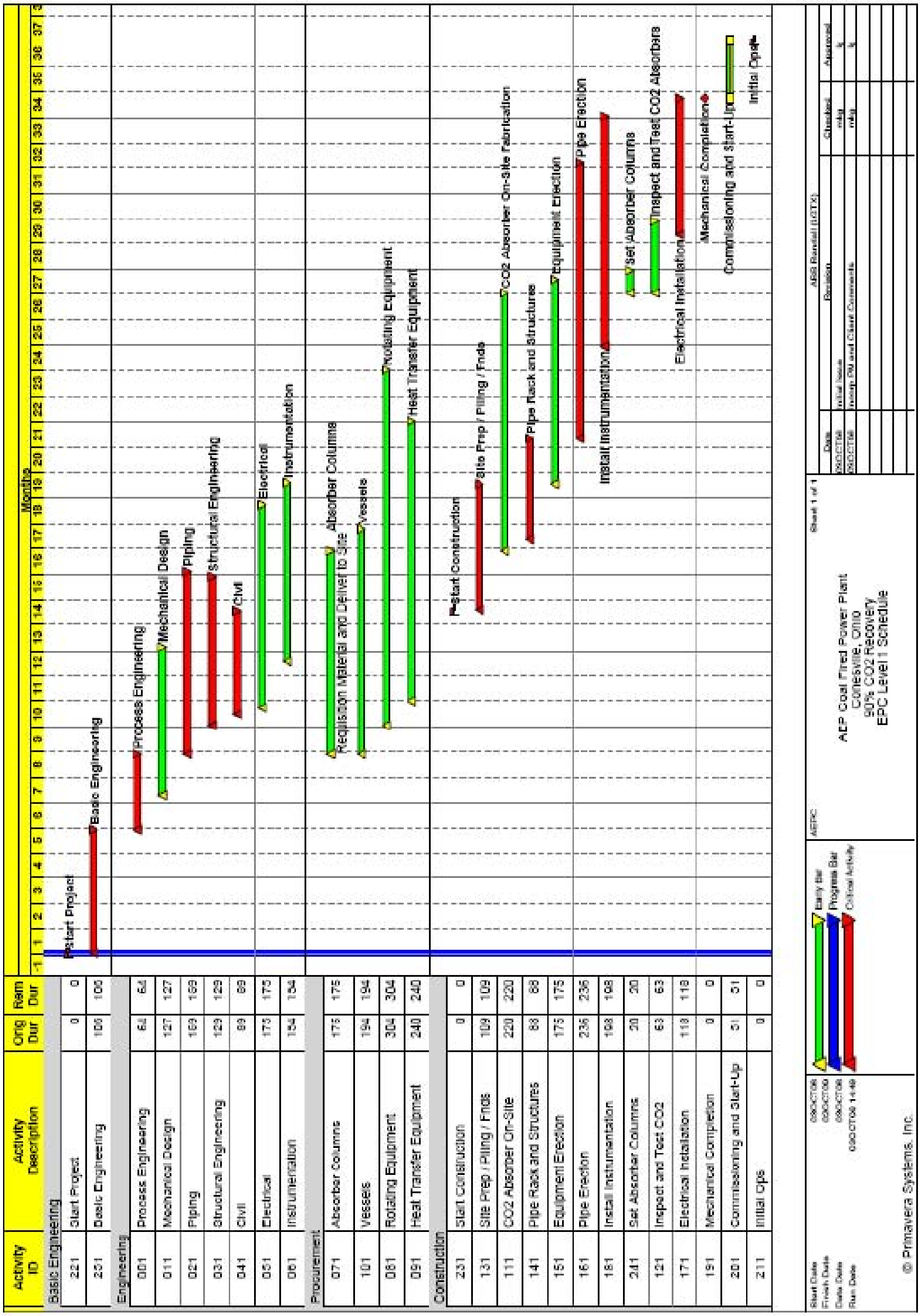


Figure 3-30: Project Construction Schedule (Cases 1-5)

3.2 Summary and Comparison of Overall Plant Performance and Carbon Dioxide Emissions (Cases 1-5)

This section summarizes overall performance and CO_2 emissions from the existing and modified power plants. Table 3-48 shows a comparison of the Conesville #5 plant performance and emissions for the CO_2 recovery cases and the Base Case that has no CO_2 recovery system. The first column shows the performance results for the Base Case. The performance shown for the Base Case is identical to what was reported in the previous study (Bozzuto et al., 2001) for this unit.

Table 3-48: Plant Performance and CO_2 Emissions Comparison (Base Case and Cases 1-5)

		(units)	Base-Case Original Plant	Case 5 Concept A MEA - 96% Capture	Case 1 Advanced MEA - 90% Capture	Case 2 Advanced MEA - 70% Capture	Case 3 Advanced MEA - 50% Capture	Case 4 Advanced MEA - 30% Capture
Boiler Parameters								
Main Steam Flow		(lbm/hr)	3131619	3131651	3131651	3131651	3131651	3131651
Reheat Steam Flow (to IP turbine)		(lbm/hr)	2853607	2853607	2848739	2848715	2848655	2848567
Main Steam Pressure		(psia)	2535	2535	2535	2535	2535	2535
Main Steam Temp		(Deg F)	1000	1000	1000	1000	1000	1000
Reheat Steam Temp		(Deg F)	1000	1000	1000	1000	1000	1000
Boiler Efficiency		(percent)	88.13	88.13	88.13	88.13	88.13	88.13
Flue Gas Flow leaving Economizer		(lbm/hr)	4014743	4014743	4014743	4014743	4014743	4014743
Flue Gas Temperature leaving Air Heater		(Deg F)	311	311	311	311	311	311
Coal Heat Input (HHV)	(HHV)	(10^6 Btu/hr)	4228.7	4228.7	4228.7	4228.7	4228.7	4228.7
	(LHV)	(10^6 Btu/hr)	4037.9	4037.9	4037.9	4037.9	4037.9	4037.9
CO_2 Removal Steam System Parameters								
CO_2 Removal System Steam Pressure		(psia)	---	65	47	47	47	47
CO_2 Removal System Steam Temp		(Deg F)	---	478	424	424	424	424
CO_2 Removal System Steam Extraction Flow		(lbm/hr)	---	1935690	1210043	940825	671949	403170
CO_2 Removal System Condensate Pressure (from reboilers)		(psia)	---	64.7	40	40	40	40
CO_2 Removal System Condensate Temperature		(Deg F)	---	292.7	267.3	267.3	267.3	267.3
CO_2 Removal System Heat to Cooling Tower		(10^6 Btu/hr)		1441.1	890.2	692.5	494.2	293.1
Natural Gas Heat Input	(HHV)[2]	(10^6 Btu/hr)	0	17.7	13.0	9.7	6.7	4.2
[2] (For Desicant Regeneration)	(LHV)	(10^6 Btu/hr)	---	16.0	11.7	8.7	6.0	3.8
		(10^6 SCF/Day)		0.417	0.312	0.232	0.161	0.101
Steam Cycle Parameters								
Total Heat Input to Steam Cycle		(10^6 Btu/hr)	3707.4	3707.4	3707.4	3707.4	3707.4	3707.4
Heat Output to CO_2 Removal System Reboilers & Reclaimer		(10^6 Btu/hr)	---	1953.0	1218.1	947.1	676.5	405.9
Existing Condenser Pressure		(psia)	1.23	1.23	1.23	1.23	1.23	1.23
Existing Condenser Heat Loss		(10^6 Btu/hr)	2102.8	603.3	1257.0	1514.7	1778.6	2047.6
Existing Steam Turbine Generator Output		(kW)	463478	269,341	342693	370700	398493	425787
CO_2 Removal System Turbine Generator Output		(kW)	0	62,081	45321	35170	25031	14898
Total Turbine Generator Output		(kW)	463478	331422	388014	405870	423524	440685
Auxiliary Power Requirements								
Condensate Pump Power		(kW)	563	450	504	515	527	540
Condenser Cooling Water Pump Power		(kW)	5562	5407	5679	5838	6011	6191
Boiler Island Auxiliary Power (Fans & Pulverizers)		(kW)	7753	7753	7753	7753	7753	7753
Coal & Ash Handling System		(kW)	1020	1020	1020	1020	1020	1020
FGD & ESP System Auxiliary Power		(kW)	8157	8157	8157	8157	8157	8157
Misc. Auxiliary Power (Lighting, HVAC, Trans, etc)		(kW)	6645	6645	6645	6645	6645	6645
CO_2 Removal System Auxiliary Power		(kW)	0	50355	54939	42697	30466	18312
Total Auxiliary Power		(kW)	29700	79788	84697	72625	60579	48618
fraction of gross output		(fraction)	0.064	0.241	0.218	0.179	0.143	0.110
Plant Performance Parameters								
Net Plant Output		(kW)	433778	251634	303317	333245	362945	392067
Normalized Net Plant Output (Relative to Base Case)		(fraction)	1.00	0.58	0.70	0.77	0.84	0.90
Net Plant Efficiency (HHV)		(fraction)	0.3501	0.2022	0.2441	0.2683	0.2925	0.3161
Net Plant Efficiency (LHV)		(fraction)	0.3666	0.2119	0.2556	0.2811	0.3063	0.3311
Normalized Efficiency (HHV; Relative to Base Case)		(fraction)	1.00	0.58	0.70	0.77	0.84	0.90
Net Plant Heat Rate (HHV)		(Btu/kWh)	9749	16875	13984	12719	11670	10796
Net Plant Heat Rate (LHV)		(Btu/kWh)	9309	16110	13351	12143	11142	10309
Plant CO_2 Emissions								
Carbon Dioxide Produced		(lbm/hr)	866102	868137	867595	867212	866872	866585
Carbon Dioxide Recovered		(lbm/hr)	0	835053	779775	607048	433606	260164
Carbon Dioxide Emissions		(lbm/hr)	866102	33084	87820	260164	433266	606422
Fraction of Carbon Dioxide Recovered		(fraction)	0	0.962	0.90	0.70	0.50	0.30
Specific Carbon Dioxide Emissions		(lbm/kWh)	1.997	0.131	0.290	0.781	1.194	1.547
Normalized Specific CO_2 Emissions (Relative to Base Case)		(fraction)	1.00	0.066	0.145	0.391	0.598	0.775
Avoided Carbon Dioxide Emissions (as compared to Base)		(lbm/kWh)	---	1.865	1.707	1.216	0.803	0.450

The second column shows results for Case 5/Concept A, also from the previous study (Bozzuto, et al., 2001), which captured ~96% of the CO_2 using the Kerr-McGee/ABB Lummus Global oxygen inhibited MEA technology. Columns 3, 4, 5, and 6 show results for Cases 1-4 of the current study, which capture 90%, 70%, 50%, and 30% of the CO_2, respectively, using an advanced MEA system.

Several comparisons have been made in these tables and throughout the report. Some of the more important comparisons are categorized and summarized in the following subsections.

3.2.1 Auxiliary Power and Net Plant Output

The auxiliary power required for the Base Case is 29,700 kW or about 6.4% of the gross electrical output. Net plant output is 433,778 kW. All the CO_2 capture options require large amounts of additional auxiliary power to the CO_2 compression systems and CO_2 capture systems, which deliver the CO_2 as a liquid at 138 barg (2,000 psig). These CO_2 capture and compression systems consume in the range of about 18-55 MWe. The total amount of auxiliary power for these plants represents a range of about 11-24% of the gross output, depending on CO_2 recovery level, as shown in Figure 3-31.

Additionally, extraction of steam from the existing steam turbine to provide energy necessary for solvent regeneration also significantly reduces steam turbine output (refer to Section 3.2.4) and, therefore, reduces net plant output. Net plant output is reduced to between 252-392 MWe for these cases or between about 58%-90% of the Base Case output as shown in Figure 3-31.

Comparison of net plant outputs for Case 5/Concept A from the original study (Bozzuto et al., 2001) and the advanced MEA 90% Capture case of the current study indicates the impact of the advanced MEA solvent. An improvement of about 51 MWe in net output (~20% greater output) is realized with the advanced MEA solvent. This represents an improvement of about 28% on output reduction. Correcting to a common CO_2 capture percentage of 96% would reduce this improvement to about 26%.

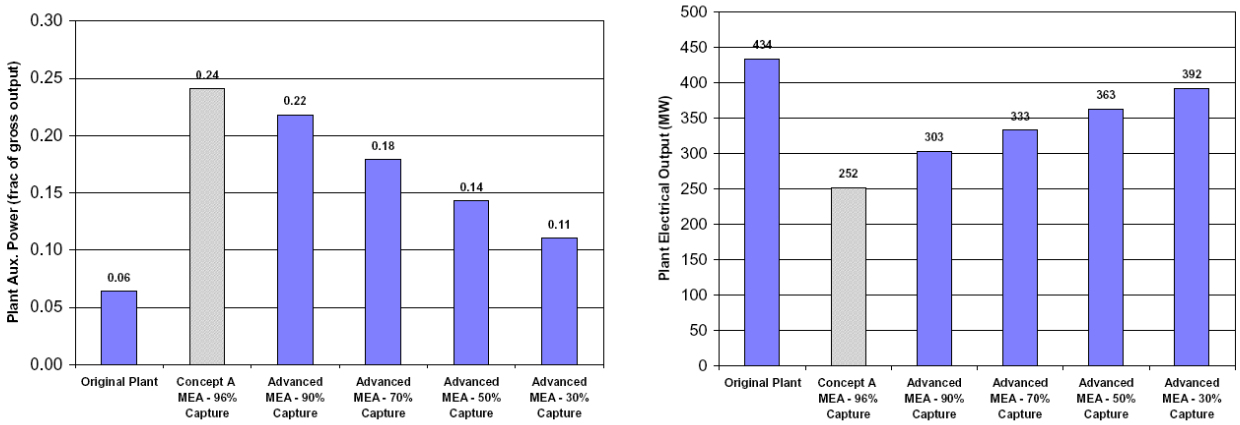


Figure 3-31: Plant Auxiliary Power & Net Electrical Output (MWe)

3.2.2 Net Plant Heat Rate and Thermal Efficiency

Because of the large energy requirements for solvent regeneration and large auxiliary power demands for the new equipment required for the CO_2 capture systems, net plant heat rate and thermal efficiency are degraded substantially relative to the Base Case as shown in Figure 3-32. Figure 3-33 shows the same results plotted as a function of the capture level. As shown in Figure 3-33, the thermal efficiency decreases linearly for the advanced amine cases as CO_2 capture level increases (Cases 1-4) and then drops sharply for Case 5 with the Kerr/McGee ABB Lummus amine.

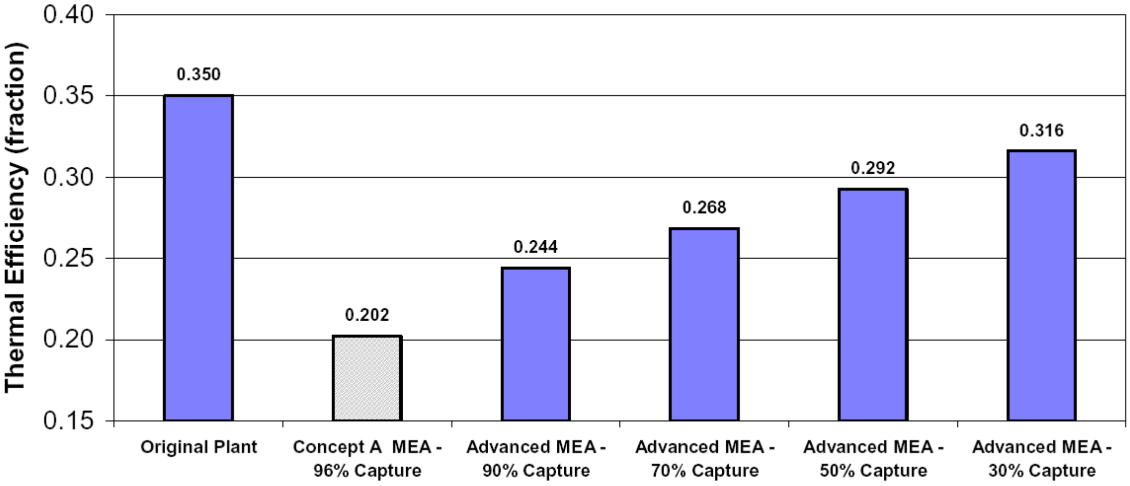


Figure 3-32: Plant Thermal Efficiency (HHV Basis)

The Base Case plant thermal efficiency (HHV Basis) is about 35%. For the CO_2 capture cases, with large amounts of steam extracted for solvent regeneration and increased auxiliary power for CO_2 compression and liquefaction systems, plant thermal efficiencies are reduced to between 31.6%-20.2% (HHV basis) depending on capture level.

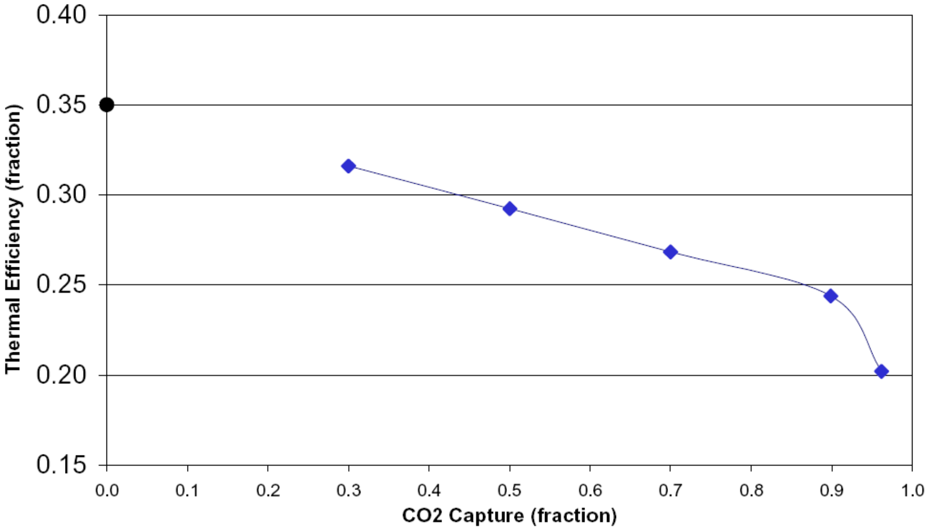


Figure 3-33: Plant Thermal Efficiency vs. Capture Level

Figure 3-34 shows the efficiency losses relative to the Base Case. Thermal efficiency losses range from about 3.4 to 14.8 percentage points.

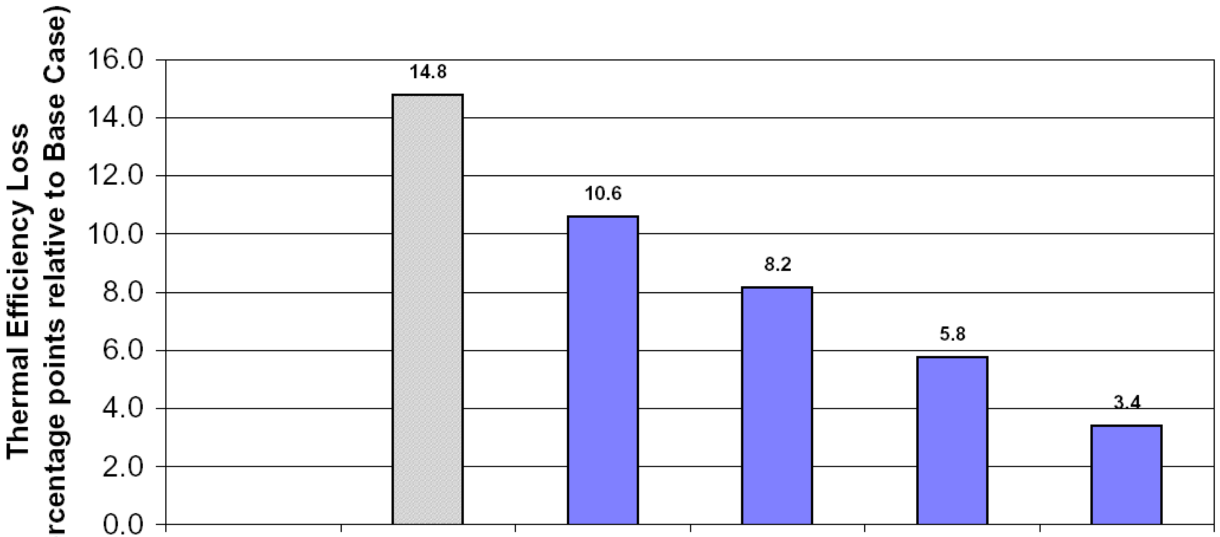


Figure 3-34: Plant Thermal Efficiency Loss Relative to Base Case (HHV Basis)

Comparison of thermal efficiency losses for Case 5/Concept A from the original study (Bozzuto et al., 2001), and the advanced MEA 90% capture case of the current study indicates the impact of using the advanced MEA solvent. A reduction of about 4.2 percentage points in thermal efficiency loss is realized with the advanced MEA solvent. This represents an improvement of about 28% with the advanced MEA solvent. Correcting to a common CO_2 capture percentage of ~96% would reduce this improvement to about 3.5 percentage points in thermal efficiency loss or about 24%.

3.2.3 CO_2 Emissions

CO_2 emissions are summarized in Table 3-48. Specific carbon dioxide emissions were reduced from 906 g/kWh (1.997 lbm/kWh) for the Base Case to between 59-702 g/kWh (0.131-1.547 lbm/kWh) depending on CO_2 capture level for these cases. This corresponds to between 6.6% and 77.5% of the Base Case carbon dioxide emissions. Figure 3-35 and Table 3-48 indicate the quantity of CO_2 captured and the avoided CO_2 emissions.

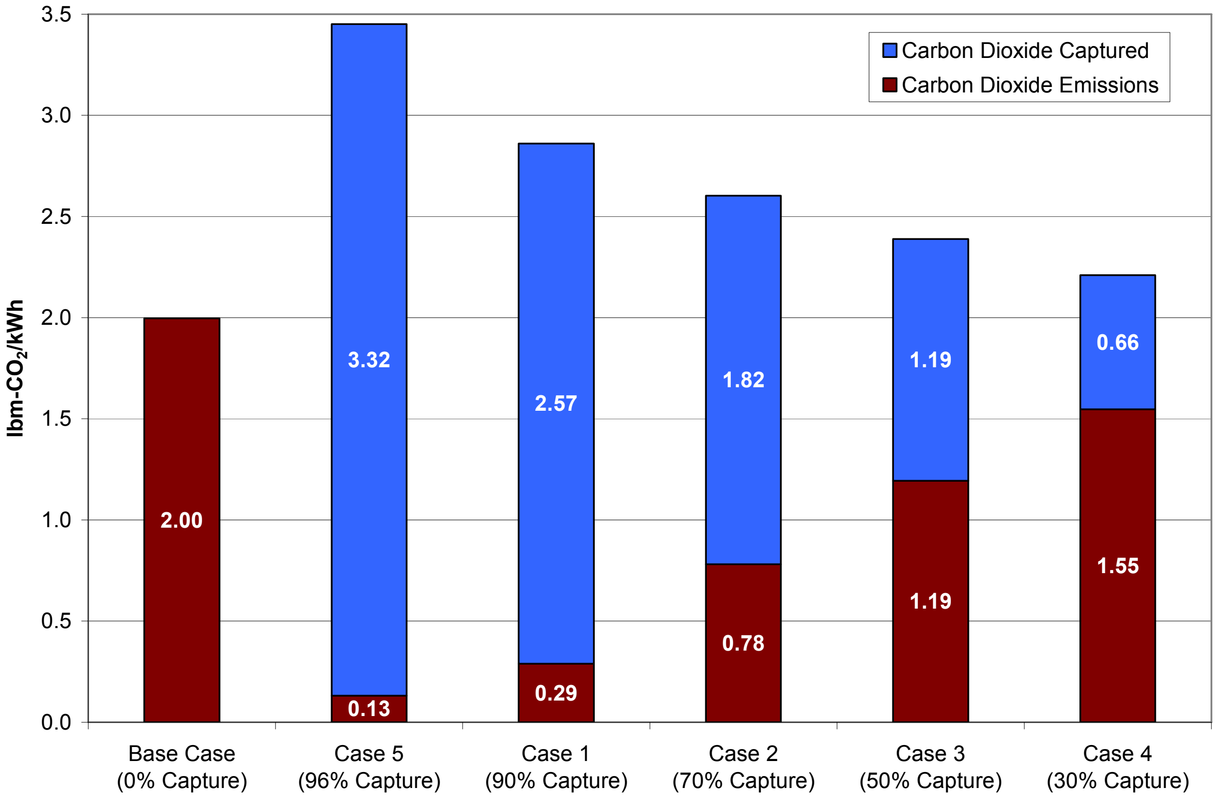


Figure 3-35: Carbon Dioxide Distribution

Figure 3-36 compares specific CO_2 emissions (lbm/kWh). Recovery of CO_2 ranged from 30% to 96% for the capture cases.

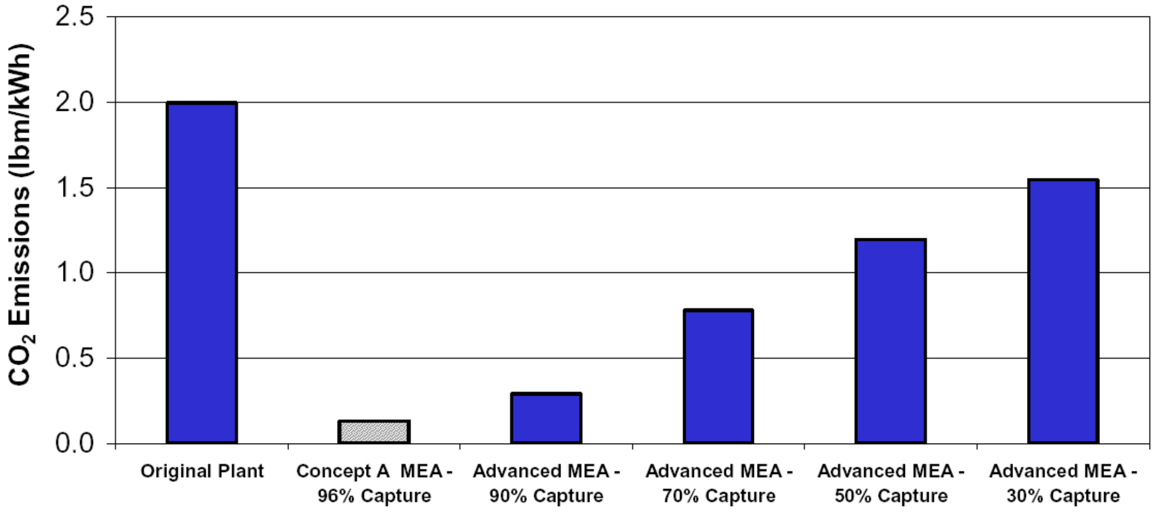


Figure 3-36: Specific Carbon Dioxide Emissions

Figure 3-37 shows these same CO_2 emission results plotted as a function of capture level.

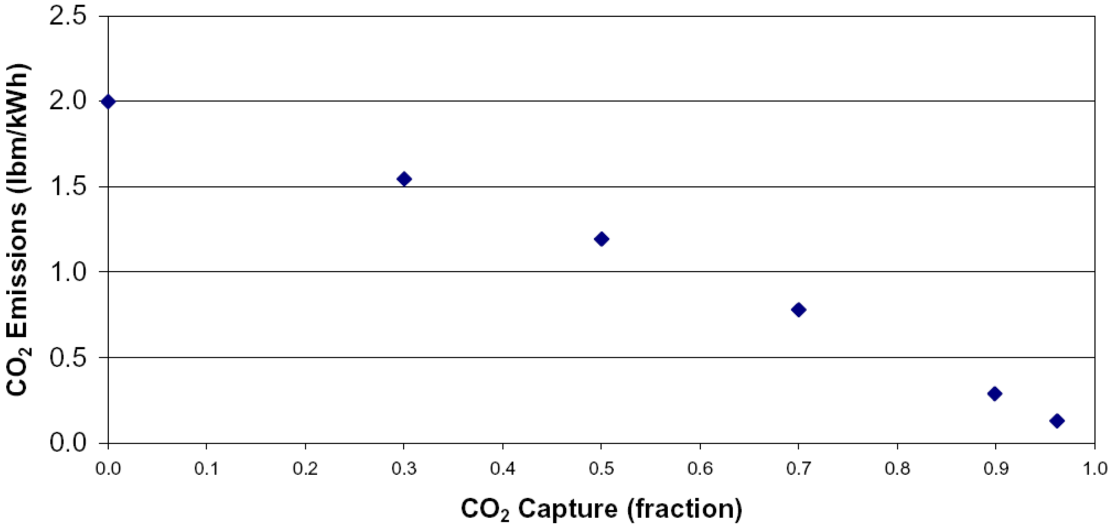


Figure 3-37: Specific Carbon Dioxide Emissions vs. CO_2 Capture Level

3.2.4 Steam Cycle Performance

The Base Case steam cycle is considered fairly typical of the U.S. fleet with subcritical steam conditions of 175 bara / 538°C / 538°C (2,535 psia / 1,000°F / 1,000°F). These represent the most common steam conditions for the existing fleet of U.S. electric utility power plant units in operation today. Six extraction feedwater heaters are used. The generator in this case produces 463,478 kWe.

The steam cycles for the five capture cases were all modified in a similar fashion. The steam cycles for the CO_2 capture cases differ from the Base Case steam cycle in that they each extract significant quantities of steam from the IP/LP crossover pipe. The extracted steam, at about 13.8 bara (200 psia) is expanded through a new “let down” steam turbine generating electric power before the steam is exhausted into the reboilers of the CO_2 recovery plant. The exhaust pressure of 4.5 bara (65 psia) was used for all the CO_2 capture cases (90%, 70%, 50%, and 30% capture) using the advanced amine of the current study (i.e., Cases 1-4).

Additionally, for Cases 1-4 of the current study, low-level heat was recovered from various areas of the CO_2 capture and compression system, and this heat was integrated with the steam cycle for overall plant efficiency improvement. This heat integration was possible in the current study because the CO_2 capture and compression equipment was able to be located relatively close to the existing unit. The absorbers were located near the existing Unit #5/6 common stack, and the strippers were located near the existing steam turbine. The CO_2 compressors were located as close as possible to the new strippers. In the previous study, all the CO_2 capture and compression equipment (absorbers, strippers, compressors, etc.) was located approximately 457 m (1,500 ft) northeast of the existing Conesville Unit #5/6 stack. Because of this relatively long distance, heat integration was determined to be impractical in the previous study.

The modified existing steam turbine generator of Case 5/Concept A, analyzed in the previous study, produces ~269 MWe and the new let down turbine produces ~62 MWe for a total generator output of ~331 MWe. The gross output for this case is reduced by ~132 MWe or about 30% as compared to the Base Case.

For Cases 1-4 of the current study using the advanced MEA solvent, the CO_2 capture levels are 90%, 70%, 50%, and 30% respectively, the modified existing steam turbine generator produces 343-426 MWe and the new letdown turbine produces 45-15 MWe for a total generator output of 388-441 MWe. The gross output is reduced by 23-75 MWe or 5%-17% for these cases. The total output is nearly a linear function of CO_2 recovery level. Figure 3-38 shows the total generator output for all the cases included in the study. The crosshatched bar shows the output of Case 5/Concept A of the previous study.

Comparison of total generator output for Case 5/Concept A from the original study (Bozzuto et al., 2001), and the advanced MEA 90% capture case of the current study indicates the impact of three primary differences between the designs as listed below:

- Reduced steam extraction required for the advanced MEA solvent regeneration
- Heat integration between the CO_2 capture/compression/liquefaction equipment and the existing steam/water cycle
- Reduced reboiler operating pressure

An improvement of about 57 MWe in total generator output is realized with the advanced MEA solvent case, which represents an improvement of about 17% on total generator output reduction. Correcting to a common CO_2 recovery percentage of ~96% would be expected to reduce this improvement to about 16%.

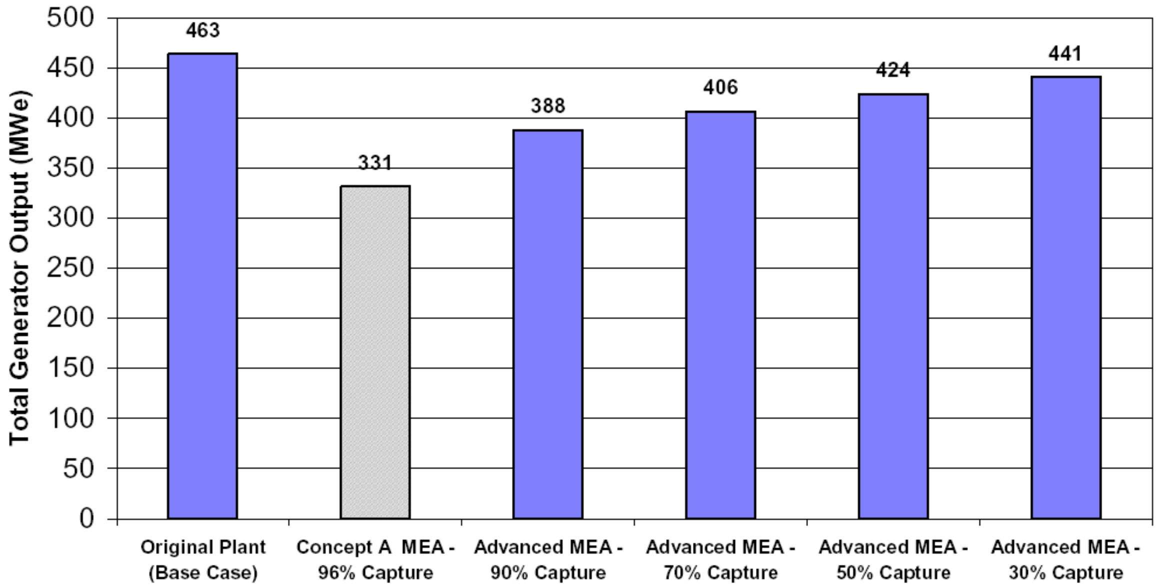


Figure 3-38: Total Generator Output (existing + new let down turbine generator)

3.2.5 Boiler Performance

The Base Case, updated Case 5/Concept A, and the four new CO_2 capture cases (Cases 1-4) were all analyzed based on the existing boiler producing a main steam flow of 395 kg/s (3,131,619 lbm/hr) at conditions of 538°C and 175 bara (1,000°F and 2,535 psia) at the steam turbine. This main steam flow represents the maximum continuous rating (MCR) for the existing unit. All six cases also provided reheat steam to the steam turbine at 538°C (1,000^{o} F). The boiler performance for the Base Case, updated Case 5/Concept A, and the four new CO_2 capture cases (Cases 1-4) was identical. Boiler efficiency for each of these six cases is 88.13%.

3.3 Cost Analysis

The project capital cost estimates (Total Investment Cost [TIC]) for all five cases, including engineering, procurement, and construction (EPC basis) and process and project contingencies, are presented in this section. All costs were estimated in July 2006 U.S. dollars. These costs include all required equipment to complete the retrofit such as the new advanced amine-based CO_2 scrubbing system, the new CO_2 compression, dehydration, and liquefaction system, the modified FGD system, the new let down steam turbine generator, and the existing steam cycle modifications.

Operating and maintenance (O&M) costs were calculated for all systems. The O&M costs for the Base Case (Conesville #5 Unit) were provided by American Electric Power (AEP). For the retrofit CO_2 capture system evaluations, additional O&M costs were calculated for the new equipment. The variable operating and maintenance (VOM) costs for the new equipment included such categories as chemicals and desiccants, waste handling, maintenance material and labor, contracted services, and make-up power cost (MUPC) from the reduction in net electricity production. The fixed operating and maintenance (FOM) costs for the new equipment includes operating labor only.

3.3.1 Cost Estimation Basis

The following assumptions were made in developing these cost estimates for each concept evaluated:

- July 2006 U.S. dollars
- Outdoor installation
- Investment in new utility systems is outside the scope
- CO_2 product pipeline is outside the scope
- No special limitations for transportation of large equipment
- No protection against unusual airborne contaminants (dust, salt, etc.)
- No unusual wind storms
- No earthquakes
- No piling required
- All releases can go to atmosphere – no flare provided
- CO_2 Product Pump designed to API standards, all other pumps conform to ANSI
- All heat exchangers designed to TEMA “C”
- All vessels are designed to ASME Section VIII, Div 1.

- Annual operating time is 7,446 hr/yr (85% capacity factor)
- The investment cost estimate was developed as a factored estimate based on in-house data for the major equipment. Such an estimate can be expected to have accuracy of +/- 30%.
- Process and project contingency were added to the EPC to derive the TIC.
- Make-up power cost was assessed at a 20-year levelized rate of 6.40 ¢/kWh (equivalent to a new Subcritical Pulverized Coal (Greenfield) Plant without carbon capture)
- No purchases of utilities or charges for shutdown time have been charged against the project

Other exclusions from the cost estimate are as follows:

- Soil investigation
- Environmental permits
- Disposal of hazardous or toxic waste
- Disposal of existing materials
- Custom's and Import duties
- Sales/use tax
- Forward escalation
- Capital spare parts
- Chemical loading facilities
- Buildings except for compressor building and electrical substation
- Financing cost
- Owners cost
- Guards during construction
- Site medical and ambulance service
- Cost & fees of authorities
- Overhead high voltage feed lines
- Cost to run a natural gas pipeline to the plant
- Excessive piling
- Contingency and risk

The costs used for consumption of fuel and chemicals in this project are shown in Table 3-49.

Table 3-49: Prices for Consumables

Consumables	($/lbm)	($/kg)
MEA	0.95	2.09
Soda Ash	0.26	0.56
Corrosion Inhibitor	3.00	6.61
Activated Carbon	1.00	2.20
Molecular Sieve	2.00	4.41
Diatomacious Earth	1.25	2.75
	($/$10^6$ Btu)	**($/GJ)**
Coal	1.80	1.90
Natural Gas	6.75	7.12

The project and process contingencies applied to the capital expenditures are shown in Table 3-50. The capital cost estimate provided for the CO_2 separation and compression system includes the let down steam turbine; therefore, the project and process contingency for carbon capture was applied to the let down steam turbine by default.

Table 3-50: Project and Process Contingencies

Capital Equipment	Project Contingency*	Process Contingency*
CO_2 Separation and Compression System	25%	18%
Flue Gas Desulfurization (FGD) System	11%	0%
Let Down Steam Turbine	25%	18%

*Percent of bare erected cost (i.e., sub-total direct cost in the investment tables for each case).

3.3.2 Carbon Dioxide Separation and Compression System Costs

This section shows both investment and operating and maintenance cost estimates for the Carbon Dioxide Separation and Compression Systems developed in this study. Five separate cost estimates for both the investment and O&M costs are provided in this section. There are four estimates provided for the 90%, 70%, 50%, and 30% CO_2 capture levels of the current study (Cases 1-4 respectively), which used an advanced amine. There is one additional cost estimate (Case 5) which is simply an update of Concept A (96% CO_2 capture) of the previous study (Bozzuto et al., 2001) to July 2006 U.S. dollars for comparison purposes. Case 5 used the Kerr McGee/ABB Lummus amine system.

3.3.2.1 Case 1 - 90% CO_2 Capture with Advanced Amine System

Investment Cost:

Table 3-51 shows investment costs for the CO_2 Separation and Compression System designed to capture 90% of the CO_2 contained in the Conesville #5 flue gas stream. Included in this table (Acc't. Code - 14200) are the steam cycle modification costs as well as the costs for the new let down turbine and associated electric generator. The steam cycle modifications were described previously in Section 3.1.3. The Total Investment Cost (TIC) of this equipment is $377,829,000. The expected level of accuracy for this cost estimate is +/-30%.

Table 3-51: Case 1 (90% Capture) CO_2 Separation and Compression System Investment Costs

Acc't Code	Description	Pieces	Direct Manhours	Labor ($1,000)	Material ($1,000)	Total ($1,000)	%
11000	Heaters						0.00%
11200	Exchangers & Aircoolers		25,200	466	19,049	19,515	5.17%
12000	Vessel / Filters		6,638	123	5,018	5,141	1.36%
12100	Towers / Internals		29,859	552	22,571	23,123	6.12%
12200	Reactors						0.00%
13000	Tanks						0.00%
14100	Pumps		4,431	82	3,350	3,432	0.91%
14200	Compressors		60,663	1,122	45,856	46,978	12.43%
18000	Special Equipment		5,070	94	3,833	3,926	1.04%
	Sub-Total Equipment	**140**	**131,862**	**2,439**	**99,676**	**102,115**	**27.03%**
21000	Civil		175,815	3,253	6,977	10,230	2.71%
21100	Site Preparation						0.00%
22000	Structures		46,152	854	4,087	4,941	1.31%
23000	Buildings		24,175	447	1,196	1,643	0.43%
30000	Piping		362,619	6,708	17,942	24,650	6.52%
40000	Electrical		186,804	3,456	7,974	11,430	3.03%
50000	Instruments		153,839	2,846	12,460	15,306	4.05%
61100	Insulation		131,862	2,439	5,183	7,623	2.02%
61200	Fireproofing		65,931	1,220	1,495	2,715	0.72%
61300	Painting		32,965	610	698	1,308	0.35%
	Sub-Total Commodities		**1,180,161**	**21,833**	**58,011**	**79,844**	**21.13%**
70000	Construction Indirects					35,228	9.32%
	Sub-Total Direct Cost (Bare Erected Cost)		**1,312,023**	**24,272**	**157,687**	**217,188**	**57.48%**
71000	Construction Management					2,000	0.53%
80000	Home Office Engineering					29,400	7.78%
80000	Basic Engineering					5,000	1.32%
95000	License Fee	Excluded					0.00%
19400	Vendor Reps					1,750	0.46%
19300	Spare parts					2,900	0.77%
80000	Training cost	Excluded					0.00%
80000	Commissioning	Excluded					0.00%
19200	Catalyst & Chemicals	Excluded					0.00%
97000	Freight					4,700	1.24%
96000	CGL / BAR Insurance						0.00%
91400	Escalation to July 2006 Dollars					7,200	1.91%
	Total Base Cost					**270,138**	**71.50%**
	Contractors Fee					14,300	3.78%
	Total (EPC):					**284,438**	**75.28%**
93000	Project Contingency					54,297	14.37%
93000	Process Contingency					39,094	10.35%
	Total Investment Cost (TIC):					**377,829**	**100.00%**

Exclusions: bonds, taxes, import duties, hazardous material handling & disposal, capital spare parts, catalyst & chemicals, commissioning and initial operations, buildings other than control room & MCC.

Operating and Maintenance Cost:

Table 3-52 shows O&M costs for the CO_2 Separation and Compression System for the 90% CO_2 Capture Case. The variable, feedstock, and make-up power costs are reported at the 85% capacity factor. The make-up power cost represents the levelized cost over a 20-year period. All other costs represent first year operating costs.

Table 3-52: Case 1 (90% Capture) CO_2 Separation and Compression System Operating & Maintenance Costs

Operating & Maintenance Costs	Subtotal ($1000/yr)	Total ($1000/yr)
Fixed O&M Costs		2,494
Operating Labor	2,494	
Variable O&M Costs		17,645
Chemicals	10,161	
Waste Handling & Contracted Services	767	
Maintenance (Materials and Labor)	6,716	
Feedstock O&M Costs		653
Natural Gas	653	
Levelized, Make-up Power Cost		62,194
Levelized, Make-up Power Cost (@ $6.40 ¢/kWh)	62,194	

3.3.2.2 Case 2 - 70% CO_2 Capture with Advanced Amine System

Investment Cost:

Table 3-53 shows investment costs for the CO_2 Separation and Compression System designed to capture 70% of the CO_2 contained in the Conesville #5 flue gas stream. Included in the table (Acc't. Code - 14200) are the steam cycle modification costs and the costs for the new let down turbine and associated electric generator. The steam cycle modifications were described previously in Section 3.1.3. The Total Investment Cost (TIC) of this equipment is $342,805,000. The expected level of accuracy for this cost estimate is +/-30%.

Table 3-53: Case 2 (70% Capture) CO_2 Separation and Compression System Investment Costs

Acc't Code	Description	Pieces	Direct Manhours	Labor ($1,000)	Material ($1,000)	Total ($1,000)	%
11000	Heaters						0.00%
11200	Exchangers & Aircoolers		20,664	382	15,620	16,002	4.67%
12000	Vessel / Filters		5,605	104	4,237	4,340	1.27%
12100	Towers / Internals		26,482	490	20,018	20,508	5.98%
12200	Reactors						0.00%
13000	Tanks						0.00%
14100	Pumps		3,402	63	2,572	2,635	0.77%
14200	Compressors		57,726	1,068	43,636	44,704	13.04%
18000	Special Equipment		4,841	90	3,659	3,749	1.09%
	Sub-Total Equipment	**133**	**118,720**	**2,197**	**89,742**	**91,938**	**23.37%**
21000	Civil		158,293	2,928	6,282	9,210	2.69%
21100	Site Preparation						0.00%
22000	Structures		41,552	769	3,679	4,448	1.30%
23000	Buildings		21,765	403	1,077	1,480	0.43%
30000	Piping		326,480	6,040	16,154	22,193	6.47%
40000	Electrical		168,187	3,111	7,179	10,291	3.00%
50000	Instruments		138,507	2,562	11,218	13,780	4.02%
61100	Insulation		118,720	2,196	4,667	6,863	2.00%
61200	Fireproofing		59,360	1,098	1,346	2,444	0.71%
61300	Painting		29,680	549	628	1,177	0.34%
	Sub-Total Commodities		**1,062,544**	**19,656**	**52,230**	**71,886**	**20.97%**
70000	Construction Indirects					31,717	9.25%
	Sub-Total Direct Cost (Bare Erected Cost)		**181,263**	**21,853**	**141,972**	**195,542**	**57.04%**
71000	Construction Management					2,000	0.58%
80000	Home Office Engineering					27,930	8.15%
80000	Basic Engineering					5,000	1.46%
95000	License Fee	Excluded					0.00%
19400	Vendor Reps					1,750	0.51%
19300	Spare parts					2,600	0.76%
80000	Training cost	Excluded					0.00%
80000	Commissioning	Excluded					0.00%
19200	Catalyst & Chemicals	Excluded					0.00%
97000	Freight					4,300	1.25%
96000	CGL / BAR Insurance						0.00%
91400	Escalation to July 2006 Dollars					6,600	1.93%
	Total Base Cost					**245,722**	**71.68%**
	Contractors Fee					13,000	3.79%
	Total (EPC):					**258,722**	**75.47%**
93000	Project Contingency					48,886	14.26%
93000	Process Contingency					35,198	10.27%
	Total Investment Cost (TIC):					**342,805**	**100.00%**

Exclusions: bonds, taxes, import duties, hazardous material handling & disposal, capital spare parts, catalyst & chemicals, commissioning and initial operations, buildings other than control room & MCC.

Operating and Maintenance Cost:

Table 3-54 shows O&M costs for the CO_2 Separation and Compression System for the 70% CO_2 Capture Case. The variable, feedstock, and make-up power costs are reported at the 85% capacity factor. The make-up power cost represents the levelized cost over a 20-year period. All other costs represent first year operating costs.

Table 3-54: Case 2 (70% Capture) CO_2 Separation and Compression System Operating & Maintenance Costs

Operating & Maintenance Costs	Subtotal ($1000/yr)	Total ($1000/yr)
Fixed O&M Costs		2,284
Operating Labor	2,284	
Variable O&M Costs		14,711
Chemicals	8,005	
Waste Handling & Contracted Services	597	
Maintenance (Materials and Labor)	6,109	
Feedstock O&M Costs		488
Natural Gas	488	
Levelized, Make-up Power Cost		47,926
Levelized, Make-up Power Cost (@ $6.40 ¢/kWh)	47,926	

3.3.2.3 Case 3 – 50% CO_2 capture with Advanced Amine Systems

Investment Cost:

Table 3-55 shows investment costs for the CO_2 Separation and Compression System designed to capture 50% of the CO_2 contained in the Conesville #5 flue gas stream. Included in this table (Acc't. Code - 14200) are the steam cycle modification costs as well as the costs for the new let down turbine and associated electric generator. The steam cycle modifications were described previously in Section 3.1.3. The Total Investment Cost (TIC) of this equipment is $258,390,000. The expected level of accuracy for this cost estimate is +/-30%.

Table 3-55: Case 3 (50% Capture) CO_2 Separation and Compression System Investment Costs

Acc't Code	Description	Pieces	Direct Manhours	Labor ($1,000)	Material ($1,000)	Total ($1,000)	%
11000	Heaters						0.00%
11200	Exchangers & Aircoolers		15,864	293	11,992	12,285	4.75%
12000	Vessel / Filters		4,051	75	3,063	3,137	1.21%
12100	Towers / Internals		23,202	429	17,538	17,968	6.95%
12200	Reactors						0.00%
13000	Tanks						0.00%
14100	Pumps		2,776	51	2,098	2,150	0.83%
14200	Compressors		38,200	707	28,876	29,583	11.45%
18000	Special Equipment		3,864	71	2,921	2,992	1.16%
	Sub-Total Equipment	**107**	**87,957**	**1,626**	**66,488**	**68,115**	**17.32%**
21000	Civil		117,276	2,170	4,654	6,824	2.64%
21100	Site Preparation						0.00%
22000	Structures		30,785	570	2,726	3,296	1.28%
23000	Buildings		16,126	298	798	1,096	0.42%
30000	Piping		241,883	4,475	11,968	16,443	6.36%
40000	Electrical		124,606	2,305	5,319	7,624	2.95%
50000	Instruments		102,617	1,898	8,311	10,209	3.95%
61100	Insulation		87,957	1,627	3,457	5,085	1.97%
61200	Fireproofing		43,979	814	997	1,811	0.70%
61300	Painting		21,989	407	465	872	0.34%
	Sub-Total Commodities		**787,218**	**14,564**	**38,695**	**53,260**	**20.61%**
70000	Construction Indirects					23,498	9.09%
	Sub-Total Direct Cost (Bare Erected Cost)					**144,874**	**56.07%**
71000	Construction Management					2,000	0.77%
80000	Home Office Engineering					22,470	8.70%
80000	Basic Engineering					5,000	1.94%
95000	License Fee	Excluded					0.00%
19400	Vendor Reps					1,750	0.68%
19300	Spare parts					1,900	0.74%
80000	Training cost	Excluded					0.00%
80000	Commissioning	Excluded					0.00%
19200	Catalyst & Chemicals	Excluded					0.00%
97000	Freight					3,200	1.24%
96000	CGL / BAR Insurance						0.00%
91400	Escalation to July 2006 Dollars					5,000	1.94%
	Total Base Cost					**186,194**	**72.06%**
	Contractors Fee					9,900	3.83%
	Total (EPC):					**196,094**	**75.89%**
93000	Project Contingency					36,219	14.02%
93000	Process Contingency					26,077	10.09%
	Total Investment Cost (TIC):					**258,390**	**100.00%**

Exclusions: bonds, taxes, import duties, hazardous material handling & disposal, capital spare parts, catalyst & chemicals, commissioning and initial operations, buildings other than control room & MCC.

Operating and Maintenance Cost:

Table 3-56 shows O&M costs for the CO_2 Separation and Compression System for the 50% CO_2 Capture Case. The variable, feedstock, and make-up power costs are reported at the 85% capacity factor. The make-up power cost represents the levelized cost over a 20-year period. All other costs represent first year operating costs.

Table 3-56: Case 3 (50% Capture) CO_2 Separation and Compression System Operating & Maintenance Costs

Operating & Maintenance Costs	Subtotal ($1000/yr)	Total ($1000/yr)
Fixed O&M Costs		2,079
Operating Labor	2,079	
Variable O&M Costs		10,876
Chemicals	5,820	
Waste Handling & Contracted Services	426	
Maintenance (Materials and Labor)	4,630	
Feedstock O&M Costs		337
Natural Gas	337	
Levelized, Make-up Power Cost		33,768
Levelized, Make-up Power Cost (@ $6.40 ¢/kWh)	33,738	

3.3.2.4 Case 4 - 30% CO_2 Capture with Advanced Amine System

Investment Cost:

Table 3-57 shows investment costs for the CO_2 Separation and Compression System designed to capture 30% of the CO_2 contained in the Conesville #5 flue gas stream. Included in this table (Acc't. Code - 14200) are the steam cycle modification costs as well as the costs for the new let down turbine and associated electric generator. The steam cycle modifications were described previously in Section 3.1.3. The Total Investment Cost (TIC) of this equipment is $189,570,000. The expected level of accuracy for this cost estimate is +/-30%.

Table 3-57: Case 4 (30% Capture) CO_2 Separation and Compression System Investment Costs

Acc't Code	Description	Pieces	Direct Manhours	Labor ($1,000)	Material ($1,000)	Total ($1,000)	%
11000	Heaters						0.00%
11200	Exchangers & Aircoolers		10,123	187	7,652	7,839	4.14%
12000	Vessel / Filters		2,413	45	1,824	1,869	0.99%
12100	Towers / Internals		12,745	236	9,634	9,870	5.21%
12200	Reactors						0.00%
13000	Tanks						0.00%
14100	Pumps		1,728	32	1,306	1,338	0.71%
14200	Compressors		34,761	643	26,276	26,919	14.20%
18000	Special Equipment		2,137	40	1,615	1,655	0.87%
	Sub-Total Equipment	**65**	**63,907**	**1,183**	**48,307**	**49,490**	**12.58%**
21000	Civil		85,208	1,576	3,382	4,958	2.62%
21100	Site Preparation						0.00%
22000	Structures		22,367	414	1,981	2,394	1.26%
23000	Buildings		11,716	217	580	796	0.42%
30000	Piping		175,742	3,251	8,695	11,947	6.30%
40000	Electrical		90,534	1,675	3,865	5,539	2.92%
50000	Instruments		74,557	1,379	6,038	7,418	3.91%
61100	Insulation		63,906	1,182	2,512	3,694	1.95%
61200	Fireproofing		31,953	591	725	1,316	0.69%
61300	Painting		15,977	296	338	634	0.33%
	Sub-Total Commodities		**101,185**	**10,581**	**28,116**	**38,696**	**20.41%**
70000	Construction Indirects					17,073	9.01%
	Sub-Total Direct Cost (Bare Erected Cost)		**635,868**	**11,764**	**76,423**	**105,259**	**55.53%**
71000	Construction Management					2,000	1.06%
80000	Home Office Engineering					15,600	8.23%
80000	Basic Engineering					5,000	2.64%
95000	License Fee	Excluded					0.00%
19400	Vendor Reps					1,750	0.92%
19300	Spare parts					1,400	0.74%
80000	Training cost	Excluded					0.00%
80000	Commissioning	Excluded					0.00%
19200	Catalyst & Chemicals	Excluded					0.00%
97000	Freight					2,300	1.21%
96000	CGL / BAR Insurance						0.00%
91400	Escalation to July 2006 Dollars					3,700	1.95%
	Total Base Cost					**137,009**	**72.27%**
	Contractors Fee					7,300	3.85%
	Total (EPC):					**144,309**	**76.12%**
93000	Project Contingency					26,315	13.88%
93000	Process Contingency					18,947	9.99%
	Total Investment Cost (TIC):					**189,570**	**100.00%**

Exclusions: bonds, taxes, import duties, hazardous material handling & disposal, capital spare parts, catalyst & chemicals, commissioning and initial operations, buildings other than control room & MCC.

Operating and Maintenance Cost:

Table 3-58 shows O&M costs for the CO_2 Separation and Compression System for the 30% CO_2 Capture Case. The variable, feedstock, and make-up power costs are reported at the 85% capacity factor. The make-up power cost represents the levelized cost over a 20-year period. All other costs represent first year operating costs.

Table 3-58: Case 4 (30% Capture) CO_2 Separation and Compression System Operating & Maintenance Costs

Operating & Maintenance Costs	Subtotal ($1000/yr)	Total ($1000/yr)
Fixed O&M Costs		1,869
Operating Labor	1,869	
Variable O&M Costs		7,019
Chemicals	3,408	
Waste Handling & Contracted Services	256	
Maintenance (Materials and Labor)	3,355	
Feedstock O&M Costs		211
Natural Gas	211	
Levelized, Make-up Power Cost		19,885
Levelized, Make-up Power Cost (@ $6.40 ¢/kWh)	19,885	

3.3.2.5 Case 5/Concept A – 96% Capture with Kerr McGee/ABB Lummus amine system (costs updated from previous study)

Investment Cost:

Table 3-59 shows investment costs for the Case 5/Concept A CO_2 Separation and Compression System, which uses the Kerr McGee/ABB Lummus amine system. The costs shown in this table are the costs from the 2000 study (Bozzuto et al., 2001) escalated to 2006 dollars (1.3017 escalation factor). Included in this table (Acc't. Code - 14200) are the steam cycle modification costs as well as the new let down turbine and associated electric generator. The steam cycle modifications were described in Section 3.1.3. The Total Investment Cost (TIC) of this equipment is $678,792,517. The expected level of accuracy for this cost estimate is +/- 30%.

Table 3-59: Case 5/Concept A (96% Capture) CO_2 Separation and Compression System Investment Costs

Acc't Code	Description	Pieces	Direct Manhours	Labor ($1,000)	Material ($1,000)	Total ($1,000)	%
11000	Heaters						0.00%
11200	Exchangers & Aircoolers		44,970	907	37,074	37,981	5.60%
12000	Vessel / Filters		5,776	117	4,762	4,879	0.72%
12100	Towers / Internals		43,200	872	35,615	36,487	5.38%
12200	Reactors						0.00%
13000	Tanks						0.00%
14100	Pumps		10,078	203	8,309	8,512	1.25%
14200	Compressors		100,925	2,036	83,203	85,239	12.56%
18000	Special Equipment		10,991	221	9,061	9,282	1.37%
	Sub-Total Equipment	**436**	**215,940**	**4,357**	**178,023**	**182,380**	**26.87%**
21000	Civil		287,919	5,809	12,461	18,271	2.69%
21100	Site Preparation						0.00%
22000	Structures		75,579	1,524	7,299	8,823	1.30%
23000	Buildings		39,589	799	2,136	2,935	0.43%
30000	Piping		593,833	11,981	32,044	44,025	6.49%
40000	Electrical		305,914	6,173	14,242	20,415	3.01%
50000	Instruments		251,929	5,083	22,253	27,336	4.03%
61100	Insulation		215,939	4,357	9,258	13,614	2.01%
61200	Fireproofing		107,970	2,179	2,670	4,849	0.71%
61300	Painting		53,985	1,090	1,246	2,335	0.34%
	Sub-Total Commodities		**1,932,657**	**38,995**	**103,608**	**142,603**	**21.01%**
70000	Construction Indirects					62,928	9.27%
	Sub-Total Direct Cost (Bare Erected Cost)					**387,911**	**57.15%**
71000	Construction Management					2,603	0.38%
80000	Home Office Engineering					57,889	8.53%
80000	Basic Engineering					6,509	0.96%
95000	License Fee	Excluded					0.00%
19400	Vendor Reps					3,254	0.48%
19300	Spare parts					5,207	0.77%
80000	Training cost	Excluded					0.00%
80000	Commission	Excluded					0.00%
19200	Catalyst & Chemicals	Excluded					0.00%
97000	Freight					1,432	0.21%
96000	CGL / BAR Insurance					8,461	1.25%
91400	Escalation to July 2001 Dollars					13,017	1.92%
	Total Base Cost					**486,283**	**71.64%**
	Contractors Fee					25,709	3.79%
	Total (EPC):					**511,991**	**75.43%**
93000	Project Contingency					96,978	14.29%
93000	Process Contingency					69,824	10.29%
	Total Investment Cost (TIC):					**678,793**	**100.00%**

Exclusions: bonds, taxes, import duties, hazardous material handling & disposal, capital spare parts, catalyst & chemicals, commissioning and initial operations, buildings other than control room & MCC.

Operating and Maintenance Cost:

Table 3-60 shows O&M costs for the Case 5/Concept A CO_2 Separation and Compression System, which captures 96% of the carbon dioxide from the Conesville #5 flue gas stream. They amount to $132,809,000/yr.

Table 3-60: Case 5/Concept A (96% Capture) CO_2 Separation and Compression System Operating & Maintenance Costs

Operating & Maintenance Costs	Subtotal ($1000/yr)	Total ($1000/yr)
Fixed O&M Costs		2,488
Operating Labor	2,488	
Variable O&M Costs		18,640
Chemicals	4,870	
Waste Handling & Contracted Services	843	
Maintenance (Materials and Labor)	$12,927	
Feedstock O&M Costs		890
Natural Gas	890	
Levelized, Make-up Power Cost		86,832
Levelized, Make-up Power Cost (@ $6.40 ¢/kWh)	86,832	

3.3.3 Boiler Modification Costs

For this project the Boiler Scope is defined as everything on the gas side upstream of the FGD System. Therefore, it includes equipment such as the steam generator, pulverizers, fans, ductwork, electrostatic precipitator (ESP), air heater, coal and ash handling systems, etc. Purposely not included in the boiler scope definition is the FGD system. The FGD system modification costs are shown separately in Section 3.3.4. For all the capture options investigated in this study (Cases 1-5), Boiler Scope is not modified from the Base Case configuration and, as such, there are no costs in this category.

3.3.4 Flue Gas Desulfurization System Modification Costs

Flue Gas Desulfurization System modification costs for these CO_2 capture options are relatively minor as compared to the other new equipment required. The Flue Gas Desulfurization System modifications, which include the addition of a secondary absorber island, building, booster fan, and ductwork, are described in Section 3.1.3. The total cost required for the Flue Gas Desulfurization (FGD) System scope modifications is $15,800,000 in January 2000 dollars. At an escalation rate of 4.12% per year for this type of equipment (Oil & Gas Journal, 2006), in July 2006 dollars EPC cost, is $20,540,000 ($[15,800,000 * 1.0412]^{6.5}$). The bare erected cost of the FGD System was estimated to be $15,680,000 in July 2006 dollars. An 11% project contingency was added to the the FGD System cost, therefore, the TIC contribution is $22,264,800. This cost is applied to all the capture options investigated in this study (i.e., Cases 1-5). This estimate includes material, engineering and construction. The expected level of accuracy for this cost estimate is +/- 10%.

3.3.5 Let Down Steam Turbine/Generator Costs

The MEA systems require significant quantities of heat for regeneration of the MEA solvent. Low-pressure steam is extracted from the existing turbine to provide the energy for solvent regeneration. The steam extraction location is the existing turbine IP/LP crossover pipe. This steam is expanded from ~200 psia to 65 psia for Case 5 or 47 psia for Cases 1-4 through a new “Let down” steam turbine/generator where electricity is produced. The exhaust steam leaving the new let down turbine provides the heat source for solvent regeneration in the reboilers of the MEA CO_2 recovery system. Table 3-61 shows the investment costs for the let down steam turbine generator (D&R cost basis). Although the costs shown for these turbines are on a D&R (Delivered and Representative) basis, construction costs and other balance of plant costs associated with these turbines are included for each case as a part of the CO_2 Separation and Compression System Investment Costs shown in Section 3.3.2.

Table 3-61: Let Down Turbine Generator Costs and Electrical Outputs for Cases 1-5 (D&R Cost Basis)

Let Down Steam Turbine Costs (D&R Basis)	OCDO-A updated	Current Study			
CO_2 Capture Percentage	96% (Case-5)	90% (Case-1)	70% (Case-2)	50% (Case-3)	30% (Case-4)
Generator Cost (10^3 $)	10,516	9,800	9,400	8,900	8,500
Generator Output (kWe)	62,081	45,321	35,170	25,031	14,898

3.3.6 Charges for Loss of Power During Construction

During the construction period for the new equipment, it is assumed the existing Conesville Unit #5 power plant will be operated in its normal way. The new CO_2 capture equipment is being located in three separate locations (see Appendix I for plant layout drawings), and it is assumed that the erection of this equipment will not impede the operation of Conesville Unit #5 or any of the other units on site. Once construction is completed, it has been assumed that the final connections between the CO_2 capture systems and the existing power plant can be completed during the annual outage for the unit. Final shakedown testing will be completed after the outage. Therefore, there are no charges for loss of power during construction.

3.3.7 Summary of Total Retrofit Investment Costs

Table 3-62 summarizes the total retrofit investment costs (TIC Basis) required for each of the five cases. The first column shows the costs for updated Case 5/Concept A from the previous study (Bozzuto et al., 2001), which captures ~96% of the CO_2. The last four columns show the costs for the current study (Cases 1-4) using the advanced MEA system. The costs include specific costs ($/kWe) on both a new and original kWe basis.

Table 3-62: Total Retrofit Investment Costs (Cases 1-5)

Retrofit Cost Summary ($1000)	2001 Study Updated	Current Study			
CO_2 Capture Percentage	96% (Case-5)	90% (Case-1)	70% (Case-2)	50% (Case-3)	30% (Case-4)
CO_2 Separation and Compression System	668,277	368,029	333,406	249,490	181,070
Flue Gas Desulfurization System	22,265	22,265	22,265	22,265	22,265
Let Down Steam Turbine Generator*	10,516	9,800	9,400	8,900	8,500
Boiler Modifications					
Total Retrofit/Investment Cost (i.e., TIC):	701,057	400,094	365,070	280,655	211,835
$/kW-new:	2,786	1,319	1,095	773	540
$/kW-original:	1,616	922	842	647	488

*Engineering, construction management, overhead, fees, and contingency are Included in the CO_2 separation and compression system cost.

Figure 3-39 shows the specific investment costs ($/kWe) for each case. Two costs are plotted for each of the cases in this figure. The upper curve specific costs are relative to the new plant output, which is lower than the original (Base Case) due to added auxiliary power and reduced steam turbine output. The lower curve specific costs are relative to the original plant output of the Base Case.

By comparing the cost for the 96% capture case from the previous study with the cost for the 90% capture case from the current study, as shown in Figure 3-39s a significant cost reduction is indicated for the current study. The current study specific costs ($/kWe-new) are about half of what the updated previous study (96% capture case) results indicate. It should be pointed out that if Case 5 (~96% recovery) was designed as a part of the current study, it would likely have equipment selections similar to Case 1 (90% recovery) and therefore significant cost reductions and improved economics would result.

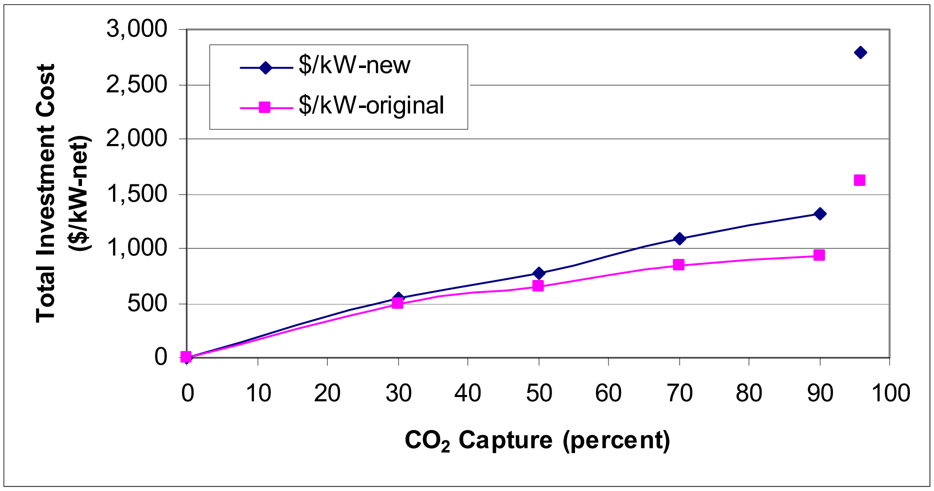


Figure 3-39: New Equipment Specific Investment Costs

The specific costs for the current study cases (Cases 1-4) are nearly a linear function of CO_2 recovery percentage. However, some economy of scale effects and other non-linearitys are evident. To help understand this non-linearity, a brief review of equipment selection is necessary.

Table 3-63 shows a summary of the major equipment selected for the CO_2 Removal, Compression, and Liquefaction Systems for all five cases. Three categories are shown in this table (Compressors, Towers/Internals, and Heat Exchangers). These three categories represent the three most costly accounts in the cost estimates for these systems. These accounts represent ~90% of the total equipment costs for these systems. A review of this table shows how the number of compression trains is reduced from two trains, for the 90% and 70% recovery cases, to one train for the 50% and 30% recovery cases. Similarly, the number of absorber/stripper trains is reduced from two trains for the 90%, 70%, and 50% recovery cases to one train for the 30% recovery. The heat selections show even more variation between the cases. Equipment sizes are also indicated in this table.

Table 3-63: CO_2 Removal, Compression, and Liquefaction System Equipment Summary (Cases 1-5)

	90% (Case-1)		70% (Case-2)		50% (Case-3)		30% (Case-4)		96% (Case-5)	
Compressors	**No.**	**HP ea**	**No.**	**HP ea**	**No.**	**HP ea**	**No.**	**HP ea**	**No.**	**HPea**
CO_2 Compressor	2	15,600	2	12,100 1		17,300	1	10,400	7	4,500
Propane Compressor	2	11,700	2	10,200 1		14,600	1	8,800	7	3,100
LP Let Down Turbine	1	60,800	1	47,200 1		33,600	1	20,000	1	82,300
Towers/Internals	**No.**	**ID/Height (ft)**	**No.**	**ID/Height (ft)**	**No.**	**ID/Height (ft)**	**No.**	**ID/Height (ft)**	**No.**	**ID/Height (ft)**
Absorber/Cooler	2	34/126	2	30/126 2		25/126	1	28/126	5	27/126
Stripper	2	22/50	2	19/50 2		16/50	1	20/50	9	16/50
Heat Exchangers	**No.**	**10^6 Btu/hr ea**	**No.**	**10^6 Btu/hr ea**	**No.**	**10^6 Btu/hr ea**	**No.**	**10^6 Btu/hr ea**	**No.**	**10^6 Btu/hr ea**
Reboilers	10	120.0	8	120.0 6		120.0	4	120.0	9	217.0
Solvent Stripper CW Condenser	12	20.0	10	20.0	7	20.0	4	20.0	9	42.0
Other Heat Exchangers / Average Duty	36	61.0	35	57.0 25		62.0	16	58.0	113	36.0
Total Heat Exchangers / Average Duty	58	62.7	53	59.5 38		63.4	24	62.0	131	48.8

It should also be noted, as shown in Table 3-63, that the design of Case 5 (See Bozzuto et al., 2001) is not totally consistent with the design of Case 1 done in the current study, although the CO_2 recovery in each case is similar. Case 1 uses two (2) absorber trains, two stripper trains, and two compression trains. Conversely, Case 5, which was designed in 2000, used five absorber trains, nine stripper trains, and seven compression trains. Because of these differences, Case 1 is able to take advantage of economy of scale effects for equipment cost with the larger equipment sizes used in each train as compared to Case 5. Additionally, Case 5 equipment was all located about 457 m (1,500 ft) from the Unit #5/6 common stack, which also contributed to the increased the cost of Case 5 relative to Case 1.

All the costs shown above were used in the economic evaluation (Section 3.4) to develop incremental Cost of Electricity (COE) values and CO_2 mitigation cost comparisons.

3.4 Economic Analysis

A comprehensive economic evaluation comparing the Base Case study unit and various retrofit CO_2 capture scenarios using an advanced amine was performed. The purpose of the evaluation was to quantify the impact of CO_2 capture on the Cost of Electricity (COE) for this existing coal-fired unit. CO_2 mitigation costs were also determined in this analysis. The economic evaluation results are presented as incremental Costs of Electricity (levelized basis). The reported costs of electricity are incremental relative to the Base Case (air fired without CO_2 capture, i.e., business as usual).

Additionally, economic sensitivity studies were developed for each of the CO_2 capture options to highlight which parameters affected the incremental COE and CO_2 mitigation cost to the greatest extents. The sensitivity parameters chosen (Investment Cost, Capacity Factor, Make-up Power Cost, and CO_2 Selling Price) were judged to be the most important parameters to vary for this project. These parameters are either site-specific or there is uncertainty in their values in looking to the future. Therefore, proper use of the sensitivity results could potentially allow extrapolation of results for application to units other than the selected study unit (Conesville Unit #5).

The economic analysis was performed by Research and Development Solutions, Inc. (RDS) using the levelized revenue requirement method (a form of discounted cash flow analysis). The model has the capability to analyze the economic effects of different technologies based on differing capital costs, operating and maintenance costs, fuel costs, and cost of capital assumptions. The primary metrics are levelized cost of electricity (LCOE) and CO_2 mitigation cost. Both are reported on an incremental cost of CO_2 capture basis within this study. All cost data were provided by Alstom (see Section 3.3).

3.4.1 Economic Study Scope and Assumptions

A total of five CO_2 capture cases were evaluated in this economic analysis in addition to the Base Case without CO_2 capture:

- Case 1: 90% CO_2 capture with advanced “State of the Art” amine
- Case 2: 70% CO_2 capture with advanced “State of the Art” amine
- Case 3: 50% CO_2 capture with advanced “State of the Art” amine
- Case 4: 30% CO_2 capture with advanced “State of the Art” amine
- Case 5: 96% CO_2 capture with Kerr-McGee/ABB Lummus amine technology (ca. 2000)

Case 5 is simply an update of Concept A of the previous study (Bozzuto et al., 2001). As shown in Section 3.3.2.5, the investment and O&M costs of Concept A of the previous study were updated to July 2006 U.S. dollars. This information was used to update the economic analysis of Case 5 to a common basis with Cases 1-4.

The primary outputs from this economic analysis are the incremental Levelized Cost of Electricity (LCOE) and CO_2 mitigation costs relative to the Base Case. These two measures of economic merit were determined for all cases evaluated.

Incremental LCOE was calculated using a simplified model derived from the NETL Power Systems Financial Model for calculating levelized cost of electricity.[3] Total Plant Cost (TPC) was replaced with Total Investment Cost (TIC) to reflect the retrofit analyzed within this study. The term “Incremental COE” and “LCOE” are used synonymously within this report. The following equation was used to calculate the LCOE over a 20-year period.

$$LCOE_P = \text{levelized annual capital charge} + \text{levelized annual operating costs}$$

$$\mathbf{LCOE_P} = \frac{\mathbf{(CCF_P)(TPC) + [(LF_{F1})(OC_{F1}) + (LF_{F2})(OC_{F2}) + \ldots] + (CF)[(LF_{V1})(OC_{V1}) + (LF_{V2})(OC_{V2}) + \ldots]}}{\mathbf{(CF)(KWH)}}$$

Where:

LCOE = levelized cost of electricity over P years
P = levelization period (e.g., 10, 20, or 30 years)
CCF = capital charge factor for a levelization period of P years
TIC = total investment cost [the sum of bare erected costs (includes costs of process equipment, supporting facilities, direct and indirect labor), detailed design costs, construction/project management costs, project contingency, process contingency and technology fees]
LF_{Fn} = levelization factor for category n fixed operating cost
OC_{Fn} = category n fixed operating cost for the initial year of operation (but expressed in “first-year-of-construction” year dollars)
CF = plant capacity factor
LF_{Vn} = levelization factor for category n variable operating cost
OC_{Vn} = category n variable operating cost at 100% capacity factor for the initial year of operation (but expressed in “first-year-of-construction” year dollars)
KWH = annual net kilowatt-hours of power generated at 100% capacity factor

All costs are expressed in “first-year-of-construction” year dollars, and the resulting LCOE is also expressed in “first-year-of-construction” year dollars (January 2007). CO_2 mitigation and capture costs were calculated according to the following equations.

$$\mathbf{CO_2\ Mitigation\ Cost} = (LCOE_{Cp} - LCOE_{Ref}) / (CO_{2Ref\ emitted} - CO_{2Cp\ emitted})$$

$$\mathbf{CO_2\ Captured\ Cost} = (LCOE_{Cp} - LCOE_{Ref}) / (CO_{2Cp\ produced} - CO_{2Cp\ emitted})$$

Where:

CO_2 Mitigation Cost = \$/ton of CO_2 avoided
CO_2 Captured Cost = \$/ton of CO_2 removed
CO_2 = Carbon dioxide (tons/kWh at plant capacity factor)
LCOE = Levelized cost of electricity (\$/kWh)
$_{Cp}$ = Capture plant
$_{Ref}$ = Reference plant

---

[3] Power Systems Financial Model Version 5.0, September 2006.

Economic Study Assumptions:

The base assumptions used to evaluate the Base Case (i.e., without CO_2 capture) and all other CO_2 capture cases (Cases 1-5) are given in Table 3-64. This approach enabled the evaluation of the impacts of CO_2 capture in terms of incremental costs of electricity and CO_2 mitigations costs.

Table 3-64: Base Economic Assumptions (Base Case and Cases 1-5)

Parameter	Unit	Value
Investment Cost	$/kW	as estimated
Capacity Factor	%	85
Income Tax Rate	%	38
Repayment Term of Debt	Years	15
Grace Period on Debt Repayment	Years	0
Debt Reserve Fund		None
Depreciation (150% declining balance)	Years	20
Working Capital (all parameters)	$	0
Investment Tax Credit	%	0
Tax Holiday	Years	0
Start-up Costs (% of EPC)	%	2
EPC Escalation	%	0
Duration of Construction	Years	3
Debt	%	45
Equity	%	55
After-tax Weighted Cost of Capital	%	9.67
Capital Charge Factor	-	0.175
Fixed O&M Levelization Factor	-	1.1568
Variable O&M Levelization Factor	-	1.1568
Natural Gas Levelization Factor	-	1.1651

Table 3-65 compares the economic analysis results for Cases 1-5 to the Base Case (0% Capture). American Electric Power (AEP) provided the assumptions pertaining to the Base Case unit (i.e., Conesville #5 Unit) operating at a 72% capacity factor. The Base Case values were adjusted to an 85% capacity factor for comparison to Cases 1-5.

Table 3-65: Economic Evaluation Study Assumptions (Base Case and Cases 1-5)

Percent CO_2 Capture (Case)	**0%** (Base Case)	**90%** (Case-1)	**70%** (Case-2)	**50%** (Case-3)	**30%** (Case-4)	**96%** (Case-5)
Power Generation						
Net Output (MW)	433.8	303.3	333.2	362.9	392.1	251.6
Capacity Factor (%)	85%	85%	85%	85%	85%	85%
Operating Hours (hrs/yr)	7,446	7,446	7,446	7,446	7,446	7,446
Net Efficiency, HHV (%)	35.0%	24.5%	26.9%	29.3%	31.7%	20.3%
Net Plant Heat Rate, HHV (Btu/kWh)	9,778	13,984	12,728	11,686	10,818	16,856
Total Fuel Heat Input at MCR (MMBtu/hr)	4,242	4,242	4,242	4,242	4,242	4,242
Coal HHV Input (MMBtu/hr)	4,229	4,229	4,229	4,229	4,229	4,229
Net Generation (MWh/yr)	3,230,075	2,258,498	2,481,342	2,702,488	2,919,331	1,873,667
Costs						
Total Investment Cost ($1000s)	NA	400,094	365,070	280,655	211,835	701,057
Total Investment Cost ($/kW)	NA	1,319	1,095	773	540	2,786
Fixed O&M Costs ($1000/yr)	0	2,494	2,284	2,079	1,869	2,488
Variable O&M Costs ($1000/yr)	0	17,645	14,711	10,876	7,019	18,640
Levelized, Make-up Power Cost						
Make-up Power Cost (¢/kWh)	NA	6.40	6.40	6.40	6.40	6.40
Make-up Power Cost ($1000/yr)	0	62,194	47,926	33,768	19,885	86,832
CO_2 By-product Revenue						
CO_2 By-product Selling Price ($/ton)	0	0	0	0	0	0
CO_2 By-product (lb/hr)	866,102	779,775	607,048	433,606	260,163	835,053
CO_2 By-product Revenue ($1000/yr)	0	0	0	0	0	0
Feedstock O&M Costs						
Coal Price ($/MMBtu)	1.80	1.80	1.80	1.80	1.80	1.80
Coal for CO_2 System (MMBtu/hr)	0	0	0	0	0	0
Coal Cost ($1000/yr)	0	0	0	0	0	0
Natural Gas Price ($/MMBtu)	6.75	6.75	6.75	6.75	6.75	6.75
Natural Gas for CO_2 System (MMBtu/hr)	0	13.00	9.70	6.70	4.20	17.70
Natural Gas Cost ($1000/yr)	0	653	488	337	211	890
LCOE Assumptions						
Levelization Term (years)	NA	20	20	20	20	20
Capital Charge Factor	NA	0.175	0.175	0.175	0.175	0.175
Fixed O&M Levelization Factor	NA	1.1568	1.1568	1.1568	1.1568	1.1568
Variable O&M Levelization Factor	NA	1.1568	1.1568	1.1568	1.1568	1.1568
Feedstock O&M Levelization Factor	NA	1.1651	1.1651	1.1651	1.1651	1.1651
LCOE Contributions						
Capital Component (¢/kWh)	NA	3.10	2.57	1.82	1.27	6.55
Fixed O&M (¢/kWh)	NA	0.13	0.11	0.09	0.07	0.15
Variable O&M (¢/kWh)	NA	3.66	2.62	1.72	0.96	5.79
Feedstock O&M (¢/kWh)	NA	0.03	0.02	0.01	0.01	0.06
Total (¢/kWh)	NA	6.92	5.32	3.64	2.31	12.54
CO_2 Mitigation Cost ($/ton)	NA	81	88	91	103	134
CO_2 Mitigation Cost ($/tonne)	NA	89	96	100	113	148
CO_2 Capture Cost ($/ton)	NA	54	58	61	70	76
CO_2 Capture Cost ($/tonne)	NA	59	64	67	77	83

Note: Make-up Power Cost (MUPC) applied to this study is already levelized over 20 years. Therefore, the annual cost represents the “levelized cost” not the “first-year cost”. The reported annual MUPC is not multiplied by the variable O&M levelization factor when calculating the LCOE. The CO_2 By-product revenue represents the “first-year cost” and is multiplied by the variable O&M levelization factor when calculating the LCOE.

Economic Sensitivity Study:

Additionally, economic sensitivity studies were developed for the five primary cases (each of the CO_2 capture options) to highlight which parameters affected the incremental LCOE and CO_2 mitigation cost to the greatest extents. A total of 40 economic evaluation cases are reported in Appendix III. The sensitivity analysis was designed to show the effects on incremental LCOE and CO_2 mitigation cost of variations in the four parameters of interest. The four parameters varied in this sensitivity study were capacity factor, total investment cost, make-up power cost (levelized), and CO_2 by-product selling price (levelized). Three points were calculated for each parameter as shown in Table 3-66. These sensitivity parameters were chosen since the base values used for these parameters are site specific to this project. Therefore proper use of these sensitivity results could potentially allow extrapolation to apply results to units other than just Conesville #5.

Table 3-66: Economic Sensitivity Study Parameters

Parameter	Units	Base	Sensitivity Analysis	
Total Investment Cost (TIC)	$	As Estimated	Base – 25%	Base +25%
Capacity Factor	%	85	72	90
CO_2 Selling Price, Levelized	$/ton	0	25	50
Make-up Power Cost, Levelized	¢/kWh	6.40	4.80	8.00

3.4.2 Economic Analysis Results

This section summarizes all the economic analysis results obtained from this study. Results discussed in subsections 3.4.2.1 and 3.4.2.2 were obtained while using a combination of economic assumptions given in Table 3-64 and Table 3-65. The results discussed in subsection 3.4.2.3 were obtained while using a combination of economic assumptions given in Table 3-64, Table 3-65, and Table 3-66. All these results are briefly discussed in the following subsections.

3.4.2.1 Economic Results for Cases 1-4 (90%-30% CO_2 capture)

Economic results for Cases 1-4 are shown in Table 3-67 and plotted in Figure 3-40 and Figure 3-41. The incremental LCOE is comprised of capital, fixed O&M, variable O&M, and fuel components. For the 90% CO_2 capture, for example, the respective LCOE values for these components are 3.10, 0.13, 3.66, and 0.03 ¢/kWh for a combined total of 6.92 ¢/kWh. The total incremental LCOE decreases almost linearly from 6.92 to 2.31 ¢/kWh as the CO_2 capture level decreases from 90% to 30%. The CO_2 mitigation cost, on the other hand, increases slightly from $89 to $113/tonne of CO_2 avoided, as the CO_2 capture level decreases from 90% to 30%, due to economy of scale effects.

Table 3-67: Economic Results (Cases 1-4)

Case	Case 1	Case 2	Case 3	Case 4
Power Generation				
Net Output (MW)	303.3	333.2	362.9	392.1
Capacity Factor (%)	85%	85%	85%	85%
Net Plant Heat Rate, HHV (Btu/kWh)	13,984	12,728	11,686	10,818
Net Efficiency, HHV (%)	24.5%	26.9%	29.3%	31.7%
Energy Penalty	10.5%	8.1%	5.7%	3.3%
CO_2 Profile				
CO_2 Captured (lb/hr)	779,775	607,048	433,606	260,163
CO_2 Captured (%)	90%	70%	50%	30%
Costs				
Total Investment Cost ($1000s)	400,094	365,070	280,655	211,835
Total Investment Cost ($/kW)	1,319	1,095	773	540
Fixed O&M Costs ($1000/yr)	2,494	2,284	2,079	1,869
Variable O&M Costs ($1000/yr)	17,645	14,711	10,876	7,019
Levelized, MUPC ($1000/yr)	62,194	47,926	33,768	19,885
CO_2 By-product Revenue ($1000/yr)	0	0	0	0
Feedstock O&M Costs ($1000/yr)	653	488	337	211
Coal Cost ($1000/yr)	0	0	0	0
Natural Gas Cost ($1000/yr)	653	488	337	211
LCOE Contributions				
Capital Component (¢/kWh)	3.10	2.57	1.82	1.27
Fixed O&M (¢/kWh)	0.13	0.11	0.09	0.07
Variable O&M (¢/kWh)	3.66	2.62	1.72	0.96
Feedstock O&M (¢/kWh)	0.03	0.02	0.01	0.01
Total (¢/kWh)	6.92	5.32	3.64	2.31
CO_2 Mitigation Cost ($/ton)	81	88	91	103
CO_2 Mitigation Cost ($/tonne)	89	96	100	113
CO_2 Capture Cost ($/ton)	54	58	61	70
CO_2 Capture Cost ($/tonne)	59	64	67	77

Note: Make-up Power Cost (MUPC) applied to this study is already levelized over 20 years. Therefore, the annual cost represents the "levelized cost" not the "first-year cost". The reported annual MUPC was not multiplied by the variable O&M levelization factor when calculating the LCOE. The CO_2 By-product revenue represents the "first-year cost" and was multiplied by the variable O&M levelization factor when calculating the LCOE.

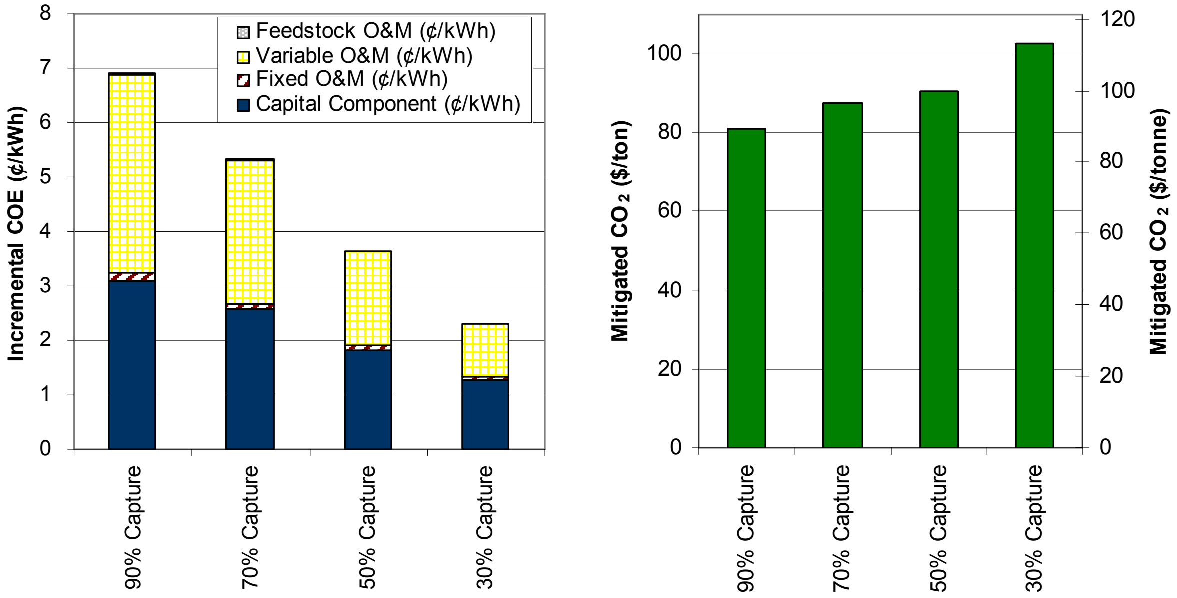


Figure 3-40: Economic Results (Cases 1-4)

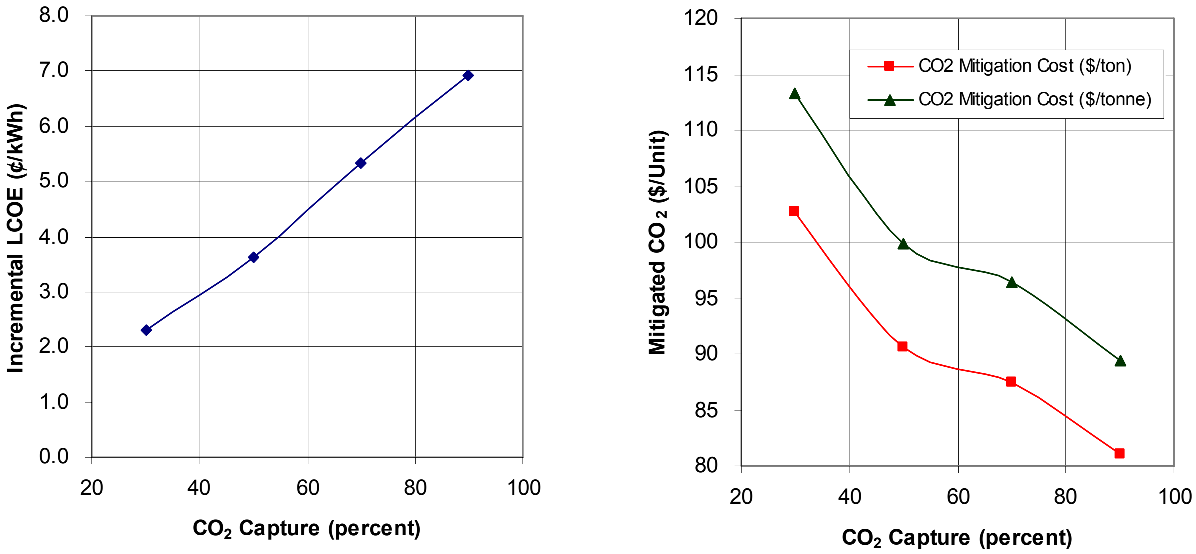


Figure 3-41: Impact of CO_2 Capture Level on Incremental LCOE and CO_2 Mitigation Cost (Cases 1-4)

3.4.2.2 Economic Results for Case 1 and Case 5 (90% and 96% CO_2 capture)

As stated in Section 3.3.2.5, the investment costs and O&M costs of Concept A (96% CO_2 Capture with MEA) from the previous study (Bozzuto et al., 2001) were updated to July 2006 dollars. The economic analysis of this case, referred to in the present study as Case 5, was then done in the same manner as Cases 1-4. Results obtained from Case 5 are compared below to those obtained from Case 1 (90% CO_2 capture). The rationale for this comparison is that the CO_2

capture percentages of both cases are close to one another, and this comparison shows the impact of using the advanced amine on economic performance parameters of merit. An equitable comparison of specific costs ($/kWe) and economics (LCOE, mitigation costs) between the advanced amine and the Kerr-McGee/ABB Lummus amine was not possible since the amine system design for the previous study was not consistent with the current designs for the advanced amine, as explained in more detail below.

Economic results for Case 1 and Case 5 are shown in Table 3-68 and Figure 3-42. The capital, fixed O&M, variable O&M, and fuel components of the incremental LCOE for Case 5 are 6.55, 0.15, 5.79, and 0.06 ¢/kWh for a total incremental LCOE value of 12.54 ¢/kWh. The corresponding values for Case 1 are 3.10, 0.13, 3.66, and 0.03 ¢/kWh for a combined total of 6.92 ¢/kWh. Extrapolating the Case 1 LCOE to 96% capture would yield an incremental COE of about 7.37 ¢/kWh. This shows an improvement of 5.17 ¢/kWh at the 96% capture level (i.e., the advanced amine vs. the Kerr-McGee/ABB Lummus amine).

The cost of electricity for Case 5 is 81% higher than that of Case 1, primarily due to its higher total investment cost ($2,786 vs. $1,319/kWe), reduced efficiency (20.3% vs. 24.5% HHV), and, to a lesser extent, higher CO_2 capture (96% vs. 90%). Consistent with incremental LCOE results, the CO_2 mitigation cost of Case 5 is more than 66% higher than that of Case 1 ($148 vs. $89/tonne).

It should be noted that the design of Case 5 (See Bozzuto et al., 2001) is not totally consistent with the design of Case 1 done in this study. Case 1 uses 2 absorbers, 2 strippers, and 2 compression trains. Case 5, which was designed in 2000, used 5 absorbers, 9 strippers, and 7 compression trains. Because of these differences, Case 1 is able to take advantage of economy of scale effects for equipment cost due to the larger equipment sizes. Additionally, Case 5 equipment was all located about 457 m (1,500 ft) from the Unit #5 stack, which also increased the costs of Case 5 relative to Case 1. It should be pointed out that if Case 5 (~96% recovery) was designed as a part of the current study, it would likely have equipment selections similar to Case 1 (i.e., a two-train system) and therefore significant cost reductions and improved economics would result.

Because of these significant design differences, an equitable comparison of specific costs ($/kWe) and economics (LCOE, mitigation costs) between the advanced amine and the Kerr-McGee/ABB Lummus amine was not possible. The results presented in Table 3-68 and Figure 3-42 must be viewed with the above context.

Table 3-68: Economic Results for Cases 1 and 5

Case	Case 1	Case 5
Power Generation		
Net Output (MW)	303.3	251.6
Capacity Factor (%)	85%	85%
Net Plant Heat Rate, HHV (Btu/kWh)	13,984	16,856
Net Efficiency, HHV (%)	24.5%	20.3%
Energy Penalty	10.5%	14.7%
CO_2 Profile		
CO_2 Captured (lb/hr)	779,775	835,053
CO_2 Captured (%)	90.0%	96.0%
Costs		
Total Investment Cost ($1000s)	400,094	701,057
Total Investment Cost ($/kW)	1,319	2,786
Fixed O&M Costs ($1000/yr)	2,494	2,488
Variable O&M Costs ($1000/yr)	17,645	18,640
Levelized Make-up Power Cost ($1000/yr)	62,194	86,832
CO_2 By-product Revenue ($1000/yr)	0	0
Feedstock O&M Costs ($1000/yr)	653	890
Coal Cost ($1000/yr)	0	0
Natural Gas Cost ($1000/yr)	653	890
LCOE Contributions		
Capital Component (¢/kWh)	3.10	6.55
Fixed O&M (¢/kWh)	0.13	0.15
Variable O&M (¢/kWh)	3.66	5.79
Feedstock O&M (¢/kWh)	0.03	0.06
Total (¢/kWh)	6.92	12.54
CO_2 Mitigation Cost ($/ton)	81	134
CO_2 Mitigation Cost ($/tonne)	89	148
CO_2 Capture Cost ($/ton)	54	76
CO_2 Capture Cost ($/tonne)	59	83

Note: Make-up Power Cost (MUPC) applied to this study is already levelized over 20 years. Therefore, the annual cost represents the “levelized cost” not the “first-year cost”. The reported annual MUPC was not multiplied by the variable O&M levelization factor when calculating the LCOE. The CO_2 by-product revenue represents the “first-year cost” and was multiplied by the variable O&M levelization factor when calculating the LCOE.

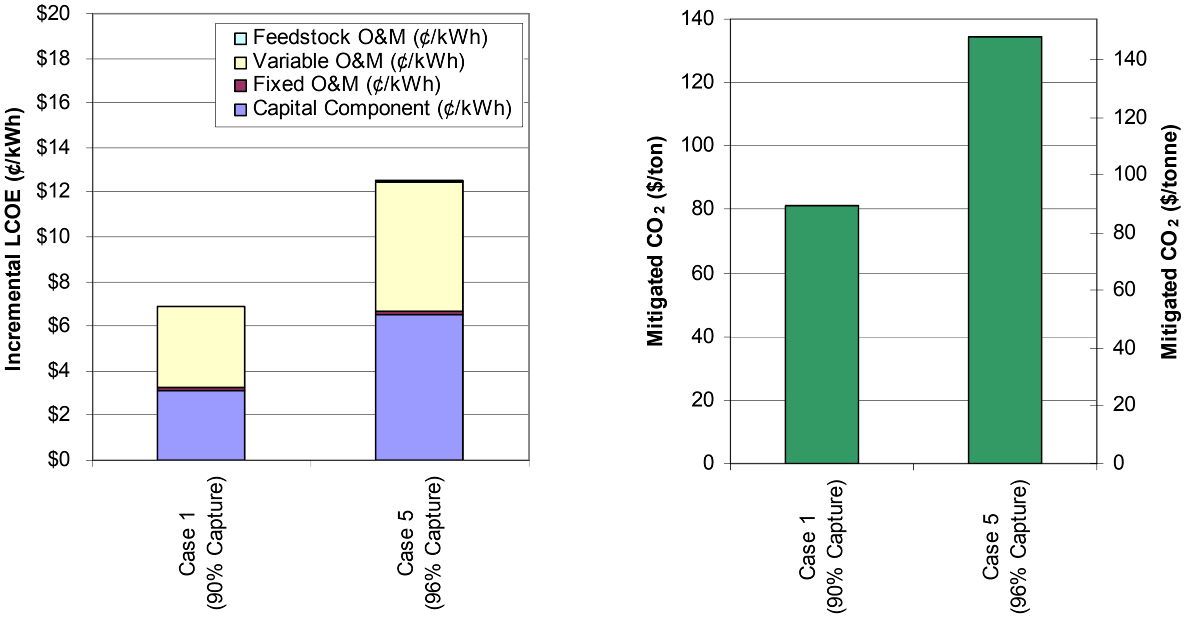


Figure 3-42: Economic Results for Case 1 and Case 5

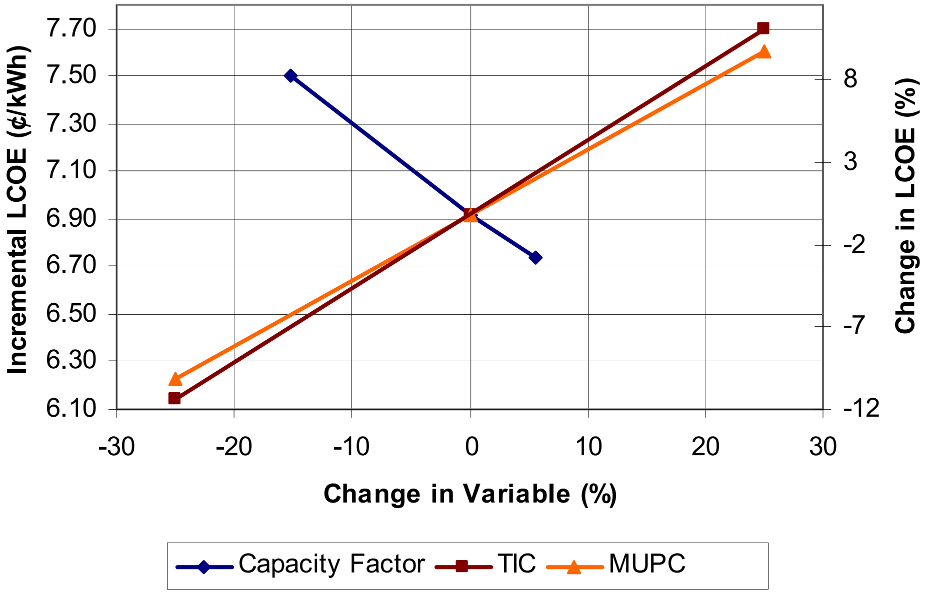
3.4.2.3 Economic Sensitivity Analysis Results

The economic sensitivity analysis was done by varying a number of parameters (Capacity Factor, Total Investment Cost, Make-up Power Cost, and CO_2 By-product Selling Price) that affect the economic results. These sensitivity parameters were chosen since the base values used for these parameters are site specific to this project. Therefore, proper use of these sensitivity results could potentially allow extrapolation to apply results to units other than just Conesville #5. The objective of this analysis was to determine the relative impacts of the sensitivity parameters and CO_2 capture level on incremental cost of electricity and CO_2 mitigation cost.

Results obtained from Cases 1, 2, 3, 4, and 5 (with 90%, 70%, 50%, 30%, and 96% CO_2 capture, respectively) are presented in tabular and graphical forms in **Appendix III.** The economic sensitivity results obtained from Case 1 (90% CO_2 capture) are briefly discussed below. Detailed economic results for Case 1 and the other cases are in Appendix III.

Economic Sensitivity Analysis Results for Case 1 (90% CO_2 Capture)

Results for the Case 1 sensitivity study are shown in Figure 3-43. This figure shows the sensitivity of incremental LCOE to capacity factor, total investment cost, make-up power cost, and CO_2 by-product selling price. The base parameter values represent the point in Figure 3-43 where all the sensitivity curves intersect (point 0.0, 0.0). The incremental LCOE ranges from a low of -0.50 ¢/kWh to a high of 7.96 ¢/kWh for the Case 1 sensitivity analysis. The order of sensitivity (most sensitive to least sensitive) of these parameters to incremental LCOE is: CO_2 by-product selling price (levelized) > capacity factor > total investment cost > make-up power cost (levelized). For Cases 2 thru 5, the total investment cost becomes more significant than the make-up power cost, but, they are approximately equivalent in Case 1.



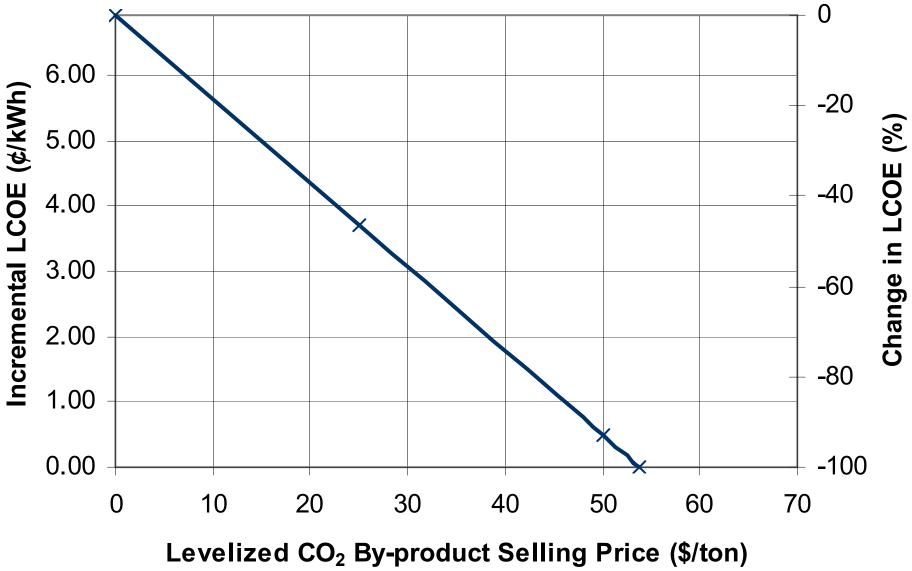


Figure 3-43: Economic Sensitivity Results (Case 1 - 90% CO_2 Capture)

4 ADVANCES IN POST COMBUSTION CO_2 CAPTURE TECHNOLOGIES

Numerous major research and developmental efforts are continually ongoing inside and outside the U.S. to further advance post-combustion CO_2 capture technologies. Such efforts seek to develop advanced/breakthrough technologies aimed at improving performance and cost, with the ultimate goal of developing cost-competitive post-combustion CO_2 capture technologies. A selected number of these technologies are listed in Table 4-1.

As can be seen, these technologies are at various stages of development, ranging from laboratory-scale to commercial-scale. More detailed information can be found on these technologies through the websites/references given in the table.

Table 4-1: List of Selected Advanced Post-Combustion CO_2 Capture Technologies

Technology	R & D by	Status	Information Source
Aqueous Ammonia	Powerspan, In-house NETL	Pilot-plant	DOE NETL Carbon Sequestration Technology Roadmap and Program Plan 2006 (www.netl.gov)
Amine-Enhanced Sorbents	In-house NETL	Laboratory scale	CO_2 Capture Systems Using Amine Enhanced Sorbents, Coal-Gen 2006, Cincinnati, OH; (www.netl.gov)
Ionic Liquids	University of Notre Dame; Sachem; Merck	Laboratory scale	DOE NETL Carbon Sequestration Technology Roadmap and Program Plan 2006 (www.netl.gov)
Enzymatic CO_2 Sorbents	Carbozyme CO_2	Laboratory scale	DOE NETL Carbon Sequestration Technology Roadmap and Program Plan 2006 (www.netl.gov)
Dry Regenerable Sorbents	Research Triangle Institute	Laboratory scale, Pilot scale	DOE Website: www.netl.gov
Chilled Ammonia	Alstom	Pilot scale (5-MWth)	http://www.power.alstom.com
KS® Solvents (KS 1, KS2, KS3)	Kansai Electric Power Co./ Mitsubishi Heavy Industry, Ltd.	Commercial scale on gas fired flue gas, Pilot scale on coal-fired flue gas	Japanese R&D on Large-Scale CO_2 Capture http://services.bepress.com/cgi/viewcontent.cgi?article=1006&context=eci/separations_technology_vi

Selected technologies are briefly described below.

Aqueous Ammonia: This joint NETL-Powerspan development entails reacting ammonia with CO_2 in the flue gas to form ammonium carbonate, and subsequently heating the ammonium carbonate to release a pure CO_2 stream. Advantages include: (1) low theoretical heat of regeneration (286 Btu/lbm CO_2 vs. 825 Btu/lbm CO_2 for MEA); and (2) multi-pollutant control with saleable by-products (ammonium sulfate and ammonium nitrate fertilizers). One technical challenge is degradation of carbonate in the CO_2 absorber leading potentially to ammonia slip in the flue gas.

Amine-Enhanced Sorbents: This technology is being developed by NETL. The principle of operation of the process entails exposing a CO_2-rich stream to a carbon material (substrate) with amine compounds attached unto it. The CO_2 absorbed on the amine sites is subsequently released upon increasing the temperature. This process has some advantages over the MEA process, e.g., higher CO_2 carrying capacity; lower heat capacity, as there is no water to heat. One technical challenge is that small particle diameters can cause high-pressure drops across the absorber. The sorbent regeneration energy has been estimated at 620 Btu/lbm CO_2, which would be a breakthrough improvement over the current state-of-the-art of about 1,600 Btu/lbm CO_2.

Chilled Ammonia: This process, being developed by Alstom, entails chilling the flue gas, recovering large quantities of water for recycle, and then utilizing a CO_2 absorber similar in design to the absorbers used in systems to reduce flue gas sulfur dioxide emissions. CO_2 is stripped at high pressure and compressed to a pressure suitable for use in EOR or sequestration. In laboratory tests co-sponsored by Alstom, EPRI, and others, the process has demonstrated a potential for capturing more than 90% CO_2 at an efficiency penalty that is much lower than other CO_2 capture technologies. This process is undergoing validation testing in a 5-MW_{th} slipstream from a plant in Wisconsin.

KS® Solvents (KS1, KS2, and KS3): Kansai Electric Power Company (KEPCO) and Mitsubishi Heavy Industry (MHI) in Japan jointly developed these sterically hindered amines. The KS1 process has been capturing 160 tonnes/day CO_2 from a steam reforming flue gas at Kedah Danul Aman in Malaysia since 1999. Hokuriku Electric Power Company has operated a test plant with KS® solvents treating 50 m^3N/hr of flue gas from a coal-fired plant at the Toyama-Shinko power station. KEPCO and MHI report that the regeneration energy for KS® solvents is much less than that of MEA (700 vs. 900 kcal/kg-CO_2 or 1,260 vs. 1,620 Btu/lbm-CO_2).

5 SENSITIVITY OF PLANT PERFORMANCE AND ECONOMICS TO SOLVENT REGENERATION ENERGY

With respect to solvent regeneration energy, process simulation results showed that the advanced amine used in this study (Cases 1-4), based on present day technology, required 1,550 Btu/lbm-CO_2. This solvent regeneration energy was 34% less than in the prior study (2,350 Btu/lbm-CO_2), which was completed six years ago. Comparatively, recent values for solvent regeneration energy in the open literature are as shown in Table 5-1:

Table 5-1: Solvent Regeneration Energy for Amine-Based CO_2 Capture Systems

Source	Kerr-McGee Lummus MEA, Bozzuto et al. 2001	Economine FG$^+$, DOE/NETL, Parsons, WorleyParsons 2006	Economine FG^{+SM}, IEA Report PH4/33 2004	KS1-IEA Report PH4/33 2004
Btu/lbm-CO_2	2,350	1,530	1,395	1,375

Numerous research and developmental efforts are ongoing to further advance post-combustion CO_2 capture technologies. These efforts seek to develop technologies that are focused on improving performance, and reducing cost with post-combustion CO_2 capture. One of the key parameters with post-combustion CO_2 capture systems that is an indicator of relative system performance is solvent regeneration energy requirement (Btu/lbm-CO_2). Hence, as a look to the future, a simplified sensitivity analysis for solvent regeneration energy and the resulting impacts on power plant performance (thermal efficiency) and economics (cost of electricity) was carried out.

It is understood that solvent regeneration energy represents a key variable for amine-based post-combustion CO_2 capture systems in terms of the impact this variable ultimately has on the most common measures of power plant performance (thermal efficiency) and economic merit (cost of electricity). Research and development in this area continues to progress and as a result, amine solvents and post-combustion capture systems are improving in performance. Future systems incorporating the improvements will have significant positive impacts on power plant performance and economics. Therefore, a sensitivity analysis showing the effect of anticipated reductions in solvent regeneration energy was performed in this study.

This sensitivity study was done at the 90% capture level only and the solvent regeneration energy levels investigated were 1,550 and 1,200 Btu/lbm-CO_2. These cases are referred to as **Cases 1 and 1a** respectively. The value of 1,550 Btu/lbm-CO_2 used in this study (Case 1) was taken as a base value and represents current technology (ca. 2006). The value of solvent regeneration energy for the sensitivity case (Case 1a) was selected, keeping in mind future technological developments/advancements. It is well known that commercial implementation of these amine-based post-combustion capture systems for power plant applications will not occur until several years in the future. This delay is due to a variety of reasons such as: these systems need to be proven at large scale, CO_2 sequestration technology needs to be proven, and policies need to be implemented to make utilization of these systems economical. It is also understood that solvent regeneration energy represents a key variable for amine-based post-combustion CO_2 capture systems in terms of the impact this variable ultimately has on the common measures of power plant performance (thermal efficiency) and economic merit (cost of electricity). Furthermore, research in this area continues, and as a result, amine solvents and post-combustion capture systems in general are improving in performance and, therefore, power plants will incur reduced impacts on power plant performance and economics. The solvent regeneration energy level selected for Case 1a was 1,200 Btu/lbm-CO_2, which represents a near future goal. This solvent regeneration energy value was agreed on by NETL, RDS, and Alstom. This more advanced amine would represent an amine that would be commercially available in the near future.

This sensitivity study was completed in a very simplified manner. Process simulations, equipment design, and cost estimates for this future amine-based capture system were not developed since physical properties and other information, which are necessary for use in process design, equipment sizing, and material selection, are unknown for this future case (Case 1a). Costs for Case 1a were assumed to be the same as for Case 1. A detailed steam turbine material and energy balance was developed for Case 1a and therefore the calculated plant performance should be quite accurate for this case.

The following basic work steps were applied for this evaluation:

- Regeneration energy requirements and heat integration requirements between the Gas Processing System and the steam cycle are defined for the new case
- Alstom Steam Turbine Group (STG) calculates new steam turbine heat balance for the reduced solvent regeneration energy case
- An estimate of amine system auxiliary power changes for the reduced solvent regeneration energy case are developed
- Overall power plant performance (thermal efficiency) is calculated for the reduced solvent regeneration energy case
- Investment costs are assumed not to change for the reduced solvent regeneration energy case as compared to Case 1 (90% CO_2 capture case with 1,550 Btu/lbm-CO_2 solvent regeneration energy requirement)
- Economics (incremental COEs and CO_2 mitigation costs) are calculated for the reduced solvent regeneration energy case
- New tables, graphs, and a new report section are developed to discuss the results from the reduced solvent regeneration energy case

This sensitivity study therefore represents a view of the potential future capabilities for amine-based post-combustion CO_2 capture systems. In summary, the results obtained from this sensitivity study enabled the quantification of the performance and economic impacts on the power plant, for the 90% CO_2 capture level, with solvent regeneration energies of 1,550 and 1,200 Btu/lbm-CO_2 **(Cases 1 and 1a** respectively). Results are discussed in the following sub-sections.

5.1 Performance Analysis

Plant performance and CO_2 emissions are summarized for the existing and modified power plants in Table 5-2. Several graphs also illustrate selected results from the table plotted as a function of solvent regeneration energy. Four cases are shown in this section. The Base Case is the “business as usual” case without CO_2 capture. Cases 1 and 1a are with 90% CO_2 capture, low level heat integration between the gas processing system and the steam cycle, and various levels of solvent regeneration energy. Comparisons between cases 1 and 1a isolate the impacts of solvent regeneration energy level. Case 5, from the earlier study (Bozzuto et al., 2001) is also shown. This case differs in that it has 96% CO_2 capture, a solvent regeneration energy requirement of 2,350 Btu/lbm-CO_2, and no heat integration between the gas processing system and the steam cycle. These four cases are listed below.

- **Base Case -** Existing power plant without CO_2 capture - refer to Section 2 for details.
- **Case 1 –** Existing power plant retrofit with an advanced “state of the art” amine system for 90% CO_2 capture (1,550 Btu/lbm-CO_2 solvent regeneration energy) - refer to Section 3 for details.
- **Case 1a –** Existing power plant retrofit with an advanced “near future” amine system for 90% CO_2 capture (1,200 Btu/lbm-CO_2 solvent regeneration energy).
- **Case 5 –** Existing power plant retrofit with a Lummus/Kerr-McGee MEA system (ca. 2000 design) for 90% CO_2 capture (2,350 Btu/lbm-CO_2 solvent regeneration energy) - refer to Section 3 for details.

Table 5-2: Plant Performance and CO_2 Emissions vs. Solvent Regeneration Energy

		(units)	Base-Case Original Plant	Case 5 Concept A MEA	Case 1 Advanced MEA	Case 1a Advanced MEA
Solvent Regeneration Energy		(Btu/lbm-CO_2)	0	2350	1550	1200
CO_2 Capture		(percent)	0	96	90	90
Boiler Parameters						
Main Steam Flow		(lbm/hr)	3131619	3131651	3131651	3131651
Reheat Steam Flow (to IP turbine)		(lbm/hr)	2853607	2853607	2848739	2848725
Main Steam Pressure		(psia)	2535	2535	2535	2535
Main Steam Temp		(Deg F)	1000	1000	1000	1000
Reheat Steam Temp		(Deg F)	1000	1000	1000	1000
Boiler Efficiency		(percent)	88.13	88.13	88.13	88.13
Flue Gas Flow leaving Economizer		(lbm/hr)	4014743	4014743	4014743	4014743
Flue Gas Temperature leaving Air Heater		(Deg F)	311	311	311	311
Coal Heat Input (HHV)	(HHV)	(10^6 Btu/hr)	4228.7	4228.7	4228.7	4228.7
	(LHV)	(10^6 Btu/hr)	4037.9	4037.9	4037.9	4037.9
CO_2 Removal Steam System Parameters						
CO_2 Removal System Steam Pressure		(psia)	---	65	47	47
CO_2 Removal System Steam Temp		(Deg F)	---	478	424	424
CO_2 Removal System Steam Extraction Flow		(lbm/hr)	---	1935690	1210043	975152
CO_2 Removal System Condensate Pressure (from reboilers)		(psia)	---	64.7	40	40
CO_2 Removal System Condensate Temperature		(Deg F)	---	292.7	267.3	267.3
CO_2 Removal System Heat to Cooling Tower		(10^6 Btu/hr)		1441.1	890.2	698.2
Natural Gas Heat Input	(HHV)[2]	(10^6 Btu/hr)	0	17.7	13.0	13.0
[2] (For Dessicant Regeneration)	(LHV)	(10^6 Btu/hr)	---	16.0	11.7	11.7
		(10^6 SCF/Day)		0.417	0.312	0.312
CO_2 produced from Natural Gas usage		(lbm/hr)			1492	1492
Steam Cycle Parameters						
Total Heat Input to Steam Cycle		(10^6 Btu/hr)	3707.4	3707.4	3707.4	3707.4
Heat Output to CO_2 Removal System Reboilers & Reclaimer		(10^6 Btu/hr)	---	1953.0	1218.1	980.6
Existing Condenser Pressure		(psia)	1.23	1.23	1.23	1.23
Existing Condenser Heat Loss		(10^6 Btu/hr)	2102.8	603.3	1260	1468
Existing Steam Turbine Generator Output		(kW)	463478	269,341	342693	367859
CO_2 Removal System Turbine Generator Output		(kW)	0	62,081	45321	36083
Total Turbine Generator Output		(kW)	463478	331422	388014	403942
Auxiliary Power Requirements						
Condensate Pump Power		(kW)	563	450	503	512
Condenser Cooling Water Pump Power		(kW)	5562	5407	5687	5730
Boiler Island Auxiliary Power (Fans & Pulverizers)		(kW)	7753	7753	7753	7753
Coal & Ash Handling System		(kW)	1020	1020	1020	1020
FGD & ESP System Auxiliary Power		(kW)	8157	8157	8157	8157
Misc. Auxiliary Power (Lighting, HVAC, Trans, etc)		(kW)	6645	6645	6645	6645
Air Separation Unit Power Requirement (Case B) **CO_2 Removal System Auxiliary Power**		(kW)	0	50355	54939	54845
Total Auxiliary Power		(kW)	29700	79788	84704	84662
fraction of gross output		(fraction)	0.064	0.241	0.218	0.210
			433.8	251.6	303.3	319.3
Plant Performance Parameters						
Net Plant Output		(kW)	433778	251634	303310	319280
Normalized Net Plant Output (Relative to Base Case)		(fraction)	1.00	0.58	0.70	0.74
Net Plant Efficiency (HHV)		(fraction)	0.3501	0.2022	0.2441	0.2569
Net Plant Efficiency (LHV)		(fraction)	0.3666	0.2119	0.2556	0.2691
Normalized Efficiency (HHV; Relative to Base Case)		(fraction)	1.00	0.58	0.70	0.73
Net Plant Heat Rate (HHV)		(Btu/kWh)	9749	16875	13985	13285
Net Plant Heat Rate (LHV)		(Btu/kWh)	9309	16110	13351	12684
Plant CO_2 Emissions						
Carbon Dioxide Produced		(lbm/hr)	866102	868137	867595	867595
Carbon Dioxide Recovered		(lbm/hr)	0	835053	779775	779775
Carbon Dioxide Emissions		(lbm/hr)	866102	33084	87820	87820
Fraction of Carbon Dioxide Recovered		(fraction)	0	0.962	0.90	0.90
Specific Carbon Dioxide Emissions		(lbm/kWh)	1.997	0.131	0.290	0.275
Normalized Specific CO_2 Emissions (Relative to Base Case)		(fraction)	1.00	0.066	0.145	0.138
Avoided Carbon Dioxide Emissions (as compared to Base)		(lbm/kWh)	---	1.865	1.707	1.722

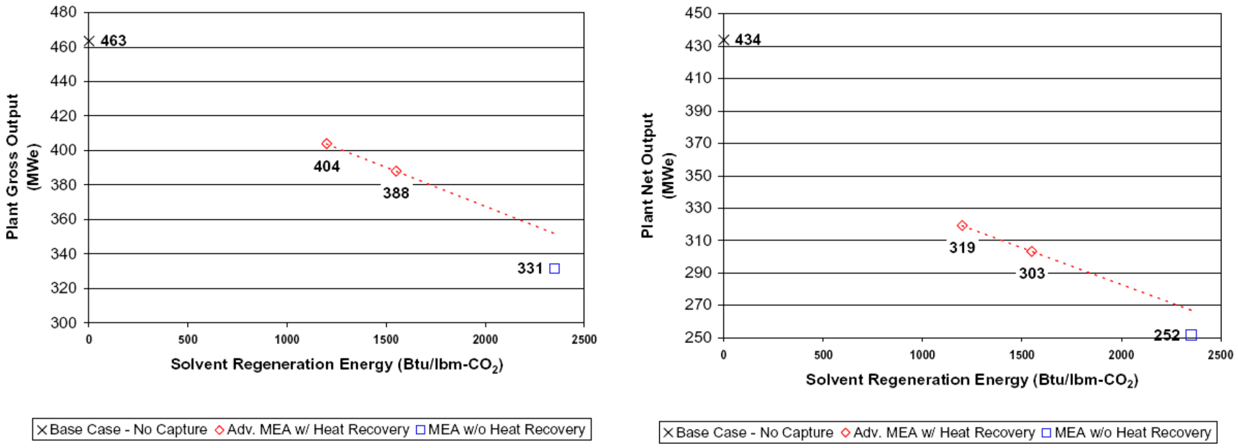


Figure 5-1: Plant Gross and Net Output versus Solvent Regeneration Energy

Plant output (both Gross and Net) is shown in Figure 5-1 as a function of solvent regeneration energy for the 90% capture level. Plant output is quite sensitive to changes in solvent regeneration energy. Plant net output was calculated to change by about 47 MWe (or about 1.5% relative to Case 1 - 1,550 Btu/lbm-CO_2) for a change in solvent regeneration energy of 1,000 Btu/lbm-CO_2.

Plant thermal efficiency and efficiency loss are shown in Figure 5-2. This figure shows, for the 90% capture level, the impacts on plant thermal efficiency of both solvent regeneration energy and low-level heat integration between the gas processing system and steam cycle.

Plant thermal efficiency is very sensitive to changes in solvent regeneration energy. Plant thermal efficiency was calculated to change by about 3.7 percentage points for a change in solvent regeneration energy of 1,000 Btu/lbm-CO_2. To help put this in perspective, Case 5 from our previous study (Bozzuto et al., 2001), which used the Kerr/McGee – ABB Lummus system, had a solvent regeneration energy of about 2,350 Btu/lbm-CO_2. This energy requirement was considered "state of the art" at the time of that study. In the current study, the advanced "state of the art" amine used for Cases 1-4 used a solvent regeneration energy requirement of 1,550 Btu/lbm-CO_2. This represents a reduction of ~800 Btu/lbm-CO_2 in 6 years.

Similarly, proper integration of the low level heat which is rejected in the gas processing system (compressor intercoolers, solvent stripper condenser, etc.) with the steam cycle condensate stream was calculated to add about 0.7 percentage points to plant thermal efficiency at the 1,550 Btu/lbm solvent regeneration energy level. This efficiency change would be lower for solvents with higher regeneration energy requirements (since less cool condensate leaving the main condenser of the steam cycle is available to recover the rejected heat from the gas processing system) and higher for solvents with lower regeneration energy requirements.

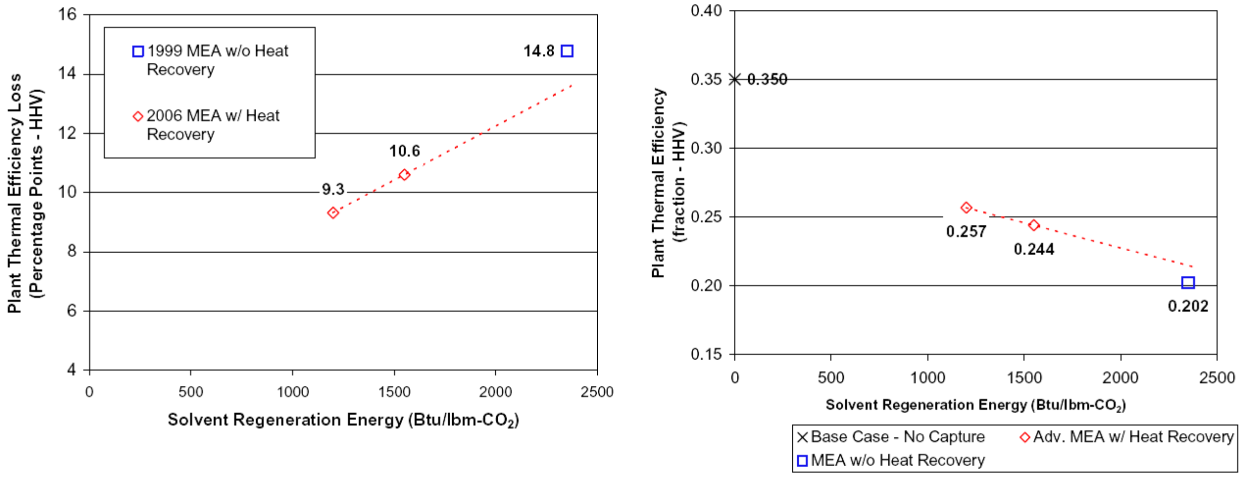


Figure 5-2: Plant Thermal Efficiency and Efficiency Loss vs. Solvent Regeneration Energy

Plant CO_2 emissions for this sensitivity study are summarized in Table 5-2 and Figure 5-3. Specific carbon dioxide emissions were reduced from 906 g/kWh (2.00 lbm/kWh) for the Base Case to between 59-132 g/kWh (0.13-0.29 lbm/kWh) depending on CO_2 capture level and solvent regeneration energy requirement for these cases. This corresponds to values between 6.6% and 14.5% of the Base Case specific carbon dioxide emissions.

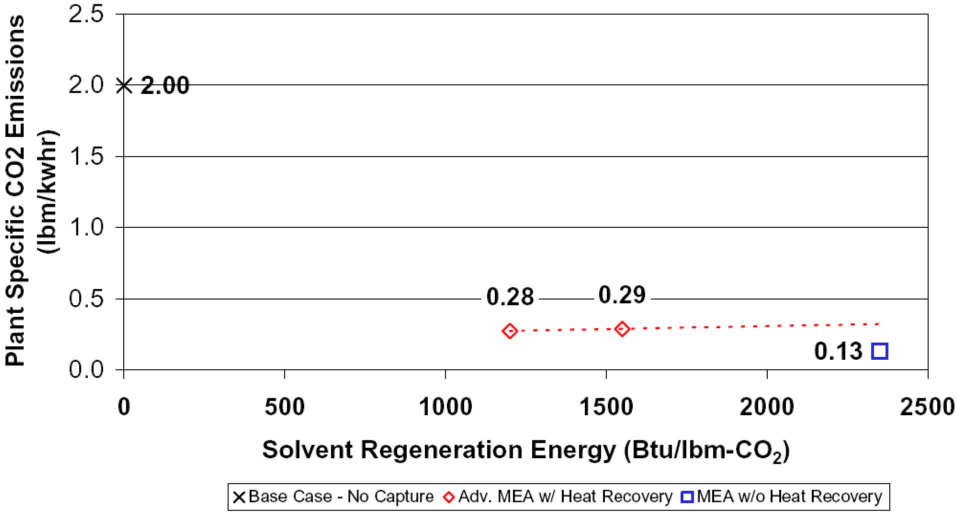


Figure 5-3: Plant CO_2 Emissions vs. Solvent Regeneration Energy

5.1.1 Steam Cycle Modifications and Performance with Reduced Solvent Regeneration Energy

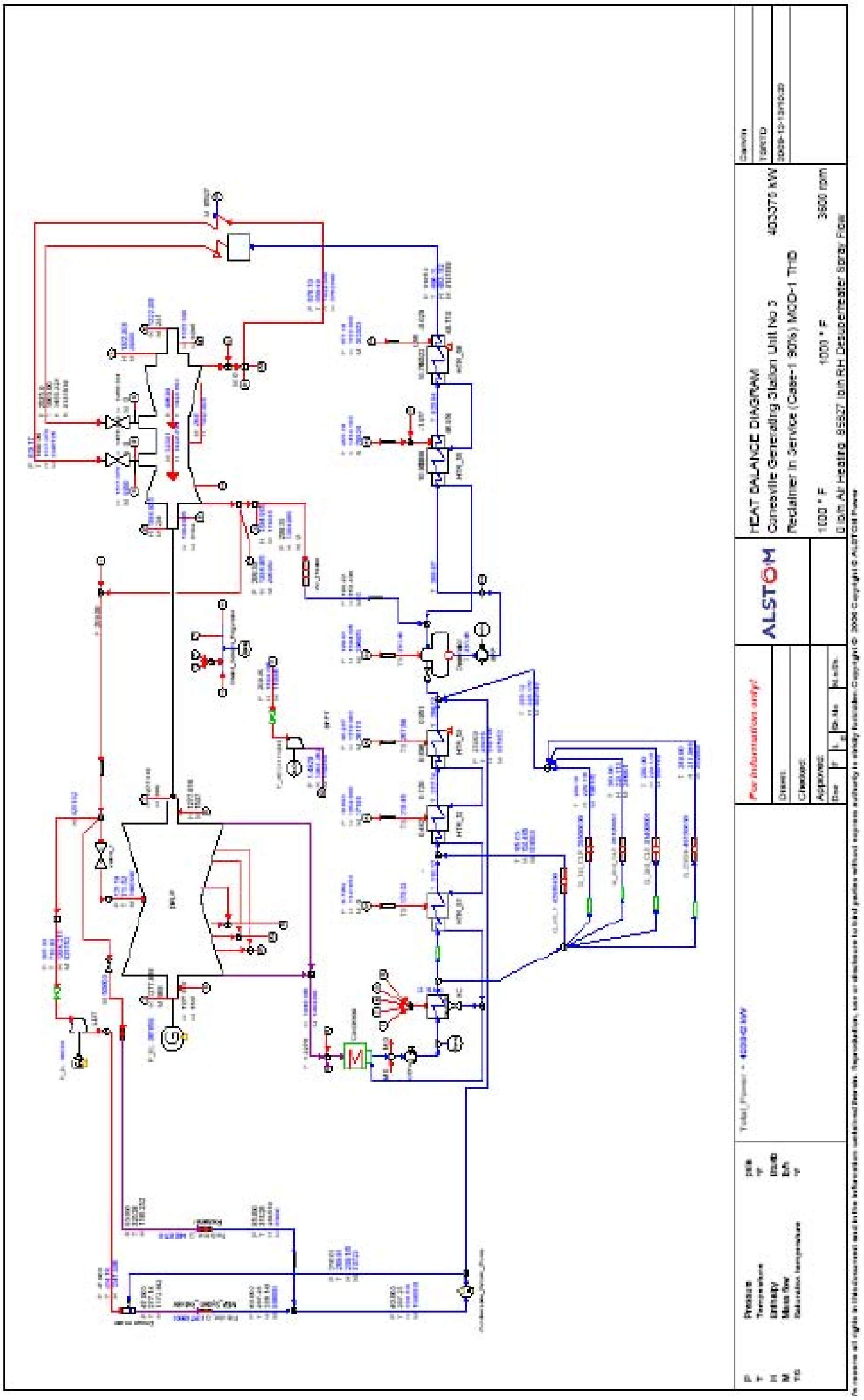
Both Case 1 (1,550 Btu/lbm-CO_2 solvent regeneration energy) and Case 1a (1,200 Btu/lbm-CO_2 solvent regeneration energy) remove 90% of the CO_2 contained in the flue gas. For a discussion of the steam cycle modifications required for Case 1a to integrate the steam cycle with the amine system please refer to Section 3.1.6 where the modifications for Cases 1-5 are discussed. Figure 5-4 shows the modified steam turbine energy and material balance for Case 1a. The steam flow required to operate the reboiler/reclaimer in the amine process for Case 1a is approximately

123.0 kg/s (975.2 x 10^3 lbm/hr), equivalent to approximately 40% of the steam that would enter the LP turbine cylinder in the absence of the amine plant. By comparison, Case1 uses 152.6 kg/s (1,210 x 10^3 lbm/hr), equivalent to approximately 50% of the steam that would enter the LP turbine cylinder in the absence of the amine plant; Case 5 uses 244.1 kg/s (1935.7 x 10^3 lbm/hr), equivalent to approximately 79% of the steam that would enter the LP turbine cylinder in the absence of the amine plant.

The higher steam flow entering the LP turbine for Case 1a would result in a correspondingly higher pressure at the LP turbine inlet. Consequently, the pressure drop across the pressure control valve would be reduced (less throttling) for this case, as compared to Case 1.

Heat integration for Case 1a is done in the same manner as for Case 1 (90% removal, 1,550 Btu/lbm-CO_2 solvent regeneration energy). Waste heat from the gas processing system (CO_2 compressor intercoolers, propane refrigeration unit compressor de-superheater, and solvent stripper overhead condenser) is recovered by preheating condensate from the steam cycle as is shown in the lower parts of Figure 5-4. The deaerator flow for this case is somewhat less than in Case 1, but still significantly higher than the flow indicated for the reference case (Base Case). This may impact the performance of the deaerator or require either modification of the deaerator or a change in the heat integration arrangement in order to reduce the duty of the deaerator.

In summary, for Case 1a as illustrated in Figure 5-4, the gross power output of the Conesville #5 Unit will decrease by approximately 12.8% (from 463.5 MW to 403.9 MW), when compared to the Base Case (please refer Section 2.2.4) after modification to remove 90% of the CO_2 contained in the flue gas with a solvent that requires 1,200 Btu/lbm solvent regeneration energy. By comparison, for Case 1, the gross power output of the Conesville #5 Unit will decrease by approximately 16.3% (from 463.5 MW to 388.0 MW), and, for Case 5, the gross power output of the Conesville #5 Unit will decrease by approximately 28.5% (from 463.5 MW to 331.4 MW), when compared to the Base Case.



We reserve all rights in this document and in the information contained therein. Reproduction, use or disclosure to third parties without express authority is strictly forbidden. Copyright © 2006 Copyright © ALSTOM Power

Figure 5-4: Case 1a – Modified Water-Steam Cycle (90% Capture: 1,200 Btu/lbm Solvent Regeneration Energy)

5.2 Cost Analysis

For the purposes of this sensitivity study, the investment cost for the new equipment associated with the reduced solvent regeneration energy case (Case 1a) is assumed to be the same as for Case 1 (i.e., 90% capture with 1,550 Btu/lbm solvent regeneration energy). This was done because the physical properties and other information, which is necessary for use in equipment sizing and material selection, are unknown for this future case (Case 1a). Referring to Table 3-65, shown previously, the total retrofit investment costs used for Case 1 was $400,094,000. This same value was also used for Case 1a of this sensitivity study. Specific investment costs are calculated to be 1,319 and 1,253 $/kWe-new for solvent regeneration energy values of 1,550 and 1,200 Btu/lbm-CO_2 respectively (Case 1 and Case 1a). The operating and maintenance costs for Case 1a are slightly lower than Case 1 due to the increase in net power.

5.3 Economic Analysis

Incremental LCOE breakdown and CO_2 mitigation costs are shown in Table 5-3 and Figure 5-5 for the two cases. Case 5 from the previous study (Bozzuto et al., 2001) is not shown in these LCOE tables or graphs because, as was discussed in Section 3.4 previously, the design and associated investment costs for Case 5 were not developed on a comparable basis to Case 1. The various components that make up the incremental LCOE (capital, fixed O&M, variable O&M, and fuel) are broken out in Table 5-3 and Figure 5-5.

Table 5-3: Incremental Cost of Electricity Breakdown & Mitigation Costs

Economic Property	Case 1 (90% Capture)	Case 1a (90% Capture)
Capital Component (¢/kWh)	3.10	2.95
Fixed O&M (¢/kWh)	0.13	0.14
Variable O&M (¢/kWh)	3.66	3.21
Feedstock O&M (¢/kWh)	0.03	0.03
Total:	6.92	6.32
Mitigated CO_2 ($/ton)	81	73
Mitigated CO_2 ($/tonne)	89	81
Captured CO_2 ($/ton)	81	73
Captured CO_2 ($/tonne)	89	81

CO_2 mitigation cost impacts are also shown in Figure 5-5 for the two solvent regeneration energy levels all with 90% CO_2 capture. The mitigation costs range from about 81-89 $/tonne for these cases.

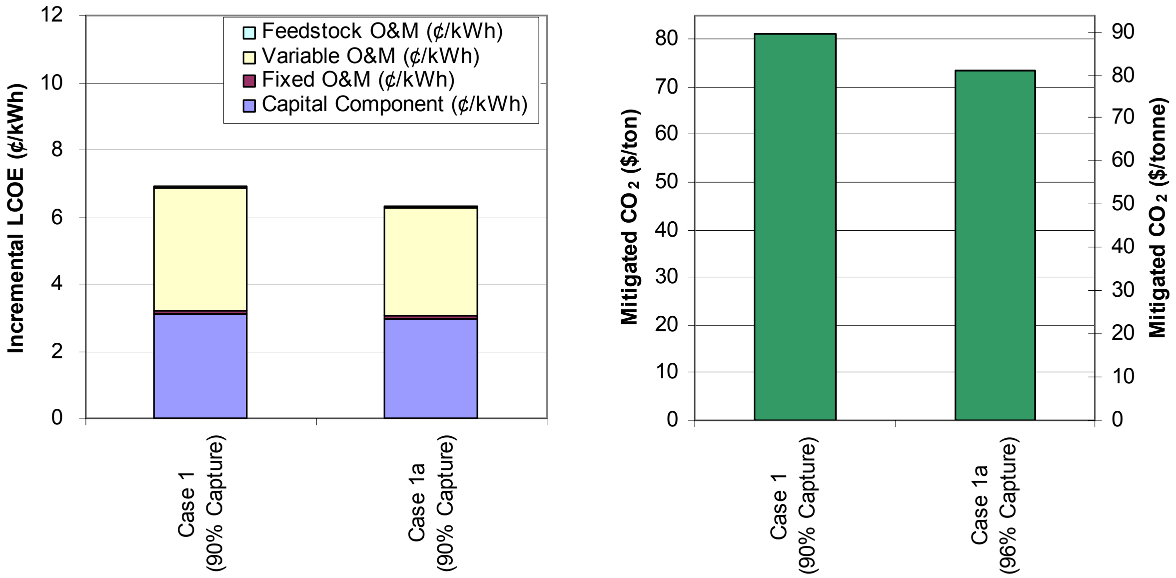


Figure 5-5: Incremental LCOE Breakdown and CO_2 Mitigation Cost

The incremental LCOE and mitigation cost results are also plotted as a function of solvent regeneration energy for these 90% capture level cases in Figure 5-6. Case 5 with 96% capture from the previous study is also shown for comparison. As shown in Figure 5-6, incremental cost of electricity is quite sensitive to changes in solvent regeneration energy. The incremental LCOE was calculated to change by about 5.41 ¢/kWh for a change in solvent regeneration energy of 1,000 Btu/lbm-CO_2. A similar impact (58.52 $/tonne) was calculated for mitigation cost.

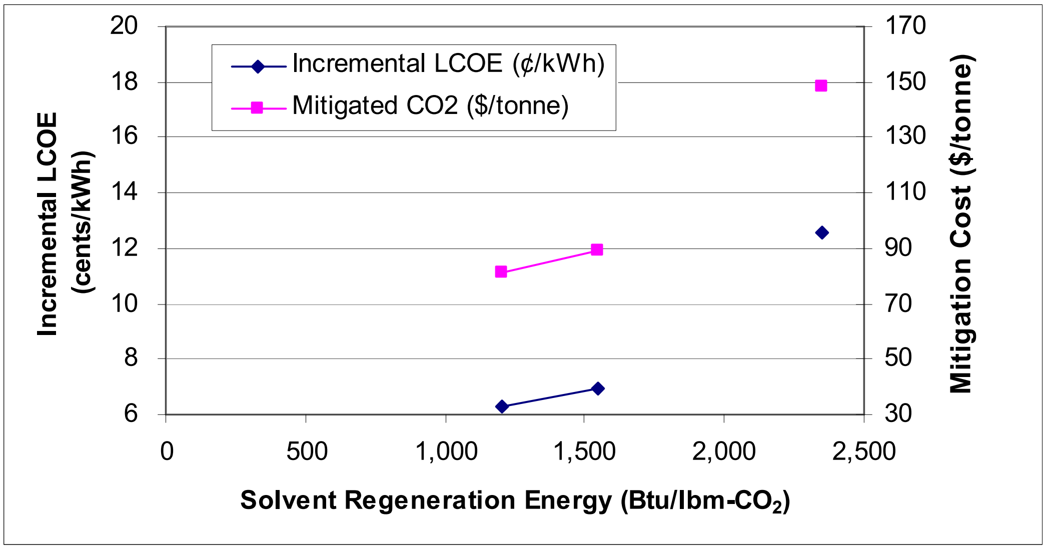


Figure 5-6: Incremental LCOE and Mitigation Cost vs. Solvent Regeneration Energy

6 REPLACEMENT POWER DISCUSSION

When CO_2 capture equipment is retrofit to a power generation plant, the net electrical output of the plant is reduced as a result of the parasitic power and/or heat requirements of the capture plant. To meet customer demand, it is necessary to replace this lost power. Therefore, sufficient replacement power should be provided to bring net plant electrical output back to the original level. Furthermore, in calculating avoided carbon amounts and costs, the cost and carbon emissions of the facilities providing the replacement power must be included in the calculations.

In considering replacement power options for the existing fleet, there is not one answer that fits every case. Each situation will be different. Factors that will vary from site to site include location, availability of unused land, surrounding land uses, climate, state and federal regulations, labor availability, etc. Some of the possible considerations for replacement (make-up) power are discussed below.

1. Purchase of replacement power from the grid from plants that have spare capacity. This is clearly the simplest option, but it is feasible only in the short range for a few plants. If a large number of plants tried to purchase replacement power from the grid, there would not be enough spare capacity without reducing the redundancy in the system to an unacceptable level. Another drawback is that most of this replacement power would come from coal-fired plants without carbon capture, and factoring this into the calculation would reduce the benefit of the CO_2 capture technology in the retrofit plant.
2. Build a new supercritical pulverized coal (SCPC) plant to provide replacement power for several CO_2 retrofit plants. SCPC plants have the highest efficiency (~40% based on HHV) of any PC plant. Thus, the replacement power would be generated at a higher efficiency than the retrofit unit. If the supercritical unit were not fitted with CO_2 capture technology, then the avoided carbon emissions of the retrofit plant would be reduced.
3. Build a natural gas combined cycle (NGCC) plant to provide replacement power. Like SCPC, NGCC has high efficiency and would replace power at a higher efficiency than the original plant being retrofit. Also, if CO_2 were not captured, the carbon-avoided penalty would be less, because natural gas combustion does not produce as much CO_2 as coal combustion. However, rising natural gas prices may hinder this option.
4. Build a nuclear plant to supply replacement power for several CO_2 retrofit plants. The advantage of nuclear is that the replacement power would be supplied by a plant that does not emit CO_2. However, under current circumstances in the U.S., it would be very difficult to get a nuclear plant built. Permitting and construction could require as long as ten years or more with attendant high costs.
5. Use some form of renewable energy as replacement power. Like nuclear, renewable energy sources do not produce any net CO_2 emissions. Of the various renewable options (wind, solar, tidal, geothermal, biomass, hydroelectric), wind appears to have the best prospects for providing replacement power. Solar and biomass (another form of solar) are dilute resources; tidal would require a large amount of engineering and most hydroelectric sites have either already been exploited or would engender so much opposition as to be infeasible. Wind power has the advantage that it can be implemented in small increments, so that economies of scale are not as important as with some of the

other options. Clearly, some sites would be more amenable to a nearby wind farm than other sites.

6. Use an integrated gasification combined cycle (IGCC) plant to produce replacement power. IGCC, with its combined cycle, has a thermal efficiency on par with SCPC (~40% based on HHV). However, because IGCC costs are significantly higher (20%-30%) than SCPC, the COE is also correspondingly higher. As with other fossil fuel fired technologies discussed above, if CO_2 is not captured, the avoided carbon emissions for the retrofit plant would be reduced.
7. Use an emerging technology to provide replacement power. Two emerging technologies that are under development are oxy-fuel combustion and chemical looping. Both these technologies are designed to produce flue gases with high CO_2 concentrations that can be easily purified to sequestration specifications. Neither technology is currently being used commercially, but both show promise. Since CO_2 capture is inherent in these technologies, they would not penalize the avoided carbon emissions of the retrofit plant.

As indicated above, which of these options is best for a particular power plant will depend on factors unique to that specific situation. In some cases, particularly with smaller or older units, it might be preferable to re-power the entire plant with SCPC, NGCC, IGCC, or some other technology rather than just retrofitting to a CO_2 capture plant. In this way, a higher efficiency option could be chosen, and CO_2 capture technology could be integrated into the design from the beginning, which is always more efficient and economic. Other factors, such as dispatching issues, could affect the entire replacement power picture.

In this study, a replacement/make-up power cost of 6.40 ¢/kWh was applied to each Case. The value reflects the levelized cost of electricity from a new subcritical pulverized bituminous coal (Greenfield) plant without carbon capture. The resulting make-up power cost was allocated to the variable O&M cost category within this study because of its dependency on the net power production and capacity factor. The MUPC of 6.40 ¢/kWh represents the lower cost perspective for a range of bituminous coal Greenfield Plant designs with and without carbon capture (i.e., ~6.33 to 11.42 ¢/kWh).

7 CONCLUSIONS AND RECOMMENDATIONS FOR FUTURE WORK

Conclusions

No major technical barriers exist for retrofitting AEP's Conesville Unit #5 to capture CO_2 with post-combustion amine-based capture systems. Lower levels of CO_2 capture can be achieved by simply bypassing some of the flue gas around the CO_2 capture system and only processing a fraction of the total flue gas in the amine-based capture systems. Flue gas bypassing was determined to be the most cost-effective approach to obtain lower CO_2 recovery levels. Nominally, 4 acres of new equipment space is needed for the amine-based capture and compression system (Case 1, 90% capture level) and this equipment is located in three primary locations on the existing 200-acre power plant site, which accommodates a total of 6 power generation units. The CO_2 absorber equipment, which occupies about 1 acre, is located just west, adjacent to the Unit #5 FGD system. The CO_2 stripper equipment, which occupies about 1 acre, is located just south of the Unit #5 turbine building with the CO_2 compression and liquefaction system, which occupies about 2 acres, is located just south of the strippers between two banks of existing cooling towers. Slightly less acreage is needed as the capture level is reduced. If all 6 units on this site were converted to CO_2 capture, it may be difficult to accommodate all the new CO_2 capture equipment on the existing site and additional land might need to be purchased.

This report is an update of a previous study (Bozzuto et al., 2001) and it demonstrates the advancement of post-combustion amine-based capture technologies. Solvent regeneration energy was reduced by ~34%, which provided an improvement in plant thermal efficiency of 4.2 percentage points (from 20.2% to 24.4%). Additionally retrofit specific investment costs ($/kWe) were reduced by 52% and incremental COE was reduced by 45%. Demonstration of advanced low cost technologies is critical to carbon capture and sequestration (CCS) for both existing and new plants.

Energy requirements and power consumption are high, resulting in significant decreases in overall power plant efficiencies, which range from about 24.4% to 31.6% as the CO_2 capture level decreases from 90% to 30% for Cases 1-4, as compared to 35% for the Base Case (all HHV basis). The efficiency decrease is essentially a linear function of CO_2 recovery level. Specific carbon dioxide emissions were reduced from about 908 g/kWh (2 lbm/kWh) for the Base Case to 132-704 g/kWh (0.29-1.55 lbm/kWh) as the CO_2 recovery level decreases from 90% to 30%. Recovery of CO_2 ranged from 30% to 90% for the new cases (Cases 1-4) and 96% for the updated case (Case 5) of the previous study.

Specific incremental investment costs are also high, ranging from about $540 to $1,319/kWe-new, depending on CO_2 capture level for the current study. The specific investment cost is also nearly a linear function of CO_2 recovery level, although equipment selections and economy of scale effects make this relationship much less linear than efficiency is.

All cases studied indicate significant increases to the LCOE as a result of CO_2 capture. The incremental COE, as compared to the Base Case (air firing without CO_2 capture), ranges from 2.31 to 6.92 ¢/kWh (depending on CO_2 capture level). Similarly CO_2 mitigation cost increases slightly from $89 to $113/tonne of CO_2 avoided as the CO_2 capture level decreases from 90% to 30%. The roughly linear decrease in LCOE with reduced CO_2 capture indicates that there is no optimum CO_2 recovery level. Economic sensitivity studies indicate the incremental LCOE is most impacted by the following parameters (in given order): CO_2 selling price, capacity factor, total investment cost, and make-up power cost.

The updated specific investment cost for Case 5/Concept A of the previous study (Bozzuto et al., 2001) was ~$2,786/kWe-new. The update of Case 5 did not include the process design or equipment selections.

The advanced amine is expected to provide significant improvement to the plant performance and economics. Use of the advanced amine in comparison to the Kerr-NMcGee/ABB Lummus amine for 90% CO_2 capture showed an improvement in thermal efficiency of about 3.5 percentage points, although, as pointed out above, the process design for Case 5 was not updated in this study. An equitable comparison of specific costs ($/kWe) and economics (LCOE, mitigation costs) was not possible since the amine system design for the previous study was not consistent with the current designs using the advanced amine, as explained in more detail in Section 3.4.

The commercial implementation of these amine-based post-combustion capture systems will be several years in the future and research is continually improving the performance of amine solvents and systems. A sensitivity analysis was completed that showed the effect of anticipated reductions in solvent regeneration energy (for the 90% capture level). The solvent regeneration energy cases investigated were 1,550 and 1,200 Btu/lbm-CO_2. Plant thermal efficiency is shown to be very sensitive to changes in solvent regeneration energy. Plant thermal efficiency was calculated to change by about 3.7 percentage points (or about 15% relative to Case 1 @ 24.5% thermal efficiency) for a change in solvent regeneration energy of 1,000 Btu/lbm-CO_2. Similarly, incremental LCOE is also quite sensitive to changes in solvent regeneration energy. The incremental LCOE was calculated to change by about 0.8 ¢/kWh (about 10% relative to Case 1 @ 6.92 ¢/kWh) for a change in solvent regeneration energy of 1,000 Btu/lbm-CO_2.

Recommendations for Future Work

Recommendations for future work for CO_2 capture from existing coal-fired utility-scale electric power plants are listed below:

- Re-do case study using best-in class-solvents. Within this context, include the use of modified steam turbine and updated process design, equipment selection, and cost to fully quantify improvements with advanced solvents
- Update the process design, equipment selections, costs, and economic analysis of the Case 5/Concept A CO_2 capture/compression/liquefaction system in order to fully quantify the improvements available with use of the advanced amine system
- Apply the results from best-in-class study to the existing U.S. coal fleet to determine the overall economic impacts and CO_2 emissions reductions, keeping in mind certain criteria:
    - Units of certain size range (large units)
    - Units of certain age group (newer units)
    - Units located near sequestration sites
    - High capacity factor units (Base Loaded)

- Because high CO_2 loadings in the rich amine accelerate corrosion, future studies should include methods or additives to reduce the corrosion to acceptable levels
- Demonstrate best-in-class solvents on a commercial scale

- Because high CO_2 loadings in the rich amine accelerate corrosion, future studies should include methods or additives to reduce the corrosion to acceptable levels

8 BIBLIOGRAPHY

Baily, D.W., and Feron, P., “Post-Combustion Decarbonization Process,” Oil and Gas Science and Technology – Rev., IFP, Vol. 60 (2005), No. 3, pp. 461-474.

Barchas, R. and Davis, R., “The Kerr-McGee/Lummus Crest Technology for the Recovery of CO_2 from Stack Gases,” Presented at the First International Conference on Carbon Dioxide Removal, Amsterdam, The Netherlands, March 4-6, 1992.

Bozzuto, C.R., Nsakala, N., Liljedahl, G., Palkes, M., Marion, J., Vogel, D., Fugate, M., Guha, M., “Engineering Feasibility and Economics of CO_2 Sequestration/Use on an Existing Coal-Fired Power Plant. Volume I: AEP’s Conesville Power Plant Unit No. 5 Retrofit Study,” Prepared for the Ohio Department of Development, Ohio Coal Development Office and US Department of Energy, National Energy Technology Laboratory (June 30, 2001).

Chapel, D. G., and Mariz, C. L., “Recovery of CO_2 from Flue Gases: Commercial Trends,” Presented at the Canadian Society of Chemical Engineers Meeting, October 4-6, 1999, Saskatoon, Saskatchewan, Canada.

Chinn, D., Eimer, D., and Hurst, P., “CO_2 Capture Project: Post-Combustion Integrated Technology (BIT) Overview.” Presented at the Third National Conference on Carbon Sequestration, National Energy Technology Laboratory, Alexandria, VA, May 3-7, 2004.

Choi, G. N., Chu, R., Degen, B., Wen, H., Richen, P. L., and Chinn, D., “Cost Effective amine Plant Design for Post Combustion CO_2 Capture from Power Plant Flue Gas,” Presented at the Third National Conference on Carbon Sequestration, National Energy Technology Laboratory, Alexandria, VA, May 3-6, 2004.

Choi, G. N., Chu, R., Degen, B., Wen, H., Richen, P. L., and Chinn, D., “CO_2 Removal from Power Plant Flue Gas – Cost Efficient Design and Integration Study,” Carbon Dioxide Capture for Storage in Deep Geologic Formations, Volume 1, D.C. Thomas and S. M. Benson (Eds.), ©2005 Elsevier Ltd.

Dakota Gasification Company, “Carbon Dioxide Specifications,” April 2005, http://www.dakotagas.com/specs/co2spec.pdf

DOE/NETL, “2006 Cost and Performance Comparison of Fossil Energy Power Plants,” Volume 1: Bituminous Coal and Natural Gas to Electricity, Draft Final, DOE/NETL – 401/053106, May 2006.

“Improvement in Power Generation with Post-Combustion Capture of CO_2,” IEA Report Number PH4/33, November 2004.

Liljedahl, G., Marion, J., Nsakala, N., Bozzuto, C., Palkes, Vogel, D., Gupta, J.C., Guha, M., Johnson, H., and Plasynski, S., “Technical and Economic Feasibility of CO_2 Capture on an Existing Coal-Fired Power Plant,” for Presentation at the 2001 International Joint Power Generation Conference, New Orleans, LA, June 4-7, 2001.

Marion, Jr., Nsakala, N., Bozzuto, C., Liljedahl, G., Palkes, Vogel, D., Gupta, J.C., Guha, M., Johnson, H., and Plasynski, S., “Engineering Feasibility of CO_2 Capture on an Existing Coal-Fried Power Plant,” Presented at the 26th International Conference on Coal Utilization & Fuel Systems, Clearwater, FL, March 5-8, 2001.

“Nelson-Farrar Refinery Construction Cost Index,” Oil & Gas Journal, May 1, 2006.

Palkes, M., Liljedahl, G., Nsakala, N., McDonald, M., and Gupta, J.C., “Preliminary Design of a CO_2/O_2 Combustion Retrofit to an Existing Coal-Fired Boiler for CO_2 Extraction,” Presented at Electric Power Gen ’99 Conference, Baltimore, MD, April 20-22, 1999.

Resnik, K.P., Yeh, J.T., Pennline, H.W.,“Aqua Ammonia process for Simultaneous Removal of CO_2, SO_2, and NOx,” International Journal of Environmental Technology and Management, Vol. 4, Nos. 1/2, 2004.

Rochelle, G., “Research Needs for CO_2 Capture from Flue Gas by Aqueous Absorption/ Stripping,” Draft of Final Report for DOE P.O. NO. DE-AF26-FT01029, University of Texas at Austin, Texas, September 2000.

Rubin, E.S., Rao, A.B., “A Technical, Economic and Environmental Assessment of Amine-based CO_2 Capture Technology for Power Plant Greenhouse Gas Control,” Annual Technical Progress Report, Contract No.: DE-FC26-00NT40935, October, 2002.

9 APPENDICES

Four appendices are included in this section, as listed below:

- Appendix I - Plant Layout Drawings
- Appendix II - Equipment Lists for the CO_2 recovery systems
- Appendix III - Economic Sensitivity Studies
- Appendix IV - Let Down Turbine Technical Information

9.1 Appendix I – Plant Drawings (Cases 1-5)

This appendix contains all layout drawings developed for this project for Cases 1-4 and Case 5/Concept A. Also included is a plot plan of the existing site without modifications for reference. The drawings provided are listed below:

Existing Plant:

66-530.00	Plot Plan – Existing Overall Conesville Site (before CO_2 unit addition)

Cases 1-4

15154-003	Plot Plan – Cases 1-4: Flue Gas Cooling & CO_2 Absorption Equipment Layout
15154-002	Plot Plan – Cases 1-4: Solvent Stripping and Compression Equipment Layout
15154-001	Plot Plan – Cases 1-4: Overall Plot Plan for Modified Conesville Unit #5

Case 5/Concept A:

U01-D-0208	Plot Plan – Case 5/Concept A: Flue Gas Cooling & CO_2 Absorption Equipment Layout
U01-D-0214	Plot Plan – Case 5/Concept A: Solvent Stripping Equipment Layout
U01-D-0204	Plot Plan – Case 5/Concept A: CO_2 Compression & Liquefaction Equipment Layout
U01-D-0211	Plot Plan – Case 5/Concept A: Overall Equipment Layout Conceptual Plan
U01-D-0200R	Plot Plan – Case 5/Concept A: Modified Overall Site Plan

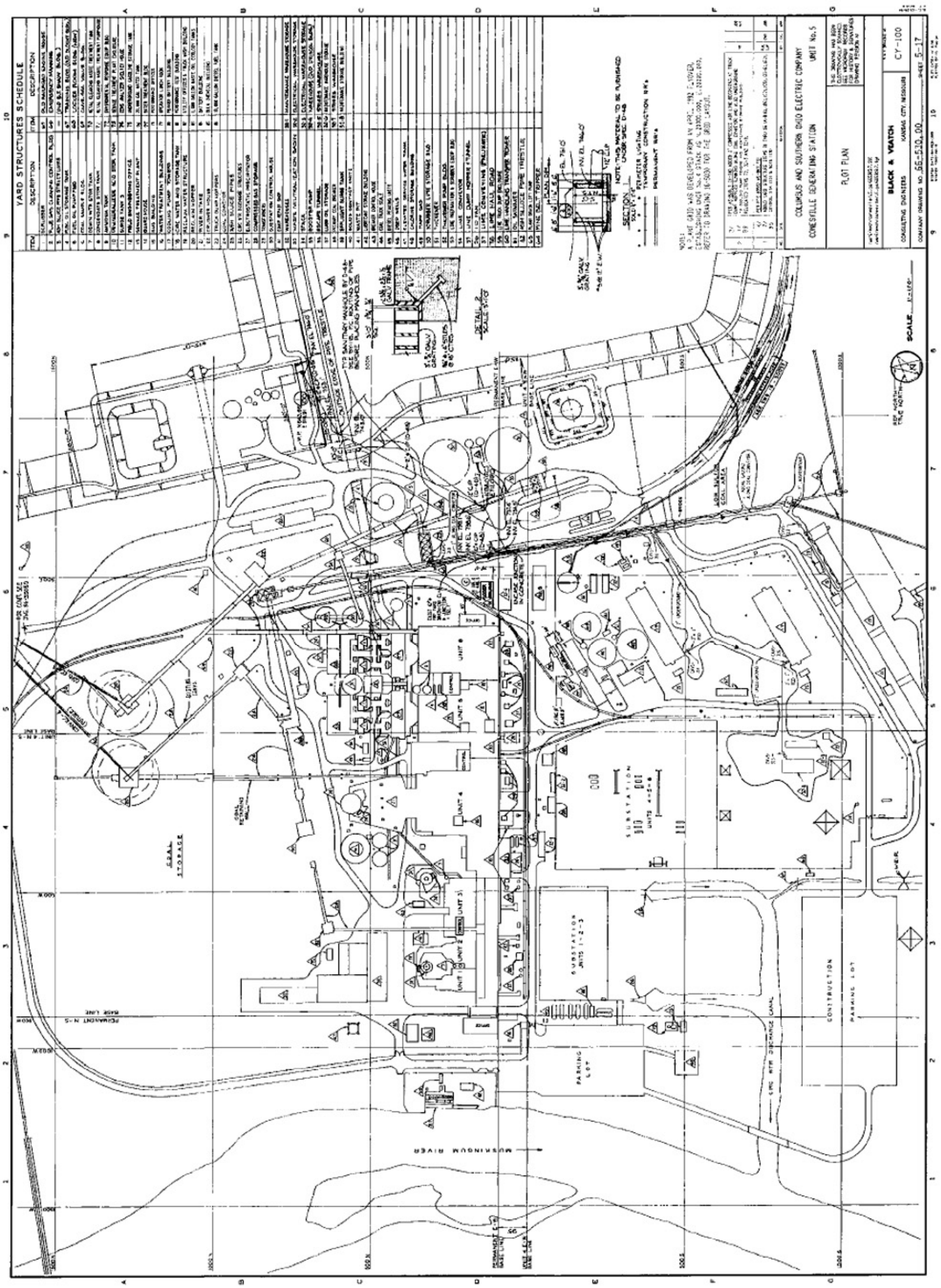


Figure 9-1: Existing Overall Site (before CO_2 Unit Addition)

Cases 1-4

The plant layout drawings prepared for the Cases 1-4 CO_2 Recovery Systems are as follows:

15154-003 Plot Plan – Cases 1: Flue Gas Cooling & CO_2 Absorption Equipment Layout
15154-002 Plot Plan – Cases 1: Solvent Stripping and Compression Equipment Layout
15154-001 Plot Plan – Cases 1: Overall Plot Plan for Modified Conesville Unit #5

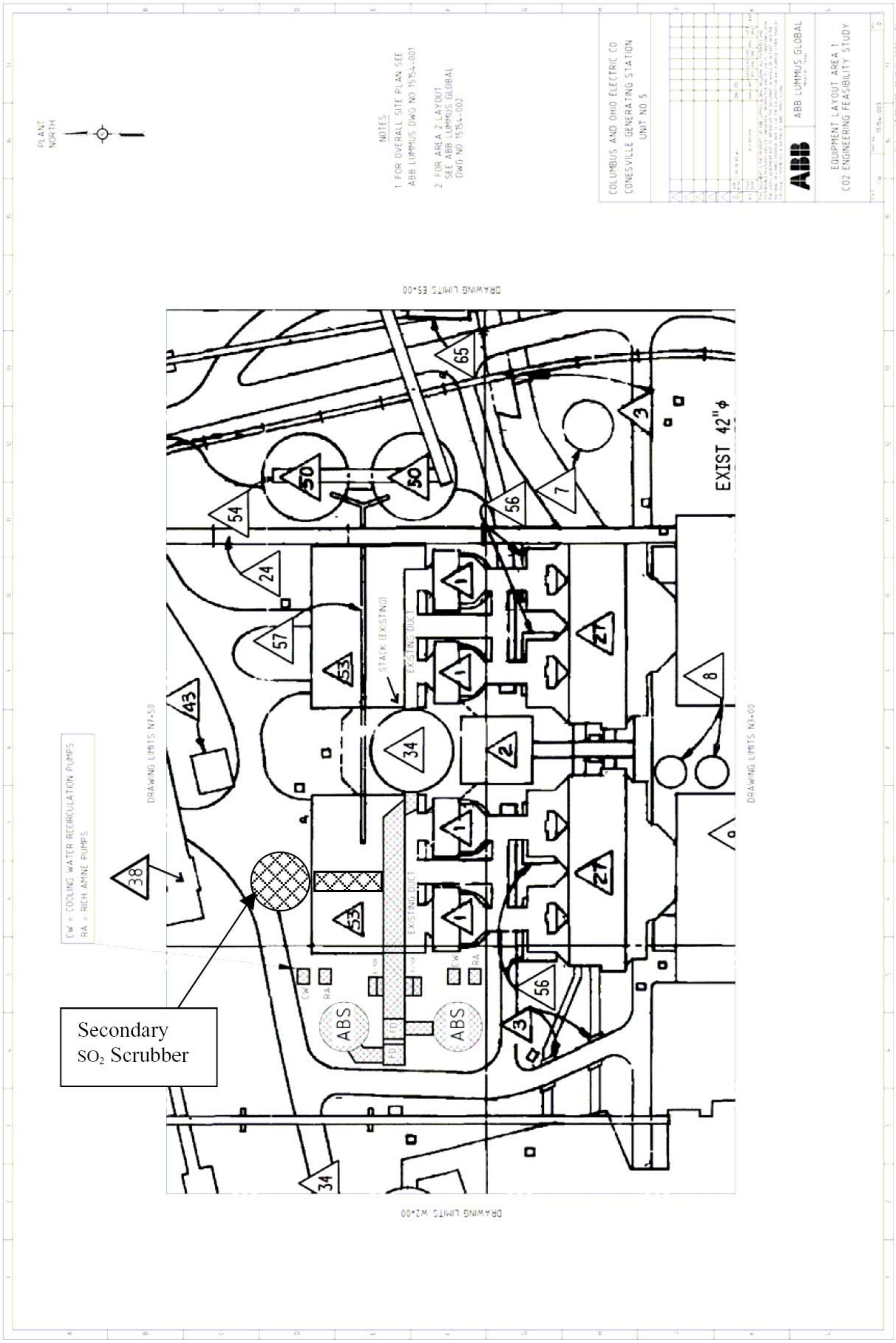


Figure 9-2: Cases 1-4 Flue Gas Cooling & CO_2 Absorption Equipment Layout

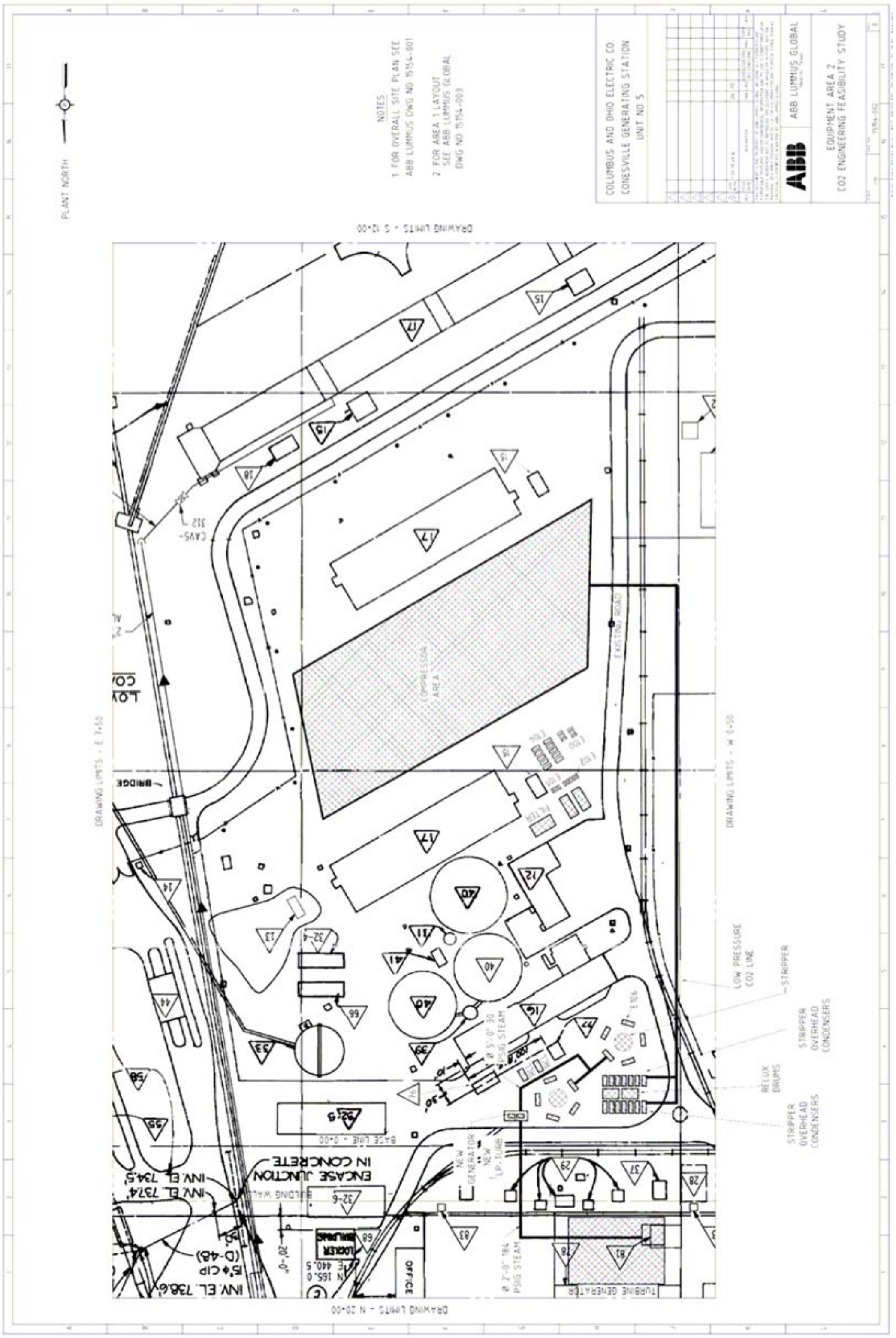


Figure 9-3: Cases 1-4 Solvent Stripping and Compression Equipment Layout

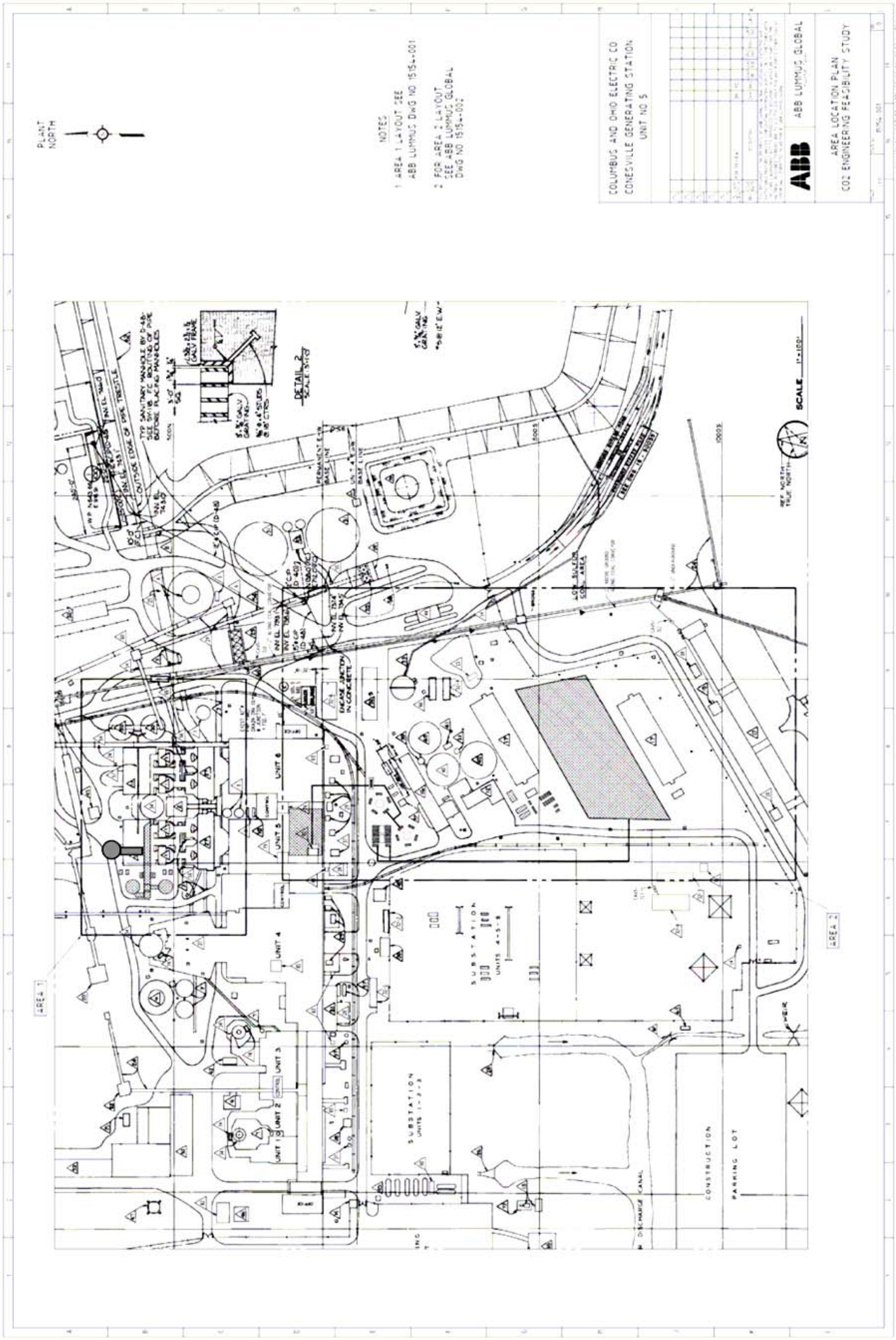


Figure 9-4: Cases 1-4 Overall Plot Plan for Modified Conesville Unit #5

Case 5/Concept A:

The plant layout drawings prepared for the Case 5/Concept A CO_2 Recovery System are as follows

U01-D-0208 Plot Plan – Case 5/Concept A: Flue Gas Cooling & CO_2 Absorption Equipment Layout
U01-D-0214 Plot Plan – Case 5/Concept A: Solvent Stripping Equipment Layout
U01-D-0204 Plot Plan – Case 5/Concept A: CO_2 Compression & Liquefaction Equipment Layout
U01-D-0211 Plot Plan – Case 5/Concept A: Overall Equipment Layout Conceptual Plan
U01-D-0200 Plot Plan – Case 5/Concept A: Modified Overall Site Plan

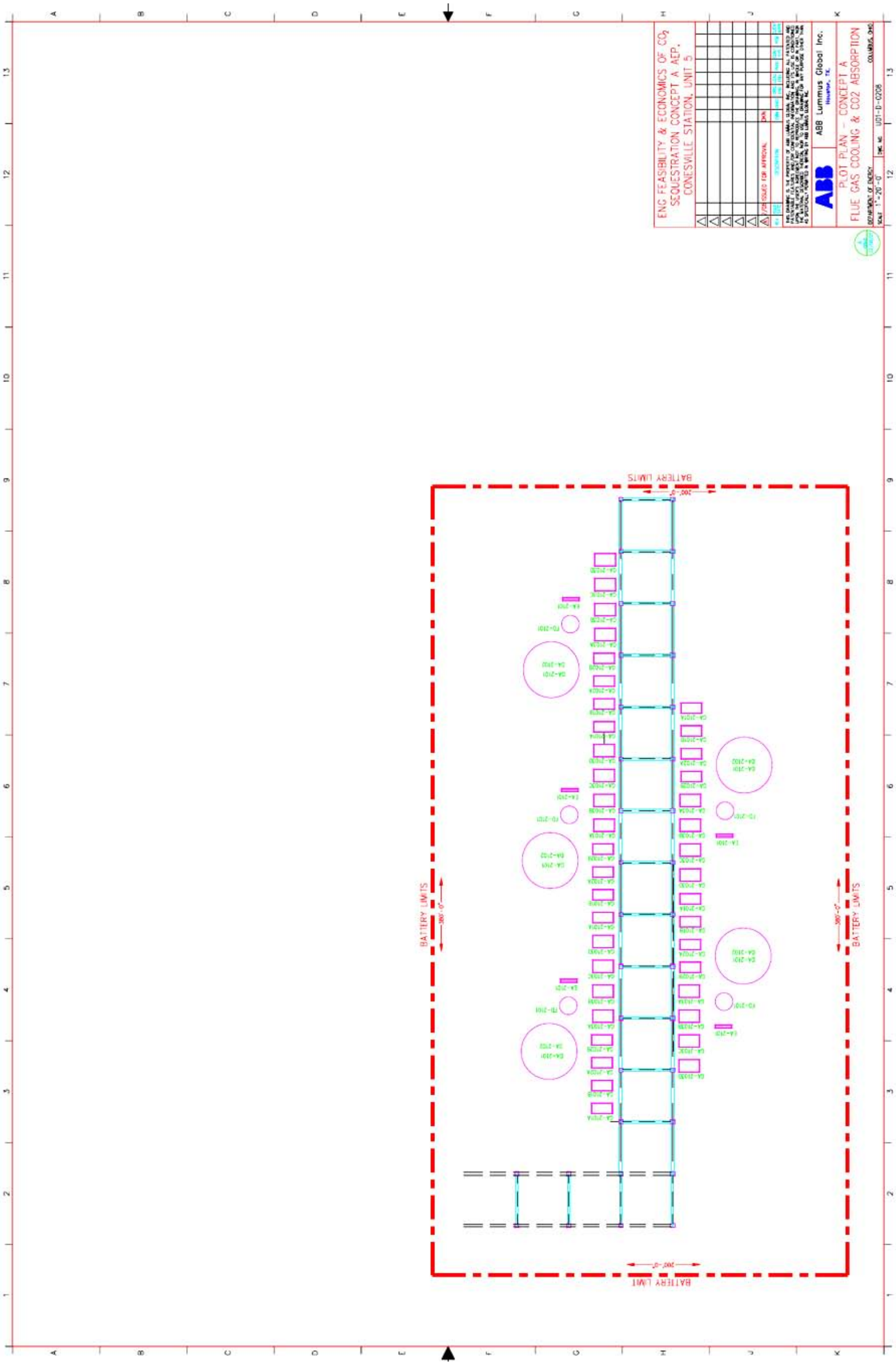


Figure 9-5: Case 5/Concept A – Flue Gas Cooling & CO_2 Absorption Equipment Layout

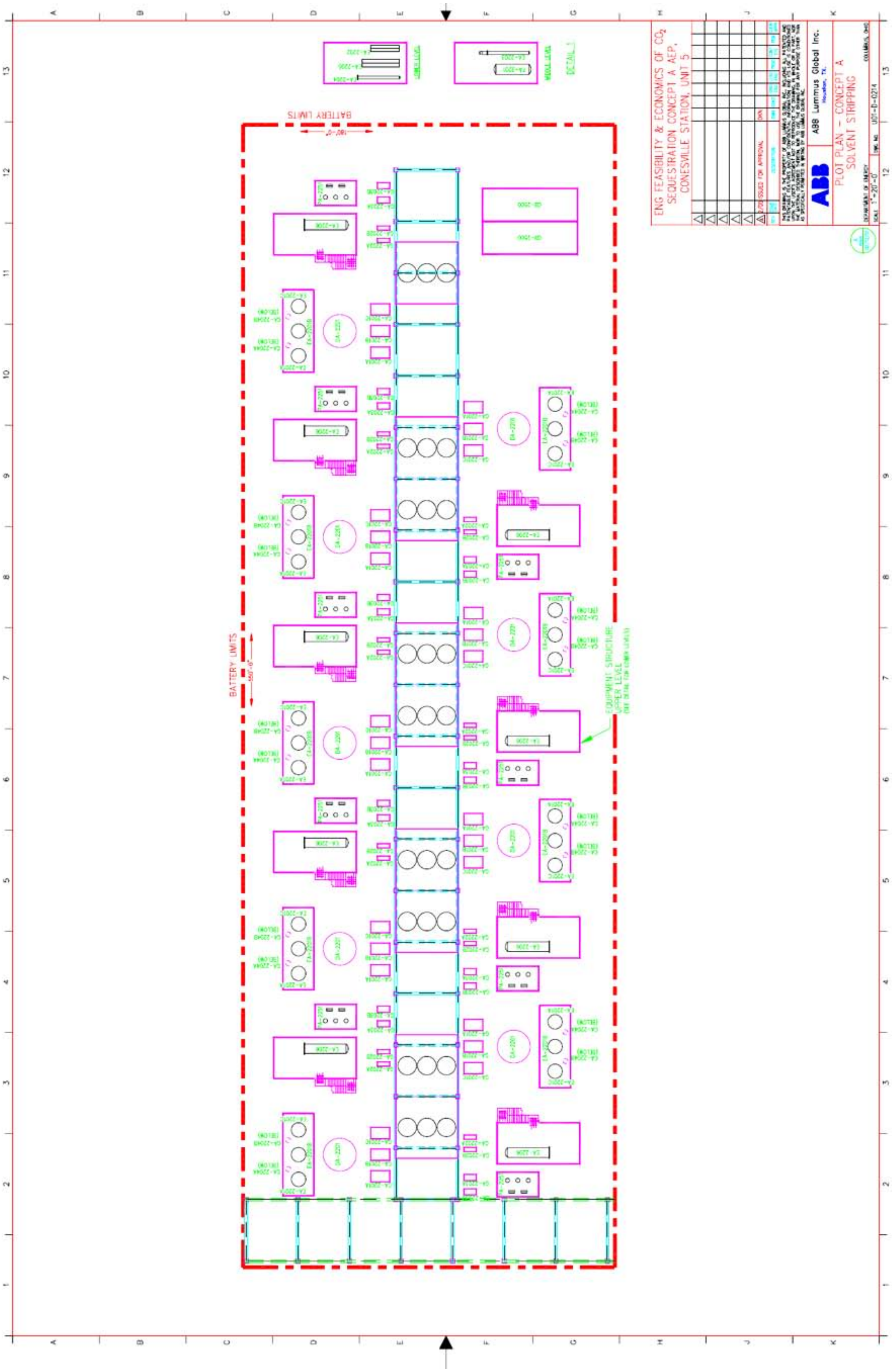


Figure 9-6: Case 5/Concept A – Solvent Stripping Equipment Layout

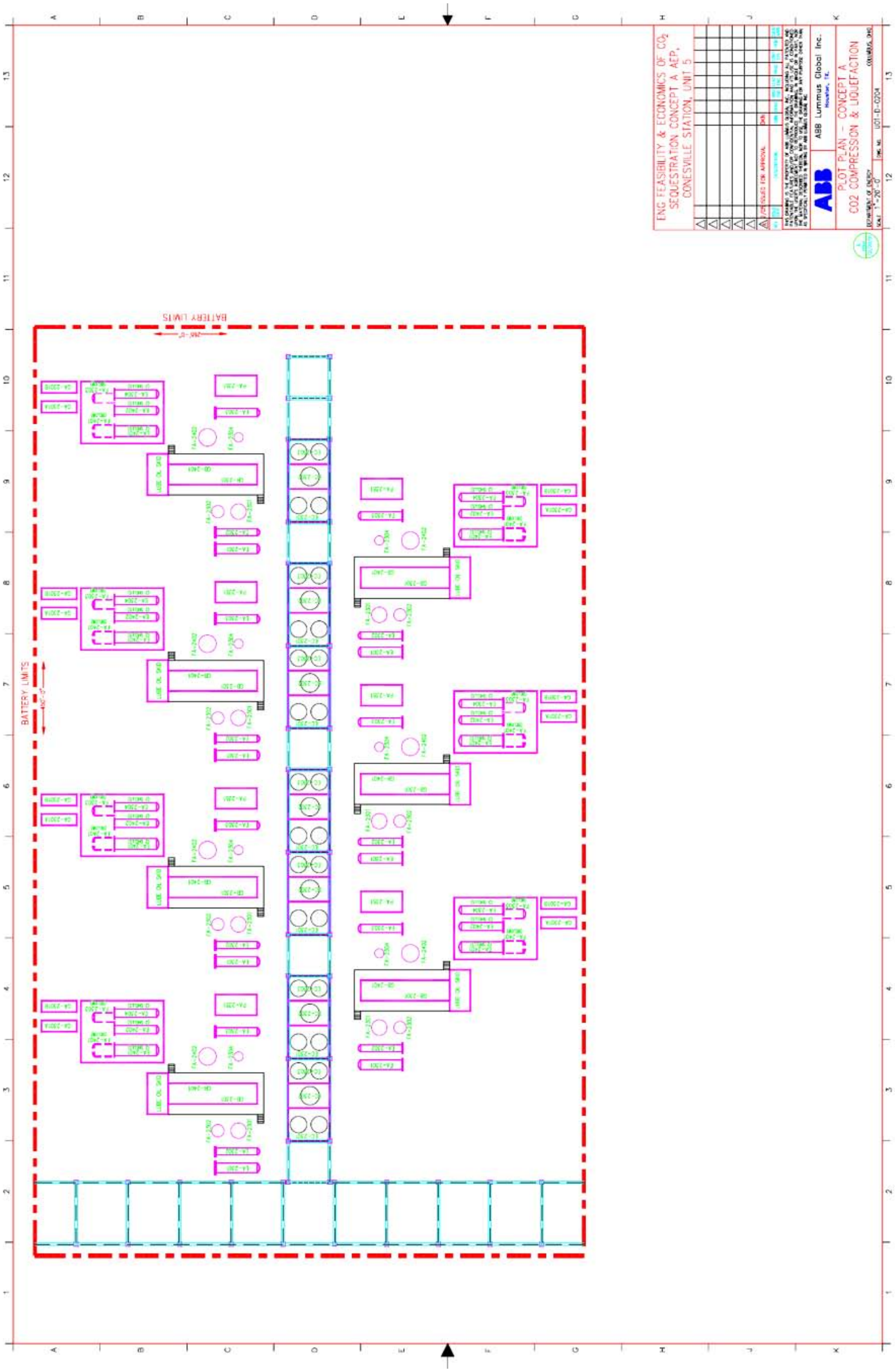


Figure 9-7: Case 5/Concept A – CO_2 Compression & Liquefaction Equipment Layout

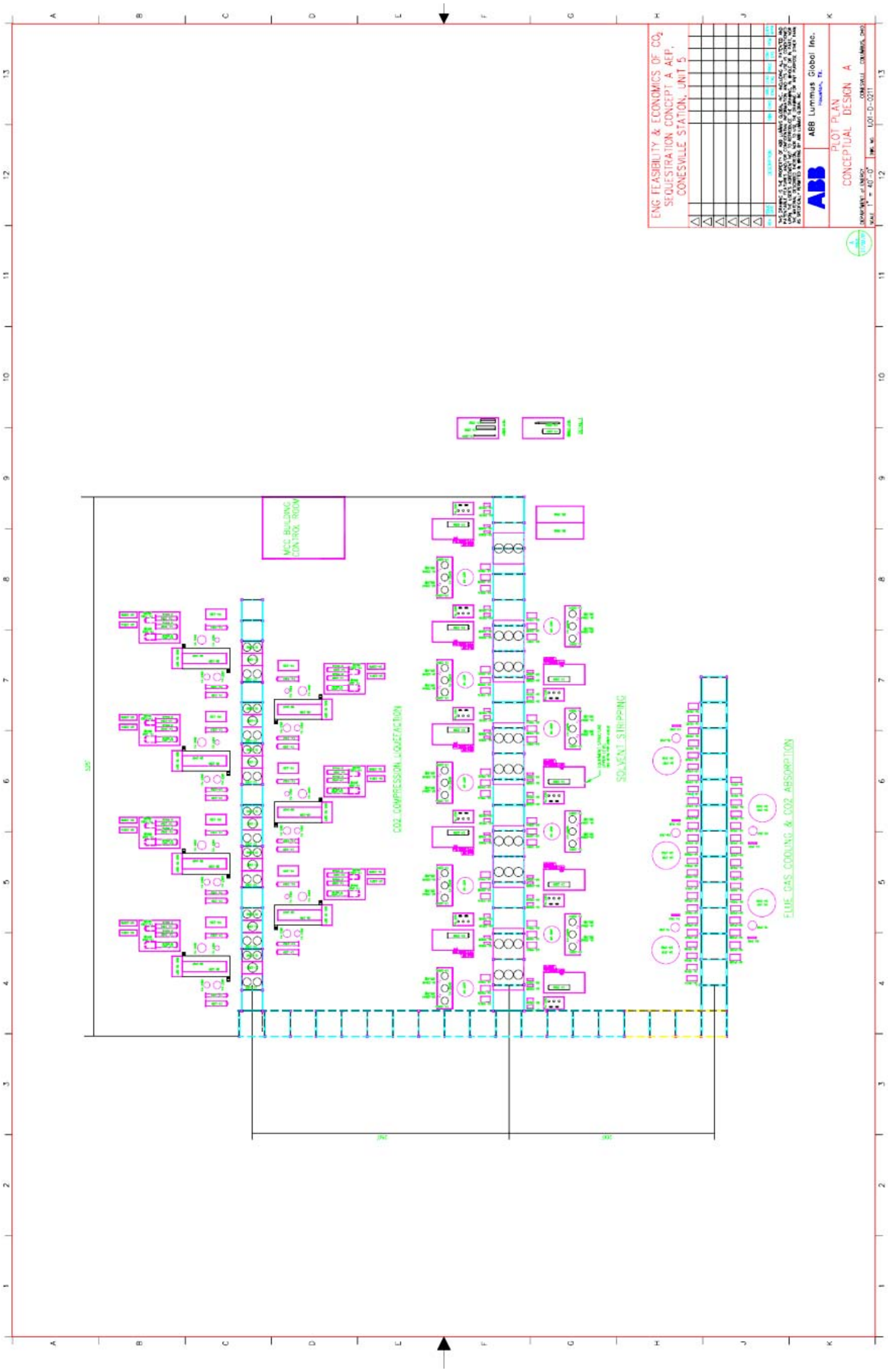


Figure 9-8: Case 5/Concept A Overall Equipment Layout Conceptual Plan

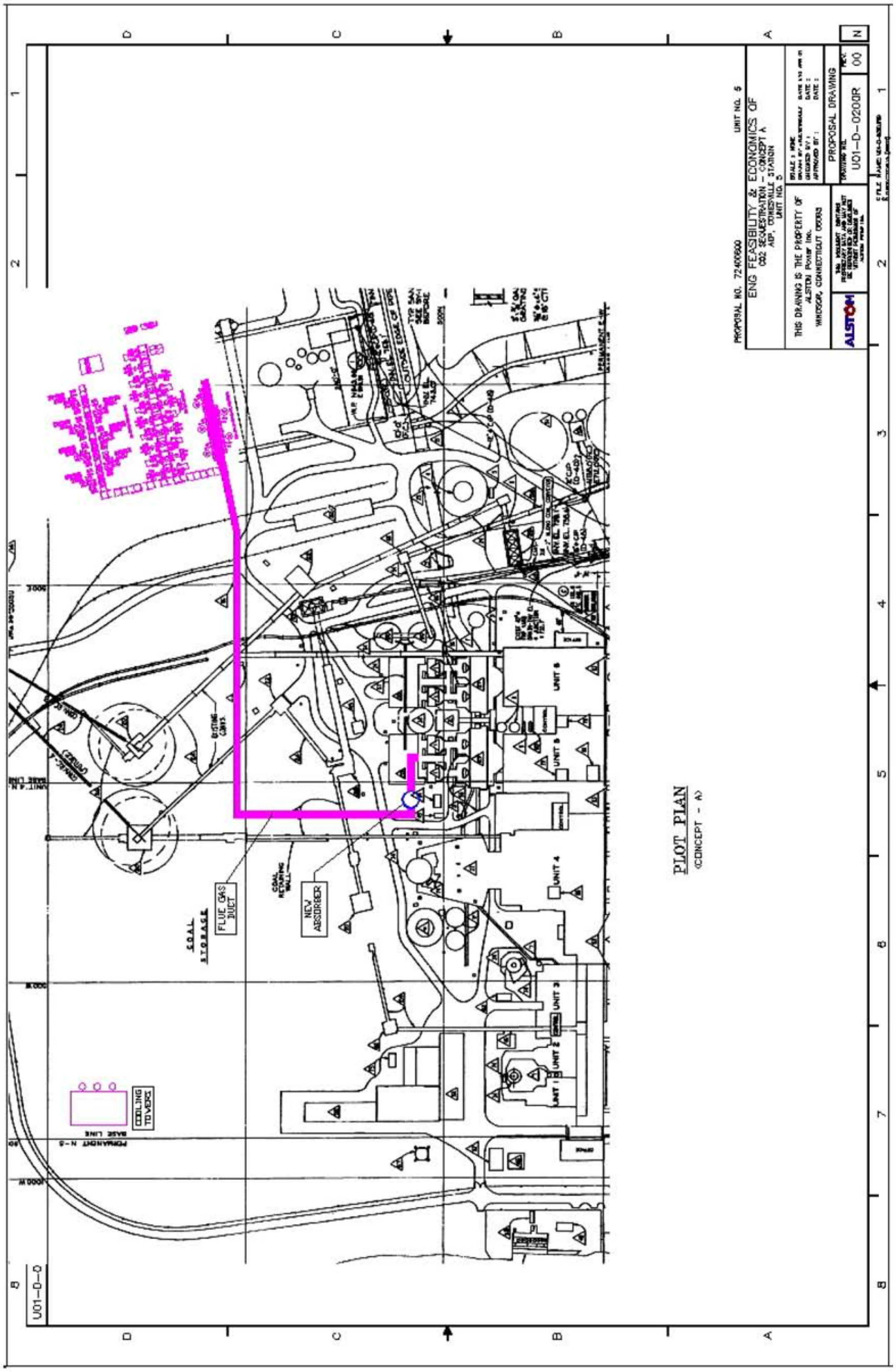


Figure 9-9: Case 5/Concept A – Overall Plot Plan for Modified Conesville Unit #5

9.2 Appendix II - Equipment Lists (Cases 1-5)

This appendix contains equipment lists for the CO_2 Capture Systems of all five cases (Cases 1-4 and Case 5/Concept A). Equipment data has been presented in the so-called “short spec” format, which provides adequate data for a factored cost estimate

Table 9-1: Case 1 CO_2 Capture System Equipment List with Data (90% CO_2 Recovery)

No. Required	Tag no.	Description	Size Parameters	Material
incl w/abs		Direct Contact Flue Gas Cooler	34' ID x 34' S/S, DP 2.5 psig/ 0.7 psi vac	CS/SS
2		CO_2 Absorber	34' ID x 92' S/S, DP 2.5 psig/ 0.7 psi vac	CS/SS
2		Solvent Stripper	22' ID x 50' S/S, DP 35 psig/ FV	CS/SS
10	E-106	Solvent Stripper Reboiler	120 MMBTU/HR PHE, 90 psig/ 90 psig	CS/SS
2	E-109	Solvent Stripper Reclaimer	21 MMBTU/HR, DP S/T, 120 psig/ 190 psig	CS/TI
2		Solvent Reclaimer Effluent Cooler	20 MMBTU/HR, DP S/T, 150 psig/ 150 psig	CS/TI
12	E-105	Solvent Stripper CW Condenser	20 MMBTU/HR, DP PHE, 150 psig/ 300 psig	SS/SS
4	E-100	Rich / Lean Solvent Exchanger	158 MMBTU/HR, PHE , 150 psig/ 150 psig	SS316
2	E-101	Rich / Semi-Lean Exchanger	119 MMBTU/HR, PHE, 150 psig/ 150 psig	SS
4	E-102	Lean / Semi-Lean Exchanger	61 MMBTU/HR, PHE, 150 psig/ 150 psig	SS
2	E-108	Absorber Feed Exchanger	117 MMBTU/HR, PHE, 150 psig/ 150 psig	SS
6	E-104	Lean Solvent Exchanger	59 MMBTU/HR, PHE 150 psig/ 150 psig	SS316
2	E-111	Propane Refrigeration De-superheater	25 MMBTU/HR, DP S/T, 300 psig/ 100 psig	CS/CS
2		Propane Refrigeration Condenser	52 MMBTU/HR, DP S/T, 300 psig/ 100 psig	CS/CS
2		Propane Refrigeration Sub-cooler	20 MMBTU/HR, DP S/T, 300 psig/ 2500 psig	CS/LTCS
2		CO_2 Compressor 1st stage Cooler	15 MMBTU/HR, DP 75 psig	SS
2		CO_2 Compressor 2nd stage Cooler	18 MMBTU/HR, DP 125 psig	SS
2		CO_2 Compressor 3rd stage Cooler	16 MMBTU/HR, DP 235 psig	SS
2		CO_2 Condenser	66 MMBTU/HR, DP S/T, 235 psig/ 300 psig	CS/TI
2		Solvent Stripper Reflux Drum	8'-6” ID x 26' S/S, DP 35 psig/ FV	304L
2		CO_2 Compressor 2nd Stage Suction Drum	11'- 6" ID x 15' S/S, DP 75 psig	CS/SS
2		CO_2 Compressor 3rd Stage Suction Drum	9' ID x 15' S/S, DP 125 psig	CS/SS
2		Liquid CO_2 Surge Drum	7' ID x 21' S/S, DP 235 psig	KCS
2		CO_2 Compressor 3rd Stage Discharge KO Drum	7' ID x 15' S/S, DP 235 psig	CS/SS
2		Propane Refrigeration Surge Drum	15' ID x 45’-6” S/S, DP 300 psig	CS
2		Propane Refrigeration Suction Scrubber	13' ID x 18' S/S, DP 300 psig	LTCS
2		Soda Ash Day Tank	2' ID x 4' S/S, DP atm	CS
4		DCC Water Filter	3532 gpm ea, DP 35 psig	SS
4	Pump-2	Wash Water Pump	2569 gpm ea, DP 29 psi	DI/SS
4	Pump-1	Direct Contact Cooler Water Pump	3532 gpm ea, DP 36 psi	SS/SS
4	P-100	Rich Solvent Pump	6634 gpm ea, DP 92 psi	SS/SS
4	P-102	Lean Solvent Pump	4870 gpm ea, DP 85 psi	SS/SS
4	P-101	Semi-Lean Pump	2168 gpm ea, DP 85 psi	SS/SS
2		Solvent Stripper Reflux Pump	212 gpm ea, DP 75 psi	DI/SS
4		Filter Circ. Pump	332 gpm ea, DP 91 psi	SS/SS
4		LP Condensate Booster Pump	650 gpm ea, DP 237 psi	CI/ SS
7		CO_2 Pipeline Pump	270 gpm ea, DP 1815 psi	CS/CS
2		Soda Ash Metering Pump	.45 gpm ea, DP 50 psi	SS
2		CO_2 Compressor (Motor driven)	15,631 hp ea	SS wheels
2		Propane Refrigeration Compressor	11,661 hp ea	LTCS
2		Corrosion Inhibitor Package	Metering, 22 lb/ hr	
4		Solvent Filter Package	184 gpm ea	
2		CO_2 Dryer Package	161 hp ea compressor, cooler, gas fired heater	
2		Crane for Compressor Bldg		
2		Flue Gas Fans and Ducting	3286 hp ea, SS blades	

Table 9-2: Case 2 CO_2 Capture System Equipment List with Data (70% CO_2 Recovery)

No. Required	Tag no.	Description	Size Parameters	Material
incl w/abs		Direct Contact Flue Gas Cooler	30' ID x 34' S/S, DP 2.5 psig/ 0.7 psi vac	CS/SS
2		CO_2 Absorber	30' ID x 92' S/S, DP 2.5 psig/ 0.7 psi vac	CS/SS
2		Solvent Stripper	19' ID x 50' S/S, DP 35 psig/ FV	CS/SS
8	E-106	Solvent Stripper Reboiler	120 MMBTU/HR PHE, 90 psig/ 90 psig	CS/SS
2	E-109	Solvent Stripper Reclaimer	17 MMBTU/HR, DP S/T, 120 psig/ 190 psig	CS/TI
2		Solvent Reclaimer Effluent Cooler	16 MMBTU/HR, DP S/T, 150 psig, 150 psig	CS/TI
10	E-105	Solvent Stripper CW Condenser	20 MMBTU/HR, DP PHE, 150 psig/ 300 psig	SS/SS
4	E-100	Rich / Lean Solvent Exchanger	158 MMBTU/HR, PHE , 150 psig/ 150 psig	SS316
2	E-101	Rich / Semi-Lean Exchanger	119 MMBTU/HR, PHE, 150 psig/ 150 psig	SS
4	E-102	Lean / Semi-Lean Exchanger	61 MMBTU/HR, PHE, 150 psig/ 150 psig	SS
2	E-108	Absorber Feed Exchanger	91 MMBTU/HR, PHE, 150 psig/ 150 psig	SS
5	E-104	Lean Solvent Exchanger	59 MMBTU/HR, PHE, 150 psig/ 150 psig	SS316
2	E-111	Propane Refrigeration De-superheater	19 MMBTU/HR, DP S/T, 300 psig/ 100 psig	CS/CS
2		Propane Refrigeration Condenser	40 MMBTU/HR, DP S/T, 300 psig/ 100 psig	CS/CS
2		Propane Refrigeration Sub-cooler	15 MMBTU/HR, DP S/T, 300 psig/ 2500 psig	CS/LTCS
2		CO_2 Compressor 1st stage Cooler	12 MMBTU/HR, DP 75 psig	SS
2		CO_2 Compressor 2nd stage Cooler	14 MMBTU/HR, DP 125 psig	SS
2		CO_2 Compressor 3rd stage Cooler	12 MMBTU/HR, DP 235 psig	SS
2		CO_2 Condenser	52 MMBTU/HR, DP S/T, 235 psig/ 300 psig	CS/TI
2		Solvent Stripper Reflux Drum	8' ID x 24' S/S, DP 35 psig/ FV	304L
2		CO_2 Compressor 2nd Stage Suction Drum	10'- 6" ID x 14' S/S, DP 75 psig	CS/SS
2		CO_2 Compressor 3rd Stage Suction Drum	8'-6" ID x 14' S/S, DP 125 psig	CS/SS
2		Liquid CO_2 Surge Drum	6'- 6" ID x 20' S/S, DP 235 psig	KCS
2		CO_2 Compressor 3rd Stage Discharge KO Drum	6'- 6" ID x 14' S/S, DP 235 psig	CS/SS
2		Propane Refrigeration Surge Drum	14' ID x 42' S/S, DP 300 psig	CS
2		Propane Refrigeration Suction Scrubber	12' ID x 17' S/S, DP 300 psig	LTCS
2		Soda Ash Day Tank	2' ID x 4' S/S, DP atm	CS
4		DCC Water Filter	2730 gpm ea, DP 35 psig	SS
4	Pump-2	Wash Water Pump	1998 gpm ea, DP 29 psi	DI/SS
4	Pump-1	Direct Contact Cooler Water Pump	2730 gpm ea, DP 36 psi	SS/SS
4	P-100	Rich Solvent Pump	5160 gpm ea, DP 92 psi	SS/SS
4	P-102	Lean Solvent Pump	3809 gpm ea, DP 85 psi	SS/SS
4	P-101	Semi-Lean Pump	1663 gpm ea, DP 85 psi	SS/SS
2		Solvent Stripper Reflux Pump	163 gpm ea, DP 75 psi	DI/SS
4		Filter Circ. Pump	258 gpm ea, DP 91 psi	SS/SS
4		LP Condensate Booster Pump	505 gpm ea, DP 237 psi	CI/SS
5		CO_2 Pipeline Pump	293 gpm ea, DP 1815 psi	CS/CS
2		Soda Ash Metering Pump	.45 gpm ea, DP 50 psi	SS
2		CO_2 Compressor (Motor driven)	12,143 hp ea	SS wheels
2		Propane Refrigeration Compressor	10,243 hp ea	LTCS
2		Corrosion Inhibitor Package	Metering, 17 lb/ hr	
4		Solvent Filter Package	258 gpm ea	
2		CO_2 Dryer Package	123 hp ea compressor, cooler, gas fired heater	
2		Crane for Compressor Bldg		
2		Flue Gas Fans and Ducting	2300 hp ea, SS blades	

Table 9-3: Case 3 CO_2 Capture System Equipment List with Data (50% CO_2 Recovery)

No. Required	Tag no.	Description	Size Parameters	Material
incl w/abs		Direct Contact Flue Gas Cooler	25' ID x 34' S/S, DP 2.5 psig/ 0.7 psi vac	CS/SS
2		CO_2 Absorber	25' ID x 92' S/S, DP 2.5 psig/ 0.7 psi vac	CS/SS
2		Solvent Stripper	16' ID x 50' S/S, DP 35 psig/ FV	CS/SS
6	E-106	Solvent Stripper Reboiler	120 MMBTU/HR PHE, 90 psig/ 90 psig	CS/SS
2	E-109	Solvent Stripper Reclaimer	12 MMBTU/HR, DP S/T, 120 psig/ 190 psig	CS/TI
2		Solvent Reclaimer Effluent Cooler	11 MMBTU/HR, DP S/T, 150 psig/ 150 psig	CS/TI
7	E-105	Solvent Stripper CW Condenser	20 MMBTU/HR, DP PHE, 150 psig/ 300 psig	SS/SS
3	E-100	Rich / Lean Solvent Exchanger	158 MMBTU/HR, PHE , 150 psig/ 150 psig	SS316
2	E-101	Rich / Semi-Lean Exchanger	119 MMBTU/HR, PHE, 150 psig/ 150 psig	SS
3	E-102	Lean / Semi-Lean Exchanger	61 MMBTU/HR, PHE, 150 psig/ 150 psig	SS
2	E-108	Absorber Feed Exchanger	66 MMBTU/HR, PHE, 150 psig/ 150 psig	SS
4	E-104	Lean Solvent Exchanger	59 MMBTU/HR, PHE 150 psig/ 150 psig	SS316
1	E-111	Propane Refrigeration De-superheater	27 MMBTU/HR, DP S/T, 300 psig/ 100 psig	CS/CS
1		Propane Refrigeration Condenser	58 MMBTU/HR, DP S/T, 300 psig/ 100 psig	CS/CS
1		Propane Refrigeration Sub-cooler	22 MMBTU/HR, DP S/T, 300 psig/ 2500 psig	CS/LTCS
1		CO_2 Compressor 1st stage Cooler	16 MMBTU/HR, DP 75 psig	SS
1		CO_2 Compressor 2nd stage Cooler	20 MMBTU/HR, DP 125 psig	SS
1		CO_2 Compressor 3rd stage Cooler	17 MMBTU/HR, DP 235 psig	SS
1		CO_2 Condenser	73 MMBTU/HR DP S/T, 235 psig/ 300 psig	CS/TI
2		Solvent Stripper Reflux Drum	7' ID x 22' S/S, DP 35 psig/ FV	304L
1		CO_2 Compressor 2nd Stage Suction Drum	12' ID x 16' S/S, DP 75 psig	CS/SS
1		CO_2 Compressor 3rd Stage Suction Drum	9' ID x 16' S/S, DP 125 psig	CS/SS
1		Liquid CO_2 Surge Drum	7' ID x 22' S/S, DP 235 psig	KCS
1		CO_2 Compressor 3rd Stage Discharge KO Drum	7' ID x 16' S/S, DP 235 psig	CS/SS
1		Propane Refrigeration Surge Drum	16' ID x 47' S/S, DP 300 psig	CS
1		Propane Refrigeration Suction Scrubber	13' ID x 19' S/S, DP 300 psig	LTCS
2		Soda Ash Day Tank	2' ID x 4' S/S, DP atm	CS
4		DCC Water Filter	1931 gpm ea, DP 35 psig	SS
4	Pump-2	Wash Water Pump	1427 gpm ea, DP 29 psi	DI/SS
4	Pump-1	Direct Contact Cooler Water Pump	1931 gpm ea, DP 36 psi	SS/SS
4	P-100	Rich Solvent Pump	3686 gpm ea, DP 92 psi	SS/SS
4	P-102	Lean Solvent Pump	2721 gpm ea, DP 85 psi	SS/SS
4	P-101	Semi-Lean Pump	1189 gpm ea, DP 85 psi	SS/SS
2		Solvent Stripper Reflux Pump	116 gpm ea, DP 75 psi	DI/SS
4		Filter Circ. Pump	184 gpm ea, DP 91 psi	SS/SS
4		LP Condensate Booster Pump	361 gpm ea, DP 237 psi	CI/SS
4		CO_2 Pipeline Pump	262 gpm ea, DP 1815 psi	CS/CS
2		Soda Ash Metering Pump	.45 gpm ea, DP 50 psi	SS
1		CO_2 Compressor (Motor driven)	17,328 hp	SS wheels
1		Propane Refrigeration Compressor	14,618 hp	LTCS
2		Corrosion Inhibitor Package	Metering, 12 lb/ hr	
4		Solvent Filter Package	184 gpm ea	
1		CO_2 Dryer Package	178 hp compressor, cooler, gas fired heater	
1		Crane for Compressor Bldg		
2		Flue Gas Fans and Ducting	1825 hp ea, SS blades	

Table 9-4: Case 4 CO_2 Capture System Equipment List with Data (30% CO_2 Recovery)

No. Required	Tag no.	Description	Size Parameters	Material
Incl w/abs		Direct Contact Flue Gas Cooler	28' ID x 34' S/S, DP 2.5 psig/ 0.7 psi vac	CS/SS
1		CO_2 Absorber	28' ID x 92' S/S, DP 2.5 psig/ 0.7 psi vac	CS/SS
1		Solvent Stripper	20' ID x 50' S/S, DP 35 psig/ FV	CS/SS
4	E-106	Solvent Stripper Reboiler	120 MMBTU/HR PHE, 90 psig/ 90 psig	CS/SS
1	E-109	Solvent Stripper Reclaimer	14 MMBTU/HR, DP S/T, 120 psig/ 190 psig	CS/TI
1		Solvent Reclaimer Effluent Cooler	13 MMBTU/HR, DP S/T, 150 psig/ 150 psig	CS/TI
4	E-105	Solvent Stripper CW Condenser	20 MMBTU/HR, DP PHE, 150 psig/ 300 psig	SS/SS
2	E-100	Rich / Lean Solvent Exchanger	158 MMBTU/HR, PHE , 150 psig/ 150 psig	SS316
1	E-101	Rich / Semi-Lean Exchanger	119 MMBTU/HR, PHE, 150 psig/ 150 psig	SS
1	E-102	Lean / Semi-Lean Exchanger	122 MMBTU/HR, PHE, 150 psig/ 150 psig	SS
1	E-108	Absorber Feed Exchanger	78 MMBTU/HR, PHE, 150 psig/ 150 psig	SS
2	E-104	Lean Solvent Exchanger	59 MMBTU/HR, PHE 150 psig/ 150 psig	SS316
1	E-111	Propane Refrigeration De-superheater	17 MMBTU/HR, DP S/T, 300 psig/ 100 psig	CS/CS
1		Propane Refrigeration Condenser	35 MMBTU/HR, DP S/T, 300 psig/ 100 psig	CS/CS
1		Propane Refrigeration Sub-cooler	13 MMBTU/HR, DP S/T, 300 psig/ 2500 psig	CS/LTCS
1		CO_2 Compressor 1st stage Cooler	10 MMBTU/HR, DP 75 psig	SS
1		CO_2 Compressor 2nd stage Cooler	12 MMBTU/HR, DP 125 psig	SS
1		CO_2 Compressor 3rd stage Cooler	11 MMBTU/HR, DP 235 psig	SS
1		CO_2 Condenser	44 MMBTU/HR DP S/T, 235 psig/ 300 psig	CS/TI
1		Solvent Stripper Reflux Drum	7' ID x 23' S/S, DP 35 psig/ FV	304L
1		CO_2 Compressor 2nd Stage Suction Drum	10' ID x 13' S/S, DP 75 psig	CS/SS
1		CO_2 Compressor 3rd Stage Suction Drum	8' ID x 13' S/S, DP 125 psig	CS/SS
1		Liquid CO_2 Surge Drum	6' ID x 19' S/S, DP 235 psig	KCS
1		CO_2 Compressor 3rd Stage Discharge KO Drum	6' ID x 13' S/S, DP 235 psig	CS/SS
1		Propane Refrigeration Surge Drum	13' ID x 40' S/S, DP 300 psig	CS
1		Propane Refrigeration Suction Scrubber	11' ID x 16' S/S, DP 300 psig	LTCS
1		Soda Ash Day Tank	3' ID x 4' S/S, DP atm	CS
2		DCC Water Filter	2286 gpm ea, DP 35 psig	SS
2	Pump-2	Wash Water Pump	1728 gpm ea, DP 29 psi	DI/SS
2	Pump-1	Direct Contact Cooler Water Pump	2286 gpm ea, DP 36 psi	SS/SS
2	P-100	Rich Solvent Pump	4420 gpm ea, DP 92 psi	SS/SS
2	P-102	Lean Solvent Pump	3220 gpm ea, DP 85 psi	SS/SS
2	P-101	Semi-Lean Pump	1480 gpm ea, DP 85 psi	SS/SS
1		Solvent Stripper Reflux Pump	140 gpm, DP 75 psi	DI/SS
2		Filter Circ. Pump	220 gpm ea, DP 91 psi	SS/SS
2		LP Condensate Booster Pump	434 gpm ea, DP 237 psi	CI/SS
3		CO_2 Pipeline Pump	210 gpm ea, DP 1815 psi	CS/CS
1		Soda Ash Metering Pump	.45 gpm, DP 50 psi	SS
1		CO_2 Compressor (Motor driven)	10,419 hp	SS wheels
1		Propane Refrigeration Compressor	8,788 hp	LTCS
1		Corrosion Inhibitor Package	Metering, 14 lb/ hr	
1		Solvent Filter Package	1870 gpm	
1		CO_2 Dryer Package	108 hp compressor, cooler, gas fired heater	
1		Crane for Compressor Bldg		
1		Flue Gas Fan and Ducting	2190 hp, SS blades	

Table 9-5: Case 5/Concept A CO_2 Capture System Equipment List with Data (96% CO_2 Recovery)

No. Required	Tag no.	Description	Size Parameters	Material
5	DA-2101	Direct Contact Flue Gas Cooler	27' ID x 34' S/S, DP 2.5 psig/ 0.7 psi vac	CS/SS
5	DA-2102	CO_2 Absorber	27' ID x 92' S/S, DP 2.5 psig/ 0.7 psi vac	CS/SS
9	DA-2201	Solvent Stripper	16' ID x 100' S/S, DP 35 psig/ FV	CS/SS
9	EA-2201	Solvent Stripper Reboiler	217 MMBTU/HR, DP S/T, 50 psig/ 60 psig	CS/SS
9	EA-2203	Solvent Stripper Reclaimer	5.6 MMBTU/HR, DP S/T, 120 psig/ 190 psig	CS/TI
9	EA-2204	Solvent Reclaimer Effluent Cooler	5 MMBTU/HR, DP S/T, 125 psig/ 100 psig	CS/TI
9	EA-2206	Solvent Stripper CW Condenser	41.6 MMBTU/HR, DP S/T, 35 psig/ 100 psig	SS/TI
7	EA-2301	CO_2 Compressor 1^{st} Stage Aftercooler	1.9 MMBTU/HR, DP S/T, 75 psig/ 100 psig	SS/TI
7	EA-2302	CO_2 Compressor 2^{nd} Stage Aftercooler	1.3 MMBTU/HR, DP S/T, 125 psig/ 100 psig	SS/TI
7	EA-2303	CO_2 Compressor 3^{rd} Stage Aftercooler	1 MMBTU/HR, DP S/T, 235 psig/ 100 psig	CS/TI
7	EA-2304	CO_2 Condenser	19 MMBTU/HR, DP S/T, 235 psig/ 300 psig	CS/TI
5	EA-2101	Direct Contact Flue Gas Water Cooler	4.8 MMBTU/HR, DP P/U, 50 psig/ 100 psig	TI
9	EA-2205	Rich / Lean Solvent Exchanger	210 MMBTU/HR, DP P/P, 135 psig/ 155 psig	SS316
9	EA-2202	Lean Solvent Cooler	101.8 MMBTU/HR, DP P/U, 135 psig/ 100 psig	TI
7	EA-2401	Propane Refrigeration Condenser	20.45 MMBTU/HR, DP S/T, 300 psig/ 100 psig	CS/CS
7	EA-2402	Propane Refrigeration Sub-cooler	5.9 MMBTU/HR, DP S/T, 300 psig/ 2500 psig	CS/LTCS
7	EC-2301	CO_2 Compressor 1^{st} Stage Air Cooler	2.94 MMBTU/HR, DP 75 psig	SS
7	EC-2302	CO_2 Compressor 2^{nd} Stage Air Cooler	3.1 MMBTU/HR, DP 125 psig	SS
7	EC-2303	CO_2 Compressor 3^{rd} Stage Air Cooler	4.6 MMBTU/HR DP 235 psig	SS
9	EC-2201	Solvent Stripper Bottoms Cooler	80.3 MMBTU/HR DP 135 psig	SS
9	FA-2201	Solvent Stripper Reflux Drum	5' ID x 16' S/S, DP 35 psig/ FV	304L
7	FA-2301	CO_2 Compressor 2^{nd} Stage Suction Drum	7' - 6" ID x 10' S/S, DP 75 psig	CS/SS
7	FA-2303	Liquid CO_2 Surge Drum	4' - 6" ID x 14' S/S, DP 235 psig	KCS
7	FA-2304	CO_2 Compr. 3^{rd} Stage Discharge KO Drum	4' - 6" ID x 10' S/S, DP 235 psig	CS/SS
7	FA-2401	Propane Refrigeration Surge Drum	10' ID x 30' S/S, DP 300 psig	CS
7	FA-2402	Propane Refrigeration Suction Scrubber	8' - 6" ID x 12' S/S, DP 300 psig	LTCS
3	FB-2503	Caustic Day Tank	2' ID x 4' S/S, DP atm	CS
5	FD-2101	DCC Water Filter	205 gpm ea, DP 35 psig	SS
5	GA-2101 A/B	Wash Water Pump	1425 gpm ea, DP 29 psi	DI/SS
5	GA-2102 A/B	Direct Contact Cooler Water Pump	205 gpm ea, DP 35 psi	SS/SS
5	GA-2103 A/B/C/D	Rich Solvent Pump	3450 gpm ea, DP 92 psi	SS/SS
9	GA-2201 A/B/C	Lean Solvent Pump	3000 gpm ea, DP 85 psi	SS/SS
9	GA-2202 A/B	Solvent Stripper Reflux Pump	210 gpm ea, DP 75 psi	DI/SS
9	GA-2203 A/B	Filter Circ. Pump	290 gpm ea, DP 91 psi	SS/SS
9	GA-2204 A/B	LP Condensate Booster Pump	512 gpm ea, DP 237 psi	CI/SS
7	GA-2301 A/B	CO_2 Pipeline Pump	217 gpm Ea, DP 1815 psi	CS/CS
3	GA-2501	Caustic Metering Pump	.45 gpm, DP 50 psi	SS
7	GB-2301	CO_2 Compressor (Motor driven)	4480 hp	SS wheels
7	GB-2401	Propane Refrigeration Compressor	3075 hp	LTCS
1	GB-2500	LP Steam Turbine/Generator	83,365 hp	
9	PA-2551	Corrosion Inhibitor Package	Metering 25 lb/ hr	
9	PA-2251	Solvent Filter Package	140 gpm	
7	PA-2351	CO_2 Dryer Package	4 driers, 200 hp compressor, electric heater, cooler	
1		Crane for Compr. Bldg. Flue Gas Ducting		
1	PA-2551	Cooling Tower	22,000 gpm, includes basin, pumps, chlorine injection	
1	PA-2552	Cooling Tower Blowdown Treatment Package	100 gpm sand filters and de-chlorinator, hypochlorite Storage Tank	

9.3 Appendix III - Economic Sensitivity Studies (Cases 1-5)

This appendix shows the results of a comprehensive economic sensitivity analysis. This analysis was done by varying a number of parameters that effect economic results for each case studied (Total Investment Cost, Capacity Factor, Make-up Power Cost [Levelized], and CO_2 by-product Selling Price [Levelized]). A total of 40 economic evaluation cases are reported in this appendix.

The sensitivity parameters listed above were chosen since the base values used for these parameters are site specific to this project or there may be some uncertainty in the value chosen when looking forward in time. Therefore proper use of these sensitivity results could potentially allow extrapolation to apply results to units other than just Conesville #5. The objective of this sensitivity analysis was to determine the relative impacts of the sensitivity parameters and CO_2 capture level on incremental cost of electricity and CO_2 mitigation cost.

The economic sensitivity results are shown in the tables and graphs, which follow in this appendix. These tables and graphs are grouped according to Case # as indicated in the following list.

- Case 1 – 90% CO_2 Capture
- Case 2 – 70% CO_2 Capture
- Case 3 – 50% CO_2 Capture
- Case 4 – 30% CO_2 Capture
- Case 5 – 96% CO_2 Capture, Updated Concept A of Previous Study

Each group includes one table and two associated graphs, which follow the table. As such, the results from this sensitivity study are summarized in Table 9-6 to Table 9-10 and plotted in Figure 9-10 to Figure 9-14.

9.3.1 Case 1 (90% CO_2 Capture)

Table 9-6: Case 1 (90% CO_2 Capture)

Power Generation									
Net Output (MW)	303.3	303.3	303.3	303.3	303.3	303.3	303.3	303.3	303.3
Capacity Factor (%)	85%	72%	90%	85%	85%	85%	85%	85%	85%
Operating Hours (hrs/yr)	7,446	6,307	7,884	7,446	7,446	7,446	7,446	7,446	7,446
Net Efficiency, HHV (%)	24.5%	24.5%	24.5%	24.5%	24.5%	24.5%	24.5%	24.5%	24.5%
Net Plant Heat Rate, HHV (Btu/kWh)	13,984	13,984	13,984	13,984	13,984	13,984	13,984	13,984	13,984
Coal HHV Input (MMBtu/hr)	4,229	4,229	4,229	4,229	4,229	4,229	4,229	4,229	4,229
Net Generation (MWh/yr)	2,258,498	1,913,081	2,391,351	2,258,498	2,258,498	2,258,498	2,258,498	2,258,498	2,258,498
Costs									
Total Investment Cost ($1000s)	400,094	400,094	400,094	300,070	500,117	400,094	400,094	400,094	400,094
Total Investment Cost ($/kW)	1,319	1,319	1,319	989	1,649	1,319	1,319	1,319	1,319
Fixed O&M Costs ($1000/yr)	2,494	2,494	2,494	2,494	2,494	2,494	2,494	2,494	2,494
Variable O&M Costs ($1000/yr)	17,645	14,947	18,683	17,645	17,645	17,645	17,645	17,645	17,645
Levelized Make-up Power Cost									
Make-up Power Cost (¢/kWh)	6.40	6.40	6.40	6.40	6.40	4.80	8.00	6.40	6.40
Make-up Power Cost ($1000/yr)	62,194	52,682	65,852	62,194	62,194	46,645	77,742	62,194	62,194
CO_2 By-product Revenue									
CO_2 By-product Selling Price ($/ton)	0	0	0	0	0	0	0	25.00	50.00
CO_2 By-product (lb/hr)	779,775	779,775	779,775	779,775	779,775	779,775	779,775	779,775	779,775
CO_2 By-product Revenue ($1000/yr)	0	0	0	0	0	0	0	(72,578)	(145,155)
Feedstock O&M Costs ($1000/yr)	653	553	692	653	653	653	653	653	653
Coal Price ($/MMBtu)	1.80	1.80	1.80	1.80	1.80	1.80	1.80	1.80	1.80
Coal for CO_2 System (MMBtu/hr)	0	0	0	0	0	0	0	0	0
Coal Cost ($1000/yr)	0	0	0	0	0	0	0	0	0
Natural Gas Price ($/MMBtu)	6.75	6.75	6.75	6.75	6.75	6.75	6.75	6.75	6.75
Natural Gas for CO_2 System (MMBtu/hr)	13.00	13.00	13.00	13.00	13.00	13.00	13.00	13.00	13.00
Natural Gas Cost ($1000/yr)	653	553	692	653	653	653	653	653	653
LCOE Contributions									
Capital Component (¢/kWh)	3.10	3.66	2.93	2.33	3.88	3.10	3.10	3.10	3.10
Fixed O&M (¢/kWh)	0.13	0.15	0.12	0.13	0.13	0.13	0.13	0.13	0.13
Variable O&M (¢/kWh)	3.66	3.66	3.66	3.66	3.66	2.97	4.35	0.44	(2.77)
Feedstock O&M (¢/kWh)	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.03
Total, Incremental COE (¢/kWh)	6.92	7.50	6.74	6.14	7.69	6.23	7.61	3.71	0.49
CO_2 Mitigation Cost ($/ton)	81	88	79	72	90	73	89	43	6
CO_2 Mitigation Cost ($/tonne)	89	97	87	79	99	80	98	48	6
CO_2 Capture Cost ($/ton)	54	58	52	48	60	48	59	29	4
CO_2 Capture Cost ($/tonne)	59	64	58	53	66	53	65	32	4

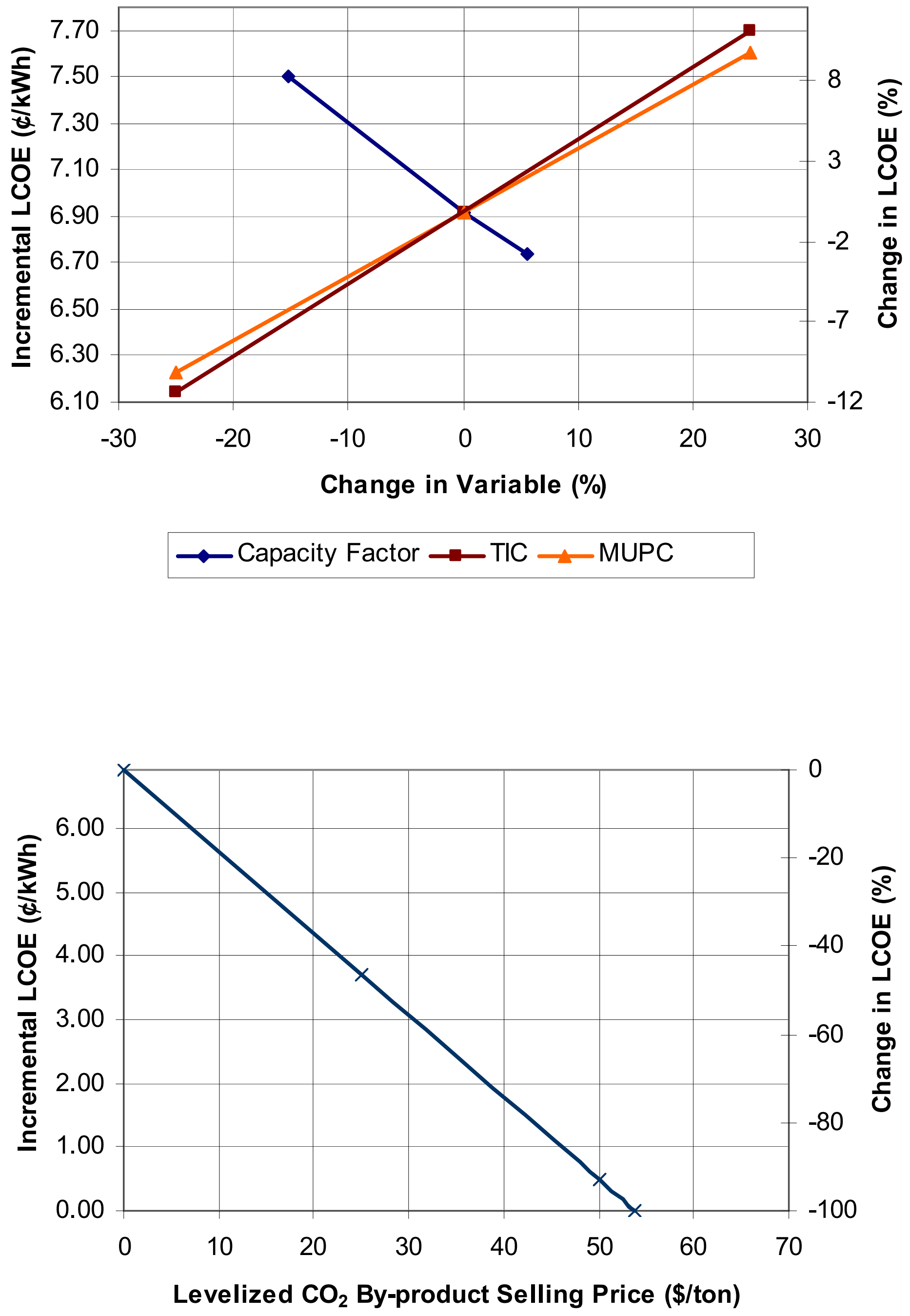
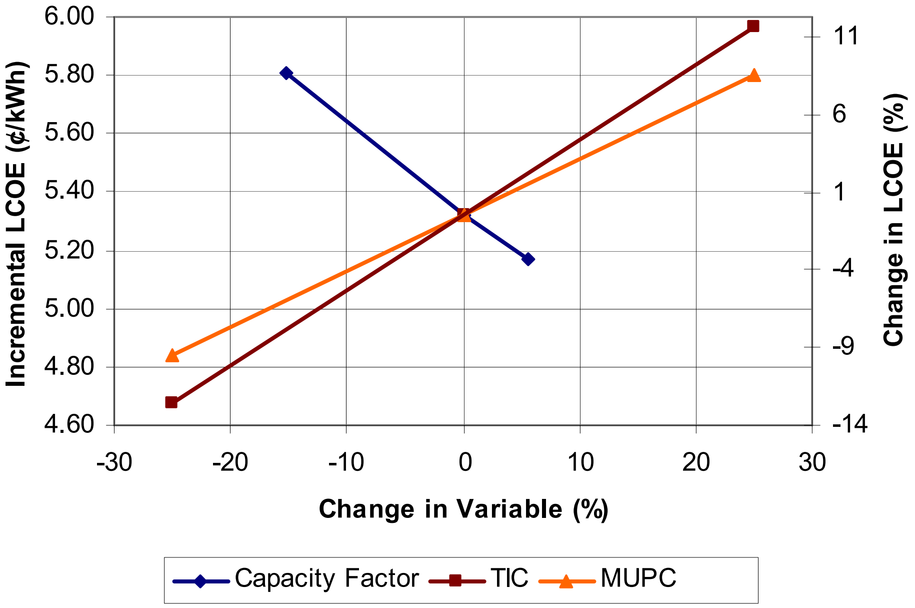


Figure 9-10: Case 1 Sensitivity Studies (90% CO_2 Capture)

9.3.2 Case 2 (70% CO_2 Capture)

Table 9-7: Case 2 (70% CO_2 Capture)

Power Generation									
Net Output (MW)	333.2	333.2	333.2	333.2	333.2	333.2	333.2	333.2	333.2
Capacity Factor (%)	85%	72%	90%	85%	85%	85%	85%	85%	85%
Operating Hours (hrs/yr)	7,446	6,307	7,884	7,446	7,446	7,446	7,446	7,446	7,446
Net Efficiency, HHV (%)	26.9%	26.9%	26.9%	26.9%	26.9%	26.9%	26.9%	26.9%	26.9%
Net Plant Heat Rate, HHV (Btu/kWh)	12,728	12,728	12,728	12,728	12,728	12,728	12,728	12,728	12,728
Coal HHV Input (MMBtu/hr)	4,229	4,229	4,229	4,229	4,229	4,229	4,229	4,229	4,229
Net Generation (MWh/yr)	2,481,342	2,101,843	2,627,304	2,481,342	2,481,342	2,481,342	2,481,342	2,481,342	2,481,342
Costs									
Total Investment Cost ($1000s)	365,070	365,070	365,070	273,802	456,337	365,070	365,070	365,070	365,070
Total Investment Cost ($/kW)	1,095	1,095	1,095	822	1,369	1,095	1,095	1,095	1,095
Fixed O&M Costs ($1000/yr)	2,284	2,284	2,284	2,284	2,284	2,284	2,284	2,284	2,284
Variable O&M Costs ($1000/yr)	14,711	12,461	15,576	14,711	14,711	14,711	14,711	14,711	14,711
Levelized Make-up Power Cost									
Make-up Power Cost (¢/kWh)	6.40	6.40	6.40	6.40	6.40	4.80	8.00	6.40	6.40
Make-up Power Cost ($1000/yr)	47,926	40,597	50,746	47,926	47,926	35,945	59,908	47,926	47,926
CO_2 By-product Revenue									
CO_2 By-product Selling Price ($/ton)	0	0	0	0	0	0	0	25.00	50.00
CO_2 By-product (lb/hr)	607,048	607,048	607,048	607,048	607,048	607,048	607,048	607,048	607,048
CO_2 By-product Revenue ($1000/yr)	0	0	0	0	0	0	0	(56,501)	(113,002)
Feedstock O&M Costs ($1000/yr)	488	413	516	488	488	488	488	488	488
Coal Price ($/MMBtu)	1.80	1.80	1.80	1.80	1.80	1.80	1.80	1.80	1.80
Coal for CO_2 System (MMBtu/hr)	0	0	0	0	0	0	0	0	0
Coal Cost ($1000/yr)	0	0	0	0	0	0	0	0	0
Natural Gas Price ($/MMBtu)	6.75	6.75	6.75	6.75	6.75	6.75	6.75	6.75	6.75
Natural Gas for CO_2 System (MMBtu/hr)	9.70	9.70	9.70	9.70	9.70	9.70	9.70	9.70	9.70
Natural Gas Cost ($1000/yr)	488	413	516	488	488	488	488	488	488
LCOE Contributions									
Capital Component (¢/kWh)	2.57	3.04	2.43	1.93	3.22	2.57	2.57	2.57	2.57
Fixed O&M (¢/kWh)	0.11	0.13	0.10	0.11	0.11	0.11	0.11	0.11	0.11
Variable O&M (¢/kWh)	2.62	2.62	2.62	2.62	2.62	2.13	3.10	0.34	(1.94)
Feedstock O&M (¢/kWh)	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02	0.02
Total, Incremental COE (¢/kWh)	5.32	5.81	5.17	4.68	5.97	4.84	5.80	3.04	0.77
CO_2 Mitigation Cost ($/ton)	88	95	85	77	98	80	95	50	13
CO_2 Mitigation Cost ($/tonne)	96	105	94	85	108	88	105	55	14
CO_2 Capture Cost ($/ton)	58	64	57	51	65	53	64	33	8
CO_2 Capture Cost ($/tonne)	64	70	63	57	72	59	70	37	9



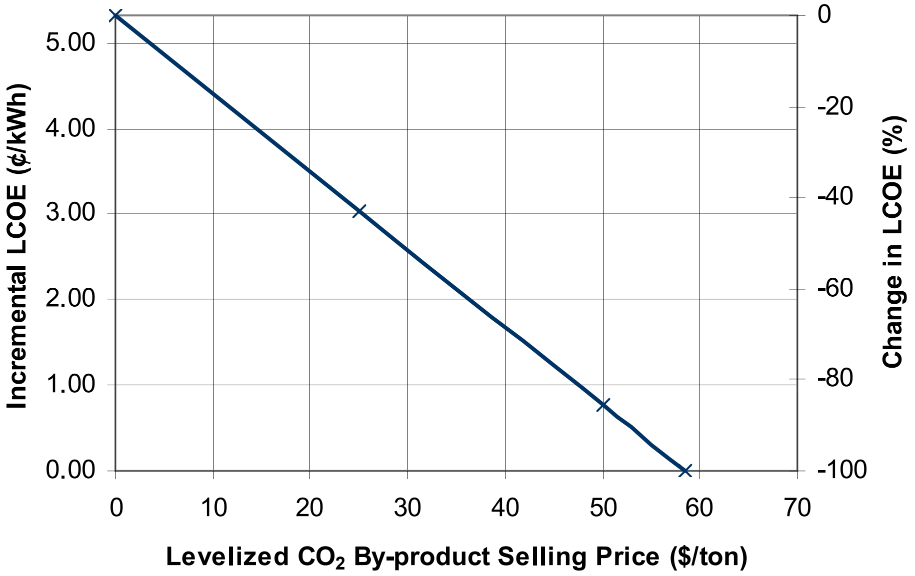


Figure 9-11: Case 2 Sensitivity Studies (70% CO_2 Capture)

9.3.3 Case 3 (50% CO_2 Capture)

Table 9-8: Case 3 (50% CO_2 Capture)

Power Generation									
Net Output (MW)	362.9	362.9	362.9	362.9	362.9	362.9	362.9	362.9	362.9
Capacity Factor (%)	85%	72%	90%	85%	85%	85%	85%	85%	85%
Operating Hours (hrs/yr)	7,446	6,307	7,884	7,446	7,446	7,446	7,446	7,446	7,446
Net Efficiency, HHV (%)	29.3%	29.3%	29.3%	29.3%	29.3%	29.3%	29.3%	29.3%	29.3%
Net Plant Heat Rate, HHV (Btu/kWh)	11,686	11,686	11,686	11,686	11,686	11,686	11,686	11,686	11,686
Coal HHV Input (MMBtu/hr)	4,229	4,229	4,229	4,229	4,229	4,229	4,229	4,229	4,229
Net Generation (MWh/yr)	2,702,488	2,289,167	2,861,458	2,702,488	2,702,488	2,702,488	2,702,488	2,702,488	2,702,488
Costs									
Total Investment Cost ($1000s)	280,655	280,655	280,655	210,491	350,818	280,655	280,655	280,655	280,655
Total Investment Cost ($/kW)	773	773	773	580	967	773	773	773	773
Fixed O&M Costs ($1000/yr)	2,079	2,079	2,079	2,079	2,079	2,079	2,079	2,079	2,079
Variable O&M Costs ($1000/yr)	10,876	9,212	11,516	10,876	10,876	10,876	10,876	10,876	10,876
Levelized Make-up Power Cost									
Make-up Power Cost (¢/kWh)	6.40	6.40	6.40	6.40	6.40	4.80	8.00	6.40	6.40
Make-up Power Cost ($1000/yr)	33,768	28,603	35,754	33,768	33,768	25,326	42,210	33,768	33,768
CO_2 By-product Revenue									
CO_2 By-product Selling Price ($/ton)	0	0	0	0	0	0	0	25.00	50.00
CO_2 By-product (lb/hr)	433,606	433,606	433,606	433,606	433,606	433,606	433,606	433,606	433,606
CO_2 By-product Revenue ($1000/yr)	0	0	0	0	0	0	0	(40,358)	(80,716)
Feedstock O&M Costs ($1000/yr)	337	285	357	337	337	337	337	337	337
Coal Price ($/MMBtu)	1.80	1.80	1.80	1.80	1.80	1.80	1.80	1.80	1.80
Coal for CO_2 System (MMBtu/hr)	0	0	0	0	0	0	0	0	0
Coal Cost ($1000/yr)	0	0	0	0	0	0	0	0	0
Natural Gas Price ($/MMBtu)	6.75	6.75	6.75	6.75	6.75	6.75	6.75	6.75	6.75
Natural Gas for CO_2 System (MMBtu/hr)	6.70	6.70	6.70	6.70	6.70	6.70	6.70	6.70	6.70
Natural Gas Cost ($1000/yr)	337	285	357	337	337	337	337	337	337
LCOE Contributions									
Capital Component (¢/kWh)	1.82	2.15	1.72	1.36	2.27	1.82	1.82	1.82	1.82
Fixed O&M (¢/kWh)	0.09	0.11	0.08	0.09	0.09	0.09	0.09	0.09	0.09
Variable O&M (¢/kWh)	1.72	1.72	1.72	1.72	1.72	1.40	2.03	0.22	(1.27)
Feedstock O&M (¢/kWh)	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01
Total, Incremental COE (¢/kWh)	3.64	3.98	3.53	3.18	4.09	3.32	3.95	2.14	0.65
CO_2 Mitigation Cost ($/ton)	91	99	88	79	102	83	98	53	16
CO_2 Mitigation Cost ($/tonne)	100	109	97	87	112	91	108	59	18
CO_2 Capture Cost ($/ton)	61	67	59	53	68	56	66	36	11
CO_2 Capture Cost ($/tonne)	67	73	65	59	75	61	73	40	12

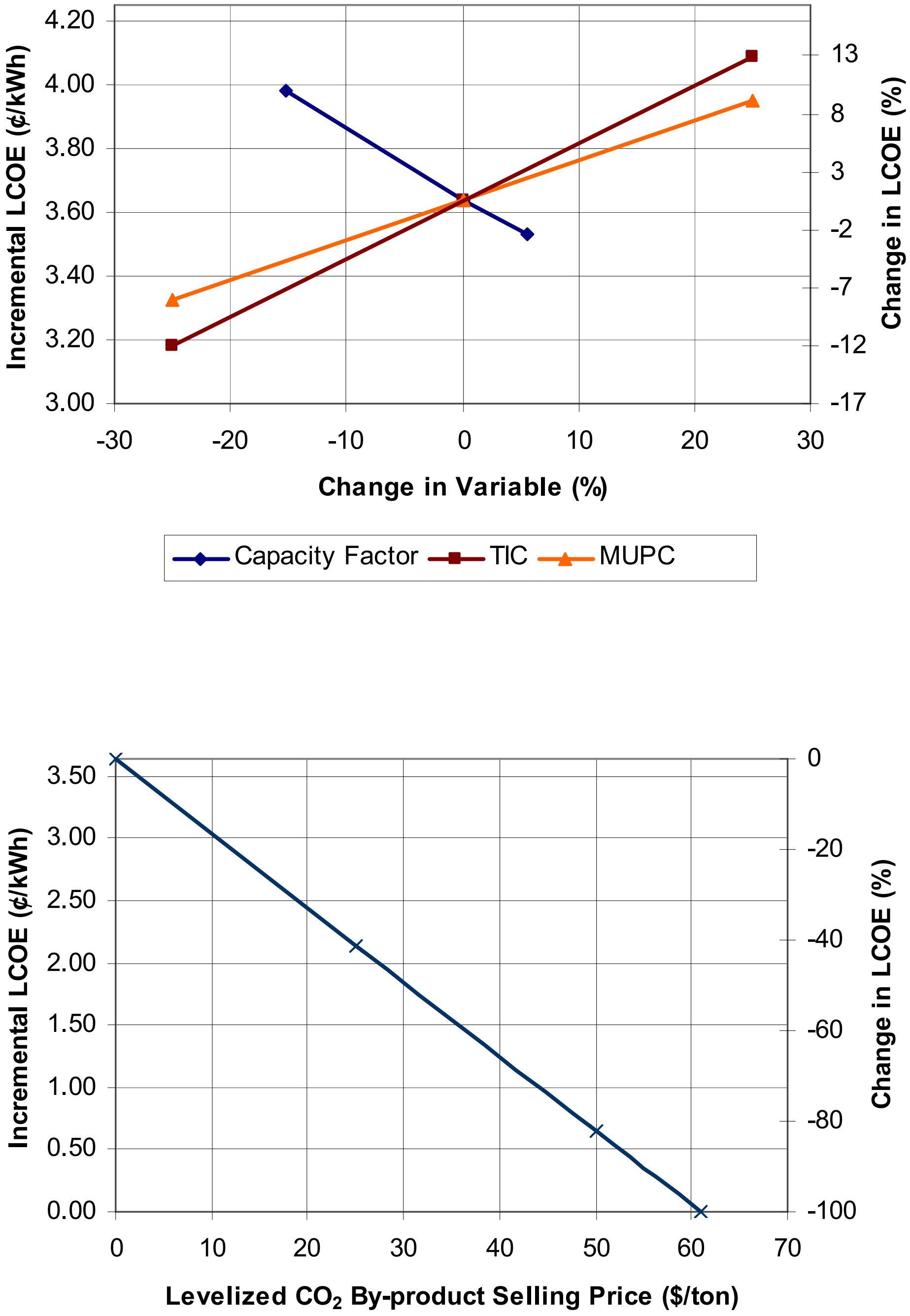
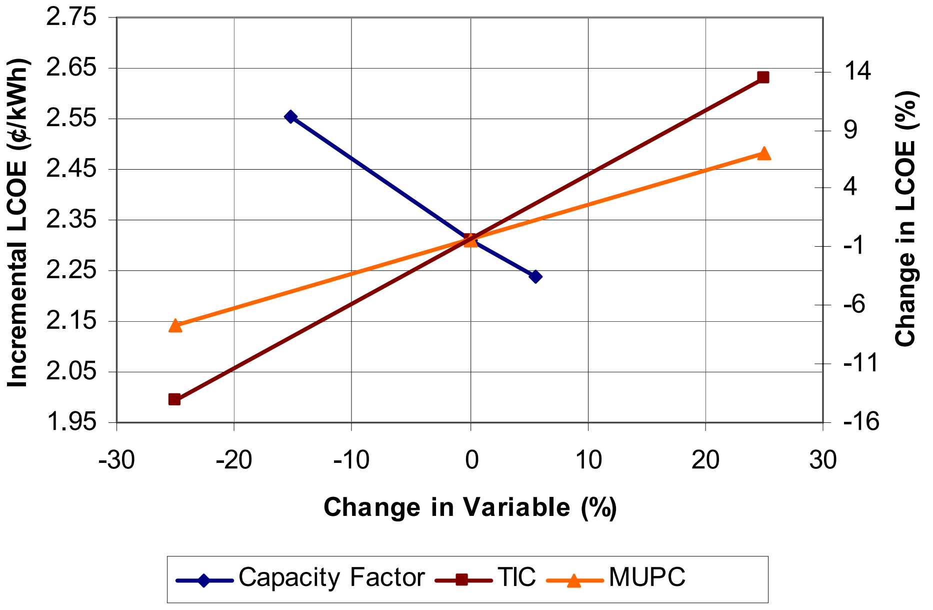


Figure 9-12: Case 2 Sensitivity Studies (50% CO_2 Capture)

9.3.4 Case 4 (30% CO_2 Capture)

Table 9-9: Case 4 (30% CO_2 Capture)

Power Generation									
Net Output (MW)	392.1	392.1	392.1	392.1	392.1	392.1	392.1	392.1	392.1
Capacity Factor (%)	85%	72%	90%	85%	85%	85%	85%	85%	85%
Operating Hours (hrs/yr)	7,446	6,307	7,884	7,446	7,446	7,446	7,446	7,446	7,446
Net Efficiency, HHV (%)	31.7%	31.7%	31.7%	31.7%	31.7%	31.7%	31.7%	31.7%	31.7%
Net Plant Heat Rate, HHV (Btu/kWh)	10,818	10,818	10,818	10,818	10,818	10,818	10,818	10,818	10,818
Coal HHV Input (MMBtu/hr)	4,229	4,229	4,229	4,229	4,229	4,229	4,229	4,229	4,229
Net Generation (MWh/yr)	2,919,331	2,472,845	3,091,056	2,919,331	2,919,331	2,919,331	2,919,331	2,919,331	2,919,331
Costs									
Total Investment Cost ($1000s)	211,835	211,835	211,835	158,876	264,794	211,835	211,835	211,835	211,835
Total Investment Cost ($/kW)	540	540	540	405	675	540	540	540	540
Fixed O&M Costs ($1000/yr)	1,869	1,869	1,869	1,869	1,869	1,869	1,869	1,869	1,869
Variable O&M Costs ($1000/yr)	7,019	5,945	7,432	7,019	7,019	7,019	7,019	7,019	7,019
Levelized Make-up Power Cost									
Make-up Power Cost (¢/kWh)	6.40	6.40	6.40	6.40	6.40	4.80	8.00	6.40	6.40
Make-up Power Cost ($1000/yr)	19,885	16,843	21,054	19,885	19,885	14,913	24,856	19,885	19,885
CO_2 By-product Revenue									
CO_2 By-product Selling Price ($/ton)	0	0	0	0	0	0	0	25.00	50.00
CO_2 By-product (lb/hr)	260,163	260,163	260,163	260,163	260,163	260,163	260,163	260,163	260,163
CO_2 By-product Revenue ($1000/yr)	0	0	0	0	0	0	0	(24,215)	(48,429)
Feedstock O&M Costs ($1000/yr)	211	179	224	211	211	211	211	211	211
Coal Price ($/MMBtu)	1.80	1.80	1.80	1.80	1.80	1.80	1.80	1.80	1.80
Coal for CO_2 System (MMBtu/hr)	0	0	0	0	0	0	0	0	0
Coal Cost ($1000/yr)	0	0	0	0	0	0	0	0	0
Natural Gas Price ($/MMBtu)	6.75	6.75	6.75	6.75	6.75	6.75	6.75	6.75	6.75
Natural Gas for CO_2 System (MMBtu/hr)	4.20	4.20	4.20	4.20	4.20	4.20	4.20	4.20	4.20
Natural Gas Cost ($1000/yr)	211	179	224	211	211	211	211	211	211
LCOE Contributions									
Capital Component (¢/kWh)	1.27	1.50	1.20	0.95	1.59	1.27	1.27	1.27	1.27
Fixed O&M (¢/kWh)	0.07	0.09	0.07	0.07	0.07	0.07	0.07	0.07	0.07
Variable O&M (¢/kWh)	0.96	0.96	0.96	0.96	0.96	0.79	1.13	0.13	(0.70)
Feedstock O&M (¢/kWh)	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01
Total, Incremental COE (¢/kWh)	2.31	2.55	2.24	1.99	2.63	2.14	2.48	1.48	0.65
CO_2 Mitigation Cost ($/ton)	103	114	99	89	117	95	110	66	29
CO_2 Mitigation Cost ($/tonne)	113	125	110	98	129	105	122	73	32
CO_2 Capture Cost ($/ton)	70	77	67	60	79	65	75	45	20
CO_2 Capture Cost ($/tonne)	77	85	74	66	87	71	82	49	22



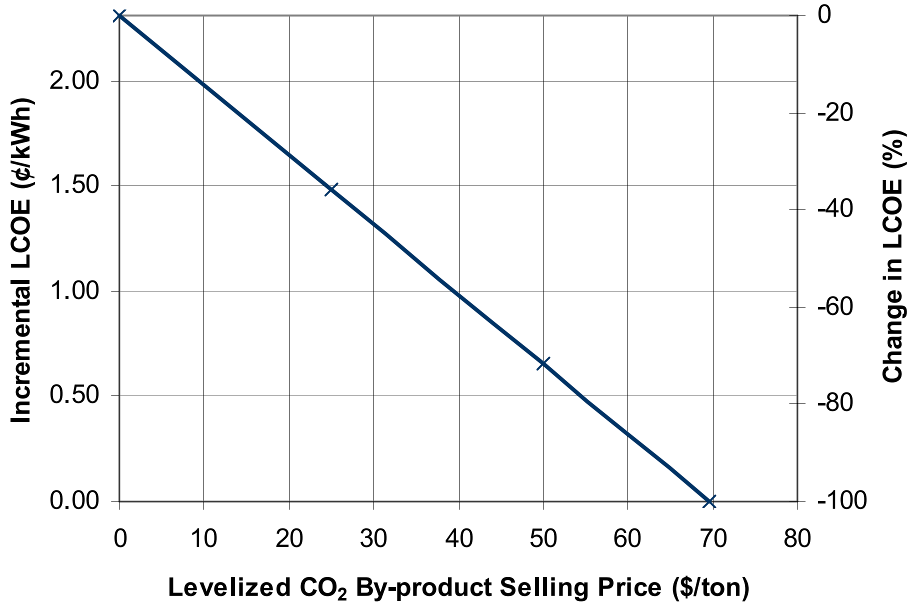
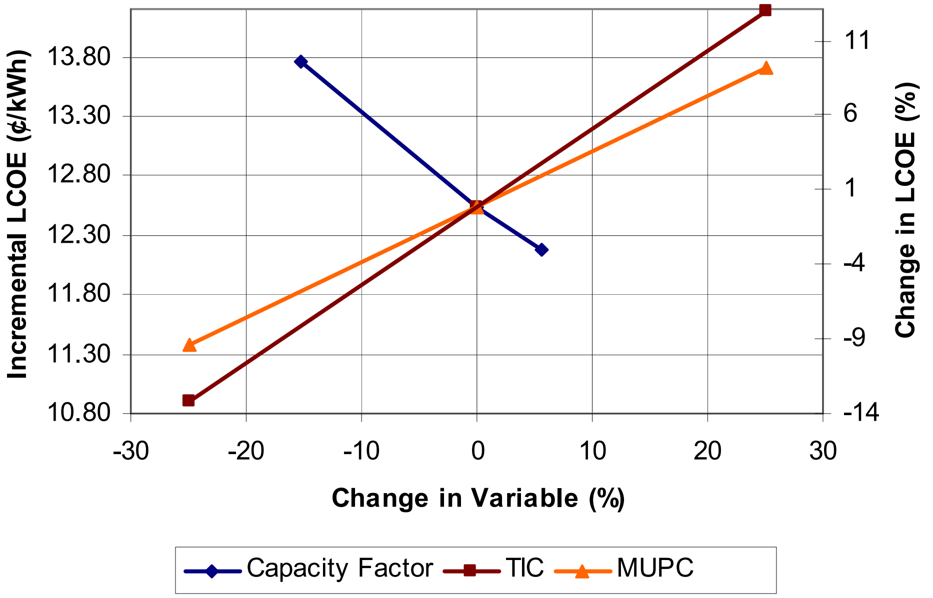


Figure 9-13: Case 4 Sensitivity Studies (30% CO_2 Capture)

9.3.5 Case 5 (96% CO_2 Capture)

Table 9-10: Case 5 (96% CO_2 Capture)

Power Generation									
Net Output (MW)	251.6	251.6	251.6	251.6	251.6	251.6	251.6	251.6	251.6
Capacity Factor (%)	85%	72%	90%	85%	85%	85%	85%	85%	85%
Operating Hours (hrs/yr)	7,446	6,307	7,884	7,446	7,446	7,446	7,446	7,446	7,446
Net Efficiency, HHV (%)	20.3%	20.3%	20.3%	20.3%	20.3%	20.3%	20.3%	20.3%	20.3%
Net Plant Heat Rate, HHV (Btu/kWh)	16,856	16,856	16,856	16,856	16,856	16,856	16,856	16,856	16,856
Total Fuel Heat Input at MCR (MMBtu/hr)	4,242	4,242	4,242	4,242	4,242	4,242	4,242	4,242	4,242
Coal HHV Input (MMBtu/hr)	4,229	4,229	4,229	4,229	4,229	4,229	4,229	4,229	4,229
Net Generation (MWh/yr)	1,873,667	1,587,106	1,983,882	1,873,667	1,873,667	1,873,667	1,873,667	1,873,667	1,873,667
Costs									
Total Investment Cost ($1000s)	701,057	701,057	701,057	525,793	876,322	701,057	701,057	701,057	701,057
Total Investment Cost ($/kW)	2,786	2,786	2,786	2,090	3,483	2,786	2,786	2,786	2,786
Fixed O&M Costs ($1000/yr)	2,488	2,488	2,488	2,488	2,488	2,488	2,488	2,488	2,488
Variable O&M Costs ($1000/yr)	18,640	15,789	19,737	18,640	18,640	18,640	18,640	18,640	18,640
Levelized Make-up Power Cost									
Make-up Power Cost (¢/kWh)	6.40	6.40	6.40	6.40	6.40	4.80	8.00	6.40	6.40
Make-up Power Cost ($1000/yr)	86,832	73,552	91,940	86,832	86,832	65,124	108,540	86,832	86,832
CO_2 By-product Revenue									
CO_2 By-product Selling Price ($/ton)	0	0	0	0	0	0	0	25.00	50.00
CO_2 By-product (lb/hr)	835,053	835,053	835,053	835,053	835,053	835,053	835,053	835,053	835,053
CO_2 By-product Revenue ($1000/yr)	0	0	0	0	0	0	0	(77,723)	(155,445)
Feedstock O&M Costs ($1000/yr)	890	754	942	890	890	890	890	890	890
Coal Price ($/MMBtu)	1.80	1.80	1.80	1.80	1.80	1.80	1.80	1.80	1.80
Coal for CO_2 System (MMBtu/hr)	0	0	0	0	0	0	0	0	0
Coal Cost ($1000/yr)	0	0	0	0	0	0	0	0	0
Natural Gas Price ($/MMBtu)	6.75	6.75	6.75	6.75	6.75	6.75	6.75	6.75	6.75
Natural Gas for CO_2 System (MMBtu/hr)	17.70	17.70	17.70	17.70	17.70	17.70	17.70	17.70	17.70
Natural Gas Cost ($1000/yr)	890	754	942	890	890	890	890	890	890
LCOE Contributions									
Capital Component (¢/kWh)	6.55	7.73	6.18	4.91	8.18	6.55	6.55	6.55	6.55
Fixed O&M (¢/kWh)	0.15	0.18	0.15	0.15	0.15	0.15	0.15	0.15	0.15
Variable O&M (¢/kWh)	5.79	5.79	5.79	5.79	5.79	4.63	6.94	1.64	(2.51)
Feedstock O&M (¢/kWh)	0.06	0.06	0.06	0.06	0.06	0.06	0.06	0.06	0.06
Total, Incremental COE (¢/kWh)	12.54	13.75	12.17	10.91	14.18	11.38	13.70	8.39	4.25
CO_2 Mitigation Cost ($/ton)	134	147	130	117	152	122	147	90	46
CO_2 Mitigation Cost ($/tonne)	148	163	144	129	168	135	162	99	50
CO_2 Capture Cost ($/ton)	76	83	73	66	85	69	83	51	26
CO_2 Capture Cost ($/tonne)	83	91	81	72	94	76	91	56	28



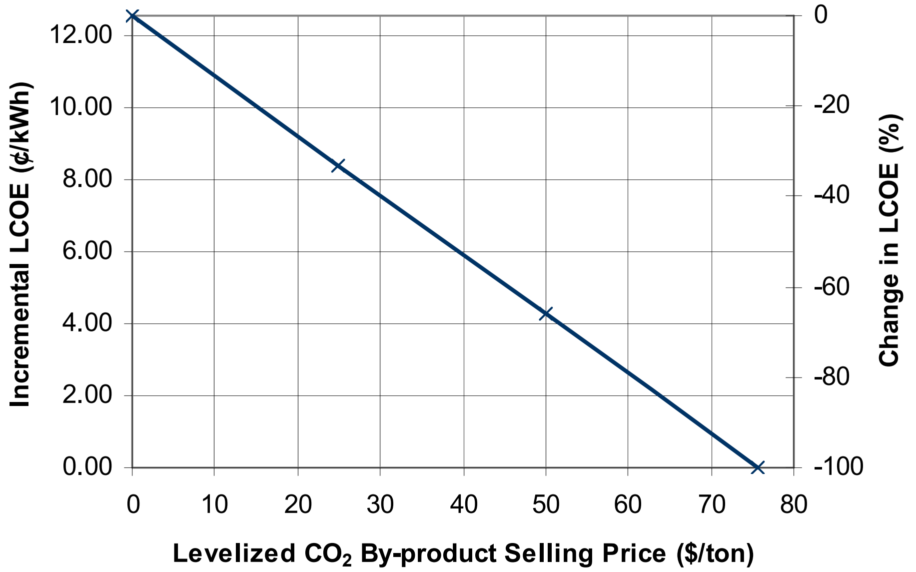


Figure 9-14: Case 5 Sensitivity Studies (96% CO_2 Capture)

9.4 Appendix IV – Let Down Turbine Technical Information (Cases 1 and 4)

This appendix provides technical information regarding the let down turbines used for Case 1 (90% CO_2 capture) and Case 4 (30% CO_2 capture). Three attachments are provided as listed below:

- Attachment A: Steam Turbine and Auxiliaries General Technical Information (applicable to both the 90% and 30% CO_2 recovery let down turbines)
- Attachment B: Information specific to the Case 1 let down turbine (90% CO_2 capture)
- Attachment C: Information specific to the Case 4 let down turbine (30% CO_2 capture turbine)

Attachment A:

Steam Turbine and Auxiliaries General Technical Information (applicable to both the 90% and 30% CO_2 recovery let down turbines)

1. GENERAL DESIGN INFORMATION

1.1 TURBINE

The turbine is a multistage straight backpressure single line type with the shaft aligned horizontally. Its casing consists of a fabricated steel structure made from welded steel plates. Steam is admitted through two inlet openings located on the top and the bottom of the inlet box, respectively. The upper part of this casing is welded to the duct (out of scope of supply).

The turbine rotor is fabricated of high chromium steel with the coupling disc at the generator side being an integral part of it.

1.2 TURBINE CHOKE VALVES

IP steam is admitted through one quick-closing choke valve and two control choke valves, located at the side of the turbine.

The quick-closing choke valves are arranged in front of the control choke valve.

1.3 BEARINGS

Turbine rotor is supported with two hydrodynamic bearings. The bearings are supplied with high pressure jacking oil at start up and in case of low speed rotor rotations.

1.4 TURNING GEAR

The turbine front pedestal will be equipped with a motor driven turning gear with automatic operation control system.

The turning gear is capable of starting the unit from standstill and rotating the turbine-generator shaft line continuously at recommended turning speed with normal lube oil pressure.

1.5 TECHNICAL DATA OF THE TURBINE

Please refer to the specific turbine under consideration (see separate attachment).

2. GENERATOR

The generator is an air-cooled generator running at 3,600 rpm.

For more specific information on the generator under consideration, please refer to the generator description in the separate attachment.

AUXILIARY SYSTEMS

3.1 TURBINE SUPERVISORY SYSTEM

The turbine supervisory system ensures supervision of turbine/generator unit shaft-line critical operating parameters (e.g.:turbine and generator journal bearings temperatures and vibration levels and turbine thrust bearing temperature and wearing).

The supervisory system is connected with the turbine safety system and may generate alarm and tripping signals through adjustable monitoring consoles.

3.2 TURBINE SAFETY AND PROTECTION SYSTEM

The safety and protection system is able to stop the steam turbine by a quick, automatic closing of choke valves.

A turbine trip may be initiated either automatically or by action of an operator under instruction. In faulty conditions of a monitored parameter, a threshold detector emits an alarm and, in the worst case, may even promote an automatic trip.

3.3 STEAM TURBINE GOVERNING SYSTEM

The Steam Turbine Governing System governs the position of the control choke valve. This control system ensures the following functions:

- Control of the turbine generator speed (frequency in island operation) when the generator is not coupled to the grid
- Control of the turbo-generator load when the generator is coupled to the grid

In normal operation the system operates with a sliding pressure at inlet at the maximum opening of the turbine with a load limitation.

3.4 GLAND STEAM SYSTEM

Correct operation of the turbine requires clearances between fixed and moving parts, through which steam tends to leak. The gland steam system ensures that no steam escapes from valves and shaft glands into the turbine room.

3.5 DRAIN SYSTEM

The drains have the following purposes:

- To eliminate the condensates in order to avoid damages to the machine,
- To ensure the thermal conditioning of the turbine by steam circulation from glands when the control valves are closed or just opened.

3.6 OIL SYSTEM

One complete combined lube and control oil system is feeding two separate circuits.

The function of this system is to ensure, on one side, the lubrication and cooling of journal bearings, and the thrust bearing, for the whole set (turbine, generator), and on the other side, the control oil of the turbine. It consists mainly of a packaged oil tank. Electrically driven positive displacement (main and auxiliary) and centrifugal (emergency) pumps are vertically submerged in this oil tank.

Two full duty oil coolers are arranged in parallel on oil and cooling water circuits with a changeover oil valve to change the cooler on duty without interruption of the oil flow to the bearings. An emergency standby pump delivers lube oil without passing through the coolers and filters.

The control and safety and protection systems use the common lube and control oil for actuation of valves.

4. SCOPE OF SUPPLY AND LIMITS OF DELIVERY

4.1 SCOPE OF DELIVERY

Table 9-11: Let Down Turbine Scope of Delivery

Item No.	Description	Quantity per one unit	Remarks
1.	Complete turbine: A) turbine casing B) bladed rotor C) blade carrier with fixed blades D) end gland seals	1 set	Including insulation
2.	Turbine steam admission system consists of quick closing and control choke valves	1 set	Including insulation
3.	Complete turbine pedestals with bearings and elements necessary for the shaft line adjustment and pedestal survey	1 set	
4.	Turbine-Generator coupling	1 set	
5.	Complete electrical turning gear with clutch and hand turning facility	1 set	
6.	Handling devices for steam turbine components	1 set	
7.	Complete gland steam system including: A) pressure reducing valve, B) piping and valves, C) gland steam condenser	1 set	
8.	Complete oil systems including: A) pumps (main, auxiliary, emergency), B) oil tank, C) coolers (2 x 100%), D) oil filter (duplex) E) piping and valves, F) oil mist and separator, G) oil tank drain piping (ending with isolating valves)	1 set	
9.	Complete air cooled generator with excitation system and AVR	1 set	

10.	Handling devices for generator components	1 set	
11.	T/G control and protection system: A) system cubicle, B) hardware, C) software, D) speed probes	1 set	
12.	T/G supervisory equipment (TSE): A) instrument rack incl. power supply B) probes and sensors with connection to local junction boxes, transmitters, etc., C) proximitors and monitors, D) software	1 set	
13.	Instrumentation and cables for the T/G and auxiliaries	1 set	Cabling up to local junction boxes
14.	Special tools	1 set	
15.	Spare parts for start-up	1 set	
16.	Mandatory spare parts	1 set	
17.	Documentation: A) quality, B) assembly, C) manuals	1 set	English versions only

4.2 LIMITS OF DELIVERY

The scope of supply as mentioned in Table 9-11 above is limited to the following boundaries:

Steam:	Inlet weld connection on IP steam admission valve Outlet weld connection on LP casing (upper exhaust)
Cooling water	Inlet/outlet of cooling water flange connections at lube oil coolers.
Condensate/Feedwater:	Inlet weld connection at LP turbine hood spray water stop valve.
Gland system:	Inlet connection at gland steam supply control valve. Outlet flange at gland steam condenser exhaust ventilator fan. Feedwater inlet/outlet flange connections at gland steam condenser. Condensate outlet flange at gland steam condenser.
Lube oil system:	Outlet flange at vapour ventilator fan of oil tank Supply and drain connections on lube oil tank.
Elec. equipment:	Terminals at motor terminal boxes. Terminals at plant mounted local junction boxes.
I&C:	Terminals at control cubicles Terminals at local junction boxes
Generator:	Output terminals of the generator and brush gear, Output terminals of the generator and brush gear measuring boxes, Output terminals of the noise hood measuring boxes, Output and input terminals in the excitation system cubicle, Output and input flanges on the coolers

Attachment B:

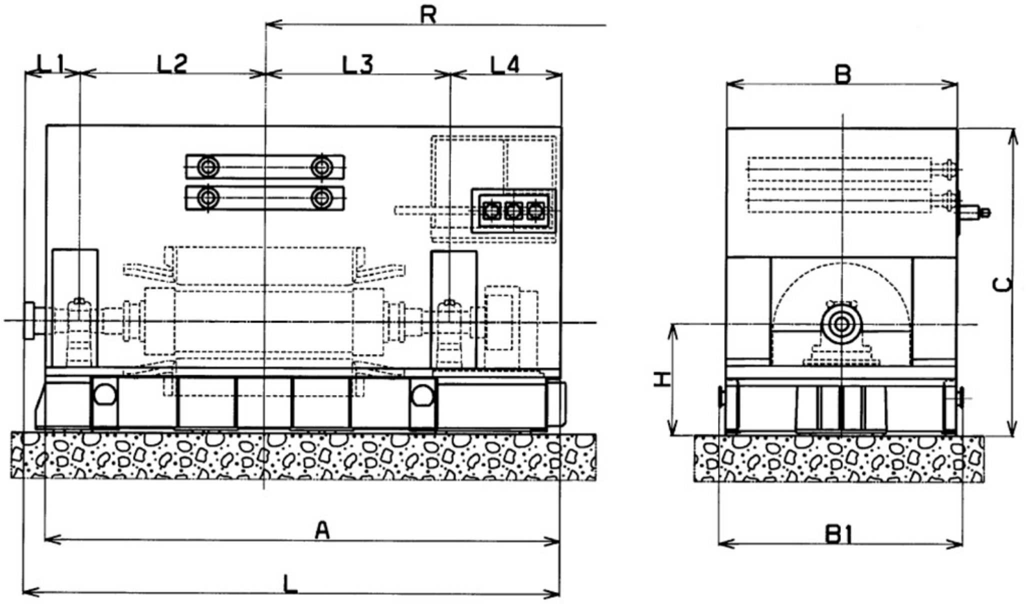
Steam Turbine and Auxiliaries for Case 1 Let Down Turbine (90% CO_2 removal)

1. TECHNICAL DATA OF THE TURBINE

Parameter	Unit	Value
Number of casings	-	1
Nominal speed	rpm	3,600
Plant cycle	-	single flash
Inlet pressure psia	200	
Temperature	°F	711
Exhaust pressure	psia	47
Gross Electric Power Output (at generator terminals)	kW	48,030

2. GENERATOR

The generator is an air-cooled generator running at 3600 rpm. It is designed for a nominal active power of 50.00 MW at a power factor of 0.9. A general arrangement drawing is shown in Figure 9-15.



MAIN FEATURES	APPROXIMATE DIMENSIONS		mm.	in.
Water / air cooled	A	:	7 252	285.5
	B	:	3 150	124.0
Brushless exciter	B1	:	3 330	131.1
	C	:	4 200	165.4
Soundproofed housing	H	:	1 500	59.1
	L	:	7 352	289.5
Protection degree IP 54	L1	:	510	20.1
	L2	:	2 530	99.6
MV equipment located	L3	:	2 530	99.6
inside the generator	L4	:	1 782	70.2
	R		11 300	444.9

APPROXIMATE WEIGHTS		tons	lbs.
Stator	:	53.0	116 800
Rotor + Exciter armature	:	18.0	39 700
Bearings	:	1.6	3 500
Base frame	:	12.1	26 700
Exciter field	:	0.7	1 500
Housing	:	12.0	26 400
Miscellaneous	:	5.0	11 000
TOTAL	:	102.4	225 600

APPROXIMATE INERTIA

MR²	Kg.m²	Lb.ft²
Generator	: 1 640	38 900

Figure 9-15: Typical General Outline Arrangement for LDT Generator for Case 1 (90% Recovery)

3. TURBINE GENERATOR ARRANGEMENT

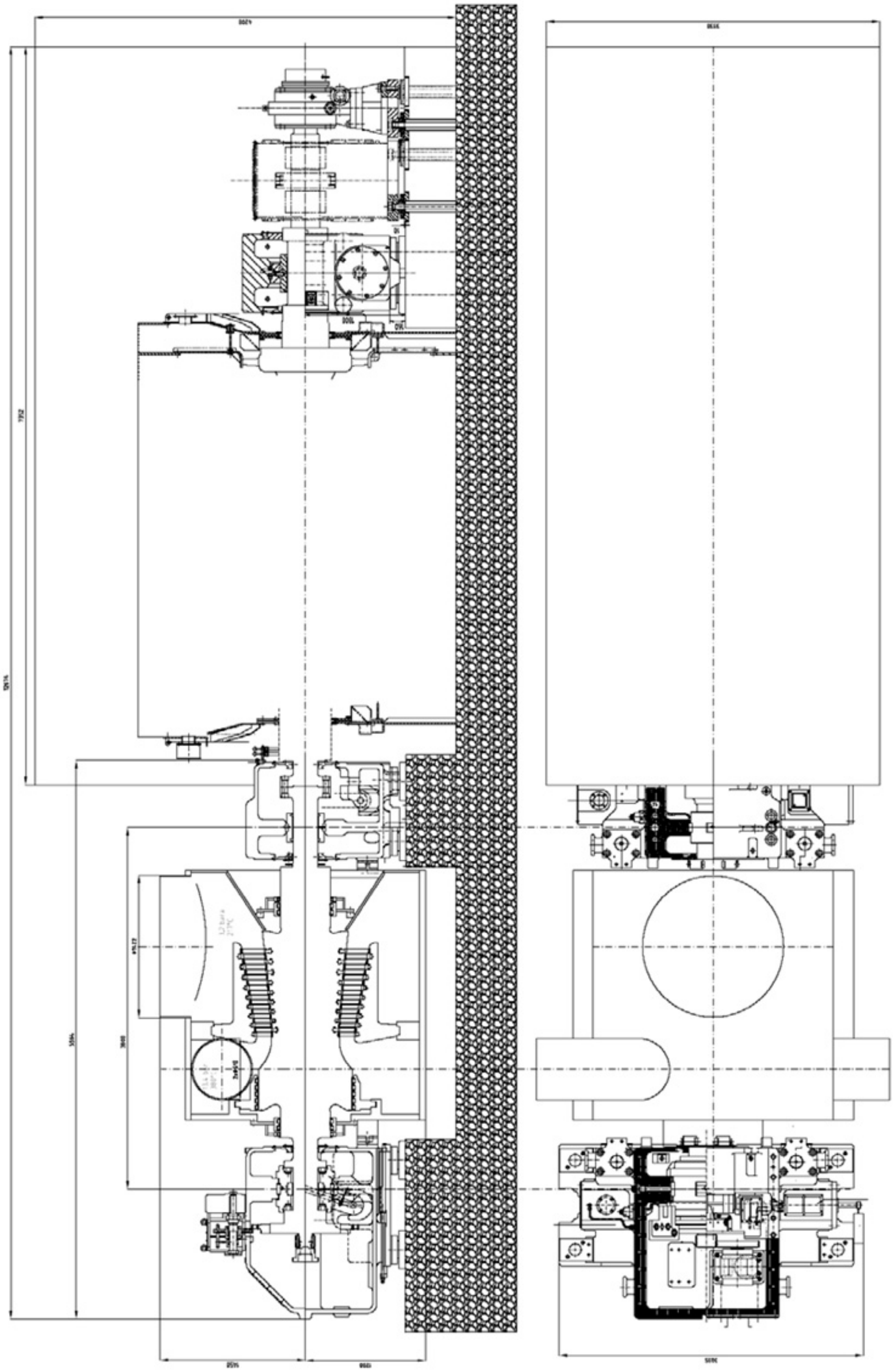


Figure 9-16: Turbine Generator General Arrangement (Case 1: 90% removal)

Attachment C:

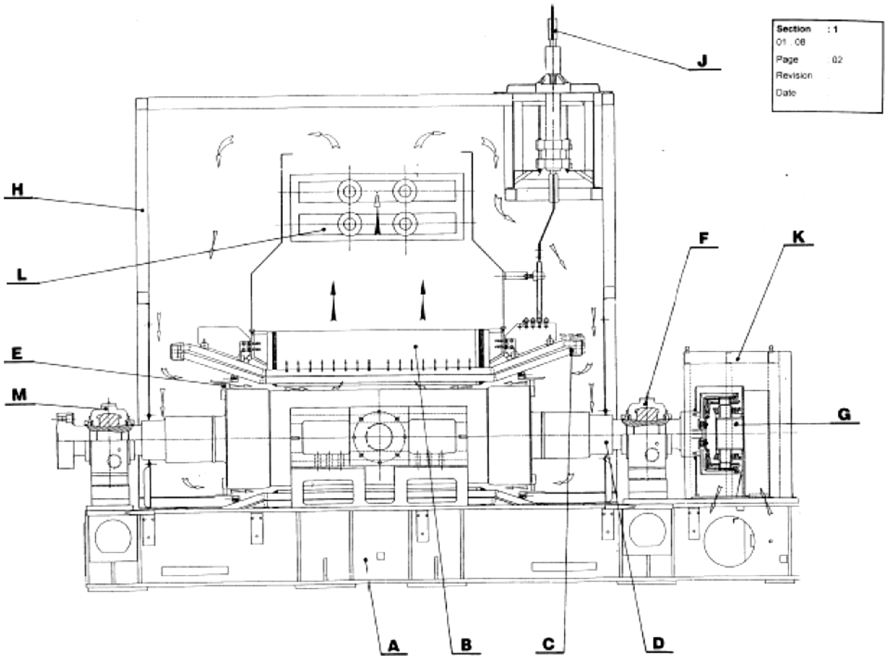
Steam Turbine and Auxiliaries for Case 4 Let Down Turbine (30% CO_2 removal)

1. TECHNICAL DATA OF THE TURBINE

Parameter	Unit	Value
Number of casings	-	1
Nominal speed Plant cycle -	rpm single flash	3600
Inlet pressure psia	195	
Temperature °F	711	
Exhaust pressure Gross Electric Power Output	psia	47
(at generator terminals)	kW	15054

2. GENERATOR

The generator is an air-cooled generator running at 3,600 rpm. It is designed for a nominal active power of 15.00 MW at a power factor of 0.9. A general arrangement drawing is shown in Figure 9-17.



A	Base
B	Magnetic core
C	Stator winding
D	Rotor
E	Fan
F	Bearing (N.E.D.)
G	Exciter
H	Noise hood
J	High voltage terminal
K	Exciter cover
L	Coolers
M	Bearing (N.E.D.)

Figure 9- 17: Typical General Outline Arrangement for LDT Generator for Case 4 (30% Recovery)

Main Features	**Approximate Weights**			
			Tons	Lbm
Water /air cooled	Stator	:	26.4	58 202
	Support base		9.4	20 723
Brushless exciter	Rotor + Exciter Rotor	:	11.5	25 353
	Exciter		0.4	882
Soundproof housing	Bearings	:	1.3	2 866
	Housing		6.0	15 212
Protection degree IP 55	Coolers	:	1.6	3 527
	Miscellaneous		2.6	5 732
MV equipment located inside the generator	TOTAL	:	60.1	132 498

3. TURBINE GENERATOR ARRANGEMENT

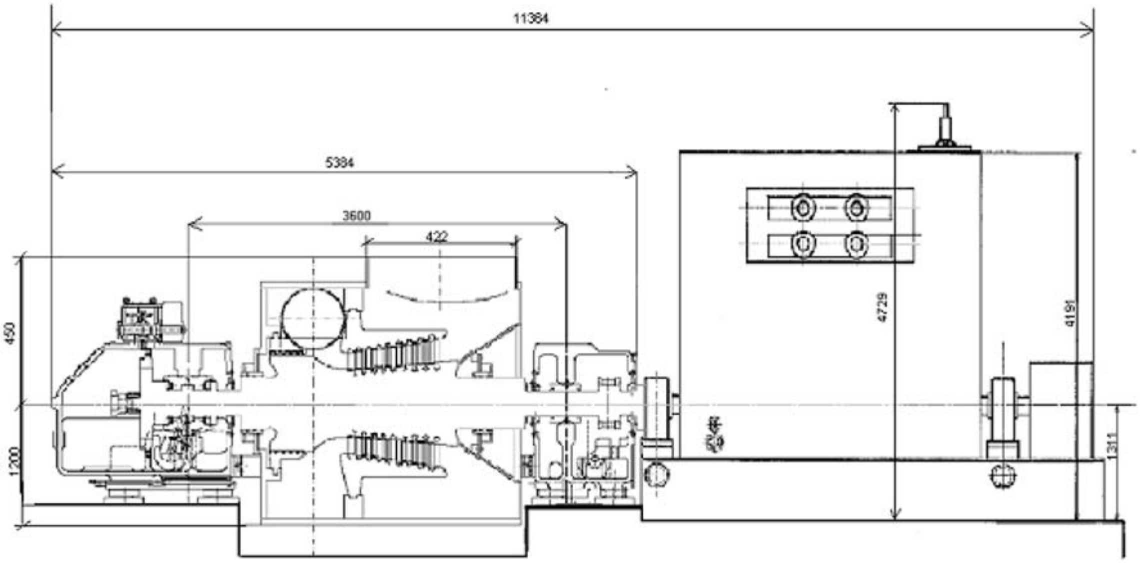


Figure 9-18: Turbine Generator General Arrangement for Case 4 (30% removal)

THE COST OF CARBON CAPTURE

Jeremy David and Howard Herzog
Massachusetts Institute of Technology (MIT), Cambridge, MA, USA

ABSTRACT

We have conducted a detailed analysis of costs associated with today's technology for CO_2 *separation and capture at three types of power plants: integrated coal gasification combined cycles (IGCC), pulverized coal-fired simple cycles (PC), and natural gas-fired combined cycles (NGCC). The analysis was based on studies from the literature that analyzed the economics of capturing* CO_2 *emitted at power plants. In this paper, we present a composite cost model and perform a sensitivity analysis to identify the cost-drivers for capture. We conclude that with new developments,* CO_2 *capture and sequestration can become a cost-effective mitigation pathway.*

INTRODUCTION

Fossil fuels currently supply over 85% of the world's energy needs and will remain in abundant supply well into the 21st century. They have been a major contributor to the high standard of living enjoyed by the industrialized world. However, their future is clouded because of the environmental and economic threat posed by possible climate change, commonly referred to as the "greenhouse effect". The major greenhouse gas is carbon dioxide (CO_2) and the major source of anthropogenic CO_2 is the combustion of fossil fuels. If we can develop technology to capture and sequester the fossil fuel CO_2 in a cost-effective and environmentally sound manner, we will be able to enjoy the benefits of fossil fuel use throughout the next century.

We have conducted a comparison of published studies from the past several years that analyzed the economics of capturing CO_2 at Integrated coal Gasification Combined Cycle (IGCC) power plants (six studies), Pulverized Coal (PC) power plants (four studies), and Natural Gas Combined Cycle (NGCC) power plants (four studies). MEA scrubbing of flue gas was used to capture the CO_2 in the PC and NGCC plants, but IGCC plants allow the use of more energy efficient scrubbing processes involving physical absorption to capture CO_2 from the high pressure synthesis gas. All studies were made using commercially available technology and include the cost of compressing the captured CO_2 to about 100 atm for pipeline transportation. The results do not include cost of CO_2 transportation and injection, which will add about $10/tonne of CO_2 avoided. Initial results were presented at GHGT-4 (Herzog, 1999), while detailed results of this analysis are presented in David (2000).

COMPOSITE COST MODEL OF CO_2 CAPTURE

Based on our analysis of the literature studies, we developed a composite cost model for CO_2 capture. The cost model developed uses six independent inputs, which were extracted from the literature studies we analyzed. Three first inputs characterize the reference (no capture) plant:

- Capital cost, in $/kW;
- Cost of electricity due to operation and maintenance, in mills/kWh;
- Heat rate, in Btu/kWh, defined on the lower heating value (LHV) basis.

We correlated the quantity of CO_2 emitted (E), in kg/kWh, as a function of heat rate for a given type of power plant (IGCC, PC or NGCC).

The second three inputs characterize the capture plant:

- Incremental capital cost, in $/kg of CO_2 processed per hour;
- Incremental cost of electricity due to operation and maintenance, in mills/kg of CO_2 processed;
- Energy requirements of the capture process, in kWh/kg of CO_2 processed.

The capture efficiency is usually about 90% in the studies reviewed. To compare the different types of capture plants on a similar basis, the capture efficiency needs to be kept constant. Consequently, we set the capture efficiency at a constant value of 90%.

The symmetry of the cost model inputs is shown in Table 1. The generation costs are normalized by the reference power plant output, while the capture costs are normalized by the quantity of CO_2 processed (which is directly related to the quantity and type of fuel burnt at the plant). These six parameters can be reasonably viewed as independent of each other. The inputs from the literature studies we analyzed are averaged for each type of power plant to obtain the composite cost model inputs shown in Table 2.

Table 1: Cost Model Inputs

	Reference Plant	**Capture Plant**
Capital Costs	$/kW	$/(kg of CO_2 processed per hour)
O&M Costs	mills/kWh	mills/kg of CO_2 processed
Energy Requirements	Btu/kWh	kWh/kg of CO_2 processed

It can be seen that NGCC power plants have the highest incremental capital cost and the highest energy requirements for the capture (0.354 kWh/kg of CO_2 processed), due to the low content of CO_2 in the flue gas (about 3%). Post-combustion decarbonization at PC plants is somewhat less energy intensive than at NGCC plants, 0.317 kWh/kg of CO_2 processed, because of the higher content of CO_2 in the flue gas (about 13%). Finally, the carbon dioxide is in a concentrated flow under a fairly high pressure at IGCC plants, so these plants have the lowest energy requirements (0.194 kWh/kg of CO_2 processed).

Table 2 reports the costs obtained for each type of power generation. We found that carbon dioxide capture increases the busbar electricity cost (COE) from 5.0 to 6.7 ¢/kWh at IGCC plants, from 4.4 to 7.7 ¢/kWh at PC plants, and, finally, from 3.3 to 4.9 ¢/kWh at NGCC plants.

Today, reference PC plants are slightly less expensive than reference IGCC plants. However, IGCC plants will become more economical than PC plants if carbon sequestration becomes necessary. Natural gas is always more competitive than coal for both reference and capture plants, assuming today's fuel prices remain constant. If gas prices rise relative to coal in the future, IGCC capture plants could then compete with NGCC capture plants.

Table 2: Cost Model for Capture Plants, in 2000 and 2012

Cycle	**IGCC**	**IGCC**	**PC**	**PC**	**NGCC**	**NGCC**
Data Description	**2000**	**2012**	**2000**	**2012**	**2000**	**2012**
Input						
Capital Cost, \$/kW	1401	1145	1150	1095	542	525
O&M, mills/kWh	7.9	6.1	7.4	6.1	2.5	2.4
Heat Rate (LHV), Btu/kWh	8081	7137	8277	8042	6201	5677
Incremental Capital Cost, \$/(kg/h)	305	275	529	476	921	829
Incremental O&M, mills/kg	2.65	2.39	5.56	5.00	5.20	4.68
Energy Requirements, kWh/kg	0.194	0.135	0.317	0.196	0.354	0.297
Basis						
Yearly Operating Hours, hrs/yr	6570	6570	6570	6570	6570	6570
Capital Charge Rate, %/yr	15	15	15	15	15	15
Fuel Cost (LHV), \$/MMBtu	1.24	1.24	1.24	1.24	2.93	2.93
Capture Efficiency, %	90	90	90	90	90	90
Reference Plant						
CO2 Emitted, kg/kWh	0.752	0.664	0.789	0.766	0.368	0.337
coe: CAPITAL, mills/kWh	32.0	26.1	26.3	25.0	12.4	12.0
coe: FUEL, mills/kWh	10.0	8.8	10.3	10.0	18.2	16.6
coe: O&M, mills/kWh	7.9	6.1	7.4	6.1	2.5	2.4
Cost of Electricity, ¢/kWh	4.99	4.10	4.39	4.10	3.30	3.10
Thermal Efficiency (LHV), %	42.2	47.8	41.2	42.4	55.0	60.1
Capture Plant						
Relative Power Output, %	85.4	91.0	75.0	85.0	87.0	90.0
Heat Rate (LHV), Btu/kWh	9462	7843	11037	9461	7131	6308
Capital Cost, \$/kW	1909	1459	2090	1718	1013	894
CO_2 Emitted, kg/kWh	0.088	0.073	0.105	0.090	0.042	0.037
coe: CAPITAL, mills/kWh	43.6	33.3	47.7	39.2	23.1	20.4
coe: FUEL, mills/kWh	11.7	9.7	13.7	11.7	20.9	18.5
coe: O&M, mills/kWh	11.6	8.4	15.7	11.6	5.1	4.4
Cost of Electricity, ¢/kWh	6.69	5.14	7.71	6.26	4.91	4.33
Thermal Efficiency (LHV), %	36.1	43.5	30.9	36.1	47.8	54.1
Comparison						
Incremental coe, ¢/kWh	1.70	1.04	3.32	2.16	1.61	1.23
Energy Penalty, %	14.6	9.0	25.0	15.0	13.0	10.0
Mitigation Cost, Capture vs. Ref., \$/tonne of CO_2 avoided	26	18	49	32	49	41

The mitigation cost (MC) in \$/tonne CO_2 avoided is given by the following equation:

$$MC = \frac{COE_{cap} - COE_{ref}}{E_{ref} - E_{cap}} \qquad (1)$$

The mitigation cost can be calculated by comparing a capture plant to any reference plant (e.g., capture IGCC vs. reference IGCC, PC or NGCC). Fig. 1 plots the cost of electricity vs. CO_2 emissions of the three reference plants and of an IGCC capture plant. The mitigation cost, which is simply the slope of the connecting lines shown on Fig. 1, varies depending on the reference plant chosen for the base case: IGCC (\$26 per tonne of CO_2 avoided), PC (\$33 per tonne of CO_2 avoided), and NGCC (\$121 per tonne of CO_2 avoided). Furthermore, the y-intercept of each line gives the cost of electricity that a zero emission technology must beat to be competitive with the IGCC sequestration option (7.76 cents per kWh based on a NGCC reference plant). It can be argued that NGCC plants should be the basis because they are the most popular plants being built today. This yields mitigation costs of \$121 per tonne of CO_2 avoided for a capture IGCC plant, \$168 per tonne of CO_2 avoided for a capture PC plant and \$49 per tonne of CO_2 avoided for a capture NGCC plant.

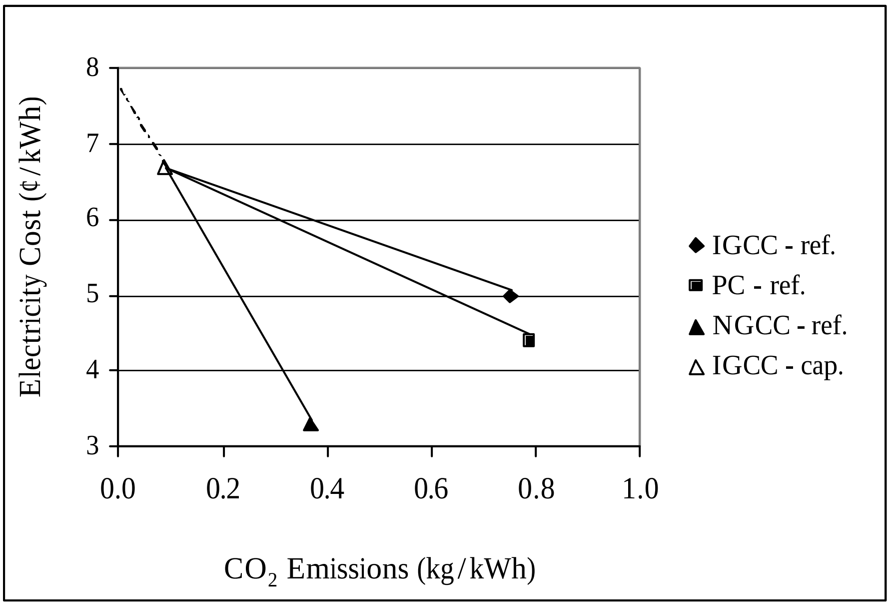


Figure 1: Calculating Mitigation Costs

IDENTIFICATION OF COST-DRIVERS AND FUTURE ECONOMICS

The six inputs of the cost model (see Table 2, year 2000 plants) are treated as independent variables. A sensitivity analysis (i.e., the inputs are decreased by 10% one by one for each type of power plant) is performed to identify the key inputs affecting the economics of the capture. Figures 2 and 3 show the change in incremental cost of electricity and mitigation cost at IGCC, PC, and NGCC power plants for a 10% decrease in each input. Note that a 10% decrease in heat rate is equivalent to an 11.1% increase in efficiency. Observations that can be drawn from Figs. 2 and 3 include:

- The key cost drivers are heat rate, energy required for capture, and capital costs of capture.
- Improving heat rates is extremely important for improving the economics of carbon sequestration. This supports a mitigation strategy that focuses on improved efficiency in the near-term, with sequestration becoming more important in the longer-term.

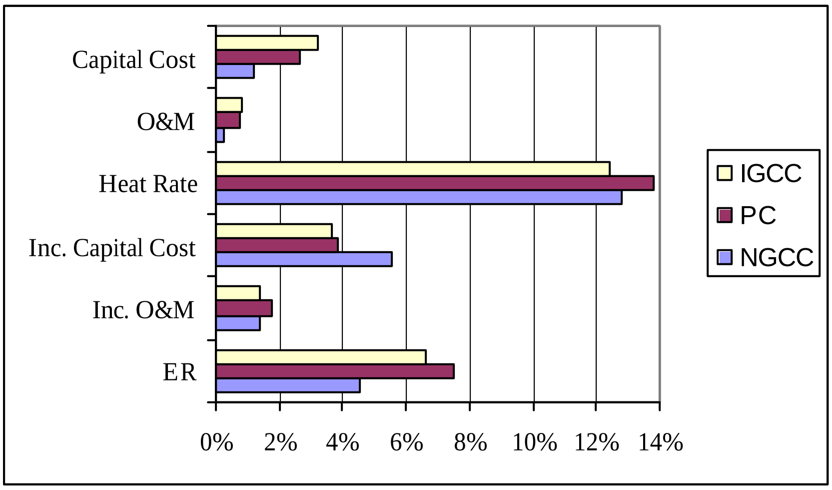


Figure 2: Incremental cost of electricity sensitivity to the cost model inputs. Decrease in incremental cost of electricity for a 10% decrease in each of the six inputs.

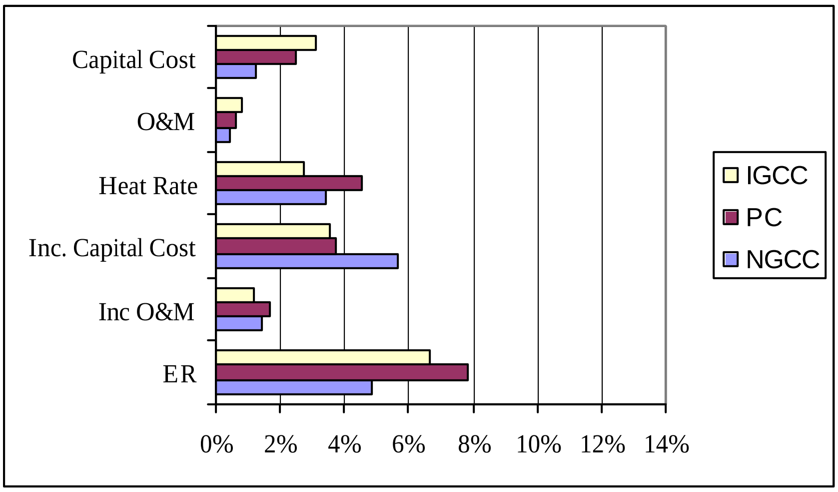


Figure 3: Mitigation cost sensitivity to the cost model inputs. Decrease in mitigation cost for a decrease in each of the six inputs..

FUTURE ECONOMICS

Technological improvements in power generation and capture technology can lower the capture costs. For instance, capital investment can be lowered and efficiency increased at the reference plant. Moreover, it is likely that improved solvents and system components will reduce the capital and energy costs for synthesis gas or flue gas treatment to separate and capture CO_2.

The capture costs in 2012 can be predicted by using the cost model. The 2012 capital costs, costs of operation and maintenance, and heat rates are taken from CURC (1998). Reductions in capital cost and gains in heat rate are significant at IGCC plants (above 10%), but limited at PC and NGCC plants (under 10%), which are more mature. The energy requirements are obtained by using the energy penalties given by Herzog and Drake (1993) at IGCC power plants, and by Mimura *et al.* (1997) at PC and NGCC power plants. The highest reductions in energy requirements for the capture processes are predicted to be at IGCC and PC plants (above 30%). Finally, it is assumed that the incremental capital cost, and the incremental cost of electricity due to operation and maintenance will be lowered by 10% from their 2000 level.

Table 2 gathers the economic performance of CO_2 capture at IGCC, PC and NGCC power plants in 2012. Although the capture costs are expected to decrease more at IGCC and PC plants than at NGCC plants, the overall economics are still more favorable at NGCC plants. New technologies like coal gasification show the most long-term promise, with incremental costs for CO_2 sequestration at IGCC power plants being potentially reduced to about 1¢/kWh in the next decade.

CONCLUSION

Based on the studies analyzed, there is a consensus that using today's capture technology would add 1.5-2¢/kWh to the busbar cost of electricity for an IGCC or NGCC power plant. For a PC plant, the incremental cost of electricity would be over 3¢/kWh. The strongest opportunities for lowering the capture costs in the future were identified as gains in heat rates and reductions in the amount of energy required by the separation. New technologies like coal gasification show the most long-term promise, with incremental costs for CO_2 sequestration at IGCC power plants being potentially reduced to about 1¢/kWh in the next decade. To put the costs presented here in context, further analysis with economic models is required (see Biggs *et al.*, 2000).

Opportunities for future cost reductions will include the investigation of innovative technologies, including new types of power plants and power cycles. Moreover, system-level analyses should be performed to minimize not only capture costs, but also the sequestration costs associated with transportation and injection.

REFERENCES

CURC (Coal Utilization Research Council), *Incentives and Research & Development for Early Commercial Applications of Clean Coal Technology* (November 1998).

Biggs S, H Herzog, H Jacoby and J Reilly, "Economic Modeling of CO_2 Capture and Sequestration," these proceedings (2000).

David J, "Economic Evaluation of Leading Technology Options for Sequestration of Carbon Dioxide," M.S. Thesis, Technology and Policy Program, MIT, Cambridge, MA (2000). Available at: <http://web.mit.edu/sequestration/JeremyDavid.pdf>

Herzog H and E Drake, "Long-Term Advanced CO_2 Capture Options," Cheltenham, U.K.: IEA/93/0E6; IEA Greenhouse Gas R&D Programme (1993).

Herzog H, "The Economics of CO_2 Capture," P Reimer, B Eliasson, A Wokaum, eds., *Greenhouse Gas Control Technologies*, Oxford: Elsevier Science Ltd., pp. 101-106 (1999).

Mimura T, H Simayoshi, T Suda, M Iijima and S Mituoka, "Development of Energy Saving Technology for Flue Gas Carbon Dioxide Recovery by Chemical Absorption Method and Steam System in Power Plant," *Energy Convers. Mgmt*. Vol. 38, Suppl., pp. S57-S62 (1997).

ACKNOWLEDGEMENTS

This work was conducted with support from the U.S. Department of Energy under award number DE-FG02-99ER62748. We gratefully acknowledge this support, as well as the encouragement of the program managers, John Houghton (BER program, Office of Science) and Bob Kane (Office of Fossil Energy).

ECONOMIC ASSESSMENT OF CARBON CAPTURE AND STORAGE TECHNOLOGIES

2011 update

2011

WorleyParsons
Schlumberger

SUPPORTED BY

GLOBAL
CCS
INSTITUTE

DISCLAIMER

This report has been prepared by WorleyParsons Services Pty Ltd (WorleyParsons) for the exclusive use of The Global CCS Institute. It is subject to and issued in accordance with the agreement between The Global CCS Institute and WorleyParsons. WorleyParsons accepts no liability or responsibility whatsoever for it in respect of any use of or reliance upon this report by any third party.

Copying this report without the permission of The Global CCS Institute or WorleyParsons is not permitted.

WorleyParsons
resources & energy

Schlumberger
Carbon Services

SUPPORTED BY

GLOBAL
CCS
INSTITUTE

PREFACE

This report presents an update of the economics of Carbon Capture and Storage (CCS) prepared in 2009. The 2009 report was commissioned by the Global CCS Institute and delivered as Foundation Report Two in the series of five studies undertaken as part of the Strategic Analysis of the Global Status of Carbon Capture and Storage.

Foundation Report Two involved a detailed analysis of the capture, transport and storage costs for power plants and a select range of industrial applications. This report presents a transparent methodology that uses updated and refined costs to reflect changes in the market since 2009.

ACKNOWLEDGEMENTS

WorleyParsons gratefully acknowledges the Global CCS Institute for funding this study and particularly the efforts of Mr Chris Short (Chief Economist).

This project was sponsored by Peter Brooks (WorleyParsons) and the Project Manager was Chai McConnell (WorleyParsons).

The report authors are Dr James Simpson, Chai McConnell and Yukiyo Matsuda (WorleyParsons).

Dr Geoff Ingram, Dwight Peters, Koji Kusaka and Arnaud van der Beken (Schlumberger) provided information on storage costs that contributed to the economic model and report.

Additional input was provided by Michael DeLallo and Vlad Vaysman (WorleyParsons).

Graphic support was provided by Anthony Holt (WorleyParsons).

ECONOMIC ASSESSMENT OF CARBON CAPTURE
AND STORAGE TECHNOLOGIES: 2011 UPDATE

For general enquires about this or any of the foundation reports, please contact
the **Global CCS Institute**, GPO Box 828, Canberra ACT 2601 Australia
or email info@globalccsinstitute.com

CONTENTS

LIST OF TABLES

LIST OF FIGURES

ABBREVIATIONS

3D	3-Dimensional
AACE	Association for the Advancement of Cost Engineering
AD, A$	Australian dollar
AFUDC	Allowance for Funds Used During Construction
AGR	acid gas removal
ANZ	Australia and New Zealand
ASU	air separation unit
CAPEX	capital expense
CCS	carbon (carbon dioxide) capture, transport and storage
CO_2	carbon dioxide
CO_2CRC	the Australian Government's Cooperative Research Centre for Greenhouse Gas Technologies
CTG	combustion turbine generator
ºC	degrees Celsius
DrillEX	drilling expenditure
ECBM, ECBMR	enhanced coal bed methane recovery
EGR	enhanced gas recovery
EIA	(United States) Energy Information Administration
EOR	enhanced oil recovery
EPC	engineering, procurement and construction contract or contractor
EPCM	engineering, procurement, construction and management contract or contractor
EPRI	Electric Power Research Institute
FEED	front-end engineering design
FGD	flue gas desulphurisation
FGR	flue gas recirculation
FID	final investment decision
FOAK	first-of-a-kind
FOM	fixed operating and maintenance
FRST	Foundation for Research Science and Technology
G8	Group of Eight (Canada, France, Germany, Italy, Japan, Russia, the United Kingdom, and the United States)
GHG	greenhouse gas
GNS	GNS Science, a New Zealand research and consultancy services organisation
GJ	gigajoules
GWe	gigawatt-electrical
HHV	higher heating value
IEA	International Energy Agency
IGCC	integrated gasification combined cycle
IPCC	Intergovernmental Panel on Climate Change
ITM	ion transfer membrane
kg	kilogram
kJ	kilojoule
km	kilometre
kW, kWe	kilowatt-electrical
LCOE	levelised busbar cost of electricity, often expressed in $/MWh
LNB	low NOx burner
m	metre
ME	Middle East
MEA	monoethanolamine
MMV	measurement, monitoring and verification
MPa	megapascal
Mtonne, Mt	million (106) tonnes
Mtpa	million (106) tonnes per annum
MW, MWe	megawatt-electrical
MWh	megawatt-hours
NA, N/A	not applicable
NETL	US DOE National Energy Technology Laboratory
NGCC	natural gas-fired combined cycle
NOAK	nth-of-a-kind
NOx	nitrous oxides
O&M	Operating and Maintenance
OFO	overfire oxygen
OPEX	operating and maintenance expense
pa	per annum
Pa	pascal = N/m^2
PC	pulverised coal
PCC	post-combustion capture
PPP	Public Private Partnership
RD&D	research, development and demonstration
R&D	research and development
ROW	right(s) of way
SCR	selective catalytic reduction
Sm^3	standard cubic metres
Svy	survey
Syngas	synthetic gas
tonne	metric ton, (1,000kg)
TPC	total plant cost
UK	United Kingdom
USA, US	United States of America
USC	ultra-supercritical
USD, US$	United States dollar
US DOE	United States Department of Energy
US EIA	United States Energy Information Administration
US EPA	United States Environmental Protection Agency
USGC	US Gulf Coast
VOM	variable operating and maintenance

1 EXECUTIVE SUMMARY

In May 2009, a consortium led by WorleyParsons and comprising Schlumberger, Electric Power Research Institute and Baker & McKenzie was engaged to undertake the Strategic Analysis of the Global Status of Carbon Capture and Storage (CCS).

The consortium was tasked with undertaking a comprehensive survey of the status of CCS and to develop a series of reports analysing CCS projects, the economics of CCS, policies supporting CCS development and existing research and development networks. A fifth report – the Synthesis Report – was also developed and this summarised the findings of the first four reports, and provided a comprehensive assessment of the gaps and barriers to the deployment of large-scale CCS projects, including strategies and recommendations to address these issues.

The second of this series of reports (Foundation Report Two) presented a detailed analysis of the capture, transport and storage costs for power plants and a select range of industrial applications. The costs of CCS were presented on a levelised cost of production basis, as well as for the cost of carbon dioxide (CO_2) captured and avoided. Foundation Report Two also considered the application of CCS in first-of-a-kind (FOAK) systems and nth-of-a-kind (NOAK) systems.

The modelling determined that the cost of CCS for power generation, based on the use of commercially available technology, was found to range from US$57-107 per tonne of CO_2 avoided or US$42-90 per tonne of CO_2 captured. The lowest cost of CO_2 avoided was at US$57 per tonne of CO_2 for the oxyfuel combustion technology, while the highest cost at US$107 per tonne of CO_2 for the natural gas-fired combined cycle (NGCC) with post-combustion capture (PCC). This compared with the lowest cost of captured CO_2 for the IGCC and oxy-combustion technologies at US$39 and US$42 per tonne of CO_2 respectively and the highest of $90 per tonne of CO_2 for NGCC technologies. The metrics were determined for the reference site in the United States of America (USA) with fuel costs based on values typical for 2010.

For the reference cases, taking into account currently available technologies, the levelised cost of electricity (LCOE) for FOAK pulverised coal (PC) supercritical technology was the greatest at US$131/MWh, while the oxy-combustion was the lowest of the commercially available technologies at US$121/MWh. While the cost of CO_2 avoided and captured range by a factor of two, the LCOE estimates ranged between US$121-131/MWh with currently available technologies.

The percentage increases in costs that the application of CCS has over non-CCS facilities were also explored. For power generation, facilities that had the lowest cost increases were IGCC (37 per cent), NGCC (40 per cent), followed by oxyfuel combustion (53 to 65 per cent) and PC supercritical (61 to 76 per cent) technologies.

The application of CCS for FOAK industrial applications showed that cost of CO_2 avoided was lowest for natural gas processing (US$19) and fertiliser production (US$20) followed by cement production and blast furnace steel production (US$54).

1 EXECUTIVE SUMMARY (CONTINUED)

The lowest cost increase was for natural gas processing (1 per cent) followed by fertiliser production (3 per cent). This was unsurprising given that these industries already have the process of capturing CO_2 as a part of their design. The production of steel (10 to 14 per cent) and cement (39 to 52 per cent) had the highest percentage cost increases with the application of CCS because the capture of CO_2 is not inherent in the design of these facilities.

The margin of error in this study made it difficult to select one technology over another based on the LCOE. Projects employing different capture technologies may be viable depending on a range of factors such as location, available fuels, regulations, risk appetite of owners and funding.

In July 2010, WorleyParsons and Schlumberger were engaged to undertake an update of Foundation Report Two. The objectives of this update were to:

- improve the regional localisation estimates;
- update and enhance capital cost estimates for power and a select range of industrial activities that could apply CCS; and
- update the economic model (having regard for the two previous items) to consider what, if any, material changed have occurred to the economics of CCS since 2009.

To meet these objectives, changes/modifications to the economic assessment methodology included:

- the revision of overnight capital costs used in the 2009 report to early 2010 US$;
- the revision of regional specific factors to move the capital costs from the reference location to the location of interest;
- the review of coal and natural gas prices on a regional basis, including the consideration of whether the fuels were locally sourced or imported and subject to international market prices;
- the adjustment of process parameters (heat rate, CO_2 emissions and CO_2 capture) according to regional coal composition and emissions requirements;
- a change in the approach for CCS on the oxyfuel combustion power generation to include an additional purification step of the CO_2 to increase the CO_2 purity to greater than 95 per cent;
- the modification of the reference pipeline length to 100km, based on findings for large-scale integrated CCS projects as identified in Foundation Report One;
- a revised approach to CO_2 storage, which considered two cases of a ‘good’ reservoir and a ‘poorer’ reservoir with either 3Mtpa or 12Mtpa injection scenarios; and
- consideration of variations in storage costs across regions. Recent data on CO_2 storage costs were obtained for Australia/New Zealand, Europe and North America for CO_2 injection wells and associated services.

The revised results of the economic assessment of CCS technologies are presented in Table 1-1.

Table 1-1 Summary results of the economic assessment of CCS technologies

		Power generation				Industrial applications			
		PC supercritical & ultra supercritical*[1]	Oxyfuel combustion standard & ITM*[1]	IGCC	NGCC	Blast furnace steel production	Cement production	Natural gas processing	Fertiliser production
	Dimensions	US$/MWh	US$/MWh	US$/MWh	US$/MWh	US$/tonne steel	US$/tonne cement	US$/GJ natural gas	US$/tonne ammonia
Levelised cost of production	Without CCS*[2]	73-76	73-76*[3]	91	88	570-800	66-88	4.97	375
	With CCS FOAK*[3]	120-131	114-123	125	123	82	34	0.056	11
	With CCS NOAK*[4]	117-129	112-121	123	121	74	31	0.056	11
	% Increase over without CCS*[5]	61-76%	53-65%	37%	40%	10-14%	39-52%	1%	3%
Cost of CO_2 avoided*[6] (\$/tonne CO_2)	FOAK	62-81	47-59	67	107	54	54	19	20
	NOAK	57-78	44-57	63	103	49	49	19	20
Cost of CO_2 captured (\$/tonne CO_2)	FOAK	53-55	42-47	39	90	54	54	19	20
	NOAK	52	41-45	38	87	49	49	19	20

Notes:

[1] The ultra-supercritical and ITM technologies are currently under development and are not commercially available. These technologies represent options with the potential for increasing the process efficiency and reducing costs.

[2] Without CCS cost of production for industrial process are typical market prices for the commodities.

[3] Oxyfuel combustion systems are not typically configured to operate in an air fired mode. Therefore, oxyfuel combustion without CCS is not an option. The values here are the PC without CCS value to be used as a reference for calculating the cost of CO_2 avoided.

[4] For industrial processes, levelised cost of production presented as cost increment above current costs.

[5] Expressed with respect to current commodity prices of industrial processes.

The updated modelling yielded the findings listed below.

- All of the coal-fired technologies showed a decrease in fuel costs related to the lower coal costs in 2010.
- For the reference cases, taking into account currently available technologies, the lowest LCOE was for oxyfuel combustion at US$114/MWh, in contrast to 2009 where LCOE for NGCC technologies was the lowest at US$112/MWh. Consistent with the findings in 2009, the LCOE for PC supercritical and IGCC technologies were the greatest at US$131/MWh and US$125/MWh respectively.
- The percentage increases in costs that the application of CCS has over non-CCS facilities have remained relatively unchanged since 2009.
- There was an increase in the capital contribution to the LCOE for oxyfuel combustion with CCS, reflecting the inclusion of an additional purification process when capturing CO_2.
- CO_2 capture still represents the greatest contribution to the cost of CCS, with the majority of the cost increases being due to changes in the capture system.
- The reduction in the length of the pipeline for the reference case has reduced the overall transport costs and the contribution of transport cost to the overall cost of CCS.

1 EXECUTIVE SUMMARY (CONTINUED)

- Consistent with the findings from Foundation Report Two in 2009, the range in coal price lead to a US$10/MWh variation in the LCOE, while for the natural gas price range the variation in the LCOE was around US$30/MWh.
- For a supercritical PC with CCS technology, for a fixed fuel cost, the sensitivity of the CO_2 capture installed capital costs and LCOE to the labour costs was reduced. The installed capital costs increased by 23 per cent (32 per cent in 2009), while the LCOE increased by 11 per cent (21 per cent in 2009). A similar trend would be observed for the other coal-fired technologies as they tend to be relatively labour-intensive installations.
- The installed CO_2 capture equipment cost and LCOE increased across all technologies in India. This was due to the consideration of a 30 per cent increase in equipment being imported into the country as well as India's typical coal heating value being very low, resulting in a greater capital cost.
- Costs increased across all technologies in Eastern Europe, primarily due to the increase in the reference coal price for the region.
- A 20 per cent increase in the technology cost in Australia which can be accounted for by the higher coal price utilised this year.
- A significant increase in the costs in Brazil, partially because of a lower labour rate being used in 2009. The revision of the coal type to one with a lower heating value also lead to a higher capital cost. Finally, additional costs associated with importing capital equipment contributed to the increase in CO_2 capture costs in Brazil.
- Only NGCC costs are displayed for Saudi Arabia, reflecting that there are no coal-fired power generation applications in the region.
- The breakpoint for the CO_2 credit value for oxyfuel has decreased from US$60/tonne of CO_2 in 2009 to US$55/tonne, which can be attributed to the lower coal costs offsetting the additional purification step included in this study. This analysis continues to indicate that oxyfuel still has the lowest CO_2 credit value breakpoint of approximately US$55/tonne of CO_2.
- The IGCC breakpoint, with respect to supercritical PC technology has decreased from $80/tonne in 2009 to $70/tonne of CO_2. This reflects the increase since 2009 in the LCOE and cost of CO_2 avoided and captured for IGCC with CCS.
- The cost breakpoint for the supercritical technologies is approximately $80/tonne of CO_2, an 11 per cent decrease from the 2009 breakpoint of $90/tonne of CO_2.
- The high breakpoint for NGCC technology has remained relatively unchanged at $112/tonne of CO_2, reflective of the lower CO_2 emission intensity of natural gas and higher cycle efficiency compared to coal-fired technologies.
- For the industrial processes, the incremental levelised product costs and the cost of CO_2 avoided/captured have increased by a small amount consistently across all applications.
- The cost to transport CO_2 is estimated to be between US$1-2 per tonne of CO_2, a decrease from US$3-4 per tonne of CO_2 in 2009. This is due to the reduction of the pipeline length in the reference case from 250km in 2009 to 100km.
- The contribution of storage cost to the LCOE was found to range from US$6-13 per tonne of CO_2 depending on whether the 'good' or 'poorer' reservoir option was considered.

Though minor changes in the costs of CCS across power generation and industrial applications have occurred, the costs of CCS still remain high. This is expected, given that it has only been 12 months since the initial Foundation Report Two, and major developments that have the potential to dramatically reduce the cost of CCS have not yet occurred.

Despite the costs of CCS being high relative to traditional power generation and industrial facilities, it is important to consider that these traditional methods currently emit large amounts of CO_2 into the atmosphere. Given the current and anticipated restrictions on facility emissions, these facilities will not be allowed to continue to operate as they have in the past.

The high costs of CCS as identified in this study should be considered with other low emission technologies to allow consideration of approaches to low emission power and industrial production. Further, if CCS is compared against the anticipated cost that may be imposed on facilities for emitting CO_2 it is likely to appear more competitive in a low carbon market.

2 INTRODUCTION

2.1 The importance of CCS

The successful development and widespread deployment of Carbon Capture and Storage (CCS) is considered by many key climate change stakeholders to be fundamental to achieving deep cuts in carbon dioxide (CO_2) emissions to atmosphere. Among a portfolio of responses, such as energy efficiency and renewable energy, CCS is required to contribute approximately 19 per cent of CO_2 emissions reductions globally by 2050 (International Energy Agency, 2008), if global emissions targets are to be achieved.

The business case to developing and deploying large-scale-integrated CCS projects (LSIPs) is challenging. One of the key challenges is the relatively high cost of CCS technology to capture, transport and safely store CO_2, compared to the same facilities without CCS. This is unsurprising, given that it costs more to capture, transport and safely store CO_2 as opposed to the current 'business as usual' scenario of venting CO_2 emissions to the atmosphere.

This report was commissioned by the Global CCS Institute as an update to the 2009 Foundation Report Two.

2.2 Background

The objective of this report is to build upon and update the cost estimates provided in 'Report 2: Economic Assessment of Carbon Capture and Storage Technologies' of the 2009 'Strategic Analysis on the Global Status of CCS'. As in 2009, the cost estimates are informed by WorleyParsons experience in the design, construction and operation of large infrastructure facilities that are likely candidates to apply CCS and direct engagement in assisting proponents to develop CCS projects. The cost estimates of storage are provided by Schlumberger and are similarly informed by their leading global position in this space.

Readers are encouraged to review the 2009 report if more background information and understanding is required. This can be found on the Global CCS Institute website at: http://www.globalccsinstitute.com/.

2.3 Scope

Building upon the 2009 study, the scope of this update is to consider the economics of CCS, based on 2010 capital, fuel and labour costs, and to assess several issues including:

- improving the regional localisation estimates;
- updating and enhancing capital cost estimates for power and a select range of industrial activities that could apply CCS; and
- updating the economic model (having regard for the two items above) to consider what, if any, material changes had occurred to the economics of CCS since 2009.

The additions to the scope were achieved by enhancing and improving operating cost estimates for capture facilities by improving fuel cost estimates at select locations and other key variables. Project location indices were also enhanced by providing greater specification that, where practicable, considered costs at a typical reference city level rather than a broad regional level.

2.4 Caveats and exclusions

As with the delivery of the Economic Assessment of Carbon Capture and Storage Technologies report (WorleyParsons, 2009), WorleyParsons and Schlumberger used its best endeavours to inform this update. Cost estimates used were observed in the global marketplace for developing large and often complex infrastructure projects and the costs of drilling for hydrocarbon were used as analogues for the cost of storage. US Gulf Coast (USGC) was the reference location.

The authors caution that when comparing costs, it is important to understand the purpose of the presented costs; are they presented to compare the costs between different technologies, or to inform how much a specific project will cost. There are several studies (e.g. the NETL Bituminous Baseline Study) which provide valuable information regarding the comparative costs of CCS technologies and the factors that impact these costs. The Global Carbon Capture Storage (CCS) Institute costing methodology and the studies prepared using it, fall into this group. These studies are typically poor predictors of project costs because they cannot accurately account for the variation in site and owner specifications included in a real project cost. Alternately, reported project costs, for specific projects, are poor sources for comparing technology costs. By the time the costs of a project are reported, only the cost of a single technology is presented which takes into account site specific requirements and owner's preferences.

The authors recognise that the economics of CCS is the subject of much conjecture and debate. Much of the conjecture is founded upon differing studies providing different results. In many cases, different results arise because key variables such as capital and operating costs, location and cost year differ. Additionally, the basis of the costs, that is which costs are included, are often not well defined or overlooked. Indeed, it was largely because of this that the Global CCS Institute commissioned the original 2009 study in an effort to provide cost estimates based on transparent and consistent variables and assumptions. The authors caution readers that care is required when applying the results of the cost estimates provided in this report. The margin of error in this study is +/- 40 per cent and the significant impact that project location and preference for CCS technology type, for example, means that the economics of CCS projects needs to be assessed on a case-by-case basis. It is important to note that all project costs are specific to that project and the figures presented in this report represent 'ball park' cost estimates for developing CCS projects as at 2011.

Furthermore, the authors note that some recent studies on the economics of integrated gasification combined cycle (IGCC) capture facilities are documenting costs that are greater than those presented in this study. Analogously, some CCS stakeholders speculate that the cost of IGCC with CCS is prohibitive based on the findings of these studies. Caution needs to be taken in considering this issue. One finding of the work performed by the Global CCS Institute is the costs can vary significantly based on location specific factors such as labor rates, fuel costs, and fuel characteristics. Additionally, with high volatility in plant construction costs and few new coal fired power (without CCS) construction starts in locations, real project costs are difficult to gauge. Therefore, the suggestion of providing a reference plant cost without CO_2 capture, using the same basis, along side of the facility with capture is suggested to give a better indication of the costs of CCS.

Additionally, more detailed engineering has been undertaken for large-scale IGCC plants with CCS than for other power generation applications. As a result, lower levels of definition for technical design and cost for oxyfuel combustion and post-combustion CO_2 capture technologies can be expected. Given that the cost estimates for IGCC increased as the projects were further defined across the asset lifecycle, it can be expected that cost estimates for the other capture technologies may also increase. In other words, the perception that oxyfuel combustion and post-combustion CO_2 capture are economically more viable than IGCC may not be observed when projects applying these technologies undergo more detailed evaluations in the future.

2.5 Following chapters

Chapter Three provides an overview of the methodology. Chapter Four presents the results and Chapter Five presents key conclusions and observations.

3 METHODOLOGY FOR CCS ECONOMIC ANALYSIS UPDATE

3.1 Overview of the economic assessment methodology

This cost analysis of CCS projects is an update of the 2009 Foundation Report Two. Building upon the 2009 report, the methodology essentially assesses the capital and operating costs over the life of an investment necessary to meet CO_2 emission reduction goals. The goal of private sector developers is to select the CCS technology that maximises profits over the long run in a sustainable way.

The methodology selected to define the economics of CCS investments is based on using capital costs and operating characteristics from published sources; WorleyParsons' and Schlumberger's in-house database of actual cost data gained from undertaking numerous designs, installations and analyses of CCS projects; and other data from corporate, government and research stakeholders. These were updated to 2010 dollars.

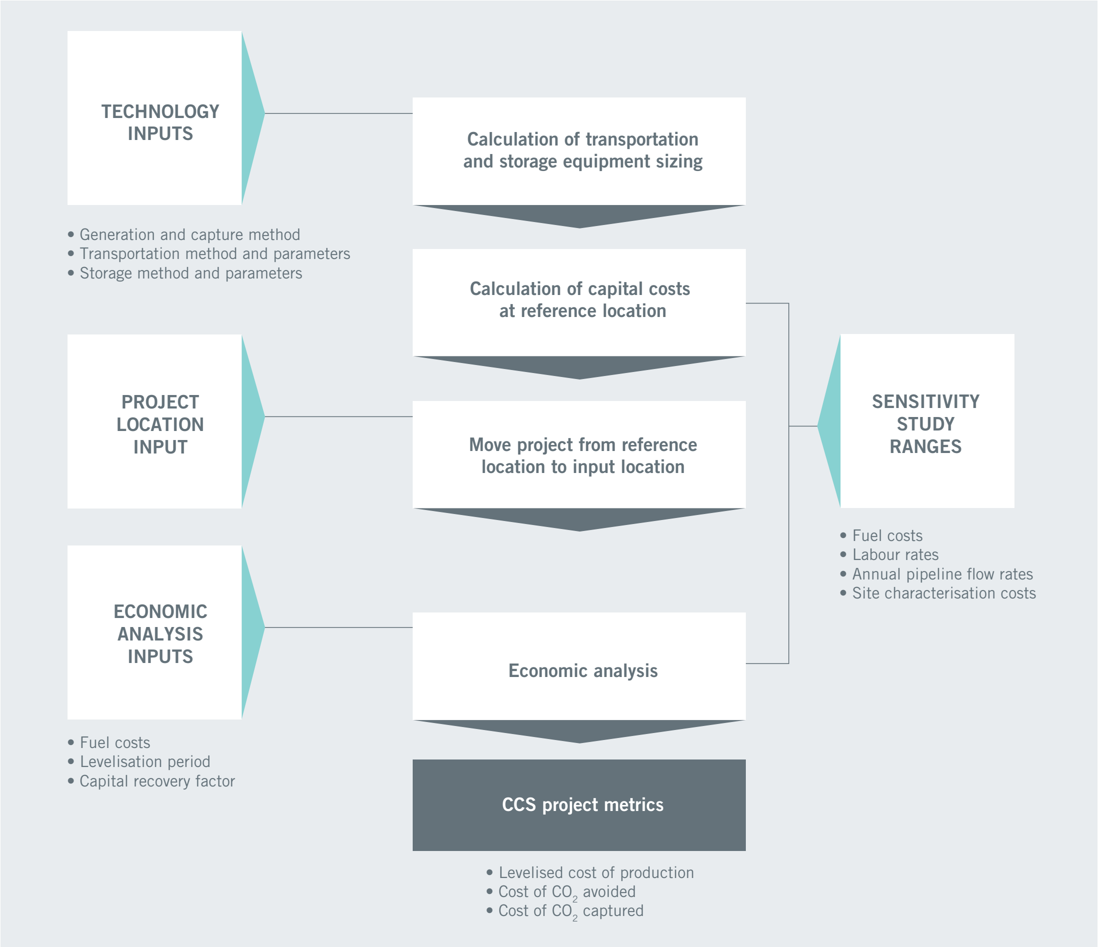
The methodology applied in this study combines these parameters to determine the appropriate metrics to be used in the economic analysis of CCS. This was conducted through:

- calculations of capital costs for the reference location;
- transposing the project to the selected location; and
- performing the subsequent economic analysis.

3 METHODOLOGY FOR CCS ECONOMIC ANALYSIS UPDATE (CONTINUED)

Figure 3-1 provides a flow chart showing, at a high level, the approach undertaken in the analysis.

Figure 3-1 Flowchart of methodology for CCS economic assessments



The CO_2 capture options considered for power generation include:

- post-combustion capture;
  - supercritical pulverised coal (PC) boiler
  - ultra-supercritical PC boiler
  - natural gas-fired combined cycle (NGCC).
- pre-combustion capture using IGCC; and
- oxy-combustion.

The application of CCS was considered on four industrial processes:

- blast furnace production of steel;
- cement kiln/furnaces;
- natural gas processing; and
- fertiliser production (ammonia).

While there are several methods of transporting CO_2 to a storage site including trucking and shipping, this study consider only the costs involved in pipelining as a means of CO_2 transport. To meet the G8 timeframes it is more than likely that transportation by pipeline will be the preferred approach, given the large volumes of CO_2 that will need to be transferred.

The 2009 study found that while the largest cost component in the CCS value chain was with the capture component, the CO_2 storage component of the value chain represents the greatest uncertainty. To model this uncertainty, the 2010 study provides a range of scenarios for ‘good’ and ‘poorer’ storage characteristics across a range of 3Mtpa and 12Mtpa reservoir capacities. This is discussed further in section 3.8.

As there have been no extensive changes in the economic and assessment methodology, refer to Foundation Report Two for the comprehensive description of the methodology. The metrics used to compare the costs of CCS projects have remained unchanged, with costs being presented in the form of:

- levelised cost of production;
- cost per tonne of CO_2 captured and injected; and
- cost per tonne of CO_2 avoided.

3.2 New methods used in the 2011 update

A change to the methodology occurred in the adjustment of process parameters of heat rate, CO_2 emissions and CO_2 capture according to regional coal composition and emissions requirements. This resulted in changes to the capital and operating and maintenance (O and M) costs for the capture facilities. This change improved the granularity of the estimates and analyses for the discrete regions given that:

- process design for coal-fired facilities is strongly dependent on local conditions including fuel composition, emissions regulations, climate and availability of water for cooling;
- design impacts sizing of equipment and thus capital costs;
- the heat rate of facilities increases with decreasing coal rank (heating value), therefore this increases the amount of fuel required and potential change in fuel costs; and
- the CO_2 emission intensity change relates to changes in fuel, potential impact on cost of CO_2 capture and avoided.

The adjustment of the process parameters was undertaken by:

- conducting a literature/study review of facility designs with different coal ranks;
- determining the heat rate and cost trend variations from reference coal designs. Trends described as simple functions from reference coal composition (Pittsburgh No. 8); and
- the adjustment of the capture technology costs at the reference location and then moving these to the region of interest.

3.3 Update of reference facility costs to 2010

The overnight capital costs used in the 2009 report were revised to early 2010 US dollars. Over the span between the reporting periods, the equipment costs have increased on the order 1 per cent while the labour costs have increased 0.5 per cent. Table 3-1 and Table 3-2 provide the operating parameters and updated capital costs used in the economic analysis. In Table 3-2, the cost of oxy-combustion without capture is set to zero based on that oxy-combustion is not considered without CCS.

A key change from the 2009 approach for CCS on the oxyfuel combustion power generation has been implemented. CCS on this facility now considers an additional purification step of the CO_2 to increase the CO_2 purity to greater than 95 per cent.

Table 3-1 Electric power generation (supercritical, ultra-supercritical and IGCC) facility parameters with and without capture.

		PC supercritical		**Supercritical 2**		**Ultra-supercritical**		**IGCC**	
	CO_2 Capture	**No**	**Yes**	**No**	**Yes**	**No**	**Yes**	**No**	**Yes**
	Fuel	**Coal**	**Coal**	**Coal**	**Coal**	**Coal**	**Coal**	**Coal**	**Coal**
Gross power output	MW	580	663	580.2	661.1	576.6	644.4	748	694
Auxiliary power	MW	30	117	30.2	111.1	26.6	94.4	112	176
Net power output	MW	550	546	550	550	550	550	636	517
Net plant HHV efficiency	%	39.10%	27.20%	39.4	28.3	44.6	33.2	41.10%	32.00%
Net plant HHV heat rate	GJ/MWh	9.20	13.22	9.14	12.73	8.07	10.83	8.76	12.61
CO_2 generated	tonne/hr	442	631	440	613	389	521	479	470
CO_2 emitted	tonne/hr	442	63	440	61	389	52	479	47
CO_2 captured	tonne/hr	0	568	0	551	0	469	0	423
Emission intensity	kg/MWh	804	115	800	112	707	95	753	90
Plant capital overnight costs ($)									
CC equipment	x1,000	539,576	957,610	542,507	950,517	559,043	930,710	948,200	962,087
CC materials	x1,000	49,823	64,954	49,079	62,080	47,287	58,829	74,721	73,827
CC labour	x1,000	276,417	437,869	275,494	449,302	267,061	422,973	256,759	296,617
Eng. CM HO & fees	x1,000	78,055	133,336	77,810	133,053	78,074	128,088	104,631	113,334
Process contingency	x1,000	0	63,711	0	61,096	40,686	104,045	55,227	74,764
Project contingency	x1,000	111,985	234,026	111,926	235,789	114,883	227,726	224,927	244,510
Total	x1,000	1,055,857	1,891,506	1,056,815	1,891,836	1,107,033	1,872,371	1,664,464	1,765,138
Total overnight	$/kW	1,919	3,464	1,921	3,440	2,013	3,404	2,618	3,413
Variable O&M ($)									
VOM equipment	$/MWh	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
VOM materials	$/MWh	3.19	6.03	3.13	6.07	2.70	5.07	1.40	2.10
VOM labour	$/MWh	1.63	2.89	1.51	2.88	1.63	2.77	3.15	3.82
Total	$/MWh	4.82	8.92	4.64	8.95	4.34	7.84	4.55	5.92
Fixed O&M ($)									
FOM equipment	x1,000	0	0	0	0	0	0	0	0
FOM materials	x1,000	10,181	17,974	10,206	18,021	10,226	17,334	26,656	26,342
FOM labour	x1,000	9,197	11,727	9,716	10,478	9,205	10,367	11,636	12,127
Total	x1,000	19,378	29,701	19,922	28,499	19,431	27,701	38,292	38,469

Table 3-2 Electric power generation (oxy-combustion and NGCC) facility parameters with and without capture.

		Oxy-combustion supercritical		Oxy-combustion ultra-supercritical		Oxy-combustion ITM supercritical		NGCC	
	Capture	**No**	**Yes**	**No**	**Yes**	**No**	**Yes**	**No**	**Yes**
	Fuel	**Coal**	**Coal**	**Coal**	**Coal**	**Coal**	**Coal**	**NG**	**NG**
Gross power output	MW	NA	786	NA	759	NA	688	570	520
Auxiliary power	MW	NA	236	NA	209	NA	138	10	38
Net power output	MW	NA	550	NA	550	NA	550	560	482
Net plant HHV efficiency	%	NA	29.3	NA	33	NA	29.3	50.80%	43.70%
Net plant HHV heat rate	GJ/MWh	NA	12.30	NA	10.92	NA	12.27	7.09	8.24
CO_2 generated	tonne/hr	NA	592	NA	525	NA	560	202	202
CO_2 emitted	tonne/hr	NA	0	NA	0	NA	47	202	20
CO_2 captured	tonne/hr	NA	592	NA	525	NA	514	0	182
Emission intensity	kg/MWh	NA	0	NA	0	NA	85	362	42
Plant capital overnight costs ($)									
CC equipment	x1,000	NA	851,291	NA	835,378	NA	950,517	245,193	391,048
CC materials	x1,000	NA	61,530	NA	60,532	NA	62,080	23,710	28,771
CC labour	x1,000	NA	503,652	NA	468,571	NA	449,302	58,152	105,552
Eng. CM HO & fees	x1,000	NA	130,217	NA	124,831	NA	133,053	29,435	47,283
Process contingency	x1,000	NA	53,683	NA	68,832	NA	61,096	0	33,450
Project contingency	x1,000	NA	195,215	NA	185,014	NA	235,789	41,892	91,356
Total	x1,000	NA	1,795,588	NA	1,743,157	NA	1,891,836	398,382	697,460
Total overnight	$/kW	NA	3,265	NA	3,169	NA	3,440	711	1,447
Variable O&M ($)									
VOM equipment	$/MWh	NA	0.00	NA	0.00	NA	0.00	0.00	0.00
VOM materials	$/MWh	NA	3.44	NA	3.05	NA	3.36	0.48	1.01
VOM labour	$/MWh	NA	2.75	NA	2.68	NA	2.46	0.69	1.29
Total	$/MWh	NA	6.19	NA	5.73	NA	5.82	1.18	2.30
Fixed O&M ($)									
FOM equipment	x1,000	NA	0	NA	0	NA	0	0.00	0.00
FOM materials	x1,000	NA	17,207	NA	16,774	NA	15,417	4,412.31	7,060.31
FOM labour	x1,000	NA	10,346	NA	10,276	NA	10,053	3,411.58	4,560.27
Total	x1,000	NA	27,553	NA	27,050	NA	25,470	7,823.89	11,620.57

3 METHODOLOGY FOR CCS ECONOMIC ANALYSIS UPDATE (CONTINUED)

3.4 Regional specific factors

This update revises factors for moving the capital costs from the reference location to the location of interest and is described in section 3.4.1. The inclusion of region specific fuel information is described in section 3.4.2.

3.4.1 Development of regional cost indices

To support conversion of the reference case costs from USGC to location specific costs (expressed in US dollars) for the selected cities/countries, conversion indices were developed for three major cost elements. These include imported equipment and materials, locally sourced equipment and materials, and labour. The conversion indices were developed using data from Richardson Products' International Cost Factor Location Manual 2009-2010 Edition. Land cost within a region is difficult to assess and is strongly dependent on the specific location within a region and the current land use. Therefore, the land cost transfer index, specifically used with the capital costs for the pipeline for CO_2 transportation, was held constant and therefore assigned a transfer index of 1.00. The regional indices developed using these methods are listed in Table 3-3.

Table 3-3 Regional indices used to transfer projects from USGC to specific locations

	Capital and O&M Costs			
Region	**Equipment**	**Materials**	**Labour**	**Land/ROW**
ANZ				
Australia	1.21	1.21	1.58	1.00
Asia				
China	0.81	0.81	0.05	1.00
Japan	1.21	1.41	1.84	1.00
India				
India	1.27	1.11	0.26	1.00
Europe				
Euro Region (Germany)	1.19	1.16	1.33	1.00
Eastern Europe (Poland)	1.01	0.81	0.79	1.00
ME and Africa				
Saudi Arabia	1.27	1.21	0.35	1.00
South Africa	1.27	1.11	1.04	1.00
Americas				
Canada	1.08	1.01	2.16	1.00
United States	1.00	1.00	1.00	1.00
Brazil	1.16	1.16	0.97	1.00

The Imported Equipment and Materials index adds a factor for freight over USGC. Taxes and duties have been specifically excluded from the calculation.

The Locally Sourced Equipment and Material index represents the cost of the locally sourced items (expressed in US dollars) relative to USGC.

The labour index considers two key elements; the relative cost of labour (on a crew rate basis) and the relative labour productivity as compared to USGC.

In general, the equipment related to power projects was considered as an imported cost except in the case for countries with significant industry such as China.

The capital costs for power generation and capture were considered for the respective regions. Based on WorleyParsons experience, the China labour index was significantly greater than that previously reported and resulted in power plant capital costs greater than those typically reported. This higher than expected index is related to the urban locations used within the Richardson's manual. Generally, the setting of most power plants is in rural areas, and as a result the labour index for China was reduced to a value more inline with the previous report of 0.05.

The reader is cautioned that the calculated indices are based on very specific locations/cities, generally major metropolitan areas, and can vary considerably, particularly with respect to labour, depending on the actual project location. Additionally, the type of labour used; non-union, union, or work camps, can greatly impact the labour costs within a given region. Further, the locally sourced equipment and materials and labour indices will vary with changes in currency exchange rates.

3.4.2 Regional coal type and fuel costs

In addition to the update of the regional factors, typical coal types for each of the regions were selected to be utilised in the economic analysis. The coal selections were based on the information provided in Projected Costs of Generating Electricity (IEA, 2010) and WorleyParsons' experience. In general, obtaining the price paid by a generator for fuel is difficult and where this data is available, is found to vary significantly between facilities. The coal pricing approach for the regions was adopted from the Projected Costs of Generating Electricity in that for countries which produce coals, local coal prices were assumed while for non-coal producing countries international market prices were assumed. The local coal prices were determined through the review of the literature such as national energy surveys, the Projected Costs of Generating Electricity, and in the case of the United States, a database of delivered fuel costs. The regional natural gas prices were determined through similar methods. A summary of the coal types and coal and natural gas prices used in this study are provided in Table 3-4.

3 METHODOLOGY FOR CCS ECONOMIC ANALYSIS UPDATE (CONTINUED)

Table 3-4 Coal types and prices used for coal and natural gas per region

Region	Coal	Coal Heating Value, HHV	Coal price (US$/GJ)	Natural gas price (US$/GJ)
ANZ				
Australia	Australian Black	24.8	1.45[1]	7.58[5]
Asia				
China	China Bituminous	24.1	3.58[2]	4.53[5]
Japan	Australian Export, Bituminous	33.3	4.95[3]	11.09[5]
India				
India	High Ash Indian Bituminous	12.6	4.44[4]	4.20
Europe				
Euro Region (Germany)	German Brown	10.5	1.37[5]	9.76[5]
Eastern Europe (Poland)	Polish Brown	10.1	0.67[6]	9.76[5]
ME and Africa				
Saudi Arabia	NA	NA	NA	2.00
South Africa	South African Coal	18.2	0.82[5]	9.76[5]
Americas				
Canada	Canadian Sub-bituminous	28.9	0.91[7]	7.40[5]
United States	Bituminous IL No #6	27.1	2.61[8]	7.40[5]
Brazil	High Ash Brazilian Bituminous	10.4	1.85[5]	7.71[5]

Notes:

[1] Average of black coal prices, 'Fuel resource, new entry and generation costs in the NEM,' (ACIL Tasman, 2009).

[2] Coal price assumed in the IEA publication in $/tonne, compares to average fuel price presented in paper, 'CO$_2$ capture from coal-fired power plants in China', (NZEC, 2009) of $3.50/GJ. Comparable to price reported by WorleyParsons China Office of $85-100/tonne.

[3] Import price of coal to Japan in Q1- 2010.

[4] Midrange of delivered coal price (range $48/t to $64/t) to power plants from 'Coal Initiative Reports, A Resource and Technology Assessment of Coal Utilization in India' (Chikkatur, 2008).

[5] Coal and Natural Gas price assumed in 'Projected Costs of Generating Electricity, 2010 Edition' (IEA, 2010).

[6] Data from WorleyParsons based on work in Eastern Europe.

[7] EIA data, Low price is reasonable based on the vertical integration and coal fired generation in Canada; see Natural Resources of Canada, www.nrca.gc.ca

[8] EIA average Form 923 (Power Plant Operations Report, delivered coal prices to power plants, EIA 2009e) data for 2009.

3.5 Power generation, CO$_2$ capture and coal type

The coal type used for power generation and its characteristics can vary significantly between locations and impact the design of the power generation facility. Typically in selecting the coal to be used as a fuel, economic analyses are performed based on a specific location, the available coals and their cost. To account for the potential variation in coal types between regions, correlations were developed between a factor to relate the capital costs, plant heat rate, and O&M costs to the coal rank and facility designs. These factors were used to adjust the reference plant operating parameters and costs, so that they would be representative of a typical coal used in the regions.

The following paragraphs provide a description of the methodology used to derive the correlations between performance and cost implications on various power generation technologies and on a variety of fired coals. The power generation technologies include supercritical (SC) and ultra-supercritical (USC) PC power generation with/without CO$_2$ capture, IGCC with/without CO$_2$ capture and SC and USC oxyfuel combustion power generation.

3.5.1 General methodology

To obtain the correlations, plant performance and cost data for different coals were extracted from various publications and the WorleyParsons database. The data was then screened and analysed. The data was selected for producing the correlations only when it was from a single study and covered a relatively wide range of coals, typically one or two bituminous coals, one sub-bituminous coal and/or a lignite coal. Selecting studies that consider multiple coal types at the same location guarantees the data is on the same basis. The factor correlations were developed by considering the ratio of a parameter, that is capital cost, at different coal heating values, to a reference coal heating value of 30.8MJ/kg, the heating value of Pittsburgh No. 8 coal. Extrapolation or interpolation was applied in some cases when necessary. The best fit curves and correlation equations were found through selected data series using regression analysis techniques in Excel (Trendline function in Chart).

The correlation equations present the percentage change in plant performance or capital cost as a function of coal heating value from the reference coal heating value. The higher heating value (HHV) was used throughout this work. It is acknowledged that coal constituents have impacts on plant performance and cost to different degrees of extent. However, only HHV is used to represent coal categories in this study.

If multiple data sets from different sources were available, the correlation equation/curves were based on all data, which basically averages the selected parameters and hence reflects generic correlations. If the correlation equations were developed based on a single data set due to limited information, the correlations reflect the conditions defined in referred studies. In such case only the studies with relatively generic design conditions were chosen.

3.5.2 Assumptions

The following assumptions were also made to derive the correlations between performance and cost implications on various power generation technologies and on a variety of fired coals.

General

a) Correlation equations are developed by data fitting technique if multiple data sets are available, which basically averages the subject parameters and hence reflects only generic trend of coal heating value. If the correlation equations are developed by single data set due to limited info availability, the equations will only reflect the data based on specific conditions of the referred study.

b) All the technical and cost data is based on studies in US region. It should be acknowledged that cost varies with the plant site, significantly sometimes. The studies are either public or from WorleyParsons database.

c) The data collected is based on the consistent system configurations for US region, and does not intend to reflect specific region or country’s pollutant emission requirements and other constraints such as water.

d) PC plant data is based on SC and USC plant. The IGCC plant data is based on Shell Technology only considering it is relatively abundant data for different coal including low rank coal.

3 METHODOLOGY FOR CCS ECONOMIC ANALYSIS UPDATE (CONTINUED)

PC case

a) All the SC and USC PC plants analysed with or without CO_2 capture are equipped with flue gas desulphurisation (FGD) for sulphur reduction and selective catalytic reduction (SCR) for NOx reduction from the stack.

b) The SC or USC PC plant performance curve does not necessarily reflect the impact of sulphur content in coal on SC steam condition. The PC heat rate curves are based on consistent steam condition for the range of coal.

c) All cases with CO_2 capture efficiencies are assumed to be approximately 90 per cent.

d) The SC/USC PC plant performance (heat rate) of each data set is based on new plants specifically designed for different coals.

e) The SC PC plant capital cost is based on sites in US. It should be acknowledged that cost varies with the plant site, dramatically sometimes.

f) PC FGD option data was based on FGD using limestone or lime in US corresponding to coals.

IGCC case

a) While specific component selection has an impact on IGCC system performance (for example an IGCC choosing Selexol as acid gas removal (AGR) system has slightly different performance from one choosing Amine system), the performance curve intends to reflect such differences.

b) IGCC with or without CO_2 capture heat rate comparisons are based on Shell technology (dry feed). It should be noted that for other technologies, the comparison result may change to different degrees of extent.

Oxyfuel PC plant

a) Based on SC/USC pulverised coal power plants (24.1MPa/599°C/621°C, 3500 psig/1110°F/1150°F)

b) The plants are equipped with low NOx burner (LNB)/overfire oxygen (OFO)/flue gas recirculation (FGR) and FGD

c) Oxygen (95 per cent pure) used for combustion is produced from cryogenic air separation unit (ASU) system.

d) CO_2 specification is for storage in saline formation (raw combustion product produced using 95 per cent oxygen and dehydrated to 0.015 per cent (by volume) H2O).

e) Assume the efficiency difference for an oxyfuel boiler firing different coal is similar to the difference for an air fired boiler firing different coal while the oxyfuel boiler efficiency has slightly higher efficiency.

f) CO_2 compression uses integrally geared, multistage centrifugal with indirect water intercoolers.

FGD option factor

a) FGD performance and cost comparisons are based on a SC PC plant firing Illinois #6 bituminous coal with a wet limestone forced oxidation positive pressure absorber.

Cooling options factor

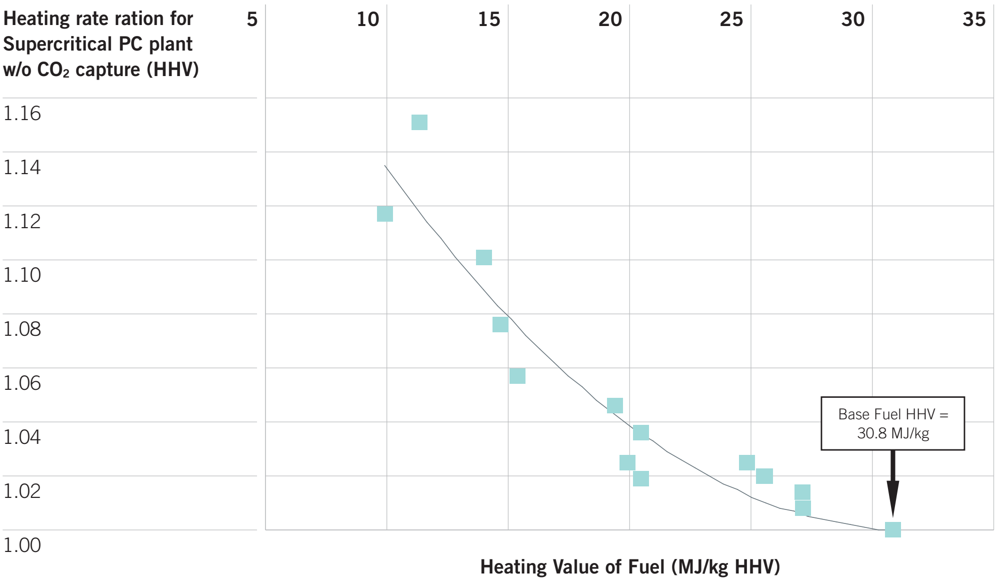
a) The factor is based on mechanical draft wet cooling options and 100 per cent dry cooling.

b) Comparison is assumed on average annual ambient temperature of about 16°C (annual average). The performance impact is of average corresponding to the ambient. It should be noted that ambient temperature has significant impact on the performance when the plant switches between wet and dry cooling options at high temperature, for instance 0°C.

3.5.3 Correlations developed

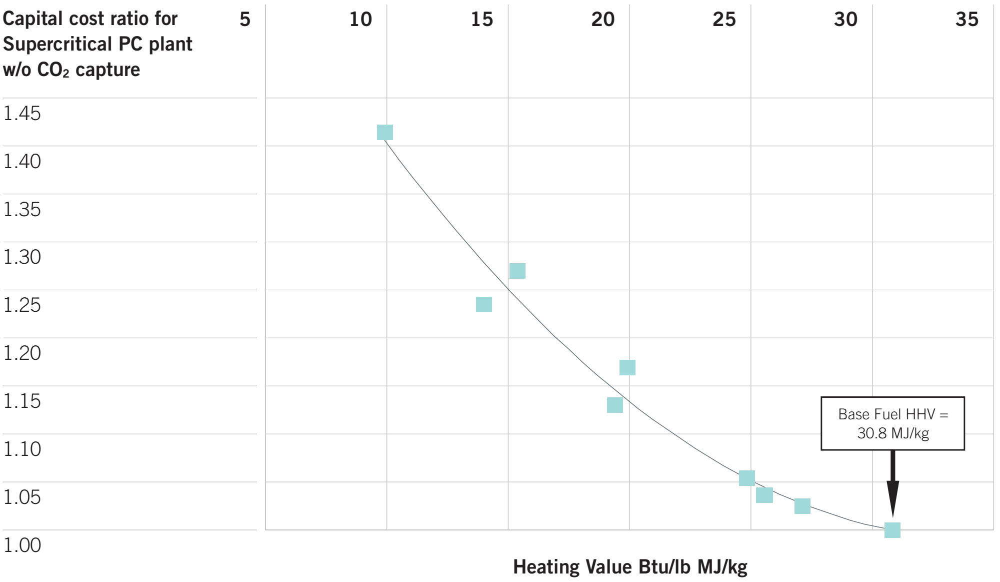
To facilitate presentation, the correlations were given in the form of a ratio of studied plant performance/cost to a referred plant performance/cost as a function of heating value difference between studied coal and the referred coal heating value. The performance and cost of plants firing bituminous coal, i.e. Pittsburgh No. 8 coal, was used as the referred data point. Correlation diagrams between net plant heat rate (HHV) and coal heating value, and plant capital costs and coal heating value for a SC PC plant without CO_2 capture are illustrated in Figure 3-2 and Figure 3-3.

Figure 3-2 Heat rate factors as a function of coal heating value



3 METHODOLOGY FOR CCS ECONOMIC ANALYSIS UPDATE (CONTINUED)

Figure 3-3 Capital cost factors as a function of coal heating value



Specifics regarding the data sources and specifics of the analyses for the various technologies are provided in the following sub-sections.

PC power plant

Correlations between performance/capital cost and coal heating values for SC-PC power plant without CO_2 capture were derived using data from several public studies and a confidential WorleyParsons study (Holt, N & Booras, G 2007; WorleyParsons 2002; EPA 2006; WorleyParsons 2009a; DOE/NETL 2007; WorleyParsons 2009b). Performance and capital cost correlations for SC-PC power plant with CO_2 capture were developed using data from various reports (WorleyParsons 2009a; DOE/NETL 2007; WP 2009b). Information were applied in correlations between coals and O&M cost for SC PC power plants with and without CO_2 capture (DOE/NETL 2007; WorleyParsons 2009b).

For the case of heat rate correlation of USC-PC without CO_2 capture, data from a WorleyParsons report (WorleyParsons 2009a) was used. Due to the sparse information the correlations for USC power plant capital cost, O&M cost, the data for SC PC plants were adopted. USC-PC plants have different capital and O&M costs from SC-PC plants. However using the ratio of the cost mitigates the impact to some extent.

IGCC power plant

Correlations for IGCC power plant were derived based on Shell coal gasification technology because of its suitability to relative wide range of coals and good information availability.

Data from several sources were used to develop the correlations for IGCC plant with and without CO_2 capture (WorleyParsons 2009a; DOE/NETL 2007; WorleyParsons 2009b; Maurstad et al 2006).

Oxycombustion power plant

Oxycombustion technology used for power generation is still in development. There is currently little published information on the impact of different coals on the plant performance and costs. However, many studies for power generation utilising oxycombustion boiler for single specific coal have been done, and some studies from technology vendors indicate that a oxycombustion PC boiler has similar or slightly higher efficiency as a conventional PC boiler, and the oxycombustion PC power plant essentially has the same system and equipment as the conventional PC power plant except the boiler island (US DOE 2008a; DOE/NETL 2006; DOE/NETL 2001). With these considerations in mind, the performance and cost for an oxycombustion power plant were estimated based on the performances and costs of an SC oxycombustion power plant based on a specific coal (US DOE 2008a) and a conventional SC PC power plant firing different coals (DOE/NETL 2007; WorleyParsons 2009b). The correlations for SC and USC PC oxycombustion power plants then were derived from these estimated performance and costs.

Flue gas desulphurisation and wet/dry cooling design options

The impact of a SC PC power plant with or without a FGD on cost and performance was estimated. The derived factor (percentage) was based on information from the 'Cost and Performance Baseline for Fossil Energy Plants' (DOE/NETL 2007). The study utilising wet limestone forced oxidation type FGD.

The impact for switching wet cooling to drying cooling on the performance and cost of PC and IGCC power plants were estimated based on information from various reports (DOE/NETL 2007; WorleyParsons 2007; Power, W.E 2005; Power, W.E 2009). It should be noted that the design ambient temperature may have significant impact on the performance and cost of a power plant. In this study the data was based on design ambient temperature of 60-65ºF, which is considered to be an annual average temperature.

3.5.4 Limitations of approach

The above approach is not intended to provide for detailed accounting of the coal type and process design requirements but rather a rapid method of assessing the impact of these parameters in an economic model. As illustrated these correlations are strictly empirical and should be treated as such. Limitations of this approach include:

- not considering variation in process designs which account for many interacting case specific factors;
- trends utilise coal heat heating as the primary driver for cost and heat rate changes, coal characteristics leading to coal heat rate are not necessarily considered; and
- impact of high coal ash content and ash composition/properties not included.

3.6 CO_2 capture from industrial processes

There were minimal changes made to the method in which the CO_2 capture cases for industrial processes were analysed. The capital and operating costs were updated from the 2009 figures, which resulted in an increase of approximately 1 per cent.

As with the initial economic assessment in 2009, the 2010 CCS economic analysis was applied to select industrial processes to determine the cost parameters for blast furnace steel production, cement production, natural gas processing, and fertiliser (ammonia) production. The specific emission intensities, typical production rates, annual CO_2 flow rates, and other parameters from current commercial practice were used as a basis (where available) and are listed in Table 3-5. Steel and cement production require both capture and compression while natural gas processing and fertiliser production are processes that require CO_2 separation from a gas stream. Thus, these processes only need the addition of drying and CO_2 compression.

Table 3-5 Emissions intensities, processing parameters, and commercial application examples for industrial processes

	Blast Furnace	Cement	Natural Gas Processing	Fertiliser Production
CO_2 emission intensity	2tonne CO_2/ tonne steel	0.83tonne CO_2/ tonne cement	0.13kg CO_2/Sm^3 natural gas	0.57tonne CO_2/ tonne ammonia
Product production rate	40steel tonne/hour	40cement tonne/hour	1,100Sm^3/hour 46GJ/hour	46tonne ammonia/hour
Annual CO_2 flow of single plant (tonne/year)	706,000	296,000	1,000,000	194,000
Commercial example	No commercial operating facility	No commercial operating facility	Sleipner	Coffeyville gasification facility

The energy for capture (primarily solvent regeneration) and CO_2 compression can be obtained through heat or electricity generated by the system or through purchased power. This uncertainty can lead to difficulties in the interpretation of the analyses. In this assessment, the following assumptions were made.

- The cost of the energy to meet the auxiliary load was determined from the natural gas cost in the region with 50 per cent conversion efficiency.
- The CO_2 from the generation of heat or electricity used by the capture system was not included in the calculation.

The CO_2 flows from the individual industrial processes, less than 1Mt per year of CO_2, are smaller than those for standard power generation plants. Therefore, it was assumed that a multi-user pipeline and storage option would be available. For storage and transportation, costs of US$4/tonne of CO_2 and US$7/tonne of CO_2 were used, respectively. The basis for the cost is discussed in the following sections.

3.7 CO_2 transport

Report One from the 2009 Strategic Analysis work identified CCS projects around the world that are in operation or under development. The study also identified a subset of CCS projects that were classified as large-scale integrated CCS projects (LSIPs). Examination of this subset of LSIPs as part of the economic assessment update found that the average length of pipelines for the LSIPs was in the order of 100km. To reflect the activities occurring in the development of CCS projects globally, this distance for an onshore pipeline is used is estimating transport costs. This differs from the reference case assumption made in the 2009 economic assessment of 250km. The other process assumptions for CO_2 transport including inlet temperature and pressures remained unchanged.

3.8 CO_2 storage

As the 2009 study showed, the economics of CO_2 storage is dependent upon the geology of the target formation. The geology will drive the storage site selection and the site will drive the commerciality of commercial scale, integrated CCS projects. In this update, the “finding” costs applied are assumed to range from US$25 million in the ideal case to US$150 million or more depending on the geology. This is consistent with the 2009 study.

Given the high variability of geologic properties within a region, let alone across nations, the assumption was made that the 'reservoir' properties were identical across all the jurisdictions for which recent market costs were available (North America, Europe and Australia/New Zealand). This allowed for a direct comparison of potential storage costs over these regions and an understanding of what the key capital and operating costs are for each region. It was also assumed that all storage projects occur onshore, an assumption that avoids the highly variable costs of drilling offshore. A further assumption is that no EOR is possible as a means of revenue offsets for CO_2 storage. Thus, the storage scenarios allow for 'reasonable' comparisons. In cases where storage costs for CO_2 were not available, the costs of hydrocarbons exploration programs were used as analogues.

Furthermore, in order to treat the inherent uncertainty of storage, two case study scenarios were considered. The two cases were for a 'good' reservoir and a 'poorer' reservoir with either 3Mtpa or 12Mtpa injection scenarios modeled. These are presented in Table 3-6. In this case, the primary differences between the two reservoirs are the absolute permeability and the reservoir thickness.

Table 3-6 Geological and well properties for 'poorer reservoir' and 'good reservoir'

	Units	'Poor reservoir'	'Good reservoir'
Net thickness	m	5.0	15.0
Absolute permeability	md	150	400
Reservoir mid-depth	m	1700	1700
Initial pore gradient	bar/m	0.1002	0.1002
Temperature gradient	°C/100m	3.0	3.0
Surface temperature	°C	20	20
Fracture gradient	bar/m	0.136	0.136
Initial pore pressure	bar	170	170
Reservoir temperature	°C	71	71
Injection pressure limit (of FG)	–	90%	90%
Initial injection pressure	bar	208	208
Relative permeability	–	0.3	0.3
Drainage radius	m	762	762
Wellbore – radius	m	0.09	0.09
Total skin	–	2.5	2.5

For both these cases, the well counts at the start of the injection period and end of the injection period were calculated using a number of factors to control well count. It was assumed that no heating of the CO_2 was required upon delivery to the injection site (this may be a simplification which would require both power and compressors at the wellhead thus increasing capital (CAPEX) and operations and maintenance (OPEX) costs) and that the injection pressure was always maintained at 90 per cent of the fracture gradient thus reducing operational risk from injection induced fracturing. The drainage radius assumes that the CO_2 forms a circular pool ('pancake') at the injection point (that is, that the reservoir is wholly homogenous throughout all the injection points).

3 METHODOLOGY FOR CCS ECONOMIC ANALYSIS UPDATE (CONTINUED)

The analysis assumed that a suitable reservoir was located and characterised and that injection could commence once the capture plant was online. It was also assumed that the exploration and appraisal program that would start in late 2010 to prove up these sites (and others as a potential portfolio of storage options) was successful and that the locations would have been shown to have suitable seals and other geological factors that would allow for safe storage of CO_2 over the longer term. Thus, the analysis assumes that the developmental timeline for the storage site starts at the FEED phase to allow injection to commence from 2016 onwards. In both cases, ‘finding costs’ are included in the site selection and. Vertical injection wells are used in both cases as the standard. The evaluation did not consider stimulation by fracturing or deviated well placement that could increase injection rates. Such well design considerations would be considered after the site characterisation and the full field development program are complete.

This analysis required that the injection well pressure was below the fracture pressure (90 per cent of fracture pressure) which provides the first order constraint on potential storage volumes. CO_2 injection above the fracture pressure could increase well injectivity rates, but at the risk of fracturing the cap-rock. Thus, an accurate determination of the actual fracture pressure can only be derived from laboratory tests on new core samples from both areas. Another assumption is that sufficient pore space is available over the life of the injection operations.

This analysis required that the injection well pressure was below the fracture pressure (90 per cent of fracture pressure) which provides the first order constraint on potential storage volumes. CO_2 injection above the fracture pressure could increase well injectivity rates, but at the risk of fracturing the cap-rock. Thus, an accurate determination of the actual fracture pressure can only be derived from laboratory tests on new core samples from both areas. Another assumption is that sufficient pore space is available over the life of the injection operations.

As previously discussed in the 2009 Foundation Report Two, the key factor determining the cost of CO_2 storage is the ‘reservoir permeability thickness product’ ($k \cdot h$) since this determines the injectivity of the site and thus controls the well count needed for a given volume of CO_2.

As can be seen in Table 3-7 below, the final well counts required for each reservoir for both injection scenarios are as follows.

Periodically new wells will need to be brought online in the poor reservoir to allow for the constant injection volume to be maintained on an annual basis. This is due to the ‘back pressure’ exerted by the reservoir as it fills with CO_2, meaning that the original injection volume on the initial injector wells will decrease over time, thus necessitating new wells for constant volumetric injection.

Table 3-7 Well counts for ‘Poor reservoir’ and ‘Good reservoir’

Reservoir quality injection volume	3Mtpa	12Mtpa
‘Poor reservoir’	16	61
‘Good reservoir’	2	8

Regional cost comparison

Recent data on the costs for CO_2 storage were available from Australia/New Zealand, Europe, and North America for CO_2 injection wells and associated services. These are listed in Table 3-8. Whilst proprietary data was available for China, confidentiality issues prevented the release of this information. As a result, Australia and New Zealand (ANZ) was used as a proxy for this region.

Table 3-8 Regional costs for CO_2 storage

Regional Costs (in 2010 US$)		ANZ	Europe	US
3D Seismic survey	mln US$/svy	18	25	18
Deep monitoring well costs	mln US$/well	6	6	5
Shallow monitoring well costs	mln US$/well	0.50	0.70	1
Injection well costs	mln US$/well	7	7	10
Injection well abandonment costs & rehab	mln US$/well	1	0.70	1
Monitoring well abandonment costs & rehab	mln US$/well	0.50	0.70	0.5
In-field flow lines	mln US$/well	0.24	0.35	0.25
Drilling cost escalation per annum	%	7 %	5 %	5 %
Well-related OPEX	% of DrillEX	5 %	5 %	5 %
Monitoring OPEX	mln $/yr	0.10	0.14	0.10
Fees & Rents OPEX	mln $/yr	0.10	0.14	0.10

There is little headline difference in the well costs between each of the regions. This recognises that the costs estimates for the complete study is in the order of +/- 40 per cent and indeed the similarity of the costs can be taken as an indication that the market for well services/rigs is worldwide. Thus, companies will seek to move staff and assets to areas of demand.

3.9 Summary of process and economic modelling assumptions

The default assumptions used for the reference CCS system during the assessments of the various technologies are listed in Table 3-9. As described in the previous sections within this chapter, primary changes from the initial report include revised parameters for the storage reservoir properties and the decrease of the pipeline length from 250km to 100km.

3 METHODOLOGY FOR CCS ECONOMIC ANALYSIS UPDATE (CONTINUED)

Table 3-9 Process modelling and financial assumptions for reference case

Capital cost estimate basis:			
	Constant 2010 US dollars (2010 US$)		
	The reference location is the US Gulf Coast.		
	Labour is based on non-union rates.		
CO_2 capture			
	Power generation: results listed for supercritical, oxy-combustion, IGCC, and NGCC technologies, system design parameters listed in Table 3-1 and Table 3-2.		
	Industrial processes: natural gas processing, cement production, blast furnace production of steel and fertiliser (ammonia) production.		
Transportation: pipeline			
	CO_2 pipeline length:	100km	
	CO_2 pipeline inlet temperature:	25°C	
	CO_2 pipeline inlet pressure:	20.2MPa	
	CO_2 pipeline outlet pressure:	15.3MPa	
Storage: saline aquifer (reference case was for 'good reservoir')			
	Reservoir pressure	17MPa	
	Reservoir thickness	15m	
	Reservoir depth	1,700m	
	Reservoir absolute permeability	400mD	
	Site screening and evaluation	US$66,000,000	
Financial: general			
	Levelisation period	30 years	
	Owners' costs	15%	
	AFUDC rate	9%	
	Labour multiplier	1	
	Coal cost	2.61US$/GJ	
	Natural gas cost	7.40US$/GJ	
Financial: cost of capital		Per cent	Rate
	Debt	40%	6%
	Equity	60%	12%
	Tax rate for debt interest deduction	33%	
Financial: escalation rates			
	Fixed O&M cost escalation rate	0%	
	Variable O&M cost escalation rate	0%	
	Real CO_2 emissions escalation rate	0%	
	Real fuel escalation rate	3%	

4 RESULTS AND COMPARATIVE SENSITIVITY ANALYSIS

An economic comparison of projects can be performed using the methodology and templates reviewed and developed in Chapters Two and Three, respectively of Foundation Report Two. Additionally, sensitivity analyses are also included in this evaluation to determine the range of results and their dependence on key cost components and design assumptions.

The assumptions used for the reference CCS system during the assessments of the various technologies are listed in Table 3-9. The following sections provide an update of the results.

4.1 Capture of CO_2

The methodology developed was applied to power generation and select industrial processes to determine the levelised production costs (US$/unit of product), cost of CO_2 avoided, and the cost of CO_2 captured.

4.1.1 Power generation

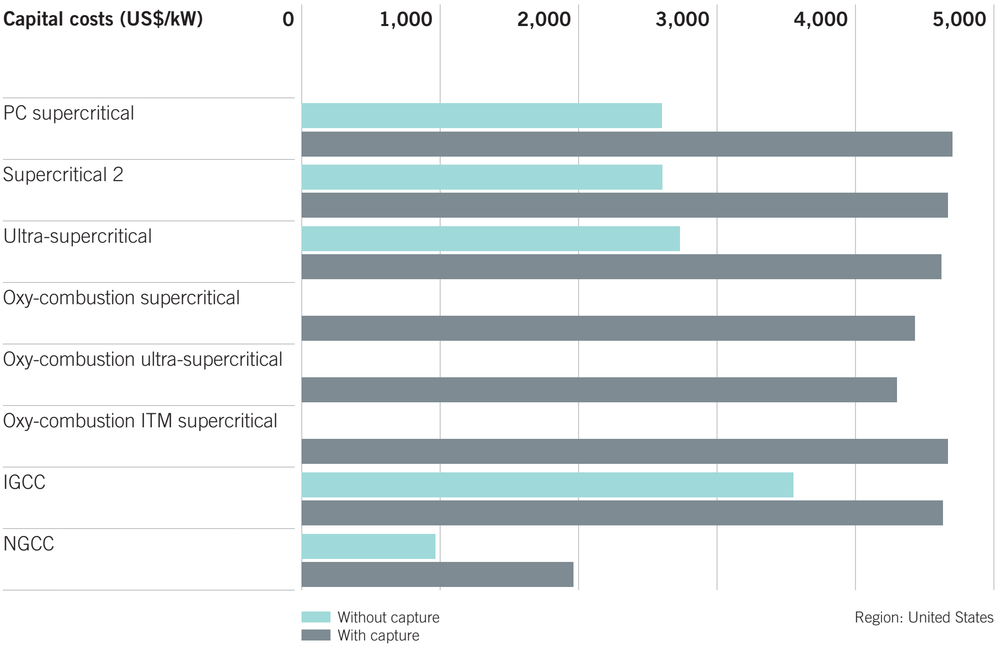
Capital costs

The impact of CCS on the cost of new electric power generation facilities was the first set of cases to be reviewed using the model developed. The parameters for the facilities are listed in Table 3-1 and Table 3-2, and under the base case assumptions listed in Table 3-9. The capital costs, including the owners' costs, for the CO_2 capture portion of the CCS system are compared on a $/kW basis in Figure 4-1 for first-of-a-kind (FOAK) systems (note that transportation and storage cost components are not included). FOAK systems are innovative and inherently involve greater uncertainty and therefore costs. Experience gained from these plants (such as design and operation) will generate learnings which will reduce the uncertainty and hence the costs. As these experiences increase, the cost reductions (for example, through process optimisation) result in nth-of-a-kind (NOAK) plant estimates.

For the non-CO_2 capture of PC supercritical and ultra-supercritical cases, there is a small cost increase related to the material requirements for the higher steam conditions. This capital cost is potentially recovered through lower fuel costs related to the greater efficiency of the ultra-supercritical unit. The selection of plant efficiency, as determined by the steam conditions, must consider a balance between the capital and fuel costs. Based on the small change in the overnight capital costs at the reference location, there is only a small increase in these values compared to the 2009 report.

4 RESULTS AND COMPARATIVE SENSITIVITY ANALYSIS (CONTINUED)

Figure 4-1 Installed costs for 550MW net generation and CO_2 capture facility (FOAK)



Levelised cost of electricity

Capital costs presented above, along with transportation and storage capital costs (based on matching the captured CO_2 flows), were used as inputs to the economic assessments. The LCOEs for FOAK and NOAK plants are presented in tabular form in Table 4-1 and graphically in Figure 4-2.

The cost parameters calculated from the LCOEs, the cost of CO_2 avoided and cost of CO_2 captured (including transportation and storage), in the economic assessment are provided for FOAK and NOAK units in Table 4-2 and shown graphically for FOAK units in Figure 4-3.

The costs of LCOE and CO_2 avoided and captured have decreased slightly since the 2009 report. This is likely to be a reflection of the revised fuel costs for this update.

Table 4-1 Breakdown of LCOE for technologies with and without CCS for FOAK and NOAK plants

	PC supercritical	PC supercritical 2	Ultra-supercritical	Oxy-combustion supercritical	Oxy-combustion ultra-supercritical	Oxy-combustion ITM supercritical	IGCC	NGCC
LCOE without capture ($/MWh)								
Generation	76	76	73	NA	NA	NA	90	88
Transportation	0	0	0	0	0	0	0	0
Storage	0	0	0	0	0	0	0	0
Total	**76**	**76**	**73**	**NA**	**NA**	**NA**	**90**	**78**
LCOE with capture for FOAK ($/MWh)								
Generation and capture	124	121	113	114	107	116	116	115
Transportation	1	1	1	1	1	1	1	1
Storage	6	6	6	6	6	6	6	6
Total	**131**	**129**	**120**	**121**	**114**	**123**	**123**	**123**
LCOE with capture for NOAK ($/MWh)								
Generation and capture	122	119	109	112	104	114	113	114
Transportation	1	1	1	1	1	1	1	1
Storage	6	6	6	6	6	6	6	6
Total	**129**	**127**	**117**	**119**	**112**	**121**	**121**	**121**

Note: The steam conditions of the ultra-supercritical technologies presented are beyond those typical used in the industry today and therefore should be considered emerging technologies. The initial deployment of this technology will be strongly dependent on the specific site and available coal characteristics. Therefore, in the near term, the PC supercritical should be considered as the reference case.

Table 4-2 Cost of CO_2 avoided and captured for FOAK and NOAK plants

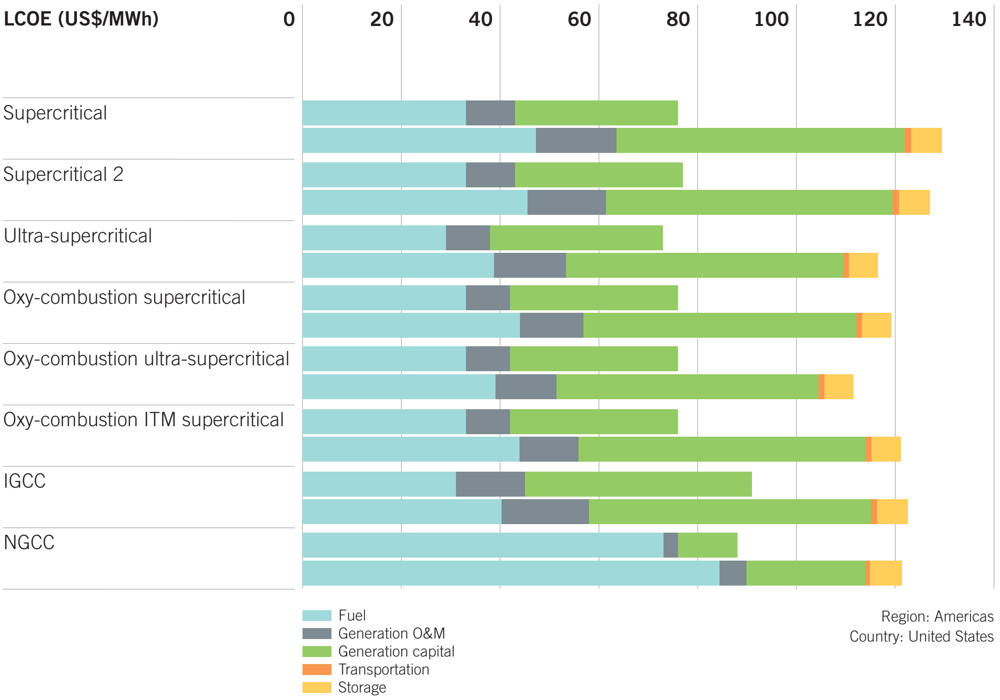
	PC supercritical	Ultra-supercritical*	Oxy-combustion supercritical	Oxy-combustion ultra-supercritical	Oxy-combustion ITM supercritical	IGCC	NGCC
FOAK							
Cost of CO_2 avoided ($/tonne CO_2)	81	62	57	47	59	67	107
Cost of CO_2 captured ($/tonne CO_2)	53	55	42	43	47	39	90
NOAK							
Cost of CO_2 avoided ($/tonne CO_2)	78	57	54	44	57	63	103
Cost of CO_2 captured ($/tonne CO_2)	52	52	41	42	45	38	87

Notes: In the cost of CO_2 avoided calculations, for the coal-fired power generation, SC-PC without CCS was used as the reference case and for the NGGC without CCS was used as the reference case for NGCC with CCS. In select previous studies (such as DOE/NETL 2007) , the cost of CO_2 avoided has been calculated with the reference plant selected as the similar technology without CCS. For IGCC, under this assumption, the FOAK and NOAK costs of CO_2 avoided are $51/tonne and $50/tonne, respectively.

* With the assumption that ultra-supercritical technology, with the designed steam conditions, is available for CCS, this becomes a low cost option for the reference facility and thus is used for the reference plant in the cost of CO_2 avoided calculation.

4 RESULTS AND COMPARATIVE SENSITIVITY ANALYSIS (CONTINUED)

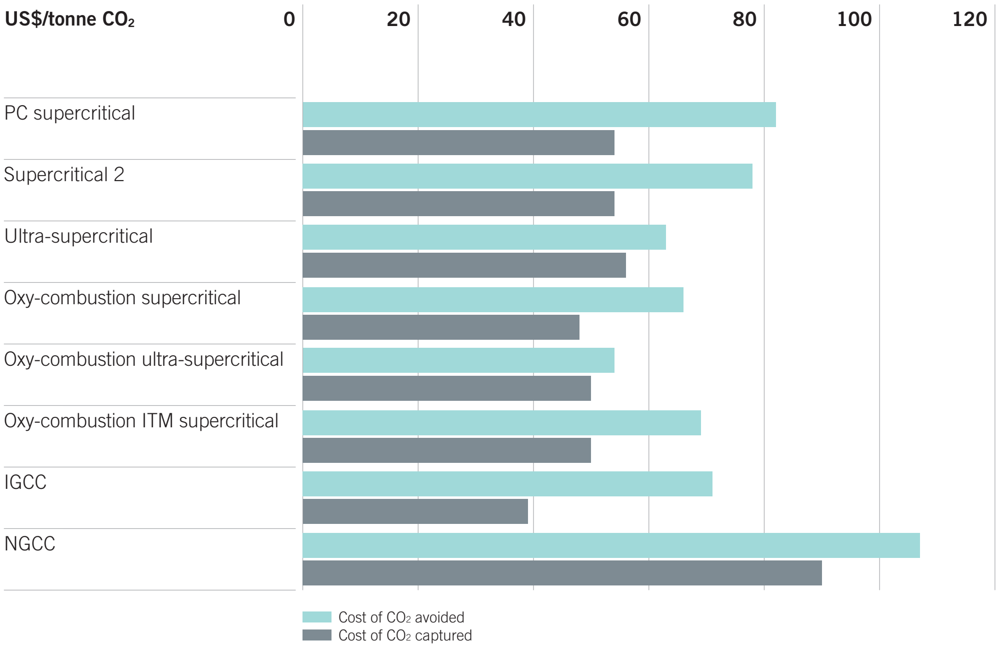
Figure 4-2 Comparison of LCOE for reference generation, with and without CCS in USGC



Region: Americas
Country: United States

Figure 4-2 indicates varied results from the 2009 analysis, with some LCOE values increasing and some decreasing. All of the coal fired technologies show a decrease in fuel costs related to the lower coal cost used in this analysis. There is an increase in the capital contribution to the LCOE for oxy-combustion with CCS which reflects the inclusion of an additional purification process when capturing CO_2 as discussed in section 3.3. Figure 4-2 also shows the relatively small contribution that CO_2 transport has to the overall cost of CCS. Capture represents the greatest cost to CCS, and the majority of the cost increases have been due to changes in the capture system. The reduction in the length of the pipeline for the reference case has also reduced the contribution of transport cost to the overall cost of CCS to the power generation applications.

Figure 4-3 Comparison of CO_2 costs avoided and captured for power generation (FOAK)



First-of-a-kind versus Nth-of-a-kind

As introduced in Foundation Report Two in 2009, it can be expected that costs for CCS will decrease in the future as the implementation of CCS progresses along the experience curve.

There does not appear to be a great benefit with further process maturity for one CCS equipped technology over another, as indicated in Table 4-1 and Table 4-2 above. The potential cost reduction from FOAK to NOAK plants is less than 5 per cent. The reason for this small decrease is that the majority of the capital costs are associated with proven and (by and large) commercially available technologies. Therefore, decreasing the cost of risk with the new technologies does not provide the potential for future cost savings through increasing maturity.

The cost reductions presented here represent decreased risk in the existing technologies and do not consider other improvements such as implementing new technologies for capture or economies of scale savings in transportation and storage.

Introduction of emerging technologies improvements

The introduction of emerging technologies has great potential to reduce the CCS costs (Rubin 2007, EPRI 2008, US DOE 2009). Potential cost saving of emerging experienced gained and the introduction of new technologies is illustrated in Table 4-3.

Table 4-3 Reduction in LCOE after deployment of 100GW capacity

Technology	Average decrease (%)	Range (%)
PC Plant	14.4	6.2 to 21.3
Oxy-combustion	9.7	3.9 to 15.4
IGCC	17.6	7.7 to 25.8
NGCC	15.5	3.3 to 22.0

The ultra-supercritical, as applied to the PC-supercritical and Oxy-combustion and the ion transfer membrane (ITM) technology as applied to the oxy-combustion technology illustrate the value of the introduction of emerging technologies to reducing the costs of CCS.

Efficiency improvements

Ultra-supercritical technology provides an efficiency improvement over supercritical combustion. Ultra-supercritical power plants achieve higher efficiencies through increasing steam temperature and pressure from the boiler. For CCS, cost savings with the ultra-supercritical technology are achieved through:

- decreasing the fuel cost per MWh of generation;
- decreasing the volume of flue gas to be treated per MWh of generation; and/or
- decreasing the amount of CO_2 to be captured and compressed per MWh of generation.

Comparing the supercritical and ultra-supercritical LCOE for the PCC and oxyfuel-combustion cases in Figure 4-2, the benefits of improving generation efficiency is evident. As summarised in Table 4-4, the LCOE, including capture, transport, and storage, decrease by 6 and 9 per cent in moving from the supercritical to ultra-supercritical technologies for the PCC and oxyfuel combustion technologies, respectively. Combined with the cost decreases achieved through lessons learnt, the cost decreases approach the range of those presented by Rubin et al (2007) and are summarised in Table 4-3.

Table 4-4 Percentage change in cost parameters from supercritical to ultra-supercritical technology.

	Percentage change from supercritical to ultra-super critical			
	Post-combustion capture		Oxyfuel-combustion	
	LCOE	Cost of CO_2 captured	LCOE	Cost of CO_2 captured
Without capture	-4%	NA	NA	NA
With capture	-9%	3%	-6%	4%

Ion transfer membrane for oxyfuel combustion

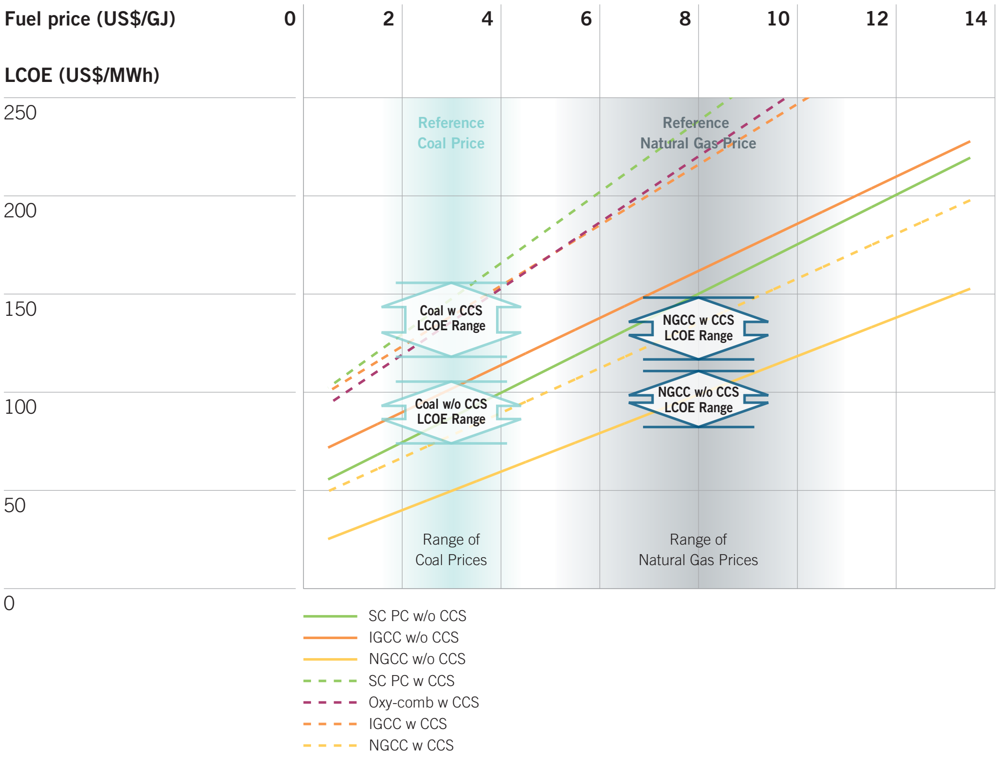
As discussed in Foundation Report Two, ITM is an emerging air separation technology that promises lower auxiliary loads and capital costs compared to cryogenic air separation. ITM presents a potential opportunity to reduce plant capital costs by 10 per cent due to a combination of reduced ASU equipment costs and decreased plant size, based on current status of development. The LCOE could also be reduced by approximately 10 per cent based on the estimations used in this model should this technology be implemented into oxy-combustion systems. This potential reduction is reported in Table 4-3 and is within the range anticipated by Rubin et al (2007).

Sensitivity study: fuel costs

As illustrated in Figure 4-2, fuel cost is a strong contributor to the LCOE. Fuel cost can vary significantly over time due to market conditions, including demand fluctuation, change in environmental regulations, speculation, natural disasters, and international disputes.

The sensitivities of the LCOE, with and without CCS, on fuel costs are illustrated in Figure 4-4. Typical fuel price ranges are included with the resulting LCOE ranges also calculated. For all coal-fired technologies, the coal price range leads to a variation in the LCOE on the order of US$10/MWh, while the natural gas price range leads to a US$30/MWh variation in the LCOE.

Figure 4-4 LCOE as a function of fuel costs



Sensitivity study: Labour costs

Labour costs for the installation and operation of a facility can vary significantly with location. The labour rates used for the base configuration are considered to be the minimum but could increase by 30 per cent within a region due to local labour conditions (that is, union labour rates or competing labour intensive projects), or up to 100 per cent in remote locations requiring labour camps. In some instances, federal government funding programs have as a condition the employment of unionised labour which can affect development costs. The sensitivity of labour to the project costs and economics in this study was the same as applied in 2009. It was investigated through applying a labour cost factor to the labour component, considering both installation and O&M labour, for each CCS component system. This factor ranged from one to two, corresponding to a labour rate increase of 0 to 100 per cent, respectively.

For a supercritical PC with CCS technology, for a fixed fuel cost, the sensitivity of the CO_2 capture installed capital costs and LCOE to the labour costs was reduced. The installed capital costs increased by 23 per cent (32 per cent in 2009), while the LCOE increased by 11 per cent (21 per cent in 2009). A similar trend would be observed for the other coal-fired technologies as they tend to be relatively labour-intensive installations.

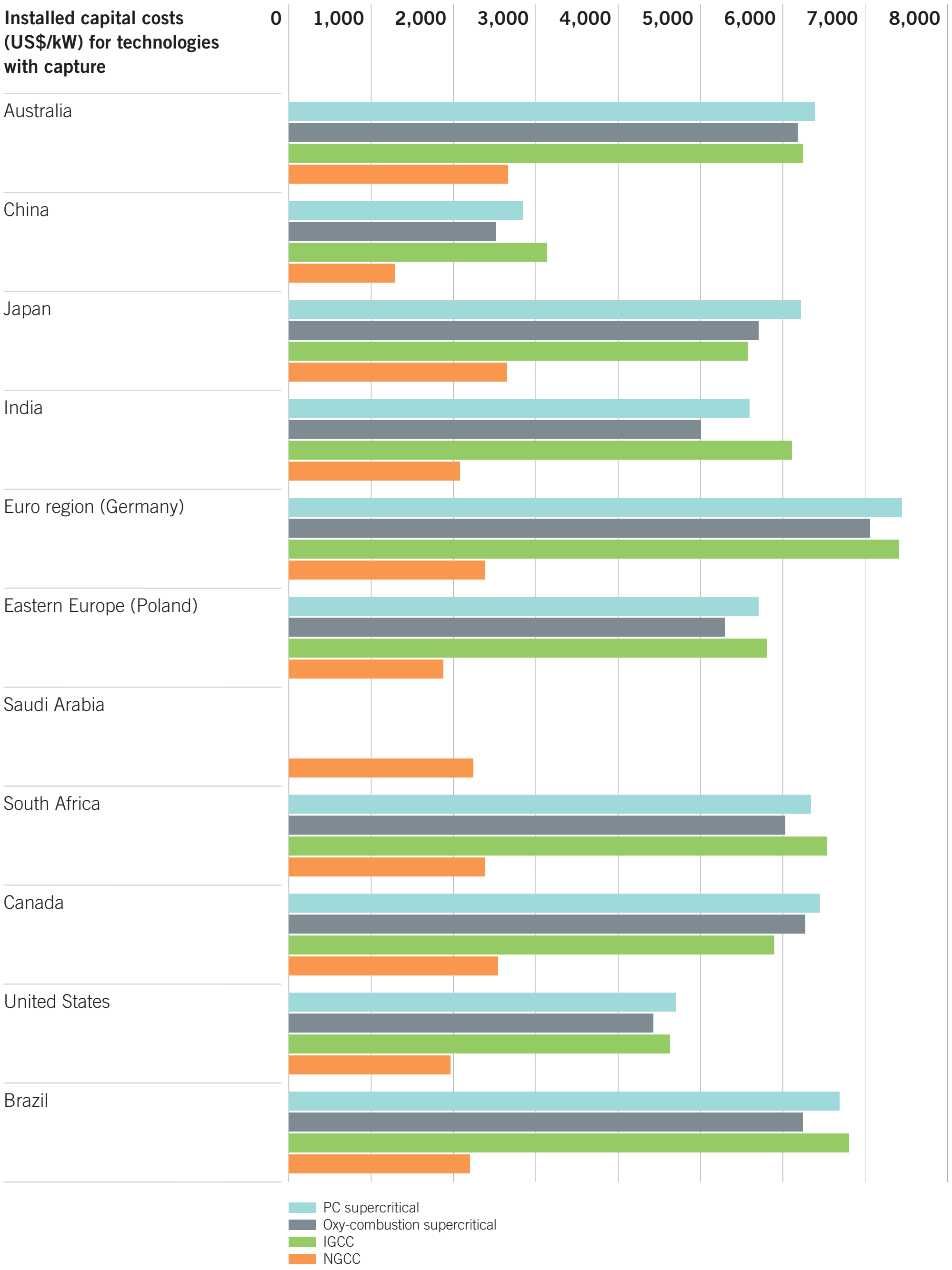
4 RESULTS AND COMPARATIVE SENSITIVITY ANALYSIS (CONTINUED)

Sensitivity study: Regional costs

The installed capital costs of the CO_2 capture technologies for the power generation applications are shown in Figure 4-5, while the LCOE for these applications with CCS are presented in Figure 4-6. A number of key variations since last year emerge, as listed below.

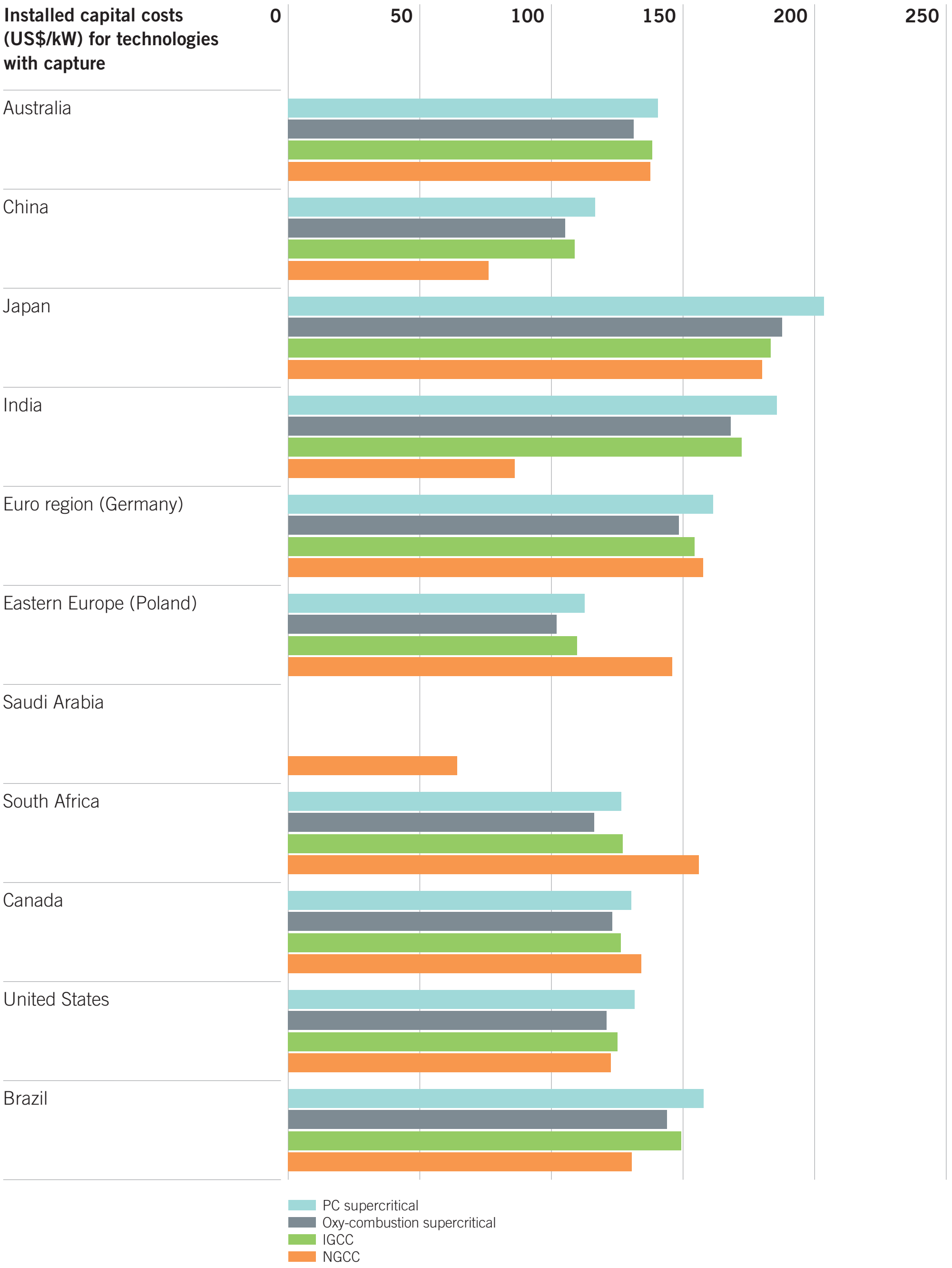
- An increase in the installed CO_2 capture equipment cost and LCOE across all technologies in India. This is due to a 30 per cent increase in equipment being imported into the country as well as India's typical coal heating value being very low, resulting in a greater capital cost.
- An increase in costs across all technologies in Eastern Europe. The increase in the reference coal price for this region has been the major contributor to this change.
- Australia has also experienced a 20 per cent increase in the technology cost which can be accounted for by the higher coal price utilised this year.
- Significant increases in costs (specifically capital costs) are seen in India and Brazil. This is partially because of a lower labour rate being used in 2009. The revision of the coal type to one with a lower heating value also leads to a higher capital cost. Finally, additional costs associated with importing capital equipment have contributed to the increase in CO_2 capture costs India and Brazil.
- Only NGCC costs are displayed for Saudi Arabia, reflecting that there are no coal-fired power generation applications in the region.

Figure 4-5 Installed CO_2 capture equipment cost as function of location



The update of the fuel costs between the regions have increased the LCOE for several of the regions including Australia, China, India and Brazil. In comparing regions, the impact of the variation in fuel costs related to the fuel source and the contract type should be remembered. In many of these regions, such as the United States, Australia, and Europe, there are different coal sources that need to be considered along with the associated variation in price. Additionally, countries may choose to source their coal from international markets in the case where it may provide an economic benefit. Therefore, the cross-regional results presented in Figure 4-6 should be considered along with Figure 4-4 which illustrates the dependence of the LCOE on fuel costs.

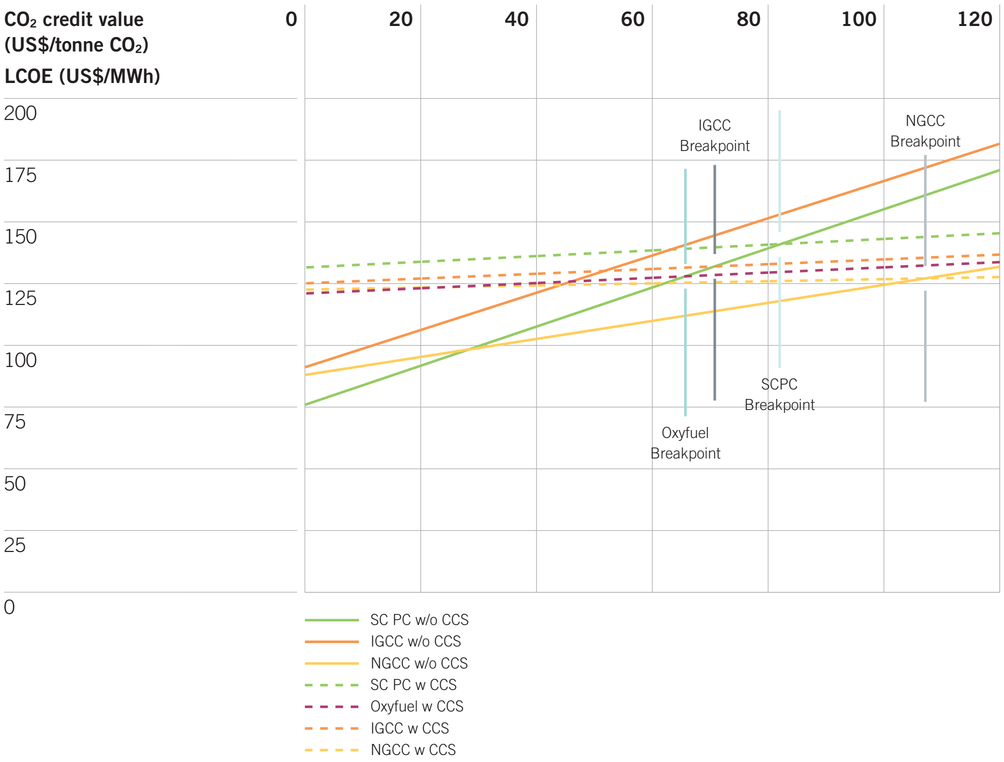
Figure 4-6 LCOE, including CCS, as a function of location



Sensitivity study: CO_2 credit value breakpoint

The CO_2 credit value breakpoint refers to the CO_2 credit value, as a $/tonne of CO_2 emitted, that drives the economics in favour of a CCS system over that without CCS. Below the breakpoint, it is more economically favourable to operate the system without CCS and for owners to pay for the emissions in the form of a tax or purchased credits. Figure 4-7 illustrates the breakpoints for the major generation and CO_2 capture technologies considered. While the oxyfuel combustion technology becomes favourable at the lowest CO_2 credit value, conversely, NGCC does not become favourable until a CO_2 credit value of greater than US$100/tonne of CO_2 is reached. This is primarily related to the lower CO_2 emission intensity of natural gas and relatively high efficiency of NGCC, as elaborated further below.

Figure 4-7 CO_2 credit value breakpoint comparison for CO_2 capture technologies.



The breakpoint for the CO_2 credit value for oxyfuel has decreased from US$60/tonne of CO_2 in 2009 to US$55/tonne, which can be attributed to the lower coal costs offsetting the additional purification step included in this study. This analysis continues to indicate that oxyfuel still has the lowest CO_2 credit value breakpoint of approximately US$55/tonne of CO_2.

The IGCC breakpoint, with respect to supercritical PC technology has decreased from US$80/tonne in 2009 to US$70/tonne of CO_2. This reflects the increase since 2009 in the LCOE and cost of CO_2 avoided and captured for IGCC with CCS.

The cost breakpoint for the supercritical technologies is approximately US$80/tonne of CO_2, an 11 per cent decrease from the 2009 breakpoint of US$90/tonne of CO_2.

The high breakpoint for NGCC technology has remained relatively unchanged at US$112/tonne of CO_2, reflective of the lower CO_2 emission intensity of natural gas and higher cycle efficiency compared to coal-fired technologies.

Care needs to be used in assessing technologies with this breakpoint approach. While the breakpoint of a technology may be the lowest, indicating a smaller cost increase to incorporate CO_2 capture, the initial cost of the technology may be high, thus still leading to a greater LCOE.

4.1.2 Industrial applications

The FOAK and NOAK cost parameters based on the methodology and assumptions described in section 3.6 for the industrial processes are listed in Table 4-5.

Table 4-5 Incremental cost of CCS for industrial processes

	Blast furnace	Cement	Natural gas processing	Fertiliser production
FOAK				
Incremental levelised product costs	US$82/tonne steel	US$34/tonne cement	US$0.056/GJ natural gas	US$11/tonne ammonia
Cost of CO_2 avoided/captured	US$54/tonne CO_2	US$54/tonne CO_2	US$19/tonne CO_2	US$20/tonne CO_2
NOAK				
Incremental levelised product costs	US$74/tonne steel	US$31/tonne cement	US$0.056/GJ natural gas	US$11/tonne ammonia
Cost of CO_2 avoided/captured	US$49/tonne CO_2	US$49/tonne CO_2	US$19/tonne CO_2	US$20/tonne CO_2

For the industrial processes, the incremental levelised product costs and the cost of CO_2 avoided/captured have increased by a small amount consistently across all applications. For PCC of CO_2 from blast furnace steel and cement production are in the order of US$50/tonne of CO_2. These costs are inclusive of the capture and compression of the CO_2 at the capture facility and the subsequent transportation and storage costs. A reduction of approximately 10 per cent is achieved through the removal of process contingency applied to the capture technology in moving from FOAK to NOAK systems.

The cost of CO_2 avoided/captured is lower than those of PCC for industrial processes that currently include a CO_2 separation/capture process, such as in natural gas processing and fertiliser production. The costs of CO_2 avoided/captured are US$19/tonne of CO_2 and US$20/tonne of CO_2 respectively for these facilities. As discussed in 2009, the additional auxiliary load required is for the compression components only, compared to the capture and compression of the PCC case. This is also a reason for the lower CCS costs for these facilities.

Table 4-6 shows the current commodity prices and the resulting increase in commodity prices, derived from the incremental levelised cost of production. The percentage increase is strongly linked to the commodity price with lower cost commodity prices being impacted to a greater extent. For example, the commodity cost for steel has increased from US$350-500/tonne in 2009 to US$570-800/tonne in 2010. As a result, the contribution that CCS has to the overall commodity cost is reduced.

Table 4-6 Commodity cost increase from CCS implementation

	Commodity cost	CCS cost increase
Steel	US$570-800/tonne	10-14%
Cement	US$66-88/tonne	39-52%
Natural gas	US$4.97/GJ	1%
Ammonia	US$375/tonne	3%

Notes:

1. Steel and natural gas commodity costs from World Bank Commodity Price Data (Pink Sheet) average or January to July 2010.
2. Portland cement prices, dependent on contract size (Portland Cement 2009).
3. Ammonia Prices and Pricing information, mid-June 2010 prices US location. (ICIS 2010)

Table 4-7 provides the cost parameters, levelised cost of production, and cost of CO_2 avoided/captured for the industrial processes in the selected regions. The cost factors used for moving the projects are listed in Table 3-3. Among the countries listed, the CCS cost parameters vary over a range of approximately 25 per cent. The countries with lower labour costs (China, India and Brazil) and low energy costs (Saudi Arabia) have the lowest cost for implementing CCS. The costs presented here are indicative of moving the same project, including the capture system and multi-user pipeline and storage, from a reference location to other regions. In assessing actual projects and comparing costs to existing projects, the characteristics and configuration of the industrial process as well as existing and proposed transportation and infrastructure must be taken into account.

Table 4-7 CCS cost parameters for NOAK industrial processes

		Blast furnace		Cement kiln		Natural gas processing		Fertiliser production	
Region	Country	ΔLCOP ($/tonne steel)	Cost of CO_2 avoided ($/tonne CO_2)	ΔLCOP ($/tonne cement)	Cost of CO_2 avoided ($/tonne CO_2)	ΔLCOP ($/GJ natural gas)	Cost of CO_2 avoided ($/tonne CO_2)	ΔLCOP ($/tonne ammonia)	Cost of CO_2 avoided ($/tonne CO_2)
ANZ	Australia	87	57	36	57	0.060	20.4	11.3	20.9
Asia	China	54	35	22	35	0.046	15.8	8.7	16.1
Asia	Japan	96	63	40	63	0.068	23.3	13.0	24.0
India	India	69	45	29	45	0.048	16.6	9.1	16.8
Europe	Euro Area	88	57	37	57	0.064	21.7	12.1	22.3
Europe	East Europe	77	51	32	51	0.061	20.9	11.6	21.4
ME/Africa	Saudi Arabia	66	43	27	43	0.044	15.0	8.2	15.2
ME/Africa	South Africa	88	57	36	57	0.063	21.6	12.0	22.1
Americas	Canada	88	57	36	57	0.060	20.6	11.4	21.1
Americas	United States	74	49	31	49	0.056	19.3	10.7	19.7
Americas	Brazil	80	52	33	52	0.058	19.8	11.0	20.2

ΔLCOP=Incremental change in levelised cost of production.

4 RESULTS AND COMPARATIVE SENSITIVITY ANALYSIS (CONTINUED)

4.2 CO_2 transportation by pipeline

As stated in the 2009 study, the pipeline transportation of CO_2 offers potential cost savings through combining the flow of CO_2 from multiple sources into a single pipeline for delivery to a single storage site.

Following the methodology used in the 2009 study, in the reference cases the CO_2 flow through the pipelines was set to the CO_2 generated by a single facility.

Figure 4-8 illustrates the cost savings achieved through increasing the CO_2 flow through a pipeline. As illustrated in Figure 4-8, the cost to transport the CO_2 will be between US$1-2 per tonne of CO_2. Through combining three or more plants, the CO_2 flow can be increased to greater than 10Mt per year, leading to a cost of less than US$1 per tonne for CO_2 transport. The resulting impact on the LCOE is illustrated in Figure 4-9.

Figure 4-8 Transportation cost saving from increasing pipeline flow for 100km pipeline

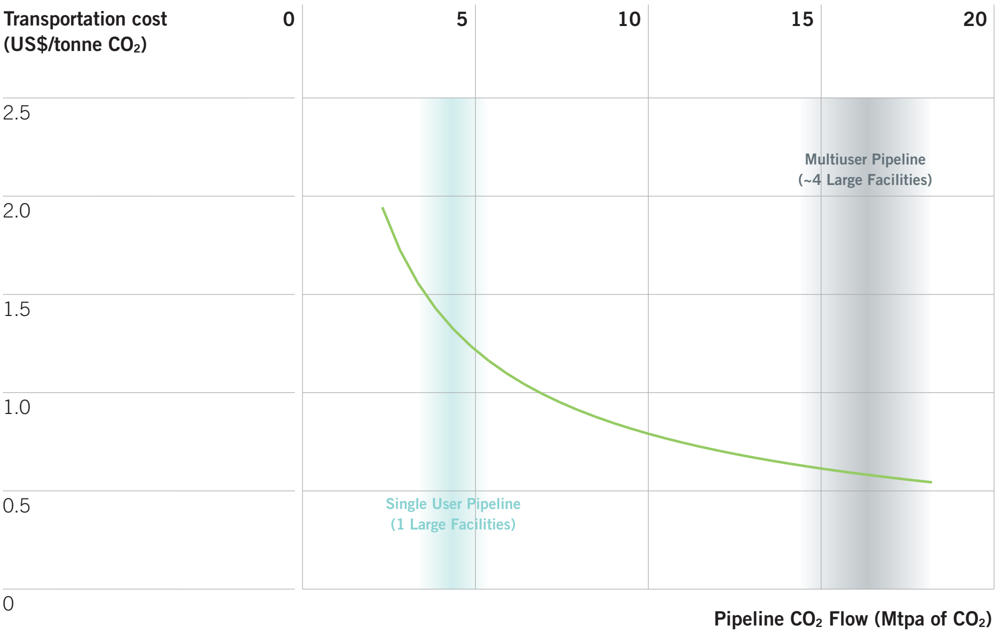
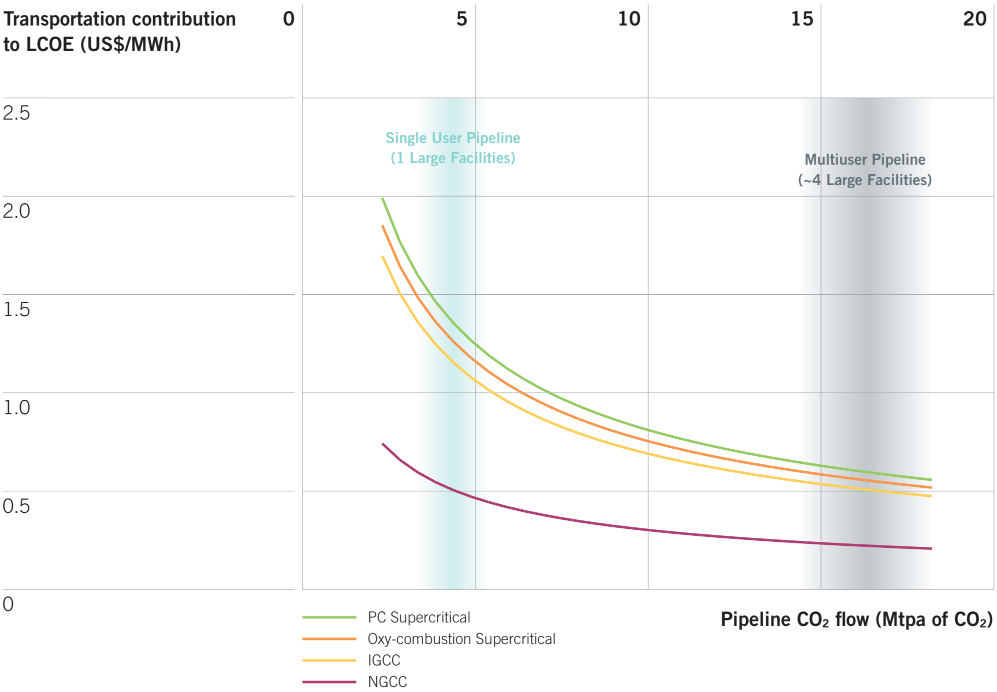


Figure 4-9 Transportation contribution to LCOE versus pipeline flow for 100km pipeline



The costs of CO_2 transport were less than those from they 2009 study as a result of the length of the pipeline being reduced from 250km to 100km.

4.3 CO_2 storage

Initial Storage Site Finding and Characterisation Costs

To illustrate the impact of this cost on the CO_2 storage cost, a sensitivity study was performed by varying the site characterisation costs over the anticipated cost range. The same cost range was used from the 2009 study (US$15 million to US$150 million) with the resulting impact on CO_2 storage cost illustrated in Figure 4-10 and the storage cost contribution to LCOE shown in Figure 4-11. The upper value of US$150 million was used as this was considered the economic threshold before proponents would abandon investigations.

4 RESULTS AND COMPARATIVE SENSITIVITY ANALYSIS (CONTINUED)

Figure 4-10 Dependence of CO_2 storage costs on initial site characterisation and identification costs

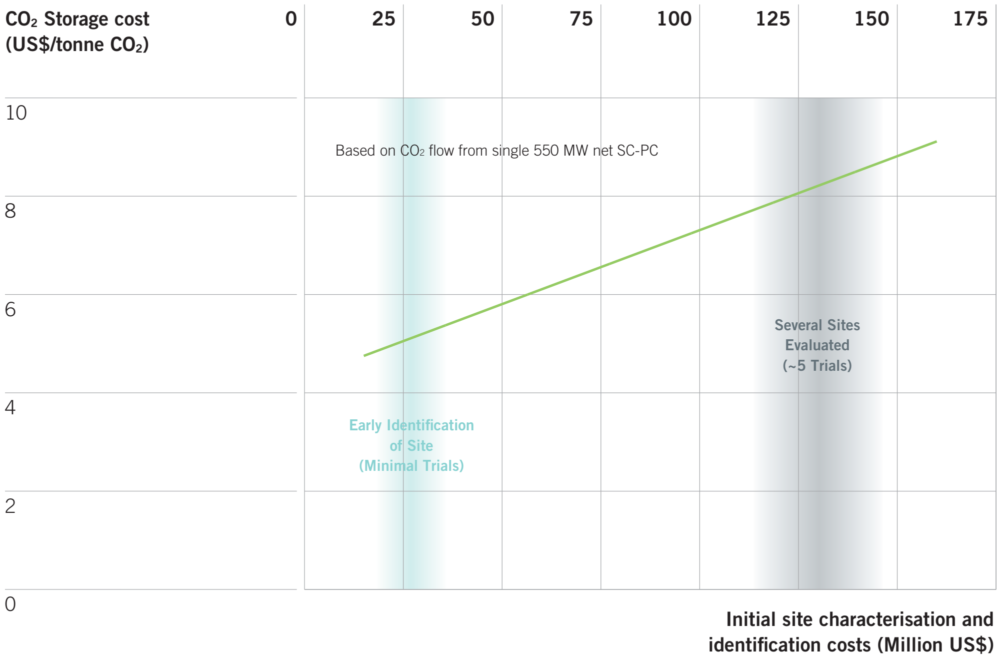
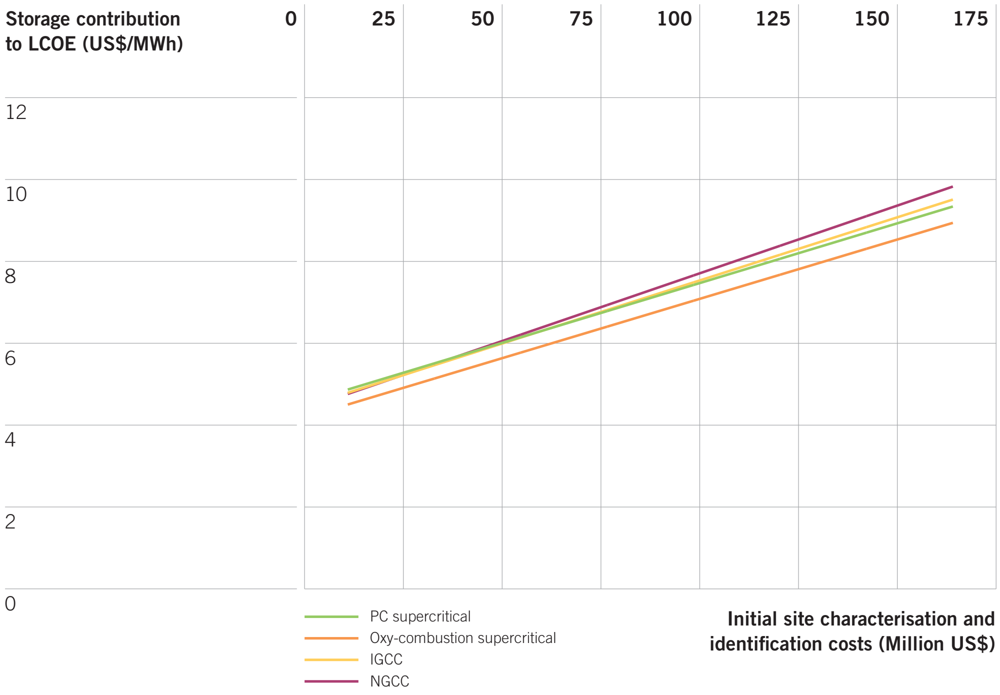


Figure 4-11 Dependence of storage contribution to LCOE on initial site characterisation and identification costs



Storage site geological characteristics

The reservoir geological properties govern the rate that the CO_2 can be injected. For reservoirs with geologic properties (low reservoir permeability thickness product) that significantly limit the injection rate, additional wells in the same area will be required to take all of the CO_2 from the pipeline. However, the number of wells, and hence maximum rate in a given area, is limited not by the performance of a single well but by pressure interference between them, such that there is a diminishing return (incremental injection rate) for additional wells. A sensitivity study for the CO_2 storage costs ($/tonne) and the LCOE ($/MWh) around the poor (absolute permeability = 150md and thickness = 5m) and the good (absolute permeability = 400md and thickness = 15m) reservoir properties are shown in Table 4-8.

In the case of the 'poor reservoir' case for NGCC, the storage cost contribution to LCOE is lower than for the other power generation applications. This can be attributed to the volumes of CO_2 from an NGCC facility being approximately half that of the others. On the other hand for the 'good reservoir' case for NGCC, the storage cost contribution to LCOE is comparable to those of the other power generation applications as the number of wells required to inject the greater volumes of CO_2 from the other applications is small.

Table 4-8 Storage cost and contribution to LCOE based on reservoir properties

	Storage cost contribution to LCOE (US$/MWh)				
Reservoir properties	PC supercritical	Oxy-combustion supercritical	IGCC	NGCC	Storage cost US$/tonne CO_2
Poor	13	13	13	9	13
Good	6	6	6	6	6

Summary of storage considerations

A number of reservoir characteristics for these conceptual case studies were assumed that allow a rapid preliminary assessment of their potential project economics. However, the reader must understand that these same assumptions remain simplifications for the purposes of these preliminary macroeconomic models. Developing an actual project will require a much deeper assessment that will require a project specific model be built once a suitable storage site has been identified and characterised.

Using these conceptual study assumptions, the overall economics show the discounted cost of storing a tonne of CO_2 is similar in both cases at approximately US$10/tonne CO_2, which shows that the storage costs under the assumptions used here could be relatively minor contributors to the overall cost of a full CCS project. However, such a statement must be read with caution that the assumption being made is that the exploration and appraisal program that must be undertaken to prove up the sites, works perfectly, and that the exploration wells can be re-used as both injection and monitoring wells. Thus there is no 'lost' exploration funds exhausted on sites that prove technically unsuitable for CO_2 storage. Likewise, should a move offshore be required, it would dramatically increase the cost of CO_2 storage and transportation over that used here.

4 RESULTS AND COMPARATIVE SENSITIVITY ANALYSIS (CONTINUED)

In reality, some key observations can be made on the cost of storing CO_2. This model only considers onshore storage. However, the authors are aware of many projects proposing to store CO_2 offshore. Storing CO_2 offshore will increase costs significantly, especially in the existing relatively tight market for offshore drilling rigs and platforms.

Water issues are also becoming apparent, not just for CCS projects but for other new hydrocarbon projects such as shale gas and coal seam methane. Thus, it is reasonable to expect that there may be more upfront (and on-going) work needed to ensure regulators are satisfied that there is little or no impact of CCS operations on water resources. This extra monitoring will increase costs.

This analysis has also only considered ‘proven’ geological storage options into saline aquifers or depleted oil and gas reservoirs, and not any potential EOR, enhanced gas recovery (EGR) or enhanced coal-bed methane recovery (ECBM) options that could provide an offsetting economic benefit of carbon capture through beneficial re-use. ECBM still has a number of technical issues needing solved prior to the implementation on large scale CCS project though progress is being made, notably in the US and beneficial re-use needs to be considered on its own merits should it prove a potential early revenue generator for CO_2 use. That said, the authors do reiterate that the sheer volumes of CO_2 being emitted from even small power stations (500MW +) places a fundamental supply and demand mismatch with many EOR projects that would be at best only able to utilise 1-2Mtpa CO_2 and indeed may not have anything like the 30 year lifespan of a power station project. This was discussed in Foundation Report Two in 2009.

As stated previously, CO_2 storage costs are site specific and the local geology will drive the costs of CO_2 storage. Injection enhancements such as deviated wells and fracturing operations may increase the injection rates, but the trade-off will need to be evaluated in the specific site context. The costs shown here are conservative. Based on currently industry trends, it is unlikely that these will decrease in the near future. Furthermore, many potential market pressures exist that are only likely to increase from this point onwards given the continued demand for oilfield services around the world.

It is important to recognise that should CO_2 storage demand increase as much as predicted, then it will place significant pressure on the supply of oilfield services competing against a mature industry (hydrocarbons). As a result, prices and price inflation could rise significantly more than that assumed in this study.

5 CONCLUSIONS

The second report as part of the Global CCS Institute's Strategic Analysis of the Global Status of CCS in 2009 (Foundation Report Two) gave a detailed analysis of the capture, transport and storage costs for power plants and a select range of industrial applications. This has been updated in 2011 to reflect updates to:

- the regional localisation estimates;
- capital cost estimates for power and select industrial CCS applications; and
- the overall economic model.

The revised results of the economic assessment of CCS technologies are summarised in Table 5-1.

The primary purpose in providing these costs is to compare the relative costs of CCS for various technologies. The costs are prepared for specific bases and when comparing to other project costs, variations are to be expected based on changes in design specification, owner's preferences and appetite for risk and how the project is financed. Great care and study is required to make these comparisons.

Table 5-1 Summary results of the economic assessment of CCS technologies

		Power generation				**Industrial applications**			
		PC supercritical & ultra super-critical[*1]	**Oxyfuel combustion standard & ITM[*1]**	**IGCC**	**NGCC**	**Blast furnace steel production**	**Cement production**	**Natural gas processing**	**Fertiliser production**
	Dimensions	**US$/MWh**	**US$/MWh**	**US$/MWh**	**US$/MWh**	**US$/tonne steel**	**US$/tonne cement**	**US$/GJ natural gas**	**US$/tonne ammonia**
Levelised cost of production	Without CCS[*2]	73-76	73-76[*3]	91	88	570-800	66-88	4.97	375
	With CCS FOAK[*3]	120-131	114-123	125	123	82	34	0.056	11
	With CCS NOAK[*4]	117-129	112-121	123	121	74	31	0.056	11
	% Increase over without CCS[*5]	61-76%	53-65%	37%	40%	10-14%	39-52%	1%	3%
Cost of CO_2 avoided[*6] (\$/tonne CO_2)	FOAK	62-81	47-59	67	107	54	54	19	20
	NOAK	57-78	44-57	63	103	49	49	19	20
Cost of CO_2 captured (\$/tonne CO_2)	FOAK	53-55	42-47	39	90	54	54	19	20
	NOAK	52	41-45	38	87	49	49	19	20

Notes:

[1] The ultra-supercritical and ITM technologies are currently under development and are not commercially available. These technologies represent options with the potential for increasing the process efficiency and reduce costs.

[2] Without CCS cost of production for industrial process are typical market prices for the commodities.

[3] Oxyfuel combustion systems are not typically configured to operate in an air fired mode. Therefore, oxyfuel combustion without CCS is not an option. The values here are the PC without CCS value to be used as a reference for calculating the cost of CO_2 avoided.

[4] For industrial processes, levelised cost of production presented as cost increment above current costs.

[5] Expressed with respect to current commodity prices industry industrial processes.

Key findings based on the updated economic modelling are presented in the following sections.

5 CONCLUSIONS (CONTINUED)

5.1 General observations

CO_2 capture still represents the greatest contribution to the cost of CCS, with the majority of the cost increases being due to changes in the capture system. The percentage increases in costs that the application of CCS has over non-CCS facilities have remained relatively unchanged since 2009.

Though minor changes in the costs of CCS across power generation and industrial applications have occurred, the costs of CCS still remain high. This is expected, given that it has only been 12 months since the initial Foundation Report Two, and major developments that have the potential to dramatically reduce the cost of CCS have not yet occurred or have been sufficiently tested for commercialisation.

Despite the costs of CCS being high relative to traditional power generation and industrial facilities, it is important to consider that these traditional methods currently emit large amounts of CO_2 into the atmosphere. Given the current and anticipated restrictions on facility emissions, these facilities will not be allowed to continue to operate as they have in the past.

The high costs of CCS as identified in this study should be considered with other low emission technologies to allow consideration of approaches to low emission power and industrial production. Further, if CCS is compared against the anticipated cost that may be imposed on facilities for emitting CO_2 it is likely to appear more competitive in a low carbon market.

5.2 Application of CCS to power generation

For the application of CCS in power generation in the United States, a decrease in fuel costs was seen across all of the coal fired technologies, related to the lower coal costs that emerged in 2010. The reduction in the length of the pipeline relative to the 2009 study reduced the contribution of transport cost to the overall cost of CCS to the power generation applications.

5.2.1 Levelised cost of electricity

For the reference cases, taking into account currently available technologies, the lowest LCOE was for oxyfuel combustion at US$114/MWh, in contrast to 2009 where LCOE for NGCC technologies was the lowest at US$112/MWh. Consistent with the findings in 2009, the LCOE for PC supercritical and IGCC technologies were the greatest at US$131/MWh and US$125/MWh respectively.

An update to the oxyfuel combustion process with CCS was the inclusion of an additional purification process when capturing the CO_2. This resulted in an increase in the capital contribution of oxyfuel combustion with CCS to the LCOE. The effects of the ranges in the coal and natural gas prices were variations in the LCOE of US$10/MWh and US$30/MWh respectively. This reflects the potential greater volatility of natural gas prices.

For a supercritical PC with CCS technology, for a fixed fuel cost, the sensitivity of the CO_2 capture installed capital costs and LCOE to the labour costs was reduced. The installed capital costs increased by 23 per cent (32 per cent in 2009), while the LCOE increased by 11 per cent (21 per cent in 2009). A similar trend would be observed for the other coal-fired technologies as they tend to be relatively labour-intensive installations.

5.2.2 Cost of CO_2 avoided

For both FOAK and NOAK plants, the cost of CO_2 avoided for the application of CCS across all power generation technologies decreased since 2009. There are various factors that resulted in this, including that most of the coal prices are lower than those modelled in the 2009 study, and that the CO_2 transport distance for the reference case was reduced from 250km to 100km, reducing the transport costs and its contribution to the overall cost of CCS;

5.2.3 Regional observations

In India, the installed CO_2 capture equipment cost and LCOE increased across all technologies. This was due to the revised consideration of a 30 per cent increase in equipment being imported into the country as well as India's typical coal heating value being very low, resulting in a greater capital cost.

The increase in costs for the coal fired technologies in Eastern Europe was primarily due to an increase in labor conversion factor from the reference location and switching to a low rank coal. Similarly, the capital cost in the euroregion increased related to a change to a low rank coal.

A higher coal price was utilised for Australia in this study, which resulted in a 20 per cent increase in the country's technology cost.

The costs in Brazil increased significantly, partially because of a lower labour rate being used in 2009. The revision of the coal type to one with a lower heating value also led to a higher capital cost. Further, additional costs associated with importing capital equipment contributed to the increase in CO_2 capture costs in Brazil.

Only NGCC costs were presented for Saudi Arabia, reflecting that there are no coal-fired power generation applications in the region.

5.2.4 CO_2 credit value breakpoint

The CO_2 credit value, on a $/tonne of CO_2 emitted basis that drives the economics of CCS in favour of a CCS system over that without CCS is known as the CO_2 value breakpoint. Once the breakpoint is exceeded, it becomes more economically favourable to operate the system with CCS.

The CO_2 credit value breakpoint for oxyfuel decreased from US$60/tonne of CO_2 in 2009 to US$55/tonne, which can be attributed to the lower coal costs offsetting the additional purification step included in this study. This analysis continues to indicate that oxyfuel still has the lowest CO_2 credit value breakpoint of approximately US$55/tonne of CO_2.

For IGCC, the CO_2 breakpoint with respect to supercritical PC technology PC technology has decreased from $80/tonne in 2009 to $70/tonne of CO_2. This reflects the increase since 2009 in the LCOE and cost of CO_2 avoided and captured for IGCC with CCS.

The cost breakpoint for the supercritical technologies is approximately $80/tonne of CO_2, an 11 per cent decrease from the 2009 breakpoint of $90/tonne of CO_2.

Finally, the high breakpoint for NGCC technology has remained relatively unchanged at $112/tonne of CO_2, reflective of the lower CO_2 emission intensity of natural gas and higher cycle efficiency compared to coal-fired technologies.

5.3 Application of CCS to select industrial applications

For the application of CCS on select industrial processes, the incremental levelised product costs and the cost of CO_2 avoided/captured have increased by a small amount consistently across all applications. A reduction of approximately 10 per cent is achieved through the removal of process contingency applied to the capture technology in moving from FOAK to NOAK systems.

Consistent with the findings from Foundation Report Two in 2009, the cost of CO_2 avoided/captured is lower for those industrial processes that currently include a CO_2 separation/capture process, compared to PCC from other industrial processes. Natural gas processing and fertiliser production, for example, already separate the CO_2 so additional expenditure is required for the compression, transport and storage of the CO_2 only.

The increases in the commodity costs due to CCS implementation also reflect the level of additional investment required to apply CCS to select industrial facilities. For example, for natural gas and fertiliser (ammonia) production, the commodity cost increases due to CCS are 1 and 3 per cent respectively. On the other hand, the commodity cost increases due to the application of CCS for steel and cement are 10-14 per cent and 39-52 per cent respectively.

5.4 CO_2 transport

The reduction in the length of the pipeline for the reference case has reduced the overall transport costs and the contribution of transport cost to the overall cost of CCS. The cost to transport CO_2 is estimated to be between US$1-2 per tonne of CO_2, a decrease from US$3-4 per tonne of CO_2 in 2009. This is due to the reduction of the pipeline length in the reference case from 250km in 2009 to 100km.

An opportunity to reduce the costs of CO_2 transport is in increasing the CO_2 flow through the pipeline, by combining the captured CO_2 from multiple sources through a larger pipeline to a common storage site. The implementation of a common user pipeline can result in a cost of less than US$1 per tonne for CO_2 transport.

5.5 CO_2 storage

The contribution of storage cost to the LCOE was found to range from US$6-13 per tonne of CO_2 depending on whether the ‘good’ or ‘poorer’ reservoir option was considered.

For the case of the ‘poor reservoir’ storage option being applied for NGCC, the storage cost contribution to LCOE us lower than for the other power generation applications. This can be attributed to the volumes of CO_2 from an NGCC facility being approximately half of that of the others. This has a lesser impact for the ‘good reservoir’ case as the number of wells required to inject greater volumes of CO_2 under this case do not vary significantly.

6 REFERENCES

ACIL Tasman 2009, *Fuel resources, new entry and generation costs in the NEM*, viewed August 2009, <http://www.aciltasman.com.au/images/pdf/419_0035.pdf>.

Chikkatur, A. P 2008, *A Resource and Technology Assessment of Coal Utilization in India*, Pew Center on Climate Change, viewed 13 October 2010, <http://www.pewclimate.org/docUploads/india-coal-technology.pdf>.

DOE/NETL 2006, *Oxyfuel Combustion Systems Analysis Study*, August 2006.

DOE/NETL 2007, *Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity*, Rev. 1, DOE/NETL-2007/1281, August 2007.

DOE/NETL 2008 Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity, Rev. 2, DOE/NETL-2007/1291, August 2008.

DOE/NETL 2008b *Storing CO_2 with Enhanced Oil Recovery*, DOE/NETL-402/1312/02-07-08. February 7.

DOE/NETL 2008c *Pulverized Coal Oxycombustion Power Plants, Volume 1: Bituminous Coal to Electricity*, Final Report, DOE/NETL 2007/1291, Revision 2, August.

DOE/NETL 2001, *Engineering Feasibility and Economics of CO_2 Capture in an Existing Coal-Fired Power Plant*

Electric Power Research Institute (EPRI) 1987-1989, TAG™ – *Technical Assessment Guide*, Vol. 1, Electricity Supply – 1989; Vol. 2, Electricity End-Use – Part 1, 1987, Parts 2 & 3, 1988; Vol. 3, Fundamentals and Methods, Supply – 1987; Vol. 4, Fundamentals and End-Use" – 1987, EPRI P-4463-SR, Palo Alto, CA.

Electric Power Research Institute (EPRI) 1999, *Enhanced Oil Recovery Scoping Study*, EPRI, Palo Alto, CA. TR-113836.

Electric Power Research Institute (EPRI) 1994, *Technical Assessment Guide*, Palo Alto, CA.

Electric Power Research Institute (EPRI) 2008, *CoalFleet Guideline for Advanced Pulverized Coal Power Plants*, Version 3, EPRI, Palo Alto, CA: March.

Electric Power Research Institute (EPRI) 2007, *CoalFleet RD&D Augmentation Plan for Integrated Gasification Combined Cycle (IGCC) Power Plants*, Report no. 1013219, EPRI, Palo Alto, CA: January.

Energy Information Administration (EIA) 2009, *Steam Coal Prices for Electricity Generation*, viewed August 2009, <http://www.eia.doe.gov/emeu/international/stmforelec.html>.

Energy Information Administration (EIA) 2008, *Natural Gas Prices for Electricity Generation*, viewed August 2009, <http://www.eia.doe.gov/emeu/international/ngasprie.html>.

Energy Information Administration (EIA) 2009b, *Coal News and Markets*, viewed August 2009, <http://tonto.eia.doe.gov/FTPROOT/coal/newsmarket/coalmar090424.html>.

Energy Information Administration (EIA) 2009c, *Natural Gas Prices*, viewed August 2009, <http://tonto.eia.doe.gov/dnav/ng/ng_pri_sum_dcu_nus_m.htm>.

Energy Information Administration (EIA) 2009d, *Monthly Fuel Receipts and Fuel Quality Database*, viewed August 2009, <http://www.eia.doe.gov/cneaf/electricity/page/eia423.html>.

Energy Information Administration (EIA) 2009e, *Form EIA-923, Power Plant Operations Report*, US DOE, Washington D.C.

Energy Information Administration (EIA) 2000, *Costs and Indices for Domestic Oil and Gas Field Equipment and Production Operations*, EIA.

6 REFERENCES (CONTINUED)

Energy Information Administration (EIA) 2008b, 'Selected Countries, Most Recent Annual Estimates, 1980-2007', *International Net Electricity Generation Tables*, viewed August 2009, data posted in 2008, <http://www.eia.doe.gov/emeu/international/electricitygeneration.html>.

Environmental Protection Agency (EPA) 2006, *Environmental Footprints and Costs of Coal-Based Integrated Gasification Combined Cycle and Pulverized Coal Technologies*, EPA, Washington D.C.

Holt, N & Booras, G 2007, *Updated Cost and Performance Estimates for Clean Coal Technologies Including CO_2 Capture – 2006*, EPRI Report No. 1013355, March 2007.

International Energy Agency 2010, *Projected Costs of Generating Electricity*, 2010 Edition, OECD, IEA, Paris, France.

Maurstad et al 2005, 'Impact of Coal Quality and Gasifier Technology on IGCC Performance', *presented at the 8th International Conference on Greenhouse Gas Control Technologies*, Trondheim, Norway.

Nygaard R & Lavoie R 2009, Project Cost Estimate – Wabamum Area CO_2 Sequestration Project (WASP), Institute for Sustainable Energy, Environment and Economy (ISEEE).

NZEC 2009, *Carbon Dioxide Capture from Coal-Fired Power Plants in China*, Summary Report for NZEC Work Package 3.

Powers, W.E 2005, 'Peak and Annual Average Energy Efficiency Penalty of Optimised Air-Cooled Condenser on 515 MW Fossil Fuel-Fired Utility Boiler', *presented at California Energy Commission/Electric Power Research Institute Advanced Cooling Technologies Symposium*, Sacramento, California, June 2005.

Powers, W.E 2009, 316(b)*Phase II Closed-Cycle Cooling Retrofit Options, Feasibility and Cost*, accessed 13 October 2010, <http://web.ead.nl.gov/bajatermoeis/documents/docs/09-30-09_cooling_retrofit_cost.pdf>.

Richardson Products 2010, *International Cost Factor Location Manual*, 2009-2010 Edition, Cost Data On Line Inc, Nevada.

Rubin, E.S, Chen, C, Rao, A.B 2007, 'Cost and Performance of Fossil Fuel Power Plants with CO_2 Capture and Storage', *Energy Policy*, Volume 35, Issue 9, September 2007, pp 4444-4454.

Rubin, E.S., Yeh, S, Antes, M, Berkenpas, M, Davison, J 2007, 'Use of experience curves to estimate the future cost of power plants with CO_2 capture', *International Journal of Greenhouse Gas Control*, Vol. 1, pp 188-197.

U.S. Department of Energy 2008a. *Pulverized Coal Oxycombustion Power Plants*, DOE/NETL-2007/1291, USA.

U.S. Department of Energy 2008b, *Enhanced Oil Recovery/CO_2 Injection Program Description*. U.S. Department of Energy, Office of Fossil Energy, Office of Oil and Natural Gas viewed July 2009, <http://fossil.energy.gov/programs/oilgas/eor/>.

US DOE 2008c, *Storing CO_2 with Enhanced Oil Recovery*, DOE/NETL-402/1312/02-07-08, USA, February 7.

US DOE 2009, *High performance materials*, U.S. Department of Energy, viewed July 2009, <www.fossil.energy.gov/programs/powersystems/advresearch/advresearch-materials.html>.

WorleyParsons 2002, Confidential Study Report 1

WorleyParsons 2007, Confidential Study Report 2

WorleyParsons 2009a, Cost Estimating Support for the Australian Government's Energy White Paper

WorleyParsons 2009b, WorleyParsons internal data for US DOE "Low Rank Coal Study", publication expected in 2010.

APPENDICES

Appendix A Breakdown of overnight capital costs of PC supercritical facility

Table A-1 Cost accounts for PC boiler (1,000 x 2010 US$)

Acct No.	Item/description	Equipment cost ($)	Material cost ($)	Labour ($)	Bare erected cost ($)	Eng'g CM H.O. & Fee ($)	Contingencies		TOTAL PLANT COST	
							Process ($)	Project ($)	($)	$/kW
1	COAL & SORBENT HANDLING	22,886	5,991	13,389	42,266	3,821	0	6,913	53,001	97
2	COAL & SORBENT PREP & FEED	15,552	875	3,809	20,237	1,776	0	3,302	25,314	46
3	FEEDWATER & MISC. BOP SYSTEMS	65,385	0	29,372	94,757	8,614	0	17,049	120,420	221
4	PC BOILER									
4.1	PC Boiler & accessories	256,449	0	123,313	379,762	34,056	0	41,382	455,200	834
4.2	SCR (w/4.1)	0	0	0	0	0	0	0	0	0
4.3	Open	0	0	0	0	0	0	0	0	0
4.4-4.9	Boiler BoP (w/ID Fans)	0	0	0	0	0	0	0	0	0
	SUBTOTAL 4	**256,449**	**0**	**123,313**	**379,762**	**34,056**	**0**	**41,382**	**455,200**	**834**
5	FLUE GAS CLEANUP	121,634	0	40,039	161,673	15,293	0	17,697	194,663	357
5B	CO_2 REMOVAL & COMPRESSION	274,753	0	79,993	354,746	33,486	62,597	90,166	540,995	991
6	COMBUSTION TURBINE/ACCESSORIES									
6.1	Combustion turbine generator	N/A	0	N/A	0	0	0	0	0	0
6.2-6.9	Combustion turbine other	0	0	0	0	0	0	0	0	0
	SUBTOTAL 6	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**0**
7	HRSG, DUCTING & STACK									
7.1	Heat recovery steam generator	N/A	0	N/A	0	0	0	0	0	0
7.2-7.9	Ductwork and stack	21,230	1,125	13,995	36,351	3,343	0	5,206	44,901	82
	SUBTOTAL 7	**21,230**	**1,125**	**13,995**	**36,351**	**3,343**	**0**	**5,206**	**44,901**	**82**
8	STEAM TURBINE GENERATOR									
8.1	Steam TG & accessories	75,946	0	8,237	84,183	6,871	0	9,105	100,159	183
8.2-8.9	Turbine plant auxiliaries and steam piping	32,386	1,317	17,111	50,814	4,385	0	7,906	63,105	116
	SUBTOTAL 8	**108,332**	**1,317**	**25,348**	**134,997**	**11,256**	**0**	**17,012**	**163,264**	**299**
9	COOLING WATER SYSTEM	25,717	13,332	22,767	61,816	5,768	0	9,143	76,726	141
10	ASH/SPENT SORBENT HANDLING SYS	6,215	186	7,849	14,250	1,364	0	1,606	17,219	32
11	ACCESSORY ELECTRIC PLANT	24,495	12,167	33,539	70,201	6,277	0	9,702	86,180	158
12	INSTRUMENTATION & CONTROL	11,209	0	11,065	22,274	2,032	1,114	3,117	28,537	52
13	IMPROVEMENTS TO SITE	3,751	2,169	7,354	13,274	1,318	0	2,918	17,510	32
14	BUILDINGS & STRUCTURES	0	27,793	26,036	53,830	4,932	0	8,814	67,576	124
	TOTAL COST	**957,610**	**64,954**	**437,869**	**1,460,433**	**133,336**	**63,711**	**234,026**	**1,891,506**	**3,464**

APPENDICES (CONTINUED)

Table A-2 Cost accounts for IGCC (1,000 x 2010 US$)

Acct No.	Item/description	Equipment cost ($)	Material cost ($)	Labour ($)	Bare erected cost ($)	Eng'g CM H.O. & fee ($)	Contingencies		TOTAL PLANT COST	
							Process ($)	Project ($)	($)	$/kW
1	COAL & SORBENT HANDLING	15,665	2,827	11,864	30,356	2,778	0	6,627	39,761	77
2	COAL & SORBENT PREP & FEED	124,146	9,723	20,168	154,037	13,362	0	33,480	200,879	388
3	FEEDWATER & MISC. BOP SYSTEMS	10,562	8,463	9,974	28,999	2,712	0	7,275	38,987	75
4	GASIFIER & ACCESSORIES									
4.1	Gasifier, syngas cooler & auxiliaries	207,320	0	50,519	257,839	15,431	36,360	47,288	356,918	690
4.2	Syngas cooling (w/4.1)	w/4.1	0	w/4.1	0	0	0	0	0	0
4.3	ASU/oxidant compression	170,291	0	w/equip.	170,291	16,174	0	18,646	205,111	397
4.4-4.9	Other gasification equipment	30,560	11,134	17,201	58,895	5,646	0	13,703	78,244	151
	SUBTOTAL 4	**408,171**	**11,134**	**67,720**	**487,024**	**37,251**	**36,360**	**79,638**	**640,273**	**1,238**
5A	GAS CLEANUP & PIPING	96,734	5,219	79,386	181,339	17,454	26,141	45,223	270,157	522
5B	CO_2 REMOVAL & COMPRESSION	20,640	0	11,692	32,332	3,091	0	7,085	42,508	82
6	COMBUSTION TURBINE/ACCESSORIES									
6.1	Combustion turbine generator	105,200	0	6,098	111,298	10,336	11,130	13,276	146,039	282
6.2-6.9	Combustion turbine other	0	784	873	1,657	158	0	545	2,360	5
	SUBTOTAL 6	**105,200**	**784**	**6,971**	**112,955**	**10,494**	**11,130**	**13,821**	**148,400**	**287**
7	HRSG, DUCTING & STACK									
7.1	Heat recovery steam generator	42,269	0	5,244	47,513	4,085	0	5,160	56,758	110
7.2-7.9	Ductwork and stack	3,851	2,653	3,482	9,987	928	0	1,768	12,682	25
	SUBTOTAL 7	**46,120**	**2,653**	**8,726**	**57,500**	**5,013**	**0**	**6,928**	**69,440**	**134**
8	STEAM TURBINE GENERATOR									
8.1	Steam TG & accessories	50,392	0	4,702	55,094	3,237	0	5,833	64,164	124
8.2-8.9	Turbine plant auxiliaries and steam piping	10,684	950	6,973	18,607	1,689	0	3,927	24,224	47
	SUBTOTAL 8	**61,076**	**950**	**11,675**	**73,701**	**4,926**	**0**	**9,760**	**88,388**	**171**
9	COOLING WATER SYSTEM	8,308	9,297	7,366	24,970	2,284	0	5,594	32,849	64
10	ASH/SPENT SORBENT HANDLING SYS	21,365	1,642	10,157	33,164	3,151	0	3,968	40,282	78
11	ACCESSORY ELECTRIC PLANT	27,704	9,505	25,910	63,120	5,847	0	13,090	82,057	159
12	INSTRUMENTATION & CONTROL	12,592	2,244	7,837	22,673	2,055	1,134	4,291	30,153	58
13	IMPROVEMENTS TO SITE	3,804	2,255	9,130	15,190	1,512	0	5,010	21,712	42
14	BUILDINGS & STRUCTURES	0	7,130	8,041	15,171	1,403	0	2,721	19,294	37
	TOTAL COST	**962,087**	**73,827**	**296,617**	**1,332,531**	**113,334**	**74,764**	**244,510**	**1,765,138**	**3,413**

Table A-3 Cost accounts for Oxyfuel Combustion (1,000 x 2010 US$)

							Contingencies		TOTAL PLANT COST	
Acct No.	**Item/description**	**Equipment Cost ($)**	**Material Cost ($)**	**Labour ($)**	**Bare Erected Cost ($)**	**Eng'g CM H.O. & Fee ($)**	**Process ($)**	**Project ($)**	**$(x1,000)**	**$/kW**
1	COAL & SORBENT HANDLING	22,026	5,762	12,876	40,664	3,662	0	6,649	50,975	93
2	COAL & SORBENTPREP & FEED	14,964	843	3,665	19,472	1,702	0	3,176	24,350	44
3	FEED WATER & MISC. BOP SYSTEMS	56,548	0	25,569	82,117	7,411	0	14,472	104,001	190
4	PC BOILER/GASIFIER									
4.1	PC boiler & accessing	243,473	0	117,076	360,549	32,008	53,683	44,624	490,865	895
4.2	SCR (w/4.1)	w/4.1	w/4.1	w/4.1	w/4.1	w/4.1	w/4.1	w/4.1	w/4.1	w/4.1
4.3	Open	0	0	0	0	0	0	0	0	0
4.4-4.9	Other equipment-ASU	136,708	0	108,574	245,282	23,930	0	26,921	296,134	540
	SUBTOTAL 4	**380,181**	**0**	**225,650**	**605,831**	**55,939**	**53,683**	**71,546**	**786,999**	**1,434**
5A	GAS CLEANUP & PIPING	109,055	0	34,759	143,814	13,556	0	15,736	173,106	315
5B	CO_2REMOVAL&COMPRESSION	68,858	0	53,971	122,829	12,282	0	27,022	162,133	295
6	COMBUSTIONTURBINE/ACCESSORIES									
6.1	Combustion turbine generator	N/A	0	N/A	0	0	0	0	0	0
6.2-6.9	Combustion turbine other	0	0	0	0	0	0	0	0	0
	SUBTOTAL 6	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**0**
7	HRSG, DUCTING & STACK									
7.1	Heat recovery steam generator	0	0	0	0	0	0	0	0	0
7.2-7.9	Ductwork and stack	0	0	0	0	0	0	0	0	0
	SUBTOTAL 7	**14,848**	**867**	**9,312**	**25,026**	**2,263**	**0**	**4045**	**31,335**	**57**
8	STEAM TURBINE GENERATOR									
8.1	Steam TG & accessories	83,235	–	9,028	92,263	7,507	0	9,977	109,746	200
8.2-8.9	Turbine plant auxiliaries and steam piping	33,091	–	17,444	50,535	4,512	0	7,769	62,816	114
	SUBTOTAL 8	**116,326**	**–**	**26,472**	**142,798**	**12,019**	**–**	**17,746**	**172,563**	**314**
9	COOLING WATER SYSTEM	15,883	8,169	14,185	38,236	3,562	0	5675	47,473	87
10	ASH/SPENTSORBENT HANDLING SYS	5,992	179	7,568	13,739	1,310	0	1548	16,597	30
11	ACCESSORY ELECTRIC PLANT	30,614	16,381	44,728	91,723	8,190	0	12808	112,720	205
12	INSTRUMENTATION & CONTROL	12,263	–	12,107	24,371	2,216	0	3,270	29,857	54
13	IMPROVEMENTS TO SITE	3,732	2,157	7,315	13,204	1,307	0	2,902	17,413	32
14	BUILDINGS & STRUCTURES	–	27,172	25,477	52,649	4,799	0	8,617	66,065	120
	Total Cost	**851,291**	**61,530**	**503,652**	**1,416,472**	**130,217**	**53,683**	**195,215**	**1,795,588**	**3,272**

U.S. Department of Energy

W.A. Parish Post-Combustion CO_2 Capture and Sequestration Project

Draft Environmental Impact Statement

Summary

September 2012

DOE/EIS-0473D

Office of Fossil Energy
National Energy Technology Laboratory

INTENTIONALLY LEFT BLANK

COVER SHEET

Responsible Federal Agency: U.S. Department of Energy (DOE)
Title: W.A. Parish Post-Combustion CO_2 Capture and Sequestration Project, Draft Environmental Impact Statement (DOE/EIS-0473D)
Location: Southeastern Texas, including Fort Bend, Wharton, and Jackson Counties
Contacts:

For further information about this Environmental Impact Statement, contact:

Mark W. Lusk
U.S. Department of Energy
National Energy Technology Laboratory
3610 Collins Ferry Road
Morgantown, WV 26507-0880
(304) 285-4145 or Mark.Lusk@netl.doe.gov
Fax: (304) 285-4403

For general information on the DOE process for implementing the National Environmental Policy Act, contact:

Carol Borgstrom, Director
Office of NEPA Policy and Compliance (GC-54)
U.S. Department of Energy
1000 Independence Avenue, SW
Washington, DC 20585-0103
(202) 586-4600 or leave message
at (800) 472-2756

Abstract:

This Environmental Impact Statement (EIS) evaluates the potential impacts associated with DOE's Proposed Action to provide financial assistance to NRG Energy, Inc. (NRG) and with NRG's proposed W.A. Parish Post-Combustion CO_2 Capture and Sequestration Project (Parish PCCS Project). DOE's Proposed Action would provide $167 million in cost-shared financial assistance to NRG under the Clean Coal Power Initiative (CCPI) Program to support construction and operation of NRG's Parish PCCS Project. The funding would be used for project design and development, procurement of capital equipment, construction, and CO_2 monitoring during the 35-month demonstration period of the integrated CO_2 capture and compression system.

NRG's proposed Parish PCCS Project would construct a CO_2 capture facility at its 4,880-acre W.A. Parish Plant in rural Fort Bend County near the small town of Thompsons, Texas. The capture facility would use an advanced amine-based carbon dioxide (CO_2) absorption technology to capture at least 90% of the CO_2 from a 250-megawatt equivalent (MWe) portion of the flue gas exhaust from Unit 8 at the W.A. Parish Plant. The project would be designed to capture approximately 1.6 million tons of CO_2 per year from the plant exhaust, which would otherwise be emitted to the atmosphere. The captured CO_2 would be compressed and transported via a new approximately 80-mile-long, 12-inch-diameter underground pipeline to the existing West Ranch oil field in Jackson County, Texas. The CO_2 would be used for enhanced oil recovery (EOR) and ultimately sequestered in geologic formations approximately 5,000 to 6,300 feet below ground surface (bgs).

DOE is the lead federal agency responsible for preparation of this EIS. DOE prepared the EIS pursuant to the National Environmental Policy Act (NEPA) and in compliance with the Council on Environmental Quality (CEQ) implementing regulations for NEPA (40 Code of Federal Regulations [CFR] 1500 through 1508) and DOE NEPA procedures (10 CFR 1021). The EIS evaluates the potential environmental impacts of the Parish PCCS Project as part of DOE's decision-making process to determine whether to provide NRG with financial assistance for its proposed project. The EIS also analyzes the No-Action Alternative, under which DOE would not provide financial assistance for the proposed project.

Comment Period:

DOE encourages public participation in the NEPA process. Comments postmarked by November 5, 2012 will be addressed in the Final EIS. DOE will consider late comments to the extent practicable.

INTENTIONALLY LEFT BLANK

TABLE OF CONTENTS

TABLES

FIGURES

INTENTIONALLY LEFT BLANK

ACRONYMS

Acronym	Definition
%	percent
°F	degrees Fahrenheit
AF	acre-feet
aka	also known as
Approx.	Approximately
Ar	argon
BEG	Texas Bureau of Economic Geology
bgs	below ground surface
BMPs	best management practices
ca.	circa
CCPI	Clean Coal Power Initiative
CCS	carbon capture and sequestration
CCTP	Climate Change Technology Program
CEQ	Council on Environmental Quality
CFR	Code of Federal Regulations
CO_2	carbon dioxide
COE	cost of electricity
CT	combustion turbine
dBA	decibel, A-weighted
DOE	U.S. Department of Energy
e.g.	for example (Latin: exempli gratia)
EIS	Environmental Impact Statement
EMT	emergency medical technician
EOR	enhanced oil recovery
EPA	U.S. Environmental Protection Agency
EPAct05	Energy Policy Act of 2005
ERC	emission reduction credit
ES&H	environmental safety and health
ETP	Energy Transfer Partners
FEMA	Federal Emergency Management Agency
FM	Farm-to-Market Road
FOA	funding opportunity announcement
FR	Federal Register
GHG	greenhouse gas
gpd	gallons per day

Acronym	Definition
H_2O	water
HAZMAT	hazardous material
HCl	hydrochloric acid
HDD	horizontal directional drilling
HEC	Hilcorp Energy Company
HF	hydrofluoric acid
HGB MSA	Houston Galveston Brazoria Metropolitan Statistical Area
HRSG	heat recovery steam generator
HVTL	high-voltage transmission line
i.e.	that is (to say); in other words (Latin: id est)
lb/hr	pounds per hour
lbs	pounds
mD	millidarcies
MECT	Mass Emission Cap & Trade
mgd	million gallons per day
MLV	main line valve
MMTA	million metric tons per annum
MP	milepost
msl	mean sea level
MSA	Metropolitan Statistical Area
MTA	metric tons per annum
MW	megawatt
MWe	megawatt equivalent
N_2	nitrogen
NAAQS	National Ambient Air Quality Standards
NEPA	National Environmental Policy Act
NETL	National Energy Technology Laboratory
NGL	natural gas liquid
NH_3	ammonia
NNSR	Nonattainment New Source Review
NO	nitrogen oxide
NO_2	nitrogen dioxide
NOI	Notice of Intent
NOx	nitrogen oxides
NRG	NRG Energy, Inc.
NRHP	National Register of Historic Places
NWI	National Wetland Inventory

Acronym	Definition
O_2	oxygen
O_3	ozone
Parish PCCS Project	W.A. Parish Post-Combustion CO_2 Capture and Sequestration Project
PCCS	Post-Combustion CO_2 Capture and Sequestration
PM_{10}	particulate matter with a diameter of 10 microns or less
ppmv	parts per million by volume (1 ppmv = 0.0001%)
psia	pounds per square inch absolute
RCRA	Resource Conservation and Recovery Act
ROD	Record of Decision
ROI	region of influence
ROW	right-of-way
RRC	Railroad Commission of Texas
SO_2	sulfur dioxide
SO_3	sulfur trioxide
STEC	South Texas Electric Cooperative
SWPPP	stormwater pollution prevention plan
TCEQ	Texas Commission on Environmental Quality
TCV	Texas Coastal Ventures LLC
THC	Texas Historical Commission
TPWD	Texas Parks and Wildlife Department
tpy	tons per year
TSDF	treatment, storage, and disposal facility
TXDOT	Texas Department of Transportation
U.S.	United States
UIC	Underground Injection Control
USACE	U.S. Army Corps of Engineers
USDW	underground source of drinking water
VOC	volatile organic compound
WWTP	wastewater treatment plant

INTENTIONALLY LEFT BLANK

GLOSSARY

Term	Definition
“A-weighted” Scale	Assigns weight to sound frequencies that are related to how sensitive the human ear is to each sound frequency. Frequencies that are less sensitive to the human ear are weighted less than those for which the ear is more sensitive. A-weighted measurements indicate the potential damage a noise might cause to hearing.
100-year floodplain	Land that becomes or will become submerged by a flood that has a chance to occur every 100 years (1% annual chance of flooding).
500-year floodplain	Land that becomes or will become submerged by a flood that has a chance to occur every 500 years (0.2% annual chance of flooding).
Ambient noise level	Background noise associated with a given environment. Ambient noise is typically formed as a composite of sounds from many near and far sources, with no particular dominant sound.
Amines	A group of organic compounds of nitrogen, typically derived from ammonia, with one or more of the hydrogen atoms in ammonia replaced by one or more organic functional groups. Amines include amino acids and a wide range of primary, secondary, and tertiary amines used for dyes, pharmaceuticals, and gas treatment.
Aquifer	Underground geologic formation composed of permeable layers of rock or sediment that holds and/or transmits water.
Best Management Practice (BMP)	Method for preventing or reducing pollution impacts resulting from an activity. BMPs include non-regulatory methods designed to minimize harm to the environment.
Carbon dioxide (CO_2)	A common chemical compound, abbreviated as CO_2, composed of two oxygen atoms covalently bonded to a single carbon atom. CO_2 is a colorless, odorless, nonpoisonous, GHG created by combustion and emitted from natural and human activities, including the burning of fossil fuels to generate electricity and operate motor vehicles.
Cultural resources	Archaeological sites, historical sites (e.g., standing structures), Native American resources, and paleontological resources.
Cumulative effects	The impact to the environment that results from the incremental effect of the action when added to other past, present, and reasonably foreseeable future actions regardless of what agency or person undertakes such other actions. Cumulative effects can result from individually minor but collectively significant actions taking place over a period of time.
Decibel (dB)	Unit used to express the intensity of sound.
Dissolution	Process of dissolving a substance into a liquid.
Effluent	Waste stream flowing into the atmosphere, surface water, groundwater, or soil.
Emergent	Erect, rooted herbaceous plants, such as cattails and bulrush, which dominate wetlands.

Term	Definition
Endangered Species	Plants or animals that are in danger of extinction. A federal list of endangered species can be found in 50 CFR 17.11 (wildlife), 50 CFR 17.12 (plants), and 50 CFR 222.23(a) (marine organisms). Texas maintains its list of endangered species with the TPWD.
Environmental justice	The fair treatment and meaningful involvement of all people regardless of race, color, national origin, or income with respect to the development, implementation, and enforcement of environmental laws, regulations, and policies. Fair treatment means that no group of people, including racial, ethnic, or socioeconomic groups, should bear a disproportionate share of the negative environmental consequences resulting from industrial, municipal, and commercial operations or the execution of federal, state, local, and tribal programs and policies. Executive Order 12898 directs federal agencies to make achieving environmental justice part of their missions by identifying and addressing disproportionately high and adverse effects of agency programs, policies, and activities on minority and low-income populations.
Erosion	The process by which particles of soils or other material are removed and transported by water, wind, and/or gravity to some other area.
Fault	A subsurface fracture or discontinuity in geologic strata, across which there is observable displacement as a result of earth movement.
Floodplain	Flat or nearly flat land adjacent to a stream or river that experiences occasional or periodic flooding.
Flue gas	Residual gases after combustion that are vented to the atmosphere through a flue or chimney.
Formation	The primary unit associated with formal geological mapping of an area. Formations possess distinctive geological features and can be combined into “groups” or subdivided into “members” or “units”.
Fossil fuel	Coal, oil, or natural gas, formed from vegetation and animals under high pressure and temperatures during a past geological age.
Fresh water	Water with bacteriological, physical, and chemical properties that make it suitable for beneficial use. (e.g., with TDS concentrations less than 1,000 mg/L).
Fugitive dust	Airborne particulate matter, typically associated with disturbance of unpaved haul roads, wind erosion of exposed surfaces, and other activities in which soil is removed and redistributed.
Greenhouse gas	Gas that contributes to the greenhouse effect by absorbing infrared radiation and ultimately warming the atmosphere. GHGs include water vapor, nitrous oxide, methane, CO_2, O_3, halogenated fluorocarbons, hydrofluorocarbons, and perfluorinated carbons.
Groundwater	Water obtained from an underground source (i.e., from an aquifer); may supply wells and/or springs.
Growth faults	Faults caused when sediment layers slump or subside at different rates. Growth faults are common along the Gulf of Mexico.

Term	Definition
Hazardous waste	Waste that exhibits at least one of four characteristics (ignitability, corrosivity, reactivity, or toxicity), or that is specifically listed by the EPA as a hazardous waste. Hazardous waste is regulated under RCRA Subtitle C.
Historic Property	Prehistoric or historic district, site, building, structure, or object included in, or eligible for inclusion in, the National Register of Historic Places.
Laydown area	Material and equipment storage area during the construction phase of a project.
Lithic scatter	Concentration of waste flakes resulting from the manufacture of stone tools.
Low income population	A community that has a proportion of low-income population greater than the respective average.
Major aquifers	Aquifers that produce large amounts of water over large areas.
Megawatt (MW)	Unit of power equal to 1 million watts. A power plant with 1 MW of capacity operating continuously for one year could supply electricity to approximately 750 households.
Minor aquifers	Aquifers that produce minor amounts of water over large areas or large amounts of water over small areas.
Minority	Individual(s) who are members of the following population groups: American Indian or Alaskan Native; Asian or Pacific Islander; Black, not of Hispanic origin; or Hispanic.
Minority population	Identified where either more than 50 percent of the population of the affected area is minority, or the affected area's minority population percentage is meaningfully greater than the minority population percentage in the general population or other appropriate unit of geographic analysis.
Mitigation	Efforts to lessen the severity or to reduce adverse impacts, including: avoiding the impact altogether by not taking a certain action or parts of an action; minimizing impacts by limiting the degree or magnitude of the action; repairing, rehabilitating, or restoring the affected environment; reducing or eliminating the impact over time by preservation; and compensating for the impact by replacing or providing substitute resources or environments.
National Ambient Air Quality Standards (NAAQS)	Uniform, national air quality standards established by EPA that restrict ambient levels of certain pollutants to protect public health (primary standards) or public welfare (secondary standards). Standards have been set for CO, lead, NO_2, O_3, particulate matter, and SO_2.
National Environmental Policy Act (NEPA)	Signed into law on January 1, 1970, the National Environmental Policy Act (NEPA) declared a national policy to protect the environment and created the Council on Environmental Quality (CEQ) in the Executive Office of the President. To implement the national policy, NEPA requires that environmental factors be considered when federal agencies make decisions, and that a detailed statement of environmental impacts be prepared for all major federal actions significantly affecting the human environment.
Nitrogen oxides (NO_x)	A product of combustion by mobile and stationary sources and a major contributor to the formation of O_3 in the troposphere.

Term	Definition
Nonattainment	An area that does not meet air quality standards set by the Clean Air Act for specified localities and time periods; locations where pollutant concentrations are greater than the NAAQS.
Notice of Intent (NOI)	Notice that an EIS will be prepared and considered. It is published in the *Federal Register* as soon as practicable after an agency knows that an EIS is required for a proposed action.
Ozone, (O_3)	A form of O_2 found naturally in the stratosphere and that provides a protective layer for shielding the Earth from ultraviolet radiation. O_3 occurring in the lower atmosphere is harmful and is classified as a criteria pollutant.
Palustrine	Living or thriving in a marshy environment.
Particulate matter (PM)	Small particles of solid or liquid materials that, when suspended in the atmosphere, constitute an atmospheric pollutant.
Permeability	Rate at which fluids flow through the subsurface; reflects the degree to which pore space is connected.
Potable water	Water that is safe and satisfactory for drinking and cooking.
Prime farmland	Land that has the best combination of physical and chemical characteristics for producing food, feed, fiber, forage, oilseed, and other agricultural crops with minimum inputs of fuel, fertilizer, pesticides, and labor, and without intolerable soil erosion.
Produced water	Brine separated from produced oil or gas at an oil field. Produced water may also be called brine, salt water, or process water.
Proposed Action	The activity proposed to accomplish a federal agency's purpose and need, often requiring an analysis of potential environmental impacts. A proposed action includes the project and its related support activities (pre-construction, construction, and operation, along with post-operational requirements).
Pulverized coal	Crushed coal used to fuel a coal power plant. Currently the principal electric generation technology in the U.S.
Region of influence (ROI)	The physical area that bounds the environmental, sociologic, economic, or cultural features of interest for the purpose of analysis.
Riparian	Pertaining to, situated, or dwelling on the bank of a river or other body of water.
Scoping meeting	An early and open process for determining the scope of issues to be addressed and for identifying the significant issues related to a proposed action.
Scrubber	A device that removes noxious gases (such as SO_2) from flue gases by using absorbents suspended in liquid solution.
Scrub-shrub	Woody vegetation less than 20 feet (6 meters) tall. Species include true shrubs, young trees, and trees or shrubs that are small or stunted because of environmental conditions.
Sediment	Material that has been eroded, transported, and deposited by erosional processes, typically wind, water, and/or glaciers.

Term	Definition
Sedimentation	The process or action of depositing sediment.
Seismic	Pertaining to, characteristic of, or produced by earthquakes or Earth vibrations.
Sensitive receptor	As used in this analysis, any specific resource (i.e., population or facility) that would be more susceptible to the effects of the impact of implementing the proposed action than would otherwise be.
Sequestration	Process of injecting the CO_2 captured from an industrial or energy-related source into deep subsurface geologic formations for long-term storage.
Slipstream	The portion or percentage of the flue gas exhaust that is diverted to another location for alternative uses, including monitoring, research, or separate testing.
Stream	A continually, frequently, or infrequently flowing body of water that follows a defined course. The three classes of streams are: ephemeral—a channel that carries water only during and immediately following rainstorms; intermittent—a watercourse that flows in a well-defined channel during the wet seasons of the year, but not the entire year; and perennial—a watercourse that flows throughout the year or more than 90 percent of the time in a well-defined channel.
Sulfur dioxide (SO_2)	A heavy, pungent, colorless, gaseous air pollutant formed primarily by the combustion of fossil fuels.
Supercritical CO_2	CO_2 usually behaves as a gas in air or as a solid known as dry ice. If the temperature and pressure are both increased (above its supercritical temperature of 88ºF [31.1ºC] and 73 atmospheres [1073 psi]), it can adopt properties midway between a gas and a liquid, such that it expands to fill its container like a gas, but has a density like that of a liquid.
Surface water	All bodies of water on the surface and open to the atmosphere, such as rivers, lakes, reservoirs, ponds, seas, or estuaries.
Topography	The configuration of a surface including its relief and position of the natural and manmade features.
Topsoil	The upper native soil layer; generally the layer that supports plant growth.
Turbidity	Capacity of material suspended in water to scatter light. Highly turbid water is often called muddy, although all manner of suspended particles contribute to turbidity.
Underground Source of Drinking Water (USDW)	Any aquifer or part of an aquifer that (1) supplies any public water system; or (2) contains a sufficient quantity of groundwater to supply a public water system, and currently supplies drinking water for human consumption or contains fewer than 10,000 milligrams per liter of total dissolved solids; and (3) is not an exempted aquifer.
Vibration	Force that oscillates about a specified reference point. Vibration is commonly expressed in terms of frequency, such as cycles per second, Hertz, cycles per minute, or strokes per minute.
Viscosity	Measure of a fluid’s resistance to flow.

Term	Definition
Volatile organic compound (VOC)	A VOC is one of a group of carbon-containing compounds that evaporate readily at room temperature. As defined in 40 CFR 51.100(s), a VOC is any compound of carbon that participates in atmospheric photochemical reactions, excluding carbon monoxide, carbon dioxide, carbonic acid, metallic carbides or carbonates, and ammonium carbonate, and other organic compounds designated by EPA as having negligible reactivity.
Wastewater	A combination of liquid and water-carried wastes from residences, commercial buildings, and/or industrial facilities.
Wetland	An area that is inundated or saturated by surface or groundwater at a frequency and duration sufficient to support, and that under normal circumstances does support, a prevalence of vegetation typically adapted for life in saturated soil conditions. Wetlands generally include swamps, marshes, bogs, and similar areas and have the following general characteristics: (1) Vegetation typically adapted to inundated or saturated soil conditions; (2) Hydric soils or soils associated with low oxygen conditions; and (3) The area is inundated either permanently or periodically at mean water depths less than 6.6 feet, or the soil is saturated to the surface at some time during the growing season of the prevalent vegetation.

INTRODUCTION

The United States (U.S.) Department of Energy (DOE) prepared this Environmental Impact Statement (EIS) to evaluate the potential impacts associated with its Proposed Action to provide financial assistance to NRG Energy, Inc. (NRG) and with NRG's proposed W.A. Parish Post-Combustion CO_2 Capture and Sequestration Project (Parish PCCS Project). DOE's Proposed Action is to provide $167 million in cost-shared funding to support construction and operation of NRG's proposed Parish PCCS Project under the Clean Coal Power Initiative (CCPI) Program. Congress established the CCPI Program to enable and accelerate the deployment of advanced technologies to promote clean, reliable, and affordable electricity for the U.S. The CCPI operates as a cost-shared partnership between government and industry to develop and demonstrate advanced coal-based power generation technology at the commercial scale. DOE selected NRG's Parish PCCS Project and four other projects during the CCPI Round 3 solicitation.

NRG's proposed Parish PCCS Project would construct a CO_2 capture facility at NRG's 4,880-acre W.A. Parish Plant in rural Fort Bend County near the small town of Thompsons, Texas. The CO_2 capture facility would use an advanced amine-based carbon dioxide (CO_2) absorption technology to capture at least 90% of the CO_2 from a 250- megawatt equivalent (MWe) portion of the flue gas exhaust from Unit 8 at the W.A. Parish Plant. The project would be designed to capture approximately 1.6 million tons of CO_2 per year from the plant exhaust that the facility would otherwise emit to the atmosphere. The captured CO_2 would be compressed and transported via a new approximately 80-mile-long, 12-inch-diameter underground pipeline to the existing West Ranch oil field in Jackson County, Texas, near the town of Vanderbilt. The CO_2 would be used for enhanced oil recovery (EOR) and ultimately sequestered in geologic formations approximately 5,000 to 6,300 feet below ground surface (bgs).

DOE is the lead federal agency responsible for preparation of this EIS. DOE prepared the EIS pursuant to the National Environmental Policy Act (NEPA) and in compliance with the Council on Environmental Quality (CEQ) implementing regulations for NEPA (40 Code of Federal Regulations [CFR] 1500 through 1508) and DOE NEPA procedures (10 CFR 1021). The EIS also evaluates the potential environmental impacts of the Parish PCCS Project as part of DOE's decision-making process to determine whether to provide NRG with financial assistance for the project.

DOE'S PURPOSE AND NEED

The *purpose* of DOE's Proposed Action under the CCPI Program is to demonstrate advanced coal-based technologies at a commercial scale that capture and geologically sequester CO_2 emissions. The principal *need* addressed by DOE's Proposed Action is to satisfy the responsibility Congress imposed on DOE to demonstrate advanced coal-based technologies that can generate clean, reliable, and affordable electricity in the U.S. The CCPI Program selects projects with the best chance of achieving the program's objectives as established by Congress: commercialization of clean coal technologies that advance efficiency, environmental performance, and cost competitiveness well beyond the level of technologies currently in commercial service.

The proposed project would help DOE, through the CCPI Program, meet its congressionally mandated mission to support advanced clean-coal technology projects. This specifically includes those projects that have progressed beyond the research and development stage to a point of readiness for operation at a scale that, once demonstrated, can be readily implemented across the commercial sector. Post-combustion CO_2 capture offers the greatest near-term potential for reducing power sector CO_2 emissions because it can be used to retrofit existing coal-based power plants and can also be tuned for various levels of CO_2 capture, which may accelerate market acceptance (NETL 2010a). A successful commercial-scale demonstration of amine-based carbon capture technology at NRG's W.A. Parish Plant with beneficial use of the CO_2 at an existing oil field would also generate technical, environmental, and financial data from the design, construction, and operation of the CO_2 capture facility, pipeline, and EOR/ CO_2 monitoring facilities at the oil field. These data would be used to evaluate whether the deployed technologies could be effectively and economically implemented at a commercial scale.

NRG'S PROJECT OBJECTIVES

Consistent with DOE's requirements under CCPI Round 3, NRG identified the following objectives for the Parish PCCS Project:

- Demonstration of an advanced amine-based CO_2 absorption technology;
- Integration of a custom-built cogeneration plant into the project to meet the specific power and steam requirements of the CO_2 capture system;
- Demonstration of EOR with CO_2 sequestration in a nearby oil field; and
- Demonstration of a CO_2 monitoring program.

ALTERNATIVES CONSIDERED BY DOE

Section 102 of NEPA requires that agencies discuss the reasonable alternatives to the Proposed Action in an EIS. The term "reasonable alternatives" is not self-defining, but rather must be determined in the context of the purpose expressed by the underlying legislation. The purpose and need for a federal action determines the reasonable alternatives for the NEPA process. Any reasonable alternative to the Proposed Action must be capable of satisfying the purpose and need of the CCPI Program.

The alternatives considered by DOE were limited to the applications submitted to DOE in response to requirements specified in the CCPI Round 3 solicitation. DOE considered all the applications that met the mandatory eligibility requirements as expressed in the funding opportunity announcement. In a competitive process, DOE can only consider site or technology combinations included in the applications received. The applicant must provide at least a 50–50 cost share and bears the responsibility for designing and executing the project. DOE's action concerning these applications was to decide which projects would receive DOE financial assistance from among the eligible applications submitted. Unlike a project owned by DOE, when projects are selected in a competitive process in response to a funding opportunity announcement, DOE does not make decisions concerning the location, layout, design, or other features of the project. In other words, DOE must select among the eligible projects submitted to DOE by the applicants. DOE cannot re-write an applicant's proposal and thereby compromise an open, fair, and competitive funding opportunity. DOE's initial decision is to select projects to receive federal financial assistance for a project definition phase, prior to DOE's final decision.

After DOE selects a project for an award, the range of reasonable alternatives becomes the project as proposed by the applicant, any alternatives still under consideration by the applicant, and the no action alternative. DOE's final decision, documented in a Record of Decision (ROD), is to either accept or reject the project as proposed by the proponent, including its proposed technology and selected sites. However, DOE may specify mitigation measures that would be required as part of the proposed project.

No-Action Alternative

Under the No-Action Alternative, DOE would not provide cost-shared funding for the proposed Parish PCCS Project. In the absence of DOE cost-shared funding, NRG could still elect to construct and operate the proposed project; therefore, the DOE No-Action Alternative could result in one of two potential scenarios:

- The proposed Parish PCCS Project would not be built; or
- The proposed Parish PCCS Project would be built by NRG without benefit of DOE cost-shared funding.

DOE assumes that if NRG proceeded with project development in the absence of DOE cost-shared funding, the project would include the features, attributes, and impacts as described for the proposed project. However, without DOE participation, it is possible that the project would be canceled. Therefore, for the purposes of analysis in this EIS, the DOE No-Action Alternative is defined as the No-Build Alternative. This means that the project would not be built and environmental conditions would not change from the current baseline (i.e., no new construction, resource use, or CO_2 capture and storage would occur).

Therefore, under the No-Action Alternative, the project technologies (i.e., large-scale CO_2 capture and geologic sequestration) may not be implemented in the near term. Consequently, commercialization of these technologies for large-scale, coal-fired electric generation facilities would be postponed and may not be realized. This scenario would not contribute to the CCPI goals to invest in the demonstration of advanced coal-based power generation technologies that capture and sequester, or put to beneficial use, CO_2 emissions. While the No-Action Alternative would not satisfy the purpose and need for the Proposed Action, this alternative was retained for comparison to the effects of the proposed project, as required under CEQ Regulations (40 CFR 15012.14). The No-Action Alternative reflects the current baseline condition and serves as a benchmark against which the effects of the Proposed Action can be evaluated.

Alternative Project Applications Considered During the CCPI Procurement Process

DOE's alternatives for CCPI – Round 3 consisted of the other eligible applications received in response to FOA DE-FOA-0000042, *Clean Coal Power Initiative - Round 3, Amendments 005 and 006*. DOE received 36 applications that met the minimum eligibility requirements listed in the FOA under Round 3 of the CCPI. These applications provided DOE with a range of options for meeting the objectives of Round 3 of the CCPI. DOE reviewed each of the 36 applications that met minimum eligibility requirements to evaluate potential environmental consequences and made preliminary determinations regarding the level of NEPA review required. DOE documented the potential environmental consequences for each application in an environmental critique that was considered by the selection official. The environmental critique was summarized in a publicly available environmental synopsis, prepared in accordance with DOE's NEPA implementing regulations (10 CFR 1021.216). DOE also reviewed each eligible application for technical and financial merit. Through this review process, DOE considered both potential environmental consequences and the ability of each application to meet DOE's purpose and need.

Considering technical and financial merit along with the potential environmental impacts associated with each application's proposal, DOE ultimately determined that the proposed Parish PCCS Project and four other proposals would best meet the goals and objectives of the CCPI Program. After selection, DOE must complete a project-specific NEPA analysis for each selected project before making a final decision. Although each of these projects is eligible for cost-shared funding under CCPI, no other relationship exists among them. The selection and potential execution of each standalone project has no effect or bearing on the other projects.

NEPA PROCESS

DOE published a Notice of Intent (NOI) to prepare an EIS in the *Federal Register* (FR) on November 14, 2011, under Docket ID No. FR Doc. 2011–29333; (76 FR 70429). The NOI identified potential issues and areas of impact that would be addressed in the EIS. DOE also published notices in local newspapers announcing the public scoping meeting locations and times. DOE held public scoping meetings on November 30, 2011, at the Needville High School in Needville, Texas, and December 1, 2011, at the Jackson County Services Building in Edna, Texas. These two meetings were attended by a total of ten members of the public, including two elected officials, along with project staff from DOE, NRG, and other project partners.

The 30-day public scoping period ended on December 15, 2011. DOE received four scoping comments at the Public Scoping Meetings. These comments, which were delivered verbally at the November 30, 2011 meeting, involved questions about ownership of the pipeline and use of eminent domain to obtain property for the pipeline; availability for inspection of a certified payroll (i.e., to report prevailing wages according to the requirements of the Davis-Bacon Act); how much DOE funding would be provided for the project; water requirements for the CO_2 capture system; and any potential impact on consumers' electricity bills.

This EIS addresses potential impacts to the areas identified during both internal planning and public scoping for the proposed project. DOE encourages public participation in the NEPA process. Public comments on the draft EIS will be solicited for 45 days from the Notice of Availability published in the *Federal Register*. Public hearings will also be held in the project area.

DESCRIPTION OF NRG'S PROPOSED PROJECT

NRG's proposed project consists of four components: a CO_2 capture facility and supporting infrastructure at the W.A. Parish Plant, a CO_2 pipeline, EOR operations at the West Ranch oil field, and a CO_2 monitoring program at the oil field. The design, construction, and operation of the CO_2 capture facility and the design and implementation of the CO_2 monitoring program are the primary focus of the DOE's Proposed Action. The CO_2 pipeline and EOR components of NRG's proposed project are connected actions that would not be conducted if not for the proposed action. Each of the four components of the Parish PCCS Project is summarized below and in Table S-1. Figure S-1 shows the general location of the proposed project components. Figure S-2 presents an overall schematic of the PCCS concept.

1. ***CO_2 Capture Facility:*** The proposed project would retrofit one of the W.A. Parish Plant's existing coal-fueled units (Unit 8) with a post-combustion CO_2 capture system constructed within the existing W.A. Parish Plant site. A new natural gas-fired cogeneration plant, estimated to be 80 MW in size, would also be constructed on the plant property to produce the auxiliary power and steam needed by the proposed CO_2 capture system. The captured CO_2 would be compressed to the pipeline pressure (i.e., 2,115 pounds per square inch absolute [psia]) and dehydrated within the CO_2 capture facility before delivery to the CO_2 pipeline. The compressed CO_2 would be a supercritical fluid (i.e., resembling a liquid but expanding to fill space like a gas) with a density heavier than air and a very low viscosity (i.e., flows readily).
2. ***CO_2 Pipeline:*** Captured CO_2 would be transported via a new, approximately 80-mile-long, 12-inch-diameter underground pipeline to the West Ranch oil field, located near the city of Vanderbilt in Jackson County, Texas. The anticipated pipeline route includes mostly rural and sparsely-developed agricultural lands in Fort Bend, Wharton, and Jackson Counties. NRG plans to use existing mowed and maintained utility rights-of-way (ROWs) to the extent practicable to minimize environmental impacts and avoid sensitive resources. As proposed, NRG's pipeline would be collocated along or within existing mowed and maintained utility ROWs (i.e., high-voltage transmission line [HVTL] and pipeline ROWs) for approximately 85% of the route. A joint venture between NRG and Hilcorp Energy Company (HEC), known as Texas Coastal Ventures LLC (TCV), would operate the pipeline.
3. ***EOR Operations:*** The Parish PCCS Project would deliver up to 1.6 million tons of CO_2 per year to the West Ranch oil field, where the CO_2 would be injected into the 98-A, 41-A, Glasscock, and Greta sand units of the Frio Formation, which lie approximately 5,000 to 6,300 feet bgs. The portions of the West Ranch oil field in which EOR operations would be conducted are currently owned or leased by TCV. HEC has been contracted to conduct the EOR operations.
4. ***CO_2 Monitoring Program:*** TCV would implement a program to monitor the injection and migration of CO_2 within the geologic formations at the EOR site based on a CO_2 Monitoring Plan developed in cooperation with the Texas Bureau of Economic Geology (BEG). In addition to satisfying the CO_2 monitoring requirements of the CCPI Program, the CO_2 monitoring program that would be conducted at the West Ranch oil field would be designed to satisfy the monitoring, sampling, and testing requirements of the Railroad Commission of Texas (RRC) certification program for tax exemptions related to use of CO_2 for EOR and use of CO_2 from anthropogenic sources.

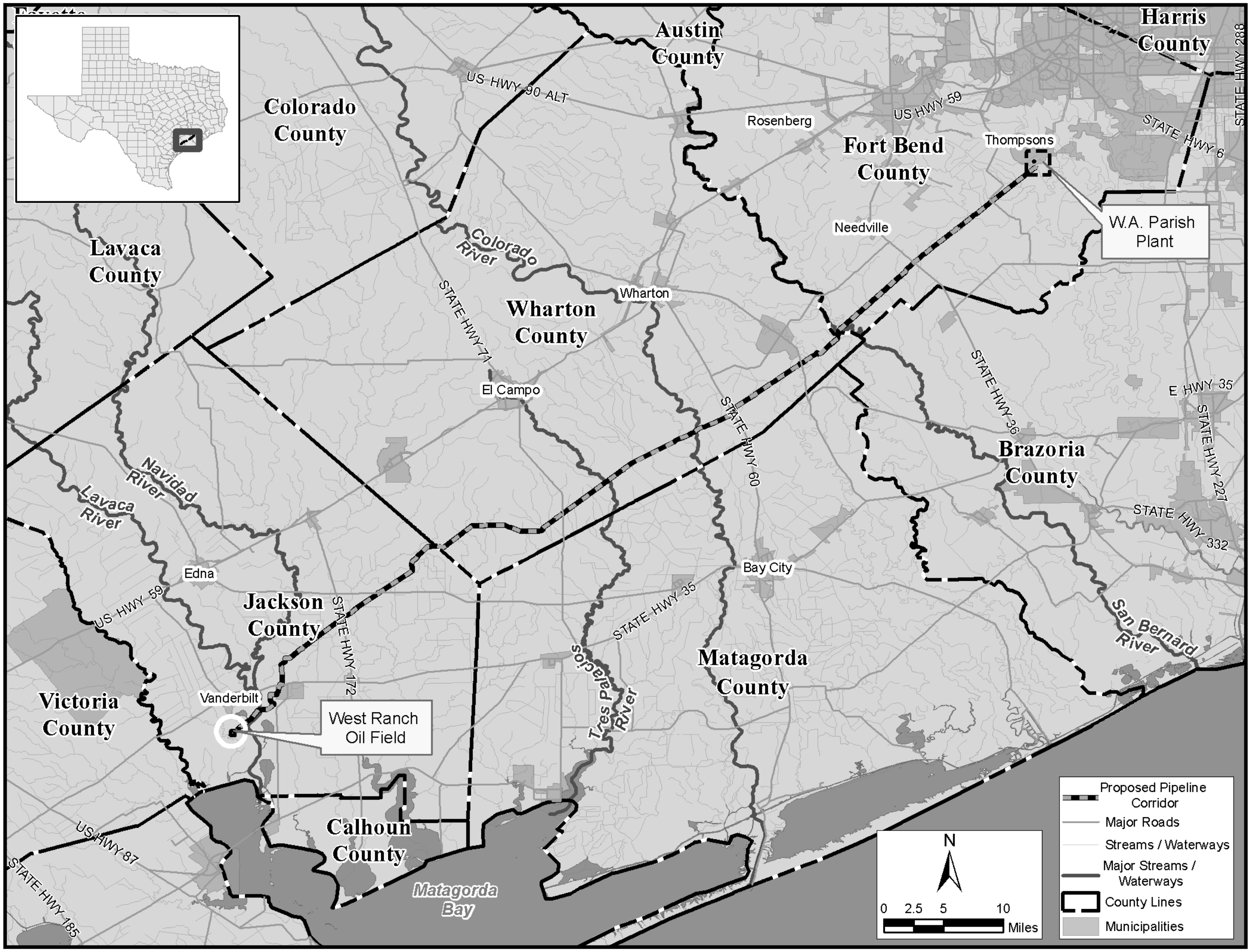


Figure S-1. Map of Project Area

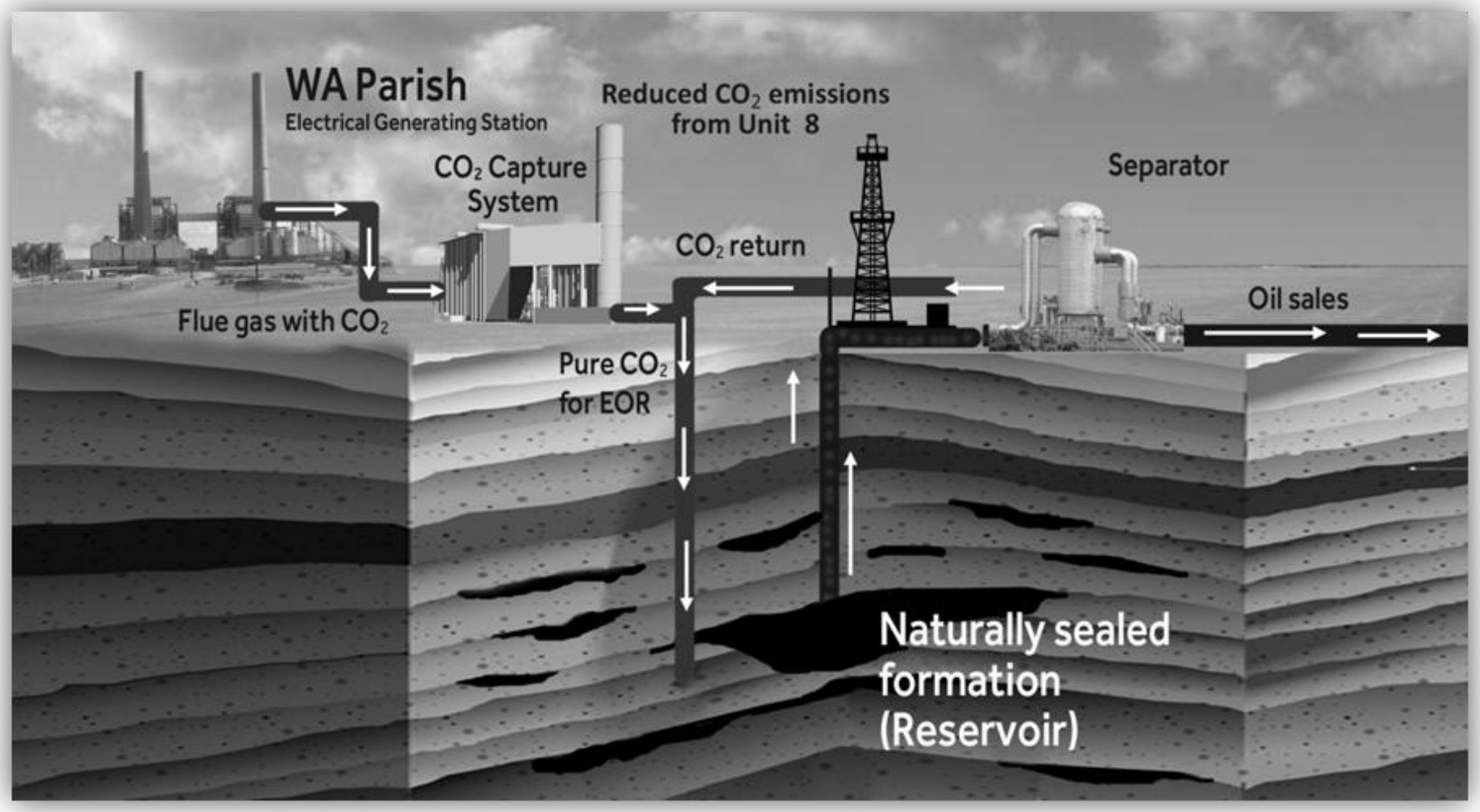


Figure S-2. Schematic of the PCCS Concept

Table S-1. Proposed Parish PCCS Project Features

Project Component	Description	Characteristics (Approximate Dimensions)
CO_2 Capture Facility	**Location:** A CO_2 capture facility would be constructed at NRG's W.A. Parish Plant. The facility would use an advanced amine-based absorption technology to capture CO_2 from a 250-MWe flue gas slipstream at the plant's 650-MW Unit 8 pulverized coal-fired electric generating unit. **Capacity:** Approximately 1.6 million tons of CO_2 per year.	**Facility Footprint:** 29 acres, as shown in Figure S-3
CO_2 Pipeline	**Route:** A new, approximately 80-mile-long, 12-inch-diameter underground pipeline would be collocated along or within existing mowed and maintained utility ROWs for approximately 85% of its length. Aboveground facilities would include approximately 12 main line valves (MLVs) and two meter stations. One meter station would be located within the W.A. Parish Plant. The second meter station would be located near milepost (MP) 76.5. **Operator:** TCV would own, operate, and maintain the pipeline.	**Construction ROW Width:** Generally 100 feet, reduced to 75 feet in some areas to minimize impacts to wetlands **Permanent ROW Width:** 30 feet **MLV Area:** 100 square feet **Meter Station Area:** 0.25 acres
EOR Operations	**Location:** The approximately 5,500-acre EOR area would be located within the portion of the West Ranch oil field currently owned/leased by TCV. A central CO_2 recycle facility would be constructed near the center of the EOR area in a disturbed area previously occupied by a gas processing facility. **Quantity:** TCV estimates that approximately 9 injection wells and 16 production wells would be used initially for EOR operations. As many as 130 injection wells and 130 production wells would be used over the 20-year span of the project. Existing wells at the West Ranch oil field would be used (i.e., refurbished or deepened as needed) to the extent practicable for the proposed project. New injection wells would be drilled if the existing wells cannot be reworked for injection. New wells would be installed on existing well pads to the extent practicable. As shown in Figure S-4, injection and production wells would be arranged in overlapping 5-spot patterns. Each 5-spot pattern would consist of four injection wells surrounding one production well. Each injection well would be installed to a sufficient depth that it could be used for injection into the 98-A, 41-A, Glasscock, and Greta sand units. A schematic illustrating the EOR process is provided in Figure S-5.	**Facility Footprint:** 5,500 acres **Well Configuration:** Overlapping, 5-spot patterns, 40 acres each pattern **Well Depths:** 5,000 to 6,300 feet bgs **Well Construction Areas:** 0.5 to 2.0 acres per well site **Well Operational Areas:** 0.01 to 0.5 acres per well site **CO_2 Recycle Facility Area:** 250 feet by 250 feet (1.5 acres)
CO_2 Monitoring Program	**Location:** Each monitoring well would be located within approximately 1,500 to 3,000 feet of an injection well. Existing wells would be used to the extent practicable to minimize the number of new wells needed. **Quantity:** Approximately 10 to 13 monitoring wells (i.e., one monitoring well for each 10 to 15 injection wells), including some wells monitoring above the injection zones and some monitoring within the injection zones.	**Well Depths:** 1,500 to 6,300 feet bgs **Well Construction Areas:** 0.5 to 2.0 acres per well site **Well Operational Areas:** 0.01 to 0.5 acres per well site
Access Roads	**W.A. Parish Plant:** As shown in Figure S-3, one road (approximately 1,000 feet long, included in 29-acre area discussed above) would be relocated. **CO_2 Pipeline:** Approximately 40 miles of existing roads would be used to access the construction ROW. Some roads may be upgraded (i.e., resurfaced and/or widened) to make them suitable for use by construction equipment. **West Ranch oil field:** Existing roads would be used to the extent practicable to access EOR and CO_2 monitoring areas. No new road construction is anticipated.	**Construction Width:** 30 feet **Permanent Width:** 20 feet

bgs = below ground surface; CO_2= carbon dioxide; EOR = enhanced oil recovery; MLV = main line valve; MP = milepost; MW = megawatt; MWe = megawatt equivalent; ROW = right-of-way; tpy = tons per year; TCV = Texas Coastal Ventures LLC; UIC = Underground Injection Control

Figure S-3. Aerial Photo of W.A. Parish Plant Showing Areas Related to CO_2 Capture Facility Construction and Operations
CO_2 = carbon dioxide; HDD = horizontal directional drill; kV = kilovolt

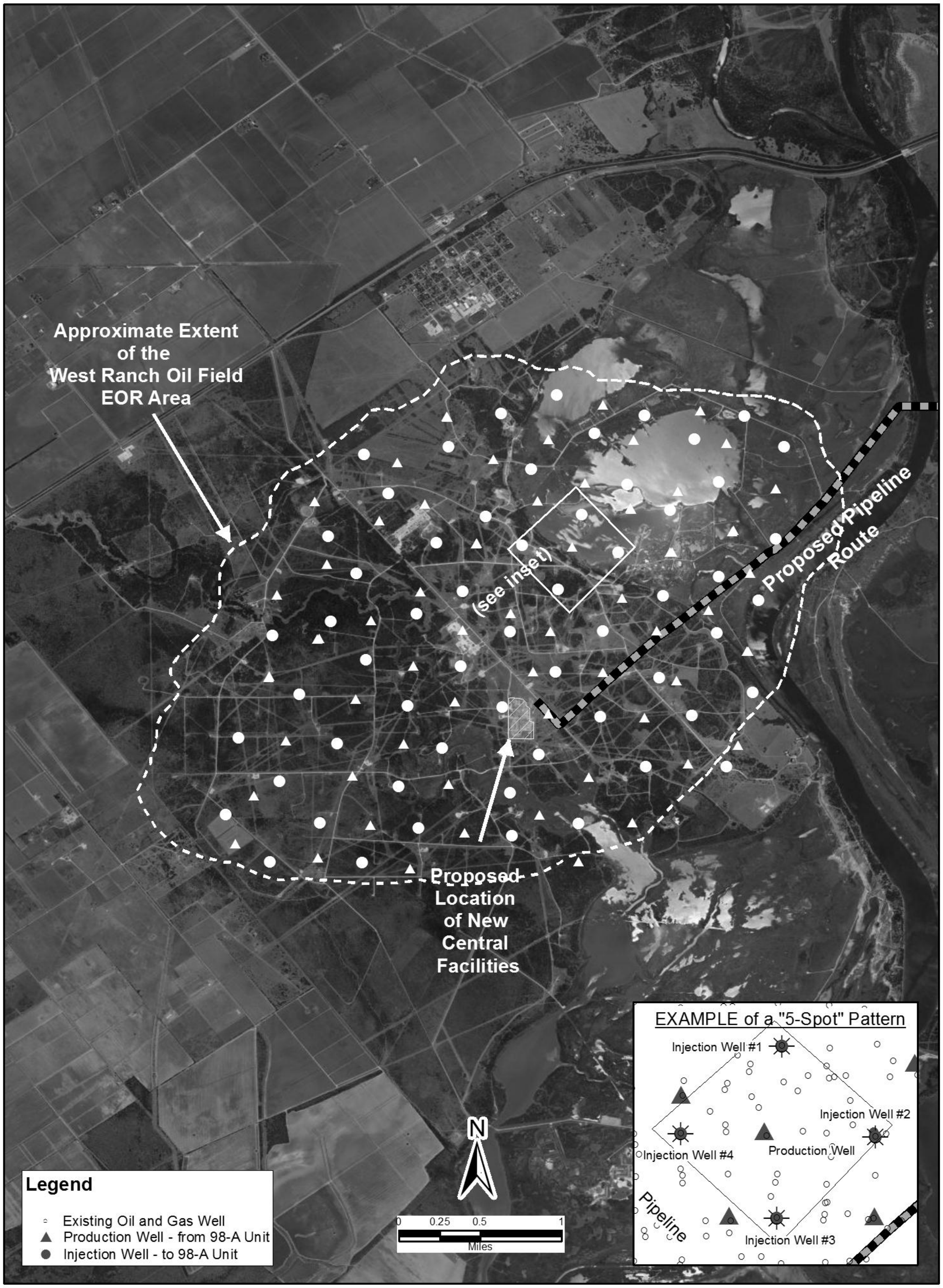


Figure S-4. Map of West Ranch Oil Field Showing Conceptual Arrangement of Injection and Production Wells for Proposed CO_2 Flood

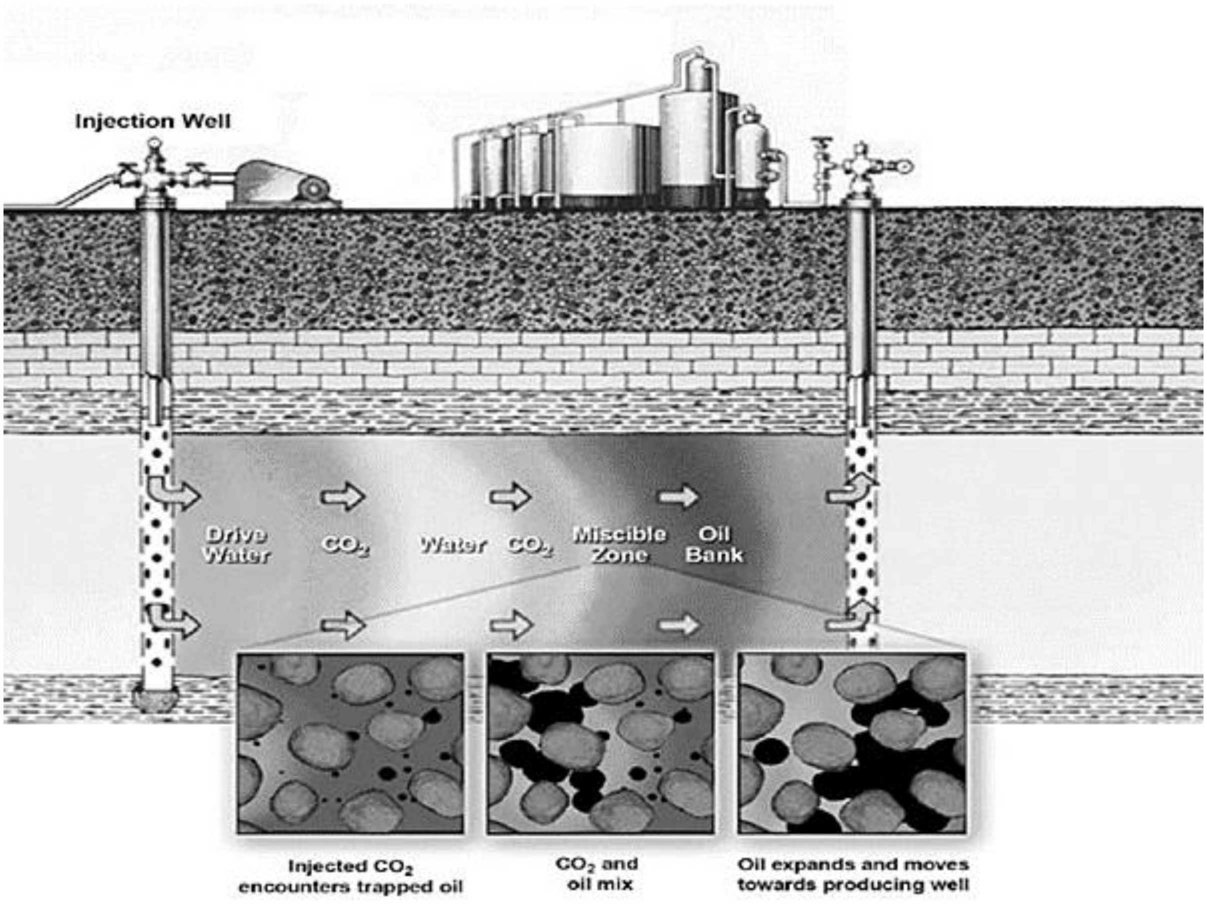


Figure S-5. Enhanced Oil Recovery Schematic

Source: NETL 2010.

The demonstration phase of the proposed project is currently scheduled to last for 35 months, according to the terms and conditions of the cooperative agreement between DOE and NRG. NRG would ultimately determine how long to continue running the CO_2 capture system following the demonstration phase based on a variety of factors, including potential future CO_2 legislation and regulations, process performance, and overall economics. For the purposes of this EIS, DOE assumes the CO_2 capture system would continue to operate for 20 years.

CO_2 capture facility construction is planned to start in late 2012, at the earliest, and take approximately 24 months to complete. Construction would be followed by a three- to six-month commissioning and start-up period to verify that all process systems function properly and achieve project requirements. The phases of construction would include site grading and preparation, the erection of administrative facilities, installation of piles and foundations, assembly of structural steel and building enclosures, and installation of mechanical and electrical systems. The number of construction workers would vary during the two-year construction period, ranging from 250 to 600 persons during the various phases of construction and averaging approximately 360 personnel. The largest demand for construction workers would likely occur approximately six months after the start of construction, when approximately 600 construction workers would be on site to construct the mechanical and electrical systems. Construction materials and equipment would be delivered by trucks and rail; however, construction-related rail traffic would be minimized to reduce the potential for disruption of coal deliveries. To operate the systems installed as part of the Parish PCCS Project, the W.A. Parish Plant would require approximately 15 additional full-time personnel.

Construction of the proposed CO_2 pipeline would take place over approximately six months beginning in April 2014. Construction techniques may include excavated trenching, boring, tunneling, and horizontal directional drilling (HDD). During pipeline construction, materials would be staged adjacent to the pipeline ROW or trucked in as necessary. The phases of construction would include clearing the ROW of trees and brush; grading the ROW; trenching or drilling, as applicable; pipe welding; pipe inspection; application of coating to welded areas; placing pipe; hydrostatic testing; backfilling; and site restoration. The pipeline would be covered by a minimum of three feet of compacted soil, but would be buried deeper (e.g., minimum of four feet in cultivated areas) or would be encased in reinforced concrete when needed to accommodate planned surface activities. During trenching in agricultural areas and wetlands, topsoil would be temporarily stored separately from other excavated material and would be replaced as the upper-most soil layer following pipeline construction. NRG's current project design would use HDD construction techniques in seven sections of the proposed pipeline corridor, including five large water body crossings (i.e., Big Creek, San Bernard River, Colorado River, Jones Creek, and Lavaca River). NRG's current design also includes conventional bores for most road crossings to minimize traffic disruption. The pipeline construction workforce would average 300 workers and reach a peak of 500 workers over the six-month construction period. Construction activities would generally be conducted 10 hours per day and six days per week. Approximately two full-time personnel would be required for pipeline operations.

The proposed pipeline would deliver CO_2 to the West Ranch oil field at a central CO_2 recycle facility, which would be constructed in an area previously occupied by an oil field gas processing facility. The CO_2 recycle facility would likely be constructed using skid-mounted equipment on gravel pads and would require a work force of approximately 12 workers during the three-month construction period. TCV estimates that approximately 9 injection wells and 16 production wells would be used initially. The number of injection and production wells would increase over the duration of the project to as many as 130 injection wells and 130 production wells. TCV plans to use existing wells (i.e., refurbished or deepened as needed) to the extent practicable for the proposed project. New injection wells would be drilled if the existing wells cannot be reworked for injection. All new injection wells would require UIC permits and TCV would install the new injection wells in accordance with the design standards specified by the RRC UIC Program. New wells would be installed on existing well pads to the extent practicable. Because some EOR operations would be automated, TCV anticipates that no additional operations personnel would be required for the EOR operations. TCV and the BEG are in the process of developing a CO_2 Monitoring Plan, scheduled for completion in early 2013, to define the particular activities that would be conducted as part of the CO_2 monitoring program for the West Ranch oil field. TCV and the BEG would conduct a variety of monitoring and modeling activities as part of this program to monitor the injection and migration of CO_2 within the geologic formations at the EOR site. As discussed in Chapter 2 of this EIS (Proposed Action and Alternatives), these monitoring and modeling activities may include preparing static and dynamic reservoir models of the proposed EOR area; conducting well integrity reviews and inspections; performing borehole seismic surveys and/or gravity surveys; performing gas tracer tests; and conducting groundwater and soil gas monitoring around the proposed EOR area. Approximately three full-time personnel would be required to implement the CO_2 monitoring program.

Table S-2 summarizes some of the key requirements and characteristics of the Parish PCCS Project.

Table S-2. Project Requirements and Characteristics Summary

Requirement/ Characteristic	Description	Source/Provider
Potable Water	**W.A. Parish Plant:** *Construction*: Approx. 10,800 gpd; *Operations*: Approx. 450 gpd **Pipeline:** *Construction*: Approx. 12,750 gpd **West Ranch:** Negligible additional water	**W.A. Parish Plant:** *Construction*: W.A. Parish Plant and other local sources; *Operations*: Existing W.A. Parish Plant groundwater wells **Pipeline:** *Construction*: local sources **West Ranch:** On-site groundwater wells
Industrial Water	**W.A. Parish Plant:** *Construction*: Approx. 12,000 gpd over 24-month construction phase for dust control and general washdown and Approx. 3.5 million gallons for hydrotesting and system startup; *Operations*: Approx. 4 to 5 mgd (approx. 3.6 to 4.9 mgd for cooling tower make-up water, 0.1 mgd for the CT/HRSG, and 0.1 mgd for CO_2 capture system) **Pipeline:** *Construction*: Approx. 1.75 million gallons **West Ranch:** Negligible additional water needed; produced water would be used for anticipated industrial purposes	**W.A. Parish Plant:** Smithers Lake (new intake point) and existing W.A. Parish Plant groundwater wells **Pipeline:** *Construction*: trucked in or obtained from surface water **West Ranch:** On-site groundwater wells
Electricity Required during Operations	**W.A. Parish Plant:** *Operations*: Approx. 50 MW (full-load) **Pipeline:** *Operations:* To be determined during detailed design (for meter station) **West Ranch:** *Operations*: Approx. 36 MW for CO_2 compressor.	**W.A. Parish Plant:** proposed 80-MW cogeneration plant **Pipeline:** Drop line from existing retail power provider **West Ranch:** Purchase from existing retail power supplier
Sanitary Wastewater	**W.A. Parish Plant:** *Construction*: Approx. 5,625 to 11,250 gpd; *Operations*: Approx. 225 to 450 gpd **Pipeline:** *Construction*: Approx. 4,500 to 9,000 gpd; *Operations*: Negligible **West Ranch:** Negligible additional wastewater	**W.A. Parish Plant:** *Construction*: portable restroom trailers and local WWTP(s); *Operations*: W.A. Parish Plant WWTP **Pipeline:** *Construction*: portable restroom trailers and local WWTP(s); **West Ranch:** On-site septic system
Industrial Wastewater	**W.A. Parish Plant:** *Construction*: Approx. 3.5 million gallons (from hydrotesting and system startup); *Operations*: Approx. 7,200 to 36,000 gpd **Pipeline:** *Construction*: Approx. 1.75 million gallons **West Ranch:** Negligible additional water	**W.A. Parish Plant:** W.A. Parish Plant WWTP **Pipeline:** Disposed to ground or surface waterbodies per RRC and EPA regulations **West Ranch:** On-site injection well
Hazardous Waste Generation	**W.A. Parish Plant:** Reclaimer effluent would be generated by the CO_2 capture system at a rate of approx. 2,712 lbs per day; Approx. 24 truck shipments per year of reclaimer effluent would be removed from the W.A. Parish Plant. **Pipeline:** None **West Ranch:** None	**W.A. Parish Plant:** Licensed and approved off-site TSDF
Solid Waste Generation	**W.A. Parish Plant:** Rate of waste generation and number of shipments to be determined during detailed design **Pipeline:** Rate of waste generation and number of shipments to be determined during detailed design **West Ranch:** Rate of waste generation and number of shipments to be determined during detailed design	**W.A. Parish Plant:** WMI Coastal Plains or WMI Conroe **Pipeline:** Organic debris to be burned under controlled conditions within ROW; other waste to nearby landfill **West Ranch:** Recycled or landfarmed on-site or disposed of at VI Wolf, Inland Environmental, or other nearby landfill

Table S-2. Project Requirements and Characteristics Summary

<table>
<tr><th>Requirement/ Characteristic</th><th>Description</th><th>Source/Provider</th></tr>
<tr>
<td>Material Transport during Operations</td>
<td>W.A. Parish Plant: Approx. number of truck shipments per year for process materials required for CO_2 capture facility operation:
<table>
<tr><th>Materials</th><th>Truck Shipments</th></tr>
<tr><td>Amine-Based Solvent</td><td>24 per year</td></tr>
<tr><td>10% carbohydrazide</td><td>2 per year</td></tr>
<tr><td>Ferric chloride coagulant</td><td>20 per year</td></tr>
<tr><td>Polymer</td><td>2 per year</td></tr>
<tr><td>Caustic</td><td>2 per year</td></tr>
<tr><td>Sulfuric Acid</td><td>50 per year</td></tr>
<tr><td>Hypochlorite</td><td>50 per year</td></tr>
<tr><td>Sodium bisulfate</td><td>2 per year</td></tr>
</table>
Pipeline: None
West Ranch: To be determined during detailed design</td>
<td>W.A. Parish Plant: Commercial vendors, shipped by commercial carriers</td>
</tr>
<tr>
<td>Construction and Operational Emissions</td>
<td>Construction Emissions (tons):
<table>
<tr><th>Compound Emitted</th><th>Pipeline (2013)</th><th>CO_2 Capture Facility (2013)</th><th>CO_2 Capture Facility (2014)</th><th>Total</th></tr>
<tr><td>CO</td><td>8.73</td><td>5.59</td><td>5.27</td><td>19.6</td></tr>
<tr><td>NO_x</td><td>22.2</td><td>17.4</td><td>13.3</td><td>53.0</td></tr>
<tr><td>PM_{10}</td><td>1.60</td><td>1.37</td><td>1.15</td><td>4.12</td></tr>
<tr><td>SO_2</td><td>124.1</td><td>119.4</td><td>92.8</td><td>336.3</td></tr>
<tr><td>VOC</td><td>1.62</td><td>1.44</td><td>1.27</td><td>4.33</td></tr>
</table>
Operational Emissions (tpy):
<table>
<tr><th>Compound Emitted</th><th>CO_2 Capture Facility</th><th>CO_2 Recycle Facility</th><th>Total</th></tr>
<tr><td>CO</td><td>102.1</td><td>9.6</td><td>111.7</td></tr>
<tr><td>NO_x</td><td>37.6</td><td>10.3</td><td>47.9</td></tr>
<tr><td>PM_{10}</td><td>75.1</td><td>0.7</td><td>75.8</td></tr>
<tr><td>$PM_{2.5}$</td><td>71.7</td><td>0.7</td><td>72.4</td></tr>
<tr><td>SO_2</td><td>6.9</td><td>0.1</td><td>7.0</td></tr>
<tr><td>VOC</td><td>65.1</td><td>14.6</td><td>79.7</td></tr>
</table>
VOC and NO_x emissions from the CO_2 capture facility exceed major source thresholds; therefore, NRG must obtain and retire VOC emission reduction credits (ERCs) and NO_x Mass Emission Cap & Trade (MECT) allowances to reduce the total net project increases of these ozone precursors (i.e., NO_x and VOC) within the Houston Galveston Brazoria Metropolitan Statistical Area (HGB MSA). NRG would be required to purchase and retire 1.3 tons of credits or allowances, as applicable, for each ton of emission increase related to the Parish PCCS project.</td>
<td>Construction emissions are from material handling (e.g. dirt moving) and emissions from combustion of fuel (i.e., gasoline and diesel) in mobile sources, which are mainly non-road construction equipment.
Operational emissions related to the CO_2 capture facility are from the CO_2 capture system, the CT/HRSG, the cooling tower, the emergency generator, and fugitive sources. Operational emissions from the CO_2 recycle facility are estimated based on reported emissions for the CO_2 recycle facility located at the West Hastings oil field in Alvin, Texas.</td>
</tr>
</table>

Table S-2. Project Requirements and Characteristics Summary

<table>
<tr><th>Requirement/ Characteristic</th><th>Description</th><th>Source/Provider</th></tr>
<tr><td>Flue Gas Inlet and Outlet Characteristics</td><td>W.A. Parish Plant: General characteristics of the flue gas feed to CO_2 capture system, the treated flue gas vent to atmosphere, and the compressed CO_2 product stream are as follows (NRG 2012g):
<table>
<tr><th>Stream Description</th><th>Flue Gas Feed</th><th>Treated Flue Gas</th><th>CO_2 Product</th></tr>
<tr><td colspan="4"><i>Component Concentrations</i></td></tr>
<tr><td>H_2O</td><td>18.0%</td><td>9.8%</td><td>100 ppmv</td></tr>
<tr><td>CO_2</td><td>11.5%</td><td>1.4%</td><td>>99.96%</td></tr>
<tr><td>N_2</td><td>65.4%</td><td>82.3%</td><td>226 ppmv</td></tr>
<tr><td>Ar</td><td>0.8%</td><td>1.0%</td><td>5 ppmv</td></tr>
<tr><td>O_2</td><td>4.3%</td><td>5.5%</td><td><10 ppmv</td></tr>
<tr><td>SO_2</td><td>60.6 ppmv</td><td>0.0 ppmv</td><td>0.0%</td></tr>
<tr><td>SO_3</td><td>1.2 ppmv</td><td>1.5 ppmv</td><td>0.0%</td></tr>
<tr><td>NO</td><td>26.0 ppmv</td><td>32.7 ppmv</td><td>0.0%</td></tr>
<tr><td>NO_2</td><td>1.4 ppmv</td><td>0.1 ppmv</td><td>0.0%</td></tr>
<tr><td>HCl</td><td>2.2 ppmv</td><td>0.0 ppmv</td><td>0.0%</td></tr>
<tr><td>HF</td><td>0.6 ppmv</td><td>0.0 ppmv</td><td>0.0%</td></tr>
<tr><td>NH_3</td><td>1.3 ppmv</td><td><1 ppmv</td><td>0.0%</td></tr>
<tr><td>Amine-Based Solvent</td><td>0.0 ppmv</td><td><1 ppmv</td><td>0.0%</td></tr>
<tr><td>Acetaldehyde</td><td>0.0 ppmv</td><td>2.1 ppmv</td><td>0.0%</td></tr>
<tr><td colspan="4"><i>Other Characteristics</i></td></tr>
<tr><td>Temperature, ºF</td><td>165</td><td>114</td><td>102</td></tr>
<tr><td>Pressure, psia</td><td>14.6</td><td>14.7</td><td>2,115</td></tr>
<tr><td>Total Flow, lb/hr</td><td>2,723,940</td><td>2,108,470</td><td>438,780</td></tr>
</table></td><td>Flue gas obtained from 250-MWe flue gas slipstream of combustion exhaust gases from existing 650-MW coal-fired Unit 8
Treated flue gas vented from a new stack (i.e., the CO_2 scrubber vent).in the CO_2 capture facility;
CO_2 product is pumped to the CO_2 pipeline</td></tr>
</table>

% = percent; °F = degrees Fahrenheit; Approx. = Approximately; Ar = argon; CO = carbon monoxide; CO_2 = carbon dioxide; CT/HRSG = combustion turbine/heat recovery steam generator; ERC = emission reduction credit; gpd = gallons per day; H_2O = water; HCl = hydrochloric acid; HF = hydrofluoric acid; lbs = pounds; lb/hr = pounds per hour; MECT = Mass Emission Cap & Trade; mgd = million gallons per day; MW = megawatt; MWe = megawatt equivalent; N_2 = nitrogen; NH_3 = ammonia; NO = nitrogen oxide; NO_2 = nitrogen dioxide; NOx = nitrogen oxides; O_2 = oxygen; ppmv = parts per million by volume (1 ppmv = 0.0001%); $PM_{2.5}$ = particulate matter with a diameter of 2.5 microns or less; PM_{10} = particulate matter with a diameter of 10 microns or less; psia = pounds per square inch absolute; SO_2 = sulfur dioxide; SO_3 = sulfur trioxide; tpy = tons per year; TSDF = treatment, storage, and disposal facility; WWTP = wastewater treatment plant; VOC = volatile organic compounds

ALTERNATIVES CONSIDERED BY NRG

NRG considered a number of design alternatives for the various components of the Parish PCCS Project while preparing its response to the FOA and during the preliminary design phase that followed DOE selection. As discussed below, these alternatives included locations for the CO_2 capture facility, methods of CO_2 capture, locations for the EOR field, and pipeline routes to the selected EOR site.

During preparation of its proposal, NRG reviewed the many power plants that it owns or operates as candidates for a large, integrated PCCS project. The primary criteria for selection were for a plant to have a sufficiently large coal-fired unit and for the plant to be located in proximity to oil fields suitable for tertiary recovery. NRG's W.A. Parish, Big Cajun II, and Limestone Plants met these criteria. NRG selected the W.A. Parish Plant because more oil fields that are suitable for tertiary recovery occur in the area.

NRG's project and technology selection process for the CO_2 capture facility focused on the set of commercially available post-combustion CO_2 capture technologies. Among the technologies currently available, only a few chemical absorption processes using ammonia or aqueous amines were determined to be sufficiently proven for a commercial-scale application. For this reason, and because an ammonia absorption project had already been selected as a demonstration project by the DOE CCPI program, NRG elected to scale up a comparatively proven advanced amine technology.

Prior to selection of the EOR site for this project, NRG approached the owners of several oil fields in the vicinity of the W.A. Parish Plant that NRG believed would be suitable for tertiary recovery using EOR. During this selection process, NRG determined that, of the prospective teaming partners who own or operate fields suitable for EOR along the Texas Gulf Coast, HEC was the candidate most aligned with the proposed project objectives. As a result, affiliates of NRG and HEC entered into a joint venture (i.e. TCV) and collectively determined that the West Ranch oil field was the most suitable candidate for the CO_2 injection and EOR component of this project.

After selecting the West Ranch oil field as the EOR site for this project, NRG considered several potential pipeline routes to convey CO_2 from the W.A. Parish Plant to the West Ranch oil field. The seven primary route alternatives considered by NRG, as described in the EIS, included:

- alternatives for routing the CO_2 pipeline through the W.A. Parish Plant from the compressor station at the CO_2 capture facility to the adjacent CenterPoint ROW;
- five alternate routes collocated with other utility or railroad ROWs from the W.A. Parish Plant to the West Ranch oil field; and
- two options for the approach to the West Ranch oil field from the adjacent South Texas Electric Cooperative (STEC) ROW.

NRG selected the pipeline route shown in Figure S-1 because it would minimize the length of the pipeline to approximately 80 miles and would be collocated along or within existing mowed and maintained utility ROWs for approximately 85% of its length, which would minimize potential environmental impacts by allowing NRG to use existing maintained ROW during construction. Additionally, this route avoids several riparian corridors and population centers that would have been crossed by more southerly routes, but it is not so far north as to impact development around the U.S. Highway 59 corridor.

CHARACTERISTICS OF THE AFFECTED ENVIRONMENT

The affected environment, also referred to as the region of influence (ROI), for the project was defined for each of 18 environmental resource areas depending on the extent of potential impacts resulting from plant and infrastructure construction and operation. The ROI includes, at a minimum, the proposed CO_2 capture facility areas at the W.A. Parish Plant, the proposed CO_2 pipeline corridor, and the proposed EOR area at the West Ranch oil field. However, the size of the ROI varies by resource depending on the extent of potential impacts on respective resources. Table S-3 summarizes the affected environment for each of the 18 resource areas. The affected environment for each of these resources is described in greater detail in Chapter 3 of the EIS.

Table S-3. Affected Environment of the Parish PCCS Project

Resource	Existing Conditions
Air Quality and Climate	With the exception of ozone in Fort Bend County, the National Ambient Air Quality Standards (NAAQS) promulgated by the EPA are being attained in the three counties in which components of the proposed project would be located (i.e., Fort Bend, Wharton, and Jackson Counties). Fort Bend County, in which the CO_2 capture facility and related infrastructure, along with a portion of the pipeline corridor, would be constructed and operated, has been classified as a severe nonattainment area for ozone.
Greenhouse Gases	In 2009, estimated U.S. CO_2 emissions totaled 5,426 million metric tons, including 2,160 million metric tons of CO_2 from generation of electricity. Emissions of CO_2 in Texas accounted for approximately 11% of total U.S. CO_2 emissions (i.e., 605.5 million metric tons) in 2009. Currently, there are no Texas regulations limiting greenhouse gas (GHG) emissions, including emissions of CO_2.
Geology	The Frio Formation is made up of several massive sand units that have created a number of highly prolific oil and gas reservoirs, including the West Ranch oil field. At the West Ranch oil field, the Frio Formation is approximately 5,000 to 7,200 feet below ground surface (bgs) and is capped by the Anahuac Formation, consisting of over 400 feet of low permeability calcareous shale with some occasional interlaminated sand lenses. A study near Beaumont, Texas, estimated the average permeability (to liquid) of the Anahuac Formation is approximately $5.2x10^{-6}$ millidarcies (mD). At the West Ranch oil field, the Frio Formation has held large quantities of buoyant fluids (i.e., oil and gas) over geologic time, indicating that very little migration occurs, if any, through the overlying Anahuac Formation. The Catahoula Sandstone is a very coarse-grained, homogenous sandstone unit found at a depth of approximately 4,250 to 4,500 feet bgs into which excess produced water is currently reinjected at the West Ranch oil field. The Burkeville confining system, which overlies the Catahoula Sandstone, consists primarily of silt and clay with a typical thickness ranging from approximately 300 to 500 feet. Southeastern Texas exhibits low seismicity and there are no major mapped faults within or near the proposed project areas. The risk of seismic events (i.e., earthquakes) occurring within the proposed project area is therefore very low. The BEG conducted a geophysical-log-based evaluation of regional structural features in the vicinity of the West Ranch oil field, which identified two growth faults in the deep subsurface to the northwest and southeast of the West Ranch oil field. The shallowest expression of the two faults is approximately 2,500 feet below mean sea level (msl) and both faults extend through the Greta, Glasscock, 41-A, and 98-A sand units of the Frio Formation. An approximately 200-foot offset of geologic strata on either side of the fault to the northwest of the oilfield reveals the simple domal structure that is responsible for hydrocarbon trapping in the West Ranch oil field. Neither of these faults extends upward to land surface nor do they lie within the boundaries of the West Ranch oil field (Appendix I, Figures 4 through 8). There are no obvious or large-scale faults within the West Ranch oil field itself.
Physiography and Soils	The project area is located in the Gulf Coastal plain, which is a low-lying area that has a gradual rise from sea level (at the Gulf of Mexico) in the south and east up to an elevation of about 900 feet above msl to the north and the west. The physiography originated from the deposition of sediments around the margins of the Gulf of Mexico in fluvial-deltaic to shallow-marine environments. In the vicinity of the project, most of the land is nearly flat (< 1 percent slope) with very small areas of slightly sloping land (< 8 percent, mostly < 3 percent). Approximately 600 acres in the construction ROW is classified as Prime Farmland and less than 20 acres classified as more than slightly erodible (i.e., moderately to severely erodible).
Groundwater	The major aquifer beneath the proposed project area is the Gulf Coast Aquifer, which is divided into four hydrostratigraphic units: the Chicot Aquifer, Evangeline Aquifer, Jasper Aquifer, and the Catahoula Confining System (aka, the Catahoula Restricted Aquifer). The Catahoula Confining System is composed of (in descending order) the Catahoula Sandstone, the Anahuac Formation, and the Frio Formation. The Chicot and Evangeline Aquifers are the primary underground sources of drinking water (USDWs) in the area. The only minor aquifer in the proposed project area is the Brazos River Alluvium Aquifer, which terminates north of the project area, but is hydraulically connected to the Brazos River, which supplies water to Smithers Lake for use by the W.A. Parish Plant.

Table S-3. Affected Environment of the Parish PCCS Project

Resource	Existing Conditions
Surface Water	The W.A. Parish Plant is located within the Brazos River Basin, immediately south of Smithers Lake, which is a 2,430-acre man-made lake with a capacity of about 18,000 acre-feet (AF) of water. The W.A. Parish Plant uses approximately 34 to 50 mgd (38,000 to 56,000 AF per year) of surface water from Smithers Lake, which receives water from the Brazos River, and discharges storm water and treated wastewater to the lake through permitted outfalls. The pipeline corridor would traverse several Texas coastal river basins. Surface water bodies drain these basins from the northwest to the southeast across the generally low topographic relief of coastal Texas towards the Gulf of Mexico. The pipeline would cross 210 waterbodies (23 perennial streams/rivers, 32 intermittent or ephemeral streams/rivers, 3 ponds, and 152 canals/ditches), including three major rivers (i.e., the San Bernard, Colorado, and Lavaca Rivers). The proposed pipeline would cross six waterbodies designated as Ecologically Significant Stream Segments (i.e., Big Creek, the San Bernard River, Cedar Lake Creek [aka Caney Creek], the Colorado River, West Carancahua Creek, and the Lavaca River) and two waterbodies designated by the State of Texas as impaired (i.e., the San Bernard River, which is listed for bacteria, and Caney Creek, which is listed for bacteria and low dissolved oxygen). The West Ranch oil field is located near the juncture of the Lavaca, Lavaca-Guadalupe, and Colorado-Lavaca River Basins. Waterbodies within the oil field ROI include the Lavaca River, the Navidad River, Venado Creek, Garcitas Creek, the Menefee Lakes, Redfish Lake, and the Venado Lakes. The Lavaca Bay/Chocolate Bay Estuary and associated tributaries, including Garcitas Creek are designated by the State of Texas as impaired for low dissolved oxygen. Additionally, the nearby oyster waters of Lavaca Bay and Chocolate Bay are designated as impaired because of bacteria.
Wetlands and Floodplains	There are no wetlands located within the area proposed for the CO_2 capture facility, but some project infrastructure may be located within approximately 50 to 200 feet of a wetland (i.e., Smithers Lake and associated canals). Approximately 117 acres of wetlands are located within the proposed pipeline construction ROW, including the following types: 99 acres of palustrine emergent, 2 acres of palustrine scrub-shrub, 1 acre of palustrine forested, 9 acres of riverine, and 6 acres of drainage ditches. These 117 acres of wetlands include several large fallow rice fields, which are categorized as palustrine emergent wetland areas, and a large gulf cordgrass (*Spartina spartinae*) marsh between the Lavaca River and the West Ranch oil field. The northern portion of the EOR area at the West Ranch oil field includes a wetland near Menefee Lake, classified as estuarine and marine wetland/deepwater by the National Wetland Inventory (NWI). Venado Creek, which crosses the EOR area at the West Ranch oil field, is classified by the NWI as estuarine and marine wetland and fresh water emergent wetland. The NWI also identifies several small fresh water ponds and a small estuarine and marine deepwater habitat within the area. There are no wetlands located within the area proposed for the CO_2 recycle facility. The area proposed for the CO_2 capture facility is located outside of the 100-year and 500-year floodplains identified by FEMA, but some project infrastructure may be located within approximately 50 to 200 feet of a floodplain (i.e., Smithers Lake and associated canals). The proposed pipeline corridor crosses FEMA 100-year and 500-year floodplains in 32 locations including areas adjacent to the following waterbodies: Colorado River, Lavaca River, Blue Creek, Juanita Creek, San Bernard River, and Tres Palacios River. The EOR area at the West Ranch oil field includes the FEMA 100-year and 500-year floodplains of Venado Creek, the Lavaca River, Menefee Lake, and Menefee Bayou. The land area proposed for the CO_2 recycle facility is located outside of the FEMA 100-year and 500-year floodplains.
Biological Resources	The CO_2 capture facility ROI has been previously disturbed (i.e., cleared and graded) and provides poor habitat quality for most wildlife species. Most of the proposed pipeline corridor consists of previously cleared utility ROW, which is maintained a minimum of once every four years, and agricultural land. Approximately 10% of the proposed pipeline construction ROW is classified as natural systems. The dominant land cover types in the West Ranch oil field EOR area, which is currently used for oil and gas production and cattle pasture, are pasture/hay/grassland/herbaceous, shrub/scrub, developed (open space/low density), emergent herbaceous wetlands, and woody wetlands. Three federally listed endangered species (Whooping crane, West Indian manatee, and Texas prairie dawn-flower) potentially occur in the three-county ROI (i.e., Fort Bend, Jackson, or Wharton Counties), which is located within the Western Gulf Coastal Plain EPA Level III

Table S-3. Affected Environment of the Parish PCCS Project

Resource	Existing Conditions
	Ecoregion. The West Indian manatee is a marine species and its occurrence in the ROI is very unlikely. The Lavaca River may provide suitable habitat, but there are no documented sightings of a West Indian manatee in the ROI and none were observed during field surveys. The ROI includes no suitable habitat (i.e., pimple mounds) for the Texas prairie dawn and no designated critical habitat for the whooping crane. A large wetland habitat is present within the proposed pipeline route between the West Ranch oil field and the Lavaca River which has the potential to provide habitat for the whooping crane. However, this area is adjacent to an active oil field, which would make it less attractive for use by whooping cranes than other wetland habitats in the vicinity. There are no documented sightings of whooping cranes within the ROI and none were observed during field surveys. The State of Texas has identified five previously used nesting areas within the ROI that are no longer in use, including a previously used colonial waterbird rookery near the W.A. Parish Plant, two previously used bald eagle nest near the W.A. Parish Plant in Fort Bend County, a previously used bald eagle nest along the boundary of Fort Bend and Wharton Counties, and a previously used bald eagle nests in Jackson County. An active bald eagle nest was observed adjacent to the proposed pipeline corridor in Wharton County near Jones Creek. The bald eagle is afforded federal protection under the Bald and Golden Eagle Protection Act and the Migratory Bird Treaty Act and is a protected species in the State of Texas. Other than the single active bald eagle nest, no state-listed species were identified during field surveys.
Cultural Resources	No State Archeological Landmarks, Texas Historic Landmarks, National Register historic buildings, or historic structures are located within the ROI. Three previously identified prehistoric lithic artifact scatters are situated within the ROI of the W.A. Parish Plant (along the southern shore of Smithers Lake and/or Dry Creek/Rabbs Bayou), but none were considered eligible by the Texas Historical Commission (THC) for listing in the National Register of Historic Places (NRHP). No known historic buildings, features, or above-ground properties listed, or eligible for listing, on the NRHP are recorded within the ROI for the proposed pipeline. Only one previously identified archaeological site is located within the ROI. This site is a prehistoric period lithic site located along the Lavaca River in Jackson County that has not been assessed for its eligibility to be listed in the NRHP. During the Phase I inventory study of the proposed pipeline construction ROW, six archaeological sites (2 prehistoric and 4 from historic periods) were identified within the pipeline construction ROW and a total of 12 buildings (6 in Fort Bend County, 5 in Wharton County, and 1 in Jackson County) greater than 45 years of age were identified within 164 feet (50 meters) of the pipeline construction ROW or associated access roads. The 12 buildings were predominantly National-style structures (6 structures), but also included two structures of undetermined design (due to inaccessibility) and single examples of a barn, a Spanish Eclectic structure, a railroad bridge, and an I-house. Most of the structures (i.e., 10 of 12) were built between ca. 1930 and the 1950s, with single examples noted from the 1890s to 1900s and 1920s to 1930s. No State Archeological Landmarks, Texas Historic Landmarks, National Register historic buildings or historic structures have been identified within the ROI of the West Ranch oil field. Nine previously identified archaeological sites have been identified within the ROI. Most of these sites are located along the boundaries of Venado Creek, with a single site by Menefee Lake. All of these sites are identified as prehistoric lithic scatters, except for one site which also contained prehistoric ceramics. Information regarding eligibility for listing on the NRHP is not available. The THC identified the following Native American Tribes that may have an interest in activities in the proposed project area: the Alabama-Coushatta Tribe of Texas, the Apache Tribe of Oklahoma, the Comanche Nation of Oklahoma, the Coushatta Tribe of Louisiana, the Kiowa Indian Tribe of Oklahoma, the Mescalero Apache Tribe of the Mescalero Reservation, the Tonkawa Tribe of Indians of Oklahoma, and the Tunica-Biloxi Indian Tribe of Louisiana. DOE sent letters to these tribes but has received no responses to date.

Table S-3. Affected Environment of the Parish PCCS Project

Resource	Existing Conditions
Land Use and Aesthetics	The proposed CO_2 capture facility would be constructed in Fort Bend County in areas within the existing W.A. Parish Plant that are currently in industrial use. The proposed CO_2 pipeline would cross lands used for utility (HVTL and pipeline) ROWs, livestock grazing, cultivated agriculture, and open space in Fort Bend, Wharton, and Jackson Counties. Approximately 85% of the pipeline corridor would be collocated with existing utility ROWs. The proposed EOR area would be constructed within the West Ranch oil field in an area used for oil and gas production and cattle pasture. Fort Bend, Wharton, and Jackson Counties have no land use plans, zoning, or development standards that would apply to the proposed project.
Traffic and Transportation	Access from Highway 59 to the W.A. Parish Plant is via Farm-to-Market Road (FM) 762 and Smithers Lake Road to the west side of the plant and via FM 2759, North Thompson Road, and Y U Jones Road to the east side of the plant. The proposed pipeline crosses several public roadways, all of which have two lanes and operate in a free-flowing manner with little congestion. Highway access to the West Ranch oil field is via State Highway 87 to FM 616 from the west or via FM 234 South to FM 616 from the east. The W.A. Parish Plant uses its rail facilities primarily for coal delivery. On average, the plant unloads two to three trainloads of coal each day, with each train averaging approximately 128 rail cars.
Noise	Dominant noise sources in the vicinity of the proposed CO_2 capture facility include power plant operation, coal train traffic and unloading, and use of heavy industrial vehicles. The nearest sensitive receptors to the proposed CO_2 capture facility are the rural residential communities near the perimeter of the W.A. Parish property (i.e., approximately 0.5 miles east, 1.5 miles to the southwest of the project site, 3 miles to the east, and 3 miles to the northwest) and a church located approximately 2.5 miles northeast of the proposed project site. The proposed pipeline would traverse primarily agricultural and rural residential areas, in which typical ambient noise levels are estimated to range between 28 and 38 dBA in calm weather conditions. Average noise levels are expected to be higher near roadways due to vehicle traffic. The existing noise at the West Ranch oil field comes from a number of sources, including truck traffic, drilling and associated activities, and well pumps and compressors. The nearest residential community is the town of Vanderbilt, located approximately 0.5 miles north of the northern perimeter of the West Ranch oil field and 2.3 miles north of the proposed location of the central CO_2 recycle facility. The nearest non-residential sensitive receptors include the Industrial Independent School District Junior and Senior High Schools, the Vanderbilt Baptist Church, and the St. John Bosco Catholic Church in the town of Vanderbilt.
Materials and Waste Management	The W.A. Parish Plant and the West Ranch oil field have current suppliers for the types of construction and operational materials that would be needed for the proposed project, including preferred providers for management of solid and hazardous wastes. The West Ranch oil field also operates injection wells permitted for the disposal of excess produced water. The W.A. Parish Plant is currently a conditionally exempt small quantity generator, but conforms to the requirements of a large quantity generator for consistency with other NRG facilities.
Human Health and Safety	Of the three counties in the ROI, Fort Bend County has the best overall health ranking (i.e., #9 of 221 Texas counties). Health rankings for Wharton and Jackson Counties are generally not as good as Fort Bend County for most indicators but neither county is consistently better than the other. Wharton County is ranked #61 of 221 Texas counties in overall health and Jackson County is ranked #57. All three counties in the ROI have better overall health rankings than over half of the counties in the state of Texas. Wharton and Jackson Counties had higher incidences of cancer deaths when compared to the average cancer rate for Texas, while Fort Bend County had a lower incidence of cancer deaths. Occupational injury data from 2008 for industries related to the proposed project (i.e., utility, pipeline, and non-residential construction; oil and gas extraction; and electric power generation) reflect total recordable incident rates of between 1.4 and 4.4 cases per 100 workers per year, including between 0.4 and 1.5 lost work day cases per 100 workers per year and between 0.7 and 2.3 days away from work, job transfer, or restriction cases per 100 workers per year. The fatality rate for the utility; construction; oil & gas extraction; and installation, repair, and maintenance industries in 2008 were between 3.9 and 23.9 fatalities

Table S-3. Affected Environment of the Parish PCCS Project

Resource	Existing Conditions
	per 100,000 workers. The population density in a small area east of the W.A. Parish Plant and a larger area southwest of the plant is 100 to 500 people per square mile. The population density west of the plant is 26 to 50 people per square mile and the areas north and south of the plant are mostly unpopulated. The majority of the pipeline traverses areas with population densities of five or less people per square mile, with certain segments that have a population density as high as 100 to 500 people per square mile. The population densities are higher within 15 miles of the W.A. Parish Plant than along the remainder of the pipeline corridor. The areas surrounding the West Ranch oil field are primarily unoccupied, except for the town of Vanderbilt, north of the oil field, which has a population density of 25 to 50 people per square mile. Areas to the southwest and northeast of the oil field have population densities of 5 to 25 people per square mile. The winds in the ROI are predominately from the direction of the Gulf of Mexico (i.e., from the south and southeast).
Utilities	The W.A. Parish Plant generates its own electricity (3,865 MW total); operates its own WWTP, treating approximately 4,000 gpd of sanitary wastewater; obtains potable water from existing groundwater wells; and obtains water for industrial use from Smithers Lake and existing groundwater wells. The combined units at the W.A. Parish Plant use a maximum of approximately 27,500 million cubic feet of natural gas per hour. The West Ranch oil field has utility service in place for potable water, produced water management, electricity, and natural gas. Crude oil produced at the West Ranch oil field is currently transported off site by truck. Existing pipelines are in place to receive crude oil shipments from the West Ranch oil field. Wastewater produced at the West Ranch oil field is primarily disposed of by underground injection along with excess produced water.
Community Services	The combined number of law enforcement officers in the project area (Fort Bend, Wharton, and Jackson Counties) is between 1.6 and 2.35 officers per 1,000 residents, as compared to the State of Texas average of 2.2 officers per 1,000 residents. The average crime rate in the three-county area is 1,680 crimes per 100,000 residents as compared to the State of Texas average of 4,239 crimes per 100,000 residents. The W.A. Parish Plant's Emergency Response Team includes Environmental, Safety, and Health (ES&H) professionals; firefighters, emergency medical technicians (EMTs); and hazardous material (HAZMAT) response personnel. Emergency response services (i.e., fire, ambulance, and HAZMAT response) within the project area are also provided by the Richmond Fire Department and several volunteer fire departments (i.e., Thompsons, Wharton County, Edna, and Vanderbilt). There are 12 hospitals in the three-county area with a total of 1,139 hospital beds (i.e., 5.62 beds per 1,000 people). The schools in the three-county area have an average of 14.28 students per teacher, as compared to the maximum of 20 students per teacher specified in the Texas Education Code.
Socioeconomics	Of the five counties in the ROI, Fort Bend County was the most populous at 585,375 persons and Jackson County was the least populous at 14,075 persons, according to the 2010 Census. Fort Bend County also had the highest population density (679.5 persons per square mile), while Jackson County had the lowest population density (17.0 persons per square mile). Fort Bend County is expected to more than triple its population to 1,917,470 persons by 2040, while Matagorda County is anticipated to shrink in population by 9% (i.e., a reduction of nearly 3,300 persons) by 2040. Brazoria County is expected to more than double its population to 664,503 persons by 2040. Jackson and Wharton Counties are anticipated to experience modest growth (6.9% and 3.7%, respectively). There are 357,884 housing units in the ROI of which 14.5% are vacant. Additionally, there are 150 hotel/motel facilities within the five county ROI. Within the ROI, the residents of Brazoria, Fort Bend, and Jackson Counties had higher average per capita incomes than the State of Texas, which was $23,863 in 2010, while Matagorda and Wharton County residents had slightly lower average per capita incomes than the State of Texas. The county with the highest unemployment rate in 2010 was Matagorda County at 6.4%. The unemployment rates for the other four counties were between 3.4% and 4.0% unemployed, all of which were lower than the unemployment rate for Texas, which was 5.7% in 2010.

Table S-3. Affected Environment of the Parish PCCS Project

Resource	Existing Conditions
Environmental Justice	Members of minority populations accounted for approximately 29.6% of the population of Texas and 27.6% for the U.S. in 2010. The percentage of minority populations in Fort Bend, Wharton, and Jackson Counties was 49.5%, 27.9%, and 18.8%, respectively, in 2010. These three counties did not exhibit minority populations that are meaningfully greater than the state or U.S. minority population percentages. Of the nine census tracts within the ROI, the largest percentage of minority populations was 42.9%. None of the nine census tracts exhibited minority populations that were meaningfully greater than the corresponding county minority population percentage. Therefore, there are no minority environmental justice areas of concern within the ROI. The median household income was $48,615 for Texas and $51,914 for the U.S. in 2010. Of the three counties analyzed, Fort Bend County had the highest median household annual income ($79,845) with 9.1% of residents below the poverty line, while Wharton County had the lowest annual income ($41,148) with 17.2% of residents below the poverty line in 2010. For the nine census tracts within the ROI, the highest median household income in 2010 was $70,321 and the lowest was $26,818. The lowest percentage below the poverty line was 6.0% and the highest percentage was 23.3%, as compared to 17.9% for the State of Texas and 15.3% for the U.S. in 2010. Neither the three counties nor the nine census tracts in the ROI exhibited a median household income below the 2010 Health and Human Services Poverty Guidelines. Also, the three counties in the ROI did not exhibit low-income populations that are meaningfully greater than the state or U.S. low-income population percentages. Additionally, none of the nine census tracts exhibited low-income populations that were meaningfully greater than the corresponding county low-income population percentage. Therefore, there are no low-income environmental justice areas of concern within the ROI.

AF = acre-feet; BEG = Texas Bureau of Economic Geology; bgs=below ground surface; ca. = circa; CO_2 = carbon dioxide; dBA = decibel, A-weighted; EMT = emergency medical technician; EOR = enhanced oil recovery; EPA = U.S. Environmental Protection Agency; ES&H = environmental safety and health; FEMA = Federal Emergency Management Agency; FM = Farm-to-Market Road; GHG=greenhouse gas; gpd = gallons per day; HAZMAT = hazardous material; HVTL = high-voltage transmission line; mD = millidarcies; mgd = million gallons per day; msl = mean sea level; NAAQS = National Ambient Air Quality Standards; NRHP = National Register of Historic Places; NWI = National Wetland Inventory; ROI = region of influence; ROW = right of way; USDWs = underground sources of drinking water

ENVIRONMENTAL IMPACTS

DOE evaluated the potential impacts of the Proposed Action and the No-Action Alternative in relation to the baseline conditions described in Chapter 3 and summarized above. More detailed discussions of potential impacts are provided in Chapter 3. Table S-4 summarizes the potential impacts for each of the 18 resource areas for the No-Action Alternative and the Proposed Action.

The EIS uses the following descriptors to qualitatively characterize impacts on respective resources:

- ***Beneficial*** – Impacts would improve or enhance the resource.
- ***Negligible*** – No apparent or measurable impacts would be expected; may also be described as "no impact" if appropriate.
- ***Minor*** – Barely noticeable but measurable adverse impacts on the resource. Mitigation measures may be considered for these impacts.
- ***Moderate*** – Noticeable and measurable adverse impacts on the resource. Mitigation measures would usually be considered for these impacts.
- ***Substantial*** – Obvious and extensive adverse effects, and potentially significant impacts on a resource. Mitigation measures would be sought to reduce these impacts.

Table S-4. Summary of Environmental Impacts

No-Action Alternative	Proposed Action
Air Quality and Climate	
No impacts. The W.A. Parish Plant, pipeline corridor, and EOR area at the West Ranch oil field would remain in their current conditions. No changes to air quality would occur.	**Negligible to minor adverse impacts with some beneficial impacts.** Construction of the CO_2 capture facility, CO_2 pipeline, and EOR/CO_2 monitoring infrastructure would result in short-term, localized increased tailpipe and fugitive dust emissions. Emission rates for criteria pollutants would be less than 1% of the total emissions in the ROI, except PM_{10} emissions during 2013, which would account for 3.1% of total ROI emissions. Emission rates for ozone precursors (i.e., volatile organic compounds [VOC] and nitrogen oxides [NO_x]) during the construction phase of the project would be lower than thresholds documented in the EPA rules for General Conformity (40 CFR 94.153). Operational emissions from the pipeline corridor would be negligible. Operational emissions of criteria pollutants from the CO_2 capture facility and related infrastructure (e.g., CT/HSRG) and the CO_2 recycle facility would be less than 1% of the total emissions in the ROI. Operational emissions of NO_x and VOC would exceed the thresholds documented in the Conformity Rules. However, as part of the NNSR permitting process, NRG would be required to provide offsets (i.e., ERCs or allowances) to reduce the total net project increases of ozone precursors (i.e., NO_x and VOC) within the HGB MSA. VOC ERCs are generated when the holder of an existing air permit reduces existing emissions and registers the emissions reduction with the TCEQ. MECT allowances were granted by the TCEQ to regulate the emissions of NO_x in the HGB MSA. Credits and allowances can either be obtained from a broker maintaining a “bank” of emissions credits and allowances generated by previously completed emissions reduction projects, or can be obtained directly from another company. In either case, the credits and allowances must be registered with the TCEQ to qualify as offsets for a new project, such as the proposed Parish PCCS Project. Also, NRG would be required to purchase and retire 1.3 tons of credits or allowances, as applicable, for each ton of emission increase related to the proposed project. Due to the 1.3 to 1 retirement ratio of ERCs and allowances, the proposed project would result in no net adverse impact on air quality in the HGB MSA with regard to ozone. Therefore, adverse impacts to air quality in the ROI due to operational emissions from the proposed project would be considered negligible to minor with some beneficial impacts in the form of elimination of SO_2 emissions from the Unit 8 flue gas slipstream, as well as reduced emissions of HCl, HF, and NH_3. As part of the state air permit application process, NRG would be required to finalize a detailed air quality analysis that includes dispersion modeling to compare predicted ambient air quality concentrations to the National Ambient Air Quality Standards (NAAQS). The detailed air quality analysis is not yet available. However, the Texas Commission on Environmental Quality (TCEQ) would not be able to issue the permit unless the modeling shows that NAAQS are met.
Greenhouse Gases	
Loss of potential beneficial impact. The W.A. Parish Plant, pipeline corridor, and EOR area at the West Ranch oil field would remain in their current conditions. However, without the project, there would be no reduction in GHG emissions from the W.A. Parish Plant and no commercial-scale demonstration of advanced coal-based power generation technologies to capture CO_2 for EOR and ultimate sequestration.	***Beneficial impacts.*** Construction of the CO_2 capture facility, CO_2 pipeline, and EOR/CO_2 monitoring infrastructure would generate up to approximately 4,900 tpy (4,400 metric tons per annum [MTA]) of CO_2 emissions over the two-year construction period. Operation of the CO_2 capture facility and CO_2 recycle facility would result in approximately 785,000 tpy (0.71 MMTA) of new CO_2 emissions. However, the proposed project would result in the capture approximately 1.6 million tpy (1.5 MMTA) of existing CO_2 emissions, resulting in a net reduction of approximately 815,000 tpy (0.74 MMTA) of CO_2 emissions during operations. The capture and geological storage of existing GHG emissions by the project would produce a minor beneficial cumulative effect on a national and global scale. The reduction in CO_2 emissions resulting from the Parish PCCS Project would incrementally reduce the rate of GHG accumulation in the atmosphere and help to incrementally mitigate climate change related to atmospheric concentrations of GHGs.

Table S-4. Summary of Environmental Impacts

No-Action Alternative	Proposed Action
Geology	
No impacts. The W.A. Parish Plant, pipeline corridor, and EOR area at the West Ranch oil field would remain in their current conditions. No changes to geologic resources would occur.	**Negligible to minor adverse impacts with some beneficial impacts.** Construction of the CO_2 capture facility, pipeline, and CO_2 recycle facility would result in negligible impacts to geologic resources. New well construction in the EOR area would result in removal of geologic media through the drilling process. This process would not be unique to the area and would not affect the availability of local geologic resources. Existing wells used by the project would be reworked, resulting in a potential beneficial impact to geologic resources by reducing the risk of leakage. Operation of the CO_2 capture facility and pipeline would not affect geologic resources. In the EOR area, the potential for CO_2 migration upward through the caprock seal is considered unlikely; however, leakage from one or more previously plugged and abandoned wells, oil-producing wells, injection wells, or observation wells might occur if any casing and/or cement placed in or around a well were to leak. To mitigate the potential for impacts related to casing or annular seal issues associated with wells in the proposed injection area, TCV and BEG would conduct a well integrity testing program prior to EOR operations and TCV would correct deficiencies prior to the use of such wells. These improvements to existing wells would result in a potential beneficial impact to geological resources by reducing the chance of leakage due to improperly sealed wells. Preliminary reservoir modeling indicates that injected CO_2 and associated zones of increased pressure would not be expected to migrate laterally outside the area at the West Ranch oil field that is leased and operated by TCV. As part of the proposed CO_2 monitoring program, TCV and BEG would conduct studies to detect migration of injected or displaced fluids, should migration occur, so that potential long term impacts to geologic resources may be minimized or avoided. No known major faults exist within the West Ranch oil field or within the area of maximum predicted EOR-induced impacts to geologic formations. Therefore, the potential for the proposed project to increase seismic activity or for seismic activity to impact proposed project activities or facilities is low. The addition of CO_2 to a geologic unit (i.e., a target geologic unit or an overlying unit, if leakage were to occur) could make the fluids within the unit more acidic. The creation of potentially more corrosive conditions could result in increased costs for later oil and gas development. However, DOE expects the injection of CO_2 to beneficially impact oil and gas resources at the West Ranch oil field by increasing production from the target geologic units. Furthermore, the presence of infrastructure for CO_2 floods may make oil production from other geologic units at the West Ranch oil field more feasible, which could result in an indirect beneficial impact.
Physiography and Soils	
No impacts. The W.A. Parish Plant, pipeline corridor, and EOR area at the West Ranch oil field would remain in their current conditions. No changes to physiography and soils would occur.	**Negligible to minor adverse impacts.** In general, potential minor impacts to physiography and soils during construction would include disturbance of soils from grading, soil excavation activities, earthwork compaction, installation of impermeable surfaces over soils at some locations, and increased soil erosion. At the CO_2 capture facility, up to 29 acres of soil within the W.A. Parish Plant boundary would be disturbed or lost. Soil in these areas is classified as Prime Farmland, but they have been previously impacted and would not be utilized for agricultural purposes. For the proposed pipeline development, up to 1,028 acres of soils would be disturbed; however, the disturbed land areas would be restored following construction and overall land use impacts would be minimized through use of existing ROW for most of its length. Approximately 600 acres in the construction ROW is classified as Prime Farmland and less than 20 acres classified as more than slightly erodible (i.e., moderately to severely erodible). In agricultural areas, impacts to soil would be minimized by segregating topsoil from underlying soil and placing the topsoil back as the top layer when trench is filled. For the EOR area, construction and operational activities would be conducted in existing operational areas; therefore, impacts to soils would be similar to existing impacts. Potential soil impacts in all construction areas would be avoided or mitigated as described in a

Table S-4. Summary of Environmental Impacts

No-Action Alternative	Proposed Action
	project-specific stormwater pollution prevention plan (SWPPP). Operational activities associated with the CO_2 capture facility, CO_2 pipeline, and EOR/CO_2 monitoring infrastructure would be anticipated to result in negligible impacts to soil resources, primarily due to disturbance of soils from vehicle traffic and an increased potential for erosion.
Groundwater	
No impacts. The W.A. Parish Plant, pipeline corridor, and EOR area at the West Ranch oil field would remain in their current conditions. No changes to groundwater resources would occur.	**Negligible to minor adverse impacts with some beneficial impacts.** The potential for groundwater contamination during construction is considered low as potential spills and unintentional releases of wastes or petroleum-based materials to groundwater would be avoided or mitigated as described in a project-specific SWPPP. Operation of the CO_2 capture facility would require an additional 0.2 to 0.3 mgd of groundwater from existing onsite wells (an approximately 13% increase as compared to current groundwater usage rates). The existing wells at the W.A. Parish Plant have capacity to supply the CO_2 capture facility with potential minor impacts to on-site groundwater supplies, such as a reduction in groundwater volumes in underlying aquifers, water level declines, and potential subsidence. There are currently no plans to withdraw groundwater or to discharge directly to groundwater during construction of the proposed pipeline. Water supply wells near the West Ranch oil field are not anticipated to be affected by injected or displaced fluids due to the relatively shallow depths of existing groundwater supply wells as compared to the depths of the proposed CO_2 injection wells in the Frio Formation (approximately 5,000 to 6,200 feet bgs) and the existing produced water injection wells in the Catahoula Sandstone (approximately 4,250 to 4,500 feet bgs); the presence of the approximately 400-foot-thick, low-permeability confining caprock formation (i.e., the Anahuac Formation) and the approximately 2,000-foot-thick low-permeability Burkeville confining system; and the absence of known faults in the EOR area. Although it is considered unlikely that CO_2 would leak from the injection zone, the possibility exists, in theory, for impacts to occur to shallower geologic units if leakage of CO_2 from the injection reservoir units were to occur. Increased groundwater acidity could result under such a hypothetical leakage scenario, potentially resulting in leaching of minerals and development of preferential flow pathways for migration of injected or displaced fluids. However, based on preliminary reservoir modeling results (Appendix H to this EIS), the probability of injected or displaced fluids migrating from the target injection zone into overlying aquifers is considered to be low. As part of the proposed CO_2 monitoring program, TCV and BEG would conduct studies to detect migration of injected or displaced fluids, should migration occur, so that potential long term impacts to groundwater resources may be minimized or avoided. In the EOR area, the potential for CO_2 to migrate upward through fractures in the caprock seal is considered unlikely; however, leakage from one or more wells (e.g., plugged and abandoned, oil-producing, injection, or observation wells) might occur if any casing and/or cement placed in or around a well were to leak. To mitigate the potential for impacts related to casing or annular seal issues associated with wells in the proposed EOR area, TCV and BEG would conduct well integrity testing prior to EOR operations and TCV would correct deficiencies prior to use of such wells. Additionally, existing wells used by the project would be reworked. Improvements to existing wells would result in a potential beneficial impact to groundwater resources by reducing the chance of leakage due to improperly sealed wells.
Surface Water	
No impacts. The W.A. Parish Plant, pipeline corridor, and EOR area at the West Ranch oil field would remain in their current conditions. No changes to surface waters would occur.	**Negligible to moderate adverse impacts.** Construction of project-related facilities has the potential to cause increased sedimentation and turbidity in adjacent waterbodies and increase the potential for surface water contamination from material spills. A SWPPP would be developed and implemented to avoid or minimize potential impacts to surface waters during construction activities. Negligible impacts to the surface water supply at W.A. Parish Plant would be

Table S-4. Summary of Environmental Impacts

No-Action Alternative	Proposed Action
	expected due to the approximately 12,000 gpd required during construction for dust suppression, vehicle wash down, and other construction-related uses. Operation of the CO_2 capture facility (including supporting infrastructure and facilities, such as the CT/HRSG and cooling water tower), would require approximately 3.5 to 4.9 mgd more surface water from Smithers Lake than is currently used by the W.A. Parish Plant. Including this approximately 10% increase in surface water usage, the W.A. Parish Plant would use a total of 38 to 55 mgd of surface water. This usage rate would be approximately 3% to 6% of the average Brazos River flow rate and approximately 8% to 13% of the Brazos River's critical low-flow rate. The portion of this water that would be related to the proposed project would account for approximately 0.5% of the average Brazos River flow rate and approximately 1% of the Brazos River's critical low-flow rate. Therefore, minor impacts on surface water supplies would be expected. NRG's projected surface water usage would also be well below NRG's current 99 mgd of surface water rights (i.e., 74 mgd from a surface water contract with the Brazos River Authority and 25 mgd of diverted Brazos River water that may be stored in Smithers Lake). During construction of the proposed pipeline, approximately 1.75 million gallons of water would be trucked in from outside sources or obtained from nearby surface water. NRG plans to discharge spent hydrotest water to upland areas according to RRC and EPA discharge permits and guidelines, as applicable. Construction would require 210 waterbody crossings. Three major rivers (i.e., the San Bernard River, the Colorado River, and the Lavaca River) and three other waterbodies (i.e., the man-made pond by FM 1994, Big Creek and Jones Creek) would be crossed by HDD. Additional mitigation measures (i.e., best management practices [BMPs], which would be specified in the project-specific SWPPP) would be employed for Ecologically Significant Stream Segments that are not crossed using HDD construction techniques (i.e., Cedar Lake Creek and West Carancahua Creek). Crossings of the San Bernard River and Caney Creek are not expected to exacerbate existing water quality impairments for these waterbodies. Construction-related impacts are expected to be negligible (i.e., for HDDs) to moderate (i.e., for open cuts) and temporary. Normal pipeline operations are not expected to impact surface waters. Negligible to minor impacts to surface water features in the West Ranch oil field ROI would be expected to occur as a result of construction activities within the proposed EOR area. During operations, the potential exists for a CO_2 well blow-out to occur, with some injected material being ejected and deposited into nearby surface waters. If that were to occur, such effects would be highly localized, minor, and readily remediated.
Wetlands and Floodplains	
No impacts. The W.A. Parish Plant, pipeline corridor, and EOR area at the West Ranch oil field would remain in their current conditions. No changes to wetlands and floodplains would occur.	**Negligible to minor adverse impacts.** There are no wetlands of floodplains located within the area proposed for the CO_2 capture facility at the W.A. Parish Plant or within the area proposed for the CO_2 recycle facility at the West Ranch oil field. However, construction of project-related facilities has the potential to cause increased sedimentation and turbidity in adjacent wetlands and increase the potential for contamination from materials spills. A SWPPP and SPCC would be developed and implemented to avoid or minimize potential impacts to wetland and floodplain areas during construction activities, resulting in negligible to minor impacts. Approximately 105 acres of wetlands would be temporarily impacted during pipeline construction and approximately 7 acres of wetlands may be permanently impacted. Topsoil in wetland areas would be segregated from other excavated material during trenching and returned to the surface to promote revegetation of disturbed areas and to restore preexisting soil conditions. NRG would reduce the width of the construction ROW in wetland areas from 100 feet to 75 feet and/or use of timber mats or low ground pressure equipment to minimize wetland impacts, as appropriate. Impacts to large riverine features and any adjacent wetlands would be avoided through the use of HDDs. Overall, the proposed project would result in minor, direct short-term

Table S-4. Summary of Environmental Impacts

No-Action Alternative	Proposed Action
	impacts to wetlands. Based on the current project design and field survey data collected to date, compensatory mitigation would not be required for NRG's proposed project by the U.S. Army Corps of Engineers (USACE) or the state of Texas. The pipeline route would cross FEMA 100-year and 500-year floodplains in 32 locations. The temporary presence of construction equipment and spoil piles would cause a minor temporary impact within the floodplain that could redirect flood flows in the event a flood occurred during construction in a floodplain. It is not expected that this impact would reach a level of endangering human health or property or conflict with any state, local, or federal floodplain ordinances as equipment and soil piles would be contained within the construction ROW and would represent relatively small, short-term obstructions as compared to the overall area of the floodplain. Following pipeline installation, the construction ROW would be returned to the original topography to the extent practicable. Three main line valves would be constructed within the FEMA 100-year floodplain in Wharton County. Changes to the flood elevation or the flow of water in the floodplain as a result of these valves would be negligible. No other aboveground facilities are planned within floodplain areas. BMPs (as specified in the site-specific SWPPP) would be implemented to avoid or minimize potential impacts to wetland and floodplain areas during construction activities, resulting in negligible to minor impacts. During operations, a 30-foot permanent ROW would be mowed and maintained along the pipeline route for pipeline inspection and maintenance activities, which could result in minor long-term impacts due to changed wetland functions in the approximately 36.6 and 42 acres of wetlands located within the proposed permanent ROW. Impacts to floodplains would be minor during pipeline operations.
Biological Resources	
No impacts. The W.A. Parish Plant, pipeline corridor, and EOR area at the West Ranch oil field would remain in their current conditions. No changes to biological resources would occur.	**Negligible to moderate adverse impacts.** The CO_2 capture facility and EOR area would be expected to have negligible impacts to biological resources as affected habitats have been previously disturbed for industrial and oil production uses. Impacts to wildlife from construction of the pipeline corridor would be negligible to minor. Approximately 85% of the proposed pipeline corridor would be constructed within or immediately adjacent to existing mowed and maintained utility corridors. Also, approximately 60% of the pipeline corridor is currently in agricultural use, which is of limited use to wildlife. The pipeline route was chosen to minimize the overall effect to wildlife and fragmentation of wildlife habitat. Construction activities, including land clearing, would cause a negligible loss of wildlife habitat. The potential would exist for invasive species to colonize newly disturbed areas following construction, which could result in long-term moderate adverse impacts to biological resources. Except in cultivated fields, unless requested by the landowner, NRG would plant areas of disturbed soil along the pipeline construction ROW following construction with an appropriate mix of seeds for perennial grasses and forbs native to the area or with a seed mixture requested by the landowner to reduce the potential for establishment of invasive plant species. Depending on the season in which construction is completed, NRG may also seed with a cold-weather annual grass species, such as Gulf Coast ryegrass (*Lolium multiflorum*), to establish a temporary vegetative cover until conditions become favorable for growth of perennial grasses and forbs. With the exception of one active bald eagle nest, no state-listed or federally listed species were identified during field surveys in the ROI. NRG would install the pipeline using HDD in the vicinity of the one observed active bald eagle nest such that construction activities would be separated by a distance of approximately 750 feet from the nest site. Therefore, no impacts to protected species would be expected. NRG would limit land-clearing activities in previously undisturbed areas to periods outside of the nesting season, to the extent practicable, to minimize the potential for impacts to migratory birds. If clearing vegetation during the nesting season is unavoidable, previously undisturbed areas within the construction area would be surveyed prior to construction to verify that nests with eggs or young would not be disturbed by construction activities.

Table S-4. Summary of Environmental Impacts

No-Action Alternative	Proposed Action
	In the process of finalizing the pipeline route and detailed design, NRG would coordinate with TPWD to identify potential waterbodies in which state-listed mussels or rare mussel habitat may be located. If suitable habitat is present in a waterbody that NRG plans to cross using open cut construction techniques, NRG would work with TPWD to determine whether a change in route or construction method (e.g., use of HDD construction techniques) would be warranted or whether surveys and other measures to avoid or minimize potential impacts to state-listed mussel species, as recommended by the TPWD, would be more appropriate.
Cultural Resources	
No impacts. The W.A. Parish Plant, pipeline corridor, and EOR area at the West Ranch oil field would remain in their current conditions. No changes to cultural resources would occur.	**Negligible adverse impacts.** DOE determined, and the THC has concurred, that no impacts to historic properties listed, or eligible for listing, in the NRHP would be expected from construction or operational activities for the CO_2 capture facility or EOR/CO_2 monitoring areas. Additionally, based on cultural resources survey data collected to date, the DOE has determined that no historic properties listed, or eligible for listing, in the NRHP would be impacted by the construction and operation of the proposed pipeline. Additional investigation activities (i.e., mechanized trenching) are pending to verify that no deeply buried archaeological deposits are present near several river crossings. DOE has submitted its findings regarding pipeline corridor surveys to the THC for review and consultation with the THC is ongoing.
Land Use and Aesthetics	
No impacts. The W.A. Parish Plant, pipeline corridor, and EOR area at the West Ranch oil field would remain in their current conditions. No changes to land use and aesthetic resources would occur.	**Negligible to moderate adverse impacts.** The proposed construction and operation of the CO_2 capture facility at the W.A. Parish Plant and EOR and CO_2 monitoring infrastructure at the West Ranch oil field is consistent with existing land use and would result in negligible to minor impacts. Construction of the proposed CO_2 pipeline would temporarily impact approximately 371 acres of agricultural lands, but no permanent loss of agricultural lands would occur. Less than 0.3 acres would be converted for aboveground pipeline facilities (one meter station and 12 main line valves). Impacts to aesthetic values would be negligible at the CO_2 capture facility and EOR field as the existing aesthetic character would generally remain unchanged. Along the proposed CO_2 pipeline route, minor to moderate aesthetic impacts to adjacent property owners would occur in some locations due to construction noise, truck traffic, fugitive dust emissions, and vegetation clearing. Operational aesthetic impacts would be negligible to minor and would be related to placement of pipeline markers, periodic vegetation clearing, and other maintenance activities. The impact of lighting during construction would be temporary and minor. The impact of lighting for operations at the proposed CO_2 capture facility, the EOR/CO_2 monitoring facilities, and the pipeline meter station would be negligible to minor as lighting would be consistent with existing operations. Lighting along the pipeline would be limited to the meter station. Meter station lighting would be down shielded to avoid interference with wildlife, which would result in minor impacts.
Traffic and Transportation	
No impacts. The W.A. Parish Plant, pipeline corridor, and EOR area at the West Ranch oil field would remain in their current conditions. Transportation resources would remain unchanged when compared to existing conditions.	**Negligible to minor adverse impacts.** The introduction of a temporary increase in traffic during construction (up to 1,100 workers) would be easily accommodated by the existing road systems with only minor temporary disruptions. Continuing operation of the W.A. Parish Plant, the pipeline, and the West Ranch oil field would have negligible effects as a relatively small number of commuting employees (20) would be added as well as a relatively small amount of additional material deliveries.

Table S-4. Summary of Environmental Impacts

No-Action Alternative	Proposed Action
Noise	
No impacts. The W.A. Parish Plant, pipeline corridor, and EOR area at the West Ranch oil field would remain in their current conditions. No changes to the noise environment would occur.	**Negligible to moderate adverse impacts.** Construction of the CO_2 capture facility would result an estimated 0.3 dB increase over existing noise levels for nearby receptors (i.e., nearby residential areas), which is below the threshold of human perception. Increased truck traffic during daytime hours may result in minor, short-term noise impacts along transportation corridors. Residences within 500 to 1,000 feet of pipeline construction would experience a short-term increase in ambient noise and vibrations from construction activity. Receptors near HDD locations could experience elevated temporary ambient noise levels as high as 78 dBA. Overall, noise and vibrations would result in minor to moderate impacts to receptors, depending on the distance from the receptor to the construction area. Construction and operations at the West Ranch oil field would result in an estimated 0.8 dB increase over existing noise levels for nearby receptors (i.e., in Vanderbilt), which is below the threshold of human perception, resulting in negligible to minor impacts to receptors.
Materials and Waste Management	
No impacts. The W.A. Parish Plant, pipeline corridor, and EOR area at the West Ranch oil field would remain in their current conditions. Conditions related to material use and waste generation would remain unchanged.	**Negligible to moderate adverse impacts.** The W.A. Parish Plant is currently a conditionally exempt small quantity generator and generates approximately 200 pounds of hazardous waste per year. Due to the generation of approximately 2,712 pounds per day of reclaimer effluent, a hazardous material, the W.A. Parish Plant would become a large quantity generator of hazardous waste. Approximately 24 shipments of reclaimer effluent would be sent to a permitted TSDF per year. The amounts sent for disposal would not substantially affect the capacities of the TSDF. Adequate waste disposal capacity exists within the ROI. Based on over 20 million tons of capacity available in waste disposal facilities that have been identified to date and the relatively low volumes of solid waste that would be generated by the proposed project (e.g., up to approximately 60 tons per year from the CO_2 capture facility), adequate capacity exists along the Texas Gulf Coast for solid waste disposal with negligible impacts to waste management service providers. Construction materials, equipment and supplies are readily available within the ROI and quantities required to support the proposed action are expected to be well within the capacity of material suppliers. Some specialized equipment may be required from outside the ROI; however, it is expected that this equipment would also be within existing supplier capacities. As a result, impacts to regional and national construction material resources and special equipment suppliers would be negligible.
Human Health and Safety	
No impacts. The W.A. Parish Plant, pipeline corridor, and EOR area at the West Ranch oil field would remain in their current conditions. Conditions related to human health and safety would remain unchanged.	**Minor adverse impacts.** The potential for worker injuries and fatalities would be present during the construction of the proposed CO_2 capture facility, CO_2 pipeline, and EOR/CO_2 monitoring infrastructure. Based on historical records for related industries, no worker fatalities would be expected. During facility operation, workers could be subject to physical and chemical hazards, which would be typical of those associated with similar power plant, pipeline, and oil field operations. An estimated nine to 12 OSHA recordable incidents would be anticipated during project construction based on national incidence rates for comparable industries. The potential for CO_2 pipeline ruptures or punctures is considered to be unlikely (i.e., the potential to occur between once in 100 years and once in 10,000 years). The upper bound impact from a pipeline release of CO_2 would be transient and reversible effects for up to 12 people. More severe impacts would affect less than one person for all other pipeline release scenarios. If a release were to occur with workers present, the workers would likely experience the physical effects of an accident (physical trauma, asphyxiation [i.e., displacement of oxygen in a small confined place], or frostbite from the rapid expansion of CO_2) or a higher concentration

Table S-4. Summary of Environmental Impacts

No-Action Alternative	Proposed Action
	exposure to CO_2 than the surrounding population. Potential exposure would be limited because the pipeline would be buried underground. Additionally, NRG plans to install 12 main line valves to stop the release of CO_2 should a puncture or rupture occur. These valves, along with pipeline pressure monitoring equipment, would be linked to the CO_2 capture system operations control room, which would be staffed at all times when the CO_2 capture system is in operation. In the event of a pressure drop indicating a pipeline rupture, the control room operator would shut down the CO_2 capture system and remotely activate the main line valves to prevent further damage to the pipeline and minimize impacts to people in the surrounding area and the environment. The potential for release of CO_2 from the EOR area is considered to range from unlikely (i.e., the potential to occur between once in 100 years and once in 10,000 years) to Incredible: (i.e., the potential to occur less than one time in 1 million years) with less than one person affected for all release scenarios. In the extremely unlikely occurrence of an injection well blowout (i.e., a sudden loss of CO_2 from failure of an injection well during operation), the main adverse outcome would be the potential for ejection of CO_2, possibly as dry ice particles, and formation fluids from the wellhead. Effects would be expected to be localized to the area around the affected wellhead and events of this type would be avoided or minimized by incorporating high pressure piping, overpressure protection (i.e., relief) valves, and blowout preventers into the design of the injection wells. A leak of amine-based solvent from a storage tank was evaluated. Such a release would be unlikely (i.e., with the potential to occur between once in 100 years and once in 10,000 years) and effects would be confined to the W.A. Parish Plant property. In this scenario, no nearby residents or the general public in the vicinity of the plant would be affected; however, plant workers would need to take appropriate response actions, since life-threatening concentrations of the solvent in air could occur within the plant site to a distance of 0.3 miles from the release. No nearby residents or general public in the vicinity of the plant would be affected beyond mild irritation if an amine-based solvent tank release occurred, although an odor may be detectable depending on the wind conditions.
Utilities	
No impacts. The W.A. Parish Plant, pipeline corridor, and EOR area at the West Ranch oil field would remain in their current conditions. No changes to utilities would occur.	**Negligible to minor adverse impacts with some beneficial impacts.** The construction and operations phases of the proposed project would increase demand for potable and industrial water; and wastewater treatment services. Construction-related impacts to water supplies would be short term and negligible to minor. Construction-related impacts to wastewater treatment would be negligible. Operations impacts to water supplies would be negligible. Operations of the CO_2 capture facility would result in negligible impacts to the natural gas supply as compared to existing use (i.e., much less than 1% of the current maximum usage). EOR operations may require additional natural gas supply and electricity, which may result in minor impacts to the local utility infrastructure. Beneficial impacts to oil supplies would be provided in the long term as a result of increased production of oil in the ROI as a result of EOR operations.
Community Services	
No impacts. The W.A. Parish Plant, pipeline corridor, and EOR area at the West Ranch oil field would remain in their current conditions. No changes to community services would occur.	**Negligible adverse impacts.** A temporary workforce of up to 1,100 workers would be required for construction of the proposed project. Long-term operation of the project would require up to 20 new employees. Many of these workers are expected to be employed from within the ROI. Negligible impacts on community services would be expected due to a relatively small population increase that would be related to the construction and operations phases of the Parish PCCS Project. Existing community services (i.e., law enforcement, emergency response, hospitals, and education) are expected to be adequate to address the needs of the population in the ROI, including project personnel.

Table S-4. Summary of Environmental Impacts

No-Action Alternative	Proposed Action
Socioeconomics	
Loss of potential benefit. The W.A. Parish Plant, pipeline corridor, and EOR area at the West Ranch oil field would remain in their current conditions. The region would lose the potential for employment, economic stimulus, and tax revenue.	***Beneficial impacts.*** The project (with approximately 1,100 construction-related jobs and up to 20 new jobs for operations) would be expected to contribute minor, long-term, beneficial impacts on the local economy and employment activities, as well as taxes and revenue through increased employment opportunities and expenditures in the local economy. Housing demand may increase slightly during construction if a portion of the 1,100 construction workers temporarily relocate to the area; however, this would be a negligible, short-term effect. The State of Texas offers many legislatively enacted production tax structures. Under the State's tax code, oil produced using methods involving the injection of CO_2 into an oil-bearing formation falls into two categories of tax rate - using CO_2 for EOR and using anthropogenic CO_2 for EOR. The Parish PCCS project would fall into both categories, which would result in an effective oil production tax rate of 1.15% (Texas Comptroller 2012c). The legislative record indicates the intent of these tax structures is to increase oil production in a manner that likewise increases tax revenue to the State. Consistent with the legislative intent, and as Section 2.3.4 of this EIS indicates, the Parish PCCS project is expected to increase the rate of oil production from the West Ranch oil. TCV's portion of the West Ranch oil field currently has approximately two million barrels of conventional proven oil reserves. TCV estimates that using CO_2 floods (i.e., EOR), the West Ranch oil field could produce an additional 55 to 75 million barrels of oil. This projected increase in oil production is expected to translate directly into additional revenues for the State of Texas, even after taking into account the tax exemptions related to use of CO_2 for EOR and use of CO_2 from anthropogenic sources.
Environmental Justice	
No impacts. The W.A. Parish Plant, pipeline corridor, and EOR area at the West Ranch oil field would remain in their current conditions. No environmental justice impacts would occur.	***No impacts.*** The project would not be expected to cause disproportionate adverse impacts on minority or low-income populations as they these populations do not occur in the ROI.

BEG = Texas Bureau of Economic Geology; BMPs = best management practices; CFR = Code of Federal Regulations; CO_2 = carbon dioxide; CT/HRSG = combustion turbine/heat recovery steam generator; EOR = enhanced oil recovery; EPA = U.S. Environmental Protection Agency; ERC = emission reduction credit; FEMA = Federal Emergency Management Agency; FM = Farm-to-Market Road; GHG=greenhouse gas; gpd = gallons per day; HCl = hydrochloric acid; HDD = horizontal directional drilling; HF = hydrofluoric acid; HGB MSA = Houston Galveston Brazoria Metropolitan Statistical Area; MECT = Mass Emission Cap & Trade; mgd = million gallons per day; MLV = main line valve; MMTA = million metric tons per annum; MTA = metric tons per annum; MW = megawatts; NAAQS = National Ambient Air Quality Standards; NH_3 = ammonia; NNSR = Nonattainment New Source Review; NOx = nitrogen oxides; NRHP = National Register of Historic Places; PCCS = Post-Combustion CO_2 Capture and Sequestration; PM_{10} = particulate matter with a diameter of 10 microns or less; ROI = region of influence; ROW = right of way; RRC = Railroad Commission of Texas; SWPPP = stormwater pollution prevention plan; TCEQ = Texas Commission on Environmental Quality; TCV = Texas Coastal Ventures LLC; tpy = tons per year; TSDF = treatment, storage, and disposal facility; UIC = Underground Injection Control; VOC = volatile organic compounds

POTENTIAL CUMULATIVE IMPACTS

DOE addressed the impacts of the Parish PCCS Project incrementally when added to the reasonably foreseeable impacts of other significant known or proposed projects within the geographic area in accordance with the cumulative impact requirements of NEPA (40 CFR 1508.7). As a result of the cumulative impacts analysis, DOE concluded that the Parish PCCS Project, in combination with other reasonably foreseeable future actions may result in cumulative impacts on the following resource areas:

- **Air Quality and Climate:** Emissions from the proposed project and other power or oil and gas projects in the same airsheds as the proposed project (e.g., the Colorado Bend Energy Center, Deer Park Energy Center Expansion, King Power Station, White Stallion Energy Center, and the Flag City Natural Gas Processing Plant) may have a cumulative impact on air quality. Due to emission limits imposed by the TCEQ as part of the Texas air permitting process, in conformity with the Texas SIP, significant adverse cumulative effects on air quality are not expected.

- **Greenhouse Gases:** The proposed project would be expected to contribute minor beneficial impacts by reducing CO_2 emissions. Other projects in the ROI that would include combustion of additional fossil fuels or other GHG emissions (e.g., Colorado Bend Energy Center, Deer Park Energy Center Expansion, King Power Station, White Stallion Energy Center, and the Flag City Natural Gas Processing Plant) would be expected to cumulatively emit additional amounts of GHGs within the ROI.
- **Physiography and Soils:** Each of the reasonably foreseeable future actions would cause some degree of soil disturbance, loss, and/or erosion, which may result in minor cumulative impacts.
- **Groundwater and Surface Water:** Each of the reasonably foreseeable future actions may require some amount of water for construction and/or operation. Minor cumulative impacts in terms of increased demand on groundwater and/or surface water, and the potential for contamination of water resources may occur in the ROI.
- **Wetlands and Floodplains:** The ETP NGL pipeline could interact with construction of the proposed CO_2 pipeline and cumulatively reduce wetland acreage by expanding the width of the mowed and maintained ROW in the existing utility corridor and/or increase the duration of temporary impacts (i.e., for wetland restoration). Cumulative impacts associated with these projects would be minor, however, as both projects would be required to avoid, minimize, and mitigate wetland impacts according to USACE permit requirements.
- **Biological Resources:** Each of the reasonably foreseeable future projects may result in some degree of wildlife habitat losses. The impacts to wildlife habitat resulting from the proposed project combined with other reasonably foreseeable future projects would be minor because comparable habitat is available throughout the region.
- **Cultural Resources:** Each of the reasonably foreseeable future actions may cause some degree of cultural resource disturbance. Thus, minor cumulative impacts would be expected on cultural resources.
- **Land Use and Aesthetics:** The ETP NGL pipeline could interact with construction of the proposed CO_2 pipeline and cumulatively make land unavailable for other uses temporarily, resulting in minor cumulative impacts on land use.
- **Transportation and Traffic, and Noise and Vibration:** Should construction of the proposed CO_2 pipeline coincide with construction of the ETP NGL pipeline and/or the bridge replacement in Fairchilds, construction-related, temporary cumulative impacts of increased traffic may occur. Additionally, minor to moderate, short-term cumulative effects of increased sound levels and perceptible vibrations may occur during project construction. However, current information suggests that the construction timeframes would be unlikely to overlap.
- **Materials and Waste Management:** Each of the foreseeable future actions would require construction materials and/or operational materials, which may result in minor cumulative adverse impacts on availability of materials and waste disposal facility capacity.
- **Utilities:** Minor beneficial cumulative impacts would be expected in terms of oil supplies. Minor cumulative adverse impacts on utility providers' supply and distribution capacities would be expected; however, the existing utility capacities within the ROI would be adequate to support the increased demand.
- **Community Services:** The planned new subdivision in Greatwood would contribute to population growth near the W.A. Parish Plant. Overall, minor impacts on community services in the Greatwood area may occur, though the contribution of the proposed project would be negligible.

- **Socioeconomics:** Construction projects could compete for skilled and unskilled labor and lodging in the short-term. However, beneficial short-term and long-term impacts result from increased employment opportunities, local spending, and related tax revenue.

The Parish PCCS Project would not contribute to adverse GHG impacts in the ROI. The estimated GHG reduction attributable to the proposed project would result in overall beneficial impacts. Cumulative impacts are also not expected for the geology, human health and safety, or environmental justice resources areas because the Parish PCCS Project is not expected to interact with other reasonably foreseeable future actions with regard to these resource areas.

CONCLUSIONS

As with the development of any large industrial project, the construction and operation of the proposed Parish PCCS Project, including the CO_2 capture facility and related infrastructure, the approximately 80-mile CO_2 pipeline, and EOR and related CO_2 monitoring activities at the West Ranch oil field, would impact the surrounding environment. Analyses included in this EIS indicate that the project could result in:

- potential beneficial impacts, primarily related to regional socioeconomics and the reduction of greenhouse gas emissions, but also related to some aspects of air quality, geology, groundwater, and utilities;
- potential moderate adverse impacts to surface water, biological resources, land use and aesthetics, noise, and materials and waste management;
- potential negligible to minor adverse impacts to air quality, geology, physiography and soils, groundwater, surface water, wetlands and floodplains, cultural resources, traffic and transportation, human health and safety, utilities, and community services; and
- no environmental justice impacts.

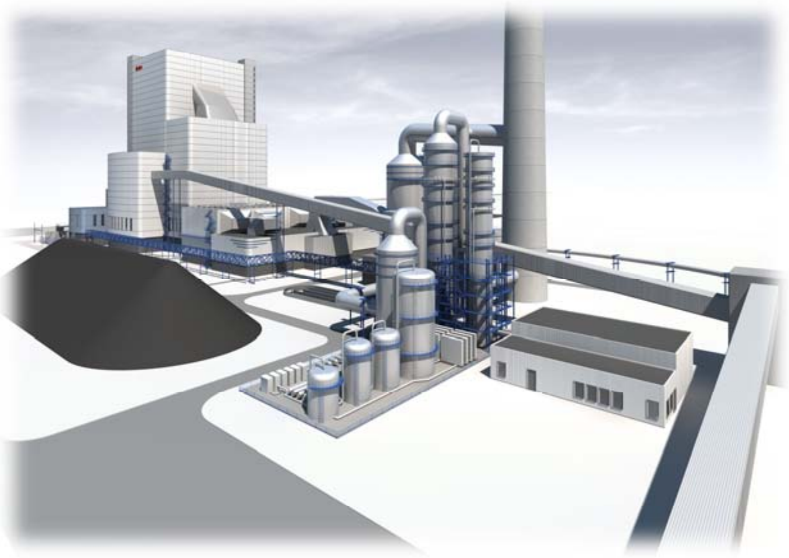
DOE’s Proposed Action would support the CCPI Program in demonstrating an advanced coal-based technology at a commercial scale that would capture, put to beneficial use, and geologically sequester CO_2 emissions. The proposed action would satisfy the responsibility Congress imposed on DOE to demonstrate advanced coal-based technologies that can generate clean, reliable, and affordable electricity in the U.S. The CCPI Program selects projects with the best chance of achieving the program’s objectives as established by Congress: commercialization of clean coal technologies that advance efficiency, environmental performance, and cost competitiveness well beyond the level of technologies currently in commercial service. Accelerated commercial use of these new or improved technologies will help to sustain economic growth, yield environmental benefits, and produce a more stable and secure energy supply.

DOE also recognizes the controversies surrounding the continued dependence on coal by the power industry and the need to address the associated environmental and climate change challenges related to the continued use of coal. However, as the most abundant fossil fuel resource in the U.S., coal will continue to play an important role in the nation’s energy supply. The proposed Parish PCCS Project would capture for EOR and ultimately sequester approximately 1.6 million tons per year of CO_2 that is currently emitted by the W.A. Parish Plant to the atmosphere. DOE considers the technological advancement and commercialization of carbon capture and storage, and beneficial use of CO_2, as important components of maintaining energy supplies while minimizing environmental impacts associated with using fossil fuel resources.

INTENTIONALLY LEFT BLANK

Non-confidential FEED study report

Special report for the Global Carbon Capture and Storage Institute



ROAD | Maasvlakte CCS Project C.V.

November 2011

SUPPORTED BY

GLOBAL CCS INSTITUTE

Government of the Netherlands

Co-financed by the European Union
European Energy Programme for Recovery

Title
Non-confidential FEED study report
Special report for the Global Carbon Capture and Storage Institute

Authors
Elvira Huizeling
Gerbert van der Weijde

Date
15 November 2011

ROAD | Maasvlakte CCS Project C.V.
Parallelweg 1
3112 NA Schiedam
The Netherlands

P.O. Box 133
3100 AC Schiedam
The Netherlands

Tel: +31 10 75 34 000
Fax: +31 10 75 34 040
E-mail: info@road2020.nl
Web: www.road2020.nl

Disclaimer
The ROAD project is co-financed by the Government of the Netherlands, the Global Carbon Capture and Storage Institute and the European Commission under the European Energy Programme for Recovery. The sole responsibility for the content of this publication lies with Maasvlakte CCS Project C.V.

This report for the ROAD project is being provided pursuant to the funding agreement between the Global Carbon Capture and Storage Institute and Maasvlakte CCS Project C.V. The provisions of that funding agreement are for the sole protection and legal benefits of the parties thereto, unless otherwise agreed in writing. No other person or entity shall have any direct cause of action or claim against any of the parties to that funding agreement arising from the publication, disclosure or distribution of this report and/or the utilization of or reliance on the content of this report.

This document is published on the Global CCS Institute's website in the interest of information exchange. The Global CCS Institute does not give any representation or warranty as to the reliability, accuracy or completeness of the information, nor does it accept any responsibility arising in any way (including by negligence) for errors in, or omissions from, the information.

Table of contents

Executive summary

In July 2009 Maasvlakte CCS Project C.V. ('MCP') submitted a project proposal to the European Commission. This marked the start of the 'ROAD project' ('Rotterdam Opslag en Afvang Demonstratieproject'; Rotterdam Storage and Capture Demonstration project). By applying for European funding, MCP committed itself to the funding program requirements, mainly related to the project schedule. The project team had to select a capture technology supplier within the given time constraints, while still respecting other important boundary conditions. For this reason, the project team developed a tailor made selection methodology.

Two parallel capture FEED studies were part of the selection. The FEED study described in this report is from the winning supplier. Not all details of the design can be disclosed as some part is proprietary information of the supplier. Nevertheless the information in this document is sufficient to form a good picture of the design of the future capture plant of the ROAD project.

Topics which are described in this report are:

- Project description - an overall description of the full chain process, including power plant, capture, compression, pipeline transport and storage
- Non-confidential process description of the capture and compression plant
- Description of power plant integration
- Simplified process flow diagrams (PFD) for the capture and compression plant
- Overall heat and material balances (complementing the PFD) for the capture and compression plants
- Overall process performance data
- CO_2 capture efficiency, total CO_2 capture capacity in tons per unit time, megawatt equivalent of facilities integrated with CO_2 capture (gross, net and explanation)
- A block flow diagram depicting the overall capture plant configuration and interconnecting streams
- Preliminary sized equipment list of major equipment (including: DCC, absorber, stripper, flue gas blower, and CO_2 compressor)
- Simplified plot plan showing capture and compression plant main equipment
- Capital cost estimates with breakdown to show equipment, engineering, indirect field costs and owner's costs.
- Operating costs broken down into variable costs, maintenance, staff costs, etc.
- CO_2 Metering – high level description of approach, locations, technology;
- CO_2 specification for transport after compression, including range of temperature, pressure and purity
- List of environmental emissions associated with the Project
- Health and safety issues in regular operation
- Construction programme showing key dates
- General project planning - summary of project progress against key milestones.

1. Project Introduction

1.1 Project Overview

ROAD is the **R**otterdam **O**pslag and **A**fvang **D**emonstratieproject (Rotterdam Capture and Storage Demonstration Project) and is one of the largest integrated Carbon Capture and Storage (CCS) demonstration projects in the world.

1.1.1 Project objectives

The main objective of ROAD is to demonstrate the technical and economic feasibility of a large-scale, integrated CCS-chain. In the power industry, to date, CCS has primarily been applied in small-scale test facilities. Large-scale demonstration projects are needed to show that CCS is an efficient and effective CO_2 abatement technology within the next 5 to 10 years. With the knowledge, experience and innovations gained by projects like ROAD, CCS could be deployed on a larger and broader scale: not only on power plants, but also within energy intensive industries. CCS is one of the transition technologies expected to make a substantial contribution to achieving climate objectives.

1.1.2 Partners

ROAD is a joint project initiated by E.ON Benelux N.V. and Electrabel Nederland N.V. (GDF SUEZ Group). Together they constitute the limited partnership Maasvlakte CCS Project C.V. The intended partners of ROAD are GDF SUEZ E&P Nederland B.V. for the CO_2 transport and TAQA Energy B.V. for the CO_2 injection and permanent storage. The ROAD-project is co-financed by the Government of the Netherlands, the European Commission within the framework of the European Energy Programme for Recovery (EEPR) and the Global CCS Institute.

1.1.3 Project specifications

ROAD applies post combustion capture technology to capture the CO_2 from the flue gases of a new 1,100 MWe coal-fired power plant (Maasvlakte Power Plant 3) in the port and industrial area of Rotterdam. The capture unit has a capacity of 250 MWe equivalent and aims to capture 1.1 million tonnes of CO_2 per year. The capture installation is planned to be operational in 2015.

Location of the ROAD-project CCS chain: Rotterdam port and industrial area and North Sea

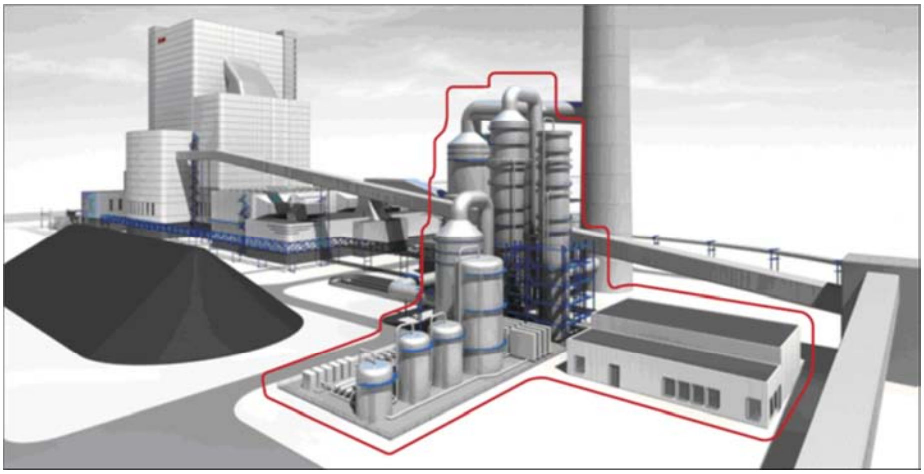
Location of the capture unit: Maasvlakte Power Plant 3 (photo: E.ON)

From the capture unit the CO_2 will be compressed and transported through a pipeline: 5 kilometers over land and 20 kilometers across the seabed to the P18 platform in the North Sea. The pipeline has a transport capacity of around 5 million tonnes per year. It is designed for a pressure of 175 bar and a maximum temperature of approximately 80 °C.

ROAD plans to store the captured CO_2 in depleted gas reservoirs under the North Sea. These gas reservoirs are located in block P18 (P18-6, P18-4 and P18-2) of the Dutch continental shelf, approximately 20 kilometers off the coast. The depleted gas reservoirs are at a depth of around 3,500 meters under the seabed of the North Sea. The CO_2 will be injected from the platform into depleted gas reservoirs. The estimated storage capacity is approximately 35 million tonnes.

1.1.4 Rationale for Rotterdam port and industrial area

The Rotterdam port and industrial area has a number of advantages that create favorable conditions to implement a CCS demonstration project like ROAD. The Rotterdam port and industrial area has many CO_2 point sources. Several new power stations prepared for the application of CCS (capture ready) are under construction. The Port of Rotterdam is relatively close to a large number of (almost) depleted gas reservoirs on the continental shelf under the North Sea, allowing for a small transport distance. These gas reservoirs meet the physical and geological properties for CO_2 storage and will become available in the next few years (from 2014 onwards). The Netherlands has a lot of knowledge and experience with both oil and gas extraction and storage of gas in aquifers and gas reservoirs. In addition, the complete CCS-chain is remote from residential areas.



250 MWe capture unit (post-combustion)

P18-A platform at the North Sea (photo: TAQA)

1.1.5 Facts & Figures

Base installation: E.ON Maasvlakte Power Plant 3 (Rotterdam, The Netherlands)

- Output : 1.070 MWe
- Efficiency : 46% (LHV)
- Operational : End 2012
- Capture ready : Carbon Capture Ready Certificate from TÜV Nord

Capture Plant

- Technology : Post-combustion
- Solvent : MEA formulation
- Capacity : 250 MWe equivalent
- Capture rate : 90%
- CO_2 captured : ~ 1.1 megatonnes / year
- Operational : 2015

Transport

- Pipeline
- Diameter : 16 inch (i.e. 40.6 cm)
- Distance : 5 km onshore, 20km offshore
- Capacity : Gas phase :1.5 megatonnes/year
  Dense phase : 5 megatonnes/year
- Design specifications : 175 bar, 80 °C

Storage

- Depleted gas reservoir : P18
- Operator : TAQA
- Depth : 3,500 meters
- Estimated capacity : ~ 35 megatonnes
- Available : 2014

1.1.6 Planning

The high level schedule of the ROAD project is as follows:

14 July 2009 : Application submitted for funding under European Energy Programme for Recovery
September 2009 : Project selected for funding by European Commission
May 2010 : Ministerial order Dutch funding published
: Grant Agreement signed by European Commission and ROAD Project
September 2010 : Front-End Engineering Design studies Capture Plant completed
Starting note Environmental Impact Assessment published
June 2011 : Submitting Environmental Impact Assessment, permit applications
Q4 2011 : Final Investment Decision
Q4 2011 : Start execution phase (procurement, construction, etc.)
2014 : CCS chain mechanically complete
2015 : Start of operation CCS chain
2015-2019 : Demonstration operation phase CCS chain
2020 : Start commercial operation CCS chain

1.2 Maasvlakte CCS Project C.V.

The initiating parties of the ROAD project are E.ON Benelux and Electrabel Nederland / GDF SUEZ Group. Together they constitute the limited partnership Maasvlakte CCS Project C.V.

1.2.1 E.ON Benelux

E.ON Benelux concentrates on the production and supply of electricity and gas to private customers and business customers in the Netherlands and Belgium. E.ON Benelux is primarily an electricity-generating company; the company can trade internationally and has its own professional sales organisation. The company was established in 1941 and since 2000 has been part of E.ON Energie AG. E.ON Benelux's power stations with a total capacity of 1,850 MW are located in the province of South Holland, the economic heart of the Netherlands. The company has approximately 600 employees. E.ON Benelux is based in Rotterdam.

1.2.2 Electrabel Nederland

Electrabel Nederland is a leading player in the Dutch energy market and part of the GDF SUEZ Group. With six state-of-the-art production locations and a total capacity of 5,103 MW Electrabel is the largest electricity producer in the Netherlands. Electrabel is a supplier of electricity and gas to both private and business customers. Electrabel Nederland has 1,250 employees.

1.3 Intended Partners

Intended partners of Maasvlakte CCS Project C.V. are GDF-SUEZ E&P Nederland B.V. for the CO_2 transport and TAQA Energy B.V. for the CO_2-injection and the permanent storage under the sea bed of the North Sea.

TAQA Energy

TAQA Energy is part of the Abu Dhabi National Energy Company PJSC (TAQA), an energy company that has worldwide interests in power generation, combined heat and water, desalination, upstream oil & gas, pipelines, services and structured finance. TAQA has a workforce of 2,800 employees and is located in Abu Dhabi, The Hague, Ann Arbor: Michigan, Aberdeen, Calgary and Amsterdam. In addition, TAQA has sustainable partnerships with companies in Africa, the Middle-East, Europe, North-America and India. TAQA is listed at the Abu Dhabi Securities Exchange (ADX).

In the Netherlands, TAQA Energy explores and produces gas and condensates from wells located onshore in the Alkmaar region and offshore in the Dutch North Sea. TAQA also operates a gas storage facility in Alkmaar and has interests in Dutch North Sea pipelines. Approximately 200 people are directly or indirectly employed by TAQA in the Netherlands, both onshore and offshore.

GDF SUEZ E&P Nederland

GDF SUEZ E&P Nederland is one of the largest operators in the Dutch sector of the North Sea. With more than thirty production platforms and 300 employees, it is at the basis of the provision of energy to the Netherlands and several other countries.

Since its first successful drilling results in the Dutch North Sea, approximately forty years ago, GDF SUEZ E&P Nederland has grown into a leading operator. It has ample expertise and experience, always chooses the safest option and is continuously working towards the development of new techniques and improved methods. Continuity is ensured through exploration, takeovers and acquisition.

1.4 Financial contributors

The ROAD-project is co-financed by the European Commission within the framework of the European Energy Programme for Recovery (“EEPR”), the Government of the Netherlands and the Global CCS Institute.

In response to the economic crisis, the European Council and the European Parliament adopted the Commission proposal for a European Energy Programme for Recovery (“EEPR”) in July 2009. The EEPR funds projects in the field of gas and electricity infrastructure as well as offshore wind energy and CO_2 capture and storage (CCS). In total 12 CCS projects applied for assistance under the EEPR. In December 2009, the European Commission granted financial assistance to six projects that could make substantial progress with project development in 2010. These projects will receive overall funding of € 1 billion under the EEPR.

2. Introduction to FEED study report

In July 2009, Maasvlakte CCS Project C.V. ('MCP') submitted its project proposal to the European Commission, to apply for funding under the framework of the European Energy Program for Recovery ('EEPR'). This marked the start of the 'ROAD project' ('Rotterdam Opslag en Afvang Demonstratieproject'; Rotterdam Storage and Capture Demonstration project).

2.1 Developing the ROAD project

In this report, the results of the front-end engineering design ('FEED') study for the capture plant as performed by Fluor will be summarized. This FEED study, however, was only part of the work performed on the ROAD project during its development phase. In parallel, various other activities were executed to develop the project to such an extent that the final investment decision could be taken by the end of 2011. This section aims to show the project development context within which the FEED studies were performed by two competing suppliers (see also [1]). It will briefly introduce the reader to phased project development approaches, with the framework developed by the Global CCS Institute taking a central role [2]. Subsequently, a rough sketch of the activities performed by ROAD in parallel with the FEED study is provided.

2.1.1 Project development framework

Using a structured stage-gated project management process is generally identified as a key factor in successfully delivering capital expenditure projects. An information-driven stage-gated process facilitates a logical sequence of project development activities, ensuring that all relevant information is available in time and that all important issues are resolved before further steps are taken [3]. Such a process is generally applied within both mother companies of Maasvlakte CCS Project C.V.

The Global Carbon Capture and Storage Institute is currently developing a project development framework that is tailored for the development of CCS projects. The framework comprises three levels:

- On the highest level, six stages in the development of a CCS project are defined, fully aligned with oil and gas industry standard practice: identify, evaluate, define, execute, operate and closure. For each of these phases the generic approach, actions to be taken and goals to be achieved are set out on one A3 page. This is done for the project in general and for each of the three steps in the CCS value chain (capture, transport, storage).
- On the second level, an overview is provided of around 20 relevant topics specific to each of the three steps in the CCS value chain, setting out the relevant steps to be taken on each of the topics in each of the six development phases.
- On the third level, the actions proposed to be taken on the second level are elaborated.

The current draft of the highest level framework as well as the current version of the second level document focusing on capture can be found in the annexes to this report.

2.1.2 ROAD –project development status during "define" phase

Phrased in terms of the Global CCS Institute's CCS project development framework, ROAD is at the time of writing this report in its "define" stage; the last phase before execution of the project starts. As the phase's name suggests, the define phase is focused on defining the project to such an extent that a final investment decision can be taken based on the available information. In this phase the technical and economic viability of a project needs to be demonstrated. Basic engineering (FEED study), cost estimating, contractor selection and permitting are the key activities in this phase of the project.

In the table below, the status achieved by the ROAD project during the ‘define’ phase is summarized, using the second level of the Institute’s framework as a basis [2]. In the remainder of the report only the FEED study for the capture plant, as made by Fluor and mentioned under “power plant – capture facility” in the table below, will be discussed.

Table 2.1: Status ROAD project

Activity	**ROAD status during ‘define’ phase**
Project Context and Opportunity Definition and Recommendation	Board Paper (incl. project status, strategic fit, organizational setup, costs & financial metrics, risks & risk mitigation, procurement, permitting, stakeholder management, execution planning) available.
Project Development Approach	Project execution plans in place.
Site Selection	Site selection not applicable for ROAD (MPP3 already under construction). Soil investigation performed. Leasehold ready for signature.
Stakeholder and External Relations	Stakeholder management plan & team in place. Public engagement activities ongoing.
Environment	Environmental studies completed. Environmental Impact Assessment submitted to competent authorities. Permit applications submitted.
Health and Safety	Several risk assessments completed, including HAZID, HAZOP, maximum credible accident study and reliability assessment. A rough outline of timing of finalizing different H&S systems and documents has been prepared. A safety management team for construction and commissioning is still to be established, as are detailed procedures on site surveys. However, coordination risks on site are being investigated.
Fuel Supply	Not applicable for ROAD (E.ON MPP3 scope).
Power Plant – Capture Facility	Power plant engineering not applicable (E.ON MPP3 scope). FEED study capture plant completed. Value engineering capture plant completed. Detailed engineering started. Basic design interfaces completed.
CO_2 Product	CO_2 specification is fixed, including impurities, pressures and temperatures. Compressor and dehydration process and design has been assessed in FEED. CO_2 transport pipeline has been specified
Infrastructure	Not applicable for ROAD (E.ON MPP3 scope).
Project Delivery	EPC contract capture plant supplier ready for signature. Contract with storage partner ready for signature.
Capital / Operating Costs	Cost estimates (+/- 10-15%) available.
Operations	Investigation performed on location of control room, combine with MPP3 or separate. Operation will be possible from both locations
Revenue and Trading (CO_2 Product Only)	Utilities agreement (incl. power, steam and utilities supply, ETS compensation, operating/dispatch principles) with E.ON in place.

Management Systems	Allocation of the responsibilities along the entire CCS chain agreed. Partner agreements ready for signature.
Ownership and Legal	Project ownership and governance structure in place.
IP Management	IP system has been set up to share info on E.ON power plant with ROAD IP of capture supplier covered in EPC contract and license agreement.
Financial Analysis	Detailed discounted cash-flow model in place.
Risk	Risk workshops held with all disciplines in project team. Risk register (incl. cost/schedule impact, mitigating measures, residual risk) in place.
Funding	Funding agreements with European Commission, Government of the Netherlands and Global CCS Institute in place.
Future Work Plan	Project execution plans in place.
Status of Studies	Various technical studies on all parts of the CCS value chain completed. Pre-FID project review performed.

2.2 Overview other ROAD related Global CCS Institute reports

Besides this FEED study report, ROAD will deliver several other special reports to the Global CCS Institute in 2011. In this section, the content of these special reports will be described. For the actual reports themselves, please refer to the Institute's website.

2.2.1 Capture technology selection methodology

The selection process for a capture technology supplier is described in ROAD's capture technology selection methodology report [1]. In that report, the selection methodology developed by the ROAD project team is described and evaluated, starting with the request for proposal for preliminary studies and ending with the final selection of the capture plant supplier. In Appendix 1 an overview of the selection process can be found.

Two parallel FEED studies were part of the selection process. The FEED study described in this report is the one conducted by the winning supplier. Not all details of the design can be disclosed as some part is proprietary information of the supplier. Nevertheless the information in this document is sufficient to get a good understanding of the design of the future capture plant of the ROAD project.

2.2.2 Permitting process

The ROAD project is a first of its kind. Therefore the permitting process is quite interesting. First the project team had to find out which permits were to be applied for. The FEED study results have been used as input for the permit application, as the required emission values could not be based on measurements on existing installations. It will be possible to find information on permitting in ROAD's permitting process report [4] and in the environmental impact assessment, which will become public by October 2011.

2.2.3 Mitigating project risk & handling business risk

The special reports on mitigating project risk [5] and handling business risks [6] will explain the risk management approach taken by ROAD. In [5] a summary of the risk register, which contains the major risks, mitigating actions and residual risks, will be included. These risks include financial and technical risks. The special report on handling and allocation of business risks ([6]), will describe on a high level the consortium structure, commercial risk analysis, overview of key risks to be addressed for the final investment decision and key commercial considerations in general.

2.2.4 Stakeholder management

In the report on stakeholder management [7] ROAD's major stakeholders will be identified. These include permitting authorities, local, national and EU government bodies and the public. Stakeholder management will be described in general and public outreach, permitting and the regulatory process will receive particular attention.

2.2.5 Execution strategy

ROAD's project execution strategy will be described in the special report on execution strategy [8]. This report will contain for example a high level project schedule, organizational structure and discussions on execution related aspects of the environmental impact assessment, contracting, finance and stakeholder management.

2.3 Structure of report

The structure of this report on the winning FEED study for ROAD's capture plant is as follows:

- Chapter 3 describes the capture FEED scope and boundaries
- Chapter 4 describes the process of the capture plant and the compression
- Chapter 5 describes the health, safety and environmental issues
- Chapter 6 describes the CO_2 stream
- Chapter 7 describes financial and scheduling aspects of capture plant and compressor.

This report is part of the knowledge to be shared under the Funding Agreement between the Global Carbon Capture and Storage Institute ('Global CCS Institute') and Maasvlakte CCS Project C.V.

3. Capture FEED scope and boundaries

The ROAD project has selected a final supplier for the capture plant based on multiple rounds of studies. The capture FEED study phase was one of these rounds. The complete description of the selection procedure can be found in the capture technology selection report [1].

As starting point of each round, the supplier received a technical specification from the ROAD project. In this description, the project was explained and boundary conditions and quality standards to be implemented in the study phase were submitted.

The result was a very detailed document, of over 150 pages, excluding appendices. In this chapter the main issues are highlighted.

3.1.1 Power plant MPP3

The capture suppliers were given information on MPP3 with more details on the general installation:

Base installation: E.ON Maasvlakte Power Plant 3 (Rotterdam, The Netherlands)

- *Output : 1.070 MWe*
- *Efficiency : 46% (LHV)*
- *Operational : End 2012*
- *Capture ready*

Besides power plant data, also information on general aspects like weather conditions, surroundings and soil investigations was provided.

3.1.2 Capture plant design

In the specification the design philosophy was described with this statement as main message:

The capture plant *process design should be optimized to minimize through life cycle costs of the project, balancing capital costs against reduced power, steam and utilities consumption without compromising plant design performance, availability and operability. The design should also be sufficiently flexible to enable stable operation at reduced flue gas flow and a rapid response to change in flue gas inlet conditions.*

As this demonstration project is expected to be a loss making project, the aim is not to have the highest possible availability, but to seek a balance between costs and plant performance.

3.1.3 Soil situation

An example of the specifications given to the supplier is the soil profile:

The general soil profile at the site is as follows:

- *Groundwater level at ca. + 2m NAP (Normaal Amsterdams Peil, reference of height)*
- *Loose to densely compacted sand with local laminations of clay from surface level to level -20 m NAP*
- *1st clay layer from level -20 m to -22 m NAP*
- *Densely compacted sand from below ca. -22 m NAP (used as bearing level for MPP3 power island piles)*
- *2nd clay layer at ca. – 40 m NAP*
- *Densely compacted sand below 2nd clay layer*

By providing such detailed information the capture plant supplier was in the best position to design a capture plant according to ROAD´s wishes.

4. Capture and compression process

In this chapter the capture and compression process are described. The information is based on the FEED study carried out by the winning FEED supplier. The information given in this report is non-confidential, but can still be valuable to assess the process.

4.1 Process description of capture and compression

This section provides the process description for the Econamine FG PlusSM (EFG+) CO_2 capture and compression plant for the ROAD project. The purpose of the proposed EFG+ plant is to recover 90% of the total CO_2 available in a 250 MW-equivalent slip stream of flue gas from Maasvlakte Power Plant 3 (MPP3), a 1,070 MW (net electric) advanced supercritical power plant. The EFG+ Plant produces approximately 4,057 t/day of CO_2 (dry basis) that will be compressed and sent to the EFG+ plant battery limit to be transported through a pipeline to off-shore storage in the North Sea.

In Appendix 2 the Process Flow Diagrams can be found, please refer to this appendix for the codes in the description below.

4.1.1 Flue gas cooling and trim SOX removal

The Process Flow Diagram of this section can be found in Appendix 2 on page 43.

The flue gas feed to the EFG+ plant comes from the Flue Gas Desulfurization (FGD) unit of MPP3. Flue gas from the FGD unit is first routed to the Direct Contact Cooler (DCC) (3 0HKB10 BB001) for both cooling and trim SOX removal. The flue gas cooling is done in the quench section of the DCC by circulating water. Circulating water is drawn from the bottom of the DCC to the DCC Water Cooler (30HKB30 AC001/2) by DCC Circulation Pump (3 0HKB20 AP001/2) and returned to the top of the column's quench section. The circulating water is heated by the cooling and condensing of water vapor in the flue gas.

Excess water produced in the DCC from the condensing of water vapor in the flue gas is neutralized using 20 wt% NaOH. The neutralized excess water stream is held temporarily in the DCC Water Neutralization Tank (3 0HKB50 BB001) and sent to the plant battery limit using the DCC Excess Water Pump (3 0HKB60 AP001). Cooled flue gas from the quench section of the DCC contacts a circulating scrubbing solution. The scrubbing solution is circulated to the top of the column by DCC Scrubbing Solution Pump (3 0HKE70 AP001/2). A 20 wt% sodium hydroxide (NaOH) solution supplied by the DCC NaOH Injection Pump (3 0HKE60 AP001/2), along with a slip stream of cooled DCC water taken downstream of the DCC Water Cooler, are fed to the suction of the DCC Scrubbing Solution Pump. A blowdown slip stream is taken from the circulating scrubbing solution loop which could be used as make-up to the power plant after treatment. The cooled, desulfurized flue gas passes through a mesh mist eliminator and exits the top of the column where it is routed to the Blower (3 0HKC10 AN001). The 20 wt% sodium hydroxide solution is made by diluting a 50 wt% NaOH stock solution, which is stored in 50 wt% NaOH Storage Tank (3 0HKE20 BB001) using demineralized water.

4.1.2 Flue gas blower and CO_2 absorption

The Process Flow Diagram of this section can be found in Appendix 2 on page 43.

The Blower (3 0HKC10 AN001), located downstream of the DCC, is used to overcome the pressure drop through the EFG+ plant. From the Blower, the flue gas enters the bottom of the Absorber (3 0HKD10 BB001) and flows upward through the packed column beds, where it reacts with the EFG+ lean solvent. A total draw of semi-rich solution is extracted from the column and sent by the Absorber Intercooler Pump (3 0HKD20 AP001/2) through the Absorber Intercooler (3 0HKD30 AC001-3) where the solution is cooled before being returned to the Absorber. Treated gas from the absorption section enters the wash section at the top of the Absorber where EFG+ solvent in the vapor phase is captured by a circulating wash water loop. Wash water is drawn from the bottom of the wash section and sent through the Wash Water Cooler (3 0HKD50 AC001/2) by Wash Water Pump (30HKD40 AP001/2). Demineralized water is also added to the top of the column which removes any entrained solvent and reduces the vapor phase solvent in the exiting flue gas.

The treated flue gas is then routed through a mesh mist eliminator and is returned to the stack of MPP3. Excess water from the wash water section is combined with lean solvent entering the Absorber above the amine packed bed.

4.1.3 Solvent regeneration

The Process Flow Diagram of this section can be found in Appendix 2 on page 44.

The rich solvent leaves the bottom of the Absorber and is pumped by the Rich Solvent Pump (3 0HKG10 AP001/2) to the solvent regeneration section of the EFG+ Plant. Rich solvent from the Absorber is routed to the Solvent Cross Exchanger (30HKG20 AC001-12) to heat the rich solution against the lean solution from the Lean Flash Drum (3 0HKH30 BB001). The hot rich solvent then enters the Stripper (30HKH10 BB0010).

Rich solvent flows down the Stripper through the packed beds counter-current to stripping steam, which removes CO_2 from the rich solvent. Solvent collects on the bottom chimney tray and is sent to the Reboiler (3 0HKH20 AC001-6). Heat input to the Reboiler is provided by condensing low pressure (LP) steam. In the wash section of the Stripper above the feed nozzle, water from the Overhead Accumulator (3 0HGA20 BB001) is used to wash entrained solvent out of the vapor stream. The resulting vapor from the top of the Stripper contains CO_2 saturated with water. The vapor is cooled and condensed by the Condenser (3 0HGA10 AC001/2) against vacuum condensate from MPP3.

The two phase mixture from the Condenser enters the Overhead Accumulator where the carbon dioxide and condensed water are separated. The condensed water is pumped from the Overhead Accumulator by the Reflux Pump (3 0HGA30 AP001/2). A portion of the condensate is returned to the Stripper as reflux and the remaining liquid from the Overhead Accumulator is sent to the Absorber. The carbon dioxide rich vapor is sent to the CO_2 Product Compressor (3 0HGB10 AN001).

The lean solvent leaving the Stripper is sent to a Lean Flash Drum where it is flashed at near atmospheric pressure. The resulting flashed vapor is returned to the Stripper via the Lean Vapor Compressor (3 0HKH40 AN001). The lean solvent from the Lean Flash Drum is pumped by the Lean Solvent Pump (30HKG30 AP001/2) to the Solvent Cross Exchanger. After passing through the Solvent Cross Exchanger, the lean solvent stream is routed back to the Absorber. A portion of the lean solvent stream is sent to the Carbon Bed (3 0HKF20 AT001).The filtered solvent returns to the main lean solvent line where it is returned back to the Absorber.

4.1.4 Solvent reclaiming

The Process Flow Diagram of this section can be found in Appendix 2 on page 45.

A slip stream of hot lean solvent from the bottom of the Stripper is taken out through a semi-continuous reclaiming operation. Also a 20 wt% NaOH solution stream is taken to the reclaimer system package. The recovered solvent is sent back to the Lean Flash Drum. The reclaimer waste is routed to the Reclaimer Waste Sump and periodically shipped off-site for disposal.

4.1.5 Solvent make-up and holding

The Process Flow Diagram of this section can be found in Appendix 2 on page 45.

The make-up solvent is stored in the Solvent Storage Tank (3 0HKJ20 BB001).Solvent make-up (normally no flow) can be injected into the lean solvent line when required. The solvent is sent from storage to the lean solvent line by Solvent Makeup Pump (3 0HKJ40 AP001) via Solvent Make-up Pump Suction Heater (3 0HKJ30AC001).

A Solvent Holding Tank (3 0HKJ50 BB001) is required to store the solvent inventory in the event of a scheduled maintenance or shutdown. The solvent can be transferred to the holding tank via the Lean Solvent Pump (to the Absorber) and Rich Solvent Pump (to the holding tank). After the maintenance or shut down is complete, the solvent in the holding tank can be transferred back to the Absorber using the Solvent Transfer Pump (3 0HKJ60 AP001).

4.1.6 CO_2 compression, oxygen removal, and dehydration

The Process Flow Diagram of this section can be found in Appendix 2 on page 46.

Following recovery, the CO_2 product must be compressed and purified to meet pipeline water and oxygen specifications. The CO_2 Product Compressor (3 0HGB10AN001) compresses the product carbon dioxide to 129 bar(a). The product gas is cooled after each stage of compression. Any condensed water is separated from the gas in knockout drums for each of the first five stages of compression.

A CATOX Unit (3 0HGC10 BB001) is used to reduce the oxygen concentration. The required hydrogen gas will be procured and stored in gas cylinders. The CATOX Unit is a reactor containing a highly reactive catalyst that allows the safe and controlled combustion of hydrogen with oxygen, producing water which is knocked out after cooling.

The gas leaving the CATOX Unit is cooled and partially condensed in the CATOX Aftercooler (3 0HGC20 AC001). Condensed water is then removed in the Water KO Drum (3 0HGC30 BB001). Overhead gas from the Water KO Drum is sent to the CO_2 Dehydration Package (3 0HGC40 AT001) to reduce the water content in the stream to below the required dew point. After the final stage of compression, the product carbon dioxide stream is cooled to 60°C and routed to the battery limit at 129 bar(a) for pipeline transportation. All water from the CO_2 Dehydration Package, Water KO Drum and compression interstage KO drums is combined and routed back to the Lean Flash Drum.

4.2 Simplified process flow diagrams of capture and compression

In Appendix 2 you will find the simplified process flow diagrams of capture plant and compressor.

4.3 Overall heat and material balances

In tables 4.1 and 4.2 the heat and mass balances can be found on a general level. The stream numbers correspond to the streams in the process flow diagrams in Appendix 2.

Within the FEED study itself more detailed confidential streams have also been considered and those streams served as input for the figures in the environmental impact assessment. However, the streams in tables 4.1 and 4.2 give a good indication of the amount and content of the streams.

Table 4.1: Capture plant heat and mass balance.

Stream Description		Flue Gas to DCC		Neutralized Excess DCC Water		Demin Water (Note 2)		50 wt% NaOH (Note 2)		Scrubbing Solution Blowdown from DCC	
Stream Numbers		101		107		109		110		115	
Temperature, °C		48.2		46.6		15.0		20.0		27.9	
Pressure, bar(a)		1.00		4.00		1.35		1.35		4.00	
Component Flows	MW	kgmol/hr	mol%	kgmol/hr	mol%	kgmol/hr	mol%	kgmol/hr	mol%	kgmol/hr	mol%
H_2O	18.02	3,491	11.2%	2,436	100.0%	8.0	100.0%	2.75	69.0%	18	96.7%
CO_2	44.01	4,268	13.7%	0	0.0%	0	0.0%	0	0.0%	0	0.0%
N_2	28.02	22,086	70.9%	0	0.0%	0	0.0%	0	0.0%	0	0.0%
Ar	39.95	249	0.8%	0	0.0%	0	0.0%	0	0.0%	0	0.0%
O_2	32.00	1,059	3.4%	0	0.0%	0	0.0%	0	0.0%	0	0.0%
SO_3	80.06	0.07	2 ppmv	0	0.0%	0	0.0%	0	0.0%	0	0.0%
SO_2	64.06	0.44	14 ppmv	0	0.0%	0	0.0%	0	0.0%	0	0.0%
NO_2	46.01	0.05	2 ppmv	0	0.0%	0	0.0%	0	0.0%	0	0.0%
NO	30.01	1.46	47 ppmv	0	0.0%	0	0.0%	0	0.0%	0	0.0%
NH_3	17.03	0.210	7 ppmv	0	0.0%	0	0.0%	0	0.0%	0	0.0%
HCl	36.46	0.06	2 ppmv	0	0.0%	0	0.0%	0	0.0%	0	0.0%
HF	20.01	0.02	1 ppmv	0	0.0%	0	0.0%	0	0.0%	0	0.0%
Sodium Salts	-	0	0.0%	0.11	0.0%	0	0.0%	1.25	31.0%	0.59	3.0%
Ammonium Salts	-	0	0.0%	0.20	0.0%	0	0.0%	0	0.0%	0	0.0%
Total Molar Flow, kgmol/hr		31,155		2,436		8.0		4.0		19	
Total Mass Flow, kg/hr		913,340		43,910		140		100		410	
Molecular Weight		29.3		18.0		18.0		23.8		21.0	
Density, kg/m^3		1.099		989.8		999.3		1,528.1		1,150.1	
Liquid Flow, m^3/hr				44.3		0.14		0.06		0.34	
Vapor Flow, $m^3_{i.N.}/hr$ (dry)		620,061									
Vapor Flow, $m^3_{i.N.}/hr$ (wet)		698,308									
Vapor Flow, m^3/hr (actual)		831,095									
pH				7.0		7.2		13.7		8.2	

1. All NO present in the FGD flue gas is assumed to be inert and passes through the EFG+ plant. All SO_3 present in the FGD flue gas is in aerosol form and passes through the EFG+ plant.
2. Stream flow rates shown above are during reclaiming operation. When reclaimer is not in operation, Stream 109 flow rate decreases by 20 kg/hr and Stream 110 flow rate decreases by 10 kg/hr.

Table 4.2: Capture plant heat and mass balance *(continued)*.

Stream Description		Treated Flue Gas to Power Plant Stack		Make-up Wash Water to Absorber		Low Pressure CO_2 to Compression		Compressed CO_2 Product		Reclaimer Effluent (Note 3)	
Stream Number		202		203		303		310		404	
Temperature, °C		34.9		15.0		40.0		60.0		139.7	
Pressure, bar(a)		1.00		1.00		1.65		129.00		1.00	
Component Flows	MW	kgmol/hr	mol%	kgmol/hr	mol%	kgmol/hr	mol%	kgmol/hr	mol%	kgmol/hr	mol%
H_2O	18.02	1,456	5.7%	409	100.0%	183	4.6%	0	< 30 ppmv	-	-
CO_2	44.01	427	1.7%	0	0.0%	3,841	95.4%	3,841	99.9%	-	-
EFG+ Solvent	61.08	0	1 ppmv	0	0.0%	0	0.0%	0	0.0%	-	-
N_2	28.02	22,085	87.4%	0	0.0%	1	261 ppmv	1	0.0%	-	-
Ar	39.95	249	1.0%	0	0.0%	0	6 ppmv	0	0.0%	-	-
O_2	32.00	1,059	4.2%	0	0.0%	0	24 ppmv	0	0.0%	-	-
SO_3	80.06	0.07	3 ppmv	0	0.0%	0	0.0%	0	0.0%	-	-
SO_2	64.06	0	0.0 ppmv	0	0.0%	0	0.0%	0	0.0%	-	-
NO_2	46.01	0	0.1 ppmv	0	0.0%	0	0.0%	0	0.0%	-	-
NO	30.01	1.46	58 ppmv	0	0.0%	0	0.0%	0	0.0%	-	-
HSS	-	0	0.0%	0	0.0%	0	0.0%	0	0.0%	-	-
Total Molar Flow, kgmol/hr		25,276		409		4,025		3,842		1.2	
Total Mass Flow, kg/hr		707,560		7,370		172,370		169,040		50	
Molecular Weight		28.0		18.0		42.8		44.0		39.0	
Density, kg/m^3		1.097		999.3		2.732		433.8		1,200.0	
Liquid Flow, m^3/hr				7.4				389.6		0.04	
Vapor Flow, m$^3_{i.N.}$/hr (dry)		533,903				86,115					
Vapor Flow, m$^3_{i.N.}$/hr (wet)		566,538				90,216					
Vapor Flow, m^3/hr (actual)		644,760				63,099					
pH				7.2						8.4	

3. Reclaimer effluent operates intermittently. Stream flow rate shown above is during reclaimer effluent operation.
4. Total Cooling load of the capture plant and compressor is 161.1 MW.
5. Overall Power of the capture plant and compressor (including the required steam power) is 45.6 MW.

4.4 CO_2 capture efficiency

In ROAD's project proposal submitted to the European Commission in 2009 the following was mentioned on the CO_2 capture rate:

- **Capture rate: 90% under design conditions.**

  The capture plant will capture 90% of the CO_2 content of the flue gas led through the capture plant in design conditions.

- **Capture capacity: 250 MWe equivalent under design conditions**

  The capture capacity is 250 MWe equivalent under design conditions. This amount equals 47 kg/s in full load. More details can be found in table 4.3.

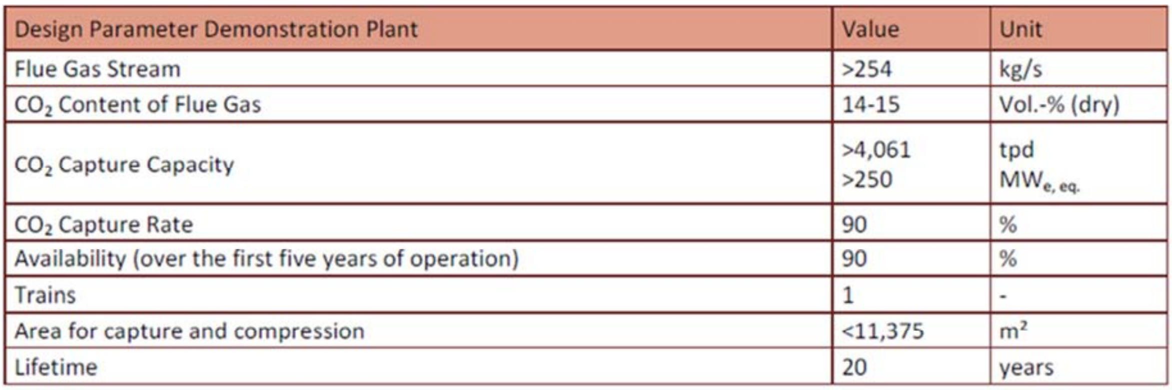
- **Energy consumption by the capture plant**

  In table 4.4 the impact of the capture plant on the MPP3 power plant process is shown. All numbers in the table are applicable to operation under design conditions. The design conditions include for example standard coal, average climate, average cooling water temperature, etc. Furthermore, these numbers are slightly changed compared to the impact as expected in 2009.

  The power plant MPP3 will produce a gross electrical output of 1117 MW. The net electrical output of MPP3 is 1069 MW without capture. At the time of applying for subsidy, with capture this number drops to 1006 MW electrical. So the MWe equivalent provided to the capture plant was expected to equal 63 MWe. After performing the FEED the total energy consumption has dropped to 57 MWe as now published in the environmental impact assessment. That means that the power output will drop to 1012 MW electrical

  The cooling demand has also dropped since the FEED was done, from the expected 196 MW thermal in 2009 to the 161 MW thermal. One thing to keep in mind is that although the capture plant requires more cooling demand, the power plant itself needs less cooling. This means that the total cooling demand rises by nearly 100 MW_{th}.

Table 4.3: Main design parameters for the capture plant.



Design Parameter Demonstration Plant	Value	Unit
Flue Gas Stream	>254	kg/s
CO_2 Content of Flue Gas	14-15	Vol.-% (dry)
CO_2 Capture Capacity	>4,061 >250	tpd $MW_{e, eq.}$
CO_2 Capture Rate	90	%
Availability (over the first five years of operation)	90	%
Trains	1	-
Area for capture and compression	<11,375	m^2
Lifetime	20	years

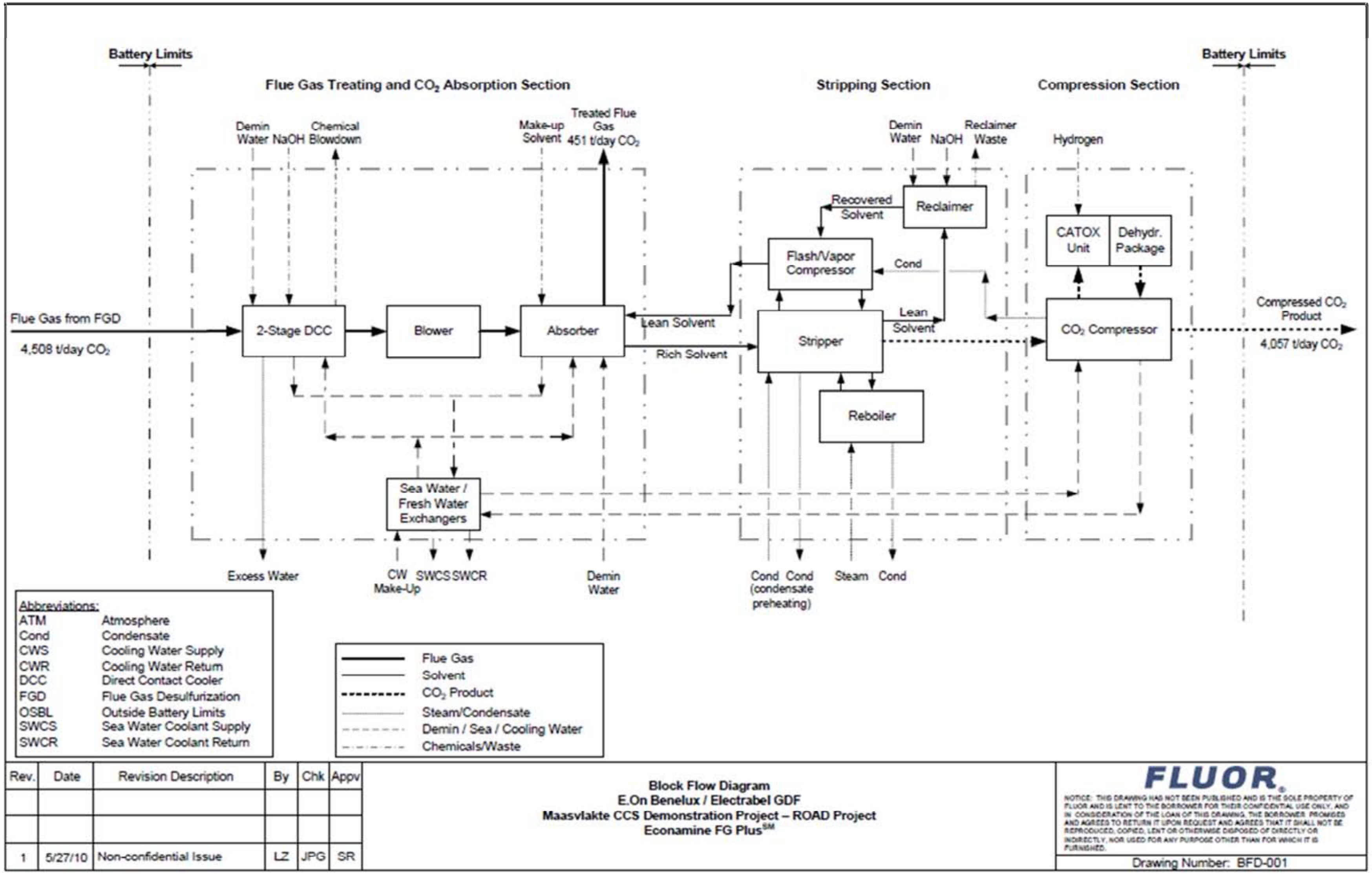
Table 4.4: Impact of the capture plant on the power plant process. Efficiency in (LHV)

Parameter	Unit	without Carbon Capture	with Carbon Capture
Electrical Output (Gross)	MW	1,117	1,086
Electrical Output (Net)	MW	1,069	**1,012**
Electrical Efficiency (Net)	%	46.3	43.9
Specific CO_2 emission	g/kWh	755	641
Cooling demand Power Plant	MW	1,090	985
Cooling demand Carbon Capture Plant	MW	0	**161**
Overall Cooling demand	MW	1,090	**1,146**

4.5 Block Flow diagram

On the next page you can find the block flow diagram of the capture and compression process (figure 4.1).

Figure 4.1: Block flow diagram of the capture and compression process.



4.6 Integration with base power plant

The design of the integration is done by E.ON New Build & Technology BV. The coal-fired unit MPP3 is already under construction for a couple of years and therefore the integration of the capture plant is in fact a retrofit instead of a completely new design. All major components (boiler, turbine, flue gas cleaning, stack etc.) have already been ordered and designed prior to the development of the capture plant. This means that in this demonstration phase some solutions have been chosen that fitted well in this situation, but might not be the solution selected when capture and power plant are designed at the same time.

4.6.1 Carbon capture plant – MPP3 interface overview

All figures and data in this report are based on full load operation under typical operating conditions. Values will differ for different load factors and operating conditions.

4.6.2 Flue gas

The capture plant was designed for an equivalent electrical output of 250 MW and will therefore only need to treat a slip stream of the flue gas produced by Maasvlakte Power Plant 3 (MPP3). In design conditions, the capture plant needs to treat 23.4 % of the flue gas from the FGD of MPP3. The flue gas slip stream has a volume flow of 698.000 Nm^3/hr at 48°C, which was considered as the maximum volume flow to the capture plant.

A flue gas slip stream will be extracted from MPP3 and routed to the capture plant. The flue gas will be extracted after the Flue Gas Desulphurisation (FGD) unit of MPP3.

The discharge of the treated flue gas, which has a volume flow of 566.000 Nm^3/hr at 35°C will be routed back to the wet stack of MPP3.

4.6.3 Electric power

The approx. 20-30 MW electrical power will be provided via a 10-kV-switchgear that will be installed at the control building of the MPP3 power plant and which is linked to a 150/10 kV transformer that is connected to the external 150 kV grid.

Since the supply to MPP3 from the external 150 kV grid is required only in emergency cases (e.g. operating failures of auxiliary power transformers), the capture plant can be supplied with electrical energy via the external grid transformer.

In case of the above mentioned emergency event, MPP3 will be supplied from the external grid. The 10 kV supply to the CCS system would be dropped automatically in case of this event. The CCS system would then not be available. A reconnection of the electrical power supply to the CCS system would be done manually.

The current CCS system design intends to supply the CCS-system with two 10kV-parts with nine or eight switchboards each.

4.6.4 Cooling water

Normally the three sea water cooling pumps of the main cooling system of MPP3 have enough capacity to provide cooling for MPP3 and the capture plant. When one pump is not available, the capture plant may have to be reduced in load or shut down.

To supply 12290 m^3/h cooling (sea)water to the capture plant, it is foreseen to connect suction lines to each of the two manholes (ID 800mm) of an inspection well in the concrete cooling water channel between the main cooling water pumps and the machine house of MPP3.

The two manhole covers will be replaced with (flanged GRP) DN800 pipe spools running to a suctionheader, which feeds the capture plant cooling water booster pumps. The DN800 spool pieces will be designed in such a way that they are removable, which will allow access to the manholes when this is needed.

The concrete superstructure around the manholes will also be modified, allowing the suction header to penetrate the superstructure and also ensuring sufficient personnel access.

The above mentioned modifications to the existing MPP3 cooling water channel manholes will be made during the operational phase of MPP3. The concrete superstructure can be modified without operational consequences. The piping from the manholes to the capture plant can also be installed while MPP3 is in operation.

The cooling water discharged from the capture plant will be routed to the outlet pond through an existing cooling water channel running adjacent to the capture plant.

4.6.5 Steam and condensate

When MPP3 is operating in full load, the steam (<100 kg/s at 2-4 bar) that is used in the desorber (stripper) of the capture plant will be extracted from the steam turbine downstream of the medium pressure turbine. In part load of MPP3 this pressure is not high enough and another (higher pressure) extraction point will have to provide the steam.

Design of this extraction system is still pending as optimization is of high importance because of the high operational costs involved. What are the scenarios – advantages and disadvantages?

Three scenarios are evaluated to have increased steam pressure in part load:

1. Modification of the crossover pipe MP to LP turbine and installing valves to keep enough pressure during part load. (High investments; operational risks, long outage)
2. Extraction from a higher pressure source (low part load efficiency).
3. Installing a steam jet booster to increase the pressure extracted from the main extraction point (more complex and extra investments).

The approx. 1200 mm steam line will be routed on a 450m long dedicated pipe bridge together with other piping needed for the utilities of the capture plant.

The condensate from the steam will be routed back to the condensate system of MPP3.

4.6.6 Condensate for cooling

The design basis for this project identifies an opportunity for heat integration between MPP3 and the capture plant via pre-heating steam condensate within the capture plant.

This condensate will be supplied at 19 bar(a) and 26°C; it will be returned at 16-18 bar(a) and 70-100°C.

The condenser that cools the stripper overhead gas is an ideal heat source for preheating this condensate. The hot-side enters the condenser at 89°C and is cooled to 40°C. This means that the full heat load (28,8 MW) of the condenser can be used for condensate pre-heating. The pre-heated condensate is returned to the MPP3 at 18 bar(a).

4.6.7 Flue gas condensate and deep FGD waste water

The waste water that is produced by the pre-scrubber section of the capture plant is divided in two split streams. The pre-scrubber combines a cooling section, the direct contact cooler (DCC) and a desulphurization unit, the deep flue gas desulphurization unit (deep FGD). Both sections produce waste water, whereas the split streams differ significantly in quantity and quality.

The DCC excess water stream amounts to 44 t/hr and consists of flue gas condensate. The split stream out of the deep FGD is smaller with only 0,3 t/hr.

The DCC excess water stream can be mixed with process water for feeding the FGD absorber of MPP3, thereby reducing the intake of fresh water from a nearby lake by more than a third.

4.6.8 Other interfaces

Several other interfaces are identified and will be engineered to ensure a safe and efficient operation. Those interfaces include:

6. Control connections
7. Telecom connections
8. Alarm connections
9. Fire water
10. Sewage water
11. Potable water

4.7 Sized equipment list

In Appendix 3 the complete equipment list can be found. It contains the specification of blowers, compressors, exchangers, filters, vessels, pumps, sumps, tanks and specialties.

- **Basis for Material Selection**

Materials of construction are selected and the corrosion allowances are determined on the basis of anticipated corrosion or material degradation under the most severe combination of process variables resulting in sustained maximum normal operating condition.

The risk of damage due to a short term condition such as an alternate operation, startup, shut down, upset and emergency conditions is evaluated as follows:

It is first determined if the short term condition will introduce or concentrate a corrodent or cause a resident corrodent to become active as a result of causing an aqueous phase to be present or crossing a temperature or partial pressure threshold.

The estimated duration of upset is used to estimate a prorated corrosion rate. The prorated rate is then used to evaluate the necessity for extra corrosion allowance and perhaps an upgrade in materials selection.

If the corrodent of concern is a crack-inducing agent, prorating is not normally permitted. The normal response to the process of even a transient active crack-inducing agent is a materials upgrade or the adoption of a preventive measure such as cladding.

Materials specified are the minimum requirement. Higher alloys may be applied due to practical considerations (i.e., availability, ease of fabrication, etc.).

- **Amine Corrosion and Stress Corrosion Cracking**

Corrosion of carbon steel (CS) components in amine services is not caused by amines themselves but by dissolved acid gases, such as CO_2, or by amine degradation products, such as heat stable salts. Variables affecting the corrosion rates are the type of acid gas, the type of amine, solvent concentration, gas loading, temperature, solution purity, velocity, use of inhibitors, design, and metallurgy.

The use of stainless steel alloys will provide excellent resistance to the non-inhibited lean and rich amine streams. These alloys will resist MEA corrosion at all temperatures being specified in the Econamine FG Plus^{SM} process.

Thermal stress relief (e.g., Post Weld Heat Treatment or PWHT) is considered the most effective means of preventing amine SCC on the unclad CS. All welds and bends must be stress relieved including external attachment welds at components with wall thicknesses less than about 25 mm, as the residual stresses from these welds can extend through the wall to the inner surface and initiate cracking from the inside diameter

(ID). In the past, some refiners and gas plant operators stress relieved only above certain operating temperatures, but due to various failures, most of the industry now stress relieves in all amine services except ambient temperature storage tanks. This represents other savings with the use of stainless steel, as PWHT is typically not needed.

- **Wet CO_2 Corrosion**

Wet CO_2 forms carbonic acid which is corrosive to carbon and low alloy steels. In this unit, this corrosion mechanism is a concern in the “wet” areas with CO_2 and no amine. For this plant, all areas with CO_2 corrosion are recommended to be 304L SS stainless steel (or better based on the chloride SCC). On some gas lines, CS can be considered if steps to prevent condensation are also implemented.

4.8 Simplified plot plan

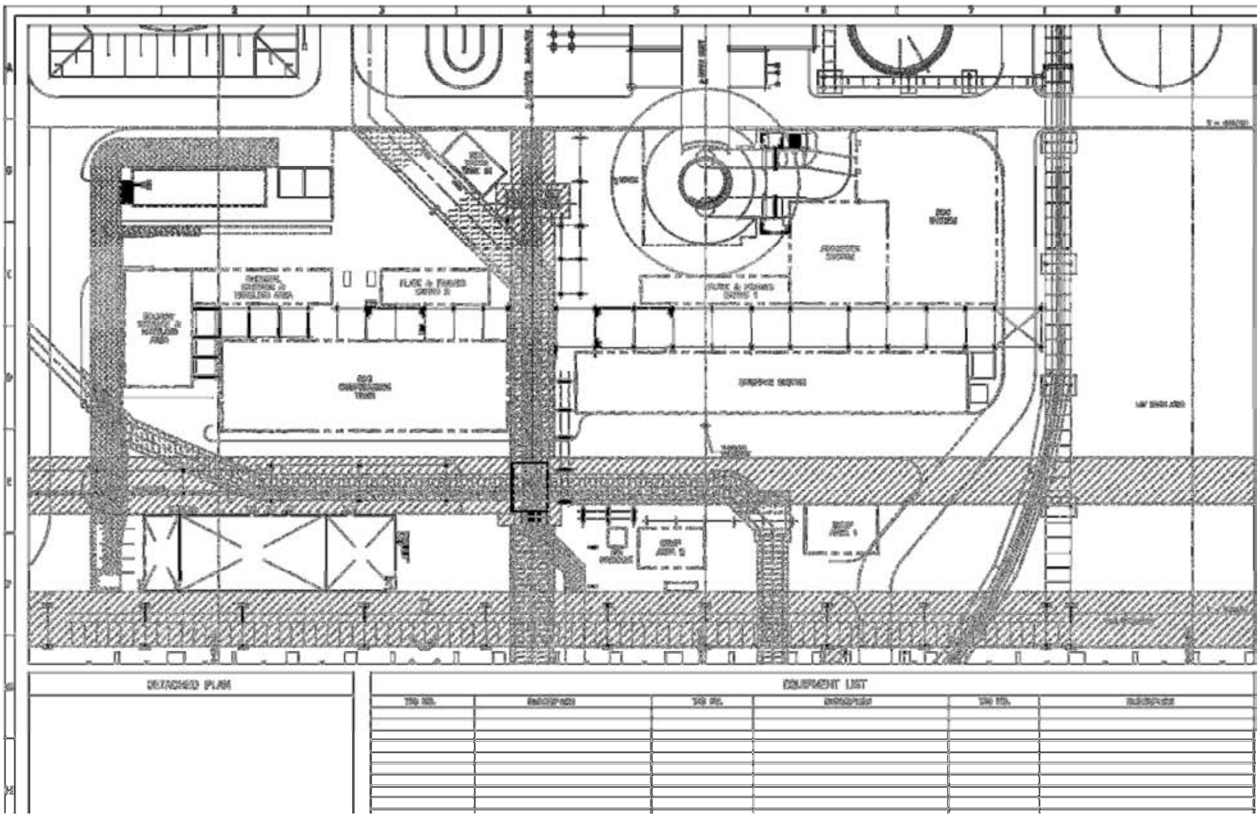
MPP3 is located in the northern part of the Maasvlakte area. The well-populated residential areas of Hoek van Holland and Oostvoorne are located at a distance of approximately 7 km from the MPP3 site. The characteristics of the Maasvlakte area are:

- Container terminals
- Ore and coal terminals (EMO)
- Chemical industry (Lyondell)

The capture installation will be built on the plot space next to the stack of MPP3. Because the MPP3 installation is designed and already under construction, the plot space has some challenges for the capture installation compared to a greenfield situation. On this plot there is an area without height limitation, an area with height limitation due to coal transport conveyor belts and an area where underground cooling water piping is located. Furthermore, 10.5 kV cables are routed on top of the cooling water duct which is crossing the site. Pipe racks on the Maasvlakte site have a passage height of 5 m and at some areas of 8 m.

The supplier managed to design the capture installation to fit all the restrictions. In Figure 4.2 a picture is placed from the simplified plot plan. The enlarged version is found in Appendix 4. The total area available for this demonstration is circa 1.2 Ha. This area is separated by the coal conveyor belt of MPP3.

Figure 4.2: Simplified plot plan.



5. Health, Safety and Environment issues

5.1 Environmental emissions

This section describes the environmental emissions associated with the ROAD project. The information in this chapter is based on the environmental impact assessment. The actuals might differ in the future, but this is the current assumption for the CCS installation (June 2011). In the Environmental Impact Assessment (EIA note: information is in Dutch), more detailed information on the emissions and noise can be found. In Appendix 5 the MSDS of MEA and NaOH can be found.

5.1.1 Emissions to air

In the capture installation circa 23% (volume) of the flue gases from MPP3 will be treated. The treated flue gas will be sent back to the stack of MPP3, mixed with the remainder of the flue gases of MPP3 and emitted to air by the existing stack of MPP3.

Three relevant operating modes are described to be able to calculate the environmental impact:

Operating mode 1. MPP3 operational without the CCS installation.

Operating mode 2. MPP3 full load, CCS fully operational, circa 23% of flue gas treated

Operating mode 3. MPP3 part load, all flue gas treated in CCS installation. Concentration will change, but environmental load stays equal.

In Table 5.1 the assumptions for the composition of the flue gas are stated.

Table 5.1: Assumptions for composition of flue gas

Component	**Flue gas MPP3** **(worst-case)** **(current situation, operating mode 1)**		**Part of Flue gases to Capture plant**	
	Concentration [mg/Nm^3][1]	**Load [kg/hour]**	**Concentration [mg/Nm^3][1]**	**Load [kg/hour]**
NH_3	0,1	0,3	0,1	0,1
NOx as NO_2	65,0	220	65[2]	45,5
SO_2	40,0	115	40	28,0
HCl	3,0	9,0	3,0	2,1
HF	0,4	1,3	0,4	0,3
CxHy	1,0	3,0	1,0	0,7
Dust	4,0	11,7	4,0	2,8
CO_2	-	857000	-	188000
Flue gas specifications				
Temperature (K)	324		321	
Water (vol%)	11,2		11,2	
O_2 (vol%)	3,4		3,4	
CO_2 (vol%)	13,7		13,7	
N_2 (vol%)	70,9		70,9	
Flow (Nm^3/hour)	3190000		700000	

[1] based on dry conditions, 6% Oxygen; [2]: 4 – 5% NO_2;

Tabel 5.2: Emissions from the installation (MPP3 and capture plant)

Component	**Operating mode 3** **Part load**		**Operating mode 2** **Full load**	
	Concentration [mg/Nm^3]	**Load [kg/hour]**	**Concentration [mg/Nm^3]**	**Load [kg/hour]**
MEA formulatino	11	6,2	2,0	6,2
NH_3	5	2,8	1,0	3,1
NOx as NO_2 1)	77,6	44,0	70,0	215,0
SO_2	0	0	31	95,0
HCl	3,7	2,1	2,9	9,0
HF	0,5	0,3	0,4	1,3
CxHy	23	13	10	31
Dust	2	1,1	3,3	10
CO_2	-	18000	-	687000
Flue gas specifications				
Temperature (K)	308		321	
Water (vol%)	5,7			
CO_2 (vol%)	1,7			
O_2 (vol%)	4,2			
Flow (Nm3/hour)	566000		3056400	

1) NO_2 based on 5% of NO_x

In Table 5.2 the emissions of Operating mode 2 (full load) and 3 (part load) are indicated. The emissions at part load are equal to the emissions at the outlet of the capture plant. The emissions at full load are representative for the combined stream of flue gases where a part is treated in the capture plant. The emissions are based on a yearly average value.

5.1.2 Water

- **Process water**

During capture of CO_2 process water is released in a number of places.

In the first place, at normal operation, circa 40-50 m^3 per hour of condensate water is released from the flue gas pretreatment. This process water occurs by cooling of the saturated flue gas. This water can be used in the flue gas desulphurisation.

The flue gases are further desulphurized by the use of a sodium hydroxide solution. The sulphur compounds react with this liquid and are removed with a surplus of water. Per hour circa 0.3 m^3 of this kind of process water is produced containing sulphuric substances.

- **Cooling water**

In the capture process cooling water is used for several cooling steps. The incoming flue gas is cooled. In the absorber the washing water is cooled, and the pumps, fans and compressor needs to be cooled as well.

The cooling water intake of MPP3 conforms to the current permit of 120000 m^3/hour of which the capture installation needs a maximum of 15000 m^3/hour. The cooling water is sea water. By raising the temperature difference between cooling water inlet and outlet, the mass flow of the cooling water remains constant.

- **Demineralized water**

In the capture process demineralized water is required to produce on spec sodium hydroxide and solvent. The demin. water will be supplied by MPP3. The amount necessary will be circa 11 m^3/hour.

5.1.3 Waste

The solvent degrades in time and needs to be regenerated periodically. In this process a part is removed as waste. This waste stream is maximum around 500 tonnes per year and is considered as chemical waste. As soon as the capture plant is operational the waste stream will be analysed to check if this consideration is correct. This chemical waste will be disposed of by a waste treatment company.

5.1.4 Other Chemicals

In the capture process several chemicals are used. These are listed in Table 5.3 below.

Table 5.3: Other chemicals

Chemical	Consumption
Sodium hydroxide (50% in water)	1.000-1.500 kg/day
Solvent	500-1.000 kg/day
Hydrogen	100-300 Nm^3/day
Nitrogen	1.000-2.000 Nm^3/day
Activated coal	50-100 ton/year

5.1.5 Noise

The design of the capture plant has been examined. All the noise making components, like pumps, fan and compressor have been included in the existing acoustic model of MPP3. The additional noise fits in the available noise limits.

5.2 Health and safety issues in regular operation

5.2.1 Introduction

HAZID (HAZard IDentification) is a technique for early identification of potential hazards and threats. A HAZID identifies and assesses hazards leading to major incidents or accidents, which provide essential input to project development decisions and early identification of required additional studies. This will lead to safer design options being adopted with a minimum cost of change penalty.

5.2.2 HAZID Methodology

The HAZID technique is:

- a means of identifying and describing hazards and threats at the earliest practicable stage of a development or venture;
- a meeting employing a highly experienced multi-discipline team using a structured brainstorming technique, based on a checklist of potential HSE issues, to assist in identifying and assessing potential hazards; and
- an identification and description process only, not a forum for trying to solve potential problems.

Many of the hazards and HSE issues are generic for a whole development and are not specific to any part of the plant or location. The typical procedure is therefore firstly to apply the technique to the whole project as a single entity. The study method is a combination of identification, analysis and brainstorming based on guidewords listed in a checklist. The guidewords are divided into three main sections:

- Section 1: external and environmental hazards;
- Section 2: facility hazards; and
- Section 3: health hazards.

Two of the sections (1 and 3) contain overall hazards and project implementation issues which are applicable to the project as a whole. The facility hazards (2) can be identified for separate units or functional blocks of the process and utilities.

5.2.3 Results and Follow Up

The HAZID resulted in a total of 20 action items of which 17 were to be resolved during FEED. The table in Appendix 5 gives an overview of all action items, and the responses from the action party.

5.2.4 Conclusion of HAZID

Seventeen action items were resolved during FEED. The three remaining items were assigned to the EPC phase, and will be followed up during the EPC phase of the project. These items are show in table 5.4.

Table 5.4: Remaining HAZID action items (assigned to the EPC phase).

No:	Description:	By:	Due:	Response:	By:	Date:
7	Discussions with authorities need to be opened to discuss impact of capture plant on operating permit, with respect to emissions.	JV Project	EPC			
10	Capture plant and MPP3 operators to be trained regarding emergency response planning.	JV operations	EPC			
20	Ensure all power lines are identified prior to any excavation or piling work.	Construction	EPC			

5.2.5 Material Safety Data Sheets

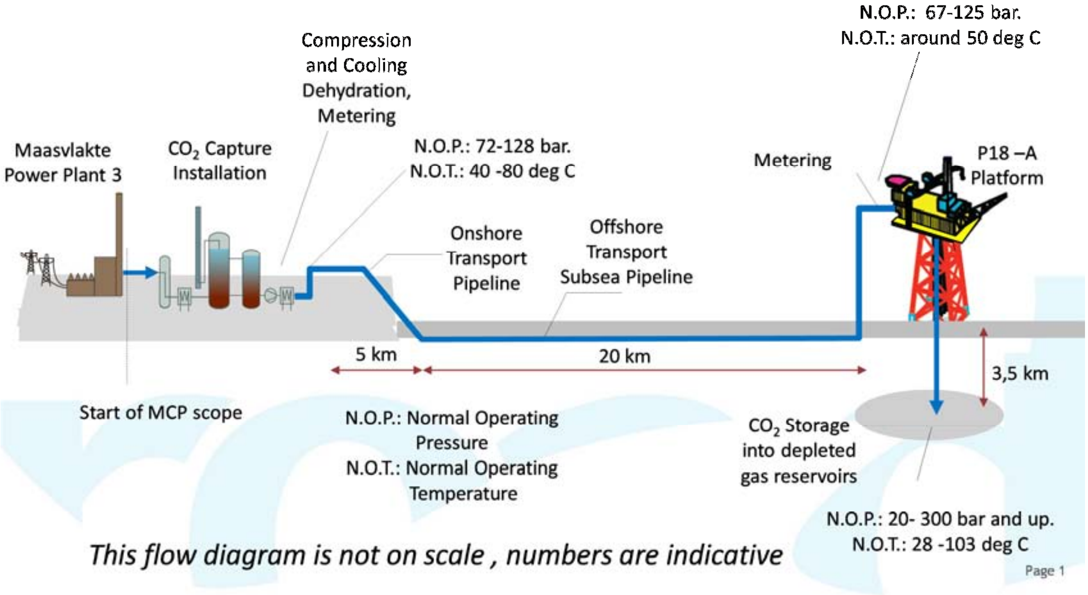
The Material Safety Data Sheets (MSDS) of sodium hydroxide and mono ethanol amine (MEA) can be found in Appendix 5

6. CO_2 stream

6.1 CO_2 specification for transport

In figure 6.1 an indication is given for the temperatures and pressures of the transported CO_2. The exact design figures over time are still under investigation in the flow assurance study and are dependent on the geology of the reservoir. However, the limitations for permits and pipeline design have been set.

Figure 6.1: indicative temperatures and pressures of CO_2 along the CCS chain



6.1.1 Purity of CO_2

The purity of the CO_2 is still under investigation. As a result of the post-combustion capture process, the CO_2 stream will be more than 99.0% pure. In the FEED study a CATOX is foreseen to remove excess oxygen and a molecular sieve to reduce water content. The composition of the compressed CO_2 is shown in table 6.1

6.2 CO_2 metering concept

6.2.1 Scope and Functional Requirements

CO_2 metering will be required in the system to measure plant performance, potentially as a means of demonstrating compliance with environmental regulation, to control flows and to provide a basis for payment under whatever transportation and sequestration performance contracts that prevail in the longer term. To achieve this, the following needs to be addressed:

- Identify the points at which metering will be required.
- Define the purpose of each metering location.
- Identify likely metering performance requirements.
- Identify the parties who may have either an interest in particular metering stations and for what reason.
- Propose accuracy specifications for discussion.
- Determine overall lifecycle costs.

Table 6.1: composition of compressed CO_2

Stream description		Compressed CO_2 product	
Stream Number		310	
Temperature, °C		60	
Pressure, bar (a)		129	
Component Flow	MW	kgmol/hr	mol%
H_2O	18.02	0	<30ppmv
CO_2	44.01	3,841	99.9%
EFG+ Solvent	61.08	0	0.0
N_2	28.02	1	0.0
Ar	39.95	0	0.0
O_2	32.00	0	0.0
SO_3	80.06	0	0.0
SO_2	64.06	0	0.0
NO_2	46.01	0	0.0
NO	30.01	0	0.0
HSS	-	0	0.0
Total Molar Flow, kgmol/hr		3,842	
Total Mass Flow, kg/hr		169,040	
Molecular Weight		44.0	
Density, kg/m^3		433.8	
Dense flow, m^3/hr		389.6	

6.2.2 Assumptions

- Fiscal or custody transfer metering should be included at only one place in the system, measuring flow into the transport system.
- Fiscal metering will be located at MPP3 in the first instance, measuring CO_2 exported into the pipeline for sequestration.
- When in future (new) CO_2 sources are desired to connect to the CO_2 collection system, then custody transfer metering for that new source will be included within the fence line of the new source measuring CO_2 exported into the pipeline system to sequestration.
- Pipeline (transport) operator will need to account for any venting losses at the pipeline transport operators cost (including cost of emissions). Such losses to be based on venting estimates maintained by the operator.
- Accuracy requirement of the custody transfer metering system will be similar to that required for fiscal metering of oil and gas +/- 0.5…1.0%, details are to be investigated.
- Fiscal accuracy metering cannot be provided at the offshore location because of maintenance limitations.

- Meters will not be used for small leak detection in the pipeline. Separately gas in/out/station pressure (+/- 0.75%), gas + ground in/out/station temperature (+/- 2.0%) and gas in/out flow measurements (+/- 2.0%) are needed in the model-based leak detection system.
- Modern fiscal meters need straight runs of pipe for around 15 x the diameter of the upstream pipeline and >=10 x the diameter of the downstream pipeline.
- Modern fiscal meters use ultrasonic transducers measuring and analysing the Doppler Effect on flowing molecules averaged across a flowing area. That does not interfere with pigging and can be located as full size meters in the main pipeline flow between the pig launcher and the fence line.
- Less accurate metering will be required at the inlet to each well. The well injection meters will be used to measure and monitor the injectivity of each well over time. CO_2 system needs to can measure gaseous and dense phase CO_2.

What do we not know

- What will the government require as an acceptable measure of sequestered CO_2 for calculating unitary payments?
- Will the government be satisfied with estimates of venting between meter and injection well? Meter and location have to be defined.
- Will there eventually be a different owner of the pipeline and of the storage field eventually requiring two transfers of custody?
- Will one fiscal metering location be enough?
- Who will carry the cost benefit of replacing meters with more accurate ones as technology advances?
- What is the next best position for leak detection?
- What metering is required in the capture plant?
- What metering is required at the platform?
- How can leakages in pipeline (onshore and offshore) be discovered?
- How can leakages out of the reservoir be detected / measured? Maybe possible to add a tracer during injection?

6.2.3 Design Requirements

Fiscal Export Meter (one required):

- Maximum Design Pressure: 175 barg.
- Measurement of saturated vapour flow at up to 40bar and later saturated liquid (dense phase) flow at pressures between 80 bar initially and up to 175 barg (design pressure).
- Temperature of flowing fluid to be finally defined.
- Design temperature is between 40 and 80C°.
- Redundancy provided by spare ultrasonic probes held on site with pre-determined calibration charts.
- Same meter to be used for vapour phase flow then converted to dense phase flow by changing the non-intrusive ultrasonic transducers only.

- Automatic sampling system required, with automatic processing of samples?
- Accuracy +/- 0.5…1.0%.
- Meter body design life 40 years.

Meter Prover (one required):

- Ballistic Type?
- Installed to operate in series with fiscal meter when flow is directed to it on one run per day? Week? Basis.
- Design and operating temperatures and pressures have to be finalised.
- No pigging while proving.
- Meter prover may need to be changed after 8 to 10 years when CO_2 vapour flow changes to dense phase.
- Capable of proving accuracy within an estimated uncertainty.

Wells Injection Meters:

- Balancing flows only – one in each active side arm
- Orifice plate type accuracy with senior orifice plate box type accessibility
- Analogue signal from transducer suitable for informing a flow control device if necessary

6.2.4 Mandatory References

ISO-5168:2005	Measurement of Fluid Flow – Procedures for the evaluation of uncertainties
ISO-10715:2001	Natural Gas – Sampling guidelines
ISO-10723:2003	Natural Gas – Performance evaluation for on-line analytical systems.
IGE/GM/4	Institution of Gas Engineers Flow metering Practices – Inlet pressure exceeding 38bar and not exceeding 100bar. Communication number 1719
EEC Directive 71/316	Gas Meters

6.2.5 Supporting References

ASME B16.5	Pipe Flanges and Flanged Fittings
ASME B31.3	Process Piping
BS EN 61000-6-4:2007	Electromagnetic compatibility (EMC). Generic standard. Emission standard for Industrial environments.

7. Financials and planning

7.1 Capital expenditure of capture plant

After performance of the FEED and asking quotations to subcontractors, the capture supplier was able to come up with a capital cost estimate for the work on the capture plant itself. This estimate does not include the integration works to MPP3.

ROAD asked for a fixed price for the EPC of the capture plant. This means that the capture supplier had to perform the FEED in sufficient detail to offer this without substantial risks for the supplier. The capture supplier provided ROAD with a capital cost estimate breakdown. For commercial reasons, the overview below is given in ranges rather than fixed numbers, but reflects the reality.

The capital cost estimate breakdown is as follows:

• Equipment:	60 – 70 MM Euro
• Materials and Labor:	65 – 75 MM Euro
• Engineering:	15 – 20 MM Euro
• Indirect Field Costs:	5 – 10 MM Euro
• Owner's costs (Operator training & first fills):	1 – 2 MM Euro
• Financial and other costs:	25 – 35 MM Euro
Total:	**170 – 210 MM Euro**

7.2 Operating costs

Based on the project budget for 2010-2019 the figures shown in table 7.1 can be provided as estimate for the operating costs for the complete project.

Table 7.1: Operating costs in million euros (2010 euro)

	2015	**2016**	**2017**	**2018**	**2019**
Staff costs	2,1	1,9	1,9	2,0	2,0
Variable costs	18,5	21,1	23,9	24,1	24,3
Maintenance	2,7	2,7	2,7	2,8	2,8
Other costs	1,5	1,5	1,5	1,5	1,5
Total operating costs	**24,7**	**27,2**	**29,9**	**30,3**	**30,7**

The assumptions behind the numbers are:

- Costs are presented as gross costs. Revenues from avoided CO_2 emission costs are not taken into account
- Power price assumed is € 55/MWh

The variable costs include the variable power consumption of the power plant at the assumed price stated above. Therefore this item is by far the largest item in table 7.1 and the final figure will be dependent on the power price at that point in time. Base load power is taken into account in other costs. Maintenance cost are estimated to amount between 1.5 and 3% of CAPEX. Other costs do only include a contingency for the capture plant.

7.3 Construction programme

In the GANTT chart of figure 7.1 the construction schedule for the capture plant is shown on headlines. The main items are the permitting schedule, engineering, procurement and construction and commissioning. The complete CCS chain should be operational by the end of Q1, 2015 (in line with the schedule presented in section 1.1.6). Start of construction will be January 2013.

The construction starts with piling and foundation works, afterwards civil works will take place. After that the equipment will be installed; piping will be made and electrical connections will be made. Finally construction ends by implementing the instrumentation systems after which commissioning can start by beginning of Q4, 2014.

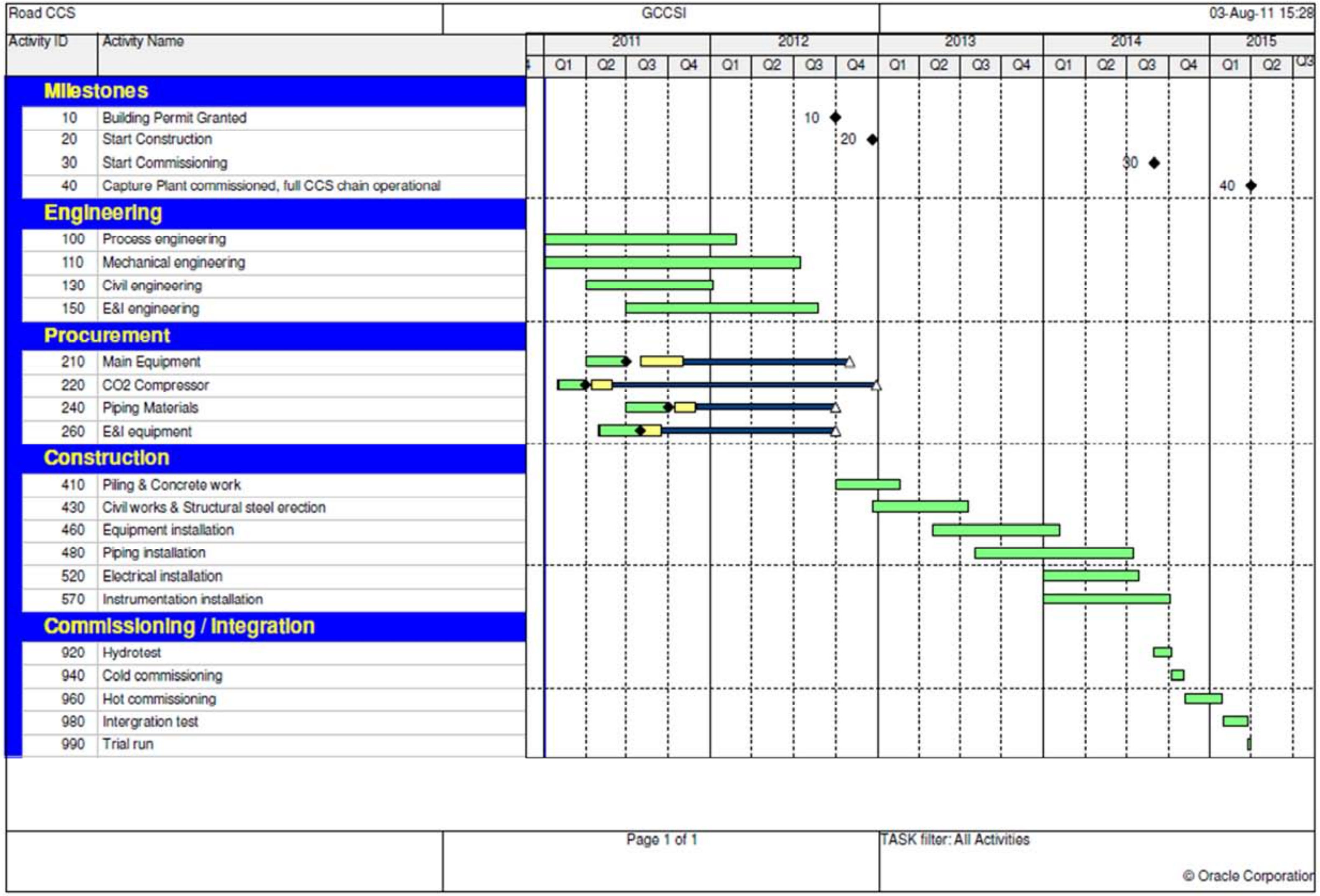


Figure 7.1: Construction schedule for the capture plant.

Road CCS | GCCSI | 03-Aug-11 15:28

Activity ID	Activity Name
Milestones	
10	Building Permit Granted
20	Start Construction
30	Start Commissioning
40	Capture Plant commissioned, full CCS chain operational
Engineering	
100	Process engineering
110	Mechanical engineering
130	Civil engineering
150	E&I engineering
Procurement	
210	Main Equipment
220	CO2 Compressor
240	Piping Materials
260	E&I equipment
Construction	
410	Piling & Concrete work
430	Civil works & Structural steel erection
460	Equipment installation
480	Piping installation
520	Electrical installation
570	Instrumentation installation
Commissioning / Integration	
920	Hydrotest
940	Cold commissioning
960	Hot commissioning
980	Intergration test
990	Trial run

Page 1 of 1

TASK filter: All Activities

© Oracle Corporation

Acronyms

CCS	Carbon capture and storage
EEPR	European Energy Programme for Recovery
EPC	Engineering, procurement and construction
FEED	Front-end engineering design
Global CCS Institute	Global Carbon Capture and Storage Institute
HSE	Health, safety and environment
Institute	Global Carbon Capture and Storage Institute
MCP	Maasvlakte CCS Project C.V.
MEA	Monoethanolamine
RFI	Request for information
RFP	Request for proposal
ROAD	Rotterdam Opslag en Afvang Demonstratie (Rotterdam Storage and Capture Demonstration)

References

[1] *CO_2 capture technology selection methodology – Special Report for the Global Carbon Capture and Storage Institute*; Van der Weijde, G. & Van de Schouw, G.; Global CCS Institute website (2011)

[2] *OpenCCS*; Global CCS Institute (2011)

[3] *Front-End Loading in the Oil and Gas Industry – Towards a Fit Front-End Development Phase*; Van der Weijde, G.; repository.tudelft.nl (2008)

[4] *ROAD's permitting process – Special Report for the Global Carbon Capture and Storage Institute*; Global CCS Institute website (expected 2011)

[5] *Mitigating project risks – Special Report for the Global Carbon Capture and Storage Institute*; Bijkerk, M. & Henry, X.; Global CCS Institute website (expected 2011)

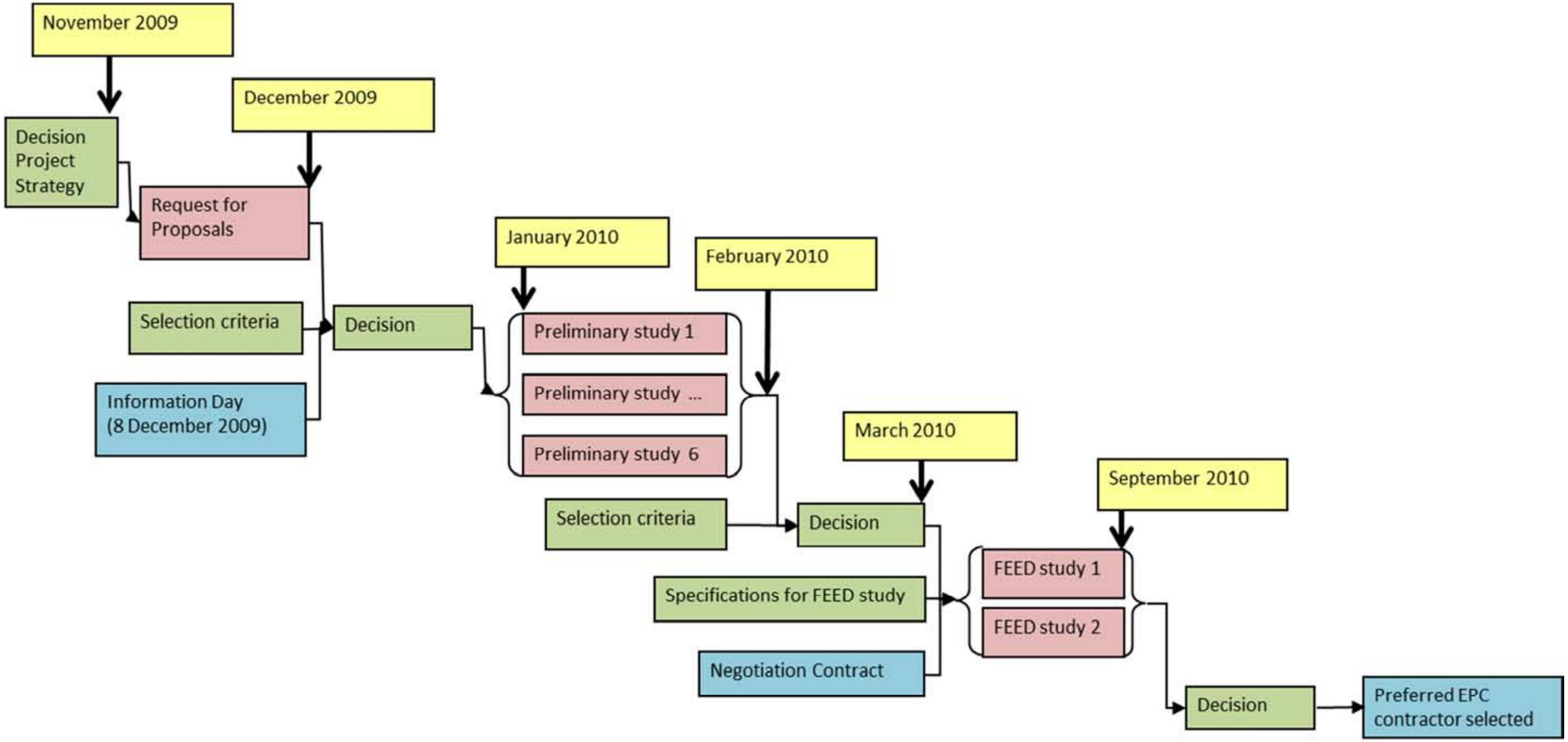
[6] *Handling and allocation of business risks – Special Report for the Global Carbon Capture and Storage Institute*; Bijkerk, M.; Global CCS Institute website (expected 2011)

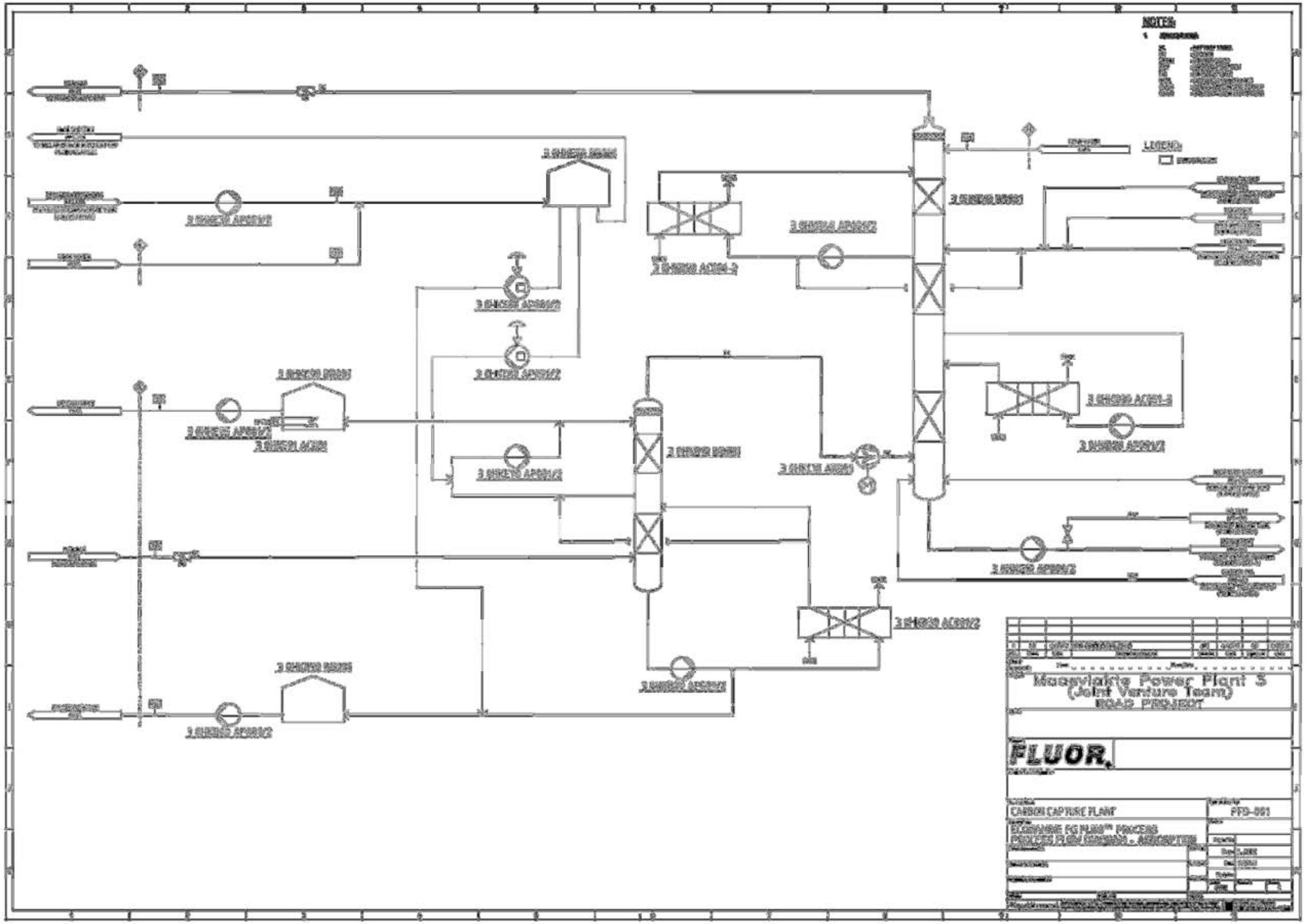
[7] *Stakeholder Management ROAD – Special Report for the Global Carbon Capture and Storage Institute*; Kombrink, M., Jonker, T. & Thonon, I.; GCCSI website (expected 2011)

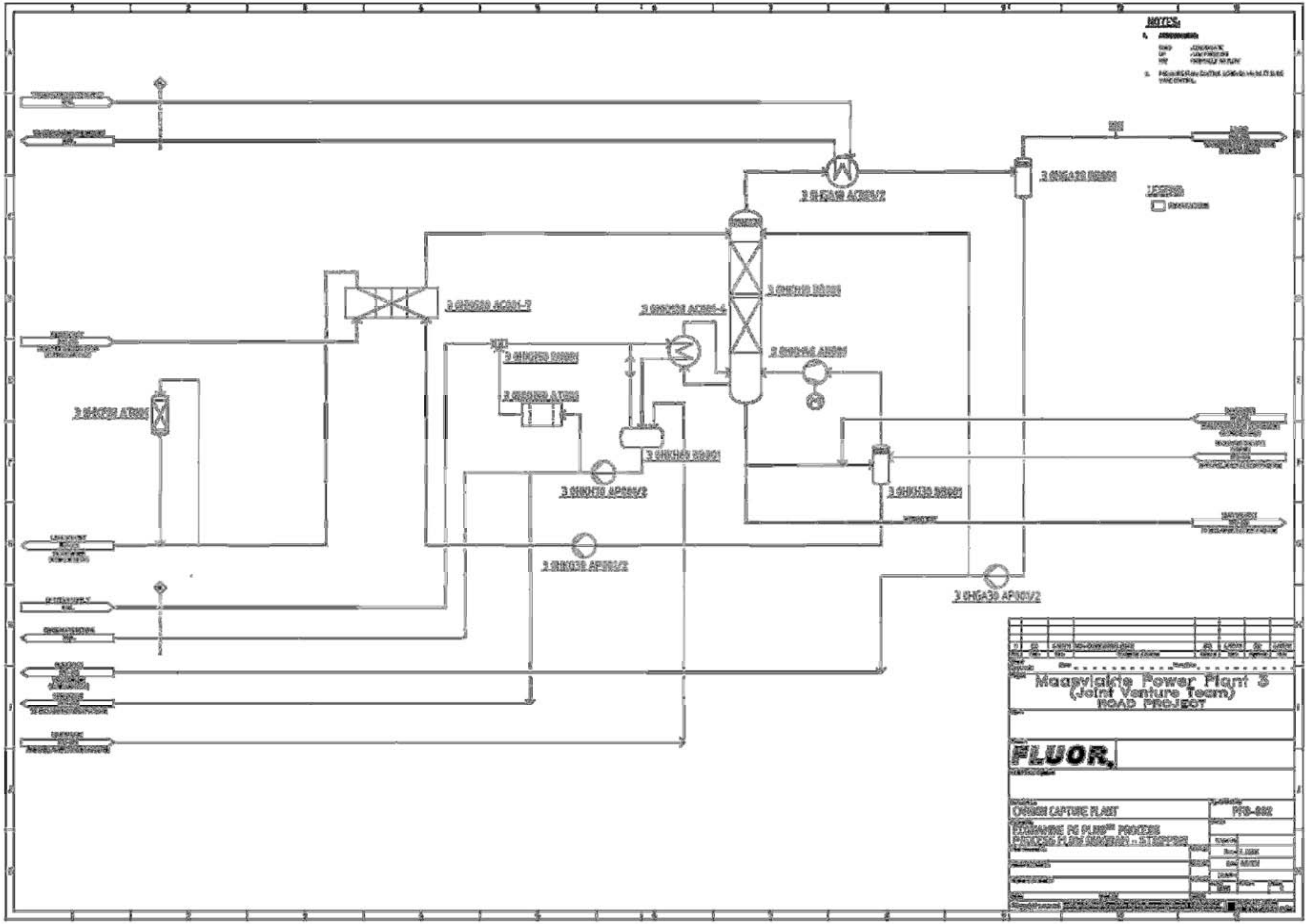
[8] *Project execution strategy – Special Report for the Global Carbon Capture and Storage Institute*; Al Azki, A.; Global CCS Institute website (expected 2011)

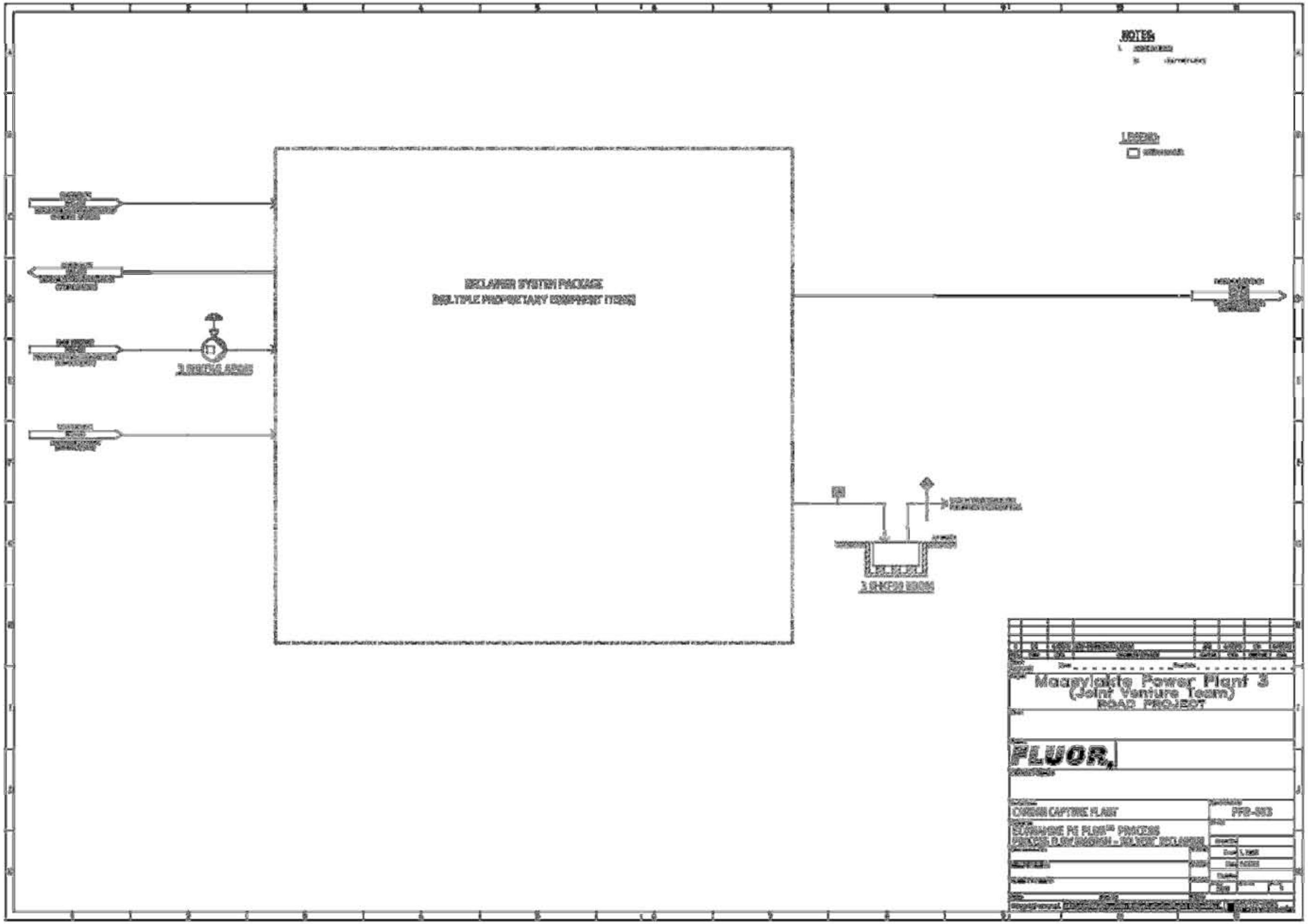
Appendix 1 Capture technology selection process

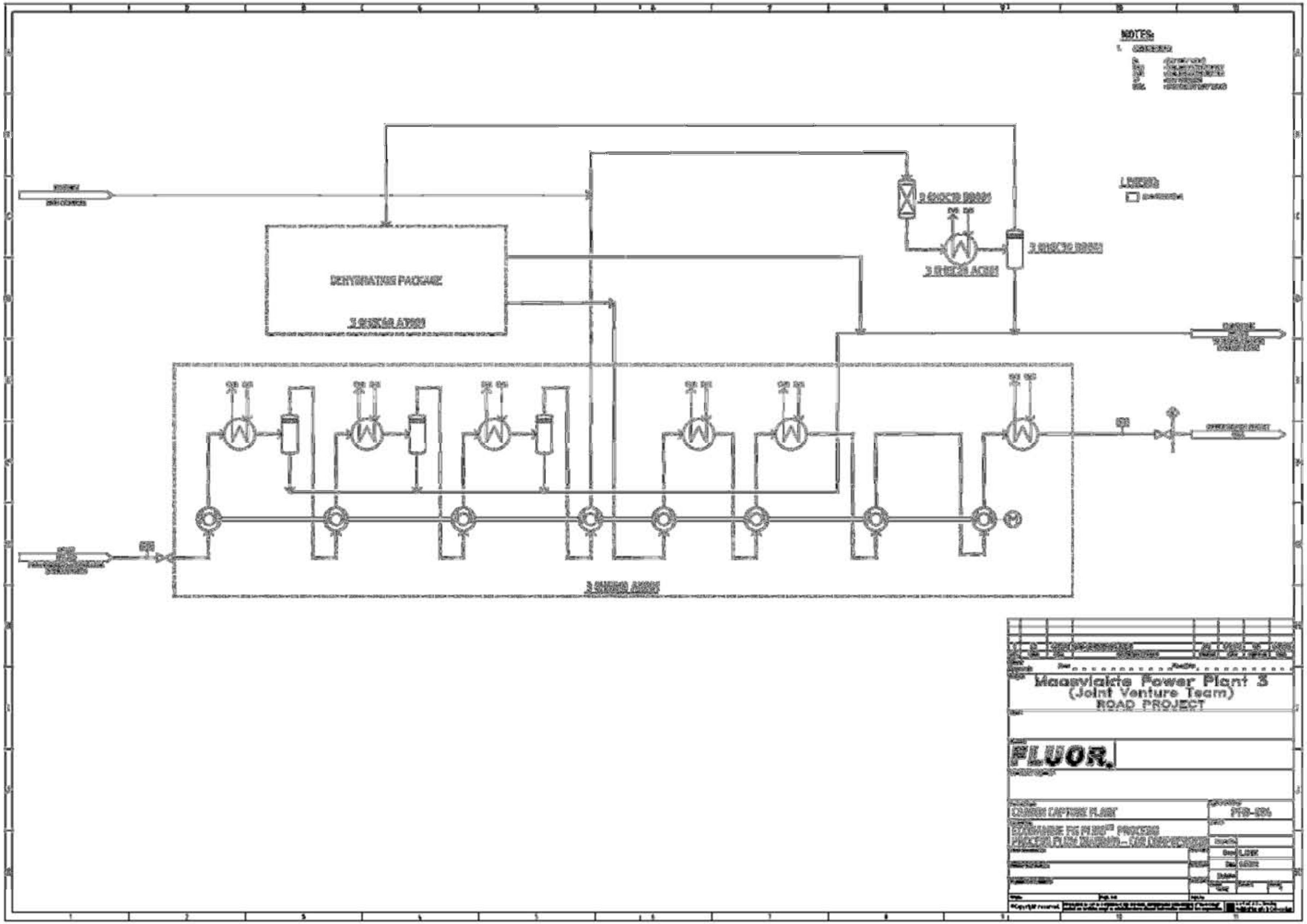
Below you find the schematic representation of the capture technology selection. More detailed information on this topic can be found in the report *CO_2 capture technology selection methodology – Special Report for the Global Carbon Capture and Storage Institute* [1]

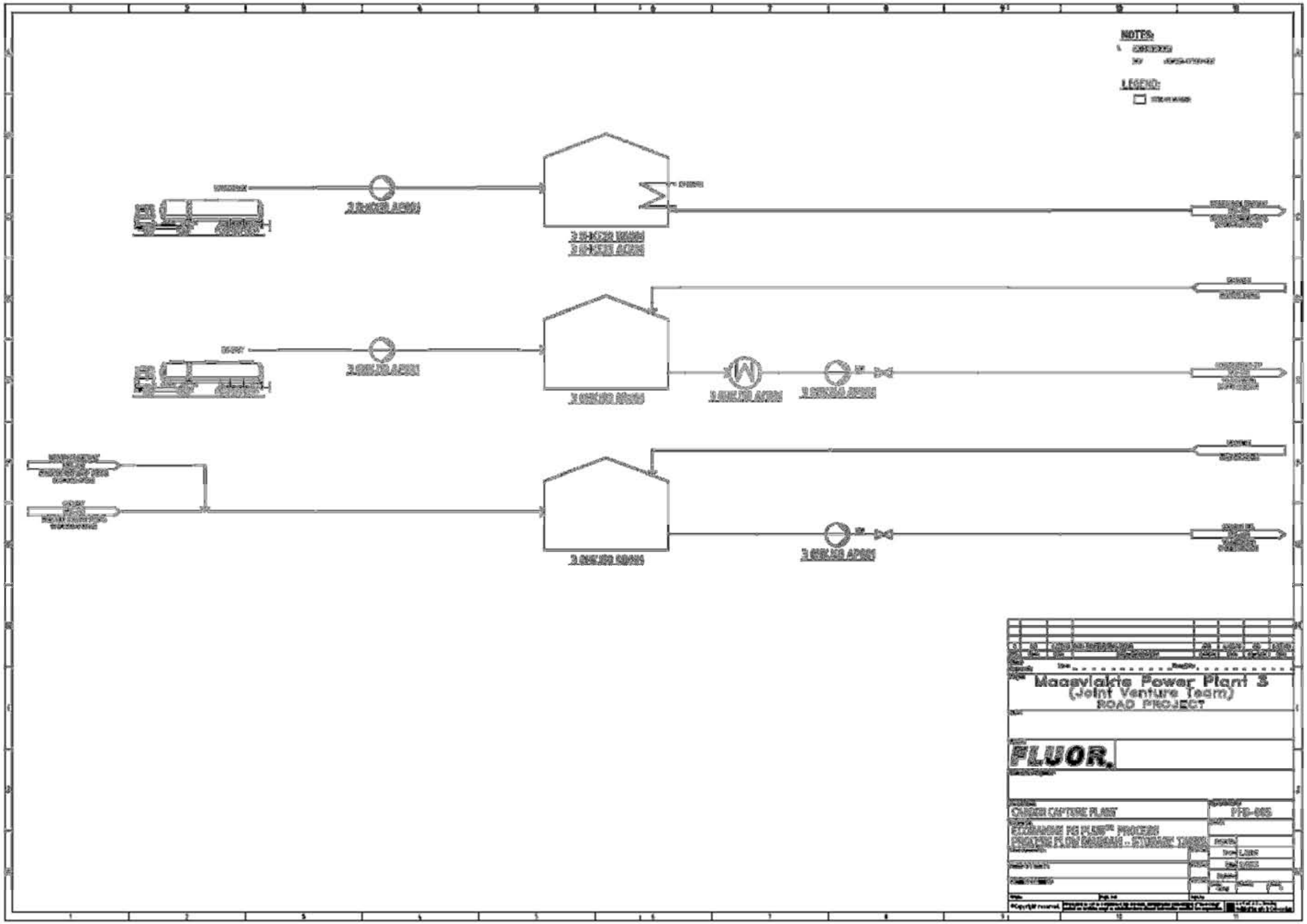


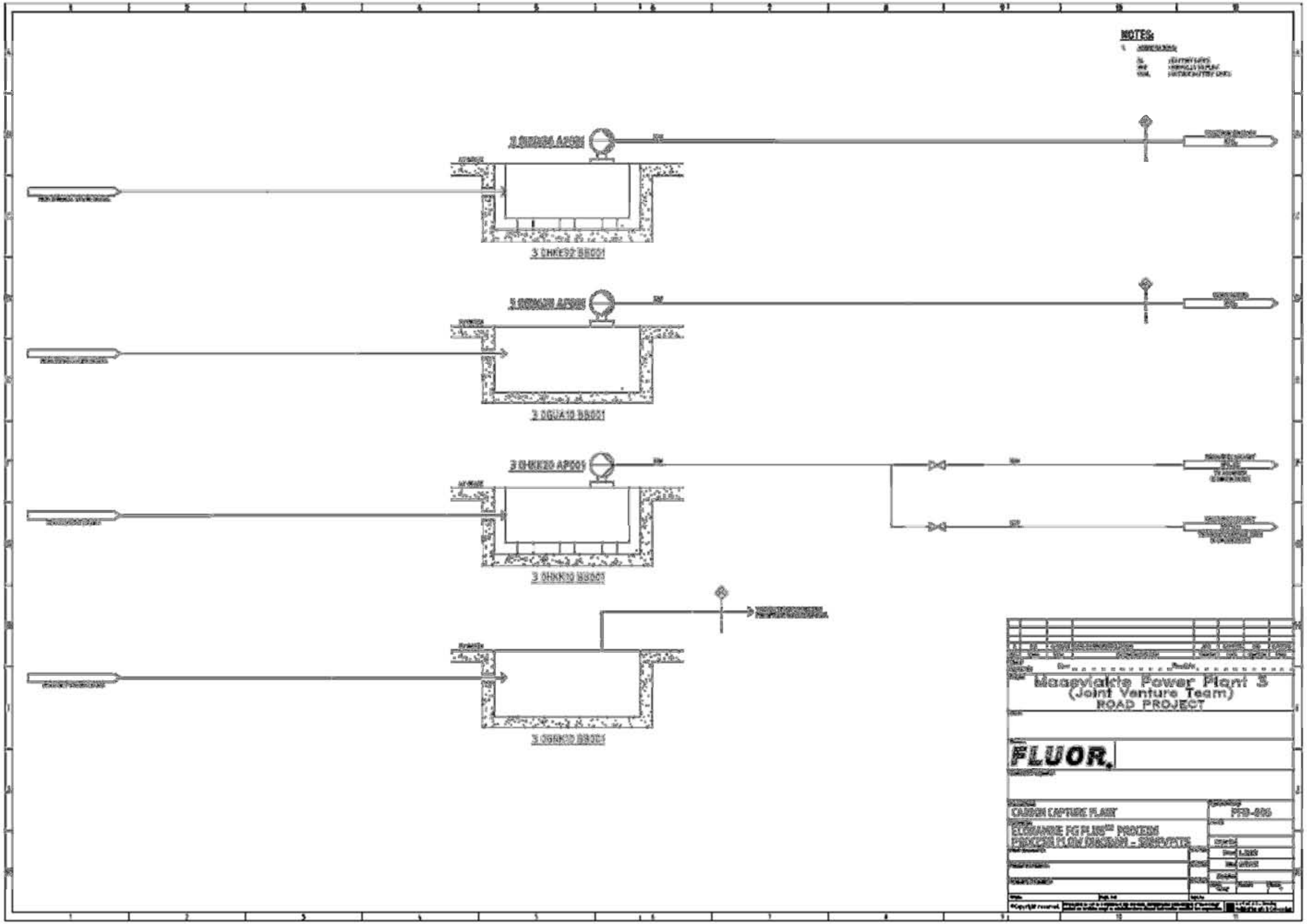

NOTES:
LEGEND:
Maasvlakte Power Plant 3
(Joint Venture Team)
ROAD PROJECT
FLUOR.
CARBON CAPTURE PLANT
PFD-001
ECONAMINE FG PLUS℠ PROCESS
PROCESS FLOW DIAGRAM - ABSORPTION

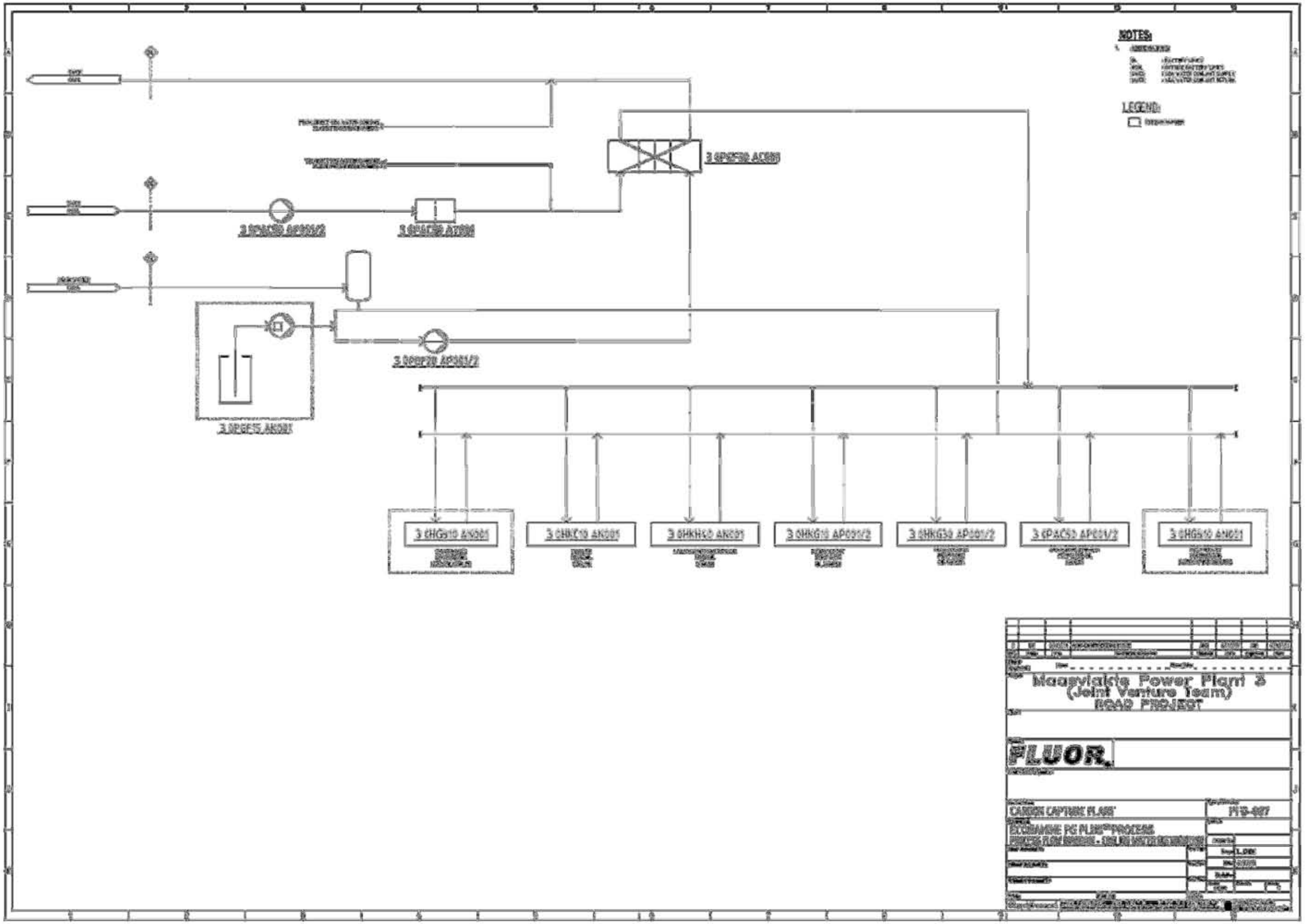

NOTES:
LEGEND
3 0HGA30 AP001/2
Maasvlakte Power Plant 3
(Joint Venture Team)
ROAD PROJECT
FLUOR
CARBON CAPTURE PLANT
ECONAMINE FG PLUS℠ PROCESS
PROCESS FLOW DIAGRAM - STRIPPER


NOTES:
LEGEND:
RECLAIMER SYSTEM PACKAGE
(MULTIPLE PROPRIETARY EQUIPMENT ITEMS)
Maasvlakte Power Plant 3
(Joint Venture Team)
ROAD PROJECT
FLUOR
CARBON CAPTURE PLANT
ECONAMINE FG PLUS SM PROCESS
PROCESS FLOW DIAGRAM - SOLVENT RECLAIMER
PFD-003


NOTES:
LEGEND:
DEHYDRATION PACKAGE
Maasvlakte Power Plant 3
(Joint Venture Team)
ROAD PROJECT
FLUOR
CARBON CAPTURE PLANT


NOTES:
LEGEND:
Maasvlakte Power Plant 3
(Joint Venture Team)
ROAD PROJECT
FLUOR.
CARBON CAPTURE PLANT
ECONAMINE FG PLUS℠ PROCESS
PROCESS FLOW DIAGRAM - STORAGE TANKS
PFD-005
©Copyright reserved.


NOTES:
3 0HKE92 BB001
3 0GUA10 BB001
3 0HKK20 AP001
3 0HKK10 BB001
3 0GNK10 BB001
Maasvlakte Power Plant 3
(Joint Venture Team)
ROAD PROJECT
FLUOR
CARBON CAPTURE PLANT
ECONAMINE FG PLUS℠ PROCESS


NOTES:
LEGEND:
3 0PAC50 AC001
3 0HKG10 AP001/2
3 0HKG30 AP001/2
3 0PAC50 AP001/2
3 0HKH60 AN001
3 0HKC10 AN001
Maasvlakte Power Plant 3
(Joint Venture Team)
ROAD PROJECT
FLUOR
CARBON CAPTURE PLANT
ECONAMINE FG PLUS℠ PROCESS

Appendix 3 Sized equipment list

Blowers / Compressors

Tag No.	Description	Motor Name plate	Size (LxW)	DN1/DN2	Metallurgy	Comments
		kW	m x m	mm		
3 0HKC10 AN001	Blower	2,250	15.6 X 4.0 + 2.5 X 1.6	(3250 x 2500) **/ 3370	316L SS with Carbon Ring Seals	Inlet Flowrate: 643530 Nm3/hr **The inlet nozzle is rectangular.
	Blower Electric Motor Driver	-			-	
3 0HKH40 AN001	Lean Vapor Compressor	2,750	8.0 x 4.0	DN 600 / DN 600	316L SS with Carbon Ring Seals	
	Lean Vapor Compressor Electric Motor Driver	-			-	
3 0HGB10 AN001	CO_2 Product Compressor	17,000	20.0 x 17.0	DN 500 / DN 80 **	316L SS	Inlet Flowrate: 90216 Nm3/hr Pressure Rise: 127.5 bar ** First stage inlet / Final stage outlet
	CO_2 Product Compressor Electric Motor Driver	-			-	

Exchangers

Tag No.	Description	TEMA Type or Plate Exch.	Size (per train)		Design Conditions				Metallurgy		Comments
					Shell/Hot (for Plate Type)		Tube/Cold (for Plate Type)		Shell/Frame	Tube/Plate	
			L m	W / D m	Pres bar(g)	Temp ℃	Pres bar(g)	Temp ℃			
3 0HKB30 AC001/2	DCC Water Cooler	Plate	4.2	1.1	6.5	125	6.5	125	CS	Titanium	
3 0HKD30 AC001-3	Absorber Intercooler	Plate	4.2	1.1	6.5	-20 / 125	6.5	-20 / 125	CS	Titanium	
3 0HKD50 AC001-3	Wash Water Cooler	Plate	4.2	1.1	7.5	125	7.5	125	CS	Titanium	
3 0HKG20 AC001-7	Solvent Cross Exchanger	Plate	4.2	1.1	11.0	140	11.0	140	CS	316L SS	
3 0HGA10 AC001/2	Condenser	BXU	7.9	2.2	30.0	160	38.0	160	CS / 304L SS cladding	316L SS	
3 0HKH20 AC001-4	Reboiler	Plate	11.8	2.0	6.0	320	6.0	320	CS	316L SS	
3 0HGC20 AC001	CATOX Aftercooler	AHU	7.70	1.5	24.0	125	18.5	125	316L SS	316L SS	
3 0PGF30 AC001	Sea Water / Fresh Water Cooler	Plate	5.2	1.1	6.5	125	6.5	125	CS	Titanium	

Filters

Tag No.	Description	Design Conditions		Metallurgy	Comments
		Pres bar(g)	Temp ℃		
3 0HKF20 AT001	Carbon Bed	11.0	140	316L SS	
3 0HKH80 AT001	Desuperheating Condensate Filter	23.0	160	316L SS	Inlet Flow Rate: 18.8 m^3/hr. Diameter: DN 200
3 0PAC50 AT001	Cooling Water Mussel Filter	6.5	125	Duplex	Inlet Flow Rate: 11963 m^3/hr. Diameter: DN 1200

Vessels

Tag No.	Description	Dimensions		Design Conditions		Metallurgy	Comments
		Diameter	T/T	Pres	Temp		
		m	m	bar(g)	°C		
3 0HKB10 BB001	Direct Contact Cooler	10.7	23.7	0.14/-0.2	125	CS / 304L SS cladding	
3 0HKD10 BB001	Absorber	12.8	40.3	0.14	125	CS / 304L SS cladding	
3 0HKH10 BB001	Stripper	7.5	26.3	3.5/FV	160	CS / 304L SS cladding	
3 0HGA20 BB001	Overhead Accumulator	3.3	5.8	3.5/FV	125	CS / 304L SS cladding	
3 0HKH30 BB001	Lean Flash Drum	7.8	12.8	3.5/FV	170	CS / 304L SS cladding	
3 0HKH60 BB001	Reboiler Condensate Drum	1.9	5.75	3.5/FV	380	CS	
3 0HGC30 BB001	Water KO Drum	1.7	3.5	23.0	125	316L SS	

Pumps

Tag No.	Description	Pump Type	Motor Name plate	Metallurgy		DN1/DN2 (Note 1)	Comments
			kW	Impeller	Casing	mm	
3 0HKB20 AP001/2	DCC Circulation Pump	Centrifugal	260	316L SS	316L SS	DN 450 / DN 400	One operating pump with one spare.
3 0HKE70 AP001/2	DCC Scrubbing Solution Pump	Centrifugal	80.0	316L SS	316L SS	DN 300 / DN 300	One operating pump with one spare.
3 0HKB60 AP001/2	DCC Excess Water Pump	Centrifugal	15.0	12% Cr	CS		One operating pump with one spare.
3 0HKE60 AP001/2	DCC NaOH Injection Pump	Positive Displacement	0..1	316L SS	316L SS		One operating pump with one spare.
3 0HKE30 AP001/2	50 wt% NaOH Pump	Centrifugal	0..1	12% Cr	CS		One operating pump with one spare. S-6.
3 0HKE10 AP001	NaOH Fill Pump	Centrifugal	7.7	12% Cr	CS		Normally not running. S-6
3 0HKD20 AP001/2	Absorber Intercooler Pump	Centrifugal	160	316L SS	316L SS	DN 400 / DN 400	One operating pump with one spare.
3 0HKD40 AP001/2	Wash Water Pump	Centrifugal	160	316L SS	316L SS	DN 300 / DN 250	One operating pump with one spare.
3 0HKG10 AP001/2	Rich Solvent Pump	Centrifugal	700	316L SS	316L SS	DN 500 / DN 350	One operating pump with one spare.
3 0HGA30 AP001/2	Reflux Pump	Centrifugal	11.0	316L SS	316L SS		One operating pump with one spare.
3 0HKG30 AP001/2	Lean Solvent Pump	Centrifugal	550	Super Duplex	Super Duplex	DN 500 / DN 350	One operating pump with one spare.
3 0HKH70 AP001/2	Reboiler Condensate Pump	Centrifugal	200	12% Cr	CS	DN 150 / DN 100	One operating pump with one spare.

Pumps *(continued)*

Tag No.	Description	Pump Type	Motor Name plate	Metallurgy		DN1/DN2 (Note 1)	Comments
			kW	Impeller	Casing	mm	
3 0HKJ10 AP001	Solvent Fill Pump	Centrifugal	2.0	316L SS	316L SS		Normally not running.
3 0HKJ40 AP001	Solvent Make-up Pump	Centrifugal	1.25	316L SS	316L SS		Normally not running.
3 0HKE80 AP001/2	Neutralization NaOH Injection Pump	Positive Displacement	0.55	316L SS	316L SS		One operating pump with one spare.
3 0HKK20 AP001	Solvent Sump Pump	Vertical	10.0	316L SS	316L SS		Normally not running.
3 0HKJ60 AP001	Solvent Transfer Pump	Centrifugal	16.0	316L SS	316L SS		Normally not running.
3 0HKF40 AP001	Reclaimer NaOH Injection Pump	Positive Displacement	0.55	316L SS	316L SS		One operating pump.
3 0PAC50 AP001/2	Sea Water Coolant Pump	Centrifugal	1,400	Super Duplex	Super Duplex	DN 1200 / DN 900	One operating pump with one spare.
3 0PGF20 AP001/2	Fresh Water Coolant Pump	Centrifugal	300	12% Cr	CS		One operating pump with one spare.
3 0HKE96 AP001	Chemical Sump Pump	Vertical	5.0	316L SS	316L SS		One operating pump.
3 0HKE95 AP001/2	DCC Blowdown Pump	Centrifugal	2.0	316L SS	316L SS		One operating pump with one spare.
3 0GUA20 AP001	Storm Water Sump Pump	Vertical	5.0	12% Cr	CS		One operating pump.

1. Nozzle sizes are shown only for pumps where vendor sizes were provided.

Sumps

Tag No.	Description	Dimensions			Metallurgy	Comments
		Length	Width	Depth		
		m	m	m		
3 0HKK10 BB001	Solvent Sump	3.3	3.3	4.0	316L SS tank in concrete sump	Located below grade. Design Temperature: 160 °C Design Pressure: 40 mbar(g)
3 0HKE92 BB001	Chemical Sump	3.5	3.5	4.0	316L SS tank in concrete sump	Located below grade. Design Temperature: 60°C Design Pressure: 40 mbar(g)
3 0HKF90 BB001	Reclaimer Waste Sump	3.8	3.8	2.3	316L SS tank in concrete sump	Located below grade. Design Temperature: 170 °C Design Pressure: 40 mbar(g)
3 0GUA10 BB001	Storm Water Sump	10.0	9.0	4.0	Concrete sump	Located below grade.
3 0GNK10 BB001	Oily Water Sump	4.0	4.0	4.0	Concrete / 316L SS lining.	Located below grade.

Tanks

Tag No.	Description	Dimensions		Metallurgy	Comments
		Diameter	T/T		
		m	m		
3 0HKJ20 BB001	Solvent Storage Tank	3.3	4.3	Duplex or CS with lining	Blanketed with Nitrogen.
3 0HKE20 BB001	50 wt% NaOH Storage Tank	4.3	5.7	CS	Tank to include 50 wt% NaOH Storage Tank Heating Coil 3 0HKE25 AC001.
3 0HKB50 BB001	DCC Water Neutralization Tank	3.7	4.9	CS	
3 0HKJ50 BB001	Solvent Holding Tank	9.9	13.1	Duplex or CS with lining	
3 0HKE50 BB001	20 wt% NaOH Storage Tank	1.3	1.7	CS	
3 0HKE90 BB001	DCC Blowdown Tank	1.6	2.1	316L SS	Tank to include DCC Blowdown Tank Heating Coil (3 0HKE91 AC001).
3 0PGF10 BB001	Circulating Water Expansion Tank	0.5	2.0	CS	

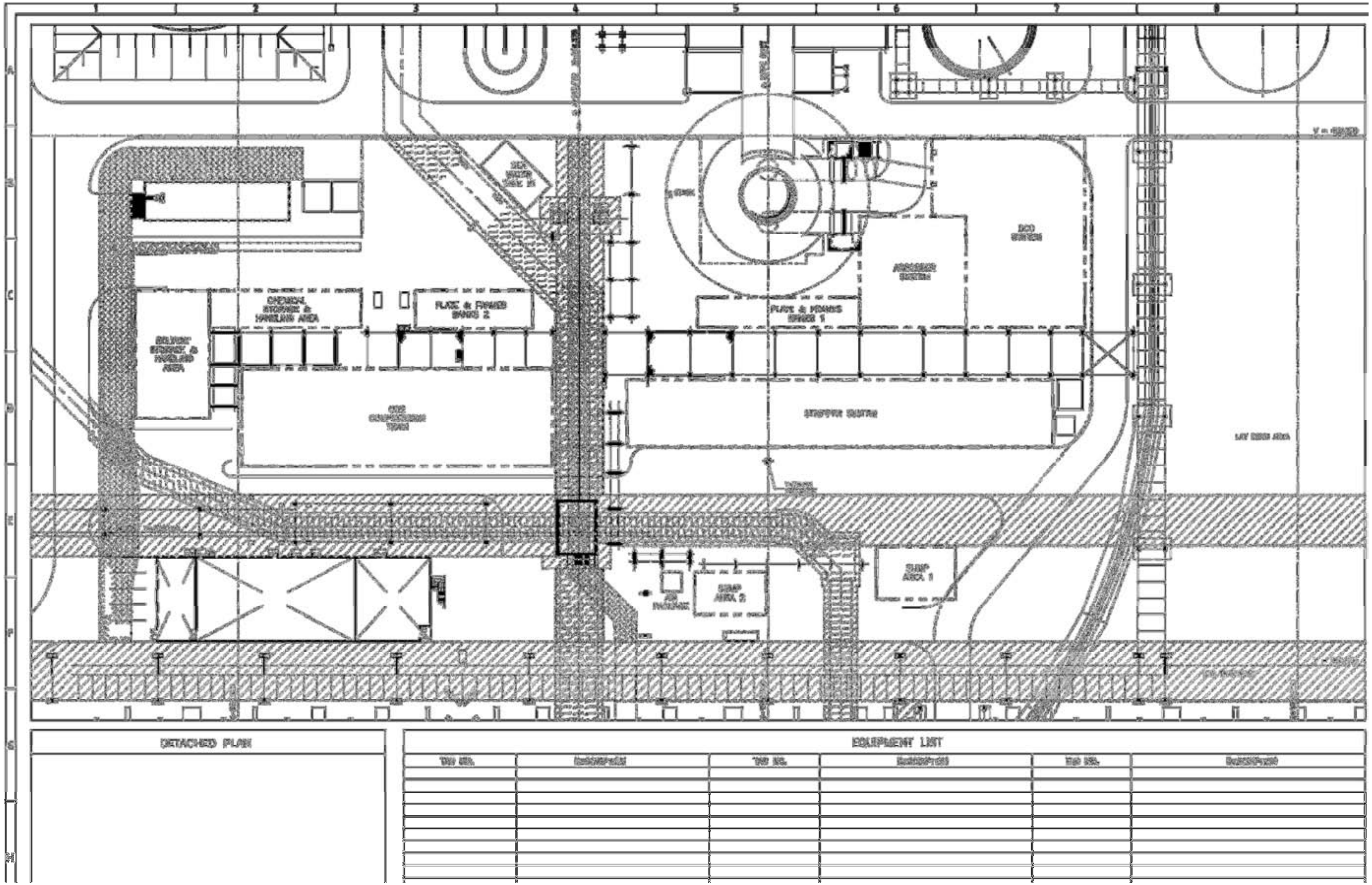
Specialty process

Tag No.	Description	Comments
3 0HKJ30 AC001	Solvent Make-up Pump Suction Heater	Electric heater with heat rate of 110 kW. CS / 304L SS cladding.
3 0HKH50 BN001	LP Steam Desuperheater	Materials of construction are as per manufacturer's standard.
3 0HGC40 AT001	CO_2 Dehydration Package	Reduces water content of product CO_2 to < 30 ppmv. MOC: per vendor standard.
3 0SCA10 AN001	Instrument and Plant Air Package	Includes instrument air compressor and dryer. Estimated capacity is 200 Nm^3/h. Supply pressure of 7 bar(g) at -20°C dewpoint. MOC: TBD
3 0HKE25 AC001	50 wt% NaOH Storage Tank Heating Coil	To be purchased as integral part of the 50 wt% NaOH storage Tank (3 0HKE20 BB001). MOC: CS

Specialty process (*continued*)

Tag No.	Description	Comments
3 0HGC10 BB001	CATOX Unit	CS / 304L SS Cladding. Reduces Oxygen content of product CO_2 to < 10 ppmw.
3 0HKE91 AC001	DCC Blowdown Tank Heating Coil	To be purchased as integral part of the DCC Blowdown Tank (3 0HKE90 BB001). MOC: Duplex
3 0PGF15 AK001	Cooling Water Treatment Package	Treatment package for circulating fresh water loop. Circulating fresh water flow rate = 1,540 m3/hr.
3 0PAC50 AK001	Sea Water Priming Package	Vacuum priming system.
3 0HKC05 BS001	Silencer	
	Reclaimer System Package	Multiple proprietary equipment items for separation of degredation products and recovery of solvent.

Appendix 4 Simplified Plot plan



Appendix 5 Material Safety Data Sheets and HAZID

Material Safety Data Sheets

Sodium hydroxide, purum, pellets
ACC# 95583

Section 1 - Chemical Product and Company Identification

MSDS Name: Sodium hydroxide, purum, pellets
Catalog Numbers: AC134070000, AC134070010, AC134070025
Synonyms: Caustic soda; Soda lye; Sodium hydrate; Lye.
Company Identification:

Acros Organics N.V.
One Reagent Lane
Fair Lawn, NJ 07410

For information in North America, call: 800-ACROS-01
For emergencies in the US, call CHEMTREC: 800-424-9300

Section 2 - Composition, Information on Ingredients

CAS#	Chemical Name	Percent	EINECS/ELINCS
1310-73-2	Sodium hydroxide	>98	215-185-5

Section 3 - Hazards Identification

EMERGENCY OVERVIEW
Appearance: white solid.
Danger! Causes eye and skin burns. May cause severe respiratory tract irritation with possible burns. May cause severe digestive tract irritation with possible burns. Hygroscopic (absorbs moisture from the air).
Target Organs: Eyes, skin, mucous membranes.

Potential Health Effects

Eye: Causes eye burns. May cause chemical conjunctivitis and corneal damage.
Skin: Causes skin burns. May cause deep, penetrating ulcers of the skin. May cause skin rash (in milder cases), and cold and clammy skin with cyanosis or pale color.
Ingestion: May cause severe and permanent damage to the digestive tract. Causes gastrointestinal tract burns. May cause perforation of the digestive tract. Causes severe pain, nausea, vomiting, diarrhea, and shock. May cause corrosion and permanent tissue destruction of the esophagus and digestive tract. May cause systemic effects.
Inhalation: Irritation may lead to chemical pneumonitis and pulmonary edema. Causes severe irritation of upper respiratory tract with coughing, burns, breathing difficulty, and possible coma. Causes chemical burns to the respiratory tract.
Chronic: Prolonged or repeated skin contact may cause dermatitis. Effects may be delayed.

Section 4 - First Aid Measures

Eyes: In case of contact, immediately flush eyes with plenty of water for a t least 15 minutes. Get medical aid immediately.
Skin: In case of contact, immediately flush skin with plenty of water for at least 15 minutes while removing contaminated clothing and shoes. Get medical aid immediately. Wash clothing before reuse.
Ingestion: If swallowed, do NOT induce vomiting. Get medical aid immediately. If victim is fully conscious, give a cupful of water. Never give anything by mouth to an unconscious person.
Inhalation: If inhaled, remove to fresh air. If not breathing, give artificial respiration. If breathing is difficult, give oxygen. Get medical aid.
Notes to Physician: Treat symptomatically and supportively.

Section 5 - Fire Fighting Measures

General Information: As in any fire, wear a self-contained breathing apparatus in pressure-demand, MSHA/NIOSH (approved or equivalent), and full protective gear. Use water spray to keep fire-exposed containers cool. Use water with caution and in flooding amounts. Contact with moisture or water may generate sufficient heat to ignite nearby combustible materials. Contact with metals may evolve flammable hydrogen gas.
Extinguishing Media: Substance is noncombustible; use agent most appropriate to extinguish surrounding fire. Do NOT get water inside containers.
Flash Point: Not applicable.
Autoignition Temperature: Not applicable.
Explosion Limits, Lower: Not available.
Upper: Not available.
NFPA Rating: (estimated) Health: 3; Flammability: 0; Instability: 1

Section 6 - Accidental Release Measures

General Information: Use proper personal protective equipment as indicated in Section 8.
Spills/Leaks: Vacuum or sweep up material and place into a suitable disposal container. Avoid runoff into storm sewers and ditches which lead to waterways. Clean up spills immediately, observing precautions in the Protective Equipment section. Avoid generating dusty conditions. Provide ventilation. Do not get water on spilled substances or inside containers.

Section 7 - Handling and Storage

Handling: Wash thoroughly after handling. Do not allow water to get into the container because of violent reaction. Minimize dust generation and accumulation. Do not get in eyes, on skin, or on clothing. Keep container tightly closed. Avoid ingestion and inhalation. Discard contaminated shoes. Use only with adequate ventilation.
Storage: Store in a tightly closed container. Store in a cool, dry, well-ventilated area away from incompatible substances. Keep away from metals. Corrosives area. Keep away from acids. Store protected from moisture. Containers must be tightly closed to prevent the conversion of NaOH to sodium carbonate by the CO_2 in air.

Section 8 - Exposure Controls, Personal Protection

Engineering Controls: Facilities storing or utilizing this material should be equipped with an eyewash facility and a safety shower. Use adequate general or local exhaust ventilation to keep airborne concentrations below the permissible exposure limits.
Exposure Limits

Chemical Name	ACGIH	NIOSH	OSHA - Final PELs
Sodium hydroxide	2 mg/m3 Ceiling	10 mg/m3 IDLH	2 mg/m3 TWA

OSHA Vacated PELs: Sodium hydroxide: No OSHA Vacated PELs are listed for this chemical.
Personal Protective Equipment
Eyes: Wear chemical splash goggles.
Skin: Wear appropriate protective gloves to prevent skin exposure.
Clothing: Wear appropriate protective clothing to prevent skin exposure.
Respirators: A respiratory protection program that meets OSHA's 29 CFR 1910.134 and ANSI Z88.2 requirements or European Standard EN 149 must be followed whenever workplace conditions warrant respirator use.

Section 9 - Physical and Chemical Properties

Physical State: Solid
Appearance: white
Odor: Odorless
pH: 14 (5% aq soln)
Vapor Pressure: 1 mm Hg @739 deg C
Vapor Density: Not available.
Evaporation Rate: Not available.
Viscosity: Not available.
Boiling Point: 1390 deg C @ 760 mm Hg
Freezing/Melting Point:318 deg C
Decomposition Temperature: Not available.
Solubility: Soluble.
Specific Gravity/Density:2.13 g/cm3
Molecular Formula: NaOH
Molecular Weight:40.00

Section 10 - Stability and Reactivity

Chemical Stability: Stable at room temperature in closed containers under normal storage and handling conditions.
Conditions to Avoid: Moisture, contact with water, exposure to moist air or water, prolonged exposure to air.
Incompatibilities with Other Materials: Acids, water, flammable liquids, organic halogens, metals, aluminum, zinc, tin, leather, wool, nitromethane.
Hazardous Decomposition Products: Toxic fumes of sodium oxide.
Hazardous Polymerization: Will not occur.

Section 11 - Toxicological Information

RTECS#:
CAS# 1310-73-2: WB4900000
LD50/LC50:
CAS# 1310-73-2:
Draize test, rabbit, eye: 400 ug Mild;
Draize test, rabbit, eye: 1% Severe;
Draize test, rabbit, eye: 50 ug/24H Severe;
Draize test, rabbit, eye: 1 mg/24H Severe;
Draize test, rabbit, skin: 500 mg/24H Severe;

Carcinogenicity:
CAS# 1310-73-2: Not listed by ACGIH, IARC, NTP, or CA Prop 65.

Epidemiology: No information found
Teratogenicity: No information found
Reproductive Effects: No information found
Mutagenicity: See actual entry in RTECS for complete information.
Neurotoxicity: No information found
Other Studies:

Section 12 - Ecological Information

No information available.

Section 13 - Disposal Considerations

Chemical waste generators must determine whether a discarded chemical is classified as a hazardous waste. US EPA guidelines for the classification determination are listed in 40 CFR Parts 261.3. Additionally, waste generators must consult state and local hazardous waste regulations to ensure complete and accurate classification.
RCRA P-Series: None listed.
RCRA U-Series: None listed.

Section 14 - Transport Information

	US DOT	**Canada TDG**
Shipping Name:	SODIUM HYDROXIDE, SOLID	SODIUM HYDROXIDE SOLID
Hazard Class:	8	8
UN Number:	UN1823	UN1823
Packing Group:	II	II

Section 15 - Regulatory Information

US FEDERAL

TSCA
CAS# 1310-73-2 is listed on the TSCA inventory.
Health & Safety Reporting List
None of the chemicals are on the Health & Safety Reporting List.
Chemical Test Rules
None of the chemicals in this product are under a Chemical Test Rule.
Section 12b
None of the chemicals are listed under TSCA Section 12b.
TSCA Significant New Use Rule
None of the chemicals in this material have a SNUR under TSCA.
CERCLA Hazardous Substances and corresponding RQs
CAS# 1310-73-2: 1000 lb final RQ; 454 kg final RQ
SARA Section 302 Extremely Hazardous Substances
None of the chemicals in this product have a TPQ.
SARA Codes
CAS # 1310-73-2: immediate, reactive.
Section 313 No chemicals are reportable under Section 313.
Clean Air Act:
This material does not contain any hazardous air pollutants.
This material does not contain any Class 1 Ozone depletors.
This material does not contain any Class 2 Ozone depletors.
Clean Water Act:
CAS# 1310-73-2 is listed as a Hazardous Substance under the CWA.
None of the chemicals in this product are listed as Priority Pollutants under the CWA.
None of the chemicals in this product are listed as Toxic Pollutants under the CWA.
OSHA:
None of the chemicals in this product are considered highly hazardous by OSHA.
STATE
CAS# 1310-73-2 can be found on the following state right to know lists: California, New Jersey, Pennsylvania, Minnesota, Massachusetts.

California Prop 65

California No Significant Risk Level: None of the chemicals in this product are listed.

European/International Regulations
European Labeling in Accordance with EC Directives
Hazard Symbols:
C
Risk Phrases:
R 35 Causes severe burns.

Safety Phrases:
S 26 In case of contact with eyes, rinse immediately with plenty of water and seek medical advice.
S 37/39 Wear suitable gloves and eye/face protection.

S 45 In case of accident or if you feel unwell, seek medical advice immediately (show the label where possible).

WGK (Water Danger/Protection)
CAS# 1310-73-2: 1
Canada - DSL/NDSL
CAS# 1310-73-2 is listed on Canada's DSL List.
Canada - WHMIS
This product has a WHMIS classification of E.
This product has been classified in accordance with the hazard criteria of the Controlled Products Regulations and the MSDS contains all of the information required by those regulations.
Canadian Ingredient Disclosure List
CAS# 1310-73-2 is listed on the Canadian Ingredient Disclosure List.

Section 16 - Additional Information

MSDS Creation Date: 8/03/2001
Revision #4 Date: 3/16/2007

INEOS Oxide

MATERIAL SAFETY DATA SHEET

Product Name: Monoethanolamine Electronics Grade
MSDS #: INEOS Oxide 55922
Revision Date: October 1, 2006

Page # 1 of 9

Protective Clothing	NFPA (USA)	EU Classification	WHMIS (Canada)	Transportation
		 Corrosive		

Section 1: Product and Company Information

Product Name:	Monoethanolamine Electronics Grade
Product Code:	55922
Manufacturer:	INEOS Oxide Block 5501 21255 A Louisiana Hwy. 1 South Plaquemine, LA 70764
Phone Number:	(866) 865-4767
24-hour Emergency:	CHEMTREC: (800) 424-9300

Section 2: Composition and Ingredient Information

Common Name	Chemical Name	CAS No.	EINECS / ELINCS	Wt.%	Symbol	R Phrases
MEA	Monoethanolamine	141-43-5	205-483-3	99 - 99.9	Xn; C	R20/21/22-34
DEA	Diethanolamine	111-42-2	203-868-0	0 - 0.1	Xn;Xi	R22-48/22-38-41

Note: See Section 8 of this MSDS for exposure limit data for these ingredients.
See Section 16 for the full text of the R-phrases above.

Section 3: Hazards Identification

Preparation Hazards and Classification:

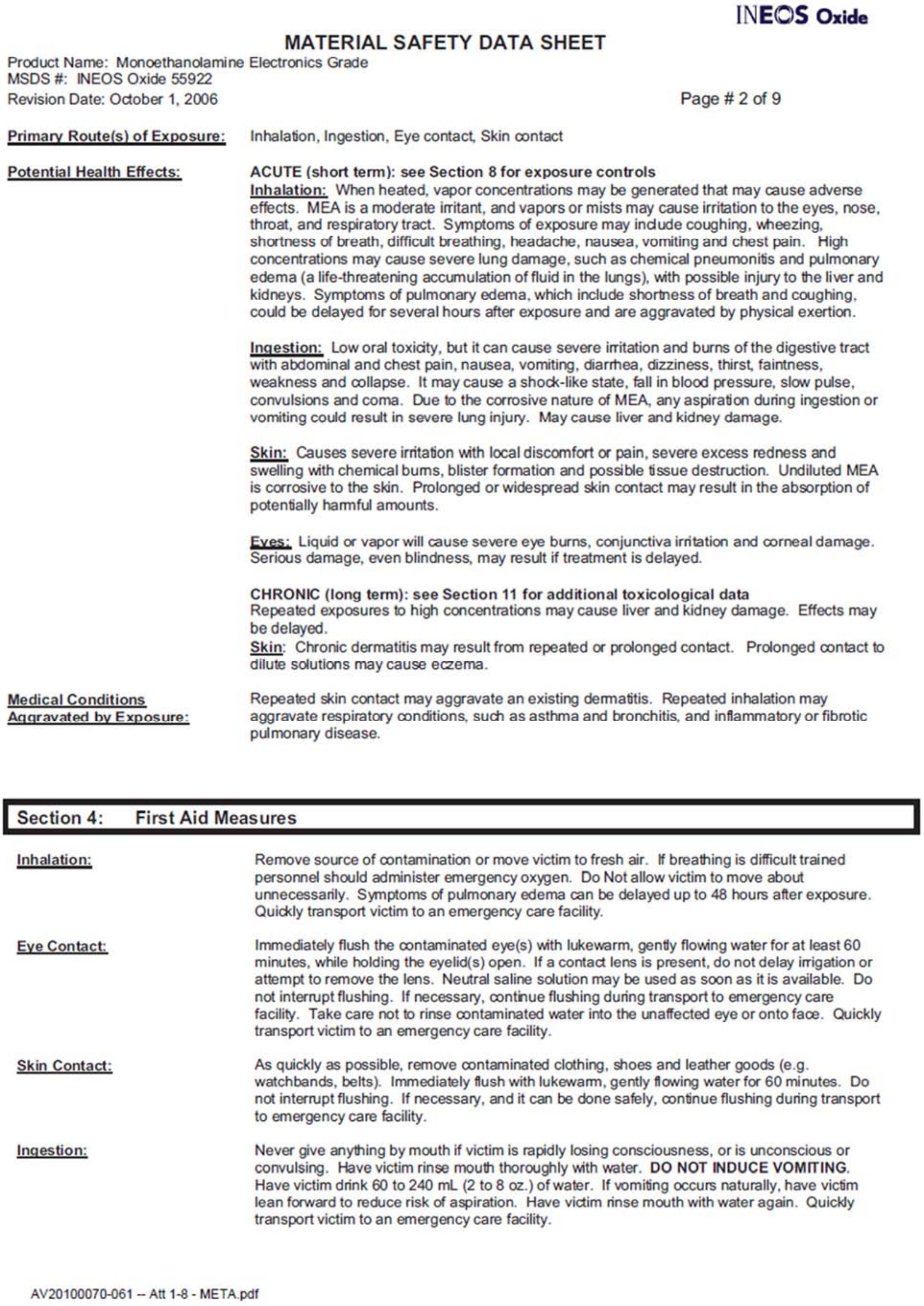
Corrosive, causes burns. Harmful by inhalation, in contact with skin and if swallowed.

USA: This product meets the criteria for hazardous material as defined by 29 CFR1910.1200, OSHA Hazard Communication Evaluation.

Canada: This is a controlled product under WHMIS.

European Communities (EC): This preparation is classified as Corrosive according to Directive 1999/45/EC and its amendments.

Appearance, Color and Odor: Colorless liquid; Slight ammonia, pyridine-like odor.



INEOS Oxide

MATERIAL SAFETY DATA SHEET

Product Name: Monoethanolamine Electronics Grade
MSDS #: INEOS Oxide 55922
Revision Date: October 1, 2006

Page # 2 of 9

Primary Route(s) of Exposure: Inhalation, Ingestion, Eye contact, Skin contact

Potential Health Effects:

ACUTE (short term): see Section 8 for exposure controls

Inhalation: When heated, vapor concentrations may be generated that may cause adverse effects. MEA is a moderate irritant, and vapors or mists may cause irritation to the eyes, nose, throat, and respiratory tract. Symptoms of exposure may include coughing, wheezing, shortness of breath, difficult breathing, headache, nausea, vomiting and chest pain. High concentrations may cause severe lung damage, such as chemical pneumonitis and pulmonary edema (a life-threatening accumulation of fluid in the lungs), with possible injury to the liver and kidneys. Symptoms of pulmonary edema, which include shortness of breath and coughing, could be delayed for several hours after exposure and are aggravated by physical exertion.

Ingestion: Low oral toxicity, but it can cause severe irritation and burns of the digestive tract with abdominal and chest pain, nausea, vomiting, diarrhea, dizziness, thirst, faintness, weakness and collapse. It may cause a shock-like state, fall in blood pressure, slow pulse, convulsions and coma. Due to the corrosive nature of MEA, any aspiration during ingestion or vomiting could result in severe lung injury. May cause liver and kidney damage.

Skin: Causes severe irritation with local discomfort or pain, severe excess redness and swelling with chemical burns, blister formation and possible tissue destruction. Undiluted MEA is corrosive to the skin. Prolonged or widespread skin contact may result in the absorption of potentially harmful amounts.

Eyes: Liquid or vapor will cause severe eye burns, conjunctiva irritation and corneal damage. Serious damage, even blindness, may result if treatment is delayed.

CHRONIC (long term): see Section 11 for additional toxicological data
Repeated exposures to high concentrations may cause liver and kidney damage. Effects may be delayed.
Skin: Chronic dermatitis may result from repeated or prolonged contact. Prolonged contact to dilute solutions may cause eczema.

Medical Conditions Aggravated by Exposure: Repeated skin contact may aggravate an existing dermatitis. Repeated inhalation may aggravate respiratory conditions, such as asthma and bronchitis, and inflammatory or fibrotic pulmonary disease.

Section 4: First Aid Measures

Inhalation: Remove source of contamination or move victim to fresh air. If breathing is difficult trained personnel should administer emergency oxygen. Do Not allow victim to move about unnecessarily. Symptoms of pulmonary edema can be delayed up to 48 hours after exposure. Quickly transport victim to an emergency care facility.

Eye Contact: Immediately flush the contaminated eye(s) with lukewarm, gently flowing water for at least 60 minutes, while holding the eyelid(s) open. If a contact lens is present, do not delay irrigation or attempt to remove the lens. Neutral saline solution may be used as soon as it is available. Do not interrupt flushing. If necessary, continue flushing during transport to emergency care facility. Take care not to rinse contaminated water into the unaffected eye or onto face. Quickly transport victim to an emergency care facility.

Skin Contact: As quickly as possible, remove contaminated clothing, shoes and leather goods (e.g. watchbands, belts). Immediately flush with lukewarm, gently flowing water for 60 minutes. Do not interrupt flushing. If necessary, and it can be done safely, continue flushing during transport to emergency care facility.

Ingestion: Never give anything by mouth if victim is rapidly losing consciousness, or is unconscious or convulsing. Have victim rinse mouth thoroughly with water. **DO NOT INDUCE VOMITING.** Have victim drink 60 to 240 mL (2 to 8 oz.) of water. If vomiting occurs naturally, have victim lean forward to reduce risk of aspiration. Have victim rinse mouth with water again. Quickly transport victim to an emergency care facility.

AV20100070-061 -- Att 1-8 - META.pdf

INEOS Oxide

MATERIAL SAFETY DATA SHEET

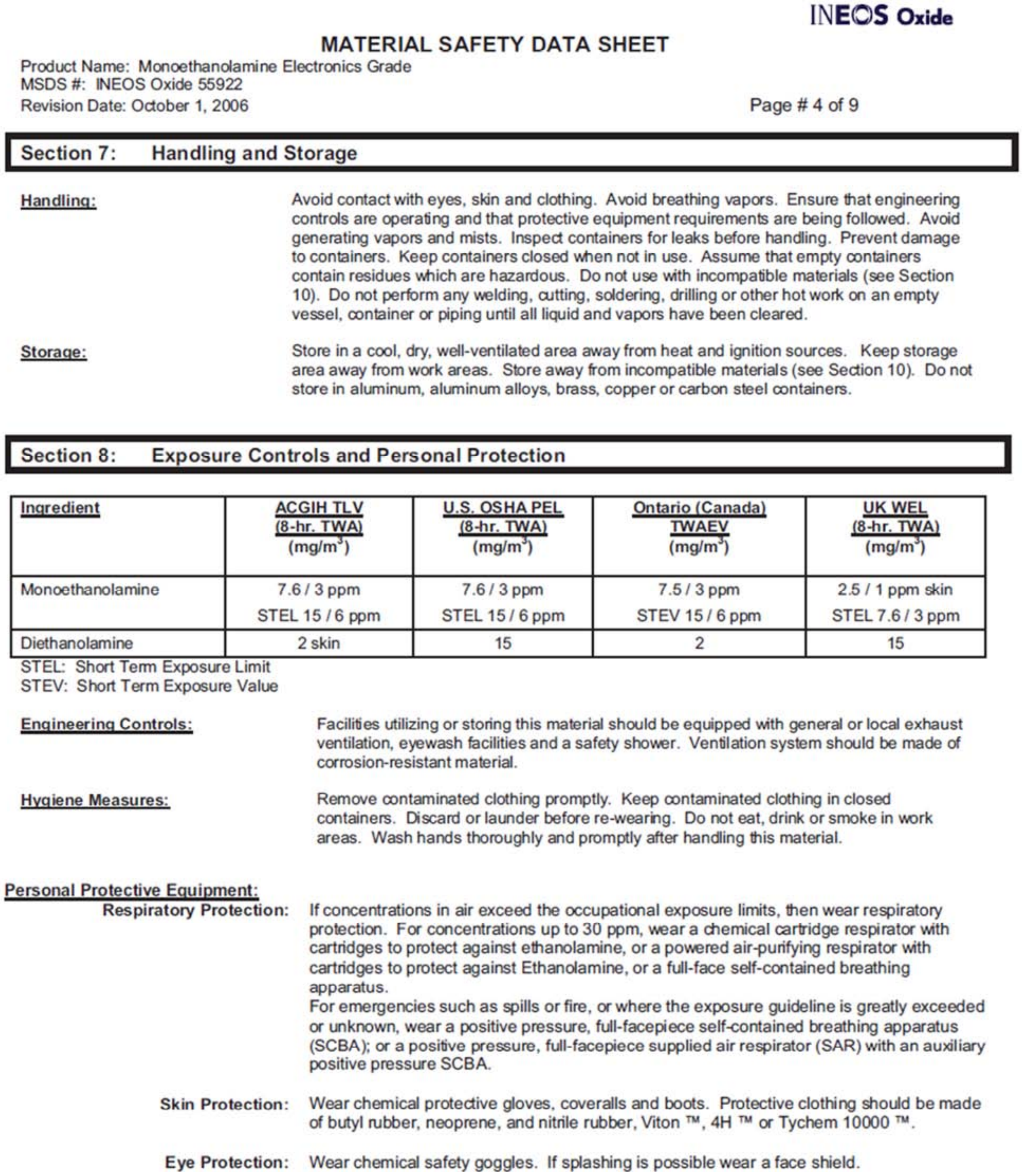
Product Name: Monoethanolamine Electronics Grade
MSDS #: INEOS Oxide 55922
Revision Date: October 1, 2006

Section 5: Fire Fighting Measures

Flash Point and Method (°C):	86 - 94°C (186 - 201°F)
Flammability Limits (%):	Lower: 3% at 140°C Upper: 23.5% at 140°C
Auto Ignition Temperature (°C):	410°C (770°F)
Extinguishing Media:	Water fog or fine spray, alcohol-resistant foam or dry chemical. Use water spray to cool fire-exposed containers.
Unusual Fire and Explosion Hazards:	Sensitivity to mechanical impact: Not available Sensitivity to static discharge: Not available
Fire Fighting Instructions:	Evacuate the area and fight fire from a safe distance or a protected location. Approach the fire from upwind to avoid hazardous vapors. Water spray may be used to flush spills away from ignition sources. Do not use a solid water stream as it may spread the fire. Avoid contact with this material during fire fighting operations. Wear chemical resistant clothing (chemical splash suit) and positive-pressure self-contained breathing apparatus.
Hazardous Combustion Products:	During a fire, smoke may contain vaporized MEA. Combustion products may include toxic nitrogen oxide, hydrogen cyanide, formaldehyde carbon monoxide, carbon dioxide and ammonia gases.
Other Flammability Information:	Violent steam generation or eruption may occur upon application of direct water stream to hot liquids. Spills of these organic liquids on hot fibrous insulation may lead to lowering of the auto ignition temperatures possibly resulting in spontaneous combustion.

Section 6: Accidental Release Measures

Personal Precautions:	Isolate the area; keep all unprotected people away from the spill area. Ventilate the area. Wear all proper personal protective equipment as indicated in Section 8. Prevent all inhalation exposures, skin and possible eye contact. Ensure clean-up is conducted by trained personnel only. Do not touch the spilled material. Extinguish or remove all ignition sources. Spilled material may pose a slipping hazard.
Environmental Precautions:	Prevent material from contaminating soil and from entering sewers or waterways.
Methods for Containment:	Isolate the spill area. Stop the spill if it is safe to do so. Keep materials which can burn away from spilled material. Contain the spill with earth, sand or suitable absorbent. Do not absorb with sawdust, woodchips or other cellulose materials.
Methods for Clean-up:	Clean up spills immediately. Scoop up spilled material and any contaminated absorbents into appropriate, labeled containers. Contaminated absorbent may pose the same hazards as the spilled product. Flush the area with water. Notify occupational and environmental authorities.



INEOS Oxide

MATERIAL SAFETY DATA SHEET

Product Name: Monoethanolamine Electronics Grade
MSDS #: INEOS Oxide 55922
Revision Date: October 1, 2006

Page # 4 of 9

Section 7: Handling and Storage

Handling: Avoid contact with eyes, skin and clothing. Avoid breathing vapors. Ensure that engineering controls are operating and that protective equipment requirements are being followed. Avoid generating vapors and mists. Inspect containers for leaks before handling. Prevent damage to containers. Keep containers closed when not in use. Assume that empty containers contain residues which are hazardous. Do not use with incompatible materials (see Section 10). Do not perform any welding, cutting, soldering, drilling or other hot work on an empty vessel, container or piping until all liquid and vapors have been cleared.

Storage: Store in a cool, dry, well-ventilated area away from heat and ignition sources. Keep storage area away from work areas. Store away from incompatible materials (see Section 10). Do not store in aluminum, aluminum alloys, brass, copper or carbon steel containers.

Section 8: Exposure Controls and Personal Protection

Ingredient	ACGIH TLV (8-hr. TWA) (mg/m^3)	U.S. OSHA PEL (8-hr. TWA) (mg/m^3)	Ontario (Canada) TWAEV (mg/m^3)	UK WEL (8-hr. TWA) (mg/m^3)
Monoethanolamine	7.6 / 3 ppm STEL 15 / 6 ppm	7.6 / 3 ppm STEL 15 / 6 ppm	7.5 / 3 ppm STEV 15 / 6 ppm	2.5 / 1 ppm skin STEL 7.6 / 3 ppm
Diethanolamine	2 skin	15	2	15

STEL: Short Term Exposure Limit
STEV: Short Term Exposure Value

Engineering Controls: Facilities utilizing or storing this material should be equipped with general or local exhaust ventilation, eyewash facilities and a safety shower. Ventilation system should be made of corrosion-resistant material.

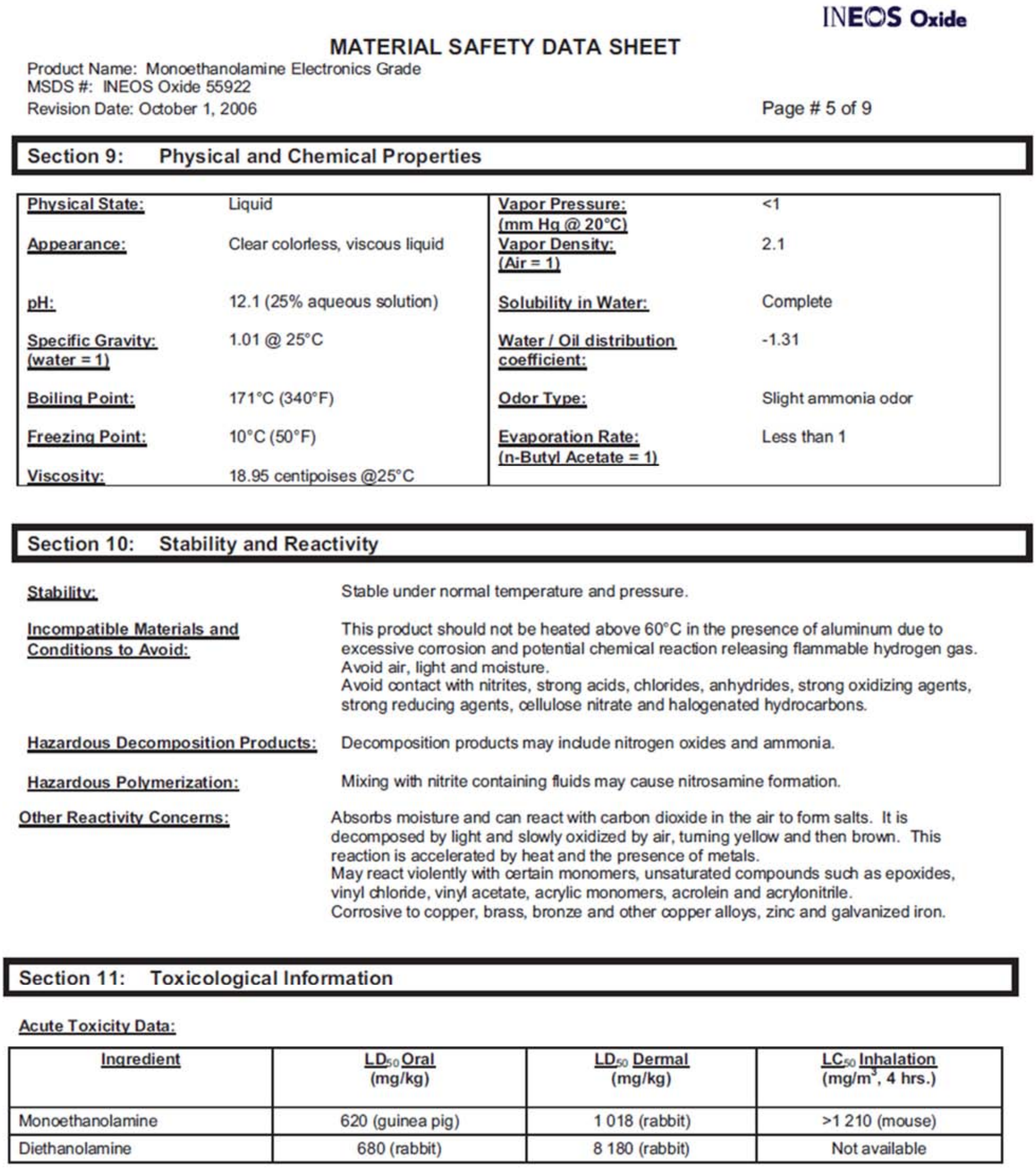
Hygiene Measures: Remove contaminated clothing promptly. Keep contaminated clothing in closed containers. Discard or launder before re-wearing. Do not eat, drink or smoke in work areas. Wash hands thoroughly and promptly after handling this material.

Personal Protective Equipment:

Respiratory Protection: If concentrations in air exceed the occupational exposure limits, then wear respiratory protection. For concentrations up to 30 ppm, wear a chemical cartridge respirator with cartridges to protect against ethanolamine, or a powered air-purifying respirator with cartridges to protect against Ethanolamine, or a full-face self-contained breathing apparatus.
For emergencies such as spills or fire, or where the exposure guideline is greatly exceeded or unknown, wear a positive pressure, full-facepiece self-contained breathing apparatus (SCBA); or a positive pressure, full-facepiece supplied air respirator (SAR) with an auxiliary positive pressure SCBA.

Skin Protection: Wear chemical protective gloves, coveralls and boots. Protective clothing should be made of butyl rubber, neoprene, and nitrile rubber, Viton ™, 4H ™ or Tychem 10000 ™.

Eye Protection: Wear chemical safety goggles. If splashing is possible wear a face shield.



INEOS Oxide

MATERIAL SAFETY DATA SHEET

Product Name: Monoethanolamine Electronics Grade
MSDS #: INEOS Oxide 55922
Revision Date: October 1, 2006

Page # 5 of 9

Section 9: Physical and Chemical Properties

Physical State:	Liquid	**Vapor Pressure: (mm Hg @ 20°C)**	<1
Appearance:	Clear colorless, viscous liquid	**Vapor Density: (Air = 1)**	2.1
pH:	12.1 (25% aqueous solution)	**Solubility in Water:**	Complete
Specific Gravity: (water = 1)	1.01 @ 25°C	**Water / Oil distribution coefficient:**	-1.31
Boiling Point:	171°C (340°F)	**Odor Type:**	Slight ammonia odor
Freezing Point:	10°C (50°F)	**Evaporation Rate: (n-Butyl Acetate = 1)**	Less than 1
Viscosity:	18.95 centipoises @25°C		

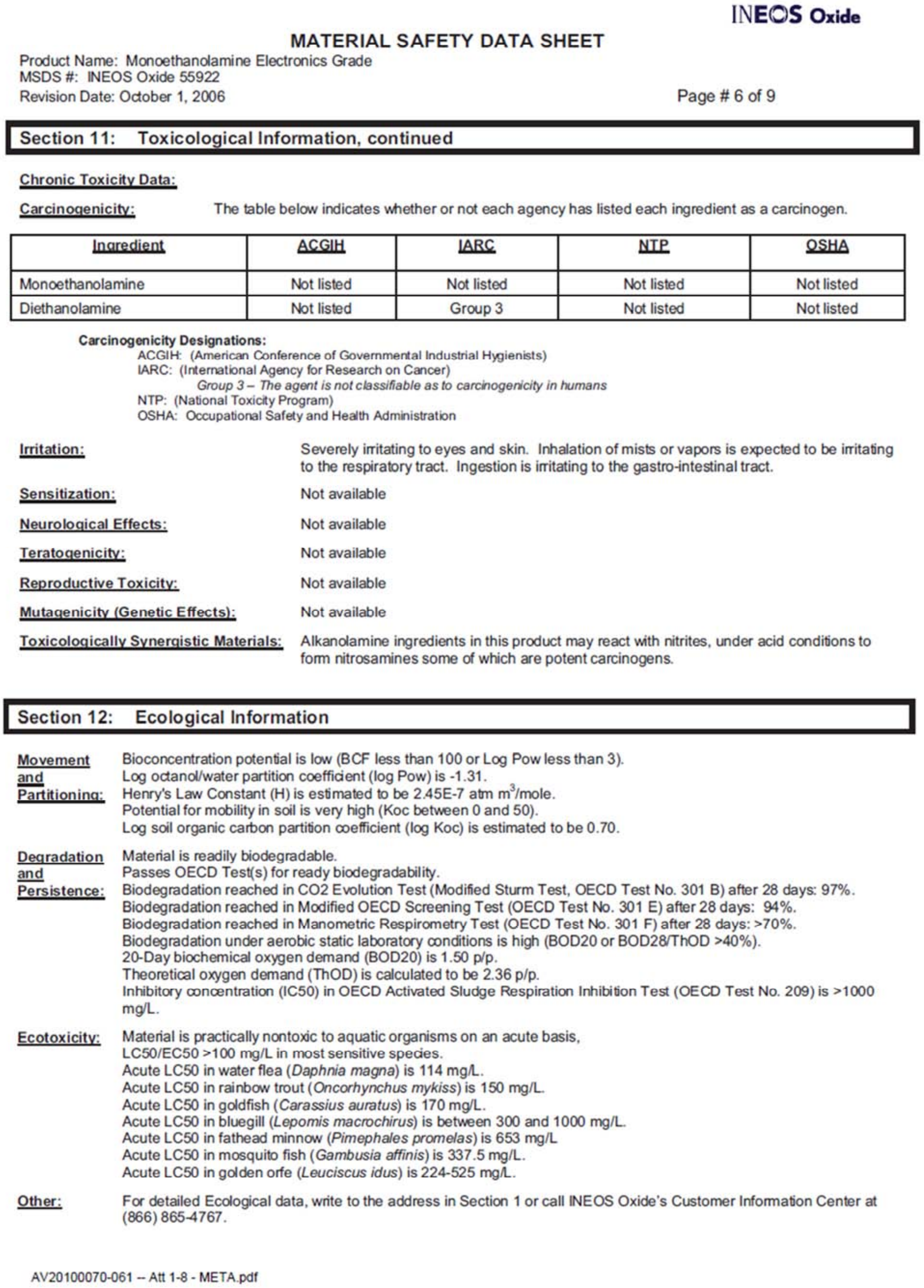
Section 10: Stability and Reactivity

Stability:	Stable under normal temperature and pressure.
Incompatible Materials and Conditions to Avoid:	This product should not be heated above 60°C in the presence of aluminum due to excessive corrosion and potential chemical reaction releasing flammable hydrogen gas. Avoid air, light and moisture. Avoid contact with nitrites, strong acids, chlorides, anhydrides, strong oxidizing agents, strong reducing agents, cellulose nitrate and halogenated hydrocarbons.
Hazardous Decomposition Products:	Decomposition products may include nitrogen oxides and ammonia.
Hazardous Polymerization:	Mixing with nitrite containing fluids may cause nitrosamine formation.
Other Reactivity Concerns:	Absorbs moisture and can react with carbon dioxide in the air to form salts. It is decomposed by light and slowly oxidized by air, turning yellow and then brown. This reaction is accelerated by heat and the presence of metals. May react violently with certain monomers, unsaturated compounds such as epoxides, vinyl chloride, vinyl acetate, acrylic monomers, acrolein and acrylonitrile. Corrosive to copper, brass, bronze and other copper alloys, zinc and galvanized iron.

Section 11: Toxicological Information

Acute Toxicity Data:

Ingredient	LD_{50} Oral (mg/kg)	LD_{50} Dermal (mg/kg)	LC_{50} Inhalation (mg/m^3, 4 hrs.)
Monoethanolamine	620 (guinea pig)	1 018 (rabbit)	>1 210 (mouse)
Diethanolamine	680 (rabbit)	8 180 (rabbit)	Not available



INEOS Oxide

MATERIAL SAFETY DATA SHEET

Product Name: Monoethanolamine Electronics Grade
MSDS #: INEOS Oxide 55922
Revision Date: October 1, 2006

Page # 6 of 9

Section 11: Toxicological Information, continued

Chronic Toxicity Data:

Carcinogenicity: The table below indicates whether or not each agency has listed each ingredient as a carcinogen.

Ingredient	ACGIH	IARC	NTP	OSHA
Monoethanolamine	Not listed	Not listed	Not listed	Not listed
Diethanolamine	Not listed	Group 3	Not listed	Not listed

Carcinogenicity Designations:
ACGIH: (American Conference of Governmental Industrial Hygienists)
IARC: (International Agency for Research on Cancer)
Group 3 – The agent is not classifiable as to carcinogenicity in humans
NTP: (National Toxicity Program)
OSHA: Occupational Safety and Health Administration

Irritation:	Severely irritating to eyes and skin. Inhalation of mists or vapors is expected to be irritating to the respiratory tract. Ingestion is irritating to the gastro-intestinal tract.
Sensitization:	Not available
Neurological Effects:	Not available
Teratogenicity:	Not available
Reproductive Toxicity:	Not available
Mutagenicity (Genetic Effects):	Not available
Toxicologically Synergistic Materials:	Alkanolamine ingredients in this product may react with nitrites, under acid conditions to form nitrosamines some of which are potent carcinogens.

Section 12: Ecological Information

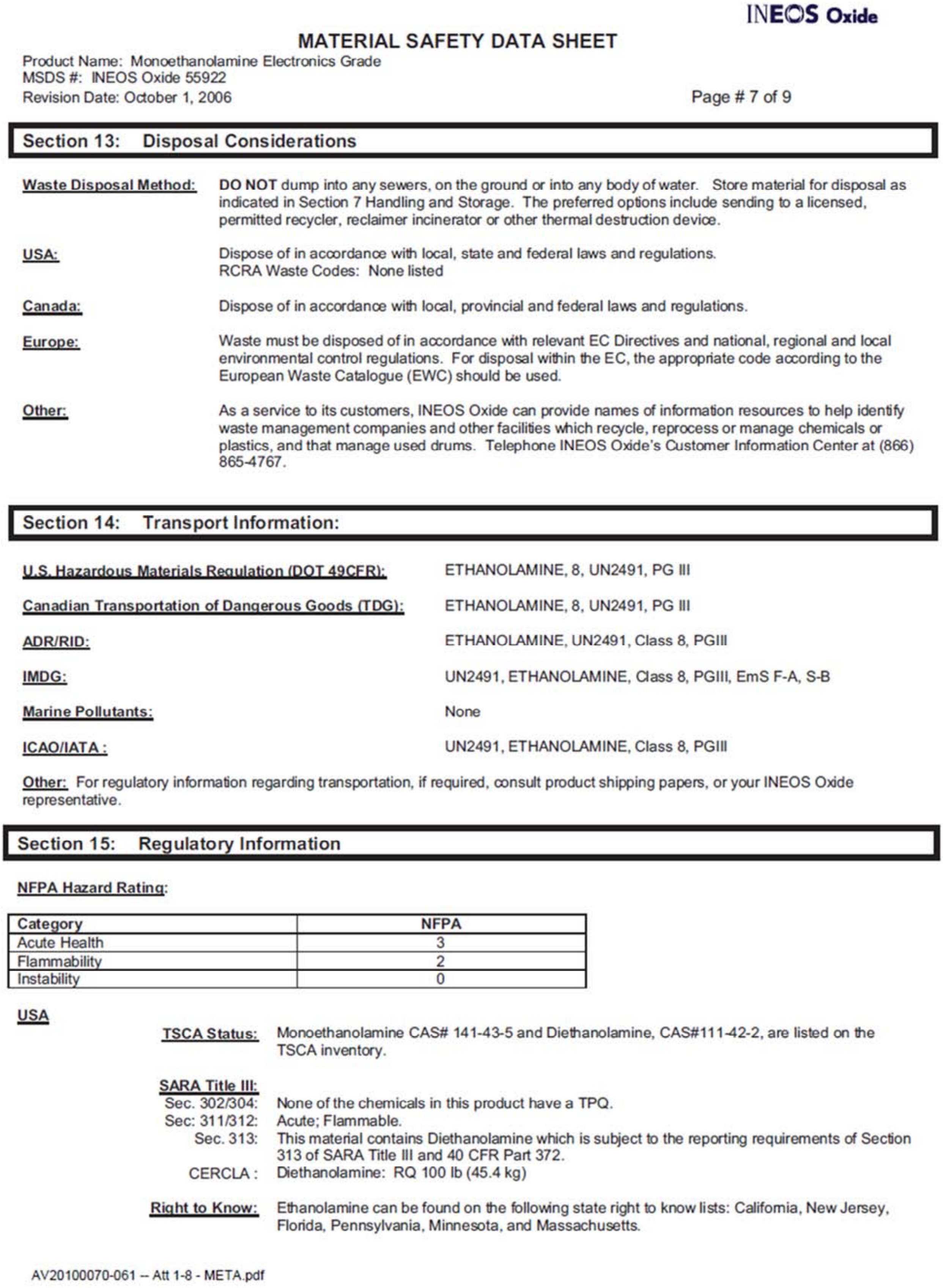
Movement and Partitioning:
Bioconcentration potential is low (BCF less than 100 or Log Pow less than 3).
Log octanol/water partition coefficient (log Pow) is -1.31.
Henry's Law Constant (H) is estimated to be 2.45E-7 atm m^3/mole.
Potential for mobility in soil is very high (Koc between 0 and 50).
Log soil organic carbon partition coefficient (log Koc) is estimated to be 0.70.

Degradation and Persistence:
Material is readily biodegradable.
Passes OECD Test(s) for ready biodegradability.
Biodegradation reached in CO2 Evolution Test (Modified Sturm Test, OECD Test No. 301 B) after 28 days: 97%.
Biodegradation reached in Modified OECD Screening Test (OECD Test No. 301 E) after 28 days: 94%.
Biodegradation reached in Manometric Respirometry Test (OECD Test No. 301 F) after 28 days: >70%.
Biodegradation under aerobic static laboratory conditions is high (BOD20 or BOD28/ThOD >40%).
20-Day biochemical oxygen demand (BOD20) is 1.50 p/p.
Theoretical oxygen demand (ThOD) is calculated to be 2.36 p/p.
Inhibitory concentration (IC50) in OECD Activated Sludge Respiration Inhibition Test (OECD Test No. 209) is >1000 mg/L.

Ecotoxicity:
Material is practically nontoxic to aquatic organisms on an acute basis,
LC50/EC50 >100 mg/L in most sensitive species.
Acute LC50 in water flea (*Daphnia magna*) is 114 mg/L.
Acute LC50 in rainbow trout (*Oncorhynchus mykiss*) is 150 mg/L.
Acute LC50 in goldfish (*Carassius auratus*) is 170 mg/L.
Acute LC50 in bluegill (*Lepomis macrochirus*) is between 300 and 1000 mg/L.
Acute LC50 in fathead minnow (*Pimephales promelas*) is 653 mg/L
Acute LC50 in mosquito fish (*Gambusia affinis*) is 337.5 mg/L.
Acute LC50 in golden orfe (*Leuciscus idus*) is 224-525 mg/L.

Other: For detailed Ecological data, write to the address in Section 1 or call INEOS Oxide's Customer Information Center at (866) 865-4767.

AV20100070-061 -- Att 1-8 - META.pdf



INEOS Oxide

MATERIAL SAFETY DATA SHEET

Product Name: Monoethanolamine Electronics Grade
MSDS #: INEOS Oxide 55922
Revision Date: October 1, 2006

Section 13: Disposal Considerations

Waste Disposal Method: **DO NOT** dump into any sewers, on the ground or into any body of water. Store material for disposal as indicated in Section 7 Handling and Storage. The preferred options include sending to a licensed, permitted recycler, reclaimer incinerator or other thermal destruction device.

USA: Dispose of in accordance with local, state and federal laws and regulations.
RCRA Waste Codes: None listed

Canada: Dispose of in accordance with local, provincial and federal laws and regulations.

Europe: Waste must be disposed of in accordance with relevant EC Directives and national, regional and local environmental control regulations. For disposal within the EC, the appropriate code according to the European Waste Catalogue (EWC) should be used.

Other: As a service to its customers, INEOS Oxide can provide names of information resources to help identify waste management companies and other facilities which recycle, reprocess or manage chemicals or plastics, and that manage used drums. Telephone INEOS Oxide's Customer Information Center at (866) 865-4767.

Section 14: Transport Information:

U.S. Hazardous Materials Regulation (DOT 49CFR):	ETHANOLAMINE, 8, UN2491, PG III
Canadian Transportation of Dangerous Goods (TDG):	ETHANOLAMINE, 8, UN2491, PG III
ADR/RID:	ETHANOLAMINE, UN2491, Class 8, PGIII
IMDG:	UN2491, ETHANOLAMINE, Class 8, PGIII, EmS F-A, S-B
Marine Pollutants:	None
ICAO/IATA :	UN2491, ETHANOLAMINE, Class 8, PGIII

Other: For regulatory information regarding transportation, if required, consult product shipping papers, or your INEOS Oxide representative.

Section 15: Regulatory Information

NFPA Hazard Rating:

Category	NFPA
Acute Health	3
Flammability	2
Instability	0

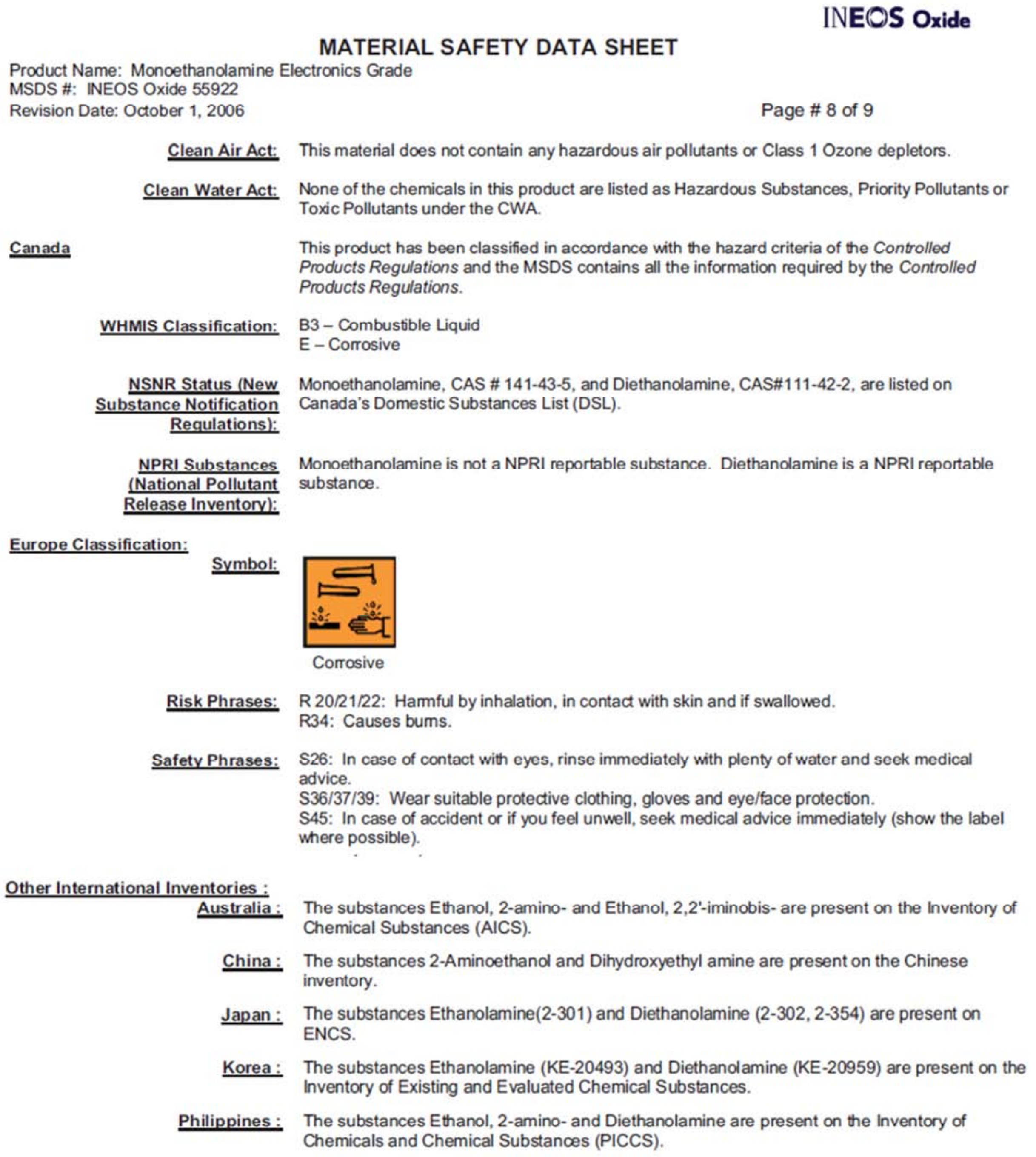
USA

TSCA Status: Monoethanolamine CAS# 141-43-5 and Diethanolamine, CAS#111-42-2, are listed on the TSCA inventory.

SARA Title III:
Sec. 302/304: None of the chemicals in this product have a TPQ.
Sec: 311/312: Acute; Flammable.
Sec. 313: This material contains Diethanolamine which is subject to the reporting requirements of Section 313 of SARA Title III and 40 CFR Part 372.
CERCLA : Diethanolamine: RQ 100 lb (45.4 kg)

Right to Know: Ethanolamine can be found on the following state right to know lists: California, New Jersey, Florida, Pennsylvania, Minnesota, and Massachusetts.

AV20100070-061 -- Att 1-8 - META.pdf



INEOS Oxide

MATERIAL SAFETY DATA SHEET

Product Name: Monoethanolamine Electronics Grade
MSDS #: INEOS Oxide 55922
Revision Date: October 1, 2006

Page # 8 of 9

Clean Air Act: This material does not contain any hazardous air pollutants or Class 1 Ozone depletors.

Clean Water Act: None of the chemicals in this product are listed as Hazardous Substances, Priority Pollutants or Toxic Pollutants under the CWA.

Canada

This product has been classified in accordance with the hazard criteria of the *Controlled Products Regulations* and the MSDS contains all the information required by the *Controlled Products Regulations*.

WHMIS Classification:
B3 – Combustible Liquid
E – Corrosive

NSNR Status (New Substance Notification Regulations): Monoethanolamine, CAS # 141-43-5, and Diethanolamine, CAS#111-42-2, are listed on Canada's Domestic Substances List (DSL).

NPRI Substances (National Pollutant Release Inventory): Monoethanolamine is not a NPRI reportable substance. Diethanolamine is a NPRI reportable substance.

Europe Classification:

Symbol:

Corrosive

Risk Phrases:
R 20/21/22: Harmful by inhalation, in contact with skin and if swallowed.
R34: Causes burns.

Safety Phrases:
S26: In case of contact with eyes, rinse immediately with plenty of water and seek medical advice.
S36/37/39: Wear suitable protective clothing, gloves and eye/face protection.
S45: In case of accident or if you feel unwell, seek medical advice immediately (show the label where possible).

Other International Inventories :

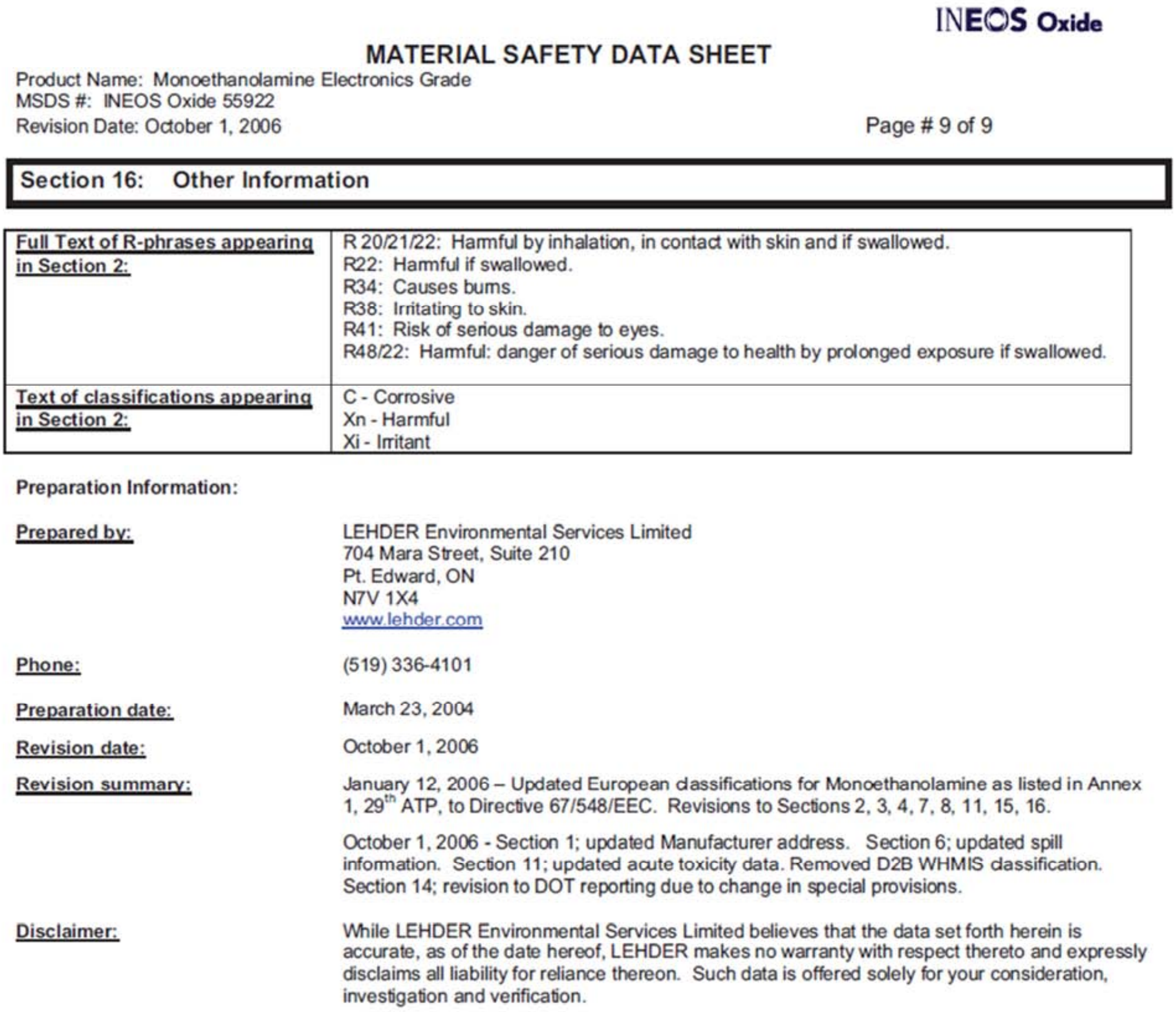
Australia : The substances Ethanol, 2-amino- and Ethanol, 2,2'-iminobis- are present on the Inventory of Chemical Substances (AICS).

China : The substances 2-Aminoethanol and Dihydroxyethyl amine are present on the Chinese inventory.

Japan : The substances Ethanolamine(2-301) and Diethanolamine (2-302, 2-354) are present on ENCS.

Korea : The substances Ethanolamine (KE-20493) and Diethanolamine (KE-20959) are present on the Inventory of Existing and Evaluated Chemical Substances.

Philippines : The substances Ethanol, 2-amino- and Diethanolamine are present on the Inventory of Chemicals and Chemical Substances (PICCS).



INEOS Oxide

MATERIAL SAFETY DATA SHEET

Product Name: Monoethanolamine Electronics Grade
MSDS #: INEOS Oxide 55922
Revision Date: October 1, 2006

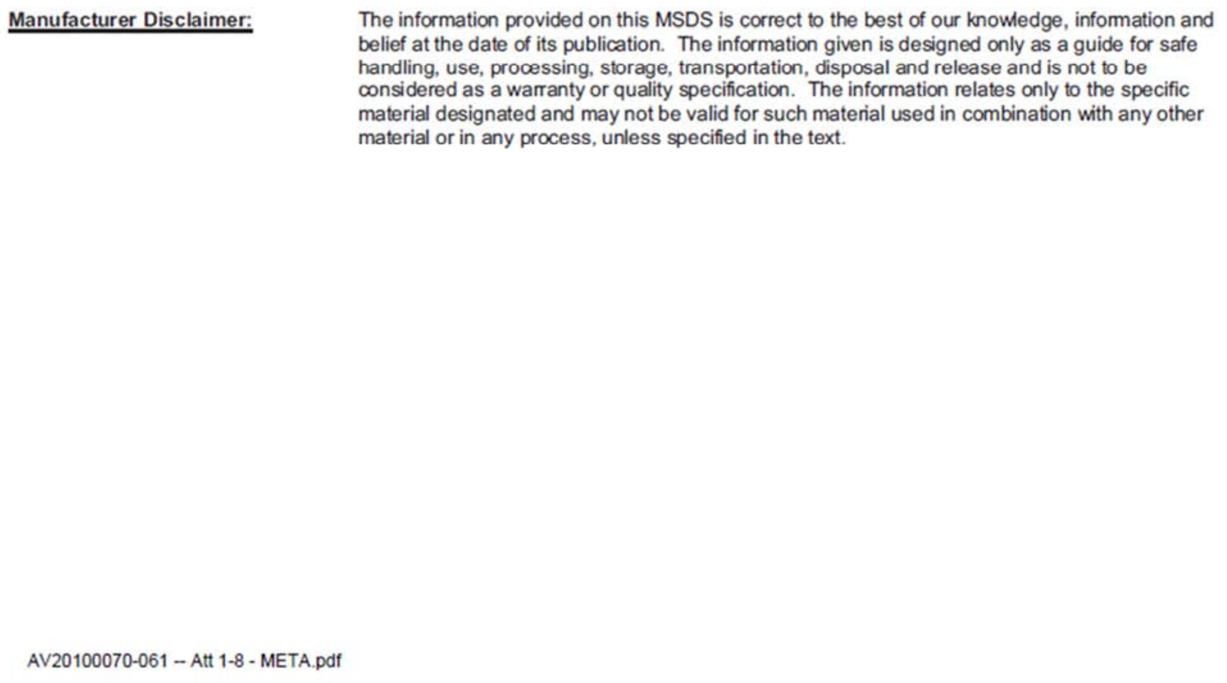
Page # 9 of 9

Section 16: Other Information

Full Text of R-phrases appearing in Section 2:	R 20/21/22: Harmful by inhalation, in contact with skin and if swallowed. R22: Harmful if swallowed. R34: Causes burns. R38: Irritating to skin. R41: Risk of serious damage to eyes. R48/22: Harmful: danger of serious damage to health by prolonged exposure if swallowed.
Text of classifications appearing in Section 2:	C - Corrosive Xn - Harmful Xi - Irritant

Preparation Information:

Prepared by:	LEHDER Environmental Services Limited 704 Mara Street, Suite 210 Pt. Edward, ON N7V 1X4 www.lehder.com
Phone:	(519) 336-4101
Preparation date:	March 23, 2004
Revision date:	October 1, 2006
Revision summary:	January 12, 2006 – Updated European classifications for Monoethanolamine as listed in Annex 1, 29th ATP, to Directive 67/548/EEC. Revisions to Sections 2, 3, 4, 7, 8, 11, 15, 16. October 1, 2006 - Section 1; updated Manufacturer address. Section 6; updated spill information. Section 11; updated acute toxicity data. Removed D2B WHMIS classification. Section 14; revision to DOT reporting due to change in special provisions.
Disclaimer:	While LEHDER Environmental Services Limited believes that the data set forth herein is accurate, as of the date hereof, LEHDER makes no warranty with respect thereto and expressly disclaims all liability for reliance thereon. Such data is offered solely for your consideration, investigation and verification.



Manufacturer Disclaimer:

The information provided on this MSDS is correct to the best of our knowledge, information and belief at the date of its publication. The information given is designed only as a guide for safe handling, use, processing, storage, transportation, disposal and release and is not to be considered as a warranty or quality specification. The information relates only to the specific material designated and may not be valid for such material used in combination with any other material or in any process, unless specified in the text.

AV20100070-061 -- Att 1-8 - META.pdf

HAZID results

No:	Description:	By:	Due:	Response:	By:	Date:
1	Determine what the limit value is for amine in stormwater according to Dutch regulations.	Fluor HSE	1-Jul	No set level exists for amine. IPPC Reference Document on Best Available Techniques in Common Waste Water and Waste Gas Treatment / Management Systems in the Chemical Sector gives limit for Chemical Oxygen Demand of 30-250mg/l. Final emission limits will be set in Environmental permit.	O. Schot	5-Jul-10
2	Investigate whether REACH Directive does require a risk assessment for amine.	Fluor HSE	1-Jul	Pre-registration for amine has been submitted. No risk assessment required.	O. Schot	5-Jul-10
3	Verify whether design is in accordance with Eon ATS 7 document.	Fluor Project	15-Jul	All client specs will be reviewed by SC and deviations will be advised to the JV. The first day of Meeting # 5 with the JV will be used to review all Fluor comments / deviations to the specifications. This action item will therefore automatically be closed out.	R. Naafs	5-Jul-10
4	Consider to merge smaller foundations into one larger piled foundation whenever practical.	Fluor CSA	1-Jul	Small foundations are placed on a common mat	A. Marfatia	22-Jul-10
5	Consider to apply measures to protect existing cooling water channels.	Fluor CSA	1-Jul	The current design location of the foundations are not closely located to the existing cooling water channels. This should be evaluated further during the EPC with the final geotech consultant.	L. Chiu	27-Jul-10
6	Provide the relevant parts of EIA (MER) to Fluor. Masterplan brandveiligheid MPP3.	JV Project	15-Jun	The EIA of MPP3 does not contain items that are relevant to the capture plant except the		

No:	Description:	By:	Due:	Response:	By:	Date:
				"masterplan brandveiligheid MPP3", which was provided.		
7	Discussions with authorities need to be opened to discuss impact of capture plant on operating permit, with respect to emissions.	JV Project	EPC	Discussions with the authorities have started but results regarding impact of emission on operation permit will take a long time. Resolution by JV to continue in EPC.		
8	Provide the Masterplan Brandveiligheid MPP3 to Fluor.	JV Project	15-Jun	Received by Fluor.	O. Schot	5-Jul-10
9	Consider to apply protection barriers at equipment close to the road.	Fluor CSA	01-Jul	We have included in the material takeoff (50 Guard Posts), some of which can be used for equipment barrier protection throughout the CO_2 plant.	H. Chang	27-Jul-10
10	Capture plant and MPP3 operators to be trained regarding emergency response planning.	JV operations	EPC			
11	Verify design, including design temperature, of system from compressor to stack to ensure depressurization of compressor can be done taking auto-refrigeration and mixing effects with flue gas into account.	Fluor Process	1-Jul	Depressurization done with manual valves in two stages from high to intermediate and intermediate to low. Fluor to confirm with compressor vendor to finalize settle-out pressure and depressurization strategy. P&IDs shall be clarified and all depressurization lines to be traced.	J. Gilmartin	5-Jul-10
12	Verify that overfilling of the 20% caustic tank due to static head from the 50% caustic tank can not occur.	Fluor process	1-Jul	Line from 50% tank to 20% tank will run up through pipe rack. Specified high point on P&IDs.	J. Gilmartin	5-Jul-10

No:	Description:	By:	Due:	Response:	By:	Date:
13	Measures should be taken to avoid galvanic corrosion when stainless steel meets carbon steel.	Fluor Piping material	1-Jul	Galv Corrosion occurs between Stainless(SS) and galv. carbon steel(CS). Where there is a connection between the two materials an INSULATING SET is provided to prevent conductivity. Where a SS pipe is being supported, galv. steel will not be used.	K. Vo	22-Jul-10
14	Ensure control room is upwind or cross wind of prevailing direction.	Fluor Piping	1-Jul	The prevailing wind is from the southwest which makes the control building location in the crosswind direction.	D. Muriilo	22-Jul-10
15	Verify Technical Specification on requirement for deluge system on lube oil system.	Fluor HSE	1-Jul	Technical specification does not require deluge system on lube oil system.	O. Schot	5-Jul-10
16	Ensure adequate escape routes from control room.	Fluor CSA	July 15th	Secondary escape route from Control room has been added on the evacuation plan and will be added in the final FEED drawings.	A. Marfatia	22-Jul-10
17	Consider portable CO_2 detection in the CO_2 handling and detection concept.	Fluor HSE	July 15th	Fixed CO_2 detection is foreseen for all buildings/enclosures that could be subject to elevated CO_2 concentrations. Outdoor acoustic leak detection is foreseen for leaks in the high pressure CO_2 system. Therefore no portable CO_2 detection is deemed required.	O. Schot	5-Jul-10
18	Fire water demand for capture plant needs to be determined. JV then confirms availability of firewater from MPP3 system.	Fluor HSE	15-Jul	Firewater demand is determined and documented in A4NA-0312-53-RP012, Fire Protection Concept	O. Schot	5-Jul-10

No:	Description:	By:	Due:	Response:	By:	Date:
19	Ensure sufficient flush medium available and handling capacity for absorber and stripper.	Fluor Process	1-Jul	We would fill the absorber/stripper system with demin water from the initial fill line. The amount of demin water needed is equal to the solvent system inventory (estimated at 923 m3). If this amount of water is not available, then water would need to be provided by water trucks.	J. Gilmartin	22-Jul-10
20	Ensure all power lines are identified prior to any excavation or piling work.	Construction	EPC			

Recovery of CO_2 from Flue Gases: Commercial Trends

Originally presented at the Canadian Society of Chemical Engineers annual meeting
October 4-6, 1999, Saskatoon, Saskatchewan, Canada

Authors: Dan G. Chapel (dan.chapel@fluor.com; 949-349-7530)
Carl L. Mariz (carl.mariz@fluor.com; 949-349-7530)
FluorDaniel
One Fluor Drive
Aliso Viejo CA, 92698

John Ernest (john.ernest@minimed.com; 818-576-4293)
Advanced Quality Services Inc
11024 Balboa Blvd. PMB154,
Granada Hills, CA 91344-5007

Recovery of CO_2 from Flue Gases: Commercial Trends

Originally presented at the Canadian Society of Chemical Engineers annual meeting
October 4-6, 1999, Saskatoon, Saskatchewan, Canada

Authors: Dan Chapel - Fluor Daniel Inc., Senior Vice President Technology; Oil, Gas & Power

John Ernest - Advanced Quality Services Inc., Validation Engineer

Carl Mariz - Fluor Daniel Inc., Director, Process Engineering; Oil, Gas & Power

1.0 ABSTRACT

Interest in recovery of carbon dioxide (CO_2) from flue gases is being propelled by multiple factors: the merchant CO_2 market, renewed interest in enhanced oil recovery (EOR), and the desire to reduce greenhouse gas emissions. This paper reviews the latest operating and capital cost data for the Fluor Daniel Econamine FG^{SM} process. It also recaps the key process design and operating issues for amine chemical solvent CO_2 recovery processes and summarizes the competitive processes for CO_2 recovery from flue gases.

The Econamine FG process has proven, reliable operations with both natural-gas and fuel oil-derived flue gases in plants ranging in size from 6 to 1000 tonne/day (te/d) CO_2 and in pilot plant operation with coal-derived flue gases. No flue gas CO_2 recovery process can compete in the merchant CO_2 market in areas where CO_2 is available in sufficient quantity from by-product sources such as fermentation, natural gas sweetening and ammonia and hydrogen manufacture, or from CO_2 wells. However, where by-product CO_2 is not available, the Econamine FG process has favorable costs for plants of 6 te/d and larger. The precise cost of a given plant will vary due to the plant size, the market price of CO_2, the flue gas source, and government regulations, Single-train plants of more than 4000 te/d are now possible, and economy of scale, together with the ongoing development of the Econamine FG process, have the potential of delivering CO_2 at a price approximately half that of a 1000 te/d plant.

2.0 INTRODUCTION

Carbon dioxide is used in the food industry in carbonated beverages, brewing, and flash drying. Its industrial uses include enhanced oil recovery (EOR), welding, chemical feedstock, inert gas, firefighting, and solvent extraction as a supercritical fluid. It is an essential ingredient in medical oxygen, where in low concentrations it acts as a breathing stimulant.

The largest potential market for CO_2 is in EOR. The major CO_2 sources that can be considered for the EOR market are:

- Natural sources
  - CO_2 wells
- Industrial byproducts
  - natural gas sweetening
  - synthesis gas production

Flue gases
  Fossil fuel-fired power plants
  Industrial furnaces
  Cement plants
  Engine exhausts
  Lime kiln exhausts

The most economical sources of CO_2 are CO_2 wells and natural gas sweetening or synthesis gas purification byproducts. The price of crude oil in 1999 ($12 to $16 per barrel) did not justify the recovery of CO_2 from flue gas sources for EOR purposes, however the recent crude price rise to $25 to $35 per barrel can justify flue gas derived CO_2 for EOR.

Flue gases have long been an important source of CO_2 for the merchant CO_2 market, especially in remote locations where by-product CO_2 sources are unavailable. In the simplest case, fuel is combusted to produce flue gas. CO_2 is then extracted from the flue gas using a dilute MEA solution, as sufficient heat is available from the combustion process to support the heat required for the CO_2 capture. This process is widely used but is wasteful of energy. Where a source of flue gas is available and lower energy consumption is desired, skid-mounted units employing the Econamine FG technology can be utilized. At present eleven such units are in operation to recover CO_2 for the food and beverage industries.

Enhanced oil recovery by CO_2 flooding has been described by Taber and others (1, 2). In many cases CO_2 flooding is the most cost-effective method for extracting the final amounts of recoverable oil from depleted fields. The Econamine FG process (known as the GAS/SPEC FT-1™ process prior to 1989 when Fluor Daniel purchased the technology from Dow Chemical) and related processes were used in several plants in West Texas to recover CO_2 from boiler flue gas for EOR purposes between 1982 and 1986. The plants were technical and commercial successes, and were shut down only because of the crude oil price collapse in 1986.

EOR by CO_2 flooding requires large quantities of CO_2 at low prices. The acceptable price of CO_2 depends critically on both the price of crude oil and the amount of CO_2 required to produce a barrel (Bbl) of oil. The market price of CO_2 varies widely, for instance $13/tonne delivered at pressure in Canada in 1997 and $55/tonne in the U.S. in 1986. In 1985, the North American EOR market was commercially successful with a US $30/Bbl oil price and a CO_2 price of approximately $1 to $2/10^{3}scf, which is equivalent to $19 to $38/tonne CO_2. The amount of CO_2 required for EOR is between 5 and 15x10^{3}scf/bbl oil recovered (3). At a makeup CO_2 rate of 6x10^{3}scf/bbl, suggested by Tontiwachwuthikul (4), and a CO_2 price of $19/tonne, CO_2 recovery from flue gas adds $6/bbl to the oil production cost.

No flue gas CO_2 recovery process can presently compete with by-product CO_2 where it is available in sufficient quantity. However, where by-product CO_2 is not available, or if the merchant price of crude oil remains at its present levels, the Econamine FG process and other similar competing processes will again become economically viable.

This paper provides the basic technical and economic guidelines for evaluating various flue gases as CO_2 sources and indicates the issues involved in an economic analysis. It also reviews important design and cost issues for flue gas recovery and the applicable processes.

3.0 ECONAMINE FG PROCESS

The Econamine FG process is today's premier commercially proven process for the recovery of CO_2 from flue gases, using an inhibited 30 wt.% MEA solution. These features allow the widespread use of carbon steel and give the process the lowest stripper reboiler steam demand of all of the well-established commercial processes. It can recover 85-95% of the CO_2 in the flue gas and produces a 99.95+% pure CO_2 product (dry basis). The inhibitor not only tolerates oxygen and NO_X-containing flue gas, but also requires oxygen to maintain its activity. The process can be used with SO_X-containing flue gas after SO_2 scrubbing. The additional SO_2 scrubbing returns an environmental benefit. Coal-derived flue gases usually have better economics as the process is more efficient with the higher CO_2 content.

The Econamine FG process is a typical, regenerable alkanolamine process. The standard process flow diagram is presented as Figure 1. The process is described by Mariz (3).

The Econamine FG process is not applicable to reducing gas streams, for instance streams containing large amounts of carbon monoxide and hydrogen, on streams that contain more than 1 ppm hydrogen sulfide, or on streams that have less than 1 vol% oxygen. The process is applicable to pressurized gas streams, but the full commercial advantage of this process lies in atmospheric pressure applications.

4.0 COMMERCIAL HISTORY OF THE ECONAMINE FG PROCESS

Dow Chemical and Union Carbide developed inhibited 30 wt.% MEA processes for recovery of CO_2 from flue gases in the 1970's and 1980's primarily for the EOR market. When the price of crude oil dropped dramatically in 1986, the EOR market disappeared and these licensors became less active. Dow Chemical sold its GAS/SPEC FT-1™ process to Fluor Daniel, Inc. in 1989. Fluor Daniel renamed the technology Econamine FG℠. The licensing history of the Econamine FG and predecessor FT-1 processes is shown in Table 1. Twenty-one commercial plants have been built, including ten large plants (60 te/d and larger) and eleven skid-mounted plants. Eight of the large plants are still operating. Fluor Daniel licensed three of these commercial plants, a 320 te/d CO_2 plant in Bellingham, Massachusetts for Northeast Energy Associates; a 150 te/d plant for Sumitomo Chemical in Chiba, Japan; and a 90 te/d plant for Prosint Produtos Sintéticos in Rio de Janeiro, Brazil. Fluor Daniel also constructed the Bellingham plant. Prior to this, Dow had licensed six commercial plants ranging from 6 to 1000 te/d. The eleven facilities employing skid-mounted Econamine FG process units have capacities ranging from 6 to 50 te/d and are manufactured by the Wittemann Company of Palm Coast, Florida and Union Engineering of Fredericia, Denmark.

All of the large commercial plants use flue gas resulting from the combustion of natural gas, except the Sumitomo plant which fires a variety of fuels including heavy fuel oil. The Econamine FG process has also been demonstrated with coal-derived flue gas in three pilot plants, a 4.5 te/d plant in Yokosuka, Japan for Tokyo Electric Power Co, a 2 te/d unit at the Sundance Generating Plant in Alberta (5, 6), and a 3 te/d unit at the Boundary Dam Power Plant in Saskatchewan (9, 10, 11). Fluor Daniel has also licensed a natural gas-fired 2 te/d pilot plant in Osaka, Japan for KEPCO (12, 13).

Of the ten large commercial plants, the 1000 te/d Carbon Dioxide Technology Corp. plant (26) serving the Garza field south of Lubbock, Texas, and the 90 te/d N-Ren Southwest plant (14) provided CO_2 for EOR; two others feed urea plants; and six serve the food and beverage industry. All the skid-mounted units serve the food and beverage industry.

5.0 GENERAL DESIGN ISSUES, RECOVERY OF CO_2 FROM FLUE GAS

The practical recovery of CO_2 from flue gas is sufficiently different from other gas treating applications such as natural gas and refinery gas sweetening and purification of ammonia and hydrogen plant synthesis gas that it requires its own specialized processes. Candidate processes must be active at low CO_2 partial pressure and tolerate oxygen and NO_X. Flue gases from coal-fired boilers also contain SO_X, soot, and fly ash, which must be dealt with. Design issues relevant to CO_2 recovery processes in general and to amine processes in particular are discussed below.

5.1 Low CO_2 Partial Pressure. Flue gases and engine exhausts have very low CO_2 partial pressures because they are typically available at or near atmospheric pressure with CO_2 concentrations of typically 3 to 13 vol%. Many acid gas treatment processes operate at thousands of kPag and cannot remove CO_2 to less than about 10 kPa partial pressure. These high-pressure processes can be used by compression to the normal operating pressure followed by recovery of a portion of the compression energy in expanders; however the overall energy consumption invariably results in unattractive economics. The only commercial absorbents active enough for recovery of dilute CO_2 from atmospheric pressure gas are monoethanolamine (MEA) and other primary amines including the newly developed hindered amines.

5.2 Regeneration Energy. Absorption processes that are effective at low pressure are those with higher reaction energies that require the most regeneration energy. The design challenges are to minimize regeneration energy by selecting a solvent with a relatively low reaction energy, and to use low-value heat sources to provide this energy.

5.3 Oxygen . Oxygen can cause corrosion and solvent degradation problems. Carbon steel is desirable from a cost standpoint, but it can rapidly corrode in the presence of oxygen, especially in hydrogen sulfide-free CO_2 recovery systems. Also, uninhibited alkanolamines such as diethanolamine (DEA) and MEA can be excessively degraded in an oxygen environment. The Econamine FG process employs an inhibitor to both passivate the metal and inhibit amine degradation. Flue gas streams with at least 1.5 vol% oxygen are preferred to maintain inhibitor activity.

Alternate approaches include the use of expensive alloys or the removal of all the oxygen with a combination of a near-stoichiometric burner waste heat boiler and a catalytic reactor. Mitchell Energy Corp.'s Bridgeport, Texas plant used the latter approach in the 1980's (11). Another alternative, continuous addition of an oxygen scavenger to the solvent system, has not been commercially demonstrated.

5.4 SO_X. Flue gases can contain significant concentrations of SO_X unless natural gas or very low sulfur fuels are being fired. SO_X reacts irreversibly with MEA to produce non-reclaimable corrosive salts that are very detrimental to plant operation. For MEA-based processes, it is less expensive to install a SO_X scrubber than to accept the solvent losses when the flue gas contains more than 10 ppmv SO_2. Coal fired boilers produce the highest concentrations of SO_X, often 300 to 5000 ppmv before flue gas desulfurization (FGD), but even oil firing can produce 100 ppmv SO_X (9, p. 475). The limestone or wet lime FGD systems in large power boilers today achieve SO_X reductions in the 90-95% range. Therefore, even the flue gas from a low-sulfur liquid or solid fuel, or from a limestone FGD system needs further SO_2 removal. The 10 ppmv SO_2 requirement is met by using the active alkali metal neutralizing agents, caustic soda or soda ash, in a relatively inexpensive spray scrubber (10).

Sulfur trioxide, SO_3, presents additional problems. SO_3, like SO_2, leads to solvent losses due to the formation of non-reclaimable heat stable salts, but it also forms a corrosive H_2SO_4 aerosol in wet scrubbers. Furthermore, less than one-third of the SO_3 may be removed by the SO_2 scrubbing system unless a special mist eliminator is used. Therefore, most of the remaining SO_3 will form heat stable salts in the absorber. The fraction of SO_X which forms SO_3 is a function of combustion, fuel composition, and flue gas processing factors, but SO_3 typically accounts for a few percent of the total sulfur. Minimization of SO_3 is a boiler design issue preferably handled upstream of the SO_2 scrubber.

5.5 Fly Ash. Fly ash in the CO_2 absorption solvent may cause foaming in the absorber and stripper, scaling and plugging of equipment, erosion, crevice corrosion, and increased solvent loss through chemical degradation and physical association with removed sludge. Furthermore, fly ash may create significant difficulties in the upstream SO_2 scrubber.

If the fuel produces fly ash, the particulates should be removed upstream of the SO_2 scrubber. A coal-fired boiler flue gas has a typical loading of 2 grains of fly ash per dry standard cubic foot (gr/dscf). This amounts to about 20 te/d of fly ash per 1000 te/d CO_2. It is recommended that the loading be reduced to 0.006 gr/dscf (24), requiring a removal efficiency of about 99.7 weight percent.

5.6 Soot. Soot presents a special problem in the absorber. Soot derived from firing gas or very light fuel oils does not cause problems and passes harmlessly through the absorber tower. However, soot derived from heavy fuel oil stabilizes an amine mist above the CO_2 absorption zone that is not captured in the water wash zone. In this instance, a special mist eliminator must be employed in the absorber to capture the micron-sized MEA mist particles.

5.7 NO_X. Thus far, oxides of nitrogen have never created problems in Econamine FG units, however they have led to corrosion of steel and amine degradation in other plants. The Bridgeport CO_2 recovery plant did not eliminate some corrosion problems until the NO_X was reduced to less than 1 ppmv in the absorber feed. NO_X is best controlled though control of the peak flame temperature in the boilers. Also, any boiler NO_X reduction (SCR) equipment will benefit the absorption process. The chief culprit in NO_X is NO_2, which reacts to form nitric acid in the amine solvent and ultimately heat stable salts. However, typically only 10% of the NO_X is NO_2 and only a fraction of the NO_2 gas is absorbed in the solvent. NO_X can be a problem in the CO_2 product if it is to be used in the food and beverage industry and steps must be taken in the liquefaction unit for its removal.

5.8 High Flue Gas Temperature. Hot flue gases can cause solvent degradation and decrease absorber efficiency. The flue gas must be cooled to a water dew point of 50 ^oC (122 ^oF) prior to entering the absorber. This is accomplished either in the SO_2 scrubber, if present, or in a direct contact water cooler.

6.0 OTHER PROCESSES FOR CO_2 RECOVERY FROM FLUE GAS

This section surveys other processes that have been proposed for CO_2 recovery from flue gas. CO_2 can also be recovered by caustic soda, ammonia, and various other weak acid-alkali salts. Additional processes are available for applications involving EOR where CO_2 is to be produced rather than removed. Pure oxygen boilers (15) and oxygen-blown gasifier derivatives can produce a nitrogen-free flue gas without CO_2 separation. Such processes are outside the scope of this article.

6.1 Higher Pressure Absorption Processes Absorption at atmospheric pressure severely handicaps processes that rely on higher pressures. This limitation is foremost on physical solvents, such as the Selexol process, where CO_2 recovery is directly proportional to the partial pressure of the CO_2 in the feed gas. This limitation also applies to a lesser degree to the less reactive chemical solvents, including methyldiethanolamine (MDEA), diethanolamine (DEA), diisopropanolamine (DIPA), triethanolamine (TEA), and hot potassium carbonate, and also to molecular sieves, membranes, and cryogenic separation (Ryan-Holmes process) (16). For these processes, the driving force for CO_2 absorption and the cyclic loading between absorption and desorption will be insufficient for economical CO_2 recovery. Alternately, compression costs to put the flue gas into the operational range of these processes are prohibitive.

6.2 Hot Potassium Carbonate. Hot potassium carbonate (HPC) or "Hot Pot" is effectively used in many ammonia, hydrogen, ethylene oxide and natural gas plants. To improve CO_2 absorption mass transfer and to inhibit corrosion, proprietary activators and inhibitors are added. These systems are known as "activated hot potassium carbonate" (AHPC) systems. The most widely licensed of these are the Benfield process, with over 675 units worldwide licensed by UOP (17, p. 93), and the Catacarb process, with over 100 units licensed as of 1992 by Eickmeyer & Associates (18, p. 91). Other commercial processes are the Exxon Flexsorb HP process, which uses a hindered amine activator, and Giammarco-Vetrocoke's new process, which uses an organic activator. The processes are designed for bulk CO_2 removal from high pressure streams but also produce CO_2 of high purity. The VLE curves for the partial pressure of CO_2 vs. solution loading of CO_2 for various activators (9, p.370), including the original Benfield activator, DEA, show that the full capacity of the "hot pot" family of processes requires a feed CO_2 partial pressure of about 700 kPa. The Benfield and Catacarb processes are commercially offered for applications at a minimum CO_2 partial pressure of 210 to 345 kPag. The Benfield process was once proposed for treating flue gases (19). Even so, that paper concluded that the optimum operating pressure for the process would be 700 kPag.

6.3 Membranes. Membranes suffer from both the cost of compression and heat exchange to obtain a high pressure feed and in that they produce an impure CO_2 product. For instance, Separex membrane systems are currently offered by UOP for feed pressures starting at 2750 kPag (17, p. 124). There are currently no commercial applications of membranes for recovery of CO_2 from flue gases, though they have been used in large EOR projects to recycle CO_2 from the associated gas. The presence of fly ash and the effects of trace components such as SO_X, NO_X, HCl, and HF are also potential complications. The most likely applications for membranes are in small skid-mounted plants where an impure CO_2 product is acceptable and offshore applications that can take advantage of their compact size and low weight. Membrane-amine and membrane-cryogenic separation hybrids have been considered for special applications such as offshore locations where again their compact size and low weight are beneficial (6, p. 1276; 18, p.104)

6.4 Other MEA. MEA has a long history of commercial CO_2 recovery with various feeds including flue gases. Uninhibited MEA is generally limited by corrosion problems to about 15-20 wt% MEA. Furthermore, many applications rely on the presence of hydrogen sulfide in the feed to provide a passivating layer of iron sulfide to reduce carbonic acid corrosion. This protection is absent in flue gas systems where H_2S is not present. The low MEA concentration raises the reboiler duty substantially. For example, the reboiler duty increases 20% when the MEA concentration decreases from 30 to 15 wt%. The required pump power increases even more. Since the reboiler heat duty is the most important key to operating costs, this is a significant handicap. However, Kerr-McGee/ABB Lummus Global has licensed four units that use 15-20 wt% MEA to recover CO_2 from coal-fired flue gas. The plant capacities vary between 180 and 720 te/day (11, p. 94).

Some corrosion inhibitors in conjunction with a quantitative oxygen and NO_X removal system allow the MEA concentration to be raised to 25-30 weight percent. The Bridgeport CO_2 plant referred to above, which used an early Amine Guard™ process, is one example. There are now over 500 Amine Guard units worldwide including the UCARSOL™ family of formulated amines, licensed by UOP. Ucarsol plants using atmospheric pressure absorption are offered; however, not on oxygen-containing gas streams.

6.5 Sterically Hindered Amines. Kansai Electric Power Company (KEPCO) and Mitsubishi Heavy Industries have been developing a proprietary hindered amine called KS-1 as an MEA replacement for flue gas applications. KS-1 has a lower circulation rate (due to its higher lean to rich CO_2 loading differential), lower regeneration temperature (110 °C), and 10-15% lower heat of reaction with CO_2. It is non-corrosive to carbon steel (less than 5 mils/yr) at 130 °C in the presence of oxygen (12, 13, 20). A second sterically hindered amine, AMP, (2-amino-1-methyl-1-propanol) may have similar properties to KEPCO's KS-1 (21). The first commercial plant using KS-1 for Petronas Fertilizer Kedah Sdn Bhd's fertilizer plant in Gurun Kedah, Malaysia is now in operation.

7.0 CAPITAL AND OPERATING COSTS OF THE ECONAMINE FG PROCESS

This section discusses the capital and operating costs for Econamine FG plants using coal-fired (high CO_2, SO_X-containing) and natural gas-fired (low CO_2, SO_X-free) flue gases. The information in this section reviews the important factors in all atmospheric pressure CO_2 recovery applications of the Econamine FG technology. This information can be used for a preliminary evaluation of Econamine FG process applicability, but the licensor should always be consulted due to the major variations in site-specific costs. Plants smaller than about 70 te/d can be purchased from licensed packaged-unit vendors as the Wittemann Co. of Palm Coast, Florida, or Union Engineering of Fredericia, Denmark.

7.1 Economic Studies. Mariz reported in 1998 on the main cost factors, potential savings, and capital and operating costs for a 1000 te/d CO_2 recovery plant over a range of flue gas CO_2 contents from 3 vol% for natural-gas-fired turbines to 13 vol% for coal-fired boilers (3). The costs from that study are the basis of this economic analysis.

A 1998 study by Tontiwachwuthikul, et. al., reexamines CO_2 recovery from coal-fired flue gas in large-scale plants for the purpose of EOR (4). An earlier and extensive study of CO_2 recovery from coal-fired power plant flue gases was commissioned by EPRI in 1991 (25). This report considered the ultimate disposal of CO_2 in the ocean as an alternative to commercial use.

Sander et. al. in 1992 studied the recovery of 5800 te/d of CO_2 (equivalent to 90% of all the CO_2 in the flue gas) from a 800 MW combined cycle power plant (25). Gosar, et. al studied CO_2 recovery from natural-gas fired gas turbine flue gases in 1986 (26).

Both Suda, et. al. (14, 15, 16) and Tontiwachwuthikul considered the integration of the utility requirements for both the amine process and the downstream steam turbine driven CO_2 compressor with steam and waste heat sources from the power generation unit to reduce the operating and capital costs.

7.2 Promising Areas for Cost Reduction. The most promising areas for operating cost savings according to Mariz (3) are (1) reduction of steam consumption, (2) integration of power generation with the stripper reboiler, and (3) reduction of the absorber packing pressure drop. The most promising areas for capital cost savings are (1) absorber and flue gas cooling vessel materials, (2) absorber vessel size, (3) economies of scale, and (4) improved oxidation inhibitors. Improved technology in the form of new chemical solvents and more sophisticated process designs will reduce both steam consumption and solvent circulation rate. Reduced steam consumption will also help reduce capital costs as the stripper reboiler and overhead condenser will become smaller.

7.3 Plant Size. Commissioned Econamine FG plants range in size from 90 te/d to 1000 te/d CO_2 recovered. Larger single-train plants are feasible, limited by the maximum diameter of the absorber and DCC vessels. Using the equations below, the largest economic single train plant is approximately 4600 te/d from coal-fired flue gas or 2400 te/d from natural gas, based on a maximum column diameter of 12.8 m. This equation is applicable to conventional circular absorber vessels. For very large plants, larger vessels may be economic, and the vessel may be more cost effective if constructed with a rectangular cross section.

Further economy of scale may be possible in other areas of the plant; e.g. one solvent stripper can serve more than one absorber. The stripper size is limited by being able to accommodate no more than four reboilers. At the other end of the scale, plants employing skid-mounted units are built as small as 6 te/d.

7.4 Equipment Size. The absorber diameter can be approximated from equation 1:

$$\text{absorber diameter, m} = A * \text{sqrt}[(\text{te/d}) / (\%\ CO_2)] \quad (1)$$

where A= 0.56 at 3% CO_2 to 0.62 for 13% CO_2

For all equations:

te/d = tonne/day CO_2 recovered
% CO_2 = volume% CO_2 in flue gas before cooling, wet basis
m = meters

The flue gas cooling Direct Contact Cooler (DCC) vessel, if present, will be slightly larger in diameter than the absorber. If the flue gas is from a coal-fired boiler, the SO_X scrubber will perform the function of the DCC and a separate DCC vessel will not be required.

The diameter of a stripper serving a single absorber can be estimated with equation 2. Note that one stripper may serve multiple absorbers.

stripper diameter, m = 0.13 * sqrt(te/d) (2)

A single stripper may include up to four reboilers as noted above.

7.5 Steam Consumption . The consumption of steam is the most important component of the operating cost and is strongly influenced by the process design. Some important parameters are the lean-rich cross-exchanger approach temperature, the stripper reflux rate, and the stripper pressure. Well-designed Econamine FG plants have been proven to use less than 4.2 GJ/t CO_2 (3.6x10^6 Btu/ton), which is equivalent to 115 kg of 345 kPag saturated steam/m^3 solvent (0.95 lb/gal). This relatively low rate reflects the high solvent concentration and high net CO_2 loading employed in the Econamine FG process. Typical steam levels are 345 kPag (50 psig) for the reboiler and 520 kPag (75 psig) for the reclaimer. Minimum steam levels are 210 kPag (30 psig) for the reboiler and 310 kPag (45 psig) for the reclaimer.

A split flow solvent process shows promise in providing significant further energy savings. The precise savings obtainable with split flow schemes are a function of the inlet CO_2 concentration with the greatest savings occurring at CO_2 levels obtained from coal-fired boilers. Additional savings from the integration of the CO_2 recovery plant steam system with a power generation cycle are discussed at the end of this section.

7.6 Solvent Loading and Circulation Rate. Unhindered primary amine solvents have a maximum theoretical loading of 0.5 mole CO_2 per mole of amine. The Econamine FG process removes a net of approximately one-quarter mol CO_2 per mol solvent between lean and rich loadings, which is somewhat dependent on the CO_2 content of the flue gas. The rich solvent flow can be determined approximately using equation 3:

rich solvent circulation rate, m^3/h = 0.71 * (te/d) (3)

7.7 Power Consumption . The chief power consumer over the entire range of CO_2 concentrations is the flue gas blower. The flue gas blower power, DCC circulation pump power, and absorber wash water pump power per te/d CO_2 are all inversely proportional to the CO_2 concentration in the flue gas. The plant electrical consumption for a water-cooled plant can be estimated using equation 4.

kW = (0.4 + 16.4/(% CO_2))*(te/d) (4)

For example, the electrical consumption is 2.45 kW/(te/d), or 60 kWh per tonne CO_2, for a flue gas containing 8 vol% CO_2.

7.8 Solvent Consumption . Solvent losses result from degradation, vapor losses, leaks, and spills. The major factors in amine degradation are temperature and solution maintenance. Thermal degradation is limited by minimizing skin and bulk temperatures and residence times in the reboiler, stripper and reclaimer. In addition, the Econamine FG process adds water to the reclaimer to reduce the reclaimer operating temperature, reducing degradation during reclaiming and reducing the required steam pressure to operate the reclaimer. The carbon beds remove degradation products to prevent the acceleration of the solvent degradation processes. Sufficient oxygenation of the solution is needed to maintain the effectiveness of the corrosion inhibitor that controls corrosion-related amine solvent losses. Solvent losses by vaporization and entrainment are controlled with well-designed water wash sections and mist eliminators. No amine is lost with the product CO_2.

Total solvent losses average 1.6 kg solvent per tonne CO_2 for gas-fired flue gas feed streams in a well-managed plant.

If SO_X is present in the flue gas, an additional two moles of MEA are lost per mol SO_X entering the absorber. SO_X scrubbers are designed to reduce SO_2 to below 10 ppmv but most of the SO_3 present will pass through the scrubber and react with the MEA in the same way as the residual SO_2. The additional solvent loss due to non-reclaimable sulfur salts is estimated by equation 5. This loss is approximately twice the stoichiometric amount because of the additional degradation in the reclaimer attributable to these salts.

$$\text{kg MEA/te } CO_2 = 0.5 * (\text{ppmv } SO_X \text{ entering the absorber/\% } CO_2) \tag{5}$$

7.9 Additive Consumption. The cost of the inhibitor additive is approximately 20% of the cost of the makeup amine.

7.10 Process Water Consumption. A CO_2 recovery plant can be a net producer or a net consumer of water, depending on the water content of the flue gas. The major process loss is the humidification loss in the absorber wash water system. The Direct Contact Cooler usually produces water, which can be filtered to remove solids and used within the plant. In most cases the temperature of the absorber vent can be adjusted in order to maintain a water balance in the solvent loop.

7.11 Process Cooling. Air cooling is a feasible option where cooling water is not available. Air cooling is, however, more expensive.

The circulating cooling water rate is approximately 4.6 m^3/h cooling water per te/d CO_2 for an 8 vol% CO_2 flue gas.

7.12 Carbon Consumption. The activated carbon consumption will typically average 0.075 kg activated carbon per tonne CO_2. This rate is based on changing out the carbon every three months.

7.13 Waste Disposal. A CO_2 recovery plant produces three waste streams: reclaimer waste, spent carbon from the activated carbon filters, and filter elements from the slip-stream solvent filters at the carbon bed. The amount of amine waste generated in the process is a function of flue gas composition and plant operating conditions. The reclaimer waste will be about 0.003 m^3/tonne (0.8 gal per tonne) of CO_2. This waste can be disposed of by incineration. The spent carbon can be returned to the vendor. The filter elements may contain flue gas particulate, though an upstream FGD scrubber will collect most of the fly ash and soot which otherwise would be filtered from the water.

Table 2
Operating Cost for a Standard Design, 1000 tonne/day, Coal-Fired Econamine FG CO_2 Recovery Plant

	US $/tonne	
Steam, 345 kPag sat.($4.40/tonne=$2.20/MMBtu)	7.90	
Electric Power ($0.07/kWh)	2.77	
Cooling Water, 11°C rise ($0.0045/$m^3$=$0.017/1000 gal)	0.34	
TOTAL UTILITIES		**11.01**
Make-up Solvent ($0.58/lb) & Inhibitor @ 20% of Make-up Solvent cost	2.40	
Caustic (soda ash @ $510/ton)	0.07	
Activated Carbon ($1.75/lb)	0.19	
TOTAL CHEMICALS		**2.66**
Labor, Maintenance (Labor & Materials), Taxes, & Insurance		
OPERATIONS AND MAINTENANCE		**3.31**
SUBTOTAL EXCLUDING SO_2 REMOVAL		**16.98**
SO_2 REMOVAL @ 250 ppmv SO_2		**1.72**
TOTAL OPERATING COST (US $/tonne CO_2)		**18.70**
Cost per 1000 scf CO_2		$0.98
Cost per 1000 Nm^3 CO_2		$36.69

This is the high end of the operating cost range because of the standard plant design. It does not take credit for any savings due to integration with a steam power system or process optimization, which are currently active areas of research and development.

The above estimate assumes a flue gas with 13 percent CO_2 and 250 ppmv SO_2. The corresponding operating cost for a 3 percent CO_2, sulfur-free flue gas plant is $25.14/tonne. This estimate assumes a net zero water consumption. It excludes reclaimer and solid waste disposal costs, SO_2 scrubber waste disposal costs, downstream CO_2 drying and compression costs, and pipeline costs. These costs can be significant. Disposal of scrubber waste has not been an important cost to date so there is no relevant historical cost basis. The cost to add and operate drying and 13.8 MPag (2000 psig) CO_2 compression facilities would increase the cost of the CO_2 by approximately US $8/tonne.

7.18 Capital Cost. Capital costs for flue gas plants using the Econamine FG process were reported by Mariz (3) for a 1000 TPD plant for a variety of investment assumptions. For the particular case of a 20 year project life and 15% Return On Investment, U.S. Gulf Coast location, and fourth quarter 1997 pricing, the capital recovery cost was U.S. $10.83/(te/d) for a coal-fired flue gas (13 vol% CO_2) and $17.85/(te/d) for a gas-turbine flue gas (3 vol% CO_2). This price includes an allowance for initial solvent and adsorbents fill, spare parts, startup costs, owners costs, licensing fee, and similar charges. Income taxes were excluded. The capital cost includes the CO_2 recovery equipment and SO_2 scrubber (if required), but not the CO_2 drying and compression equipment and pipelines.

For larger plants, Mariz suggests a 40% to 50% capital cost increase for doubling the plant size, equivalent to a power law exponent between 0.5 and 0.6.

7.19 SO_2 Scrubber Capital Cost. When SO_2 is present in the flue gas, the cost of the SO_2 removal equipment, the cost of caustic soda or soda ash, and the cost of waste disposal from the SO_2 scrubber must be added to the capital and operating costs. However, the extra capital cost is offset by the elimination of the flue gas cooling vessel (named the Direct Contact Cooler vessel (DCC) in the Econamine FG process) and the necessity of pH control in the scrubbing system allows the use of a carbon steel scrubber instead of a stainless steel DCC. In fact, Mariz claims that the net impact of adding a simple, nonregenerable SO_2 scrubber may be a slight capital savings (3).

7.20 Total Cost of a Standard Design Econamine FG Plant. The total operating plus capital cost of a 1000 te/d CO_2 Econamine FG plant of current standard design, including operating cost, capital cost recovery, and SO_2 scrubbing, for a low sulfur coal-fired flue gas would be:

US \$29.50/tonne = \$26.80/ton = \$1.55/1000 scf = \$57.90/1000 Nm^3.

The total cost for a 1000 te/d plant with gas turbine-fired flue gas:

US \$43.50/tonne = \$39.50/ton = \$2.29/1000 scf = \$85.40/1000 Nm^3.

The assumptions, including those about site-specific factors, are detailed in the operating cost, capital cost, and SO_2 scrubber capital cost sections above. A number of assumptions concerning owner's cost are implicit in the capital cost recovery. For more accurate costing, consult the specific process licensor.

7.21 Future Reductions in the Total Cost of an Econamine FG Plant. There are still large opportunities for future reductions in the total costs of large Econamine FG plants. The capital cost of a maximum single train size plant would be reduced on a per-tonne basis by 30 to 36% for a 2400 te/d natural gas-fired plant and 47 to 54% for a 4600 te/d coal-fired plant. Fluor Daniel is performing engineering studies on large plants in this size range.

Suda (14, 15, 16) has studied the savings from thermal integration of a CO_2 recovery unit processing the entire flue gas stream with a 600 MW, LNG-fired boiler. When the CO_2 recovery unit uses steam extracted from the low pressure steam turbine to reboil the stripper, and preheats boiler feedwater using the stripper overhead cooler, and uses the KS-1 proprietary amine with a regeneration energy 12% lower than MEA, the steam demand of the CO_2 recovery unit drops from 20% to 7.3% of the total usable power plant steam enthalpy. Most of this 64% energy savings is applicable to the Econamine FG process as well. Furthermore, Suda also concludes that use of KP-1, a proprietary low pressure drop structured absorber packing under development by KEPCO, could save an amount equivalent to 1.4% of the total power plant electrical output.

Tontiwachwuthikul (4) has studied a similar plant integration scheme for an Econamine FG or AMP-solvent plant in western Canada. This study concludes that a CO_2 production cost of US \$0.90-1.10/1000 scf (reported as Canadian \$1.25-1.50) is achievable. This is the equivalent of US \$17.50-20.90/tonne. This study also reexamines EOR pricing factors and concludes that the resulting total production cost of an incremental barrel of Saskatchewan oil would be US \$13-15.

Additional operating cost savings on the order of several percent have been identified in recent licensor studies when a split flow solvent (lean and semi-lean) process is applicable.

These opportunities when taken together would cut the price of CO_2 approximately in half.

8.0 CONCLUSIONS

The technology to recover CO_2 from flue gases is commercially available and, though mature, is being significantly improved. The Econamine FG process can be utilized to produce liquid CO_2 for the merchant CO_2 market at reasonable prices for projects where by-product CO_2 is not available. The process is most attractive for very large plants and is also routinely used for small, skid-mounted units for remote site applications. This paper provides an up-to-date review of the factors that determine Econamine FG process plant costs. A 1000 te/d plant of conservative design can produce CO_2 for $29.50/tonne or $1.55/1000 scf from coal-fired flue gas. However, 4600 te/d single-train plants are possible. Economy of scale, together with the ongoing development of the Econamine FG process, have the potential of delivering CO_2 at a price approximately half that of the reference 1000 te/d plant.

9.0 NOTICES

Econamine FG℠ is a service mark of Fluor Daniel, Inc.
GAS/SPEC FT-1™ is a trademark of the Dow Chemical Company.
KS-1 is a proprietary hindered amine developed by the Kansai Electric Power Company.
KP-1 is a proprietary structured packing developed by the Kansai Electric Power Company.
Amine Guard^R is a registered trademark of the UOP Company.

10.0 REFERENCES

1 Taber, J. J., "Need, Potential and Status of CO_2 for Enhanced Oil Recovery," conference paper, U.S.D.O.E. and Argonne National Laboratory, February 1985.

2 Taber, J. J., Martin, F. D., Seright, R. S., "EOR screening criteria revisited-Part 1: Introduction to screening criteria and enhanced recovery field projects," SPE Reservoir Engineering, Vol. 12, No. 3, August 1997, pp. 189-198.

3 Mariz, C. L., "Carbon Dioxide Recovery: Large Scale Design Trends," Journal of Canadian Petroleum Technology, Vol. 37, No. 7, July 1998, pp. 42-47.

4 Tontiwachwuthikul, Paitoon, Chan, Christine W., Kritpiphat, Weerapong, deMontigny, David, Skoropad, David., Gelowitz, Don, Aroonwilas, Adisorn, "Large Scale Carbon Dioxide Production from Coal-Fired Power Stations for Enhanced Oil Recovery: A New Economic Feasibility Study," Journal of Canadian Petroleum Technology, in publication.

5 Wilson, M.A., Wrubleski, R.M., and Yarborough, L.. "Recovery of CO_2 from Power Plant Flue Gases Using Amines," First International Conference on Carbon Dioxide Removal, Amsterdam, The Netherlands, Pergamon Press, March, 1992, pp. 325-332.

6 Horner, W. Norval, "The Sundance Carbon Dioxide Project," conference paper, Canadian Gas Processors Association, Westin Hotel, Calgary, Alberta, June 9, 1982.

7 Saskatchewan Heavy Oil/ Fossil Fuels Research Program, Boundary Dam CO_2 Extraction Pilot Plant-Final Report, Canada 1989.

8 Amoco Canada Petroleum Company Limited, Boundary Dam CO_2 Extraction Pilot Plant; Final Report, December, 1989.

9 Kohl, Arthur L., Nielsen, Richard B., Gas Purification, fifth edition, Gulf Publishing Co, Houston, Texas, 1997, ISBN 0-88415-220-0.

10 Brady, Jack D., "Flue Gas Scrubbing Process for Sulfur Dioxide and Particulate Emission Preceding Carbon Dioxide Absorption," conference paper No. 31d, American Institute of Chemical Engineers Spring National Meeting, April 6-10, 1986, New Orleans, Louisiana.

11 Hopson, Steve, "Amine Inhibitor Copes with Corrosion," Oil & Gas Journal, Vol. 83, No. 26, July 1, 1985, pp. 44-47.

12 Suda, T., Fujii, M., Mimura, T., Shimojo, S., Iijima, M., Mitsuoka, S., "Development of Flue Gas Carbon Dioxide Recovery Technology," International Symposium on CO_2 Fixation and Efficient Utilization of Energy, November 29-December 1, 1993.

13 Suda, T., Fujii, M., Yoshida, K., Iijima, M., Seto, T., Mitsuoka, S., "Development of Flue Gas Carbon Dioxide Recovery Technology," First International Conference on Carbon Dioxide Removal, Amsterdam, The Netherlands, Pergamon Press, March, 1992, pp. 317-324.

14 Pauley, C. Rich, Pat L. Simiskey and Scott Haigh, "N-Ren Recovers CO_2 from Flue Gas Economically," Oil & Gas Journal, Vol. 82, No. 20, May 14, 1984, pp. 87-92.

15 Abraham, B. M., Asbury, J. G., Lynch, E. P., Teotia, A. P. S., "Coal-Oxygen Process Provides CO_2 for Enhanced Recovery," Oil & Gas Journal, Vol. 80, No. 11, March 15, 1982, pp. 68-70, 75.

16 Holmes, Arthur S., "Recovery of CO_2 from Man-made Sources Using Cryogenic Distillation Techniques," conference paper, U.S.D.O.E. and Argonne National Laboratory, February 1985.

17 "Gas Processes '98," Hydrocarbon Processing, Vol. 77, No. 4, April 1998.

18 "Gas Process Handbook '92," Hydrocarbon Processing, Vol. 71, No. 4, April 1992.

19 Grover, B.S. and E. L. Bergen, "CO_2 Recovery from Flue Gas by the Benfield Process," Petroenergy 1983, Houston, Texas, September 15, 1983

20 Mimura, Tomio, Shimojo, S., Suda, T., Iijima, M., Mitsuoka, S., "Research and development on energy saving technology for flue gas carbon dioxide recovery and steam system in power plant," Energy Conversion and Management, Vol. 36., No. 6-9, June-September 1995, pp. 397-400.

21 Gelowitz, Don, Carbon Dioxide Extraction: Optimization of Facility and Solvent Technologies, Chapter 6, pp. 28-33, Regina, Saskatchewan, April, 1993.

22 Sander, Matthew T., and Mariz, Carl L., "The Fluor Daniel Econamine FG^{SM} PROCESS: Past Experience and Present Day Focus," Energy Conversion Management, Vol. 33, No. 5-8, 1992, pp. 341-348.

23 Gosar, D.M., Miller, D.B., and Soychak, T.J., "Economics of Recovering CO_2 from Exhaust Gases of Cogeneration Facilities Utilizing Gas Turbines or Gas Engines," conference paper No. 21d, American Institute of Chemical Engineers Spring National Meeting, April 6-10, 1986, New Orleans, Louisiana.

24 Simiskey, Patrick L., "The Recovery of CO_2 from Flue Gases," American Institute of Chemical Engineers Spring National Meeting, April 6-10, 1986, New Orleans, Louisiana.

25 Smelser, S.C., Stock, R.M., and McCleary, G.J., Engineering and Economic Evaluation of CO_2 Removal from Fossil-Fuel-Fired Power Plants, Vol. 1: Pulverized-Coal-Fired Power Plants, EPRI IE-7365, 1991

26 St. Clair, J.H., and Simister, W.F., "Process to Recover CO_2 from Flue Gas Gets First Large Scale Tryout in Texas," Oil and Gas Journal, Vol. 81, No. 6, Feb 14, 1983, pp. 109-113.

Table 1

FLUOR DANIEL ECONAMINE FG^{sm}
COMMERCIAL PLANTS

OWNER	LOCATION	SIZE, Te/D	CO_2 USE
Plants no longer in operation			
N-Ren Southwest	Carlsbad, New Mexico	90	EOR
Carbon Dioxide Tech. Corp.	Lubbock, Texas	1000	EOR
Paca	Israel	25	Food Industry
Plants in operation			
Liquid Air Australia	Altona, Australia	60	Food Industry
Liquid Air Australia	Botany, Australia	60	Food Industry
*Industrial de Gaseoses Cia. Ltda.	Quito, Ecuador	6.0	Food Industry
*Pepsi Cola	Manila, Philippines	6.0	Food Industry
*Pepsi Cola	Quezon City, Philippines	6.0	Food Industry
*Cosmos Bottling Co.	San Fernando, Philippines	6.0	Food Industry
*San Miguel Corp.	San Fernando, Philippines	45	Food Industry
Indo Gulf Fertilizer Co.	Uttar Pradesh, India	150	Urea Plant Feed
Luzhou Natural Gas	Sechuan Province, PRC	160	Urea Plant Feed
Northeast Energy Associates	Bellingham, Mass.	320	Food Industry
Kansei Electric Power Co.	Osaka, Japan	2.0	Pilot Plant
Tokyo Electric Power Co.	Japan	5.0	Pilot Plant
Sumitomo Chem/Nippon Oxygen	Chiba, Japan	160	Food Industry
*Cervezaria Baveria	Barranquilla, Colombia	25	Food Industry
Prosint	Rio de Janeiro, Brazil	90	Food Industry
*Coca Cola	Cairo, Egypt	6.0	Food Industry
*Azucar Liquida SA	Santo Domingo, Dom. Rep.	6.0	Food Industry
# European Drinks	Sudrigiu, Bihor County, Romania	36	Food Industry
*Messer Greisheim do Brazil Ltda	Sao Paulo, Brazil	50	Food Industry
# Messer Greisheim do Brazil/SPAL	Sao Paulo, Brazil	80	Food Industry

Skid-mounted plants built by Union Engineering A/S
*Skid-mounted plants built by Wittemann

North America:

5995 Rogerdale Road
Houston, Texas 77072 United States
+1.832.351.7887 (fax)
+1.832.351.7800 (phone)

525 West Monroe, Suite 1350
Chicago, Illinois 60661 United States
+1.312.655.9706 (fax)
+1.312.655.9207 (phone)

205 Quarry Park Boulevard SE
Calgary, AB T2C 3E7 Canada
+1.403.385.1543 (fax)
+1.403.258.6896 (phone)

Europe:

Tower Bridge Court
226 Tower Bridge Road
London SE1 2UP United Kingdom
+44.(0)20.7089.7750 (fax)
+44.(0)20.7403.3330 (phone)

100 Plesmanlaan
2332 CB Leiden, The Netherlands
+31.71.582.7050 (fax)
+31.71.582.7144 (phone)

Asia:

B-1523, Nanxincang Business Tower
No. 22 Dongsishitiao
Dongcheng District
Beijing, 100007
+86.10.6409.6636 (phone)

6F South Tower B, China Diamond Exchange Center
1701 Century Avenue, Pudong New District
Shanghai, 200122
+86.21.5081.6006 (fax)
+86.21.5081.8811 * 607 (phone)

JACOBS® Consultancy

Visit our website (www.jacobsconsultancy.com) or e-mail us (pce@jacobsconsultancy.com) for more information regarding the services we offer.