JACOBS Consultancy

Evaluation of Integrating a Molten Carbonate Fuel Cell (MCFC) with a SAGD Facility

Prepared For

Alberta Innovates – Energy and Environment Solutions

July 2015

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For Jacobs Consultancy

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Section A.

Executive Summary

Background

Alberta Innovates – Energy and Environment Solutions (“AI-EES”) and an industry consortium of three companies commissioned Jacobs Consultancy Canada Inc. (“Jacobs Consultancy”) to perform an assessment of the benefits of integrating a Steam Assisted Gravity Drainage (“SAGD”) central processing facility (“CPF”) with a molten carbonate fuel cell coupled with CO_2 capture and compression (“MCFC+CO_2”). The integration assessment (the “Study”) is a follow-up to a techno-economic comparison study (“Phase 1”) performed by Jacobs Consultancy in 2013 that found that the MCFC+CO_2 process had the potential to capture and compress CO_2 with lower costs of captured and avoided CO_2 than competing commercial amine-based technologies.

AI-EES directed the Study, which was funded and directed by a Steering Committee consisting of the following companies:

- AI-EES (Sponsor and Steering Committee Chair)
- MEG Energy
- Shell Canada
- Suncor Energy (“Suncor”)

Fuel Cell Energy (“FCE”) has developed a technology configuration using MCFC called *Combined Electric Power and Carbon Dioxide Separation* (“CEPACS”) that utilizes the MCFC to concentrate CO_2 which is later purified and compressed, forming a complete MCFC plus CO_2 concentration and compression system that for the purposes of this Study will be referred to interchangeably as *MCFC* or *MCFC+CO_2*. We have used FCE’s technology as the representative MCFC+CO_2 for the purposes of this Study.

This report documents our methodology, basis, findings, and recommendations for the Study.

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.

In Phase 1, we found that MCFC+CO_2 had the potential to capture and compress CO_2 from a SAGD plant at a lower cost than commercially available amine-based technologies.

Therefore, the objective of the Study was to determine the benefits of integrating MCFC technology and CO_2 capture and compression with a Warm Lime Softening (“WLS”) / Once-Through Steam Generator (“OTSG”) based SAGD CPF via the following metrics:

- Energy efficiency
- Make-up water intensity
- Direct / indirect CO_2 emitted and captured
- Capital and operating costs

Based on the analysis above, we were also asked to provide comments regarding the integration of the MCFC+CO_2 facility into SAGD CPF using evaporators and drum boilers for water treatment and steam generation, respectively and a SAGD CPF generating steam and power via a combination of conventional cogen and boilers.

Study Highlights

The following list summarizes the highlights of the Study:

- Integrating streams between a SAGD CPF and MCFC+CO_2 unit needs to be focused on improving the efficiency of the SAGD plant to reduce capital costs and operating costs, because it is less expensive to reduce the generation of CO_2 than to generate it and then capture CO_2.

- Integration streams between the SAGD CPF and MCFC+CO_2 are limited to flue gas from the SAGD to the MCFC+CO_2 and water, heat, and power from the MCFC+CO_2 unit back to SAGD.
- Both the SAGD plant and the MCFC+CO_2 unit are extensively heat integrated on an individual basis; therefore, the potential for economical heat integration between the MCFC+CO_2 unit and SAGD is limited.
- Water and power from the MCFC+CO_2 are readily integrated into the SAGD CPF flow scheme with very little change in the process flow of the SAGD CPF.

Study Basis

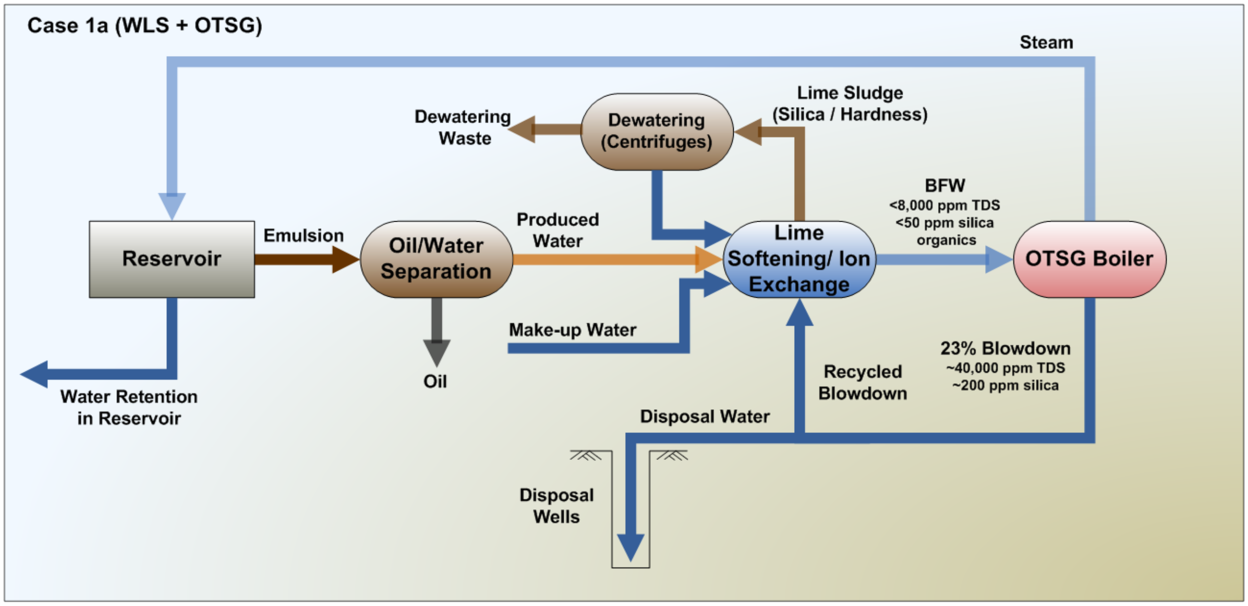
The request for proposal originally envisioned evaluating the integration potential of the MCFC+CO_2 unit with three different SAGD CPF configurations based on water treatment and steam generation technology differences. However, to reduce costs, the evaluation was narrowed to a single SAGD CPF flow-scheme (Group 1) plus any qualitative comments regarding integration with the other two SAGD CPF configurations (Groups 2 and 3) if applicable. For the purposes of this report, Case 1a refers to the Group 1 configuration using warm lime softening (WLS) and once-through steam generators (OTSG) without CO_2 capture, while Case 1b refers to the same SAGD CPF configuration but includes MCFC+CO_2 for CO_2 capture.

From a SAGD perspective, the Study basis was the same as Phase 1. The most notable difference is that the emulsion temperature to the CPF as reflected in the Request for Proposal (RFP) was lower than in Phase 1.

The Steering Committee decided at the kick-off meeting to size the MCFC+CO_2 unit to handle the entire flue gas stream and capture and compress 90 percent of the CO_2 in the flue gas.

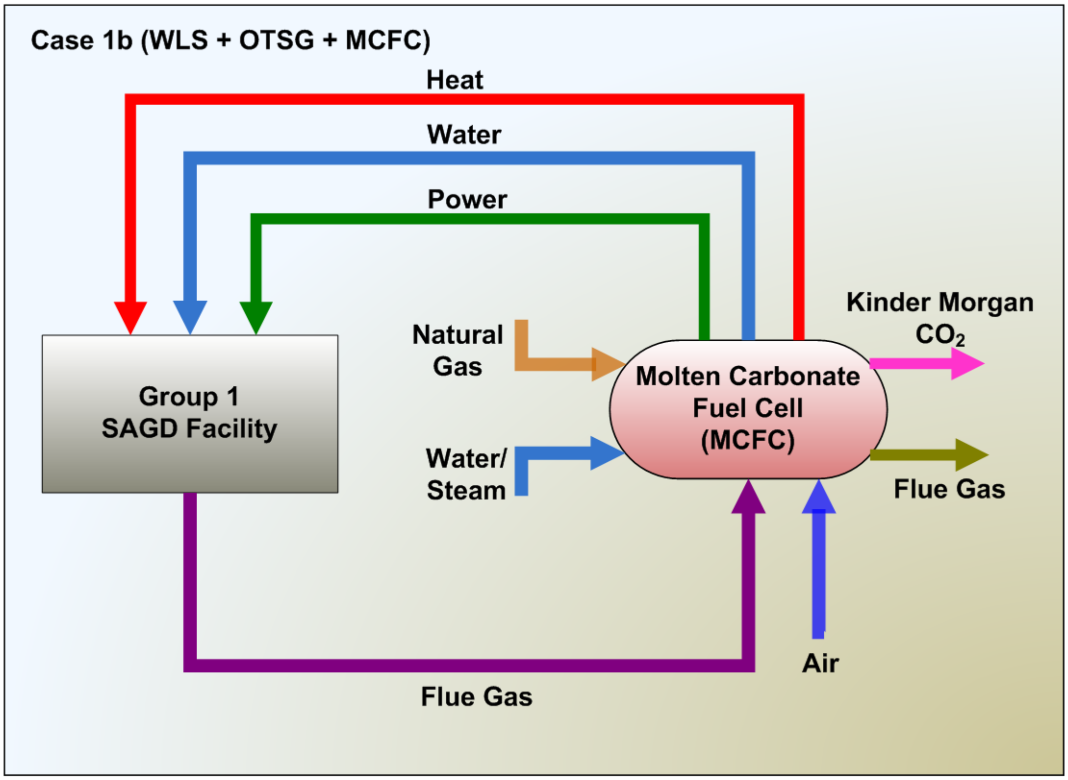
The SAGD CPF configuration evaluated is represented in Figure A-1 as a conceptual flow diagram (“CFD”). The SAGD flow scheme is conventional. The first step is oil water separation of the incoming emulsion. The produced water is de-oiled, treated with a warm lime softener and ion exchange. Make-up water is added and treated along with the produced water. The combined treated water becomes boiler feed water. The boiler feed water is heated then fed at high pressure to the OTSG, which produces steam at 77 percent quality. The blowdown is recycled back through water treatment to a recycle limit of 65 percent. The excess blowdown is cooled, treated for silica, and disposed via deep well injection.

Figure A-1.
Case 1a - SAGD CPF Configuration



Conceptually, the integration of the SAGD CPF (represented in Figure A-1) with the MCFC+CO_2 is shown in Figure A-2. As discussed above, flue gas from the SAGD CPF is routed to the MCFC+CO_2 system for concentration and compression. Power, water, and heat from the MCFC+CO_2 are then available for use by the SAGD CPF.

Figure A-2.
Case 1b - Simplified Integration Diagram



Methodology

At a high level, Jacobs Consultancy performed the following main activities for Cases 1a and 1b:

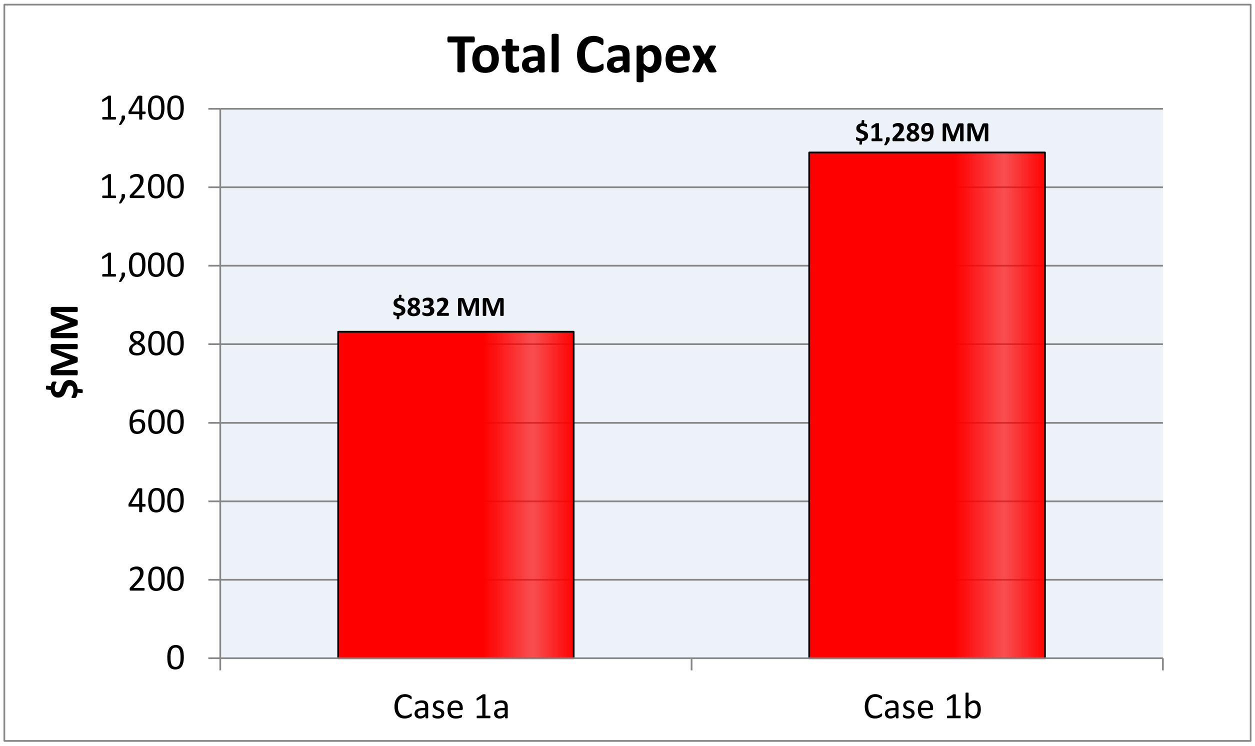
1. Determined the integration basis
2. Brainstormed integration opportunities
3. Evaluated the options
4. Selected the optimum integration option
5. Finalized the high-level material balances
6. Estimated the Capex and Opex for the selected option
7. Tabulated results

Results

Our analysis showed that it was only economic to integrate power and water from the MCFC+CO_2 into the SAGD CPF configuration. Heat produced by the MCFC+CO_2 was not at a temperature sufficient to invest any capital for using the heat at the SAGD CPF. Therefore, only the water and power were used at the SAGD CPF, which reduced the amount of makeup water required, met all the power requirements for the SAGD facility—including the production well pumps—and converted the whole facility into a net power exporter.

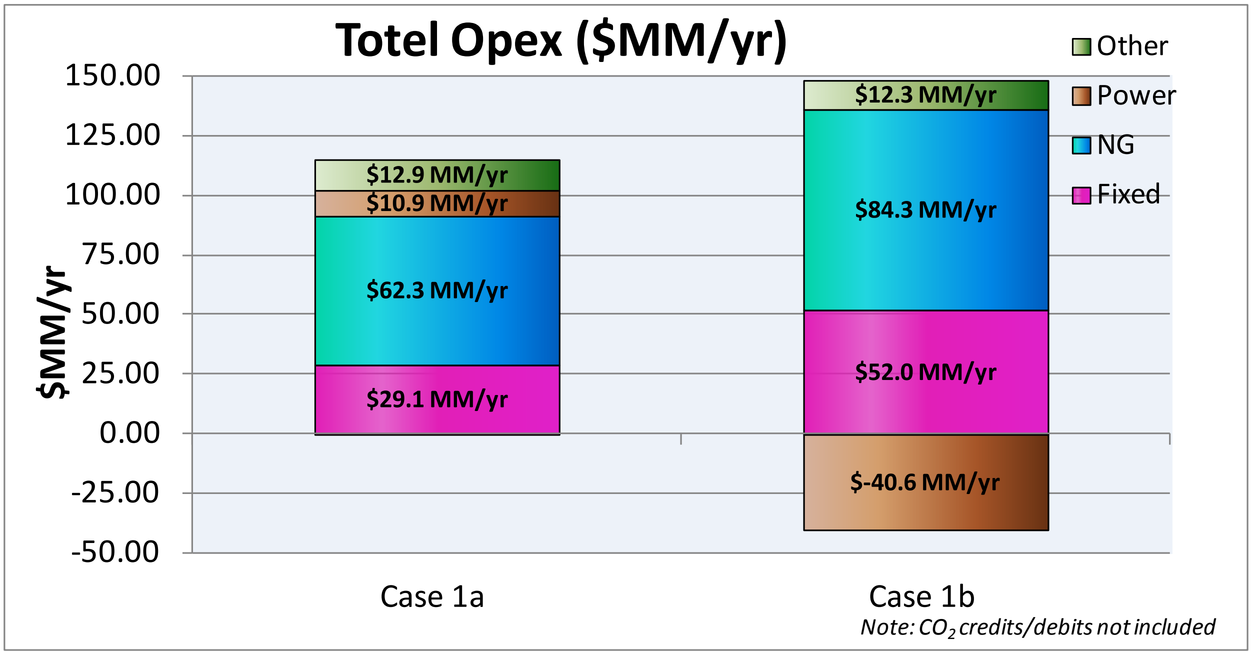
Capital costs (Capex) increased by about $450 MM for the addition of the MCFC+CO_2 on the SAGD CPF, with 99 percent of the additional costs resulting from the fuel cells and carbon capture and compression equipment. Costs to integrate the power and water are estimated to be less than $10 MM. Figure A-3 summarizes the capital costs for both the SAGD CPF and the SAGD CPF with the integrated MCFC+CO_2.

Figure A-3.
Capex



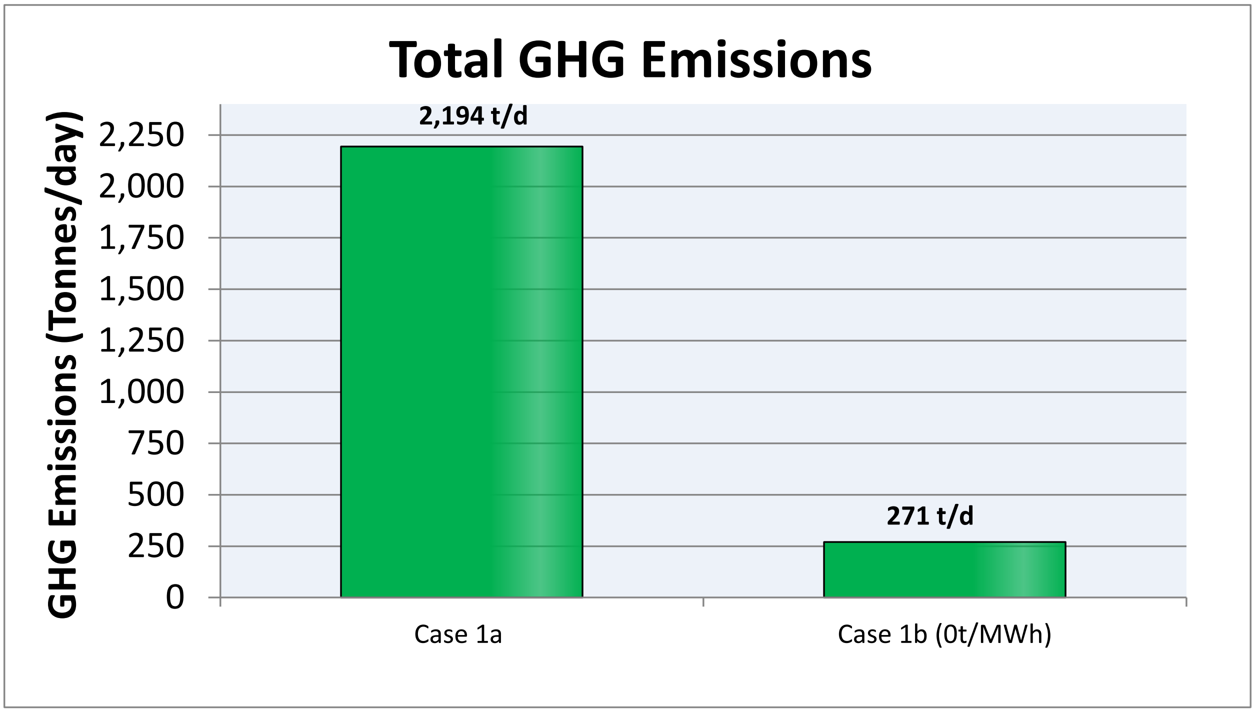
As shown in Figure A-4, estimated operating costs (Opex) actually decreased with the integration of the MCFC+CO_2 as a result of the 51 MW of power exported to the grid at a price of $90/MWhr. The estimated operating costs were $115 MM/yr for the SAGD complex and $108 MM/yr with the integration of the MCFC+CO_2 plant, even with fixed operating costs increasing an estimated $23 MM/yr due to the increased capital cost and an increase in natural gas usage with a cost of $22 MM/yr.

Figure A-4.
Opex



Note: CO_2 credits/debits not included

Figure A-5.
GHG Emissions



The primary reason for integrating the MCFC+CO_2 facility in a SAGD CPF is to reduce the amount of GHG emissions, assuming that the compressed and concentrated CO_2 is sequestered. In this regard, the MCFC+CO_2 has a large impact on GHG emissions. As shown in Figure A-5, where we assume 90 percent capture, dilute CO_2 emitted by a SAGD CPF is reduced more than 1,920 tonnes/day, even without any GHG credits for the power exported to the grid. This case is identified as Case 1b (0t/MWh) which shows zero tonnes per mega-watt hour used for power export.

Qualitative Comments Regarding Other SAGD CPF Configurations

As indicated in the objectives, we were asked to provide qualitative comments regarding the integration of the MCFC+CO_2 into other water treatment and steam generation configurations for the SAGD CPF, based on our analysis of the WLS + OTSG SAGD configuration. The two configurations were labeled Group 2 and Group 3. Simplified representations of each are shown in Figures A-6 and A-7.

Figure A-6.
Case 2a Configuration

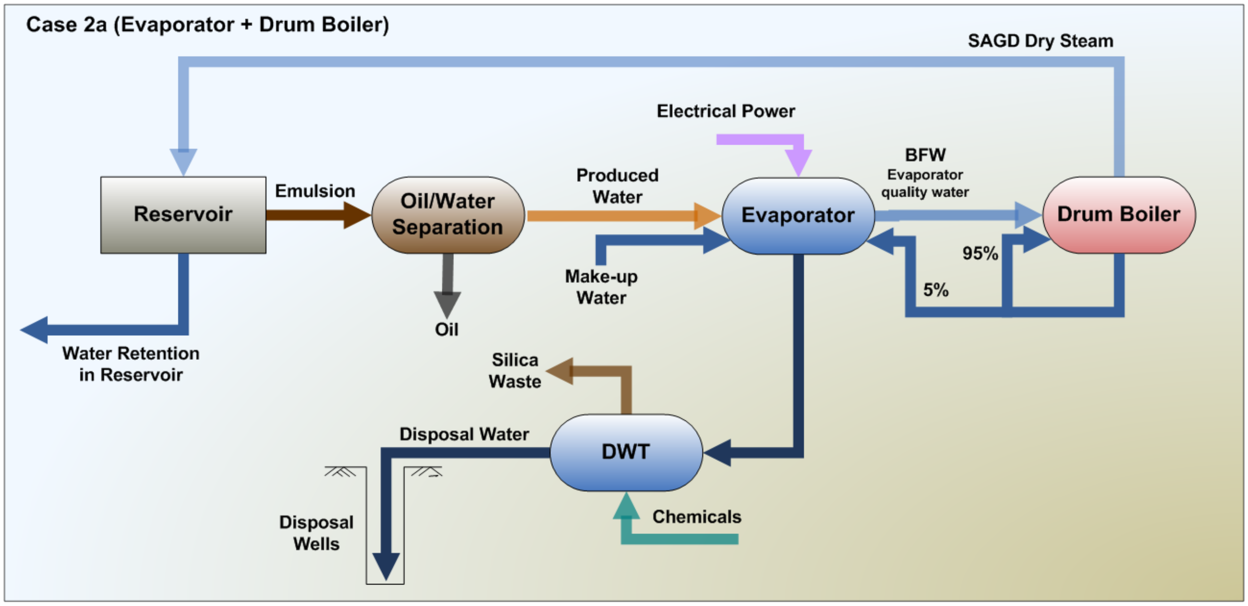
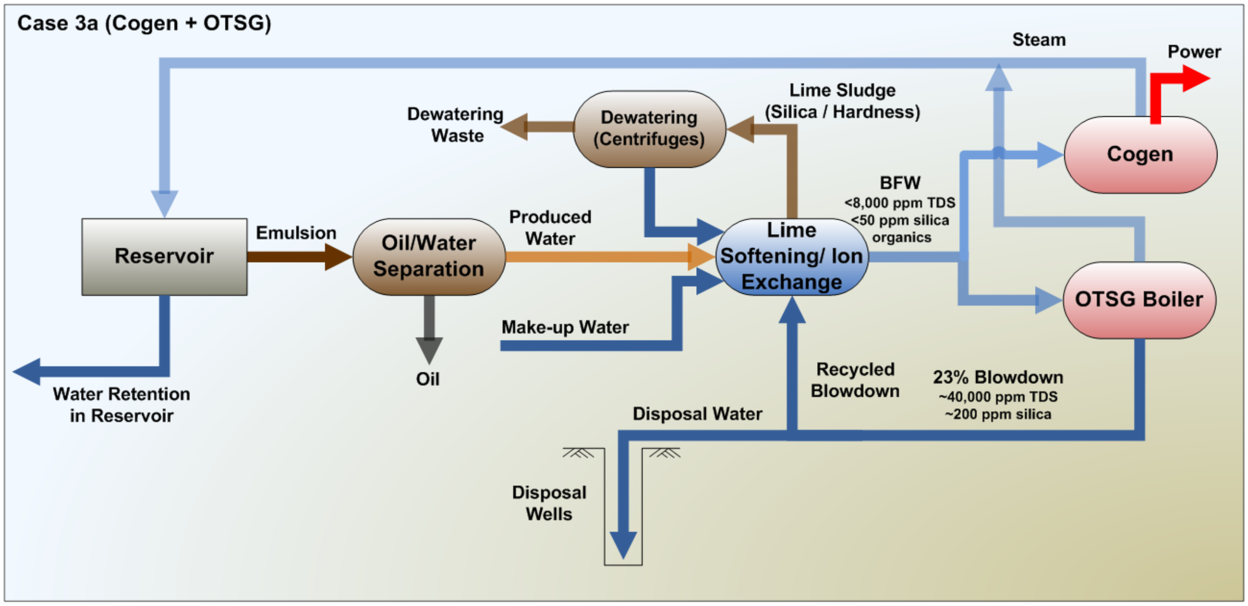


Figure A-7.
Case 3a Configuration



Our comments regarding the integration of the MCFC+CO_2 into an evaporator+drum boiler SAGD CPF represented in Figure A-6 are as follows:

- From a SAGD perspective, drum boilers offer slightly better efficiency than OTSG, so natural gas combustion and therefore flue gas production are lower. Evaporator-based water treatment uses more power than WLS, so the configuration consumes more power. In addition, make-up water is lower because the disposal volume is less. The heat exchange scheme is similar to Group 1.
- The MCFC+CO_2 system is slightly smaller due to decreased flue gas
- Therefore, in our opinion:
  - Power consumption on-site increases and MCFC production is lower, so the amount of power exported goes down.
  - Water requirements from the CPF are lower but with reservoir retentions close to 10 percent, all the MCFC+CO_2 water can be utilized by the CPF and make-up water is lower than in Group 1.
  - There is no change in heat integration. The level of heat produced by the MCFC is too low to be used economically.

Our comments regarding the integration of the MCFC+CO_2 into a cogen + OTSG SAGD CPF represented in Figure A-7 are as follows:

- The SAGD CPF has more flue gas because of the cogen. There will be additional heat, O_2 and NOx in the flue gas. However, make-up water requirements should be the same as Group 1 and the heat integration is similar to Group 1. The biggest change is that the SAGD CPF is a net exporter of power before the addition of the MCFC+CO_2
- The MCFC+CO_2 system is larger because the quantity of flue gas and the flue gas cooling requirements increased due to higher temperature flue gas. The fuel cell, however, should be fine with increases in O_2 and NOx.
- Therefore, in our opinion:
    - The site becomes an even larger exporter of power with the addition of the MCFC+CO_2. Infrastructure costs associated with switching from a net importer to net exporter of power may be lower.
    - The complex may be long on water due to increased water from the MCFC+CO_2.
    - Similar to Group 1, heat integration is likely to have little value.

Findings and Recommendations

In our opinion, integrating an MCFC+CO_2 facility into a SAGD CPF does the following:

- Reduces dilute CO_2 emission by 88 percent not including GHG credits for exporting 51 MW of power to the grid.
- Increases capital cost by over 50 percent compared to the cost of the SAGD CPF primarily due to the cost of installing the MCFC+CO_2 facility. Integration costs by themselves are less than $10 MM.
- Reduces operating costs by 6 percent assuming that the power can be sold to the grid at $90/MWhr and does not include GHG credits.
- Reduces make-up water requirements and eliminates the need for purchased power for the SAGD CPF.
- Is enabled by the uncomplicated nature of integrating power and water. Integrating the heat from the MCFC+CO_2 in the SAGD complex has little benefit.

Integrating the MCFC+CO_2 represented in the Study as the Combined Electric Power and Carbon Dioxide Separation ("CEPACS") by Fuel Cell Energy into a SAGD CPF is economically limited to power and water, and therefore presents very little technical risk. In addition, the

MCFC itself is commercialized. The addition of an MCFC+CO_2 system into a SAGD CPF, however, does present the following primary risks:

- Capital cost: Installing an MCFC+CO_2 of this size has not been done. In addition, there are capital cost risks associated with installing a plant of this magnitude in northern Alberta.
- Reliability, operability, and maintenance costs: Although most of the components of the MCFC+CO_2 are commercial or near commercial, the reliable operation and maintenance costs resulting from an installation of this size in northern Alberta are risks. There are likely to be unforeseen challenges regarding the operation and maintenance of this type of facility.

Therefore, in our opinion, this Study supports the findings from Phase 1, in that the MCFC+CO_2 addition to SAGD CPF can offer substantially lower costs of capture than commercially available amine systems and is a promising technology for CO_2 capture and compression for the purposes of producing CO_2 for EOR or sequestration. We recommend in the near term:

- Feasibility studies to investigate the benefits of MCFC+CO_2 for capturing CO_2 from the flue gas of Steam Methane Reformers (SMR) and smaller fired heaters present in upgraders and refineries.
- A detailed study in conjunction with Fuel Cell Energy focused on the trade-offs of increasing the amount and temperature of the heat generated by the fuel cell for the purposes of reducing natural gas consumption for SAGD.
- A design study to determine location, feasibility, and estimated capital cost of an MCFC+CO_2 demonstration facility in Alberta to confirm reliability, operability, and commercial readiness, and to highlight other development issues.
- A study to investigate the economics and emissions resulting from the co-production of hydrogen with an MCFC+CO_2. The MCFC operating in Orange County, CA is configured to produce and concentrate hydrogen in addition to generating power.

Section B.

Study Background

The Alberta government and the oil sands industry have collectively been committed to evaluating techniques in efforts to reduce greenhouse gas (“GHG”) emissions from oil sands production. The development of low-cost carbon capture technologies has been a prime focus to bringing producers closer in reaching Alberta’s GHG emission reduction goals. As a part of this initiative, Jacobs Consultancy completed the Phase 1 Study “Evaluation of Electrochemical Membranes for Carbon Capture at a SAGD Facility” in 2012, which examined the potential of combining Electrochemical Membrane (“ECM”)—generically referred to as Molten Carbonate Fuel Cell (“*MCFC”)*—for carbon capture from SAGD operations. The study showed significantly lower costs of capture for this technology than for commercial amine-based systems. Due to the encouraging results, AI-EES and a number of oil sands companies were interested in evaluating the integration of MCFC into a SAGD CPF as opposed to just adding the MCFC to the back end of the SAGD CPF.

Fuel Cell Energy (“FCE”) has developed a technology configuration using MCFC called *Combined Electric Power and Carbon Dioxide Separation* (“CEPACS”) that utilizes the MCFC to concentrate CO_2 which is later purified and compressed, forming a complete MCFC plus CO_2 concentration and compression system that for the purposes of this Study will be referred to interchangeably as *MCFC* or *MCFC+*CO_2.

Study Objectives

The objective of the Study was to determine the benefits of integrating MCFC technology and CO_2 capture and compression with a Warm Lime Softening (“WLS”) / Once-Through Steam Generator (“OTSG”) based SAGD CPF via the following metrics:

- Energy efficiency
- Make-up water intensity
- Direct / indirect CO_2 emitted and captured
- Capital and operating costs

Based on the analysis above, we were also asked to provide comments regarding the integration of the MCFC+CO_2 facility into SAGD CPF using evaporators and drum boilers for water treatment and steam generation, respectively, and a SAGD CPF generating steam and power via a combination of conventional cogen and boilers.

Configuration Options

The SAGD CPF configurations were split into three Groups as follows:

- **Group 1**
    a. **Case 1a: Warm Lime Softener ("WLS") + OTSG**

    Case 1a is the reference case of a SAGD production facility that employs WLS for water treatment and OTSG to generate steam by burning natural gas and imports electricity from the grid.

    b. **Case 1b: WLS + OTSG + MCFC+CO_2 capture and compression**

    Case 1b is the case of a SAGD facility that employs WLS for water treatment and OTSG to generate steam by burning natural gas. The power is generated by MCFC. MCFC is used to separate OTSG CO_2 and is integrated with the whole SAGD system. The captured CO_2 is compressed to Kinder Morgan pipeline specs.

Based on the results found for the Group 1 cases, qualitative comments regarding integration for Groups 2 and 3 are found in Section E. Groups 2 and 3 are summarized below.

- **Group 2**
    a. **Case 2a: Evaporator + Drum Boiler**

    Case 2a is the reference case of a SAGD production facility that employs an evaporator for water treatment and drum boiler to generate steam by burning natural gas and imports electricity from the grid.

    b. **Case 2b: Evaporator + Drum Boiler + MCFC+CO_2 capture and compression**

    Case 2b is the case of a SAGD production facility that employs an evaporator for water treatment and drum boiler to generate steam by burning natural gas. The power is generated by MCFC. MCFC is used to separate CO_2 and is fully integrated with the whole SAGD system. The captured CO_2 is compressed to Kinder Morgan pipeline specs.

- **Group 3**
    a. **Case 3a: WLS + [Cogen (50%) + OTSG (50%)]**

    Case 3a is the case of a SAGD production facility that employs WLS for water treatment and cogeneration and OTSG to generate steam by burning natural gas. Cogeneration ("Cogen") and OTSG 50:50 split refers to steam, each to meet 50 percent total steam demand. Power is generated by the gas turbine generator ("GTG") part of Cogen.

b. **Case 3b: WLS + [Cogen (50%) + OTSG (50%)] + MCFC+CO_2 capture and compression**

Case 3b is the case of a thermal in-situ production facility that employs WLS for water treatment and cogeneration and OTSG to generate steam by burning natural gas. Power is generated by both the Cogen (GTG) and MCFC. MCFC is used to separate CO_2 from a combined OTSG flue gas and Cogen flue gas. MCFC is fully integrated with the entire SAGD surface facility and lifting system. The captured CO_2 is compressed to Kinder Morgan pipeline specs.

Basis and Assumptions

For consistency purposes, the set of basis and assumptions developed for the Phase 1 Study was used in this Study, with the exception of a lower produced emulsion temperature as indicated in the request for proposal ("RFP").. The SAGD basis is summarized as follows:

- 33,000 barrels per calendar day ("bpcd") SAGD production facility
- Steam-to-oil ratio ("SOR") = 3.0
- Gas-to-oil ratio ("GOR") = 5.0; produced gas will offset natural gas purchases
- Produced Water to Steam Ratio (PWSR) = 0.9
- Cases were based on an "Efficient SAGD Design" which includes Electrical submersible pumps ("ESPs") for oil lifting, OTSG air economizer, efficient design, and minimization of Ethylene Glycol ("EG")
- OTSG steam quality = 77 percent at 100 barg
- Imported electrical power to meet on-site power demands
- Exported power transmission charges: 5 percent of net sales to AESO grid
- Produced fluid at 160°C
- De-oil process identical for the cases: free water knockout, followed by skim tank and Induced Gas Flotation ("IGF"), and then Oil Removal Filter ("ORF")
- Blowdown disposal by deep well injection
- Air preheat with flue gas economizer is economic
- Service factor of 94 percent
- Send 100 percent of flue gas to MCFC; capture 90 percent of CO_2 emitted from OTSG stack and MCFC and CO_2 purification and compression

- CO_2 that goes into pipeline is compressed to 2000 psi per Kinder Morgan specs
- Make-up water is from a saline aquifer with total dissolved solids > 4,000 ppm. Water qualities for the Study are shown in Table B-1:

Table B-1.
Produced and Make-up Water Qualities

Quality	Units	Produced Water	Make-up Water
TDS	ppm	1290	6961
Hardness	ppm	14	204
M-Alkalinity ($CaCO_3$)	ppm	173	665
Ca	ppm	4	34
Mg	ppm	1	29
TOC	ppm	300	6
Silica	ppm	188	7

- Heat recovery and rejection assumes the following:
    - Minimum atmospheric temperature = - 40°C
    - Maximum atmospheric temperature = 40°C
    - Average atmospheric temperature = 16°C
    - Maximum temperature for disposal = 80°C

GHG Emissions Basis

Direct and indirect GHG emissions were comprised of the following:

- In Case 1a, the direct GHG source is primarily the CO_2 from the OTSGs used to produce steam; in Case 1b, the direct GHG source is the vented CO_2 post MCFC and CO_2 capture and compression.
- Indirect GHG emissions are comprised of the following:
    - Imported power generation at 0.66 tonne of CO_2/MWh
    - Trucking of wastes at 0.018 tonne of CO_2/tonne
    - Production of imported chemicals at 1 tonne of CO_2/tonne of lime
- GHG credit for exported power in the MCFC+CO_2 integrated case is shown at 0.00 and 0.66 tonne of CO_2/MWh with sensitivities at 0.42 and 0.88 tonne of CO_2/MWh

Overall Methodology

Our methodology involved the following high-level tasks:

1. Prepare a design basis document and agree on basis and assumptions. Refer to Appendix 1.
2. Brainstorm options to integrate the MCFC+CO_2 into the SAGD CPF regarding heat, power, and water use
3. Analyze the integration options
4. Select an integration option for heat, water, and power
5. Develop BFDs with the following information for the base case and selected option:
    a. Flows of key streams
    b. Utility summaries
    c. Disposal volumes
    d. GHG emissions
    e. Water use
    f. Estimated variable operating costs
6. Estimate rough order of magnitude ("ROM") capital costs based on the selected integration option. For Case 1b, the Total Installed Cost ("TIC") of the MCFC was added to the TIC of the CPF to determine the total TIC.
7. Estimate fixed costs based on 3.5 percent of the TIC for the SAGD CPF and 5 percent of the TIC for the MCFC+CO_2 portion, which includes an allowance for periodic fuel cell stack replacement.
8. Prepare comparison graphs and report findings.
9. Based on the Group 1 analysis, provide qualitative comments regarding MCFC+CO_2 integration for Groups 2 and 3.

Operating Cost Estimate Basis and Methodology

Operating costs are a sum of both fixed and variable operating costs:

- Fixed operating costs include items such as operating labour, general and administrative expenses, maintenance, property taxes, and insurance; these are based on a factor of the TIC.
- Variable costs include utilities, fresh water, chemicals, and disposal.

The commodity prices used are shown in Table B-2 below. We did not account for CO_2 credits for exported power or CO_2 debits for indirect emissions associated with imported power in the variable operating costs. The quantities of each of the commodities are presented in the Block Flow Diagrams ("BFDs") and Utility Summaries presented in Appendices 2 and 3 and serve as a basis for the participating companies to perform individual economic evaluations.

Table B-2.
Commodity Prices

Commodity	Units	Price ($ CND)
Power	$/ MWhr	$90
Gas	$/ GJ	$5
Chemicals	$/ m3 water	$0.77
Disposal costs including trucking	$/ tonne	$200
Fixed operating costs	% of TIC	3.5 (CPF); 5.0 (MCFC)

SAGD CPF Capital Cost Estimate Basis and Methodology

The capital costs presented in this study represent an "unclassified" estimate with an accuracy of +100%/-50%. The capital costs are representative only and are considered sufficient for comparison purposes only.

For the SAGD CPF estimate, we used Jacobs' proprietary models to estimate capital costs for the Group 1 cases. The model includes historical information for costs and a parametric model to represent the SAGD facility including plot plan, modules, piping details, and a representative sized equipment list. Indirect costs, engineering, and other costs are factored off the total direct costs. Costs for well pads, drilling and completions, gathering and steam lines, and source and disposal water wells are not included in the capital cost estimate.

The following assumptions were used:

- Estimates are on a 1st half 2015 basis

- Location: Northern Alberta
- 100 percent Canadian sourced equipment
- 25 percent contingency for SAGD CPF facility consistent with the accuracy level of the estimate
- Escalation excluded
- No Owner’s Cost
- No Commissioning and Startup

MCFC+CO_2 Capital Cost Estimate Basis and Methodology

The MCFC+CO_2 cost estimate was developed utilizing a combination of unit capacity factored and budget quotes from licensors/vendors. These methods utilize historical data from plants with similar units or equipment. For Case 1b, the TIC for the MCFC+CO_2 was added to the TIC for the CPF.

Capacity-factored estimating is based on multiplying the cost of a similar unit, for which the direct construction costs are known from a prior project, by the ratio of the new unit's capacity to the capacity of the known unit. Capacity ratios are adjusted by an exponent chosen on the basis of the unit type. Adjustments are made as required to allow for differences in unit design bases, location, and time frame.

The base estimates were developed on a United States Gulf Coast (“USGC”) during the 1st half of 2015 time period broken down into labour, equipment, and materials. The USGC costs were then adjusted to a local site basis using location-specific information. The site is Fort McMurray, Alberta. The USGC to site adjustment factors include an efficiency factor applied to the construction labour hours, use of a local construction wage rate, a 10percent increase in equipment costs and a 17 percent increase in bulk material costs to account for winterization, and use of local factors for estimating indirect field costs. An exchange rate of 1.2678 CAD$/US$ was utilized. These adjustments were made for units where costs were developed by Jacobs as well as to the USGC cost data supplied by vendors.

Costs for CO_2 export pipeline are not included.

The estimate includes EPC contractor profit but does not include an EPC contractor risk fee for a Lump Sum-Turn Key estimate.

The capital cost estimates are comprised of the following components:

- Direct Field Costs
    - Materials: equipment and bulk materials
    - Labour
- Indirect Field Costs
    - Construction management
    - Temporary facilities and services
- Home Office Costs
- Contingency

The sum of these components is defined as the plant installed cost or EPC cost of the project because it contains the items typically within the scope of the EPC contractor. Owner's costs that are typically outside of the EPC cost are discussed in a later subsection.

Other Cost Estimate Information

Direct Field Costs

Materials

Direct field material costs are the costs of the permanent physical plant facilities and include the following elements:

- Excavation and Civil—Includes piling, roads, asphalt paving, and clean structural fill.
- Concrete and Fireproofing—Includes foundations, concrete structures, retaining walls, floor slabs, concrete area paving, and concrete fireproofing.
- Structural Steel—Includes steel structures, pipe racks, handrails, ladders, stairs, platforms, and miscellaneous supports.
- Buildings—Includes control, electrical substation, office, compressor, laboratory, warehouse, canteen, workshop and gatehouse buildings including framing, walls, exterior cladding and roofing, HVAC, electrical lighting and power, and interior fixtures and finishes.
- Equipment—Includes tanks, vessels, fuel cell stacks combustion turbines, steam turbines, compressors, heat exchangers, heat recovery steam generators ("HRSGs"),

pumps, material processing equipment, material handling equipment, and miscellaneous equipment (e.g., filters, strainers).

- Piping—Includes aboveground and underground pipe, fittings, valves, flanges, gaskets, shoes and guides, specialty items, and non-destructive examination.
- Electrical—Includes major electrical equipment (e.g., transformers, breakers, switches, bus bars, etc.), power wire and cable, conduit, cable tray, push-button stations, welding and power receptacles, lighting, grounding, and instrument wire and cable.
- Instrumentation—Includes process instruments, analysers, process connections, control house consoles, distributed control systems, and instrument mountings.
- Paint—Includes paint and coatings for plant equipment, piping, and structures.
- Insulation—Includes hot, cold, and personnel protection insulation for plant equipment and piping.

Materials costs have been adjusted for the site locations and include freight. The direct field material costs are based on worldwide procurement of material. For that material, charges have been included for inland freight to the shipping port, port charges, ocean freight, shipping insurance, and inland freight to the project site. A winterisation allowance of 17 percent has been applied to the bulk materials cost component, and a winterisation allowance of 10 percent has been applied to the equipment except for the fuel cell stacks. As the fuel cell stacks are located in a building, no winterization is required.

Labour

For the MCFC TIC, the direct field labour costs at the Canadian site locations were derived by multiplying the estimates of USGC field labour man-hours by a labour efficiency factor of 1.43 with a representative field rate.

For the SAGD CPF, labour hours were estimated based on historical information related to the amount of bulk materials such as concrete, piping, steel, and electrical cable.

Indirect Field Costs

As indicated above, indirect field costs related to camp, engineering, and consumables are factored from the total direct labour costs.

Contingency

Contingency is a special monetary provision in the estimate to cover uncertainties or unforeseeable elements of time/cost within the scope of the project. Costs associated with the following items are included in contingency:

- Material cost changes (other than scope changes)
- Labour rate changes
- Labour efficiency changes
- Design changes (other than scope changes)
- Schedule slippage
- Estimate methodology inaccuracies

Contingency excludes the items listed below because, with the exception of scope changes, they are unusual items that are not expected to occur on every project and their impacts can vary significantly. Including contingency for all of these items would result in an exceptionally high cost that would not represent the most likely cost for the project.

- Scope changes
- Changes in government regulations
- Major design changes
- Catastrophic events
- Extreme weather
- Unknown field conditions
- Labour shortages or strikes
- Significant exchange rate fluctuations

An overall contingency of 20 percent of the bottom line cost has been assumed for the MCFC+CO_2 and 25 percent for the SAGD CPF. Contingency has not been applied to individual estimate components at this early stage of the project. This level of contingency is considered an appropriate value when taking into account the extent of engineering and the factored approach used in developing the cost estimate.

Excluded Costs

Excluded costs include taxes, financing fees, interest during construction, working capital, and escalation.

Section C.

Case Descriptions

SAGD CPF Process Description

At the heart of any SAGD CPF is water treatment and steam generation. The product of any SAGD plant is diluted bitumen, but most of the operating and capital cost of the CPF is involved in water treatment and steam generation. Seven main processing areas are incorporated into the SAGD CPF:

- Oil Treatment—Emulsion is cooled and diluted, and two stages of water separation occur. Oil is degassed and sent to product tankage.
- Deoiling—Produced water is deoiled via gravity separation, gas flotation, and sometimes filtration.
- Water Treatment—Produced and make-up water is treated to reduce hardness and silica.
- Steam Generation—Steam is generated from high-pressure boiler feed water.
- Disposal Water Treating—Water with concentrated contaminants is pH adjusted, treated to reduce silica precipitation and disposed, typically via deep well injection.
- Gas Processing—Produced gas is recovered, compressed, separated from liquid, and combined with natural gas for combustion. If required, the produced gas may be treated to remove sulfur.
- Glycol and Heat Exchange—Heat is transferred either via direct heat exchange or via a glycol intermediary.

In all cases, emulsion is received into the CPF from the well pads. The emulsion is cooled and diluted with a naphtha-boiling-range diluent to roughly a 16 API. The diluted emulsion is separated into water and oil in the free water knock-out. The oil phase is dewatered, degassed, cooled, and diluted to meet pipeline specifications. The water phase, however, receives much more attention. The water is first de-oiled via another round of gravity separation and then gas flotation. In many cases, the de-oiled water is further filtered to remove additional oil. Once the majority of the free oil is removed from the water, the water is treated to reduce hardness, silica and some of the dissolved organics. The treated water is then reheated, pumped to high pressure, and fed to the boilers to produce high-pressure steam for the SAGD operations. Some water is disposed of to purge contaminants. Produced gas is recovered, compressed, and used for fuel. A glycol system and direct heat exchange are used to transfer heat from streams that are cooled to those that are heated.

Water requirements are typically 2–4 times the volume of oil. Most of the water is recycled but some make-up water is required to offset losses to the reservoir and disposal.

In the base case SAGD CPF, the flue gas is cooled against both incoming boiler feed water and combustion air before being vented at approximately 130˚C. For the purpose of integrating the MCFC+CO_2 system in Case 1b, the produced gas was desulphurized, which allowed for the flue gas to be cooled further to 95˚C via air preheat without sulphidic water condensation.

Group 1

The objective of the Study is to provide a comparison for the Group 1 cases, which is Case 1a (Base Case SAGD with WLS + OTSG) with Case 1b (Base Case + MCFC+CO_2). In Case 1b, the intent is to provide the most cost effective integration option of the MCFC+CO_2 with the SAGD CPF.

Case 1a, which serves as the reference case, employs WLS for water treatment and OTSGs for steam generation. Electricity is imported from the grid. Case 1b employs WLS for water treatment and OTSGs for steam generation. The MCFC is integrated with the SAGD facility to remove and concentrate the CO_2 from the OTSG flue gas. The MCFC generates power, heat, and water. Following the removal of the CO_2, it is purified and compressed to Kinder Morgan specs. Details of the MCFC+CO_2 integration and process description follow in this section.

The Conceptual Flow Diagrams ("CFD") for Cases 1a and 1b are shown in Figures C-1 and C-2, respectively.

Figure C-1.
Case 1a (SAGD with WLS + OTSG)

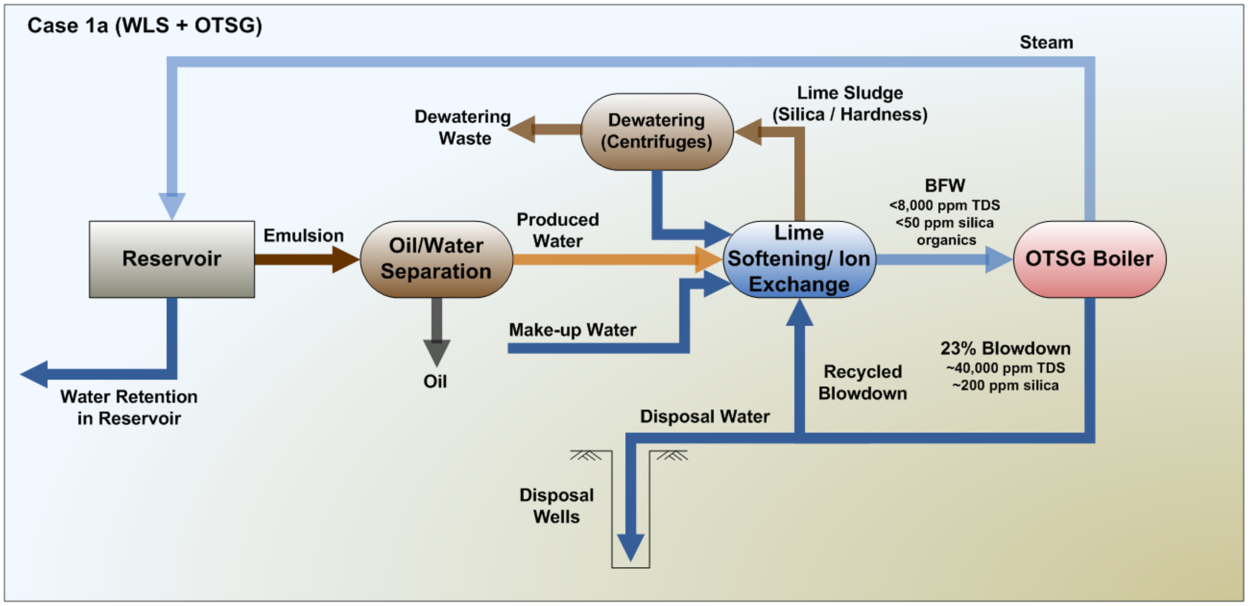
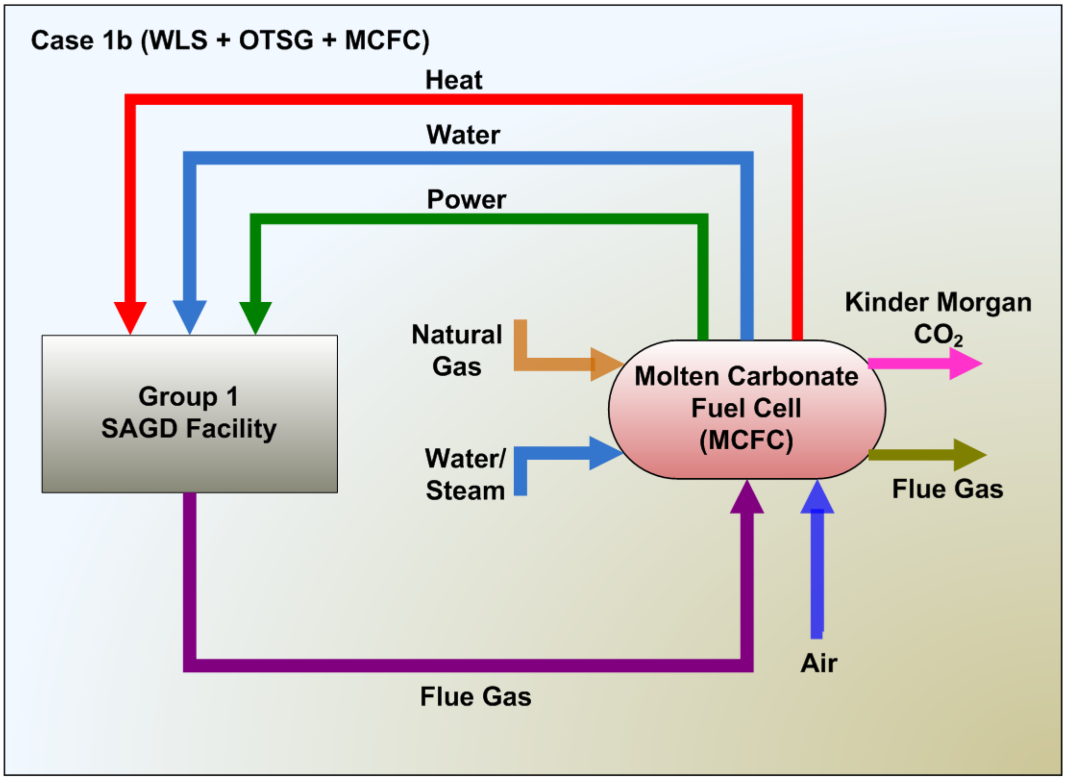


Figure C-2.
Case 1b (WLS + OTSG + MCFC+CO_2)



Groups 2 and 3

A qualitative discussion of integrating the MCFC+CO_2 facility in two other SAGD configurations, based on our analysis of Group 1, is provided in Section E of this report.

The Group 2 SAGD CPF configuration uses an evaporator for water treatment and a drum boiler for steam generation. As in Case 1a, power is imported from the grid.

The Group 3 SAGD CPF configuration uses WLS as the water treatment technology, just as in Group 1. In Group 3, 50 percent of the steam is produced by heat recovery steam generators (HRSGs) associated with a cogeneration unit and 50 percent from OTSGs. The main difference is that in Group 3, the SAGD CPF is already a power exporter due to the large amount of power produced via the cogeneration gas turbine generators.

The SAGD CPF configurations for Groups 2 and 3 are represented in Figures C-3 and C-4, respectively.

Figure C-3.
Case 2a (Evaporator + Drum Boiler)

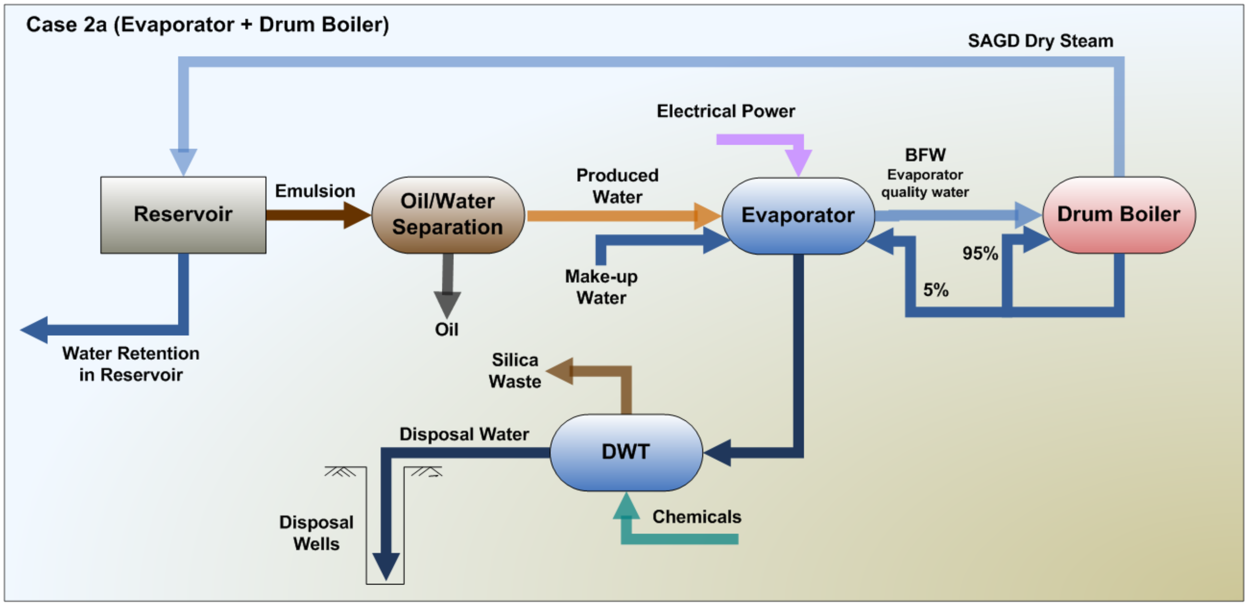
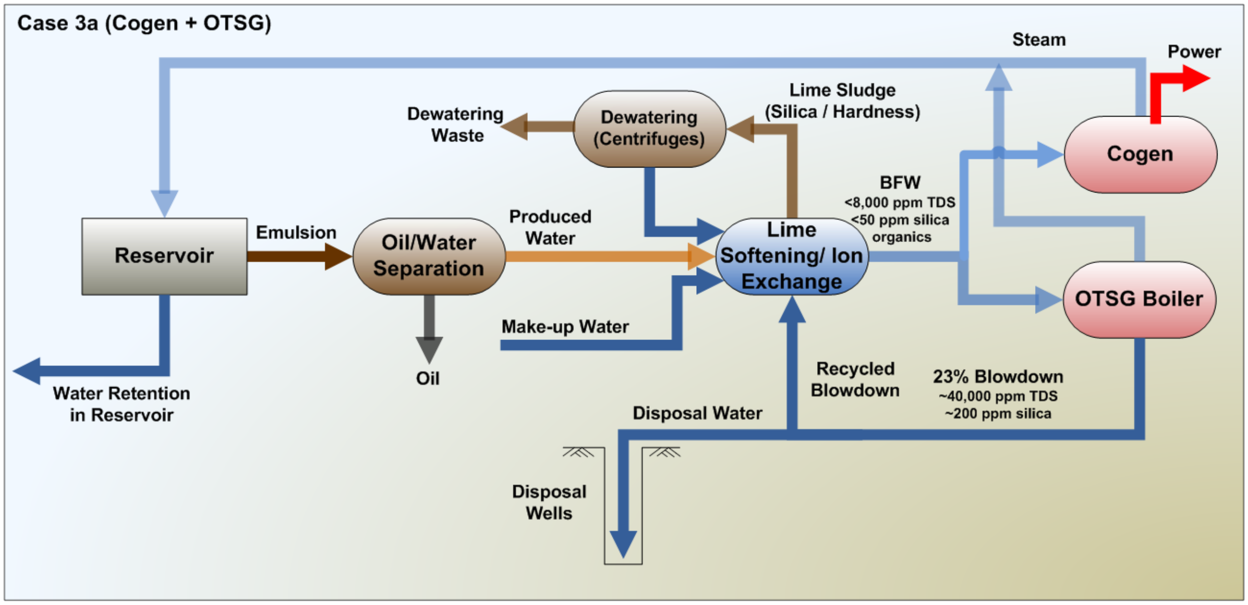


Figure C-4.
Case 3a (WLS + Cogen + OTSG)



MCFC and CO_2 Capture and Compression Process Description

Plant Configuration

The CO_2 recovery section of the plant consists of three primary sections:

- Flue gas collection and cooling—This system collects the flue gas from multiple OTSGs, boosts the pressure, and cools the flue gas using a direct contact cooler (“DCC”) to condense excess water contained in the flue gas, and then distributes the flue gas to the power generation units. The configuration used for the DCC (i.e., single or multiple trains) is dependent on vessel size and transportation limits for the project. As the fuel gas to the OTSG has been desulphurised and the flue is essentially sulphur free, no flue gas scrubbing for SO_2 is required.
- Power Generation and CO_2 removal—This section consists of molten carbonate fuel cells configured for carbon. The OTSG flue gas is fed to the cathode side of the ECM, and the CO_2 in the flue gas migrates across the membrane generating power and concentrating the CO_2 in the Anode discharge stream.
- CO_2 purification and compression—The CO_2 is recovered from the anode discharge by compressing and cooling the stream to first condense water and dry the stream and then to liquefy the CO_2. The non-condensable gases—primarily unconverted fuel and H_2—are then separated from the CO_2 by simple gas-liquid separation and recycled to the fuel cell. The liquid CO_2 is pumped to pipeline conditions and exported. To maximize efficiency and take advantage of low level heat in the MCFC, an ammonia adsorption refrigeration system is used to cool and liquefy the CO_2.

As stated earlier, Fuel Cell Energy (“FCE”) calls the power generation and CO_2 purification and compression system *Combined Electric Power and Carbon Dioxide Separation* (“CEPACS”). FCE also markets its MCFC as an Electrochemical Membrane (“ECM”). For the purposes of this Study we have used CEPACS to represent the MCFC+CO_2 system.

Unlike previous studies, in which FCE proposed using standard 1.5 MW fuel cell modules containing 4 individual stacks, FCE is now proposing to house up to 200 individual stacks in a single purpose-built building to simplify distribution piping and reduce required plot space. There is a single train of heat integration and support equipment for the stacks contained in each building. In addition to the ECM buildings, one additional building is required to house the refrigeration, CO_2 compression, and liquefaction equipment.

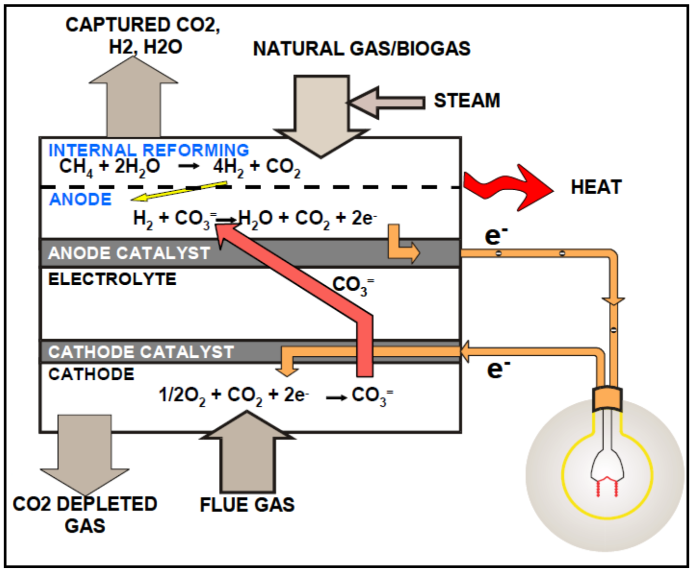
Process Description

The flue gas at 95°C from the OTSGs is collected in a common feed header and fed to a flue gas blower that boosts the pressure of the flue gas to overcome the pressure drops of the direct contact cooler and the MCFC system. The flue gas then passes through the DCC where the flue gas is cooled to about 40°C against a circulating water stream. As the flue gas is cooled below its dew point, water vapour will condense out of the flue gas and enter the circulating water stream. Excess water from the MCFC+CO_2 unit is used as makeup to the top stage of the DCC to provide additional cooling of the flue. The heat picked up by the circulating water stream is rejected via an air cooler. The excess water from the DCC at 62°C is discharged and used as makeup to the OTSG boiler feed water system.

The MCFC+CO_2 unit is installed downstream of the DCC system, where the flue gas is routed to the cathode side of the MCFC. The MCFC unit separates the CO_2 from the other components in the flue gas, and concentrates it in the anode offgas stream in addition to the CO_2 generated by the fuel fed to the MCFC. The anode offgas stream is then compressed, dried, and cooled to condense the CO_2 which is separated from the non-condensable gases—principally unconverted fuel and hydrogen. The non-condensable gases are recycled to the anode and cathode streams. The liquid CO_2 is then pumped and reheated to export conditions.

Figure C-5 below provides a representation of the MCFC with the flows and reactions that occur.

Figure C-5.
MCFC



The flue gas and air streams are preheated first by heat exchange with the cathode and anode offgas before combining them and then catalytically oxidising a portion of the non-condensables (predominantly unreacted methane and H_2) separated from the anode offgas, in the air and flue gas stream. This also increases the overall CO_2 recovery.

At the cathode, the CO_2 in the flue gas reacts with the oxygen and two electrons from the external circuit according to the reaction:

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

The carbonate ions migrate across the electrolyte to the anode driven by the electrochemical potential.

The CO_2-depleted flue gas exits the cathode and is cooled by heat exchange with the anode feed streams, used to raise the steam used at the anode, preheat the flue gas feed, and to regenerate the refrigerant for the absorption chillers used in the purification unit. The cooled cathode offgas is then vented.

Natural gas is desulphurised, preheated against the cathode offgas, and mixed with steam before being fed to the anode side of the MCFC along with the remaining portion of the inerts stream. There it passes over a catalyst that promotes the steam methane reforming reaction, converting the methane (as well as the higher hydrocarbons present in the natural gas) to carbon monoxide and hydrogen according to the equation:

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

The same catalyst promotes the water gas shift reaction converting the produced carbon monoxide to carbon dioxide and more hydrogen according to the equation:

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

These are equilibrium reactions, and some methane and CO will be present in the anode offgas stream.

At the anode, the hydrogen generated by the steam reforming and water gas shift reactions reduces carbonate ions to CO_2 and water and releases the two electrons to the external circuit, according to the equation:

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

The anode offgas, which still contains some unreacted natural gas, CO, and H2, is cooled by preheating the incoming air and passed over a low temperature shift catalyst to convert the remaining CO to CO_2, and producing additional hydrogen, according to the equation:

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

The outlet stream from the low temperature shift reactor is cooled to ambient temperature by heat exchange with cathode air, product CO_2, heat integration with the OTSG, or excess water heating and heat rejection via air cooling and an ammonia chiller. The water in the stream is condensed. Most is reused within the process; the excess is used in the DCC to provide additional cooling before being exported with the DCC excess condensate for use in the OTSG boiler.

The raw gaseous CO_2 product stream is compressed in multiple stages to about 20 bar, cooled with refrigerant, dried, and then refrigerated to condense most of the CO_2. The product CO_2 is separated from the impurities (predominantly methane, hydrogen, and nitrogen) by simple gas liquid separation. The liquid CO_2 is pumped to discharge pressure and warmed to ambient temperature. The gas stream containing the ‘impurities’ is flashed to low pressure and then heated against the incoming compressed CO_2 before being split. Part is recycled to the anode inlet, and the remainder is used to preheat the inlet to the cathode in a catalytic oxidiser.

Section D.

Results

Power, Water, and Heat Integration

The integration of the MCFC+CO_2 facility into a WLS+OTSG-based SAGD complex was guided by the overarching basis, assumptions, and definitions as follows:

- Integration is the use of heat, water, and power from MCFC for the purposes of reducing operating costs on the SAGD, which in turn will reduce the amount of flue gas produced, and, therefore, reduce the amount of CO_2 which must be concentrated, purified, and compressed by the MCFC+CO_2 facility.
- There are three integration streams to consider: power, heat, and water.
- The capital cost in Alberta is a dominating factor that leads to efforts to reduce the size of the MCFC, by reducing flue gas production, minimizing equipment count for integration, and understanding that the direct costs associated with bulk items such as piping, concrete, and electric cabling are material. The cost of the MCFC+CO_2 is significant for the integrated case; therefore, reducing the amount of flue gas produced by the SAGD OTSG has a measurable impact on the overall cost of carbon capture.
- The flue gas has to be cooled to reduce H_2O partial pressure and enhance the driving force for the transfer of ions across the molten carbonate membrane.
- The infrastructure required to support the SAGD CPF is the same between SAGD and SAGD + MCFC+CO_2 CPF and therefore is not considered material for the comparison of Case 1a and Case 1b.
- Remaining surface facilities and sub-surface costs are unchanged.

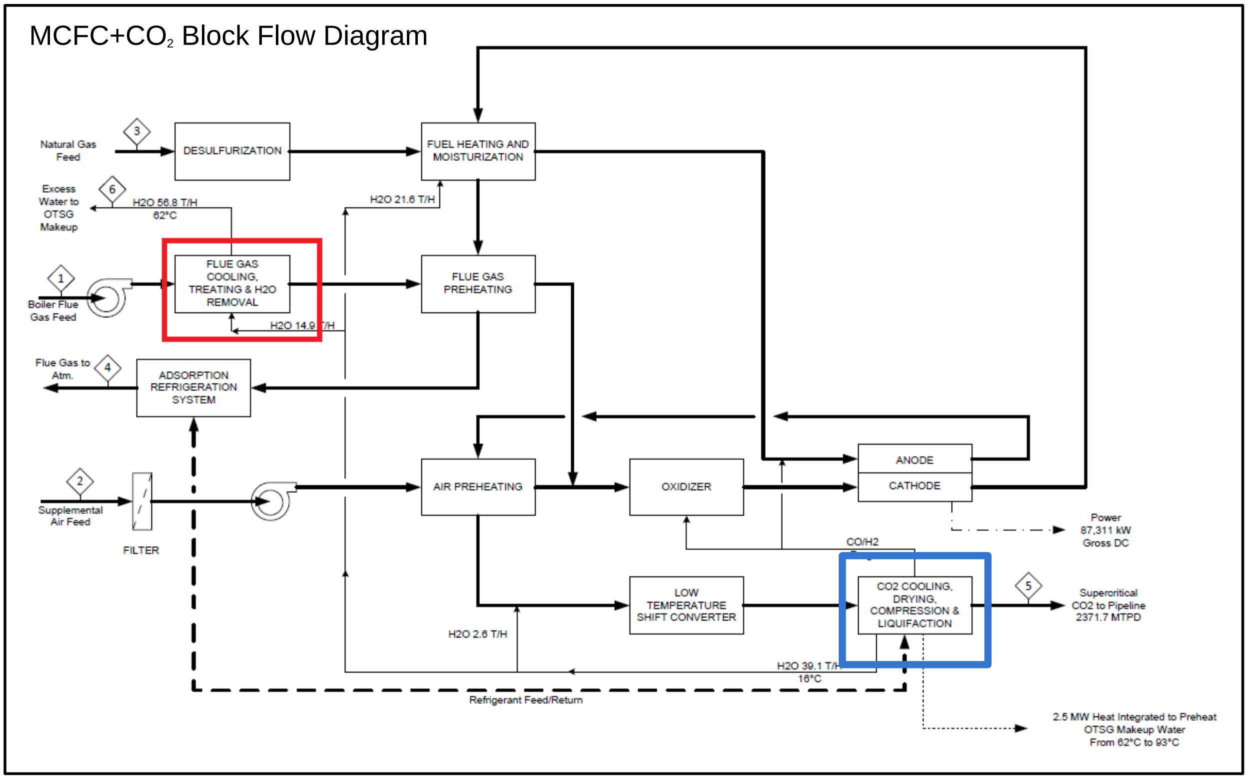
In brainstorming the integration options, Jacobs Consultancy found it helpful to consider power, water, and heat individually. Our thought process for each was as follows:

- Power—Integrating the power from the MCFC was straightforward and could be completed without materially impacting the overall capital costs. In summary, a total of 87 MW is generated by the MCFC. Power uses are as follows: 19 MW for CO_2 purification and compression and 14 MW for the SAGD operations. This leaves 51.4 MW available for export (assuming 5 percent transmission losses).
- Water—Integrating water was straightforward. Water generated in the MCFC+CO_2 area totals about 57 tonnes/hr of relatively clean water, with approximately 40 tonnes/hr being generated by the flue gas condensation and 17 tonnes/hr from the MCFC operation. The entire water stream can be used by the SAGD facility and backs out brackish make-up water. It was assumed that the cost of piping the water from the MCFC+CO_2 area to the SAGD area was offset by the reduction in make-up water wells, piping, and treatment.

- Heat—Integrating heat from the MCFC facility into the SAGD facility was challenging. The MCFC+CO_2 as configured for the Study is highly heat integrated, and the remaining heat available for export is limited to a temperature of 139°C with only about 2.5 MW available down to 80˚C.

As shown in the blue box in Figure D-1, the option for recovering heat revolved around the CO_2 purification and compression area where the concentrated CO_2 is cooled and compressed. The red area highlights where the water is combined and produced via flue gas condensation as well as the MCFC operation.

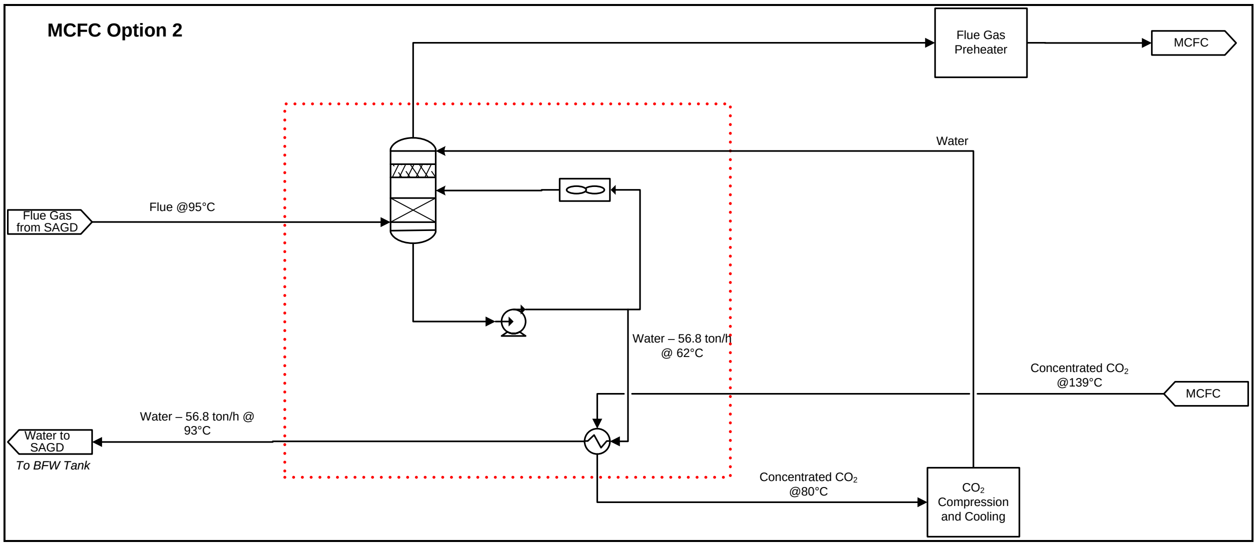
Figure D-1.
MCFC Block Flow Diagram



We identified three options for recovering heat:

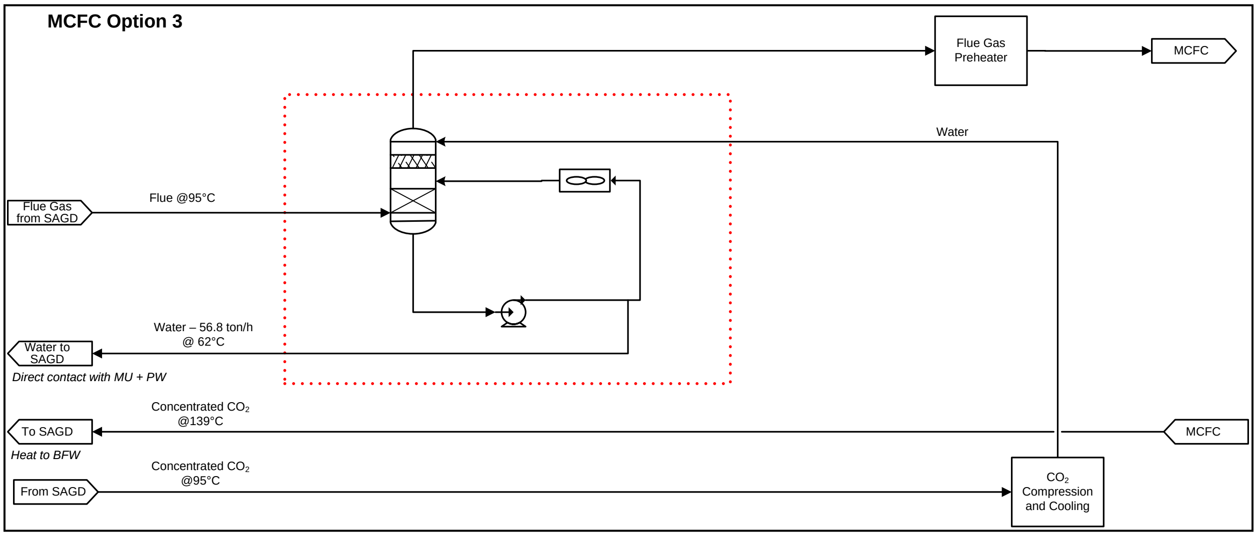
1. No heat integration—reject the 2.5 MW available to the MCFC+CO_2 configuration to the onsite air cooler.
2. Heat produced water—use the heat available to heat the 57 tonnes/hr of water to 93°C and inject the water into the BFW tank. This option is illustrated in Figure D-2 below. Heat from the CO_2 is used to preheat water from the DCC area.

Figure D-2.
Option 2 (Heat Produced Water)



3. Heat BFW—integrate the heat via direct exchange with the BFW pre-heat. This is illustrated in Figure D-3. Heat in the CO_2 area is exchanged directly with BFW and water is recovered and injected without preheat as in Option 1.

Figure D-3.
Option 3 (Heat BFW)



We evaluated all three options by incorporating them into our SAGD model. As shown in Table D-1, there was little benefit to using the heat in the SAGD CPF, and certainly not enough to justify any additional capital investment. Therefore, we selected Option 1, in which the heat from the MCFC+CO_2 was rejected via the existing air cooler. In all three cases, the produced gas is treated to remove sulphur to enable water condensation in the flue gas. Therefore, we were able to recover more heat in the OTSG stack gas via air preheat, without running into sulphidic corrosion issues.

Table D-1.
Integration Options

Group 1 Options	BFW Temp (°C)	Natural Gas Flow (tonne/h)	Glycol Air Cooler Duty (MW)	Stack Temp (°C)	Integration Equipment
Base Case (WLS+OTSG)	189.4	29.8	8.0	130.0	
Option 1- Integrate power, water but no heat integration	189.5	29.3	11.8	95.6	Desulfurizer 1 pipe
Option 2 - Integrate power, water and heat 57t/h to 93°C with 139°C process gas and route to BFW Tank	189.9	29.3	13.3	95.6	Desulfurizer 1 pipe 1 exchanger
Option 3 - Integrate heat via direct exchange with BFW	179.7	29.3	12.5	95.6	Desulfurizer 3 pipes 1 exchanger

Figures D-4 and D-5 illustrate the configuration of Option 1. At the MCFC+CO_2 facility, the water is recovered via direct contact with the flue gas and routed to the SAGD CPF in the place of cool make-up water. Make-up water can be added to the SAGD CPF in various places. Two of the most common configurations are to inject the water into the produced water upstream of deoiling or between deoiling and water treatment. The advantage of adding the water upstream of deoiling is that it can assist in cooling the produced water, reducing the size of the produce water / glycol exchanger. The disadvantage is that the make-up water consumes a small amount of volumetric capacity in the deoiling circuit. Alternatively, the make-up water can be added between deoiling and water treatment. The advantage of this location is that all the volumetric capacity in the deoiling section is reserved for produced water. Practically speaking, there is little difference in performance and capital cost, and regardless of the injection point, the water from the MCFC+CO_2 facility can be used as a one-for-one replacement of brackish make-up water. A side benefit is the water that is recovered is of better quality than brackish make-up water and directionally reduces chemical consumption in water treatment, even though the volume of water is small relative to the total volume of water treated.

Figure D-4.
DCC Configuration for Option 1

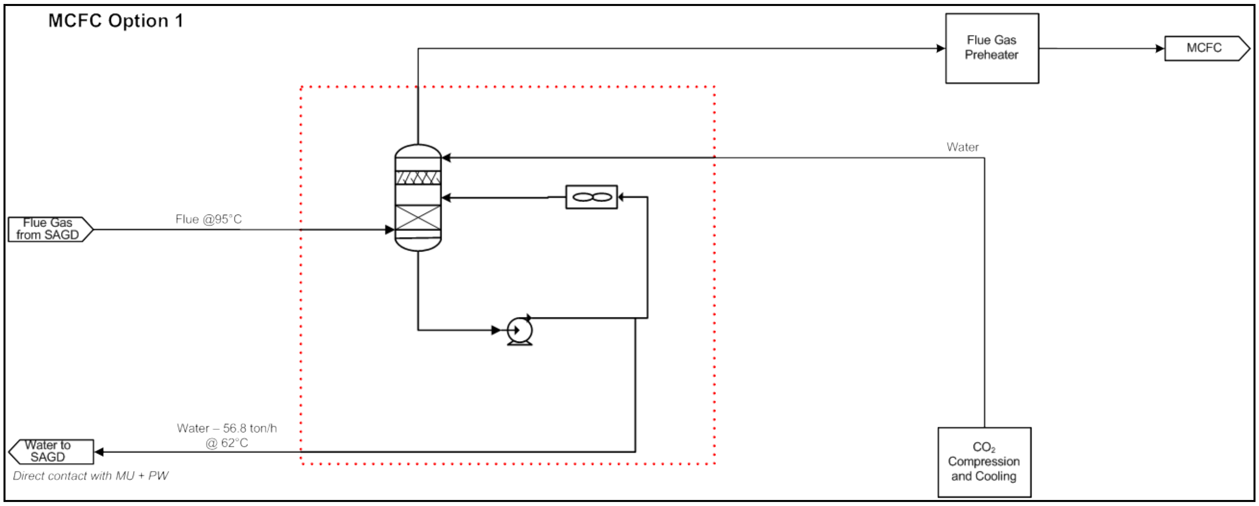
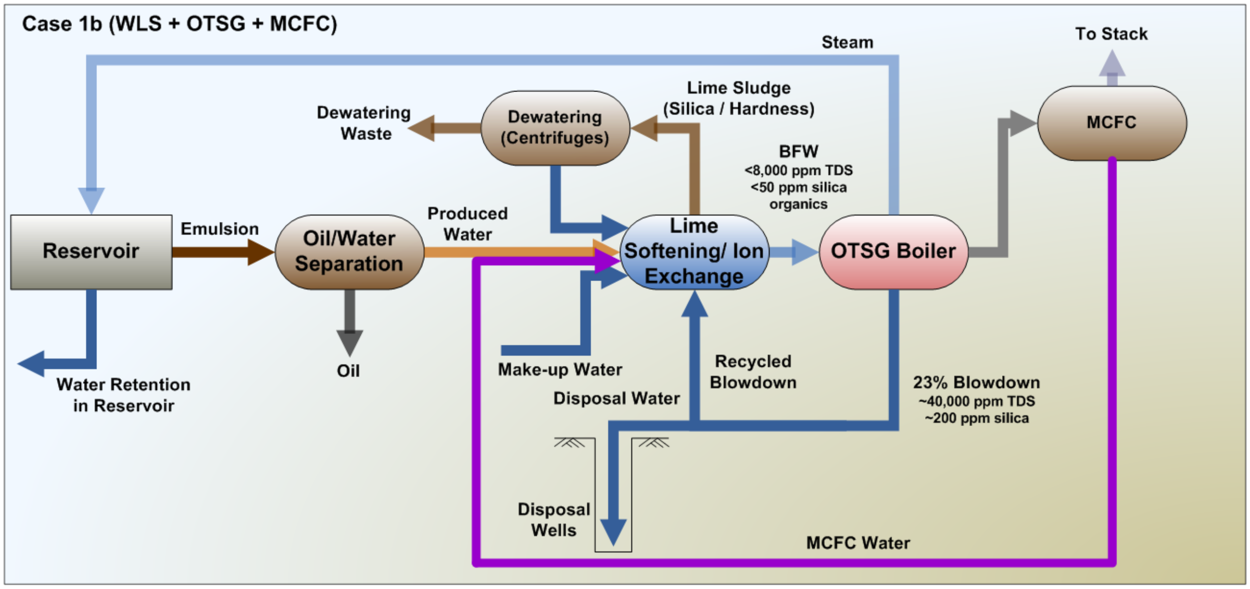


Figure D-5.
MCFC Water Routing into SAGD Facility



Figures D-6 and D-7 below summarize the impact that the integrated MCFC+CO_2 has on the overall CPF balance in terms of utilities and waste streams. As discussed, natural gas increases due to the consumption of fuel by the MCFC; make-up water decreases due to recovery of flue gas water and water produced by the MCFC; power is exported; and a concentrated and compressed CO_2 stream is produced. Most importantly, for the purposes of this Study, Direct

CO_2 emissions are reduced from 81 tonnes/hr to 10 tonnes/hr. There may be additional GHG credits due to the export of power, that in theory reduce the GHG emissions related to power production for the Alberta grid.

Figure D-6.
Case 1a Overall Balance

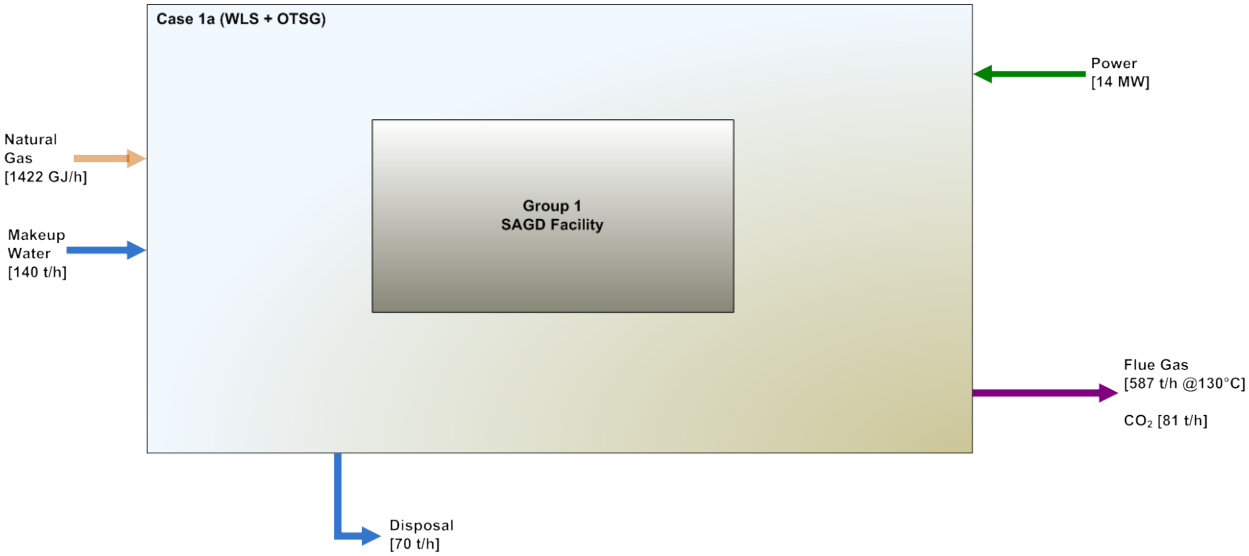
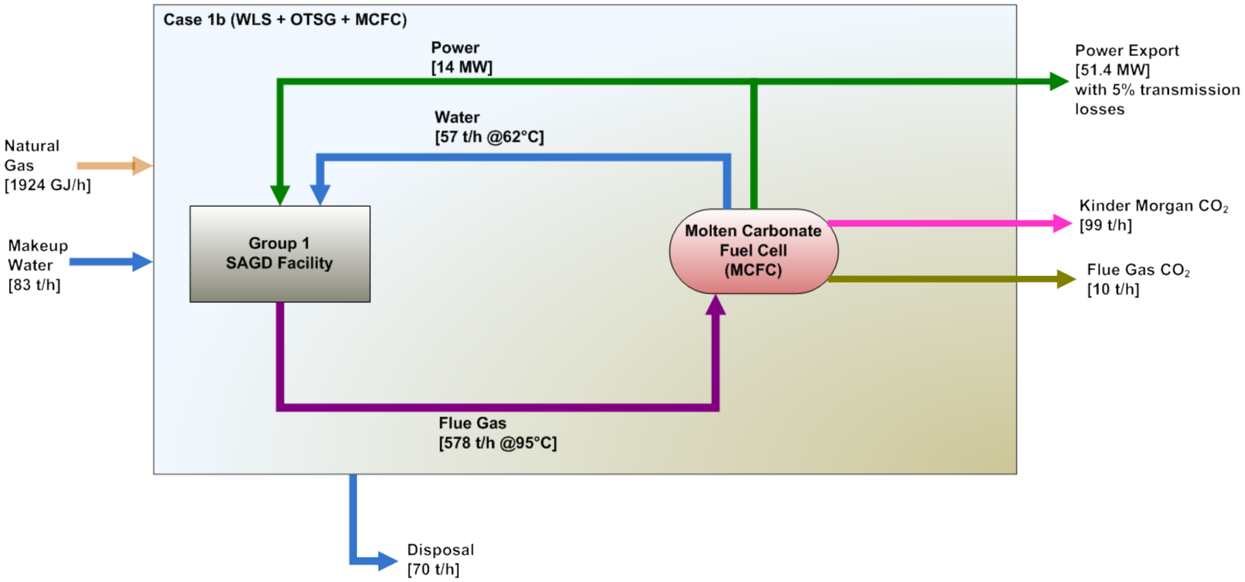


Figure D-7.
Case 1b Overall Balance



Performance Results

One of Jacobs Consultancy's objectives was to assess the integration of the MCFC+CO_2 facility into a SAGD complex by comparing the performance in terms of the following performance metrics:

- Energy efficiency
- Make-up water intensity
- Direct / indirect CO_2 emitted and captured
- Natural gas usage
- Power
- Capital and operating costs

Table D-2 summarizes the impact of integrating the MCFC+CO_2 facility into the SAGD CPF in terms of key metrics per barrel of bitumen (sometimes called intensity).

The first row provides an overall energy intensity which includes natural gas consumed as well as power imported or exported at 38 percent efficiency (equivalent to a heat rate of 9,000 BTU/kWhr). The integration of the MCFC+CO_2 reduces energy intensity because the power produced by the MCFC is more efficient than the 38 percent assumed, even after including the energy for CO_2 purification and compression.

Make-up water intensity shows a 40 percent improvement in Case 1b because of the integration of water produced by the flue gas condensation on the MCFC operation.

Table D-2.
Group 1 Cases Comparison

	Case 1a - SAGD (WLS + OTSG)	**Case 1b - SAGD (WLS + OTSG) + MCFC+CO_2**
Energy intensity *(net energy purchased/exported per bbl of bitumen produced)*	1.13 GJ/bbl	1.02 GJ/bbl
Make-up water intensity *(volume of water per bbl bitumen produced)*	0.10 m^3/bbl	0.06 m^3/bbl

Total GHG Emissions

For the purposes of this Study, the reason to integrate the MCFC+CO_2 facility into a SAGD facility is to reduce the amount of dilute CO_2 emitted to the atmosphere. As shown in Figure D-8, the technology reduces atmospheric admissions by 88 percent, not including a GHG credit for exporting power. It should be noted that total CO_2 generated by the integration of the MCFC+CO_2 facility is increased by 35 percent due to the natural gas consumed by the MCFC.

Figure D-8.
Total GHG Emissions

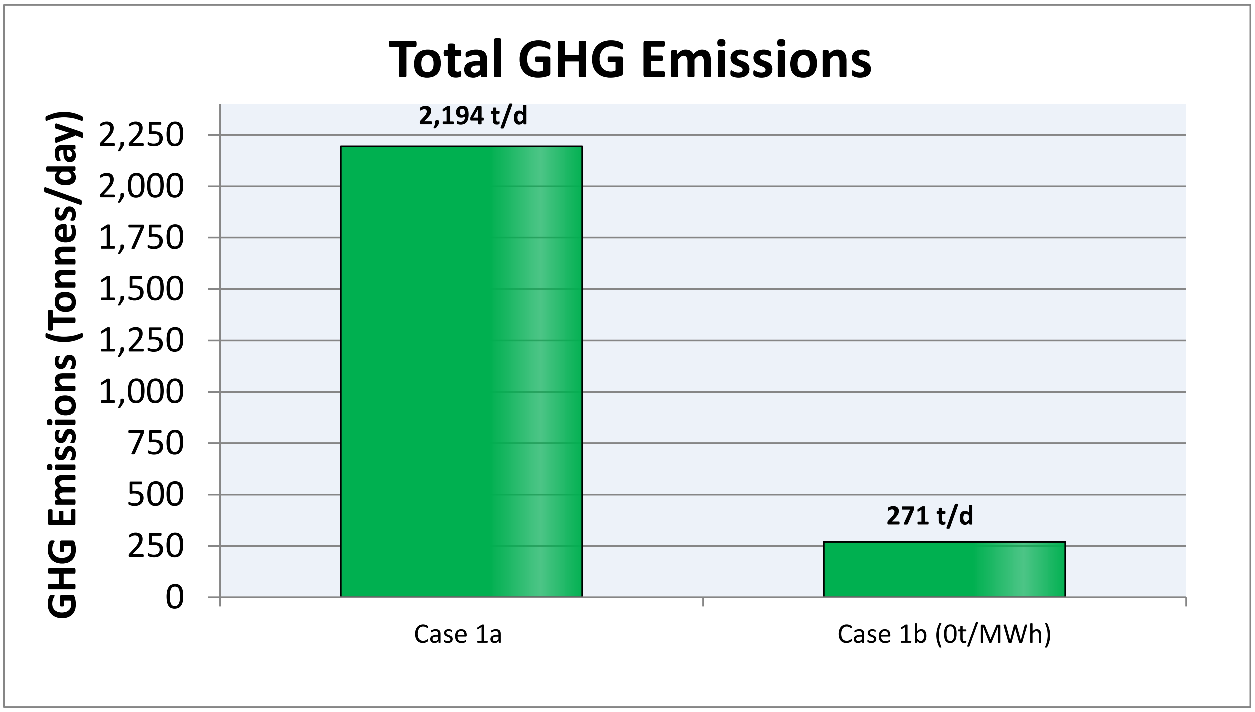


Table D-3 below summarizes the direct and indirect CO_2 emissions for Case 1b assuming a CO_2 credit for power export ranging from 0.42 tonne/MWhr to 0.88 tonne/MWhr. In all cases, the credit could more than offset the emissions of the SAGD facility, which makes the MCFC+CO_2 one of the few technologies that avoid more GHG emissions than the amount captured.

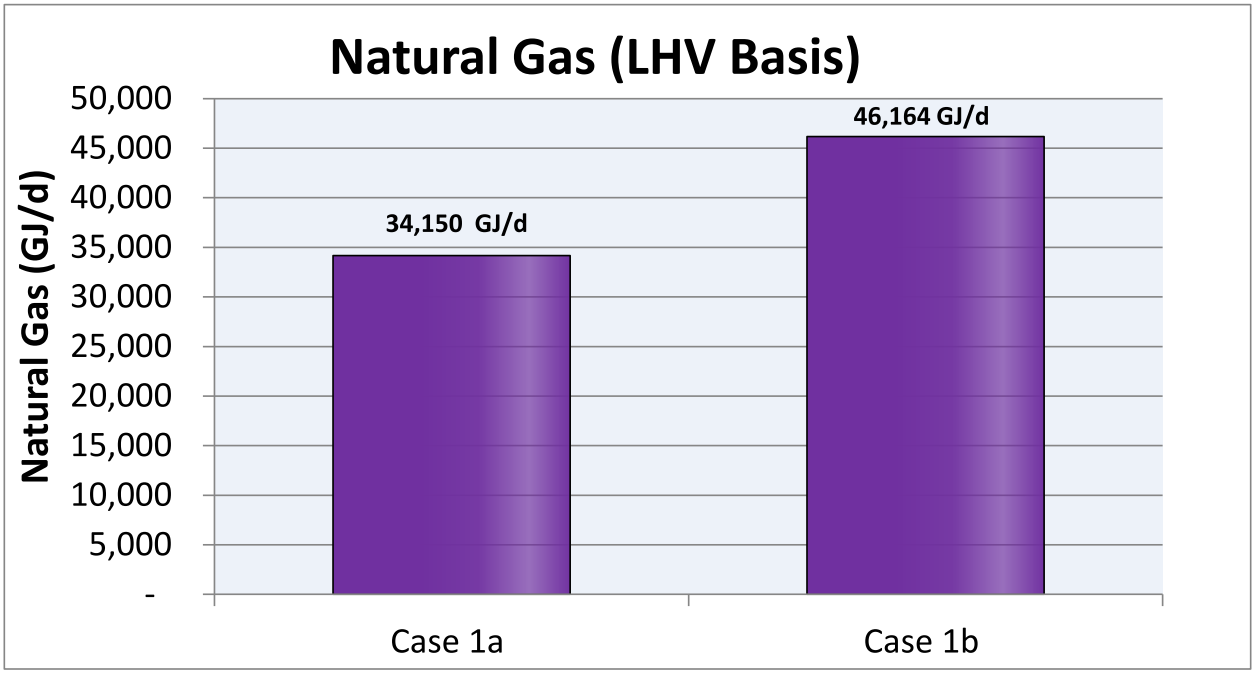
Table D-3.
Case 1b – Total GHG Emissions

	Direct CO_2 (MT/day)	Indirect CO_2 (MT/day)	Total CO_2 (MT/day)
0 tonne of CO_2/MWh	251	20	271
0.42 tonne of CO_2/MWh	251	-498	-247
0.66 tonne of CO_2/MWh	251	-795	-544
0.88 tonne of CO_2/MWh	251	-1,066	-815

Natural Gas Consumption

In Case 1a, natural gas is consumed by the OTSGs for the purposes of steam generation. The flue gas from the OTSGs was kept to 130°C as discussed above. In Case 1b a stack temperature of 95°C could be achieved due to desulphurization of the produced gas. This resulted in better boiler efficiency with the MCFC+CO_2 in Case 1b. The MCFC, however, consumed 525 GJ/hr of natural gas resulting in a net increase in natural gas consumption as indicated in Figure D-9. After factoring in the power losses for CO_2 purification and compression, the MCFC+CO_2 facility produces power at an efficiency of about 44 percent, which is still much better than a coal-fired power plant operating at roughly 38 percent efficiency without CO_2 capture and compression.

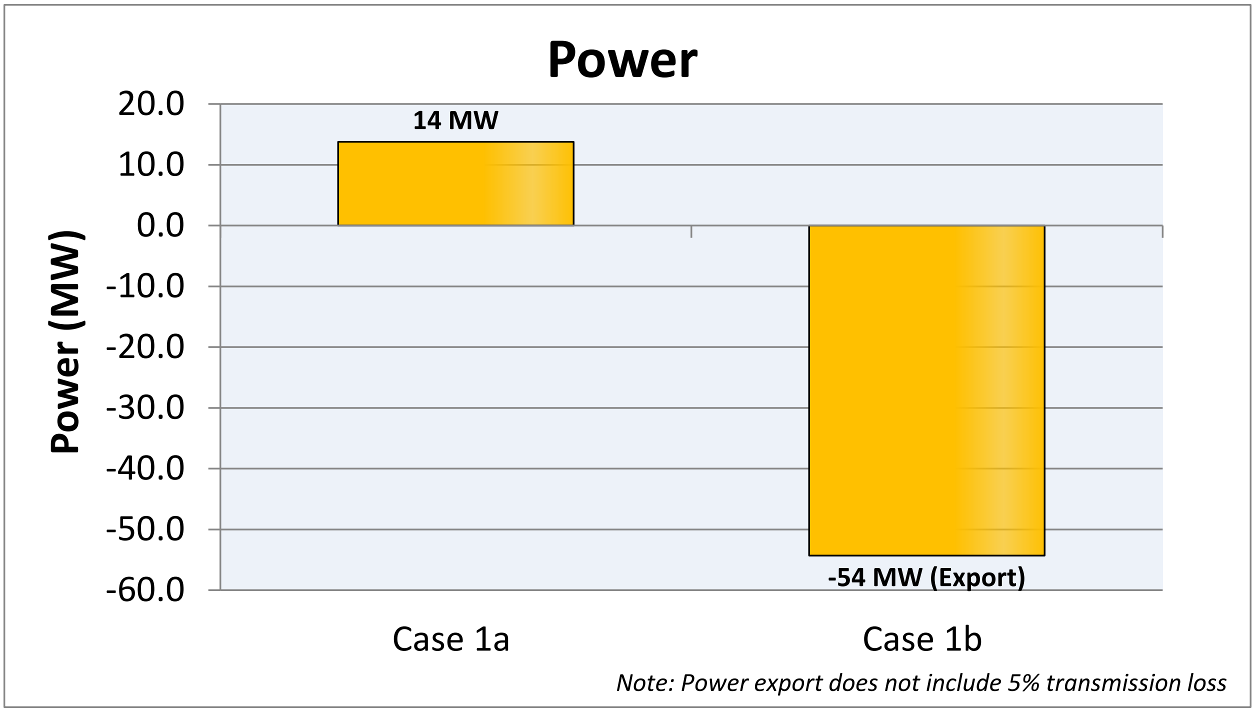
Figure D-9.
Natural Gas Consumption



Power

Case 1a requires approximately 14MW for the SAGD operation including power for the electric submersible pumps (“ESPs”). In Case 1b, the power generated by the MCFC operation is 87 MW. Of that, 19 MW is required for CO_2 purification and compression, 14 MW for the SAGD operation and 5 percent transmission losses. As indicated in Figure D-10, power available is 54 MW and after 5 percent transmission losses becomes 51.4 MW for export in Case 1b.

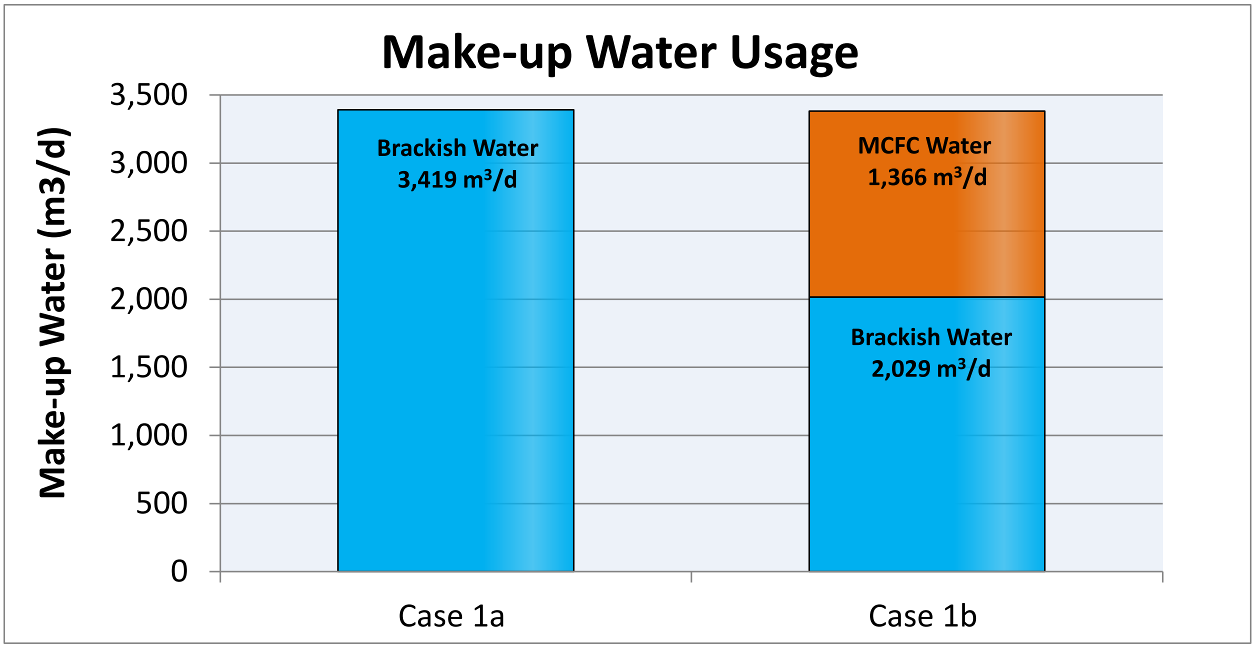
Figure D-10.
Power Import / Export



Make-up Water Usage

The total amount of make-up water needed to offset losses to the reservoir and disposal losses is the same in both Case 1a and 1b because the SAGD steam generation is the same. As indicated in Figure D-10, the water generated by the flue gas condensation and operation of the MCFC in Case 1b offsets about 40 percent of the make-up water requirements via integration of the water.

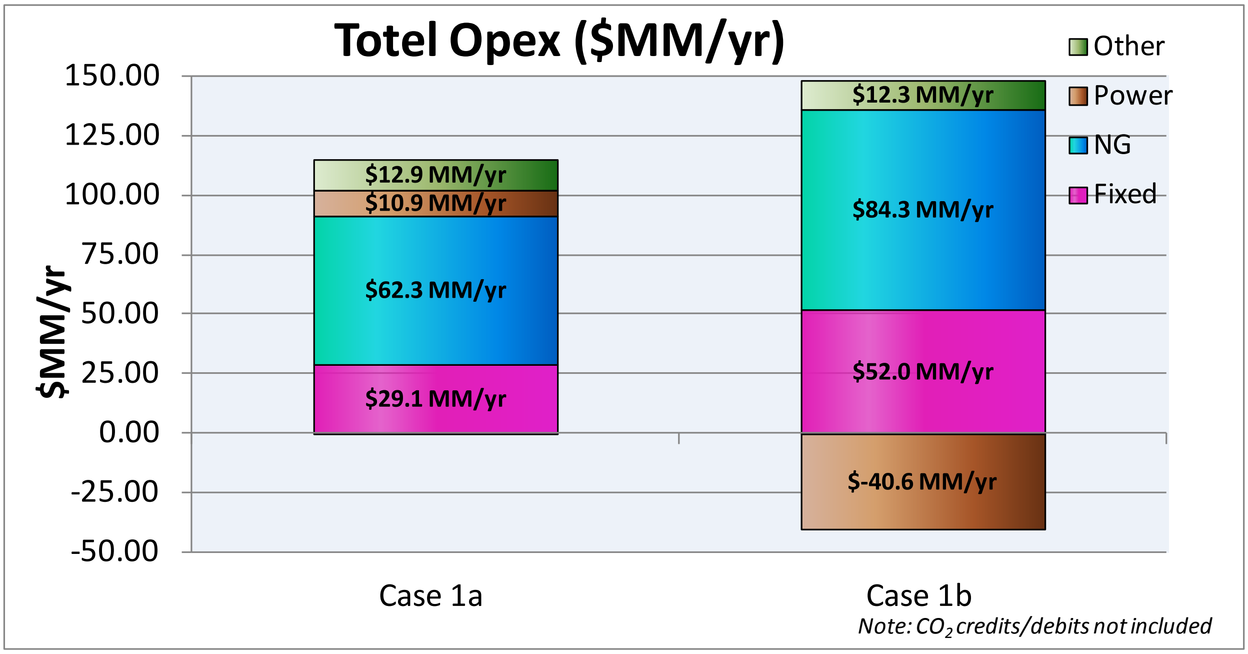
Figure D-11.
Make-up Water Usage



Operating Costs

Operating costs are the sum of the estimated fixed and variable operating costs. The variable operating costs are estimated based on natural gas, power, solids disposal, and oil and water treatment chemicals requirements. The cost basis used can be found in Section B. Fixed operating costs are estimated based on capital costs. For SAGD facilities, historical information indicates that fixed costs run about 3.5 percent of capital costs an annual basis. There is no historical basis for the MCFC+CO_2, but based on estimates provided by FCE and in-house analysis, we estimate that fixed costs for the MCFC+CO_2 are around 5 percent of the capital costs, including the need to periodically replace the fuel cell stacks. Figure D-11 shows the operating cost comparison for Case 1a and Case 1b. In spite of increased fixed operating costs and additional natural gas, Case 1b is estimated to have slightly lower operating costs than Case 1a, due to the sale of exported power. This does not include any GHG credits for the 51.4 MW of exported power.

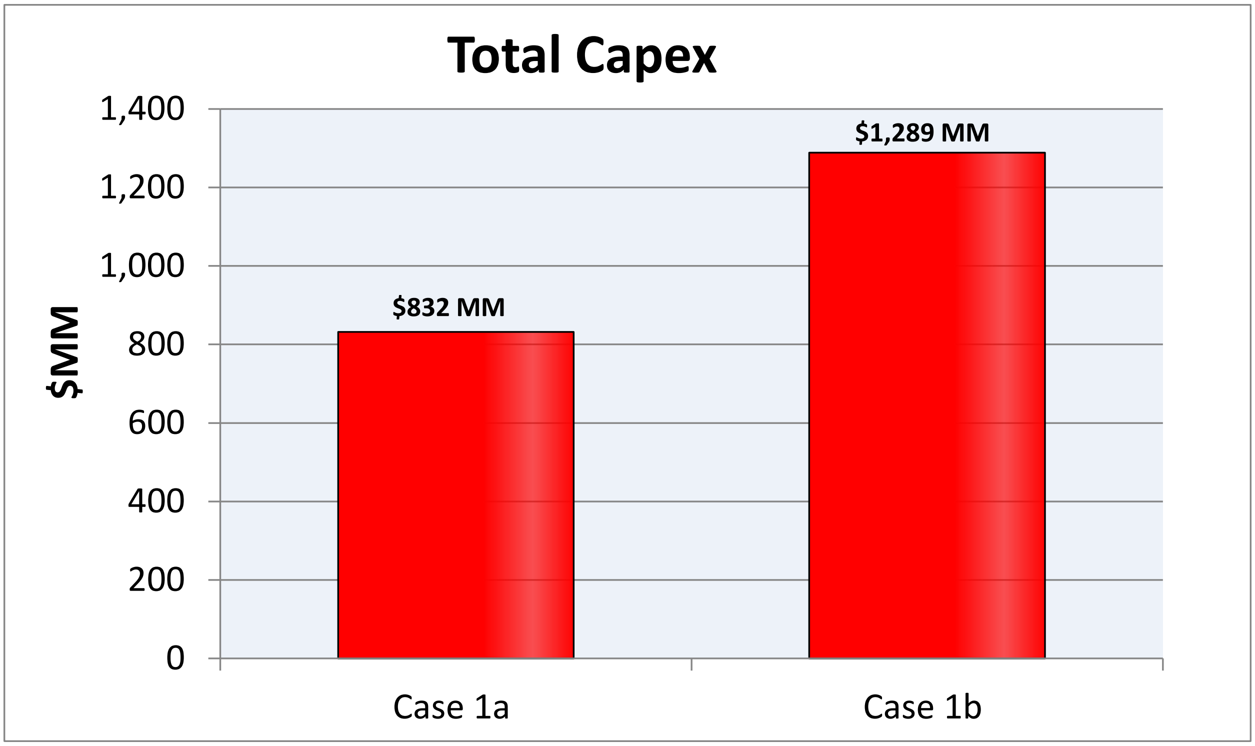
Figure D-12.
Total Operating Costs



Capital Costs

As shown in Figure D-13, estimated capital costs are higher for Case 1b than Case 1a by approximately $450MM, primarily due to the addition of the MCFC+CO_2 unit. Integration costs for the sulphur removal of the produced gas and the water line from the MCFC+CO_2 unit to the SAGD CPF are estimated to be less than $10 MM.

Figure D-13.
Total Capital Costs



Comparison to Phase 1

As indicated in Table D-4, compared to the Phase 1 study, performance results of the integrated SAGD CPF and MCFC+CO_2 of this Study are similar, primarily because we found no advantage to do any heat integration.

Table D-4.
Performance Relative to Phase 1

		Phase 1		Integration	
Calendar Day Rates by Case (Rev 1)		**1a OTSG**	**1c OTSG+ECM**	**1a OTSG**	**1b OTSG+MCFC +CO2**
Makeup Water Rate					
Steam Production for SAGD	**m3/day**	3,375	3,375	3,392	3,382
PCC	**m3/day**		0	0	0
ECM/SOFC Water Produced	**m3/day**		-1,488	0	(1,366)
Total	**m3/day**	3,375	1,887	3,392	2,016
Water Treating Chemicals					
Lime	**tonnes/day**	6.9	6.3	8.37	6.79
MagOx	**tonnes/day**	2.6	2.6	2.27	2.27
NaOH				0.85	0.69
Soda Ash	**tonnes/day**	0.8	0.7	0.00	0.00
Total Water Treating Chemicals	**CAD/day**	15,841	15,049	15,949	14,876
Produced Gas Consumed					
Bitumen Production (LHV)	**GJ/day**	898	898	821	821
Bitumen Production (HHV)	**GJ/day**	1,002	1,002	911	910
Natural Gas Consumed					
Bitumen Production (LHV)	**GJ/day**	30,712	30,712	34,151	33,572
Bitumen Production (HHV)	**GJ/day**	34,073	34,073	37,887	37,244
ECM (LHV)	**GJ/day**		10,946	0	12,602
ECM (HHV)	**GJ/day**		12,145	0	13,968
Total (LHV)	**GJ/day**	30,712	41,658	34,151	46,174
Total (HHV)	**GJ/day**	34,073	46,218	37,887	51,212
Power Consumed					
Bitumen Production	**MW**	14	14	14	14
Carbon Capture	**MW**				
CO2 Compression	**MW**	0			16
ECM	**MW**		11		
Inverter losses			3		3.5
Total Power Consumed	**MW**	14	28		33
Power Generated	**MW**	0	76	0	87
Power Imported	**MW**	14	0	14	0
Power Exported	**MW**	0	48	0	51
CO2 Direct					
Steam Production for SAGD	**tonnes/day**	1,765	1,765	1,953	1,920
Carbon Capture Steam Generation	**tonnes/day**				
ECM Operation	**tonnes/day**		608		703
CO2 Direct Total	**tonnes/day**	1,765	2,373	1,953	2,623
CO2 Indirect	**tonnes/day**				
Without Power	**tonnes/day**	5	5	5	5
CO2 Captured	**tonnes/day**	0	2,205		2372
CO2 Generated (Direct)	**tonnes/day**	1,765	168	1,953	2,623
% CO2 Capture (Direct Only)		0%	93%	0%	90%

Section E.

Findings and Recommendations

Findings

In our opinion, integrating an MCFC+CO_2 facility into a SAGD CPF does the following:

- Reduces dilute CO_2 emission by 88 percent even without GHG credits for exporting 51 MW of power to the grid. It should be noted that the total amount of CO_2 produced by integrating the MCFC+CO_2 is 35 percent more than for the SAGD alone, due to the natural gas consumed by the MCFC.
- Increases capital cost by over 50 percent compared to the cost of the SAGD CPF, primarily due to the cost of installing the MCFC+CO_2 facility. Integration costs by themselves are less than $10 MM.
- Reduces operating costs by 6 percent assuming that the power can be sold to the grid at $90/MWhr not including GHG credits
- Reduces make-up water requirements by 40 percent and eliminates the need for purchased power for the SAGD CPF.
- Is enabled by the uncomplicated nature of integrating power and water. Integrating the heat from the MCFC+CO_2 in the SAGD CPF has little benefit.

Integrating the MCFC+CO_2 represented in the Study as the Combined Electric Power and Carbon Dioxide Separation ("CEPACS") by Fuel Cell Energy into a SAGD CPF is economically limited to power and water, and therefore presents very few technical risks. In addition, the MCFC itself is commercialized. The addition MCFC+CO_2 system into a SAGD CPF, however, does present the following primary risks:

- Capital cost: Installing an MCFC+CO_2 of this size has not been done. In addition, there are additional capital cost risks associated with installing a plant of this magnitude in northern Alberta.
- Reliability, operability, and maintenance costs: The MCFC+CO_2 is a completely new operating facility within a SAGD CPF. Like anything new, there are likely to be unforeseen challenges regarding the operation and maintenance of this type of facility.

Qualitative Comments Regarding Group 2 and 3 CPF Configurations

Based on our analysis of the WLS + OTSG SAGD configuration, qualitative comments are provided for Groups 2 and 3. Simplified representations of each are shown in Figures E-1 and E-2.

Figure E-1.
Case 2a Configuration

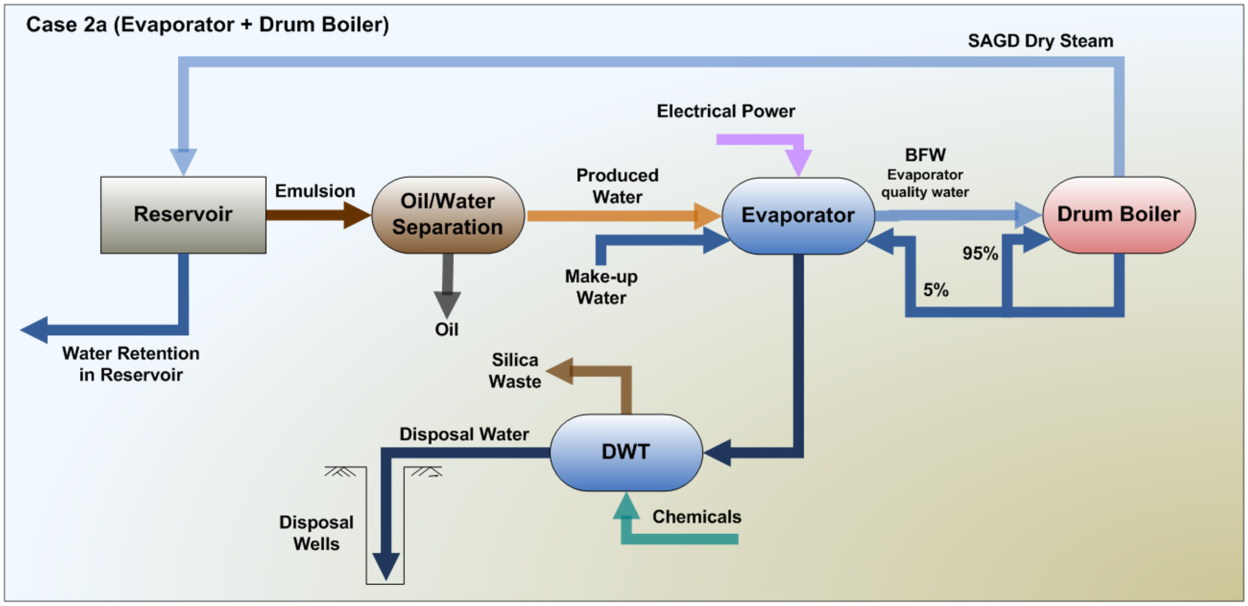
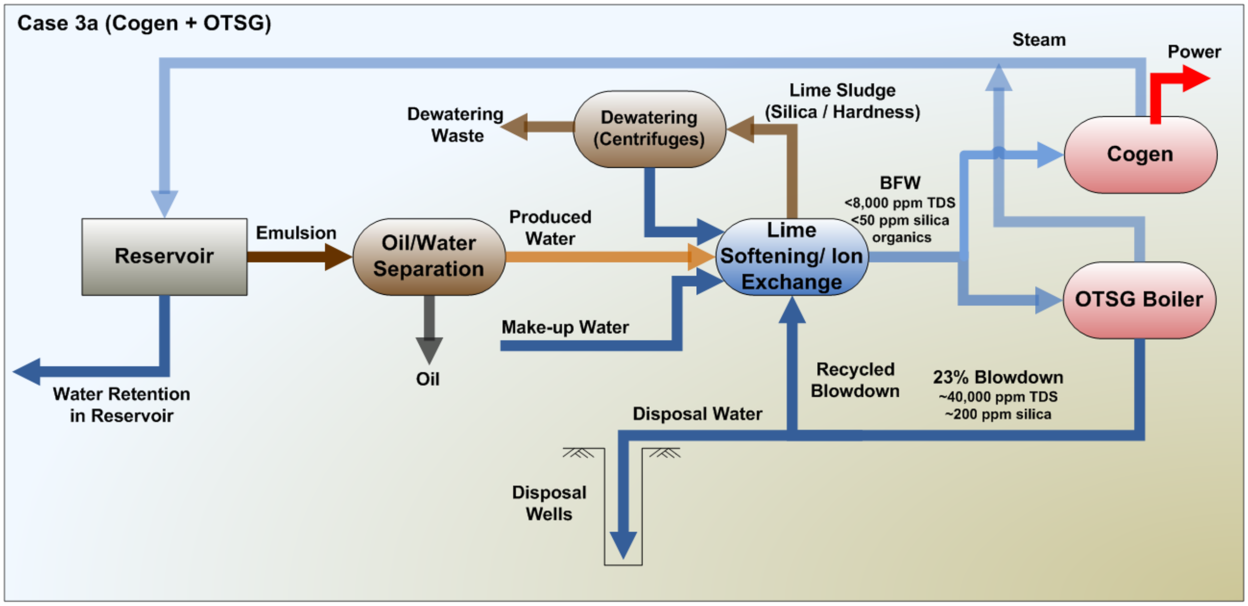


Figure E-2.
Case 3a Configuration



Our comments regarding the integration of the MCFC+CO_2 into an evaporator+drum boiler SAGD CPF represented in Figure E-1 are as follows:

- From a SAGD perspective, drum boilers offer slightly better efficiency than OTSG, so natural gas combustion and therefore flue gas production are lower. Evaporator-based water treatment uses more power than WLS, so the configuration consumes more power. In addition, make-up water is lower because the disposal volume is less. The heat exchange scheme is similar to Group 1.
- The MCFC+CO_2 system is slightly smaller due to decreased flue gas
- Therefore, in our opinion:
  - Power consumption on-site increases and MCFC production is lower, so the amount of power exported goes down.
  - Water requirements from the CPF are lower but with reservoir retentions close to 10 percent, all the MCFC+CO_2 water can be utilized by the CPF and make-up water is lower than in Group 1.
  - There is no change in heat integration. The level of heat produced by the MCFC is too low to be used economically.

Our comments regarding the integration of the MCFC+CO_2 into a cogen + OTSG SAGD CPF represented in Figure E-2 are as follows:

- The SAGD CPF has more flue gas because of the cogen. There will be additional heat, O_2 and NOx in the flue gas. However, make-up water requirements should be the same as Group 1 and the heat integration is similar to Group 1. The biggest change is that the SAGD CPF is a net exporter of power before the addition of the MCFC+$CO_{2.}$
- The MCFC+CO_2 system is larger because the quantity of flue gas and the flue gas cooling requirements increased due to higher temperature flue gas. The fuel cell, however, should be fine with increases in O_2 and NOx.
- Therefore, in our opinion:
  - The site becomes an even larger exporter of power with the addition of the MCFC+CO_2. Infrastructure costs associated with switching from a net importer to net exporter of power may be lower.
  - The complex may be long on water due to increased water from the MCFC+C_{O2}.
  - Similar to Group 1, heat integration is likely to have little value.

Recommendations

In our opinion, this Study supports the findings from Phase 1: the MCFC+CO_2 addition to a SAGD CPF can offer substantially lower costs of capture than commercially available amine systems and is a promising technology for CO_2 capture and compression for the purposes of producing CO_2 for EOR or sequestration. We recommend in the near term:

- Feasibility studies to investigate the benefits of CEPACs for capturing CO_2 from the flue gas of Steam Methane Reformers ("SMR") and smaller fired heaters present in upgraders and refineries.
- A detailed study in conjunction with Fuel Cell Energy focused on the trade-offs of increasing the amount and temperature of the heat generated by the fuel cell for the purposes of reducing natural gas consumption for SAGD.
- A design study to determine location, feasibility, and estimated capital cost of an MCFC+CO_2 demonstration facility in Alberta to confirm reliability, operability, and commercial readiness, and to highlight other development issues.
- A study to investigate the economics and emissions resulting from the co-production of hydrogen with an MCFC+CO_2. The MCFC operating in Orange County, CA is configured to produce and concentrate hydrogen in addition to generating power.

Appendix 1

DBM

AI-EES – Evaluation of Integrating Molten Carbonate Fuel Cell (MCFC) Technology with a SAGD Facility

Design Basis Document

Revision Record

Rev. No.	Date	By	Comments
1	14-April-15	RRH	Draft – for review

1. Introduction

The objective of this study is to determine the benefits of integrating a molten carbonate fuel cell (MCFC) with a Steam Assisted Gravity Drainage (SAGD) central processing facility (CPF) consisting of warm lime softening for the primary water treatment and once through steam generators (OTSG) for steam generation. The purpose of this document is to define the basis of the study.

The main objective of the study is to determine the benefits of integrating MCFC technology with a WLS / OTSG based SAGD facility using the following metrics:

- Energy efficiency
- Makeup water intensity
- Direct / indirect CO_2 emitted and captured
- Capital and operating costs

2. Case Descriptions

As discussed above, the objective is to compare the Base Case (Case A) with the Base Case + MCFC (Case B). Collectively, both cases are referred to as Group 1.

Case A:

- WLS for water treatment; OTSG for steam generation; electricity imported from the grid

Case B:

- WLS for water treatment; OTSG for steam generation; power, heat and water generated by MCFC. MCFC is integrated with SAGD system and used to concentrate OTSG CO_2

The Conceptual Flow Diagrams for Cases A and B are shown below.

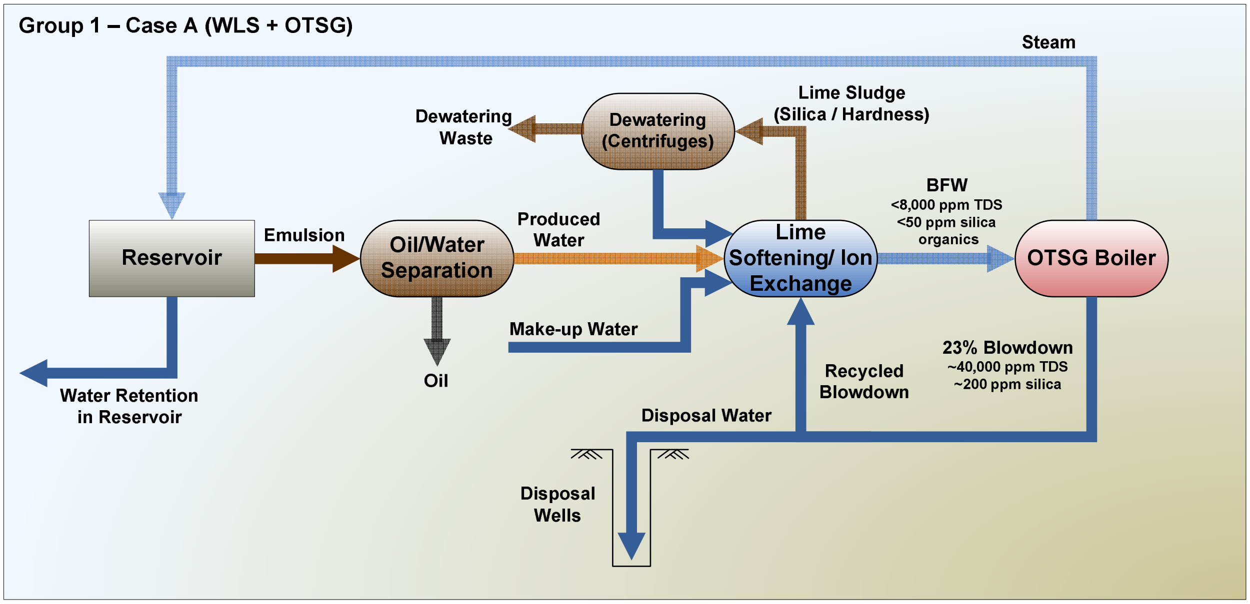


Figure 2-1: Group 1 – Case A CFD

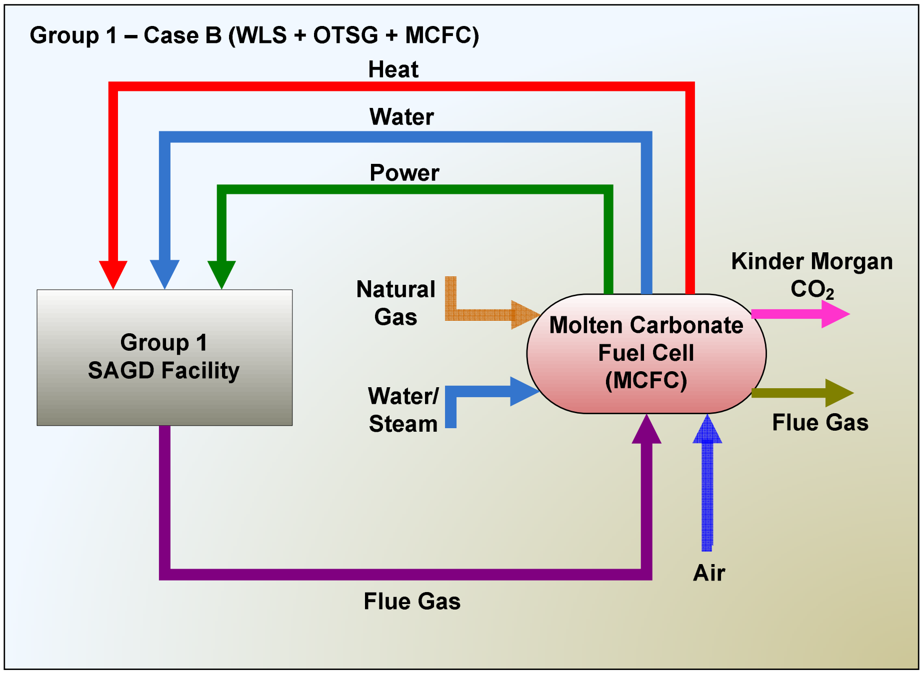


Figure 2-2: Group 1 – Case B CFD

In addition to the quantitative comparisons for the Cases above, we will provide qualitative comments on two other Groups of cases to the extent possible based on inferences from the Group 1 case comparison. The two other Groups of cases are represented by the CFDs for the A cases in Figures 2-3 and 2-4 below. Case B for Groups 2 and 3 are similar to the Group1 cases above in that a MCFC has been added to concentrate CO_2 from the flue gas from the Drum Boiler in Group 2 and the HRSG and OTSGs in Group 3.

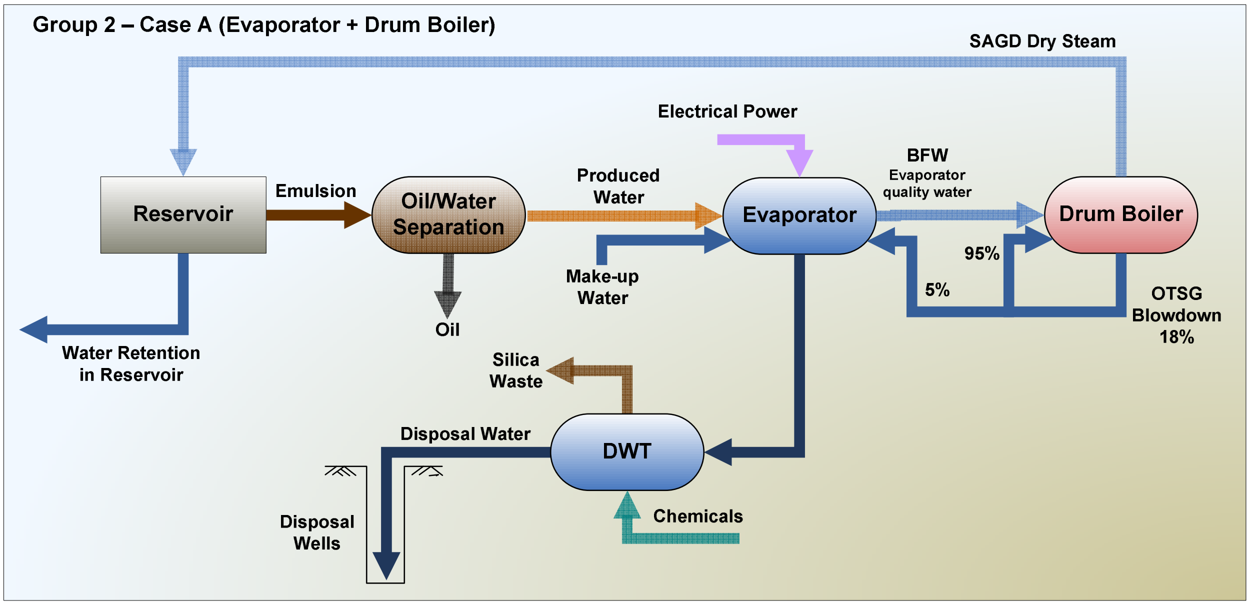


Figure 2-3: Group 2 – Case A CFD (Qualitative purposes only)

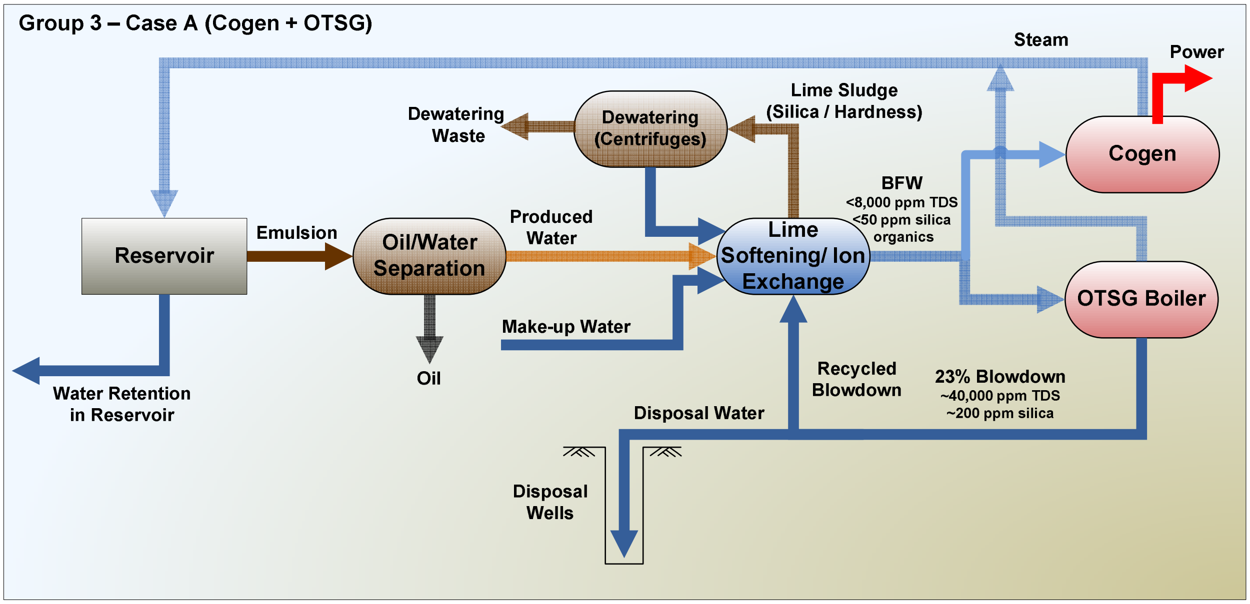


Figure 2-4: Group 3 – Case A CFD (Qualitative purposes only)

3. Base Case Assumptions

Key assumptions for the SAGD facility are as follows:

- Facility Size – For purposes of this study, the configuration will be a SAGD facility producing 33,000 bpcd of undiluted bitumen.
- Steam to Oil Ratio (SOR) = 3.0 (fixed)
- Gas to Oil Ratio (GOR) = 5.0 (fixed), produced gas will offset natural gas (NG) purchases
- Produced Water to Steam Ratio (PWSR) = 0.9
- Cases will be based on an “Efficient SAGD Design” which includes ESP for oil lifting, OTSG air economizer, efficient design, minimization of Ethylene Glycol (EG)
- OTSG Steam Quality = 77% at 100 barg
- Imported electrical power to meet on-site power demands
- Produced fluid at 160°C
- De-oil process identical for cases
- Blowdown disposal by deep well injection
- GHG calculations include CO_2 credits for power export based on 0.65 tonne/MWhr with sensitivities for 0.42 tonne/MWhr and 0.88 tonne/MWhr. The important part is to quantify the MWhr exported.
- Produced and makeup water qualities are shown below:

Quality	Produced Water (ppm)	Makeup Water (ppm)
TDS	1290	6961
Hardness	14	204
M-Alkalinity (CaCO3)	173	665
Ca	4	34
Mg	1	29
TOC	300	6
Silica	188	7

4. Key Commodity and Cost Assumptions

Operating costs will be based on the commodity prices shown in Table 4-1. The important part again is to show quantities for each of the commodities for each of the participating companies to perform their own economics.

Table 4-1: Commodity Prices

Commodity	Units	Price ($ CND)
Power	$/ MWhr	$90
Gas	$/ GJ	$5
CO_2 Credits	$/ tonne	$15
Chemicals	$/ m3 water	$0.77
Disposal costs including trucking	$/ tonne	$200
Fixed operating costs	% of TIC	3.5

Other important cost components for SAGD facility such as drilling and completions, well pads, roads, infrastructure, pipelines, gathering lines, steam lines, source and disposal wells are assumed to be the same regardless of the integration of a MCFC within the SAGD CPF and therefore are not included in the comparison.

5. Key Design Criteria

First and foremost, this is a high level, techno-economic comparison of a representative site, so detailed design criteria is not necessary. However, a few key items related to heat recovery and rejection are assumed to be as follows:

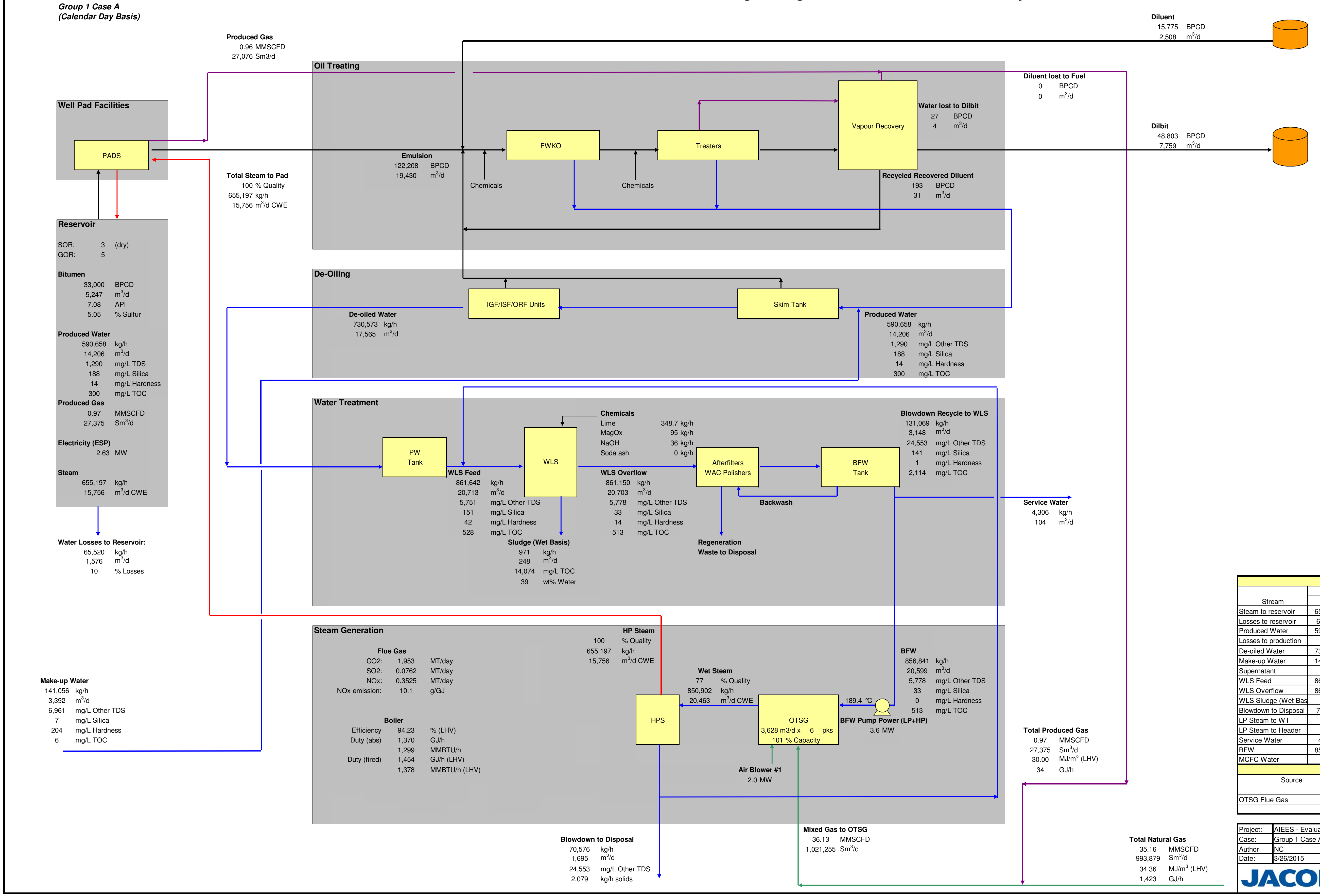
- Minimum atmospheric temperature = - 40°C
- Maximum atmospheric temperature = 40°C
- Average atmospheric temperature = 16°C
- Maximum temperature for disposal = 80°C

Appendix 2

IEs and BFDs

AIEES - Evaluation of Integrating MCFC with SAGD Study

Group 1 Case A
(Calendar Day Basis)



Summary Table	
MU TDS (ppm)	6,961
PW TDS (ppm)	1,290
PW TOC (ppm)	300
LP Flash BD (%)	-
BD Recycle (%)	65%
TDS to Boiler (ppm)	6,434
Boiler TOC (ppm)	513
MU Flowrate (kg/h)	141,056
WLS Sludge, Wet (kg/h)	971
Disposal Type (L,S)	L
Disposal Rate (kg/h)	70,576
Disposal Solids (kg/h)	2,079

Water Balance

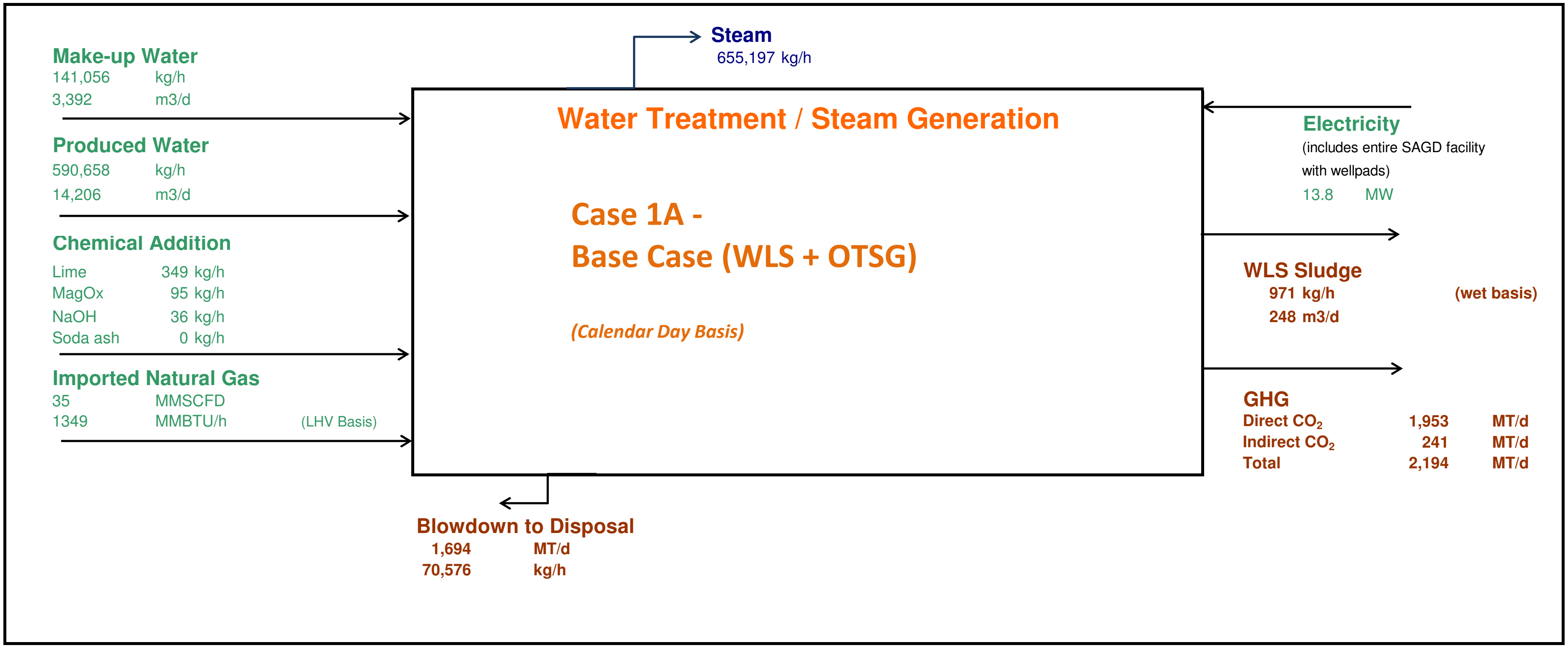
Stream	Flow kg/h	Flow m³/d	TDS ppm	Silica ppm	Hardness ppm
Steam to reservoir	655,197	15,756	-	-	-
Losses to reservoir	65,520	1,576	-	-	-
Produced Water	590,658	14,206	1,290	188	14
Losses to production	176	4	-	-	-
De-oiled Water	730,573	17,565	0	0	0
Make-up Water	141,056	3,392	6,961	7	204
Supernatant	-	-	-	-	-
WLS Feed	861,642	20,713	5,751	151	42
WLS Overflow	861,150	20,703	5,778	33	14
WLS Sludge (Wet Bas	971	248	-	-	-
Blowdown to Disposal	70,576	1,695	24,553	141	1
LP Steam to WT	0	0	0	0	0
LP Steam to Header	0	0	0	0	0
Service Water	4,306	104	5,778	33	0
BFW	856,841	20,599	5,778	33	0
MCFC Water	0	0	-	-	-

Emissions Summary

Source	SO2 metric t/d	S metric t/d	CO2 metric t/d	NOx metric t/d
OTSG Flue Gas	0.0762	0.0381	1953	0.35

Project:	AIEES - Evaluation of Integrating MCFC with SAGD Study
Case:	Group 1 Case A (Calendar Day Basis)
Author	NC
Date:	3/26/2015

JACOBS Consultancy



Water Treatment / Steam Generation

Case 1A -
Base Case (WLS + OTSG)

(Calendar Day Basis)

Make-up Water

141,056	kg/h
3,392	m3/d

Produced Water

590,658	kg/h
14,206	m3/d

Chemical Addition

Lime	349 kg/h
MagOx	95 kg/h
NaOH	36 kg/h
Soda ash	0 kg/h

Imported Natural Gas

35	MMSCFD	
1349	MMBTU/h	(LHV Basis)

Steam
655,197 kg/h

Electricity
(includes entire SAGD facility
with wellpads)
13.8 MW

WLS Sludge
971 kg/h (wet basis)
248 m3/d

GHG

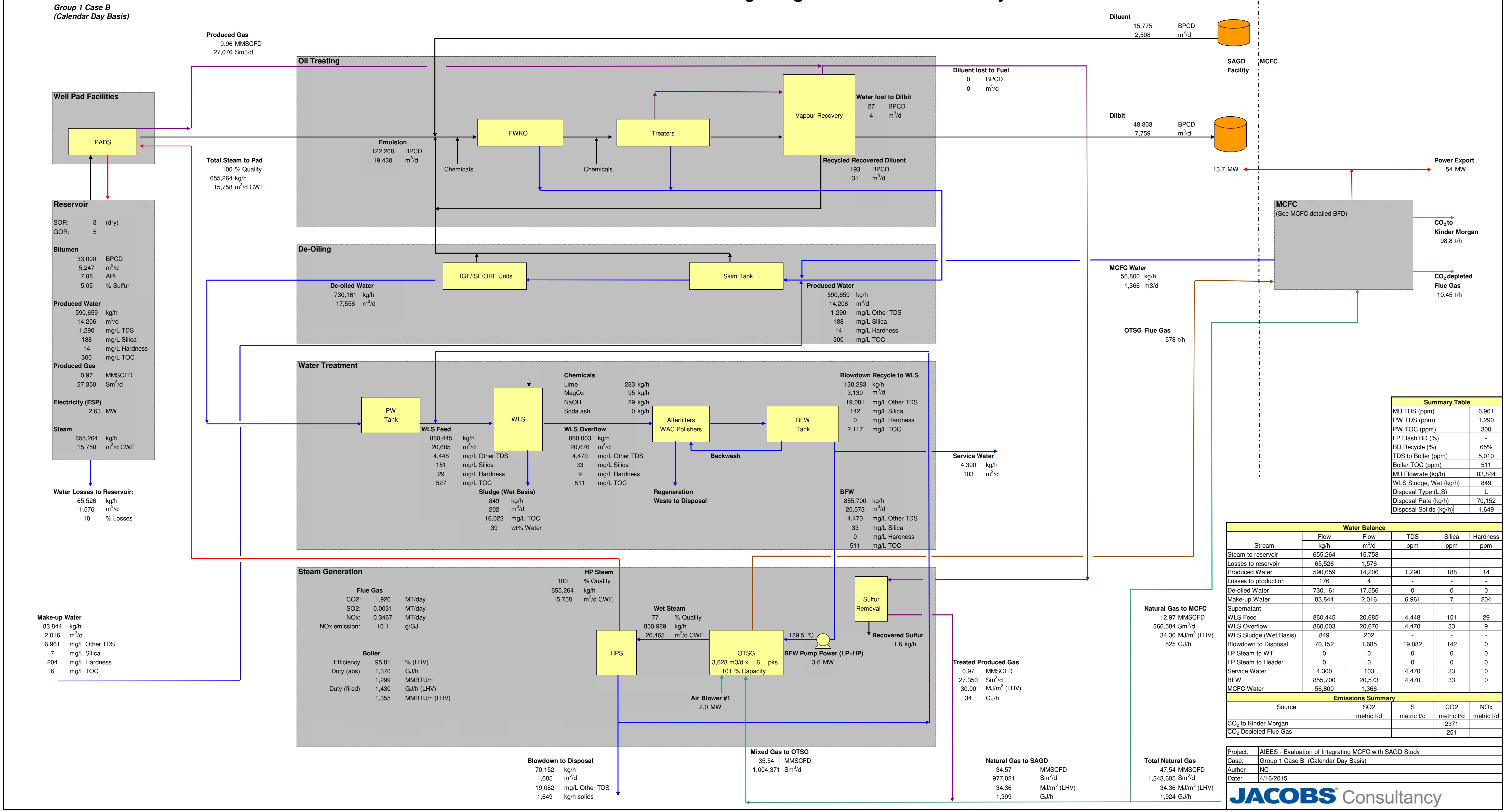
Direct CO_2	1,953	MT/d
Indirect CO_2	241	MT/d
Total	2,194	MT/d

Blowdown to Disposal

1,694	MT/d
70,576	kg/h

AIEES - Evaluation of Integrating MCFC with SAGD Study

Group 1 Case B
(Calendar Day Basis)



Summary Table	
MU TDS (ppm)	6,961
PW TDS (ppm)	1,290
PW TOC (ppm)	300
LP Flash BD (%)	-
BD Recycle (%)	65%
TDS to Boiler (ppm)	5,010
Boiler TOC (ppm)	511
MU Flowrate (kg/h)	83,844
WLS Sludge, Wet (kg/h)	849
Disposal Type (L,S)	L
Disposal Rate (kg/h)	70,152
Disposal Solids (kg/h)	1,649

Water Balance

	Flow	Flow	TDS	Silica	Hardness
Stream	kg/h	m^3/d	ppm	ppm	ppm
Steam to reservoir	655,264	15,758	-	-	-
Losses to reservoir	65,526	1,576	-	-	-
Produced Water	590,659	14,206	1,290	188	14
Losses to production	176	4	-	-	-
De-oiled Water	730,161	17,556	0	0	0
Make-up Water	83,844	2,016	6,961	7	204
Supernatant	-	-	-	-	-
WLS Feed	860,445	20,685	4,448	151	29
WLS Overflow	860,003	20,676	4,470	33	9
WLS Sludge (Wet Basis)	849	202	-	-	-
Blowdown to Disposal	70,152	1,685	19,082	142	0
LP Steam to WT	0	0	0	0	0
LP Steam to Header	0	0	0	0	0
Service Water	4,300	103	4,470	33	0
BFW	855,700	20,573	4,470	33	0
MCFC Water	56,800	1,366	-	-	-

Emissions Summary

Source	SO2	S	CO2	NOx
	metric t/d	metric t/d	metric t/d	metric t/d
CO_2 to Kinder Morgan			2371	
CO_2 Depleted Flue Gas			251	

Project:	AIEES - Evaluation of Integrating MCFC with SAGD Study
Case:	Group 1 Case B (Calendar Day Basis)
Author	NC
Date:	4/16/2015

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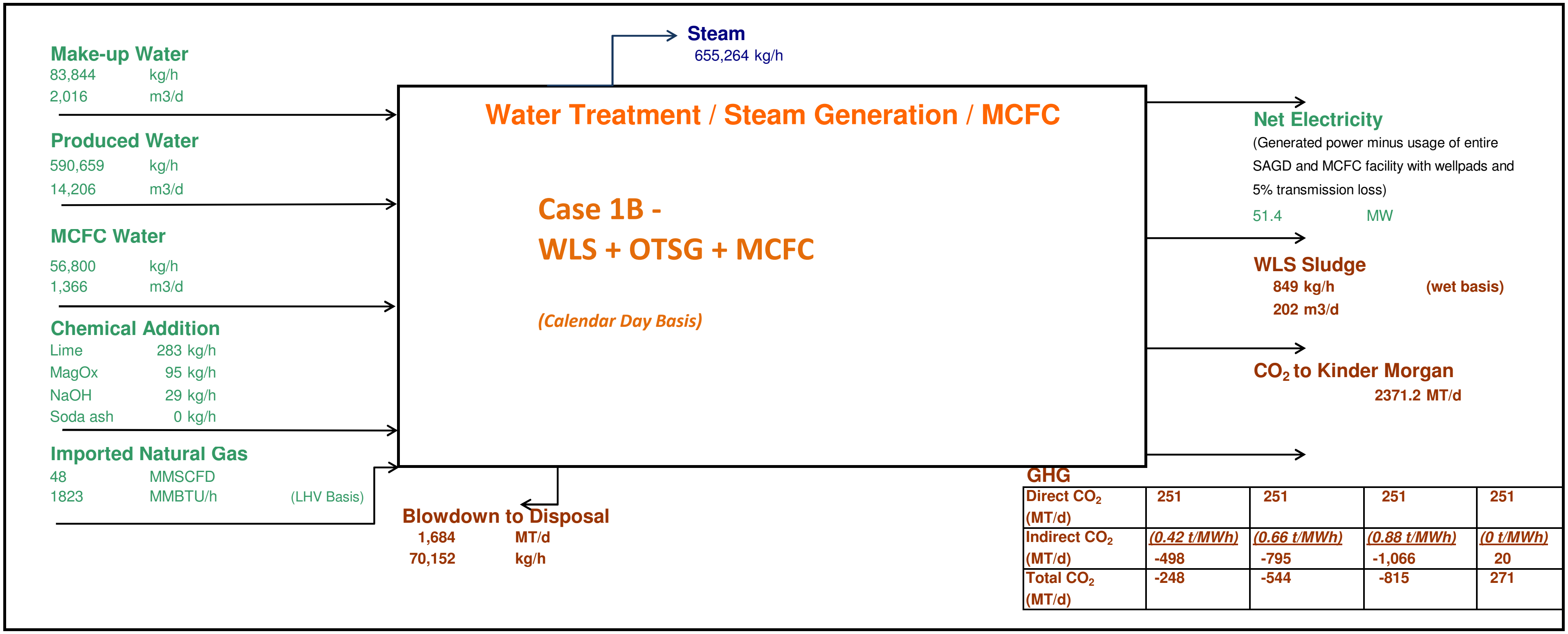


Make-up Water
83,844 kg/h
2,016 m3/d

Produced Water
590,659 kg/h
14,206 m3/d

MCFC Water
56,800 kg/h
1,366 m3/d

Chemical Addition

Lime	283 kg/h
MagOx	95 kg/h
NaOH	29 kg/h
Soda ash	0 kg/h

Imported Natural Gas
48 MMSCFD
1823 MMBTU/h (LHV Basis)

Steam
655,264 kg/h

Water Treatment / Steam Generation / MCFC

Case 1B -
WLS + OTSG + MCFC

(Calendar Day Basis)

Net Electricity
(Generated power minus usage of entire SAGD and MCFC facility with wellpads and 5% transmission loss)
51.4 MW

WLS Sludge
849 kg/h (wet basis)
202 m3/d

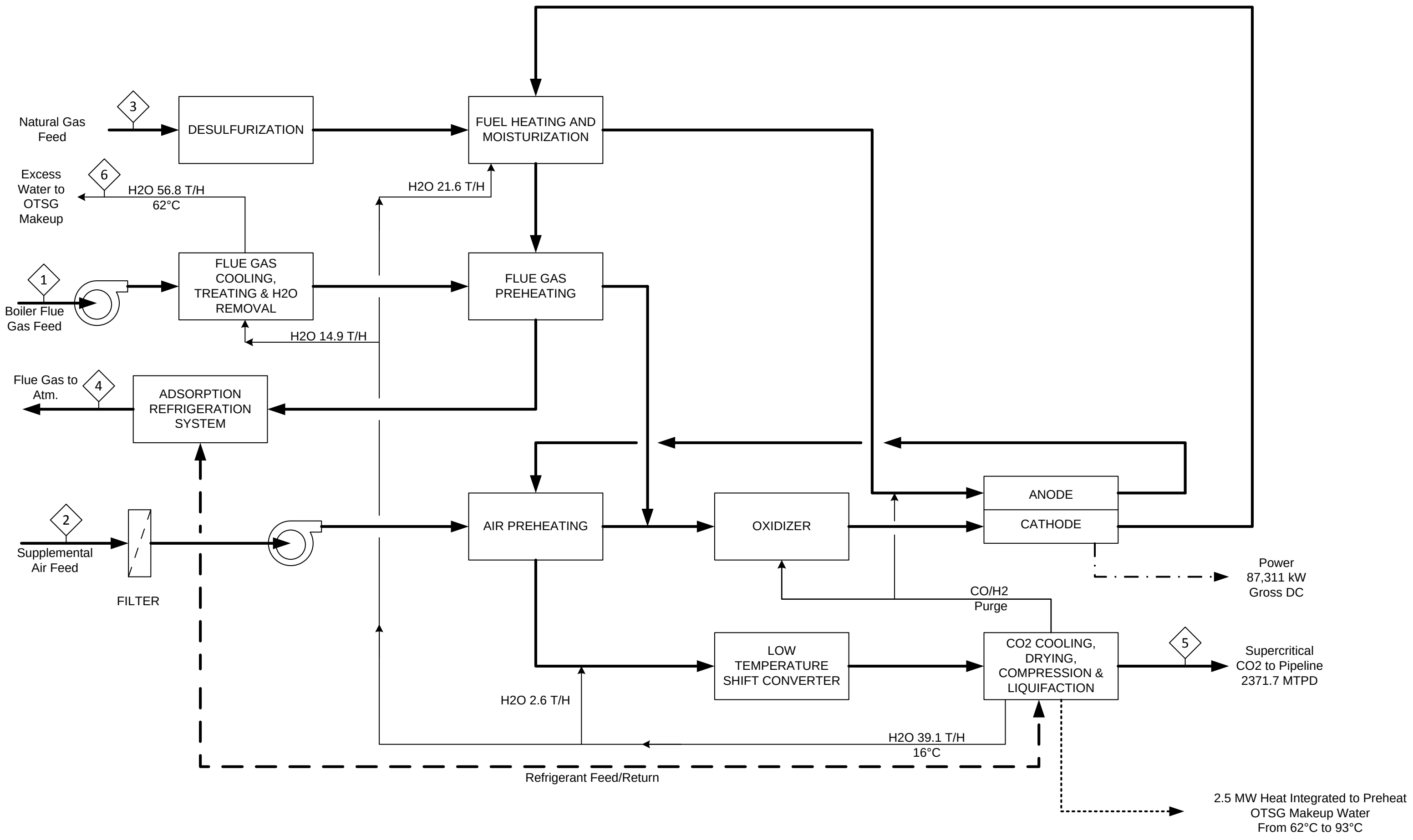
CO_2 to Kinder Morgan
2371.2 MT/d

Blowdown to Disposal
1,684 MT/d
70,152 kg/h

GHG

Direct CO_2 (MT/d)	251	251	251	251
Indirect CO_2 (MT/d)	*(0.42 t/MWh)* -498	*(0.66 t/MWh)* -795	*(0.88 t/MWh)* -1,066	*(0 t/MWh)* 20
Total CO_2 (MT/d)	-248	-544	-815	271

MCFC+CO_2 BFD


Natural Gas Feed
3
DESULFURIZATION
FUEL HEATING AND MOISTURIZATION
Excess Water to OTSG Makeup
6
H2O 56.8 T/H
62°C
H2O 21.6 T/H
1
Boiler Flue Gas Feed
FLUE GAS COOLING, TREATING & H2O REMOVAL
FLUE GAS PREHEATING
H2O 14.9 T/H
Flue Gas to Atm.
4
ADSORPTION REFRIGERATION SYSTEM
2
Supplemental Air Feed
FILTER
AIR PREHEATING
OXIDIZER
ANODE
CATHODE
Power
87,311 kW
Gross DC
CO/H2
Purge
LOW TEMPERATURE SHIFT CONVERTER
CO2 COOLING, DRYING, COMPRESSION & LIQUIFACTION
5
Supercritical CO2 to Pipeline
2371.7 MTPD
H2O 2.6 T/H
H2O 39.1 T/H
16°C
Refrigerant Feed/Return
2.5 MW Heat Integrated to Preheat OTSG Makeup Water From 62°C to 93°C

MCFC+CO_2 Material Balance

JPD94/1 Printed at : 9:44 4/30/2015 9:44 4/30/2015 Filename: O:\Gasification\JC-157760 Alberta Innovates SAGD CO2 Phase 2\23Process\05Calculations\01Original Calculations\[summary table rev AI 11.xlsx]Summary

JACOBS ENGINEERING

MATERIAL BALANCE

Document No.: **JC-1577-60/P.05/0001/A**

Project No.: JC157760 Plant: OTSG-ECM

Client: AI-EES Location: Alberta, Canada

Option 2 - ECM Material Balance

Page 1 of 1

STREAM NUMBER		1		2		3		4		5		6	
STREAM NAME		OTSG Flue Gas		Supplemental Air Feed		Natural Gas		Flue Gas to Atmosphere		CO2 Product to Pipeline		Excess Water	
COMPONENT	MW	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	6.32	0.28	0.00	0.00
Methane	16.042	0.00	0.00	0.00	0.00	615.79	95.06	0.00	0.00	0.24	0.01	0.00	0.00
Carbon Monoxide	28.010	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.28	0.01	0.00	0.00
Carbon Dioxide	44.010	1815.89	8.75	2.12	0.03	5.12	0.79	237.46	1.03	2245.46	99.67	0.01	0.00
Water	18.015	3517.63	16.95	14.84	0.21	0.00	0.00	1668.44	7.23	0.00	0.00	3151.24	100.00
Nitrogen	28.013	14994.04	72.25	5507.04	77.93	9.26	1.43	20509.79	88.86	0.55	0.02	0.00	0.00
Oxygen	31.999	425.44	2.05	1477.12	20.90	0.00	0.00	599.09	2.60	0.00	0.00	0.00	0.00
Argon	39.948	0.00	0.00	65.73	0.93	0.00	0.00	65.73	0.28	0.00	0.00	0.00	0.00
Ethane	30.069	0.00	0.00	0.00	0.00	11.47	1.77	0.00	0.00	0.00	0.00	0.00	0.00
Propane	44.096	0.00	0.00	0.00	0.00	3.69	0.57	0.00	0.00	0.00	0.00	0.00	0.00
I-Butane	58.122	0.00	0.00	0.00	0.00	1.75	0.27	0.00	0.00	0.00	0.00	0.00	0.00
N-Butane	58.122	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
I-Pentane	72.149	0.00	0.00	0.00	0.00	0.71	0.11	0.00	0.00	0.00	0.00	0.00	0.00
Temperature	°C	95.60		20.00		10.00		159.44		50.00		93.05	
Pressure	bara	1.00		1.01		6.89		1.01		152.99		2.00	
Total Molar Flow	(kg.mol/h)	20753.00	100.00	7066.85	100.00	647.79	100.00	23080.51	100.00	2252.85	100.00	3151.25	100.00
Total Mass Flow (kg/h)		576,940		204,520		11,020		636,850		98,860		56,770	
Molecular Weight		27.80		28.94		17.02		27.59		43.88		18.02	

Notes :	Issue:	A	Date		Date		Date		Date		Date
	Description:	For Issue									
	Made By:	DGL	21-Apr-15								
	Checked:										
	Approved:										

Appendix 3

Utility Summary

Project:	AIEES - Evaluation of Integrating MCFC with SAGD Study
Case:	Group 1 Case A (Calendar Day Basis)
Author:	NC
Date:	3/26/2015

Utilities Summary

	Capacity	Unit	Commodity	Power	Steam (positive=generation)		Fuel		CO2
Process Unit					HP	LP	Produced Gas	Natural Gas	
				kW	kg/h	kg/hr	MMSCFD	MMSCFD	metric t/d
Well Pads	33,000	BPCD	Bitumen						
Steam Injected to Wells					-655,197				
ESP Pumps				2630					
Pumpjacks									
Oil Treatment	122,208	BPCD	Emulsion						
Pumps				28					
De-oiling	110,481	BPCD	De-oiled Water						
Pumps				190					
VRU Compressors				15					
Water Treating	89,355	BPCD	Produced Water						
Pumps				49					
WLS				429		0			
Steam Generation	20,463	m3/d CWE	Wet Steam						
Air Blower				3374					
OTSG					655,197		0.97	35.16	1,953
Pumps				5566					
Offsites									
Sulphur	0.00	metric t/d	Sulfur						
Glycol	14,044	m3/d	Glycol	239					
Misc				1252					
Totals				13772					

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Project:	AIEES - Evaluation of Integrating MCFC with SAGD Study
Case:	Group 1 Case B (Calendar Day Basis)
Author:	NC
Date:	4/16/2015

Utilities Summary

Process Unit	Capacity	Unit	Commodity	Power kW	Steam (positive=generation) HP kg/h	LP kg/hr	Fuel Produced Gas MMSCFD	Natural Gas MMSCFD	CO2 metric t/d	
Well Pads	33,000	BPCD	Bitumen							
Steam Injected to Wells					-655,264					
ESP Pumps				2630						
Pumpjacks										
Oil Treatment	122,208	BPCD	Emulsion							
Pumps				28						
De-oiling	110,422	BPCD	De-oiled Water							
Pumps				190						
VRU Compressors				16						
Water Treating	89,355	BPCD	Produced Water							
Pumps				29						
WLS				428						
Steam Generation	20,465	m3/d CWE	Wet Steam							
Air Blower				3318						
OTSG					655,264		0.97	47.54	271	(based on 0 tonne of CO_2/MWh)
Pumps				5566						
Offsites										
Sulphur	0.04	metric t/d	Sulfur							
Glycol	14,642	m3/d	Glycol	250						
Misc				1246						
Totals				13701						
Net Power from MCFC				-68000						
Power Export (does not include 5% transmission loss)				-54299						