Alternative Fuels Study
Configuration Review

Prepared For

Alberta Energy Research Institute &
Alberta Environment

December 2008

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Section A.

Introduction

With the progression of fields in the Alberta Oil Sands region, producers are increasingly relying on in-situ methods of recovery, particularly for deposits too far down to allow mining methods. The most common method of bitumen recovery is Steam Assisted Gravity Draining (SAGD), which currently consumes large amounts of natural gas in the production of steam. In fact, natural gas makes up nearly two-thirds of the entire operating expense of a SAGD facility. Producers are looking for a more economic alternative, while regulatory bodies such as Alberta Environment (AENV) would like to encourage producers to reduce usage of this relatively clean energy source.

Alberta Energy Research Institute (AERI) and Alberta Environment (AENV) commissioned Jacobs Consultancy Inc. ("Jacobs Consultancy") to conduct this alternative fuels study to better understand the economic and environmental implications of alternative fuel sources. The first phase of the study (documented in Volume 1 of 2) was a general review of all technologies related to combustion or gasification of alternative fuels for use in SAGD bitumen production. In the second phase, the project team assembled technologies selected from the first phase to create a production facility configuration. The configuration study is documented in this report (Volume 2 of 2).

Modeling of the Alternative Fuels Process Configuration is intended to achieve the following:

1. Establish a basis by which to compare natural gas combustion with alternative fuels in terms of economic performance.

2. Define expected externalities such as foot print, air emissions, solid wastes, and waste water generation for the different fuel types.

3. Verify Capital Expenditure (CAPEX) and Operating Expense (OPEX) assumptions posited for different technologies in Phase 1 of this study.

The results of the configuration study were used both to serve as a basis for a decision matrix analysis and to allow economic comparisons of the technologies. Comparisons are intended to provide an indication of the relative incentives to producers for alternative fuels and establish the marginal costs of emission controls.

Section B.

Methodology & Assumptions

Methodology

Jacobs Consultancy has performed multiple Bitumen production and upgrading configuration studies that are generally intended to provide a method for producers to evaluate alternative technologies to burning natural gas, without commissioning a more detailed engineering study. The following are the key features of a Configuration Study:

- Block Flow Diagrams—Block Flow Diagrams (BFDs) are a means of creating a graphical representation of a SAGD facility without showing individual pieces of equipment. For this level of study, we do not delineate equipment; rather, we represent costs, utility requirements, feeds and products for plants or blocks. A BFD will show key feed and product flow rates along with utility requirements, but will generally not indicate intermediate streams or ancillary service flows.

- Utility Balances—A critical portion of the economics involve accounting for heat inputs as well as other utility requirements such as raw water, cooling water, power and steam.

- CAPEX calculations—Jacobs Consultancy has significant experience with engineering and construction of SAGD and bitumen upgrading facilities. We have a database of costs that can be used to represent the costs of similarly sized facilities by means of exponential curves.

- Economic Comparison—Based on the inputs from the BFD and Utility Balances, the economics of the cases (competing technology configurations) are compared side-by-side against each other and the natural gas base case.

- Decision Matrix Analysis—Failure Modes and Effects Analysis (FMEA) is a Six Sigma-based tool that is used to rank technologies by key parameters, both financial and non-financial, to determine the relative benefits of each configuration towards meeting project objectives.

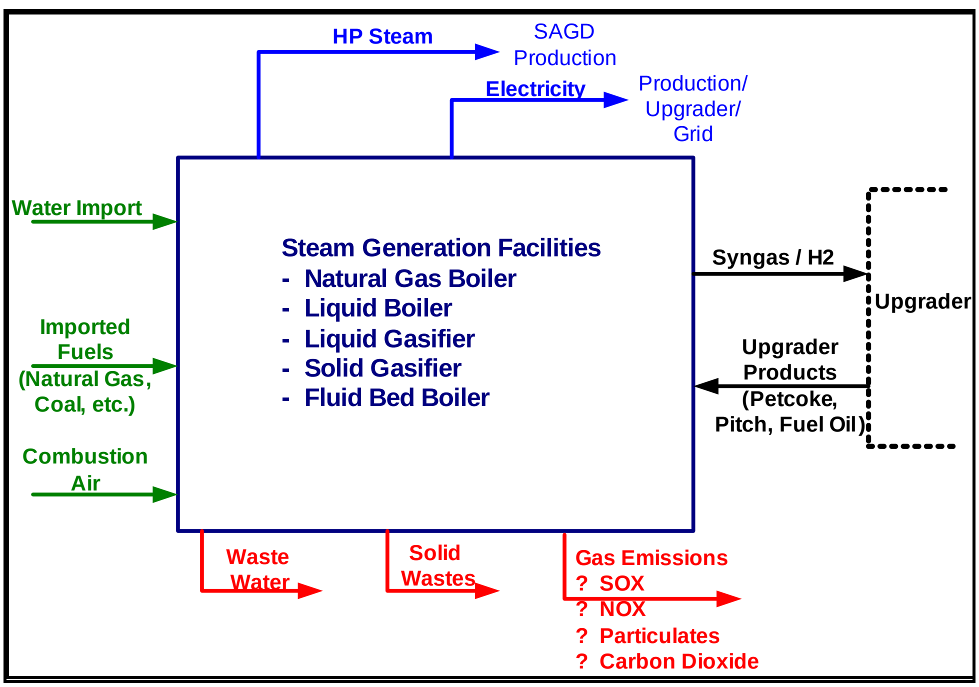
- Sensitivities—Configurations are made up of several small parts or blocks, and it is sometimes effective to analyze a change to a given block without reconstructing the whole configuration. Using sensitivity analysis, we recalculate the cash flow by adjusting a small number of parameters reflecting changes associated with an alternative configuration. As an example, to determine the economic impact of wet versus dry flue gas scrubbing, we would adjust the capital and operating costs associated with flue gas desulphurization and recalculate the cash flow.

Assumptions

Key assumptions in the creation of configurations independent of prices and other economic bases are as follows:

- Facility Size—For purposes of this study we have defined the plant as a SAGD facility feeding an upgrader. Based on past experience we defined the size of the upgrader to be 200 KBPD. We further assumed three SAGD facilities would supply the upgrader, making each SAGD facility 66 KBPD. The equipment required to burn or gasify alternative fuels are complex and expensive, and lend themselves more to a larger facility (>50 kbpd).
- Facility Scope—The objective of this study is to investigate alternative energy sources for SAGD steam generation facilities. As such, we have ignored the upgrading piece, assuming instead that the bitumen (as dilbit) will be sold to the local market. We selected to eliminate co-location (*i.e.,* onsite upgrading) and cogeneration from the scope. Thus, facility imports include bitumen (from the reservoir), utilities (water, steam, electricity), and diluent (see Figure B-1). The exports include any net energy (steam or power), syngas (carbon monoxide and hydrogen), hydrogen, flue gas, waste materials, and dilbit.

Figure B-1.
Facility Imports



- Steam to Oil Ratio—Steam to oil ratio (SOR) is the process parameter that most impacts site economics. This study, however, is not addressing the risks and variability in steam to oil ratio associated with reservoir performance. An SOR of 3.0 was selected for this study to represent the range of reservoir performance for operating and planned SAGD facilities.

- Bitumen Assay—The bitumen assay we have used is a blend of various publically available assays and is consistent with producer information. The assay is shown in Appendix 1.

- Diluent Requirement—Diluent requirements are essentially defined by targeting a maximum pipeline viscosity of 350 cSt @ 4˚C, which typically also results in a viscosity and API gravity for dilbit (mixed bitumen and diluent) suitable for oil water separation. For purposes of this study we have targeted a dilbit API of 20˚-22˚ (which also meets the pipeline viscosity requirements).

- Power Generation—While power generation can be incorporated into the SAGD facility, export of electricity from a Cogeneration plant skews economics and creates a new set of risk parameters. We have included a steam turbine in the gasification configurations to provide a home for gasification quench steam. However, for all cases we have assumed a net import of electricity with a constant price of $85.0 per MWhr.

- Water Utilization—Raw water import costs are assumed to be source well water and therefore not charged. We have included the costs for water treating and disposal water treating and injection. Current ERCB regulations require a water recycle of at least 90 percent. Should there be limitations on the amount of this water source, the use of brackish water may be required. This will require a membrane or evaporator system to allow use of this source of water. This equipment was not included in the scope of this study.

- Reservoir Characterization—Jacobs Consultancy estimated the reservoir design criteria, shown in Table B-1, based on past project experience; these are not from any particular reservoir. It is understood that each reservoir is unique and will change over time, but for the purposes of this study the typical or average reservoir characteristics have been compiled for the design of the well pads and Central Processing Facility (CPF).

Table B-1.
SAGD CPF and Reservoir Design Criteria

Reservoir Data	
Artificial Lift Mechanism	LP-SAGD / ESP's
Steam Oil Ratio (SOR)	3.0
Gas Oil Ratio (GOR)	4.0
Produced Gas Sulphur Content (mol%)	3.0
Reservoir Pressure (kPag)	1,400
Reservoir Production Subcooling (°C)	15
Water losses to Reservoir (% of injected)	10%
Oil Treating	
Bitumen and Naphtha Assays	Refer to Appendix 1
Treater Operating Temperature (°C)	130
Minimum API for Pipeline (°API)	22
Max Viscosity for Pipeline (cSt)	350 cSt @ 4°C
Target BS&W (%wt)	0.5
Water Treating	
Minimum Water Recycle (ERCB formula)	90%
Maximum TDS to OTSG	8000 ppm
Maximum Silica to OTSG	50 ppm
Package Drum Boiler Water Quality	Assume Evaporator water with post treatment is adequate
Disposal water maximum silica	200 ppm
Dewatered Sludge disposal	Local Landfill (Class 1)
Boilers	
OTSG Steam Quality	75%
Package Boiler Steam Quality	97%
Neat Bitumen/DRU Bottoms atomizing Steam (lb steam /lb oil)	0.2

* assuming enough pressure to return to CPF
** assumes flue gas / combustion air heat exchange

Environmental Standards

Every oil production facility is governed by the ERCB Standards. Most specifications outlined in these standards will be addressed in the detailed design phase of this project. However, a key requirement is the total amount of sulfur emissions allowed. According to Interim Directive # ID 2001-3 "Sulphur Recovery Guidelines for the Province of Alberta," the following table must be used for determining how much sulphur to recover from the gases produced in the plant. This guideline was used in the selection of the sulfur recovery options selected.

Table B-2.
Alberta Sulphur Recovery Guidelines

Sulphur Inlet Rate (tonnes/Day)	% of Sulphur inlet that must be recovered	
	Design sulphur recovery criteria	Calendar quarter-year sulphur recovery guidelines
1-5	70	69.7
>5 - 10	90	89.7
> 10 - 50	96.2	95.9
> 50 - 2000	96.5 – 98.8[1]	98.2 – 98.5[2]
> 2000	99.8	99.5

[1] **Recovery = 98.18185 + 0.18259 log10(sulphur inlet rate)**
[2] **Calendar quarter-year recovery = 97.88185 + 0.187259 log10 (sulphur inlet rate)**
Table taken from Alberta Sulphur Recovery Guidelines

Alberta Environment is in the process of reviewing Policy 1-A and will issue updated emission targets for alternative fuels in the near future.

Section C.

Configuration Descriptions

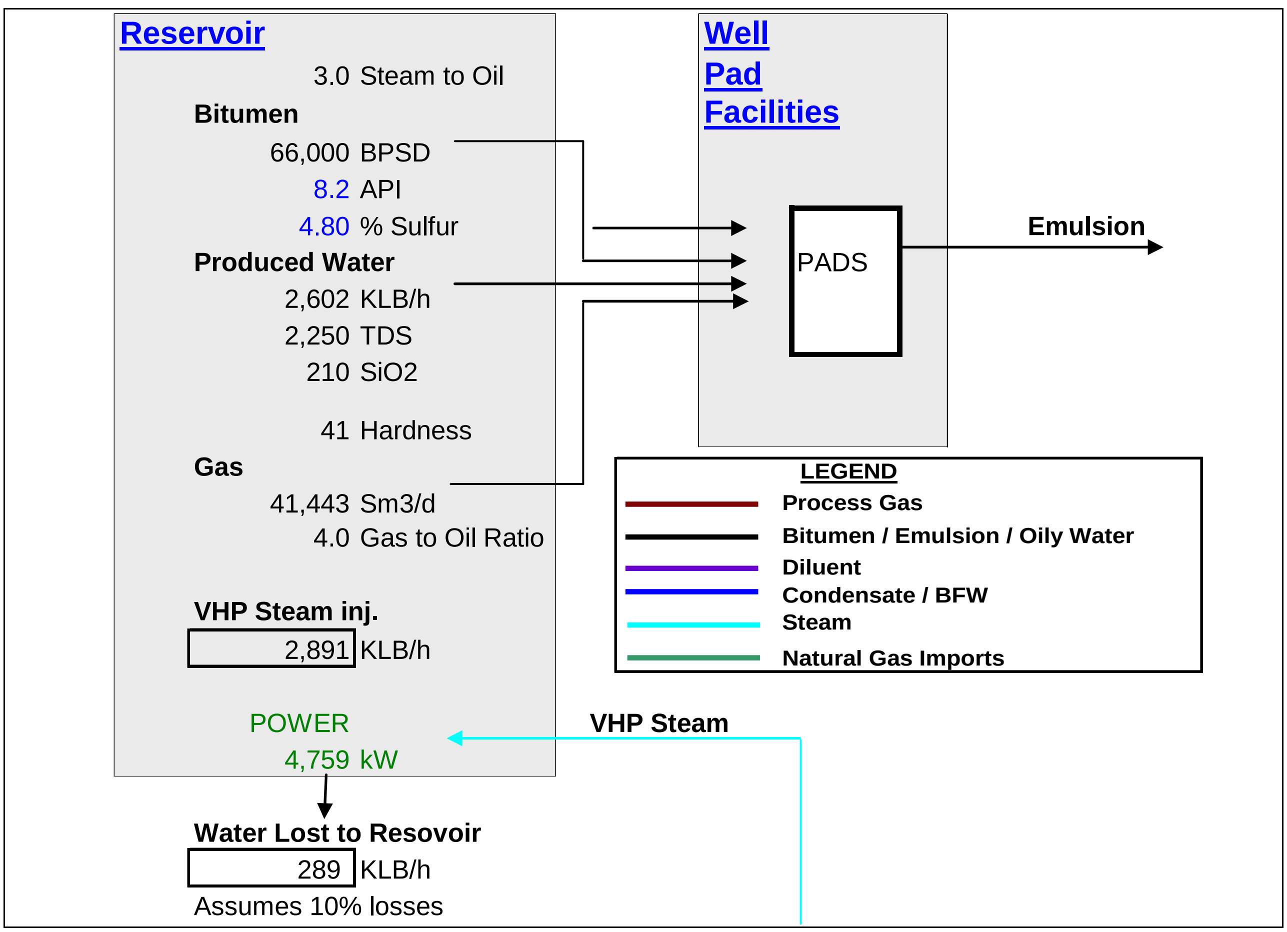
Block flow diagrams and utility balances for all cases are shown in Appendix 2.

Common Facilities

Reservoir and Well Pads

Figure C-1 depicts the blocks shown for the bitumen production reservoir and well pad. High Pressure Steam is injected into the reservoir to recover bitumen with a 3.0 Steam to Oil Ratio (3 Barrels of Steam Condensate for every barrel of oil). Lost water is assumed to be 10% of total steam injection. Power consumption reflects requirements of down-hole pumps.

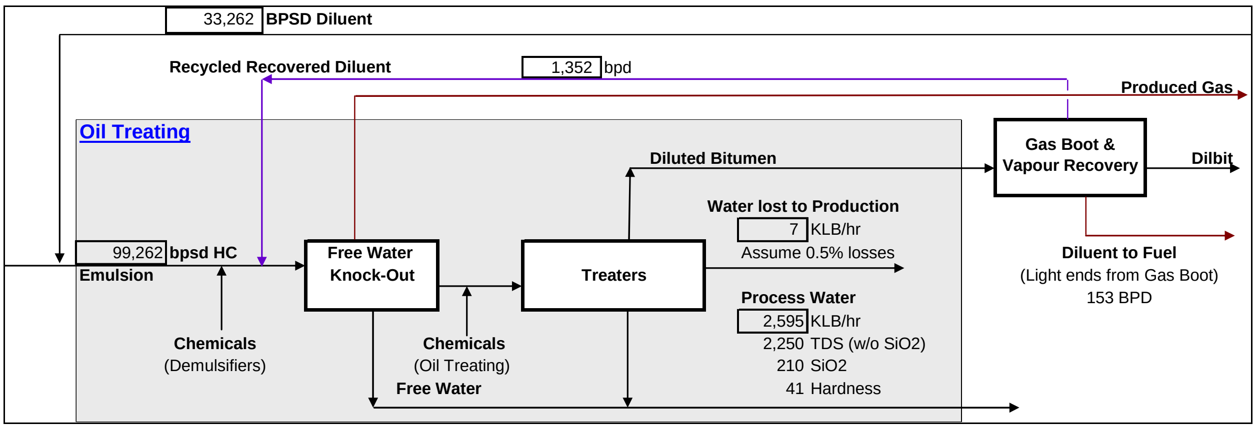
Figure C-1.
Bitumen Recovery Facilities



Oil Treating Section

The oil treating section block, as shown in Figure C-2, separates the 3-phase mixture (bitumen, gas and condensate). Diluent is added to reduce bitumen viscosity and density, assisting in the water-oil separation.

Figure C-2.
Oil Treating Section



De-Oiling and Water Treating

All facilities have water treating, but the configuration is slightly different. Once-Through Steam Generation (OTSG) boilers can tolerate lower quality water (at a steam quality of 75%), meaning that the less expensive Warm Lime Softening water treatment can be used. Modular Package Boilers and Circulating Fluidized Bed Boilers require higher quality water generated from Mechanical Vapor Recompression (MVR) evaporators.

Steam Generation

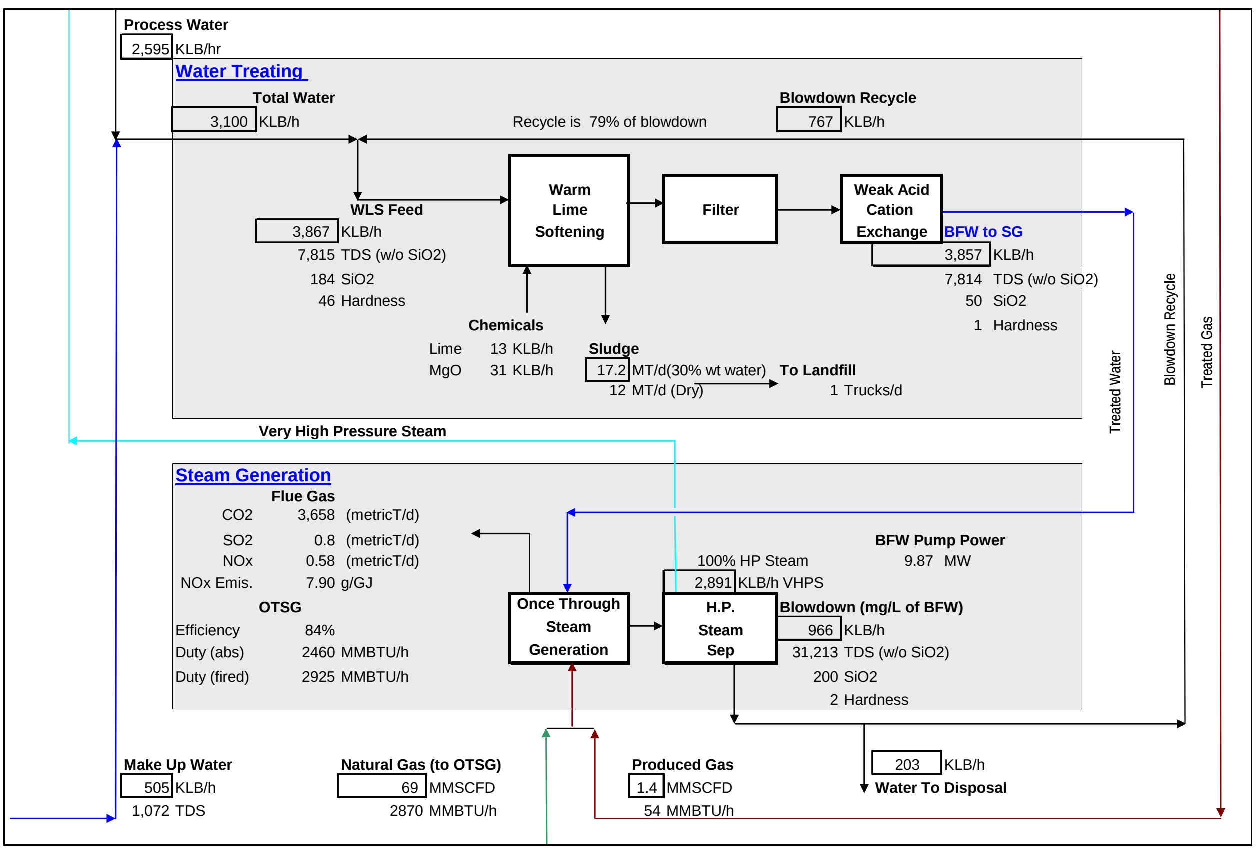
The primary differences between the configurations, aside from fuel handling and production sections, is the steam generation sections in which alternative fuels and boiler designs are used to generate the steam required for bitumen production. The commonality of this section is the import of boiler feed water and fuel with the export of flue gas and Very High Pressure (VHP) Steam.

Case 1—Natural Gas Configuration

Natural gas was selected as the base configuration (Figure C-3) and is representative of the overwhelming majority of SAGD facilities currently in operation. The features of the Natural Gas Configuration are as follows:

- Warm Lime Softening Water Treatment
- Once-Through Steam Generation
- Pretreatment of produced gas to remove sulfur
- NOx Control through Low NOx burners

Figure C-3.
Water Treating and Steam Generation Blocks—Natural Gas Case



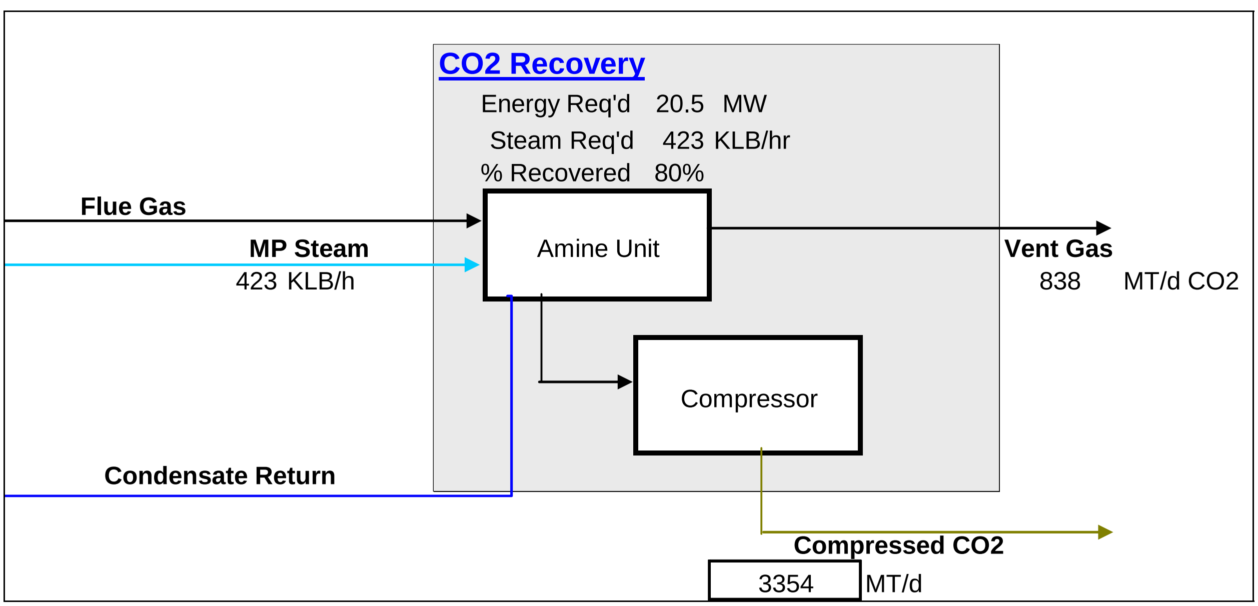
Case 2—Natural Gas Combustion with Carbon Capture

Even-numbered cases represent the previous odd-number case with a carbon capture facility added. In the case of Natural Gas, the technology chosen for carbon capture is amine absorption and regeneration. We have assumed an 80% recovery. This percentage represents recovering the majority of the CO_2 without resorting to expensive or untested methods. The project team chose an 80% target to avoid diminishing carbon capture economics by additional capital spending for marginal CO_2 recovery. An additional 15% steam is generated in the boilers to provide amine regeneration heat, further increasing the CO_2 production before capture.

Figure C-4 shows the carbon capture block for Natural Gas Case 2. Power costs reflect the cost of CO_2 compression and amine solvent circulation. Steam import for solvent regeneration increases the net heat load on the SAGD boilers.

Notably, since the flue gas must be cooled to knock out the water before compression (to avoid compressing water vapour), there is a significant amount of water condensed from the flue gas, which helps nearly offset the amount of make-up water required with 10% water losses to the reservoir and a portion of boiler blowdown to disposal. If the reservoir losses are less the facility would be a net producer of clean water.

Figure C-4.
Carbon Capture Block—Natural Gas Case 2

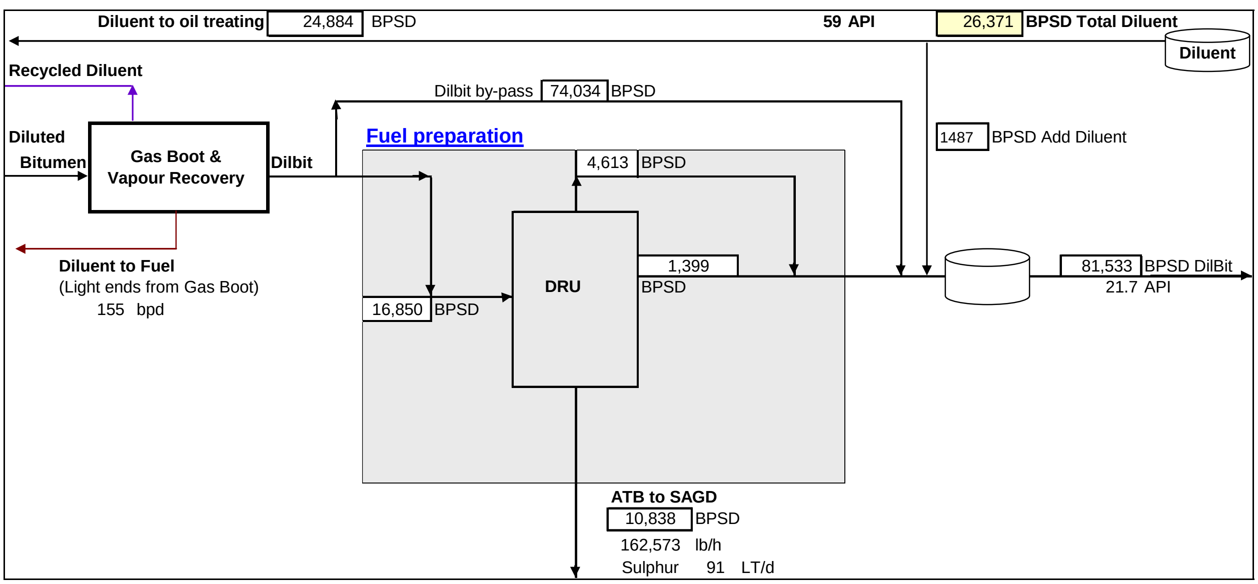


Case 3—Bitumen ATB Combustion

Since bitumen production is the end product of SAGD, burning a portion of the bitumen to generate steam is the first logical choice for alternative fuels. As shown in Figure C-5, treated bitumen is routed to a diluent recovery unit (DRU) to avoid sending diluent or diesel material to the boiler. The bottoms material from the DRU is typically referred to as Atmospheric Tower Bottoms (ATB). For purposes of this report, the terms "bitumen", "bitumen ATB," and "ATB" are used interchangeably when referring to energy generation.

Diluent is blended upstream of treating/desalting to allow for oil-water separation. While bitumen could be used prior to treating/desalting, boiler reliability would suffer due to the high chloride content of the fuel oil. The bitumen required for steam generation is approximately 16% of the total bitumen produced. Note that the export of dilbit for this case will be less than for the solid and natural gas combustion cases.

Figure C-5.
Diluent Recovery Unit Block



Other features of the bitumen combustion case include:

- Mechanical Vapor Recompression Evaporator water treating (modular boilers require higher water quality)
- Modular Package Boiler—with liquid/gas burners

- Wet Limestone Scrubber for Flue Gas Desulphurization (FGD)
- NOx control through Selective Catalytic Reduction (SCR)

Case 4—Bitumen ATB Combustion with Carbon Capture

As for natural gas, we have assumed the post combustion carbon capture for bitumen to be amine absorption (refer to Figure C-4). Some amine solvents may be less tolerant to SOx and therefore require increased flue gas desulphurization, but the project team has assumed that a suitable amine composition can be found for this application. An additional 23% steam must be generated in the boilers to provide amine regeneration heat, increasing the CO_2 production before capture. There are potential savings in heat integration with carbon capture facilities that need to be considered as part of the initial design. This factor will make carbon capture retro-fits more costly and less efficient.

Case 5—Petcoke Combustion

Petcoke, or coke produced from petroleum refining, was selected as a solid fuel because of the number of cokers being used locally for bitumen upgrading. Upgrading is primarily about carbon rejection to increase the quality of the synthetic crude, and coking is the oldest and best understood carbon rejection method. Combustion of coke is similar to coal combustion, except that the contaminants, particularly sulfur and nitrogen, are typically in a higher concentration.

A major complication to coke combustion in production areas is transport. With no train lines it will be necessary to truck in coke, potentially from the Suncor Upgrader sites or other locations in Edmonton. The amount of coke required to generate steam for a 66 KBPD bitumen facility will be 2,400 tons per day requiring 60 Super B (40 MT/load) trucks a day to supply. In addition, limestone supply for flue gas desulphurization will require twenty-one trucks a day. This large traffic increase will impact the environment (spills and noise) and road safety.

The circulating fluidized bed has a number of advantages over modular boiler technology, particularly regarding emissions. NOx emissions are substantially limited through distributing the burning zone throughout the bed, though nitrogen in the fuel may still oxidize. Limestone is added to the combustion bed, reducing SOx emissions, and making post combustion dry scrubbing (with at least 98% recovery) possible even with high sulphur petcokes.

Other features of the CFB Boiler configuration include:

- Mechanical Vapor Recompression (MVR) Evaporator Water Treating
- SNCR facilities for NOx control
- Fabric filters (baghouse) to control particulate matter
- No additional emissions control required for SOx and NOx

Case 6—Petcoke Combustion with Carbon Capture

Petcoke carbon capture can be the same as that for bitumen and natural gas, *i.e*. an amine flue gas scrubber with steam solvent regeneration. An additional 37% more steam (Case 5) is required for amine regeneration to achieve the 80% reduction in CO_2 emissions. As a consequence this case also produces 37% more CO_2 (before capture) than the non capture Case 5. Note that since the CFB is a single boiler, to retrofit CO_2 capture the producer must either design the boiler with excess capacity or cut facility production. Adding a second boiler for carbon capture will further reduce economic return.

This case also has the highest electrical power consumption for combustion cases because of the evaporator compressors and the CO_2 compressors. It is a total of 96 MW of power consumption between the two blocks, which is equivalent to an additional 2.2 MT/D of indirect CO_2 emissions.

Case 7—Bitumen Gasification

Gasification technologies are detailed in Volume 1. The advantage to gasification in energy generation is that the resulting syngas (Carbon Monoxide and Hydrogen) can be easily desulphurized and burned as with natural gas. Thus, SOx, NOx and particulate emissions are much lower than for liquid or solid combustion.

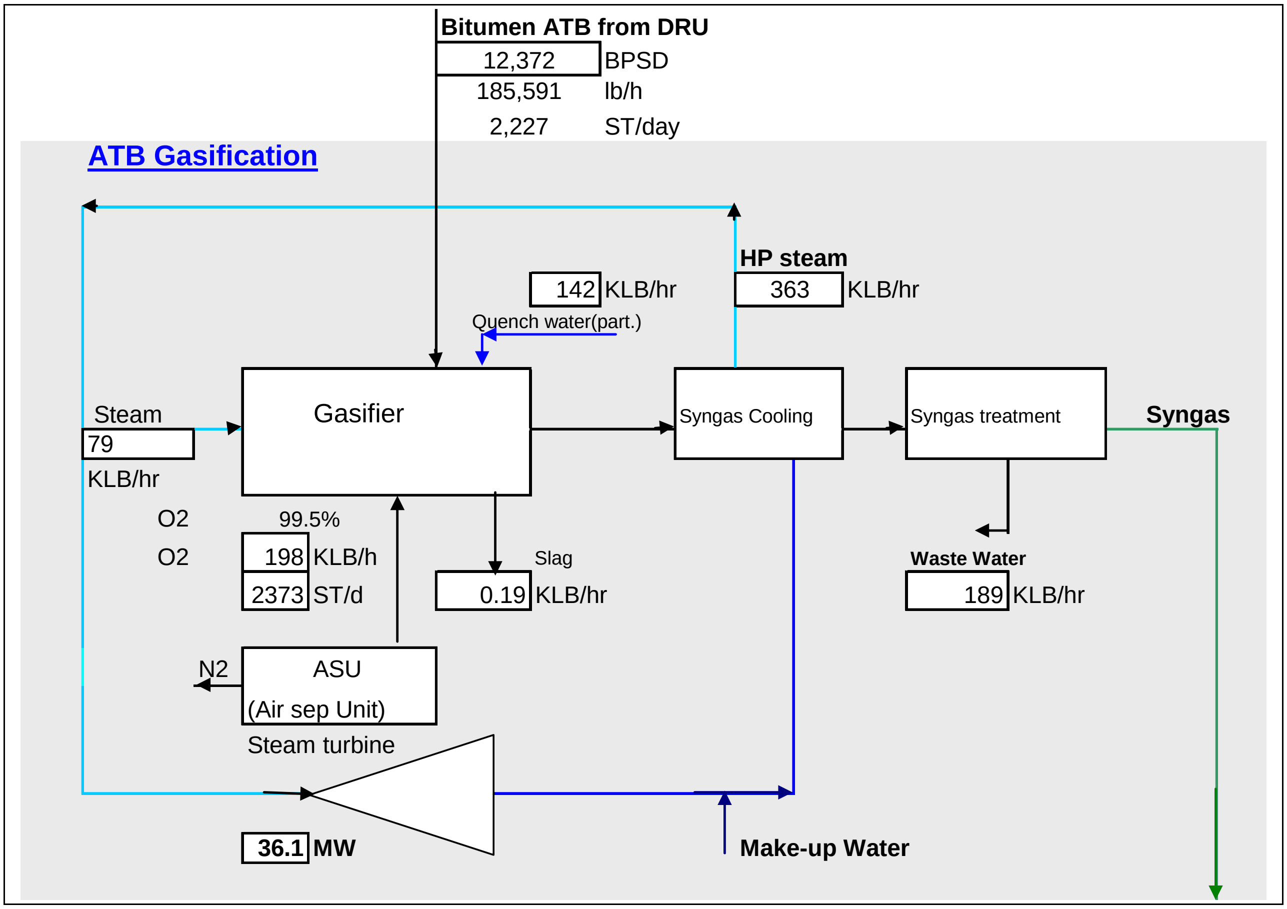
A significant disadvantage to gasification is that equipment requirements are much more significant than those for liquid or solid combustion. There is also a large power consumption to produce oxygen for the gasification step. For bitumen gasification the feed supply has been assumed to be the same as for bitumen (atmospheric resid) combustion. Heavier liquids such as Solvent De-Asphalting pitch are more typical gasification feeds, but generating this material was not considered for this high level study. The gasification process being used is the Entrained Flow gasifier, which can be used for both liquid and solid gasification.

Features of the gasification step include:

- Air Supply Unit
- Gasification Reactor
- Syngas Cooling and Treating
- Acid Gas Removal and Sulfur Production (Amine Contacting with SuperClaus®)
- Power Recovery Turbine

The power recovery turbine is required because syngas cooling from the gasification step requires higher quality water than the WLS can deliver. This condensate is segregated from high solids water used for OTSG. Generating internal power reduces import power demand and keeps quench water treating costs low.

Figure C-6.
Bitumen ATB Gasification Block



The steam generation and water treating facility for gasification are the same as for natural gas combustion (Case 1) including:

- Warm Lime Softening
- Once-Through Steam Generators
- Low NOx burners with recirculation for NOx control

Case 8—ATB Gasification with Carbon Capture

We did not produce a block flow diagram for the gasification carbon capture cases. The carbon separation facility is integrated with the acid gas removal amine system. A solvent, such as Selexol or Rectisol, can typically be used to absorb Carbon Dioxide and Acid Gas after shifting Carbon Monoxide and water to make hydrogen. Recovery capital and energy costs are reduced from those of post combustion or oxyfuel capture technologies.

Costs for carbon capture are significantly less for gasification than combustion for the following reasons:

- Syngas can be shifted to a mixture of carbon dioxide and hydrogen, which is much more easily separated than a mixture of nitrogen, oxygen, water vapor and carbon dioxide.
- Solvents used for syngas separation (Selexol, Rectisol, *etc*.) can absorb and regenerate at higher pressures, reducing compression costs.
- Energy input for solvent regeneration is much lower for these solvents.

Case 9—Petcoke Gasification

Solids gasification is similar in configuration, energy requirements and process conditions to the bitumen gasification cases. As with petcoke combustion, a major challenge to implementing petcoke gasification is related to supply and the massive increase in plant site truck traffic. Since gasification is less efficient than combustion, the amount of coke supply increases, as does carbon dioxide generation. While capital costs are higher and thermal efficiency is lower for solids gasification than liquid gasification, the gross margin is higher because all bitumen recovered is sold as dilbit. As with petcoke combustion, gasification of petcoke at a remote site will require extensive infrastructure and transport to supply the required fuel.

Case 10—Solid Gasification with Carbon Capture

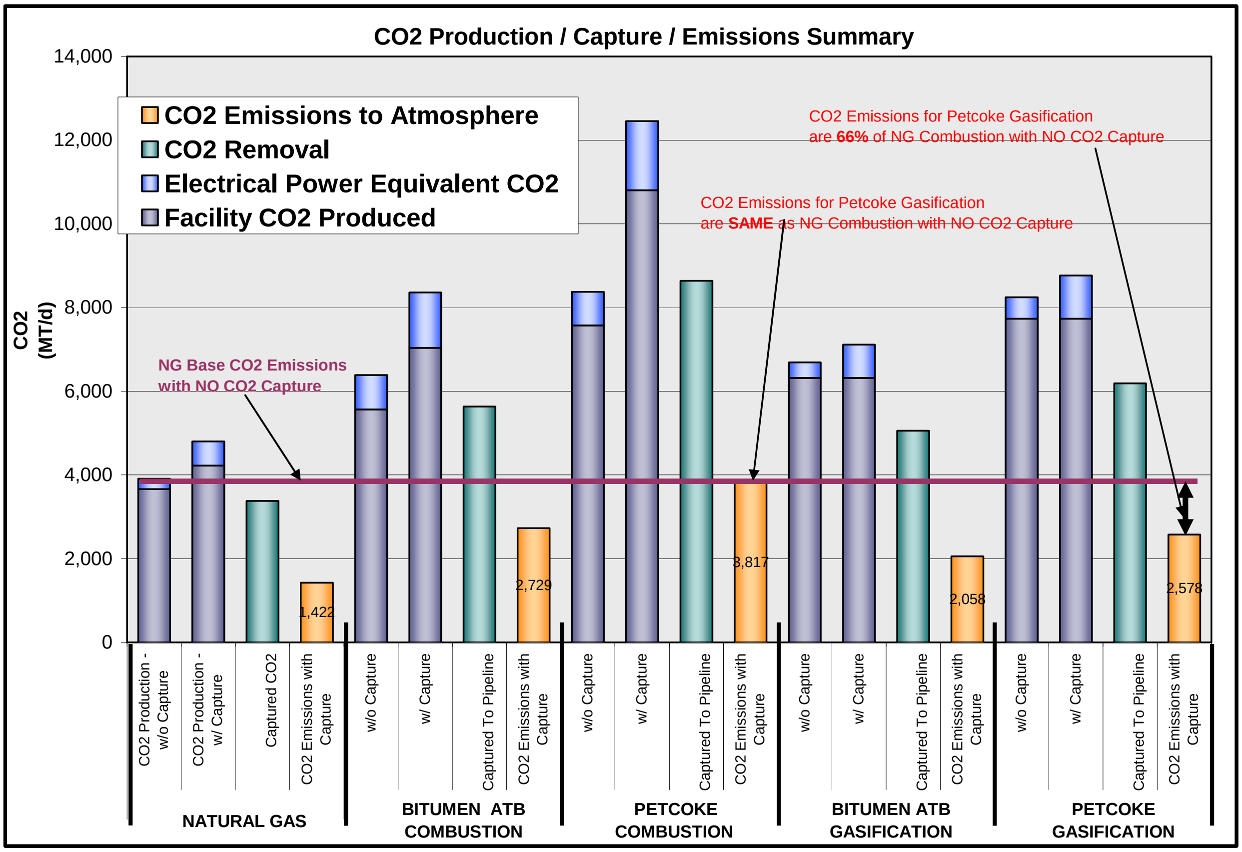
As with Case 8, the carbon capture block is integrated with the process. More CO_2 will be produced for petcoke gasification, and power requirements for disposal will be slightly higher. Still the energy requirements and net carbon emissions are much lower for petcoke gasification than the petcoke combustion case with capture.

CO_2 Emissions Summary

Figure C-7 shows the direct and indirect CO_2 generated for each type of fuel along with the recovered CO_2 for the carbon capture cases. It is clear from the above graph that alternative fuels produce much more CO_2 than natural gas to generate the same amount of steam, which is a function of the hydrogen content of the fuels. One striking result, demonstrated in Figure C-7, is that even with 80% carbon capture, the total emissions of petcoke in a CFB are the same as burning natural gas WITHOUT carbon capture. Similarly, the petcoke gasification case has 66% of burning natural gas WITHOUT carbon capture.

It should also be noted that the amount of CO_2 that must be pipelined and sequestered for petcoke gasification is roughly double the amount for natural gas. ATB is roughly 66% more CO_2 captured than natural gas. CO_2 transport and storage costs (not considered for this study) are normally small compared to capture costs, but these costs will further reduce the return for alternative fuels.

Figure C-7.
CO_2 Emissions by Fuel Type



Carbon Emissions on a common oil recovery basis (constant steam to reservoir)
Indirect Emissions from power imports are based on 0.613 MT/MWHr (US Average per DOE)

Section D.

Economic Analyses

Economic Assumptions and Qualifications

Economic assumptions are summarized in Table D-1. Information is provided with the following qualifications.

- All costs are presented on a 2008 basis, assuming construction in Fort McMurray with start-up in 2013. Cost figures are highly variable and should not be used for purposes other than technology and fuel type comparison.
- Capital Expenditures are over 4 years, ending in 2013.
- Cash Flow Discounted at 10 percent.
- Royalties, Depreciation, and Taxes included in Cash Flow (refer to Table D-1).
- All costs are curve costs developed from Jacobs' projects and should be considered +50% / - 25% in accuracy.
- Economic calculations are based on a 35-year plant life with zero terminal value.
- All figures are presented in Canadian dollars. An exchange rate of 1.0 USD per Canadian Dollar has been used in all calculations.

Feed and Product Pricing

Hydrocarbon prices are calculated from the following:

- Bitumen = 0.52 * WTI—This basis is low basis current market conditions, reflecting the best economic conditions for alternative fuels combustion.
- Diluent = WTI + $5
- Purchased Coke $15 / MT including transport costs. Coke price is assumed independent of natural gas and bitumen price (*i.e*. excess supply in region).
- Base Prices: Natural Gas = $10 / MMBtu; WTI = $80 / Bbl
- CO_2 Penalty = $15 / MT (Basis Alberta Government guidelines).

CAPEX

- TIC (SAGD Steam Generation and Production) based on Jacobs Curve Costs

- OSBL, EPCM, Owner Costs Based on Factors
- 25% Contingency has been applied to all capital numbers

OPEX

- Utility and Chemical Prices (Oil Treating, water treating, *etc*.) developed from 2008 pricing data and technology utility balances.
- Ongoing Maintenance based on CAPEX
- Fixed Costs (Salaries, Insurance, Taxes) factored from CAPEX

Table D-1.
Economic Analysis Base Assumptions

Parameter		Comments
Plant Operating Parameters		
Operating days per year	343 Days	
Onstream Availability	94.0%	
Ongoing Maintenance Costs (% of TIC)	3% per year	
Insurance (% of TIC)	0.3% per year	
Staffing Costs (% of TIC)	1.2% per year	
Basic Economic Parameters		
Federal Corporate Tax Rate	na	
State or Provincial Corporate Tax Rate	na	
Combined Corporate Tax rate	30.0%	
Exchange Rate ($US/%Can)	1 $US / Can	
Discount Rate	10.0%	
Return for royalty calculation	5.0%	
Long Term Inflation Rates		
Feedstocks	3.0% per year	
Products	3.0% per year	
Fixed and Variable Costs	3.0% per year	
Capital	2.5% per year	Construction Cost Escalation
Fuel	3.0% per year	
Base Case Costs		
Electricity	$85.00 / MWhr	
Water	$0.00 / MT	
FGD Limestone	$98.00 / MT	
Water Treating Chemicals	$0.85 /$M^3$ Water	
Oil Treating Chemicals	0.5 / Bbl oil	
Solid Waste Disposal	$50 / MT	
Purchased Petcoke	$15 / MT	Includes Transportation Costs
Natural Gas	$10 / GJ	Base Case (Varied in sensitivities)
West Texas Intermediate (WTI) Crude	$80 / Bbl	Base Case (Varied in sensitivities)
Carbon Dioxide Penalty	$15 / MT (CO2)	Base Case (Varied in sensitivities)
Pipeline Bitumen Transfer Cost (To Edmonton)	$1.63 / Bbl	
Pipeline Diluent Transfer Cost (To Edmonton)	$1.31 / Bbl	
Pipeline Diluent Transfer Cost (From Edmonton)	$1.31 / Bbl	
Diluent	Calc	WTI Price + $5
Bitumen	Calc	WTI Price * 0.52
Sulfur Product	-$15.00	

Indicates items varied in sensitivities

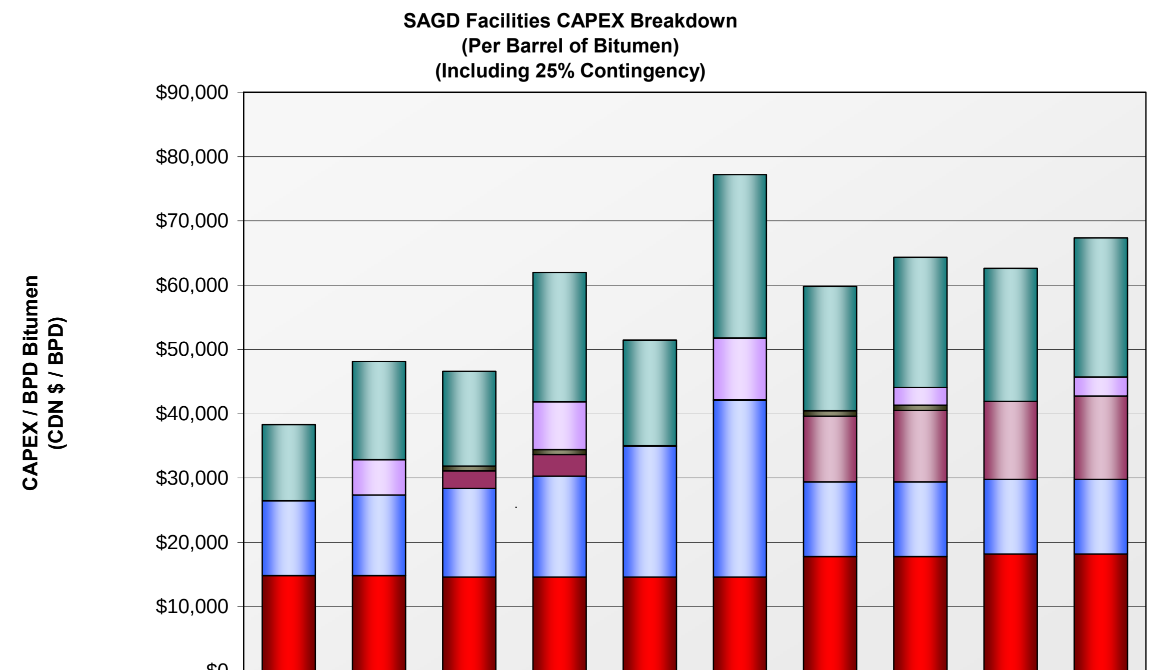
Capital Expenditures

Figure D-1 shows a capital expenditures breakdown for each case as well as the overall costs. Note that these figures are dollar per barrel of bitumen extracted, which includes any combusted bitumen. We can make the following conclusions about capital expenditures (CAPEX) based on this figure:

- Natural Gas combustion (Case 1) is the lowest cost of the alternative studied. The savings for Natural Gas are primarily in boiler and water treating costs and the benefit of limited emissions control investment.

- Of the combustion technologies, petcoke combustion (Case 5) is the most expensive, with most of the cost differential in the more expensive CFB boiler.
- Gasification costs (Cases 7 and 9) are more expensive than the combustion technologies without carbon capture, attributable primarily to high gasifier costs.
- Costs to integrate carbon capture are much lower for gasification. When comparing carbon capture cases, solids gasification is less expensive than petcoke (CFB) combustion.
- Carbon capture facility costs increase notably for combustion technologies depending on the amount of carbon generated (*i.e.* higher costs for coke combustion). Carbon capture is based on capturing 80% of the produced CO_2.
- Costs for Carbon capture are broken out as a separate line item. In addition to these direct costs, other parts of the facility (Steam Generation, water treating, *etc*.) have to increase in size to meet the heat requirements of carbon capture.
- Oil treating costs for gasification are higher because of the inclusion of the Acid Gas Removal facilities.
- Flue gas emissions control costs are included for petcoke and bitumen combustion. The costs for CFB petcoke combustion are small as desulphurization and NOX control are incorporated into the boiler.
- Emission control costs for bitumen combustion are significant, but removing these costs will not make costs comparable to the natural gas OTSG costs.

Figure D-1.
Configuration Capital Costs—Total Expenditures



	Case 1 NG	Case 2 NG + CC	Case 3 Bitumen ATB	Case 4 Bitumen ATB + CC	Case 5 PetCoke	Case 6 Coke + CC	Case 7 Bitumen ATB Gasif.	Case 8 Bit. ATB Gasif + CC	Case 9 Petcoke Gasif	Case 10 Petcoke Gasif + CC
Other Costs (EPCM, Owner Ocsts, Offsites)	11,838	15,268	14,744	20,113	16,435	25,443	19,360	20,232	20,671	21,623
CO2 Capture	0	5,481	0	7,447	0	9,628	0	2,767	0	2,989
DRU	0	0	763	776	0	0	838	838	0	0
Emission Controls (NOx and SOx)	0	0	2,739	3,353	75	96	0	0	0	0
Gasification	0	0	0	0	0	0	10,217	11,083	12,127	12,931
Steam Generation	11,650	12,550	13,769	15,682	20,343	27,450	11,639	11,639	11,650	11,650
Oil Production Facilties	14,816	14,816	14,600	14,600	14,600	14,600	17,764	17,764	18,157	18,157

Operating Expenditures

Figure D-2 shows the case operating expenditures total and breakdown. We can make the following conclusions regarding the operating expenditures (OPEX) of cases:

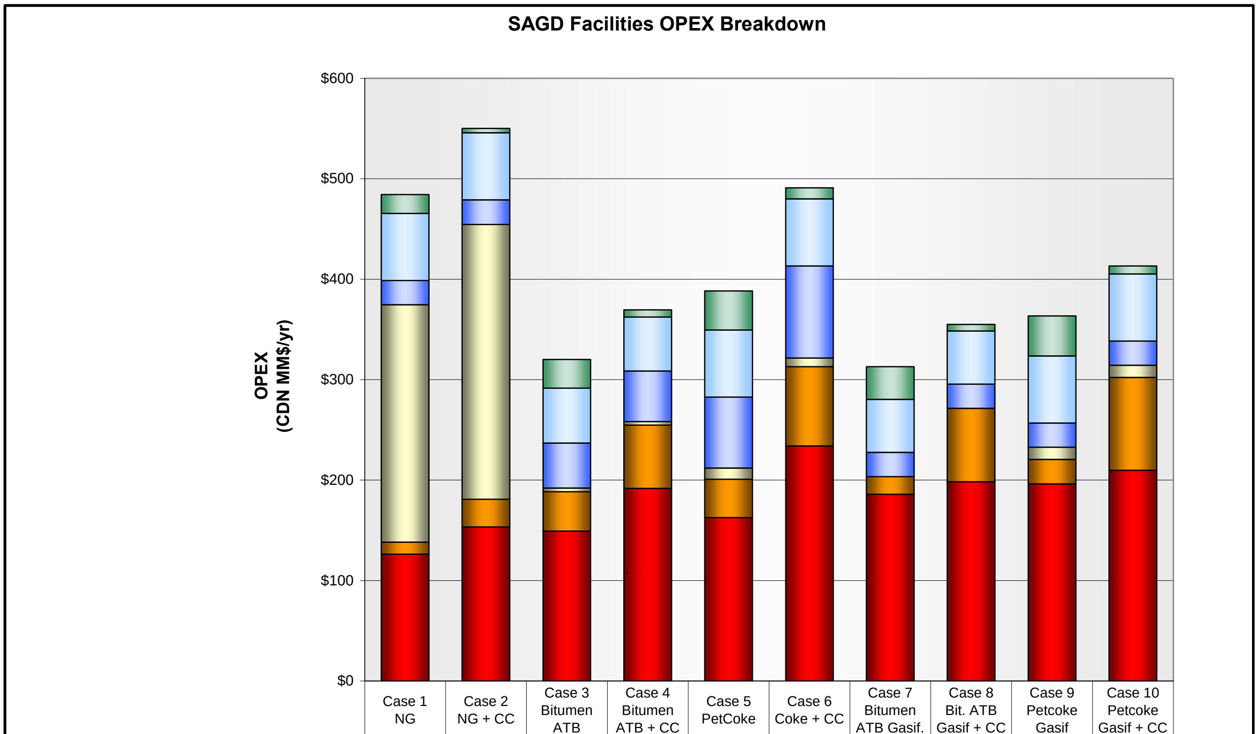
- As expected, the natural gas cases (Case 1 and 2) have the highest operating expenses, reflecting the cost of natural gas imports. However, the OPEX savings realized by going to alternative fuels is less than 20% of the natural gas case total operating costs (primarily because of higher costs to run evaporators).
- For comparison purposes, bitumen combustion is indicated as an operating cost. In fact, this amount is actually an opportunity cost, reflecting lost revenues and not a third party payment.

- For combustion costs, the savings realized from carbon capture (reduced CO_2 penalty) do not cover the increased operating costs (Power, steam, Maintenance, manpower *etc*.) of capturing CO_2. Thus, at the assumed $15/MT penalty for CO_2 emissions, there would be no economic incentive for a producer to recover CO_2, even assuming the capital costs are sunk.

- For gasification cases, the benefit of recovering carbon roughly offsets the increased OPEX. The largest operating expense associated with carbon capture for both combustion and gasification is the power costs for CO_2 compression.

- Gasification cases have higher fixed costs, reflecting increased equipment maintenance and staffing. Fuel costs are also slightly higher compared to the corresponding combustion cases, reflecting lower efficiency for gasification. These increases are offset by lower chemical costs (no FGD limestone) and electrical costs (gasifier includes steam turbine generation).

Figure D-2.
SAGD Facility Operating Expenses per barrel of bitumen (2008 Basis)



	Case 1 NG	Case 2 NG + CC	Case 3 Bitumen ATB	Case 4 Bitumen ATB + CC	Case 5 PetCoke	Case 6 Coke + CC	Case 7 Bitumen ATB Gasif.	Case 8 Bit. ATB Gasif + CC	Case 9 Petcoke Gasif	Case 10 Petcoke Gasif + CC
CO2 Penalty ($15/MT)	19	4	29	7	39	11	33	7	40	8
Pipeline transfer costs	67	67	55	54	67	67	53	53	67	67
Chemicals	24	25	45	50	71	92	24	24	24	24
Natural Gas / Petcoke	236	274	4	4	11	9	0	0	12	12
Electrical Power	12	27	39	63	38	79	18	73	24	92
Fixed OPEX (Staff, Maintenance, Taxes, Insurance)	126	153	149	192	163	234	186	198	196	210

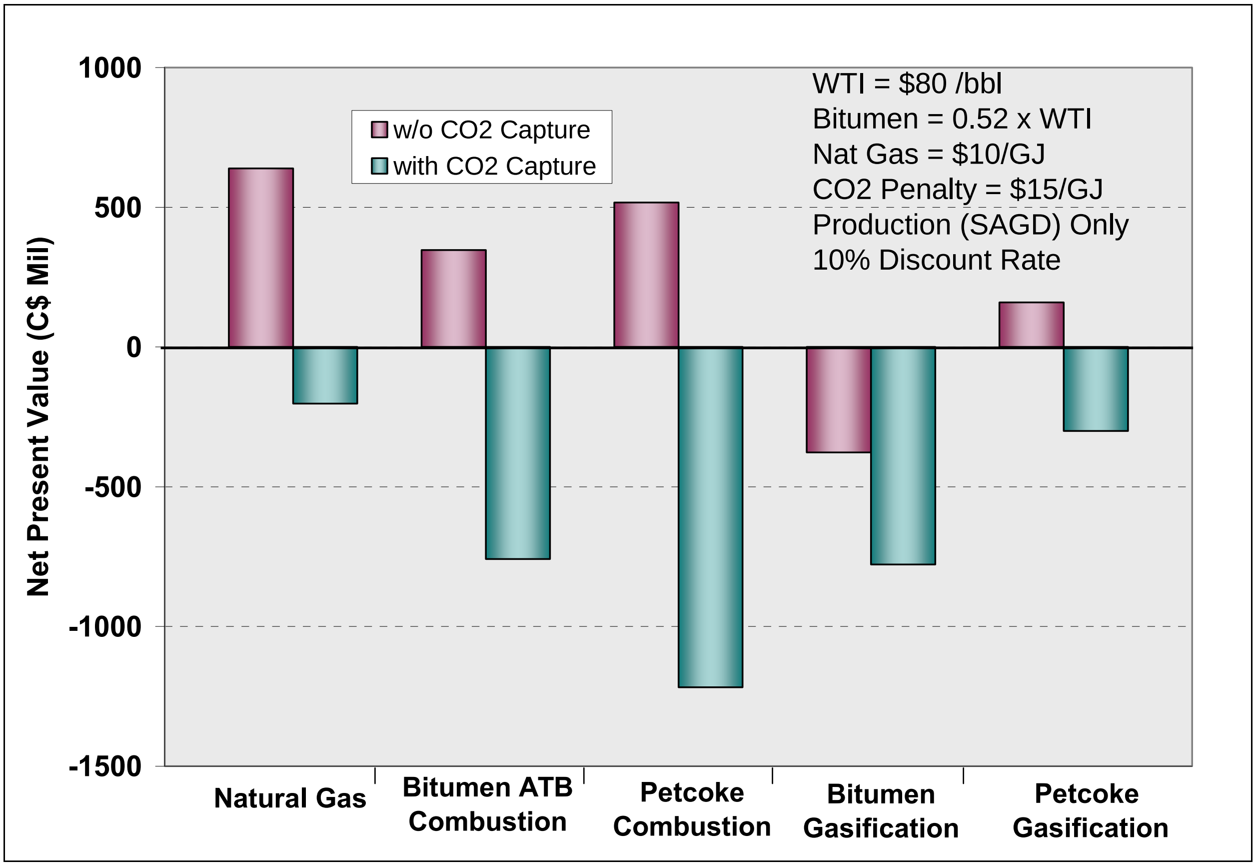
Economic Return

Figure D-4 shows a comparison of the internal rate of return for all configurations. The internal rate of return assumes a 30-year plant life, and a 10% discount rate. Royalties, depreciation, working capital changes, and taxes are all considered in the economic analysis. Product and utility prices are correlated to Natural Gas price (base - 10 $/MMBtu) and a benchmark crude price (West Texas Intermediate or WTI; base - $80 / bbl : Bitumen = 52% WTI price). The results of our economic comparison suggest the following:

- A higher natural gas price would be required to invest in alternative fuels.
- Petcoke combustion offers a better return than bitumen combustion for these configurations. The advantage is primarily related to lost production associated with combusting bitumen.

- Bitumen combustion and gasification offer poor returns compared to other fuels. Gasification of liquids usually requires a less valuable product (*e.g.* Solvent De-asphalted pitch) to be economic. Pitch gasification was considered for this study, but economics for pitch must either include pipeline costs from Edmonton (currently non-existent) or the addition of a Solvent De-asphalting facility, both of which are outside this study scope.

Figure D-3.
Configuration Net Present Value



A summary of economics at base conditions is shown in Table D-2. These results are highly dependent on the assumptions outlined in Table D-1 and will shift as these parameters, particularly crude and natural gas prices, are adjusted.

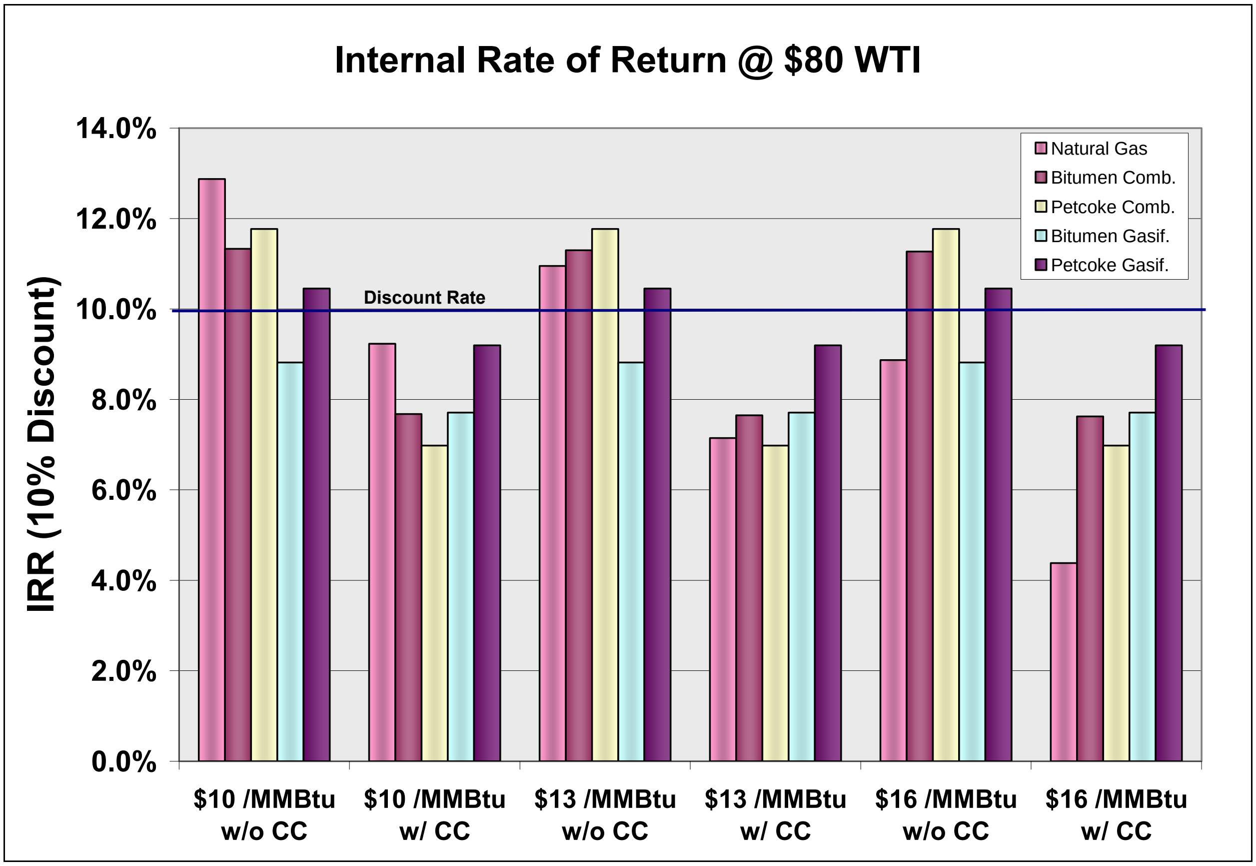
Table D-2.
Base Case Economic Results

		Natural Gas	Bitumen ATB Comb.	Petcoke CFB	Bitumen ATB Gasif.	Petcoke Gasif.
Without Carbon Capture		**Case 1 w/o Capture**	**Case 3 w/o Capture**	**Case 5 w/o Capture**	**Case 7 w/o Capture**	**Case 9 w/o Capture**
Capital Costs	$ Million	2,432	2,980	3,299	3,852	4,099
Product Revenue	$ Mil / Yr	2,034	1,656	2,034	1,605	2,033
Feed Costs	$ Mil / Yr	1,032	818	1,032	791	1,032
Total Expenses	$ Mil / Yr	484	320	388	313	363
Margin (Before Tax)	$ Mil / Yr	358	171	226	127	167
Net Present Value (2008)	$ Million	639	347	517	-377	159
Internal Rate of Return		12.87%	11.33%	11.77%	8.82%	10.45%
With Carbon Capture		**Case 2 w/ Capture**	**Case 4 w/ Capture**	**Case 6 w/ Capture**	**Case 8 w/ Capture**	**Case 10 w/ Capture**
Capital Costs	$ Million	3,079	3,994	5,000	4,149	4,424
Product Revenue	$ Mil / Yr	2,034	1,632	2,034	1,605	2,033
Feed Costs	$ Mil / Yr	1,032	818	1,032	791	1,032
Total Expenses	$ Mil / Yr	550	370	491	355	413
Margin (Before Tax)	$ Mil / Yr	397	178	257	157	203
Net Present Value (2008)	$ Million	-202	-758	-1,217	-778	-300
Internal Rate of Return		9.23%	7.68%	6.98%	7.71%	9.20%

Figure D-4 demonstrates the variation in economics as natural gas price increases. As expected, alternative fuels become more attractive at higher gas prices. However, the fuel price required for economic parity is greater than $10/MMBtu, which exceeds Alberta's highest historic natural gas price (based on annual average). The current Jacobs Consultancy projection is that above $10/MMBtu, liquefied natural gas (LNG) transfer by ship becomes economic, thus providing a cap on local natural gas prices.

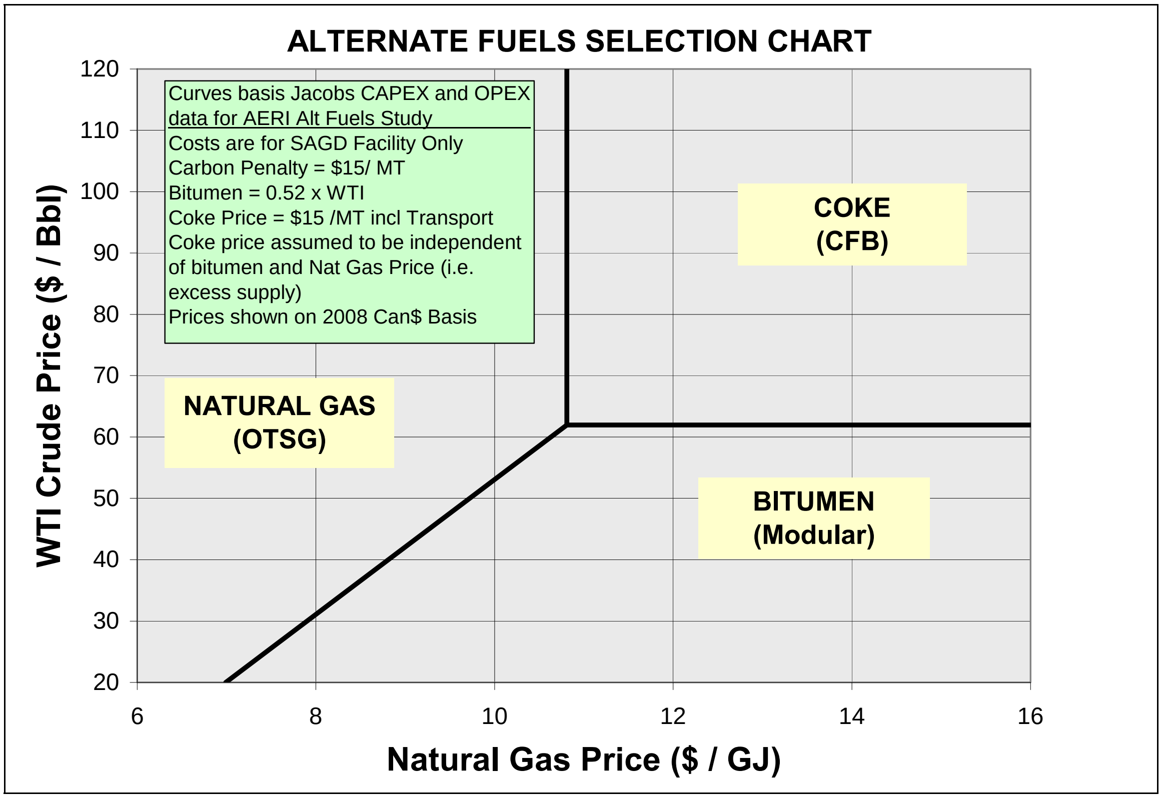
Alternative fuel cases are essentially independent of natural gas prices, with the exception of a small amount of pilot gas imports for the bitumen combustion cases. This fuel gas is required for operation of the Diluent Recovery Unit (DRU).

Figure D-4.
Internal Rate of Return as a Function of Natural Gas Price



Based on the capital and operating expenses calculated for each configuration, it is possible to determine the preferred fuel type at different natural gas and crude prices. Figure D-5 shows a tipping chart produced for the combustion configurations. A producer looking to select a fuel type can use this chart to determine preferred fuel assuming crude and natural gas forecasts.

Figure D-5.
Alternative Fuels Selection Chart without Carbon Capture



Without carbon capture, bitumen combustion is favored at low crude prices and high gas prices. As crude prices increase, petcoke becomes economically attractive. At lower gas prices, natural gas is the preferred fuel.

Based on Figure D-5, petcoke production would not be pursued unless:

- Fuel gas prices were projected to be greater than $10.8 / MMBtu; and
- Crude prices were projected to be greater than $61 / Bbl

Bitumen combustion would not be economically favored unless:

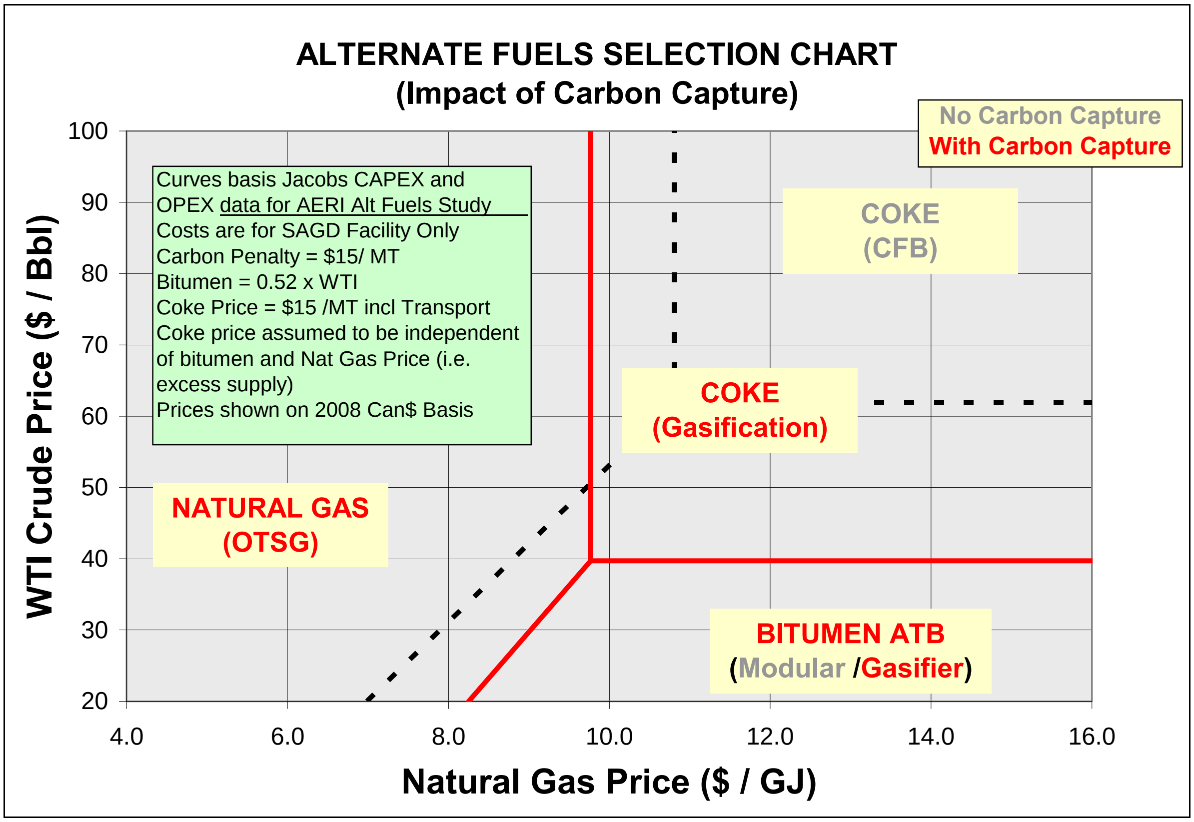
- The Ratio of WTI to fuel gas price is less than 6.0; or
- Crude prices are less than $61 / bbl and natural gas prices are greater than $10.2/MMBtu.

The following additional conclusions can be ascertained based on Figure D-5:

- Historically, none of these scenarios outlined have been in effect for more than a few months. During periods of high natural gas prices other sources are identified and the price drops. Based on monthly average data, natural gas prices in Alberta have not been above $10 /GJ (constant dollar basis) since 2005, and then only for three months.
- The tipping chart does not show gasification because the economics for gasification are never better than combustion unless carbon capture is added. More importantly, the economics of gasification versus combustion are not a function of fuel costs but capital costs.
- Unless the capital and operating costs associated with gasification can be lowered relative to combustion, gasification will not become economic under any pricing scenario that excludes carbon capture benefits.
- Figure D-5 shows that bitumen combustion is favored at WTI prices less than $61/bbl, but given the assumed bitumen price ratio a facility would lose money (*i.e.* negative Net Present Value) at these conditions. Bitumen may be preferred but only for an infeasible scenario.

Figure D-6 shows the same tipping chart representing how the break points shift when we add carbon capture to the facilities.

Figure D-6.
Alternative Fuels Selection Chart with Carbon Capture



Gasification of bitumen and coke replaces combustion as the economic energy choice. Carbon capture makes natural gas less attractive for the following reasons:

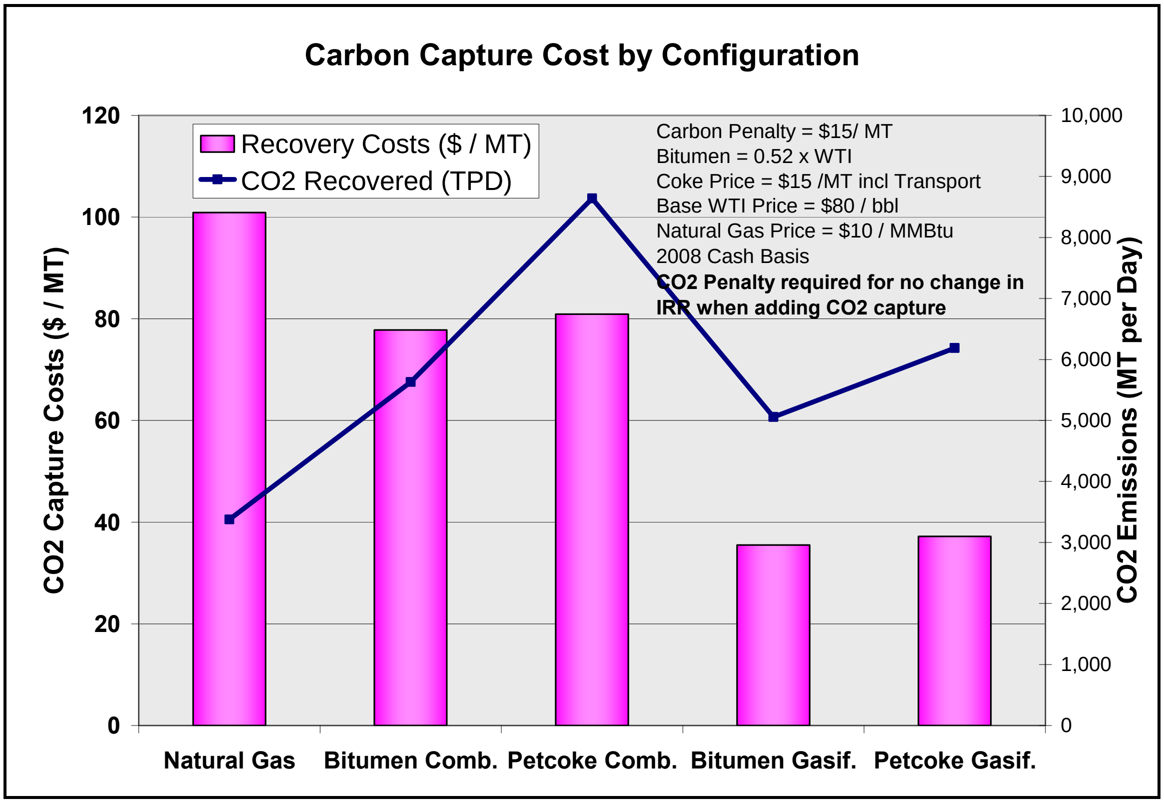
- Natural gas is the more expensive fuel source and carbon capture requires a ~10% increase in fuel firing.
- Natural gas benefits less from the reduction in CO_2 penalty (*i.e.* recovering 80% of natural gas CO_2 is worth less than recovering 80% of coke CO_2).
- Natural gas remains a post-combustion solvent extraction technology that is very energy intensive. Liquid and Solid gasification can utilize low energy precombustion separation.

The bitumen break point also reduces for carbon capture, primarily because bitumen gasification costs, while less than bitumen combustion capture costs, reduce feasibility compared to solids gasification.

Emission Control Costs

Figure D-7 shows the incremental cost of carbon capture for the five configurations. This figure is, in effect, the cost a carbon penalty would have to be to give a carbon capture case the same economic return as the non-capture case. Natural gas has the highest carbon capture costs because of higher fuel costs and less net reduction in CO_2 emissions. While costs on a per ton barrel basis are higher for natural gas, costs on a per barrel or per gigajoule basis are considerably lower.

Figure D-7.
Carbon Capture Penalty Required

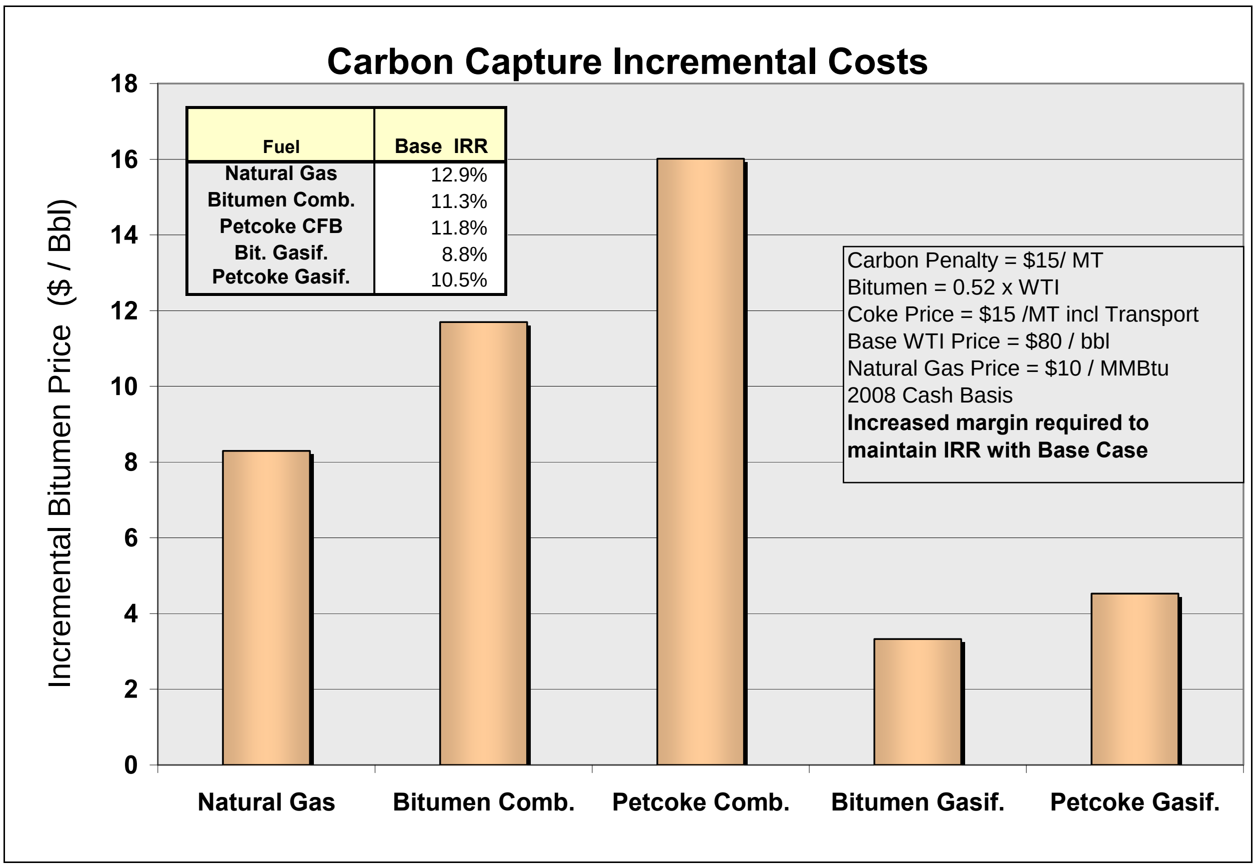


Another way to compare technologies is to calculate the increase in product price (in this case bitumen) required to make the project return (IRR) the same. In other words, "how much does oil price have to go up to cover the costs of CO_2 Capture?" In a sense, this is a way to quantify the economic "pain" that will be felt by a SAGD producer faced with carbon capture. If the project is economic at $60/bbl without carbon capture, how much higher does the oil price have to be with carbon capture?

Figure D-8 shows the cost of carbon capture in terms of the required increase in bitumen price. Thus, to maintain economic return for natural gas with carbon capture, bitumen price must increase by $8 per barrel. With reference to Table D-2, to increase Case 2 IRR from 9.2% to the Case 1 return of 12.9%, the product margin for Case 2 must be $8 per barrel higher. Assuming a 0.52 ratio, this delta corresponds to a WTI price increase of $15/bbl to $95/bbl. The cost increases significantly for alternative fuel combustion, reflecting the higher carbon dioxide generation per barrel of bitumen.

As expected, the crude price increase required for the gasification cases is much less, but this ignores the higher initial cost for gasification. Bitumen gasification cost to implement carbon capture is the smallest, but this increment is to maintain a rate of return that is already less than the discount rate of 10 percent.

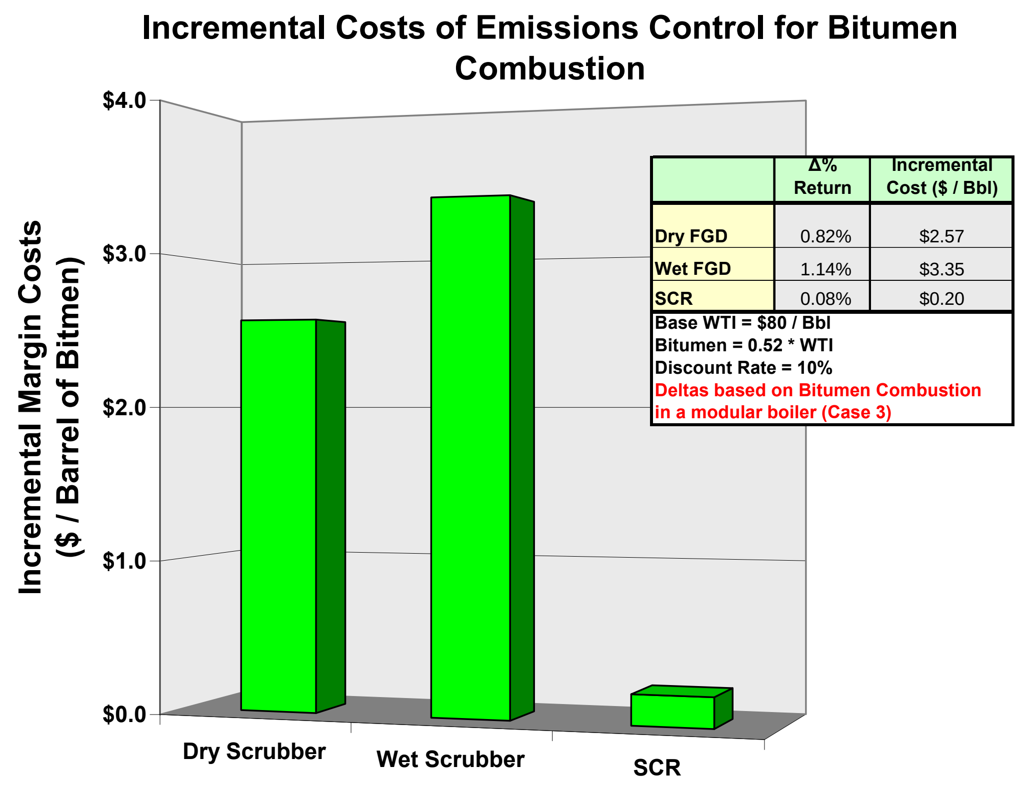
Figure D-8.
Carbon Capture Costs Basis Product Price Differential



In a similar fashion, it is possible to isolate emission treating requirements. Figure D-9 shows the relative costs of emission treating used for the bitumen ATB combustion case. Compared to carbon capture, controlling SOx and NOx is relatively inexpensive if they are viewed in terms of the overall facility construction. Adding an SCR to the project scope creates an incremental cost of $0.02/bbl bitumen. Flue Gas Desulphurization costs are higher, primarily because of the chemical costs for the limestone. However, the costs of going from a dry scrubber to a wet FGD is relatively small, requiring a bitumen price increase of less than a dollar per barrel.

Emission control costs and rate of return impact will be a much larger portion of the total economics for a revamp case (refer to Section E).

Figure D-9.
Emission Control Costs Basis Product Price



	Δ% Return	Incremental Cost ($ / Bbl)
Dry FGD	0.82%	$2.57
Wet FGD	1.14%	$3.35
SCR	0.08%	$0.20

Section E.

Configuration Sensitivities

Retrofit

As natural gas prices increase, producers may consider retrofitting natural gas boilers to burn alternative fuels. To address this scenario, we looked at retrofitting a typical natural gas SAGD facility to burn bitumen. We assumed that the OTSG boilers can be converted to bitumen ATB combustion, though this process has not been proven at scale. The retrofit configuration includes the DRU, the Wet Gas Scrubber, and the SCR NOx treating facility. Costs for emissions treating are increased in recognition of the additional labor and shutdown time required for a major SAGD retrofit. Likewise, some costs are included for boiler changes (new burners, repiping, controls, *etc*.).

Figure E-1.
Bitumen ATB Retrofit Economic Break-Even Point

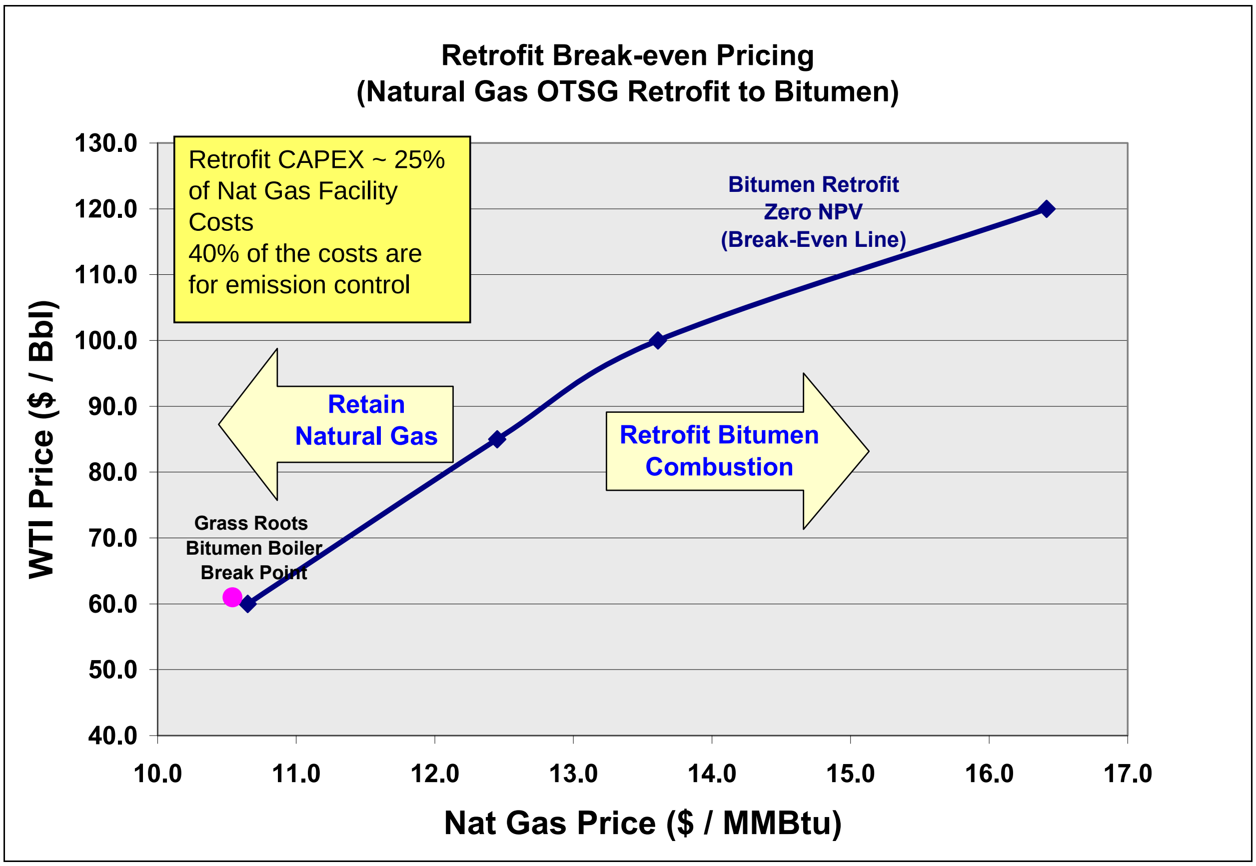


Figure E-1 shows the economic breakpoint for a retrofit (*i.e.* the crude and natural gas price necessary for a zero Net Present Value). The break-even WTI:NG price ratio for the retrofit case is approximately 5.5-6.0 (*i.e*. at a WTI price of $60, gas prices would need to be $10-

$11/MMBtu). This is similar to the break point for a new unit (refer Figure D-5). Typically, economics for retrofitting are less favorable than for a new facility (*i.e.* a lower crude price to gas price is required) because:

- Capital costs for a retrofit are higher than capital costs for a new unit.
- To make a retrofit economic, a producer would have to retain existing capacity and reduce operating costs sufficiently to cover the capital investment.
- Burning bitumen reduces production, so the improvement in operating costs has to cover lost production.

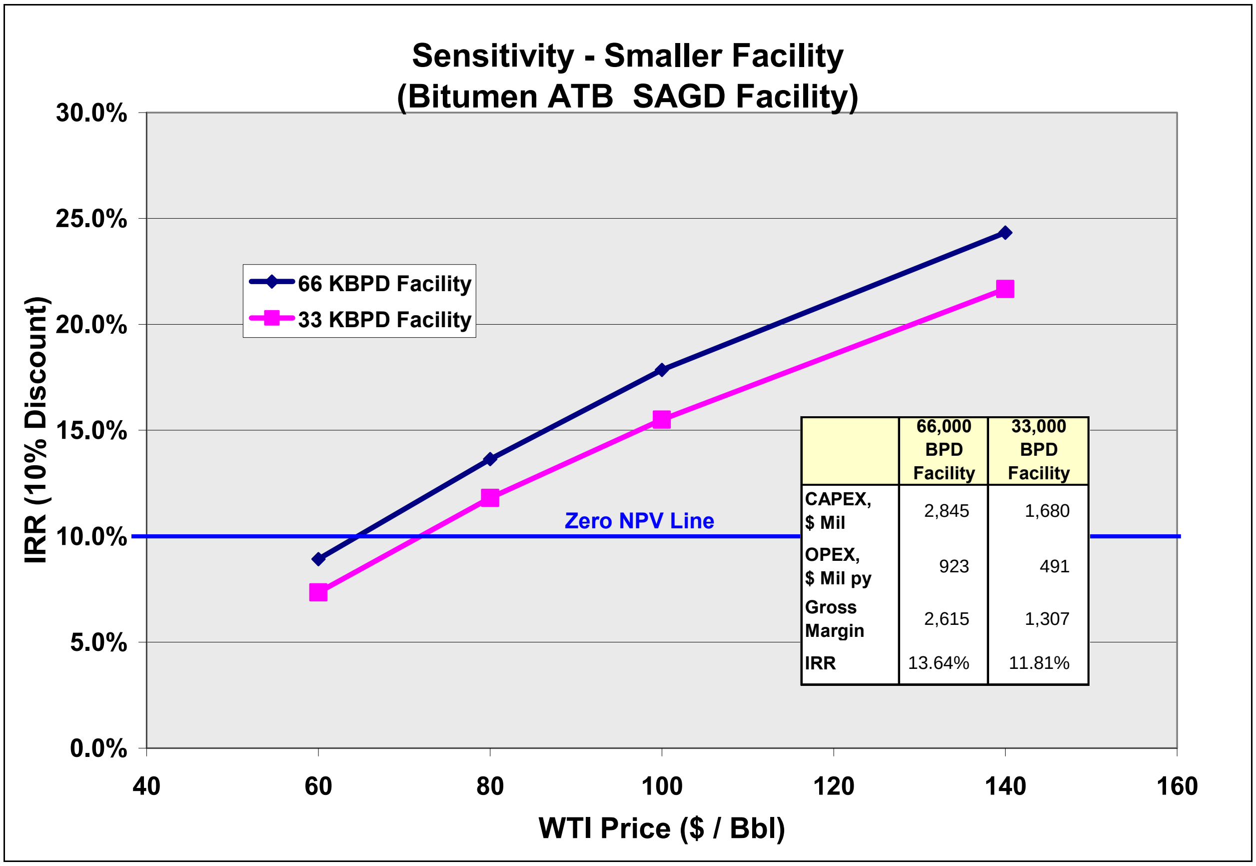
Economics for retrofitting in this case are improved by assuming an OTSG can be used as opposed to the modular boiler used in Case 3.

Given that the current spot market bitumen-to-crude price ratio is much higher than we have used (>0.7 vs. 0.52 used in the study), it is difficult to see how a retrofit to burn bitumen can make economic sense. Even if treating requirements are eliminated, a retrofit to bitumen would not be economic in the current environment. Retrofitting an OTSG facility to burn other alternative fuels (*e.g.* petcoke or coal) are well beyond current technical capabilities.

Reduced Scale Facilities

In past configuration and facility studies Jacobs Consultancy has conducted within and outside the Oils Sands region and for all major oil producers; the one consistent conclusion has been that bigger is usually better. Economies of scale strongly favor increasing facility sizes to a practical maximum, either in terms of upgrader capacity or oil handling capability. Figure E-2 demonstrates this conclusion and the level of lost return caused by selecting a facility half the size (33,000 BPD) as compared to Case 3 Bitumen ATB combustion. At base conditions there is a penalty of 1.83%, which is equivalent to an $11 differential in crude price. This means that for a smaller facility to be profitable, crude price needs to be substantially higher than for a larger facility.

Figure E-2.
Small Facility Economic Disadvantage



Carbon Capture Technology

Carbon capture technologies are still relatively unproven at this scale of Alberta SAGD facilities. Even the most developed technologies, such as Econamine®, are still in the demonstration phase for large energy producers. To address this uncertainty, we have looked at the economic impact of considering alternative technologies. Based on the screening work done in Phase 1 (refer to Volume 1), the project team selected Oxyfuel combustion and chilled ammonia technology for additional review.

Oxyfuel Combustion

A common alternative to post combustion carbon capture is oxygen combustion, which replaces combustion air with oxygen, usually after mixing with flue gas. Combusting with oxygen increases the concentration of CO_2 in flue gas to the point where condensing water and removing the trace contaminants (SOx and NOx) leaves a concentrated CO_2 stream suitable for compression and sequestration. Power requirements are significantly higher for the Air Separation Unit, but because there is no solvent regeneration requiring additional steam, bitumen combustion (and thus production) remains the same as in the non-capture case. For coke and natural gas facilities the savings would be in reduced fuel costs.

The large uncertainty in Oxyfuel combustion is the capital cost relative to amine flue gas treating. Air separation units are expensive, as are pressurized boilers and flue gas condensers. However, if the lower flue gas rate can be translated to reduced boiler size and increased CO_2 compressor suction pressure, there could be significant CAPEX and OPEX savings. Also, it may be possible to eliminate flue gas treating, recovering SOx and NOx as weak acids. Based on discussions with vendors there may not be significant boiler savings for oxyfuel combustion due to the need to maintain a maximum of 25% oxygen concentration at the boiler mixing point.

Chilled Ammonia Carbon Capture

Chilled ammonia carbon capture is a post combustion absorption technology similar to amine contacting but using chilled ammonia as a CO_2 solvent. This technology, offered by boiler manufacturer Alstom, offers the following advantages over amine absorption:

- Chilled ammonia requires less regeneration energy than amine absorption.
- Chilled ammonia can be regenerated at higher pressures, substantially reducing CO_2 compression capital and operating costs.
- Chilling costs in Alberta will be much lower than in the US, further increasing the OPEX advantage relative to amine technologies.
- Reduced heat requirements reduce boiler load, lowering net facility capital and operating costs.
- Power consumption and capital costs are lower for Carbon Capture. There is also a potential to integrate chilled ammonia carbon capture with flue gas desulphurization, utilizing the ammonia desulphurization technology.

Chilled Ammonia technology has a few issues that must be addressed before implementing in a full-scale SAGD facility:

- Ammonia slip—Loss of ammonia with stack gas needs to be controlled to avoid safety and environmental issues.
- Ammonia Handling—Delivery and storage of anhydrous ammonia always poses a significant safety hazard. However, fertilizer and petrochemical facilities throughout the world deal with this risk year round.

Economic Comparison

Table E-1 compares the economics of Oxyfuel and chilled ammonia against the base amine capture technology. According to an EPRI./ DOE study,[1] chilled ammonia can be installed for less than 50% of the capital costs of an Econamine® facility of the same capacity; however, the net capital savings are less since there is still a requirement for energy generation. Capital savings for Oxyfuel are even greater, because there is no additional heat load imposed on the steam system. However, Oxyfuel combustion has substantially higher operating costs, primarily for oxygen generation in the air separation unit. Based on making the appropriate adjustments to the Case 4 cash flow statement, we were able to calculate the impact these changes would have on project return and net present value.

[1] Chilled-Ammonia Post Combustion CO_2 Capture System, Electric Power Research Institute – Nov 2006

Table E-1.
Carbon Capture Sensitivities

	Amine CO2 Recovery (Case 4)	OxyFuel Combustion	Chilled Ammonia
ΔOPEX for CO2 Capture, Milion $/yr	40	52	23
ΔCAPEX for CO2 Capture, Million $	968	873	884
ΔNPV for CO2 Capture, Million $	-906	-708	-668
CO2 Capture Costs, $ / MT	71.2	85.5	58.7
CO2 Captured, MT/Day	5,631	4,757	5,194
Indirect CO2 (Imported Power), MT/Day	1,764	2,746	1,433
Total CO2 Emissions, MT/Day	3,172	3,935	2,731
ΔIRR for CO2 Capture, 10% Discount Rate	-3.91%	-3.21%	-3.07%

Economics shown at WTI Crude Price = $80 / Bbl
Delta Basis is Case 3 - Bitumen ATB Combustion

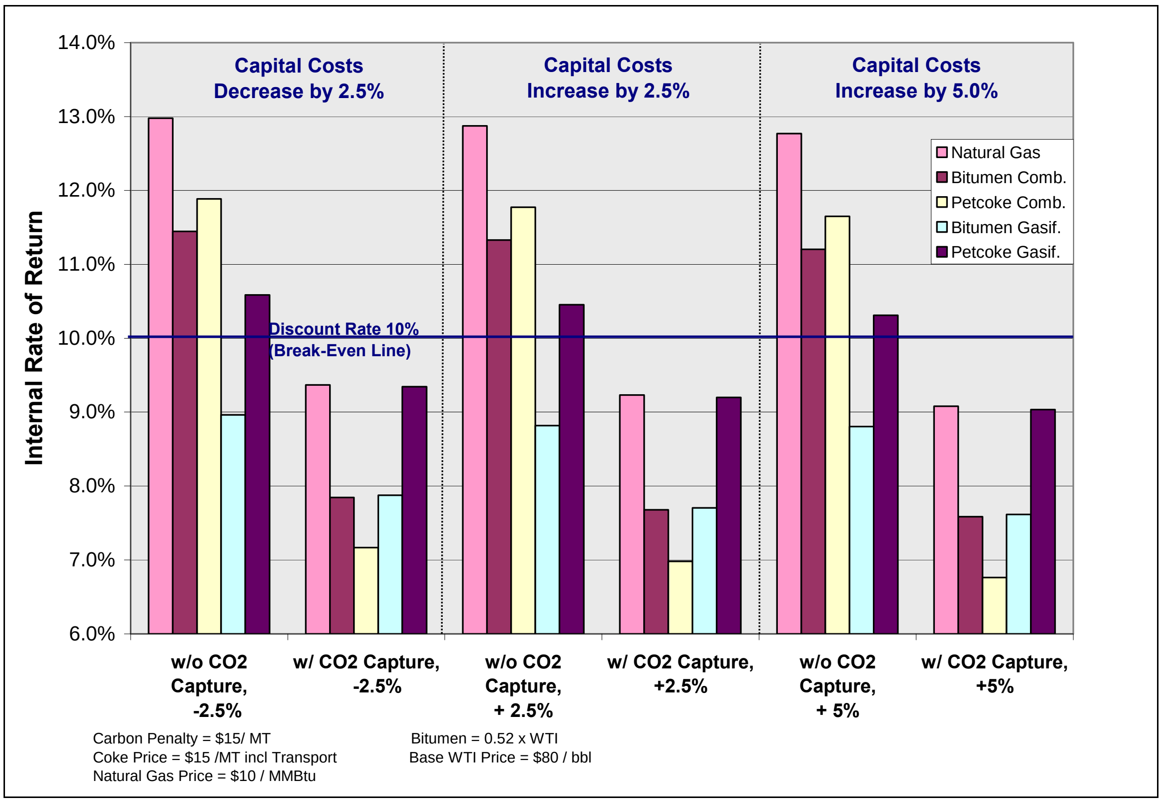
Capital Cost Escalation

Jacobs Consultancy has recently conducted several studies correlating recent trends in capital costs for new facilities. The general conclusions that we have noted are the following:

- Capital costs increases in the past five years have exceeded all historic projections of trends. Escalation has been higher than 10% a year, based on some indices.
- Much of the pressure on capital costs has come from the sudden rise in crude prices following several years of stagnant prices. Thus, project requests exceed the availability of engineers and manufacturers.
- Engineering firms have increased staffing as the number of proposed projects (along with crude price) has begun to decline. It is reasonable to assume capital costs trends will return to historic levels or even decline.

For this evaluation we have assumed a 2.5% capital escalation from 2008 to end of construction in 2014. Figure E-3 demonstrates the relative impact of differing escalation assumptions on economics. In effect, reducing escalation improves project economics, but the change is less than the accuracy of our results and much less significant than the price of natural gas and bitumen. Lower escalation favors the higher capital configurations such as gasification and petcoke combustion, but the impact is too small to significantly impact capital decisions for SAGD facilities. Unless there is a decline of 10% or more, it is not likely that capital escalation will have a significant impact on the relative returns as compared to the impact of bitumen and natural gas prices.

Figure E-3.
Impact of Cost Escalation

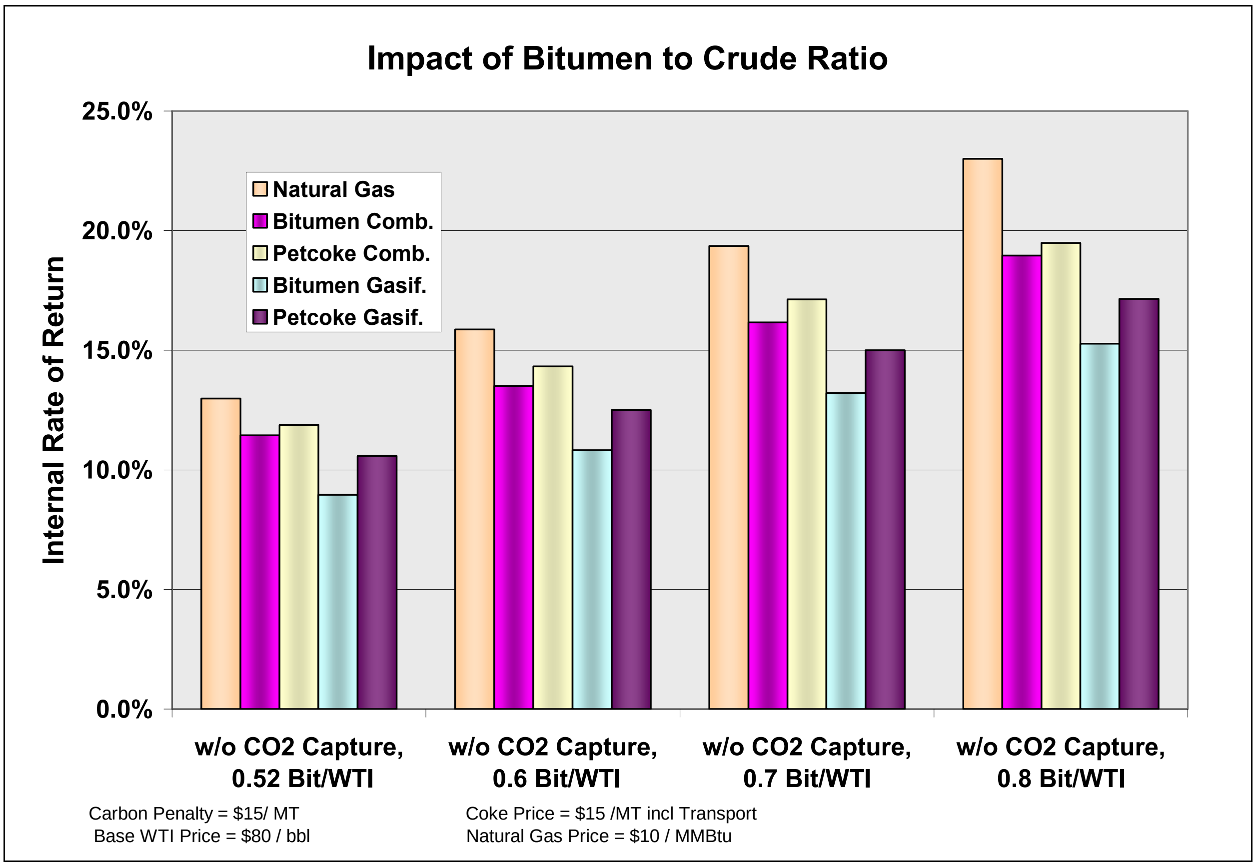


Bitumen Price Ratio

As noted, the project team has assumed a low bitumen to crude price ratio of 0.52, compared to the current ratio of greater than 0.7 (*i.e.* if WTI price is $50 we would currently project a bitumen price of $35 or greater). The low ratio was selected to provide bitumen combustion the best economic scenario. As demonstrated in Figure E-4, economics for bitumen combustion improve with a lower relative bitumen price.

The relative benefit of natural gas versus alternative fuels improves at higher bitumen price. This trend reflects high operating costs and low capital costs. Offsetting the operating costs with revenues significantly improves project return. The impact is more subtle for alternative fuels. Bitumen combustion and gasification benefit from the higher revenue with a smaller capital investment, but revenue gains are less than for the corresponding petcoke cases. The impacts appear to offset. Overall, a higher bitumen-to-crude price ratio improves the case for natural gas combustion. As expected, the IRR for all cases is greatly improved with higher bitumen prices.

Figure E-4.
Impact of Bitumen Price on Project Return



Section F.

Decision Matrix Analysis

Decision Matrix Methodology

The project team used the Failure Means and Effects Analysis (FMEA) to create a decision matrix ranking of the alternative fuels configurations (Refer Figure F-1). The purposes in ranking were to establish a preferred configuration as well as the criteria to be used as evaluating future configurations.

The Decision Matrix process consists of the following steps:

- *Assemble Team*—Assemble a multi-disciplinary audit team composed of stake-holders and technical experts, both from within and outside the client organization. It is critical to include as many people directly involved with the project facility as possible. The goal is both to create buy-in among stake-holders and to identify and address concerns surrounding the project.

- *Define Key Impacts*—Considering the proposed scope of the project, the team should define all parameters that could have an impact on any part of the organization environment as a whole. The project team will provide the majority of these impacts with the audit team adding any additional concern.

- *Assign Importance Ranking to Impacts*—The audit team will rank the importance of each impact relative to the refinery and organizational culture. This step is critical to identifying issues that are most important to a specific site. For purposes of this analysis, the numeric scale was as follows:

Must Have	➔	Project must meet this criteria
10	➔	Considered critical to project success
7	➔	Important to the site but not critical
4	➔	Somewhat important to the site
1	➔	Not of importance to the site

- *Rate Each Project for Each Impact*—Independent of the importance, criteria ratings are to be determined for each impact. The criteria should be specific to the impact, quantifiable and capable of differentiating the projects. The rating criteria will be selected to create an appropriate spread in project ratings, and the more positive the criteria the higher the rating. The project with the lowest operating cost should have the highest ranking for OPEX impact.

- *Calculate and Sum the Impacts*—For each project, the project rating for an impact is multiplied by the site importance rating for that impact. These products are then summed to determine the total rating for the project

Figure F-1.
Sample Decision Matrix

		Case 3 Bitumen Modular Boiler	Case 5 Petcoke Circulating Fluidized Bed	Case 7 Liquid Gasifi	Weighting	Case 5 Product (Rank x Weighting)
Carbon Capture Ready						
	CAPEX		1	10	4	4
	Energy Requirement	4	1	10	7	7
	Ease of Retrofit	4	1	7	7	7
		144.00	**18.00**	**159.00**	**6.00**	**18.00**

Impact

Case Rank

Weighting

Combined Score

Average Weighting

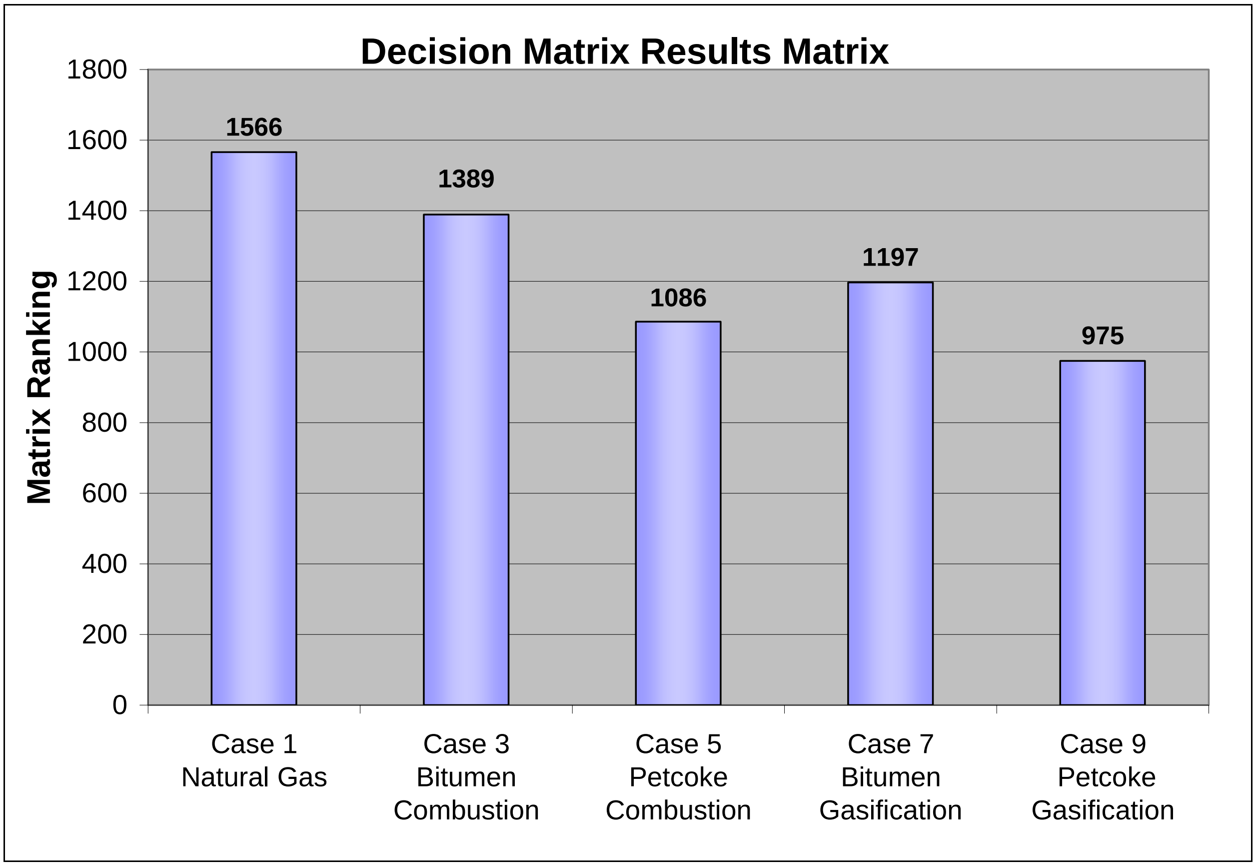
The Alternative Fuels FMEA compared the five base configurations along with their carbon capture capabilities. The categories selected for rating were as follows:

- Net Efficiency—Technologies that achieve the best performance in terms of bitumen production per Gigajoule of energy input should be favored. (Weighting: 10)
- Environmental Impacts—Several subcategories were used to define emission issues, foot print, waste water generation, and solid waste generation. (Weighting: 9.25)
- Safety Issues—This category addresses risk to both facility workers and the general public. (Weighting: 9.25)
- Independent of Natural Gas Use—This category favors configurations that do not require natural gas import. (Weighting: 7.0)
- Risk of Proliferation—This category was selected to give preferences to technologies that cannot be easily installed in a small scale. The team felt that environmental policy should favor fewer large-scale projects as opposed to multiple smaller-scale facilities. (Weighting: 7.0)
- Capital Cost—Recognizing none of the project team represented an owner company, capital cost was still considered critical to configuration selection because economic viability and the capability to implement environmental controls are related to capital cost and are an important part of the AENV/AERI objectives. (Weighting: 7.0)
- Operating Expense—Controlling operating costs is an important part of economic viability. (Weighting: 7.0)

- Operational Complexity—More complex configurations will have higher requirements for operating, maintenance and technical support, and are more prone to safety and environmental incidents. (Weighting: 7.0)
- Reliability—Facility reliability (estimated by equipment number and technical risk) refers to the capability of a configuration to meet target capacity while avoiding unplanned shutdowns. (Weighting: 6.0)
- Infrastructure Requirements—Recognizing that most oil sands projects are located in remote areas of Alberta, technologies requiring minimal infrastructure (roads, electrical supply, pipelines, *etc*.) are favored. (Weighting: 5.8)
- Carbon Capture Ready—Subcategories including CAPEX and ease of retrofit were selected to rank the capability of a given configuration to be adapted for carbon capture. (Weighting: 5.5)

Utilizing these categories, each configuration was ranked (note Carbon Capture cases were considered a category defining each case and therefore were not ranked independently). The full decision matrix is shown in Appendix 4. Figure F-2 shows the results of the ranking summing all categories. In effect, we give each subcategory full weighting as assigned by the FMEA team.

Figure F-2.
Decision Matrix Results



We can make the following conclusions regarding FMEA results:

- Case 1 and Case 2 (Natural Gas and Bitumen Combustion) are the top two options. Natural Gas is favored in Figure F-2 (all subcategories) because of a much stronger environmental profile, particularly regarding emissions. A modular boiler bitumen combustion has other benefits including smaller foot print, less equipment, and independence from natural gas.
- The overall difference in performance of all cases is relatively small (less than 50% of the maximum), suggesting that all configurations could be considered, provided certain areas of concern are addressed.
- Solid combustion and gasification were downgraded due to the difficult logistics of supplying more than 2000 TPD of petcoke using 40-ton trucks. If the logistics can be worked out (*e.g.* co-location of upgrader), the score of these case will improve.

- Gasification technologies, both liquid and solid, scored poorly in some categories because of their complexity and higher capital cost.
- The Carbon Capture category favors liquid gasification because of the relative ease of implementation and lower parasitic energy costs.

It should be noted that FMEA does not suggest that a technology or configuration is implicitly better, but instead points out the limitations of that configuration relative to other alternatives. The results should be with regard to the criteria that established the rankings. Rather than saying that a Petcoke CFB is not as desirable as a bitumen boiler, the FMEA should be used to identify the shortcomings of the CFB (carbon emissions, road traffic, *etc.*) and look into ways to correct these problems.

Section G.

Areas for Future Study

Strategic Considerations

Results of this study have shown that there is not a strong economic incentive to drive new SAGD facilities to alternative fuels until natural gas prices rise to historical highs (above $10/GJ). Emission treating is a relatively small part of the extra cost, and reducing the capital and operating expense of emissions control will have little impact on the economic viability of alternative fuels, especially when compared to the impact of natural gas and crude oil pricing. To reduce dependence on natural gas, the following high level issues will need to be investigated.

Address Water Treating Considerations

While evaporative water treating costs are higher than once-through steam generation, the waste water generation is less. Encouraging the industry (*e.g.* tax incentives, waste water reduction targets) to adopt evaporative water technologies will reduce waste water and improve the economics of alternative fuels.

Improve Gasification Capital Costs

Gasification has a number of positive environmental advantages such as removal of sulfur in elemental form and easier integration of CO_2 Capture. However, capital costs and complexity make it uncompetitive. Making gasification less expensive through technology research and centralization of facilities will be an important step towards implementing a government alternative fuels strategy.

Reduce Energy Costs of Carbon Capture

Even for natural gas facilities, the energy costs for carbon capture are currently more than the economic incentive to recover CO_2 (assumed $15/MT) meaning most producers will chose to pay the penalties. Technologies that can reduce regeneration and compression costs for CO_2 will be required to make carbon capture compatible with alternative fuels combustion.

Investigate Alternative Methods of Petcoke Transport

Based on projections of natural gas and crude prices, it is probable that petcoke will be the most economic fuel in the near future, especially if capital costs come down. To make large-scale utilization of petcoke or other solid fuels, a suitable means of transport must be determined. Potential solutions to be investigated are rail-lines, gasification at a central facility, and a slurry pipeline.

Consider Transport of Stranded Liquid Fuels

While bitumen combustion can be competitive, reducing product exports greatly diminishes facility profitability. An alternative method of liquid combustion is to transfer low value liquid products (*e.g.* SDA pitch, unconverted hydrocracker bottoms) from the upgraders in the industrial heartland to Oil Sands SAGD facilities. Like petcoke combustion, identifying lower cost fuels will improve the viability of alternative fuels.

Reduce CO_2 Emissions by Improving SAGD Efficiency

The SAGD process rejects significant amounts of “waste heat,” the heat with temperatures below what can be recovered in a useful form. The key to increasing heat recovery is to increase the temperature of the waste heat. New technologies, including reservoir operation, have made this step possible, and likely economically feasible. Further study is required to evaluate and rank these technologies and configurations.

Technologies to Be Investigated

Consistent with improving the viability of alternative fuels and increasing the efficiency of SAGD facilities, AERI and AENV should encourage research and implementation of the following technologies:

Circulating Dry Scrubbing

Wet scrubbing is technically and economically feasible, but the reliability in Alberta has not been established. There is a high potential for line freezing in sub-zero weather. Traditional dry scrubbing will not guarantee the recovery required for high sulfur fuels. Circulating dry scrubbing (CDS) may offer the increased recovery of wet scrubbing without the high water circulation risks. CDS is a commercial technology, but the facilities are few in number and have not demonstrated the reliability expected of other scrubbing technologies.

Improved Gasification

Several gasification technologies in development, such as Rocketdyne and Bluegas™, have the potential to greatly reduce the costs and reliability issues of gasification systems.

MSAR Technology

MSAR technology, licensed by Quadrise, has been shown to dramatically improve NOX emissions from liquid fuels, reducing emissions up to 50 percent. MSAR can also facilitate the transfer of low quality or stranded fuel oils to SAGD facilities, and may also allow producers to burn bitumen in OTSG facilities.

Chilled Ammonia

The key to improving the economics of carbon capture will be the ability to minimize parasitic energy costs, namely solvent regeneration and CO_2 compression. Chilled ammonia is a technology that can reduce both of these costs. Furthermore, chilling costs in Alberta will be significantly less than other locations, providing additional reductions in OPEX and CAPEX. .

Downhole Pump Technologies

While not specific to alternative fuels, improvements in the reliability of downhole pumps will improve the overall energy efficiency of SAGD facilities, reducing the amount of waste heat lost to the atmosphere.

Section H.

Conclusions

From our alternative fuels configuration study, we can make the following conclusions about the use of alternative fuels in Oil Sands production applications:

1. Alternative fuels combustion is economically feasible (*i.e.* yields a return greater than 10%) if product price is sufficiently high, but natural gas combustion will be favored in most pricing scenarios, based on historical trends.

2. Natural gas has the lowest air emissions and capital costs of all the configurations evaluated.

3. Bitumen combustion achieves many of the AERI and AENV objectives (Independent of Natural Gas, Lower Capital Costs, *etc.*) and can be retrofit with existing natural gas facilities. However, economic returns are less than natural gas and petcoke combustion because of lost revenue from burning bitumen. Currently bitumen price is high relative to crude and natural gas, which makes the economics of bitumen combustion difficult to justify. In effect, the bitumen configuration is squeezed between low bitumen prices where returns are too low to justify investment and high prices when other fuels (petcoke and natural gas) offer better returns.

4. Petcoke combustion in a CFB will become competitive with natural gas at higher crude and natural gas prices. Fuel transport logistics will be a major challenge to petcoke utilization.

5. Petcoke combustion produces double the CO_2 of natural gas while bitumen combustion produces 55% more CO_2 than natural gas. Even with 80% carbon capture, the petcoke combustion case has similar CO_2 emissions to the natural gas case **without** capture.

6. Gasification is not economically competitive with combustion technologies unless carbon capture is a requirement. In addition, increased plant complexity and capital costs will make it difficult to implement in Oil Sands production facilities.

7. For new plants, air emission control investments do not significantly impact economics.

8. **Carbon capture has the equivalent impact of an $8/bbl drop in the price of bitumen** for a natural gas SAGD facility. The drop increases to $12 for bitumen combustion and $16/bbl for petcoke combustion. CO_2 capture configurations, gasification and combustion are all economically infeasible (below a 10% IRR at $80/bbl WTI).

9. The assumed economic penalty for CO_2 emissions ($15 per MT) does not cover the increased operating expense associated with carbon capture.

The following steps can be taken to make alternative fuel combustion more competitive:

1. Investigate alternative supply methods for bringing liquid and solid fuels to the Oil Sands production areas.
2. Encourage research into lower energy-intensive methods of carbon capture such as chilled ammonia scrubbing and Oxyfuel combustion.
3. Test Circulating Dry Scrubbers in Alberta flue gas treating applications to determine their effectiveness and reliability relative to wet flue gas scrubbers.
4. Evaluate technology to make gasification more cost effective and less complex relative to combustion.
5. Study process and operating modifications to improve the overall energy efficiency of steam generation and bitumen production.

While escalation did not shift comparative results, lower capital costs will substantially improve economic viability. Government and the industry can best meet objectives for conserving natural gas resources by encouraging and coordinating alternative fuels investment during economic downturns, when labor rates and equipment prices are more reasonable.

Appendix 1.

Assay Information

Bitumen Assay—Jacobs' Representative Athabasca Bitumen

The bitumen assay is a blend of various athabasca bitumens to give a representative assay for both the SAGD CPF and the upgrader.

Table B-1.
Jacobs Representative Athabasca Bitumen Assay

General Properties		Distillation Curve (TBP Distillation)		
Properties		**Distillation (Volume)**	**Units**	**Results**
API Gravity	8.2	IBP	ºC / ºF	217 / 423
Relative density @ 15ºC, g/ml	1.0129	5%	ºC / ºF	313 / 595
Absolute Density@15ºC, kg/m3	1012.3	10%	ºC / ºF	348 / 658
Ash content, mass %	0.04	15%	ºC / ºF	375 / 707
Asphaltene C5 insoluble mass%	14.7	30%	ºC / ºF	450 / 843
Asphaltene C7 insoluble mass%	7.9	50%	ºC / ºF	556 / 1033
Microcarbon Residue, mass %	18.6	70%	ºC / ºF	670 / 1238
Total Nitrogen, ppm wt	4851	90%	ºC / ºF	840 / 1544
Basic Nitrogen, ppm wt	1598	95%	ºC / ºF	919 / 1868
Particulates, mass %	0.036			
Sediment & Water, volume fraction	< 0.001			
Total Sulphur, mass%	4.8			
Viscosity				
cSt @ 15.5ºC/ 60ºF	1.1E+06			
cSt @ 37.8º / 100ºF	4.6E+04			
cSt @ 100ºC / 212ºF	296.5			
Water Content, mass %	0.2			
Metals				
Copper, ppm wt	<0.03			
Iron, ppm wt	31			
Nickel, ppm wt	77			
Sodium, ppm wt	5.4			
Vanadium, ppm wt	222			
Manganese, ppm wt	0.08			

Diluent Properties—Treated Naphtha

The diluent properties used in the study are as follows:

Table B-2.
Diluent Properties

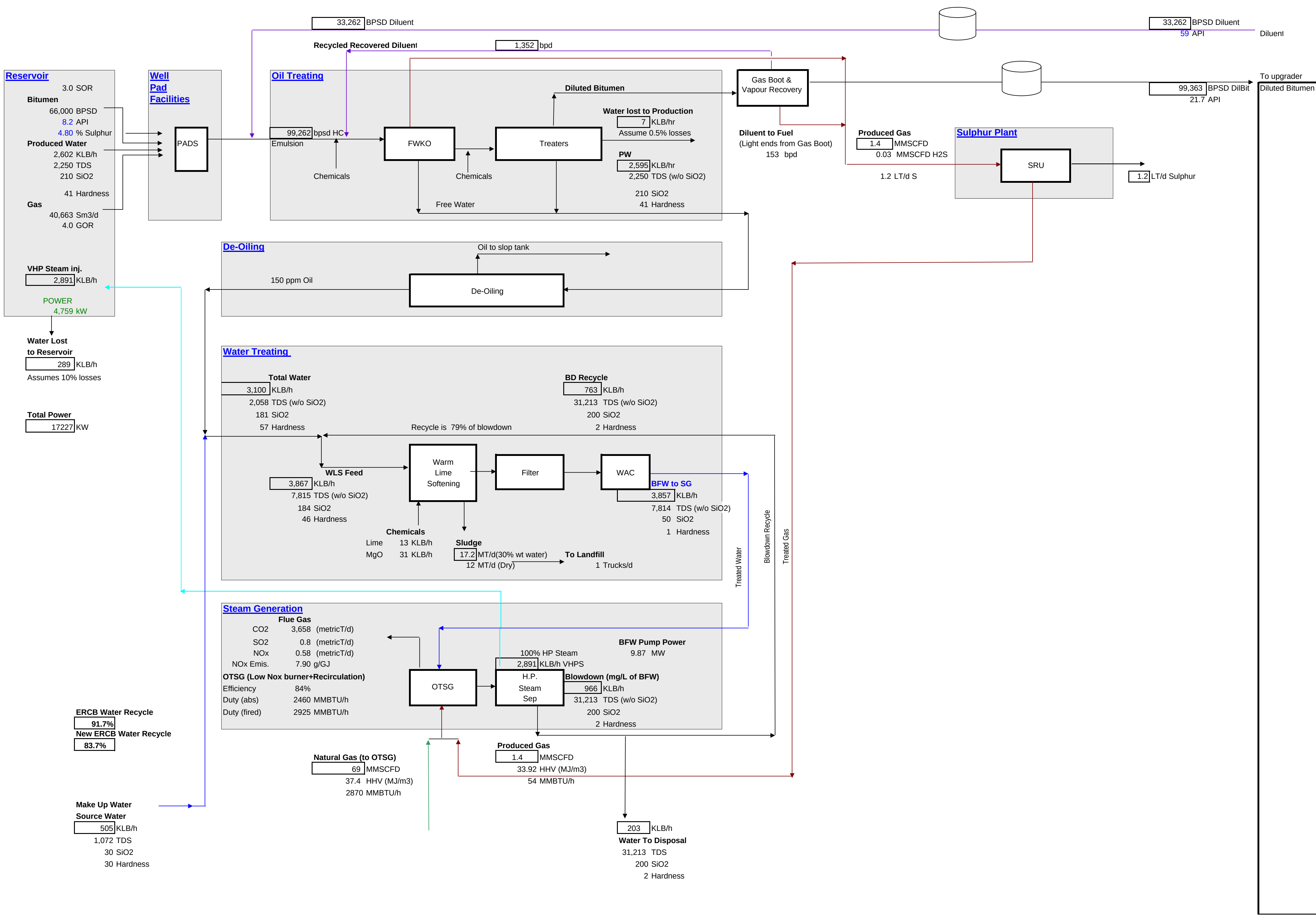
Treated Naphtha		
SG (15.6°C)		0.7443
API		59
Sulfur, wppm		20
Nitrogen, wppm		≤1
PONA		
Paraffins, Vol%		65
Olefins, Vol%		--
Naphthenes, Vol%		28
Aromatics, Vol%		7
RVP, kPa (a)		11
Water Content, wppm		Nil
Distillation		
TBP Dist., °C		
	5% vol	67
	10%	82
	30%	111
	50%	130
	70%	148
	90%	169
	95%	182

Appendix 2.

Block Flow Diagrams

AERI / AENV Alternative Fuel Study BFD

Case 1 - SAGD 66,000 bpsd
Base Case: Natural Gas - OTSG



Water Balance

Stream	Flow (m^3/sd)	TDS (w/o SiO2) (ppm)	SiO2 (ppm)	Hardness (ppm)
Steam to Reservoir	31,471	0	0	0
Losses to Reservoir	(3,147)			
Produced Water	28,324			
Losses to Production	(79)			
BD Recycle	8,302	7,815	184	46
M.U. Water	5,500	1,072	30	141
WLS Feed	42,088	7,815	184	46
AF Recycle	40	7,815	184	46
Regen Water	(20)	7,815	184	46
Sludge	(28)			
BFW	42,080	7,814	50	1
Steam to Pads	31,552	0	0	0
Blowdown	10,528	31,213	200	2

Emissions Summary

	SO2	S	CO2	NOX	
Emissions From OTSG	0.8	0.38	3,658	0.58	mt/d
Recovered Sulphur		1.2			mt/d
Imported Power			253		
Total CO2 Emissions			3911		mt/d

LEGEND
- Process Gas
- Bitumen / Emulsion / Process Water
- Diluent
- Condensate / BFW
- Steam
- Natural Gas Imports
- CO2 for Export

Rev 9
By: Ian Buchanan / D.Westphalen
Date: 12/22/2008
Project #: JC102910

APPENDIX 2: BLOCK FLOW DIAGRAMS

Utility Summary for AERI Alternative Fuel Study

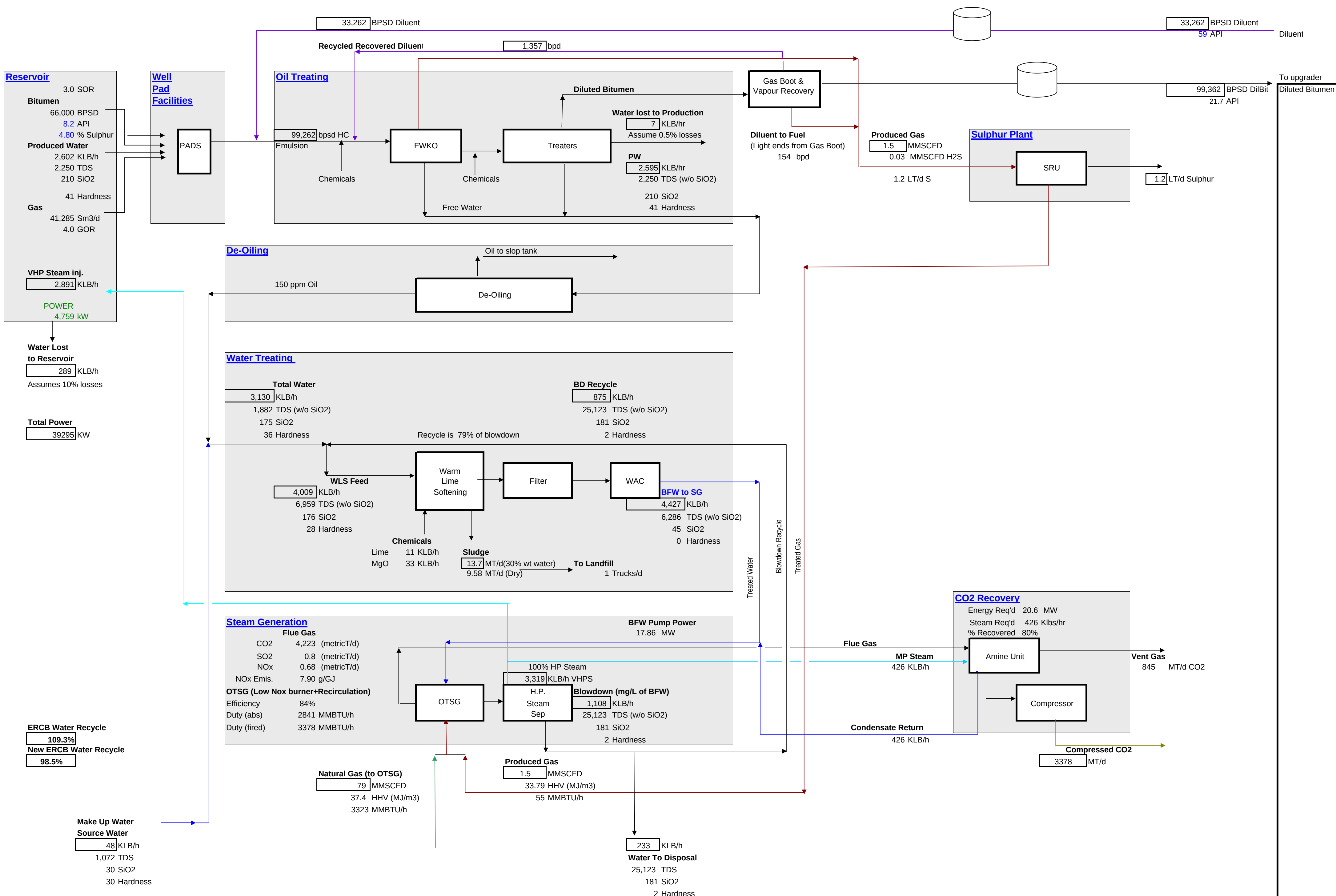
Case 1

Utility Summary Project: AERI Alternative Fuel Study Case 1 - Natural Gas fed Steam generator (OTSG),WLS Revision: 0	Capacity		Elec	Steam									Fuel Consumed		CO2 Generated	Recovered CO2
Process Unit	**Capacity**	**Units**	**Power KW**	**Steam** 1600 psig k#/hr	**Steam** 600 psig k#/hr	**(positive = generation)** 250 psig k#/hr	50 psig k#/hr	BFW k#/hr	Raw Water k#/hr	Cond. k#/hr	Waste W k#/hr	Loss k#/hr	**Produced Gas** MMBTU/hr	**Natural Gas** MMBTU/h	**CO2** MT/day	**CO2** MT/d
Oil Treating/Deoiling																
Pumps			398	0	0	0	0	0	0	0	0	0	0	0	0	0
Steam Generation																
OTSG	31471	m3/day CWE Steam	9,866	2,891	0	0	0	0	0	0	0	0	54	2870	3,658	0
Water Treating																
WLS	42088	m3/day CWE BFW	398	0	0	0	0	0	0	0	0	0	0	0	0	0
Well Pads																
Steam Injection to Wells	31471	m3/day CWE BFW		-2,891	0	0	0	0	0	0	0	0				
Wellbore Pumps	244139	BPD	4,759													
Offsites																
Sulphur Plant	1.2	MT/d	5	0	0	0	0	0	0	0	0	0	0	0	0	0
Glycol System	7636	gpm	825													
Misc	-	-	975	0	0			0	0	0	0	0	0	0	0	0
Steam Balance																
Dearators	0	KLB BFW	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Demin Plant		KLB/h	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HP MP Letdowns	0	KLB/h	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MP LP Letdowns	0	KLB/h	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Indirect CO2 (Imported Power)															253	
TOTALS			**17,227**	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**54**	**2,870**	**3,911**	**0**

Notes

AERI / AENV Alternative Fuel Study BFD

Case 2 - SAGD 66,000 bpsd
Fuel: Natural Gas - OTSG



Water Balance

Stream	Flow (m^3/sd)	TDS (w/o SiO2) (ppm)	SiO2 (ppm)	Hardness (ppm)
Steam to Reservoir	31,471	0	0	0
Losses to Reservoir	(3,147)			
Produced Water	28,324			
Losses to Production	(79)			
BD Recycle+ AF Recycle	9,565	6,959	176	28
Condensate from Flue gas	5,308			
M.U. Water	517	1,072	30	141
WLS Feed	43,635	6,959	176	28
AF Recycle	(40)	6,959	176	28
Condensate return	4,658			
Regen Water	(40)	6,959	176	28
Sludge	(28)			
BFW	48,185	6,286	45	0
Steam to Users	36,129	0	0	0
Blowdown	12,056	25,123	181	2

Emissions Summary 0.68

	SO2	S	CO2	NOX	
Emissions From OTSG	0.8	0.38	4,223	0.68	mt/d
Recovered Sulphur		1.2			mt/d
Recovered CO2			-3,378		
Imported Power			578		
Total CO2 Emissions			1,422		mt/d
Recovered CO2			3,378		mt/d

LEGEND

- Process Gas
- Bitumen / Emulsion / Process Water
- Diluent
- Condensate / BFW
- Steam
- Natural Gas Imports
- CO2 for Export

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By: Ian Buchanan / M. Armstrong
Date: 12/22/2008
Project #: JC102910

APPENDIX 2: BLOCK FLOW DIAGRAMS

Utility Summary for AERI Alternative Fuel Study

Case 2

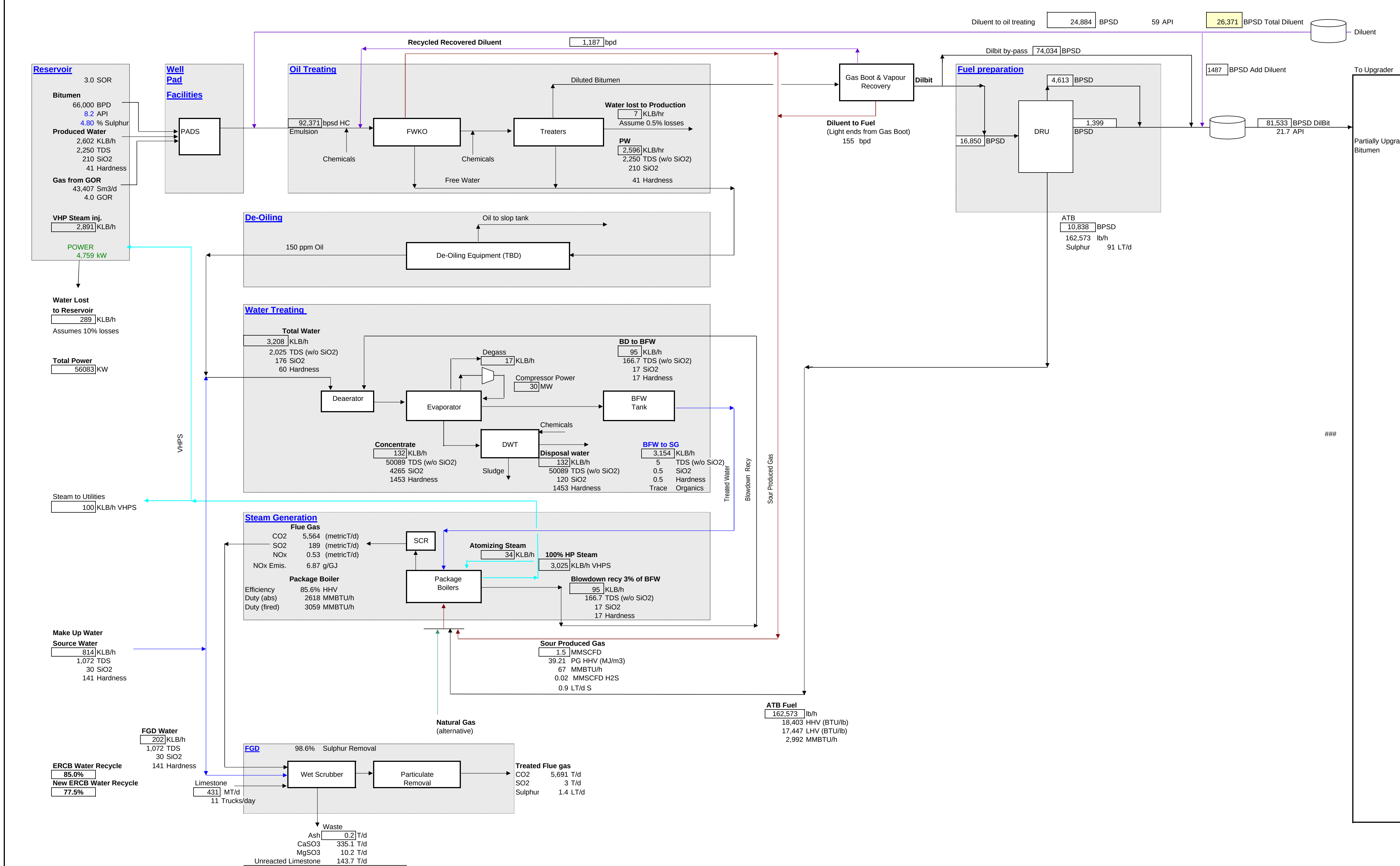
Utility Summary Project: AERI Alternative Fuel Study Case 1 - Natural Gas fed Steam generator (OTSG),WLS Revision: 0	Capacity		Elec	Steam									Fuel Consumed		CO2 Generated	Recovered CO2
Process Unit	**Capacity**	**Units**	**Power** KW	**Steam** 1600 psig k#/hr	**Steam** 600 psig k#/hr	**(positive = generation)** 250 psig k#/hr	50 psig k#/hr	BFW k#/hr	Raw Water k#/hr	Cond. k#/hr	Waste W k#/hr	Loss k#/hr	**Produced Gas** MMBTU/hr	**Natural Gas** MMBTU/h	**CO2** MT/day	**CO2** MT/d
Oil Treating/Deoiling																
Pumps			418	0	0	0	0	0	0	0	0	0	0	0	0	0
Steam Generation																
OTSG	31471	m3/day CWE Steam	11,316	3319	0	0	0	0	0	0	0	0	55	3323	4223	3378
Water Treating																
WLS	43635	m3/day CWE BFW	402	0	0	0	0	0	0	0	0	0	0	0	0	0
Well Pads																
Steam Injection to Wells	31471	m3/day CWE BFW		-2891	0	0	0	0	0	0	0	0				
Wellbore Pumps	244139	BPD	4,759													
Offsites																
Sulphur Plant	1.2	MT/d	5	0	0	0	0	0	0	0	0	0	0	0	0	0
Glycol System	7342	gpm	729													
Misc	-	-	1,058	0	0			0	0	0	0	0	0	0	0	0
CO2 Recovery																
Amine Unit	4223	MT/d CO2	1,351			-426										
CO2 Compressor	3378	MT/d CO2	19,257													
Steam Balance																
Dearators	0	KLB BFW	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Demin Plant		KLB/h	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Boiler Feed Water Letdown				-332	0	332										
HP MP Letdowns	0	KLB/h	0	-96	93		0	0	0	0	0	3	0	0	0	0
MP LP Letdowns	0	KLB/h	0	0	-93	93	0	0	0	0	0	0	0	0	0	0
Indirect CO2 (Imported Power)															578	
TOTALS			**39,295**	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**3**	**55**	**3,323**	**4,801**	**3,378**

Notes

% CO2 Recovery **80%**

AERI / AENV Alternative Fuel Study BFD

Case 3 - SAGD 66,000 bpsd
Fuel: ATB Neat



Water Balance

Stream	Flow (m³/sd)	TDS (w/o SiO2) (ppm)	SiO2 (ppm)	Hardness (ppm)
Steam to Reservoir	31,471	0	0	0
Losses to Reservoir	(3,147)	0	0	0
Produced Water	28,324	2,250	210	41
Losses to Production	(72)			
BD Recycle	1,030	0	0	0
M.U. Water	6,662	1,072	30	141
Evaporator Feed	35,943	2,025	176	60
Degass Loss	(180)	0	0	0
Concentrate	(1438)	50,089	4265	1453
BFW to SG	34,325	0	0	0
Steam to Users	33,308	0	0	0
Blowdown	1,030	0	0	0

ATB Analysis

BPSD	10,838
lb/d	3,901,764
SG	1.027
API	6.262
Sulphur, wt%	5.17

Emissions Summary

	SO2	S	CO2	NOX	
From PB	189.2	94.6	5,564	0.5	mt/d
From DRU			51		
Treated Flue Gas	2.7	1.4	5,615	0.5	mt/d
FGD			127		mt/d
Imported Power			824		
Total CO2 Emissions			6,567		mt/d

Notes

LEGEND
- Process Gas
- Bitumen / Emulsion / Process Water
- Diluent
- Condensate / BFW
- Steam
- Natural Gas Imports
- CO2 for Export

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By: Ian Buchanan / D.Westphalen
Date: 12/22/2008
Project #: JC2910

APPENDIX 2: BLOCK FLOW DIAGRAMS

Utility Summary for AERI Alternative Fuel Study

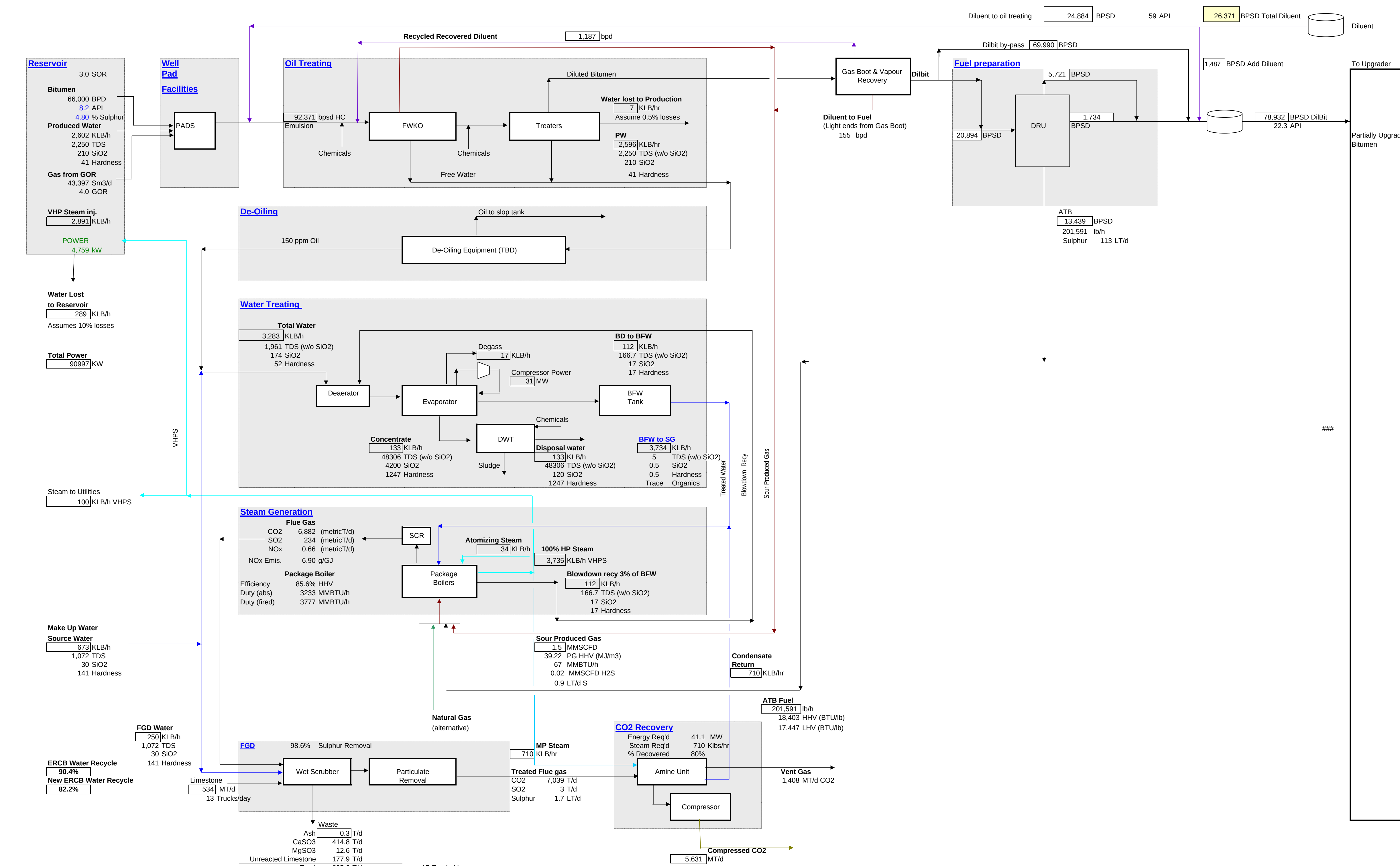
Case 3

Utility Summary Project: AERI Alternative Fuel Study Phase 2 Case 3 - Atmospheric Resid Neat Alternative Fuel Revision: 0	Capacity		Elec	Steam									Fuel Consumed		ATB Burned		CO2 Generated	Recovered CO2
Process Unit	**Capacity**	**Units**	**Power KW**	**Steam** 1600 psig k#/hr	**Steam** 600 psig k#/hr	**(positive = generation)** 250 psig k#/hr	50 psig k#/hr	BFW k#/hr	Raw Water k#/hr	Cond. k#/hr	Waste W k#/hr	Loss k#/hr	**Produced Gas** MMBTU/hr	**Natural Gas** MMBTU/h	lb/h	lb/h	**CO2** MT/day	**CO2** MT/d
Oil Treating/Deoiling																		
Pumps			343	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Steam Generation																		
Package Boilers	34332	m3/day CWE Steam	10,826	3025	0	0	-34	0	0	0	0	0	67	0	162,573	0	5,564	0
FGD			5,650														127	0
Water Treating																		
Evaporators	35950	m3/day CWE BFW	30,370	0	0	0	-74	0	0	0	0	0	0	0		0	0	0
Well Pads																		
Steam Injection to Wells	31471	m3/day CWE BFW		-2891	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Wellbore Pumps	244156	BPD	4,759	0	0	0	0	0	0	0	0	0						
Offsites																		
Atm Flash (DRU)	16850	BPD Bitumen	456	0	0	0	-26	0	0	0	0	0	0	44	0	0	51	0
Glycol System	4668	gpm	504															
Misc	-	-	3,175	0	0			0	0	0	0	0						
Steam Balance																		
Dearators	0	KLB BFW	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Demin Plant		KLB/h	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
VHP HP Letdowns		KLB/h	0	-134	134													
HP MP Letdowns		KLB/h	0	0	-134	134	0	0	0	0	0	0	0	0	0	0	0	0
MP LP Letdowns		KLB/h	0	0	0	-134	134	0	0	0	0	0	0	0	0	0	0	0
Indirect CO2																	824	
TOTALS			**56,083**	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**67**	**44**	**162,573**	**0**	**6,567**	**0**

Notes

AERI / AENV Alternative Fuel Study BFD

Case 4 - SAGD 66,000 bpsd
Fuel: ATB Neat



Water Balance

Stream	Flow (m³/sd)	TDS (w/o SiO2) (ppm)	SiO2 (ppm)	Hardness (ppm)
Steam to Reservoir	31,471	0	0	0
Losses to Reservoir	(3,147)	0	0	0
Produced Water	28,324	2,250	210	41
Losses to Production	(72)			
BD Recycle	1,219	0	0	0
Condensate from flue gas	2,068			
M.U. Water	4,608	1,072	30	141
Evaporator Feed	36,153	1,937	168	50
Degass Loss	(181)	0	0	0
Condensate Returen	6,120			
Concentrate	(1446)	48,306	4200	1247
BFW to SG	40,646	0	0	0
Steam to Users	39,426	0	0	0
Blowdown	1,219	0	0	0

ATB Analysis

BPSD	13,439
lb/d	4,838,178
SG	1.027
API	6.26
Sulphur, wt%	5.17

Emissions Summary

	SO2	S	CO2	NOX	
From PB	233.8	116.9	6,882	0.7	mt/d
From DRU			51		
Treated Flue Gas	3.3	1.7	6,933	0.7	mt/d
FGD			157.3		mt/d
Recovered CO2			-5,631		mt/d
Direct CO2 Emissions			1,459		mt/d
Imported Power			1,338		mt/d
Total CO2 Emissions			2,797		mt/d
Recovered CO2			5,631		mt/d

Notes

LEGEND
- Process Gas
- Bitumen / Emulsion / Process Water
- Diluent
- Condensate / BFW
- Steam
- Natural Gas Imports
- CO2 for Export

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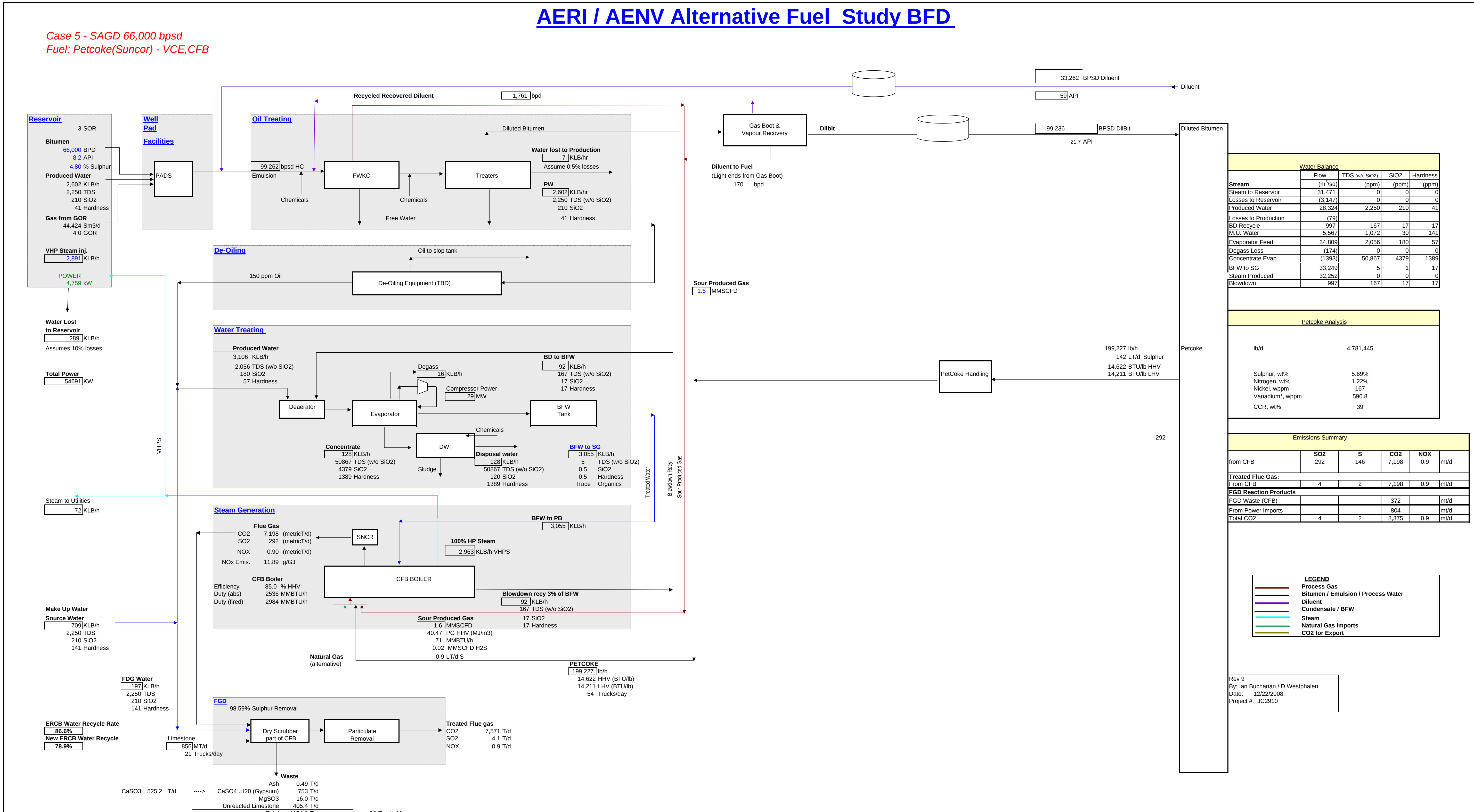
Utility Summary for AERI Alternative Fuel Study

Case 4

Utility Summary Project: AERI Alternative Fuel Study Phase 2 Case 3 - Atmospheric Resid Neat Alternative Fuel Revision: 0	Capacity		Elec	Steam									Fuel Consumed		ATB Burned		CO2 Generated	Recovered CO2
Process Unit	**Capacity**	**Units**	**Power KW**	**Steam** 1600 psig k#/hr	**Steam** 600 psig k#/hr	**(positive = generation)** 250 psig k#/hr	50 psig k#/hr	BFW k#/hr	Raw Water k#/hr	Cond. k#/hr	Waste W k#/hr	Loss k#/hr	**Produced Gas** MMBTU/hr	**Natural Gas** MMBTU/h	lb/h	lb/h	**CO2** MT/day	**CO2** MT/d
Oil Treating/Deoiling																		
Pumps			343	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Steam Generation																		
Package Boilers	40646	m3/day CWE Steam	10,826	3,735	0	0	-34	0	0	0	0	0	67	0	201,591	0	6,882	5,506
FGD			5,650														157	126
Water Treating																		
Evaporators	36153	m3/day CWE BFW	30,370	0	0	0	-74	0	0	0	0	0	0	0		0	0	0
Well Pads																		
Steam Injection to Wells	31471	m3/day CWE BFW		-2,891	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Wellbore Pumps	244156	BPD	4,759	0	0	0	0	0	0	0	0	0						
Offsites																		
Atm Flash (DRU)	16850	BPD Bitumen	456	0	0	0	-26	0	0	0	0	0	0	44	0	0	51	0
Glycol System	4973	gpm	504															
Misc	-	-	3,175	0	0			0	0	0	0	0						
CO2 Capture																		
Amine Unit	7039	MTD CO2	2,816			-710												
Compressor	5631	MTD CO2	32,099															
Steam Balance																		
Dearators	0	KLB BFW	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Demin Plant		KLB/h	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Condensate Flash		KLB/h		-374		374												
VHP HP Letdowns		KLB/h	0	-471	471													
HP MP Letdowns		KLB/h	0	0	-471	471	0	0	0	0	0	0	0	0	0	0	0	0
MP LP Letdowns		KLB/h	0	0	0	-135	135	0	0	0	0	0	0	0	0	0	0	0
Indirect CO2 (Power Import)																	1,338	
TOTALS			**90,997**	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**67**	**44**	**201,591**	**0**	**8,428**	**5,631**

Notes

% Recovery 80%



Water Balance

Stream	Flow (m^3/sd)	TDS (w/o SiO2) (ppm)	SiO2 (ppm)	Hardness (ppm)
Steam to Reservoir	31,471	0	0	0
Losses to Reservoir	(3,147)	0	0	0
Produced Water	28,324	2,250	210	41
Losses to Production	(79)			
BD Recycle	997	167	17	17
M.U. Water	5,567	1,072	30	141
Evaporator Feed	34,809	2,056	180	57
Degass Loss	(174)	0	0	0
Concentrate Evap	(1393)	50,867	4379	1389
BFW to SG	33,249	5	1	17
Steam Produced	32,252	0	0	0
Blowdown	997	167	17	17

Emissions Summary

	SO2	S	CO2	NOX	
from CFB	292	146	7,198	0.9	mt/d
Treated Flue Gas:					
From CFB	4	2	7,198	0.9	mt/d
FGD Reaction Products					
FGD Waste (CFB)			372		mt/d
From Power Imports			804		mt/d
Total CO2	4	2	8,375	0.9	mt/d

Petcoke Analysis

lb/d	4,781,445
Sulphur, wt%	5.69%
Nitrogen, wt%	1.22%
Nickel, wppm	167
Vanadium*, wppm	590.8
CCR, wt%	39

APPENDIX 2: BLOCK FLOW DIAGRAMS

Utility Summary for AERI Alternative Fuel Study

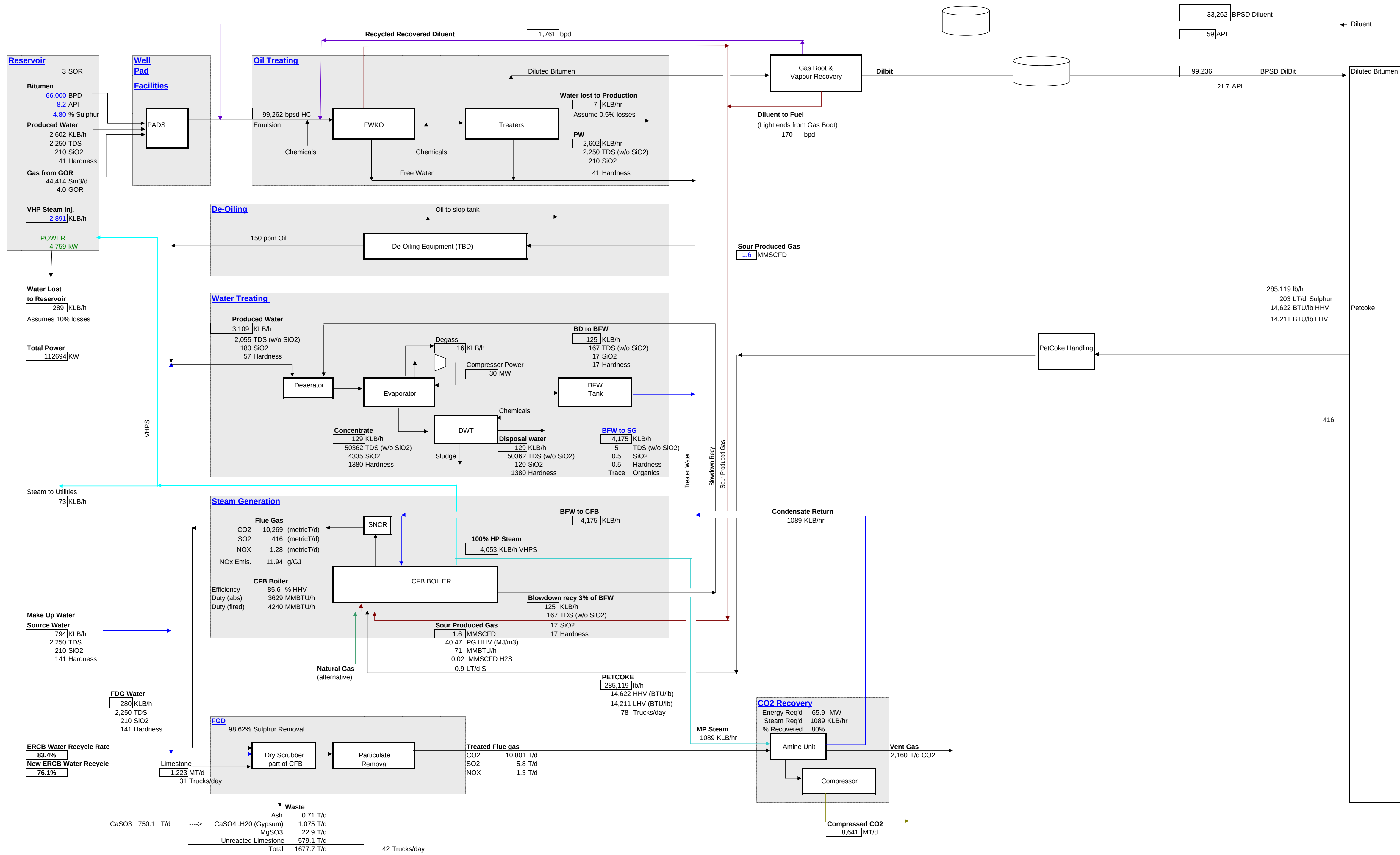
Case 5

Utility Summary Project: AERI Alternative Fuel Study Case 5 - Petcoke Alternative Fuel Revision: 0	Capacity		Elec	Steam									Fuel Consumed		PETCOKE Burned		CO2 Generated	Recovered CO2
Process Unit	**Capacity**	**Units**	**Power KW**	**Steam** 1600 psig k#/hr	**Steam** 600 psig k#/hr	**(positive = generation)** 250 psig k#/hr	50 psig k#/hr	BFW k#/hr	Raw Water k#/hr	Cond. k#/hr	Waste W k#/hr	Loss k#/hr	**Produced Gas** MMBTU/hr	**Natural Gas** MMBTU/h	lb/h	lb/h	**CO2** MT/day	**CO2** MT/d
Oil Treating/Deoiling																		
Pumps			403	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Steam Generation																		
CFB Boiler	33249	m3/day CWE Steam	10,827	2963	0	0	0	0	0	0	0	0	71	0	199,227	0	7,198	0
FGD			5,650														372	
Water Treating																		
Evaporators	34816	m3/day CWE BFW	29,412	0	0	0	-72	0	0	0	0	0	0	0		0	0	0
Well Pads																		
Steam Injection to Wells	31471	m3/day CWE BFW		-2891	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Wellbore Pumps	244156	BPD	4,759	0	0	0	0	0	0	0	0	0						
Offsites																		
Atm Flash (DRU)	0	BPD Bitumen	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Glycol System	5032	gpm	544															
Misc	-	-	3,096	0	0			0	0	0	0	0						
Steam Balance																		
Dearators	0	KLB BFW	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Demin Plant		KLB/h	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
VHP HP Letdowns		KLB/h	0	-72	72													
HP MP Letdowns		KLB/h	0	0	-72	72	0	0	0	0	0	0	0	0	0	0	0	0
MP LP Letdowns		KLB/h	0	0	0	-72	72	0	0	0	0	0	0	0	0	0	0	0
Indirect CO2 (Power Import)																	804	
TOTALS			**54,691**	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**71**	**0**	**199,227**	**0**	**8,375**	**0**

Notes

AERI / AENV Alternative Fuel Study BFD

Case 6 - SAGD 66,000 bpsd
Fuel: Petcoke - VCE,CFB



Water Balance

Stream	Flow (m³/sd)	TDS (w/o SiO2) (ppm)	SiO2 (ppm)	Hardness (ppm)
Steam to Reservoir	31,471	0	0	0
Losses to Reservoir	(3,147)	0	0	0
Produced Water	28,324	2,250	210	41
Losses to Production	(79)			
BD Recycle	1,363	167	17	17
M.U. Water	5,593	1,072	30	141
Evaporator Feed	35,201	2,019	174	56
Degass Loss	(176)	0	0	0
Concentrate Evap	(1408)	50,362	4335	1380
Condensate Reture	11814			
BFW to SG	45,437	5	1	17
Steam Produced	44,074	0	0	0
Blowdown	1,363	167	17	17

Petcoke Analysis

lb/d	6,842,856
Sulphur, wt%	5.69%
Nitrogen, wt%	1.22%
Nickel, wppm	167
Vanadium*, wppm	590.8
CCR, wt%	39

Emissions Summary

	SO2	S	CO2	NOX	
from CFB	416 0	208 0	10,269 0	1.3 0.0	mt/d
Treated Flue Gas:					
From CFB	6	3	10,269	1.3	mt/d
FGD Reaction Products					
FGD Waste (CFB)			532		mt/d
Recovered CO2			-8,641	1.3	mt/d
Direct CO2 Emissions			2,160		mt/d
Imported Power			1,657		mt/d
Total CO2 Emissions			3,817		mt/d
Recovered CO2			8,641		mt/d

LEGEND
- **Process Gas**
- **Bitumen / Emulsion / Process Water**
- **Diluent**
- **Condensate / BFW**
- **Steam**
- **Natural Gas Imports**
- **CO2 for Export**

Rev 9
By: Ian Buchanan / M. Armstrong
Date: 12/22/2008
Project #: JC2910

APPENDIX 2: BLOCK FLOW DIAGRAMS

Utility Summary for AERI Alternative Fuel Study

Case 6

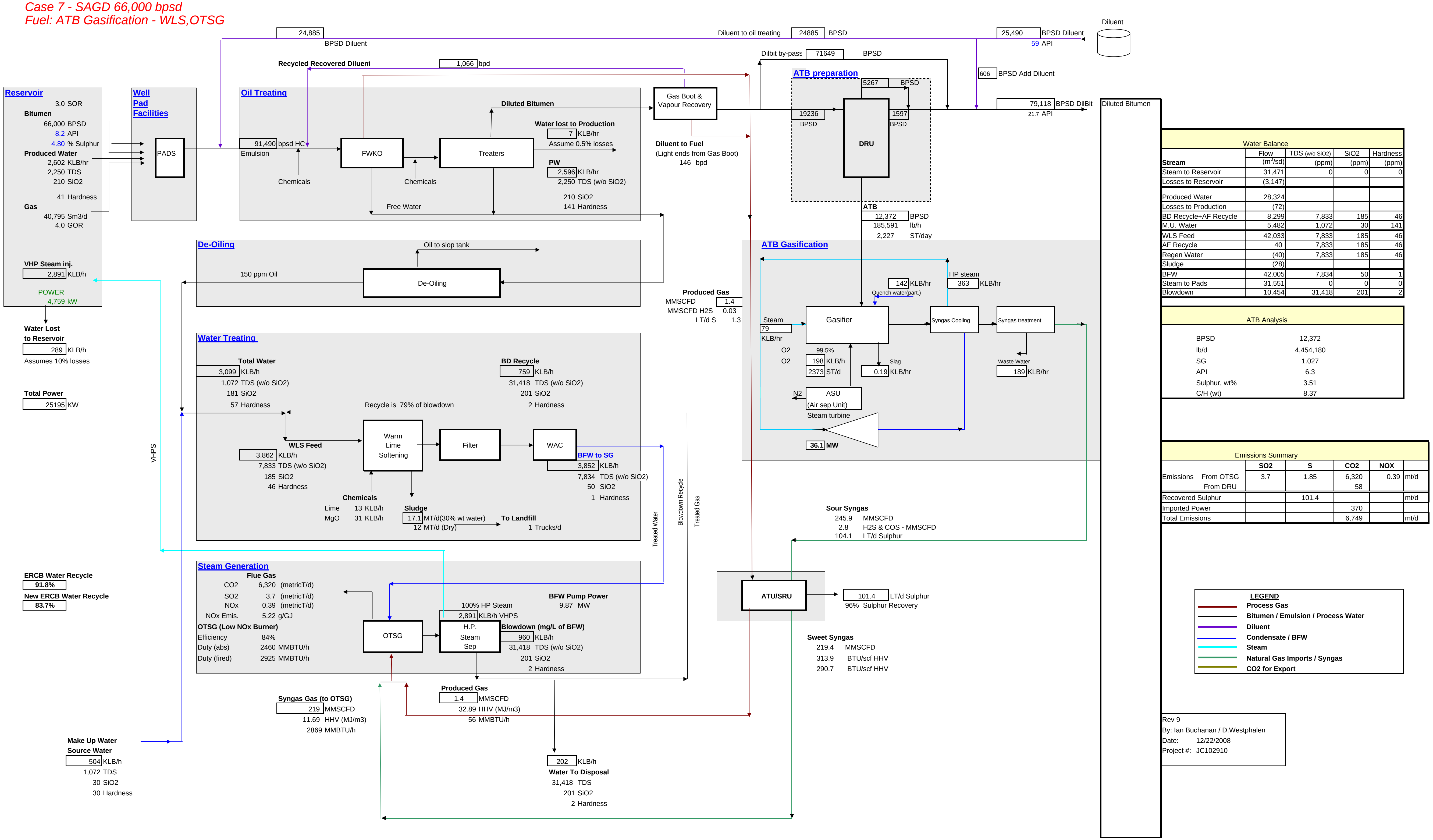
Utility Summary Project: AERI Alternative Fuel Study Case 5 - Petcoke Alternative Fuel Revision: 0	Capacity		Elec	Steam									Fuel Consumed		PETCOKE Burned		CO2 Generated	Recovered CO2
Process Unit	**Capacity**	**Units**	**Power KW**	**Steam** 1600 psig k#/hr	**Steam** 600 psig k#/hr	**(positive = generation)** 250 psig k#/hr	50 psig k#/hr	BFW k#/hr	Raw Water k#/hr	Cond. k#/hr	Waste W k#/hr	Loss k#/hr	**Produced Gas** MMBTU/hr	**Natural Gas** MMBTU/h	lb/h	lb/h	**CO2** MT/day	**CO2** MT/d
Oil Treating/Deoiling																		
Pumps			404	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Steam Generation																		
CFB Boiler	45,437	m3/day CWE Steam	14,624	4,053	0	0	0	0	0	0	0	0	71	0	285,119	0	10,269	8,215
FGD			5,650														532	425
Water Treating																		
Evaporators	35,208	m3/day CWE BFW	29,743	0	0	0	-72	0	0	0	0	0	0	0		0	0	0
Well Pads																		
Steam Injection to Wells	31,471	m3/day CWE BFW		-2,891	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Wellbore Pumps	244,156	BPD	4,759	0	0	0	0	0	0	0	0	0						
Offsites																		
Atm Flash (DRU)	-	BPD Bitumen	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Glycol System	5,512	gpm	596															
Misc	-	-	3,347	0	0			0	0	0	0	0						
CO2 Capture																		
Amine Unit	10,801	MTD CO2	4,320			-1,089												
Compressor	8,641	MTD CO2	49,251															
Steam Balance																		
Dearators	-	KLB BFW	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Demin Plant		KLB/h	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Condensate Flash		KLB/h	0	-405		405												
VHP HP Letdowns		KLB/h	0	-756	755							1						
HP MP Letdowns		KLB/h	0	0	-755	755	0	0	0	0	0	0	0	0	0	0	0	0
MP LP Letdowns		KLB/h	0	0	0	-72	72	0	0	0	0	0	0	0	0	0	0	0
Indirect CO2 (Power Import)																	1,657	
TOTALS			**112,694**	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**0**	**1**	**71**	**0**	**285,119**	**0**	**12,457**	**8,641**

Notes

% Recovery 80%

AERI/AENV Alternative Fuel Study BFD

Case 7 - SAGD 66,000 bpsd
Fuel: ATB Gasification - WLS,OTSG



Water Balance

Stream	Flow (m^3/sd)	TDS (w/o SiO2) (ppm)	SiO2 (ppm)	Hardness (ppm)
Steam to Reservoir	31,471	0	0	0
Losses to Reservoir	(3,147)			
Produced Water	28,324			
Losses to Production	(72)			
BD Recycle+AF Recycle	8,299	7,833	185	46
M.U. Water	5,482	1,072	30	141
WLS Feed	42,033	7,833	185	46
AF Recycle	40	7,833	185	46
Regen Water	(40)	7,833	185	46
Sludge	(28)			
BFW	42,005	7,834	50	1
Steam to Pads	31,551	0	0	0
Blowdown	10,454	31,418	201	2

ATB Analysis

BPSD	12,372
lb/d	4,454,180
SG	1.027
API	6.3
Sulphur, wt%	3.51
C/H (wt)	8.37

Emissions Summary

	SO2	S	CO2	NOX	
Emissions From OTSG	3.7	1.85	6,320	0.39	mt/d
From DRU			58		
Recovered Sulphur		101.4			mt/d
Imported Power			370		
Total Emissions			6,749		mt/d

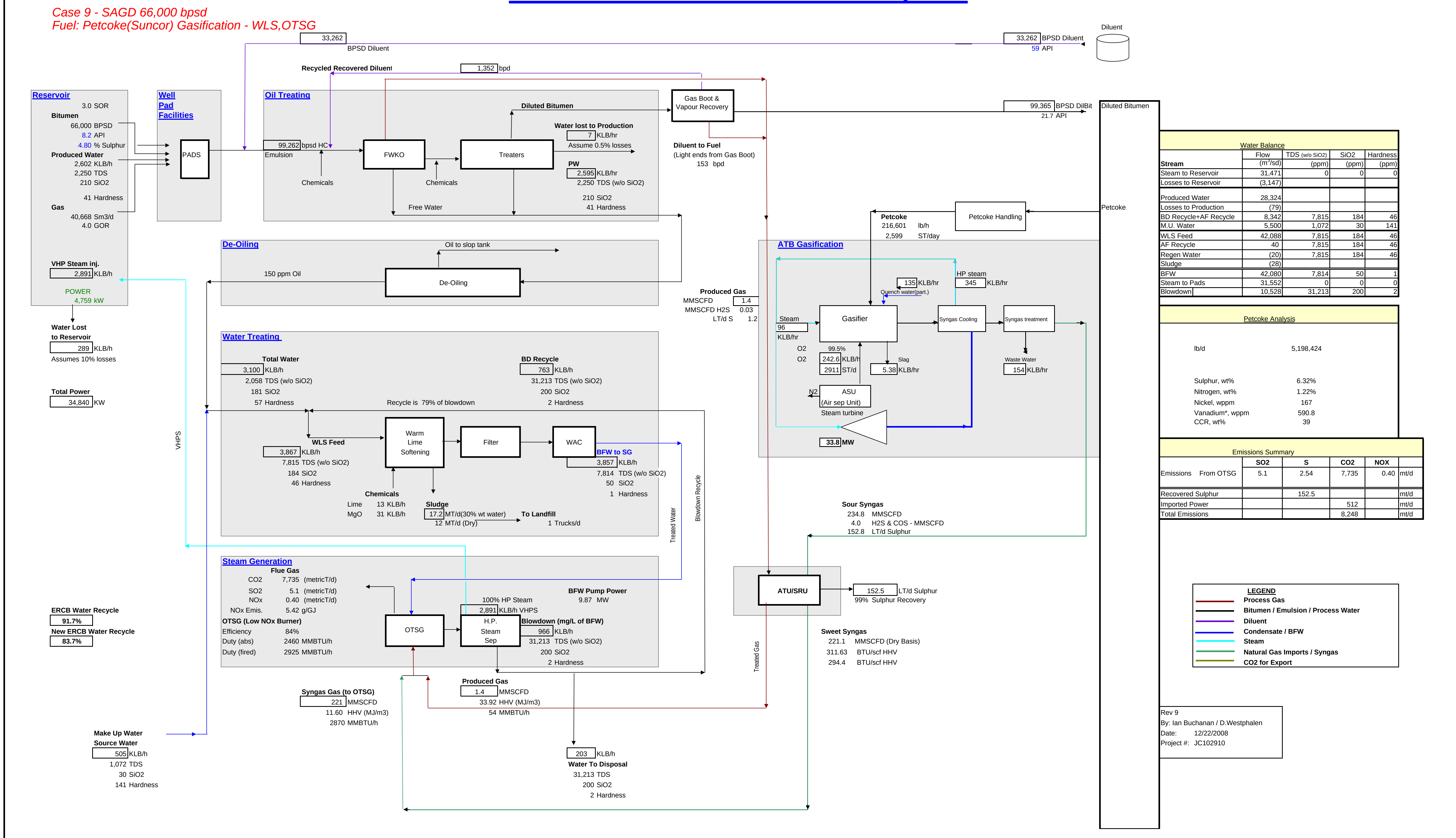
Utility Summary for AERI Alternative Fuel Study Phase 2

Case 7

Utility Summary Project: AERI Alternative Fuel Study Phase 2 Case 7 - ATB Gasification Alternative Fuel Revision: 0	Capacity		Elec											Fuel Consumed		CO2 Generated	Recovered CO2
				Steam	Steam	(positive = generation)							Produced	Gasifier	Natural		
Process Unit	**Capacity**	**Units**	**Power KW**	1600 psig k#/hr	600 psig k#/hr	150 psig k#/hr	50 psig k#/hr	BFW k#/hr	Raw Water k#/hr	Cond. k#/hr	Waste W k#/hr	Loss k#/hr	**Gas** MMBTU/hr	**Syn Gas** MMBTU/hr	**Gas** MMBTU/h	**CO2** MT/day	**CO2** MT/d
Oil Treating/Deoiling																	
Pumps			286	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Steam Generation																	
OTSG	31471	m3/day CWE Steam	10,132	3,104	0	0	0	0	0	0	0	0	56	2,869	0	6,320	
Water Treating																	
WLS	42033	m3/day CWE BFW	398	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Well Pads																	
Steam Injection to Wells		m3/day CWE BFW		-2,891	0	0	0	0	0	0	0	0	0	0	0	0	0
Downhole Pumps	244139	BPD	4,759	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Offsites																	
Sulphur Plant	0	MT/d	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Amine Regen	0.00	MM SCFD AG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Atm Flash (DRU)	19236	BPD	401	0	0	-180	0	0	0	100	80	0	0	0	50	58	0
Glycol System	7257	gpm	784														
Misc	-	-	1,426	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Gasifier Plant																	
Gasifier (Rx, Steam Gen, Shift)	2227	ST/d of Pitch	42,760	363	0	0	0	-367	-142	0	189	32	0	0	0	0	0
ASU		ST/D of O2	in above	-79	0	0	0	0	0	0	0	0	0	0	0	0	0
PSA		MM SCFD H2	in above	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Acid Gas Removal		MMSCFD Syngas	in above	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Power Plant (GT/HRSG)		MW	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HP Steam Turbine Generator	284	KLB/h	-36,096	-284	0	0	0	0	0	284	0	0	0	0	0	0	0
Dearators	367	KLB BFW	345	0	0	0	-33	367	0	-334	0	0	0	0	0	0	0
Demin Plant	269	KLB/h	0	0	0	0	0	0	0	0		0	0	0	0	0	0
VHP HP Letdowns		KLB/h		-213	213												
HP MP Letdowns		KLB/h	0	0	-213	213	0	0	0	0	0	0	0	0	0	0	0
MP LP Letdowns		KLB/h	0	0	0	-33	33	0	0	0	0	0	0	0	0	0	0
Indirect CO2 (Power Import)																370	
TOTALS			**25,195**	**0**	**0**	**0**	**0**	**0**	**142**	**51**	**269**	**32**	**56**	**2,869**	**50**	**6,749**	**0**

AERI/AENV Alternative Fuel Study BFD

Case 9 - SAGD 66,000 bpsd
Fuel: Petcoke(Suncor) Gasification - WLS,OTSG



Water Balance

Stream	Flow (m^3/sd)	TDS (w/o SiO2) (ppm)	SiO2 (ppm)	Hardness (ppm)
Steam to Reservoir	31,471	0	0	0
Losses to Reservoir	(3,147)			
Produced Water	28,324			
Losses to Production	(79)			
BD Recycle+AF Recycle	8,342	7,815	184	46
M.U. Water	5,500	1,072	30	141
WLS Feed	42,088	7,815	184	46
AF Recycle	40	7,815	184	46
Regen Water	(20)	7,815	184	46
Sludge	(28)			
BFW	42,080	7,814	50	1
Steam to Pads	31,552	0	0	0
Blowdown	10,528	31,213	200	2

Petcoke Analysis

lb/d	5,198,424
Sulphur, wt%	6.32%
Nitrogen, wt%	1.22%
Nickel, wppm	167
Vanadium*, wppm	590.8
CCR, wt%	39

Emissions Summary

	SO2	S	CO2	NOX	
Emissions From OTSG	5.1	2.54	7,735	0.40	mt/d
Recovered Sulphur		152.5			mt/d
Imported Power			512		mt/d
Total Emissions			8,248		mt/d

Utility Summary for AERI Alternative Fuel Study

Case 9

Utility Summary Project: AERI Alternative Fuel Study Case 9 - Petcoke Gasification Alternative Fuel Revision: 0	Capacity		Elec											Fuel Consumed		CO2 Generated	Recovered CO2
				Steam	Steam	(positive = generation)							Produced	Gasifier	Natural		
Process Unit	**Capacity**	**Units**	**Power KW**	1600 psig k#/hr	600 psig k#/hr	150 psig k#/hr	50 psig k#/hr	BFW k#/hr	Raw Water k#/hr	Cond. k#/hr	Waste W k#/hr	Loss k#/hr	Gas MMBTU/hr	Syn Gas MMBTU/hr	Gas MMBTU/h	CO2 MT/day	CO2 MT/d
Oil Treating/Deoiling																	
Pumps			332	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Steam Generation																	
OTSG	31471	m3/day CWE Steam	10131	3103	0	0	0	0	0	0	0	0	54	2870	0	7735	0
Water Treating																	
WLS	42088	m3/day CWE BFW	398	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Well Pads																	
Steam Injection to Wells		m3/day CWE BFW		-2891	0	0	0	0	0	0	0	0	0	0	0	0	0
Downhole Pumps	244139	BPD	4759	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Offsites																	
Sulphur Plant		MT/d	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Amine Regen		MM SCFD AG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Atm Flash (DRU)		BPD	0	0	0	-180	0	0	0	100	80	0	0	0	0	0	0
Glycol System	7637	gpm	825														
Misc	-	-	1972	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Gasifier Plant																	
Gasifier (Rx, Steam Gen, Shift)	2599	ST/d of Petcoke	49905	345	0	0	0	-348	-135	0	154	78	0	0	0	0	0
ASU		ST/D of O2	in above	-96	0	0	0	0	0	0	0	0	0	0	0	0	0
PSA		MM SCFD H2	in above	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Acid Gas Removal		MMSCFD Syngas	in above	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Power Plant (GT/HRSG)		MW	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HP Steam Turbine Generator	249	KLB/h	-33810	-249	0	0	0	0	0	249	0	0	0	0	0	0	0
Dearators	348	KLB BFW	327	0	0	0	-31	348	0	-317	0	0	0	0	0	0	0
Demin Plant	234	KLB/h	0	0	0	0	0	0	0	0		0	0	0	0	0	0
VHP HP Letdowns		KLB/h		-211	211												
HP MP Letdowns		KLB/h	0	0	-211	211	0	0	0	0	0	0	0	0	0	0	0
MP LP Letdowns		KLB/h	0	0	0	-31	31	0	0	0	0	0	0	0	0	0	0
Indirect CO2 (Power Import)																512	
TOTALS			**34,840**	**0**	**0**	**0**	**0**	**0**	**135**	**32**	**234**	**78**	**54**	**2870**	**0**	**8248**	**0**

Notes

Appendix 3.

Decision Matrix Analysis

APPENDIX 3 : DECISION MATRIX ANALYSIS

	Case 1 Natural Gas OTSG	Case 3 Bitumen Modular Boiler	Case 5 Petcoke Circulating Fluidized Bed	Case 7 Liquid Gasification	Case 9 Solid Gasification	Weighting	Metric
Independent of Natural Gas Usa	**1**	**7**	**10**	**10**	**10**	**7**	
Carbon Capture Ready							
CAPEX	7	4	1	10	7	4	$ / Bbl Investment for CO2 Capture
Energy Requirement	7	4	1	10	7	7	$ / Bbl Energy OPEX for CO2 Capture
Ease of Retrofit	7	4	1	7	7	7	$ / Bbl Investment for CO2 Capture
Operating Cost						4	$ / Bbl OPEX for CO2 Capture
Risk of Proliferation	1	4	7	7	7	7	Difficult to convert technology to smaller scale
Environmental Impacts							
Footprint	4	7	7	1	1	10	Sq. Ft Plot Space (No. of Vessels)
Waste Water	4	7	7	4	4	10	usgpm
Carbon Emissions	7	4	1	1	1	10	MTPD
Solid Wastes	10	4	1	10	7	7	kg/day
SOX	7	4	1	1	1	10	g / GJ or lbs/KW
NOX	7	4	4	7	7	10	g / GJ or lbs/KW
Particulates	7	4	4	7	7	10	g / GJ or lbs/KW
Road Traffic	10	7	1	10	1	7	Trucks per day
Reliability							
Equipment #	4	7	7	1	1	4	pieces of equipment
Rotating Equipment	4	4	7	1	1	4	# of pumps and compressor
Technical Risk	10	7	7	7	7	10	10 - Demonstrated in Alberta, 7- Demostrated at scale, 4 - Demonstrated at smaller scale
Capital Cost	7	7	4	1	1	7	$ / GJ
Operating Expense							
Maintenance Costs	7	7	4	1	1	7	$ / GJ
Power	7	1	1	4	4	7	$ / GJ
Chemical Costs	10	4	1	10	10	7	$ / GJ
Fuel Costs	1	4	10	4	10	7	$ / GJ
Other Costs	7	7	4	1	1	7	$ / GJ
Net Efficiency	7	7	7	4	4	10	Steam Gen / Fuel Consumed
Safety Issues							
Operability	4	7	7	1	1	10	No. of Control Loops (est.), (Equip Count)
Pressure Equipment	4	7	7	1	1	10	No. of Vessels, (Equip Count)
Road Traffic	10	7	1	10	1	7	No. of Trucks
Toxic Chemicals	7	7	7	4	4	10	Primarily H2S
Operational Complexity	10	7	4	4	1	7	Overall Difficulty of Process Operation
Infrastructure Requirements							
Fuel and Chemical Supply	4	7	1	7	1	7	7 - <5 Trucks per day, 4 - Pipeline fuel, 1 > 50 TPD
Product Sales, (Excluding Bitumen and Dilbit)	7	7	7	4	4	4	10 - None, 7 - Occasional, 4 - Frequent
Diluent	4	7	4	7	4	7	BPD of Diluent
Electrical Demand	7	1	1	4	4	7	KW
OSBL Costs	7	7	4	1	1	4	
Overall Totals	**1566**	**1389**	**1086**	**1197**	**975**		
Grouped Totals	**470**	**482**	**430**	**398**	**332**		

Case 2, 4, 6, 8, and 10 are the above cases with Carbon Capture added

APPENDIX 3 : DECISION MATRIX ANALYSIS

		Case 1 Natural Gas OTSG	Case 3 Bitumen Modular Boiler	Case 5 Petcoke Circulating Fluidized Bed	Case 7 Liquid Gasification	Case 9 Solid Gasification	Weighting
Independent of Natural Gas Usage		**1**	**7**	**10**	**10**	**10**	**7**
		7	**49**	**70**	**70**	**70**	
Carbon Capture Ready							
	CAPEX	7	4	1	10	7	4
	Energy Requirement	7	4	1	10	7	7
	Ease of Retrofit	7	4	1	7	7	7
	Operating Cost						4
		28.875	**16.5**	**4.125**	**37.125**	**28.875**	5.5
Risk of Proliferation		1	4	7	7	7	7
		7	**28**	**49**	**49**	**49**	
Environmental Impacts							
	Footprint	4	7	7	1	1	10
	Waste Water	4	7	7	4	4	10
	Carbon Emissions	7	4	1	1	1	10
	Solid Wastes	10	4	1	10	7	7
	SOX	7	4	1	1	1	10
	NOX	7	4	4	7	7	10
	Particulates	7	4	4	7	7	10
	Road Traffic	10	7	1	10	1	7
		64.75	**47.40625**	**30.0625**	**47.40625**	**33.53125**	9.25
Reliability							
	Equipment #	4	7	7	1	1	4
	Rotating Equipment	4	4	7	1	1	4
	Technical Risk	10	7	7	7	7	10
		36	**36**	**42**	**18**	**18**	6
Capital Cost		7	7	4	1	1	7
		49	**49**	**28**	**7**	**7**	
Operating Expense							
		44.8	**32.2**	**28**	**28**	**36.4**	7
Net Efficiency		7	7	7	4	4	10
		70	**70**	**70**	**40**	**40**	
Safety Issues							
	Operability	4	7	7	1	1	10
	Pressure Equipment	4	7	7	1	1	10
	Road Traffic	10	7	1	10	1	7
	Toxic Chemicals	7	7	7	4	4	10
		57.8125	**64.75**	**50.875**	**37**	**16.1875**	9.25
Operational Complexity		10	7	4	4	1	7
		70	**49**	**28**	**28**	**7**	
Infrastructure Requirements							
	Fuel and Chemical Supply	4	7	1	7	1	7
	Product Sales, (Excluding Bitumen and Dilbit)	7	7	7	4	4	4
	Diluent	4	7	4	7	4	7
	Electrical Demand	7	1	1	4	4	7
	OSBL Costs	7	7	4	1	1	4
		33.64	**33.64**	**19.72**	**26.68**	**16.24**	5.8
Grouped Totals		**470**	**482**	**430**	**398**	**332**	

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