EVALUATION OF THE ECONOMIC FEASIBILITY OF HEAVY OIL PRODUCTION PROCESSES FOR WEST SAK FIELD

by

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STATEMENT OF THESIS APPROVAL

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ABSTRACT

The West Sak heavy oil reservoir on the North Slope of Alaska represents a large potential domestic oil source which has not been fully developed due to difficulties with producing viscous oil from a cold reservoir. Past studies have evaluated the economic viability of producing from West Sak, but given the rising demand for oil, a fresh evaluation of the economic feasibility of heavy oil production processes from West Sak is warranted. Therefore, the objective of this project was to design a set of possible processes for recovery of heavy oil from West Sak and identify any economic barriers to production.

Discounted cash flows were used to determine the investor's rate of return (IRR) for each process assuming oil sold for either a fixed price or followed a given price forecast. Capital and operating costs were estimated primarily using the methodology suggested by Seider *et al*. (2008). Three different scenarios were analyzed using this methodology: a base case and two alternatives for oil transport (dilution with gas-to-liquids and upgrading via hydrotreating). Polymer flooding was selected as the recovery method for all scenarios and production rates were estimated from recovery curves published by Seright (2011). Each scenario also investigates the possibility of using oxy-firing for CO_2 capture as an alternative method for providing process heating.

Results of the economic analysis show that the base case would produce an IRR of 41% (dilution would produce a 45% IRR, and upgrading a 6% IRR). A sensitivity analysis performed on the model's inputs gave a range of possible IRRs for the base case of 30% to 50%, dilution's range was 24% to 62%, and upgrading ranged from -2% to 29%. Both the base case and dilution scenarios have no economic barriers to development. If West Sak heavy oil as produced can be delivered via pipeline, then the base case would be the economically preferable scenario. Upgrading is not economically feasible due to high capital costs which drive up the required oil price and result in large severance tax liabilities.

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NOMENCLATURE

Symbol or Abbreviation	Units	Description
A	ft^2	Cross-sectional area
ACES		Alaska's Clear and Equitable Share law
AHO		Alaskan Heavy Oil crude
ANS		Alaskan North Slope crude
bbl		Barrel
B_o		Oil formation volume factor
b		Scaling power
bpd	bbl/day	Barrels per day
C	$	Cost or bare-module cost
C_{DPI}	$	Total direct permanent investment
C_{drill}	$	Capital cost for drilling
CEPCI		Chemical Engineering Plant Cost Index
C_f	$/yr	Total fixed operating costs (i.e. costs at that are not a function of production capacity such as labor, administration, and insurance)
CF_n	$/yr	Annual cash flow in year n
C_L	$	Capital cost for purchasing and/or leasing land
C_o	$	Base cost
COM	$/yr	Cost of manufacture
C_p	$	Direct purchase cost

C_P	$	Capital cost for permitting
CPFB	$/bpd	Capital per flowing barrel
C_{pipe}	$	Capital cost for pipeline
C_R	$	Capital cost for intellectual property royalties
C_S	$	Capital cost for startup
C_{TBM}	$	Total bare module investment
C_{TCI}	$	Total capital investment
C_{TDC}	$	Total depreciable capital investment
C_{TPI}	$	Total permanent investment
C_v	$/yr	Total variable operating costs (i.e. operating costs that are a function of production capacity such as water, electricity, fuel and other utilities) at full production capacity
C_{WC}	$	Working capital
d	$	Depletion
D	$	Depreciation
DEA		Diethanol amine
D_{econ}	inch	Economic pipeline diameter
DOR		Alaska Department of Revenue
E		Efficiency of the pipe's motor and pump
EIA		Energy Information Administration
ENR		Engineering News and Record
EOR		Enhanced oil recovery
F		Ratio of total cost for fittings and installation to purchase cost for new pipe
F_{BM}		Bare-module factor
F_d		Design factor

f_d		Direction drilling factor
f_E		Fraction of operating costs associated with extracting oil
F_m		Material factor
f_n		Discount factor
F_p		Pressure factor
f_t		Well type factor
GTL		Gas to liquids
H	ft	Reservoir thickness
HC		Hydrocarbon
HPAM		Polyacrylamide
H_y	hr/yr	Hours of operation per year
i	1/yr	Annual interest rate
I		Current index value
I_o		Base index value
IRR		Investor’s rate of return
J		Fractional loss due to fittings and bends
JAS		API Joint Association Survey
K	$/kWh	Cost of electricity
K_F		Annual fixed charges for financing and maintenance expressed as a fraction of total pipe cost
L	ft	Well spacing
LS	$/yr	Labor salary and benefits
LW	$/yr	Labor wages and benefits
m		OOIP recovery curve slope
MACRS		Modified accelerated cost recovery system

MCF		Thousand standard cubic feet
MCFD		Thousand standard cubic feet per day
md		Millidarcy
MS		Maintenance salary and benefits
MW		Maintenance wages and benefits
n	yr	Project year
NGL		Natural gas liquids
n_{inj}		Number of injection wells
N_p	bpd	Total oil production rate
NPV	$	Net present value
OOIP	bbl	Original oil in place
P	psi	Pressure
P_{inj}	psig	Injection pressure
P_L	psig	Pressure at producer well
P_n		Production capacity fraction (days operated per days in one year) for year n
P_o	psig	Pressure at injection well
PSA		Pressure-swing adsorption
PV		Pore volume
Q		Capacity
Q	bpd	fluid injection rate
q_f	ft^3/s	Fluid flow rate
Q_o		Base capacity
R	$/yr	Royalties, including both oil and intellectual property royalties
rb		Reservoir barrel

Re		Reynolds number
RFG		Recycled flue gas
R_{IP}	\$/yr	Royalties for intellectual property
ROI		Return on investment
R_{oil}	\$/yr	Royalties for oil
S	\$/yr	Total sales at full production capacity
SO_X		Sulfur oxides
ST	\$/yr	Severance taxes
stb		Stock tank barrel
S_{wc}		Connate water saturation
T	\$/yr	Taxes, including state, federal, severance, and property taxes
TAPS		Trans-Alaska Pipeline System
t_F		Federal corporate income tax rate
T_F	\$/yr	Federal corporate income tax
TI	\$/yr	Taxable income
t_S		State corporate income tax rate
T_S	\$/yr	State corporate income tax
U		Utility requirement
u	ft/s	Fluid velocity
U_o		Base utility requirement
USGS		United States Geological Survey
VAPEX		Vapor extraction processes
VGO		Vacuum gas oil
VRWAG		Viscosity reducing water alternating gas
VSA		Vacuum-swing adsorption

W	ft	Length of lateral well segment
WAG		Water alternating gas
WHP	\$/bbl	Wellhead profit
WTI		West Texas Intermediate crude
X	\$/ft	Purchase cost of new 1” diameter pipe per foot of pipe length
κ	md	Permeability
μ	cP	Viscosity
μ_c	cP	Fluid viscosity
μ_o	cP	Viscosity of oil
μ_w	cP	Viscosity of water
ρ	lb/ft^3	Fluid density
ϕ		Porosity

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INTRODUCTION

Recent surges in the price of oil have renewed interest in developing U.S. domestic unconventional oil resources. One such resource is heavy oil, which is defined by the U.S. Department of Energy as having an API gravity between 10.0° – 22.3° (Nehring, Hess and Kamionski 1983). The size of the heavy oil resource in the U.S. has been estimated to be on the order of nine billion barrels (bbl), one-third of which are located in the West Sak field on the North Slope of Alaska (Hinkle and Batzle 2006). However, despite the size of the resource, West Sak still remains largely undeveloped. A number of publicly available studies from the early 1990s have analyzed the economic feasibility of increasing oil production from West Sak, but given the increased demand for oil, a fresh evaluation of the subject with a focus on current economic conditions is warranted.

Therefore, the purpose of this project was to evaluate the economic feasibility of heavy oil production processes from the West Sak field. Specifically, the objectives were to:

- Define and design a representative set of possible processes for recovery of heavy oil from West Sak.
- Evaluate the economics of each process using standard engineering cost estimation methodologies.
- Identify the major economic barriers to the production of heavy oil from West Sak.
- Investigate the sensitivity of each process' profitability to pricing and other important economic modeling assumptions.

A detailed description of the resource, its production history, and a review of previous studies is given below. Section 2 describes the economic and cost estimating methodologies used to evaluate the feasibility of the production process scenarios described in Section 3. The results of that analysis are given in Section 4, followed by a discussion of the results in Section 5 with conclusions and recommendations for future work in Section 6.

1.1 Geology of West Sak

West Sak is considered a satellite of the Kuparuk River Field and is located above Kuparuk at depths of 2,500 ft – 4,600 ft in six major layers ranging in thickness from 10 ft – 50 ft (Gondouin and Fox 1991). A map showing the location of West Sak relative to other oil fields on the North Slope is shown in Figure 1-1 and a generalized cross section of the resource is shown in Figure 1-2. Reservoir properties are given in Table 1-1.

Various claims about the size of the reservoir have been published, ranging from 3 billion bbl of original oil in place (OOIP) (Hinkle and Batzle 2006) to 25 billion bbl OOIP (Panda, *et al*. 1989). The lithology of the reservoir has been reported as fine-grained quartzitic shaly sandstone (very friable) with some swelling clays and glauconite (Panda, *et al*. 1989). The reservoir was deposited during the Upper Cretaceous period approximately 65 million years ago.

1.2 West Sak Production History

The first pilot development in West Sak began in 1983 using conventional verticals wells. The development included fracturing and waterflooding to improve production rates and recovery but was ultimately abandoned in 1986 as uneconomic (Hartz, *et al*. 2004). Development was restarted in 1997, again using waterflooding. By 2004, production from the formation had reached approximately 10,000 barrels per day (bpd) (BP America 2004). The large increase in production was due primarily to advances in horizontal and multilateral drilling which brought well production rates from 200-300 bpd to 1,000-2,000 bpd (Hartz, *et al*. 2004). In 2004, several major oil companies (ConocoPhillips, BP, Unocal, ExxonMobil, and Chevron Texaco) planned a 30,000 bpd expansion to be completed in 2007 (Nelson 2007). However, according to the reported production data (AOGCC 2004-2011), production from West Sak has yet to reach planned levels. Production rates and cumulative production of crude oil, water, and gas are shown in Figure 1-3 and Figure 1-4.

1.3 Review of Previous Studies

Since the closure of the initial pilot development in 1986, several studies have been published that review or propose the feasibility of producing oil from West Sak. The primary difficulty identified in all of the studies is that West Sak heavy oil has very low mobility at reservoir conditions because of its high viscosity, resulting in low production rates. In other heavy oil plays, viscosity is typically reduced by injecting steam into the reservoir (Nehring, Hess and Kamionski 1983), which increases the temperature of oil in the reservoir and reduces viscosity, as shown in Figure 1-5.

However, injecting steam into West Sak is difficult because of the nearly 2,000 ft of permafrost overburden (Gondouin and Fox 1991). Heat transfer from any potential steam injection well to the permafrost would both reduce the quality of any steam injected and melt the surrounding permafrost, reducing the structural support of the well. Hallam *et al*. (1992) investigated the issue using computer simulation, and calculated the resulting strain experienced by the well from the rapid decrease in pore pressure that occurs during permafrost melt, finding that safety limits were only exceeded for uninsulated tubing. Regardless, steam injection into West Sak has not been used by any producer operating in the North Slope.

A variety of papers have been published suggesting different methods for extracting heavy oil from West Sak. Sharma, Kamath, Godbole, & Patil (1990) published a large report that analyzed the simultaneous injection of steam and other gases, including N_2, CH_4, and CO_2, as well as the economic feasibility of a steam flooding process. Recovery rates for simultaneous injection ranged from 77% (steam only) –to 92% (steam and CO_2) of OOIP. The results of their economic analysis determined that an oil market price between $18-25/bbl (1990 dollars) was necessary for steamflooding of West Sak to generate a 20% rate of return assuming a 2,000 bpd steam injection rate (43.82% - 56.20% OOIP recovery over 10 years). The effect of steam flooding on the permafrost layer was not considered.

Gondouin & Fox (1991) proposed using a downhole catalytic methanator. Syngas produced at the surface would be pumped downhole to the methanator to produce steam below the permafrost layer; extraction would then proceed following traditional cyclic steam injection methods. The authors claimed that steam was preferable to miscible displacement processes (CO_2 injection) because of the potential for asphaltene precipitation and reduced reservoir permeability. Gondouin & Fox also analyzed the economic feasibility of their proposed extraction method, finding (in 1991 dollars) that a

65,000 bpd operation required a $16,185 capital per flowing barrel (CPFB) investment and produced a 12% IRR with an oil price of $17/bbl over a project life of 30 years. The authors did not report what ultimate recovery of the OOIP they expected to achieve, but they did cite steam recoveries from the literature of around 70% - 80%.

Hornbrook, Dehghani, Qadeer, Ostermann, & Ogbe (1991) conducted a laboratory displacement study to evaluate the effectiveness of simultaneous CO_2 and steam injection. They found that a 1:3 mixture of CO_2 – to – steam recovered 90.0% of OOIP compared to 77.2% OOIP with steam flooding. Both recovery rates cited are after injecting six pore volumes (PV) of fluid. The authors did not evaluate the economic feasibility of their process.

Ogbe, Zhu, & Kovscek (2004) conducted experimental and numerical studies of the feasibility of using vapor extraction processes (VAPEX) to enhance oil recovery from West Sak. VAPEX is similar to steam injection, except that a solvent (ethane, propane, or butane) is used in place of steam as the injection fluid. The group found that VAPEX recovered 15% - 20% of OOIP in the equivalent of 15 years extraction time. The authors did not evaluate the economic feasibility of VAPEX for West Sak.

Mohanty (2004) investigated the use of water-alternating-gas (WAG) to find the optimal solvent, injection schedule, and well geometry for producing from heavy oil reservoirs on Alaska’s North Slope (such as West Sak). WAG alternates injections of water with a miscible solvent or gas such as CO_2 or natural gas liquids (NGL) to improve the recovery rates of water flooding. Using a variety of solvent mixtures, the author was able to achieve recoveries of 60% - 100% with injection of two PV of fluid, but optimal WAG process parameters (solvent, injection schedule, etc.) are not given. Instead, Mohanty (2004) states that parameters will depend on the specific economic analysis of a given scenario.

Revana & Erdogan (2007) present a review of many of the widely used heavy oil production methods and an economical optimized steam injection process for a single well in a generic reservoir. The authors recommend cold production (nonthermal artificial pumping techniques) for heavy oil with West Sak-like properties, citing the low capital investment involved and potential recoveries of up to 10% OOIP.

Seright (2011) recently published a report detailing the potential oil recovery from unconventional reservoirs by polymer flooding, including heavy oil on the North Slope. Polymer floods are similar to waterflooding except that a polymer additive is used to increase the viscosity of the mixture so that both fluids (water and oil) have the same viscosity; having similar viscosities reduces viscous fingering and channeling. The author used a fractional flow analysis to determine the OOIP recovery as a function of PV of fluid injected. The results of polymer flooding ($\mu_w = \mu_o$) for a homogeneous single layered reservoir representative of North Slope heavy oil reservoirs is given in Figure 1-6.

A comprehensive review of the economic feasibility of heavy oil production from Alaska was written by Olsen, Taylor, & Mahmood (1992). The authors determined that most of the heavy oil resources on the North Slope were uneconomical for a variety of reasons. Due to legislative constraints, Alaskan North Slope (ANS) and Alaskan Heavy Oil (AHO) crude must be sold in the United States, placing heavy oil from West Sak in direct competition with heavy oil from California, with the added burden of transporting AHO from the North Slope to refineries in California (transportation costs were estimated to be near $10/bbl in 1991 dollars). The authors expected a low recovery factor, 5% OOIP, given the high oil viscosity and absence of natural pressure-maintenance mechanisms such as gas-cap or water-drive. The authors also expressed concern with the ability of the Trans-Alaska Pipeline System (TAPS) to deliver viscous heavy oil. Finally, Olsen, Taylor, & Mahmood reviewed the economic results reported by Sharma *et al.* (1990) and were quite critical of both the simplifying assumptions in their reservoir model (ignoring reservoir heterogeneities and rock-fluid interactions) and the assumption of an “unrealistic” transportation cost of $4.08/bbl.

More recently, Targac *et al*. (2005) reviewed production from West Sak and industry plans for future development of the reservoir. The authors noted the same trends cited in Hartz, Decker, Houle, & Swenson (2004) as being primarily responsible for the viability of production from West Sak, namely increased well production rates from the use of horizontal and multilateral drilling techniques. Using pilot results based on the new drilling techniques, Targac *et al*. predicted an ultimate recovery of 15%-20% OOIP with waterflooding. Future development is expected to utilize a viscosity-reducing water-alternating-gas (VRWAG) process.

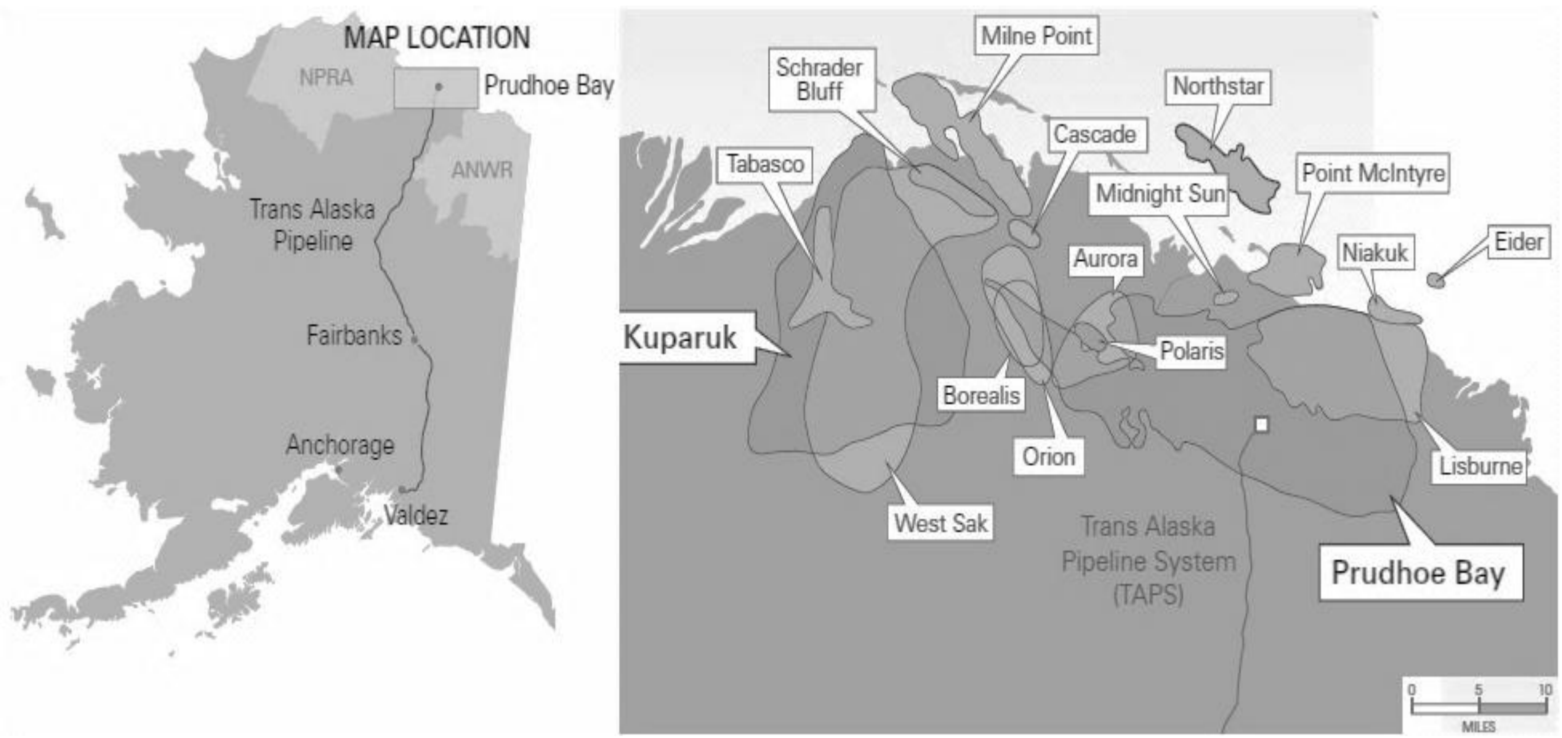

Figure 1-1: Location of West Sak (ConocoPhillips, BP 2006).

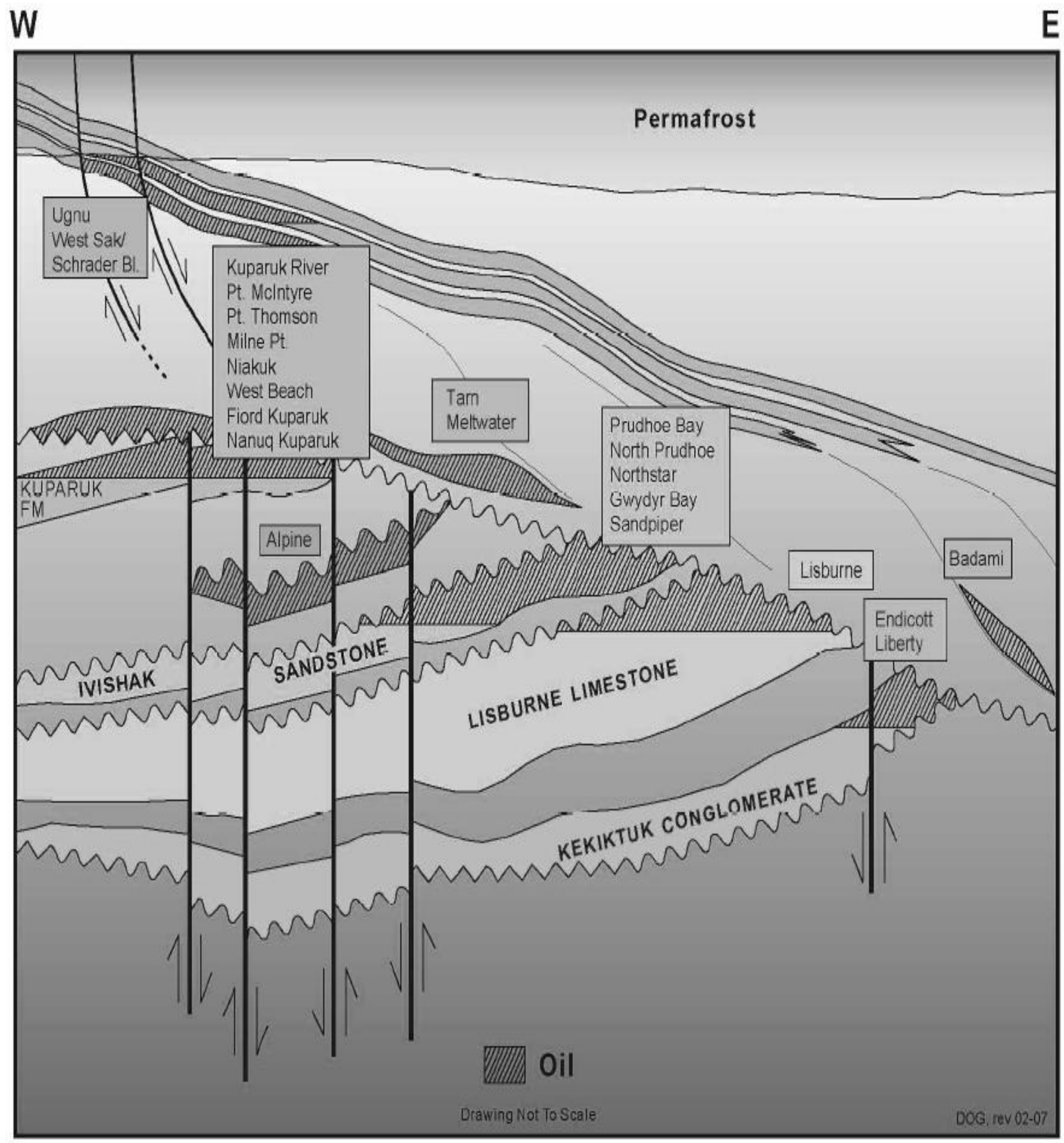

Figure 1-2: Generalized cross section of Central Artic Slope fields (Hartz, *et al*. 2004).

Table 1-1: West Sak reservoir and crude oil properties (Seright 2011, Hinkle and Batzle 2006, Gondouin and Fox 1991).

Property (units)	Value
Original Oil in Place (billion bbl)	3 – 25
Areal Extent (square miles)	300
Oil Gravity (° API)	10.5 – 23
Reservoir Depth (ft)	2,500 – 4,600
Reservoir Temperature (°F)	45 – 100
Number of Separate Layers	6
Layer Thickness (ft)	10 – 50
Porosity (vol. %)	20 – 30
Oil Saturation (% pore volume)	60% – 88%
Permeability (md)	150
Oil Viscosity (cP)	20 – 90
Solution GOR (scf/stb)	210
Bubble Point (psi)	1,690
Oil Formation Volume Factor (bbl/stb)	1.069
Gas Composition (% CH_4)	98
C_{21+} Fraction (mol %)	38.82
Molecular Weight (C_{21+} Fraction)	455
Sulfur (wt. %)	1.82
Asphaltene (wt. %)	2.8

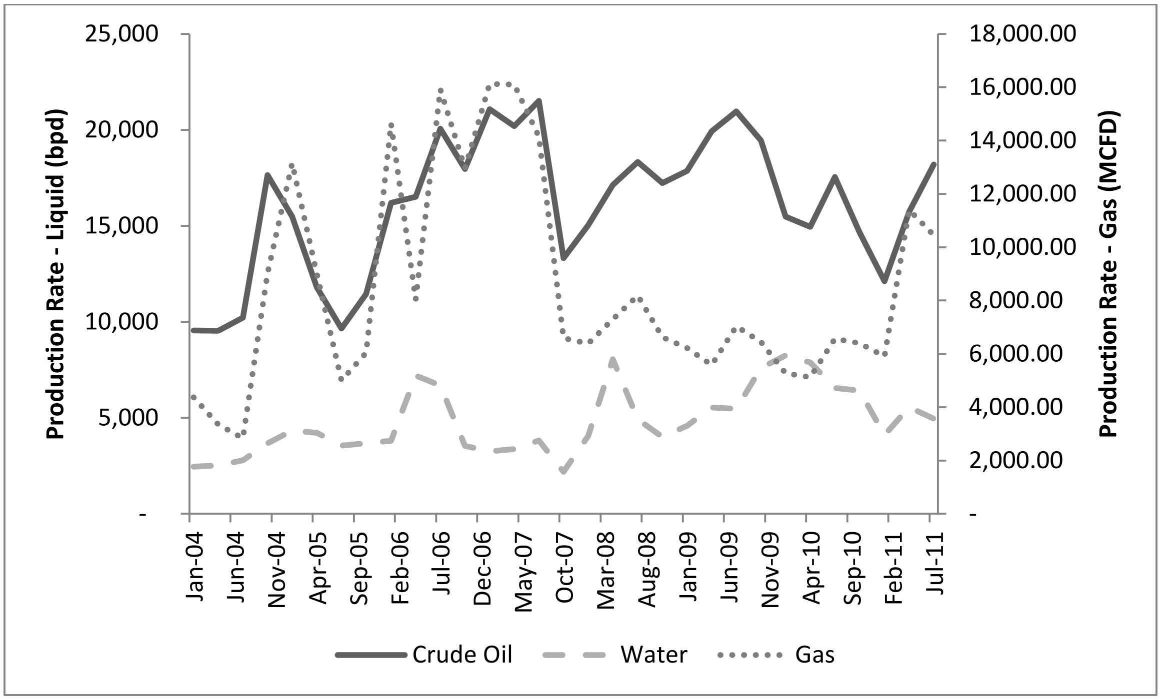

Figure 1-3: West Sak production rates from 2004 to present (AOGCC 2004-2011).

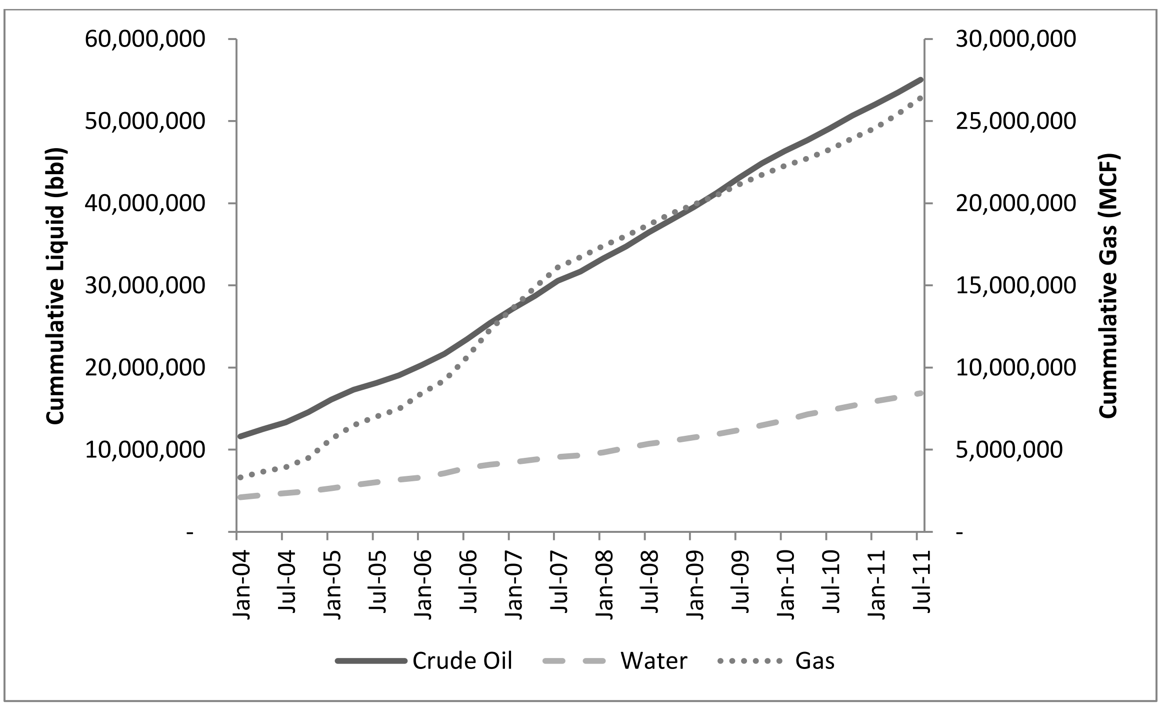

Figure 1-4: West Sak total production from 2004 to present (AOGCC 2004-2011).

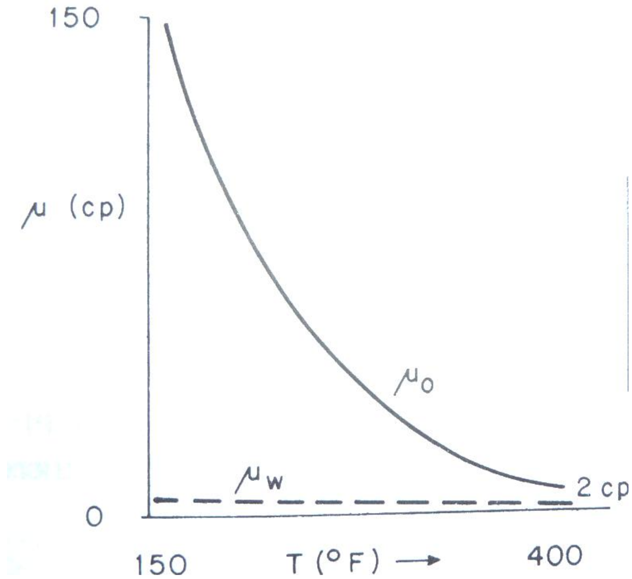

Figure 1-5: Typical oil (μ_o) and water (μ_w) viscosities as a function of temperature (Dake 1978).

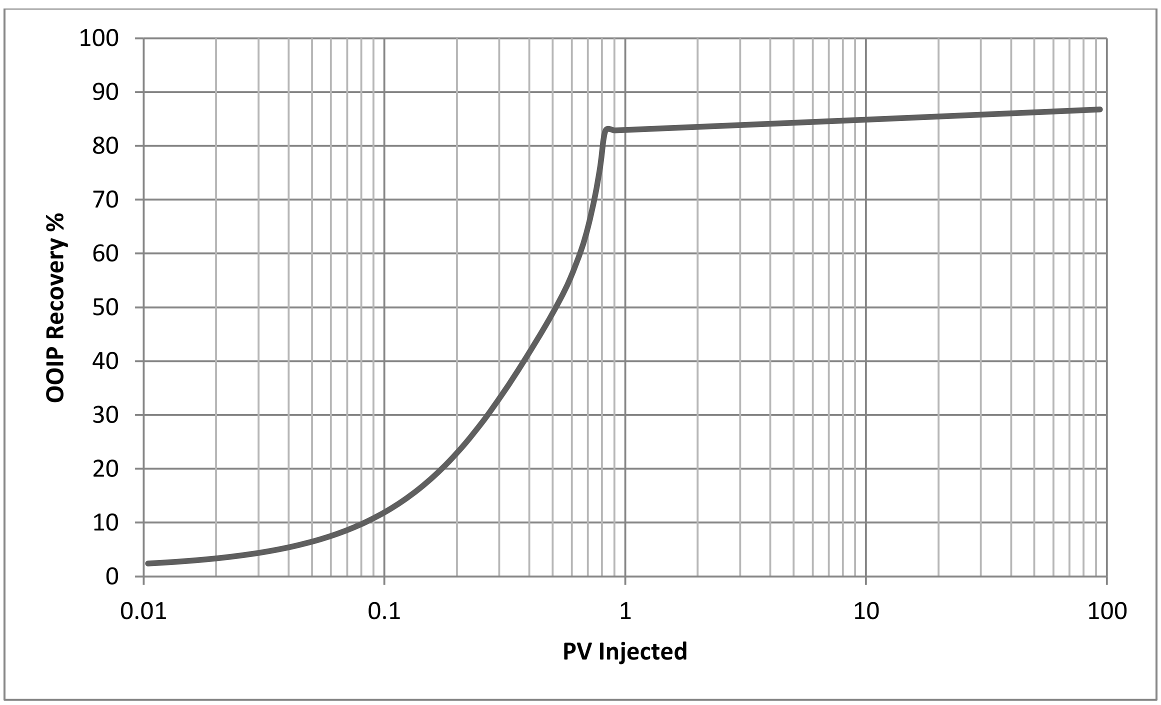

Figure 1-6: Polymer flood OOIP recovery vs. PV injected (Seright 2011).

ECONOMIC AND COST ESTIMATION METHODOLOGY

Discounted cash flows are used as the basic methodology to evaluate the profitability (i.e. economic feasibility) of production process scenarios in this study. This approach is primarily based on the economic analysis method described by Seider *et al*. (2008). As defined by Seider *et al*. (2008), the cash flow is defined as the sum of all costs and revenue in a given amount of time. In this study, cash flows are calculated annually. On this basis, the cash flow for any given year *n* can be calculated using Eq. 2-1:

$$CF_n = P_n(S - C_v) - C_f - T - R \pm C_{WC} - C_{TDC} - C_L - C_R - C_P - C_S \qquad 2\text{-}1$$

where the variables above are defined as:

CF_n Annual cash flow in year *n*

P_n Production capacity fraction (days operated per days in one year) for year *n*

S Total sales at full production capacity

C_v Total variable operating costs (i.e. operating costs that are a function of production capacity such as water, electricity, fuel, and other utilities) at full production capacity

C_f Total fixed operating costs (i.e. costs that are not a function of production capacity such as labor, administration, and insurance)

T Taxes for year *n*, including state, federal, severance, and property taxes

R	Royalties for year *n*, including both oil and intellectual property royalties
C_{WC}	Working capital
C_{TDC}	Total depreciable capital investment
C_L	Capital cost for purchasing and/or leasing land
C_R	Capital cost for intellectual property royalties
C_P	Capital cost for permitting
C_S	Capital cost for startup

To account for the time value of money, the cash flow for each year of a project is multiplied by a discount factor *f*, defined as:

$$f_n = \frac{1}{(1+i)^n} \tag{2-2}$$

where *i* is desired annual interest rate that the entity financing the project wishes to make and *n* is the year of the project. Summing the discounted cash flows for each year of a project gives the net present value of the project (NPV):

$$NPV = \sum_{n=1}^{k} f_n CF_n \tag{2-3}$$

When Eq. 2-3 equals zero (i.e. the net present value of a project is zero), the interest rate *i* is defined as the investor's rate of return (IRR). The IRR is a particularly useful measure of profitability because it accounts for both the time value of money and it normalizes the cash flows for any project. For these reasons, the IRR is used as the primary metric for quantifying profitability in this study.

The single most important assumption that must be made in evaluating the IRR of any of the scenarios for West Sak is picking the sales price for oil. Two different methods are used in this report:

1. Specify an oil price forecast. Given the price of oil each year, calculate cash flows using Eq. 2-1. Solve for the IRR by varying the interest i in Eq. 2-2 so that the NPV in Eq. 2-3 equals zero.
2. Specify the IRR. Given the interest rate i, calculate discount factors from Eq. 2-2. Assume that oil sells for a fixed average price over all years of the project. Solve for the fixed oil price by varying the sale price of oil then calculating the resulting cash flows in Eq. 2-1 and NPV in Eq. 2-3 until the NPV equals zero.

The first method is a better reflection of reality in that it accounts for the time value of oil sales, but it is limited by the accuracy of oil price forecasts. However, like weather forecasts (and other methods of predicting the future), oil price forecasts are notoriously inaccurate predictors of future market prices, especially over the timespan of 20 to 30 years. Therefore, the intent of the second method of analysis is to determine what the price of oil would have to be, on average, in order to make a specified profit.

The project timeline used in all scenarios is discussed below, followed by a more detailed discussion of how individual terms in Eq. 2-1 were estimated.

2.1 Project Timeline

Each scenario evaluated in this study is scheduled to last 20 years, ramping up to an oil production rate of 50,000 bpd. The scheduled activity and spending plan for each year is outlined in Table 2-1.

Design and construction work are each assumed to take one year to complete. Ramping up to the full production capacity of 335 days of operation per year (assuming 30 days of downtime for annual maintenance) is assumed to take two years. The fractions given under the "Investment" column of Table 2-1 represent the fraction of that item spent in the given year. For example, 0.5 or 50% of the total depreciable capital (C_{TDC}) is spent in 2010, but C_{TDC} would be neglected in calculating the cash flow for 2012 using Eq. 2-1. Capital royalties (C_R) are discussed in Section 2.5. Note that working capital (C_{WC}) is accounted for as a cost in 2012 and as a credit in 2030.

2.2 Capital Cost Estimation

Capital costs are one-time expenses that are paid for land acquisition, drilling, equipment, construction, etc. The various capital costs included in this analysis are given in Table 2-2, followed by a more detailed description of the methodologies used to estimate certain capital cost components in Sections 2.2.1 – 2.2.5.

2.2.1 Equipment Costs

Equipment costs are estimated using either the Method of Guthrie or by scaling according to William's six-tenths rule. The Method of Guthrie can be used for calculating the capital cost of individual pieces of process equipment using equations of the form:

$$C = C_P(x)\left[F_{BM} + \left(F_d F_p F_m - 1\right)\right]\left(\frac{I}{I_o}\right) \qquad 2\text{-}4$$

where $C_p(x)$ is the total direct price for a specific category of process equipment as a function of a size factor x. The various F_i are factors for are for bare-module (F_{BM}, which covers indirect costs such as delivery, insurance, taxes, installation, etc.), equipment design (F_d), pressure (F_p), and material (F_m). Finally, the capital cost estimate is adjusted for inflation (I and I_o) using the Chemical Engineering Plant Cost Index (CEPCI). When this costing method was used, the specific form of Eq. 2-4 was taken from Seider *et al*. (2008). Calculating the size factor *x* for each piece of process equipment typically requires a detailed process design. When this information was not available, a scaling rule with the following form was used:

$$C = C_o \left(\frac{Q}{Q_o}\right)^b \left(\frac{I}{I_o}\right) \qquad \text{2-5}$$

where *C* is the cost of a unit or entire process designed for a throughput Q, I is an appropriate cost index, *b* is a scaling power, and the subscript "*o*" refers to the base value of the subscripted variable. Equation 2-4 is referred to as William's six-tenths rule because, according to (Williams 1947), the value of *b* that resulted in the best fit for most pieces of processing equipment was 0.6. This study assumes that $b = 0.6$ wherever Eq. 2-5 is used.

2.2.2 Drilling Costs

The costs for drilling are estimated as a function of total well length based on data from API's 2003 Joint Association Survey (JAS) on drilling costs as published in Augustine *et al*. (2006) and reproduced in Figure 2-1. Based on the data in Figure 2-1, the cost for drilling any well was calculated as:

$$C_{drill} = y(x) f_d f_t \left(\frac{I}{I_o}\right) \qquad \text{2-6}$$

where $y(x)$ is a fourth order polynomial fitted to the data in Figure 2-1, x is the length of the well in feet, f_d is a directional drilling factor, f_t is a well type factor, and I is a cost index. The use of a fourth order polynomial to fit the cost data has no theoretical justification but is useful for interpolating inside the bounds of the data set (well costs calculated with Eq. 2-6 are limited to $1{,}800 \leq x \leq 18{,}500$ where x is well length in feet). If the well is drilled horizontally, it is assumed that $f_d = 1.5$ (for vertical wells, $f_d = 1$). If the well is drilled using coiled tubing, $f_t = 0.6$ (for conventional wells, $f_t = 1$). To compute time-adjusted drilling costs, the Bureau of Labor and Statistics Producers Price Index (PPI) for drilling was used.

2.2.3 Pipeline Costs

The costs for pipelines (water and oil) are based on the costing methodology published by Boyle Engineering Corporation (2002). The cost of the pipeline (in $ / foot of length / inch of diameter) is given by Eq. 2-7:

$$C_{pipe} = 1.05[3.3 + 0.027(D_{econ} - 36)]\left(\frac{I}{6300}\right) \qquad 2\text{-}7$$

where D_{econ} is the optimal economic diameter of the pipeline (which balances the tradeoff between operating and capital costs) in inches and I is the current Engineering News and Record (ENR) index value. Equation 2-7 covers the costs for a pipeline with the following assumptions:

- Buried, with 7 feet or less of cover
- Easily rippable soil
- Undeveloped open and flat terrain with no congestion

- Neutral bidding climate

In order to select the optimal economic diameter for the pipeline, one of the following relations should be used (Peters and Timmerhaus 1991):

Turbulent:
$$D_{econ} = q_f^{0.448} \rho^{0.132} \mu_c^{0.025} \left[\frac{0.88K(1+J)H_y}{(1+F)XEK_F} \right]^{0.158} \quad \text{2-8}$$

Laminar:
$$D_{econ} = q_f^{0.364} \mu_c^{0.182} \left[\frac{0.064K(1+J)H_y}{(1+F)XEK_F} \right]^{0.182} \quad \text{2-9}$$

The values and definitions for terms in Eq. 2-8 and Eq. 2-9 are defined in Table 2-3. In practice, the laminar D_{econ} equation (Eq. 2-9) must first be solved; D_{econ} can then be used to calculate the Reynolds number:

$$Re = \frac{4\rho q_f}{\pi \mu D_{econ}} \quad \text{2-10}$$

If $Re \geq 2{,}000$, the flow is considered turbulent and D_{econ} is calculated from Eq. 2-8; otherwise, the flow is laminar and the result from Eq. 2-9 is the optimal diameter.

In addition to the cost of the pipeline, the cost for pumping stations was calculated using the methodology given by Boyle (2002):

$$C = 46{,}000 \left(\frac{Q}{100} \right)^{0.75} \left(\frac{H}{300} \right)^{0.66} \left(\frac{I}{6300} \right) \quad \text{2-11}$$

where Q is the flowrate in gallons per minute (gpm), H is the pump head (ft), and I is the current ENR index value. The number of pumping stations required is given by:

$$n = \frac{\rho H}{P_{max}} \qquad 2\text{-}12$$

where n is the number of pumping stations, ρ the density of the fluid in the pipeline, and P_{max} is the maximum design pressure of the pipeline (400 psig in this study).

2.2.4 Water Reservoir Costs

The capital cost for constructing the reservoir is based on the estimating methodology published by R S Means Co (2002), which gives guidelines for the costs of specific construction activities on a per unit basis (i.e. per cubic yard, per square foot, etc.). Construction activities included in the cost of constructing the reservoir are given in Table 2-4. The shape of the reservoir is assumed to be the base two-thirds of an inverted square pyramid with a 30% grade. Assuming this geometry, the dimensions of any reservoir capacity can be calculated and its costs computed using the values in Table 2-4. Reported reservoir costs in this study are sufficient for 90 days of process operation.

2.2.5 Utility Plant Costs

All scenarios analyzed in this study include the capital costs of constructing utility plants (steam, electricity, natural gas, etc.) required for their operation following the guidelines given by Seider *et al*. (2008), as summarized in Table 2-5. In addition to the costs given by Seider *et al*. (2008), estimates of the costs for electrical and natural gas lines were solicited from private industry (SageGeotech 2010), as shown in Table 2-6. Note that costs given in Table 2-6 are for the U.S. Midwest Region. Therefore, it is assumed that the investment site factors given by Seider *et al*. (2008) are sufficient for adjusting the costs of the utility lines.

2.3 Sales

Revenue from the sale of oil is calculated as a fraction of the value of West Texas Intermediate (WTI) crude oil based on the historical market price differences between ANS and WTI crudes, as shown in Figure 2-2 (data from U.S. Energy Information Administration (EIA), (EIA 2011)). Since the differential in prices between the two crudes has been steadily decreasing over the past two decades, only the last five years are considered in determining the average price differential of 0.918, which is assumed in the rest of the report for sales of West Sak oil without upgrading.

As discussed in the introduction to this section, the price for oil is assumed to be either fixed at a constant price or specified by a price forecast. With the price forecast option, one of three EIA price forecasts for WTI, based on economic growth rate projections (low, reference, and high), are used to determine oil sales revenue. The different price forecasts are shown in Figure 2-3 (EIA 2010).

For scenarios involving upgrading, the end product is a WTI equivalent crude. As a result, no price differential is assumed and the value of the upgraded crude is assumed to be the same as WTI. Additionally, upgrading produces excess steam and elemental sulfur as byproducts. Any excess steam is sold to the offsite steam utility plant at 50% of the cost of purchasing high pressure (600 psig, 700 °F) steam, a price of $3.48 / k lb. Sulfur is assumed to be sold at 2010 market prices as reported in USGS (2011), $100 /

metric ton. Finally, for scenarios involving oxy-firing, CO_2 is assumed to be sold at $25 / ton (NETL 2010)

2.4 Operating Costs

The operating costs in each scenario can be differentiated into variable (C_V) and fixed (C_F) costs based on whether or not they are functions of the operation of the process. In this report, variable costs are defined as a combination of utilities (water, fuel, electricity, etc.) and other expenses related indirectly to production such as research and administration. Utility requirements are either taken directly from the appropriate process design flow sheet or scaled from base scenario process data using a variant of Eq. 2-5 given below:

$$U = U_o \left(\frac{Q}{Q_o}\right)^b \qquad \text{2-13}$$

where U is the utility requirement, the scaling exponent b is always set to 1, and all other variables are the same as in Eq. 2-5. Most utility costs are estimated from price data given by Seider *et al*. (2008), with supplementary price data coming from EIA (2010), (DOR 2010), (Erturk 2011), and others; see Table 2-7. EIA forecasts for natural gas and electricity are used whenever EIA price forecasts for oil are used to estimate oil sales; otherwise, these prices are fixed at the values given in Table 2-7 from the sources cited above.

Since the majority of the water used in the process is for polymer flooding (see Section 3.1 Production), it is assumed that brine is used as the primary water source.

Where higher quality water is needed, treatment is required and those costs are listed in Table 2-7.

In addition to the utility costs given in Table 2-7, costs for conducting research of \$0.74/bbl of oil produced are also included as a variable expense based on estimates of research spending in Alberta, Canada (Heidrick and Godin 2006).

The fixed expenses in the present scenarios include the cost of labor, property taxes, and insurance, all of which are estimated as suggested by Seider *et al*. (2008). Labor is assumed to be a fixed expense because the large amount of manpower required during maintenance and downtime implies that operational labor would be participating in work during shut downs. Labor related to operations is estimated according to assumed hourly wages and the number of operators required for a sequence of unit operations based on the type of process (solids/fluids) they handle and their throughput. Maintenance-related labor is estimated as a percentage of C_{TDC}, again based upon the type of process. All processes are assumed to require operators 24 hours per day, 7 days per week. Operators will be paid \$30 per hour on average. Maintenance personnel, also paid \$30 per hour, are required for one shift per day. In addition to operators and maintenance personnel, a team of process engineers will be required. The salaries for all process engineers, \$52,000 per operator per shift per year, are accounted for in the technical assistance to manufacturing. Next, workers in the control laboratory are budgeted at \$57,000 per operator per shift per year. Finally, management, including accounting and business services, supervisors, human relations, and the mechanical department, is budgeted as operating overhead based on specific percentages of the total salaries, wages, and benefits of the operators, maintenance personnel, lab personnel, and engineers. Property taxes and insurance are assumed to be a percentage of C_{TPI}. These and other fixed costs are defined in Table 2-8.

2.5 Royalties

Two types of royalties are considered in this study, royalties for intellectual property and royalties for oil. Royalties for intellectual property (R_{IP}) cover the use of patented processes or technology through licensing fees. Following the suggestions given in Seider *et al*. (2008), it is assumed that the licensing of any patented technologies in use in these scenarios is covered by a one-time capital royalty payment of 2% of C_{TDC} and an annual royalty fee of 3% of the cost of manufacture (COM – defined as the sum of all operating costs except for general expenses, i.e. research, administration, and management incentive compensation). Royalty payments for oil (R_{oil}) are paid to the land owner for removing mineral wealth from the leased property. The predominant land owners in the North Slope are the Federal and State government, and both charge the same fixed rate of 12.5% (one-eighth) of the sales value of oil (S_{oil}). The total amount paid in any given year in royalties is thus:

$$R = R_{IP} + R_{oil} = 0.03(COM) + 0.125(S_{oil}) \qquad \text{2-14}$$

2.6 Taxes

Four different taxes are calculated in this study, corporate income tax (federal and state), severance tax, and property tax. Severance taxes (ST) are taxes imposed by a state on the extraction of natural resources, including oil, regardless of land ownership. The current Alaskan severance tax policy is based on the “Alaska’s Clear and Equitable

Share" (ACES) law passed in 2007 and is administered by Alaska's Department of Revenue (DOR). According to DOR, ACES consists of a base 25% severance tax rate on the wellhead profit (WHP, defined as the value of the oil at the wellhead after deducting all costs related to its extraction) with a progressive surcharge on wellhead profit above $30/bbl of 0.4% for each additional $1 increase in per barrel WHP (DOR 2010). Once the base rate and progressive surcharge reach 50% of the WHP, the rate of growth of the surcharge reduces to 0.1% until capping out at a maximum nominal tax rate of 75%. In this report, it is assumed that the WHP is equal to the value of ANS crude less the cost of oil royalty payments and the proportion of operating costs assumed to be associated with extracting the oil (f_E). Stated mathematically:

$$f_E = \frac{sum\ of\ all\ capital\ spent\ on\ extraction}{C_{TDC}} \qquad 2\text{-}15$$

$$WHP = S - R_{oil} - f_E(C_V + C_F) \qquad 2\text{-}16$$

Corporate tax rates are 35% and 9.4% of taxable income for federal (t_F) and state (t_S) government, respectively. State corporate taxes are deductible from federal corporate taxes. Additional deductions can be taken from corporate tax liability for royalties, severance taxes, all expenses, depreciation, and depletion. Depreciation (D) is assumed to follow the ten-year Modified Accelerated Cost Recovery System (MACRS) shown in Table 2-9.

Depletion (d) is calculated following the percentage depletion method appropriate for small oil and gas producers (i.e. 15% of sales revenue). Accounting for all deductions, the taxable income (TI) for the project in a given year is:

$$TI = P(S - C_V - d) - C_F - D - R - ST \tag{2-17}$$

Therefore, the total federal and state corporate income tax is:

$$T = T_F + T_S = TI[t_F(1 - t_S) + t_S] \tag{2-18}$$

or 41.11% of TI.

Property tax has already been described in Section 2.4 and is counted as a fixed operating expense, following the accounting approach suggested by Seider *et al*. (2008).

Table 2-1: Project timeline

Chronology			**Investment**					
Action	**Year**	**P_n**	**C_{TDC}**	**C_{WC}**	**C_L**	**C_R**	**C_P**	**C_S**
Design	2010	0.000	0.5				0.5	
Construction	2011	0.000	0.5		1.0	1.0	0.5	
Startup	2012	0.450		-1.0				1.0
Startup	2013	0.677						
Production	2014	0.904						
Production	2015	0.904						
Production	2016	0.904						
Production	2017	0.904						
Production	2018	0.904						
Production	2019	0.904						
Production	2020	0.904						
Production	2021	0.904						
Production	2022	0.904						
Production	2023	0.904						
Production	2024	0.904						
Production	2025	0.904						
Production	2026	0.904						
Production	2027	0.904						
Production	2028	0.904						
Production	2029	0.904						
Production	2030	0.904		1.0				

Table 2-2: Capital cost categories and their estimation methodologies. Unless otherwise noted, specific values are from Seider *et al*. (2008).

Category	Component	Description
Total Bare Module Investment (C_{TBM})	Equipment	Capital cost of all equipment required for extracting, processing, and transporting heavy oil as defined by production scenario. Includes all direct (material, installation labor, etc.) and indirect (construction overhead, engineering, etc.) costs for each piece of equipment. See Section 2.2.1 for more detail.
C_{TBM} =	(Sum of all equipment)	
Total Direct Permanent Investment (C_{DPI})	Drilling	Cost for drilling all wells. See Section 2.2.2 for more detail.
	Site Preparation	10% of C_{TBM}, covers land surveys, dewatering and drainage, surface clearing, excavation, grading, piling, fencing, roads, sidewalks, railroad sidings, sewer lines, fire protection facilities, and landscaping.
	Service Facilities	20% of C_{TBM}, covers utility lines, control rooms, laboratories for feed and product testing, maintenance shops, etc.
	Oil Pipeline	See Section 2.2.3 for more detail.
	Water Pipeline	See Section 2.2.3 for more detail.
	Water Reservoir	Construction of a reservoir large enough to hold all the water required for 90 days of process operation. See Section 2.2.4 for more detail.
	Allocated Costs for Utility Plants	Includes utility plants for steam, electricity (substation, line, switch gear, and tap), water, refrigeration, and natural gas (line, metering, and regulation facility). See Section 2.2.5 for more detail.
C_{DPI} =	(Sum of the above) + C_{TBM}	

Table 2–2: Continued

Category	Component	Description
Total Depreciable Capital (C_{TDC})	Contingency	15% of C_{DPI}, accounts for any higher than expected capital cost components listed above.
C_{TDC} =	Contingency + C_{DPI}	
Total Permanent Investment (C_{TPI})	Land	2% of C_{TDC}, covers all land purchases and leasing costs.
	Permitting	$0.10 per bbl of oil produced to cover all permitting requirements (Snarr 2010).
	Capital Royalties	2% of C_{TDC}, covers initial licensing fees for any proprietary technology used in process.
	Startup	10% of C_{TDC}, covers additional costs of getting process into steady-state operation.
	Site Factor	1.25, represents fractional increase in cost of capital cost components listed above compared to U.S. Gulf Coast region.
C_{TPI} =	(Site Factor) [(Sum of the above) + C_{TDC}]	
Total Capital Investment (C_{TCI})	Working Capital (C_{WC})	15% of C_{TPI}, represents funds required on hand to cover business accounting.
C_{TCI} =	C_{WC} + C_{TPI}	

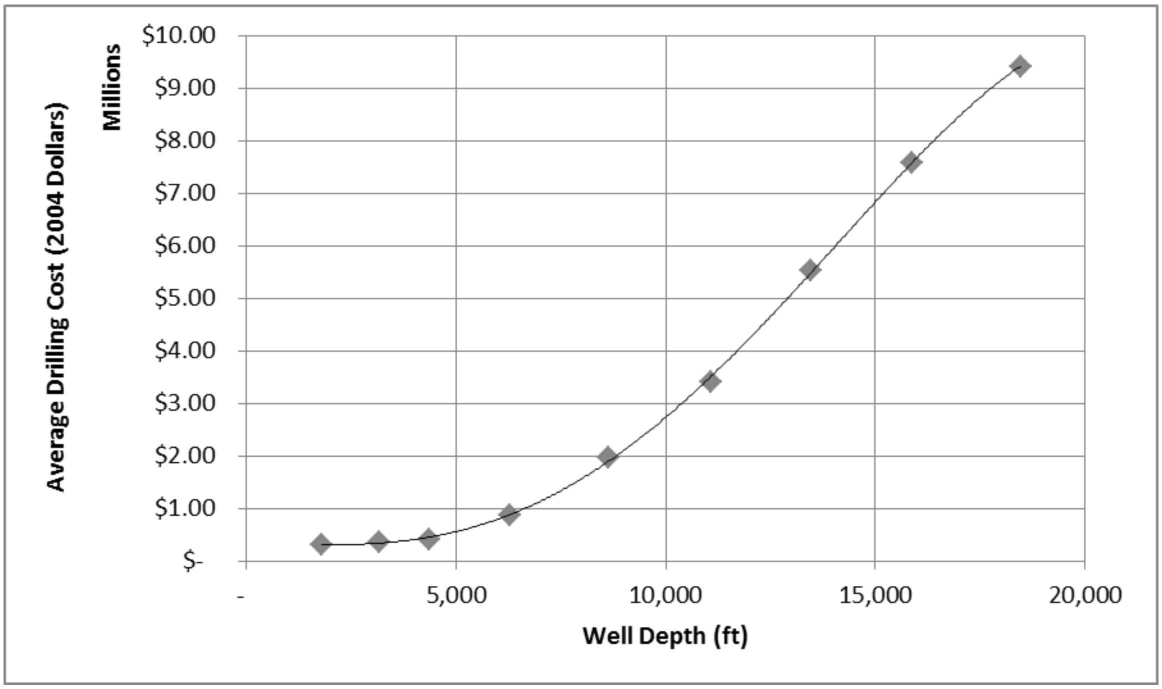

Figure 2-1: Drilling cost for a conventional vertical well as a function of depth in 2004 dollars. Based on data from API's 2003 JAS published in Augustine *et al*. (2006).

Table 2-3: Variables for pipeline sizing.

Variable	Value	Units	Definition
q_f	Independent Variable	ft^3/s	Fluid flow rate
ρ	62.3 (water), 52.75* (oil)	lb/ft^3	Fluid density
μ_c	1.002 (water), 1.41* (oil)	cP	Fluid viscosity
K	0.07	$/kWh	Cost of electricity
J	0.35	---	Fractional loss due to fittings and bends
H_y	7920	hr/yr	Hours of operation per year
F	1.4	---	Ratio of total cost for fittings and installation to purchase cost for new pipe
X	1.14	$/ft	Purchase cost of new 1” diameter pipe per foot of pipe length
E	0.72	---	Efficiency of the pipe’s motor and pump
K_F	0.2	---	Annual fixed charges for financing and maintenance expressed as a fraction of total pipe cost

* From ProMax process simulator data.

Table 2-4: Water reservoir construction methods and costs.

Step	**Task**	**Cost***	**Per**
Excavate Reservoir	Excavating	$1.07	Cubic yard
	Truck Loading	----	15% of Excavating Cost
	Hauling	$2.15	Cubic yard
Line with Clay (1 ft. thick)	Backfill	$1.36	Cubic yard
	Compaction	$1.57	Cubic yard
	Clay Purchase	$6.50	Cubic yard
Line with Plastic	Sheet Waterproofing	$1.81	Square foot

* Note: costs listed are in 2002 dollars and must be adjusted using the Means Historical Cost Index

Table 2-5: Allocated costs for utility plants (Seider, *et al*. 2008).

Utility	Capital Cost Rate
Steam	$50 / lb / hr
Water (cooling)	$58 / gpm
Refrigeration	$1,330 / ton

Table 2-6: Costs for electrical and natural gas lines (SageGeotech 2010). Costs given are in 2010 dollars for U.S. Midwest Region (site factor = 1.15).

Line	**Item**	**Cost (per mile)**
Electricity	Line	$425,000
	Switching Gear and Tap	$10,000
Natural Gas	Line	$1,056,000
	Metering and Regulation Facility*	$1,000,000

* Note: Flat fee, not a function of line length.

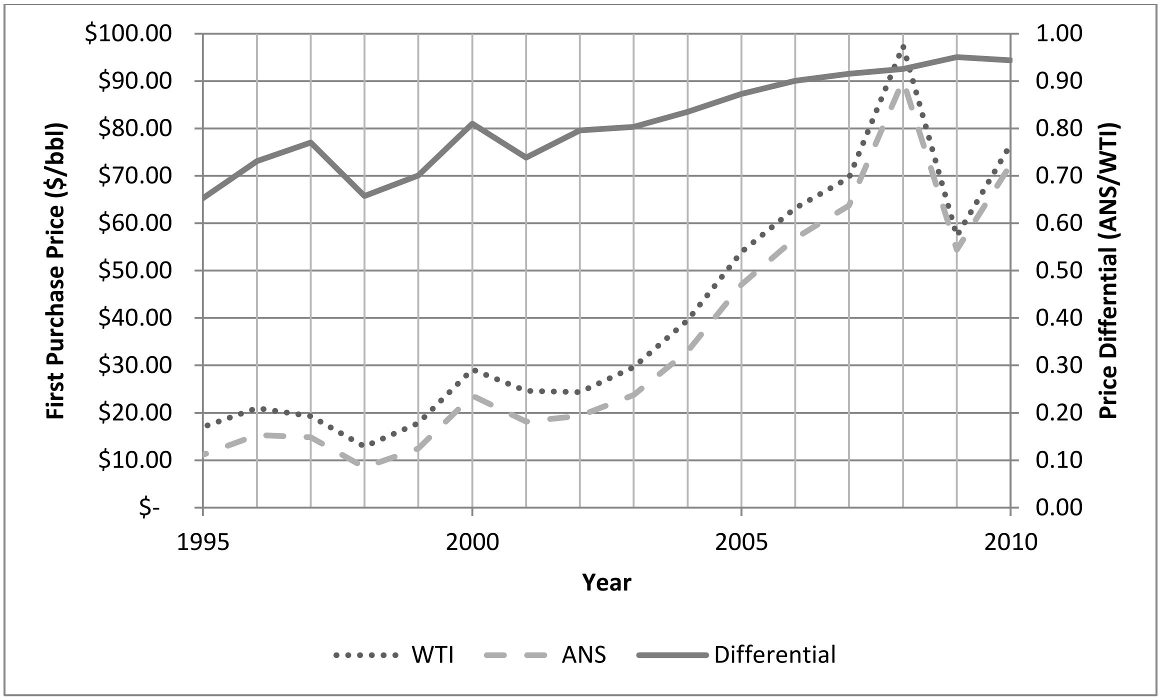

Figure 2-2: Historical price trends for WTI and ANS crude (EIA 2011).

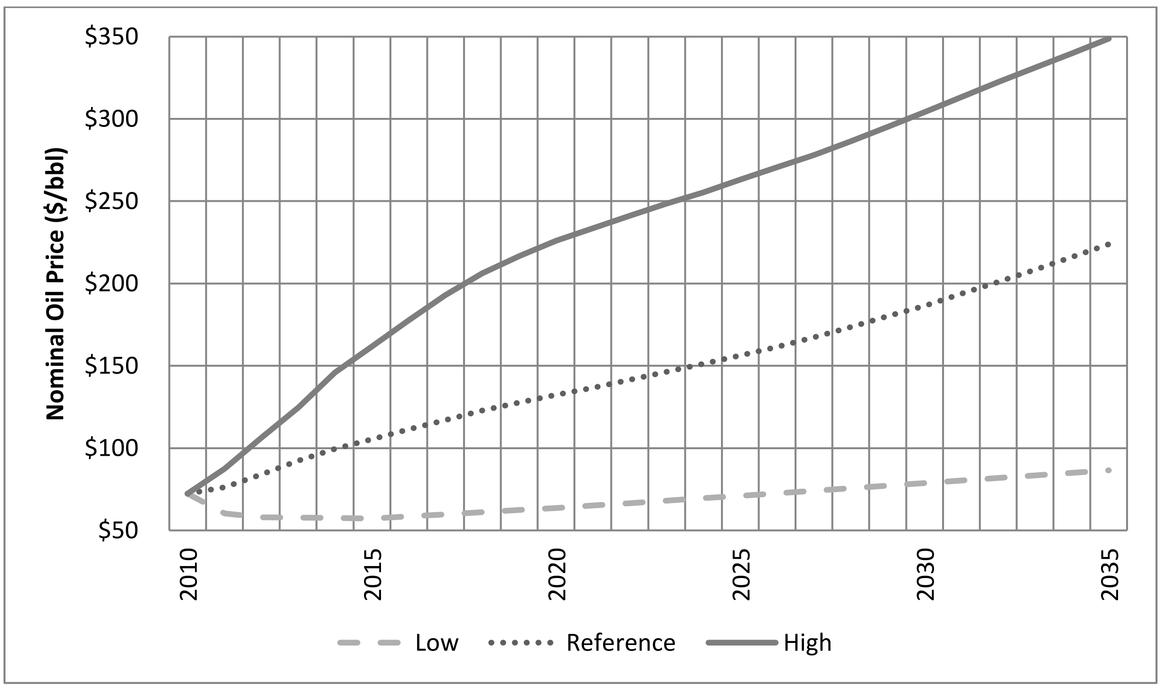

Figure 2-3: EIA WTI oil price forecasts for low, reference, and high economic growth rates (EIA 2010).

Table 2-7: Utility pricing.

Utility	Price	Per
Catalyst	$4.24	kg
CO_2		
Tax Rate	$25.00	ton
Sale Rate	$25.00	ton
Diluent	$70.00	bbl
Electricity	$0.058	kWh
Fuel (natural gas)		
Purchase price	$6.202	MMBtu
Transmission fee	$0.18	MMBtu
Fuel reimbursement fee	1.37%	of annual purchase cost
Oxygen	$70.00	ton
Polymer	$1.17	lb
Refrigerant (R-134a)	$7.90	GJ
Steam		
150 psig	$3.00	k lb
450 psig	$6.60	k lb
Tanker Fee	$2.05	bbl
TAPS Tariff	$4.10	bbl
Water Treatment		
Boiler feed	$1.80	k gal
Process	$0.50	k gal
Cooling	$0.08	k gal

Table 2-8: Fixed costs based on Seider, *et al*. (2008) with modifications.

Cost	Method of Calculation
Labor for Operations	
Wages and benefits (LW)	LW = \$30/operator-hr
Salary and benefits (LS)	LS = 15% of LW
Operating supplies and services	6% of LW
Technical assistance to manufacturing	\$52,000/(operator/shift)/yr
Control laboratory	\$57,000/(operator/shift)/yr
Maintenance	
Wages and benefits (MW)	MW = $F_P * C_{TDC}$
Fluid processing	F_P = 3.5%
Solids and fluids processing	F_P = 4.5%
Solids processing	F_P = 5.0%
Salary and benefits (MS)	MS = 25% of MW
Materials and services	100% of MW
Maintenance overhead	5% of MW
Operating Overhead	
General plant overhead	7.1% of (LW + LB + MW + MB)
Mechanical department services	2.4% of (LW + LB + MW + MB)
Employee relations department	5.9% of (LW + LB + MW + MB)
Business services	7.4% of (LW + LB + MW + MB)
Property Tax	1.0% of C_{TPI}
Insurance	0.4% of C_{TPI}
General Expenses	
Administrative expense	\$200,000/(20 employees)/yr
Management incentive compensation	1.25% of net profit

Table 2-9: Ten-year MACRS depreciation schedule.

Year	Percent of C_{TDC} Claimed
1	10.00
2	18.00
3	14.40
4	11.52
5	9.22
6	7.37
7	6.55
8	6.55
9	6.56
10	6.55
11	3.28

PROCESS DESCRIPTION

Three scenarios were analyzed in this study for extracting heavy oil from West Sak. The two main steps in each are drilling/production and delivery to market. Based on the reservoir properties reported in the literature, we have assumed that polymer flooding is used in all scenarios. Produced oil is then transported to the TAPS terminal in Prudhoe Bay by a feeder pipeline. A flat TAPS tariff and marine shipping costs are paid and the heavy oil is presumably sold to a refinery on the U.S. West Coast. However, since the pipeline compatibility of heavy oil from West Sak has been questioned (Olsen, Taylor and Mahmood 1992), two possible alternative scenarios are considered: (1) diluting the heavy oil with either gas-to-liquids (GTL) oil products or natural gas liquids (NGL), or (2) upgrading the heavy oil through hydrotreating. Finally, oxy-firing is considered as an alternative combustion system for providing heating in each scenario to address the potential impact of regulation on CO_2 emissions. The steps included in each scenario are summarized in Table 3-1. A process flow diagram is shown in Figure 3-1.

3.1 Production

Heavy oil is produced using a line-drive (alternating injector/producer wells) polymer flood from horizontally drilled wells, as shown in Figure 3-2. Based on data published in Sorbie (1991), a concentration of about 1,720 ppm polyacrylamide (HPAM), or 0.48 lb HPAM per cubic meter of water, would be sufficient to give the water a viscosity of 35.4 cP. Seright's (2011) injection vs. recovery curve (see Figure 1-6) is used as the basis for determining oil production. At the beginning of the flood, each unit volume of water injected into the reservoir displaces an equivalent volume of reservoir fluid. This trend continues until approximately 0.8 PV have been injected, at which point the injected fluid front reaches the producer well and "breaks through," dramatically decreasing the oil production rate.

For the purpose of determining OOIP based on PV, PV is defined as:

$$PV = \phi WHL \tag{3-1}$$

where ϕ is the porosity of the reservoir, *W* is the length of one lateral well segment, *H* is the thickness of the reservoir, and *L* is the distance between wells. Assuming that there is no gas present in the reservoir, the OOIP (in stock tank barrels, stb) is:

$$OOIP = PV(1 - S_{WC}) \tag{3-2}$$

where S_{WC} is the connate water saturation. If fluid is injected at a rate Q through n_{inj} injector wells, then the total oil production N_P rate is:

$$N_P = 4n_{inj}mQ\frac{OOIP}{B_o} \tag{3-3}$$

where *m* is the slope of the linear section of the OOIP recovery curve prior to breakthrough given by Seright (2011) and B_o is the oil formation volume factor. Equation 3-3 is multiplied by four since four PV are flooded by each injector well.

Darcy's law for fluid flow through porous media is used to determine the pumping pressure and work required to meet the injection rate Q through each injection well. Fluid flowing through the reservoir moves with a velocity *u* given by Darcy's law:

$$u = -\frac{\kappa}{\mu}\frac{dP}{dx} \qquad \text{3-4}$$

where κ is permeability, μ is viscosity, and *dP/dx* is the pressure gradient in the reservoir. Equation 3-4 is written assuming 1-D flow through porous media with no changes in elevation (i.e. that there is no potential energy difference between any point in the reservoir). Integrating Eq. 3-4 with the boundary conditions at the injection well $x = 0$ and $P = P_o$ and at the producer well $x = L$ and $P = P_L$ gives:

$$u = -\frac{\kappa}{\mu}\frac{(P_L - P_o)}{L} \qquad \text{3-5}$$

The velocity *u* given by Eq. 3-5 is the velocity of the mixture of oil and water flowing through the cross-sectional area between the injector and producer. If the total horizontal length of each well is *W* and the thickness of the reservoir is *H*, then the cross-sectional area is $A = WH$ and the volumetric flowrate Q is:

$$Q = uWH \qquad \text{3-6}$$

Substituting Eq. 3-5 into Eq. 3-6 for *u* and solving for the pressure change gives:

$$P_o - P_L = \frac{Q\mu L}{\kappa WH} \qquad \text{3-7}$$

The pressure at the producer well is the hydrostatic pressure of the column of fluid in the well. The pressure at the injector well is the sum of both the hydrostatic pressure of the fluid column and whatever pressure is applied by pumping (P_{inj}). Assuming that the producer and injector wells are at the same depth and neglecting the density difference between water and oil, the hydrostatic pressures in each well cancel each other out, reducing Eq. 3-7 to:

$$P_{inj} = \frac{Q\mu L}{\kappa W H} \qquad 3\text{-}8$$

The injection pressure can then be used to determine capital and operating costs for pumping following the costing methodology given by Seider *et al*. (2008).

In addition to the main steps for production outlined above, several other components are required. Water (brine) required for injection is pumped from Smith Bay to West Sak (approximately 26 miles). Mixing injection water and polymer and separating produced oil, water, and gas is accomplished with mixing and separating tanks. Both mixing and separating tanks are sized to accommodate five minutes of holdup time. It is assumed that any gas produced is reinjected into the reservoir. Natural gas and electrical lines are assumed to run straight from Atqasuk, AK to West Sak (approximately 57 miles). Two different combustion systems are considered to supply the heating required to keep the mixing and separating tanks at process operating temperatures: air-fired and oxy-fired. Both systems are shown in Figure 3-1; the dashed lines are for processes that only apply to oxy-firing. In the air-fired system, natural gas is combusted with air and the effluent is sent to a stack. In the oxy-fired system, natural gas is combusted with pure O_2 that has been mixed with recycled flue gas (RFG). The design specifications for production are summarized in Table 3-2.

3.2 Delivery

Delivery is accomplished in three stages: a feeder pipeline from the location of the reservoir to the TAPS terminal at Prudhoe Bay, TAPS pumping from Prudhoe Bay to a tanker at Valdez, AK, and finally, tanker delivery to a refinery on the West Coast. The costs for the feeder pipeline are calculated according to the procedure outlined in Section 2.2.3, assuming a straight-line path of approximately 154 miles. Costs for TAPS and tanker delivery are calculated on a per barrel basis from data reported by the Alaska Department of Revenue (2010).

3.3 Alternative Scenarios

3.3.1 Dilution

In this scenario, a lighter, miscible hydrocarbon such as NGL or GTL is added to the heavy oil produced from West Sak to reduce its viscosity. Based on the rheology of GTL and ANS crudes (Inamdar, *et al*. 2006), a mixing ratio of 1 to 2.5 GTL / crude was selected. Market prices reported by Erturk (2011) were used to establish the cost of purchasing GTL for use as a diluent, and the sales price for the mixture is assumed to be the same as that for WTI crude. Since a significant volume of GTL (14,286 bpd) is used, a smaller production rate of heavy oil is sufficient to meet the desired production volume of 50,000 bpd.

3.3.2 Upgrading

For this scenario, hydrotreating is used to improve the quality and pipeline compatibility of heavy oil produced from West Sak by removing impurities such as sulfur and nitrogen and saturating hydrocarbons (HC) in the heavy oil. The process leads to an overall decrease in oil density (i.e. increase in API gravity), and as such, only 45,279 bpd of heavy oil is required to produce 50,000 bpd of WTI-like crude.

Most of the process design for the hydrotreating section is based on process flowsheet simulations performed by Castro (2010) with ProMax software. Supplemental costing data for certain process steps were based on data from Maples (2000) and scaled as described in Section 2.2 and 2.4. A process flow diagram for the hydrotreating process is shown in Figure 3-3. A detailed process description for each step is given below.

3.3.2.1 Fractionation

The fractionator is an atmospheric distillation column that separates the condensed HC vapors and various gases (CO, CO_2, NH_3, H_2S, H_2O, and H_2) coming from the heavy oil feed. The oil and gases from the retort are separated into the following streams:

- Gases
- Fouled water
- Naptha - boiling range 100°F - 400°F
- Vacuum Gas Oil (VGO) - boiling range 400°F - 950°F
- Wax - boiling range > 950°F

The three different distillation cuts (naptha, VGO, and wax) comprising the heavy oil product from the fractionator are stored in heated surge tanks until they are moved to the hydrotreater for upgrading. Sour gases and fouled water are sent to the ammonia scrubber and sour water stripper, respectively. Capital and operating costs for the fractionator are scaled from data given by Maples (2000).

3.3.2.2 Hydrotreater

The process of hydrotreating, depicted in Figure 3-4, takes place in a catalytic reactor where H_2 is reacted with the heavy oil. Aromatic components of the oil are converted to aliphatic components, nitrogen to NH_3, and sulfur to H_2S. Heavy metals are confined to the coke residue. The process begins by pumping raw oil from storage and heating it to reactor entrance conditions (450°C). The raw oil enters the top of the reactor and trickles down through the catalyst where it reacts. The reaction products are given by Eq. 3-9:

$$Feed + H_2 \xrightarrow{catalyst} Saturated\ HC + H_2S + NH_3 + H_2 \tag{3-9}$$

Consumption of H_2 in the hydrotreater is determined from Figure 3-5 based on the composition of the oil feed. For this scenario, H_2 consumption is estimated to be 26 m^3 (2.14 kg) per barrel of heavy oil feed.

Gaseous byproducts (H_2S and NH_3) are removed from the hydrotreating unit in the purge stream, which is sent to the ammonia scrubber as described in Section 3.3.2.4. A sour water stripping unit is also included to remove these same byproducts from the hydrotreater's recycled cooling water (see Section 3.3.2.7). The annual production of H_2S and NH_3 is noted in Table 3-3.

Heat requirements for the catalytic reactor are supplied by a natural gas combustion system. Heat integration is used to lower process energy requirements. After the reactor, the gas stream passes through a flash unit to remove condensable gases (mostly H_2) that are recycled back to the reactor. The upgraded oil is cooled down and sent to storage awaiting pipeline transportation. More detailed information, including mass and energy flows associated with the hydrotreater, can be found in Castro (2010).

The properties of the raw and upgraded heavy oil are given in Table 3-4. The upgraded oil is of high quality: 35°API, low pour point, low in sulfur, and low in nitrogen. Table 3-4 also shows a direct comparison between the upgraded oil and three common reference crude oils: WTI, Brent Crude oil, and Arabian Light Crude.

3.3.2.3 Hydrogen Plant

The hydrogen required for the hydrotreater is produced by the steam reformation of natural gas. A schematic of the overall process is shown in Figure 3-6. The process uses natural gas, O_2, and water as feedstocks to produce H_2 in two steps. The key step for producing H_2, the steam reforming reaction, is given by:

$$CH_4 + H_2O \rightarrow CO + 3H_2 \qquad \text{3-10}$$

where H_2O is introduced to the reactor as steam. This reaction is endothermic and requires a large amount of heat, which is generated by the combustion of natural gas and tail gases (H_2, CO, CH_4) from the pressure swing adsorption unit (PSA). The byproduct CO is used to produce additional H_2 in the water-gas shift reactor in the slightly exothermic reaction:

$$CO + H_2O \rightarrow CO_2 + H_2 \qquad \text{3-11}$$

While water-gas shift reactions are typically carried out in two stages with a high (350°C) and a low (200°C) temperature step (Fleshman 2004), in this work, acceptable levels of CO conversion were achieved with only the high temperature step.

Following the water-gas shift reactor, the raw gas stream is cooled and scrubbed prior to entering the PSA. The PSA produces an H_2 product stream that is 99.9% pure and contains 50% of the H_2 present in the inlet raw H_2 stream. The waste gas stream from the PSA containing the other 50% of the H_2 and other tail gases is sent back to the steam reformer for combustion. For additional details, including the catalysts employed in the reformer and in the shift reactor, see Fleshman (2004). This PSA-based H_2 production system produces significant amounts of excess steam, generated from various heat exchangers. In the present analysis, steam that is generated is sold back to the offsite steam utility at 50% of the cost of purchasing high pressure steam (600 psig, 700^{o}F).

Capital and operating costs for the hydrogen plant are scaled based on data from Fleshman (2004).

3.3.2.4 Ammonia Scrubber

Sour gases separated from the fractionator and generated as byproducts in the hydrotreater are fed to a wet scrubber with dilute sulfuric acid. Ammonia passing through the scrubber reacts with the acid to form ammonium sulfate (a fertilizer):

$$2NH_3 + H_2SO_4 \rightarrow (NH_4)_2SO_4 \qquad 3\text{-}12$$

Based on the amount of ammonia produced by the hydrotreater, approximately 190,000 tons of ammonium sulfate would be generated by the scrubber annually, which is assumed to be sold at a price sufficient to cover the expenses of operating the scrubber (both the scrubber's costs and the ammonium sulfate revenue are neglected in this analysis). After passing through the ammonia scrubbers, the waste gas stream is sent to the amine treatment unit for H_2S removal as described in Section 3.3.2.5.

3.3.2.5 Amine Treatment Unit

Acid gases are scrubbed from the waste gas streams by contacting them with an amine, such as diethanol amine (DEA), in an absorber column. The amine reacts with the acid gases such as H_2S to produce a water soluble salt:

$$(OHCH_2CH_2)_2NH + H_2S \leftrightarrow (OHCH_2CH_2)_2NH_2^+HS^- + Heat \tag{3-13}$$

This reaction is reversed in the amine regeneration column to produce a concentrated acid gas stream, which is sent to the sulfur recovery unit (see Section 3.3.2.6). The sweet gas and fuel gas streams are then burned to recover their heating value. A process flow diagram for a typical amine treatment unit is shown in Figure 3-7.

Capital and operating costs for the amine treatment unit are scaled from data in Maples (2000).

3.3.2.6 Sulfur Recovery Unit

Elemental sulfur is recovered from the acid gas waste stream in the sulfur recovery unit using the Claus process, which involves the following chemical reactions:

$$H_2S + \frac{3}{2}O_2 \rightarrow SO_2 + H_2O \tag{3-14}$$

$$2H_2S + SO_2 \rightarrow 3S + 2H_2O \qquad 3\text{-}15$$

In the first reaction step, one-third of the H_2S in the acid gas stream is burned in a thermal reactor; the fraction of H_2S participating in the combustion reaction is controlled by limiting the amount of oxygen present. The result is a stoichiometric mixture of H_2S and SO_2. This mixture is then passed over a catalyst which allows for the second reaction to occur, creating gaseous elemental sulfur. The gaseous sulfur is then removed by condensation (the waste heat is used to generate steam). This process can be repeated up to four times by reheating the gas stream after condensation and passing the gases over another catalyst bed to achieve sulfur recoveries of up to 98% (Maples 2000). Further sulfur removal requires the use of a tail gas treating unit (see Section 3.5). A typical example of a process flow diagram for a sulfur recovery unit is given in Figure 3-8.

Capital and operating costs for the sulfur recovery unit were scaled from data in Maples (2000). We have assumed a sulfur recovery rate of 95% and that any sulfur recovered is sold at market prices (USGS 2011).

3.3.2.7 Sour Water Stripper

Fouled water from the fractionator and recycled cooling water from the hydrotreater is processed through a sour water stripper to remove any NH_3, H_2S, or other dissolved contaminants that have collected in the water. Contaminants are removed from the water using steam generated from the sour water itself in a stripping column, as shown in Figure 3-9. Stripped water is then sent to the water reservoir for reuse. Any acid gases removed from the water are sent to the ammonia scrubbers described in Section 3.3.2.4.

Capital and operating costs for the sulfur recovery unit were scaled from data in Maples (2000).

3.4 Labor Utilization

Skilled and maintenance labor as well as management are required for each scenario. The number of employees is determined for each unit operation of the process as listed in Table 3-5 on a per shift basis, following the labor estimating guidelines given by Seider *et al*. (2008). Assuming that five shifts per week are used for 24/7 operation, the total number of employees for the base case or the diluent scenarios is 35. The labor requirement for upgrading, however, is much higher, 325 employees with air-firing or 360 with oxy-firing due to the greater number of unit operations involved in upgrading. However, because of the uncertainty associated with the labor estimating methods of Seider *et al*. (2008), actual labor requirements could be quite different from those predicted here.

3.5 Environmental Aspects of Heavy Oil Production

While the cost analysis in this report does not include all of the externalities associated with heavy oil production, the costs for air pollution control, carbon management, and water management are included as described below.

3.5.1 Air Pollution Control

As discussed in Sections 3.3.2.4 through 3.3.2.7, this scenario includes the costs of removing H_2S from the various sour gas streams generated by the upgrading of heavy oil. Capital and operating expenses for removing NH_3 are assumed to be offset by the sale of ammonium sulfate; see Section 3.3.2.4. All other capital costs for air pollution control equipment for this scenario could be computed based on flow rate estimates of the waste air streams, but that information is not available for all unit operations in this study. In addition, operating costs are extremely difficult to estimate. Hence, these additional costs for air pollution control are assumed to be covered by the scenario's contingency cost. Given its low cost impact, this assumption is not seen as a serious omission for the purposes of this analysis.

3.5.2 Carbon Management

Given the uncertainty of the regulatory climate with respect to carbon, a careful accounting of CO_2 production, possible mitigation methods, and potential costs are an essential part of this scenario. To accomplish these objectives, two different combustion systems (each with a different carbon emissions strategy) are considered to supply heat for the various unit operations and the steam plant.

In the conventional system, natural gas is combusted with air and the resulting combustion gases are sent to a stack. Combustion stack gases are scrubbed for sulfur oxides (SO_X) removal when the fuel has significant quantities of sulfur in it. For this system with no CO_2 capture, two cases are considered in the supply cost analysis that follows: (1) no tax on CO_2 and (2) a $25 per ton tax on CO_2.

In the oxy-combustion system, natural gas is combusted with a mixture of O_2 and recycled flue gas consisting primarily of CO_2 and water. Using a ProMax simulation, the product gases are then cooled to cryogenic conditions in a series of heat exchangers so that condensable gases such as water, H_2S and NH_3 can be removed; see Figure 3-10. The nearly pure CO_2 stream that remains after cryogenic treatment is compressed to pipeline conditions. Equipment sizes and operating requirements for the CO_2 compression system are calculated using ProMax.

The O_2 used in the process is purchased from a supplier at a given price per kilogram. These costs are then offset by the sale of CO_2 for enhanced oil recovery (EOR). The sale price is assumed to be $25 per ton at pipeline conditions. The costs for any CO_2 pipeline are assumed to be the responsibility of the purchaser and are not included in our analysis. Additional details about the CO_2 compression and cleanup plant can be found in Castro (2010).

Carbon dioxide is produced from the heat requirements of the mixing and separating tanks, fractionator, hydrotreater, and hydrogen plant. However, not all scenarios utilize all of the listed unit operations. The total CO_2 production from each scenario with air and oxy-firing is listed in Table 3-6. Note that only the CO_2 directly produced by combustion of natural gas is accounted for in this analysis (i.e. CO_2 produced for generation of steam, electricity, etc. are excluded).

3.5.3 Water Management

Water is used primarily for polymer flooding in all scenarios. As noted in Section 3.1, water is pumped from Smith Bay to the injection site. It is assumed that there is no cost for using this water other than the costs of pumping it to the injection site. Any water produced from the reservoir is pumped back through the injection wells. However, since breakthrough does not occur until OOIP recovery is about 80% and the maximum recovery in any of the scenarios (within the 20-year operating range of this analysis) is only 55%, produced water volumes are expected to be much smaller than the required injection rates.

In the upgrading scenario, additional water is used in several process units. Brine is still used as the feedstock, but it is cleaned at offsite water treatment plants to the extent required for its usage. Water in the form of steam is used in the hydrogen plant as a reactant; cooling water and process water are also used. The hydrotreater and sulfur recovery unit use water as steam and as cooling water. For oxy-fired scenarios, the CO_2 compression plant uses cooling water in interstage coolers. Other small water uses include evaporation of cooling water from the cooling tower (assumed to be 3% of the cooling water flow rate), steam losses, and water used for various scrubbers. Dirty water is cleaned in the sour water stripper and recycled through the treatment plants.

In order to buffer the process from water supply disruptions, a reservoir is constructed onsite of sufficient size to hold an equivalent of 90 days of water (see Section 2.2.4 for more details about the reservoir's construction and geometry). The total annual water usage and reservoir size for each scenario are given in Table 3-7.

Table 3-1: Summary of scenario production steps.

Scenario	Base Case	Dilution	Upgrading
Extraction	Line-drive polymer flood from horizontal wells		
Delivery	Feeder pipeline from site → TAPS → Tanker→ Market		
Heating	Air-fired and oxy-fired variants		
Additional Steps	None	Dilution with GTL or NGL purchased on North Slope	Hydrotreating

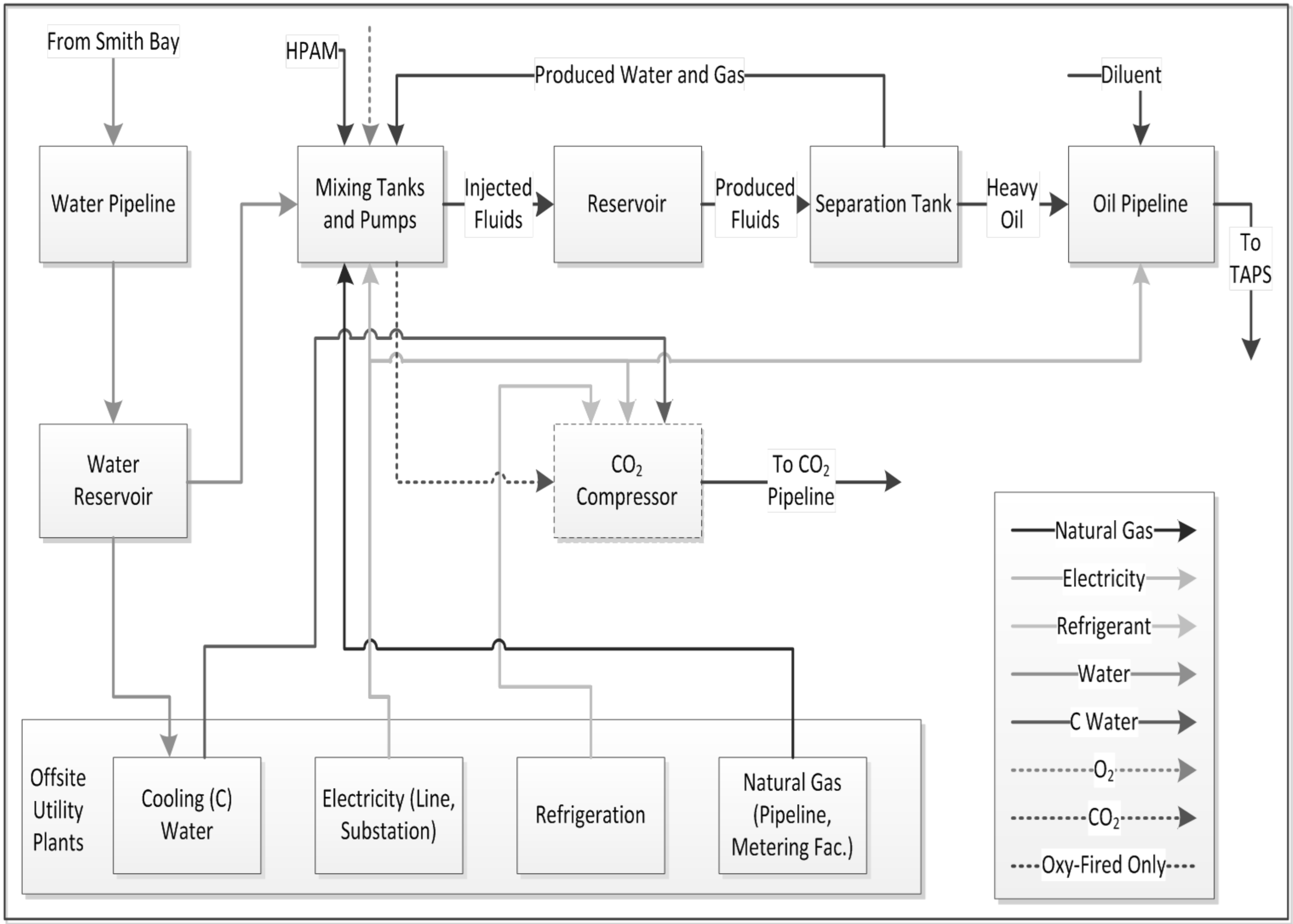

Figure 3-1: Process flow diagram for base case and dilution scenarios. Dashed lines are for oxy-firing only. Note that diluent is only applicable for dilution scenario.

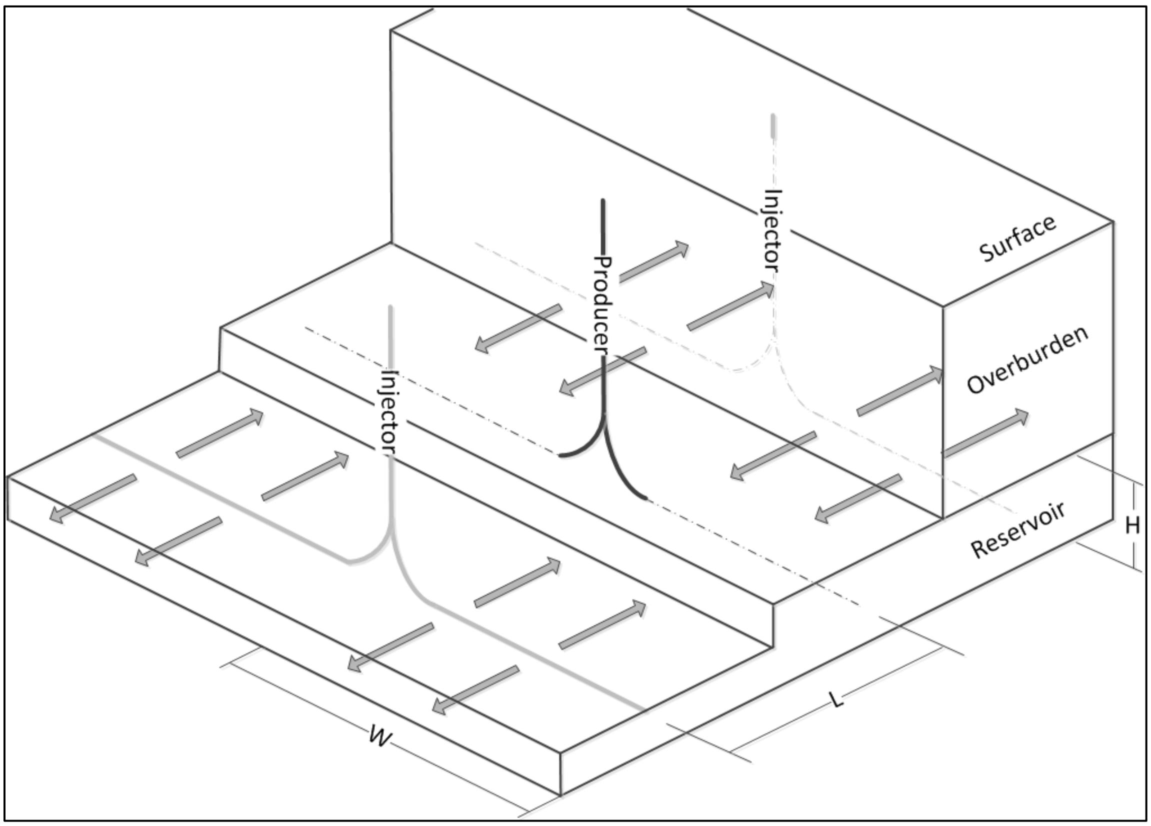

Figure 3-2: Horizontal line-drive polymer flood diagram. Viscous water is pumped through injectors into the reservoir, displacing heavy oil which is produced through producer well. Wells are spaced L lengths apart, each lateral segment is W long, and the reservoir is H thick.

Table 3-2: Production design, specifications, and assumptions.

Item	**Value**	**Units**	**Notes**
Reservoir			
Location			
Latitude	70°31'26.029"	N	
Longitude	154°59'39.92"	W	
Depth	4,000	ft	
H	50	ft	Thickness (single homogeneous layer)
κ	150	md	Permeability
Porosity (ϕ)	0.30		Porosity
B_o	1.069	rb/stb	Oil formation volume factor
S_{WC}	0.12		Connate water saturation (unitless number?)
μ	35.4	cP	Oil and water viscosity
Well Design			
L	2,000	ft	Well spacing
Well length			
Vertical	2,292	ft	
Transition	2,723	ft	Assuming 1° buildup rate per 30 ft of pipe and 90° turn
W	3,369	ft	Length of each lateral segment
Injector Wells	10		
Producer Wells	11		
Production			
Q			Injection rate
Base Case	6,185	bpd/well	
Dilution	4,418	bpd/well	
Upgrading	5,601	bpd/well	
M	0.982	OOIP/PV	Slope of linear recovery line

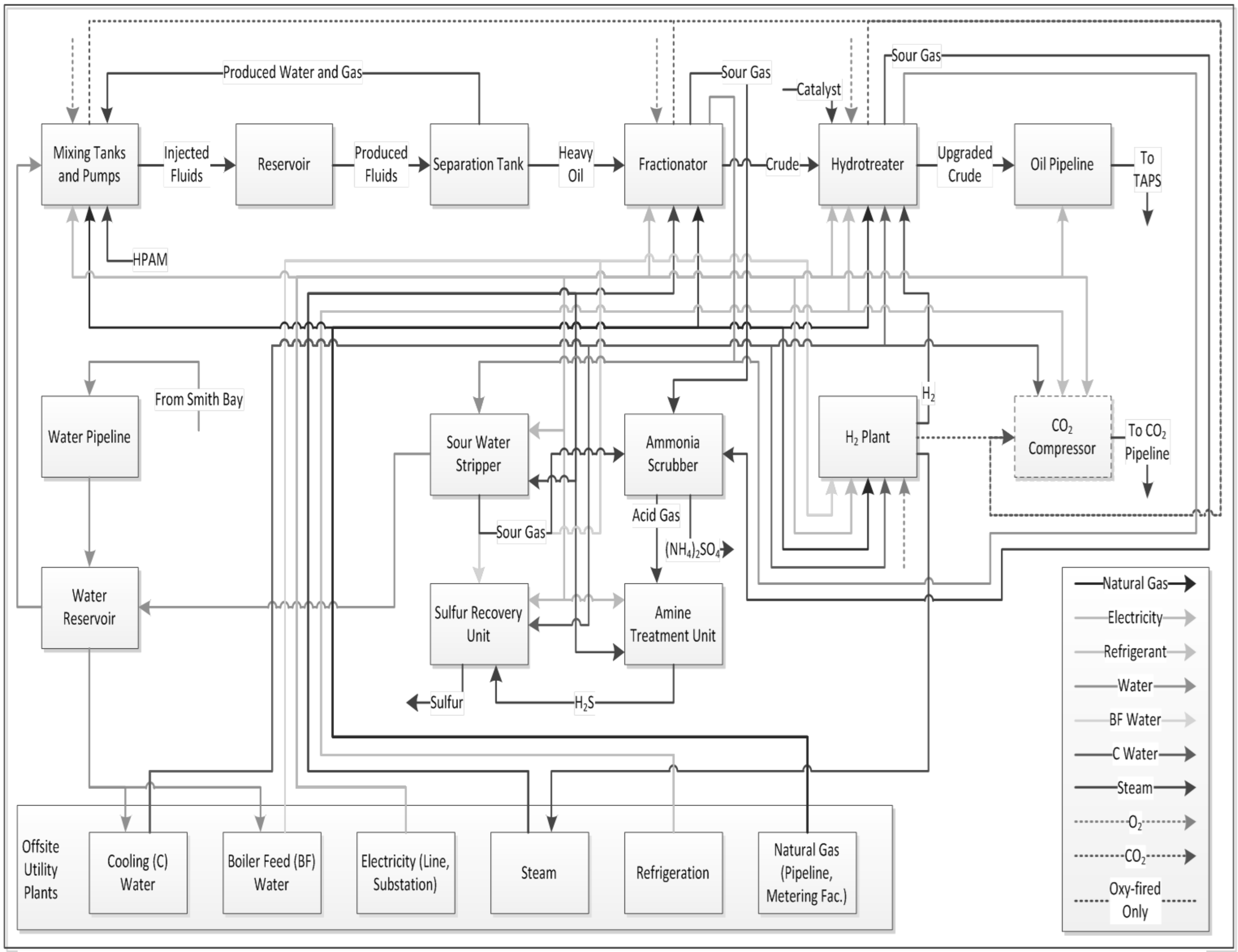

Figure 3-3: Upgrading process flow diagram. Dashed lines indicate steps that are only applicable to oxy-firing.

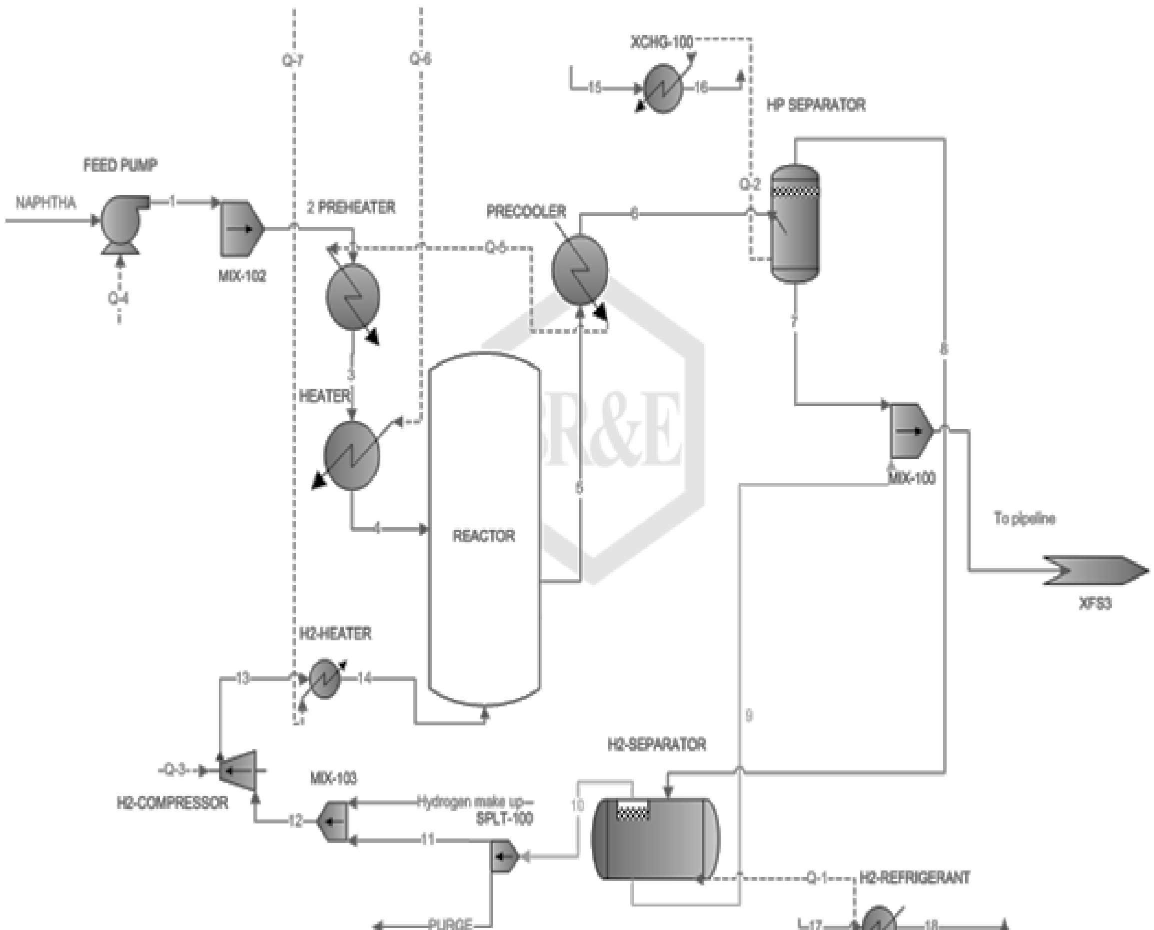

Figure 3-4: Hydrotreater process flow diagram developed in ProMax by Castro (2010).

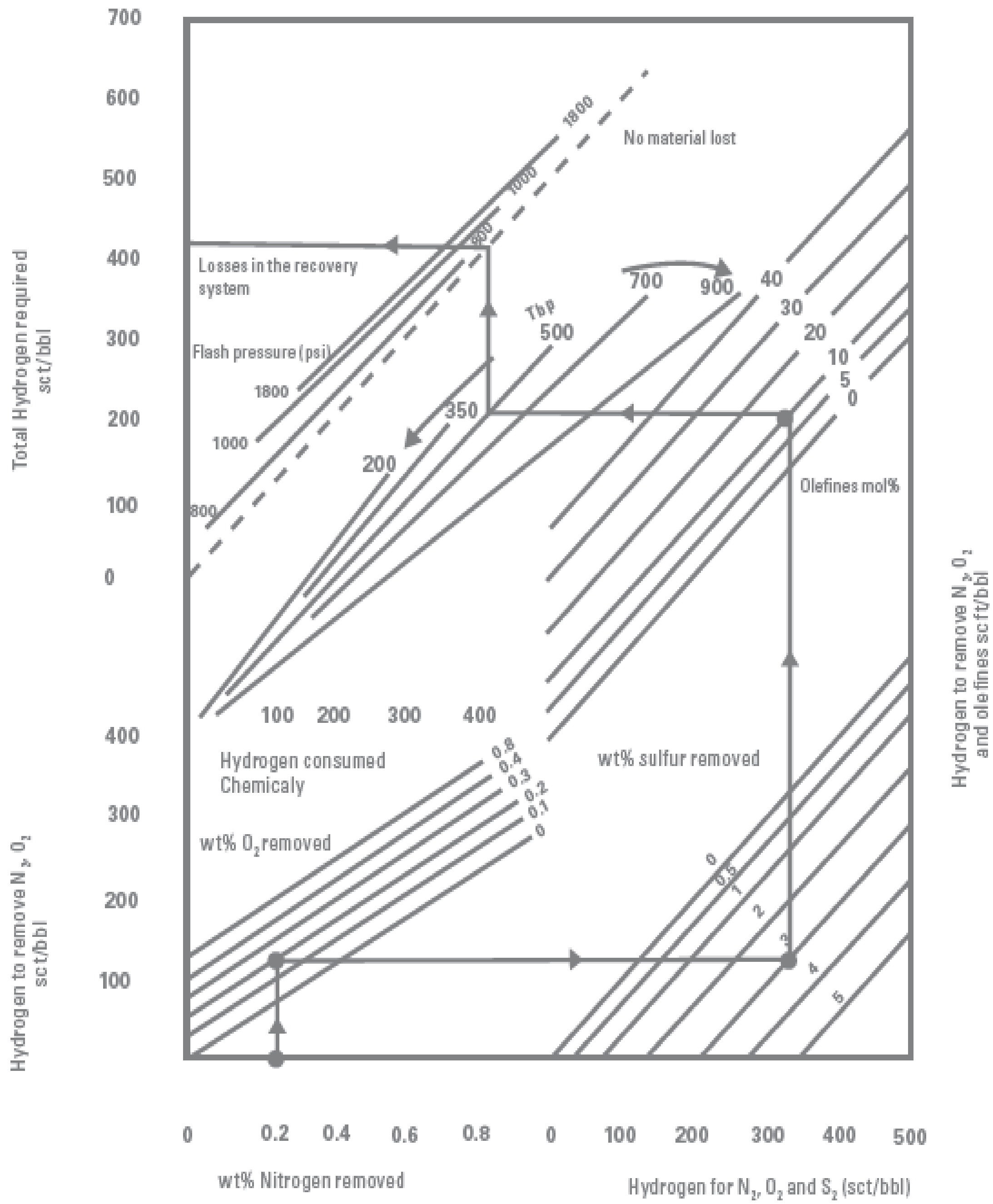

Figure 3-5: Hydrogen consumption for hydrotreating crude oils with various properties (Owusu 2005).

Table 3-3: Annual production of gaseous byproducts.

Gaseous Byproduct	Production (ton/yr)
Hydrogen Sulfide (H_2S)	20,000
Ammonia (NH_3)	49,096

Table 3-4: Properties of raw and upgraded oil in comparison to three benchmark crudes (Wang, *et al*. 2003) (Enivronment Canada 2011).

		Heavy Oil	Upgraded Oil	WTI	Brent Crude	Arabian Light Crude
Oil Properties	API Gravity	18.5	35.0	39.6	38.0	34.0
	Sulfur (wt%)	0.70	0.01	0.24	0.37	1.70
	Nitrogen (wt%)	1.9	0.10		0.10	0.07
	Pour Point (°F)		0	-18	45	-10
Distillate Cuts	Boiling Range (°F)	(vol %)				
Naptha	100 - 400	0	73	56		
	104 - 800				78	67
Vacuum Gas Oil	400 - 950	65	26	32		
	800 +				21.7	32
Wax	950 +	35	2	9		
	1000 +				10.2	17

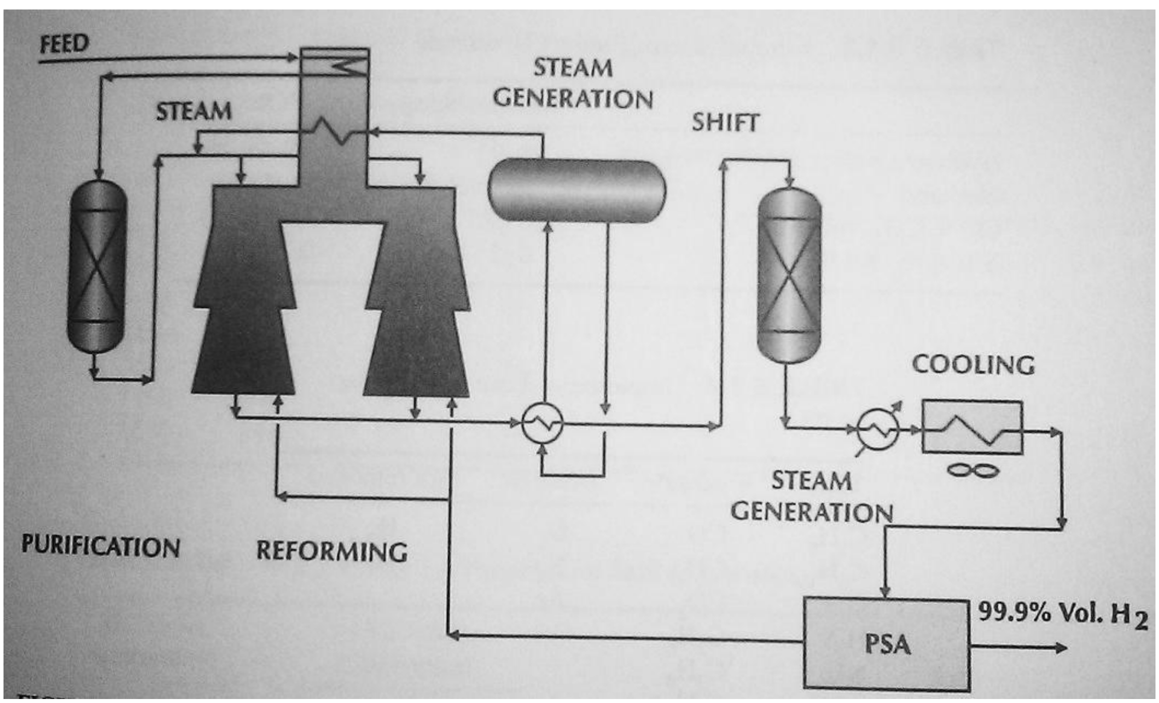

Figure 3-6: Hydrogen production system from Fleshman (2004).

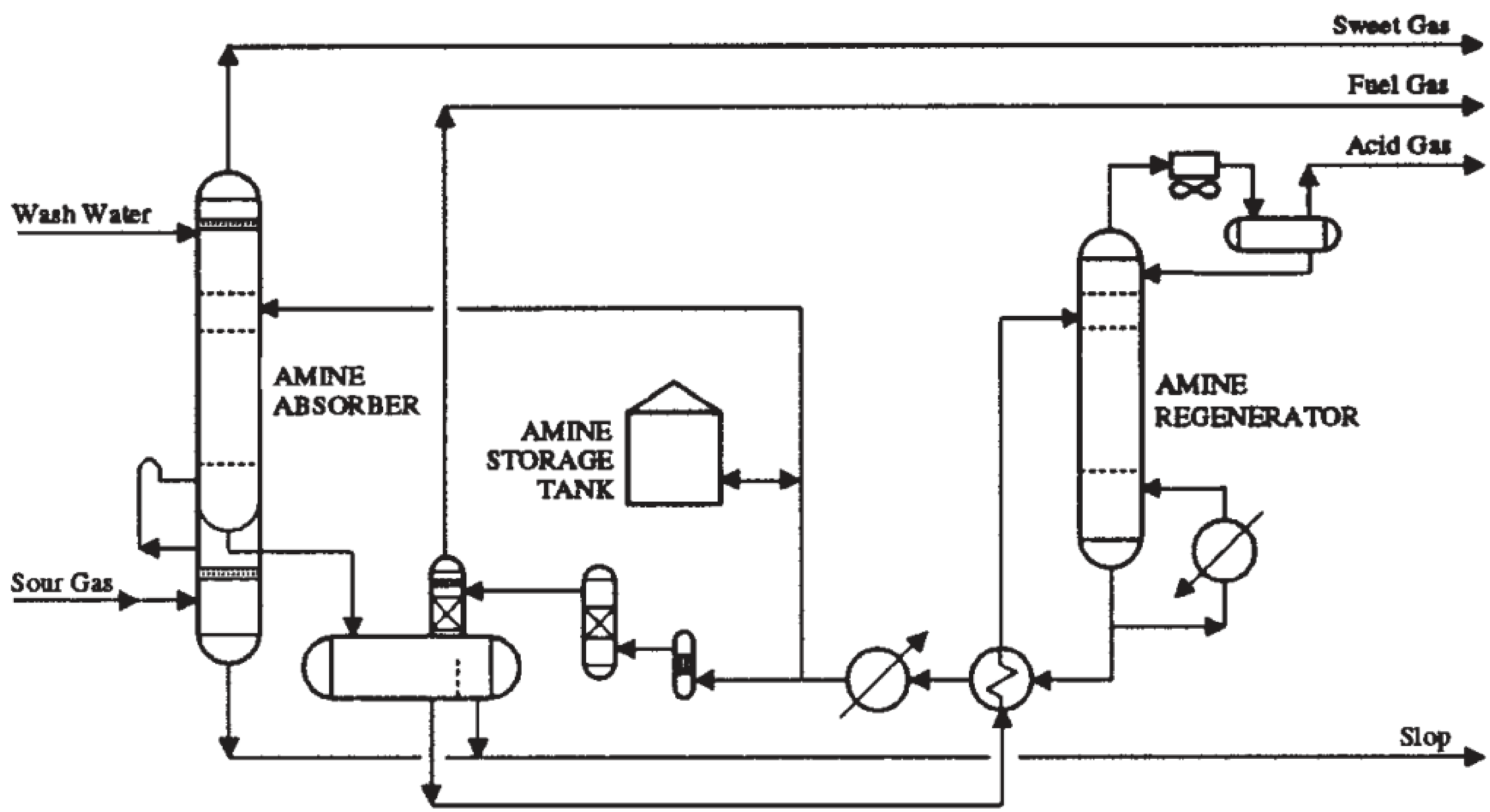

Figure 3-7: Amine treatment unit process flow diagram (Maples 2000).

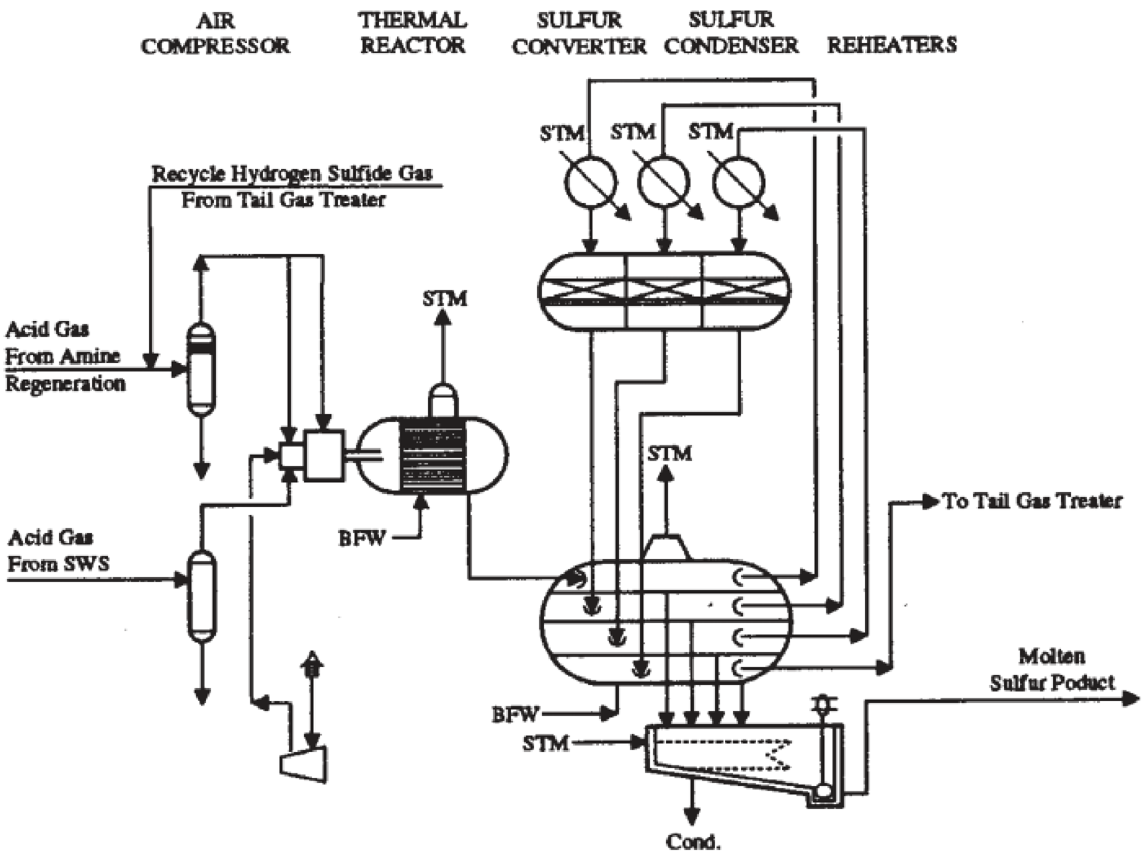

Figure 3-8: Sulfur recovery unit utilizing the Claus process (Maples 2000). Note that the abbreviation “BFW” is boiler feed water and “STM” is steam.

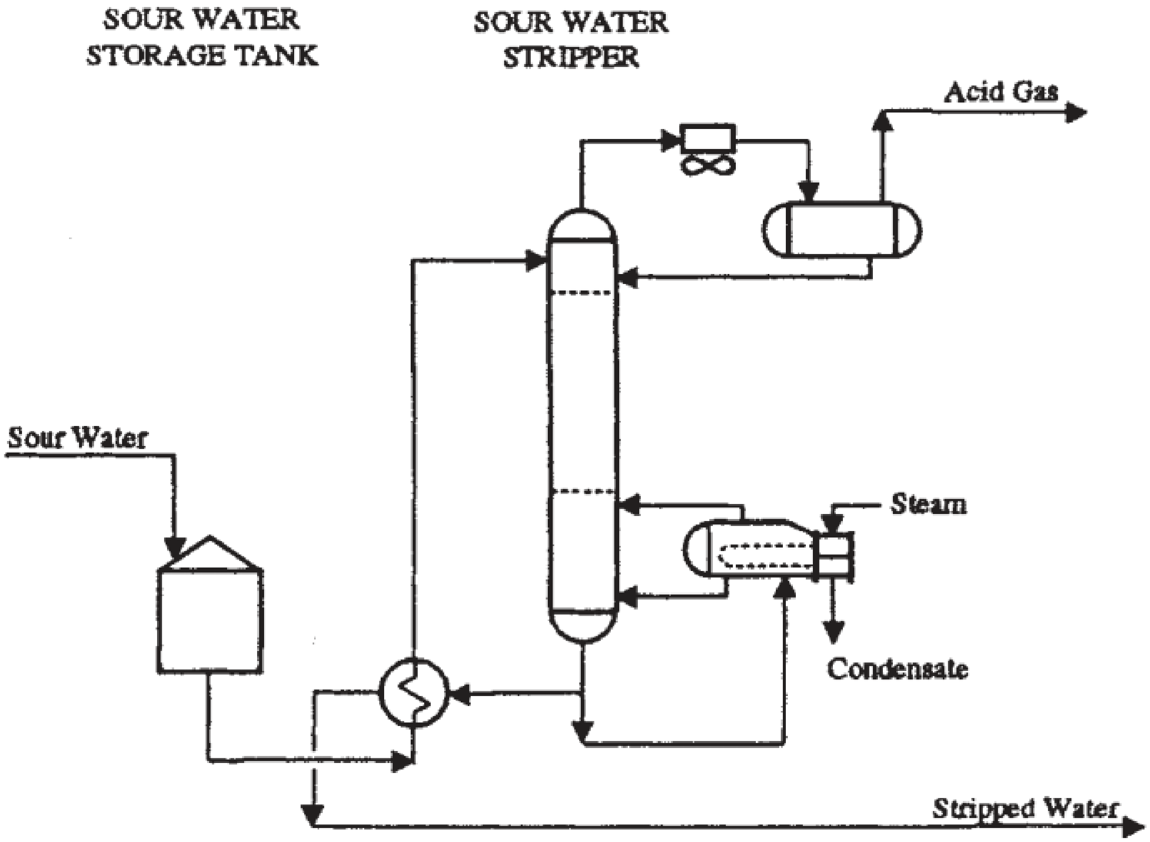

Figure 3-9: Sour water stripper process diagram (Maples 2000).

Table 3-5: Labor requirements for heavy oil production (per shift).

Process	Operators	Lab & Engineering	Management
Polymer Flood	4	2	1
Upgrading Only			
Fractionator	2	2	1
Hydrotreater	18	2	1
Hydrogen Plant	6	2	1
Sour Water Stripper	4	2	1
Amine Treatment Unit	4	2	1
Sulfur Recovery Unit	6	2	1
Oxy-Fired Only			
CO2 Compression	4	2	1

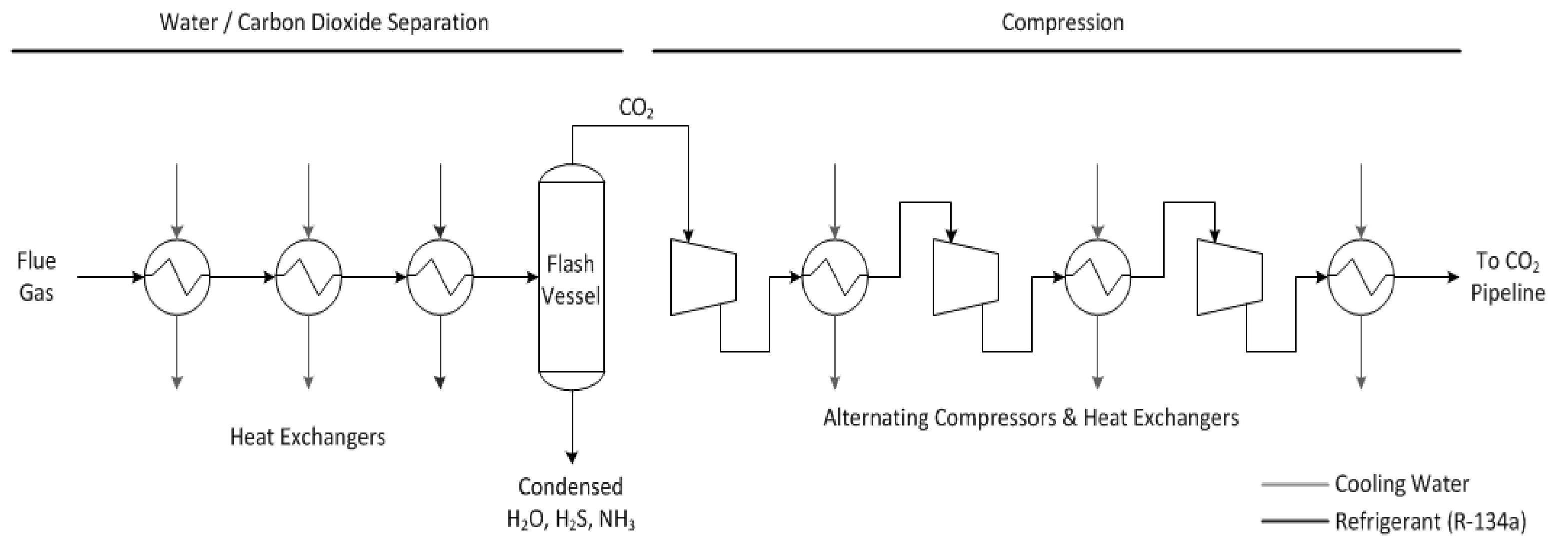

Figure 3-10: Process flow diagram for CO_2 compression system.

Table 3-6: CO_2 production by scenario.

Scenario	CO2 Production (10^3 ton/yr)	
	Air-Fired	Oxy-Fired
Base Case	38	36
Dilution	27	26
Upgrading	1,564	1,540

Table 3-7: Water usage and reservoir size.

Scenario	Water Usage (10^6 gal/yr)		Reservoir Size (acre-ft)	
	Air-Fired	Oxy-Fired	Air-Fired	Oxy-Fired
Base Case	948	956	717	723
Dilution	677	683	512	517
Upgrading	1,167	1,484	883	1,123

RESULTS

Results of the economic evaluation of each scenario are given below. Capital costs are shown first, followed by an itemized list of all costs on a dollar per barrel basis (defined as supply costs), then annual cash flows and finally, a sensitivity analysis. As discussed in Section 2, all calculations are based on discounted cash flows. However, for the sake of clarity, all results are shown without applying the discount factor (i.e. total 2010 U.S. dollars). Furthermore, all results (unless stated otherwise) are for the specified IRR solution method (IRR = 15%).

4.1 Base Case

4.1.1 Capital Costs

The total capital investment (C_{TCI}) and capital per flowing barrel (CPFB) for the base case scenario for air and oxy-firing are shown in Table 4-1. CPFB is a common oil industry metric for capital costs, and is defined as the C_{TCI} divided by the oil production rate (bpd). For air firing, the largest capital costs are for drilling (25%), the feeder oil pipeline (11%), and utility plants (10%); results for oxy-firing are similar. While the initial expense for working capital is larger than either the feeder pipeline or the utility plants, the entire amount is taken as a credit at the end of the project, resulting in a present value cost of only $77 million dollars. A detailed capital cost breakdown is given in Table 4.2.

4.1.2 Supply Costs

The total supply cost for the base case scenarios is \$29.44/bbl and \$30.41/bbl for air and oxy-firing, respectively. The largest supply costs (neglecting net earnings, i.e. the investor's return) are taxes (\$8.61/bbl, 29%), TAPS and tanker transportation (\$6.15/bbl, 21%), and royalties (\$4.11, 14%). Itemized supply costs for air-firing are shown in Table 4-3. Note that the costs for working capital are zero in Table 4-3 because without applying the discount factor for the time value of money, the same amount of capital is returned as a credit at the end of the project as was invested at the beginning of the project. Simplified supply costs are shown in Figure 4-1 and a comparison of base case scenario supply costs under various conditions, including oxy-firing and the low, reference, and high EIA energy forecasts, is given in Figure 4-2. Air and oxy cases in Figure 4-2 refer to the combustion method and use the specified IRR solution method (IRR = 15%). EIA cases use the specified price forecast solution method.

4.1.3 Cash Flow

The annual cash flow for the base case scenario is given in Figure 4-3 for air-firing; oxy-firing cash flows are approximately the same.

4.1.4 Sensitivity Analysis

The sensitivities of (1) the specified IRR oil price for the fixed price case and (2) the ROI and IRR computed from a specified forecast to a variety of parameters are shown in Table 4-4. ROI, return on investment, is added as an additional profitability metric commonly used in evaluating profitability, and is defined as:

$$ROI = \frac{(net\ earnings)}{(total\ capital\ investment)} = \frac{(1 - t_F - t_S)(S - C_V - C_F)}{C_{TCI}} \qquad 4\text{-}1$$

Only air-firing is considered for this sensitivity analysis. The first value listed in the range column is the default value, followed by the range of values for that parameter and their resultant fixed oil price (from the specified IRR method), ROI, and IRR for that parameter value (both from the specified forecast method). Finally, the combined results of all favorable and unfavorable parameter changes are shown. Values for the base case are shown at the bottom of the table for comparison.

The parameters investigated for Table 4-4 were selected due to their impact on the economic results and to the range of uncertainty associated with their assumed values. The OOIP recovery factor *m* has a key impact on production (see Eq. 3-3) and an unproven performance (no references to field tests of polymer flooding in West Sak were found in the literature). No high value of *m* was selected since $m \cong 1$ already and producing more than a 1:1 ratio of PV injected to PV produced with an incompressible fluid is not possible. The net impact of lower values of *m* is that more wells have to be drilled to reach the same level of production. The fee paid for delivery is one of the major costs in the base case scenario (second only to production) and TAPS fees have varied over time. Finally, it has been historically demonstrated that through creative (but entirely legal) accounting practices, the tax liabilities of major corporations can be greatly reduced or avoided entirely. Royalties and severance taxes, however, are harder to avoid and thus they are not changed here.

4.2 Dilution

Results for the dilution scenario are shown below following the same format as presented previously with the base case scenario. To avoid repetition in the text, tables and figures in this and subsequent sections are introduced without reference unless new information needs to be conveyed.

4.2.1 Capital Costs

With air-firing, the largest capital costs for dilution are drilling (26%), the feeder oil pipeline (12%), and utility plants (11%); results for oxy-firing are similar. The capital costs are nearly identical to the base case scenario, except that since less oil volume is needed to produce 50,000 bpd, smaller injection and water handling equipment is called for, resulting in an overall reduction in C_{TCI}.

4.2.2 Supply Costs

The total supply costs for dilution are \$48.41/bbl and \$49.23/bbl for air and oxy-firing, respectively. The largest supply costs (neglecting net earnings) are diluent (\$20.00/bbl, 41%), taxes (\$6.43/bbl, 13%), and TAPS and tanker transportation (\$6.15/bbl, 13%).

4.2.3 Cash Flow

The annual cash flow for dilution is given in Figure 4-6 for air-firing; oxy-firing cash flows are approximately the same.

4.2.4 Sensitivity Analysis

The diluent to oil ratio and the diluent price are introduced as new parameters here to investigate the largest supply cost item for this scenario. In addition to the high/low variations, a fixed fraction variant is included for the value of diluent. The fixed fraction assumption is that the price differential between diluent and WTI in 2010 (88%) stays constant as oil prices vary (either forecasted or fixed).

4.3 Upgrading

4.3.1 Capital Costs

With air-firing, the largest capital costs are for the hydrotreater (29%), contingency (8%), and service facilities (6%); results for oxy-firing are similar. Both the contingency and service facilities are high as a result of the size of the total bare module investment (C_{TBM}), since both are calculated as percentages of C_{TBM}.

4.3.2 Supply Costs

The total supply costs for upgrading are \$195.94/bbl and \$214.34/bbl for air and oxy-firing, respectively. The largest supply costs (neglecting net earnings) are taxes (\$95.88/bbl, 49%), royalties (\$26.17/bbl, 13%), and C_{TCI} (\$10.72/bbl, 5%, excluding drilling). Note that for the EIA low forecast, no non-negative interest rate exists that gives an NPV = 0. Therefore, results are shown for the low forecast in Figure 4-8 with $i = 0\%$.

4.3.3 Cash Flow

The annual cash flow for upgrading is given in Figure 4-9 for air-firing; oxy-firing cash flows are approximately the same.

4.3.4 Sensitivity Analysis

The fuel costs, combined capital and operating costs of the hydrotreater, and the method of assessing the wellhead value of produced heavy oil are added as parameters in the sensitivity analysis for upgrading. Fuel costs are added since they represent the largest utility cost and natural gas prices on the North Slope could vary widely from the national average industrial price used in the base analysis. The hydrotreater's capital and operating costs are varied since it is the largest cost for this scenario. The wellhead value of heavy oil plays a major role in determining severance tax liability. An alternative method of assessing value based on the fraction of the API gravity of West Sak heavy oil (19.2°) over that of WTI (39.6°) is used instead of the ANS price differential discussed in Section 2.3.

Table 4-1: Base case C_{TCI} and CPFB.

Firing Method	C_{TCI} (10^6 dollars)	CPFB ($/bpd)
Air	$839.00	$16,781
Oxy	$856.50	$17,130

Table 4-2: Base case capital cost summary (millions of dollars).

Category	Item	Base case			
		Air-fired		Oxy-fired	
Total Bare Module Investment (C_TBM)	Polymer Flood	$	3.5	$	3.5
	Fractionator	$	-	$	-
	Hydrotreater	$	-	$	-
	Hydrogen Plant	$	-	$	-
	Sour Water Stripper	$	-	$	-
	Amine Treatment Unit	$	-	$	-
	Sulfur Recovery Unit	$	.0	$	.0
	CO2 Compressor	$	-	$	6.8
	C_TBM Subtotal	**$**	**3.5**	**$**	**10.3**
Total Direct Permanent Investment (C_DPI)	Site Preparation	$	.4	$	1.0
	Service Facilities	$	.7	$	2.1
	Oil Pipeline	$	93.6	$	93.6
	Water Pipeline	$	17.9	$	17.9
	Water Reservoir	$	8.6	$	8.6
	Drilling	$	210.8	$	210.8
	Allocated Costs for Utility Plants	$	86.9	$	87.1
	C_DPI Subtotal	**$**	**422.3**	**$**	**431.5**
Total Depreciable Capital (C_TDC)	Contingency	$	63.3	$	64.7
	C_TDC Subtotal	**$**	**485.6**	**$**	**496.2**
Total Permanent Investment (C_TPI)	Land	$	9.7	$	9.9
	Permitting	$	30.1	$	30.1
	Royalties for Intellectual Property	$	9.7	$	9.9
	Startup	$	48.6	$	49.6
	Investment Site Factor		1.25		1.25
	C_TPI Subtotal - US Midwest	**$**	**729.6**	**$**	**744.8**
Total Capital Investment (C_TCI)	Working Capital	$	109.4	$	111.7
	Total ($)	**$**	**839.0**	**$**	**856.5**

Table 4-3: Base case itemized supply costs.

Category	Item	Capital	Labor	Electricity	Fuel	Water	Steam	Other*	Total
Extraction	Drilling	$ 0.88	$ -	$ -	$ -	$ -	$ -	$ -	$ 0.88
	Polymer Flooding	$ 0.01	$ 0.17	$ 0.03	$ 0.21	$ -	$ -	$ 0.11	$ 0.54
Delivery	Oil Pipeline	$ 0.39	$ -	$ 0.12	$ -	$ -	$ -	$ -	$ 0.51
Other	Water Pipeline	$ 0.07	$ -	$ 0.03	$ -	$ -	$ -	$ -	$ 0.10
Notes	* Other includes:	Polymer						Allocated Costs for Utility Plants	$ 0.36
								Water Reservoir	$ 0.04
	** Taxes includes:	State Tax						Site Preparation	$ 0.00
		Federal Tax						Service Facilities	$ 0.00
		Severance Tax						Contingency	$ 0.26
		Property Tax						Permitting	$ 0.13
								Research	$ 0.74
								Administration	$ 0.01
								Incentive Compensation	$ 0.11
								Insurance	$ 0.18
								TAPS & Tanker	$ 6.15
								Taxes**	$ 8.61
								Royalties - oil	$ 3.68
								Royalties - IP	$ 0.43
								Working Capital	$ -
								Land	$ 0.04
								Startup	$ 0.20
								Net Earnings	$ 6.47
Total									**$ 29.44**

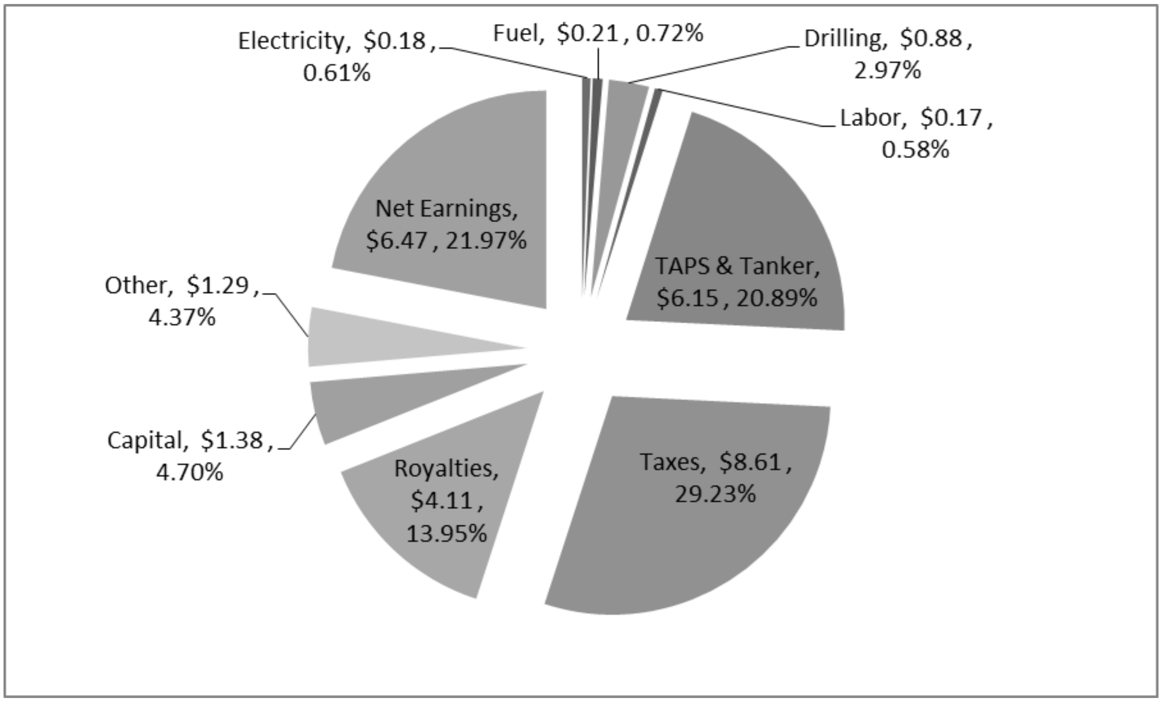

Figure 4-1: Base case simplified supply costs summary.

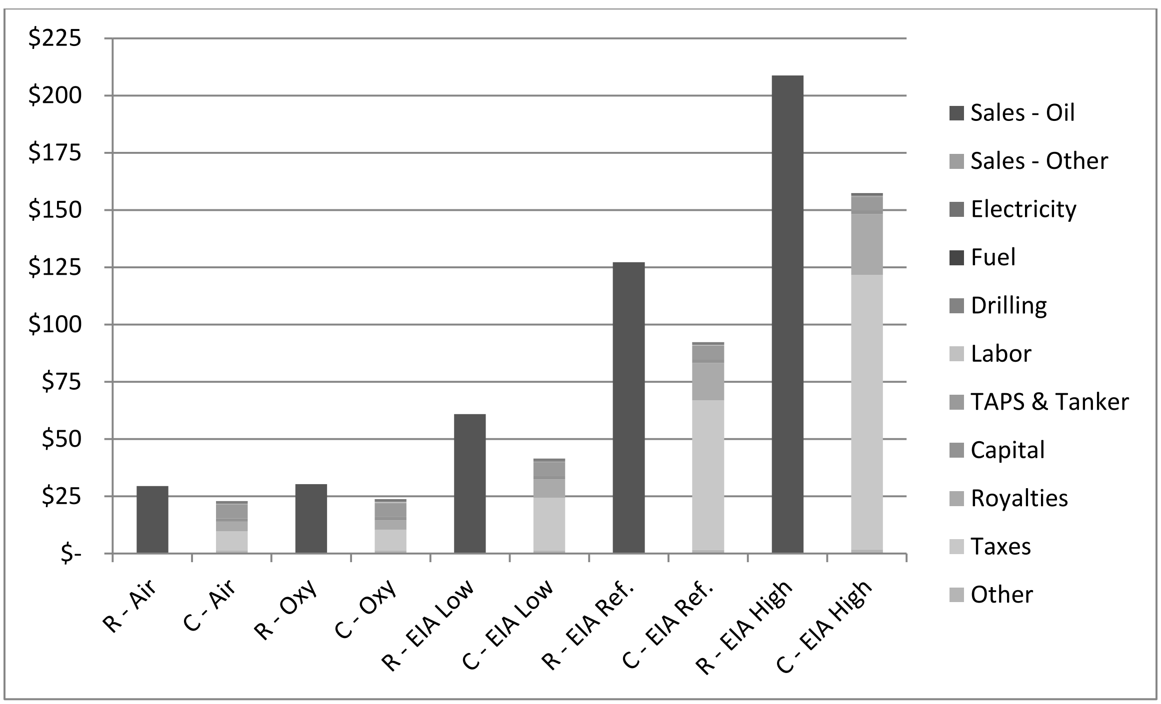

Figure 4-2: Base case revenue (R) and supply costs (C) variations. Air and oxy cases have specified IRR = 15%. EIA cases use price forecasts (resultant IRR values: low = 28%, reference = 41%, high = 51%).

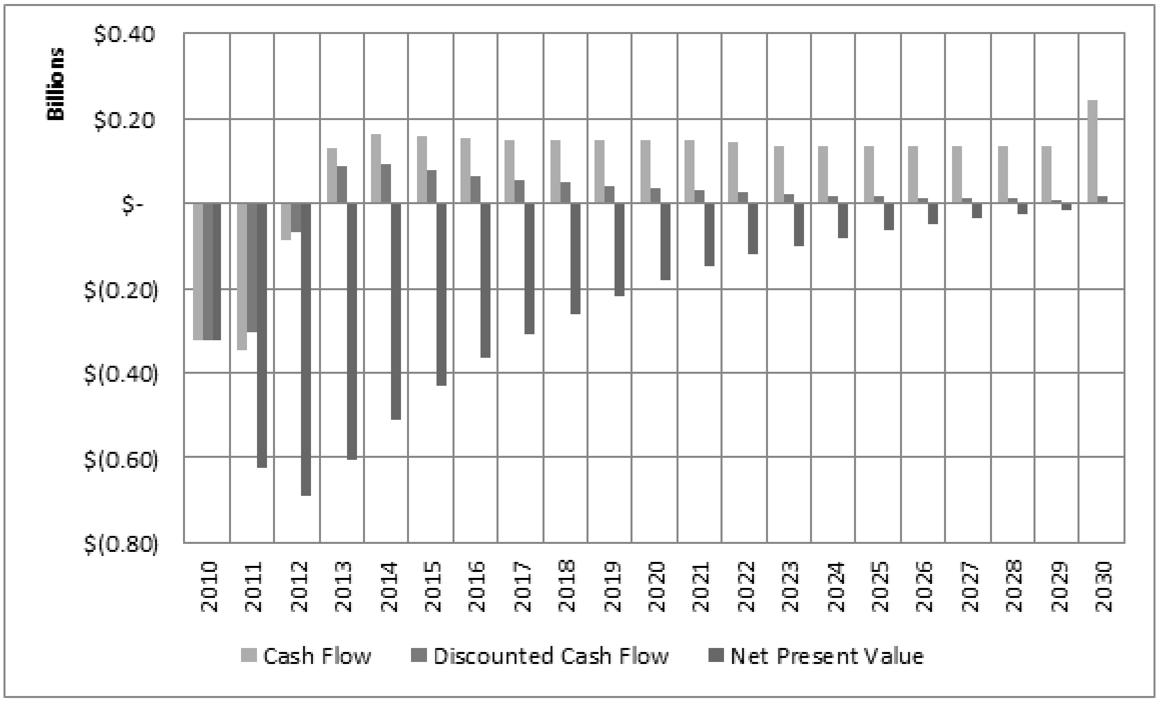

Figure 4-3: Base case annual cash flow.

Table 4-4: Base case sensitivity analysis.

Base Case		**Air-Fired**		
Oil Price Model		**Fixed Price (15% IRR)**	**EIA Reference Forecast**	
Variable	**Range**	**Oil ($/bbl)**	**ROI**	**IRR**
OOIP Recovery - m	**0.982**			
Low (50%)	0.491	$ 39.18	93.94%	30.44%
TAPS & Tanker ($/bbl)	**$6.15**			
Low (50%)	$ 3.08	$ 25.66	147.82%	42.53%
High (150%)	$ 9.23	$ 33.32	140.42%	39.81%
Combined Corporate Tax Rate	**44.4%**			
Zero	0.0%	$ 26.71	259.22%	49.02%
Low (50%)	22.2%	$ 27.92	201.67%	45.05%
Combined	---			
Con (↓m, ↑TAPS, full tax)	---	$ 42.82	91.46%	29.54%
Pro (base m, ↓TAPS, zero tax)	---	$ 22.46	265.87%	51.05%
Base	**---**	**$ 29.44**	**144.13%**	**41.18%**

Table 4-5: Dilution C_{TCI} and CPFB.

Firing Method	C_{TCI} (10^6 dollars)	CPFB ($/bpd)
Air	$813.9	$16,279
Oxy	$830.8	$16,615

Table 4-6: Dilution capital cost summary (millions of dollars).

Category	Item	Dilution			
		Air-fired		Oxy-fired	
Total Bare Module Investment (C_TBM)	Polymer Flood	$	2.1	$	2.1
	Fractionator	$	-	$	-
	Hydrotreater	$	-	$	-
	Hydrogen Plant	$	-	$	-
	Sour Water Stripper	$	-	$	-
	Amine Treatment Unit	$	-	$	-
	Sulfur Recovery Unit	$	-	$	-
	CO2 Compressor	$	-	$	6.7
	C_TBM Subtotal	**$**	**2.1**	**$**	**8.8**
Total Direct Permanent Investment (C_DPI)	Site Preparation	$	.2	$	.9
	Service Facilities	$	.4	$	1.8
	Oil Pipeline	$	93.6	$	93.6
	Water Pipeline	$	15.3	$	15.4
	Water Reservoir	$	6.4	$	6.4
	Drilling	$	210.8	$	210.8
	Allocated Costs for Utility Plants	$	86.7	$	86.9
	C_DPI Subtotal	**$**	**415.5**	**$**	**424.4**
Total Depreciable Capital (C_TDC)	Contingency	$	62.3	$	63.7
	C_TDC Subtotal	**$**	**477.8**	**$**	**488.1**
Total Permanent Investment (C_TPI)	Land	$	9.6	$	9.8
	Permitting	$	21.5	$	21.5
	Royalties for Intellectual Property	$	9.6	$	9.8
	Startup	$	47.8	$	48.8
	Investment Site Factor		1.25		1.25
	C_TPI Subtotal - US Midwest	**$**	**707.8**	**$**	**722.4**
Total Capital Investment (C_TCI)	Working Capital	$	106.2	$	108.4
	Total ($)	**$**	**813.9**	**$**	**830.8**

Table 4-7: Dilution itemized supply costs.

Category	Item	Capital	Labor	Electricity	Fuel	Water	Steam	Other*	Total
Extraction	Drilling	$ 0.88	$ -	$ -	$ -	$ -	$ -	$ -	$ 0.88
	Polymer Flooding	$ 0.01	$ 0.16	$ 0.02	$ 0.15	$ -	$ -	$ 0.08	$ 0.42
Delivery	Oil Pipeline	$ 0.39	$ -	$ 0.12	$ -	$ -	$ -	$ -	$ 0.51
Other	Water Pipeline	$ 0.06	$ -	$ 0.02	$ -	$ -	$ -	$ -	$ 0.08
	* Other includes:	Polymer				Allocated Costs for Utility Plants			$ 0.36
						Water Reservoir			$ 0.03
	** Taxes includes:	State Tax				Site Preparation			$ 0.00
		Federal Tax				Service Facilities			$ 0.00
		Severance Tax				Contingency			$ 0.26
		Property Tax				Permitting			$ 0.09
						Research			$ 0.74
						Administration			$ 0.01
						Incentive Compensation			$ 0.11
						Insurance			$ 0.18
						Diluent			$ 20.00
						TAPS & Tanker			$ 6.15
						Taxes**			$ 6.43
						Royalties - oil			$ 4.32
						Royalties - IP			$ 1.17
						Working Capital			$ -
						Land			$ 0.04
						Startup			$ 0.20
						Net Earnings			$ 6.43
Total									**$ 48.41**

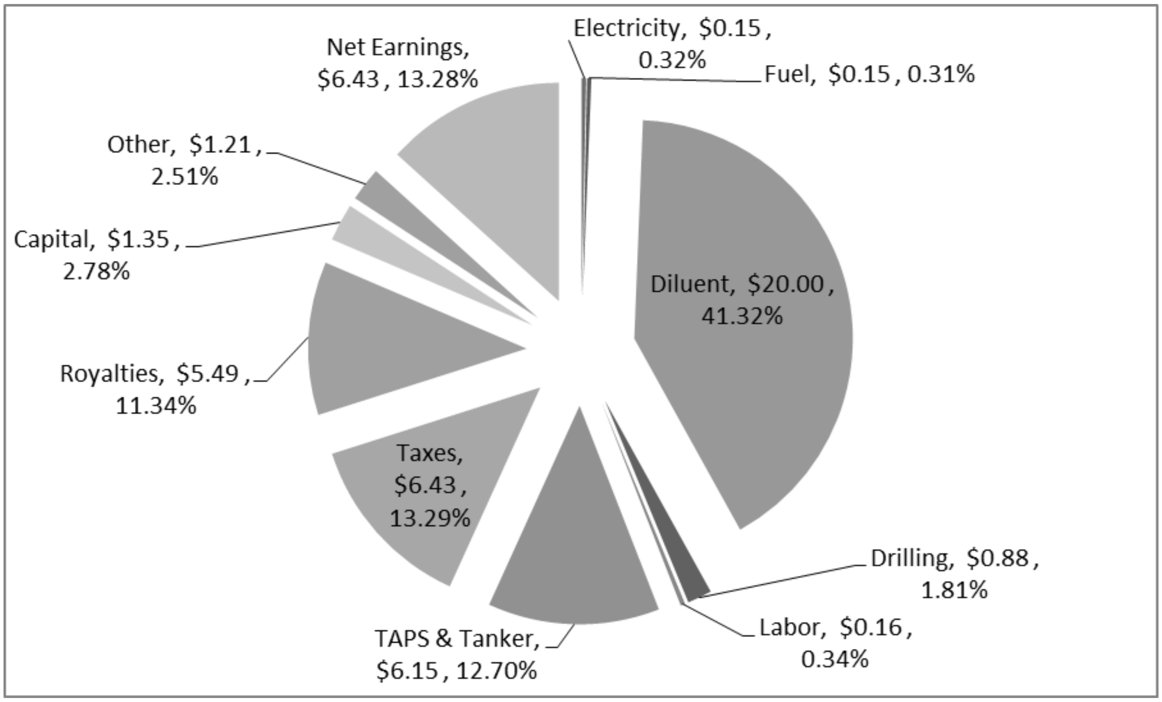

Figure 4-4: Dilution simplified supply costs summary.

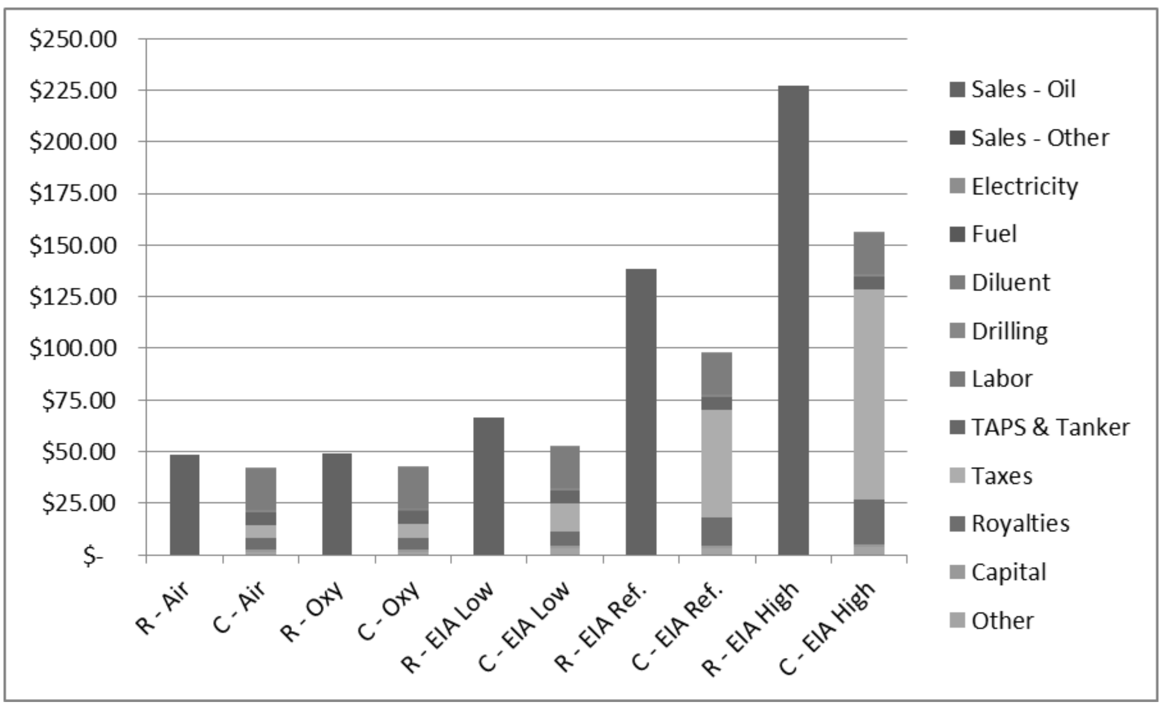

Figure 4-5: Dilution revenue (R) and supply costs (C) variations. Air and oxy cases have specified IRR = 15%. EIA cases use price forecasts (resultant IRR values: low = 23%, reference = 45%, high = 60%).

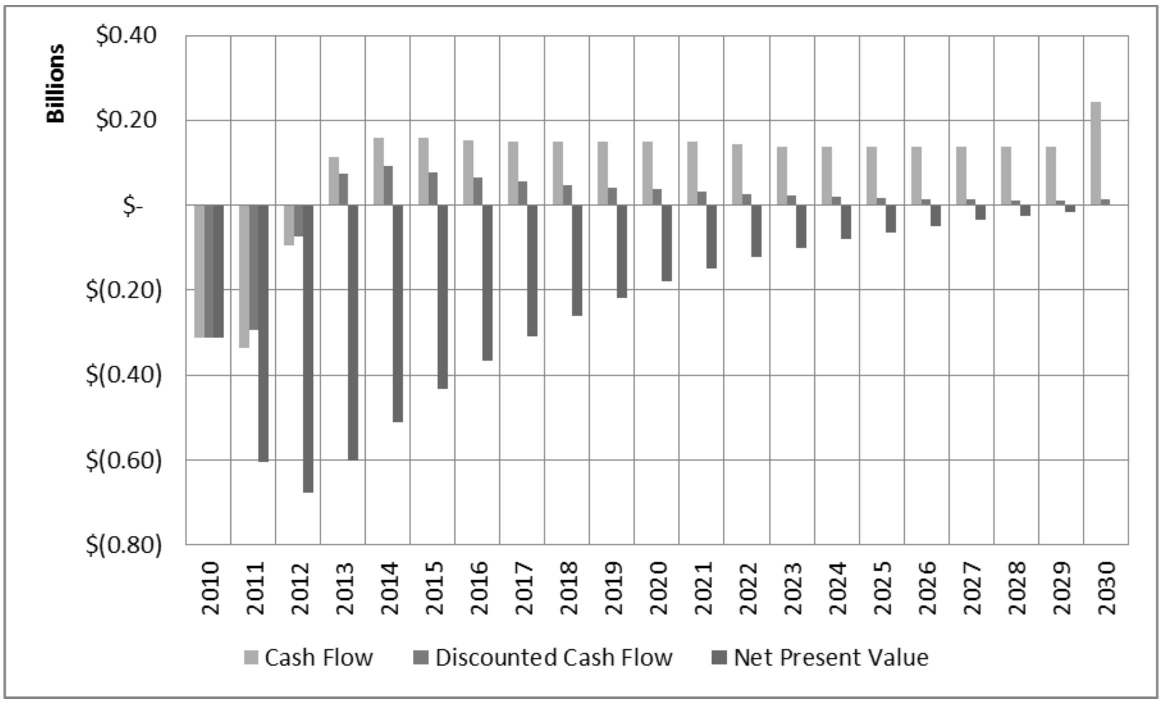

Figure 4-6: Dilution annual cash flow.

Table 4-8: Dilution sensitivity analysis.

Dilution		Air-Fired		
Oil Price Model		Fixed Price (15% IRR)	EIA Reference Forecast	
Variable	Range	Oil ($/bbl)	ROI	IRR
OOIP Recovery - m	**0.982**			
Low (50%)	0.491	$ 56.00	81.41%	34.34%
TAPS & Tanker ($/bbl)	**$6.15**			
Low (50%)	$ 3.08	$ 44.84	128.77%	46.59%
High (150%)	$ 9.23	$ 52.01	121.12%	43.59%
Combined Corporate Tax Rate	**44.4%**			
Zero	0.0%	$ 47.62	224.73%	53.82%
Low (50%)	22.2%	$ 47.98	174.84%	49.44%
Diluent Ratio (oil to diluent)	**2.5**			
Low	1	$ 62.14	109.72%	42.99%
High	4	$ 42.72	130.79%	45.71%
Diluent Price ($/bbl)	**$ 70**			
Low	$ 35	$ 36.87	137.38%	49.90%
High	$ 105	$ 60.24	112.52%	40.13%
Fixed fraction of WTI	88%	$ 35.71	108.96%	42.20%
Combined	---			
Con (↓m, ↑TAPS, full tax, ↓ratio, ↑price)	---	$ 91.81	54.11%	23.99%
Pro (base m, ↓TAPS, zero tax, ↑ratio, ↓price)	---	$ 28.67	257.54%	61.76%
Base	**---**	**$ 48.41**	**124.95%**	**45.11%**

Table 4-9: Upgrading C_{TCI} and CPFB.

Firing Method	C_{TCI} (10^6 dollars)	CPFB ($/bpd)
Air	$4,129.0	$82,580
Oxy	$4,255.8	$85,115

Table 4-10: Upgrading capital cost summary (millions of dollars).

Category	Item	Upgrading	
		Air-fired	Oxy-fired
Total Bare Module Investment (C_TBM)	Polymer Flood	$ 2.7	$ 2.7
	Fractionator	$ 41.1	$ 41.1
	Hydrotreater	$ 1,206.2	$ 1,212.5
	Hydrogen Plant	$ 51.9	$ 51.9
	Sour Water Stripper	$ 10.0	$ 10.0
	Amine Treatment Unit	$ 2.0	$ 2.0
	Sulfur Recovery Unit	$ 5.7	$ 5.7
	CO2 Compressor	$ -	$ 32.3
	C_TBM Subtotal	**$ 1,319.6**	**$ 1,358.2**
Total Direct Permanent Investment (C_DPI)	Site Preparation	$ 132.0	$ 135.8
	Service Facilities	$ 263.9	$ 271.6
	Oil Pipeline	$ 93.6	$ 93.6
	Water Pipeline	$ 20.0	$ 23.2
	Water Reservoir	$ 10.3	$ 12.8
	Drilling	$ 210.8	$ 210.8
	Allocated Costs for Utility Plants	$ 117.8	$ 129.3
	C_DPI Subtotal	**$ 2,168.0**	**$ 2,235.3**
Total Depreciable Capital (C_TDC)	Contingency	$ 325.2	$ 335.3
	C_TDC Subtotal	**$ 2,493.2**	**$ 2,570.5**
Total Permanent Investment (C_TPI)	Land	$ 49.9	$ 51.4
	Permitting	$ 30.1	$ 30.1
	Royalties for Intellectual Property	$ 49.9	$ 51.4
	Startup	$ 249.3	$ 257.1
	Investment Site Factor	1.25	1.25
	C_TPI Subtotal - US Midwest	**$ 3,590.4**	**$ 3,700.7**
Total Capital Investment (C_TCI)	Working Capital	$ 538.6	$ 555.1
	Total ($)	**$ 4,129.0**	**$ 4,255.8**

Table 4-11: Upgrading itemized supply costs.

Category	Item	Capital	Labor	Electricity	Fuel	Water	Steam	Other*	Total
Extraction	Drilling	$ 0.88	$ -	$ -	$ -	$ -	$ -	$ -	$ 0.88
	Polymer Flooding	$ 0.01	$ 0.17	$ 0.03	$ 0.19	$ -	$ -	$ 0.10	$ 0.50
Upgrading	Hydrotreater	$ 5.01	$ 7.44	$ 0.45	$ 1.72	$ 0.00	$ -	$ 0.13	$ 14.75
	H2 Plant	$ 0.22	$ 0.50	$ 0.02	$ 7.67	$ 0.03	$ -	$ -	$ 8.44
	Fractionator	$ 0.17	$ 0.33	$ 0.03	$ 0.59	$ -	$ 0.07	$ -	$ 1.18
	Sour Water Stripper	$ 0.04	$ 0.21	$ 0.01	$ -	$ -	$ 0.04	$ -	$ 0.29
	Amine Treatment Unit	$ 0.01	$ 0.16	$ 0.00	$ -	$ -	$ 0.65	$ -	$ 0.82
	Sulfur Recovery Unit	$ 0.02	$ 0.24	$ 0.00	$ -	$ 0.00	$ -	$ -	$ 0.27
Delivery	Oil Pipeline	$ 0.39	$ -	$ 0.12	$ -	$ -	$ -	$ -	$ 0.51
Other	Water Pipeline	$ 0.08	$ -	$ 0.03	$ -	$ -	$ -	$ -	$ 0.12
	* Other includes:	Catalyst						Allocated Costs for Utility Plants	$ 0.49
		R-134a						Water Reservoir	$ 0.04
		Polymer						Site Preparation	$ 0.55
								Service Facilities	$ 1.10
	** Taxes includes:	State Tax						Contingency	$ 1.35
		Federal Tax						Permitting	$ 0.13
		Severance Tax						Research	$ 0.74
		Property Tax						Administration	$ 0.10
								Incentive Compensation	$ 0.58
								Insurance	$ 0.91
								TAPS & Tanker	$ 6.15
								Taxes**	$ 95.88
								Royalties - oil	$ 24.40
								Royalties - IP	$ 1.77
								Working Capital	$ -
								Land	$ 0.21
								Startup	$ 1.04
								Net Earnings	$ 32.78
Total									**$ 195.94**

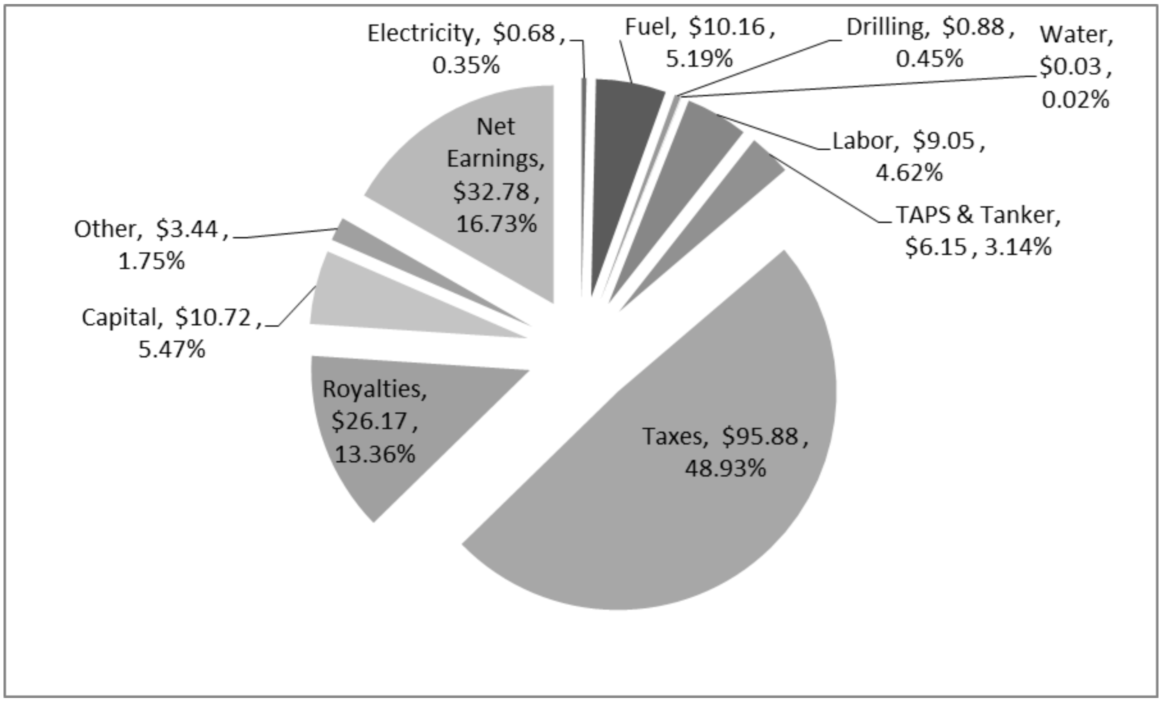

Figure 4-7: Upgrading simplified supply costs summary.

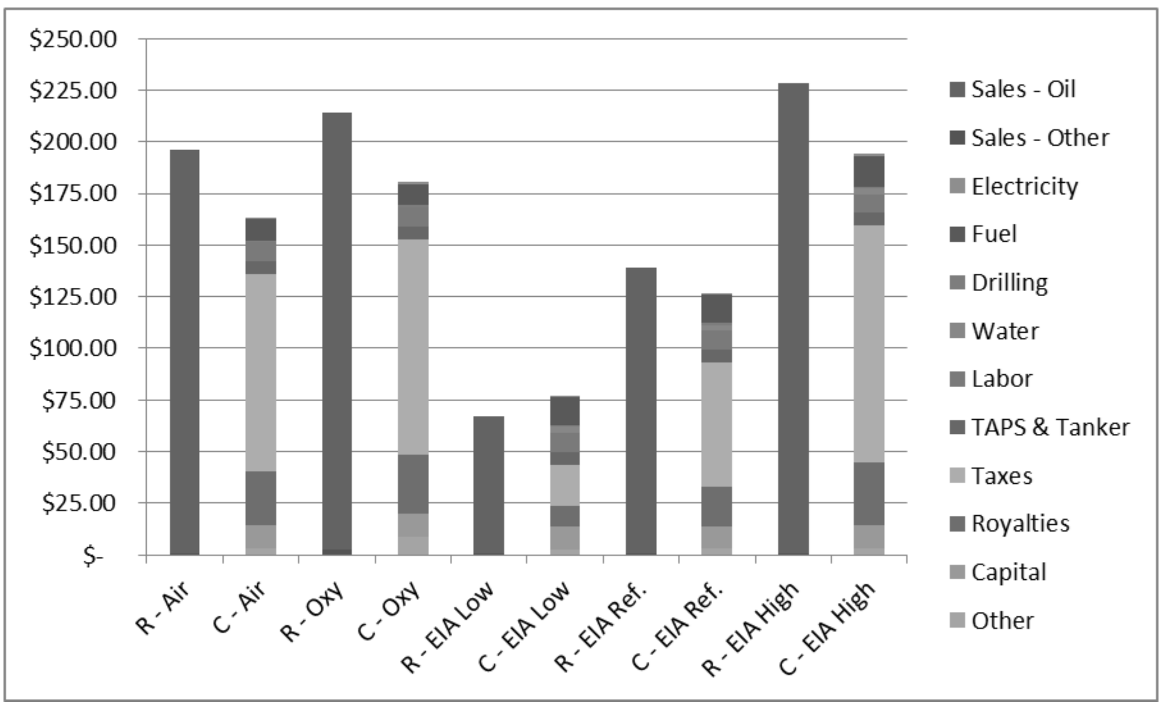

Figure 4-8: Upgrading revenue (R) and supply costs (C) variations. Air and oxy cases have specified IRR = 15%. EIA cases use price forecasts (resultant IRR values: low = 0%, reference = 6%, high = 13%). Note that for the EIA low forecast NPV $\neq$ 0.

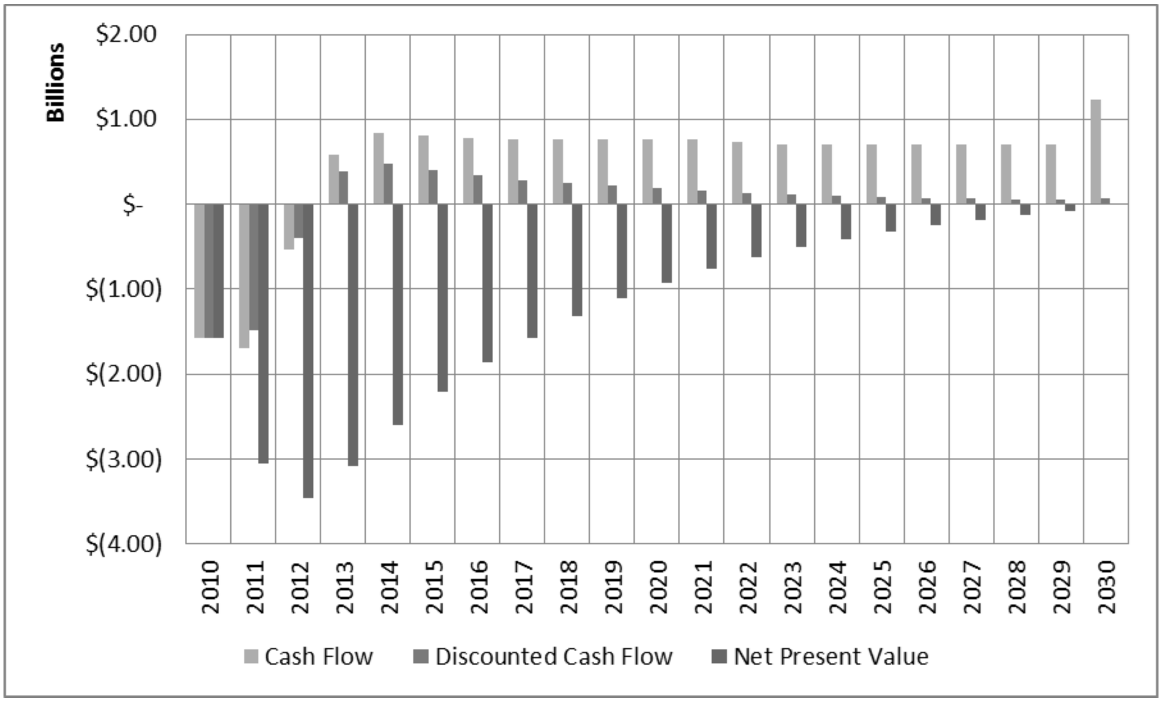

Figure 4-9: Upgrading annual cash flow.

Table 4-12: Upgrading sensitivity analysis.

Upgrading		**Air-Fired**		
Oil Price Model		**Fixed Price (15% IRR)**	**EIA Reference Forecast**	
Variable	**Range**	**Oil ($/bbl)**	**ROI**	**IRR**
OOIP Recovery - m	**0.982**			
Low (50%)	0.491	$ 208.20	20.32%	4.67%
Fuel Costs	**100%**			
Low	50%	$ 182.80	24.68%	7.58%
High	150%	$ 208.14	21.36%	3.36%
Combined Corporate Tax Rate	**44.4%**			
Zero	0.0%	$ 180.20	41.43%	6.22%
Low (50%)	22.2%	$ 187.17	32.23%	5.90%
Hydrotreater Capital Cost	**100%**			
Low	50%	$ 122.18	38.27%	13.30%
High	150%	$ 291.00	15.84%	0.68%
Wellhead vs. WTI Price Differential (ANS)	**0.918**			
Proportional to API gravity (API)	0.485	$ 118.50	23.73%	14.21%
Combined	---			
Con (↓m, ↑fuel, full tax, ↑hydrotreater, ANS)	---	$ 331.42	13.03%	-2.27%
Pro (base m, ↓fuel, zero tax, ↓hydrotreater, API)	---	$ 66.24	75.02%	29.26%
Base	**---**	**$ 195.24**	**23.03%**	**5.59%**

DISCUSSION

The results presented in Section 4 are discussed below according to each of the categories (capital costs, supply costs, cash flow, and sensitivity analysis) introduced previously. A summary of the results from Section 4 is presented in Table 5-1.

5.1 Capital Costs

The capital costs for the base case and dilution scenarios are nearly identical. The primary difference between the two is that since the dilution scenario requires a smaller amount of heavy oil production, smaller pumps, motors, storage tanks, etc. are required for dilution compared to the base case. The primary capital expense in both scenarios is drilling (~$200 million); however, the costs for utility plants (which includes the cost of natural gas and electrical lines) and the feeder oil pipeline are also sizeable (~$90 million) because of the distances traversed.

The capital costs for upgrading can be primarily attributed to the hydrotreater. At a total cost of $1.2 billion, it is by far the most expensive component in this study. The capital cost of the hydrotreater cascades to each subsequent category defined as a percentage of previous cost categories (site preparation, service facilities, startup, etc.), so that upgrading's C_{TCI} is a little over five times larger than that for the base case and dilution scenarios. Costs also cascade in the other scenarios (see the discussion of changing values of *m* in Section 5.4), but since the largest costs are incurred in C_{DPI} for the base and dilution cases instead of C_{TBM}, the inflationary effect is not as widespread.

5.2 Supply Costs

Taxes (severance, property, state, and federal) are one of the largest supply costs in all scenarios. Most of the taxes paid are for severance taxes (base case – 62%, dilution – 71%, upgrading – 88%) because very few deductions are available for severance taxes. As discussed in Section 2.6, the only deductions that can be taken for severance taxes are royalties and operating costs. As costs rise in a scenario, more money must be earned to maintain the same IRR, requiring a higher oil price, which leads to a larger wellhead profit and more severance taxes. This mechanism is responsible for the variation in the percentage of severance taxes in each scenario and explains why nearly 50% of the supply costs for upgrading are taxes.

Looking at each scenario individually, the base case has few supply costs besides delivery. Royalties and taxes are paid on the remainder of the profit generated and the initial capital investment in production ($1.38/bbl for capital and $0.88/bbl for drilling) is easily paid off.

Dilution's largest supply cost is clearly diluent. However, the use of diluent has a number of interesting impacts because it reduces the required production rate of heavy oil. As noted previously, the smaller production scale results in reduced capital costs. Additionally, since royalties and severance taxes are only paid on heavy oil produced from West Sak, the dilution scenario pays less royalties and taxes per barrel of product than the base case when the forecast solution method is used. Overall, the scenario is still more expensive due to the cost of purchasing approximately 30% of the product volume at $70/bbl, but the smaller production scale dampens the impact of the diluent purchase.

Aside from what has already been noted about taxes, the upgrading scenario is the only scenario where significant capital and utility costs (fuel) are a major factor.

5.3 Cash Flow

The base case and dilution scenarios have similar cash flows, with the largest negative present value occurring in year three of the project in the amount of $690 million and $680 million for the base case and dilution, respectively. Annual positive cash flows without applying discount factors and neglecting depreciation are $140 million for both scenarios. The cash flow for the upgrading scenario also has its largest negative present value occur in year three in the amount of $3.5 billion, with annual cash flows of $700 million (without applying discount factors and neglecting depreciation). The cash flows in all scenarios follow the same shape because of the constant oil sale price and 15% IRR assumption they share.

5.4 Sensitivity Analysis

The base case sensitivity analysis is extremely positive for all of the parameters investigated. Reducing the efficiency of production by cutting *m* in half requires a doubling of the number of wells drilled to meet the same production schedule and has a substantially greater impact on reducing the IRR than increasing delivery expenses. As discussed in Section 5.1, increased capital costs for drilling and other items inflate other estimated capital expenses defined as a percentage of previous categories. Reducing or eliminating corporate income taxes is the greatest potential IRR gain.

Similar to the base case scenario, the biggest potential loss of IRR for dilution is reduced recovery. Diluent mixing ratio has a smaller impact on profitability than diluent purchase price due to the tradeoff between reduced diluent cost and higher capital, royalty, and severance tax costs. Overall, the dilution scenario is also very positive economically. Even the worst case scenario results in an IRR > 15% and a constant oil price of $92/bbl that is close to the current market price for WTI.

For upgrading, the parameter that the economic results are most sensitive to is the wellhead value of West Sak heavy oil. If the crude's value is assumed to be the same as ANS with a price differential of 91.8% of WTI, then several billion dollars are being spent to get the crude into a pipeline, severance taxes are paid on nearly the full sale price of the crude, and the sale value is only increased by 8%. Dropping the assumed value of heavy oil compared to WTI substantially reduces the severance tax liability while still allowing for sale at the fully upgraded WTI price. Changes to the cost of building and operating the hydrotreater also substantially affect the overall economics of the scenario. As noted previously, large capital costs cascade into other costs defined as fractions of previous cost categories. Assuming that the hydrotreater's capital cost could be reduced by 50% ($600 million), C_{TCI} is reduced from $4.1 billion to $2.7 billion. Modifying the overall fuel costs has a modest ± 2% impact on IRR. Unlike previous scenarios, reducing the corporate tax rate or recovery efficiency has very little impact compared to the parameters already discussed. The combined positive parameters give an economic result that is similar to dilution, but overall, upgrading appears to have little chance of making an acceptable rate of return.

Table 5-1: Results summary (air-firing). The lowest present value refers to the maximum negative present value for that scenario.

	Base Case	**Dilution**	**Upgrading**
C_{TCI} (10^6 dollars)	\$839	\$814	\$4,129
CPFB (\$/bpd)	\$16,781	\$16,279	\$82,580
Supply Cost (for 15% IRR, \$/bbl)	\$29.44	\$48.41	\$195.94
Lowest Present Value (10^6 dollars)	\$690	\$680	\$3,500
IRR			
Low	28%	23%	---
Reference	41%	45%	6%
High	51%	60%	13%

CONCLUSIONS AND RECOMMENDATIONS

Based on the results of this study, there appear to be no significant economic barriers to the production of heavy oil from West Sak via polymer flooding with or without dilution. Both the base case and dilution scenarios produce an IRR > 15% under all of the conditions investigated in the sensitivity analysis. Dilution is riskier than the base case scenario and should only be pursued if heavy oil is incompatible with TAPS and its current oil shipments.

Upgrading heavy oil via hydrotreating is not economically feasible. A number of unrealistic assumptions must be made for the scenario to make a reasonable rate of return, and it is possible that the scenario could be a loss. The largest barrier to the economic feasibility of this scenario is a combination of capital costs and severance tax policy. Large capital costs require significant net earnings which imply high oil prices and considerable severance tax liabilities. Capital costs must either be cut drastically or severance tax policy changed in order for upgrading to be viable. Even so, if dilution (or another similar pipeline compatibility method) were available, upgrading would still have a poorer economic outlook than the alternative.

Carbon management is largely irrelevant for the scenarios studied here. Both the base case and dilution scenario are relatively small industrial sources. Due to the large amounts of heat required for upgrading, that scenario does produce a significant amount of CO_2. However, a carbon tax of at least $86 per ton of CO_2 would be necessary to incentivize oxy-firing; otherwise, it is less expensive to just pay the CO_2 tax. Given the current political climate, no laws or regulations taxing CO_2 emissions at that rate (or any rate) is likely for the foreseeable future.

A number of assumptions made in this study could be improved upon in future work. Some categories of equipment were selected because they were the only equipment for which costing information was available, even if their use is not applicable for the environment on the North Slope. Included in this list would be buried pipelines, utility service lines costs, drilling cost data based on national averages for vertical wells, and storage tanks for separating and mixing reservoir fluids. Several capital cost categories defined as percentages of other capital costs might not be applicable for heavy oil production. For example, it is difficult to believe that it takes $50 million dollars for startup in the base case scenario when there are effectively no major pieces of equipment to manage. While differential pricing data were available for ANS crude, no pricing data were found specifically for West Sak (or West Sak like) heavy oil. Given the sensitivity of all the scenarios to severance taxes, determining wellhead value of produced heavy oil accurately should be a priority. In terms of production, reservoir simulations or actual field tests could improve upon the accuracy of fractional flow calculations. Plans announced in 2004 by major oil companies to expand production from West Sak to 40,000 bpd have all failed to produce at predicted levels, and it is not clear from the literature what obstacles to production have been encountered.

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