NATIONAL ENERGY TECHNOLOGY LABORATORY

Baseline Analysis of Crude Methanol Production from Coal and Natural Gas

Revised October 15, 2014

DOE/NETL-341/101514

U.S. DEPARTMENT OF ENERGY

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Final Report

October 15, 2014

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DOE Contract DE-FE0004001

Task 04001.341.02, Activity 022

Acknowledgments

This report was updated by Energy Sector Planning and Analysis (ESPA) for the United States Department of Energy (DOE), National Energy Technology Laboratory (NETL). This work was completed under DOE NETL Contract Number DE-FE0004001. This work was performed under ESPA Task 04001.341.02, Activity 022.

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

William Summers, NETL, Office of Program Performance and Benefits

Kristin Gerdes, NETL, Office Director, Office of Program Performance and Benefits

John Wimer, NETL, Office Director, Office of Coal and Power R&D

Acknowledgments for permissions to use figures

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List of Exhibits

List of Acronyms and Abbreviations

AGR	Acid gas removal
ASU	Air separation unit
ATR	Autothermal reformer
BACT	Best available control technology
BAH	Booz Allen Hamilton Inc.
bbl	Barrel
BEC	Bare erected cost
BOP	Balance of plant
BPD	Barrel per day
Btu	British thermal unit
C-MTG	Coal-to-methanol-to-gasoline
CB-MTG	Coal and biomass-to-methanol-to-gasoline
CBTL	Coal and biomass-to-liquids
CCF	Capital charge factor
CCS	Carbon capture and sequestration
CCUS	Carbon capture, utilization, and storage
CH_3OH	Methanol
Cm	centimeter
CMM	Coal mine methane
CMT	Constant Maturity Treasury
CO	Carbon monoxide
CO_2	Carbon dioxide
CO_2E	Carbon dioxide equivalent
COE	Cost of oil equivalent
COP	Crude oil price
COS	Carbonyl sulfide
CTL	Coal to liquids
CTM	Coal-to-methanol
DB	Daily barrel
DME	Di-methyl ether
DOE	Department of Energy
DSCR	Debt service coverage ratio
EIA	Energy Information Administration
EISA	Energy Independence and Security Act
EOR	Enhanced oil recovery
EPA	Environmental Protection Agency
EPC	Engineering/procurement/construction
EPCM	Engineering/procurement/construction management
ERC	Emission reduction credits
FCR	Fixed charge rate
ft	feet
GHG	Greenhouse gas
GHGEV	Greenhouse gas emission value
GJ	Gigajoule
gpm	Gallon per minute
GWP	Global warming potential
H_2	Hydrogen
H_2S	Hydrogen sulfide
HGT	Heavy gasoline treatment
HHV	Higher heating value
HP	High pressure
hp	Horsepower
hr	hour
HRSG	Heat recovery steam generator
HVAC	Heating, Ventilation, & Air Conditioning
IEA	International Energy Agency
IGCC	Integrated gasification combined cycle
IOU	Investor-owned utility
IPCC	Intergovernmental Panel on Climate Change
IRR	Internal rate of return
IRROE	Internal rate of return on equity
ISO	International Organization for Standardization
kJ	Kilojoule
kPa	Kilopascal absolute
kW, kWe	Kilowatt (electric)
kWh	Kilowatt-hour
kWt	Kilowatt (thermal)
LAER	Lowest achievable emission rate
lb	Pound
lbmole	Pound mole
LHV	Lower heating value
LIBOR	London Interbank Offered Rate
LNB	Low NOx burner
LOX	Liquid oxygen
LPG	Liquefied petroleum gas
m	Meter
MJ	Megajoules
MMBtu/hr	Million British thermal units (also shown as 106 Btu) per hour
mol%	Mole percent
MON	Motor octane number
MPa	Megapascal absolute
MTG	Methanol to gasoline
MW, MWe	Megawatt (electric)
MWh	Megawatt-hour
MWt	Megawatt (thermal)
N_2	Nitrogen
N_2O	Nitrous oxide
Neg.	Negligible
NETL	National Energy Technology Laboratory
NGCC	Natural gas combined cycle
NGPA	Natural Gas Policy Act
NGTM	Natural gas to methanol
NOx	Oxides of nitrogen
NSR	New Source Review
N/A	Not applicable
O_2	Oxygen
O&GJ	Oil and Gas Journal

O&M	Operations and maintenance
PFD	Process flow diagram
PM	Particulate matter
POTW	Publicly owned treatment works
ppb	Parts per billion
ppbv	Parts per billion volume
ppm	Parts per million
ppmd	Parts per million, dry
ppmv	Parts per million volume
ppmvd	Parts per million volume, dry
PRB	Powder River Basin coal
PSA	Pressure swing adsorption
PSFM	Power systems financial model
psia	Pound per square inch absolute
psig	Pound per square inch gauge
QGESS	Quality Guidelines for Energy System Studies
RAND	RAND Corporation
ROE	Return on equity
RON	Research octane number
RSP	Required selling price
scf	standard cubic feet
scm	standard cubic meter
SCR	Selective catalytic reduction
SO_2	Sulfur dioxide
TASC	Total as-spent cost
TOC	Total overnight cost
tonne	Metric ton (1000 kg)
TPC	Total plant cost
T&S	Transport and storage
U.S.	United States
vol%	Volume percent
WGSR	Water gas shift reactor
wt%	Weight percent
WTW	Well-to-wheels
°C	Degrees Celsius
°F	Degrees Fahrenheit
$M	Millions of dollars
μS	microSiemens

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Executive Summary

Gasification of coal, in addition to generating syngas for power production, has the potential to produce a diverse array of high-value products. It is a challenge to understand the optimal use of this domestic coal resource amidst the potential technology options, product slates (including co-production of power), and competing feedstocks (natural gas, petroleum). This analysis seeks to begin addressing that challenge by focusing on one primary product, methanol, which also serves as a readily-transportable intermediate to many other products including olefins, gasoline, and di-methyl ether (DME). This report presents the design configuration, performance summaries, and cost estimates of three cases which generate crude methanol.

- Case 1: Coal-to-crude methanol without carbon capture and sequestration (CCS) (i.e., representing a building block to other derivatives, not chemical-grade methanol)
- Case 2: Coal-to-crude methanol with CCS
- Case 3: Natural gas-to-crude methanol with CCS

The required selling price (RSP) of crude methanol was estimated for each case and the results are listed in Exhibit ES-1. The RSP for the natural gas case is significantly lower than the coal cases due primarily to the lower capital requirements, which more than offset the higher feedstock cost.

Exhibit ES-1 Cost estimation results

Case	1	2	3
Total Plant Cost (2011 million$)	4,586	4,882	2,172
Total Overnight Cost (2011 million$)	5,615	5,973	2,644
Total As-spent Capital (2011 million$)[B]	6,440 6,631	6,851 7,054	3,033 3,123
RSP[A] Component Details ($/gal)			
Capital[B]	0.85 1.09	0.91 1.16	0.40 0.52
Fixed O&M	0.14	0.15	0.07
Variable O&M	0.09	0.09	0.03
Coal	0.21	0.21	0
Natural gas	0.03	0.05	0.57
Power	0	0	-0.05
CO_2 T&S	0	0.06	0.01
RSP[B] Total ($/gal)	**1.31 1.56**	**1.46 1.72**	**1.03 1.14**
RSP[E] Total ($/ton)	**396.70 469.29**	**441.44 518.67**	**311.17 345.39**
Costs of CO_2 captured[BC] ($/tonne)	**N/A**	**16.36 19.20**	**N/A**
Cost of CO_2 avoided[BD] ($/tonne)	**N/A**	**29.67 32.75**	**N/A**

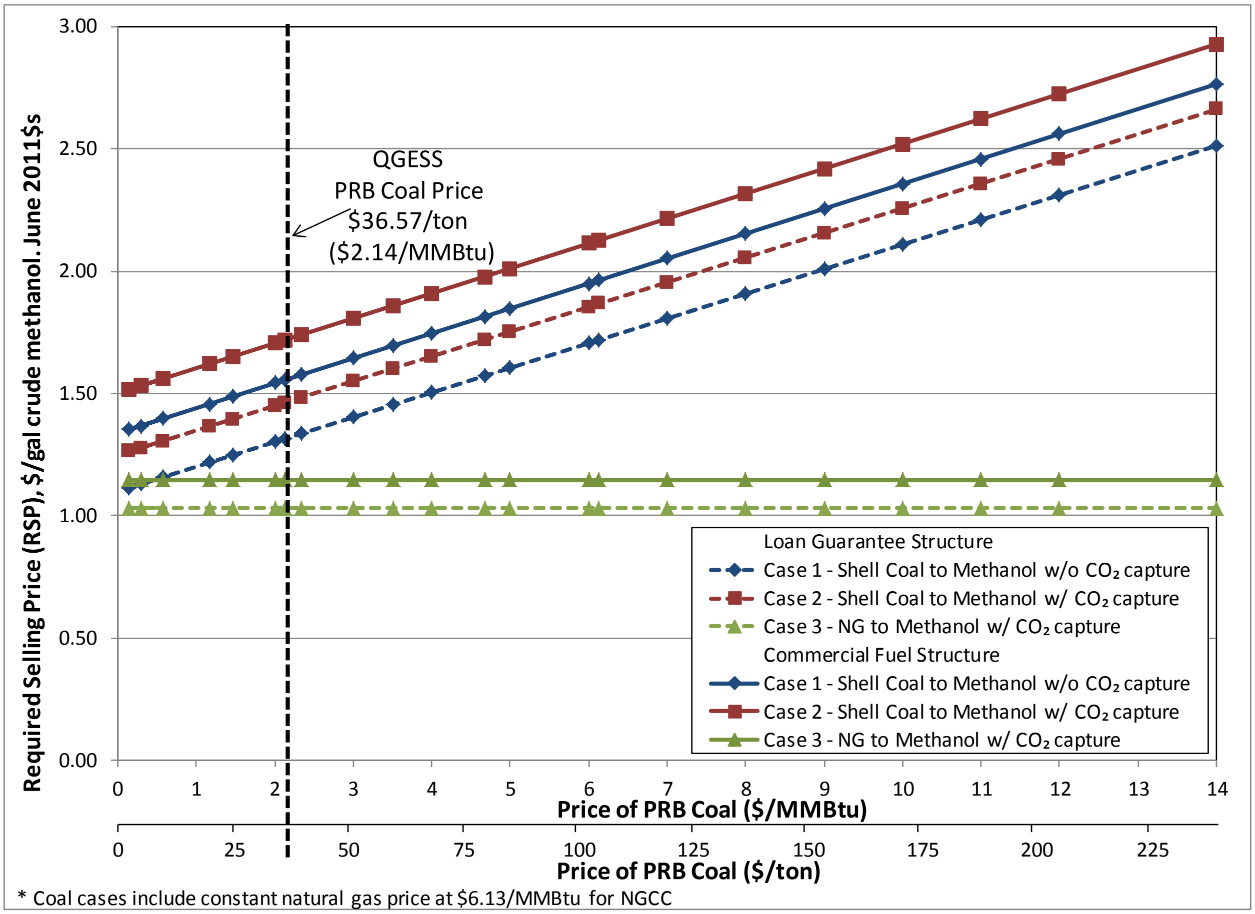
[A] Capacity factor assumed to be 90 percent.
[B] Values are shown for two financial structures.
The first (lower) value is based on the loan guarantee finance structure.
The second (higher) value is based on the commercial fuels finance structure.
[C] Excludes CO_2 transport and storage (T&S).
[D] Includes CO_2 T&S.
[E] Based on 332.6 gal/tonne or 301.73 gal/ton.
All costs are in June 2011 dollars.

The sensitivity of the RSP to the coal and natural gas feedstock prices is illustrated in Exhibit ES-2 and Exhibit ES-3, respectively. The prices used in this study are based on the National Energy Technology Laboratory (NETL) Quality Guidelines for Energy System Studies (QGESS) Recommended Fuel Prices. [1] Cases 1 and 2 use coal from the Powder River Basin (PRB) region in Montana, priced at $36.57/ton ($2.1351/MMBtu) including delivery to the Midwestern site, as the feedstock. Case 3 uses natural gas, priced at $6.13/MMBtu. Cases 1 and 2 also burn some natural gas to provide additional power required by the process.[1] The expected RSPs for the case using natural gas as feedstock are below those for the coal-based cases because of the lower capital and fixed operating charges. On an energy content basis, the RSPs for all cases increase at approximately the same rate with increases in the feedstock prices due to the similarity between the feedstock requirements per unit of product. Cases 1 and 2 require 9.75 MMBtu of PRB coal per gallon of methanol generated and Case 3 requires 9.27 MMBtu of natural gas per gallon of methanol generated.

Given the higher cost per MMBtu of natural gas relative to coal and the similar energy input requirements of the conversion process, a 100 percent increase in natural gas prices leads to a 50 to 55 percent increase in RSP, for commercial and loan guarantee financing respectively, versus a 100 percent increase in coal prices leading to a 13 to 16 percent increase in RSP consistent with the natural gas–based methanol production being the less capital intense but more operating margin dependent technology choice. Hence, the natural gas route is more exposed to feedstock price volatility.

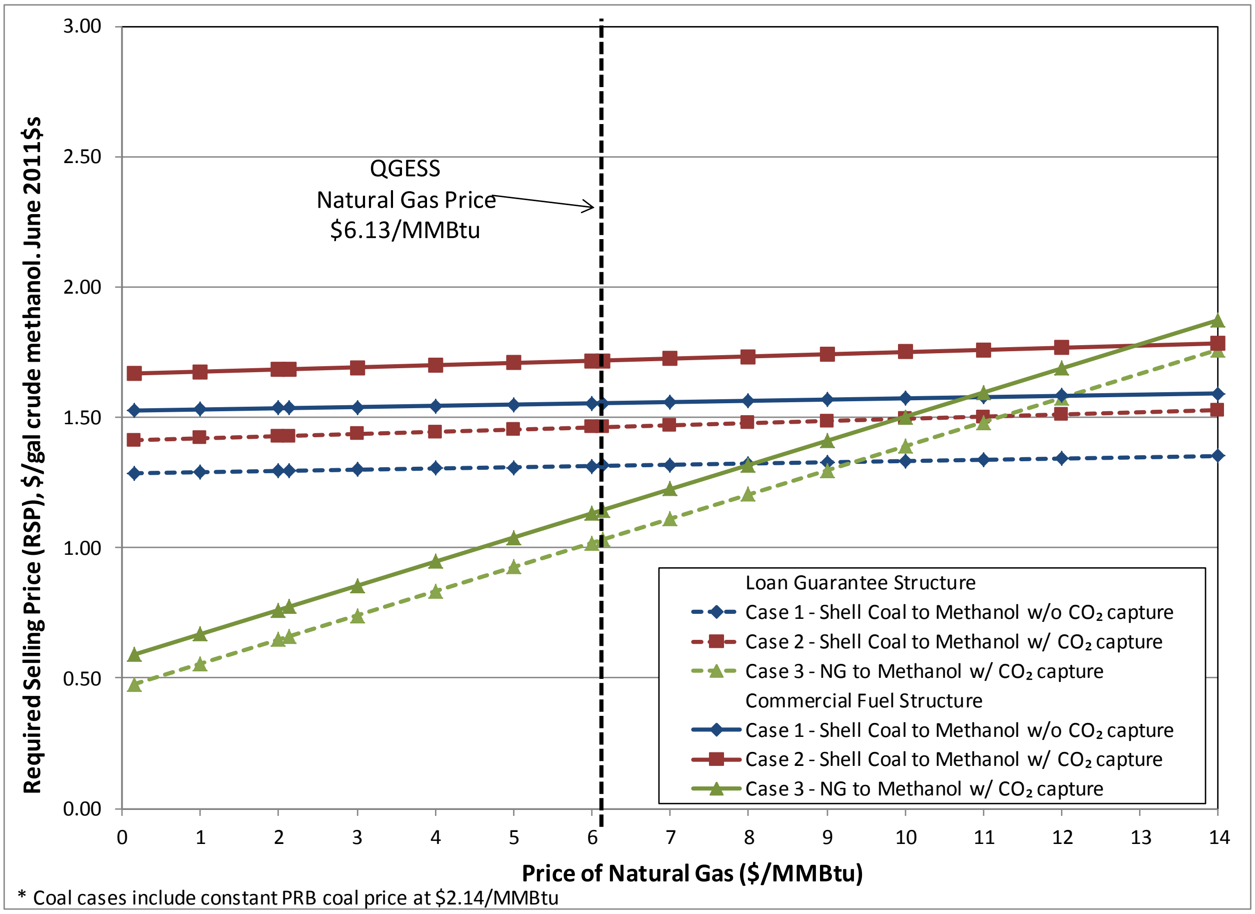
[1] A natural gas combined cycle plant was used to generate additional power in the coal gasification cases to make the plant approximately power neutral while maximizing the production of crude methanol (i.e., no syngas is diverted for power production).

Exhibit ES-2 Sensitivity of the required selling price to the coal price



Based on Methanol density of 332.6 gal/tonne or 301.73 gal/ton

Exhibit ES-3 Sensitivity of the required selling price to the natural gas price



Based on Methanol density of 332.6 gal/tonne or 301.73 gal/ton

1 Overview

This report presents the design configuration, performance, and cost of the three crude methanol cases.

1.1 Background

Gasification of coal, in addition to generating syngas for power production, has the potential to produce a diverse array of high-value products. It is a challenge to understand the optimal use of this domestic coal resource amidst the potential technology options, product slates (including co-production of power), and competing feedstocks (natural gas, petroleum). This analysis seeks to begin to address that challenge by focusing on one primary product, methanol, which also serves as a readily-transportable intermediate to many other products including olefins, gasoline, and di-methyl ether (DME). The information from this analysis and previous studies can be used, along with available data on the production of methanol derivatives, in developing a framework for evaluating and optimizing the utilization of domestic coal and natural gas resources.

1.2 Case Descriptions

For each of the cases listed in Exhibit 1-1, and also described below, a system study was completed in accordance with the May 2010 version of the report, "Scope and Reporting Requirements for NETL System Studies." [2] The cases in this report are limited to the following:

- Coal-to-crude methanol with and without carbon capture and sequestration (CCS) (i.e., representing a building block to other derivatives, not chemical-grade methanol)
- Natural gas-to-crude methanol with CCS

All three cases are sized to produce approximately 10,000 metric tons of methanol per day (3,326,000 gal/day based on 332.6 gal/tonne [3]). This plant size is considered large scale but typical for the current design of new plants. [4] This report presents the results of the performance modeling and cost estimating for the crude methanol product cases (cases 1, 2, and 3).

Exhibit 1-1 Case descriptions

Case	Feedstock	Steam Cycle, psig/°F/°F	Combustion Turbine*	Gasifier Technology	Oxidant	Sulfur Removal/ Recovery	CO_2 Separa-tion	Product(s)
1	Coal/Syngas	1800/1050/ 1050	SGT-1000F	Shell	95 mol% O_2	Rectisol/ Claus Plant	Rectisol**	Crude Methanol
2	Coal/Syngas	1800/1050/ 1050	SGT6-2000E	Shell	95 mol% O_2	Rectisol/ Claus Plant	Rectisol & Amine***	Crude Methanol
3	Natural Gas	1800/1050/ 1050	SGT6-3000E	N/A	95 mol% O_2	N/A	Amine	Crude Methanol

**Turbine selection based on vendor data included in GT PRO software by Thermoflow, Inc. [5]*
***CO_2 removed by the Rectisol process and not used for coal transportation is vented.*
****Amine process added to NGCC system only.*

2 Design Criteria

2.1 Site Description

All plants in this study are assumed to be located at a generic plant site in Midwestern U.S., with ambient conditions and site characteristics as presented in Exhibit 2-1 and Exhibit 2-2. These assumptions are identical to the National Energy Technology Laboratory (NETL) baseline studies. [6,7] The ambient conditions are the same as International Organization for Standardization (ISO) conditions.

Exhibit 2-1 Site ambient conditions

Description	Values
Elevation, m, (ft)	0, (0)
Barometric Pressure, MPa (psia)	0.10 (14.696)
Design Ambient Temperature, Dry Bulb, °C (°F)	15 (59)
Design Ambient Temperature, Wet Bulb, °C (°F)	11 (51.5)
Design Ambient Relative Humidity, %	60

Exhibit 2-2 Site characteristics

Description	Conditions
Location	Greenfield, Midwestern USA
Topography	Level
Size, acres	300
Transportation	Rail
Ash Disposal	Off Site
Water	Municipal (50%) / Groundwater (50%)
Access	Land locked, having access by rail and highway
CO_2 Storage	Compressed to 15.3 MPa (2,215 psia), transported 80 kilometers (50 miles) and sequestered in a saline formation at a depth of 1,239 m (4,055 ft)

As assumed for gasification-based cases in the NETL baseline studies, the land area for these cases (including the natural gas reforming case) is estimated as 30 acres required for the plant proper with the balance providing a buffer of approximately 0.25 miles to the fence line. [6] The extra land could also provide for a rail loop, if required.

In all cases, it was assumed that the steam turbine is enclosed in a turbine building. The gasifiers, reformers, methanol synthesis reactors, and the combustion turbines are not enclosed.

Allowances for normal conditions and construction are included in the cost estimates. The following design parameters are considered site-specific, and are not quantified for this study. Costs associated with the site specific parameters can have significant impact on capital cost estimates.

- Flood plain considerations
- Existing soil/site conditions
- Water discharges and reuse
- Rainfall/snowfall criteria
- Seismic design

- Buildings/enclosures
- Local code height requirements
- Noise regulations – Impact on site and surrounding area

2.2 Design Feedstocks

2.2.1 Natural Gas Characteristics

The natural gas composition used in this analysis, representative of natural gas after going through standard midstream processing, is presented in Exhibit 2-3. [8]

Exhibit 2-3 Natural gas composition

Component		Volume Percentage
Methane	CH_4	93.1
Ethane	C_2H_6	3.2
Propane	C_3H_8	0.7
n-Butane	C_4H_{10}	0.4
Carbon Dioxide	CO_2	1.0
Nitrogen	N_2	1.6
	Total	100.0
Units	**LHV**	**HHV**
kJ/kg	47,454	52,581
MJ/scm	34.71	38.46
Btu/lb	20,410	22,600
Btu/scf	932	1,032

Note: Feedstock composition is normalized and heating values are calculated. For the purposes of this study no sulfur content was considered.

2.2.2 Coal Characteristics

The coal properties from National Energy Technology Laboratory's (NETL) Quality Guidelines for Energy System Studies (QGESS): Specifications for Selected Feedstocks, for a subbituminous Powder River Basin (PRB) coal from Montana are shown in Exhibit 2-4. [8]

Exhibit 2-4 Montana Rosebud PRB, Area D, Western Energy Co. Mine, subbituminous design coal analysis

Proximate Analysis	Dry Basis, %	As Received, %
Moisture	0.0	25.77
Ash	11.04	8.19
Volatile Matter	40.87	30.34
Fixed Carbon	48.09	35.70
Total	100.0	100.0
Ultimate Analysis	**Dry Basis, %**	**As Received, %**
Carbon	67.45	50.07
Hydrogen	4.56	3.38
Nitrogen	0.96	0.71
Sulfur	0.98	0.73
Chlorine	0.01	0.01
Ash	11.03	8.19
Moisture	0.00	25.77
Oxygen[1]	15.01	11.14
Total	100.0	100.0
Heating Value	**Dry Basis**	**As Received**
HHV, kJ/kg	26,787	19,920
HHV, Btu/lb	11,516	8,564
LHV, kJ/kg	25,810	19,195
LHV, Btu/lb	11,096	8,252
Hardgrove Grindability Index		57
Ash Mineral Analysis		**%**
Silica	SiO_2	38.09
Aluminum Oxide	Al_2O_3	16.73
Iron Oxide	Fe_2O_3	6.46
Titanium Dioxide	TiO_2	0.72
Calcium Oxide	CaO	16.56
Magnesium Oxide	MgO	4.25
Sodium Oxide	Na_2O	0.54
Potassium Oxide	K_2O	0.38
Sulfur Trioxide	SO_3	15.08
Phosphorous Pentoxide	P_2O_5	0.35
Barium Oxide	Ba_2O	0.00
Strontium Oxide	SrO	0.00
Unknown	---	0.84
Total	---	100.0
Trace Components		**ppmd**
Mercury[2]	Hg	0.081

[1] *By difference*
[2] *Mercury value is the mean plus one standard deviation using EPA's ICR data*

2.3 Environmental Requirements

The environmental limits presented in this section refer to the gasification/power cycle only, because the environmental requirements for the methanol plant are considered beyond the scope of the study.

The environmental targets for the study were considered on a technology- and fuel-specific basis. Since all the cases are located at a greenfield site, permitting a new plant would involve the New Source Review (NSR) permitting process. The NSR process requires installation of emission control technology, meeting either the best available control technology (BACT) determinations for new sources located in areas meeting ambient air quality standards (attainment areas), or the lowest achievable emission rate (LAER) technology for sources located in areas that do not meet ambient air quality standards (non-attainment areas). This study uses the BACT guidelines, which are summarized in Exhibit 2-5.

Exhibit 2-5 BACT environmental design basis

Environmental Design Basis		
Pollutant	**Control Technology**	**Limit**
Sulfur Oxides (SO_2)	Rectisol® + Claus Plant/ Econamine Plus	≤ 0.050 lb/10^6Btu
Nitrogen Oxides (NOx)	LNB and N_2 Dilution primarily with humidification as needed	15 ppmvd (@ 15% O_2)
Particulate Matter (PM)	Cyclone/Barrier Filter/Wet Scrubber/AGR Absorber	0.006 lb/10^6Btu
Mercury (Hg)	Activated Carbon Bed	95% removal

Selection of the process technology accounts for obtaining minimum sulfur content in the syngas and final product.

The following regulatory assumptions are used for assessing environmental control technologies:

- NOx Emission Reduction Credits (ERC) and allowances are not available for the project emission requirements when located in the ozone attainment area
- Solid waste disposal is either offsite at a fixed $/ton fee or is classified as a byproduct for reuse, claiming no net revenue ($/ton) or cost
- Raw water is available to meet technology needs
- Wastewater discharge will meet effluent guidelines rather than water quality standards for this screening

Based on published vendor literature, it was assumed that low NOx burners (LNB) and nitrogen dilution can meet 15 ppmv (dry) limits on NOx emissions from the combustion turbines at 15 percent O_2. This value was used for all cases.

The acid gas removal (AGR) process in the coal gasification cases must have a sulfur capture efficiency of about 99.7 percent to reach the environmental target and to also produce sulfur-free syngas to avoid poisoning of catalysts in methanol synthesis process. Vendor data on AGR processes used indicate that this level of sulfur removal is possible.

In the CO_2 capture cases, the two-stage Rectisol process in the coal gasification cases was designed for 90 percent plant CO_2 removal. The amine capture process for the natural gas case and the natural gas combined cycle (NGCC) system in the coal fired gasification case with capture was also designed for 90 percent plant CO_2 removal. The 90 percent capture design for both CO_2 point sources results in an overall 90 percent capture rate for the plant.

For the coal feedstock cases, most of the coal ash is removed from the gasifier as slag. Any ash that remains entrained in the syngas is captured in the downstream equipment, including the syngas scrubber and a cyclone followed by either ceramic or metallic candle filters. The environmental target can be achieved with the combination of particulate control devices.

Mercury capture is not required for the case with processed natural gas as the starting feedstock. For the coal feedstock cases, however, the environmental target for mercury capture is greater than 90 percent capture. Eastman Chemical's operating experience at its coal-to-methanol plant in Kingsport, Tennessee has shown mercury removal efficiency of 95 percent. This value was used as the assumed performance level for this study. Sulfur-impregnated activated carbon is used by Eastman as the adsorbent in the packed beds operating at 30°C (86°F) and 6.2 MPa (900 psig). Mercury removal between 90 and 95 percent has been reported with a bed life of 18 to 24 months. Removal efficiencies may be even higher, but at 95 percent the measurement precision limit is reached. Mercury removals of greater than 99 percent can be achieved by the use of dual beds, i.e., two beds in series. However, this study assumes that the use of sulfur-impregnated carbon in a single carbon bed achieves 95 percent reduction of mercury emissions, which meets the environmental target limits and the New Source Performance Standards limit in all cases.

For the cases that feature carbon capture, CO_2 transport and storage (T&S) was modeled based on the specifications in the NETL QGESS: Estimating Carbon Dioxide Transport and Storage Costs. [9] The CO_2 is supplied to a 100 km (62 mi) pipeline at the plant fence line at a pressure of 15.3 MPa (2,215 psia). The CO_2 product gas composition varies in the cases presented, but is expected to meet the specification described in Exhibit 2-6. [10] A glycol dryer located near the mid-point of the compression train is used to meet the moisture specification.

Exhibit 2-6 CO_2 pipeline specification

Parameter	Units	Parameter Value
Inlet Pressure	MPa (psia)	15.3 (2,215)
Outlet Pressure	MPa (psia)	10.4 (1,515)
Inlet Temperature	°C (°F)	35 (95)
N_2 Concentration	Ppmv	< 300
O_2 Concentration	Ppmv	< 40
Ar Concentration	Ppmv	< 10
H_2O Concentration	Ppmv	< 150

2.4 Balance of Plant Requirements

Assumed balance of plant requirements are listed in Exhibit 2-7.

Exhibit 2-7 Balance of plant design requirements

Feedstock and Other Storage	
Coal	30 days
Slag	30 days
Sulfur	30 days
Natural Gas	Pipeline delivery (no on site storage)
Plant Distribution Voltage	
Motors below 1 hp	110/220 volt
Motors 250 hp and below	480 volt
Motors above 250 hp	4,160 volt
Motors above 5,000 hp	13,800 volt
Steam and Gas Turbine Generators	24,000 volt
Grid Interconnection Voltage	345 kV
Water and Waste Water	
Cooling system	Recirculating, Evaporative Cooling Tower
Makeup water	The water supply is 50 percent from a local POTW and 50 percent from groundwater, and is assumed to be in sufficient quantities to meet plant makeup requirements. Makeup for potable, process, and de-ionized (DI) water is drawn from municipal sources.
Feed water	Process water treatment is included and will produce boiler feed quality water for the gasification, natural gas and steam cycle systems.
Process Wastewater	Process wastewater and storm water that contacts equipment surfaces will be collected and treated for recycle. Selected blowdown will be discharged through a permitted discharge permit.
Water Discharge	Most of the wastewater is to be recycled for plant needs. Blowdown will be treated for chloride and metals, and discharged.
Sanitary Waste Disposal	Design will include a packaged domestic sewage treatment plant with effluent discharged to the industrial wastewater treatment system. Sludge will be hauled off site.
Solid Waste	Gasifier slag is assumed to be a solid waste that is classified as non-hazardous. An offsite waste disposal site is assumed to have the capacity to accept waste generated throughout the life of the facility. Solid waste sent to disposal is at an assumed nominal fee per ton, even if the waste is hauled back to the mine. Solid waste generated that can be recycled or reused is assumed to have a zero cost to the technology.

2.5 Crude Methanol – Deviations from the C-MTG Study

This study is based on the coal-to-methanol-to-gasoline (C-MTG) cases in the Baseline Analysis of Subbituminous Coal and Biomass to Gasoline (Indirect Liquefaction by Methanol Synthesis) [11] with the methanol purification and gasoline synthesis processes removed. A natural gas to methanol case was added for comparison. The cases covered in this report simply consider a relaxation of the product purity specifications typically associated with solvent-grade (Grade A) and chemical-grade (Grade AA) methanol. Methanol is the raw material for many chemical products such as formaldehyde, dimethyl terephthalate, methylamines, methyl halides, methyl methacrylate, and acetic acid. Most importantly, methanol provides the foundation of the methanol-to-hydrocarbon platform which has commercially proven processes to convert methanol into gasoline and olefins (such as ethylene, the building block of the petrochemical industry).

Since the methanol-to-hydrocarbon mechanism is fairly forgiving with respect to oxygenated hydrocarbons and limited amounts of water, cases were considered where methanol purification (i.e. distillation columns) was removed. This is consistent with the methanol being fed to the commercial methanol-to-gasoline (MTG) process in New Zealand following mild equilibration over an acidic alumina bed (to bring the methanol-DME-water into equilibrium)[2]. Skipping this step (removal of methanol purification) allows for potential savings in capital investment, so it was eliminated from the cases in this report.

Use of natural gas instead of coal significantly reduces the amount of capital investment. Besides elimination of coal handling functions, the syngas production function of the Shell gasifier in the coal cases is replaced by steam methane reforming, partial oxidation, or autothermal reforming for natural-gas-based applications. Based on experience, autothermal reforming is considered the proper technology choice for any reasonable-scale, natural-gas-fed methanol plant. The autothermal reformer produces a H_2/CO ratio of two, as preferred for methanol synthesis, similar to the case where coal is fed to a Shell gasifier and a water-gas-shift process.

The methanol synthesis is essentially the same following the production of syngas of an appropriate stoichiometric number.[3] The designation crude merely conveys not purifying the methanol beyond the requirements for (essentially) immediate consumption in a methanol-to-products (gasoline or similar) facility. This study did investigate high level adjustments to the reactor that could lead to reduced capital expenditure, such as lowering the pressure, but the corresponding impact on process yields made the decision to not investigate such options further obvious. Long-range opportunities to make coal-based systems more competitive, such as revisiting non-copper-based catalytic systems, were not analyzed. Non-copper-based catalytic systems (i.e., zinc oxide and chromium oxide systems) are more tolerant to common coal contaminants and, therefore, may require less capital investment. Technology scoping of these opportunities was beyond the scope of this study.

[2] An allocation of the cost of the equilibration bed should be included in evaluating this option; however, this cost is outside the scope of this study and would be part of a more complex integrated facility cost optimization if implemented.

[3] The stoichiometric number or "S" is defined as [moles of hydrogen – moles of carbon dioxide]/[moles of carbon monoxide + moles of carbon dioxide].

3 Crude Methanol Model Performance Results

3.1 Technology Background and Readiness Assumptions

The current status of the major technology systems included in this study is highlighted below.

- Coal gasification is a mature technology that has been deployed throughout the world, including in conjunction with methanol synthesis plants.
- Natural gas reforming is a mature technology that has been deployed throughout the world, including in conjunction with methanol synthesis plants.
- Reductions in the price of natural gas and increases in available quantities have made natural gas the preeminent feedstock for commercial methanol production since the mid-twentieth century
- Methanol synthesis from syngas is a commercial process first used in 1923. The current catalytic process has been in widespread use since the late 1960s and is considered to be a proven and robust technology. The current catalytic process is optimized for the dominant natural gas feedstocks rather than coal, but is proven to be economically viable for either feedstock. [12]
- Elemental sulfur is generated by the well-established Claus process for the coal cases. The sulfur is collected with a purity of 99 percent.

In Cases 2 and 3, carbon dioxide is captured, treated, and compressed to meet the CO_2 pipeline specifications. CO_2 specifications are negotiated and determined on a site-by-site basis. Drying was included in the CO_2 capture and compression cost estimations, but additional cleanup that could be needed to meet more rigorous pipeline specifications was considered to be beyond the scope of this study. [13]

3.2 Modeled Performance Summary

The three cases analyzed in this report were modeled using Aspen Plus® (Aspen). Performance and process estimates were based upon published reports, presentations, information obtained from vendors, cost and performance data from design/build utility projects, and/or best engineering judgment.

The mass balances for the two coal-based cases are summarized in Exhibit 3-1. All three of the cases are sized to produce approximately 10,000 metric tons of methanol per day (3,326,000 gal/day based on 332.6 gal/tonne [14]). This plant size is considered large scale but typical for the current design of new plants. [4] The Case 2 mass balance is the same as Case 1, except that the CO_2 recovered from the Rectisol unit is not vented, but rather compressed and sequestered. The mass balance for the natural-gas-based case is shown in Exhibit 3-2.

A comparison of the syngas feed to the methanol synthesis system as well as the raw methanol product stream is shown in Exhibit 3-3. The methanol synthesis systems for the three cases are virtually identical.

A summary of the auxiliary power requirements for the crude methanol plants in each of the three cases is given in Exhibit 3-4 and Exhibit 3-5. Some of the auxiliary power requirements for each case are met by adding a heat recovery steam generation (HRSG) system to power a steam turbine. Small gas turbines are combined with the HRSGs in each case to supply additional power, and the performance estimates for the additional gas turbine plants are shown in Exhibit 3-6.

In the natural gas case, the combustion turbine is fueled by the process exhaust gases solely to recover energy value from this stream. Additional natural gas is not necessary to supply sufficient power because the combustion of the tail gases generates excess power beyond auxiliary power load requirements. The excess power can be sold to the grid; however, the sale may be at a steep discount as entities that are negotiating a power purchase agreement will know the power production is an inherent by-product of core methanol production operations. Consequently, the actual achieved transfer price for excess power will be a significant risk in natural gas feedstock projects and would be highly project dependent. The sensitivity of the results to the electricity selling price received for the excess power is included in the results section of this report and indicates that the impact on the methanol RSP is fairly small. High excess power is endemic to the production of methanol from natural gas because there is considerable heat recovery from the exothermic synthesis process, the production of significant amounts of tail gas without use such as coal drying, and lower energy requirements than coal-based processes.

In the coal cases, the exhaust gases are used in coal drying; so the combustion turbine is fueled by natural gas to generate additional auxiliary power while maximizing the production of crude methanol (i.e., no syngas is diverted for power production). The NGCCs are sized to meet the auxiliary power load requirements that are beyond the capabilities of the HRSG alone and eliminate the need for power from the local electrical grid for the coal cases. The performance estimates for the addition of the gas turbines is shown in Exhibit 3-6

Exhibit 3-1 Mass balance summary of coal to crude methanol plant

Stream		Case 1 and 2	
		Mole Flow lbmole/hr	Mass Flow lb/hr
Input	Wet coal	N/A	1,618,190
	Dried coal	N/A	1,277,850
	Oxygen from ASU containing 95% O_2	31,437	1,010,968
	Steam to gasifier	8,788	158,326
	Air for coal drying	15,000	434,322
	N_2 to coal drying	110,661	3,100,000
	Air to direct-fired boiler	4,197	121,518
	Total makeup water	310,943	5,596,974
Intermediate Products	Raw syngas from gasifier to scrubber	318,181	6,543,540
	Clean syngas for methanol production	95,912	1,183,080
	Fuel gas (for coal drying)	724	27,080
	Flash gas (for coal drying)	2,279	90,917
	Tail gas from Claus unit (for coal drying)	2,116	61,476
	Purge Gas (for coal drying and power Generation)	4,693	148,223
Final Products	Crude Methanol	29,476	941,823
	Sulfur (S_8)	45	11,576
	CO_2 (Case 2 only)	35,668	1,569,410
	Water discharge	124,933	2,248,798

Exhibit 3-2 Mass balance summary of natural gas to crude methanol plant

Stream		Case 3	
		Mole Flow lbmole/hr	Mass Flow lb/hr
Input	Natural Gas to reformer	33,685	583,677
	Oxygen from ASU at 95% O_2	21,464	682,554
	Steam to reformer	7,258	130,751
Intermediate Products	Clean syngas for methanol production	96,861	1,183,090
Final Products	Crude Methanol	29,382	940,989
	CO_2	5,358	235,808
	Water discharge	22,127	1,398,202

Exhibit 3-3 Stream compositions to and from methanol synthesis process

Case	1 and 2			3	
Feedstock	PRB coal			Natural Gas	
Description	Raw Syngas	HP Sweet Syngas To Methanol Plant	Raw Methanol Product	HP Sweet Syngas To Methanol Plant	Raw Methanol Product
Mole Fractions					
AR	0.0050	0.0098	0.0001	0.0048	0.0001
N_2	0.0055	0.0108	0.0000	0.0087	0.0000
CO	0.3492	0.3120	0.0000	0.3169	0.0000
CO_2	0.0362	0.0359	0.0116	0.0313	0.0116
H_2	0.1299	0.6284	0.0000	0.6218	0.0000
H_2O	0.4723	0.0031	0.0168	0.0031	0.0168
CH_4	0.0001	0.0001	0.0000	0.0134	0.0000
CH_3OH	0.0000	0.0000	0.9711	0.0000	0.9711
C_2H_6O	0.0000	0.0000	0.0002	0.0000	0.0002
C_3H_8O	0.0000	0.0000	0.0001	0.0000	0.0001
TOTAL	1.0000	1.0000	1.0000	1.0000	1.0000
Total Flow (lbmole/hr)	**318,181**	**95,912**	**29,317**	**96,861**	**29,317**
Total Flow (lb/hr)	**6,543,540**	**1,183,090**	**936,736**	**1,183,090**	**936,736**
Total Flow (ft^3/hr)	**5,882,680**	**957,495**	**19,523**	**966,372**	**19,523**
Temperature (°F)	**603.32**	**228.60**	**118.29**	**228.18**	**118.29**
Pressure (psia)	**605.34**	**755.00**	**380.00**	**755.00**	**380.00**
Density (lb/ft^3)	**1.11**	**1.24**	**47.98**	**1.22**	**47.98**
Average MW (lb/lbmole)	**20.57**	**12.34**	**31.95**	**12.21**	**31.95**
Molar Ratios					
H_2/CO_2	3.58	17.50	-	19.87	-
H_2/CO	0.37	2.01	-	1.96	-
$(H_2-CO_2)/(CO+CO_2)$	0.24	1.70	-	1.70	-

Exhibit 3-4 Auxiliary power load summary of CTM plants

Auxiliary Power Loads(kW)	Case 1	Case 2
Coal handling and milling	9,090	9,090
Slag handling	1,940	1,940
ASU	179,940	179,940
Syngas recycle compressor	6,600	6,600
Incinerator air blower	2,680	2,680
Direct-fired boiler air blower	310	310
Flash bottoms pump	720	720
Scrubber pumps	1,070	1,070
Rectisol auxiliary	51,270	51,270
Claus plant auxiliary	250	250
CO_2 compressor auxiliary	9000*	68,820
Syngas compressor	20,760	20,760
Recycle gas compressor	3,370	3,370
Water treatment	3,530	3,530
Air cooler fans	1,800	1,800
Circulating water pump	9,110	9,430
Boiler feedwater pump	1,500	1,500
Cooling tower fans	360	510
Steam turbine auxiliary	100	100
Miscellaneous BOP	5,000	5,000
Total Gasification + Methanol plants	308,400	368,690

**CO_2 used for coal transportation only.*

Exhibit 3-5 Auxiliary power load summary of NGTM plant

Auxiliary Power (kW)	Case 3
ASU	122,104
Direct-fired boiler air blower	310
Syngas compressor	20,760
Recycle gas compressor	3,370
Water treatment	3,530
Air cooler fans	1,800
Circulating water pump	9,110
Boiler feedwater pump	1,500
Cooling tower fans	360
Steam turbine auxiliary	100
Miscellaneous BOP	5,000
Total Reformer + Methanol plants	167,944

Exhibit 3-6 Performance summary of NGCC plants

Plant Output				
Parameter	**Case 1**	**Case 2**	**Case 3**	**Units**
Gas Turbine Power	64,800	113,700	105,000	kW_e
Steam Turbine Power	256,000	276,400	204,000	kW_e
Total	**320,800**	**390,100**	**309,000**	kW_e
Auxiliary Load				
Condensate Pumps	200	210	140	kW_e
Boiler Feedwater Pumps	3,550	3,910	3,160	kW_e
Amine System Auxiliaries	0	3,000	5,600	kW_e
NGCC CO_2 Compression	0	4,800	8,970	kW_e
Circulating Water Pump	3,350	3,730	3,450	kW_e
Ground Water Pumps	300	350	310	kW_e
Cooling Tower Fans	1,730	2,030	1,780	kW_e
SCR	10	10	10	kW_e
Gas Turbine Auxiliaries	700	700	700	kW_e
Steam Turbine Auxiliaries	100	100	100	kW_e
Miscellaneous Balance of Plant[2]	500	500	500	kW_e
Transformer Losses	1,840	2,140	1,410	kW_e
Total Methanol Synthesis Auxiliary Power	308,400	368,690	167,940	kW_e
Total	**320,680**	**390,170**	**194,070**	**kW_e**
Plant Performance				
Net Plant Power	**120**	**-70**	**114,930**	**kW_e**
Plant Capacity Factor	90%	90%	90%	
Combustion Turbine Fuel	Natural Gas	Natural Gas	Tail Gases	
Fuel Flow Rate	13,586 (29,953)	23,833 (52,543)	109,777 (242,0157)	kg/hr (lb/hr)
Thermal Input (HHV)[1]	197,940	347,226	1,599,358	kW_t
Thermal Input (LHV)[1]	178,469	313,069	1,442,027	kW_t
Condenser Duty	1,382 (1,310)	1,456 (1,380)	971 (920)	GJ/hr (MMBtu/hr)
Raw Water Withdrawal	42.1 (11,109)	42.1 (11,109)	12.9 (3,408)	m^3/min (gpm)
Raw Water Consumption	18.9 (4,994)	18.9 (4,994)	9.9 (2,612)	m^3/min (gpm)

[1] *Higher heating value (HHV) of Natural Gas 52,314 kJ/kg (22,491 Btu/lb)*
Lower heating value (LHV) of Natural Gas 47,220 kJ/kg (20,301 Btu/lb)

[2] *Includes plant control systems; lighting; heating, ventilation, and air conditioning (HVAC); and miscellaneous low voltage loads*

3.3 Process Descriptions

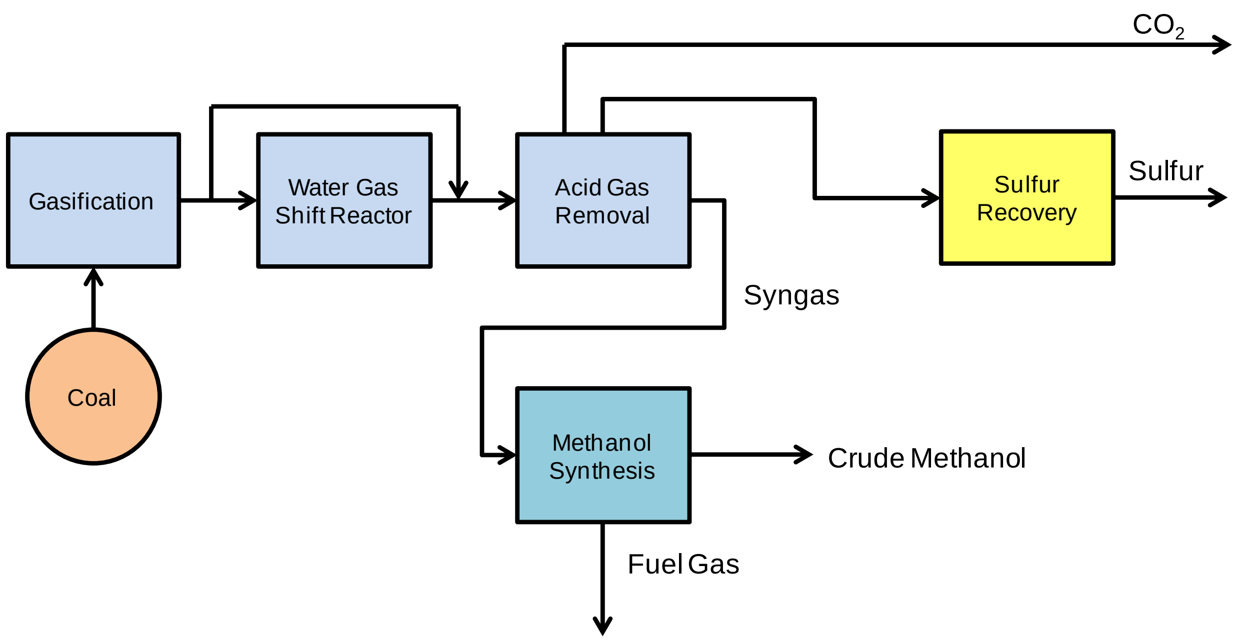
This section describes the crude methanol production process, divided into the functions (as "blocks") of each process step. The section includes process flow diagrams (PFD) and stream tables as referred to in the text. The diagrams and tables are grouped together at the end of this section.

3.3.1 General Process Descriptions

The conversion of coal-to-methanol is a two-step process: first conversion of the coal to the appropriate quality syngas via gasification and applying the water gas shift reaction, and second, catalytically converting the syngas to methanol.

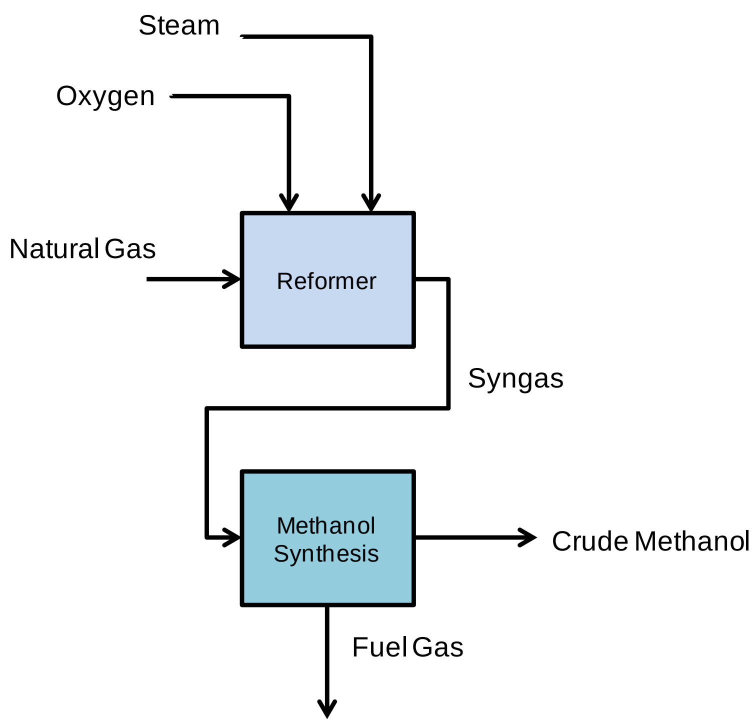
For the coal-to-methanol (CTM) cases 1 and 2, the plant configuration is illustrated in Exhibit 3-7. Syngas is generated from the gasification of PRB coal in a high-pressure, oxygen-blown Shell quench-type gasifier. The high temperature entrained-bed gasifier uses a partial water quench and syngas cooler to cool the hot syngas stream and generate steam for the water gas shift reactors and power generation. Crude raw syngas (post quench) from the gasification unit is scrubbed and split into two streams. The first stream is fed to a sour water gas shift reactor (WGSR) to increase the hydrogen content so that a H_2/CO molar ratio of 2:1 in the feed stream to the synthesis reactor can be achieved, and the second stream bypasses the WGSR. The streams are combined to achieve the desired composition. This partial bypass mode of operation allows the shift reactor to operate at a higher conversion ratio resulting in a smaller size. The syngas is cooled in a low-temperature heat-recovery system and then cleaned of mercury, sulfur, and CO_2 in preparation for methanol synthesis. Clean syngas is fed into a fixed-bed process to generate methanol.

Exhibit 3-7 Simplified process flow diagram of coal to crude methanol



For the natural gas to methanol (NGTM) case, the plant configuration is illustrated in Exhibit 3-8. Syngas is generated by combining natural gas, steam, and oxygen in an autothermal reformer (ATR). The reformed syngas is cooled to recover heat at a useful temperature and fed into a fixed-bed process to generate methanol. For this study, no additional cleanup of the syngas was included based on the assumption that contaminants were removed from the natural gas feedstock by the supplier. Site specific circumstances may require sulfur polishing and other cleanup processes in a more detailed design basis, but they were considered beyond the scope and accuracy of this study.

Exhibit 3-8 Simplified process flow diagram of natural gas to crude methanol



3.3.2 Air Separation Unit

All cases include an air separation unit (ASU) for generating oxygen. The ASU is a conventional, cryogenic, pumped liquid oxygen (LOX) unit that provides oxygen for the gasification and reforming processes, as well as nitrogen for ancillary equipment. The ASU is designed to produce 95 mole percent O_2 for use in the gasifier and Claus plant in the coal cases and for use in the reformer in the natural gas case. The air compressor is powered by an electric motor. Nitrogen is recovered and used as a diluent for coal drying in the coal cases and vented in the natural gas case.[4]

The ASU process for the coal fed cases is shown in Exhibit 3-11 with the gasification, quench, and dry solid removal processes. Cases 1 and 2 share a common PFD. The mass balances of this process for the two cases are presented by Exhibit 3-12. Since Case 2 uses the same specifications as Case 1, they both share a common mass balance in the exhibit. The stream numbers on the PFD correspond to the stream numbers in the mass balance tables. The ASU for

[4]In the natural gas cases, no disposition was assigned to the nitrogen from the ASU and no potential credit from its sale was applied to the process economics.

the natural gas case was not modeled, but the performance was estimated from similar ASU systems.

3.3.3 Coal Gasification Syngas Process Description

Cases 1 and 2 use coal gasification as the initial processing step. The gasifier block utilizes coal, oxygen, and steam to produce raw syngas. The following units in this subsection are included in this block.

3.3.3.1 Coal Milling, Grinding, and Drying

The coal drying PFD is shown in Exhibit 3-9. Case 1 and Case 2 share a common PFD. The mass balances of the coal drying process are presented in Exhibit 3-10. Since Case 2 uses the same specifications as Case 1, they both share a common mass balance in the exhibit. The stream numbers on the PFD correspond to the stream numbers in the mass balance tables.

The Shell process uses a dry feed system that is sensitive to the coal moisture content. Coal moisture consists of two parts: surface moisture and inherent moisture. For coal to flow smoothly through the lock hoppers, the surface moisture must be removed. The PRB coal used in this study contains 25.77 percent total moisture on an as-received basis. It was assumed that the coal must be dried to 6 percent moisture to allow for smooth flow through the dry feed system. The coal is simultaneously crushed and dried in the coal mill, then delivered to a surge hopper.

The drying heat is provided by burning the tail-gas from the Claus plant and the flash and purge gas from the methanol synthesis process in an incinerator. The hot incinerator flue gas mixes with N_2 from the ASU and exhaust is recycled from the mill to maintain a drying gas temperature of less than 500°F with oxygen content lower than 8 vol%. The dried coal is drawn from the surge hoppers and fed through a pressurization lock hopper system to a dense-phase pneumatic conveyor, which uses CO_2 to convey the coal to the gasifier. Using CO_2 rather than N_2 as the transport gas has the benefit that CO_2 is removed in the downstream acid gas removal (Rectisol) process, minimizing the buildup of inert species in the methanol generation recycle loop, thus reducing the size of the equipment needed. For this study, it was assumed that there was no impact of hot, concentrated CO_2 in the presence of moisture on standard materials of construction.

3.3.3.2 Gasification, Gas Quench, and Water Quench

The gasification and quench processes for the coal fed cases are shown in Exhibit 3-11. Cases 1 and 2 share a common PFD. The mass balances of this process for the two cases are presented by Exhibit 3-12. Since Case 2 uses the same specifications as Case 1, they both share a common mass balance in the exhibit. The stream numbers on the PFD correspond to the stream numbers in the mass balance tables.

Syngas is generated from the gasification of PRB coal in a high-pressure, oxygen-blown Shell quench-type gasifier. The high temperature entrained-bed gasifier uses a partial water quench and syngas cooler to cool the hot syngas stream and generate steam for the water gas shift reactors and power generation.

The coal feedstock is gasified in the presence of O_2 and superheated process steam. The O_2 requirement (O_2 required/dried feedstock) depends on the moisture content of the dried feedstock feeding the Shell gasifier. The coal is gasified at 2,550°F and 650 psia in a membrane wall reactor installed inside a pressure vessel, forming syngas, fly ash, and slag. The reactor design includes entrained flow, high temperature, recycled ash particulates, and slagging gasification and achieves carbon conversion greater than 99 percent.

The syngas leaving the gasifier is quenched to 850°F before entering a cyclone system for initial particulate removal described in the next section. The majority of the slag leaves the gasifier via the bottom as molten slag and is quenched and scattered to small glassy granulates in the slag (water) bath. In addition to the syngas and slag, the gasifier produces medium pressure saturated steam. Steam is produced from water that is circulated over the membrane wall of the gasifier to remove the heat of reaction and maintain an operating temperature of 2,550°F in the gasifier.

For this study, the design size requires eight operating trains with one spare train for a total of nine gasifiers. The facility contains one spare gasifier train to allow operation at a 90 percent capacity factor and to generally improve availability. The spare gasifier train feeds into the same gas clean-up trains as the other gasifier trains so that start-up/operation is transparent to downstream processes. The gasifier costs were scaled from previous studies which assumed two trains plus one spare train to achieve a high availability. [6,7,15]

3.3.3.3 Dry Solids Removal and Wet Scrubbing

The dry solid removal processes are shown in Exhibit 3-11 with the gasification and quench processes. Cases 1 and 2 share a common PFD. The mass balances of this process for the two cases are presented by Exhibit 3-12. Since Case 2 uses the same specifications as Case 1, they both share a common mass balance in the exhibit. The stream numbers on the PFD correspond to the stream numbers in the mass balance tables.

After passing through the water quench system, the syngas passes through a cyclone and a raw gas candle filter where a majority of the fine particles are removed and returned to the gasifier with the coal feedstock. Fines produced by the gasification system are recirculated to extinction. Final dust removal is achieved in the wet scrubbing section, to lower the dust content of the syngas to <1 mg/Nm³, and to lower its halide content to <1 ppmv. The wet scrubbing system consists of a Venturi scrubber followed by a packed bed wash column. A three percent by weight slurry bleed is fed to the primary wastewater treatment. Syngas is water saturated and leaves the wet scrubbing system at a temperature of 425°F.

3.3.3.4 Water Gas Shift

The water gas shift process is shown in Exhibit 3-13, where both coal cases share a common PFD. The material balances for the two cases are presented in Exhibit 3-14. Since Case 2 uses the same specifications as Case 1, they both share a common mass balance in the exhibit. The stream numbers on the PFD correspond to the stream numbers in the mass balance tables.

Coal-derived syngas from the wet scrubber enters the sour shift and cooling section. In order to achieve a 2:1 ratio of H_2 to CO in the final syngas, approximately 55 to 60 percent of the coal-derived syngas is shifted. The syngas to be shifted is heated in a countercurrent exchanger by the effluent of the first shift reactor to a temperature of 530°F. This syngas passes through the first stage shift reactor where the shift reaction exothermically converts CO and H_2O into H_2 and CO_2, leading to an outlet syngas temperature of 940°F. The syngas is then cooled to 495°F by a heat recovery steam generating exchanger followed by the counter current exchanger with the first stage feed stream.

The syngas/steam mixture passes through the second stage shift reactor where the shift reaction converts additional CO and H_2O into H_2 and CO_2, with a resulting second-stage shift reactor outlet syngas temperature of approximately 600°F. After cooling, the shifted syngas from the second-stage shift reactor outlet mixes with the bypass syngas and is sent to low-temperature gas cooling before being sent to the downstream Rectisol unit for removal of sulfur and CO_2.

3.3.3.5 Low-Temperature Gas Cooling

The low-temperature gas cooling process is shown in Exhibit 3-15 where both coal cases share a common PFD. The material balance for this process is presented in Exhibit 3-16. Since Case 2 uses the same specifications as Case 1, they both share a common mass balance in the exhibit. The stream numbers on the PFD correspond to the stream numbers in the mass balance tables.

Syngas is cooled in a number of steps to recover heat at useful temperatures. As the shifted syngas is cooled, process condensate, feed water, and a mix of clarified water and raw water are being heated.

The condensate from the air cooler and trim cooler knock-out drums absorbs nearly all of the NH_3 from the syngas. This condensate is mixed and sent to the process condensate stripper where low pressure steam is used to strip NH_3 and other absorbed gases from the condensate. The stripped gases are sent to the sulfur recovery unit to be treated with other sour gas streams.

The stripped condensate mixes with process condensate separated from the syngas. The mixed temperature is 325 °F. The process condensate is then pumped and heated to 390°F before being fed into the wet scrubbers.

3.3.3.6 Mercury Removal

The mercury removal process is shown in Exhibit 3-15 where both coal cases share a common PFD. The material balance for this process is presented by Exhibit 3-16. Since Case 2 uses the same specifications as Case 1, they both share a common mass balance in the exhibit. The stream numbers on the PFD correspond to the stream numbers in the mass balance tables.

Mercury removal is achieved by a packed bed of sulfur-impregnated activated carbon. The low-temperature syngas must be heated to 105°F prior to feeding it to the mercury-removal bed. The packed carbon bed vessels located upstream of the sulfur recovery unit with 20-second superficial gas residence time would achieve more than 90 percent removal of mercury (based on mercury content in the gasifier feedstock) in addition to removal of some portion of other volatile heavy metals such as arsenic. Mercury-removal systems using sulfur-impregnated activated carbon downstream of a coal gasifier have a reported bed life of 18 to 24 months, and usually replacement is required due to fouling of the bed rather than mercury saturation.

3.3.3.7 Acid Gas Removal and CO_2 Compression

The acid gas removal and CO_2 compression processes are shown in Exhibit 3-17, where Case 1 and Case 2 are shown in separate PFDs. Exhibit 3-19 represents the material balances of this process for the cases. Case 1 and Case 2 are shown in separate tables, as the recovered CO_2 in Case 1 is not compressed and sequestered. The stream numbers on the PFD correspond to the stream numbers in the mass balance tables.

A feature of this plant configuration is that H_2S and CO_2 are removed within the same process, via the Rectisol unit. The Rectisol acid gas removal (AGR) process was specified primarily because the methanol synthesis catalyst requires an H_2S level below 100 ppbv in order to maintain an adequate catalyst lifetime.

The Rectisol process uses chilled methanol as a solvent. Because of high vapor pressure of methanol, the process is operated at a temperature in the range of -30 to -100 °F. There are many possible process configurations for the Rectisol process depending on process requirements, product specifications, and scalability. In this study, the methanol solvent contacting the feed gas in the first stage of the absorber is stripped in two stages of flashing via pressure reduction. The acid gas leaving the first stage solvent regenerator is suitable for processing in a Claus plant. The regenerated solvent is virtually free of sulfur compounds but contains some CO_2. The second stage of absorption then removes the remaining CO_2 present. The rich solvent from the bottom of the second stage of the absorber is stripped in a steam-heated regenerator and returned to the top of the absorption column after cooling and refrigeration.

In Case 1, sufficient CO_2 for use in transporting the coal is compressed as needed; the remaining recovered CO_2 is vented. In Case 2, the low-pressure CO_2 stream recovered from the Rectisol unit is compressed to 2,200 psig in a multiple-stage, intercooled compressor to supercritical conditions, which is then ready for pipeline transport.

3.3.3.8 Sulfur Recovery Unit

The sulfur recovery unit is shown in Exhibit 3-18, where both coal cases share a common PFD. Exhibit 3-19 represents the material balances. Case 1 and Case 2 material balances are shown separately. The stream numbers on the PFD correspond to the stream numbers in the mass balance tables.

The purpose of the sulfur recovery unit is to treat the acid gas from the Rectisol unit and sour gas streams from the sour water strippers to make an effluent gas acceptable for venting to the atmosphere or burning.

Currently, the Claus process remains the mainstay for sulfur recovery. Conventional three-stage Claus plants, with indirect reheat and feeds with a high H_2S content, can approach greater than 98 percent sulfur recovery.

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

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

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

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

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

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

In this process, one-third of the H_2S is burned in the furnace with oxygen from the ASU to give sufficient SO_2 to react with the remaining H_2S. Since these reactions are highly exothermic, a waste heat boiler that recovers high-pressure steam follows the furnace. Sulfur is recovered in a condenser that follows the high-pressure steam recovery section. Low-pressure steam is raised in the condenser. The tail gas from the first condenser then goes to three catalytic conversion stages, where the remaining sulfur is recovered via the Claus reaction. Each catalytic stage consists of gas preheat, a catalytic reactor, and a sulfur condenser. The liquid sulfur goes to the sulfur pit, while the tail gas proceeds to the incinerator for coal drying.

3.3.4 Natural Gas / Steam Reforming

In case 3, a methane reformer is used to convert natural gas to a syngas suitable for methanol synthesis. Natural gas is combined with oxygen in the autothermal reformer (ATR). The ATR represents a process intensification of syngas production where partial oxidation of the feedstock provides the energy to drive the endothermic reforming of the feedstock to syngas. The reforming is accomplished through contacting the reaction mixture with a nickel supported on alumina catalyst. The ATR is fed 95 percent pure oxygen from the ASU; a purified oxygen feed was chosen to minimize the amount of inert gases introduced into the methanol synthesis loop. The ATR operates at 355.3 psia and 1,935ºF (24.5 bar and 1,057°C). Steam and oxygen feeds were adjusted to the reformer to obtain a H_2/CO ratio of approximately 2/1 and to provide enough steam to mitigate the risk of excessive coking. The reformed syngas is cooled to recover heat at a useful temperature. As the syngas is cooled, process condensate, feed water, and raw water are being heated. For this study, it was assumed that no additional cleanup of the syngas was necessary before the production of crude methanol. Sulfur polishing and other cleanup processes may be required in a more detailed design, but they were considered beyond the scope and accuracy of this study.

3.3.5 Methanol Reactor and Synthesis Loop - All Cases

The Methanol Synthesis Loop PFD is shown in Exhibit 3-22, where all the cases share a common PFD and the material balances for the cases are shown in Exhibit 3-23. Since Case 2 uses the same specifications as Case 1, they both share a common mass balance in the exhibit while the natural gas-case balance is shown on the second page. The stream numbers on the PFD correspond to the stream numbers in the mass balance tables.

There are two routes for the production of methanol: vapor phase and liquid phase. A vapor-phase methanol process was chosen over a liquid-phase methanol process due to the breadth of operating experience with vapor-phase production units and the lack of commercial operating experience with liquid-phase methanol production. [16] However, the advent of mega- methanol projects in stranded gas fields may rapidly provide the experience curve to justify revisiting this design choice.

The methanol reactor converts hydrogen and carbon monoxide to methanol. The reactor is a catalytic packed-bed reactor. The primary side reactions produce ethanol, propanol, and formaldehyde. Acetone and acetaldehyde are also common impurities in the methanol product and are captured in this analysis.

CO_2-lean syngas with a H_2/CO ratio of 2:1 from the AGR process or from the natural gas reformer is compressed from 490 psia to the synthesis loop operating pressure of 755 psia in the syngas compressor. The compressed syngas is mixed with the recycled gas, heated to 400ºF, and routed to the methanol reactor. The reactor is steam cooled to facilitate near isothermal operation at 475ºF and 735 psia. In-line blowers, coolers, and knock-out drums are used within the synthesis loop to maintain pressure and remove crude methanol.

Because the methanol synthesis reaction is equilibrium limited, in order to promote continued production of products, methanol reactor effluent is cooled to condense out the product crude methanol that is removed in a phase separator. Ninety-six percent of the separated gas is compressed to reactor pressure and readmitted along with fresh syngas to the methanol reactor. This recycling elevates the overall conversion of carbon, overcoming the low per-pass conversion of CO (60 percent). A small purge-gas stream (approximately four percent) is taken to limit the build-up of inert gas species. In the coal cases, the exhaust gas is burned in the coal-drying process where some heat is recovered as steam for power generation. In the natural gas case the exhaust gas is combusted in the combustion turbine for power generation.

3.3.6 Heat Recovery and Power Generation

The power cycle heat and material balance diagrams for all cases are shown in Exhibit 3-24. Both the gasification and the methanol synthesis processes generate a large amount of heat that can be recovered. The recovered heat is used to generate steam for process requirements and power generation. The process steam is generated at three different pressure levels. Additional power is generated by adding a small (50 to 100 MW) gas turbine for combined cycle power generation in each of the cases. The gas turbine in the natural gas case is fed process tail gases only; but the coal cases, which use the tail cases for coal drying, are fed natural gas. The total power generated is designed to equal the total estimated auxiliary loads for the coal fed cases 1 and 2. Excess power is generated from the process heat recovered and tail gas combustion in the natural gas case. No additional natural gas is consumed in the power cycle for the natural gas case. The model performance results are discussed above in section 3.2.

3.3.7 Water Balance

Water required for the operation of the facility is obtained from a source such as a lake, river, or well. If the quality of the water is adequate, raw water is used directly as makeup to the cooling tower and the gasifier quench. To meet the rest of the plant's water needs, makeup must be treated first by filtration to create service-quality water. This quality of water serves as makeup to the plant's potable water, demineralized water, fire water, and service water systems. Higher quality boiler feedwater is treated by a typical reverse osmosis and electrodeionization package. Water rejected by the system is of an acceptable quality to be used as makeup to the cooling tower.

In addition to meeting the makeup water needs of the facility, water treatment systems must be capable of capturing and treating on-site waste streams. Wastewater created by the gasification process must pass through a number of pretreatment steps before being combined with other wastewater streams. Metals, ammonia, and suspended solids are removed from the stream through the use of a clarifier and a biological treatment unit. Once processed, the wastewater can be combined with the cooling tower blowdown as well as other plant waste streams in a final clarifier. Dechlorination and pH adjustment are performed as needed at this step of the process in order to meet all local discharge regulations. Solids separated out in this process are dried by means of a filter press and taken away for offsite disposal.

Exhibit 3-25 shows the water balances for the coal cases. Since Case 2 uses the same specifications as Case 1, they both share a common mass balance in the exhibit. The water usage for the natural gas case was not modeled, but the water consumption and cooling requirements were estimated from similar systems.

Exhibit 3-9 Coal drying process flow diagram
Case 1 and Case 2

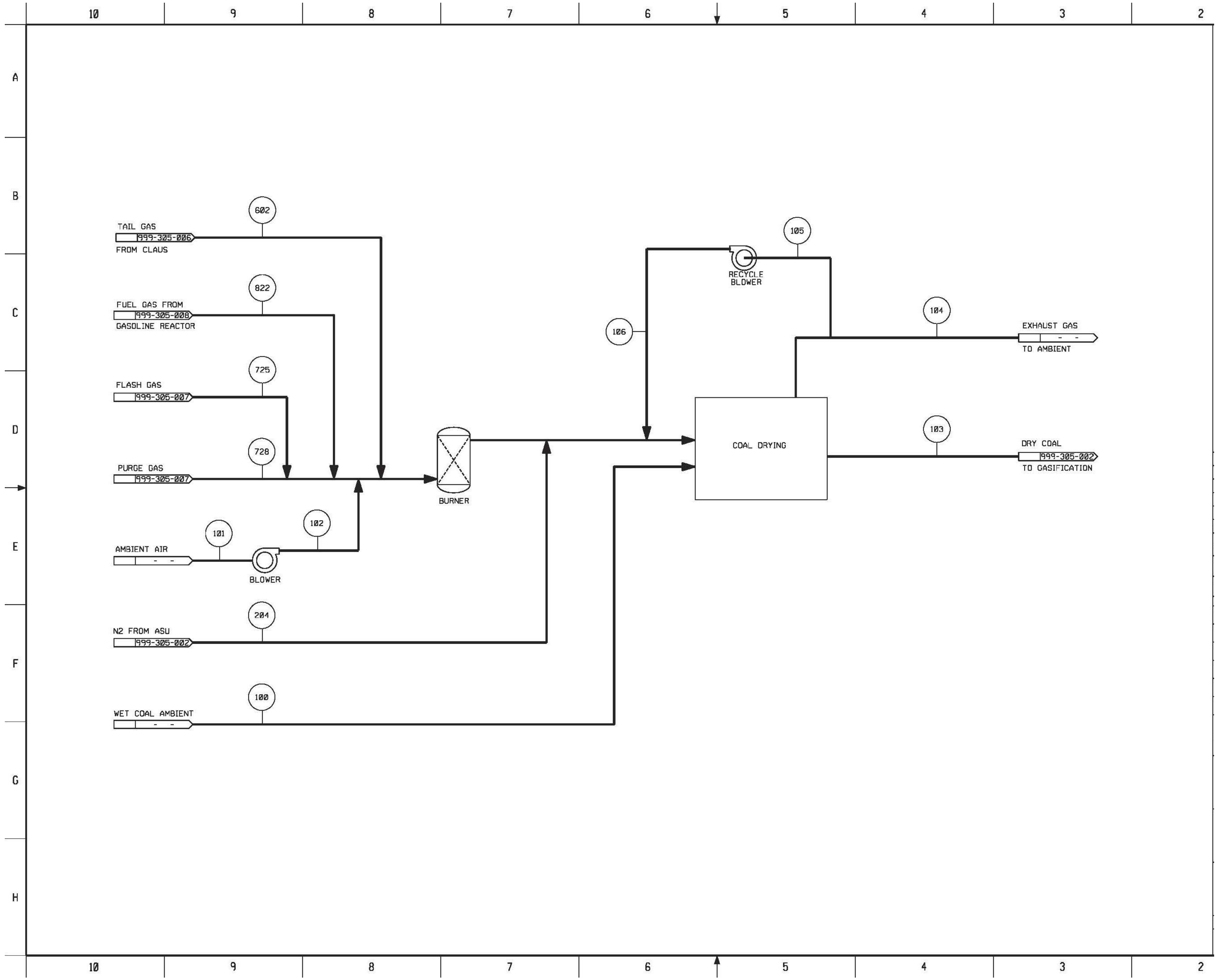


Exhibit 3-10 Mass balance of coal drying process
Case 1 and Case 2 – coal drying

STREAM		100	101	102	103	104	105	106
Description		Wet Coal	Ambient Air	Feed Air	Dry Coal	Exhaust Gas	LP Recycle Gas	HP Recycle Gas
Mole Flow (Vapor/Liquid)	lbmol/hr							
AR		0	139	139	0	418	147	147
CH_4		0	11,711	11,711	0	122,651	43,094	43,093
CO		0	3,143	3,143	0	462	162	162
CO_2		0	0	0	0	0	0	0
H_2S		0	0	0	0	4,586	1,611	1,611
COS		0	0	0	0	0	0	0
HCN		0	7	7	0	22,890	8,043	8,042
H_2		0	0	0	0	0	0	0
H_2O		0	0	0	0	0	0	0
N_2		0	0	0	0	1	0	0
NH_3		0	0	0	0	0	0	0
SO_2		0	0	0	0	5	2	2
Mass Flow (solids)	lb/hr							
COAL		1,618,190			1,277,850			
Total Flow	lbmol/hr	0	15,000	15,000	0	151,013	53,059	53,058
Total Flow	lb/hr	1,618,190	434,322	434,322	1,277,850	4,081,910	1,434,180	1,434,160
Total Flow	ACFH	0	5,679,650	5,200,550	0	67,445,300	23,697,000	22,769,000
Temperature	°F	59.00	59.00	89.56	157.00	164.60	164.60	180.15
Pressure	psia	14.70	14.70	17.00	15.00	15.00	15.00	16.00
Density	lb/ft³	0.00	0.08	0.08	0.00	0.06	0.06	0.06
Average MW	lb/lbmol	0.00	28.95	28.95	0.00	27.03	27.03	27.03

Exhibit 3-11 Shell gasification process flow diagram
Case 1 and Case 2

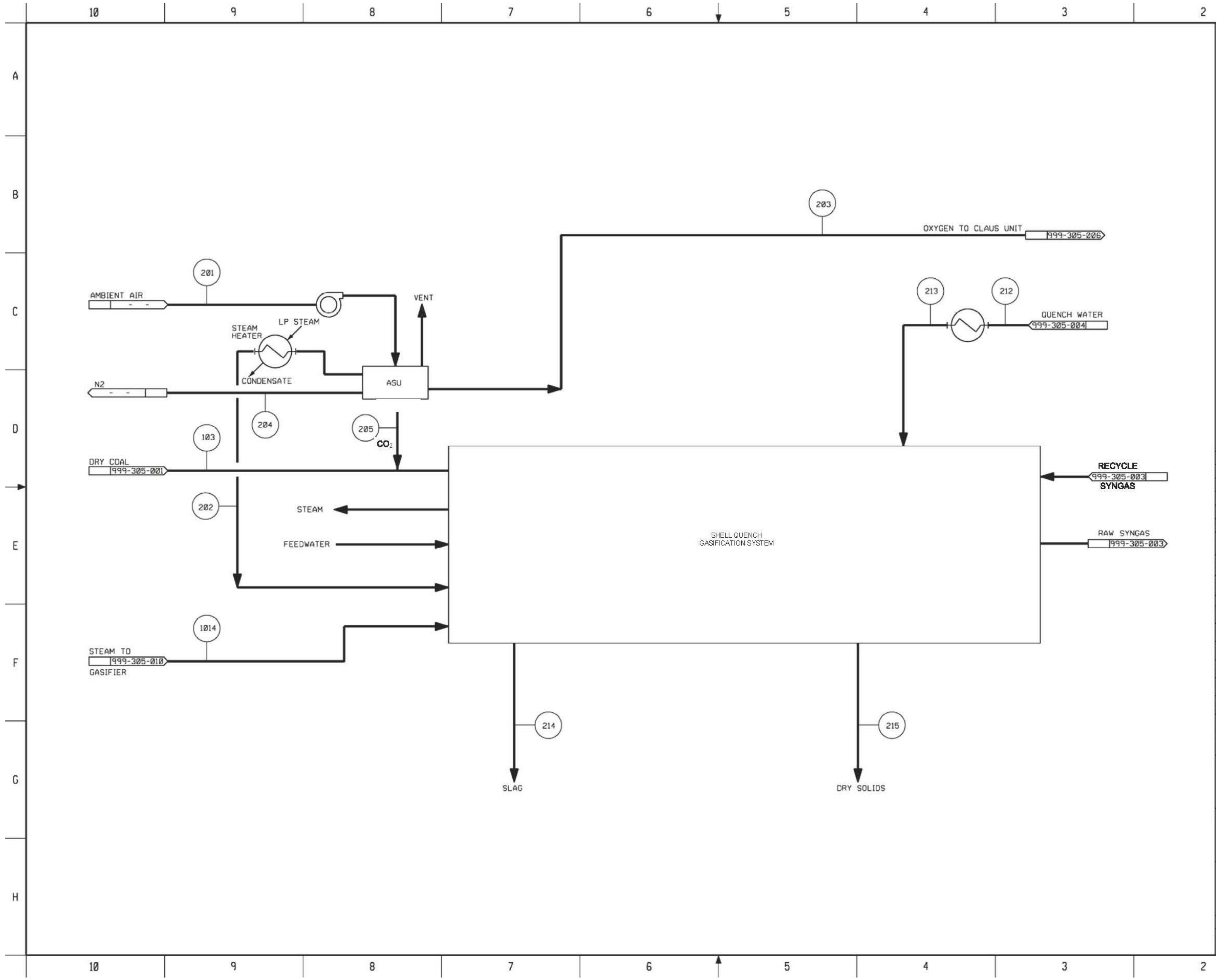


Exhibit 3-12 Mass balance of gasification process
Case 1 and Case 2

STREAM		201	202	203	204	205	209	212	213	214	215
Description		ASU Air	Hot O2	Claus O2	Coal Drying N2	Gasifier CO2	Raw Syngas	Cold Quench Water	Hot Quench Water	Slag	Dry Solids
Mole Flow (Vapor/Liquid)	lbmol/hr										
AR		1,332	938	5	0	0	1,580	0	0	0	0
CH4		0	0	0	0	0	10	0	0	0	0
CO		0	0	0	0	37	111,095	0	0	0	0
CO2		47	0	0	0	5,564	11,503	0	0	0	0
H2S		0	0	0	0	0	571	0	0	0	0
COS		0	0	0	0	0	51	0	0	0	0
HCN		0	0	0	0	0	2	0	0	0	0
H2		0	0	0	0	10	41,343	0	0	0	0
H2O		1,422	0	0	0	0	150,261	61,508	61,508	0	0
N2		111,363	626	3	110,661	0	1,743	0	0	0	0
NH3		0	0	0	0	0	16	0	0	0	0
O2		29,874	29,714	151	0	0	0	0	0	0	0
SO2		0	0	0	0	0	0	0	0	0	0
Mass Flow (solids)	lb/hr										
ASH		0	0	0	0	0	0	0	0	0	26,506
SLAG		0	0	0	0	0	0	0	0	84,949	0
Total Flow	lbmol/hr	144,038	31,278	159	110,661	5,611	318,181	61,508	61,508	0	0
Total Flow	lb/hr	4,156,505	1,005,840	5,128	3,100,000	245,922	6,543,540	1,108,080	1,108,080	84,949	26,506
Total Flow	ACFH	54,498,694	244,402	61,865	34,944,900	51,612	5,882,680	17,725	20,469	0	0
Temperature	°F	59	300.00	72.14	70.00	269.76	603.32	59.00	392.00	2,650.00	612.32
Pressure	psia	14.70	710.88	14.70	18.00	768.89	605.34	700.00	685.00	650.00	630.00
Density	lb/ft³	0.08	4.12	0.08	0.09	4.76	1.11	62.52	54.13	0.00	0.00
Average MW	lb/lbmol	28.86	32.16	32.16	28.01	43.83	20.57	18.02	18.02	0.00	0.00

Exhibit 3-13 CO shift reaction process flow diagram
Case 1 and Case 2

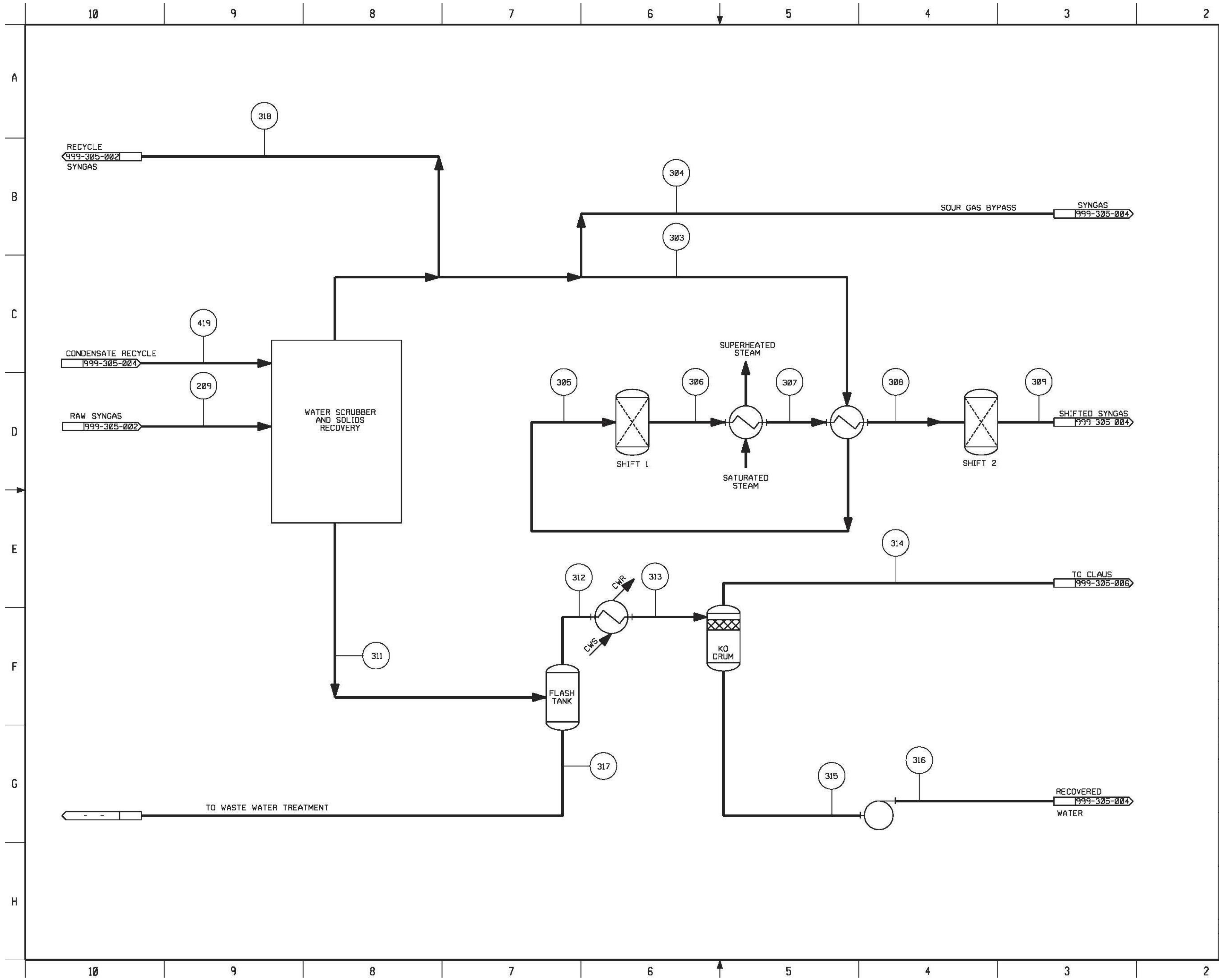


Exhibit 3-14 Mass balance of shift reaction process
Case 1 and Case 2

STREAM		303	304	305	306	307	308	309	311	312	313	314	315	316	317	318
Description		Syngas to Shift	Bypass Syngas	Hot Syngas to Shift	Hot 1st Shift Syngas	Warm 1st Shift Syngas	Cold 1st Shift Syngas	Hot 2nd Shift Syngas	Scrubber Water	Scrubber Hot Vapor	Scrubber Cold Vapor	Scrubber Sour Gas	LP Recycle Water	HP Recycle Water	Bottoms to WWT	Recycle Syngas
Mole Flow (Vapor/Liquid)	lbmol/hr															
AR		573	366	573	573	573	573	573	0	0	0	0	0	0	0	642
CH_4		4	2	4	4	4	4	4	0	0	0	0	0	0	0	4
CO		40,254	25,721	40,254	12,003	12,003	12,003	4,457	9	9	9	9	0	0	0	45,111
CO_2		4,192	2,679	4,192	32,460	32,460	32,460	40,007	5	5	5	0	0	0	0	4,698
H_2S		208	133	208	225	225	225	226	1	1	1	0	0	0	0	233
COS		18	12	18	2	2	2	0	0	0	0	0	0	0	0	21
HCN		1	1	1	0	0	0	0	0	0	0	0	0	0	0	1
H_2		14,984	9,575	14,984	43,236	43,236	43,236	50,782	5	5	5	5	0	0	0	16,793
H_2O		65,562	41,892	65,562	37,293	37,293	37,293	29,745	62,790	9,511	9,511	1	9,502	9,502	53,279	73,473
N_2		632	404	632	632	632	632	632	0	0	0	0	0	0	0	709
NH_3		13	8	13	14	14	14	14	1	0	0	0	0	0	0	14
O_2		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO_2		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mass Flow (solids)	lb/hr															
ASH		0	0	0	0	0	0	0	8,985	0	0	0	0	0	8,985	0
Total Flow	lbmol/hr	126,440	80,792	126,440	126,440	126,440	126,440	126,440	62,818	9,531	9,531	24	9,502	9,502	53,288	141,699
Total Flow	lb/hr	2,572,400	1,643,710	2,572,400	2,572,400	2,572,400	2,572,400	2,572,400	1,140,895	171,837	171,837	651	171,186	171,186	969,056	2,882,840
Total Flow	ACFH	1,863,100	1,190,480	2,163,830	3,206,240	2,436,760	2,211,220	2,532,320	21,323	1,465,660	5,299	3,003	2,781	2,777	16,591	2,087,930
Temperature	°F	415.31	415.31	530.00	938.81	600.00	495.16	606.40	415.31	281.93	130.00	130.00	130.00	130.96	281.93	415.31
Pressure	psia	605.34	605.34	600.34	593.09	584.39	575.69	567.69	605.34	50.76	50.76	50.76	50.76	550.00	50.76	605.34
Density	lb/ft³	1.38	1.38	1.19	0.80	1.06	1.16	1.02	53.51	0.12	32.43	0.22	61.56	61.64	58.41	1.38
Average MW	lb/lbmol	20.34	20.34	20.34	20.34	20.34	20.34	20.34	0.00	18.03	18.03	26.60	18.02	18.02	0.00	1.00

Exhibit 3-15 Low temperature gas cooling process flow diagram
Case 1 and Case 2

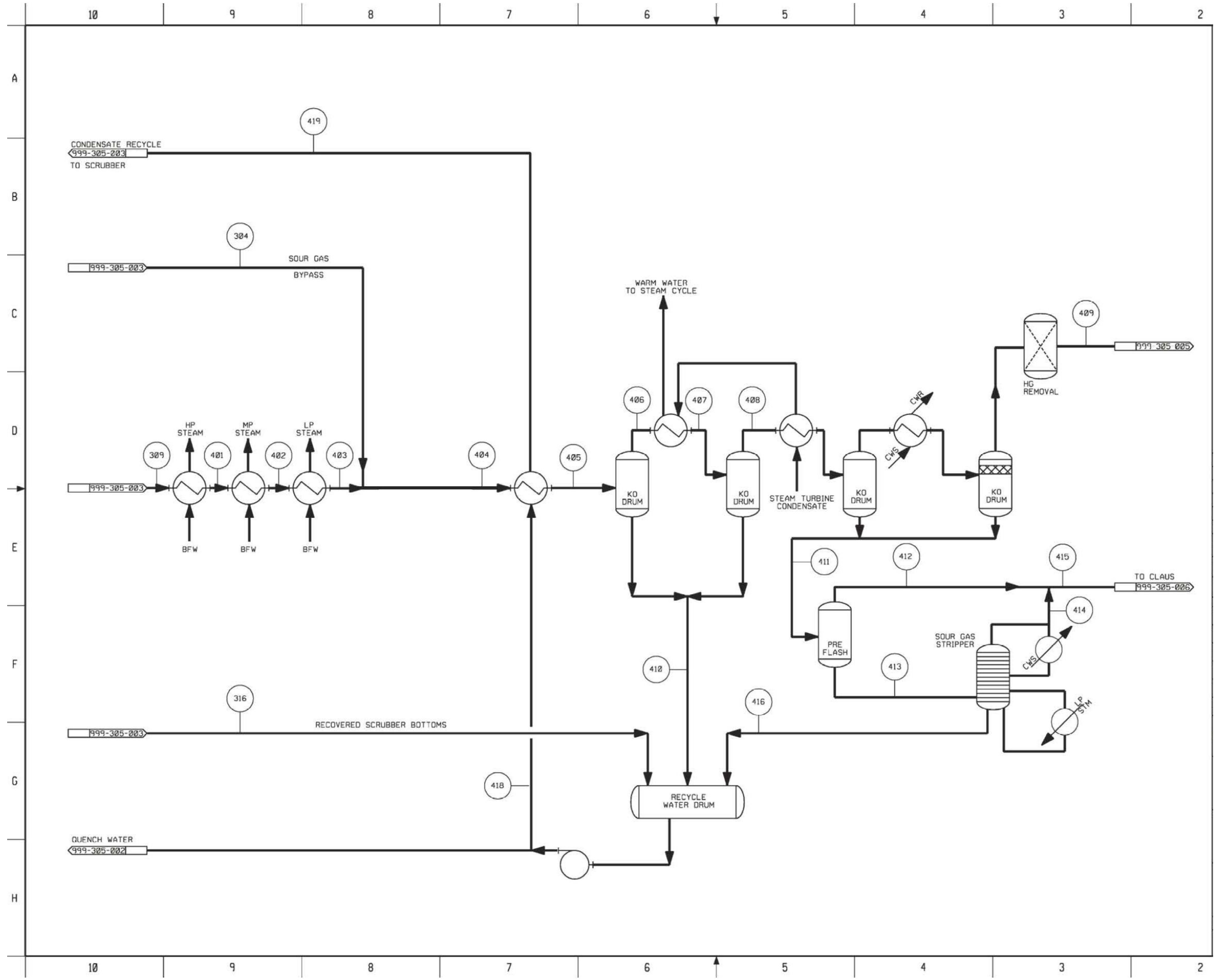


Exhibit 3-16 Mass balance of low temperature gas cooling
Case 1 and Case 2

STREAM		401	402	403	404	405	406	407	408	409
Description		Shifted Syngas	Shifted Syngas	Shifted Syngas	Shifted Syngas	Shifted Syngas	Shifted Syngas	Shifted Syngas	Shifted Syngas	Shifted Syngas
Mole Flow (Vapor/Liquid)	lbmol/hr									
AR		573	573	573	938	938	938	938	938	938
CH_4		4	4	4	6	6	6	6	6	6
CO		4,457	4,457	4,457	30,178	30,178	30,177	30,177	30,175	30,173
CO_2		40,007	40,007	40,007	42,686	42,686	42,671	42,667	42,640	42,568
H_2S		226	226	226	359	359	358	358	358	356
COS		0	0	0	12	12	12	12	12	12
HCN		0	0	0	1	1	1	1	1	0
H_2		50,782	50,782	50,782	60,356	60,356	60,352	60,352	60,347	60,342
H_2O		29,745	29,745	29,745	71,637	71,637	50,570	50,566	24,073	313
N_2		632	632	632	1,037	1,037	1,037	1,037	1,036	1,036
NH_3		14	14	14	22	21	20	16	15	0
O_2		0	0	0	0	0	0	0	0	0
SO_2		0	0	0	0	0	0	0	0	0
Total Flow	lbmol/hr	126,440	126,440	126,440	207,232	207,232	186,142	186,138	159,600	135,745
Total Flow	lb/hr	2,572,400	2,572,400	2,572,400	4,216,110	4,216,110	3,835,830	3,835,830	3,356,910	2,925,330
Total Flow	ACFH	2,376,400	2,277,470	2,087,460	3,389,870	3,172,460	2,904,840	2,401,550	2,393,160	1,569,710
Temperature	°F	536.00	489.29	409.06	406.45	366.96	350.00	305.78	305.78	110.63
Pressure	psia	562.69	557.69	552.69	552.69	549.69	543.24	540.34	540.34	515.83
Density	lb/ft³	1.08	1.13	1.23	1.24	1.33	1.32	1.60	1.40	1.86
Average MW	lb/lbmol	20.34	20.34	20.34	20.34	20.34	20.61	20.61	21.03	21.55

STREAM		410	411	412	413	414	415	416	418	419
Description		Knock Out Water	Sour Water	Sour Flash Gas	Sour Stripper Feed	Sour Stripper Gas	Sour Gas	Sour Stripper Bottoms	Cold Condensate	Hot Condensate to Scrubber
Mole Flow (Vapor/Liquid)	lbmol/hr									
AR		0	0	0	0	0	0	0	0	0
CH_4		0	0	0	0	0	0	0	0	0
CO		3	2	0	2	2	2	0	3	3
CO_2		41	57	0	57	57	57	0	46	55
H_2S		1	2	0	2	0	0	1	2	2
COS		0	0	0	0	0	0	0	0	0
HCN		0	0	0	0	0	0	0	0	0
H_2		10	5	0	5	5	5	0	10	10
H_2O		47,560	23,745	0	23,745	216	216	23,529	80,595	80,604
N_2		0	0	0	0	0	0	0	0	0
NH_3		2	0	0	0	0	0	0	6	15
O_2		0	0	0	0	0	0	0	0	0
SO_2		0	0	0	0	0	0	0	0	0
Total Flow	lbmol/hr	47,627	23,840	0	23,840	280	280	23,560	80,694	80,703
Total Flow	lb/hr	859,195	431,577	7	431,569	6,472	6,480	425,097	1,455,480	1,455,480
Total Flow	ACFH	15,248	7,160	4	7,156	35,539	35,579	7,445	25,365	26,938
Temperature	°F	327.11	178.95	178.95	178.95	276.97	276.97	307.02	299.48	392.00
Pressure	psia	540.34	524.53	524.53	524.53	61.11	61.11	74.16	768.89	761.64
Density	lb/ft³	56.35	60.28	1.71	60.31	0.18	0.18	57.10	57.38	54.03
Average MW	lb/lbmol	18.04	18.10	22.12	18.10	23.12	23.11	18.04	18.04	18.04

Exhibit 3-17 Rectisol process flow diagram
Case 1

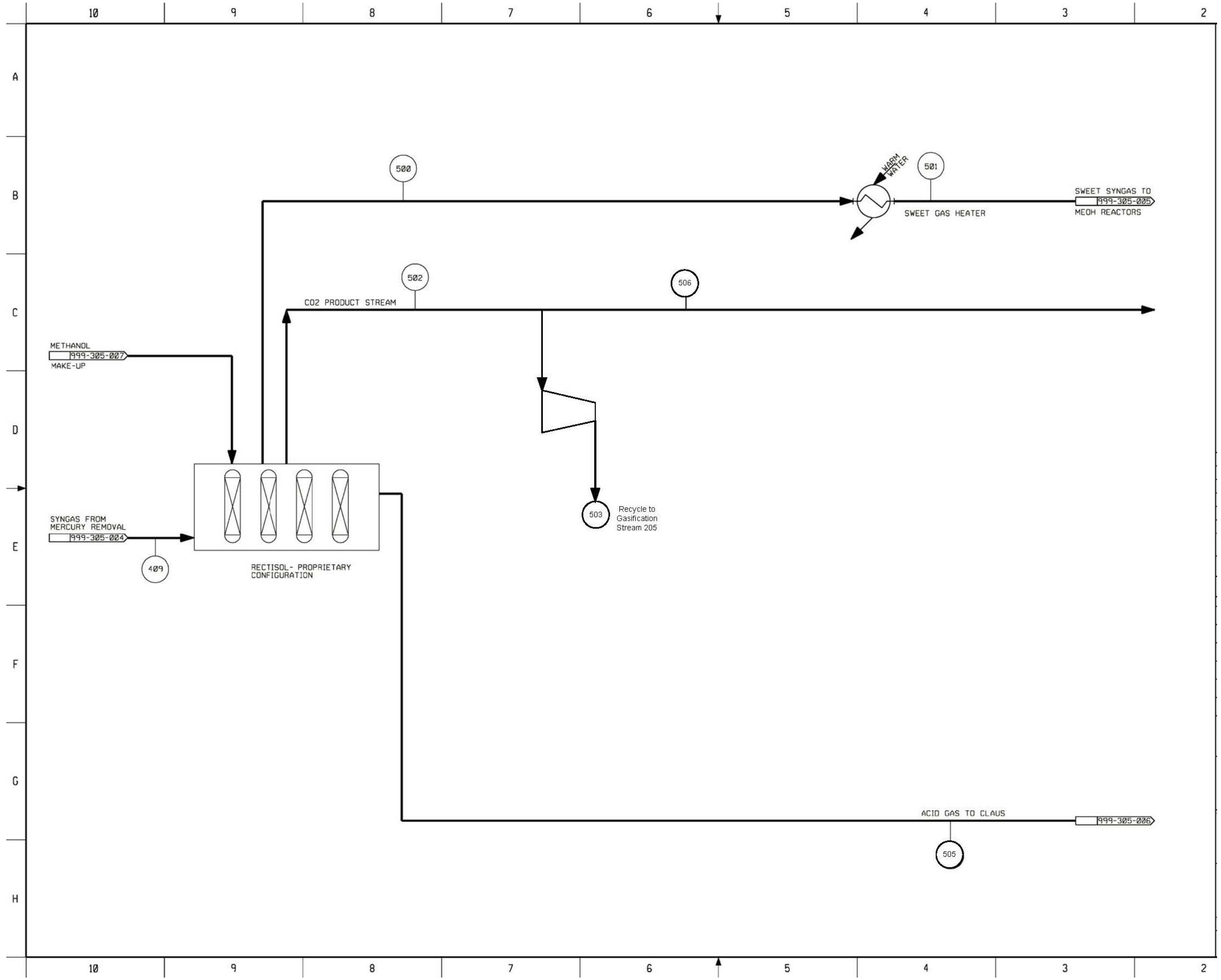


Exhibit 3-17 Rectisol process flow diagram (continued)
Case 2

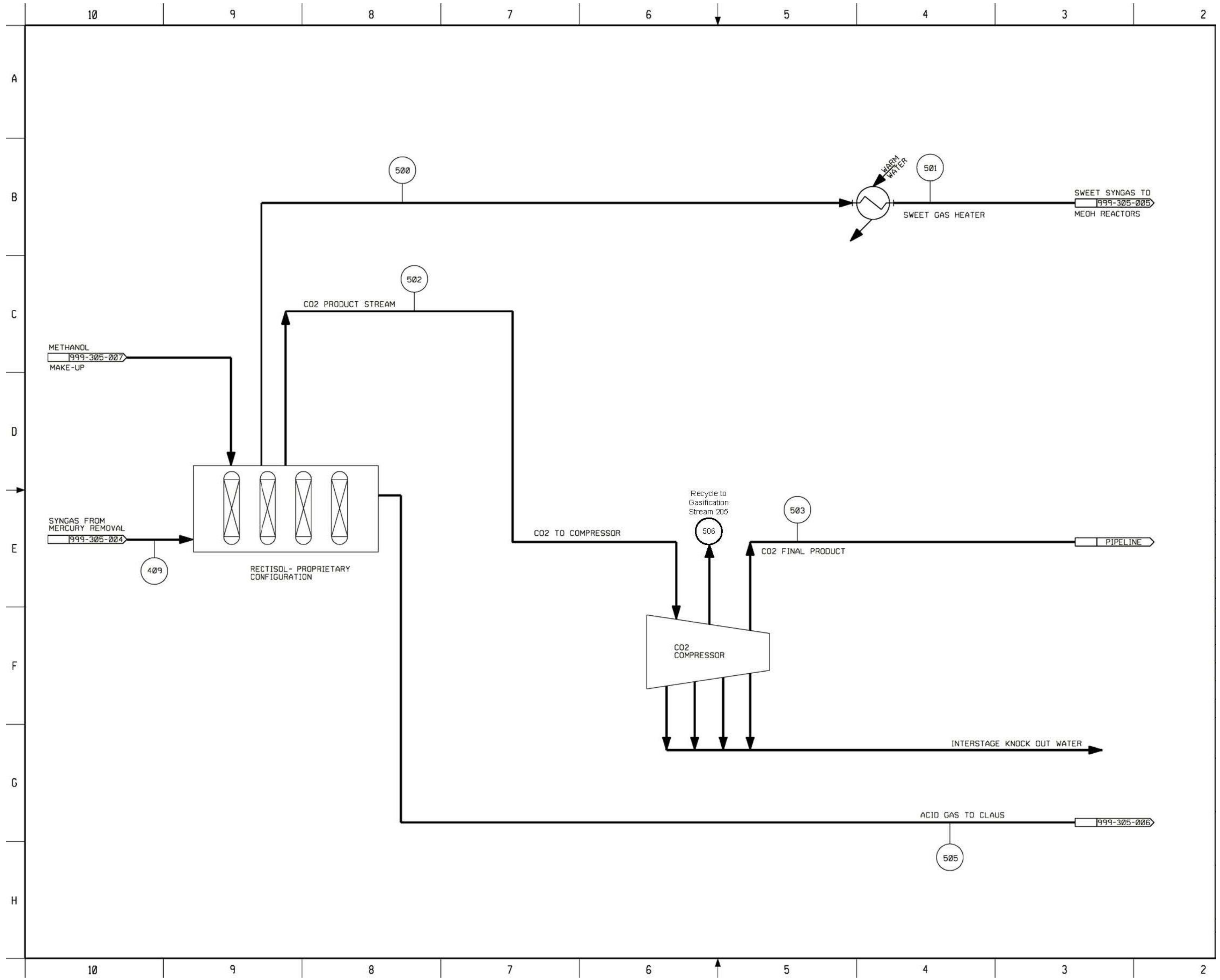


Exhibit 3-18 Claus process flow diagram
Case 1 and Case 2

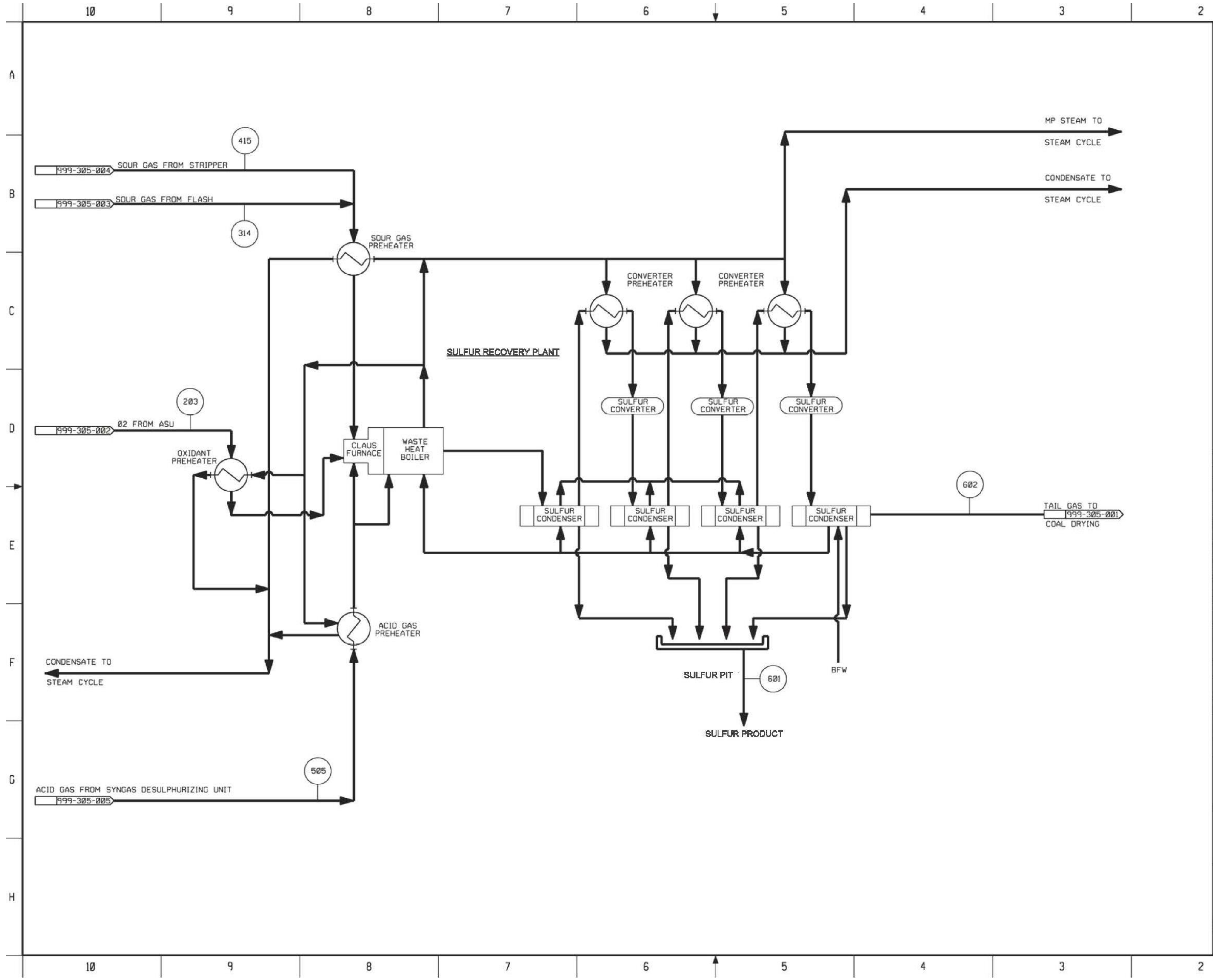


Exhibit 3-19 Mass balance of Rectisol and Claus processes
Case 1

STREAM		500	501	502	503	504	505	506		601	602
Description		Cold Sweet Syngas	Hot Sweet Syngas	LP CO_2	IP CO_2	KO Water	Acid Gas to Claus	CO_2 Vent		Sulfur Product	Tail Gas to Coal Drying
Mole Flow (Vapor/Liquid)	lbmol/hr										
AR		938	938	0	0	0	0	0		0	5
CH_4		6	6	0	0	0	0	0		0	0
CO		29,920	29,920	251	37	0	1	215		0	65
CO_2		3,443	3,443	38,219	5,564	0	906	32,655		0	925
H_2S		0	0	0	0	0	356	0		0	4
COS		0	0	0	0	0	12	0		0	1
HCN		0	0	0	0	0	0	0		0	0
H_2		60,270	60,270	71	10	0	1	60		0	23
H_2O		298	298	0	0	0	16	0		0	582
N_2		1,036	1,036	0	0	0	0	0		0	4
NH_3		0	0	0	0	0	0	0		0	0
O_2		0	0	0	0	0	0	0		0	0
SO_2		0	0	0	0	0	0	0		0	2
Mass Flow (solids)	lb/hr										
Sulfur		0	0	0	0	0	0	0		11,576	0
Total Flow	lbmol/hr	95,912	95,912	38,541	5,611	0	1,292	32,930		45	1,611
Total Flow	lb/hr	1,183,080	1,183,080	1,689,200	245,922	0	53,049	1,443,280		11,576	53,731
Total Flow	ACFH	1,096,130	1,225,740	14,650,000	51,483	0	257,856	12,517,700		35	671,652
Temperature	°F	68.00	125.00	57.20	269.76	100	86.00	56.88		387.64	320.00
Pressure	psia	495.83	493.83	14.50	769.00	15	29.01	14.50		20.01	20.01
Density	lb/ft³	1.08	0.97	0.12	4.78	61	0.21	0.12		327.05	0.08
Average MW	lb/lbmol	12.34	12.34	43.83	43.83	18	41.06	43.83		256.53	33.34

Exhibit 3-19 Mass balance of Rectisol and Claus processes (continued)
Case 2

STREAM		500	501	502	503	504	505	506		601	602
Description		Cold Sweet Syngas	Hot Sweet Syngas	LP CO_2	HP CO_2	Knock Out Water	Acid Gas to Claus	IP CO_2		Sulfur Product	Tail Gas to Coal Drying
Mole Flow (Vapor/Liquid)	lbmol/hr										
AR		938	938	0	0	0	0	0		0	5
CH_4		6	6	0	0	0	0	0		0	0
CO		29,920	29,920	251	215	0	1	37		0	65
CO_2		3,443	3,443	38,219	32,655	0	906	5,564		0	925
H_2S		0	0	0	0	0	356	0		0	4
COS		0	0	0	0	0	12	0		0	1
HCN		0	0	0	0	0	0	0		0	0
H_2		60,270	60,270	71	60	0	1	10		0	23
H_2O		298	298	0	0	0	16	0		0	582
N_2		1,036	1,036	0	0	0	0	0		0	4
O_2		0	0	0	0	0	0	0		0	0
SO_2		0	0	0	0	0	0	0		0	2
Mass Flow (solids)	lb/hr										
Sulfur		0	0	0	0	0	0	0		11,576	0
Total Flow	lbmol/hr	95,912	95,912	38,541	32,930	0	1,292	5,611		45	1,611
Total Flow	lb/hr	1,183,080	1,183,080	1,689,200	1,443,280	1	53,049	245,922		11,576	53,731
Total Flow	ACFH	1,096,130	1,225,740	14,650,000	52,523	0	257,856	51,483		35	671,652
Temperature	°F	68.00	125.00	57.20	166.80	100	86.00	269.76		387.64	320.00
Pressure	psia	495.83	493.83	14.50	2,214.70	15	29.01	769.00		20.01	20.01
Density	lb/ft³	1.08	0.97	0.12	27.48	61	0.21	4.78		327.05	0.08
Average MW	lb/lbmol	12.34	12.34	43.83	43.83	18	41.06	43.83		256.53	33.34

Exhibit 3-20 Natural gas reforming process flow diagram
Case 3

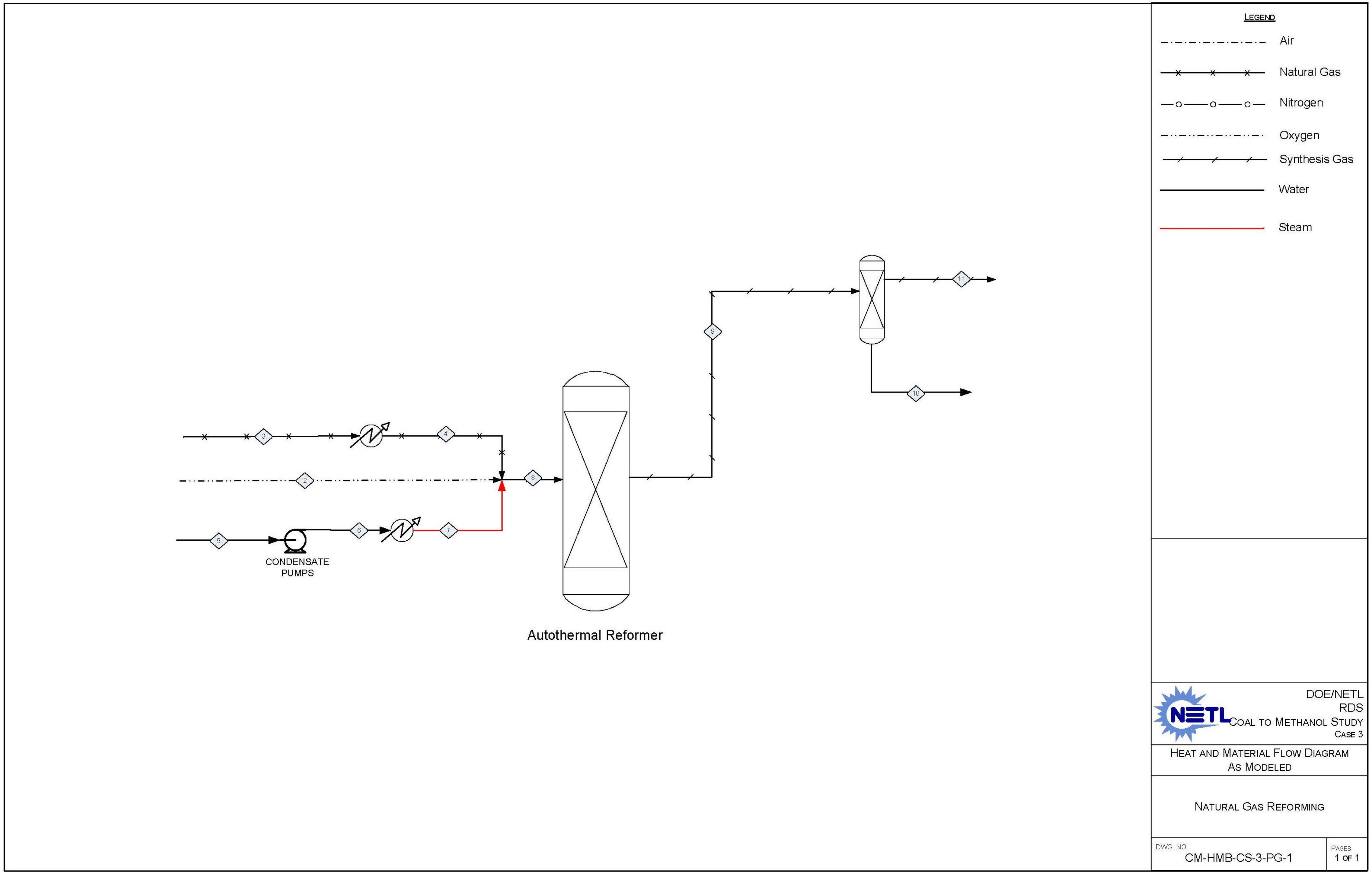


Exhibit 3-21 Mass balance of natural gas reforming
Case 3

STREAM	-	2	3	4	5	6	7	8	9	10	11
Description	ASU Air	O_2	Natural Gas Feed	Warm Natural Gas Feed	H_2O Feed	H_2O	Steam	Reformer Feed	Reformer Exit	Condensate	Raw Syngas
Mole Flow (Vapor/Liquid) lbmol/hr											
AR	905	0	0	0	0	0	0	0	0	0	0
CH_4	0	0	31,361	31,361	0	0	0	31,361	1,301	0	1,301
CO	0	0	0	0	0	0	0	0	30,758	0	30,758
CO_2	0	0	337	337	0	0	0	337	3,041	0	3,041
C_2H_6	0	0	1,078	1,078	0	0	0	1,078	0	0	0
C_3H_8	0	0	236	236	0	0	0	236	0	0	0
$N\text{-}C_4H_{10}$	0	0	135	135	0	0	0	135	0	0	0
H_2	0	0	0	0	0	0	0	0	60,356	0	60,356
H_2O	44	0	0	0	7,258	7,258	7,258	7,258	11,873	11,873	0
N_2	75,990	1,073	539	539	0	0	0	1,612	1,612	0	1,612
O_2	20,391	20,391	0	0	0	0	0	20,391	0	0	0
Total Flow lbmol/hr	97,330	21,464	33,685	33,685	7,258	7,258	7,258	62,407	108,941	11,873	97,068
Total Flow lb/hr	2,818,166	682,554	583,677	583,677	130,751	130,751	130,751	1,396,980	1,396,980	213,898	1,183,080
Total Flow ACFH	36,853,318	469,137	421,602	1,900,690	2,089	2,089	405,066	3,088,390	7,908,740	3,437	7,972,290
Temperature °F	59	305	100	1,500	59	59	1,500	1,261	1,935	70	494
Pressure psia	14.70	375.00	450.00	375.00	14.70	375.00	375.00	375.00	355.34	14.70	125.00
Density lb/ft³	0.08	1.45	1.38	0.31	62.59	62.59	0.32	0.45	0.18	62.23	0.15
Average MW lb/lbmol	28.95	31.80	17.33	17.33	18.02	18.02	18.02	22.38	12.82	18.02	12.19

Exhibit 3-22 Methanol synthesis process flow diagram
Case 1, Case 2, and Case 3

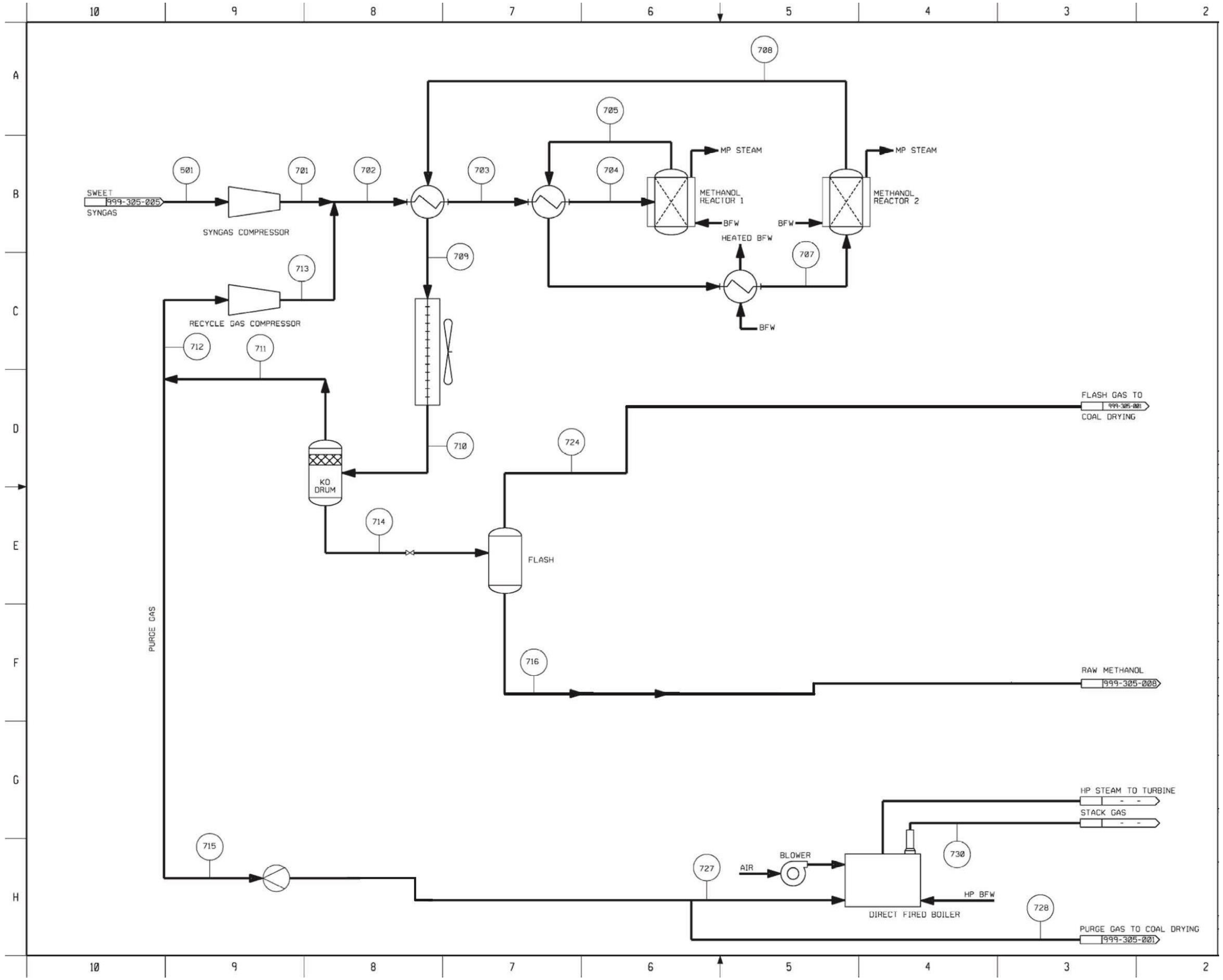


Exhibit 3-23 Mass balance of methanol synthesis
Case 1 and Case 2

STREAM		701	702	703	704	705	706	707	708	709	710	711	712	713
Description		HP Sweet Syngas	Cold Mixed Feed	Warm Mixed Feed	Hot Mixed Feed	Stage 1 Product	N/A	Stage 2 Cooled Feed	Hot Stage 2 Product	Warm Stage 2 Product	Cool Stage 2 Product	Flash Gas	LP Recycle Gas	HP Recycle Gas
Mole Flow (Vapor/Liquid)	lbmol/hr													
AR		938	20,449	20,449	20,449	20,449		20,449	20,449	20,449	20,449	20,324	19,511	19,511
N_2		1,036	23,453	23,453	23,453	23,453		23,453	23,453	23,453	23,453	23,350	22,416	22,416
CO		29,920	50,718	50,718	50,718	31,099		31,099	21,712	21,712	21,712	21,666	20,798	20,798
CO_2		3,443	36,311	36,311	36,311	35,856		35,856	36,130	36,130	36,130	34,237	32,867	32,867
H_2		60,270	100,585	100,585	100,585	59,984		59,984	42,030	42,030	42,030	41,994	40,315	40,315
H_2O		298	314	314	314	777		777	513	513	513	17	16	16
CH_4		6	119	119	119	119		119	119	119	119	118	113	113
CH_3OH		0	2,757	2,757	2,757	22,814		22,814	31,912	31,912	31,912	2,872	2,757	2,757
C_2H_6O		0	35	35	35	41		41	47	47	47	36	35	35
C_3H_8O		0	0	0	0	2		2	3	3	3	0	0	0
Total Flow	lbmol/hr	95,912	234,742	234,742	234,742	194,595		194,595	176,367	176,367	176,367	144,616	138,830	138,830
Total Flow	lb/hr	1,183,090	4,792,870	4,792,870	4,792,870	4,792,870		4,792,870	4,792,870	4,792,870	4,792,870	3,760,250	3,609,780	3,609,780
Total Flow	ACFH	957,495	2,118,840	3,059,710	2,940,930	2,659,350		2,449,880	2,291,550	1,638,940	1,638,940	1,246,230	1,196,360	1,160,650
Temperature	°F	228.60	173.42	437.73	400.00	475.00		400.00	430.00	230.00	130.00	130.00	130.00	141.50
Pressure	psia	755.00	755.00	750.00	747.00	737.00		732.00	727.00	722.00	720.00	717.00	717.00	755.00
Density	lb/ft³	1.24	2.26	1.57	1.63	1.80		1.96	2.09	2.92	2.92	3.02	3.02	3.11
Average MW	lb/lbmol	12.34	20.42	20.42	20.42	24.63		24.63	27.18	27.18	27.18	26.00	26.00	26.00

STREAM		714	715	716	718	720	722	724	725	726	727	728	729	730
Description		Raw Methanol	PSA Feed	Raw Methanol	N/A	N/A	N/A	Methanol Flash Gas	N/A	N/A	Purge Gas	Purge Gas to Coal Drying	N/A	Stack Gas
Mole Flow (Vapor/Liquid)	lbmol/hr													
AR		125	813	2	0	0	0	123	0	0	665	148	0	704
N_2		102	934	1	0	0	0	101	0	0	764	170	0	4,041
CO		45	867	0	0	0	0	45	0	0	709	158	0	0
CO_2		1,893	1,369	343	0	0	0	1,550	0	0	1,120	249	0	1,930
H_2		36	1,680	0	0	0	0	36	0	0	481	107	0	0
H_2O		496	1	493	0	0	0	2	0	0	1	0	0	683
CH_4		1	5	0	0	0	0	1	0	0	4	1	0	0
CH_3OH		29,039	115	28,626	0	0	0	413	0	0	94	21	0	0
C_2H_6O		11	1	6	0	0	0	4	0	0	1	0	0	0
C_3H_8O		3	0	3	0	0	0	0	0	0	0	0	0	0
Total Flow	lbmol/hr	31,752	5,785	29,476	0	0	0	2,276	0	0	3,839	854	0	7,489
Total Flow	lb/hr	1,032,610	150,410	941,823	0	0	0	90,790	0	0	121,246	26,977	0	242,765
Total Flow	ACFH	21,589	49,849	19,614	0	0	0	346,987	0	0	1,212,280	269,725	0	3,780,830
Temperature	°F	130.00	130.00	117.46	0.00	0.00	0.00	117.46	0.00		70.00	70.00		246.00
Pressure	psia	717.00	717.00	40.00	0.00	0.00	0.00	40.00	0.00		18.00	18.00		15.00
Density	lb/ft³	47.83	3.02	48.02	0.00	0.00	0.00	0.26	0.00		0.10	0.10		0.06
Average MW	lb/lbmol	32.52	26.00	31.95	0.00	0.00	0.00	39.89	0.00		31.58	31.58		32.42

Exhibit 3-23 Mass balance of methanol synthesis (continued)
Case 3

STREAM		701	702	703	704	705	706	707	708	709	710	711	712	713
Description		HP Sweet Syngas	Cold Mixed Feed	Warm Mixed Feed	Hot Mixed Feed	Stage 1 Product		Stage 2 Cooled Feed	Hot Stage 2 Product	Warm Stage 2 Product	Cool Stage 2 Product	Flash Gas	LP Recycle Gas	HP Recycle Gas
Mole Flow (Vapor/Liquid)	lbmol/hr													
AR		463	10,255	10,255	10,255	10,255		10,255	10,255	10,255	10,255	10,200	9,792	9,792
N_2		845	19,330	19,330	19,330	19,330		19,330	19,330	19,330	19,330	19,255	18,485	18,485
CO		30,692	61,690	61,690	61,690	42,060		42,060	32,351	32,351	32,351	32,292	30,998	30,998
CO_2		3,035	36,177	36,177	36,177	35,944		35,944	36,189	36,189	36,189	34,524	33,143	33,143
H_2		60,227	97,485	97,485	97,485	57,525		57,525	38,839	38,839	38,839	38,810	37,258	37,258
H_2O		301	313	313	313	555		555	319	319	319	12	12	12
CH_4		1,298	27,368	27,368	27,368	27,368		27,368	27,368	27,368	27,368	27,156	26,070	26,070
CH_3OH		0	3,295	3,295	3,295	23,142		23,142	32,590	32,590	32,590	3,432	3,295	3,295
C_2H_6O		0	40	40	40	46		46	52	52	52	41	40	40
C_3H_8O		0	0	0	0	2		2	3	3	3	0	0	0
Total Flow	lbmol/hr	96,861	255,955	255,955	255,955	216,228		216,228	197,298	197,298	197,298	165,725	159,094	159,094
Total Flow	lb/hr	1,183,090	5,019,980	5,019,980	5,019,980	5,019,980		5,019,980	5,019,980	5,019,980	5,019,980	3,996,850	3,836,890	3,836,890
Total Flow	ACFH	966,372	2,283,970	3,207,910	3,203,200	2,953,770		2,720,560	2,563,470	1,864,960	1,864,960	1,418,000	1,361,250	1,319,340
Temperature	°F	228.18	168.92	404.57	400.00	475.00		400.00	430.00	230.00	130.00	130.00	130.00	140.93
Pressure	psia	755.00	755.00	750.00	747.00	737.00		732.00	727.00	722.00	720.00	717.00	717.00	755.00
Density	lb/ft³	1.22	2.20	1.56	1.57	1.70		1.85	1.96	2.69	2.69	2.82	2.82	2.91
Average MW	lb/lbmol	12.21	19.61	19.61	19.61	23.22		23.22	25.44	25.44	25.44	24.12	24.12	24.12

STREAM		714	715	716	718	720	722	724	725	726	727	728	729	730
Description		Raw Methanol	PSA Feed	Raw Methanol	Raw Methanol	Refined Methanol	Pure Methanol to Rectisol	Methanol Flash Gas	Gas to Coal Drying	H_2 Product	Purge Gas	Purge Gas to Coal Drying	H_2 to Hydrogenate	Stack Gas
Mole Flow (Vapor/Liquid)	lbmol/hr													
AR		55	408	1	0	0	0	54	0	0	665	148	0	704
N_2		75	770	1	0	0	0	74	0	0	764	170	0	4,041
CO		60	1,292	0	0	0	0	59	0	0	709	158	0	0
CO_2		1,664	1,381	312	0	0	0	1,352	0	0	1,120	249	0	1,930
H_2		30	1,552	0	0	0	0	30	0	0	481	107	0	0
H_2O		307	0	306	0	0	0	1	0	0	1	0	0	683
CH_4		211	1,086	6	0	0	0	205	0	0	4	1	0	0
CH_3OH		29,158	137	28,746	0	0	0	411	0	0	94	21	0	0
C_2H_6O		10	2	6	0	0	0	4	0	0	1	0	0	0
C_3H_8O		3	0	3	0	0	0	0	0	0	0	0	0	0
Total Flow	lbmol/hr	31,573	6,629	29,382	0	0	0	2,191	0	0	3,839	854	0	7,489
Total Flow	lb/hr	1,023,130	159,874	940,989	0	0	0	82,143	0	0	121,246	26,977	0	242,765
Total Flow	ACFH	21,563	56,720	19,637	0	0	0	334,731	0	0	1,212,280	269,725	0	3,780,830
Temperature	°F	130.00	130.00	118.53	0.00	0.00	0.00	118.53	0.00		70.00	70.00		246.00
Pressure	psia	717.00	717.00	40.00	0.00	0.00	0.00	40.00	0.00		18.00	18.00		15.00
Density	lb/ft³	47.45	2.82	47.92	0.00	0.00	0.00	0.25	0.00		0.10	0.10		0.06
Average MW	lb/lbmol	32.41	24.12	32.03	0.00	0.00	0.00	37.49	0.00		31.58	31.58		32.42

Exhibit 3-24 Power cycle heat and material balances
Case 1

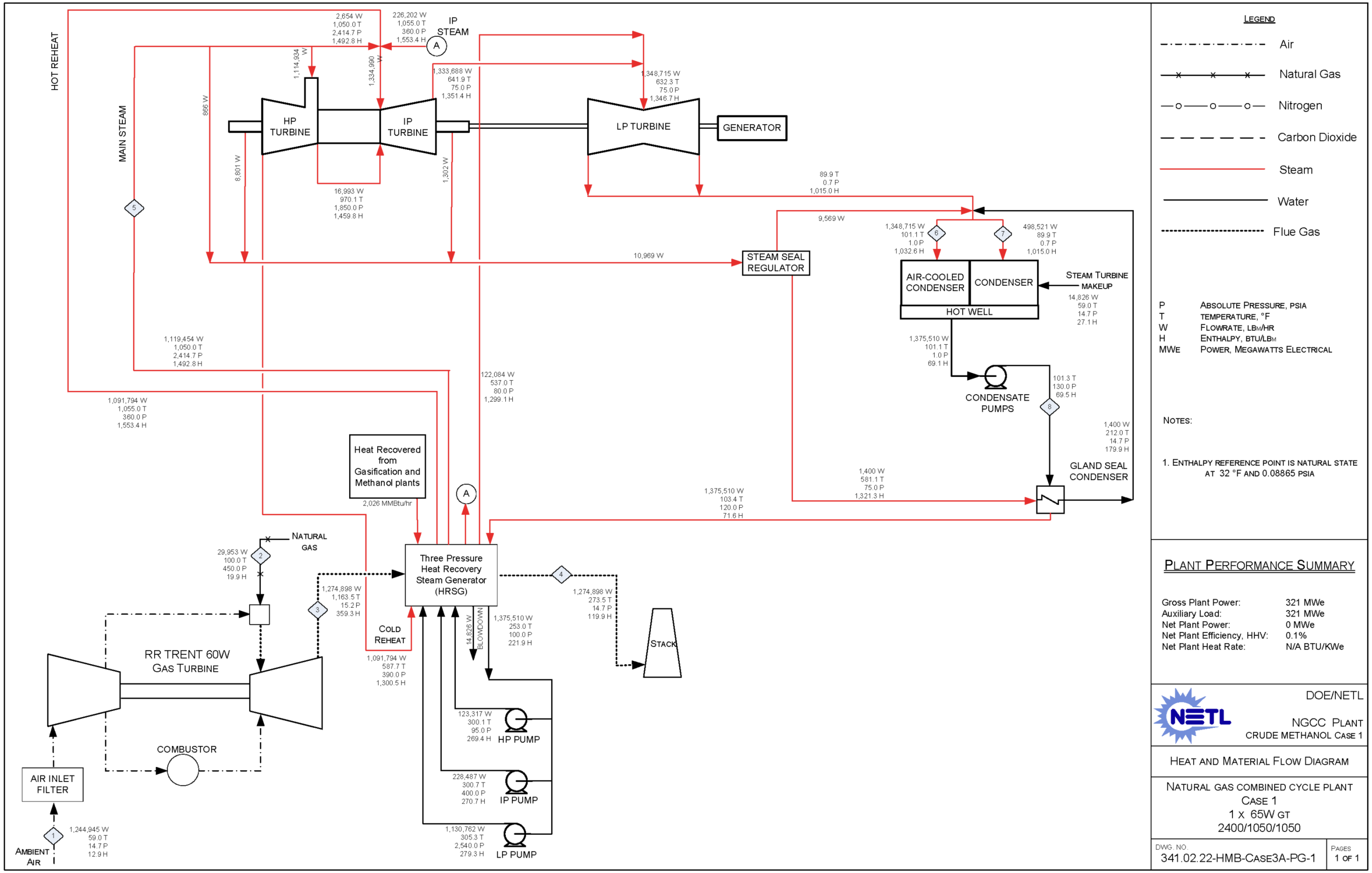


Exhibit 3-24 Power cycle heat and material balances (continued)
Case 2

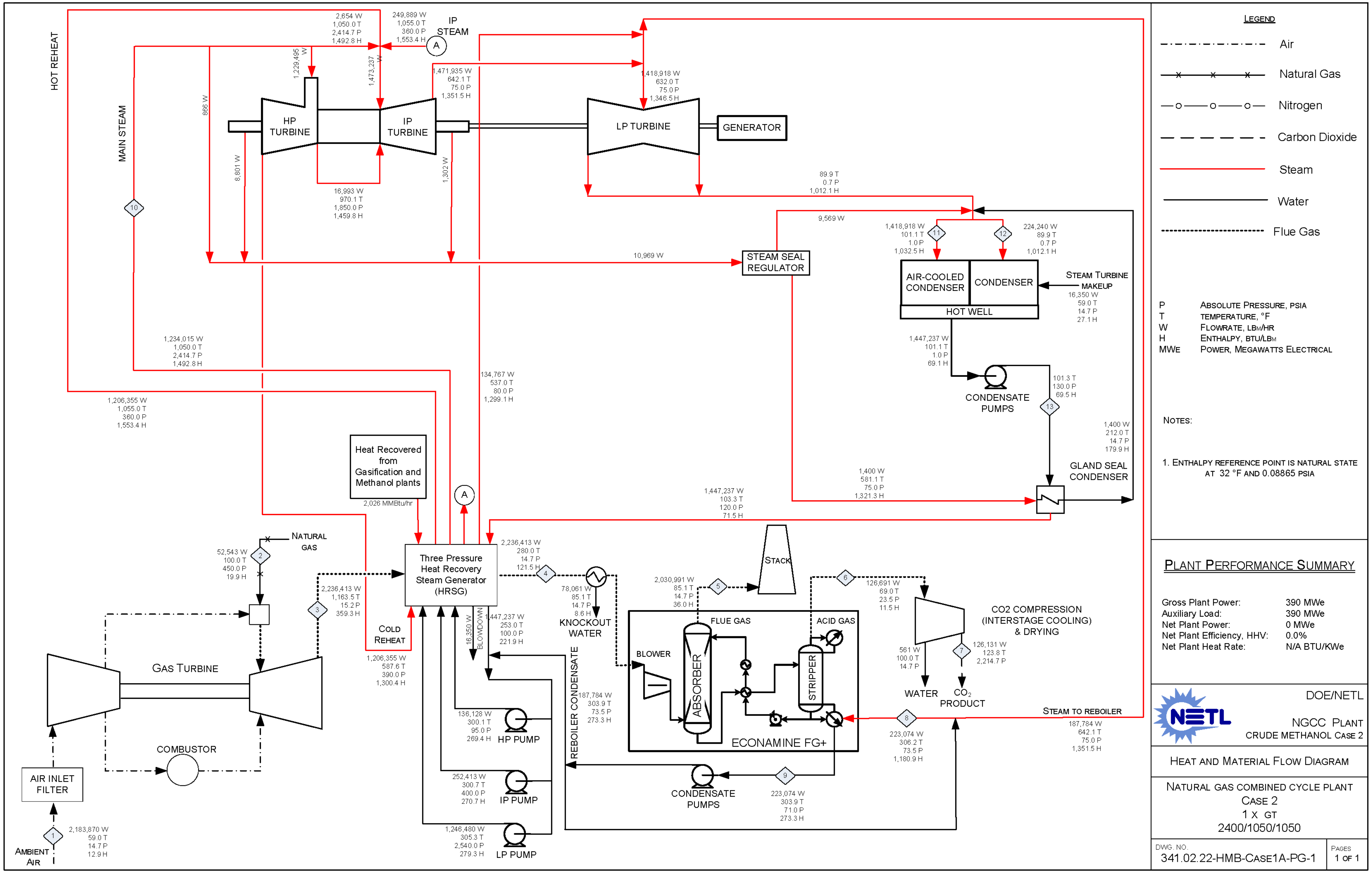


Exhibit 3-24 Power cycle heat and material balances (continued)
Case 3

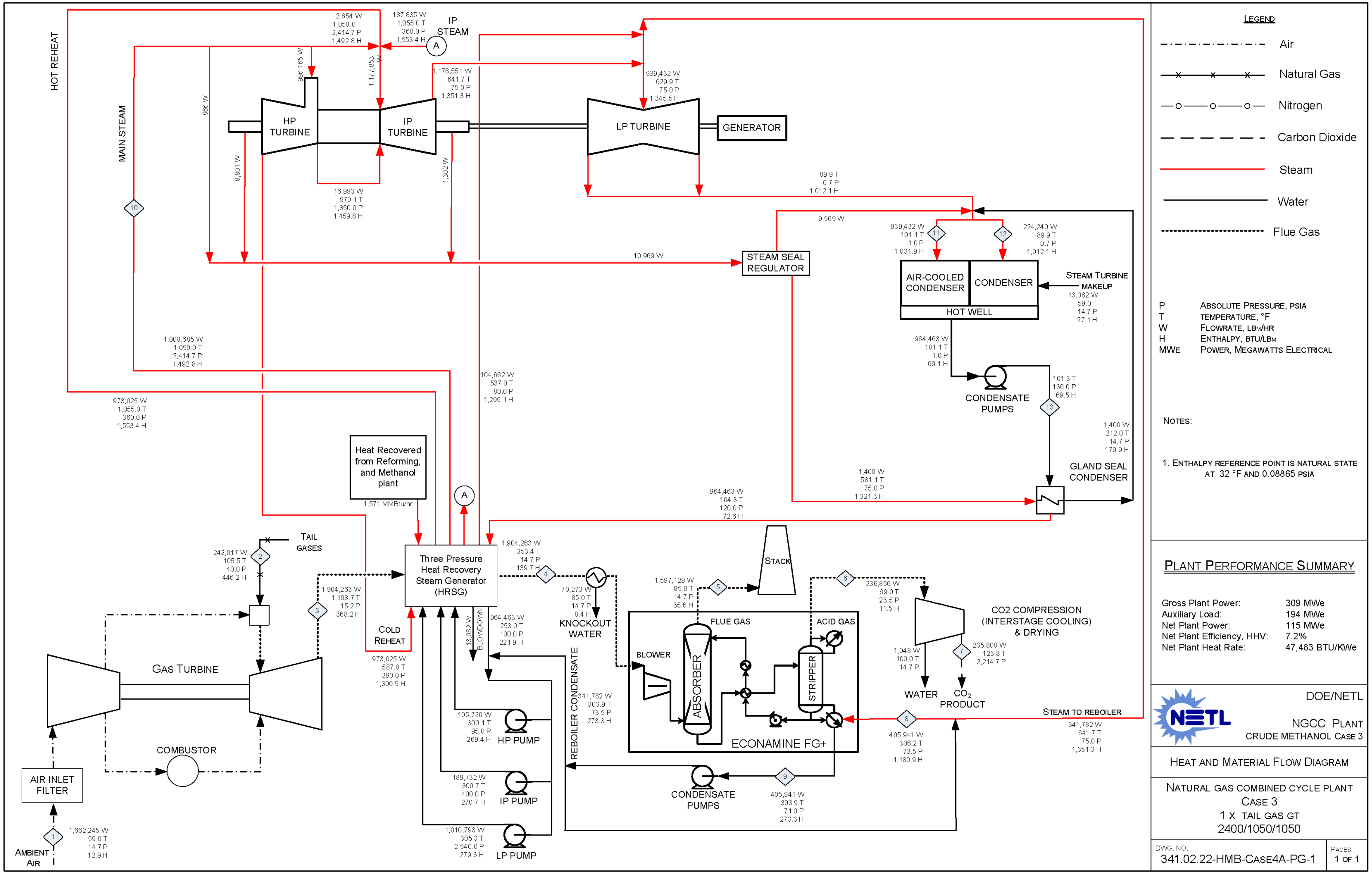
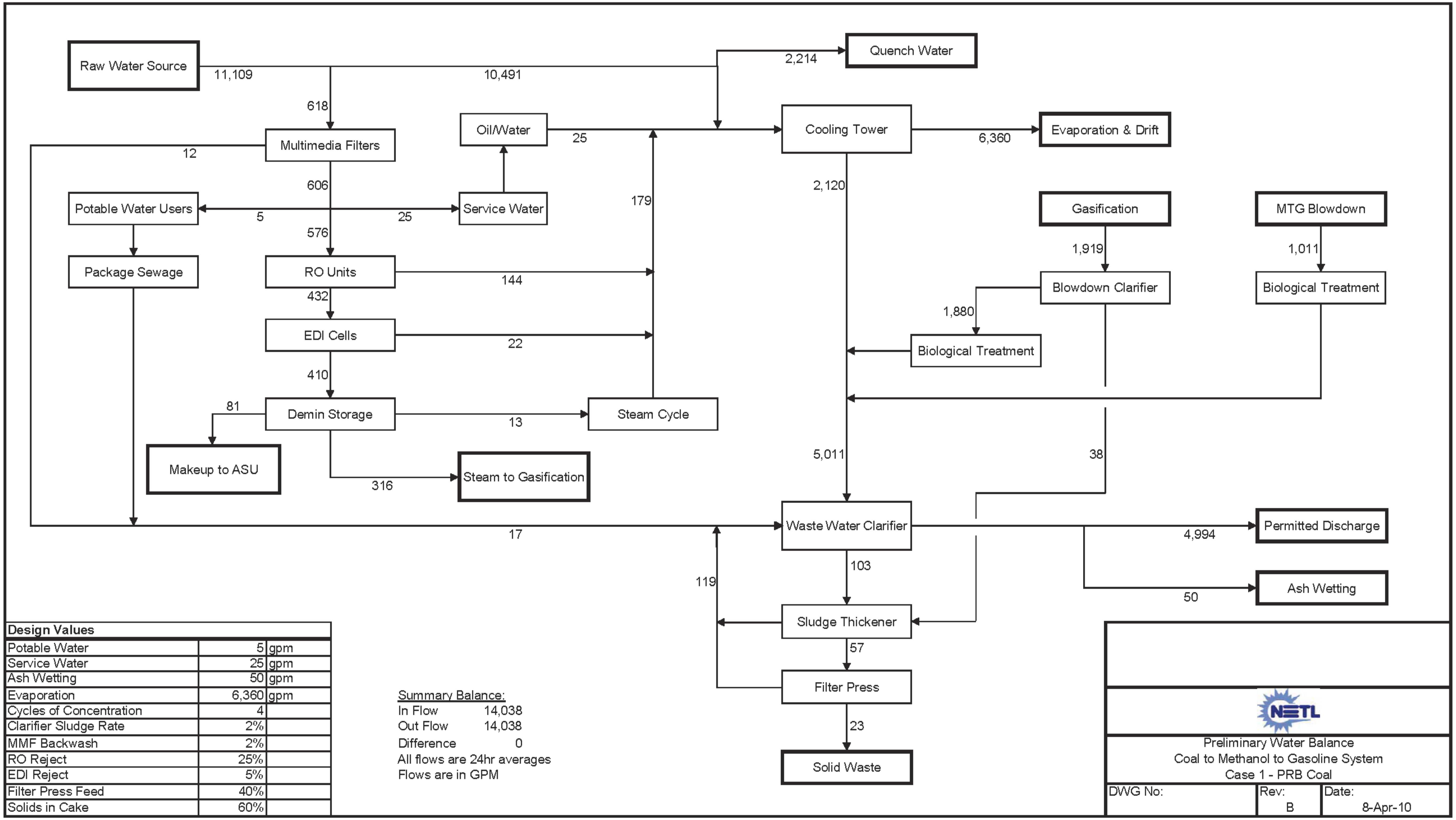
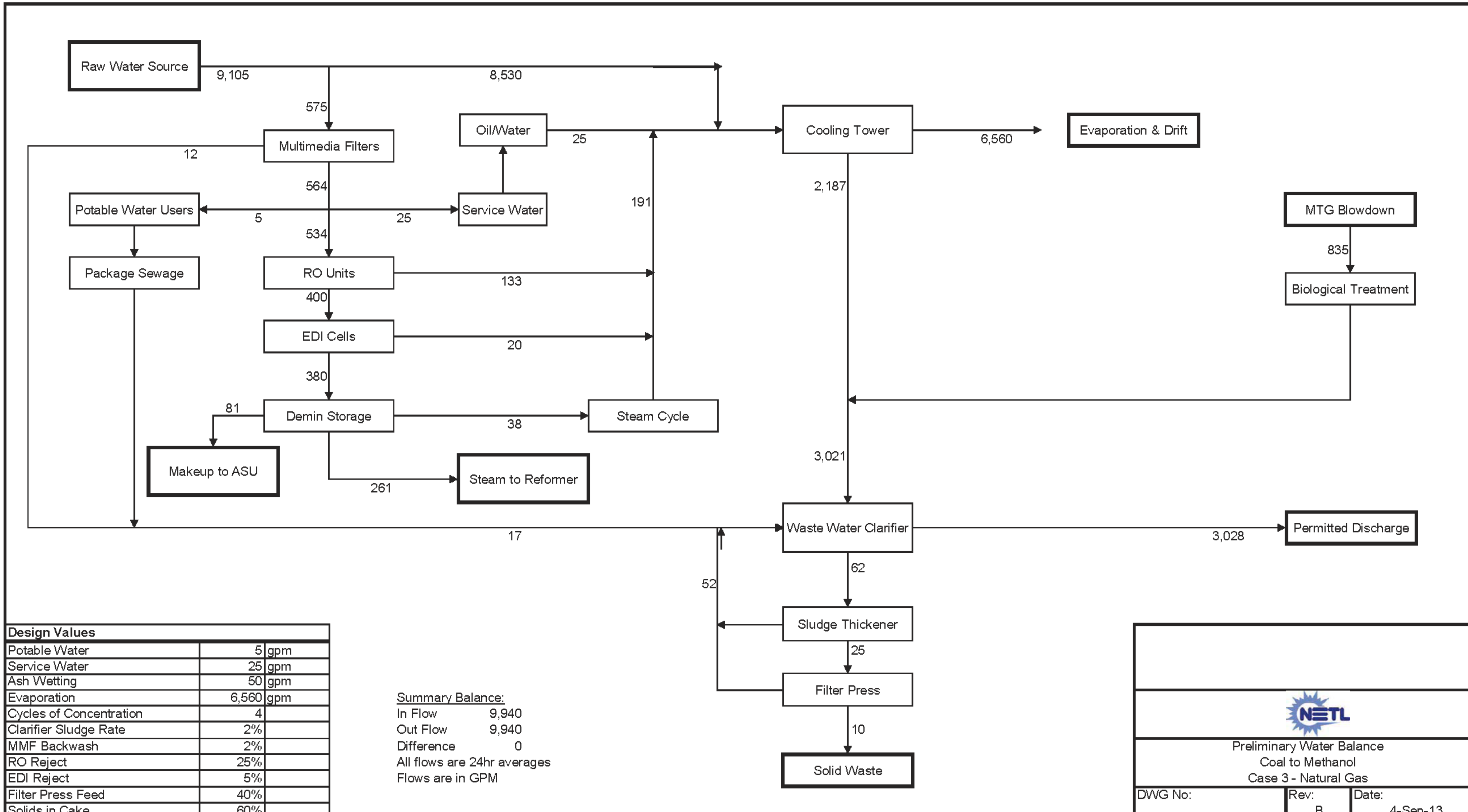


Exhibit 3-25 Water balances
Case 1 and 2



Design Values		
Potable Water	5	gpm
Service Water	25	gpm
Ash Wetting	50	gpm
Evaporation	6,360	gpm
Cycles of Concentration	4	
Clarifier Sludge Rate	2%	
MMF Backwash	2%	
RO Reject	25%	
EDI Reject	5%	
Filter Press Feed	40%	
Solids in Cake	60%	

Exhibit 3-25 Water balances (continued)
Case 3



Design Values		
Potable Water	5	gpm
Service Water	25	gpm
Ash Wetting	50	gpm
Evaporation	6,560	gpm
Cycles of Concentration	4	
Clarifier Sludge Rate	2%	
MMF Backwash	2%	
RO Reject	25%	
EDI Reject	5%	
Filter Press Feed	40%	
Solids in Cake	60%	

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4 Economic Analysis

4.1 Cost Estimating Methodology

Capital and operating cost estimates developed for these cases were based on adjusted vendor-furnished quotes, previous studies, actual cost data, or the best available information. All estimates are expressed in June 2011 dollars consistent with NETL methodology documented in Quality Guidelines and baseline reports. [17,18,19]

4.1.1 Capital Costs

The total overnight cost (TOC) for each plant was calculated by adding owner's costs to the total plant cost (TPC). TPC includes all equipment (complete with initial chemical and catalyst loadings), materials, labor (direct and indirect), engineering and construction management, and contingencies (process and project).

The capital costs have an estimated accuracy of +30/-15 percent, consistent with the screening study level of design engineering applied to the various cases in the study. The value of the study lies not in the absolute accuracy of the individual cases, but in the fact that all cases were evaluated under the same set of technical and economic assumptions. The consistency of the approach allows meaningful comparisons among the cases evaluated.

Process contingency was added to cost accounts to compensate for uncertainty in cost estimates caused by performance uncertainties associated with the development status of a technology. Project contingency was added to the engineering/procurement/construction management (EPCM) capital accounts to cover project uncertainty and the cost of any additional equipment that would result from a detailed design. The contingencies represent costs that are expected to occur. Each bare erected cost (BEC) account was evaluated against the level of estimate detail and field experience to determine project contingency.

TOC values are expressed in June 2011 dollars. The estimate represents current commercial offerings for the gasification and methanol synthesis and conversion processes. The estimates represent a complete fuels plant facility. The boundary limit is defined as the total plant facility within the fence line, including the coal receiving facilities and water supply system. Costs were grouped according to a process/system-oriented code of accounts; all reasonably allocable components of a system or process are included in the specific system account in contrast to a facility, area, or commodity account structure.

4.1.2 Feedstock prices

4.1.2.1 Coal Price

The coal type assumed for this study is a Powder River Basin (PRB) subbituminous coal supplied from the Montana Rosebud mine. The coal price was assumed to be $36.57 per short ton (year 2011 dollars) based on Montana Rosebud PRB Coal delivered to the Midwest as specified in the QGESS: Fuel Prices for Selected Feedstocks in NETL Studies. [1]

4.1.2.2 Natural Gas Price

The natural gas price was assumed to be $6.13/MMBtu (2011 dollars) based on natural gas prices delivered to Midwest power plants as specified in the QGESS: Fuel Prices for Selected Feedstocks in NETL Studies. [1]

4.1.3 Production Costs and Expenses

The production, or operations and maintenance (O&M), costs described in this section pertain to charges associated with operating and maintaining the methanol plant over its expected life.

O&M costs are determined an annual basis for the first year of operation. Quantities for major consumables, such as feedstock and fuel, were taken from the heat and mass balance developed for this application. Using reference data, other consumables were evaluated on the basis of the quantity required. Operating labor costs were determined on the basis of the number of operators. Maintenance costs were evaluated on the basis of requirements for each major plant section.

The O&M costs and expenses associated with the plant include the following:

- Operating labor
- Maintenance – material and labor
- Administrative and support labor
- Consumables
- Fuel/Feedstock cost
- Taxes and insurance

These costs and expenses are estimated on a reference basis and escalated to June 2011 dollars. The costs assume normal operation and do not include the initial startup costs. The operating labor, maintenance material and labor, and other labor-related costs were combined and then divided into two components: fixed O&M costs, which are independent of liquids production, and variable O&M costs, which are proportional to liquids production. The variable O&M cost estimate allocation is based on the plant capacity factor.

The other operating costs, consumables, and feedstock, are determined on a daily, 100-percent operating-capacity basis and are adjusted to an annual plant operation basis. The inputs for each category of operating costs and expenses are identified in the succeeding subsections, along with more specific discussion of the evaluation processes.

4.1.4 Required Selling Price

The figure-of-merit in this report is the required selling price (RSP) expressed in $/gal of crude methanol. The RSP values were calculated using the Power Systems Financial Model (PSFM) [20] and estimated to be the value calculated when the required return on equity (ROE) equals the internal rate of return (IRR) for 30 years of operation based on the assumed financial structure and escalations. RSP was assumed to escalate at three percent per year for the thirty-year economic life of the plant. All costs are expressed in June 2011 dollars.

For the natural gas case which generates excess electricity, the price of electricity was assumed to be $59.59/MWh expressed in June 2011 dollar. This is based on the lowest cost option from the Bituminous Baseline study results [19], which was considered to be typical of baseline 2011 plant designs. A sensitivity analysis of the RSP to this assumed value is included in the results discussion. While the excess power can be sold to the grid; the sale may be at a steep discount as entities that are negotiating a power purchase agreement will know the power production is an inherent by-product of core methanol production operations. Consequently, the actual achieved transfer price for excess power will be a significant risk in natural gas feedstock projects and would be highly project dependent. The impact of the price on the methanol RSP is reflected in the sensitivity analysis results.

The capital and operating costs for CO_2 T&S were independently estimated by NETL at $11/metric ton of CO_2 in 2011 dollars. [9]

The RSP was calculated for each case assuming (i) a financial structure representative of a commercial fuels project, and (ii) a financial structure with loan guarantees or other government subsidies. The financial assumptions and structures used to estimate the RSPs are shown in Exhibit 4-1, Exhibit 4-2, and Exhibit 4-3. [21] A sensitivity of the RSP to this assumed structure was included in the results discussion.

Exhibit 4-1 Financial assumptions for PSFM [17]

Parameter	Value
TAXES	
Income Tax Rate	38% (Effective, 34% Federal, 6% State)
Capital Depreciation	20 years, 150% declining balance
Investment Tax Credit	None
Tax Holiday	None
FINANCING TERMS	
Repayment Term of Debt	30 years
Grace Period on Debt Repayment	None
Debt Reserve Fund	None
TREATMENT OF CAPITAL COSTS	
Capital Cost Escalation During Construction (nominal annual rate)	3.6% [a]
Distribution of Total Overnight Capital over the Capital Expenditure Period (before escalation)	5-Year Period: 10%, 30%, 25%, 20%, 15%
Working Capital	zero for all parameters
% of Total Overnight Capital that is Depreciated	100% *(this assumption introduces a very small error even if a substantial amount of TOC is actually non-depreciable)*
INFLATION	
RSP, O&M, Fuel Escalation (nominal annual rate)	3.0% [b] RSP, O&M, COE, Fuel

[a] A nominal average annual rate of 3.6 percent is assumed for escalation of capital costs during construction. This rate is equivalent to the nominal average annual escalation rate for process plant construction costs between 1947 and 2008, according to the Chemical Engineering Plant Cost Index.
[b] An average annual inflation rate of 3.0 percent is assumed. This rate is equivalent to the average annual escalation rate between 1947 and 2008 for the U.S. Department of Labor's Producer Price Index for Finished Goods, the so-called "headline" index of the various Producer Price Indices. (The Producer Price Index for the Electric Power Generation Industry may be more applicable, but that data does not provide a long-term historical perspective since it only dates back to December 2003.)

Exhibit 4-2 Financial structure for commercial fuels projects [22]

Type of Security	Percent of Total	Current (Nominal) Dollar Cost
Debt	50	8% (LIBOR=3.5% + 4.5%)
Equity	50	20%

Exhibit 4-3 Financial structure for loan guarantee projects [23]

Type of Security	Percent of Total	Current (Nominal) Dollar Cost
Debt	60	4.56% (CMT=4.34% + 0.22%)
Equity	40	20%

4.2 Cost Estimation results

The capital and O&M costs for each of the cases are shown in Exhibit 4-4 through Exhibit 4-9.

Exhibit 4-4 Capital cost summary for Case 1

Client:	USDOE/NETL	**Report Date:**	2013-Sep-12
Project:	Baseline Study of Crude Methanol Plants		
	TOTAL PLANT COST SUMMARY		
Case:	Case 1 - Shell Coal to Methanol w/o CO_2 capture		
Plant Size:	941,823 Methanol (lb/hr):	**Estimate Type:** Conceptual	**Cost Base (Jun)** 2011 ($x1000)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	TOTAL PLANT COST $
1	COAL & SORBENT HANDLING	$36,523	$6,549	$28,312	$0	$0	**$71,384**	$6,479	$0	$15,573	**$93,436**
2	COAL & SORBENT PREP & FEED	$440,636	$31,753	$70,335	$0	$0	**$542,724**	$47,045	$0	$117,954	**$707,724**
3	FEEDWATER & MISC. BOP SYSTEMS	$29,200	$9,205	$20,034	$0	$0	**$58,439**	$5,478	$0	$14,635	**$78,552**
4	GASIFIER & ACCESSORIES										
4.1	Gasifier, Syngas Cooler & Auxiliaries (Shell)	$429,138	$0	$194,177	$0	$0	**$623,315**	$55,887	$89,766	$118,739	**$887,708**
4.2	Syngas Cooling	w/4.1	$0	w/4.1	$0	$0	**$0**	$0	$0	$0	**$0**
4.3	Air Separation Unit	$608,392	$0	w/equip.	$0	$0	**$608,392**	$58,971	$0	$66,736	**$734,099**
4.4-4.9	Other Gasification Equipment	$107,905	$32,835	$60,116	$0	$0	**$200,855**	$19,296	$0	$46,764	**$266,915**
	SUBTOTAL 4	**$1,145,435**	**$32,835**	**$254,293**	**$0**	**$0**	**$1,432,563**	**$134,154**	**$89,766**	**$232,239**	**$1,888,722**
5A	GAS CLEANUP & PIPING										
5A.1	Rectisol System	$350,917	$0	w/equip	$0	$0	**$350,917**	$33,153	$70,183	$90,851	**$545,104**
5A.2-5A.9	Other Gas cleanup & Piping Equipment	$72,962	$8,963	$69,557	$0	$0	**$151,482**	$14,617	$667	$33,312	**$200,079**
	SUBTOTAL 5A	**$423,879**	**$8,963**	**$69,557**	**$0**	**$0**	**$502,399**	**$47,769**	**$70,851**	**$124,163**	**$745,182**
5B	CO_2 Removal and COMPRESSION	$12,189	$0	$4,595	$0	$0	**$16,784**	$1,608	$0	$3,678	**$22,070**
5C	METHANOL PRODUCTION	$133,185	$56,033	$112,067	$0	$0	**$301,285**	$30,129	$0	$66,283	**$397,696**
6	COMBUSTION TURBINE/ACCESSORIES										
6.1	Combustion Turbine Generator	$20,055	$0	$1,310	$0	$0	**$21,365**	$1,797	$0	$2,356	**$25,518**
6.2-6.9	Combustion Turbine Other	$0	$181	$196	$0	$0	**$377**	$32	$0	$82	**$491**
	SUBTOTAL 6	**$20,055**	**$181**	**$1,506**	**$0**	**$0**	**$21,743**	**$1,829**	**$0**	**$2,437**	**$26,009**
7	HRSG, DUCTING & STACK										
7.1	Heat Recovery Steam Generator	$44,163	$0	$8,308	$0	$0	**$52,471**	$4,357	$0	$5,683	**$62,511**
7.2-7.9	HRSG Accessories, Ductwork, & Stack	$2,477	$1,226	$1,658	$0	$0	**$5,361**	$449	$0	$883	**$6,694**
	SUBTOTAL 7	**$46,640**	**$1,226**	**$9,966**	**$0**	**$0**	**$57,833**	**$4,806**	**$0**	**$6,566**	**$69,205**
8	STEAM TURBINE GENERATOR										
8.1	Steam TG & Accessories	$35,686	$0	$4,793	$0	$0	**$40,479**	$3,881	$0	$4,436	**$48,797**
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	$20,278	$1,114	$9,648	$0	$0	**$31,040**	$2,790	$0	$6,975	**$40,805**
	SUBTOTAL 8	**$55,964**	**$1,114**	**$14,441**	**$0**	**$0**	**$71,519**	**$6,671**	**$0**	**$11,411**	**$89,601**
9	COOLING WATER SYSTEM	$17,962	$19,584	$14,636	$0	$0	**$52,182**	$4,828	$0	$11,587	**$68,597**
10	ASH/SPENT SORBENT HANDLING SYS	$80,651	$3,590	$75,947	$0	$0	**$160,188**	$15,498	$0	$18,339	**$194,026**
11	ACCESSORY ELECTRIC PLANT	$34,017	$19,371	$34,182	$0	$0	**$87,571**	$7,605	$0	$18,897	**$114,073**
12	INSTRUMENTATION & CONTROL	$11,382	$3,440	$9,546	$0	$0	**$24,369**	$2,187	$1,218	$4,978	**$32,753**
13	IMPROVEMENTS TO SITE	$5,818	$2,834	$15,212	$0	$0	**$23,865**	$2,356	$0	$7,866	**$34,088**
14	BUILDINGS & STRUCTURES	$0	$7,832	$10,959	$0	$0	**$18,791**	$1,703	$0	$3,333	**$23,827**
	TOTAL COST	**$2,493,538**	**$204,510**	**$745,590**	**$0**	**$0**	**$3,443,638**	**$320,147**	**$161,835**	**$659,940**	**$4,585,560**
	Total Overnight Costs (TOC)										**$5,614,724**
	Total As-Spent Cost (TASC)										**$6,630,989**

Exhibit 4-5 Capital cost summary for Case 2

Client: USDOE/NETL — **Report Date:** 2014-Sep-22
Project: Baseline Study of Crude Methanol Plants
TOTAL PLANT COST SUMMARY
Case: Case 2 - Shell Coal to Methanol w/ CO_2 capture
Plant Size: 941,823 Methanol (lb/hr): — **Estimate Type:** Conceptual — **Cost Base (Jun)** 2007 ($x1000)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	TOTAL PLANT COST $
1	COAL & SORBENT HANDLING	$36,523	$6,549	$28,312	$0	$0	**$71,384**	$6,479	$0	$15,573	**$93,436**
2	COAL & SORBENT PREP & FEED	$440,636	$31,753	$70,335	$0	$0	**$542,724**	$47,045	$0	$117,954	**$707,724**
3	FEEDWATER & MISC. BOP SYSTEMS	$31,578	$9,286	$20,105	$0	$0	**$60,970**	$5,722	$0	$15,190	**$81,882**
4	GASIFIER & ACCESSORIES										
4.1	Gasifier, Syngas Cooler & Auxiliaries (Shell)	$429,138	$0	$194,177	$0	$0	**$623,315**	$55,887	$89,766	$118,739	**$887,708**
4.2	Syngas Cooling	w/4.1	$0	w/4.1	$0	$0	**$0**	$0	$0	$0	**$0**
4.3	Air Separation Unit	$608,392	$0	w/equip.	$0	$0	**$608,392**	$58,971	$0	$66,736	**$734,099**
4.4-4.9	Other Gasification Equipment	$107,905	$32,835	$60,116	$0	$0	**$200,855**	$19,296	$0	$46,764	**$266,915**
	SUBTOTAL 4	**$1,145,435**	**$32,835**	**$254,293**	**$0**	**$0**	**$1,432,563**	**$134,154**	**$89,766**	**$232,239**	**$1,888,722**
5A	GAS CLEANUP & PIPING										
5A.1	Rectisol System	$350,917	$0	w/equip	$0	$0	**$350,917**	$33,153	$70,183	$90,851	**$545,104**
5A.2-5A.9	Other Gas cleanup & Piping Equipment	$72,962	$8,963	$69,557	$0	$0	**$151,482**	$14,617	$667	$33,312	**$200,079**
	SUBTOTAL 5A	**$423,879**	**$8,963**	**$69,557**	**$0**	**$0**	**$502,399**	**$47,769**	**$70,851**	**$124,163**	**$745,182**
5B	CO_2 Removal and COMPRESSION	$144,222	$0	$50,113	$0	$0	**$194,335**	$17,686	$15,082	$45,420	**$272,523**
5C	METHANOL PRODUCTION	$133,185	$56,033	$112,067	$0	$0	**$301,285**	$30,129	$0	$66,283	**$397,696**
6	COMBUSTION TURBINE/ACCESSORIES										
6.1	Combustion Turbine Generator	$29,881	$0	$1,952	$0	$0	**$31,832**	$2,678	$0	$3,510	**$38,019**
6.2-6.9	Combustion Turbine Other	$0	$270	$292	$0	$0	**$562**	$47	$0	$122	**$731**
	SUBTOTAL 6	**$29,881**	**$270**	**$2,244**	**$0**	**$0**	**$32,394**	**$2,725**	**$0**	**$3,631**	**$38,750**
7	HRSG, DUCTING & STACK										
7.1	Heat Recovery Steam Generator	$44,914	$0	$8,449	$0	$0	**$53,363**	$4,431	$0	$5,779	**$63,573**
7.2-7.9	HRSG Accessories, Ductwork, & Stack	$2,519	$1,301	$1,737	$0	$0	**$5,557**	$465	$0	$921	**$6,943**
	SUBTOTAL 7	**$47,433**	**$1,301**	**$10,186**	**$0**	**$0**	**$58,920**	**$4,896**	**$0**	**$6,700**	**$70,516**
8	STEAM TURBINE GENERATOR										
8.1	Steam TG & Accessories	$36,292	$0	$4,875	$0	$0	**$41,167**	$3,947	$0	$4,511	**$49,626**
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	$20,623	$1,133	$9,812	$0	$0	**$31,567**	$2,838	$0	$7,094	**$41,498**
	SUBTOTAL 8	**$56,915**	**$1,133**	**$14,686**	**$0**	**$0**	**$72,734**	**$6,785**	**$0**	**$11,605**	**$91,124**
9	COOLING WATER SYSTEM	$18,267	$19,916	$14,885	$0	$0	**$53,068**	$4,910	$0	$11,784	**$69,762**
10	ASH/SPENT SORBENT HANDLING SYS	$80,651	$3,590	$75,947	$0	$0	**$160,188**	$15,498	$0	$18,339	**$194,026**
11	ACCESSORY ELECTRIC PLANT	$39,363	$23,108	$40,531	$0	$0	**$103,002**	$8,955	$0	$22,315	**$134,271**
12	INSTRUMENTATION & CONTROL	$12,555	$3,795	$10,530	$0	$0	**$26,880**	$2,413	$1,344	$5,491	**$36,128**
13	IMPROVEMENTS TO SITE	$6,009	$2,891	$15,764	$0	$0	**$24,665**	$2,435	$0	$8,130	**$35,231**
14	BUILDINGS & STRUCTURES	$0	$8,022	$11,375	$0	$0	**$19,396**	$1,754	$0	$3,431	**$24,581**
	TOTAL COST	**$2,646,532**	**$209,446**	**$800,930**	**$0**	**$0**	**$3,656,908**	**$339,355**	**$177,042**	**$708,249**	**$4,881,555**
	Total Overnight Costs (TOC)										**$5,973,133**
	Total As-Spent Cost (TASC)										**$7,054,270**

Exhibit 4-6 Capital cost summary for Case 3

Client: USDOE/NETL **Report Date:** 2014-Sep-22
Project: Baseline Study of Crude Methanol Plants

TOTAL PLANT COST SUMMARY

Case: Case 3 - NG to Methanol w/ CO_2 capture
Plant Size: 940,989 Methanol (lb/hr): **Estimate Type:** Conceptual **Cost Base (Jun)** 2007 ($x1000)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	TOTAL PLANT COST $
1	COAL & SORBENT HANDLING	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**
2	COAL & SORBENT PREP & FEED	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**
3	Feedwater, Natural Gas & MISC. BOP SYSTEMS	$51,389	$5,804	$10,963	$0	$0	**$68,156**	$6,490	$0	$15,841	**$90,487**
4	GASIFIER & ACCESSORIES										
4.1	Gasifier, Syngas Cooler & Auxiliaries (Shell)	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**
4.2	Syngas Cooling	w/4.7	$0	w/4.7	$0	$0	**$0**	$0	$0	$0	**$0**
4.3	Air Separation Unit	$482,112	$0	w/equip.	$0	$0	**$482,112**	$46,731	$0	$52,884	**$581,728**
4.4-4.9	Reformer and Other Gas processing Equipment	$329,244	$0	w/equip.	$0	$0	**$329,244**	$31,913	$0	$36,116	**$397,273**
	SUBTOTAL 4	**$811,357**	**$0**	**$0**	**$0**	**$0**	**$811,357**	**$78,644**	**$0**	**$89,000**	**$979,001**
5A	GAS CLEANUP & PIPING	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**
5B	CO_2 Removal and COMPRESSION	$107,192	$0	$92,382	$0	$0	**$199,574**	$17,057	$33,368	$50,000	**$299,998**
5C	METHANOL PRODUCTION	$133,067	$55,984	$111,967	$0	$0	**$301,018**	$30,102	$0	$66,224	**$397,344**
6	COMBUSTION TURBINE/ACCESSORIES										
6.1	Combustion Turbine Generator	$31,501	$0	$2,057	$0	$0	**$33,559**	$2,823	$0	$3,700	**$40,081**
6.2-6.9	Combustion Turbine Other	$0	$285	$308	$0	$0	**$593**	$50	$0	$128	**$771**
	SUBTOTAL 6	**$31,501**	**$285**	**$2,365**	**$0**	**$0**	**$34,151**	**$2,872**	**$0**	**$3,828**	**$40,852**
7	HRSG, DUCTING & STACK										
7.1	Heat Recovery Steam Generator	$39,776	$0	$7,482	$0	$0	**$47,258**	$3,924	$0	$5,118	**$56,301**
7.2-7.9	HRSG Accessories, Ductwork, & Stack	$2,231	$1,181	$1,565	$0	$0	**$4,976**	$417	$0	$828	**$6,221**
	SUBTOTAL 7	**$42,006**	**$1,181**	**$9,048**	**$0**	**$0**	**$52,235**	**$4,341**	**$0**	**$5,946**	**$62,521**
8	STEAM TURBINE GENERATOR										
8.1	Steam TG & Accessories	$32,141	$0	$4,317	$0	$0	**$36,458**	$3,496	$0	$3,995	**$43,948**
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	$18,263	$1,003	$8,689	$0	$0	**$27,956**	$2,513	$0	$6,282	**$36,751**
	SUBTOTAL 8	**$50,404**	**$1,003**	**$13,006**	**$0**	**$0**	**$64,413**	**$6,009**	**$0**	**$10,277**	**$80,699**
9	COOLING WATER SYSTEM	$16,177	$17,638	$13,182	$0	$0	**$46,997**	$4,348	$0	$10,436	**$61,781**
10	ASH/SPENT SORBENT HANDLING SYS	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**
11	ACCESSORY ELECTRIC PLANT	$29,449	$15,268	$27,013	$0	$0	**$71,730**	$6,201	$0	$15,362	**$93,293**
12	INSTRUMENTATION & CONTROL	$8,260	$2,497	$6,928	$0	$0	**$17,684**	$1,587	$884	$3,613	**$23,768**
13	IMPROVEMENTS TO SITE	$8,744	$2,241	$6,842	$0	$0	**$17,828**	$1,760	$0	$5,876	**$25,464**
14	BUILDINGS & STRUCTURES	$0	$5,272	$7,758	$0	$0	**$13,030**	$1,175	$0	$2,325	**$16,529**
	TOTAL COST	**$1,289,547**	**$107,172**	**$301,454**	**$0**	**$0**	**$1,698,173**	**$160,586**	**$34,253**	**$278,728**	**$2,171,740**
	Total Overnight Costs (TOC)										**$2,644,295**
	Total As-Spent Cost (TASC)										**$3,122,912**

Exhibit 4-7 O&M cost summary for Case 1

INITIAL & ANNUAL O&M EXPENSES					Cost Base (Jun):	2011
Case 1 - Shell Coal to Methanol w/o CO_2 capture						
					MWe-net:	0
Methanol (lb/hr):	941,823				Capacity Factor (%):	90
OPERATING & MAINTENANCE LABOR						
Operating Labor						
Operating Labor Rate(base):	39.70	$/hour				
Operating Labor Burden:	30.00	% of base				
Labor O-H Charge Rate:	25.00	% of labor				
			Total			
Skilled Operator	2.0		2.0			
Operator	10.0		10.0			
Foreman	1.0		1.0			
Lab Tech's, etc.	3.0		3.0			
TOTAL-O.J.'s	16.0		16.0			
					Annual Cost	Annual Unit Cost
					$	$/lb
Annual Operating Labor Cost					$7,233,658	$0.001
Maintenance Labor Cost					$44,717,809	$0.006
Administrative & Support Labor					$12,987,867	$0.002
Property Taxes and Insurance					$91,711,204	$0.012
TOTAL FIXED OPERATING COSTS					**$156,650,538**	**$0.021**
VARIABLE OPERATING COSTS						
						$/lb
Maintenance Material Cost					$67,076,714	**$0.00903**
Consumables	Consumption		Unit	Initial Fill		
	Initial Fill	/Day	Cost	Cost		
Water(/1000 gallons)	0	7,998	1.67	**$0**	**$4,398,421**	**$0.00059**
Chemicals						
MU & WT Chem. (lb)	0	47,653	0.27	$0	$4,192,723	$0.00056
Carbon (Mercury Removal) (lb)	263,903	402	1.63	$428,905	$214,452	$0.00003
MEA Solvent (ton)	0	0.0	3,751.70	$0	$0	$0.00000
Water Gas Shift Catalyst (ft^3)	10,552	7.23	771.99	$8,146,310	$1,832,920	$0.00025
Claus Catalyst (ft^3)	w/equip	2.06	203.15	$0	$137,210	$0.00002
NG ATR Reformer Catalyst (ft^3)	0	0.00	650.00	$0	$0	$0.00000
Methanol Synthesis Catalyst (ft^3)	4,056	3.70	534.68	$2,168,500	$650,550	$0.00009
Aromatics Hydrotreater Catalyst (ft^3)	0	0.00	507.94	$0	$0	$0.00000
Subtotal Chemicals				**$10,743,716**	**$7,027,856**	**$0.00095**
Other						
Supplemental Electricity (for consumption)	0	0	62.33	$0	$0	$0.00000
Gases, N2 etc. (/100scf)	0	0	0.00	$0	$0	$0.00000
L.P. Steam (/1000 pounds)	0	0	0.00	$0	$0	$0.00000
Subtotal Other				**$0**	**$0**	**$0.00000**
Waste Disposal						
Spent Mercury Catalyst (lb.)	0	402	0.65	$0	$85,781	$0.00001
Flyash (ton)	0	0	0.00	$0	$0	$0.00000
Slag (ton)	0	2,081	25.11	$0	$17,167,503	$0.00231
Subtotal-Waste Disposal				**$0**	**$17,253,285**	**$0.00232**
By-products & Emissions						
Sulfur (tons)	0	140	0.00	$0	$0	$0.00000
Supplemental Electricity (for sale) (MWh)	0	0	-59.59	$0	$0	$0.00000
Subtotal By-Products				**$0**	**$0**	**$0.00000**
TOTAL VARIABLE OPERATING COSTS				**$10,743,716**	**$95,756,275**	**$0.01290**
Fuel/Feedstock						
Coal (ton)	0	19,418	36.57	**$0**	**$233,276,555**	**$0.03142**
Natural Gas (MMBtu)	0	16,210	6.13	**$0**	**$32,641,812**	**$0.00440**

Exhibit 4-8 O&M cost summary for Case 2

INITIAL & ANNUAL O&M EXPENSES					Cost Base (Jun):	2011
Case 2 - Shell Coal to Methanol w/ CO_2 capture						
					MWe-net:	0
Methanol (lb/hr):	941,823				Capacity Factor (%):	90
OPERATING & MAINTENANCE LABOR						
Operating Labor						
Operating Labor Rate(base):	39.70	$/hour				
Operating Labor Burden:	30.00	% of base				
Labor O-H Charge Rate:	25.00	% of labor				
			Total			
Skilled Operator	2.0		2.0			
Operator	10.0		10.0			
Foreman	1.0		1.0			
Lab Tech's, etc.	3.0		3.0			
TOTAL-O.J.'s	16.0		16.0			
					Annual Cost	Annual Unit Cost
					$	$/lb
Annual Operating Labor Cost					$7,233,658	$0.001
Maintenance Labor Cost					$47,038,006	$0.006
Administrative & Support Labor					$13,567,916	$0.002
Property Taxes and Insurance					$97,631,091	$0.013
TOTAL FIXED OPERATING COSTS					**$165,470,670**	**$0.022**
VARIABLE OPERATING COSTS						
						$/lb
Maintenance Material Cost					$70,557,009	**$0.00950**
Consumables	Consumption		Unit	Initial Fill		
	Initial Fill	/Day	Cost	Cost		
Water(/1000 gallons)	0	7,998	1.67	**$0**	**$4,398,421**	**$0.00059**
Chemicals						
MU & WT Chem. (lb)	0	47,653	0.27	$0	$4,192,723	$0.00056
Carbon (Mercury Removal) (lb)	263,903	402	1.63	$428,905	$214,452	$0.00003
MEA Solvent (ton)	117	0.16	3,751.70	$439,595	$202,615	$0.00003
Water Gas Shift Catalyst (ft^3)	10,552	7.23	771.99	$8,146,310	$1,832,920	$0.00025
Claus Catalyst (ft^3)	w/equip	2.06	203.15	$0	$137,210	$0.00002
Triethylene Glycol (gal)	0	592.60	6.57	$0	$1,279,570	$0.00017
NG ATR Reformer Catalyst (ft^3)	0	0.00	650.00	$0	$0	$0.00000
Methanol Synthesis Catalyst (ft^3)	4,056	3.70	534.68	$2,168,500	$650,550	$0.00009
Aromatics Hydrotreater Catalyst (ft^3)	0	0.00	507.94	$0	$0	$0.00000
Subtotal Chemicals				**$11,183,311**	**$8,510,041**	**$0.00115**
Other						
Supplemental Electricity (for consumption)	0	0	62.33	$0	$0	$0.00000
Gases, N2 etc. (/100scf)	0	0	0.00	$0	$0	$0.00000
L.P. Steam (/1000 pounds)	0	0	0.00	$0	$0	$0.00000
Subtotal Other				**$0**	**$0**	**$0.00000**
Waste Disposal						
Spent Mercury Catalyst (lb.)	0	402	0.65	$0	$85,781	$0.00001
Flyash (ton)	0	0	0.00	$0	$0	$0.00000
Slag (ton)	0	2,081	25.11	$0	$17,167,503	$0.00231
Subtotal-Waste Disposal				**$0**	**$17,253,285**	**$0.00232**
By-products & Emissions						
Sulfur (tons)	0	140	0.00	$0	$0	$0.00000
Supplemental Electricity (for sale) (MWh)	0	0	-59.59	$0	$0	$0.00000
Subtotal By-Products				**$0**	**$0**	**$0.00000**
TOTAL VARIABLE OPERATING COSTS				**$11,183,311**	**$100,718,756**	**$0.01356**
Fuel/Feedstock						
Coal (ton)	0	19,418	36.57	**$0**	**$233,276,555**	**$0.03142**
Natural Gas (MMBtu)	0	28,435	6.13	**$0**	**$57,259,665**	**$0.00771**

Exhibit 4-9 O&M cost summary for Case 3

INITIAL & ANNUAL O&M EXPENSES					Cost Base (Jun):	2011
Case 3 - NG to Methanol w/ CO_2 capture						
					MWe-net:	0
Methanol (lb/hr):	940,989				Capacity Factor (%):	90
OPERATING & MAINTENANCE LABOR						
Operating Labor						
Operating Labor Rate(base):	39.70	$/hour				
Operating Labor Burden:	30.00	% of base				
Labor O-H Charge Rate:	25.00	% of labor				
			Total			
Skilled Operator	2.0		2.0			
Operator	4.0		4.0			
Foreman	1.0		1.0			
Lab Tech's, etc.	3.0		3.0			
TOTAL-O.J.'s	10.0		10.0			
					Annual Cost	Annual Unit Cost
					$	$/lb
Annual Operating Labor Cost					$4,521,036	$0.001
Maintenance Labor Cost					$20,926,592	$0.003
Administrative & Support Labor					$6,361,907	$0.001
Property Taxes and Insurance					$43,434,792	$0.006
TOTAL FIXED OPERATING COSTS					**$75,244,327**	**$0.010**
VARIABLE OPERATING COSTS						
						$/lb
Maintenance Material Cost					$31,389,888	**$0.00423**
Consumables	Consumption		Unit	Initial Fill		
	Initial Fill	/Day	Cost	Cost		
Water(/1000 gallons)	0	2,454	1.67	**$0**	**$1,349,340**	**$0.00018**
Chemicals						
MU & WT Chem. (lb)	0	14,619	0.27	$0	$1,286,236	$0.00017
Carbon (Mercury Removal) (lb)	0	0	1.63	$0	$0	$0.00000
MEA Solvent (ton)	219	0.31	3,751.70	$821,845	$378,799	$0.00005
Water Gas Shift Catalyst (ft^3)	0	0.00	771.99	$0	$0	$0.00000
Claus Catalyst (ft^3)	w/equip	0.00	203.15	$0	$0	$0.00000
Triethylene Glycol (gal)	0	88.93	6.57	$0	$192,014	$0.00003
NG ATR Reformer Catalyst (ft^3)	14,670	13.40	650.00	$9,535,505	$2,860,651	$0.00039
Methanol Synthesis Catalyst (ft^3)	4,074	3.72	534.68	$2,178,434	$653,530	$0.00009
Aromatics Hydrotreater Catalyst (ft^3)	0	0.00	507.94	$0	$0	$0.00000
Subtotal Chemicals				**$12,535,784**	**$5,371,231**	**$0.00072**
Other						
Supplemental Electricity (for consumption)	0	0	62.33	$0	$0	$0.00000
Gases, N2 etc. (/100scf)	0	0	0.00	$0	$0	$0.00000
L.P. Steam (/1000 pounds)	0	0	0.00	$0	$0	$0.00000
Subtotal Other				**$0**	**$0**	**$0.00000**
Waste Disposal						
Spent Mercury Catalyst (lb.)	0	0	0.65	$0	$0	$0.00000
Flyash (ton)	0	0	0.00	$0	$0	$0.00000
Slag (ton)	0	0	25.11	$0	$0	$0.00000
Subtotal-Waste Disposal				**$0**	**$0**	**$0.00000**
By-products & Emissions						
Sulfur (tons)	0	0	0.00	$0	$0	$0.00000
Supplemental Electricity (for sale) (MWh)	0	**2,758**	-59.59	$0	-$53,994,889	-$0.00728
Subtotal By-Products				**$0**	**-$53,994,889**	**-$0.00728**
TOTAL VARIABLE OPERATING COSTS				**$12,535,784**	**-$15,884,429**	**-$0.00214**
Fuel/Feedstock						
Coal (ton)	0	0	36.57	**$0**	**$0**	**$0.00000**
Natural Gas (MMBtu)	0	315,872	6.13	**$0**	**$636,072,345**	**$0.08574**

4.3 Summary Comparisons

The cost of product (crude methanol) was estimated for each case and the results are listed in Exhibit 4-10 and illustrated in Exhibit 4-11. The RSP for the natural gas case is significantly lower than the coal cases. While the coal case values are dominated by the capital costs, the natural gas case is dominated by the feedstock costs.

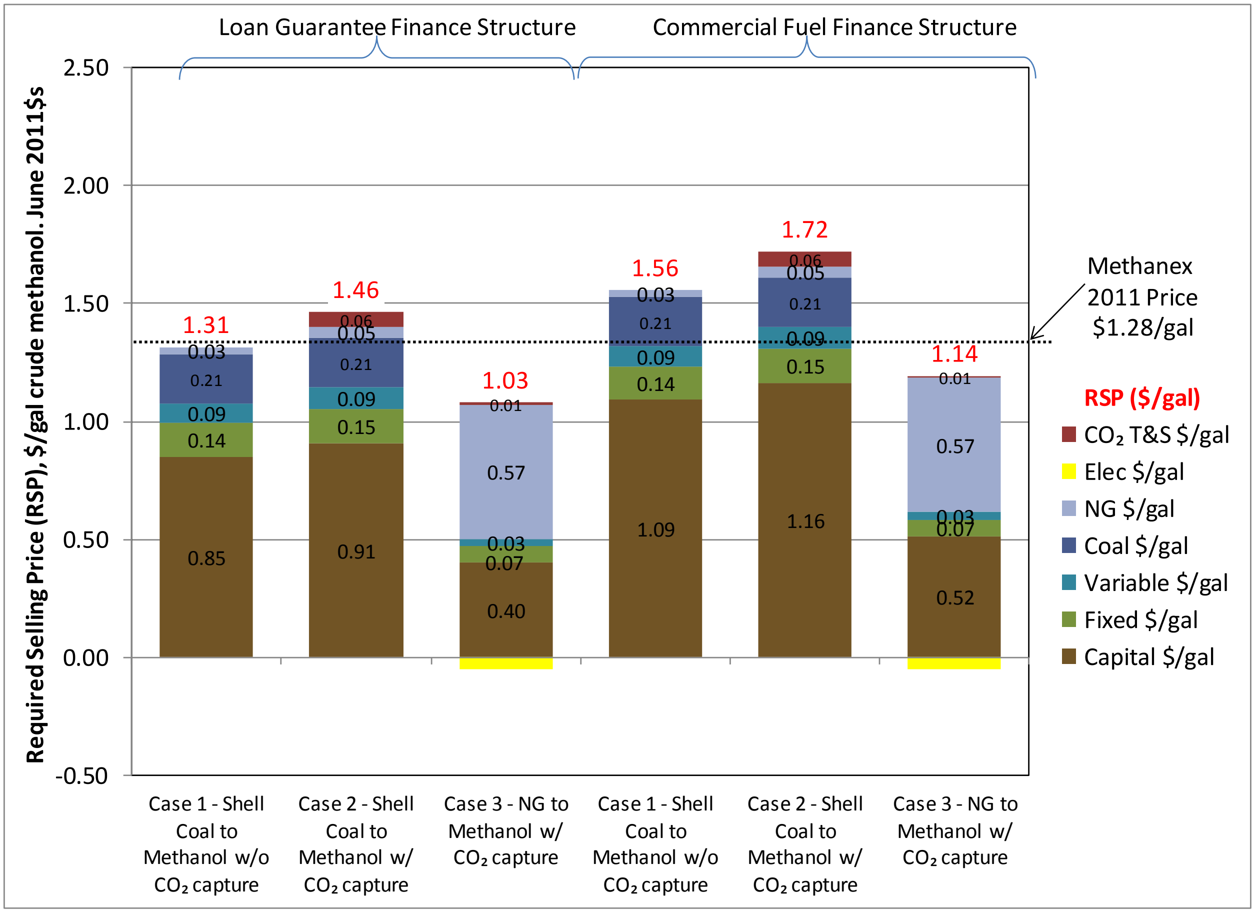
Exhibit 4-10 Cost estimation results

Case	1	2	3
Total Plant Cost (2011 million$)	4,586	4,882	2,172
Total Overnight Cost (2011 million$)	5,615	5,973	2,644
Total As Spent Capital (2011 million$)[B]	6,440 6,631	6,851 7,054	3,033 3,123
RSP[A] Component Details ($/gal)			
Capital[B]	0.85 1.09	0.91 1.16	0.40 0.52
Fixed O&M	0.14	0.15	0.07
Variable O&M	0.09	0.09	0.03
Coal	0.21	0.21	0
Natural gas	0.03	0.05	0.57
Power	0	0	-0.05
CO_2 T&S	0	0.06	0.01
RSP[B] Total ($/gal)	**1.31 1.56**	**1.46 1.72**	**1.03 1.14**
RSP[E] Total ($/ton)	**396.70 469.29**	**441.44 518.67**	**311.17 345.39**
Costs of CO_2 captured[BC] ($/tonne)	**N/A**	**16.36 19.20**	**N/A**
Cost of CO_2 avoided[BD] ($/tonne)	**N/A**	**29.67 32.75**	**N/A**

[A] Capacity factor assumed to be 90 percent.
[B] Values are shown for two financial structures.
The first (lower) value is based on the loan guarantee finance structure.
The second (higher) value is based on the commercial fuels finance structure.
[C] Excluded CO_2 T&S.
[D] Includes CO_2 T&S.
[E] Based on 332.6 gal/tonne or 301.73 gal/ton.
All costs are in June 2011 dollars.

The RSP for the natural gas case is consistent with the information provided in Methanex's 2011 annual report. [24] Specifically, the margin inferred from their financial statements is on par with the weighted average cost of capital used in the PSFM for this study. The average sales prices of methanol reported by Methanex and other vendors (~ 1.28 dollars per gallon or $426 dollars per metric ton in Methanex historical data reference [25]) are higher than the RSP of the natural gas cases in this study because their prices also cover other significant costs of methanol sales and distribution associated with transporting the methanol from production locations to markets and distribution within markets[5]. The Methanex price may also include the costs of purification into refined methanol.

Exhibit 4-11 Components of required selling price



Based on Methanol density of 332.6 gal/tonne or 301.73 gal/ton

The sensitivity of the RSP to the coal and natural gas feedstock prices is illustrated in Exhibit 4-12 and Exhibit 4-13, respectively. The prices used in this study are based on the NETL QGESS Recommended Fuel Prices. [1] Cases 1 and 2 use coal from the Powder River Basin (PRB) region in Montana, priced at $36.57/ton ($2.1351/MMBtu) including delivery to the Midwestern site, as the feedstock. Case 3 uses natural gas, priced at $6.13/MMBtu. Cases 1 and 2 also burn some natural gas to provide additional power required by the process.[6] The cost of the natural gas used for the NGCCs in the coal cases is less than four percent of the RSP values, so it was assumed constant at $6.13/MMBtu for the coal cases in this chart. The expected RSPs for the case using natural gas as feedstock are below those for the coal-based cases because of the lower capital and fixed operating costs. On an energy content basis, the

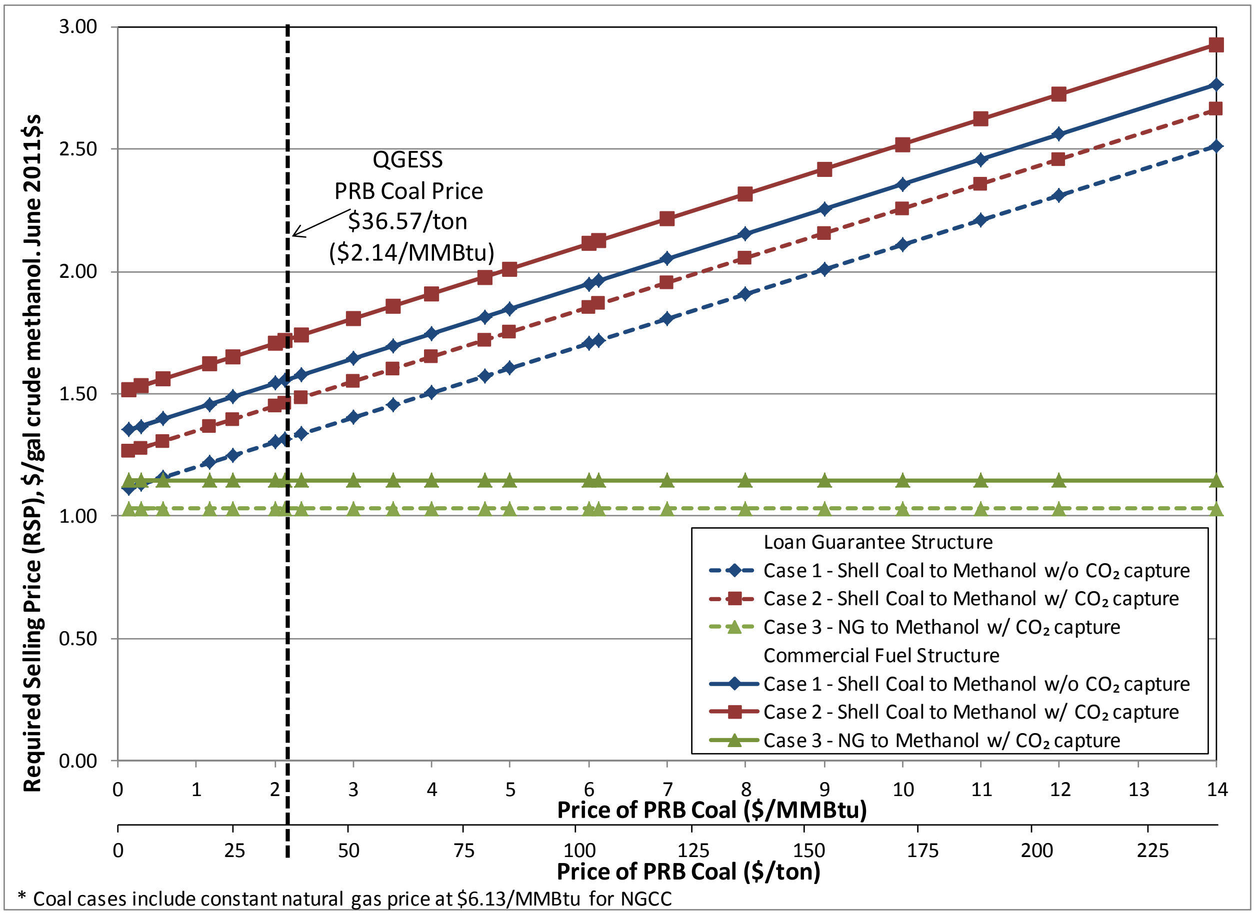
[5] These costs are a key reason why chemicals and fuels projects derivative of potential mega-methanol projects are generally co-located with methanol production.

[6] A natural gas combined cycle plant was used to generate additional power in the coal gasification cases to make the plant approximately power neutral while maximizing the production of crude methanol (i.e., no syngas is diverted for power production).

RSPs for all cases increase at approximately the same rate with increases in the feedstock prices due to the similarity between the feedstock requirements per unit of product. Cases 1 and 2 require 9.75 MMBtu of PRB coal per gallon of Methanol generated and Case 3 requires 9.27 MMBtu of natural gas per gallon of Methanol generated.

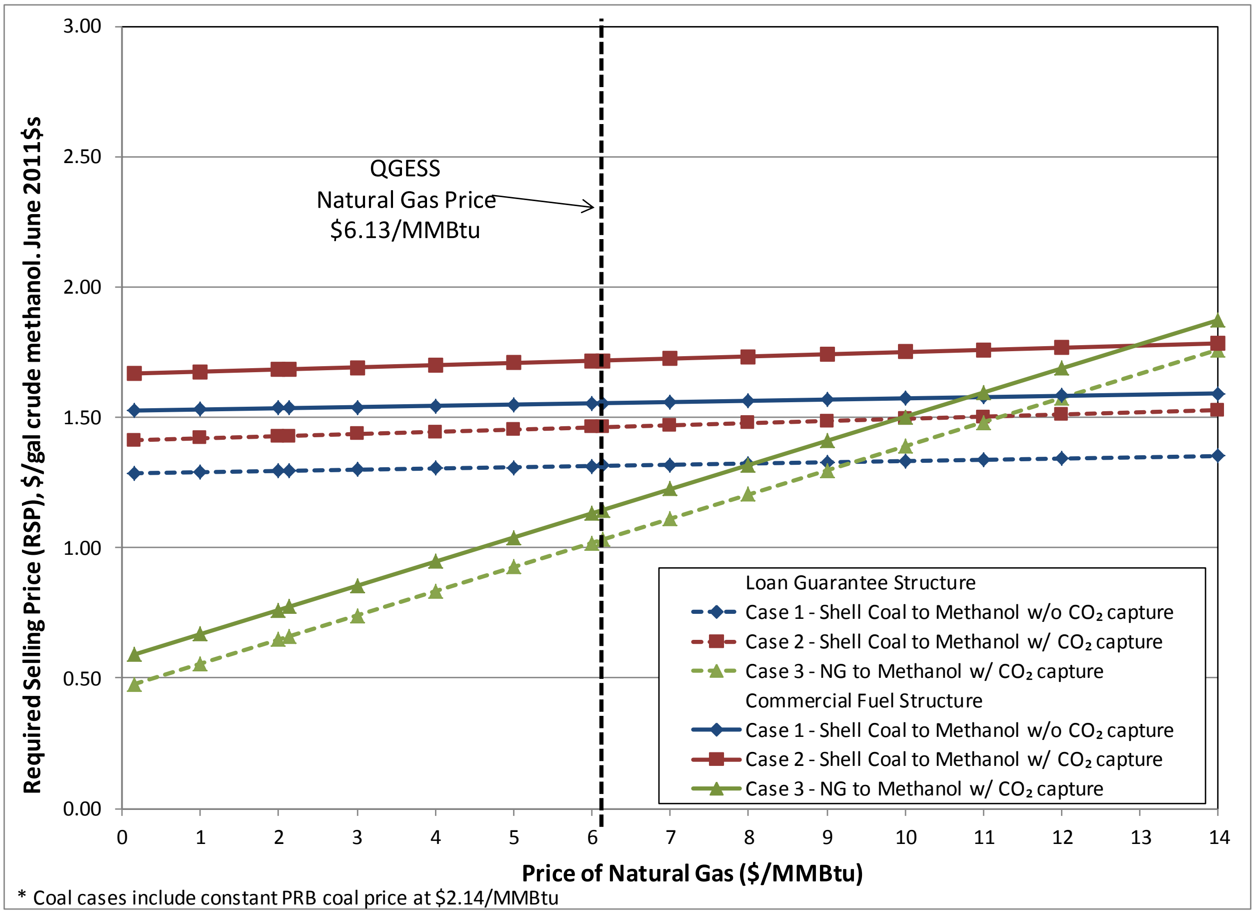
Given the relatively higher cost per MMBtu of natural gas compared to coal and the similar energy input requirements of the conversion process, a 100 percent increase in natural gas prices leads to a 50 to 55 percent increase in RSP, for commercial and loan guarantee financing respectively. However, the same increase of 100 percent in coal prices leads to a 13 to 16 percent increase in RSP for the same financing structures, respectively. These results are consistent with the natural gas–based methanol production being the less capital intense but more operating margin dependent technology choice. Hence, the natural gas route is more exposed to feedstock price volatility.

Exhibit 4-12 Sensitivity of the required selling price to the coal price



Based on Methanol density of 332.6 gal/tonne or 301.73 gal/ton

Exhibit 4-13 Sensitivity of the required selling price to the natural gas price

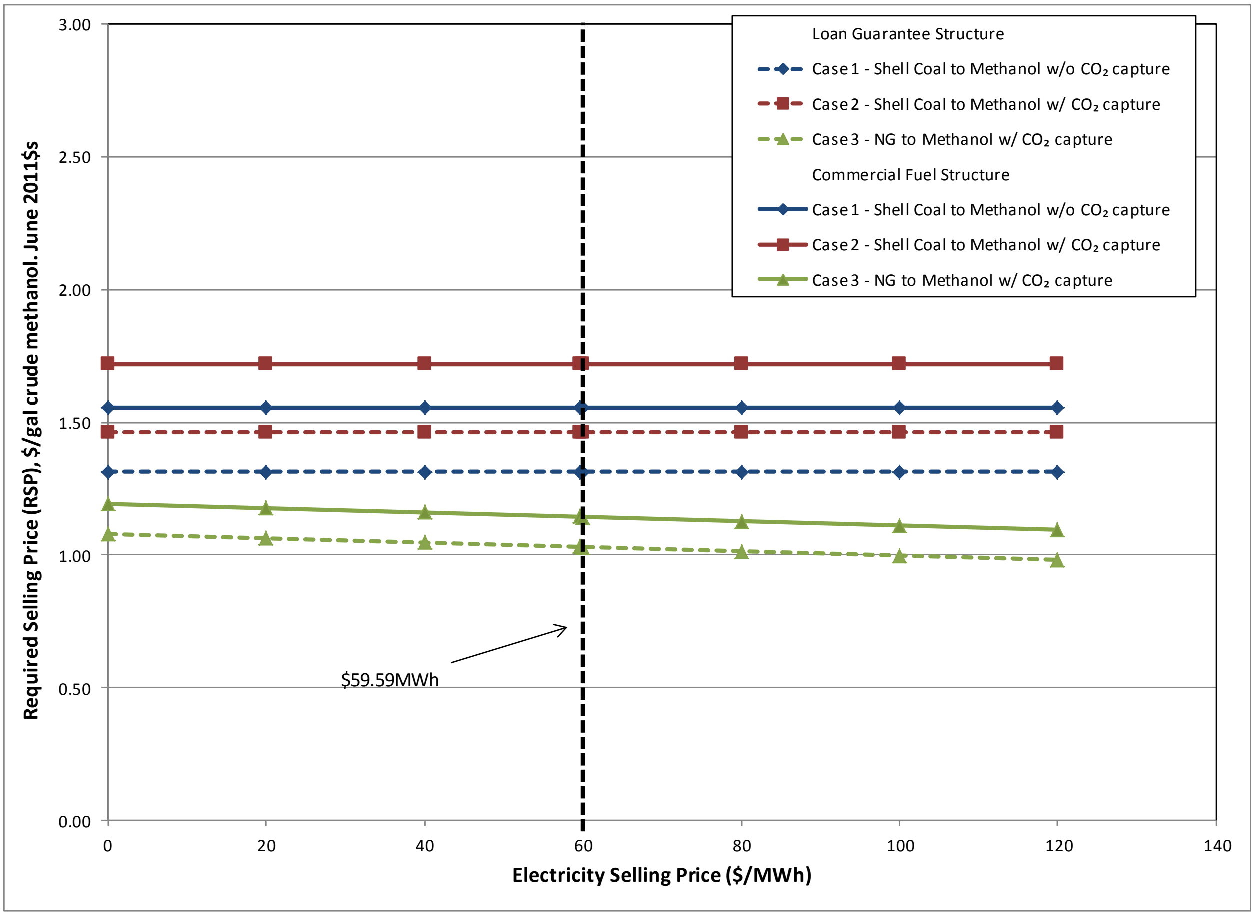


Based on Methanol density of 332.6 gal/tonne or 301.73 gal/ton

The heat recovery from the exothermic methanol synthesis reaction combined with burning the excess tail gases in a combustion turbine result in a substantial amount of excess electricity being generated in the natural gas case. If this electricity is sold, the revenue is applied to the cash flow and results in a lower RSP. The sensitivity of the RSPs to the selling price of electricity is illustrated in Exhibit 4-14. The RSPs for the coal feedstock cases remain constant but are shown for comparison. The RSPs for the natural gas feedstock case decease by approximately 0.08 cents per gallon for each dollar per MWh increase in electricity selling price.

While the excess power can be sold to the grid; the sale may be at a steep discount as entities that are negotiating a power purchase agreement will know the power production is an inherent by-product of core methanol production operations. Consequently, the actual achieved transfer price for excess power will be a significant risk in natural gas feedstock projects and would be highly project dependent. However, as shown in Exhibit 4-14, the impact of reduced electricity revenues has a fairly small impact on the methanol RSP. Decreasing the electricity sell price to $0/MWh only increases the methanol RSP by about $0.05/gal.

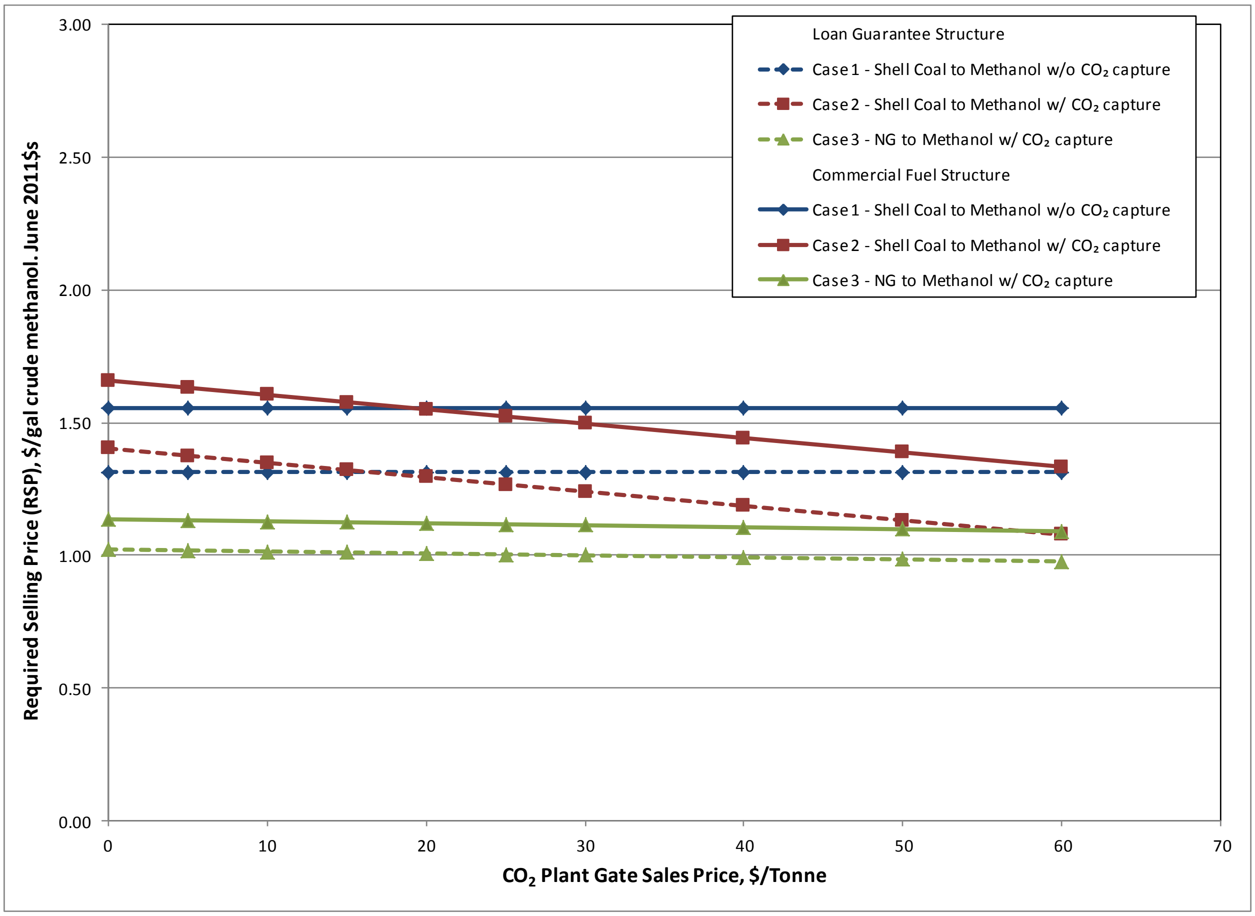
Exhibit 4-14 Sensitivity of the required selling price to the electric selling price



Based on Methanol density of 332.6 gal/tonne or 301.73 gal/ton

Options for carbon dioxide sequestration include both storage in a saline reservoir (i.e., carbon capture and sequestration (CCS)) and usage in enhanced oil recovery (EOR) (i.e., carbon capture, utilization and storage (CCUS)). The EOR option may be attractive even without the passage of carbon regulations. The impact of selling the captured CO_2 for EOR or other uses at various plant gate sale prices was estimated, and the results shown in Exhibit 4-15. The horizontal lines represent the without-capture case RSP values. As the plant gate sale price increases, the RSP values decrease and approach the without-capture values. The plant gate sale price at the point where each capture case line crosses the corresponding without-capture line is equal to the cost of CO_2 captured for that capture case. The cost of capture can be interpreted as the breakeven plant gate sale price where the cost of capture equals the revenue generated by selling the recovered CO_2.

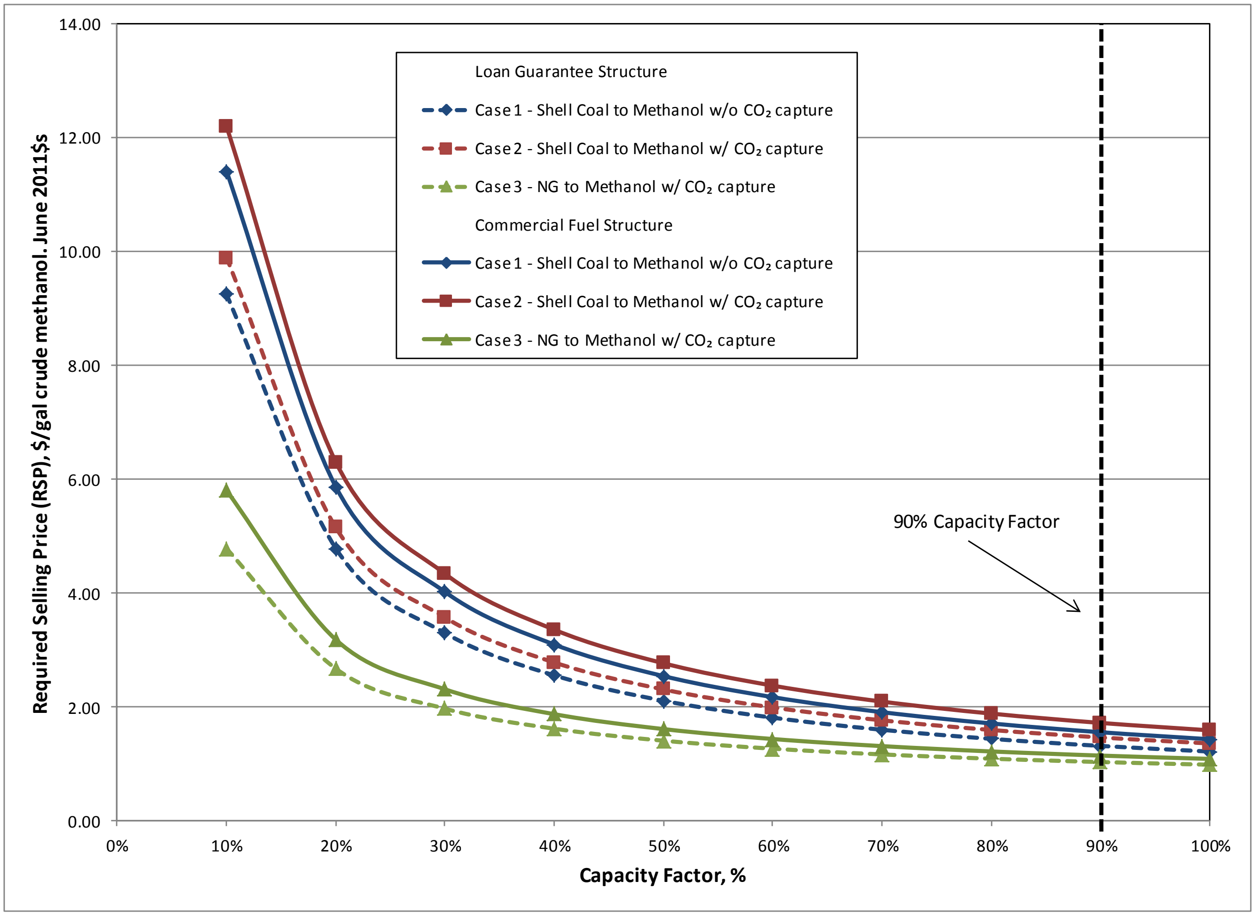
Exhibit 4-15 Sensitivity of RSP to CO_2 plant gate sales price



Based on Methanol density of 332.6 gal/tonne or 301.73 gal/ton

The capacity factor was assumed to be 90 percent for the cases in this study. The sensitivity of the RSP to the capacity factor is illustrated in Exhibit 4-16. The RSPs for the coal cases increase at a faster rate than the natural gas cases as the capacity factor falls because the coal-based cases have more dollars of capital underutilized when the capacity factor is reduced.

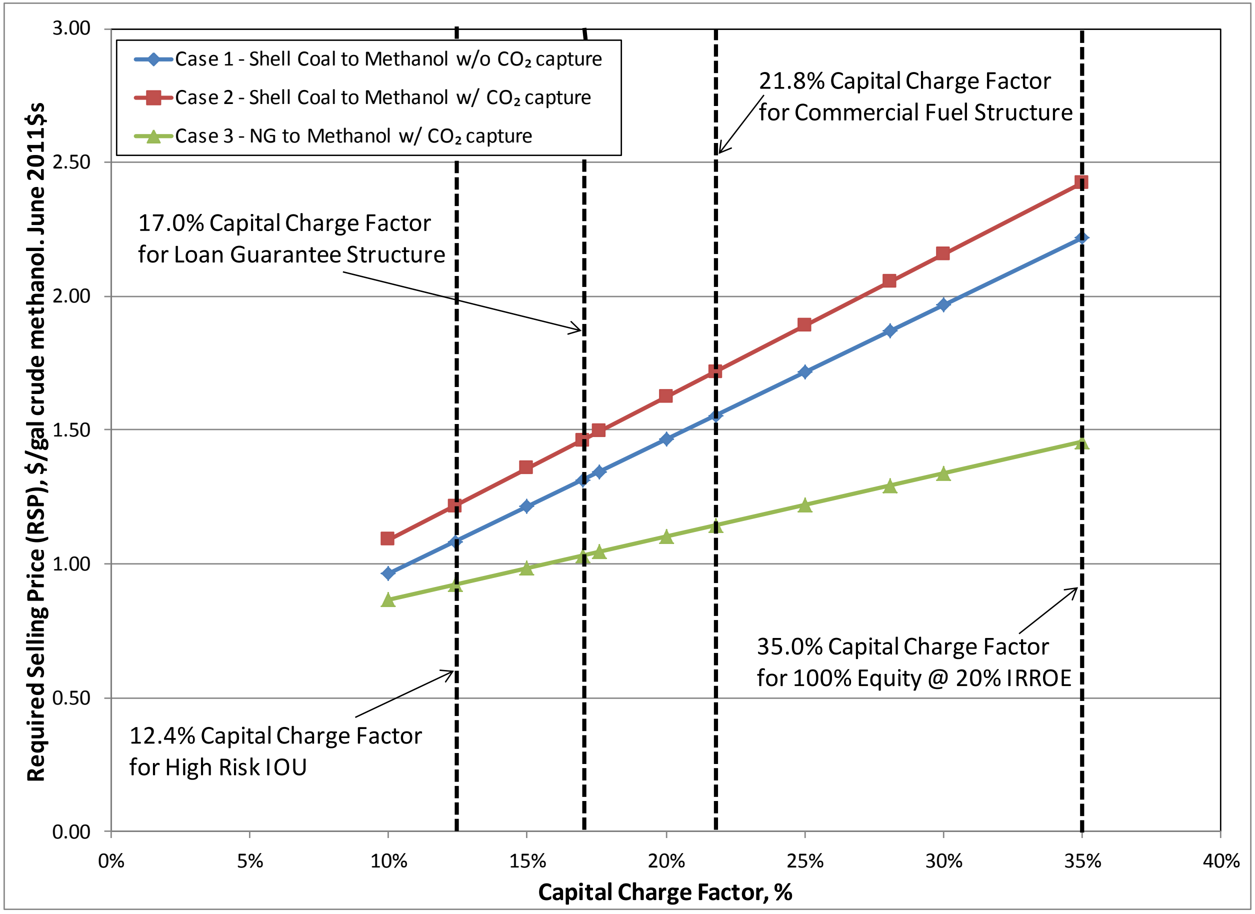
Exhibit 4-16 Sensitivity of the required selling price to the capacity factor



Based on Methanol density of 332.6 gal/tonne or 301.73 gal/ton

Two financial structures were assumed for calculating RSPs in this activity as described in Section 4.1.4. These structures are based on typical values for fuel projects with and without loan guarantees or government subsidies. The assumed values were used in the PSFM to establish capital charge factors (CCF), the portion of the total overnight capital cost to include in the annual cost of producing a product, for each financial structure. The sensitivity of the RSP to the CCF is illustrated in Exhibit 4-17. The RSP values were calculated for CCFs ranging from 10 percent to 35 percent (the value estimated for a project assuming 100 percent equity and 20 percent internal rate of return on equity). The RSP values calculated for a CCF of 12.4 percent (the value estimated for a high risk investor-owned utility (IOU) project assuming 45 percent debt at 5.5 percent interest and 55 percent equity and 12 percent internal rate of return on equity) are also included in the chart. The RSPs for the coal cases increase at a faster rate than the natural gas cases as the capital charge factor increases because the coal-based cases have more dollars of capital to be included in the cost of production. The coal case values increase by approximately 5 cents per gallon for each one absolute percent increase in the capital charge factor. The natural gas case values increase by approximately 2 cents per gallon for each one absolute percent increase in the capital charge factor. Loan guarantees and/or government subsidies could reduce the RSPs for the coal-based cases closer to the values of commercial natural gas cases.

Exhibit 4-17 Sensitivity of the required selling price to the capital charge factor



Based on Methanol density of 332.6 gal/tonne or 301.73 gal/ton

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