COMPARISON OF COMMERCIAL, STATE-OF-THE-ART, FOSSIL-BASED HYDROGEN PRODUCTION TECHNOLOGIES

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ACRONYMS AND ABBREVIATIONS

AACE	Association for the Advancement of Cost Engineering
abs	Absolute
AC	Alternating current
acfm	Actual cubic feet per minute
ACI	Activated carbon injection
ADIP	Aqueous di-isopropanol
AF	Availability factor
AFB	Ash Agglomerating Fluidized Bed
AGR	Acid gas removal
ANSI	American National Standards Institute
Ar	Argon
Aspen	Aspen Plus®
ASU	Air separation unit
atm	Atmosphere
ATR	Autothermal reforming
B&V	Black & Veatch
BACT	Best Available Control Technology
Bara	Bar absolute
barg	Bar gauge
BDL	Blowdown losses
BEC	Bare erected cost
BEP	Breakeven emissions penalty
BFD	Block flow diagram
BFW	Boiler feedwater
BOP	Balance of plant
BSP	Breakeven sales price
Btu	British thermal unit
Btu/hr	British thermal units per hour
Btu/kWh	British thermal units per kilowatt hour
Btu/lb	British thermal units per pound
Btu/scf	British thermal units per standard cubic foot
$CaCO_3$	Calcium carbonate
$CaSO_4$	Calcium sulfate
CB&I	Chicago Bridge & Iron Company
CCS	Carbon capture and storage
CCUS	Carbon capture, utilization, and storage
CDR	Carbon dioxide recovery
CF	Capacity factor
CFR	Code of Federal Regulations
CGCU	Cold gas cleanup
CGE	Cold gas efficiency
CH_4	Methane
CH_4S	Methanethiol
Cl^-	Chloride ion
CL	Closed-loop
cm	Centimeter
CO	Carbon monoxide
CO_2	Carbon dioxide
CO_2e	Carbon dioxide equivalent
COE	Cost of electricity
COS	Carbonyl sulfide
CS	Carbon steel
CT	Combustion turbine
CTG	Combustion turbine-generator
CWP	Circulating water pump
CWR	Cooling water return
CWS	Circulating water system
CWT	Cold water temperature
DAC	Direct air capture
db, DB	Dry basis
DBQ	Dry Bottom Quench
DC	Direct current
DCS	Distributed control system
DI	De-ionized
DIPA	Diisopropanolamine
DLN	Dry low NOx
DME	Di-methyl ether
DOE	Department of Energy
DSC	Dry Syngas Cooler
DSI	Dry sorbent injection
E-Gas™	CB&I gasifier technology
EAF	Equivalent availability factor
ECUST	East China University of Science and Technology
EGU	Electric utility steam generating unit
ELG	Effluent Limitation Guidelines

EMF	Emission modification factors
Eng'g CM H.O.& Fee	Engineering, construction management, home office and fees
EOR	Enhanced oil recovery
EPA	Environmental Protection Agency
EPC	Engineering, procurement and construction
EPCC	Engineering, procurement and construction cost
EPRI	Electric Power Research Institute
ERC	Emission reduction credit
ESP	Electrostatic precipitator
ETE	Effective thermal efficiency
FC	Flow controller
FD	Forced draft
FE	Fossil Energy
FG	Flue gas
FGD	Flue gas desulfurization
FRC	Flow rate controller
FRP	Fiberglass-reinforced plastic
FSQ	Full-slurry quench
ft	Foot, feet
ft^3	Cubic feet
FW	Feedwater
GADS	Generating Availability Data System
gal	Gallon
GDP	Gross domestic product
GEP	General Electric Power
GFB	Gas-fired boiler
GHG	Greenhouse gas
GJ	Gigajoule
GJ/hr	Gigajoules per hour
gpd	Gallons per day
gpm	Gallons per minute
gr/100 scf	Grains per one hundred standard cubic feet
GSR	Gas switching reforming
Gt	Gigatonne
GWP	Global warming potential
h, hr	Hour
H_2	Hydrogen
H_2O	Water
H_2S	Hydrogen sulfide
HCl	Hydrochloric acid
HCO_3	Bicarbonate
HDPE	High-density polyethylene
Hg	Mercury
HHV	Higher heating value
hp	Horsepower
HP	High-pressure
HRSG	Heat recovery steam generator
HSS	Heat stable salt
HTL	Hantian Lu
HVAC	Heating, ventilating, and air conditioning
HWT	Hot water temperature
HX	Heat exchanger
Hz	Hertz
I&C	Instrumentation and control
ICR	Information Collection Request
ID	Induced draft
IGCC	Integrated gasification combined cycle
IGV	Inlet guide vane
in.	Inch
in. Hg	Inch mercury
IOU	Investor-owned utility
IP	Intermediate pressure
IPCC AR5	International Panel on Climate Change 5th Assessment Report
IPM	Integrated Planning Model
IRROE	Internal rate of return on equity
ISO	International Organization for Standardization
kg/GJ	Kilograms per gigajoule
kg/hr	Kilograms per hour
kg/s	Kilograms per second
kgmol	Kilogram mole
kgmol/hr	Kilogram moles per hour
kJ	Kilojoule
kJ/hr	Kilojoules per hour
kJ/kg	Kilojoules per kilogram
kJ/m^3	Kilojoules per cubic meter
kJ/Nm^3	Kilojoules per normal cubic meter
km	Kilometer

KO	Knockout
kV	Kilovolt
kW, kWe	Kilowatt electric
kWh	Kilowatt-hour
kWt	Kilowatt thermal
LAER	Lowest Achievable Emission Rate
lb	Pound
lb/gal	Pound per gallon
lb/ft^2	Pounds per square foot
lb/ft^3	Pounds per cubic foot
lb/hr	Pounds per hour
lb/MMBtu	Pounds per million British thermal units
lb/MWh	Pounds per megawatt hour
lb/s	Pounds per second
lb/TBtu	Pounds per trillion British thermal units
lbmol/hr	Pound moles per hour
LCA	Life cycle analysis
LCC	Levelized capital cost
LCOH	Levelized cost of hydrogen
LCV	Low calorific value
LFP	Levelized fuel price
LHV	Lower heating value
LIW	Loss-in-weight
LNB	Low NOx burner
LOM	Levelized operating and maintenance cost
LP	Low pressure
lpm	Liters per minute
LTHR	Low temperature heat recovery
m	Meter
M	Thousand
m/min	Meters per minute
m^3/min	Cubic meters per minute
MA-ATR	Membrane assisted autothermal reforming
MAC	Main air compressor
MAF	Moisture and ash free
MATS	Mercury and Air Toxics Standards
MCR	Maximum continuous rate
MCSF	Multi-component slurry gasifier
MDEA	Methyl diethanolamine
MEA	Monoethanolamine
mi	Mile
MISO	Midwest Independent System Operator
MJ	Megajoule
MJ/Nm^3	Megajoules per normal cubic meter
MJ/scm	Megajoule per standard cubic meter
mm	Millimeter
MM	Million
MMacf	Million actual cubic feet
MMBtu	Million British thermal units
MMBtu/hr	Million British thermal units per hour
MNQC	Multi Nozzle Quiet Combustor
MMSCFD	Million standard cubic feet per day
mol%	Percent by mole
MPa	Megapascal
MSW	Municipal solid waste
MVA	Mega volt-amps
MW	Megawatt
MWe	Megawatt electric
MWh	Megawatt-hour
N_2	Nitrogen
N/A	Not applicable
NaCl	Sodium chloride
NaOH	Sodium hydroxide
NEMA	National Electrical Manufacturers Association
NERC	North American Electric Reliability Council
NETL	National Energy Technology Laboratory
NFPA	National Fire Protection Association
NG	Natural gas
NGCC	Natural gas combined cycle
NH_3	Ammonia
NH_4Cl	Ammonium chloride
Nm^3	Normal cubic meter
Nm^3/hr	Normal cubic meter per hour
NOAK	Nth-of-a-kind
NO_2	Nitrogen dioxide
NOx	Nitrogen oxides

NSPS	New Source Performance Standards
NSR	New Source Review
NTU	Nephelometric turbidity unit
O_2	Oxygen
O&M	Operation and maintenance
OEM	Original equipment manufacturers
OFA	Overfire air
O-H, OH	Overhead
OP/VWO	Over pressure/valve wide open
P	Absolute pressure
PA	Primary air
PAC	Powdered activated carbon
PC	Pulverized coal
PEM	Proton-exchange membrane
p.f.	Power factor
ph	Phase
pH	Power of hydrogen
PM	Particulate matter
POTW	Publicly-owned treatment works
POX	Partial oxidation
ppb	Parts per billion
ppm	Parts per million
ppmv	Parts per million volume
ppmvd	Parts per million volume, dry
ppmw	Parts per million weight
ppmwd	Parts per million weight, dry
ppt	Parts per trillion
PRB	Powder River Basin
PSA	Pressure swing absorption
PSD	Prevention of Significant Deterioration
psi	Pounds per square inch
psia	Pound per square inch absolute
psid	Pound per square inch differential
psig	Pound per square inch gauge
QGESS	Quality Guidelines for Energy System Studies
R&D	Research and development
R+Q	Radiant plus quench
RD&D	Research, development, and demonstration
RH	Reheater
RO	Reverse osmosis
RR	Ramp rate
SC	Supercritical
scfh	Standard cubic feet per hour
scfm	Standard cubic feet per minute
Sch.	Schedule
SCOT	Shell Claus Off-gas Treating
SCR	Selective catalytic reduction
SCWG	Supercritical water gasification
SDA	Spray dryer absorber
SDE	Spray dryer evaporator
SGC	Synthesis gas cooler
Shell	Shell Global Solutions
SMR	Steam methane reforming
SNCR	Selective non-catalytic reduction
SNG	Synthetic natural gas
SO_2	Sulfur dioxide
SO_3	Sulfur trioxide
SRU	Sulfur recovery unit
SS	Stainless steel
STE	Steam turbine efficiency
STG	Steam turbine generator
STHR	Steam turbine heat rate
STP	Steam turbine power
SubC	Subcritical
SWS	Sour water stripper
T	Temperature
T&S	Transport and storage
TASC	Total as-spent cost
TBtu	Trillion British thermal units
TDS	Total dissolved solids
TEG	Triethylene glycol
TEWAC	Totally Enclosed Water-to-Air Cooled
TG	Tail gas
TGTU	Tail gas treating unit
TI	Thermal input
TO	Thermal output
TOC	Total overnight cost
tonne	Metric ton (1,000 kg)
TPC	Total plant cost
tpd	Ton per day
tph	Tons per hour

U.S.	United States
UAN	Urea ammonium nitrate
UCC	United Conveyor Corporation
USC	Ultra-supercritical
V	Volt
V-L	Vapor liquid portion of stream (excluding solids)
vol%	Percent by volume
VSA	Vacuum swing adsorption
VT	Voltage transformers
WGS	Water gas shift
wt%	Percent by weight
wtr	Water
yr	Year
ZLD	Zero liquid discharge
$/GJ	Dollars per gigajoule
$/kW	Dollars per kilowatt
$/MMBtu	Dollars per million British thermal units
$M	Millions of dollars
µS/cm	Micro Siemens per cm
°C	Degrees Celsius
°F	Degrees Fahrenheit
5-10s	50-hour work week

EXECUTIVE SUMMARY

This report presents an independent assessment of the cost and performance of select hydrogen production plants utilizing fossil fuel resources as the primary feedstocks – specifically, natural gas (NG), steam methane reforming (SMR), NG autothermal reforming (ATR), coal gasification, and coal/biomass co-gasification – using a systematic, transparent technical and economic approach. Study cases were selected to reflect the capabilities of current, commercial technologies within plant configurations, and at scales, representative of next commercial offerings facing no fundamental research and development (R&D) obstacles. Additionally, several areas of R&D are identified as potential pathways for performance improvements and cost reductions.

Attributional global warming potential (GWP) profiles of the impactful energy and material streams entering and exiting the plant boundaries were used to develop life cycle greenhouse gas (GHG) emissions results for each case based on the quantity of each stream. The results are deemed to be commensurate with an International Organization for Standardization (ISO) 14040/14044 life cycle analysis (LCA) [1, 2] and are expressed as carbon dioxide equivalents (CO_2e). A life cycle GHG emissions target of 0 pounds (lb) CO_2e/lb H_2 produced, or "net-zero," was targeted in one plant.

The cost and performance of fossil fuel-based hydrogen production technologies represented in this report are important inputs to assessments and determinations of technology combinations to be utilized to meet the projected demands of the nascent hydrogen economy. In addition to informing technology comparisons, the reference plant configurations found in this report provide perspective for regulators and policy makers. From an R&D perspective, this report may be used to assess goals and metrics and to provide a consistent basis for comparing developing technologies.

Six hydrogen plant configurations are analyzed in this report. A summary of the configurations is shown in Exhibit ES-1:

- Three natural gas reforming configurations—two NG SMR cases (with and without carbon dioxide [CO_2] capture) and one NG ATR case (with CO_2 capture).
- Three gasification configurations — two coal gasification cases (with and without CO_2 capture) and one coal/biomass co-gasification case (with CO_2 capture).

The CO_2 capture strategy employed for each case recovers greater than 90 percent of the carbon entering the plant boundary, minus the slag formed in gasification cases. High levels of capture are achieved by using a combination of water gas shift (WGS) reactors and solvent-based CO_2 separation technologies. The solvent technologies used include methyl diethanolamine (MDEA) (SMR and ATR), Shell Cansolv (SMR), and two-stage Selexol (coal and coal/biomass gasification).

This report reflects varying degrees of technology vendor input for each case. However, the final assessment of performance and cost was determined independently and is not endorsed by individual vendors.

Exhibit ES-1. Case configuration summary

Case[A]	Plant Type	Feedstock(s)	Reformer Type	Gasifier Type	CO_2 Capture (%)	H_2 Purification	Hydrogen Production Capacity
1	Reforming	Natural Gas	SMR	N/A	0	Pressure Swing Absorber (PSA)	200 MMSCFD 483,000 kg/day 44,400 lb/hr
2					96.2		
3			ATR		94.5		274 MMSCFD 660,000 kg/day 60,600 lb/hr
4	Gasification	Illinois No. 6 Coal	N/A	Shell[B]	0		
5					92.5		
6		Illinois No. 6 Coal/ Torrefied Woody Biomass			92.6		55 MMSCFD 133,000 kg/day 12,200 lb/hr

[A]All plants in this report are assumed to be located at a generic plant site in the midwestern United States.
[B]The Shell gasifier has been used in multiple prior NETL studies. As of May 2018, Air Products has acquired the coal gasification technology licensing business from Shell. To be consistent with prior NETL studies and avoid confusion, the gasifier will continue to be labeled the “Shell” gasifier.

To ensure methodologically-sound, consistent, and transparent technology assessments and comparisons, NETL relies upon its Quality Guidelines for Energy System Studies (QGESS) reports, which provide guidance on topics such as recommended feedstock specifications and pricing, [3] [4] recommended performance modeling assumptions, [5] and guidance for cost estimation methodology. [6]

The methodology for developing the performance results presented in this report included performing steady-state simulations of the hydrogen plant configurations at the design capacity using the Aspen Plus® (Aspen) process modeling software. The major plant equipment performance and process limits were based on published reports, information obtained from vendors and users of the technology, performance data from design/build utility projects, and/or best engineering judgment. [5] Mass and energy balance data from the Aspen models were used to size major pieces of equipment, which formed the basis for developing the cost estimates presented.

Full, bottom-up estimates for each case in this study were not pursued. Instead, the capital and operation and maintenance (O&M) costs for each of the cases were estimated using a combined bottom-up and scaling approach. Black & Veatch (B&V) leveraged an in-house database and conceptual estimating models. Costs were further calibrated using a combination of adjusted vendor-furnished data and scaled estimates from previous design/build projects. In addition, cost scaling was performed on select plant areas using reference costs from prior NETL studies according to the 2019 revision of the QGESS document "Capital Cost Scaling Methodology: Revision 4 Report." [7] The cost results are reported in 2018 dollars.

The baseline NG and coal costs for this analysis are specified in the 2019 revision of the QGESS report on "Fuel Prices for Selected Feedstocks in National Energy Technology Laboratory [NETL] Studies." [8] The levelized price for Illinois No. 6 coal delivered to the Midwest is $2.11/ gigajoule (GJ) ($2.23/million British thermal units [MMBtu]), on a higher heating value (HHV) basis and in 2018 United States (U.S.) dollars. The levelized price for natural gas delivered to the Midwest is $4.19/GJ ($4.42/MMBtu), on an HHV basis and in 2018 U.S. dollars. The levelized price of torrefied, woody biomass was estimated using an NETL model based on vendor data and is $5.14/GJ ($5.43/MMBtu), on an HHV basis and in 2018 U.S. dollars.

The cost metric used in this report is the levelized cost of hydrogen (LCOH) reported in real 2018 dollars, which is the revenue that must be received by the producer per kilogram of hydrogen produced to meet the desired return on equity after meeting all debt and tax obligations and operating expenses. Detailed information pertaining to LCOH calculations is available in the 2019 revision of the QGESS report "Cost Estimation Methodology for NETL Assessment of Power Plant Performance." [6] The cost of CO_2 transport and storage (T&S), on an equivalent dollar per kilogram basis, is added to the LCOH and represents a 62 km (100 mile) CO_2 pipeline and storage in a deep saline formation in the Midwest.[a] On a unit basis, the cost of CO_2 T&S applied is $10 per tonne ($9/ton) of CO_2.

[a] Estimated using the Office of Fossil Energy (FE)/NETL CO_2 Transport Cost Model and the FE/NETL CO_2 Saline Storage Cost Model. Additional detail on development of these costs is available in the 2019 revision of the QGESS report "Carbon Dioxide Transport and Storage Costs in NETL Studies." [56]

The LCOH results shown for the cases considered in this study are not intended to reflect all the potential market pressures experienced by plants operating today, or the price consumers can expect to pay. Rather, the primary focus is on a sound, transparent, consistent methodology to develop those results, built with industry and vendor input, which ultimately leads to an independent benchmark for the cases considered. This outcome is of significant value to internal and external stakeholders.

LCA models considered a cradle-to-gate system boundary with a functional unit of 1 lb of hydrogen at the plant gate. Results were reported as GWP in units lb CO_2e/lb H_2 for all hydrogen cases considered. Similar to the approach for the cost analysis, full models of each case were not pursued. Instead, the general approach to generating the GHG emissions results in this study was to leverage the engineering results to identify the upstream energy inputs needed, and to apply GHG emissions factors to those inputs in order to estimate cradle-to-gate emissions. Facility level emissions were directly estimated from the engineering analyses and are also part of the overall cradle-to-gate emissions.

The upstream NG emissions for this analysis are specified in the NETL natural gas baseline report [9], and extended to the year 2017 [10], and the boundaries are from natural gas extraction through transmission (i.e., for a large-scale industrial user). Electricity emissions are modeled using the NETL electricity baseline [11, 12], represented by the U.S. average consumption mix at user. Emissions from the upstream production of biomass, specifically the torrefied Southern Yellow Pine (SYP) used in this study, followed work previously done for the Connecticut Center for Advanced Technology (CCAT) [13] with some modifications. The upstream production of coal leverages modeling and default assumptions from NETL's Supercritical Pulverized Coal (SCPC) Power Plant study [14], assuming Illinois underground bituminous coal. Emissions for storage of CO_2 in saline aquifer followed modeling and assumptions used in the SCPC Power Plant Study [14]. Overall, the greenhouse gas emissions data is representative of the time period 2016 to 2017.

Additionally, the cases considered are not meant to depict all available technologies, process arrangements, and potential scales. Therefore, the study results only reflect the technologies and arrangements considered and can change based on alternatives. Ultimately, selection of new hydrogen production technologies will depend on many factors, including, but not limited to:

- Capital and operating costs
- Overall efficiency
- Operational flexibility (e.g., ramp rate, turndown, start-up time)
- Feedstock prices
- Project financial requirements
- Availability, reliability, and environmental performance
- Current and potential regulations governing air, water, and solid waste discharges from fossil-based hydrogen plants
- Specific site and application constraints and requirements

RESULTS ANALYSIS

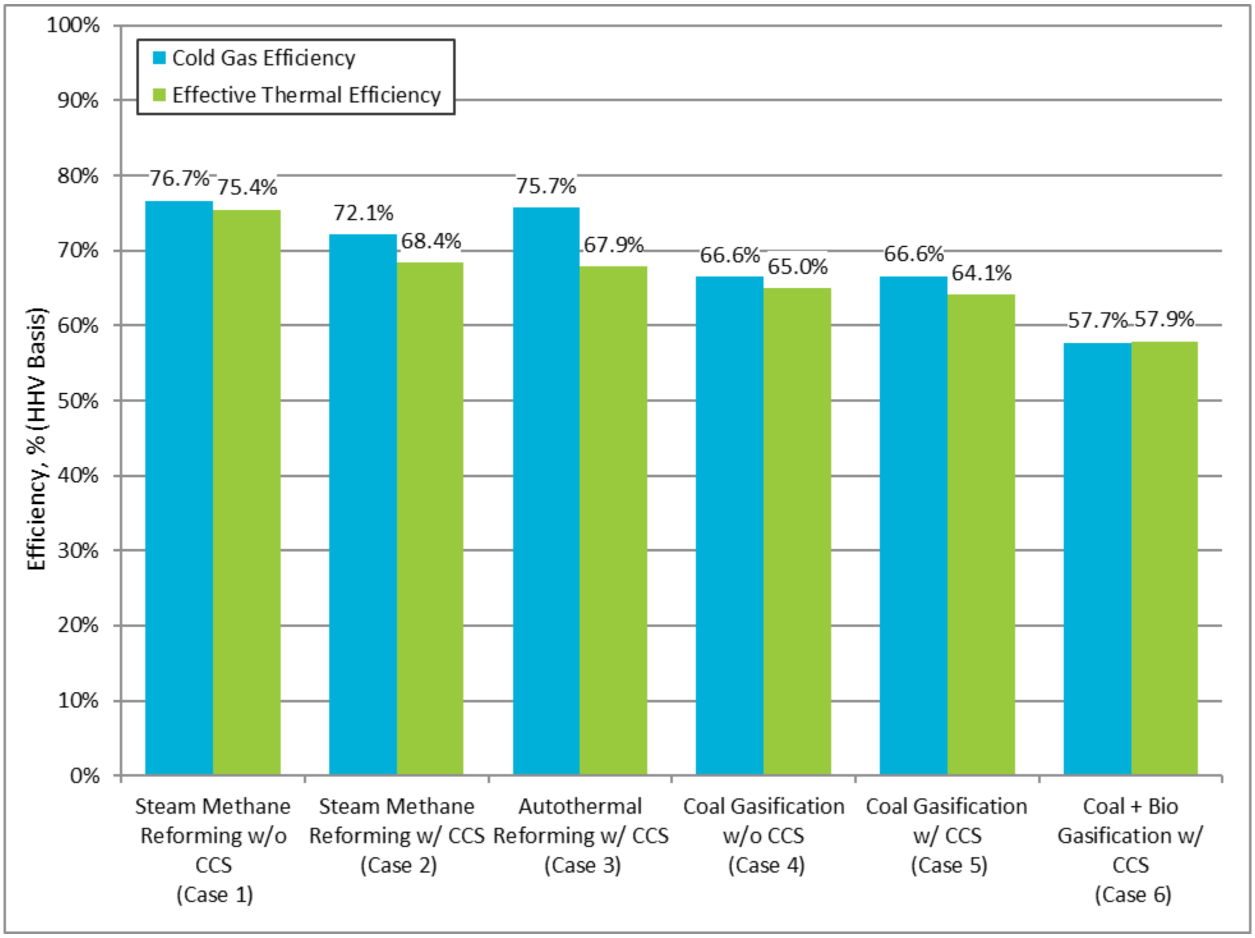
Exhibit ES-2 shows the performance and environmental profile summary for all cases. A graph of the net plant efficiency (HHV basis) is provided in Exhibit ES-3. The results are described in greater detail in the following paragraphs.

Exhibit ES-2. Performance summary and environmental profile for all cases

Case Name	Reforming			Gasification		
	SMR		ATR	Shell Gasifier		
	1	2	3	4	5	6
CO_2 Capture	0%	96.2%	94.5%	0%	92.5%	92.7%
Capacity Factor	90%	90%	90%	80%	80%	80%
Hydrogen Production Rate (lb/hr)	44,369	44,369	60,627	60,622	60,627	12,194
Gross Power Output (MWe)	0	0	0	89	109	37
Auxiliary Power Requirement (MWe)	13	41	110	114	148	37
Net Power Output (MWe)	-13	-41	-110	-25	-39	1
Coal Flowrate (lb/hr)	N/A	N/A	N/A	467,308	467,308	65,917
Biomass Flowrate (lb/hr)	N/A	N/A	N/A	N/A	N/A	50,750
Natural Gas Flowrate (lb/hr)	156,482	166,387	213,694	N/A	N/A	N/A
HHV Thermal Input (kW_t)	1,031,068	1,096,328	1,408,040	1,597,711	1,597,711	370,367
HHV Effective Thermal Efficiency (%)	75.4%	68.4%	67.9%	65.0%	64.1%	57.9%
HHV Cold Gas Efficiency (%)	76.7%	72.1%	75.7%	66.6%	66.6%	57.7%
Raw Water Withdrawal (gpm)	1,457	2,727	3,720	3,188	3,638	1,035
Process Water Discharge (gpm)	63	586	775	672	773	220
Raw Water Consumption (gpm)	1,395	2,140	2,945	2,516	2,866	814
CO_2 Emissions (lb/MMBtu)	118	5	7	199	15	15
SO_2 Emissions[A] (lb/MMBtu)	0.000	0.000	0.000	0.027	0.000	0.000
NOx Emissions (lb/MMBtu)	0.003	0.002	0.000	0.017	0.007	0.010
PM Emissions (lb/MMBtu)	0.000	0.000	0.000	0.007	0.007	0.007
Hg Emissions (lb/TBtu)	0.000	0.000	0.000	0.571	0.571	0.348

[A] Trace amounts of sulfur emissions may exist in the flue gas stream to the stack in capture cases

Exhibit ES-3. Net plant efficiency (HHV basis)



Cold gas efficiency (CGE) is calculated as the ratio of the thermal content of the hydrogen product to the thermal input to the system. The effective thermal efficiency (ETE) is calculated similarly to the CGE but the numerator is adjusted for the net electrical power imported or exported. Therefore, importing electrical power lowers the ETE.

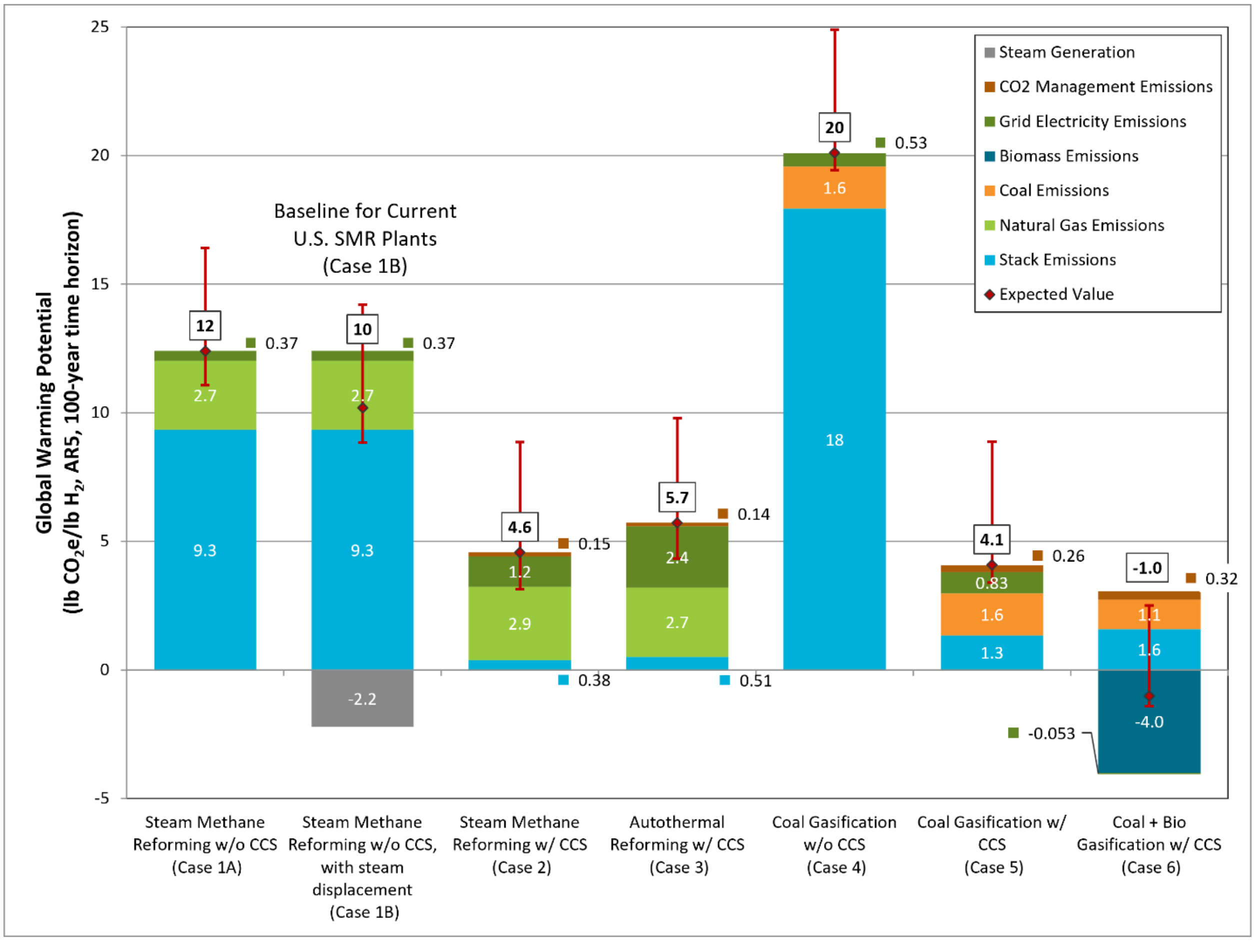
- The reforming cases have on average about 10 percentage points higher cold gas efficiencies than the gasification cases. The highest CGE (76.7 percent) is from the SMR plant without capture (Case 1), and the lowest CGE (57.7 percent) is from the coal/biomass co-gasification case with capture (Case 6).
- Adding capture to the SMR plant decreases the ETE by about seven percentage points due to a greater percentage of the NG feedstock being combusted to satisfy the thermal demands of the CO_2 removal processes; while adding capture to the coal gasification plant decreases the CGE by less than one percentage point since additional feedstock is not required to support CO_2 removal.
- The coal/biomass co-gasification plant with carbon capture and storage (CCS) (Case 6) is the only case in which the ETE is higher than the cold gas efficiency due to the positive net power production of the system. This reflects the design assumption made for this

case to eliminate power imports, so life cycle GHG emissions associated with power imports are eliminated.

Life cycle GHG emissions are reported for each study case. A greenhouse gas life cycle assessment (LCA) was performed in accordance with ISO 14040/14044 [1, 2], and results have been divided into emissions sources, including stack emissions, downstream CO_2 management, and key upstream processes: natural gas, coal, grid electricity, and biomass. Since Case 1 (SMR without CCS) produces excess steam as a co-product, parallel models of Case 1 (A and B) were created to estimate effects with and without consideration of the steam displacing production from other sources. Note that Case 1A aligns with Case 1's levelized cost of hydrogen result which represents a conservative scenario in which no end use nor corresponding value of the steam byproduct is considered. In Case 1B, SMR hydrogen production is integrated or co-located with an industrial operation that can utilize the excess steam, thus a commensurate emissions credit based on system expansion using a natural gas boiler to produce steam of equivalent quality is used. Case 1B is only represented with LCA results within this report and represents current industrial hydrogen production with co-located use of the steam product. Case 1B best represents the baseline for current U.S. SMR plants at the modeled compressed hydrogen product pressure of 925 psig.

Exhibit ES-4 shows the life cycle global warming potential (GWP) with a 100-year time horizon, in lb CO_2e per lb H_2, for each of the cases. The error bars represent overall ranges of variability in the emissions results calculated from a Monte Carlo simulation with 5,000 iterations, as well as regional variability in natural gas scenarios based upon high and low basin values.

Exhibit ES-4. CO_2e life cycle emissions for all cases



The variability in the natural gas-based Cases (1-3) is largely from the known variability within the average of upstream processes used to extract, process, and transport natural gas. Cases 2 and 3 have variability that is also influenced by the higher amount of grid electricity needed for the CCS systems. Cases involving only coal (4-5) have comparable error bars, largely associated with known variability in coal mine methane emissions based on the methane content of the coal seams in the Illinois basin. Finally, error bars for Case 6 of coal with biomass is due to variation in yields and fertilization rates along with the coal variability.

The following observations can be made about the life cycle emissions:

- The configuration with the highest global warming potential on a life cycle basis is the coal gasification plant without CCS (Case 4, 20 lb CO_2e/lb H_2). The configuration with the lowest GWP is the coal/biomass co-gasification plant with CCS (Case 6, -1.0 lb CO_2e/lb H_2).

  The average life cycle GHG emissions of the reforming cases with capture (Case 2, 4.6 lb CO_2e/lb H_2 and Case 3, 5.7 lb CO_2e/lb H_2) is approximately 10 to 40 percent higher than the life cycle GWP of the coal gasification case with capture (Case 5, 4.1 lb CO_2e/lb H_2). This is primarily due to the life cycle GHG emissions of the grid electricity and natural gas feedstock procurement, which make up approximately 90 percent of the life cycle emissions of the reforming cases with capture.

- Considering steam displacement for Case 1 leads to a notable 17 percent reduction in GWP. This suggests that the value of steam as a co-product has a significant impact on life cycle results.
- The life cycle emissions of Case 6 include a small credit (0.053 lb CO_2e/lb H_2) for electricity sold back to the grid. This credit is further described in Section 2.8.

Exhibit ES-5 shows the cost summary for all cases. A graph of the levelized cost of hydrogen as $/kilogram (kg) is provided in Exhibit ES-6. The error bars included represent the potential LCOH range relative to the maximum and minimum capital cost uncertainty ranges. The LCOH ranges presented are not reflective of other changes, such as variations in fuel price, labor price, CF, or other factors.

Exhibit ES-5. Cost summary for all cases

Case Name	Reforming[A]			Gasification[A]		
	SMR		ATR	Shell Gasifier		
	1	2	3	4	5	6
Total Plant Cost (2018$/[kg H_2/day])	554	1,394	1,092	4,264	4,901	5,241
Total Overnight Cost (2018$/[kg H_2/day])	713	1,735	1,372	5,243	6,044	6,515
Owner's Costs	159	341	280	979	1,142	1,274
Process Contingency	0	52	1	231	304	211
Project Contingency	94	252	114	628	734	765
Home Office Expenses	77	182	163	456	516	538
Bare Erected Cost	384	908	814	2,949	3,347	3,726
Total As-Spent Cost (2018$/[kg H_2/day])	763	1,856	1,467	5,853	6,747	7,273
LCOH ($/kg H_2) (excluding T&S)	1.06	1.54	1.51	2.58	2.92	3.44
Capital Costs	0.14	0.33	0.26	1.17	1.35	1.46
Fixed Costs	0.07	0.15	0.11	0.53	0.60	0.75
Variable Costs	0.09	0.24	0.36	0.43	0.52	0.43
Fuel Costs	0.77	0.82	0.77	0.44	0.44	0.80
LCOH ($/kg H_2) (including T&S)	1.06	1.64	1.59	2.58	3.09	3.64
CO_2 T&S Costs	0.00	0.10	0.09	0.00	0.17	0.20
Breakeven CO_2 Sales Price (ex. T&S), $/tonne[B]	N/A	50.1	50.4	N/A	20.8	42.9
Breakeven CO_2 Emissions Penalty (incl. T&S), $/tonne[B]	N/A	64.1	60.3	N/A	30.8	65.0

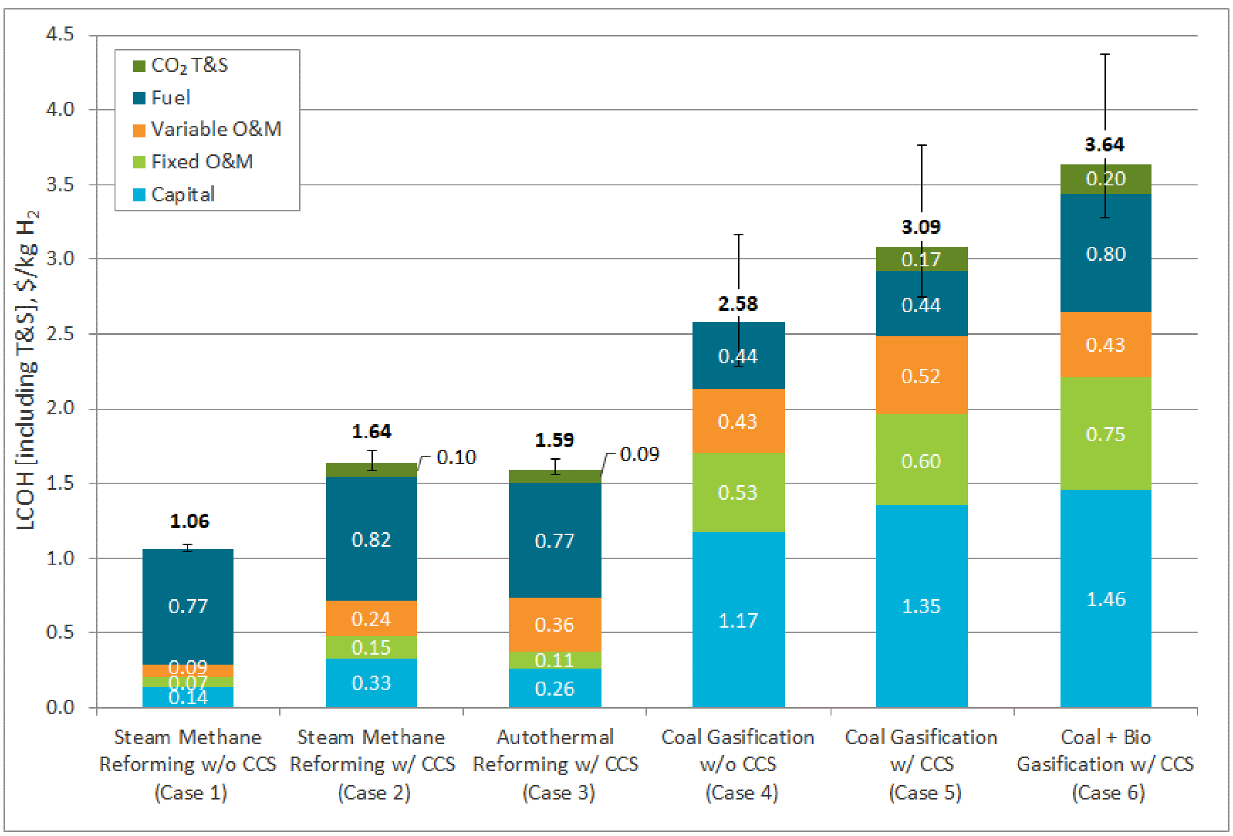
[A]Financing structures are presented in NETL's "QGESS: Cost Estimation Methodology for NETL Assessments of Power Plant Performance" [6]

[B]Both the breakeven CO_2 sales price and emissions penalty were calculated based on the respective non-capture cases within the reforming cases (Case 1) and gasification cases (Case 4)

The primary observations that can be made are:

- The lowest LCOH is \$1.06/kg H_2 from the SMR plant without capture (Case 1). The highest LCOH is \$3.64/kg H_2 from the coal/biomass co-gasification case with capture (Case 6).
- The average LCOH for the gasification cases is about two times greater than the average LCOH for the reforming cases. This is mainly due to the higher capital and fixed costs needed in the gasification cases compared to the reforming cases.
- The largest contributor to the LCOH for the reforming cases is the fuel cost, accounting for between 49 and 73 percent of the total LCOH. The largest contributor for the gasification cases is the capital cost, accounting for between 40 and 46 percent of the total LCOH. The cost of the feedstock, on a per thermal content unit basis, is about double for natural gas compared to coal. On the other hand, the greater complexity of the gasification plants explains the large contribution of capital costs on the LCOH.
- The addition of CO_2 capture technology impacts the reforming plant's LCOH more than the gasification plant's LCOH. Adding CCS to the reforming cases increases the LCOH by 54 percent for the SMR plant (Case 2). Adding capture to the coal gasification plant (Case 5) increases the LCOH by 20 percent.

Exhibit ES-6. LCOH by cost component



*Financing structures are presented in NETL's "QGESS: Cost Estimation Methodology for NETL Assessments of Power Plant Performance" [6]

Costs of Mature Technologies and Designs

The cost estimates for plant designs that only contain fully mature technologies, which have been widely deployed at commercial scale (e.g., SMR and gasification without CO_2 capture), reflect N^{th}-of-a-kind (NOAK) on the technology commercialization maturity spectrum. The costs of such plants have dropped over time due to "learning by doing" and risk reduction benefits that result from serial deployments as well as from continuing R&D.

Costs of Emerging Technologies and Designs

The cost estimates for plant designs that include technologies that are not yet fully mature (e.g., any plant with CO_2 capture) use the same cost estimating methodology as for mature plant designs, which does not fully account for the unique cost premiums associated with the initial, complex integrations of emerging technologies in a commercial application. Thus, it is anticipated that early deployments of these plants may incur costs higher than those reflected within this report.

Other Factors

Actual reported project costs for all the plant types are also expected to deviate from the cost estimates in this report due to project- and site-specific considerations (e.g., contracting strategy, local labor costs, seismic conditions, water quality, financing parameters, local environmental concerns, weather delays) that may make construction more costly. Such variations are not captured by the reported cost uncertainty.

Future Cost Trends

Continuing research, development, and demonstration (RD&D) is expected to result in designs that are more advanced than those assessed by this report, leading to costs that are lower than those estimated here.

OPPORTUNITIES FOR FUTURE R&D

This report also identifies several potential areas for R&D and supporting analysis that may open avenues for performance improvements and/or cost reductions beyond what the technologies included in this study provide. Opportunities are classified as either approaches to drive down the life cycle GHG emission profiles of existing technologies or incorporation of advanced technology concepts. Reducing the life cycle GHG emissions, for example, can be accomplished by blending renewable natural gas (RNG), pairing direct air capture (DAC), sourcing low-carbon auxiliary power, and increasing CO_2 capture rates.

1 INTRODUCTION

This report presents an independent assessment of the cost and performance of select hydrogen production plants utilizing fossil fuel resources as the primary feedstocks – specifically, natural gas (NG) steam methane reforming (SMR), NG autothermal reforming (ATR), coal gasification, and coal/biomass co-gasification – using a systematic, transparent technical and economic approach. Study cases were selected to reflect the capabilities of current, commercial technologies within plant configurations, and at scales, representative of next commercial offerings facing no fundamental research and development (R&D) obstacles. Additionally, several areas of R&D are identified as potential pathways for performance improvements and cost reductions.

Global warming potential (GWP) profiles of the impactful energy and material streams entering and exiting the plant boundaries were used to develop life cycle greenhouse gas (GHG) emissions results for each case based on the quantity of each stream. The results are deemed to be commensurate with an International Organization for Standardization (ISO) 14040/47 life cycle analysis (LCA) and are expressed as carbon dioxide equivalents (CO_2e). A life cycle GHG emissions target of 0 lb CO_2e/lb H_2 produced, or "net-zero," was targeted in the coal/biomass co-gasification case.

The cost and performance of fossil fuel-based hydrogen production technologies represented in this report are important inputs to assessments and determinations of technology combinations to be utilized to meet the projected demands of the nascent hydrogen economy. In addition to informing technology comparisons, the reference plant configurations found in this report provide perspective for regulators and policy makers. From a R&D perspective, this report may be used to assess goals and metrics and to provide a consistent basis for comparing developing technologies.

Six hydrogen plant configurations are analyzed in this report. A summary of the configurations is shown in Exhibit 1-5:

- Three natural gas reforming configurations—two NG SMR cases (with and without carbon dioxide [CO_2] capture) and one NG ATR case (with CO_2 capture).
- Three gasification configurations — two coal gasification cases (with and without CO_2 capture) and one coal/biomass co-gasification case (with CO_2 capture).

The CO_2 capture strategy employed for each case recovers greater than 90 percent of the carbon entering the plant boundary. High levels of capture are achieved by using a combination of water gas shift (WGS) reactors and solvent-based CO_2 separation technologies. The solvent technologies used include methyl diethanolamine (MDEA) (SMR and ATR), Shell Cansolv (SMR) and two-stage Selexol (coal and coal/biomass gasification).

This report reflects varying degrees of technology vendor input for each case. However, the final assessment of performance and cost was determined independently and is not endorsed by individual vendors.

1.1 Unit Configurations

A summary of study cases and their corresponding plant configurations considered in this report is presented in Exhibit 1-5. Components of each plant configuration are described in more detail in the corresponding report sections.

As previously stated, study cases were selected to represent the capabilities of current, commercial technologies within plant configurations, and at scales, representative of next commercial offerings facing no fundamental R&D obstacles. Additionally, a plant configuration capable of achieving net-zero greenhouse gas emissions over the plant life cycle using a carbon negative biomass feedstock, while adhering to these criteria, was evaluated.

1.1.1 Literature Review

A literature review was undertaken to inform the study case selection process. The objectives of the literature review were to:

- Survey a market sample of commercially operating, and commercially available, natural gas-, coal-, and alternative feedstock-derived, dedicated, high-purity hydrogen production sources, both within the U.S. and abroad
- Gather information on commercial H_2/CO_2 separation technologies

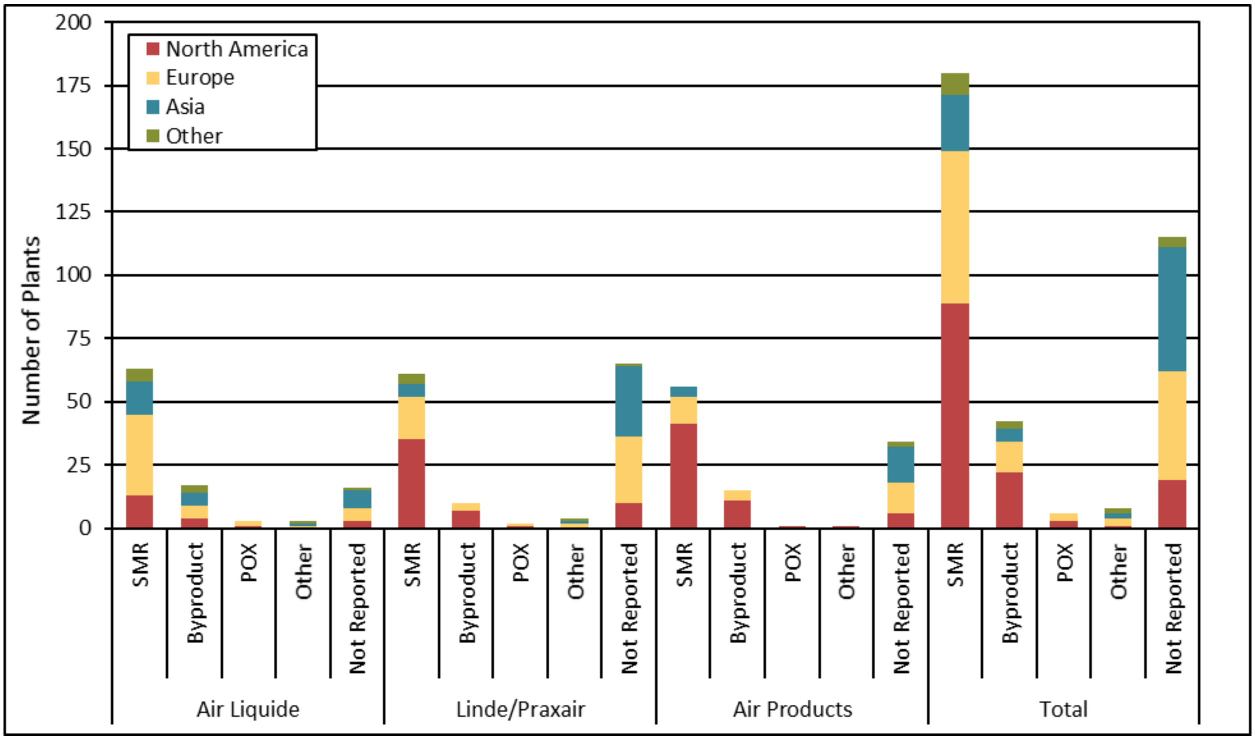
The primary data sources for the literature review included:

- Merchant hydrogen production data representing approximately 10 percent of the annual, global pure hydrogen market [15]
- Production data representing approximately 10 percent of the annual, global pure hydrogen produced from coal for ammonia [16]
- Global data of gasification facilities operating on biomass and alternative feedstocks (e.g., municipal solid waste (MSW)) [17]
- Miscellaneous sources including conference proceedings, recent project press releases, and discussions with industry contacts

The primary findings from the literature review follow:

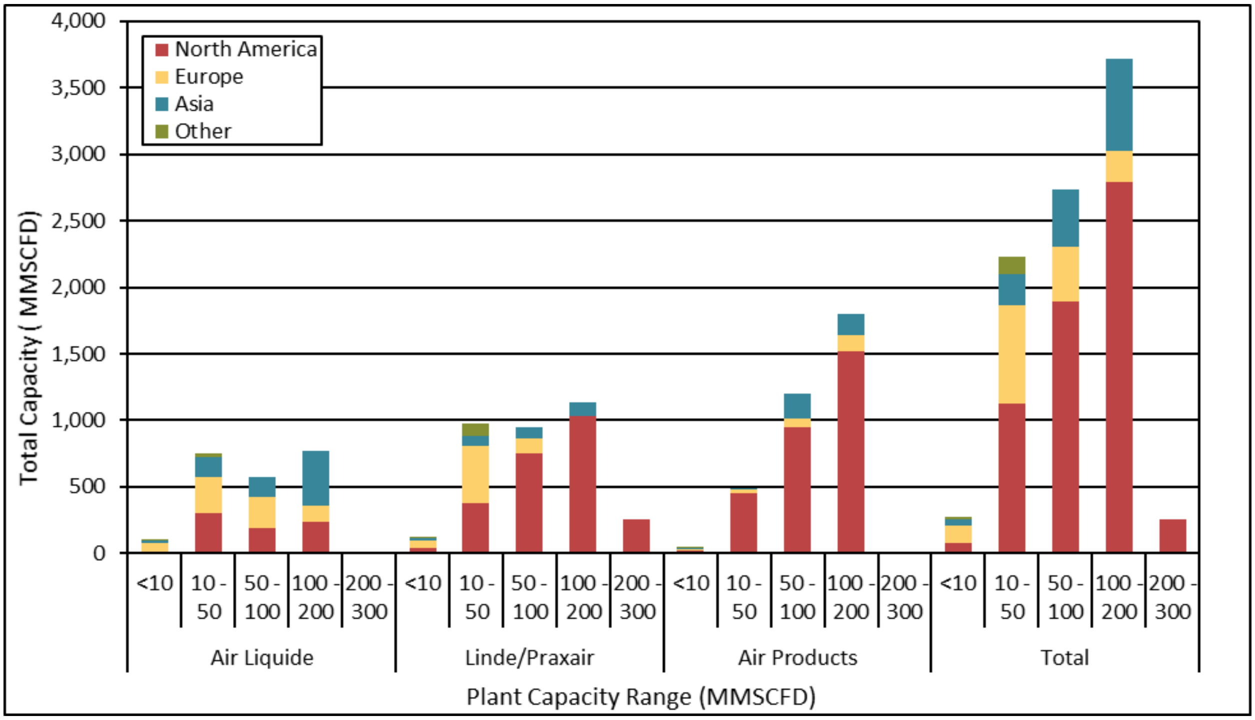
- As seen in Exhibit 1-1, merchant, high-purity hydrogen production facilities are spread globally and predominantly use NG SMR, with additional production from processing of chlor-alkali byproducts and partial oxidation (POX)

Exhibit 1-1. Global merchant hydrogen fleet production route profile (by supplier)



- Exhibit 1-2 shows the distribution of plant capacity ranges for hydrogen production facilities, broken out by technology supplier, Air Liquide, Linde/Praxair, Air Products, and by location, North America, Europe, Asia, or other
  - The chart shows that the majority of plants produce between 100 and 200 million standard cubic feet per day (MMSCFD) of high-purity hydrogen. The total capacity of plants drops significantly at capacities above 200 MMSCFD, indicating that this is likely an upper-limit of SMR plant capacity, which is the most common hydrogen production route
  - Even flow distribution of combustion gasses across the furnace reactor tubes is a primary scale-up limitation of SMR plants which limits the largest, single-train SMR facilities to approximately 200 MMSCFD [18]

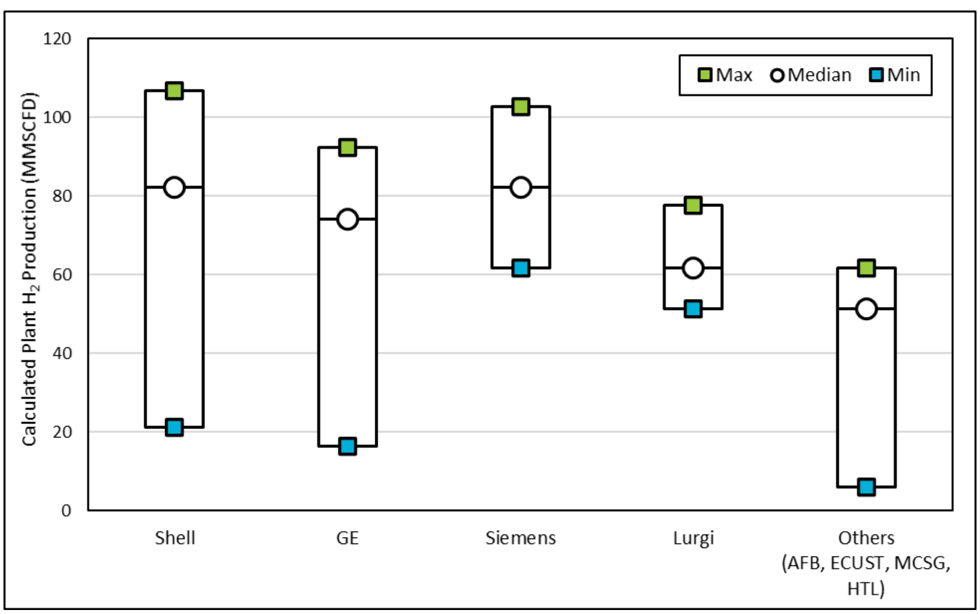
Exhibit 1-2. Global merchant hydrogen fleet capacity profile (by supplier)



Note: Total shown reflects the combined capacity of the three suppliers shown.

- Exhibit 1-3 shows high-purity hydrogen from coal for ammonia produced primarily in China using various gasification technologies in facilities with estimated median production rates of 50-100 MMSCFD. Looking at ammonia production serves as a proxy for hydrogen production capacity, as high-purity hydrogen is an intermediate in ammonia production.

Exhibit 1-3. Chinese hydrogen from coal (for ammonia) technology profile



- Exhibit 1-4 contains the results of a literature review search conducted in early 2022 of operating and planned hydrogen production facilities with CCUS. As can be seen, only a few hydrogen production facilities are currently operating with carbon capture, utilization, and storage (CCUS) (<90 percent capture); several others are in development (>90 percent capture)

Exhibit 1-4. Hydrogen production with CCUS (operating and planned)

	Plant/Project Name	Location	H_2 Production kNm³/hr (MMSCFD)	H_2 Production Platform	AGR Technology	Overall CO_2 Capture Rate (%)
Operating	Air Products Port Arthur [19]	USA	220 (200)	SMR	VSA	60
	Air Liquide Port Jerome [20]	France	50 (45)	SMR	Membranes, CRYOCAP™ H2	60
	Shell Quest [21] [22]	Canada	210 (191)	SMR	ADIP-X	50
Under Development	H-Vision	Netherlands	700 (636)	ATR	Rectisol™	88.0
	HyNet [23]	United Kingdom	100 (90)	ATR	TBD	97.2
	H21 [24] [25]	United Kingdom	3,200 (2,900) from 9 units	ATR	aMDEA	94.2
	Acorn [26]	Scotland	53 (48)	ATR	Amine	98.7
	H2Teesside [27]	United Kingdom	275 (250)	TBD	TBD	98
	Air Products Alberta [28]	Canada	>695 (>623)	ATR	TBD	95
	Air Products Louisiana [29]	USA	>837 (>750)	ATR/POX	TBD	95

- There are currently no commercial examples of facilities gasifying alternative feedstocks (e.g., biomass, MSW) to produce high-purity hydrogen as an end-product
- The decommissioned Buggenum integrated gasification combined cycle (IGCC) (coal/biomass co-gasification) and Eastman Kingsport (coal/waste plastics) facilities are the only examples of commercially operated facilities to co-gasify coal with an alternative feedstock. Neither facility produces high-purity hydrogen as an end-product

1.1.2 Case Selection Rationale

Several fossil-based hydrogen pathways were considered for inclusion into this study. Final selection of study cases emphasized the current landscape of the global, high-purity hydrogen production fleet, as well as the outlook for fossil-based, hydrogen projects with carbon capture and storage (CCS) currently under development. As a consequence, no screening, or down selection, of potential study cases based on performance and/or economic criteria was performed. Additionally, it is recognized that not all fossil-based, hydrogen production pathways and configurations meeting the selection criteria are represented in this study.

1.1.2.1 Reforming Cases

SMR without CCS is included to provide a current, commercial reference because it is the predominant high-purity production route in the current market. SMR and ATR with CCS are each included to assess future R&D benefits relative to the non-capture commercial reference in future analyses. The 200 MMSCFD production capacity basis of the SMR plants maximizes economies of scale with a single-reactor arrangement. This capacity reflects the upper range of commercially operating SMR plants today.

The single-train design capacity of ATR with CCS, unconstrained by the scale-up limitations inherent to SMR, matches the hydrogen production output of the coal gasification cases and results in a hydrogen production rate of 274 MMSCFD. This capacity falls within the expected plant sizes of the projects under development shown in Exhibit 1-4.

All reforming cases operate on pipeline natural gas. Assessment of the potential for net-zero greenhouse gas (GHG) emissions from blending alternative feedstocks (e.g., renewable natural gas [RNG]) is ongoing and are not considered in this study.

1.1.2.2 Gasification Cases

The three gasification cases are based on performance data for an oxygen-blown, entrained-flow gasifier that is intended to be similar to the Shell gasifier, by which it is referred to from here on out. This gasifier has been included in numerous prior NETL studies, including most-recently in, the Fossil Energy Baseline. The estimates were derived from public literature and relied heavily on conference presentations given by Shell at the Gasification Technologies Conference in 2009. It is not intended to faithfully reproduce the operating point of any specific gasifier either currently operating, previously operating, or planned. It is a conceptual design that extends some of the features of the gasifier island, such as integration with other plant components. It was intended to keep the design basis comparable to how other gasifiers were represented in previous studies like the Fossil Energy Baseline. In 2018, the Shell gasifier technology was purchased and is now licensed by Air Products. It is recognized that the technology is now owned by Air Products, not Shell, but the gasifier will continue to be called the “Shell” gasifier to be consistent with the Fossil Energy Baseline and avoid confusion.

The primary motivation for the selection of the Shell gasifier was the historical, commercial experience of the Shell gasifier continuously co-gasifying coal with 30 percent woody biomass at the Buggenum IGCC facility with prior plans to increase to 50 percent. [30] Additional considerations were the prevalence of the Shell technology in the Chinese coal-to-ammonia market.

Consistent with the SMR cases, coal gasification without CCS was included as a current, commercial reference for coal gasification with CCS. The capacity basis of the coal gasification cases is the equivalent as-received coal feed rate from Case B1B from the NETL Fossil Energy

Baseline study. [31] As indicated previously, the resulting hydrogen production rate for these cases is 274 MMSCFD and equal to the ATR with CCS case. Although larger than the commercially operating examples found, process design studies have been completed for coal-to-ammonia plants at comparable scales as indicated by conversations with Black and Veatch (B&V).

Co-gasification of coal and biomass is included as the only study case achieving net-zero greenhouse gas life cycle emissions. A torrefied, woody biomass is considered due to the compatibility with the dry-fed Shell gasifier, as well as the experience at Buggenum. [30] The capacity basis of 1,400 ton per day (tpd) of combined feed matches that of the Buggenum facility.

The remainder of this report is organized as follows:

- Section 2 provides the basis for technical, environmental, and cost evaluations
- Section 3 describes the reforming technologies modeled and presents the results for the three cases
- Section 4 describes the gasification technologies modeled and presents the results for the three cases
- Section 5 provides results analysis for all six cases
- Section 6 provides suggestions for future work

Exhibit 1-5. Case descriptions

<table>
<tr><th>Case[A]</th><th>Plant Type</th><th>Feedstock</th><th>Reformer Type</th><th>Gasifier Type</th><th>CO_2 Capture[B]</th><th>H_2 Purification</th><th>Capacity Basis</th></tr>
<tr><td>1</td><td rowspan="3">Reforming</td><td rowspan="3">Natural Gas</td><td rowspan="2">SMR</td><td rowspan="3">N/A</td><td>0%</td><td rowspan="6">PSA</td><td rowspan="2">Single Train SMR Max (200 MMSCFD H_2)</td></tr>
<tr><td>2</td><td>Max</td></tr>
<tr><td>3</td><td>ATR</td><td>Max</td><td>Match H_2 output of Cases 4 & 5</td></tr>
<tr><td>4</td><td rowspan="3">Gasification</td><td rowspan="2">Illinois No. 6 Coal</td><td rowspan="3">N/A</td><td rowspan="3">Shell</td><td>0%</td><td rowspan="2">5,608 AR tpd coal</td></tr>
<tr><td>5</td><td>Max</td></tr>
<tr><td>6</td><td>Illinois No. 6 Coal/ Torrefied Woody Biomass</td><td>Max</td><td>1,400 tpd total gasifier feedstock</td></tr>
</table>

[A]All plants in this report are assumed to be located at a generic plant site in the midwestern United States
[B]CO_2 capture targets the maximum amount of feedstock carbon captured from the syngas (ATR and gasification cases) and syngas plus furnace flue gas (SMR case)

2 General Evaluation Basis

For each of the plant configurations analyzed in this report, an Aspen Plus® (Aspen) model was developed and used to generate material and energy balances which were, in turn, used to provide a design basis for items in the major equipment list. The equipment list and material balances were used as the basis for generating the capital and operating cost estimates. Performance and process limits were based upon published reports, information obtained from vendors and users of the technology, performance data from design/build projects, and/or best engineering judgment. Capital and operating costs include scaled estimates from prior studies using National Energy Technology Laboratory (NETL) Quality Guidelines for Energy System Studies (QGESS) methodology as well as estimates provided by B&V based on simulation results using an in-house database and conceptual estimating models. The estimating models are based on a United States (U.S.) Gulf Coast location and the labor cost was factored to reflect a Midwest location. Costs were further calibrated using a combination of adjusted vendor-furnished data and scaled estimates from previous design/build projects. All costs are in 2018 dollars. Ultimately, a levelized cost of hydrogen (LCOH) was calculated for each of the cases and is reported as the revenue requirement figure-of-merit.

The balance of this section discusses the design basis common to all technologies, as well as environmental targets and cost assumptions used in this report. Technology specific design criteria are covered in subsequent chapters.

2.1 Site Characteristics

All plants in this report are assumed to be located at a generic plant site in the midwestern United States, with site characteristics and ambient conditions as presented in Exhibit 2-1 and Exhibit 2-2. The ambient conditions are the same as International Organization for Standardization (ISO) conditions.

Exhibit 2-1. Site characteristics

Parameter	Value
Location	Greenfield, Midwestern U.S.
Topography	Level
Size (Gasification), acres	300
Size (Reforming), acres	100
Transportation	Rail or Highway
Slag Disposal	Off-Site
Water	50% Municipal and 50% Ground Water

Exhibit 2-2. Site ambient conditions

Parameter	Value
Elevation, m (ft)	0 (0)
Barometric Pressure, MPa (psia)	0.101 (14.696)
Average Ambient Dry Bulb Temperature, °C (°F)	15 (59)
Average Ambient Wet Bulb Temperature, °C (°F)	10.8 (51.5)
Design Ambient Relative Humidity, %	60
Cooling Water Temperature, °C (°F)[A]	15.6 (60)
Air composition based on published psychrometric data, mass %	
N_2	75.055
O_2	22.998
Ar	1.280
H_2O	0.616
CO_2	0.050
Total	100.00

[A]The cooling water temperature is the cooling tower cooling water exit temperature. This is set to 4.8 °C (8.5 °F) above ambient wet bulb conditions in ISO cases

The land area for gasification cases assumes that 30 acres are required for the plant proper, and the balance provides a buffer of approximately 0.4 km (0.25 mi) to the fence line. The extra land could also provide for a rail loop if required (rail loop is not included in this report). In the reforming cases it was assumed the plant proper occupies about 10 acres leaving a buffer of 0.24 km (0.15 mi) to the plant fence line.

The quality of plant makeup water will vary dramatically from source-to-source (municipal versus groundwater), as well as from site to site, and can be expected to vary significantly throughout any given site, particularly if ground water is utilized. In this study, 50 percent of the makeup water to the plants is sourced from a publicly owned treatment works, with the balance of the makeup water sourced from groundwater. The assumed design makeup water composition is provided in Exhibit 2-3.

The makeup water composition reported in the following table is based on water qualities from actual operations. The design concentration of each constituent is individually representative of a plant configuration comparable to those in this study. However, due to the interaction and interdependencies of each constituent and the multitude of potential species, the makeup water quality cannot be considered representative as a whole. The makeup water quality is intended to inform users of the contaminants likely present, and at what concentrations they may be expected, to facilitate appropriate equipment selection and design.

Exhibit 2-3. Design makeup water quality

Parameter	Groundwater (Range)	POTW Water (Range)	Makeup Water (Design Basis)
pH	6.6 – 7.9	7.1 – 8.0	7.4
Specific Conductance, µS/cm	1,096 – 1,484	1,150 – 1,629	1312
Turbidity, NTU		<50	<50
Total Dissolved Solids, ppm			906
M-Alkalinity as $CaCO_3$, ppm[A]	200 – 325	184 – 596	278
Sodium as Na, ppm	102 – 150	172 – 336	168
Chloride as Cl, ppm	73 – 100	205 – 275	157
Sulfate as SO_4, ppm	100 – 292	73 – 122	153
Calcium as Ca, ppm	106 – 160	71 – 117	106
Magnesium as Mg, ppm	39 – 75	19 – 33	40
Potassium as K, ppm	15 – 41	11 – 21	18
Silica as SiO_2, ppm	5 – 12	21 – 26	16
Nitrate as N, ppm	0.1 – 0.8	18 – 34	12
Total Phosphate as PO_4, ppm	0.1 – 0.2	1.3 – 6.1	1.6
Strontium as Sr, ppm	2.48 – 2.97	0.319 – 0.415	1.5
Fluoride as F, ppm	0.5 – 1.21	0.5 – 0.9	0.8
Boron as B, ppm	0.7 – 0.77		0.37
Iron as Fe, ppm	0.099 – 0.629	0.1	0.249
Barium as Ba, ppm	0.011 – 0.52	0.092 – 0.248	0.169
Aluminum as Al, ppm	0.068 – 0.1	0.1 – 0.107	0.098
Selenium as Se, ppm	0.02 – 0.15	0.0008	0.043
Lead as Pb, ppm	0.002 – 0.1		0.026
Arsenic as As, ppm	0.005 – 0.08		0.023
Copper as Cu, ppm	0.004 – 0.03	0.012 – 0.055	0.018
Nickel as Ni, ppm	0.02 – 0.05		0.018
Manganese as Mn, ppm	0.007 – 0.015	0.005 – 0.016	0.009
Zinc as Zn, ppm	0.005 – 0.024		0.009
Chromium as Cr, ppm	0.01 – 0.02		0.008
Cadmium as Cd, ppm	0.002 – 0.02		0.006
Silver as Ag, ppm	0.002 – 0.02		0.006
Mercury as Hg, ppm	0.0002 – 0.001		3E-04

[A]Alkalinity is reported as $CaCO_3$ equivalent, rather than the concentration of HCO_3. The concentration of HCO_3 can be obtained by dividing the alkalinity by 0.82

In the gasification cases, it was assumed that the steam turbine is enclosed in a turbine building. The gasifiers are not enclosed. In the reforming cases, it was assumed that the steam methane reformer is enclosed. The autothermal reformer is not enclosed.

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

- Flood plain considerations
- Existing soil/site conditions
- Water discharges and reuse
- Rainfall/snowfall criteria
- Seismic design
- Buildings/enclosures
- Local code height requirements
- Weather delays
- Other local environmental concerns
- Noise regulations

2.2 Coal Characteristics

The design coal is Illinois No. 6 with characteristics presented in Exhibit 2-4. The coal properties are from the 2019 revision of the QGESS document "Detailed Coal Specifications." [3]

Exhibit 2-4. Design coal

Rank	Bituminous	
Seam	Illinois No. 6	
Proximate Analysis (weight %)[A]		
	As Received	Dry
Moisture	11.12	0.00
Ash	9.70	10.91
Volatile Matter	34.99	39.37
Fixed Carbon	44.19	49.72
Total	100.00	100.00
Sulfur	2.51	2.82
HHV, kJ/kg (Btu/lb)	27,113 (11,666)	30,506 (13,126)
LHV, kJ/kg (Btu/lb)	26,151 (11,252)	29,544 (12,712)
Ultimate Analysis (weight %)		
	As Received	Dry
Moisture	11.12	0.00
Carbon	63.75	71.72
Hydrogen	4.50	5.06
Nitrogen	1.25	1.41
Chlorine	0.15	0.17
Sulfur	2.51	2.82
Ash	9.70	10.91
Oxygen[B]	7.02	7.91
Total	100.00	100.00

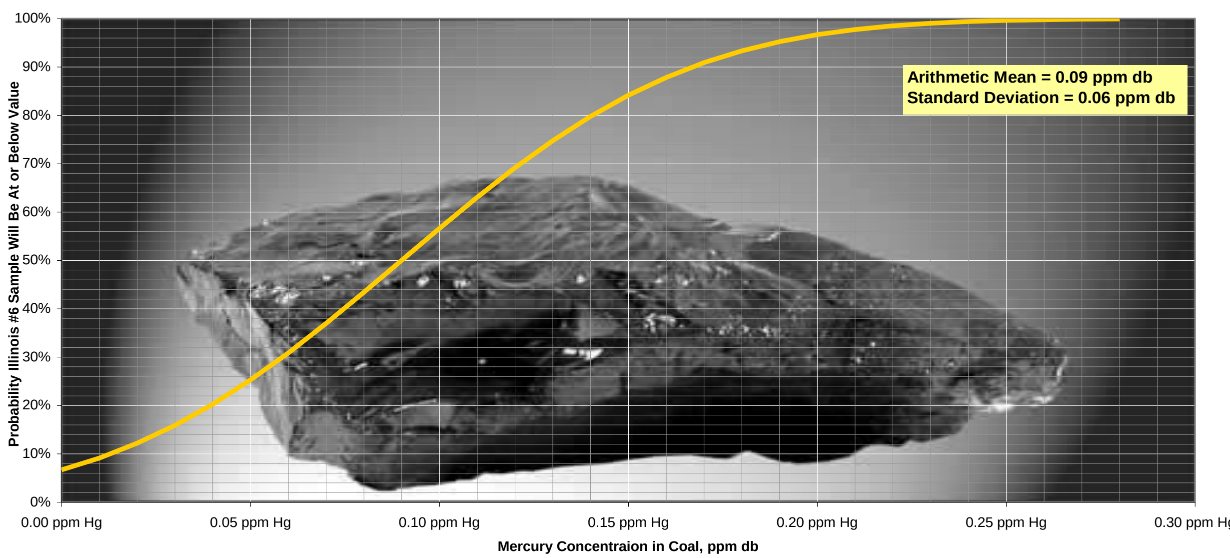
[A]The proximate analysis assumes sulfur as volatile matter
[B]By difference

The chlorine content of 34 samples of Illinois No. 6 coal has an arithmetic mean value of 1,671 parts per million weight, dry (ppmwd) with a standard deviation of 1,189 ppmwd based on coal samples shipped by Illinois mines. [32]

Based on the location of the Illinois No. 6 mine, along with the Herrin coal chlorine map published by the Illinois State Geological Survey [33], it was determined that Illinois No. 6 coal could be expected to have a chlorine content between 0.1 and 0.2 percent, on a dry basis. Therefore, the coal chloride content for this report was assumed to be the arithmetic mean value of 1,671 ppmwd.

The mercury (Hg) content of 34 samples of Illinois No. 6 coal has an arithmetic mean value of 0.09 ppmwd with standard deviation of 0.06 based on coal samples shipped by Illinois mines. [32] Hence, as illustrated in Exhibit 2-5, there is a 50 percent probability that the Hg content in the Illinois No. 6 coal would not exceed 0.09 ppmwd. The coal Hg content for this report was assumed to be 0.15 ppmwd, which corresponds to the mean plus one standard deviation and encompasses about 84 percent of the samples. It was further assumed that all the coal Hg enters the gas phase and none leaves with the slag in gasification cases. [34]

Exhibit 2-5. Probability distribution of mercury concentration in the Illinois No. 6 coal



Fuel costs used in this report are specified according to the 2019 QGESS document "Fuel Prices for Selected Feedstocks in NETL Studies." [8] The current levelized coal price is $2.11/gigajoule (GJ) ($2.23/million British thermal units [MMBtu]) on a higher heating value (HHV) basis for Illinois No. 6 bituminous coal delivered to the Midwest and reported in 2018 dollars. Fuel costs are levelized over an assumed 30-year plant operational period with an assumed on-line year of 2023.

2.3 Natural Gas Characteristics

Natural gas is utilized as the feedstock in the reforming cases, and its composition is presented in Exhibit 2-6. The natural gas properties are from the 2019 revision of the QGESS document "Specification for Selected Feedstocks" [4] including the addition of methanethiol (mercaptan). [35]

The current levelized natural gas price is \$4.19/GJ (\$4.42/MMBtu) on an HHV basis, delivered to the Midwest, and reported in 2018 U.S. dollars.[b] Fuel costs are levelized over an assumed 30-year plant operational period with an assumed on-line year of 2023.

Exhibit 2-6. 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	
Methanethiol[A]	CH_4S	$5.75x10^{-6}$	
	Total	100.0	
	LHV		**HHV**
kJ/kg (Btu/lb)	47,201 (20,293)		52,295 (22,483)
MJ/scm (Btu/scf)	34.52 (927)		38.25 (1,027)

[A]The sulfur content of natural gas is primarily composed of added mercaptan (methanethiol [CH_4S]) with trace levels of hydrogen sulfide (H_2S) [35]
Note: Fuel composition is normalized, and heating values are calculated using Aspen

2.4 Biomass Characteristics

A torrefied, non-pelletized, short rotation, woody biomass is utilized as a co-feedstock in Case 6, and its composition is presented in Exhibit 2-7.

It is assumed that the torrefaction of the biomass is accomplished in dedicated torrefaction facilities beyond the boundaries of the hydrogen production facility. Assumptions pertaining to the torrefied biomass supply and cost are explained in Section 2.5.

[b] As specified in the 2019 QGESS document on "Fuel Prices for Selected Feedstocks in NETL Studies." [8]

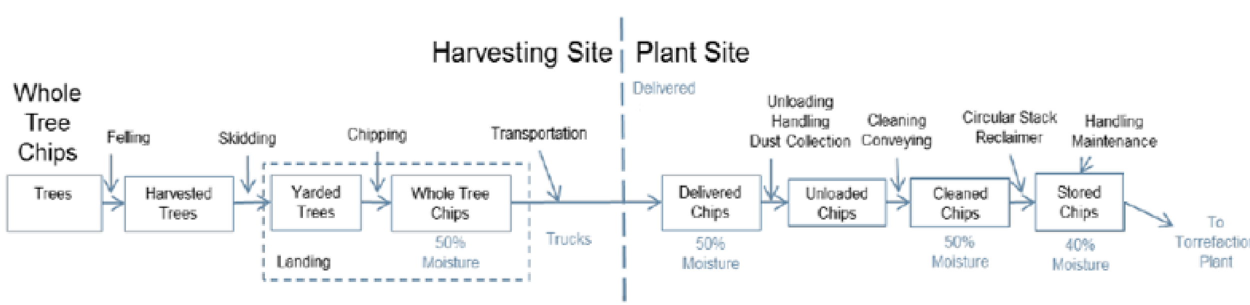
Exhibit 2-7. Design biomass

Torrefied, Woody Biomass		
Ultimate Analysis (weight %)		
	As Received	Dry
Moisture	5.72	0.00
Carbon	59.89	63.52
Hydrogen	5.11	5.42
Nitrogen	0.41	0.44
Chlorine	0.00	0.00
Sulfur	0.00	0.00
Ash	0.51	0.54
Oxygen	28.36	30.08
Total	100.00	100.00
HHV, kJ/kg (Btu/lb)	22,676 (9,749)	24,051 (10,340)
LHV, kJ/kg (Btu/lb)	21,406 (9,203)	22,853 (9,825)

2.5 Biomass Supply and Cost

Torrefied biomass is delivered to the Case 6 hydrogen production facility gate in trucks and consists of torrefied chips. The delivered cost of the torrefied biomass was determined using a performance and economic model developed in a prior NETL study which leveraged work completed by the Idaho National Laboratory (INL). [36] [37] The model considers a wood chip supply chain design that uses conventional technologies and operations prior to delivery to the torrefaction facilities. The supply chain steps, and the pre-processing steps undertaken at the torrefaction plants are shown in Exhibit 2-8.

Exhibit 2-8. Wood chip supply chain



The model incorporates the cost of the wood chips delivered to the torrefaction facility as well as vendor-provided performance and cost data for a 63 kiloton per year woody biomass torrefaction system. Modeling parameter adjustments were made to reflect the total required torrefied biomass supply rate, capital charge factor, capital cost escalation, and cost for electric power to maintain consistency with the assumptions for Case 6. The equivalent output of 3.2 torrefaction systems operating at a capacity factor of 90 percent are required to meet the torrefied biomass demand of Case 6. The results of the model are shown in Exhibit 2-9.

Exhibit 2-9. Woody biomass torrefaction system economic summary

Process Parameters/Category	Value	Units
Feed Prep and Drying		
Raw Wood Feed Rate	1,315	Tons/day
Raw Wood Moisture	43.3[A]	wt. %
Dried Wood Moisture	8.2[B]	wt. %
Dryer Thermal Capacity	52.2	MMBtu/hr
Torrefaction		
Feed Mass Flow	812	Tons/day
Mass Yield	.753	lb torrefied solid/lb feed
Energy Yield to Mass Yield Ratio	1.185	N/A
Torrefied Product HHV (dry basis)	10,340	Btu/lb
Torrefied Solids Product	609	Tons/day
Capacity Factor	90	%
Cost Estimate		
Capital Cost of Torrefaction Equipment	50.9	MM$
Capital Charge Factor	0.0654	N/A
Capital Component	3.3	MM$
Raw Wood (@$22.63/ton)	9.8	MM$
Natural Gas ($4/mcf)	1.4	MM$
Electric Power (2.2 MW @ $71.7/MWh)	1.2	MM$
Operations and Maintenance Costs	2.5	MM$
Annual Revenue Requirement	18.3	MM$
Required Selling Price (Torrefaction Plant Gate)	91.5	$ per ton

[A] The chips enter the torrefaction facility with about 50% moisture content. After storage at the plant, moisture is lost and on reclaiming the woody biomass is assumed to have an average moisture content of 43.3%. [36]

[B] The process of torrefaction dries the wood to 8.2% moisture before torrefaction. The torrefied material has a moisture content of 5.72%. [36]

The required selling price of torrefied biomass is $91.5/ton at the torrefaction plant gate. An additional cost of $5.73/ton must be added for transportation of the torrefied biomass to the hydrogen production facility. A 40-mile transport distance at a cost of $0.14 per mile-ton is used. This brings the total price of as-received, torrefied biomass to \$97.2 per ton. The levelized price is \$117/tonne (\$106/ton) or \$5.15/GJ (\$5.43/MMBtu) on an HHV basis, delivered to the site, and reported in 2018 U.S. dollars. Fuel costs are levelized over an assumed 30-year plant operational period with an assumed on-line year of 2023.

2.6 Hydrogen Product Specification

All plant configurations in this study produce a hydrogen product at the plant gate meeting or exceeding the specification shown in Exhibit 2-10. No site-specific, single end-use is considered. Instead, the study goal was to consider a hydrogen product reflective of current, centralized, fossil-based production facilities and suitable for a wide range of potential energy applications. The maximum impurity concentrations shown in the specification are suitable for ammonia grade hydrogen to avoid catalyst poisoning. The specification also results in a product suitable for the following applications per ISO: [38]

- Type I Grade A – Gaseous hydrogen; internal combustion engines for transportation; residential/commercial combustion appliances (e.g., boilers, cookers, and similar applications)
- Type I Grade B – Gaseous hydrogen; industrial fuel for power generation and heat generation except proton-exchange membrane (PEM) fuel cell applications

The hydrogen product pressure specification was determined based on feedback from a commercial owner/operator of large-scale, centralized merchant hydrogen production facilities and a hydrogen pipeline transportation and distribution network. The owner/operator advised that pipeline operating pressures vary nominally from 800 to 900 pound per square inch gauge (psig) depending on the location within the network. A nominal hydrogen product pressure at the plant fence is greater than 900 psig, and compression normally occurs within the plant boundary.

Exhibit 2-10. Hydrogen product specification

Characteristic	Specification
H_2 Purity (vol%)	99.90 (min.)
CO_2 (ppm)	A
CO (ppm)	A
H_2S (ppb)	10 (max.)
H_2O (ppm)	A
O_2 (ppm)	A
Pressure (psia)	939.7

[A] The maximum total concentration of all oxygen-containing species is 10 ppm

2.7 Environmental Targets

2.7.1 Air Emissions Targets

The environmental approach for the study was to evaluate each case on the same regulatory design basis, considering differences in fuel and technology. Since all cases are located at a green-field site, permitting a new plant would likely invoke the New Source Review (NSR) permitting process. The NSR process requires installation of emission control technology meeting either Best Available Control Technology (BACT) determinations for new sources being located in areas meeting ambient air quality standards (attainment areas) or Lowest Achievable Emission Rate (LAER) technology for sources being located in areas not meeting ambient air quality standards (non-attainment areas).

BACT guidelines are used for the cases evaluated in this study. Low NOx burners will be included in the SMR and ATR cases to minimize the NOx emissions from the fired equipment to meet BACT requirements. Sulfur emissions are eliminated due to the pre-treatment of the feedstock to remove sulfur. For the production of hydrogen from coal gasification, the primary control standards that are expected to apply include emissions of particulates, nitrogen oxides, sulfur species, and mercury. Process technology is directed toward minimum sulfur content in the syngas and product. BACT for NOx emissions from the off-gas fired boilers will be utilized. BACT control technologies and emission limits are summarized in Exhibit 2-11 and Exhibit 2-12.

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

- NOx Emission Reduction Credits (ERCs) 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.
- 90+ percent removal of carbon in design fuel.

Exhibit 2-11. BACT environmental design basis for reforming cases

Pollutant	Control Technology	Limit
Sulfur Dioxide (SO_2)	Zinc oxide guard bed	Negligible
Nitrogen Oxides (NOx)	Low NOx Burners	2.5 ppmv (dry) @ 15% O_2
Particulate Matter (PM)	N/A	Negligible
Mercury (Hg)	N/A	Negligible

Exhibit 2-12. BACT environmental design basis for gasification cases

Pollutant	Control Technology	Limit
Sulfur Dioxide (SO_2)	Selexol/MDEA + Claus Plant or equivalent performing system	99+% or ≤ 0.050 lb/10^6Btu
Nitrogen Oxides (NOx)	Low NOx Burners	15 ppmvd (@ 15% O_2)
Particulate Matter (PM)	Cyclone/Barrier Filter/Wet Scrubber/AGR Absorber	0.015 lb/10^6Btu
Mercury (Hg)	Activated Carbon Bed or equivalent performing system	95% removal

2.7.2 Water Emissions Targets

EPA issued updated Effluent Limitation Guidelines (ELG) and standards for the steam electric power generation point source category in November 2015 to strengthen controls on wastewater discharges.[c] [39] The ELG are national technology-based New Source Performance Standards (NSPS) derived from data collected from industry. They are intended to provide flexibility in implementation through use of technologies already installed and operating in the power industry. The federal standards established by this rule are the minimum discharge standards. As ELG are enforced under the National Pollutant Discharge Elimination System [40], more stringent water quality-based standards may be established by the local permitting authority; however, these additional requirements were not considered in this report.

The final ELG rule established new wastewater categories and discharge limits and updated discharge requirements for existing wastewater categories. Wastewater from facilities gasifying fuels such as coal and petroleum coke are covered by the final ELG rule. Although not explicitly categorized as a steam electric power generation point source under the final ELG rule, the rule is assumed to apply to the gasification cases in this study due to the identical wastewater sources.

For the gasification cases in this study, the gasification wastewater from the balance of plant is recycled within the gasification and syngas cleanup process, ultimately being utilized as makeup to the syngas scrubber. Therefore, only the syngas scrubber blowdown requires treatment.

Non-chemical metal cleaning wastewater was established as a new wastewater category in the updated ELG. However, new limits were not established for this category; therefore, treatment of this stream has not been evaluated in this report.

The landfill of plant byproducts is assumed to be outside the scope of the plants considered in this study; therefore, landfill leachate is not evaluated in this report.

Intermittent discharges (e.g., chemical metal cleaning wastewater), coal pile runoff, low volume waste (e.g., boiler blowdown), and cooling tower blowdown were assumed to be compliant with all applicable regulations with no additional treatment beyond conventional considerations.

[c] In April 2017, EPA announced plans to reconsider the power plant ELG rule—as they apply to existing sources—and their intent to request a stay of the regulations, pending litigation. [40]

Under the assumptions established in this section, no additional control technology considerations are required for natural reforming compliance with the ELG rule.

The applicable wastewater discharge limits for the gasification cases are shown in Exhibit 2-13, respectively.

Exhibit 2-13. New source treated gasification wastewater discharge limits [39]

Effluent Characteristic	Long-Term Average	Daily Maximum Limit	Monthly Average Limit[A]
Arsenic, ppb	4.0	4	-
Mercury, ppt	1.08	1.8	1.3
Selenium, ppb	147	453	227
Total Dissolved Solids, ppm	15.2	38	22

[A]Monthly Average Limit refers to the highest allowable average of daily discharges over 30 consecutive days

The limits are applied at the discharge, prior to commingling with other plant water systems.

2.7.3 Carbon Dioxide

Global CO_2 emissions from the production of high-purity hydrogen are significant, accounting for 830 million tonnes (915 million tons) of CO_2 annually. This is roughly equivalent to the CO_2 emissions of Indonesia and the United Kingdom combined. [24] Large-scale CO_2 separation is commonly recovered from commercially-operating, hydrogen production processes and utilized for enhanced oil recovery, sold into the beverage market, or for synthesis of fertilizer products such as urea ammonium nitrate (UAN). However, the absence of environmental regulations targeting CO_2 emissions from industrial sources such as hydrogen production plants limits the overall CO_2 capture rate at such facilities as well as industry-wide adoption of CO_2 capture.[d] However, due to recent international government initiatives, several large-scale, fossil-based hydrogen production projects targeting high levels of CO_2 capture and storage (90+ percent) are currently under development in anticipation of future CO_2 regulations. Exhibit 2-14 provides an overview of some these projects and the targeted level of CO_2 capture.

Exhibit 2-14. Fossil-based hydrogen projects with CO_2 capture

Project Name	Location	H_2 Production kNm³/hr (MMSCFD)	H_2 Production Platform	AGR Technology	Targeted Overall CO_2 Capture Rate (%)
HyNet [23]	United Kingdom	100 (90)	ATR	TBD	97.2
H21 [24] [25]	United Kingdom	3,200 (2,900)	ATR	aMDEA	94.2

[d] The Infrastructure Investment and Jobs Acts (IIJA) of 2021 establishes a definition of clean hydrogen to mean hydrogen produced with a carbon intensity (CI) less than or equal to 2 kg CO_2e produced at the site of production per kg of H_2 produced. The IIJA further requires an initial CI standard be developed in mid-2022 with consideration for adjustment in 2027.

Project Name	Location	H_2 Production kNm³/hr (MMSCFD)	H_2 Production Platform	AGR Technology	Targeted Overall CO_2 Capture Rate (%)
Acorn [26]	Scotland	53 (48)	ATR	Amine	98.7
H2Teesside [27]	United Kingdom	275 (250)	TBD	TBD	~98
Air Products Alberta [28]	Canada	>695 (>623)	ATR	TBD	95
Air Products Louisiana [29]	USA	>837 (>750)	ATR/POX	TBD	95

As led by the growing number of such projects, an objective of this study was to examine plant arrangements capable of achieving capture rates greater than 90 percent of the carbon in the plant feedstock. A study definition of "maximum CO_2 capture" was established which considers the following:

- A 2x3 water gas shift (WGS) reactor arrangement was incorporated into each plant to increase the CO conversion efficiency to 97.2 percent. The conversion rate matches that of the E-Gas™ IGCC cases studied in the NETL Fossil Energy Baseline study which employed the same reactor arrangement. [31]
- SMR with CCS incorporates CO_2 capture technologies to treat the pressurized syngas stream as well as the reactor furnace low-pressure flue gas
- The ATR and gasification cases with CCS incorporate CO_2 capture technologies treating only the pressurized syngas stream. No low-pressure CO2 capture technologies are incorporated.
- No partial bypasses of the acid gas removal (AGR) technologies are included as control mechanisms for targeting specific levels of CO_2 capture in the plant flowsheets
- No adjustments have been made to vendor-provided CO_2 separation efficiencies of the AGR technologies

2.8 Life Cycle Analysis of Greenhouse Gas Emissions

The goal of the LCA performed was to generate attributional life cycle GHG emissions estimates for each study case.

Life cycle GHG emissions are reported for select processes occurring outside of the plant boundary as well as atmospheric CO_2 uptake during the regrowth of biomass. While LCA can include the inventory of many types of flows, such as criteria air pollutants, the focus of this LCA was greenhouse gases.

The final life cycle results were generated on a functional unit basis of 1 lb of H_2 produced. The life cycle impact assessment method used is from the International Panel of Climate Change 5th Assessment Report (IPCC AR5), [41] which, amongst other features, provides GWP values that can be used to convert emissions of various greenhouse gases into equivalent emissions of carbon dioxide, or CO_2e. The results reflect a 100-year timeframe with atmospheric carbon feedback, using the values shown in Exhibit 2-15. As such, the life cycle GHG emissions results are generated in units of lb CO_2e/lb H_2.

Exhibit 2-15. Fossil GWP values for select GHGs

Greenhouse Gas	Formula	GWP Value
Carbon dioxide	CO_2	1
Methane	CH_4	36
Nitrous oxide	N_2O	298

The system boundary for the LCA of the different hydrogen pathways includes natural resources consumed, utilities, relevant waste, products, and co-products where applicable. Case 1 was modeled in two ways given the presence of steam as a co-product, once by ignoring the effect and once by using an LCA method called system expansion with displacement to credit the emissions that would otherwise be needed to produce the extra steam. Similarly, Case 6's model performs system expansion to credit net negative electricity emissions due to the small amount of electricity that is sold back to the grid.

See Exhibit 2-16 through Exhibit 2-21 for diagrams of each study case's system boundary. The parallelogram shapes indicate upstream emissions data whereas the rectangle shapes indicate downstream processes. Various non-GHG inventory flows are represented in the diagrams but were not included as results in this study.

Exhibit 2-16. Case 1 system boundary, steam methane reforming plant without CO_2 capture

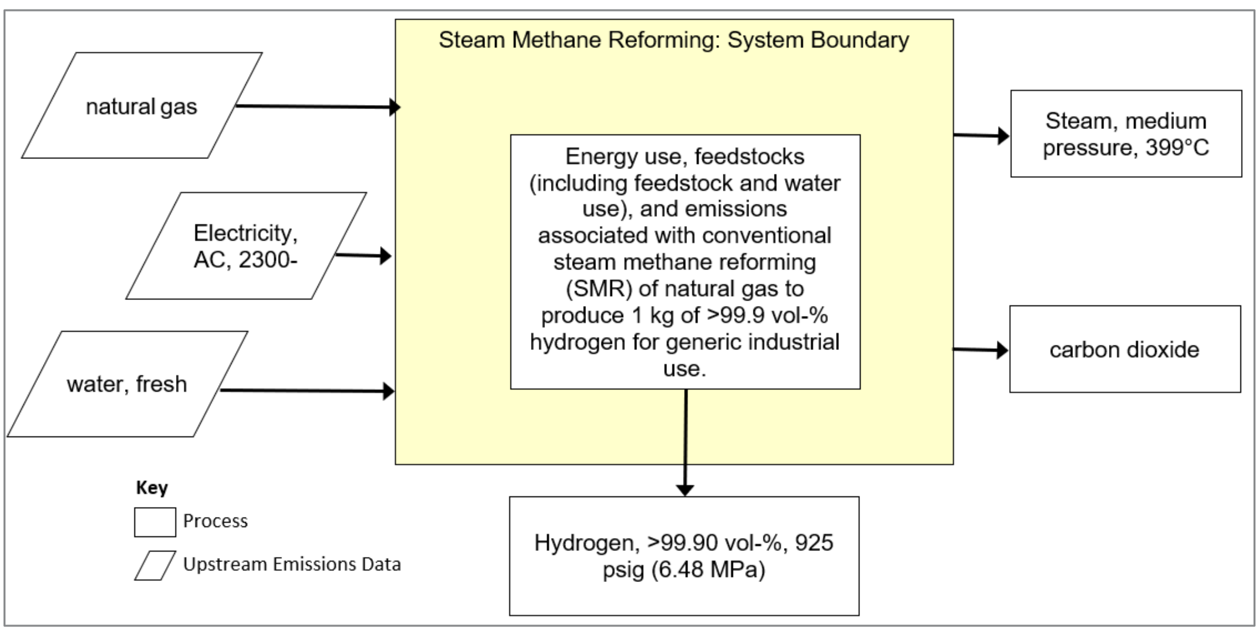


Exhibit 2-17. Case 2 system boundary, steam methane reforming plant with CO_2 capture

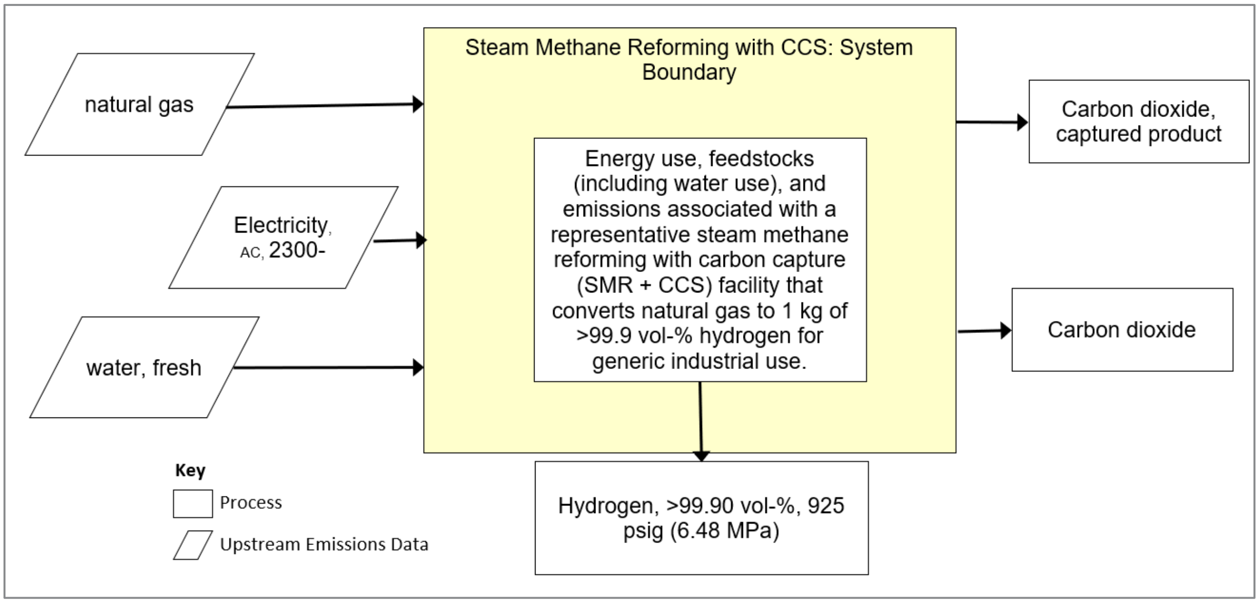


Exhibit 2-18. Case 3 system boundary, autothermal reforming plant with CO_2 capture

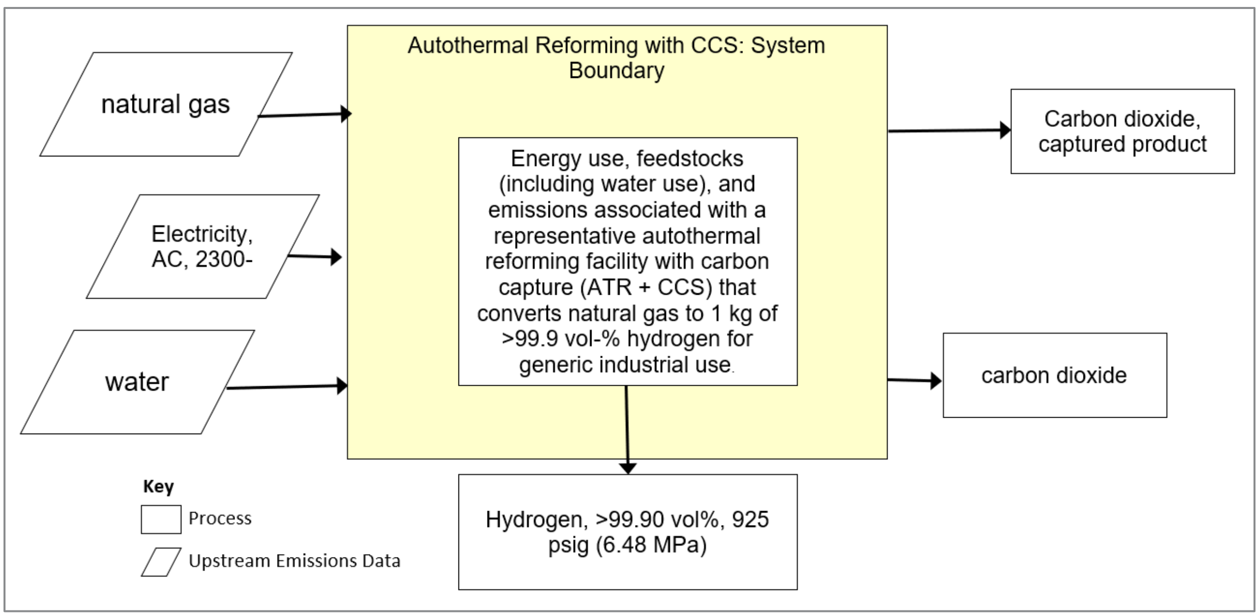


Exhibit 2-19. Case 4 system boundary, coal gasification plant without CO_2 capture

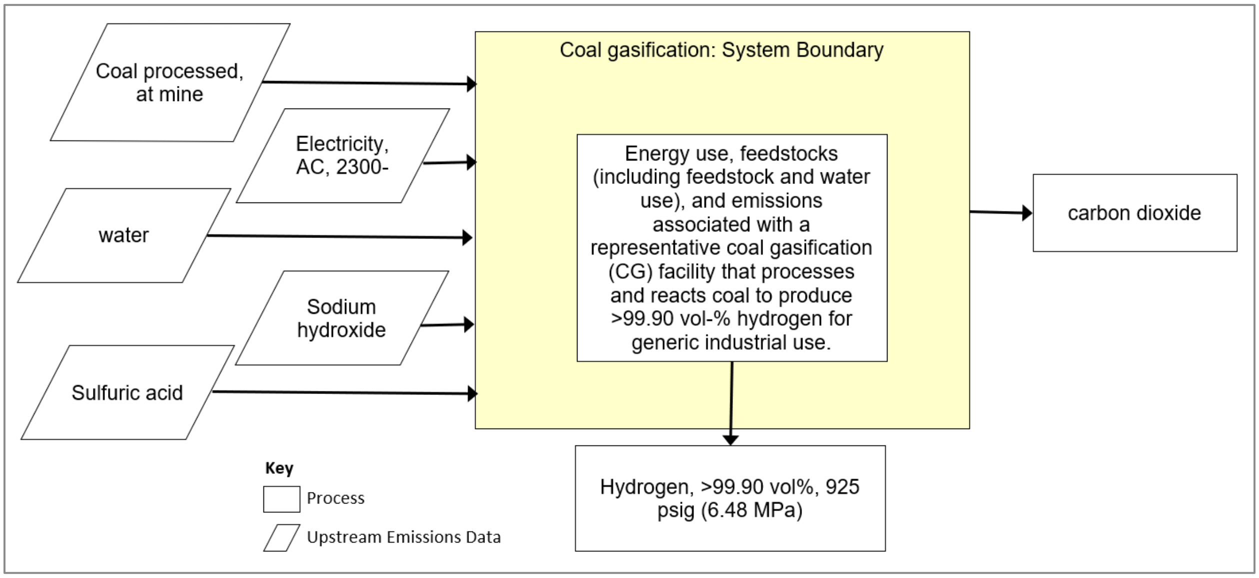


Exhibit 2-20. Case 5 system boundary, coal gasification plant with CO_2 capture

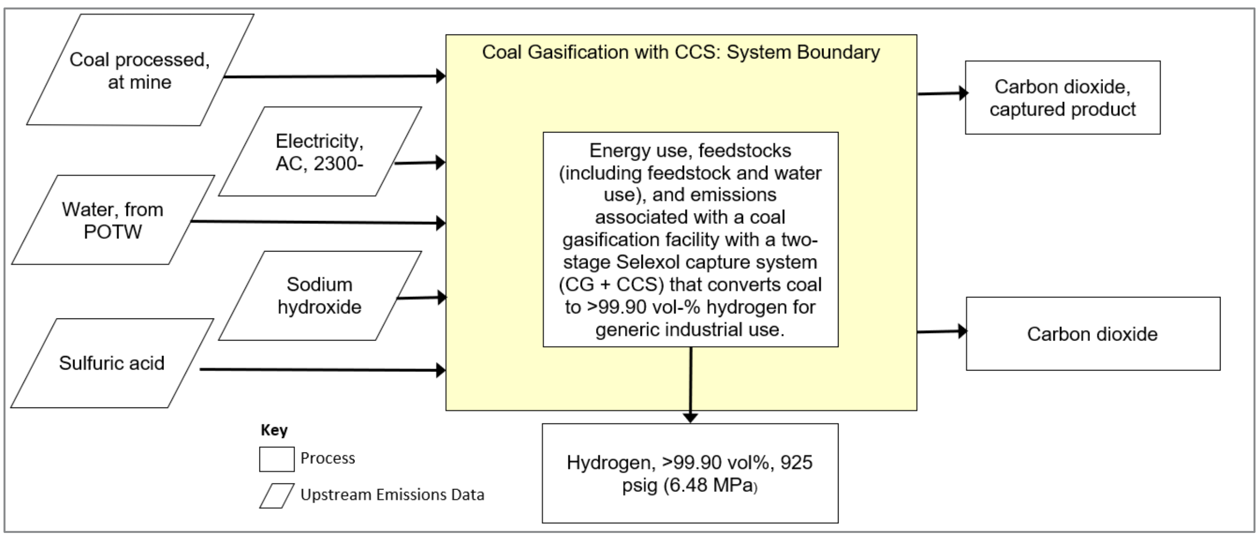
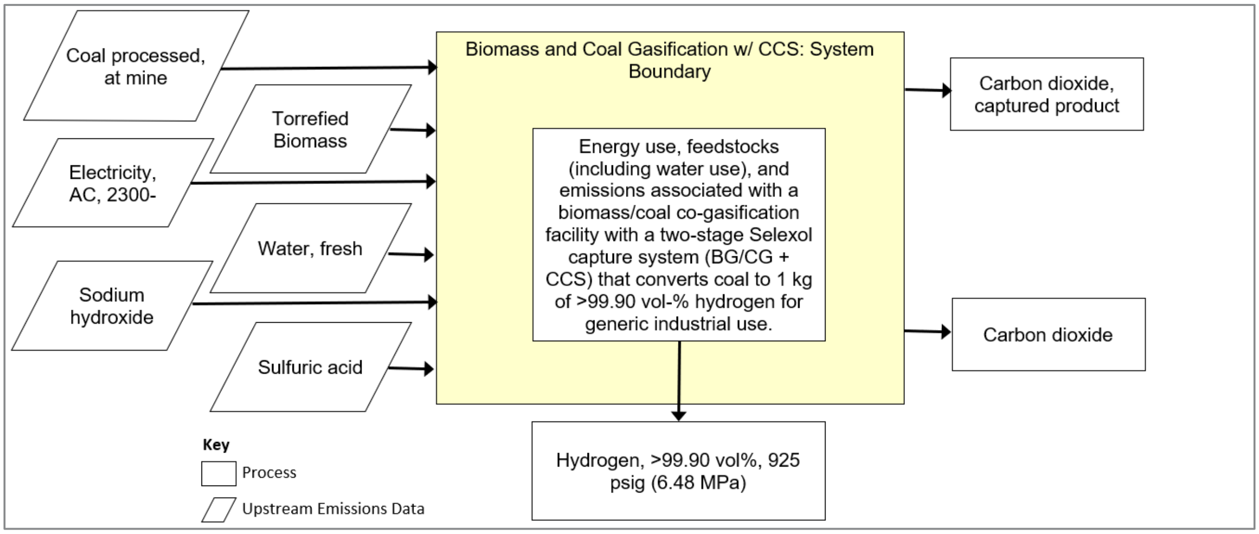


Exhibit 2-21. Case 6 system boundary, coal/biomass co-gasification plant with CO_2 capture



OpenLCA software and various existing NETL LCA resources were used to model the systems defined above. Notably, this effort builds upon past NETL work in which life cycle models were developed for producing various feedstocks for hydrogen production plants. The following gives more information about emissions data related to all modeled upstream and downstream processes:

- Upstream production of natural gas was modeled using the NETL natural gas model, which is documented in the NETL 2019 natural gas baseline report. [9] The boundaries for the model start at natural gas extraction and end at transmission, the same boundaries used in evaluating the electricity generation cases in Section 7 of the natural gas report. All parameters and assumptions used in the model are documented in the report. The U.S. average and basin-level data was updated to the year 2017 based on a more recent NETL report [10].
- Upstream production of biomass, specifically the torrefied Southern Yellow Pine (SYP) used in this study, followed work previously done for the Connecticut Center for Advanced Technology (CCAT). [13] All but two parameters used in the modeling of biomass came from this previous work. The biomass loss rate during harvesting was adjusted to be 5%. Upstream potash fertilizer production was updated to a more-recent value found in a 2018 journal article. [42]
- The upstream production of coal leverages modeling and default assumptions from the NETL Supercritical Pulverized Coal (SCPC) Power Plant study. [14] The SCPC study considers upstream production of coal from the Illinois basin, and Illinois underground bituminous coal was assumed to represent the Illinois #6 coal used in this study. The coal mine methane emissions found in the SCPC study were adjusted to 0.0086 lb CH_4/lb coal. Transportation of coal from mine to plant assumes U.S. average distances of 3.8

miles by truck, 35.1 miles by barge, 42.1 miles by ocean/lake vessel, and 577 miles by train.

- Process information for the storage of CO_2 in a saline aquifer followed modeling and assumptions used in the NETL SCPC Power Plant study. [14]
- Electricity impacts are modeled using the NETL electricity baseline, represented by the U.S. average consumption mix at the user. [11] [12]

Excluded from the factors considered are life cycle GHG emissions associated with plant construction, additional upstream material and energy infrastructure necessary for deliveries to the hydrogen production facility (e.g., connector rail spurs, roads, pipelines), and emissions from water conditioning at the hydrogen production facility. Based on other GHG LCAs performed by NETL, impacts from these areas are assumed to be minor contributors to the overall life cycle GHG emission profiles of the study cases compared to the other emissions sources which occur throughout the operating lifetime of the plant. [43] [14]

2.9 CAPACITY FACTOR

2.9.1 Capacity Factor Assumptions

Availability is the percent of time during a specific period that a generating unit is capable of producing hydrogen. This report assumes that each new plant would be dispatched any time it is available and would be capable of generating the nameplate capacity when online. Therefore, the capacity factor (CF) and availability are equal. The calculations assume that the CF and availability are constant over the life of the plant, but in practice, a plant will have a higher peak availability to counter lower availability in the first several years of operation.

2.9.2 Capacity Factor for Reforming Plants

The overall availability of the reforming plants was assumed to be 90 percent, which is consistent with commercial plants. Each reforming plant assumes a single reactor operating at full capacity.

2.9.3 Capacity Factor for Gasification Plants

The Electric Power Research Institute (EPRI) has reported an availability goal for IGCC plants of 85 percent. [44] Plants built before 2000 have achieved availability of 80 percent for limited periods. Common projections from technology suppliers and EPRI are that IGCC plants are capable of 80–85 percent availability without a spare gasifier and could achieve greater than 90 percent availability with a spare gasifier. [45], [46], [47], [48], [49], [50]

While an availability of 85 percent is the goal, given the technology experience of the similar IGCC process to that of the gasification plants in this study, a CF of 80 percent was selected for gasification plants with no spare gasifier.

2.9.4 Capacity Factor for Plants with Carbon Capture

The implementation of carbon capture and storage adds extra equipment to the hydrogen plant. Preliminary reliability analyses show that small reductions in reliability may occur with the addition of carbon capture and storage technology. The solvent-based carbon capture and CO_2 compression technologies used in this study have commercial operating experience, although at smaller scales, and have demonstrated high reliability. Therefore, given the report basis and use of commercial technology, the assumption is made that the CFs for a given plant with and without carbon capture are the same.

2.10 Raw Water Withdrawal and Consumption

A water balance was performed for each case on the major water consumers in the process. The total water demand for each subsystem was determined, and internal recycle water available from various sources like condensate from syngas, or from flue gas (in SMR with CO_2 capture cases) was applied to offset the water demand. The difference between demand and recycle is raw water withdrawal. Raw water withdrawal is the water removed from the ground or diverted from a municipal source for use in the plant. Raw water consumption is also accounted for as the portion of the raw water withdrawn that is evaporated, transpired, incorporated into products, or otherwise not returned to the water source from which it was withdrawn.

Raw water makeup was assumed to be provided 50 percent by a publicly owned treatment works (POTW) and 50 percent from groundwater. Raw water withdrawal is defined as the water metered from a raw water source and used in the plant processes for all purposes, such as cooling tower makeup, boiler feedwater (BFW) makeup, and reforming steam, depending on the technology examined. The difference between withdrawal and process water returned to the source is consumption. Consumption represents the net impact of the process on the water source.

BFW blowdown and air separation unit (ASU) knockout were assumed to be treated and recycled to the cooling tower. The cooling tower blowdown was assumed to be treated and 90 percent returned to the water source.

The largest consumer of raw water for SMR without CCS (Case 1) is steam generation for reforming and export. In all the remaining cases it is cooling tower makeup. It was assumed that all cases utilized a mechanical draft, evaporative cooling tower. The design ambient wet bulb temperature of 11 °C (51.5 °F) (Exhibit 2-1 and Exhibit 2-2) was used to achieve a cooling water temperature of 16 °C (60 °F) using an approach of 5 °C (8.5 °F). The cooling water range was assumed to be 11 °C (20 °F). The cooling tower makeup rate was determined using the following [51]:

- Evaporative losses of 0.8 percent of the circulating water flowrate per 5.5 °C (10 °F) of range
- Drift losses of 0.001 percent of the circulating water flowrate
- Blowdown losses (BDL) were calculated as follows:

$$BDL = \frac{EL}{CC - 1}$$

Where:

EL – Evaporative Losses

CC – Cycles of concentration

The cycles of concentration are a measure of water quality and a mid-range value of four was chosen for this report.

The water balances presented in subsequent sections include the water demand of the major water consumers within the process, the amount provided by internal recycle, the amount of raw water withdrawal by difference, the amount of process water returned to the source, and the raw water consumption, again by difference.

2.11 Cost Estimating Methodology

Detailed information pertaining to topics such as contracting strategy; engineering, procurement, and construction (EPC) contractor services; estimation of capital cost contingencies; owner's costs; cost estimate scope; economic assumptions; finance structures; and LCOHs are available in the 2019 revision of the QGESS document "Cost Estimation Methodology for NETL Assessment of Power Plant Performance." [6] Select portions are repeated in this report for completeness.

Capital Costs:

The capital cost estimates documented in this report reflect different uncertainty ranges depending on the technology considered as shown in Exhibit 2-22.

Exhibit 2-22. Capital cost uncertainty ranges

Technology	Uncertainty Range	AACE Classification
Gasification	-25/+50	Class 5
Reforming	-15/+25	Class 4

Gasification cases carry an uncertainty range of -25 percent/+50 percent, consistent with Association for the Advancement of Cost Engineering (AACE) Class 5 cost estimates (i.e., feasibility study) [6] [52] [53], based on the level of engineering design performed. This range has been deemed reflective of recent commercial power IGCC experience in the NETL Fossil Energy Baseline study. [31] This uncertainty range is considered applicable to the gasification cases in this study given the number of common process systems with IGCC plants and the limited recent development of coal gasification projects in the United States in recent years.

Reforming cases carry smaller uncertainty ranges, and both fall within AACE Class 4 estimates. Given recent experience with SMR and ATR plants, the reforming uncertainty range is smaller than gasification.

In all cases, this report intends to represent the next commercial offering and relies on vendor cost estimates for component technologies. It also applies process contingencies at the appropriate subsystem levels in an attempt to account for expected but undefined costs, which can be a challenge for emerging technologies.

Costs of Mature Technologies and Designs:

The cost estimates for plant designs that only contain fully mature technologies, which have been widely deployed at commercial scale (e.g., SMR and gasification without CO_2 capture), reflect NOAK on the technology commercialization maturity spectrum. The costs of such plants have dropped over time due to "learning by doing" and risk reduction benefits that result from serial deployments as well as from continuing R&D.

Costs of Emerging Technologies and Designs:

The cost estimates for plant designs that include technologies that are not yet fully mature (e.g., any plant with CO_2 capture) use the same cost estimating methodology as for mature plant designs, which does not fully account for the unique cost premiums associated with the initial, complex integrations of emerging technologies in a commercial application. Thus, it is anticipated that early deployments of these plants may incur costs higher than those reflected within this report.

Other Factors:

Actual reported project costs for all the plant types are also expected to deviate from the cost estimates in this report due to project- and site-specific considerations (e.g., contracting strategy, local labor costs, seismic conditions, water quality, financing parameters, local environmental concerns, weather delays) that may make construction more costly. Such variations are not captured by the reported cost uncertainty.

Future Cost Trends:

Continuing research, development, and demonstration (RD&D) is expected to result in designs that are more advanced than those assessed by this report, leading to costs that are lower than those estimated here.

2.11.1 Capital Costs

As illustrated in Exhibit 2-23, this report defines capital cost at five levels: BEC, EPCC, TPC, TOC, and TASC. BEC, EPCC, TPC, and TOC are "overnight" costs and are expressed in "base-year" dollars. The base year is the first year of capital expenditure. TASC is expressed in mixed, current-year dollars over the entire capital expenditure period, which is assumed in most NETL studies to last five years for coal plants and three years for natural gas plants.

The Bare Erected Cost (BEC) comprises the cost of process equipment, on-site facilities and infrastructure that support the plant (e.g., shops, offices, labs, road), and the direct and indirect labor required for its construction and/or installation. The cost of EPC services and contingencies are not included in BEC.

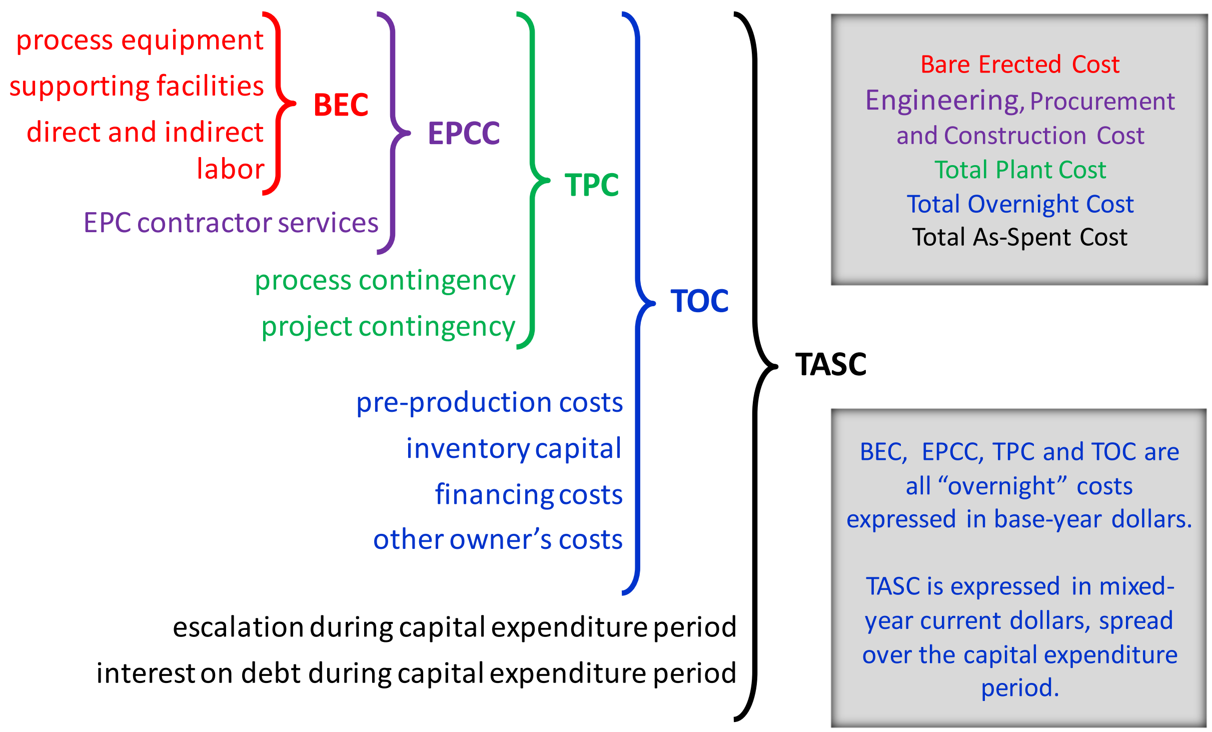
The <u>Engineering, Procurement and Construction Cost</u> (EPCC) comprises the BEC plus the cost of services provided by the EPC contractor. EPC services include: detailed design, contractor permitting (i.e., those permits that individual contractors must obtain to perform their scopes of work, as opposed to project permitting, which is not included here), and project/construction management costs.

The <u>Total Plant Cost</u> (TPC) comprises the EPCC plus project and process contingencies.

The <u>Total Overnight Cost</u> (TOC) comprises the TPC plus all other overnight costs, including owner's costs. TOC does not include escalation during construction or interest during construction.

The <u>Total As-Spent Cost</u> (TASC) is the sum of all capital expenditures as they are incurred during the capital expenditure period including their escalation. TASC also includes interest during construction, comprising interest on debt and a return on equity.

Exhibit 2-23. Capital cost levels and their elements



2.11.1.1 Cost Estimate Basis and Classification

Full, bottom-up estimates for each case in this study were not pursued. Instead, the TPC and operation and maintenance (O&M) costs for each of the cases were estimated using a combined bottom-up and scaling approach. Black & Veatch leveraged an in-house database and conceptual estimating models. Costs were further calibrated using a combination of adjusted vendor-furnished data and scaled estimates from previous design/build projects. Costs were further calibrated using a combination of adjusted vendor-furnished data and scaled estimates from previous design/build projects. In addition, cost scaling was performed on select plant areas using reference costs from prior NETL studies according to the 2019 revision of the QGESS document “Capital Cost Scaling Methodology: Revision 4 Report.” [7]

2.11.1.2 System Code-of-Accounts

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

2.11.1.3 Estimate Scope

The estimates represent a complete hydrogen production facility on a generic site. The plant boundary limit is defined as the total plant facility within the “fence line” including feedstock receiving and water supply system but terminating at the injection point into the hydrogen pipeline network. Hydrogen pipeline and storage is beyond the estimate scope. CO_2 transport and storage (T&S) cost is not included in the reported capital cost or O&M costs but is treated separately and added to the LCOH.

2.11.1.4 Capital Cost Assumptions

Black & Veatch developed select portions of the capital cost estimates for each plant using the company’s in-house database and conceptual estimating methodology for each of the specific technologies. A reference bottom-up estimate for each major component provides the basis for the estimating models. The following assumptions are reflected in the bottom-up costs as well as the costs scaled according to NETL QGESS methodology:

- Labor costs are based on Midwest, Merit Shop. The estimating models are based on a U.S. Gulf Coast location and the labor cost has been factored to a Midwest location. Labor cost data were sourced from recent projects and proprietary Black & Veatch in-house references/cost databases.
- The estimates are based on a competitive bidding environment, with adequate skilled craft labor available locally.
- Labor is based on a 50-hour work week (5 days x 10 hours). No additional incentives such as per- diem allowances or bonuses have been included to attract craft labor.
- While not included at this time, labor incentives may ultimately be required to attract and retain skilled labor depending on the amount of competing work in the region, and the availability of skilled craft in the area at the time the projects proceed to construction.
- The estimates are based on a greenfield site.
- The site is considered to be Seismic Zone 1, relatively level, and free from hazardous materials, archeological artifacts, or excessive rock. Soil conditions are considered adequate for spread footing foundations. The soil bearing capability is assumed adequate such that piling is not needed to support the foundation loads.

- Engineering and Construction Management are estimated based on Black & Veatch's historical experience in designing and building power projects. The cost, as a percentage of BEC, varies by technology; 20 percent for reforming and 15 percent for gasification. The percentages were selected based on those used for natural gas combined cycle (NGCC) and IGCC plant cost estimation in the NETL Fossil Energy Baseline study. [31] These costs consist of all home office engineering and procurement services as well as field construction management costs. Site staffing generally includes construction manager, resident engineer, scheduler, and personnel for project controls, document control, materials management, site safety, and field inspection.

2.11.1.5 Price Fluctuations

During the writing of this report, the prices of equipment and bulk materials fluctuated as a result of various market forces. Some reference quotes pre-dated the 2018-year cost basis while others may be considered more historical. All vendor quotes used to develop these estimates were adjusted to December 2018 dollars accounting for the price fluctuations. Price indices, e.g., The Chemical Engineering Plant Cost Index [54] and the Gross Domestic Product Chain-type Price Index [55], were used as needed for these adjustments. While these overall indices are nearly constant, it should be noted that the cost of individual equipment types may still deviate from the December 2018 reference point.

2.11.1.5.1 Process Contingency

Process contingencies were applied to the gasification and reforming with capture estimates in this report, with justification provided, as follows:

- Gasifiers and Syngas Coolers: 14 percent on all cases—next-generation commercial offering
- Two-Stage Selexol: 20 percent on gasification with capture cases—unproven technology at commercial scale in gasification service
- Mercury Removal: 5 percent on all gasification cases—minimal commercial scale experience in gasification applications
- Instrumentation and Controls: 5 percent on most accounts—integration issues

Process contingencies were applied to the reforming estimates in this report as follows:

- Cansolv System: 18 percent on SMR with capture case—post-combustion capture process unproven at commercial scale for SMR applications. No contingency added for MDEA in any case due to the extensive commercial experience in hydrogen applications.
- Instrumentation and Controls: 5 percent on most accounts within reforming with capture cases—integration issues

2.11.1.6 Owner's Costs

Detailed explanation of the owner's costs is available in the 2019 revision of the QGESS document "Cost Estimation Methodology for NETL Assessment of Power Plant Performance."

[6] Owner's costs are split into three categories: pre-production costs, inventory capital, and other costs.

Pre-production allocations are expected to carry the specific plants through substantial completion, and to commercial operation. Substantial completion is intended to represent the transfer point of the facility from the EPC contractor (development entity) to the end user or owner, and is typically contingent on mutually acceptable equipment closeout, successful completion of facility-wide performance testing, and full closeout of commercial items.

Two examples of what could be included in the "other" owner's costs are rail spur and switch yard costs. Rail spur costs would only be applied to the gasification cases; however, the switch yard costs would be included in all cases.

Switch yard costs are dependent on voltage, configuration, number of breakers, layout, and air-insulated versus gas-insulated. As a rule of thumb, a 345-kilovolt (kV) switchyard (air-insulated, ring bus) would cost roughly $850,000 per breaker.

On-site only rails (excludes long runs) would be expected to cost in the range of $850,000 to $950,000 per mile (relatively flat level terrain) plus the costs of any switches/turnouts (approximately $50,000 each) and road crossings (approximately $300 per linear foot).

2.11.2 Operation and Maintenance Costs

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

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

There are two components of O&M costs: fixed O&M, which is independent of hydrogen production, and variable O&M, which is proportional to hydrogen production. Taxes and insurance are included as fixed O&M costs, totaling 2 percent of the TPC.

2.11.2.1 Operating Labor

Operating labor cost was determined based on the number of operators required for each technology. The average base labor rate used to determine annual cost is $38.50/hour. The associated labor burden is estimated at 30 percent of the base labor rate.

2.11.2.2 Maintenance Material and Labor

Maintenance cost was evaluated on the basis of relationships of maintenance cost to initial capital cost. This represents a weighted analysis in which the individual cost relationships were considered for each major plant component or section.

2.11.2.3 Administrative and Support Labor

Labor administration and overhead charges are assessed at a rate of 25 percent of the burdened O&M labor.

2.11.2.4 Consumables

The cost of consumables, including fuel, was determined based on each consumable's individual rate of consumption, the unit cost of the specific consumable commodity, and the plant annual operating hours. The consumption rate is determined from process data for fuel and water, the initial fill amount and lifetime of the consumable for catalysts and adsorbents, or vendor input for other consumables.

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

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

Initial fills of the consumables, fuels, and chemicals may be accounted for directly in the O&M tables or included with the equipment pricing in the capital cost. Where applicable, the O&M tables state where this cost is included on a case-by-case basis.

2.11.2.5 Waste Disposal

Waste quantities and disposal costs were determined/evaluated similarly to the consumables. Chemical and catalyst waste streams are individually reported, in addition to others such as slag. Spent reforming catalysts are assumed to carry a residual value in buyback from the supplier. No resale value is credited. Instead, a \$0/ft^3 disposal cost is assumed. Waste disposal costs were separated into two categories: non-hazardous and hazardous waste. Non-hazardous waste is disposed of at a rate of \$41.9/tonne (\$38.0/ton). Hazardous waste is disposed of at a rate of \$88.2/tonne (\$80.0/ton).

2.11.2.6 Co-Products and By-Products

By-product quantities were also determined similarly to the consumables. For cases exporting power, a grid sales price of \$71.7/MWh was assumed, the same price at which cases importing power must purchase it. This price is the average price paid by an industrial consumer in the Midwest Independent System Operator (MISO) region in 2019[e]. However, due to the variable

[e] As reported in "Annual Electric Power Industry Report," Form EIA-861

marketability of other by-products, specifically export steam and sulfur, no credit was taken for their potential salable value.

It should be noted that by-product credits and/or disposal costs could potentially be an additional determining factor in the choice of technology for some companies and in selecting some sites. A high local value of the product can establish whether added capital should be included in the plant costs to produce a particular co-product. Slag is a potential by-product in certain markets. However, as stated above, these material streams are considered waste in this report with a concomitant disposal cost.

2.11.3 CO_2 Transport and Storage

The cost of CO_2 T&S in a deep saline formation is estimated using the Fossil Energy (FE)/NETL CO_2 Transport Cost Model (CO_2 Transport Cost Model) and the FE/NETL CO_2 Saline Storage Cost Model (CO_2 Storage Cost Model). Additional detail on development of these costs is available in the 2019 revision of the QGESS document "Carbon Dioxide Transport and Storage Costs in NETL Studies." [56]

Due to the variances in the geologic formations that make up saline formations across the United States, the cost to store CO_2 will vary depending on location. Storage cost results from the CO_2 Storage Cost Model align with generic plant locations from the NETL studies that utilize the coal found in those particular basins:

- Midwest plant location – Illinois Basin
- Texas plant location – East Texas Basin
- North Dakota plant location – Williston Basin
- Montana plant location – Powder River Basin

The far-right column of Exhibit 2-24 shows the total T&S costs used in NETL system studies for each plant location rounded to the nearest whole dollar. Only the \$10/tonne (\$9/ton) value is used in this study report since all cases are in the Midwest.

Exhibit 2-24. CO_2 transport and storage costs

Plant Location	Basin	Transport (2018 \$/tonne)	Storage Cost at 25 Gt (2018 \$/tonne)	T&S Value for System Studies[A] (2018 \$/tonne)
Midwest	Illinois	2.07	8.32	10
Texas	East Texas		8.66	11
North Dakota	Williston		12.98	15
Montana	Powder River		19.84	22

[A]The sum of transport and storage costs rounded to the nearest whole dollar

2.11.4 LCOH and Breakeven CO_2 Sales Price and Stack Emissions Penalty

The LCOH is the amount of revenue required per kilogram of H_2 produced during the plant's operational life to meet all capital and operational costs. The real LCOH can be obtained from the following formula:

$$LCOH = LCC + LOM + LFP$$

Where:

LCOH – the levelized cost of hydrogen, reported in $/kg

LCC – the levelized capital cost

LOM – the levelized operating and maintenance cost

LFP – the levelized fuel price

The method used to determine capital recovery factor and levelization factors for operating and maintenance and fuel costs is found in the Cost Estimating Quality Guideline. [6] No changes were made to the standard global economic assumptions. However, changes to the financial structure assumed were made to reflect the financing of hydrogen production facilities. These customized assumptions are shown in Exhibit 2-25.

Exhibit 2-25. Nominal and real rates financial structure for hydrogen production

Type of Security	% of Total	Current Dollar Cost	Weighted Average Cost of Capital	After-Tax Weighted Average Cost of Capital
Nominal				
Debt	38%	7.25%	2.76%	2.05%
Equity	62%	5.16%	3.20%	3.20%
Total			5.96%	5.25%
Real (based on 2.01% average real GDP deflator, 1990-2018) [57]				
Debt	38%	5.15%	1.96%	1.45%
Equity	62%	3.10%	1.92%	1.92%
Total			3.88%	3.37%

The breakeven CO_2 sales price represents the minimum CO_2 plant gate sales price that will incentivize carbon capture relative to a defined reference non-capture plant. The breakeven CO_2 sales price is calculated using the following formula:

$$Breakeven\ CO_2\ Sales\ Price\ (\frac{\$}{tonne}) = \frac{(LCOH_{CCS} - LCOH_{Non\ CCS})}{CO_2\ Captured}$$

The breakeven stack CO_2 emissions penalty represents the minimum stack CO_2 emissions price, when applied to both the capture and non-capture plant, that will incentivize carbon capture relative to a defined reference non-capture plant. The breakeven stack CO_2 emissions penalty is calculated using the following formula:

$$Breakeven\ Stack\ CO_2\ Emissions\ Penalty\ (\frac{\$}{tonne}) = \frac{(LCOH_{CCS\ with\ T\&S} - LCOH_{Non\ CCS})}{CO_2\ Emissions_{Non\ CCS} - CO_2\ Emissions_{CCS}}$$

Where:

CCS – the capture plant for which the breakeven CO_2 sales price/emissions penalty is being calculated (excluding T&S unless otherwise noted)

Non-CCS – the reference non-capture plant, as described below

LCOH – the levelized cost of hydrogen, reported in $/kg

The CCS plant includes CO_2 compression to 15.3 MPa at 2,215 pounds per square inch absolute (psia)

For CO_2 Sales Price, the levelized cost of hydrogen (LCOH) excludes T&S costs

For CO_2 Emissions Penalty, the LCOH includes T&S costs

CO_2 Captured – the rate of CO_2 captured, reported in tonne/kg H_2

Stack CO_2 Emissions – the rate of CO_2 emitted out the stack, reported in tonne/kg H_2

For a greenfield reforming plant (SMR or ATR) with CCS, the reference plant used to calculate the breakeven CO_2 sales price/emission penalty is a non-capture SMR plant.

For today's greenfield gasification plant with CCS, the reference non-capture plant used to calculate the breakeven CO_2 sales price/emission penalty is a coal gasification plant without capture.

REFORMING PLANTS

3 Reforming Plants

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

Cases 1 and 2 use a steam methane reformer and differ in that Case 1 does not include CO_2 capture while Case 2 does. The cases were sized based on the maximum hydrogen production capacity for a single-train steam methane reforming plant of 200 MMSCFD. The third case, Case 3, uses an autothermal reformer and employs CO_2 capture. This case was sized to match the hydrogen production rate of coal gasification Cases 4 and 5, which is 274 MMSCFD.

3.1 Key System Assumptions

System assumptions for cases 1, 2, and 3 are compiled in Exhibit 3-1.

Exhibit 3-1. Reforming plant study configuration matrix

Case	1	2	3
Feedstock	Natural Gas		
Fuel Pressure at Plant Battery Limit, MPa (psia)	3.1 (450)		
Reforming	Prereformer and single-train, vertical tube steam methane reformer (externally heated)		Prereformer and single train autothermal reformer (O_2-blown)
Plant Capacity	222 kNm^3/hr (200 MMSCFD) H_2 – single train maximum		304 kNm^3/hr (274 MMSCFD) H_2 – match Cases 4 & 5
H_2 Product Purity	99.90 vol%		
H_2 Product Pressure	939.7 psia		
Oxidant	N/A		95 vol% O_2
Water Gas Shift	2x3 train configuration, high-temperature, 97.2% conversion		
H_2 Purification	Pressure Swing Adsorption		
PSA Off-Gas	Recycled as reformer fuel		Recycle to amine unit and fuel
Auxiliar Power Block	None		
Sulfur Control	Zinc Oxide Guard Bed		
NOx Control	LNB		
Particulate Control	N/A		
Mercury Control	N/A		
CO_2 Control	N/A	MDEA and Cansolv	MDEA
CO_2 Storage	N/A	Off-site Saline Formation	
CO_2 Product Pressure	2,215 psia		

3.1.1 Balance of Plant

The balance of plant assumptions are common to all three cases and presented in Exhibit 3-2.

Exhibit 3-2. Balance of plant assumptions

Parameters	Values
Cooling System	Recirculating Wet Cooling Tower
Fuel and Other Storage	
Natural Gas	Pipeline supply at 3.0 MPa (450 psia) and 15 °C (59 °F)
Plant Distribution Voltage	
Motors below 1 hp	110/220 V
Motors between 1 hp and 250 hp	480 V
Motors between 250 hp and 5,000 hp	4,160 V
Motors above 5,000 hp	13,800 V
Grid Interconnection Voltage	345 kV
Water and Wastewater	
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
Process Wastewater	Storm water that contacts equipment surfaces is collected and treated for discharge
Sanitary Waste Disposal	Design includes a packaged domestic sewage treatment plant with effluent discharged to the industrial wastewater treatment system. Sludge is hauled off site. Packaged plant is sized for 5.68 cubic meters per day (1,500 gallons per day)
Water Discharge	Blowdown from the cooling tower is softened and passed through a two-stage reverse osmosis (RO) with pre-treatment and demineralizer before being discharged

3.2 Sparing Philosophy

The sparing philosophy for cases 1, 2, and 3 is provided below. Only a single train is necessary at the scales of hydrogen production considered. There is no redundancy other than normal sparing of rotating equipment. The plant design consists of the following major subsystems:

- One train of reforming, including prereformer and reformer (1 x 100 percent)
- One train ASU in Case 3 (1 x 100 percent)
- Two trains of three water gas shift reactors in series (2 x 50 percent)

- One train of MDEA in Case 2 and Case 3 and one train of Cansolv in Case 2 (1 x 100 percent)
- One train of CO_2 compression and drying (1 x 100 percent) in Case 2 and Case 3
- One train of H_2 production and compression systems (1 x 100 percent)

3.3 Reforming Common Process Areas

The reforming cases have process areas which are common to each plant configuration. As detailed descriptions of these process areas for each case would be repetitious, they are presented in this section for general background information. Where there is case-specific performance information, the performance features are presented in relevant case sections.

3.3.1 Natural Gas Supply System

It was assumed that a natural gas main with adequate capacity is near to the site fence line (within 16 km [10 mi]) and that a suitable right-of-way is available to install a branch line to the site. For the purposes of this report, it was also assumed that the gas will be delivered to the plant custody transfer point at sufficient pressure such that natural gas is available at a pressure of 3.1 MPa (450 psia) and a temperature of 15 °C (59 °F). Hence, neither a pressure reducing station, nor a booster compressor is required.

A new gas metering station is assumed to be added on the site. The meter may be of the rate-of-flow type, with input to the plant computer for summing and recording or may be of the positive displacement type. In either case, a complete timeline record of gas consumption rates and cumulative consumption is provided.

3.3.2 Sulfur Polishing

The natural gas feed contains 5.7 parts per million volume (ppmv) of a mercaptan additive methanethiol (CH_4S). Since even a small amount of sulfur present can poison the reformer catalysts, sulfur removal before the prereformer is necessary.

A zinc oxide sulfur guard bed is employed to polish the natural gas feed. The sulfur guard bed is a fixed bed, catalytic reactor type. The NG feedstock is preheated to 370 °C (698 °F) and 98 percent of the CH_4S in the natural gas is removed, resulting in a maximum sulfur content of 0.1 ppm exiting the sulfur guard bed.

3.3.3 Prereformer

The desulfurized natural gas feedstock is preheated further to a temperature of 500 °C (932 °F) before being mixed with process steam and sent to the prereformer unit. The prereformer is a fixed-bed, catalytic, adiabatic reactor. It serves the primary purpose of reforming the C_{2+} hydrocarbons, such as ethane (C_2H_6), propane (C_3H_8), and butane (C_4H_{10}) in the natural gas feed prior to the primary reformer to reduce carbon deposition on the downstream catalyst. This has the additional impacts of reducing the load on the primary reformer, increasing carbon capture ability, and improving product recovery. The following reactions take place in the prereformer:

$$C_2H_6 + 2H_2O \leftrightarrow 2CO + 5H_2;\ \Delta H°_{rxn} = 346.4\ kJ/mol$$

$$C_3H_8 + 3H_2O \leftrightarrow 3CO + 7H_2;\ \Delta H°_{rxn} = 498.6\ kJ/mol$$

$$C_4H_{10} + 4H_2O \leftrightarrow 4CO + 9H_2;\ \Delta H°_{rxn} = 651.0\ kJ/mol$$

$$CO + 3H_2 \leftrightarrow CH_4 + H_2O;\ \Delta H°_{rxn} = -205.8\ kJ/mol$$

$$CO + H_2O \leftrightarrow CO_2 + H_2;\ \Delta H°_{rxn} = -41.2\ kJ/mol$$

The prereformer has been designed to ensure that the effluent stream contains less than 500 ppm of total C_{2+} hydrocarbons. The primary reformer is discussed within each case's individual process descriptions.

3.3.4 Water Gas Shift Reactors

In all cases, the raw syngas exiting the reforming step is converted to a CO_2 and H_2-rich syngas to maximize H_2 product yield and maximize the extent of CO_2 separation that can be performed on the high-pressure syngas stream. CO is converted to CO_2 by reacting with steam over a bed of catalyst. The exit steam to dry gas molar ratio of the shift reactors is maintained at 0.25 to prevent carbon deposition and deactivation of the catalyst. Ample steam is added to the process gas mixture prior to the prereformer to ensure this steam-to-dry gas ratio is met. There is also a loss of chemical energy associated with the exothermic conversion of water (H_2O) and CO to CO_2 and H_2. In the WGS configuration employed, intercooling is applied between stages and the recovered heat is used to generate steam for use elsewhere in the plant, thus offsetting some of this loss. The following chemical equation is the water gas shift reaction:

$$CO + H_2O \leftrightarrow CO_2 + H_2;\ \Delta H°_{rxn} = -41.2\ kJ/mol$$

The CO shift converter can be located either upstream of the AGR step (sour gas shift) or immediately downstream (sweet gas shift). In the present cases, the WGS is located upstream of the AGR (i.e., sour gas shift) to achieve high levels of carbon capture.

The WGS consists of two paths of parallel fixed-bed reactors arranged in series. Three reactors in series are used in each parallel path to achieve conversion up to 97.2 percent, since one of the goals of this study is to maximize the carbon capture rates of the applicable analyzed configurations. Ample steam is added to the system prior to the prereformer so that the product stream from the reformer contains enough water vapor to drive the WGS reaction and to meet the necessary outlet steam to dry gas ratio.

3.3.5 Hydrogen Production

3.3.5.1 Pressure Swing Adsorber

The pressure swing adsorber unit purifies the syngas into the highly pure hydrogen product meeting the specifications provided in Section 2.6. A design H_2 recovery efficiency of 85 percent is considered for all reforming cases. An H_2-rich off-gas stream is produced as a byproduct which is sent to the primary reformer furnace in the Cases 1 and 2. In Case 3, 31 percent of the mass flow is recycled to the AGR unit for additional CO_2 capture and H_2 recovery. The remaining amount is combusted in a gas-fired heater where process heat is recovered.

3.3.5.2 Hydrogen Compressor

The hydrogen compressor is an integrally geared, multi-stage centrifugal compressor. Interstage cooling to a temperature of 30 °C (86 °F) is required to control the temperature increase. The high-purity hydrogen from the pressure swing adsorber (PSA) is compressed to a pipeline-ready pressure of 6.48 MPa (925 psig).

3.3.6 Accessory Electric Plant

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

3.3.7 Process Water Systems

3.3.7.1 Feedwater

The function of the feedwater (FW) system is to pump the various FW streams to the boiler feedwater heaters before entering the medium pressure (MP) steam drum for reforming steam demands. Feedwater is recovered from multiple sources within the plant configurations as internal recycle. Two 100 percent capacity motor-driven feed pumps are provided for each FW stream. Each pump is provided with inlet and outlet isolation valves, outlet check valves, and individual minimum flow recirculation lines. The recirculation flow is controlled by pneumatic flow control valves. In addition, the suctions of the boiler feed pumps are equipped with strainers.

3.3.7.2 Circulating Water System

The function of the circulating water system (CWS) is to supply cooling water to the low temperature syngas coolers, for the AGR processes, and for the auxiliary cooling system. The system consists of two 50 percent capacity vertical circulating water pumps (CWPs), a mechanical draft evaporative cooling tower, and interconnecting piping.

The auxiliary cooling system is a closed-loop (CL) system. Plate and frame heat exchangers (HXs) with circulating water as the cooling medium are provided. The system provides cooling water to the following systems:

1. BFW pumps
2. Air compressors
3. Sample room chillers
4. Blowdown coolers

The AGR systems in Case 2 and Case 3 require a substantial amount of cooling water that is provided by the plant CWS. The additional cooling load imposed by the AGR system is reflected in the significantly larger CWPs and cooling tower in those cases compared to in Case 1.

3.3.7.3 Buildings and Structures

Structures assumed for natural gas cases can be summarized as follows:

1. CWP House
2. Administration/Office/Control Room/Maintenance Building
3. Water Treatment Building
4. Fire Water Pump House

3.3.8 Instrumentation and Control

An integrated plant-wide distributed control system (DCS) is provided. The DCS is a redundant microprocessor-based, functionally distributed system. The control room houses an array of video monitors and keyboard units. The monitor/keyboard units are the primary interface between the generating process and operations personnel. The DCS incorporates plant monitoring and control functions for all the major plant equipment. The DCS is designed to provide 99.5 percent availability.

3.3.9 Miscellaneous Systems

An onsite water treatment facility treats all runoff, cleaning wastes, blowdown, and backwash. It is anticipated that the treated water will be suitable for discharge into existing systems and be within EPA standards for suspended solids, oil and grease, pH, and miscellaneous metals.

The waste treatment system is minimal and consists, primarily, of neutralization and oil/water separators (along with the associated pumps, piping, etc.).

Miscellaneous systems consisting of service air, instrument air, and service water are provided. All truck roadways and unloading stations inside the fence area are provided.

3.3.10 Performance Summary Metrics

3.3.10.1 Effective Thermal Efficiency

The effective thermal efficiency is useful for comparing the overall system performance of hydrogen production systems while considering the electrical power net electrical power requirements to operate the facility. Facilities generating power sufficient to export to the grid will exhibit higher effective thermal efficiencies while configurations not exporting power to the grid will exhibit lower effective thermal efficiencies.

This metric is calculated by adding the thermal content of the hydrogen product to the net power and dividing that difference by the thermal input to the system, expressed as a percentage. The efficiency can be calculated on both a HHV and a lower heating value (LHV) basis. This calculation is represented by the following equation:

$$ETE = \frac{(TP + P_{net})}{TI} * 100\%$$

Where:

ETE – effective thermal efficiency

TP – thermal content of the hydrogen product

TI – thermal input to the system

P_{net} – net power of the system

The thermal input to the system is calculated by taking the feed rate into the system and multiplying by the heating value of the feedstock.

The thermal output is calculated by taking the mass flowrate of the hydrogen product and multiplying by the heating value of the product.

The net power of the system is the total gross power generated minus the total auxiliary load of the system. The net power is a negative value in all three reforming cases because none of these configurations generate power and require grid imports.

3.3.10.2 Cold Gas Efficiency

The cold gas efficiency is useful for comparing the overall system performance of hydrogen production alone and is calculated as the ratio of the thermal content of the hydrogen product to the thermal input to the system, expressed as a percentage. The efficiency can be calculated on both a HHV and a LHV basis. This calculation is represented by the following equation:

$$CGE = \frac{TP}{TI} * 100\%$$

Where:

CGE – cold gas efficiency

TP – thermal content of the hydrogen product

TI – thermal input to the system

The thermal input to the system is calculated by taking the feed rate into the system and multiplying by the heating value of the feedstock.

The thermal output is calculated by taking the mass flowrate of the hydrogen product and multiplying by the heating value of the product.

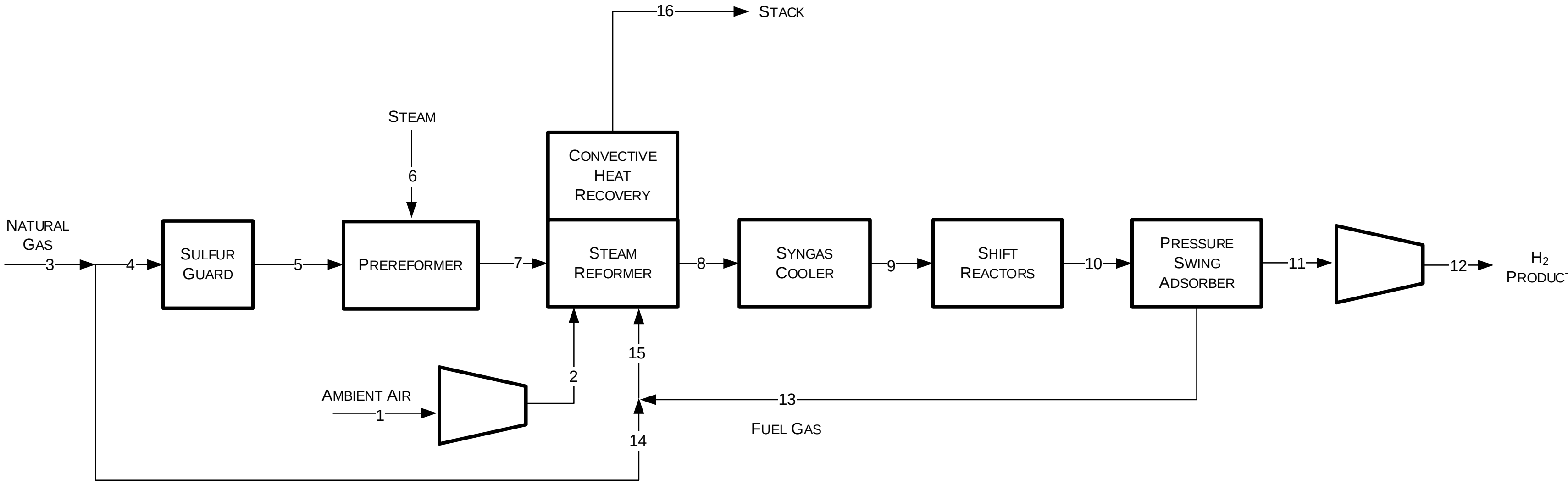
3.4 Case 1 – Steam Methane Reforming Plant Without CO_2 Capture Process Description

This section contains an evaluation of a plant design for Case 1, which is based on a natural gas steam methane reforming plant without CO_2 capture. The system descriptions follow the BFD provided in Exhibit 3-3 with the associated stream tables that show process data provided in Exhibit 3-4.

The balance of this section is organized as follows:

- Process Description – provides an overview of the technology operation as applied to Case 1. The systems that are common to all reforming cases were covered in Section 3.3 and only features that are unique to Case 1 are discussed further in this section
- Performance Results – provides the main modeling results from Case 1, including the performance summary, environmental performance, carbon balance, sulfur balance, water balance, energy balance, mass and energy balance diagrams. Note – some rounding error may be present in results reporting
- Equipment List – provides an itemized list of major equipment for Case 1
- Cost Estimates – provides a summary of capital and operating costs for Case 1

Exhibit 3-3. Case 1 block flow diagram, steam methane reforming plant without CO_2 capture



Note: Block Flow Diagram is not intended to represent a complete material balance. Only major process streams and equipment are shown.

Exhibit 3-4. Case 1 stream table, steam methane reforming plant without capture

	1	2	3	4	5	6	7	8
V-L Mole Fraction								
Ar	0.0092	0.0092	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.0000	0.9310	0.9310	0.9310	0.0000	0.2583	0.0411
CH_4S	0.0000	0.0000	0.0000[A]	0.0000	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0000	0.0000	0.0320	0.0320	0.0320	0.0000	0.0000	0.0000
C_3H_8	0.0000	0.0000	0.0070	0.0070	0.0070	0.0000	0.0000	0.0000
C_4H_{10}	0.0000	0.0000	0.0040	0.0040	0.0040	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.1432
CO_2	0.0003	0.0003	0.0100	0.0100	0.0100	0.0000	0.0174	0.0124
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0439	0.4609
H_2O	0.0099	0.0099	0.0000	0.0000	0.0000	1.0000	0.6760	0.3393
N_2	0.7732	0.7732	0.0160	0.0160	0.0160	0.0000	0.0042	0.0030
O_2	0.2074	0.2074	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	17,806	17,806	4,096	3,687	3,687	9,830	13,930	19,519
V-L Flowrate (kg/hr)	513,824	513,824	70,979	63,881	63,880	177,086	240,966	240,966
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0
Temperature (°C)	15	24	15	15	370	399	395	871
Pressure (MPa, abs)	0.10	0.11	3.10	3.10	3.00	3.10	2.90	2.83
Steam Table Enthalpy (kJ/kg)[B]	30.23	39.78	-6.77	-6.77	969.07	3,217.36	2,587.64	3,639.90
AspenPlus Enthalpy (kJ/kg)[C]	-97.58	-88.02	-4,516.00	-4,516.00	-3,540.22	-12,753.76	-10,182.33	-6,233.86
Density (kg/m^3)	1.2	1.3	24.4	24.4	9.7	10.4	9.3	3.7
V-L Molecular Weight	28.857	28.857	17.328	17.328	17.328	18.016	17.299	12.345
V-L Flowrate (lb_{mol}/hr)	39,255	39,255	9,031	8,128	8,128	21,670	30,710	43,032
V-L Flowrate (lb/hr)	1,132,789	1,132,789	156,482	140,834	140,832	390,407	531,239	531,239
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0
Temperature (°F)	59	76	59	59	698	750	743	1,600
Pressure (psia)	14.7	16.0	450.0	450.0	435.3	450.0	421.0	411.0
Steam Table Enthalpy (Btu/lb)[A]	13.0	17.1	-2.9	-2.9	416.6	1,383.2	1,112.5	1,564.9
AspenPlus Enthalpy (Btu/lb)[B]	-42.0	-37.8	-1,941.5	-1,941.5	-1,522.0	-5,483.1	-4,377.6	-2,680.1
Density (lb/ft^3)	0.076	0.080	1.525	1.525	0.605	0.652	0.578	0.229

[A]Natural gas entering battery limits contains 5.74 ppm of the mercaptan additive methanethiol (CH_4S)
[B]Steam table reference conditions are 32.02 °F & 0.089 psia
[C]AspenPlus thermodynamic reference state is the component's constituent elements in an ideal gas state at 25 °C and 1 atm

Exhibit 3-4. Case 1 stream table, steam methane reforming plant without capture (continued)

	9	10	11	12	13	14	15	16
V-L Mole Fraction								
Ar	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0072
CH_4	0.0411	0.0513	0.0000	0.0000	0.1410	0.9310	0.1940	0.0000
CH_4S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0320	0.0022	0.0000
C_3H_8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0070	0.0005	0.0000
C_4H_{10}	0.0000	0.0000	0.0000	0.0000	0.0000	0.0040	0.0003	0.0000
CO	0.1432	0.0050	0.0000	0.0000	0.0137	0.0000	0.0128	0.0000
CO_2	0.0124	0.1891	0.0000	0.0000	0.5198	0.0100	0.4855	0.1858
H_2	0.4609	0.7483	0.9998	0.9998	0.3085	0.0000	0.2878	0.0000
H_2O	0.3393	0.0025	0.0000	0.0000	0.0069	0.0000	0.0065	0.1913
N_2	0.0030	0.0038	0.0002	0.0002	0.0100	0.0160	0.0104	0.6012
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0146
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	19,519	15,654	9,959	9,959	5,695	410	6,104	23,005
V-L Flowrate (kg/hr)	240,966	171,320	20,126	20,126	151,195	7,098	158,293	672,117
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0
Temperature (°C)	204	38	38	30	38	15	35	121
Pressure (MPa, abs)	2.80	2.45	2.39	6.48	0.14	3.10	0.14	0.10
Steam Table Enthalpy (kJ/kg)[A]	1,740.51	109.39	544.72	446.66	59.24	-6.77	56.28	426.36
AspenPlus Enthalpy (kJ/kg)[B]	-8,133.24	-7,225.03	188.98	90.92	-8,204.10	-4,516.00	-8,038.73	-3,980.71
Density (kg/m^3)	8.8	10.4	1.8	5.0	1.4	24.4	1.4	0.9
V-L Molecular Weight	12.345	10.945	2.021	2.021	26.549	17.328	25.931	29.217
V-L Flowrate (lb_{mol}/hr)	43,032	34,510	21,955	21,955	12,555	903	13,458	50,717
V-L Flowrate (lb/hr)	531,239	377,697	44,369	44,369	333,328	15,648	348,976	1,481,765
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0
Temperature (°F)	400	100	100	86	100	59	95	250
Pressure (psia)	406.0	356.0	346.0	939.7	20.0	450.0	20.0	14.7
Steam Table Enthalpy (Btu/lb)[A]	748.3	47.0	234.2	192.0	25.5	-2.9	24.2	183.3
AspenPlus Enthalpy (Btu/lb)[B]	-3,496.7	-3,106.2	81.2	39.1	-3,527.1	-1,941.5	-3,456.0	-1,711.4
Density (lb/ft^3)	0.549	0.647	0.115	0.314	0.089	1.525	0.087	0.056

[A]Steam table reference conditions are 32.02 °F & 0.089 psia

[B]AspenPlus thermodynamic reference state is the component's constituent elements in an ideal gas state at 25 °C and 1 atm

3.4.1 Steam Methane Reformer

Partially reformed gas and steam effluent exiting the prereformer enter the primary steam methane reformer in Case 1 and Case 2. The mixture is reacted over a nickel-based catalyst contained inside a system of high alloy steel tubes. The following reaction takes place in the reformer:

$$\mathrm{CH_4 + H_2O \leftrightarrow CO + 3H_2;\ \Delta}H°_{\mathrm{rxn}} = 205.8\ \mathrm{kJ/mol}$$

The reforming reaction is strongly endothermic, with energy supplied by firing the reformer on the outside of the catalyst tubes with recycled syngas from the downstream hydrogen purification process plus supplemental natural gas. The metallurgy of the tubes usually limits the reaction to 760-925 °C (1,400-1,700 °F).

The reforming reaction quickly reaches equilibrium at all points in the catalyst bed. The equilibrium composition of the reformed gas is favored by the high steam-to-carbon ratio, low pressure, and high temperature. The process employs a molar steam-to-carbon ratio of 2.56 entering the reactor. The raw syngas exits the reactor at a temperature of 870 °C (1,600 °F) and pressures of 2.8 MPa (411 psia) to convert more than 77 percent of the methane in the primary reformer feed.

The flue gas path of the fired reformer is integrated with additional boiler surfaces to produce export steam in Case 1 or to provide steam for the regeneration of CO_2-removal solvents from the AGR processes in Case 2.

3.4.2 Convective Heat Recovery Section

The steam methane reforming configuration (Case 1 and Case 2) has an integrated heat recovery section as a part of the primary reformer which allows for convective heat recovery of the reformer flue gas. The hot flue gas from the primary reformer is cooled by providing heat to various sections of the plant and improving plant efficiency, thereby minimizing the need to fire supplemental NG. This section generates superheated steam for the reforming reactions, provides feedwater preheating, and provides preheating of the natural gas feed prior to the sulfur guard and prereformer. Approximately 53 percent of the superheated steam generated is used for reforming while the remaining is exported from the plant.

3.5 Case 1 – Performance Results

Overall performance for the entire plant is summarized in Exhibit 3-5. The plant produces 20,126 kilogram (kg)/hr (44,369 lb/hr) of hydrogen at an effective thermal efficiency of 75.4 percent (HHV basis). The total auxiliary load for the plant is 13 megawatts of electricity (MWe).

Exhibit 3-6 provides a detailed breakdown of the auxiliary power requirements. The hydrogen compressor accounts for about 79 percent of the total auxiliary load. The BFW pumps and cooling water system, including the circulating water pumps and cooling tower fan, accounts for approximately 6 percent of the auxiliary load. All other systems together constitute the remaining 15 percent of the auxiliary load.

Exhibit 3-5. Case 1 plant performance summary

Performance Summary	
Steam Turbine Power, MWe	0
Total Gross Power, MWe	**0**
CO_2 Capture/Removal Auxiliaries, kWe	0
CO_2 Compression, kWe	0
Balance of Plant, kWe	12,650
Total Auxiliaries, MWe	**13**
Net Power, MWe	**-13**
Hydrogen Production, kg/hr (lb/hr)	**20,126 (44,369)**
CO_2 Capture, %	**–**
HHV Effective Thermal Efficiency[A], %	75.4%
HHV Cold Gas Efficiency[B], %	76.7%
LHV Effective Thermal Efficiency[A], %	70.5%
LHV Cold Gas Efficiency[B], %	71.9%
AGR Cooling Duty, GJ/hr (MMBtu/hr)	–
Natural Gas Feed Flow, kg/hr (lb/hr)	70,979 (156,482)
HHV Thermal Input, kWt	1,031,068
LHV Thermal Input, kWt	930,641
Raw Water Withdrawal, m^3/min (gpm)	5.5 (1,457)
Raw Water Consumption, m^3/min (gpm)	5.3 (1,395)

[A]ETE = (Hydrogen Heating Value + Net Power) / Fuel Heating Value
[B]CGE = Hydrogen Heating Value / Fuel Heating Value

Exhibit 3-6. Case 1 plant power summary

Power Summary	
Steam Turbine Power, MWe	0
Total Gross Power, MWe	**0**
Auxiliary Load Summary	
Air Blower, kWe	1,410
Circulating Water Pumps, kWe	280
Cooling Tower Fans, kWe	140
CO_2 Capture/Removal Auxiliaries, kWe	0
CO_2 Compression, kWe	0
Feedwater Pumps, kWe	320
Ground Water Pumps, kWe	130
Hydrogen Compressor, kWe	10,010
Miscellaneous Balance of Plant[A], kWe	290
Transformer Losses, kWe	70
Total Auxiliaries, MWe	**13**
Net Power, MWe	**-13**

[A]Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

3.5.1 Environmental Performance

The environmental targets for emissions of Hg, NOx, sulfur dioxide (SO_2), and PM were presented in Section 2.6. A summary of the plant air emissions for Case 1 is presented in Exhibit 3-7.

Exhibit 3-7. Case 1 air emissions

	kg/GJ (lb/MMBtu)	Tonne/year (ton/year)[A]	lb/lb H_2
SO_2	0.000 (0.000)	1.4 (1.5)	0.000
NOx	0.001 (0.003)	35 (39)	0.000
Particulate	0.000 (0.000)	5 (6)	0.000
Hg	0.00E+0 (0.00E+0)	0.000 (0.000)	0.00E+0
CO	0.000 (0.000)	5 (6)	0.000
CO_2[B]	51 (118)	1,482,893 (1,634,610)	9.3
CO_2e[C]	72 (166)	2,092,674 (2,306,778)	12

[A]Calculations based on a 90 percent capacity factor
[B]Plant stack emissions
[C]Life cycle emissions

The natural gas feed contains 5.74 ppmv of the mercaptan additive methanethiol (CH_4S). A zinc oxide sulfur guard bed was used to remove 98 percent of the sulfur in the feed, resulting in a maximum sulfur content of 0.1 ppm post-sulfur guard bed.

NOx emissions are limited by the use of low-NOx burners in the design. The production of PM is a result of system inefficiencies and is not produced or emitted in any significant amount. The pipeline natural gas was assumed to contain no Hg nor hydrochloric acid (HCl), resulting in zero emissions.

Plant stack CO_2 emissions represent the uncontrolled discharge from the process. Life cycle CO_2 emissions represent the total GWP, in CO_2e emissions, of the process over the plant life cycle, as detailed in Section 2.8.

The carbon balance for the plant is shown in Exhibit 3-8. The carbon input to the plant consists of carbon in the air in addition to carbon in the natural gas. Carbon leaves the plant as a small portion of the export steam and hydrogen product as well as in the stack gas.

Exhibit 3-8. Case 1 carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Natural Gas	51,267 (113,024)	Stack Gas	51,333 (113,169)
Air (CO_2)	70 (154)	Export Steam	1 (3)
		Hydrogen Product	2 (5)
		Sulfur Guard Bed	0 (1)
		CO_2 Product	–
		CO_2 Dryer Vent	–
		CO_2 Knockout	–
Total	**51,337 (113,178)**	**Total**	**51,337 (113,178)**

Exhibit 3-9 shows the sulfur balance for the plant. Sulfur input comes solely from the mercaptan additive in the natural gas. Sulfur output includes sulfur recovered in the sulfur guard bed and small amounts of sulfur emitted in the stack gas.

Exhibit 3-9. Case 1 sulfur balance

Sulfur In		Sulfur Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Natural Gas	1 (2)	Stack Gas	0 (0)
		Sulfur Guard Bed	1 (1)
Total	**1 (2)**	**Total**	**1 (2)**

Exhibit 3-10 shows the overall water balance for the plant. Water demand represents the total amount of water required for a particular process. Some water is recovered within the process, primarily as syngas condensate, and is re-used as internal recycle.

Exhibit 3-10. Case 1 water balance

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)
CO_2 Drying	–	–	–	–	–
CO_2 Capture System Makeup	–	–	–	–	–
CO_2 Capture Recovery	–	–	–	–	–
CO_2 Compression Recovery	–	–	–	–	–
Steam Generation	**5.6 (1,485)**	**1.2 (307)**	**4.5 (1,178)**	**–**	**4.5 (1,178)**
Export Steam	2.7 (704)	–	2.7 (704)	–	2.7 (704)
SMR Steam	3.0 (781)	–	3.0 (781)	–	3.0 (781)
Syngas Condensate	–	1.2 (307)	-1.2 (-307)	–	-1.2 (-307)
Cooling Tower	**1.1 (280)**	**–**	**1.1 (280)**	**0.2 (63)**	**0.8 (217)**
Total	**6.7 (1,765)**	**1.2 (307)**	**5.5 (1,457)**	**0.2 (63)**	**5.3 (1,395)**

An overall plant energy balance is provided in tabular form in Exhibit 3-11.

Exhibit 3-11. Case 1 overall energy balance (0 °C [32 °F] reference)

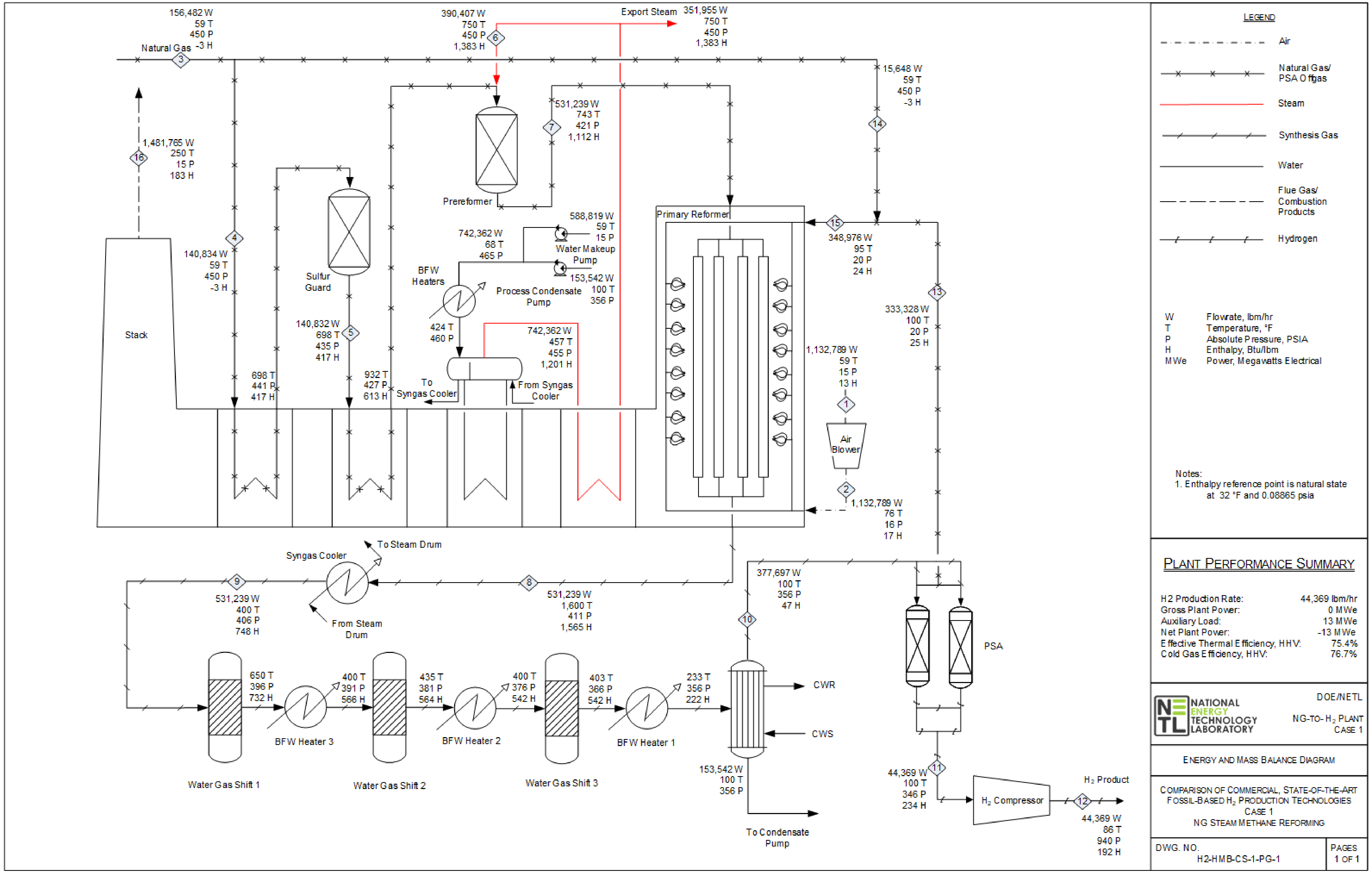
	HHV	Sensible + Latent	Power	Total
Heat In, GJ/hr (MMBtu/hr)				
Natural Gas	3,712 (3,518)	2.5 (2.4)	–	3,714 (3,521)
Air	–	16 (15)	–	16 (15)
Raw Water Makeup	–	21 (20)	–	21 (20)
Auxiliary Power	–	–	46 (43)	46 (43)
Total	**3,712 (3,518)**	**39 (37)**	**46 (43)**	**3,796 (3,598)**
Heat Out, GJ/hr (MMBtu/hr)				
Hydrogen Product	2,846 (2,697)	9.0 (8.5)	–	2,855 (2,706)
Export Steam	–	514 (487)	–	514 (487)
Stack Gas	–	287 (272)	–	287 (272)
Sulfur	–	0.0 (0.0)	–	0.0 (0.0)
Motor Losses and Design Allowances	–	–	1.4 (1.3)	1.4 (1.3)
Cooling Tower Load[A]	–	138 (131)	–	138 (131)
CO_2 Product Stream	–	–	–	–
Power	–	–	0.0 (0.0)	0.0 (0.0)
Total	**2,846 (2,697)**	**947 (898)**	**1.4 (1.3)**	**3,795 (3,597)**
Unaccounted Energy[B]	–	1.3 (1.2)	–	1.3 (1.2)

[A]Includes compression and last stage syngas cooling loads
[B]By difference

3.5.2 Energy and Mass Balance Diagrams

The energy and mass balance diagram for Case 1 is shown in Exhibit 3-12.

Exhibit 3-12. Case 1 steam methane reforming plant without CO_2 capture energy and mass balance



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3.6 Case 1 – Major Equipment List

Major equipment items for the steam methane reforming plant without CO_2 capture are shown in the following tables. In general, the design conditions include a 10 percent design allowance for flows and heat duties and a 21 percent design allowance for heads on pumps and fans.

Case 1 – Account 3: Feedwater and Miscellaneous Balance of Plant Systems

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Process Condensate Pumps	Horizontal, split case, multi-stage, centrifugal	640 lpm @ 90 m H_2O (170 gpm @ 300 ft H_2O)	2	1
2	Water Makeup Pumps	Horizontal, split case, multi-stage, centrifugal	2,470 lpm @ 380 m H_2O (650 gpm @ 1,260 ft H_2O)	2	2
3	Auxiliary Boiler	Shop fabricated, water tube	18,000 kg/hr, 2.8 MPa, 343 °C (40,000 lb/hr, 400 psig, 650 °F)	1	0
4	Service Air Compressors	Flooded Screw	13 m^3/min @ 0.7 MPa (450 scfm @ 100 psig)	2	1
5	Instrument Air Dryers	Duplex, regenerative	13 m^3/min (450 scfm)	2	1
6	Engine-Driven Fire Pump	Vertical turbine, diesel engine	3,785 lpm @ 110 m H_2O (1,000 gpm @ 350 ft H_2O)	1	1
7	Fire Service Booster Pump	Two-stage horizontal centrifugal	2,650 lpm @ 80 m H_2O (700 gpm @ 250 ft H_2O)	1	1
8	Raw Water Pumps	Stainless steel, single suction	3,000 lpm @ 20 m H_2O (800 gpm @ 60 ft H_2O)	2	1
9	Filtered Water Pumps	Stainless steel, single suction	150 lpm @ 50 m H_2O (40 gpm @ 160 ft H_2O)	2	1
10	Filtered Water Tank	Vertical, cylindrical	145,000 liter (38,000 gal)	1	0
11	Makeup Water Demineralizer	Multi-media filter, cartridge filter, RO membrane assembly and electro-deionization unit	330 lpm (90 gpm)	1	0
12	Liquid Waste Treatment System	–	10 years, 24-hour storm	1	0
13	Gas Pipeline	Underground, coated carbon steel, wrapped cathodic protection	53 m^3/min @ 3.1 MPa (1,882 acfm @ 450 psia) 39 cm (16 in) standard wall pipe	16 km (10 mile)	0
14	Gas Metering Station	–	53 m^3/min (1,882 acfm)	1	0

Case 1 – Account 4: Reformer and Accessories

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Primary Reformer	Steam Methane Reformer (SMR), side fired, fixed bed, catalytic	Syngas Production: 265,000 kg/hr @ 2.8 MPa, 871 °C (584,000 lb/hr @ 411 psia, 1600 °F)	1	0
2	Air Blower	Centrifugal	283,000 kg/hr @ 4,000 m^3/min, 3 cm WG (623,000 lb/hr @ 129,000 acfm, 1 in. WG)	2	1
3	Sulfur Guard Bed	Fixed Bed, catalytic (ZnO)	Inlet: 70,000 kg/hr @ 3.0 MPa, 370 °C (155,000 lb/hr @ 441 psia, 698 °F)	1	0
4	Prereformer	Fixed Bed, catalytic	NG In: 70,000 kg/hr @ 2.9 MPa, 500 °C (155,000 lb/hr @ 427 psia, 932 °F) Steam In: 195,000 kg/hr @ 3.1 MPa, 399 °C (429,000 lb/hr @ 450 psia, 750 °F)	1	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
5	Syngas Coolers	Shell and tube heat exchanger	Syngas Cooler: 457 GJ/hr (434 MMBtu/hr) BFW Heater 3: 93 GJ/hr (88 MMBtu/hr) BFW Heater 2: 13 GJ/hr (12 MMBtu/hr) BFW Heater 1: 179 GJ/hr (170 MMBtu/hr) AGR Precooler: 101 GJ/hr (96 MMBtu/hr)	5	0

Case 1 – Account 6: Syngas Cleanup

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Water Gas Shift Reactors	Fixed bed, catalytic	88,000 kg/hr (292,000 lb/hr) 204 °C (400 °F) 2.8 MPa (410 psia)	6	0

Case 1 – Account 7: Hydrogen Production

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Pressure Swing Adsorber	Polybed Proprietary	Syngas: 171,320 kg/hr (377,697 lb/hr) 38 °C (100 °F) 2.5 MPa (356.0 psia) Hydrogen: 20,126 kg/hr (44,369 lb/hr) 38 °C (100 °F) 2.4 MPa (346.0 psia) Off Gas: 151,195 kg/hr (333,328 lb/hr) 38 °C (100 °F) 0.1 MPa (20.0 psia)	1	0
2	Hydrogen Compressor	Integrally geared, multi-stage centrifugal	73 m^3/min @ 6.5 MPa (2,587 acfm @ 940 psia)	1	1

Case 1 – Account 9: Cooling Water System

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Circulating Water Pumps	Vertical, wet pit	27,000 lpm @ 30 m (7,000 gpm @ 100 ft)	2	1
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	11 °C (52 °F) wet bulb/16 °C (60 °F) CWT/27 °C (80 °F) HWT/ 150 GJ/hr (140 MMBtu/hr) heat duty	1	0

Case 1 – Account 11: Accessory Electric Plant

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	High Voltage Auxiliary Transformer	Oil-filled	345 kV/13.8 kV, 0 MVA, 3-ph, 60 Hz	2	0
2	Medium Voltage Transformer	Oil-filled	18 kV/4.16 kV, 12 MVA, 3-ph, 60 Hz	1	1
3	Low Voltage Transformer	Dry ventilated	4.16 kV/480 V, 2 MVA, 3-ph, 60 Hz	1	1
4	Medium Voltage Switchgear	Metal clad	4.16 kV, 3-ph, 60 Hz	1	1
5	Low Voltage Switchgear	Metal enclosed	480 V, 3-ph, 60 Hz	1	1
6	Emergency Diesel Generator	Sized for emergency shutdown	750 kW, 480 V, 3-ph, 60 Hz	1	0

Case 1 – Account 12: Instrumentation and Control

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	DCS – Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W)	Operator stations/printers and engineering stations/printers	1	0
2	DCS – Processor	Microprocessor with redundant input/output	N/A	1	0
3	DCS – Data Highway	Fiber optic	Fully redundant, 25% spare	1	0

3.7 Case 1 – Cost Estimating

The cost estimating methodology was described previously in Section 2.11. Exhibit 3-13 shows a detailed breakdown of the capital costs; Exhibit 3-14 shows the owner's costs, TOC, and TASC; Exhibit 3-15 shows the initial and annual O&M costs; and Exhibit 3-16 shows the LCOH breakdown.

The estimated TOC of the steam methane reforming plant without CO_2 capture is $713/[kg H_2/day]. Process contingency represents 0 percent of the TOC and project contingency represents 13.1 percent. The LCOH is $1.06/kg H_2.

Exhibit 3-13. Case 1 total plant cost details

	Case:	1	– Steam Methane Reforming w/o CCS					Estimate Type:		Conceptual	
	Plant Size (kg H_2/day):	483,014						Cost Base:		Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/[kg H_2/day]
3	Feedwater & Miscellaneous BOP Systems										
3.1	Feedwater System	$25	$52	$139	$109	$325	$65	$0	$58	$448	$1
3.2	Water Makeup & Pretreating	$2,272	$227	$1,287	$0	$3,786	$757	$0	$909	$5,452	$11
3.3	Other Feedwater Subsystems	$741	$243	$231	$0	$1,215	$243	$0	$219	$1,677	$3
3.4	Service Water Systems	$689	$1,316	$4,261	$0	$6,267	$1,253	$0	$1,504	$9,024	$19
3.5	Other Boiler Plant Systems	$230	$84	$209	$0	$523	$105	$0	$94	$721	$1
3.6	Natural Gas Pipeline and Start-Up System	$9,304	$400	$300	$0	$10,005	$2,001	$0	$1,801	$13,807	$29
3.7	Wastewater Treatment Equipment	$945	$0	$579	$0	$1,524	$305	$0	$366	$2,195	$5
3.8	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
3.9	Miscellaneous Plant Equipment	$14,217	$1,865	$7,225	$0	$23,306	$4,661	$0	$5,594	$33,561	$69
	Subtotal	**$28,424**	**$4,186**	**$14,232**	**$109**	**$46,951**	**$9,390**	**$0**	**$10,544**	**$66,885**	**$138**
4	Reformer & Accessories										
4.1	Primary Reformer	$25,327	$756	$8,076	$6,299	$40,458	$8,092	$0	$12,137	$60,687	$126
4.2	Air Blower	$634	$507	$821	$640	$2,602	$520	$0	$468	$3,590	$7
4.3	Sulfur Guard Bed	$192	$236	$362	$283	$1,073	$215	$0	$193	$1,481	$3
4.4	Prereformer	$1,273	$302	$462	$361	$2,398	$480	$0	$719	$3,597	$7
4.5	Syngas Coolers	$9,113	$1,318	$1,413	$1,102	$12,946	$2,589	$0	$3,884	$19,418	$40
4.6	Air Separation Unit/Oxidant Compression	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.7	Steam Drum(s)	w/4.1	w/4.1	w/4.1	w/4.1	$0	$0	$0	$0	$0	$0
	Subtotal	**$36,539**	**$3,119**	**$11,134**	**$8,684**	**$59,476**	**$11,895**	**$0**	**$17,402**	**$88,773**	**$184**
5	Flue Gas Cleanup										
5.1	Cansolv Carbon Dioxide (CO_2) Removal System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.4	Carbon Dioxide (CO_2) Compression & Drying	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.5	Carbon Dioxide (CO_2) Compressor Aftercooler	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.12	Gas Cleanup Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	Subtotal	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**

Case:		1	– Steam Methane Reforming w/o CCS				Estimate Type:			Conceptual	
Plant Size (kg H_2/day):		483,014					Cost Base:			Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	\$/1,000	\$/[kg H_2/day]
6	Syngas Cleanup										
6.1	Recycle Compressor	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0
6.2	Methyl Diethanolamine (MDEA) – Low Temperature Acid Gas Removal	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0
6.7	Water Gas Shift (WGS) Reactors	\$1,764	\$1,882	\$2,769	\$2,160	\$8,576	\$1,715	\$0	\$2,058	\$12,349	\$26
6.12	Gas Cleanup Foundations	w/ 6.7	w/ 6.7	w/ 6.7	w/ 6.7	\$0	\$0	\$0	\$0	\$0	\$0
	Subtotal	**\$1,764**	**\$1,882**	**\$2,769**	**\$2,160**	**\$8,576**	**\$1,715**	**\$0**	**\$2,058**	**\$12,349**	**\$26**
7	Hydrogen Production										
7.1	Pressure Swing Adsorber	\$16,112	\$23	\$730	\$569	\$17,434	\$3,487	\$0	\$4,184	\$25,105	\$52
7.2	Hydrogen Compressor	\$6,844	\$187	\$501	\$391	\$7,922	\$1,584	\$0	\$1,901	\$11,408	\$24
	Subtotal	**\$22,956**	**\$211**	**\$1,230**	**\$960**	**\$25,356**	**\$5,071**	**\$0**	**\$6,085**	**\$36,513**	**\$76**
8	Fired Heater, Ductwork, & Stack										
8.1	Fired Heater	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0
8.2	Fired Heater Accessories	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0
8.3	Ductwork	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0
8.4	Stack	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0
8.5	Fired Heater, Ductwork & Stack Foundations	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0
	Subtotal	**\$0**	**\$0**	**\$0**	**\$0**	**\$0**	**\$0**	**\$0**	**\$0**	**\$0**	**\$0**
9	Cooling Water System										
9.1	Cooling Towers	\$1,485	\$0	\$453	\$0	\$1,938	\$388	\$0	\$349	\$2,674	\$6
9.2	Circulating Water Pumps	\$200	\$0	\$12	\$0	\$212	\$42	\$0	\$38	\$292	\$1
9.3	Circulating Water System Auxiliaries	\$2,605	\$0	\$344	\$0	\$2,949	\$590	\$0	\$531	\$4,069	\$8
9.4	Circulating Water Piping	\$0	\$569	\$515	\$0	\$1,084	\$217	\$0	\$195	\$1,496	\$3
9.5	Make-up Water System	\$230	\$0	\$295	\$0	\$525	\$105	\$0	\$95	\$725	\$2
9.6	Component Cooling Water System	\$84	\$0	\$64	\$0	\$148	\$30	\$0	\$27	\$205	\$0
9.7	Circulating Water System Foundations	\$0	\$132	\$219	\$0	\$350	\$70	\$0	\$84	\$504	\$1
	Subtotal	**\$4,604**	**\$700**	**\$1,902**	**\$0**	**\$7,206**	**\$1,441**	**\$0**	**\$1,318**	**\$9,965**	**\$21**
11	Accessory Electric Plant										
11.2	Station Service Equipment	\$0	\$840	\$63	\$49	\$951	\$190	\$0	\$171	\$1,313	\$3
11.3	Switchgear & Motor Control	\$0	\$436	\$133	\$104	\$673	\$135	\$0	\$121	\$929	\$2
11.4	Conduit & Cable Tray	\$0	\$189	\$237	\$185	\$612	\$122	\$0	\$110	\$845	\$2
11.5	Wire & Cable	\$0	\$249	\$180	\$140	\$569	\$114	\$0	\$102	\$785	\$2
11.6	Protective Equipment	\$0	\$235	\$443	\$346	\$1,024	\$205	\$0	\$184	\$1,413	\$3

Case:		1	– Steam Methane Reforming w/o CCS						Estimate Type:		Conceptual	
Plant Size (kg H_2/day):		483,014							Cost Base:		Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/[kg H_2/day]
11.7	Standby Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
11.8	Main Power Transformers	$0	$226	$71	$56	$353	$71	$0	$63	$487	$1
11.9	Electrical Foundations	$0	$5	$14	$11	$30	$6	$0	$7	$43	$0
	Subtotal	**$0**	**$2,179**	**$1,142**	**$890**	**$4,212**	**$842**	**$0**	**$760**	**$5,814**	**$12**
12						Instrumentation & Control					
12.4	Other Major Component Control Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
12.5	Signal Processing Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
12.6	Control Boards, Panels & Racks	$0	$159	$16	$13	$188	$38	$0	$34	$259	$1
12.7	Distributed Control System Equipment	$6,704	$0	$204	$0	$6,908	$1,382	$0	$1,243	$9,533	$20
12.8	Instrument Wiring & Tubing	$0	$137	$252	$197	$585	$117	$0	$105	$808	$2
12.9	Other Instrumentation & Controls Equipment	$0	$1,001	$240	$188	$1,429	$286	$0	$257	$1,973	$4
	Subtotal	**$6,704**	**$1,297**	**$713**	**$397**	**$9,111**	**$1,822**	**$0**	**$1,640**	**$12,573**	**$26**
13						Improvements to Site					
13.1	Site Preparation	$0	$502	$10,650	$0	$11,152	$2,230	$0	$2,676	$16,058	$33
13.2	Site Improvements	$0	$199	$794	$619	$1,612	$322	$0	$387	$2,321	$5
13.3	Site Facilities	$1,548	$0	$1,624	$0	$3,173	$635	$0	$761	$4,569	$9
	Subtotal	**$1,548**	**$701**	**$13,068**	**$619**	**$15,937**	**$3,187**	**$0**	**$3,825**	**$22,949**	**$48**
14						Buildings & Structures					
14.2	Interconnecting Pipe	$0	$425	$990	$772	$2,188	$438	$0	$394	$3,019	$6
14.4	Administration Building	$0	$334	$227	$0	$561	$112	$0	$101	$774	$2
14.5	Circulation Water Pumphouse	$0	$8	$4	$0	$12	$2	$0	$2	$17	$0
14.6	Water Treatment Buildings	$0	$198	$181	$0	$379	$76	$0	$68	$523	$1
14.7	Machine Shop	$0	$499	$319	$0	$818	$164	$0	$147	$1,129	$2
14.8	Warehouse	$0	$390	$236	$0	$626	$125	$0	$113	$863	$2
14.9	Structural Steel Pipe Racks	$0	$1,041	$598	$466	$2,105	$421	$0	$379	$2,905	$6
14.10	Waste Treating Building & Structures	$0	$658	$1,173	$0	$1,831	$366	$0	$330	$2,527	$5
	Subtotal	**$0**	**$3,554**	**$3,727**	**$1,239**	**$8,520**	**$1,704**	**$0**	**$1,534**	**$11,758**	**$24**
	Total	**$102,539**	**$17,831**	**$49,917**	**$15,057**	**$185,344**	**$37,069**	**$0**	**$45,166**	**$267,578**	**$554**

Exhibit 3-14. Case 1 owner's costs

Description	$/1,000	$/[kg H_2/day]
Pre-Production Costs		
6 Months All Labor	$2,641	$5
1 Month Maintenance Materials	$282	$1
1 Month Non-Fuel Consumables	$993	$2
1 Month Waste Disposal	$2	$0
25% of 1 Months Fuel Cost at 100% CF	$2,838	$6
2% of TPC	$5,352	$11
Total	**$12,107**	**$25**
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	$540	$1
0.5% of TPC (spare parts)	$1,338	$3
Total	**$1,878**	**$4**
Other Costs		
Initial Cost for Catalyst and Chemicals	$15,295	$32
Land	$300	$1
Other Owner's Costs	$40,137	$83
Financing Costs	$7,225	$15
Total Overnight Costs (TOC)	**$344,520**	**$713**
TASC Multiplier (IOU, 35 year)	1.070	
Total As-Spent Cost (TASC)	**$368,568**	**$763**

Exhibit 3-15. Case 1 initial and annual operating and maintenance costs

Case:	1	– Steam Methane Reforming w/o CCS			Cost Base:	Dec 2018
Plant Size (kg H_2/day):	483,014				Capacity Factor (%):	90
Operating & Maintenance Labor						
Operating Labor				Operating Labor Requirements per Shift		
Operating Labor Rate (base):		38.50	$/hour	Skilled Operator:		1.0
Operating Labor Burden:		30.00	% of base	Operator:		2.0
Labor O-H Charge Rate:		25.00	% of labor	Foreman:		1.0
				Lab Techs, etc.:		1.0
				Total:		**5.0**
Fixed Operating Costs						
					Annual Cost	
					($)	($/[kg H_2/day])
Annual Operating Labor:					$2,192,190	$4.539
Maintenance Labor:					$2,033,595	$4.210
Administrative & Support Labor:					$1,056,446	$2.187
Property Taxes and Insurance:					$5,351,567	$11.080
Total:					**$10,633,798**	**$22.015**
Variable Operating Costs						
					($)	($/kg H_2)
Maintenance Material:					$3,050,393	$0.02
Consumables						
	Initial Fill	Per Day	Per Unit	Initial Fill		
Water (gal/1000):	0	1,049	$1.90	$0	$654,947	$0.00413
Makeup and Wastewater Treatment Chemicals (ton):	0	3.1	$550	$0	$564,762	$0.00356
Zinc Oxide Sulfur Guard Catalyst (ft^3):	705	0.5	$600	$423,000	$95,175	$0.00060
Prereformer Catalyst (ft^3):	818	0.6	$1,250	$1,022,500	$230,063	$0.00145
Primary Reformer Catalyst (ft^3):	2770	1.5	$525	$1,454,250	$261,765	$0.00165
Water Gas Shift Catalyst (ft^3):	13949	9.6	$480	$6,695,503	$1,506,488	$0.00949
Methyl Diethanolamine Solution (gal):	0	0	$2.80	$0	$0	$0.00000
Post-Combustion CO_2 Capture System[A]:			Proprietary		$0	$0.00000
PSA Unit Adsorbent (ft^3):	38000	5.2	$150	$5,700,000	$256,500	$0.00162
Triethylene Glycol (gal):	w/equip.	0	$6.80	$0	$0	$0.00000
Electricity (MWh):	0	304	$71.70	$0	$7,150,827	$0.04507
Subtotal:				**$15,295,253**	**$10,720,528**	**$0.06756**
Waste Disposal						
Zinc Oxide Sulfur Guard Catalyst (ft^3):	0	0.5	$40.00	$0	$6,345	$0.00004
Prereformer Catalyst (ft^3):	0	0.6	$0.00	$0	$0	$0.00000
Primary Reformer Catalyst (ft^3):	0	1.5	$0.00	$0	$0	$0.00000
Water Gas Shift Catalyst (ft^3):	0	9.6	$2.50	$0	$7,846	$0.00005
Methyl Diethanolamine Solution (gal):	0	0	$0.35	$0	$0	$0.00000
PSA Unit Adsorbent (ft^3):	0	5.2	$1.50	$0	$2,565	$0.00002
Triethylene Glycol (gal):	0	0	$0.35	$0	$0	$0.00000
Amine Purification Unit Waste (ton):	0	0	$38.0	$0	$0	$0.00000
Thermal Reclaimer Unit Waste (ton):	0	0	$38.0	$0	$0	$0.00000
Subtotal:				**$0**	**$16,756**	**$0.00011**

Case:	1	– Steam Methane Reforming w/o CCS			Cost Base:	Dec 2018
Plant Size (kg H_2/day):	483,014				Capacity Factor (%):	90
			By-Products			
					(\$)	(\$/kg H_2)
Steam (ton):	0	4,223	\$0.00	\$0	\$0	\$0.00000
Electricity (MWh):	0	0	\$71.70	\$0	\$0	\$0.00000
Subtotal:				**\$0**	**\$0**	**\$0.00000**
Variable Operating Costs Total:				**\$15,295,253**	**\$13,787,677**	**\$0.08690**
			Fuel Cost			
Natural Gas (MMBtu):	0	84,436	\$4.42	\$0	\$122,603,526	\$0.77269
Total:				**\$0**	**\$122,603,526**	**\$0.77269**

[A]Post-combustion CO_2 capture system chemicals includes ion exchange resin, sodium hydroxide (NaOH), and Cansolv solvent

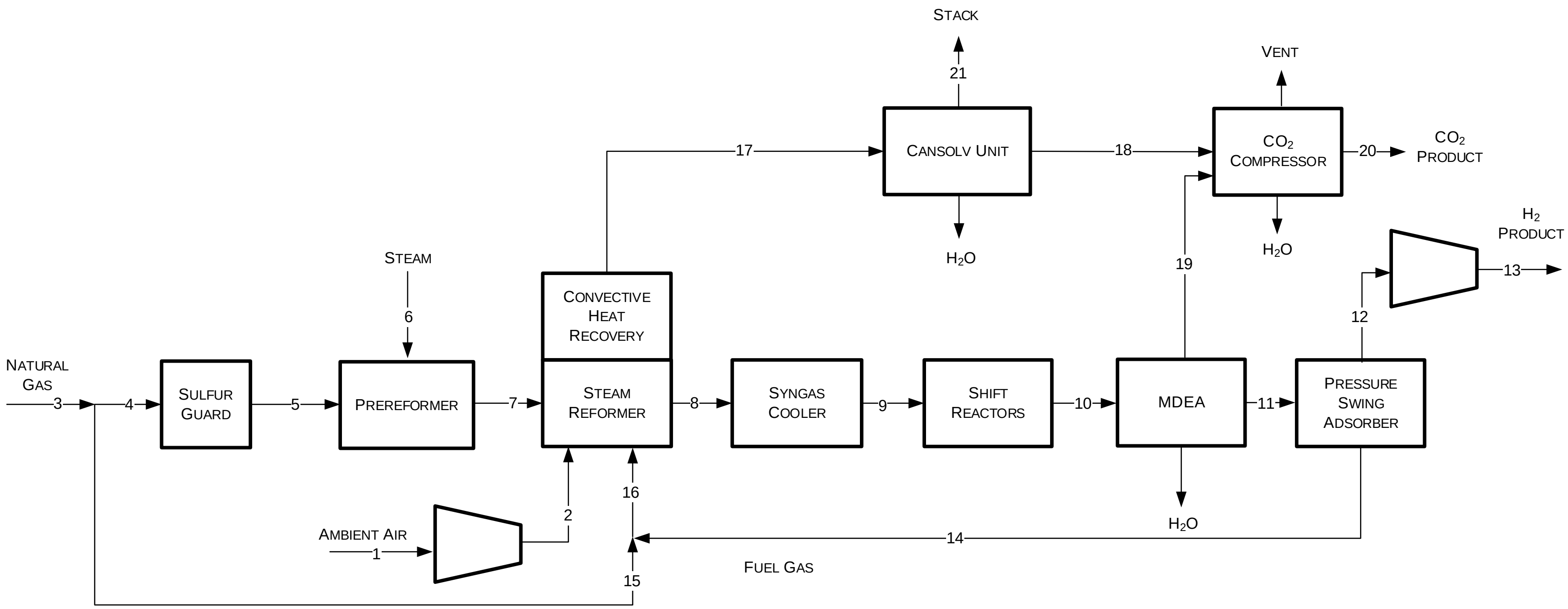
Exhibit 3-16. Case 1 LCOH breakdown

Component	Value, \$/kg H_2	Percentage
Capital	0.14	13%
Fixed	0.07	6%
Variable	0.09	8%
Fuel	0.77	73%
Total (Excluding T&S)	**1.06**	**N/A**
CO_2 T&S	0.00	0%
Total (Including T&S)	**1.06**	**N/A**

3.8 Case 2 – Steam Methane Reforming Plant with CO_2 Capture Process Description

This section contains an evaluation of a plant design for Case 2, which is based on a natural gas steam methane reforming plant with CO_2 capture. The plant configuration is nearly identical to that of Case 1, with the exception that this case is configured to produce hydrogen with CO_2 capture. The hydrogen production rate of Case 2 is equal to the Case 1 capacity. The system descriptions follow the BFD provided in Exhibit 3-17 with the associated stream tables that show process data provided in Exhibit 3-18. Rather than repeating the entire process description, only differences from Case 1 are reported here.

Exhibit 3-17. Case 2 block flow diagram, steam methane reforming plant with CO_2 capture



Note: Block Flow Diagram is not intended to represent a complete material balance. Only major process streams and equipment are shown.

Exhibit 3-18. Case 2 stream table, steam methane reforming plant with capture

	1	2	3	4	5	6	7	8	9	10
V-L Mole Fraction										
Ar	0.0092	0.0092	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.0000	0.9310	0.9310	0.9310	0.0000	0.2583	0.0411	0.0411	0.0411
CH_4S	0.0000	0.0000	0.0000[A]	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0000	0.0000	0.0320	0.0320	0.0320	0.0000	0.0000	0.0000	0.0000	0.0000
C_3H_8	0.0000	0.0000	0.0070	0.0070	0.0070	0.0000	0.0000	0.0000	0.0000	0.0000
C_4H_{10}	0.0000	0.0000	0.0040	0.0040	0.0040	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.1432	0.1432	0.0040
CO_2	0.0003	0.0003	0.0100	0.0100	0.0100	0.0000	0.0174	0.0124	0.0124	0.1517
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0440	0.4609	0.4609	0.6001
H_2O	0.0099	0.0099	0.0000	0.0000	0.0000	1.0000	0.6760	0.3393	0.3393	0.2000
N_2	0.7732	0.7732	0.0160	0.0160	0.0160	0.0000	0.0042	0.0030	0.0030	0.0030
O_2	0.2074	0.2074	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	20,603	20,603	4,356	3,687	3,687	9,830	13,930	19,519	19,519	19,519
V-L Flowrate (kg/hr)	594,529	594,529	75,472	63,883	63,882	177,085	240,967	240,967	240,967	240,967
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	15	24	15	15	370	399	395	871	204	38
Pressure (MPa, abs)	0.10	0.11	3.10	3.10	3.00	3.10	2.90	2.83	2.80	2.45
Steam Table Enthalpy (kJ/kg)[B]	30.23	39.78	-6.77	-6.77	969.07	3,216.15	2,587.78	3,639.94	1,740.54	94.92
AspenPlus Enthalpy (kJ/kg)[C]	-97.58	-88.02	-4,516.00	-4,516.00	-3,540.22	-12,753.89	-10,182.19	-6,233.87	-8,133.27	-9,735.53
Density (kg/m^3)	1.2	1.3	24.4	24.4	9.7	10.4	9.3	3.7	8.8	14.5
V-L Molecular Weight	28.857	28.857	17.328	17.328	17.328	18.015	17.298	12.345	12.345	12.345
V-L Flowrate (lb_{mol}/hr)	45,421	45,421	9,602	8,128	8,128	21,671	30,711	43,033	43,033	43,033
V-L Flowrate (lb/hr)	1,310,711	1,310,711	166,387	140,838	140,836	390,405	531,241	531,241	531,241	531,241
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	59	76	59	59	698	750	743	1,600	400	100
Pressure (psia)	14.7	16.0	450.0	450.0	435.3	450.0	421.0	411.0	406.0	356.0
Steam Table Enthalpy (Btu/lb)[A]	13.0	17.1	-2.9	-2.9	416.6	1,382.7	1,112.5	1,564.9	748.3	40.8
AspenPlus Enthalpy (Btu/lb)[B]	-42.0	-37.8	-1,941.5	-1,941.5	-1,522.0	-5,483.2	-4,377.6	-2,680.1	-3,496.7	-4,185.5
Density (lb/ft^3)	0.076	0.080	1.525	1.525	0.605	0.652	0.578	0.229	0.549	0.906

[A]Natural gas entering battery limits contains 5.74 ppm of the mercaptan additive methanethiol (CH_4S)
[B]Steam table reference conditions are 32.02 °F & 0.089 psia
[C]AspenPlus thermodynamic reference state is the component's constituent elements in an ideal gas state at 25 °C and 1 atm

Exhibit 3-18. Case 2 stream table, steam methane reforming plant with capture (continued)

	11	12	13	14	15	16	17	18	19	20	21
V-L Mole Fraction											
Ar	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0082	0.0000	0.0000	0.0000	0.0110
CH_4	0.0627	0.0000	0.0000	0.2824	0.9310	0.4059	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0000	0.0000	0.0000	0.0000	0.0320	0.0061	0.0000	0.0000	0.0000	0.0000	0.0000
C_3H_8	0.0000	0.0000	0.0000	0.0000	0.0070	0.0013	0.0000	0.0000	0.0000	0.0000	0.0000
C_4H_{10}	0.0000	0.0000	0.0000	0.0000	0.0040	0.0008	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0061	0.0000	0.0000	0.0274	0.0000	0.0222	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.0116	0.0000	0.0000	0.0521	0.0100	0.0441	0.0746	0.9835	0.9728	0.9995	0.0100
H_2	0.9150	0.9998	0.9998	0.6180	0.0000	0.5003	0.0000	0.0000	0.0002	0.0000	0.0000
H_2O	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2114	0.0165	0.0270	0.0005	0.0358
N_2	0.0046	0.0002	0.0002	0.0201	0.0160	0.0193	0.6890	0.0000	0.0000	0.0000	0.9209
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0167	0.0000	0.0000	0.0000	0.0224
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	12,802	9,959	9,959	2,843	669	3,512	23,216	1,585	2,891	4,373	17,371
V-L Flowrate (kg/hr)	46,850	20,125	20,125	26,725	11,588	38,313	632,842	69,097	125,180	192,413	486,994
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	38	38	30	38	15	29	121	30	49	30	30
Pressure (MPa, abs)	2.35	2.28	6.48	0.14	3.10	0.14	0.10	0.20	0.14	15.27	0.10
Steam Table Enthalpy (kJ/kg)[A]	302.69	544.39	446.67	124.41	-6.77	84.73	485.92	41.78	71.11	-231.09	88.99
AspenPlus Enthalpy (kJ/kg)[B]	-2,601.43	188.79	91.07	-4,698.89	-4,516.00	-4,643.57	-2,844.39	-8,969.58	-8,972.06	-9,194.65	-443.21
Density (kg/m^3)	3.3	1.8	5.0	0.5	24.4	0.6	0.8	3.5	2.2	630.1	1.1
V-L Molecular Weight	3.660	2.021	2.021	9.400	17.328	10.909	27.258	43.581	43.301	43.997	28.035
V-L Flowrate (lb_{mol}/hr)	28,223	21,955	21,955	6,268	1,474	7,743	51,184	3,495	6,373	9,642	38,296
V-L Flowrate (lb/hr)	103,287	44,369	44,369	58,918	25,548	84,466	1,395,178	152,333	275,974	424,198	1,073,638
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	100	100	86	100	59	84	250	86	120	86	87
Pressure (psia)	341.0	331.0	939.7	20.0	450.0	20.0	14.7	28.9	20.0	2,214.7	14.8
Steam Table Enthalpy (Btu/lb)[A]	130.1	234.0	192.0	53.5	-2.9	36.4	208.9	18.0	30.6	-99.4	38.3
AspenPlus Enthalpy (Btu/lb)[B]	-1,118.4	81.2	39.2	-2,020.2	-1,941.5	-1,996.4	-1,222.9	-3,856.2	-3,857.3	-3,953.0	-190.5
Density (lb/ft^3)	0.206	0.110	0.314	0.031	1.525	0.037	0.053	0.217	0.140	39.338	0.071

[A]Steam table reference conditions are 32.02 °F & 0.089 psia

[B]AspenPlus thermodynamic reference state is the component's constituent elements in an ideal gas state at 25 °C and 1 atm

3.8.1 Steam Methane Reformer

No differences from Case 1.

3.8.2 Convective Heat Recovery Section

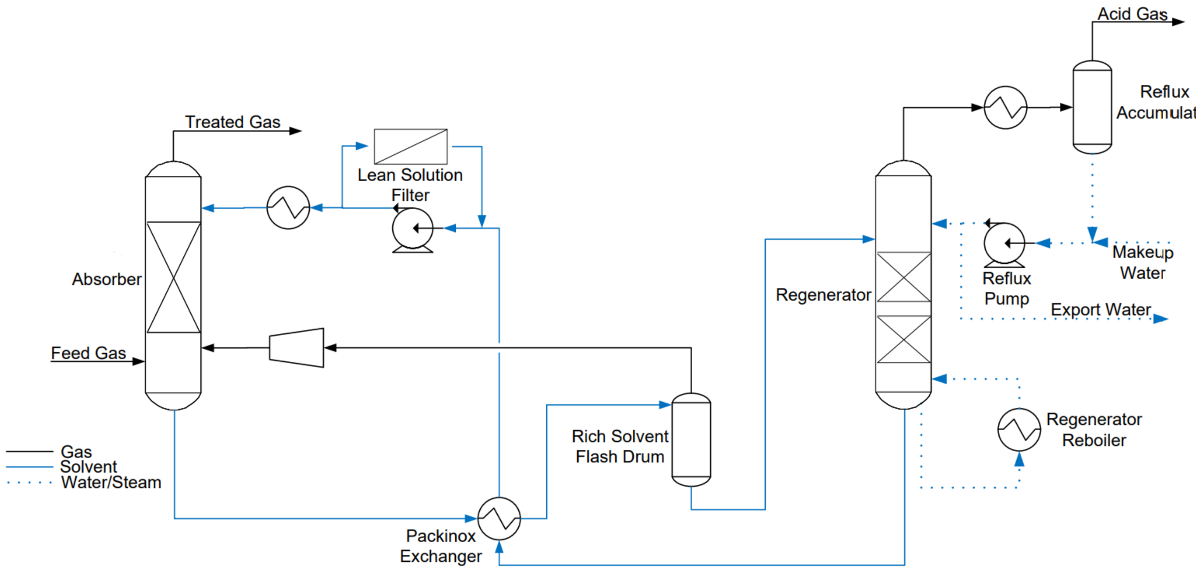
The steam methane reforming configuration (Case 1 and Case 2) has an integrated heat recovery section as a part of the primary reformer which allows for convective heat recovery from the reformer flue gas. The hot flue gas from the primary reformer is cooled by providing heat to various sections of the plant, improving plant efficiency. This section generates steam for the reforming reactions, provides feedwater preheating, and provides preheating of the natural gas feed prior to the sulfur guard and prereformer. Unlike Case 1, all of the steam generated is utilized within the plant due to the additional process steam demands of the AGR units.

3.8.3 Acid Gas Removal

3.8.3.1 MDEA

Refrigerated methyl diethanolamine (MDEA) was selected as the pre-combustion CO_2 capture system for the reforming cases and is used in both the steam methane and autothermal reforming designs with capture, Case 2 and Case 3 respectively. The CO_2 is removed from the syngas stream prior to the PSA unit in both configurations by chemical absorption with a highly selective, hybrid amine. Cool, particulate-free syngas from the shift reactors enters the absorber unit at approximately 2.5 MPa (356 psia) and 38 °C (100 °F). The gas is passed through amine tower where it is contacted counter-currently with a circulating stream of lean aqueous amine solution. The absorber column is operated at 44 °C (112 °F) by refrigerating the lean MDEA solvent. CO_2 in the feed is removed from the gas stream by the circulating lean amine. The rich amine from the absorber is then sent to a stripper column where the amine is regenerated with a steam reboiler to remove the CO_2 by fractionation. Because of the steam load required to regenerate CO_2, there is no steam export from these plant configurations. Regenerated lean amine is then cooled and sent back to the amine tower. The regenerated CO_2 stream is recovered at 20 psia and 120 °F and is sent to the CO_2 compressor for shipment off-site. A diagram of the MDEA CO_2 capture process for reforming plants is shown in Exhibit 3-19.

Exhibit 3-19. MDEA CO_2 capture process typical flow diagram for reforming plants



If the CO_2 were only captured from the shifted syngas stream, the overall CO_2 recovery would be about 62 percent for the steam methane reforming case. To increase the overall carbon recovery, a second CO_2 removal process is utilized in the reformer heater stack to remove CO_2 resulting from fuel gas combustion.

3.8.3.2 Cansolv

The CO_2 absorption/stripping/solvent reclamation process designed for post-combustion capture in the steam methane reforming with capture case (Case 2) is based on the Cansolv system. The following are characteristics of the Cansolv system used for post-combustion capture:

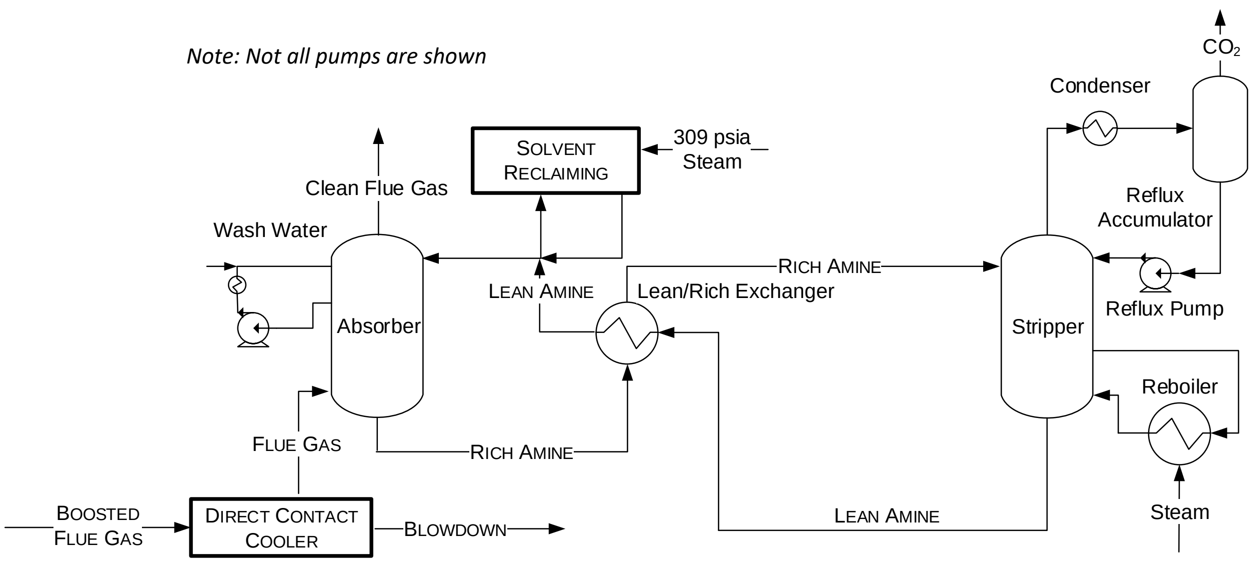
- The Cansolv solvent used for post-combustion reforming configurations provides increased reactivity in the low CO_2 content reforming flue gas environment. This is feasible given the lack of solvent contaminants (i.e., sulfur) present in the reforming plant flue gas stream.
- No SO_2 polishing step is required in reforming cases, as the pipeline natural gas sulfur content produces a flue gas with an SO_2 content below that which is required by the Cansolv process.
- No absorber inter-stage solvent cooling is employed in the reforming plant case.
- No lean solvent flash or vapor recovery, compression, and reinjection is employed in the reforming plant case.
- Solvent reclaiming considers an additional purification step, beyond thermal reclaiming, for the reforming plant case. In addition to the thermal reclaimer, an ion exchange reclaimer is also applied in the amine purification section. The acids formed by the oxidative degradation of the amine, as well as through reactions with nitrogen dioxide

(NO_2) and SO_2 heat stable salt (HSS), neutralize a portion of the amine making it inactive to further CO_2 absorption. Therefore, excess HSS are removed via an ion exchange (resin bed contained within a column) before continuing to the thermal reclaimer.

- For the steady-state case described here, the low-pressure steam requirement for the reboiler only is calculated as approximately 2.8 megajoule (MJ)/kg (1,217 Btu/lb) CO_2 for the Cansolv process.

A diagram of the Cansolv CO_2 capture process for reforming plants is provided in Exhibit 3-20.

Exhibit 3-20. Cansolv CO_2 capture process typical flow diagram for reforming plants



The Cansolv system in the reforming plant case discharges CO_2 at a temperature of 30 °C (86 °F) and a pressure of 0.2 MPa (28.9 psia).

3.8.4 CO_2 Compression and Drying System

The CO_2 compressor specifications, stage pressure ratios and outlet stage pressures are identical for each of the CO_2 capture cases with the major differentiator being the inlet CO_2 flowrates.

The compression system was modeled based on vendor supplied data and using elements of the compressor design presented in the Carbon Capture Simulation Initiative's paper "Centrifugal Compressor Simulation User Manual." [58] The design was assumed to be an eight-stage front-loaded integrally geared centrifugal compressor. The stage discharge pressures are presented in Exhibit 3-21.

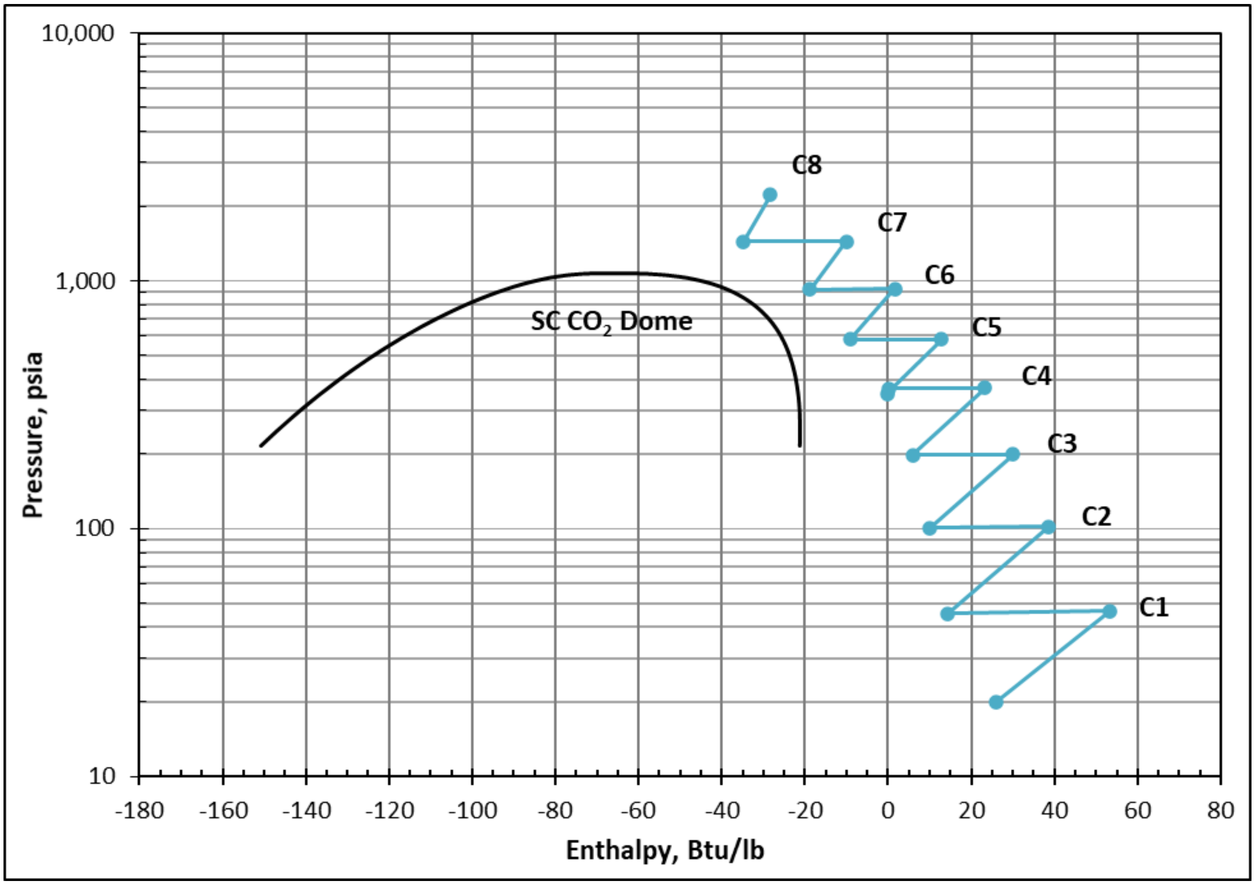
Exhibit 3-21. CO_2 compressor interstage pressures

Stage	Outlet Pressure, MPa (psia)	Stage Pressure Ratio
1	0.32 (47)	2.33
2	0.70 (102)	2.24
3	1.38 (199)	1.99
4	2.54 (369)	1.86
5	4.00 (581)	1.66
6	6.38 (925)	1.60
7	9.91 (1,437)	1.56
8	15.28 (2,217)	1.54

The CO_2 streams produced by the AGR processes contain approximately 99 percent CO_2 and are mixed prior to the compression train. The mixed CO_2 stream enters the first stage of the CO_2 compressor at 0.14 MPa (20 psia) and is compressed to 2.54 MPa (369 psia) in the first four compression stages with intercooling. Next, the CO_2 stream is dehydrated using a triethylene glycol (TEG) dryer. The dried CO_2 stream is then further compressed in the final four stages, with intercooling, to the target product pressure of 15.28 MPa (2,217 psia).

Intercooling is included for each stage with the first three stages including water knockout. The first five intercoolers cool the CO_2 to 29 °C (85 °F), the sixth intercooler cools the CO_2 to 40 °C (104 °F), and the final intercooler cools the CO_2 to 55 °C (131 °F). The increased temperature is utilized in the final two stages of intercooling to provide a suitable buffer between the compressor operating profile and SC CO_2 dome which is shown in Exhibit 3-22. A CO_2 product aftercooler is also included to cool the CO_2 30 °C (86 °F). CO_2 transportation and storage costs assume that the CO_2 enters the transport pipeline as a dense phase liquid; thus, a pipeline inlet temperature of 30 °C (86 °F) is considered.

Exhibit 3-22. Case 2 CO_2 vapor dome



A TEG dehydration unit is included between stages 4 and 5, operating at 2.53 MPa (367 psia), to reduce the moisture concentration of the CO_2 stream to 500 ppmv, consistent with QGESS design point specifications for CO_2 for pipeline export. [59] The dryer was designed based on a paper published by the Norwegian University of Science and Technology. [60] In an absorption process, such as in a TEG dehydration unit, the gas containing water flows up through a column while the TEG flows downward. The solvent preferentially binds the water by physical absorption. The dried gas exits at the top of the column, while the solvent rich in water exits at the bottom. After depressurization to around atmospheric pressure, the solvent is regenerated by heating it and passing it through a regeneration column where the water is boiled off. A TEG unit is capable of reducing water concentrations to meet the QGESS design point of 500 ppmv. [59]

3.9 Case 2 – Performance Results

Overall performance for the entire plant is summarized in Exhibit 3-23. The plant produces 20,125 kg/hr (44,369 lb/hr) of hydrogen at an effective thermal efficiency of 68.4 percent (HHV basis). The total auxiliary load for the plant is 41 MWe.

Exhibit 3-24 provides a detailed breakdown of the auxiliary power requirements. The hydrogen compressor accounts for about 26 percent of the total auxiliary load. About 57 percent of the auxiliary load is from the CO_2 capture, removal, and compression systems. The BFW pumps and cooling water system, including the circulating water pumps and cooling tower fan, accounts for approximately 10 percent of the auxiliary load. All other systems together constitute the remaining 7 percent of the auxiliary load.

Exhibit 3-23. Case 2 plant performance summary

Performance Summary	
Steam Turbine Power, MWe	0
Total Gross Power, MWe	**0**
CO_2 Capture/Removal Auxiliaries, kWe	6,710
CO_2 Compression, kWe	16,460
Balance of Plant, kWe	17,500
Total Auxiliaries, MWe	**41**
Net Power, MWe	**-41**
Hydrogen Production, kg/hr (lb/hr)	**20,125 (44,369)**
CO_2 Capture, %	**96.2%**
HHV Effective Thermal Efficiency[A], %	68.4%
HHV Cold Gas Efficiency[B], %	72.1%
LHV Effective Thermal Efficiency[A], %	63.5%
LHV Cold Gas Efficiency[B], %	67.6%
AGR Cooling Duty, GJ/hr (MMBtu/hr)	947 (898)
Natural Gas Feed Flow, kg/hr (lb/hr)	75,472 (166,387)
HHV Thermal Input, kWt	1,096,328
LHV Thermal Input, kWt	989,544
Raw Water Withdrawal, m^3/min (gpm)	10.3 (2,727)
Raw Water Consumption, m^3/min (gpm)	8.1 (2,140)

[A]ETE = (Hydrogen Heating Value + Net Power) / Fuel Heating Value
[B]CGE = Hydrogen Heating Value / Fuel Heating Value

Exhibit 3-24. Case 2 plant power summary

Power Summary	
Steam Turbine Power, MWe	0
Total Gross Power, MWe	**0**
Auxiliary Load Summary	
Air Blower, kWe	1,640
Circulating Water Pumps, kWe	2,500
Cooling Tower Fans, kWe	1,290
CO_2 Capture/Removal Auxiliaries, kWe	6,710
CO_2 Compression, kWe	16,460
Feedwater Pumps, kWe	140
Ground Water Pumps, kWe	250
Hydrogen Compressor, kWe	10,430
Miscellaneous Balance of Plant[A], kWe	1,000
Transformer Losses, kWe	250
Total Auxiliaries, MWe	**41**
Net Power, MWe	**-41**

[A]Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

3.9.1 Environmental Performance

The environmental targets for emissions of Hg, NOx, SO_2, and PM were presented in Section 2.7. A summary of the plant air emissions for Case 2 is presented in Exhibit 3-25.

Exhibit 3-25. Case 2 air emissions

	kg/GJ (lb/MMBtu)	Tonne/year (ton/year)[A]	lb/lb H_2
SO_2	0.000 (0.000)	0 (0)	0.000
NOx	0.001 (0.002)	31 (34)	0.000
Particulate	0.000 (0.000)	0 (0)	0.000
Hg	0.00E+0 (0.00E+0)	0.000 (0.000)	0.00E+0
CO	0.000 (0.000)	0 (0)	0.000
CO_2[B]	2 (5)	60,298 (66,467)	0.4
CO_2e[C]	28 (64)	858,396 (946,220)	4.6

[A]Calculations based on a 90 percent capacity factor
[B]Plant stack emissions
[C]Life cycle emissions

The natural gas feed contains 5.74 ppmv of the mercaptan additive methanethiol (CH_4S). A zinc oxide sulfur guard bed was used to remove 98 percent of the sulfur in the feed, resulting in a maximum sulfur content of 0.1 ppm post-sulfur guard bed.

NOx emissions are limited by the use of low-NOx burners in the design. The production of PM is a result of system inefficiencies and is not produced or emitted in any significant amount. The pipeline natural gas was assumed to contain no Hg or HCl, resulting in zero emissions.

Ninety five percent of the CO_2 from the syngas is captured in the pre-combustion AGR system and ninety percent of the CO_2 from the stack gas is captured in the post-combustion AGR system and compressed for storage.

The carbon balance for the plant is shown in Exhibit 3-26. The carbon input to the plant consists of carbon in the air in addition to carbon in the natural gas. Carbon leaves the plant as a small portion of the hydrogen product, the captured CO_2 product, the carbon in the CO_2 dryer vent and knockout, and the CO_2 in the stack gas. The overall carbon capture efficiency is 96.2 percent, defined as one minus the mass of carbon in the stack gas and CO_2 dryer vent relative to the total carbon, represented by the following equation:

$$Carbon\ Capture\ Efficiency\ (\%) = \left(1 - \left(\frac{Carbon\ in\ Stack\ and\ CO_2\ Dryer\ Vent}{Total\ Carbon\ In}\right)\right) * 100\%$$

Exhibit 3-26. Case 2 carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		**kg/hr (lb/hr)**
Natural Gas	54,512 (120,178)	Stack Gas	2,081 (4,588)
Air (CO_2)	81 (178)	Export Steam	–
		Hydrogen Product	2 (5)
		Sulfur Guard Bed	0 (1)
		CO_2 Product	52,502 (115,747)
		CO_2 Dryer Vent	6.9 (15)
		CO_2 Knockout	0.1 (0.3)
Total	**54,593 (120,356)**	**Total**	**54,593 (120,356)**

Exhibit 3-27 shows the sulfur balance for the plant. Sulfur input comes solely from the mercaptan additive in the natural gas. Sulfur output includes sulfur recovered in the sulfur guard bed and small amounts of sulfur recovered in the AGR system.

Exhibit 3-27. Case 2 sulfur balance

Sulfur In		Sulfur Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Natural Gas	1 (2)	AGR Sulfur	0 (0)
		Sulfur Guard Bed	1 (1)
Total	**1 (2)**	**Total**	**1 (2)**

Exhibit 3-28 shows the overall water balance for the plant.

Exhibit 3-28. Case 2 water balance

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)
CO_2 Drying	–	–	–	0.0 (0.9)	0.0 (-0.9)
CO_2 Capture System Makeup	0.1 (17)	–	0.1 (17)	–	0.1 (17)
CO_2 Capture Recovery	–	–	–	–	–
CO_2 Compression Recovery	–	–	–	0.0 (7.3)	0.0 (-7.3)
Steam Generation	**3.0 (781)**	**2.4 (642)**	**0.5 (138)**	**–**	**0.5 (138)**
Export Steam	–	–	–	–	–
SMR Steam	3.0 (781)	–	3.0 (781)	–	3.0 (781)
Syngas Condensate	–	2.4 (642)	-2.4 (-642)	–	-2.4 (-642)
Cooling Tower	**9.7 (2,571)**	**–**	**9.7 (2,571)**	**2.2 (578)**	**7.5 (1,993)**
Total	**13 (3,369)**	**2.4 (642)**	**10 (2,727)**	**2.2 (586)**	**8.1 (2,140)**

An overall plant energy balance is provided in tabular form in Exhibit 3-29.

Exhibit 3-29. Case 2 overall energy balance (0 °C [32 °F] reference)

	HHV	Sensible + Latent	Power	Total
Heat In, GJ/hr (MMBtu/hr)				
Natural Gas	3,947 (3,741)	2.6 (2.5)	–	3,949 (3,743)
Air	–	18 (17)	–	18 (17)
Raw Water Makeup	–	39 (37)	–	39 (37)
Auxiliary Power	–	–	146 (139)	146 (139)
Total	**3,947 (3,741)**	**59 (56)**	**146 (139)**	**4,153 (3,936)**

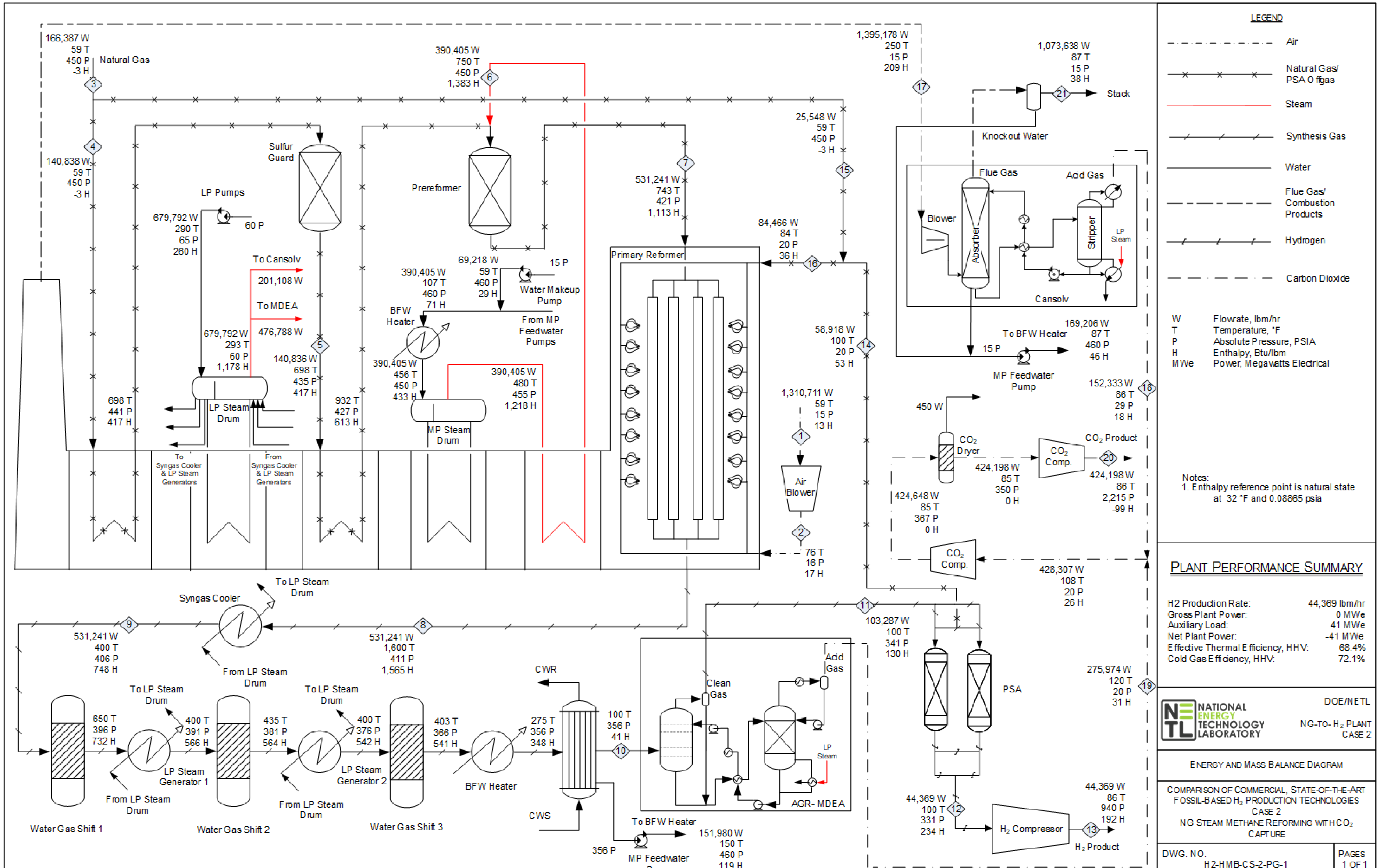
	HHV	Sensible + Latent	Power	Total
Heat Out, GJ/hr (MMBtu/hr)				
Hydrogen Product	2,846 (2,697)	9.0 (8.5)	–	2,855 (2,706)
Stack Gas	–	43 (41)	–	43 (41)
Sulfur	–	0.0 (0.0)	–	0.0 (0.0)
Motor Losses and Design Allowances	–	–	4.5 (4.3)	4.5 (4.3)
Cooling Tower Load[A]	–	1,271 (1,205)	–	1,271 (1,205)
Cooling Tower Blowdown	–	16 (15)	–	16 (15)
CO_2 Product Stream	–	-44 (-42)	–	-44 (-42)
Power	–	–	0.0 (0.0)	0.0 (0.0)
Total	**2,846 (2,697)**	**1,295 (1,228)**	**4.5 (4.3)**	**4,146 (3,929)**
Unaccounted Energy[B]	–	6.9 (6.6)	–	6.9 (6.6)

[A]Includes AGR, compression, and last stage syngas cooling loads
[B]By difference

3.9.2 Energy and Mass Balance Diagrams

The energy and mass balance diagram for Case 2 is shown in Exhibit 3-30.

Exhibit 3-30. Case 2 steam methane reforming plant with CO_2 capture energy and mass balance



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3.10 Case 2 – Major Equipment List

Major equipment items for the steam methane reforming plant with CO_2 capture are shown in the following tables. In general, the design conditions include a 10 percent design allowance for flows and heat duties and a 21 percent design allowance for heads on pumps and fans.

Case 2 – Account 3: Feedwater and Miscellaneous Balance of Plant Systems

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	LP Feedwater Pumps	Horizontal centrifugal	2,850 lpm @ 4 m H_2O (750 gpm @ 10 ft H_2O)	2	1
2	MP Feedwater Pumps	Horizontal, split case, multi-stage, centrifugal	710 lpm @ 379 m H_2O (190 gpm @ 1,240 ft H_2O)	2	1
			640 lpm @ 88 m H_2O (170 gpm @ 290 ft H_2O)	2	1
3	Water Makeup Pumps	Horizontal, split case, multi-stage, centrifugal	290 lpm @ 380 m H_2O (80 gpm @ 1,240 ft H_2O)	2	2
4	Auxiliary Boiler	Shop fabricated, water tube	18,000 kg/hr, 2.8 MPa, 343 °C (40,000 lb/hr, 400 psig, 650 °F)	1	0
5	Service Air Compressors	Flooded Screw	13 m^3/min @ 0.7 MPa (450 scfm @ 100 psig)	2	1
6	Instrument Air Dryers	Duplex, regenerative	13 m^3/min (450 scfm)	2	1
7	Engine-Driven Fire Pump	Vertical turbine, diesel engine	3,785 lpm @ 110 m H_2O (1,000 gpm @ 350 ft H_2O)	1	1
8	Fire Service Booster Pump	Two-stage horizontal centrifugal	2,650 lpm @ 80 m H_2O (700 gpm @ 250 ft H_2O)	1	1
9	Raw Water Pumps	Stainless steel, single suction	5,700 lpm @ 20 m H_2O (1,500 gpm @ 60 ft H_2O)	2	1
10	Filtered Water Pumps	Stainless steel, single suction	190 lpm @ 50 m H_2O (50 gpm @ 160 ft H_2O)	2	1
11	Filtered Water Tank	Vertical, cylindrical	182,000 liter (48,000 gal)	1	0
12	Makeup Water Demineralizer	Multi-media filter, cartridge filter, RO membrane assembly and electro-deionization unit	400 lpm (110 gpm)	1	0
13	Liquid Waste Treatment System	–	10 years, 24-hour storm	1	0
14	Gas Pipeline	Underground, coated carbon steel, wrapped cathodic protection	57 m^3/min @ 3.1 MPa (2,001 acfm @ 450 psia) 39 cm (16 in) standard wall pipe	16 km (10 mile)	0
15	Gas Metering Station	–	57 m^3/min (2,001 acfm)	1	0

Case 2 – Account 4: Reformer and Accessories

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Primary Reformer	SMR, side fired, fixed bed, catalytic	Syngas Production: 265,000 kg/hr @ 2.8 MPa, 871 °C (584,000 lb/hr @ 411 psia, 1600 °F)	1	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
2	Air Blower	Centrifugal	654,000 kg/hr @ 4,000 m^3/min, 3cm WG (1,442,000 lb/hr @ 150,000 acfm, 1in. WG)	2	1
3	Sulfur Guard Bed	Fixed Bed, catalytic (ZnO)	Inlet: 70,000 kg/hr @ 3.0 MPa, 370 °C (155,000 lb/hr @ 441 psia, 698 °F)	1	0
4	Prereformer	Fixed Bed, catalytic	NG In: 70,000 kg/hr @ 2.9 MPa, 500 °C (155,000 lb/hr @ 427 psia, 932 °F) Steam In: 195,000 kg/hr @ 3.1 MPa, 399 °C (429,000 lb/hr @ 450 psia, 750 °F)	1	0
5	Syngas Coolers	Shell and tube heat exchanger	Syngas Cooler: 457 GJ/hr (434 MMBtu/hr) LP Steam Generator 1: 93 GJ/hr (88 MMBtu/hr) LP Steam Generator 2: 13 GJ/hr (12 MMBtu/hr) BFW Heater: 108 GJ/hr (103 MMBtu/hr) AGR Precooler: 172 GJ/hr (163 MMBtu/hr)	5	0

Case 2 – Account 5: Flue Gas Cleanup

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Cansolv	Amine-based CO2 capture technology	696,000 kg/hr (1,535,000 lb/hr) 12.0 wt % CO2 concentration	1	0
2	CO_2 Dryer	Triethylene glycol	Inlet: 63 m^3/min @ 2.5 MPa (2,222 acfm @ 367 psia) Outlet: 2.4 MPa (350 psia) Water Recovered: 204 kg/hr (450 lb/hr)	1	0
3	CO_2 Compressor	Integrally geared, multi-stage centrifugal	7.0 m^3/min @ 15.3 MPa, 85 °C (264 acfm @ 2,217 psia, 185 °F)	1	0
4	CO_2 Aftercooler	Shell and tube heat exchanger	Outlet: 15.3 MPa, 30 °C (2,215psia, 86 °F) Duty: 32 GJ/hr (30 MMBtu/hr)	1	0

Case 2 – Account 6: Syngas Cleanup

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Water Gas Shift Reactors	Fixed bed, catalytic	88,000 kg/hr (292,000 lb/hr) 204 °C (400 °F) 2.8 MPa (410 psia)	6	0
2	Acid Gas Removal Plant	MDEA	265,000 kg/hr (584,000 lb/hr) 38 °C (100 °F) 2.5 MPa (356 psia)	1	0

Case 2 – Account 7: Hydrogen Production

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Pressure Swing Adsorber	Polybed Proprietary	Syngas: 46,850 kg/hr (103,287 lb/hr) 38 °C (100 °F) 2.4 MPa (341.0 psia) Hydrogen: 20,125 kg/hr (44,369 lb/hr) 38 °C (100 °F) 2.3 MPa (331.0 psia) Off Gas: 26,725 kg/hr (58,918 lb/hr) 38 °C (100 °F) 0.1 MPa (20.0 psia)	1	0
2	Hydrogen Compressor	Integrally geared, multi-stage centrifugal	73 m^3/min @ 6.5 MPa (2,587 acfm @ 940 psia)	1	1

Case 2 – Account 9: Cooling Water System

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Circulating Water Pumps	Vertical, wet pit	251,000 lpm @ 30 m (66,000 gpm @ 100 ft)	2	1
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	11 °C (52 °F) wet bulb/16 °C (60 °F) CWT/ 27 °C (80 °F) HWT/ 1400 GJ/hr (1330 MMBtu/hr) heat duty	1	0

Case 2 – Account 11: Accessory Electric Plant

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	High Voltage Auxiliary Transformer	Oil-filled	345 kV/13.8 kV, 9 MVA, 3-ph, 60 Hz	2	0
2	Medium Voltage Transformer	Oil-filled	18 kV/4.16 kV, 20 MVA, 3-ph, 60 Hz	1	1
3	Low Voltage Transformer	Dry ventilated	4.16 kV/480 V, 7 MVA, 3-ph, 60 Hz	1	1
4	Medium Voltage Switchgear	Metal clad	4.16 kV, 3-ph, 60 Hz	1	1
5	Low Voltage Switchgear	Metal enclosed	480 V, 3-ph, 60 Hz	1	1
6	Emergency Diesel Generator	Sized for emergency shutdown	750 kW, 480 V, 3-ph, 60 Hz	1	0

Case 2 – Account 12: Instrumentation and Control

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	DCS – Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser black and white)	Operator stations/printers and engineering stations/printers	1	0
2	DCS – Processor	Microprocessor with redundant input/output	N/A	1	0
3	DCS – Data Highway	Fiber optic	Fully redundant, 25% spare	1	0

3.11 Case 2 – Cost Estimating

The cost estimating methodology was described previously in Section 2.11. Exhibit 3-31 shows a detailed breakdown of the capital costs; Exhibit 3-32 shows the owner's costs, TOC, and TASC; Exhibit 3-33 shows the initial and annual O&M costs; and Exhibit 3-34 shows the LCOH breakdown.

The estimated TOC of the steam methane reforming plant with CO_2 capture is $1,735/[kg H_2/day]. An 8 percent increase in the primary reformer above Case 1 is observed due to the higher fired duty of the furnace from the addition of CO_2 capture equipment. Process contingency represents 3 percent of the TOC and project contingency represents 14.5 percent. The LCOH is $1.64/kg H_2 with CO_2 T&S.

Exhibit 3-31. Case 2 total plant cost details

	Case:	2	– Steam Methane Reforming w/ CCS						Estimate Type:	Conceptual	
	Plant Size (kg H_2/day):	483,007							Cost Base:	Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/[kg H_2/day]
	3	Feedwater & Miscellaneous BOP Systems									
3.1	Feedwater System	$233	$195	$484	$377	$1,289	$258	$0	$232	$1,779	$4
3.2	Water Makeup & Pretreating	$3,553	$355	$2,014	$0	$5,922	$1,184	$0	$1,421	$8,528	$18
3.3	Other Feedwater Subsystems	$965	$316	$300	$0	$1,581	$316	$0	$285	$2,182	$5
3.4	Service Water Systems	$1,078	$2,059	$6,666	$0	$9,803	$1,961	$0	$2,353	$14,116	$29
3.5	Other Boiler Plant Systems	$230	$84	$209	$0	$523	$105	$0	$94	$722	$1
3.6	Natural Gas Pipeline and Start-Up System	$9,304	$400	$300	$0	$10,005	$2,001	$0	$1,801	$13,807	$29
3.7	Wastewater Treatment Equipment	$4,611	$0	$2,826	$0	$7,437	$1,487	$0	$1,785	$10,709	$22
3.9	Miscellaneous Plant Equipment	$14,217	$1,865	$7,225	$0	$23,306	$4,661	$0	$5,594	$33,561	$69
	Subtotal	**$34,191**	**$5,273**	**$20,024**	**$377**	**$59,866**	**$11,973**	**$0**	**$13,564**	**$85,403**	**$177**
	4	Reformer & Accessories									
4.1	Primary Reformer	$27,110	$1,165	$8,736	$6,814	$43,825	$8,765	$0	$13,147	$65,737	$136
4.2	Air Blower	$634	$564	$888	$693	$2,778	$556	$0	$500	$3,834	$8
4.3	Sulfur Guard Bed	$192	$236	$362	$283	$1,073	$215	$0	$193	$1,481	$3
4.4	Prereformer	$1,273	$302	$462	$361	$2,398	$480	$0	$719	$3,597	$7
4.5	Syngas Coolers	$9,007	$1,425	$1,504	$1,173	$13,108	$2,622	$0	$3,932	$19,662	$41
4.6	Air Separation Unit/Oxidant Compression	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.7	Steam Drum(s)	w/4.1	w/4.1	w/4.1	w/4.1	$0	$0	$0	$0	$0	$0
	Subtotal	**$38,214**	**$3,693**	**$11,953**	**$9,322**	**$63,182**	**$12,636**	**$0**	**$18,493**	**$94,311**	**$195**
	5	Flue Gas Cleanup									
5.1	Cansolv Carbon Dioxide (CO_2) Removal System	$136,003	$160	$482	$376	$137,022	$27,404	$24,664	$47,272	$236,362	$489
5.4	Carbon Dioxide (CO_2) Compression & Drying	$40,280	$1,383	$3,438	$2,681	$47,783	$9,557	$0	$14,335	$71,674	$148
5.5	Carbon Dioxide (CO_2) Compressor Aftercooler	$1,662	$396	$708	$552	$3,318	$664	$0	$597	$4,578	$9
5.12	Gas Cleanup Foundations	w/ 5.1,4,&5	w/ 5.1,4,&5	w/ 5.1,4,&5	w/ 5.1,4,&5	$0	$0	$0	$0	$0	$0
	Subtotal	**$177,945**	**$1,939**	**$4,628**	**$3,610**	**$188,122**	**$37,624**	**$24,664**	**$62,204**	**$312,615**	**$647**
	6	Syngas Cleanup									

	Case:	2	– Steam Methane Reforming w/ CCS						Estimate Type:	Conceptual	
	Plant Size (kg H_2/day):	483,007							Cost Base:	Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/[kg H_2/day]
6.1	Recycle Compressor	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.2	Methyl Diethanolamine (MDEA) – Low Temperature Acid Gas Removal	$19,133	$0	$539	$420	$20,093	$4,019	$0	$4,822	$28,933	$60
6.7	Water Gas Shift (WGS) Reactors	$1,764	$1,882	$2,769	$2,160	$8,576	$1,715	$0	$2,058	$12,349	$26
6.12	Gas Cleanup Foundations	w/ 6.2 & 7	w/ 6.2 & 7	w/ 6.2 & 7	w/ 6.2 & 7	$0	$0	$0	$0	$0	$0
	Subtotal	**$20,897**	**$1,882**	**$3,308**	**$2,580**	**$28,668**	**$5,734**	**$0**	**$6,880**	**$41,282**	**$85**
	7					Hydrogen Production					
7.1	Pressure Swing Adsorber	$16,112	$23	$730	$569	$17,434	$3,487	$0	$4,184	$25,105	$52
7.2	Hydrogen Compressor	$6,844	$187	$501	$391	$7,922	$1,584	$0	$1,901	$11,408	$24
	Subtotal	**$22,956**	**$211**	**$1,230**	**$960**	**$25,356**	**$5,071**	**$0**	**$6,085**	**$36,513**	**$76**
	8					Fired Heater, Ductwork, & Stack					
8.1	Fired Heater	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
8.2	Fired Heater Accessories	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
8.3	Ductwork	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
8.4	Stack	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
8.5	Fired Heater, Ductwork & Stack Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	Subtotal	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
	9					Cooling Water System					
9.1	Cooling Towers	$7,438	$0	$2,252	$0	$9,690	$1,938	$0	$1,744	$13,373	$28
9.2	Circulating Water Pumps	$985	$0	$60	$0	$1,046	$209	$0	$188	$1,443	$3
9.3	Circulating Water System Auxiliaries	$7,716	$0	$1,018	$0	$8,735	$1,747	$0	$1,572	$12,054	$25
9.4	Circulating Water Piping	$0	$2,153	$1,949	$0	$4,102	$820	$0	$738	$5,661	$12
9.5	Make-up Water System	$294	$0	$378	$0	$671	$134	$0	$121	$927	$2
9.6	Component Cooling Water System	$317	$0	$244	$0	$561	$112	$0	$101	$774	$2
9.7	Circulating Water System Foundations	$0	$498	$827	$0	$1,325	$265	$0	$318	$1,908	$4
	Subtotal	**$16,751**	**$2,651**	**$6,728**	**$0**	**$26,130**	**$5,226**	**$0**	**$4,783**	**$36,139**	**$75**
	11					Accessory Electric Plant					
11.2	Station Service Equipment	$0	$1,122	$94	$74	$1,291	$258	$0	$232	$1,781	$4
11.3	Switchgear & Motor Control	$0	$659	$220	$172	$1,051	$210	$0	$189	$1,451	$3
11.4	Conduit & Cable Tray	$0	$266	$481	$375	$1,121	$224	$0	$202	$1,548	$3
11.5	Wire & Cable	$0	$477	$345	$269	$1,092	$218	$0	$197	$1,507	$3

Case:		2	– Steam Methane Reforming w/ CCS						Estimate Type:		Conceptual	
Plant Size (kg H_2/day):		483,007							Cost Base:		Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/[kg H_2/day]
11.6	Protective Equipment	$0	$444	$769	$600	$1,813	$363	$0	$326	$2,501	$5
11.7	Standby Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
11.8	Main Power Transformers	$0	$295	$88	$69	$452	$90	$0	$81	$623	$1
11.9	Electrical Foundations	$0	$7	$22	$17	$46	$9	$0	$11	$66	$0
	Subtotal	**$0**	**$3,270**	**$2,020**	**$1,575**	**$6,865**	**$1,373**	**$0**	**$1,239**	**$9,477**	**$20**
12	Instrumentation & Control										
12.4	Other Major Component Control Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
12.5	Signal Processing Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
12.6	Control Boards, Panels & Racks	$0	$234	$23	$18	$275	$55	$14	$52	$396	$1
12.7	Distributed Control System Equipment	$7,800	$0	$238	$0	$8,038	$1,608	$402	$1,507	$11,555	$24
12.8	Instrument Wiring & Tubing	$0	$207	$346	$270	$822	$164	$41	$154	$1,182	$2
12.9	Other Instrumentation & Controls Equipment	$0	$1,771	$317	$247	$2,336	$467	$117	$438	$3,358	$7
	Subtotal	**$7,800**	**$2,212**	**$925**	**$535**	**$11,472**	**$2,294**	**$574**	**$2,151**	**$16,491**	**$34**
13	Improvements to Site										
13.1	Site Preparation	$0	$525	$11,143	$0	$11,667	$2,333	$0	$2,800	$16,801	$35
13.2	Site Improvements	$0	$307	$1,303	$1,016	$2,627	$525	$0	$630	$3,783	$8
13.3	Site Facilities	$1,620	$0	$1,700	$0	$3,320	$664	$0	$797	$4,780	$10
	Subtotal	**$1,620**	**$832**	**$14,146**	**$1,016**	**$17,614**	**$3,523**	**$0**	**$4,227**	**$25,364**	**$53**
14	Buildings & Structures										
14.2	Interconnecting Pipe	$0	$656	$1,558	$1,215	$3,430	$686	$0	$617	$4,734	$10
14.4	Administration Building	$0	$333	$226	$0	$559	$112	$0	$101	$771	$2
14.5	Circulation Water Pumphouse	$0	$50	$25	$0	$75	$15	$0	$14	$104	$0
14.6	Water Treatment Buildings	$0	$298	$271	$0	$569	$114	$0	$102	$785	$2
14.7	Machine Shop	$0	$497	$318	$0	$816	$163	$0	$147	$1,126	$2
14.8	Warehouse	$0	$389	$235	$0	$624	$125	$0	$112	$861	$2
14.9	Structural Steel Pipe Racks	$0	$1,726	$982	$766	$3,474	$695	$0	$625	$4,795	$10
14.10	Waste Treating Building & Structures	$0	$655	$1,170	$0	$1,826	$365	$0	$329	$2,519	$5
	Subtotal	**$0**	**$4,606**	**$4,785**	**$1,981**	**$11,373**	**$2,275**	**$0**	**$2,047**	**$15,694**	**$32**
	Total	**$320,374**	**$26,569**	**$69,748**	**$21,958**	**$438,648**	**$87,730**	**$25,237**	**$121,673**	**$673,289**	**$1,394**

Exhibit 3-32. Case 2 owner's costs

Description	$/1,000	$/[kg H_2/day]
Pre-Production Costs		
6 Months All Labor	$4,934	$10
1 Month Maintenance Materials	$711	$1
1 Month Non-Fuel Consumables	$2,838	$6
1 Month Waste Disposal	$7	$0
25% of 1 Months Fuel Cost at 100% CF	$3,018	$6
2% of TPC	$13,466	$28
Total	**$24,972**	**$52**
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	$1,191	$2
0.5% of TPC (spare parts)	$3,366	$7
Total	**$4,558**	**$9**
Other Costs		
Initial Cost for Catalyst and Chemicals	$15,867	$33
Land	$300	$1
Other Owner's Costs	$100,993	$209
Financing Costs	$18,179	$38
Total Overnight Costs (TOC)	**$838,157**	**$1,735**
TASC Multiplier (IOU, 35 year)	1.070	
Total As-Spent Cost (TASC)	**$896,662**	**$1,856**

Exhibit 3-33. Case 2 initial and annual operating and maintenance costs

Case:	2	– Steam Methane Reforming w/ CCS			Cost Base:	Dec 2018
Plant Size (kg H_2/day):	**483,007**				**Capacity Factor (%):**	**90**
			Operating & Maintenance Labor			
Operating Labor				**Operating Labor Requirements per Shift**		
Operating Labor Rate (base):		38.50	**$/hour**	Skilled Operator:		1.0
Operating Labor Burden:		30.00	**% of base**	Operator:		3.3
Labor O-H Charge Rate:		25.00	**% of labor**	Foreman:		1.0
				Lab Techs, etc.:		1.0
				Total:		**6.3**
			Fixed Operating Costs			
					Annual Cost	
					($)	**($/[kg H_2/day])**
Annual Operating Labor:					$2,776,628	$5.749
Maintenance Labor:					$5,116,993	$10.594
Administrative & Support Labor:					$1,973,405	$4.086
Property Taxes and Insurance:					$13,465,770	$27.879
Total:					**$23,332,796**	**$48.307**
			Variable Operating Costs			
					($)	**($/kg H_2)**
Maintenance Material:					$7,675,489	$0.05
			Consumables			
	Initial Fill	**Per Day**	**Per Unit**	**Initial Fill**		
Water (gal/1000):	0	1,963	$1.90	$0	$1,225,337	$0.00772
Makeup and Wastewater Treatment Chemicals (ton):	0	5.8	$550.00	$0	$1,056,610	$0.00666
Zinc Oxide Sulfur Guard Catalyst (ft^3):	705	0.5	$600.00	$423,000	$95,175	$0.00060
Prereformer Catalyst (ft^3):	818	0.6	$1,250.00	$1,022,500	$230,063	$0.00145
Primary Reformer Catalyst (ft^3):	3160	1.7	$525.00	$1,659,000	$298,620	$0.00188
Water Gas Shift Catalyst (ft^3):	13983	9.6	$480.00	$6,711,929	$1,510,184	$0.00952
Methyl Diethanolamine Solution (gal):	392963	163	$2.80	$1,100,298	$149,834	$0.00094
Post-Combustion CO_2 Capture System[A]:			Proprietary		$2,364,483	$0.01490
PSA Unit Adsorbent (ft^3):	33000	4.5	$150.00	$4,950,000	$222,750	$0.00140
Triethylene Glycol (gal):	w/equip.	226	$6.80	$0	$504,574	$0.00318
Electricity (MWh):	0	976	$71.70	$0	$22,990,051	$0.14489
Subtotal:				**$15,866,727**	**$30,647,681**	**$0.19316**
			Waste Disposal			
Zinc Oxide Sulfur Guard Catalyst (ft^3):	0	0.5	$40.00	$0	$6,345	$0.00004
Prereformer Catalyst (ft^3):	0	0.6	$0.00	$0	$0	$0.00000
Primary Reformer Catalyst (ft^3):	0	1.7	$0.00	$0	$0	$0.00000
Water Gas Shift Catalyst (ft^3):	0	9.6	$2.50	$0	$7,866	$0.00005
Methyl Diethanolamine Solution (gal):	0	163	$0.35	$0	$18,729	$0.00012
PSA Unit Adsorbent (ft^3):	0	4.5	$1.50	$0	$2,228	$0.00001
Triethylene Glycol (gal):	0	226	$0.35	$0	$25,971	$0.00016
Amine Purification Unit Waste (ton):	0	0.8	$38.00	$0	$10,388	$0.00007
Thermal Reclaimer Unit Waste (ton):	0	0.07	$38.00	$0	$923	$0.00001
Subtotal:				**$0**	**$72,448**	**$0.00046**

Case:	2	– Steam Methane Reforming w/ CCS			Cost Base:	Dec 2018
Plant Size (kg H_2/day):	483,007				Capacity Factor (%):	90
By-Products						
					($)	($/kg H_2)
Steam (ton):	0	0	$0	$0	$0	$0.00000
Electricity (MWh):	0	0	$71.70	$0	$0	$0.00000
Subtotal:				**$0**	**$0**	**$0.00000**
Variable Operating Costs Total:				**$15,866,727**	**$38,395,618**	**$0.24199**
Fuel Cost						
Natural Gas (MMBtu):	0	89,780	$4.42	$0	$130,363,541	$0.82161
Total:				**$0**	**$130,363,541**	**$0.82161**

[A]Post-combustion CO_2 capture system chemicals includes ion exchange resin, NaOH, and Cansolv solvent.

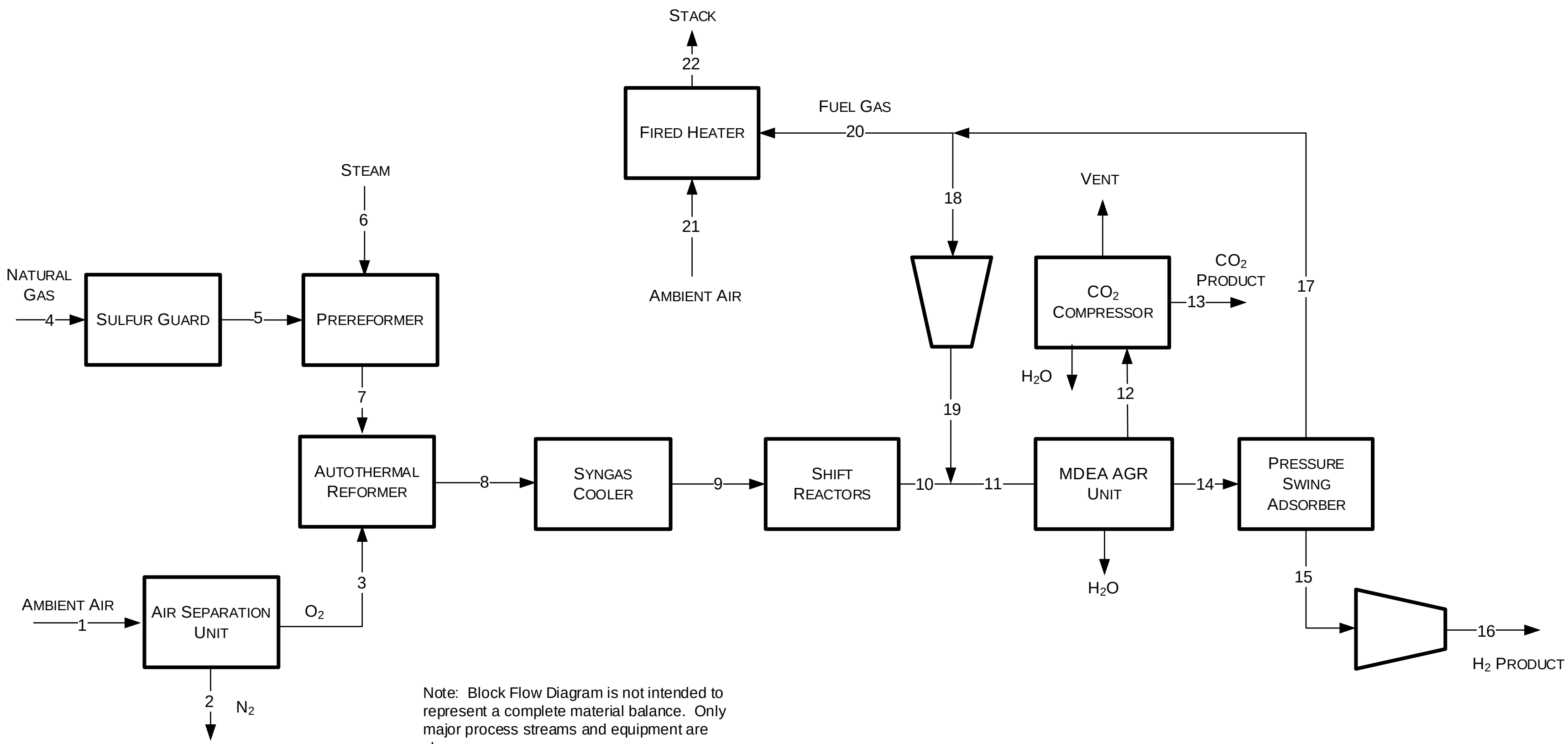
Exhibit 3-34. Case 2 LCOH breakdown

Component	Value, $/kg H_2	Percentage
Capital	0.33	20%
Fixed	0.15	9%
Variable	0.24	15%
Fuel	0.82	50%
Total (Excluding T&S)	**1.54**	**N/A**
CO_2 T&S	0.10	6%
Total (Including T&S)	**1.64**	**N/A**

3.12 Case 3 – Autothermal Reforming Plant with CO_2 Capture Process Description

This section contains an evaluation of a plant design for Case 3, which is based on a natural gas autothermal reforming plant with CO_2 capture. The system descriptions follow the BFD provided in Exhibit 3-35 with the associated stream tables that show process data provided in Exhibit 3-36.

Exhibit 3-35. Case 3 block flow diagram, autothermal reforming plant with CO_2 capture



Note: Block Flow Diagram is not intended to represent a complete material balance. Only major process streams and equipment are shown.

Exhibit 3-36. Case 3 stream table, autothermal reforming plant with capture

	1	2	3	4	5	6	7	8	9	10	11
V-L Mole Fraction											
Ar	0.0092	0.0024	0.0343	0.0000	0.0000	0.0000	0.0000	0.0051	0.0051	0.0051	0.0071
CH_4	0.0000	0.0000	0.0000	0.9310	0.9310	0.0000	0.3603	0.0007	0.0007	0.0007	0.0010
CH_4S	0.0000	0.0000	0.0000	0.0000[A]	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0000	0.0000	0.0000	0.0320	0.0320	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_3H_8	0.0000	0.0000	0.0000	0.0070	0.0070	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_4H_{10}	0.0000	0.0000	0.0000	0.0040	0.0040	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.1385	0.1385	0.0039	0.0055
CO_2	0.0003	0.0004	0.0000	0.0100	0.0100	0.0000	0.0210	0.0811	0.0811	0.2157	0.2111
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0486	0.4342	0.4342	0.5689	0.5749
H_2O	0.0099	0.0086	0.0000	0.0000	0.0000	1.0000	0.5640	0.3345	0.3345	0.1999	0.1925
N_2	0.7732	0.9850	0.0157	0.0160	0.0160	0.0000	0.0059	0.0057	0.0057	0.0057	0.0080
O_2	0.2074	0.0035	0.9501	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	18,428	14,401	3,969	5,594	5,594	9,150	15,279	26,451	26,451	26,451	27,467
V-L Flowrate (kg/hr)	531,787	402,901	127,850	96,930	96,929	164,844	261,772	389,623	389,623	389,623	400,813
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	15	21	27	15	370	399	407	1,093	204	38	38
Pressure (MPa, abs)	0.10	0.45	3.31	3.10	3.00	3.10	2.90	2.83	2.80	2.45	2.45
Steam Table Enthalpy (kJ/kg)[B]	30.23	28.12	14.67	-6.77	969.07	3,226.40	2,378.90	3,617.35	1,446.86	78.31	78.80
AspenPlus Enthalpy (kJ/kg)[C]	-97.58	-92.37	-7.83	-4,516.00	-3,540.22	-12,753.90	-9,165.98	-6,160.81	-8,331.31	-9,664.72	-9,497.97
Density (kg/m^3)	1.2	5.2	43.9	24.4	9.7	10.4	8.9	3.7	10.5	17.4	17.1
V-L Molecular Weight	28.857	27.976	32.209	17.328	17.328	18.015	17.133	14.730	14.730	14.730	14.593
V-L Flowrate (lb_{mol}/hr)	40,628	31,750	8,751	12,333	12,332	20,173	33,685	58,314	58,314	58,314	60,554
V-L Flowrate (lb/hr)	1,172,390	888,245	281,862	213,694	213,691	363,418	577,109	858,971	858,971	858,971	883,642
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	59	70	80	59	698	750	765	2,000	400	100	100
Pressure (psia)	14.7	65.0	480.0	450.0	435.3	450.0	421.0	411.0	406.0	356.0	356.0
Steam Table Enthalpy (Btu/lb)[A]	13.0	12.1	6.3	-2.9	416.6	1,387.1	1,022.7	1,555.2	622.0	33.7	33.9
AspenPlus Enthalpy (Btu/lb)[B]	-42.0	-39.7	-3.4	-1,941.5	-1,522.0	-5,483.2	-3,940.7	-2,648.7	-3,581.8	-4,155.1	-4,083.4
Density (lb/ft^3)	0.076	0.322	2.743	1.525	0.605	0.652	0.558	0.228	0.656	1.086	1.066

[A]Natural gas entering battery limits contains 5.74 ppm of the mercaptan additive methanethiol (CH_4S)
[B]Steam table reference conditions are 32.02 °F & 0.089 psia
[C]AspenPlus thermodynamic reference state is the component's constituent elements in an ideal gas state at 25 °C and 1 atm

Exhibit 3-36. Case 3 stream table, autothermal reforming plant with capture (continued)

	12	13	14	15	16	17	18	19	20	21	22
V-L Mole Fraction											
Ar	0.0000	0.0000	0.0116	0.0007	0.0007	0.0569	0.0569	0.0569	0.0569	0.0092	0.0278
CH_4	0.0000	0.0000	0.0016	0.0000	0.0000	0.0082	0.0082	0.0082	0.0082	0.0000	0.0000
CH_4S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_3H_8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_4H_{10}	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0091	0.0000	0.0000	0.0466	0.0466	0.0466	0.0466	0.0000	0.0000
CO_2	0.9811	0.9995	0.0174	0.0000	0.0000	0.0895	0.0895	0.0895	0.0895	0.0003	0.0525
H_2	0.0001	0.0000	0.9471	0.9991	0.9991	0.7313	0.7313	0.7313	0.7313	0.0000	0.0000
H_2O	0.0188	0.0005	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0099	0.2785
N_2	0.0000	0.0000	0.0132	0.0001	0.0001	0.0674	0.0674	0.0674	0.0674	0.7732	0.6265
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2074	0.0147
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	5,614	5,510	16,671	13,433	13,433	3,238	1,016	1,016	2,222	4,780	6,138
V-L Flowrate (kg/hr)	244,297	242,411	63,164	27,500	27,500	35,664	11,190	11,191	24,473	137,947	162,420
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	49	30	38	38	30	38	38	38	38	15	121
Pressure (MPa, abs)	0.14	15.27	2.35	2.28	6.48	0.14	0.14	2.76	0.14	0.10	0.10
Steam Table Enthalpy (kJ/kg)[A]	61.61	-231.09	289.28	537.24	440.76	99.65	99.65	96.04	99.65	30.23	614.39
AspenPlus Enthalpy (kJ/kg)[B]	-8,956.76	-9,194.65	-2,002.40	186.44	89.95	-3,688.61	-3,688.61	-3,692.21	-3,688.61	-97.58	-3,215.03
Density (kg/m^3)	2.3	630.1	3.4	1.8	5.1	0.6	0.6	11.7	0.6	1.2	0.8
V-L Molecular Weight	43.516	43.997	3.789	2.047	2.047	11.013	11.013	11.013	11.013	28.857	26.460
V-L Flowrate (lb_{mol}/hr)	12,377	12,147	36,754	29,614	29,614	7,140	2,240	2,240	4,899	10,539	13,533
V-L Flowrate (lb/hr)	538,583	534,424	139,252	60,627	60,627	78,625	24,671	24,671	53,954	304,121	358,075
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	120	86	100	100	86	100	100	100	100	59	250
Pressure (psia)	20.0	2,214.7	341.0	331.0	939.7	20.0	20.0	400.0	20.0	14.7	14.7
Steam Table Enthalpy (Btu/lb)[A]	26.5	-99.4	124.4	231.0	189.5	42.8	42.8	41.3	42.8	13.0	264.1
AspenPlus Enthalpy (Btu/lb)[B]	-3,850.7	-3,953.0	-860.9	80.2	38.7	-1,585.8	-1,585.8	-1,587.4	-1,585.8	-42.0	-1,382.2
Density (lb/ft^3)	0.141	39.338	0.213	0.112	0.318	0.037	0.037	0.728	0.037	0.076	0.051

[A]Steam table reference conditions are 32.02 °F & 0.089 psia

[B]AspenPlus thermodynamic reference state is the component's constituent elements in an ideal gas state at 25 °C and 1 atm

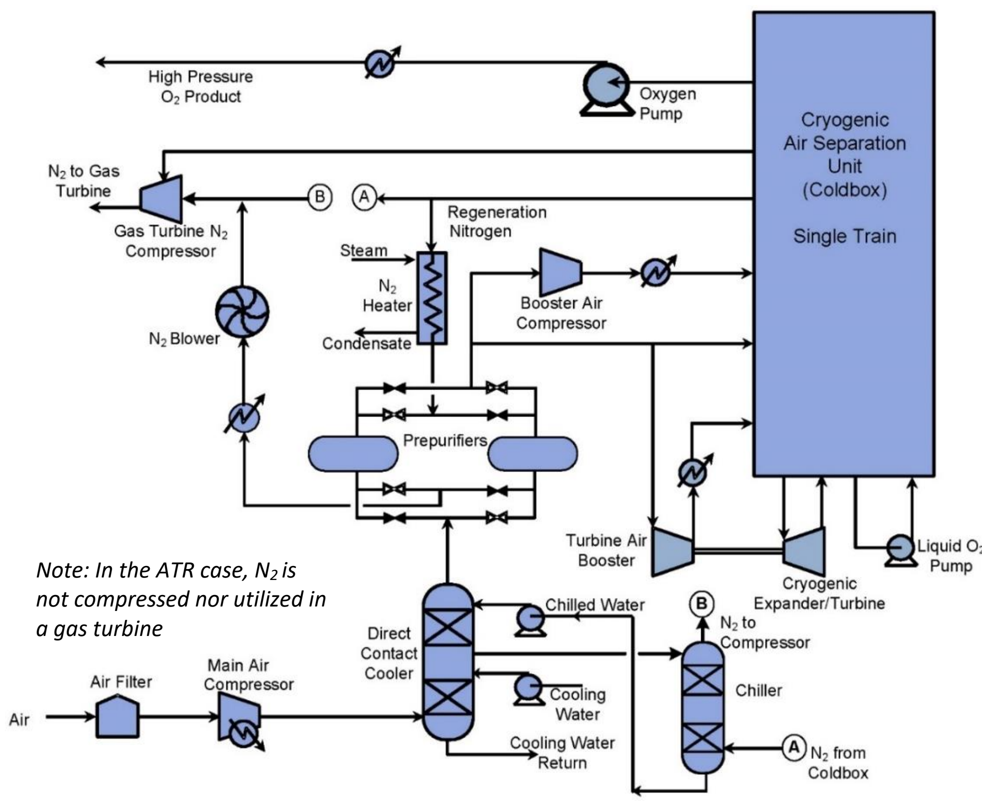
3.12.1 Air Separation Unit

An elevated pressure air separation design producing a 95 percent by volume (vol%) O_2 product at 3.3 MPa (480 psia) has been considered for Case 3. This pressure is needed to match the operating pressure of the ATR reactor and results in a main air compressor (MAC) discharge pressure of 1.6 MPa (236 psia). This is much greater than the MAC in a traditional ASU plant operating at about 0.7 MPa (105 psia). [61] The ASU power requirement for this Case 3 is 420 kWh/ton-O_2 (including the MAC, booster compressor, and auxiliaries, on a 100 percent pure O_2 basis). The amount of oxygen that is supplied to the ATR depends on the heating value of the fuel gas. Enough O_2 must be supplied to combust enough of the fuel gas to reach the reformer operating temperature of 2,000 °F. The residual nitrogen (N_2) that is produced is vented.

3.12.1.1 Air Separation Plant Process Description

The air separation plant is designed to produce 95 vol% O_2 for use in the ATR. The ASU is a single train design. The air compressor is powered by an electric motor. A process schematic of a typical elevated pressure ASU, which is based on vendor discussions and quotes, is shown in Exhibit 3-37. [62]

Exhibit 3-37. Typical ASU process schematic



The air feed to the ASU is supplied from the stand-alone MAC. Air to the compressor is first filtered in a suction filter upstream of the compressor. This air filter removes particulates, which tend to cause compressor wheel erosion and foul intercoolers. The filtered air is then compressed in the centrifugal compressor, with intercooling and aftercooling.

Air from the MAC is cooled in a two-stage direct contact cooler. Cooling for the first stage is provided with plant-cooling water. Cooling for the second stage is provided by chilled water generated by contact with cold N_2 exiting the cold box.

Chilled air is fed to a molecular sieve adsorber pre-purifier system. The adsorbent removes water, CO_2, and C_4+ saturated hydrocarbons in the air. After passing through the adsorption beds, the air is filtered with a dust filter to remove any adsorbent fines that may be present. Downstream of the dust filter a small stream of air is withdrawn to supply the instrument air requirements of the ASU.

Regeneration of the adsorbent in the pre-purifiers is accomplished by passing a hot N_2 stream through the off-stream bed(s) in a direction counter-current to the normal airflow. The N_2 is heated against extraction steam (1.7 MPa [250 psia]) in a shell and tube heat exchanger (HX). [62] The regeneration N_2 drives off the adsorbed contaminants. Following regeneration, the heated bed is cooled to near normal operating temperature by passing a cool N_2 stream through the adsorbent beds. The bed is re-pressurized with air and placed on stream so that the current on-stream bed(s) can be regenerated.

The air from the pre-purifier is then split into three streams. About 70 percent of the air is fed directly to the cold box. About 25 percent of the air is compressed in an air booster compressor. This boosted air is then cooled in an aftercooler against cooling water. The remaining five percent of the air is fed to a turbine-driven, single-stage, centrifugal booster compressor. This stream is cooled in a shell and tube aftercooler against cooling water before it is fed to the cold box. [62]

All three air feeds are cooled in the cold box to cryogenic temperatures against returning product O_2 and N_2 streams in plate-and-fin HXs. The large air stream is fed directly to the first distillation column to begin the separation process. The second largest air stream is liquefied against boiling liquid O_2 before it is fed to the distillation columns. The third, smallest air stream is fed to the cryogenic expander to produce refrigeration to sustain the cryogenic separation process.

Inside the cold box the air is separated into O_2 and N_2 products. The O_2 product is withdrawn from the distillation columns as a liquid and is pressurized by a cryogenic pump. The pressurized liquid O_2 is then vaporized against the high-pressure (HP) air feed before being warmed to ambient temperature. The HP liquid O_2 exits the cold box and is pumped to the desired pressure before being heated to 27 °C (80 °F) and fed to the autothermal reformer. N_2 is produced from the cold box at two pressure levels; however, since N_2 is not needed elsewhere in the plant, it is vented to the atmosphere.

3.12.2 Autothermal Reformer

In Case 3, the partially reformed gas from the prereformer enters an autothermal reformer. The ATR is a refractory lined vessel consisting of a partial oxidation section and a fixed bed catalytic reforming section. The feed gas is mixed with oxygen from the ASU and partially oxidized via the following reaction:

$$\mathrm{CH_4} + \frac{1}{2}\mathrm{O_2} \leftrightarrow \mathrm{CO} + 2\mathrm{H_2};\ \Delta H^{\circ}{}_{\mathrm{rxn}} = -36.0\ \mathrm{kJ/mol}$$

This exothermic reaction is used to generate the heat needed to drive the following endothermic methane reforming reaction:

$$\mathrm{CH_4} + \mathrm{H_2O} \leftrightarrow \mathrm{CO} + 3\mathrm{H_2};\ \Delta H^{\circ}{}_{\mathrm{rxn}} = 205.8\ \mathrm{kJ/mol}$$

Unlike the steam methane reformer configuration, where reaction heat is generated outside of the catalyst tubes, both the heat-generating reactions and methane reforming reaction occurs within the single autothermal reformer unit. No supplemental natural gas is required in this configuration.

The process employs a molar steam-to-carbon ratio of 1.57 and oxygen-to-carbon ratio of 1.29 entering the reactor. Raw syngas exits the ATR at a temperature of 1,090 °C (2,000 °F) and a pressure of 2.8 MPa (411 psia). The methane conversion of the process is greater than 99 percent.

Leaving the reformer, the syngas passes through a syngas cooler which is used to generate LP steam before being fed into the series of WGS reactors.

3.12.3 Acid Gas Removal

The autothermal reforming plant with capture (Case 3) uses an MDEA unit for pre-combustion capture. There are no differences between the MDEA unit employed from Case 3 and the one from Case 2. In the ATR plant, there is no need for a post-combustion Cansolv system as in steam methane reforming plant because the volumetric flow of low-pressure flue gas emitted from the plant is significantly reduced due to the absence of an externally-heated reactor. This fundamental difference in technologies enables high levels of CO_2 capture exclusively through separation of CO_2 from the high-pressure syngas.

3.12.4 CO_2 Compression and Drying System

No differences from Case 2.

3.12.5 Fired Heater

The autothermal reforming configuration (Case 3) employs a fired heater for heat recovery. The off-gas from the PSA is combined with air and combusted in the fired heater to take advantage of the waste heat available. The hot flue gas is cooled by providing heat to various sections of the plant, improving plant efficiency. This section generates steam for the reforming reactions, provides feedwater preheating, and provides preheating of the natural gas feed prior to the sulfur guard and prereformer.

3.13 Case 3 – Performance Results

Overall performance for the entire plant is summarized in Exhibit 3-38. The plant produces 27,500 kg/hr (60,627 lb/hr) of hydrogen at an effective thermal efficiency of 67.9 percent (HHV basis). The total auxiliary load for the plant is 110 MWe.

Exhibit 3-39 provides a detailed breakdown of the auxiliary power requirements. The hydrogen compressor accounts for about 13 percent of the total auxiliary load. About 25 percent of the auxiliary load is from the CO_2 capture, removal, and compression systems. The air separation unit power requirements make up about 51 percent of the auxiliary load. The BFW pumps and cooling water system, including the circulating water pumps and cooling tower fan, accounts for approximately 5 percent of the auxiliary load. All other systems together constitute the remaining 6 percent of the auxiliary load.

Exhibit 3-38. Case 3 plant performance summary

Performance Summary	
Steam Turbine Power, MWe	0
Total Gross Power, MWe	**0**
Air Separation Unit Main Air Compressor, kWe	51,070
Air Separation Unit Booster Compressor, kWe	4,020
CO_2 Capture/Removal Auxiliaries, kWe	6,680
CO_2 Compression, kWe	20,810
Balance of Plant, kWe	27,600
Total Auxiliaries, MWe	**110**
Net Power, MWe	**-110**
Hydrogen Production, kg/hr (lb/hr)	**27,500 (60,627)**
CO_2 Capture, %	**94.5%**
HHV Effective Thermal Efficiency[A], %	67.9%
HHV Cold Gas Efficiency[B], %	75.7%
LHV Effective Thermal Efficiency[A], %	62.3%
LHV Cold Gas Efficiency[B], %	70.9%
AGR Cooling Duty, GJ/hr (MMBtu/hr)	1,044 (989)
Natural Gas Feed Flow, kg/hr (lb/hr)	96,930 (213,694)
HHV Thermal Input, kWt	1,408,040
LHV Thermal Input, kWt	1,270,895
Raw Water Withdrawal, m^3/min (gpm)	14.1 (3,720)
Raw Water Consumption, m^3/min (gpm)	11.1 (2,945)

[A]ETE = (Hydrogen Heating Value + Net Power) / Fuel Heating Value
[B]CGE = Hydrogen Heating Value / Fuel Heating Value

Exhibit 3-39. Case 3 plant power summary

Power Summary	
Steam Turbine Power, MWe	0
Total Gross Power, MWe	**0**
Auxiliary Load Summary	
Air Blower, kWe	390
Air Separation Unit Auxiliaries, kWe	780
Air Separation Unit Main Air Compressor, kWe	51,070
Air Separation Unit Booster Compressor, kWe	4,020
Circulating Water Pumps, kWe	3,310
Cooling Tower Fans, kWe	1,710
CO_2 Capture/Removal Auxiliaries, kWe	6,680
CO_2 Compression, kWe	20,810
Feedwater Pumps, kWe	120
Fuel Gas Recycle Compressor, kWe	3,530
Ground Water Pumps, kWe	340
Hydrogen Compressor, kWe	14,070
Miscellaneous Balance of Plant[A], kWe	2,550
Oxygen Pump, kWe	150
Transformer Losses, kWe	650
Total Auxiliaries, MWe	**110**
Net Power, MWe	**-110**

[A]Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

3.13.1 Environmental Performance

The environmental targets for emissions of Hg, NOx, SO_2, and PM were presented in Section 2.6. A summary of the plant air emissions for Case 3 is presented in Exhibit 3-40.

Exhibit 3-40. Case 3 air emissions

	kg/GJ (lb/MMBtu)	Tonne/year (ton/year)[A]	lb/lb H_2
SO_2	0.000 (0.000)	0 (0)	0.000
NOx	0.000 (0.000)	8 (9)	0.000
Particulate	0.000 (0.000)	0 (0)	0.000
Hg	0.00E+0 (0.00E+0)	0.000 (0.000)	0.00E+0
CO	0.000 (0.000)	0 (0)	0.000
CO_2[B]	3 (7)	112,140 (123,614)	0.5
CO_2e[C]	35 (82)	1,404,284 (1,547,959)	5.6

[A]Calculations based on a 90 percent capacity factor
[B]Plant stack emissions
[C]Life cycle emissions

The natural gas feed contains 5.74 ppmv of the mercaptan additive methanethiol (CH_4S). A zinc oxide sulfur guard bed was used to remove 98 percent of the sulfur in the feed, resulting in a maximum sulfur content of 0.1 ppm post-sulfur guard bed.

NOx emissions are limited by the use of low-NOx burners in the design. The production of PM is a result of system inefficiencies and is not produced or emitted in any significant amount. The pipeline natural gas was assumed to contain no Hg or HCl, resulting in zero emissions.

Ninety five percent of the CO_2 from the syngas is captured in the pre-combustion AGR system and compressed for storage. Additionally, an LCA was conducted to determine the total GWP of the plant.

The carbon balance for the plant is shown in Exhibit 3-41. The carbon input to the plant consists of carbon in the air in addition to carbon in the natural gas. Carbon leaves the plant as a small portion of the hydrogen product, the captured CO_2 product, the carbon in the CO_2 dryer vent and knockout, the carbon in the ASU vent, and the CO_2 in the stack gas. The overall carbon capture efficiency is 94.5 percent, defined as one minus the mass of carbon in the stack gas and CO_2 dryer vent relative to the total carbon, represented by the following equation:

$$Carbon\ Capture\ Efficiency\ (\%) = \left(1 - \left(\frac{Carbon\ in\ Stack\ and\ CO_2\ Dryer\ Vent}{Total\ Carbon\ In}\right)\right) * 100\%$$

Exhibit 3-41. Case 3 carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Natural Gas	70,011 (154,347)	Stack Gas	3,874 (8,541)
Air (CO_2)	91 (201)	ASU Vent	72 (159)
		Hydrogen Product	3 (6)
		Sulfur Guard Bed	0 (1)
		CO_2 Product	66,144 (145,823)
		CO_2 Dryer Vent	8.0 (18)
		CO_2 Knockout	0.2 (0.3)
Total	**70,102 (154,548)**	**Total**	**70,102 (154,548)**

Exhibit 3-42 shows the sulfur balance for the plant. Sulfur input comes solely from the mercaptan additive in the natural gas. Sulfur output includes sulfur recovered in the sulfur guard bed and small amounts of sulfur in the stack gas.

Exhibit 3-42. Case 3 sulfur balance

Sulfur In		Sulfur Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Natural Gas	1 (2)	Stack Gas	0 (0)
		Sulfur Guard Bed	1 (2)
Total	**1 (2)**	**Total**	**1 (2)**

Exhibit 3-43 shows the overall water balance for the plant.

Exhibit 3-43. Case 3 water balance

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)
CO_2 Drying	–	–	–	0.0 (1.1)	0.0 (-1.1)
CO_2 Capture System Makeup	–	–	–	–	–
CO_2 Capture Recovery	–	–	–	–	–
CO_2 Compression Recovery	–	–	–	0.0 (7.2)	0.0 (-7.2)
Steam Generation	**2.8 (727)**	**1.6 (412)**	**1.2 (315)**	**–**	**1.2 (315)**
Export Steam	–	–	–	–	–
ATR Steam	2.8 (727)	–	2.8 (727)	–	2.8 (727)
Syngas Condensate	–	1.6 (412)	-1.6 (-412)	–	-1.6 (-412)
Cooling Tower	**13 (3,409)**	**0.0 (4.6)**	**13 (3,404)**	**2.9 (767)**	**10.0 (2,638)**
ASU Knockout	–	0.0 (4.6)	0.0 (-4.6)	–	0.0 (-4.6)
Total	**16 (4,136)**	**1.6 (416)**	**14 (3,720)**	**2.9 (775)**	**11 (2,945)**

An overall plant energy balance is provided in tabular form in Exhibit 3-44.

Exhibit 3-44. Case 3 overall energy balance (0 °C [32 °F] reference)

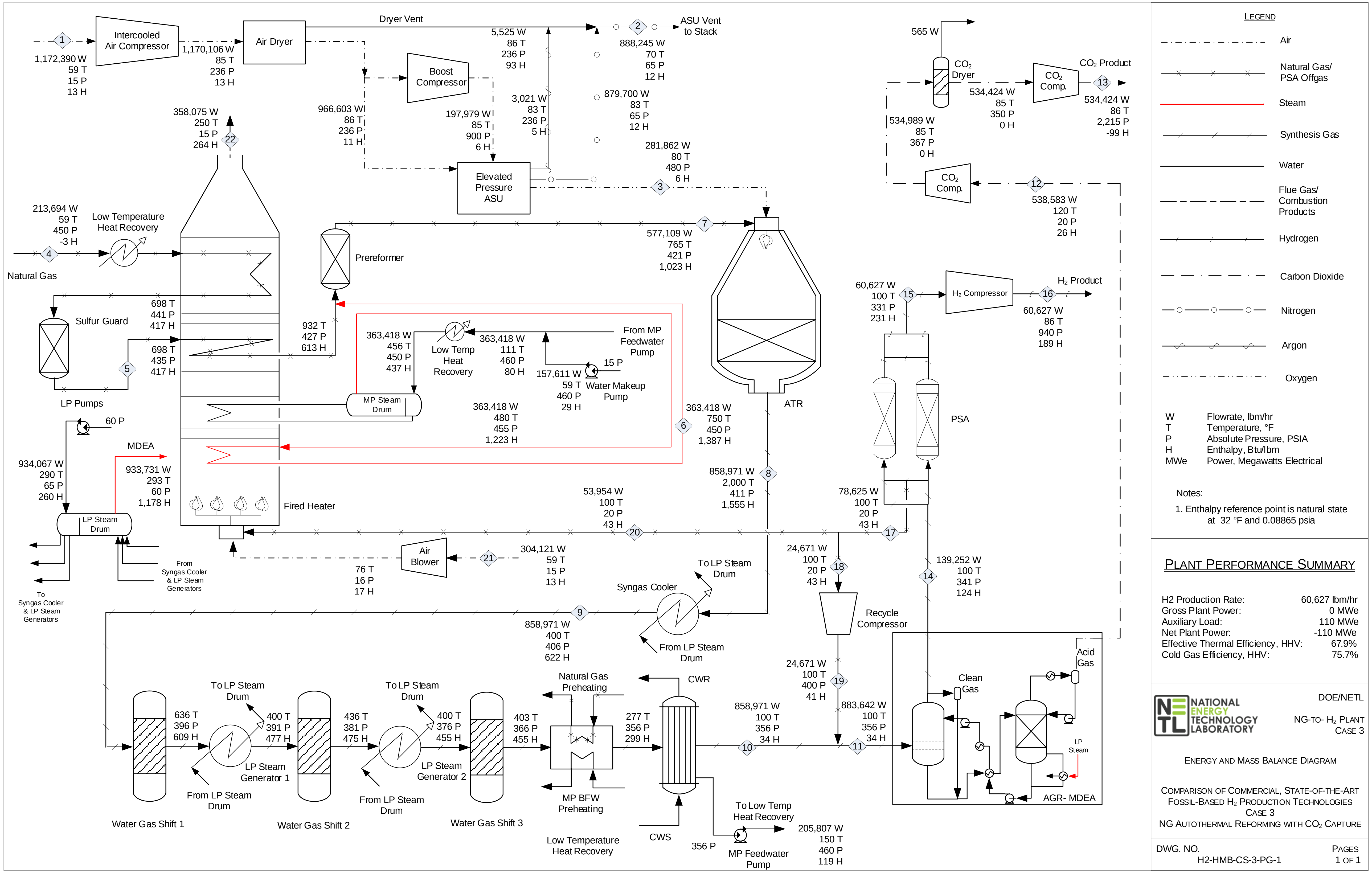
	HHV	Sensible + Latent	Power	Total
Heat In, GJ/hr (MMBtu/hr)				
Natural Gas	5,069 (4,804)	3.4 (3.2)	–	5,072 (4,808)
Air	–	20 (19)	–	20 (19)
Raw Water Makeup	–	53 (50)	–	53 (50)
Auxiliary Power	–	–	397 (376)	397 (376)
Total	**5,069 (4,804)**	**77 (73)**	**397 (376)**	**5,542 (5,253)**
Heat Out, GJ/hr (MMBtu/hr)				
Hydrogen Product	3,836 (3,636)	12 (11)	–	3,848 (3,647)
Stack Gas	–	111 (105)	–	111 (105)
Sulfur	–	0.0 (0.0)	–	0.0 (0.0)
Motor Losses and Design Allowances	–	–	12 (11)	12 (11)
Cooling Tower Load[A]	–	1,685 (1,597)	–	1,685 (1,597)
Cooling Tower Blowdown	–	22 (20)	–	22 (20)
CO_2 Product Stream	–	-56 (-53)	–	-56 (-53)
Power	–	–	0.0 (0.0)	0.0 (0.0)
Total	**3,836 (3,636)**	**1,774 (1,681)**	**12 (11)**	**5,622 (5,329)**
Unaccounted Energy[B]	–	-80 (-76)	–	-80 (-76)

[A]Includes AGR, compression, and last stage syngas cooling loads
[B]By difference

3.13.2 Energy and Mass Balance Diagrams

The energy and mass balance diagram for Case 3 is shown in Exhibit 3-45.

Exhibit 3-45. Case 3 autothermal reforming plant with CO_2 capture energy and mass balance



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3.14 Case 3 – Major Equipment List

Major equipment items for the autothermal reforming plant with CO_2 capture are shown in the following tables. In general, the design conditions include a 10 percent design allowance for flows and heat duties and a 21 percent design allowance for heads on pumps and fans.

Case 3 – Account 3: Feedwater and Miscellaneous Balance of Plant Systems

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	LP Feedwater Pumps	Horizontal centrifugal	3,910 lpm @ 4 m H_2O (1,030 gpm @ 10 ft H_2O)	2	1
2	MP Feedwater Pumps	Horizontal, split case, multi-stage, centrifugal	860 lpm @ 90 m H_2O (230 gpm @ 290 ft H_2O)	2	1
3	Water Makeup Pumps	Horizontal, split case, multi-stage, centrifugal	660 lpm @ 90 m H_2O (170 gpm @ 290 ft H_2O)	2	2
4	Auxiliary Boiler	Shop fabricated, water tube	18,000 kg/hr, 2.8 MPa, 343 °C (40,000 lb/hr, 400 psig, 650 °F)	1	0
5	Service Air Compressors	Flooded Screw	13 m^3/min @ 0.7 MPa (450 scfm @ 100 psig)	2	1
6	Instrument Air Dryers	Duplex, regenerative	13 m^3/min (450 scfm)	2	1
7	Engine-Driven Fire Pump	Vertical turbine, diesel engine	3,785 lpm @ 110 m H_2O (1,000 gpm @ 350 ft H_2O)	1	1
8	Fire Service Booster Pump	Two-stage horizontal centrifugal	2,650 lpm @ 80 m H_2O (700 gpm @ 250 ft H_2O)	1	1
9	Raw Water Pumps	Stainless steel, single suction	6,600 lpm @ 20 m H_2O (1,700 gpm @ 60 ft H_2O)	2	1
10	Filtered Water Pumps	Stainless steel, single suction	150 lpm @ 50 m H_2O (40 gpm @ 160 ft H_2O)	2	1
11	Filtered Water Tank	Vertical, cylindrical	145,000 liter (38,000 gal)	1	0
12	Makeup Water Demineralizer	Multi-media filter, cartridge filter, RO membrane assembly and electro-deionization unit	330 lpm (90 gpm)	1	0
13	Liquid Waste Treatment System	–	10 years, 24-hour storm	1	0
14	Gas Pipeline	Underground, coated carbon steel, wrapped cathodic protection	73 m^3/min @ 3.1 MPa (2,570 acfm @ 450 psia) 39 cm (16 in) standard wall pipe	16 km (10 mile)	0
15	Gas Metering Station	–	73 m^3/min (2,570 acfm)	1	0

Case 3 – Account 4: Reformer and Accessories

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Primary Reformer	ATR, fixed bed, catalytic	Syngas Production: 429,000 kg/hr @ 2.8 MPa, 1093 °C (945,000 lb/hr @ 411 psia, 2000 °F)	1	0
2	Sulfur Guard Bed	Fixed Bed, catalytic (ZnO)	Inlet: 107,000 kg/hr @ 3.0 MPa, 370 °C (235,000 lb/hr @ 441 psia, 698 °F)	1	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
3	Prereformer	Fixed Bed, catalytic	NG In: 107,000 kg/hr @ 2.9 MPa, 500 °C (235,000 lb/hr @ 427 psia, 932 °F) Steam In: 181,000 kg/hr @ 3.1 MPa, 399 °C (400,000 lb/hr @ 450 psia, 750 °F)	1	0
4	Syngas Coolers	Shell and tube heat exchanger	Syngas Cooler: 845 GJ/hr (802 MMBtu/hr) LP Steam Generator 1: 120 GJ/hr (114 MMBtu/hr) LP Steam Generator 2: 18 GJ/hr (17 MMBtu/hr) LT Heat Recovery Exchanger: 141 GJ/hr (134 MMBtu/hr) AGR Precooler: 240 GJ/hr (228 MMBtu/hr)	5	0
5	ASU Main Air Compressor	Centrifugal, multi-stage	8,000 m^3/min @ 1.6 MPa (283,000 scfm @ 236 psia)	1	0
6	Cold Box	Vendor design	3,400 tonne/day (3,700 tpd) of 95% purity oxygen	1	0
7	Oxygen Pump	Centrifugal, multi-stage	2,000 m^3/min (61,000 scfm) Suction – 1.0 MPa (130 psia) Discharge – 3.3 MPa (480 psia)	1	0

Case 3 – Account 5: Flue Gas Cleanup

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	CO_2 Dryer	Triethylene glycol	Inlet: 79 m^3/min @ 2.5 MPa (2,799 acfm @ 367 psia) Outlet: 2.4 MPa (350 psia) Water Recovered: 256 kg/hr (565 lb/hr)	1	0
2	CO_2 Compressor	Integrally geared, multi-stage centrifugal	9.0 m^3/min @ 15.3 MPa, 85 °C (333 acfm @ 2,217 psia, 185 °F)	1	0
3	CO_2 Aftercooler	Shell and tube heat exchanger	Outlet: 15.3 MPa, 30 °C (2,215psia, 86 °F) Duty: 40 GJ/hr (38 MMBtu/hr)	1	0

Case 3 – Account 6: Syngas Cleanup

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Recycle Compressor	Centrifugal	12,000 kg/hr (27,000 lb/hr) 2.8 MPa (400 psia)	1	1
2	Water Gas Shift Reactors	Fixed bed, catalytic	143,000 kg/hr (472,000 lb/hr) 204 °C (400 °F) 2.8 MPa (410 psia)	6	0
3	Acid Gas Removal Plant	MDEA	441,000 kg/hr (972,000 lb/hr) 38 °C (100 °F) 2.5 MPa (356 psia)	1	0

Case 3 – Account 7: Hydrogen Production

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Pressure Swing Adsorber	Polybed Proprietary	Syngas: 63,164 kg/hr (139,252 lb/hr) 38 °C (100 °F) 2.4 MPa (341.0 psia) Hydrogen: 27,500 kg/hr (60,627 lb/hr) 38 °C (100 °F) 2.3 MPa (331.0 psia) Off Gas: 35,664 kg/hr (78,625 lb/hr) 38 °C (100 °F) 0.1 MPa (20.0 psia)	1	0
2	Hydrogen Compressor	Integrally geared, multi-stage centrifugal	99 m^3/min @ 6.5 MPa (3,490 acfm @ 940 psia)	1	1

Case 3 – Account 8: Fired Heater, Ductwork & Stack

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Fired Heater	Bottom-Fired, PSA off-gas	Fired Duty: 510 GJ/hr, (484 MMBtu/hr)	1	0

Case 3 – Account 9: Cooling Water System

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Circulating Water Pumps	Vertical, wet pit	333,000 lpm @ 30 m (88,000 gpm @ 100 ft)	2	1
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	11 °C (52 °F) wet bulb/ 16 °C (60 °F) CWT/27 °C (80 °F) HWT/ 1850 GJ/hr (1760 MMBtu/hr) heat duty	1	0

Case 3 – Account 11: Accessory Electric Plant

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	High Voltage Auxiliary Transformer	Oil-filled	345 kV/13.8 kV, 12 MVA, 3-ph, 60 Hz	2	0
2	Medium Voltage Transformer	Oil-filled	18 kV/4.16 kV, 80 MVA, 3-ph, 60 Hz	1	1
3	Low Voltage Transformer	Dry ventilated	4.16 kV/480 V, 18 MVA, 3-ph, 60 Hz	1	1
4	Medium Voltage Switchgear	Metal clad	4.16 kV, 3-ph, 60 Hz	1	1
5	Low Voltage Switchgear	Metal enclosed	480 V, 3-ph, 60 Hz	1	1
6	Emergency Diesel Generator	Sized for emergency shutdown	750 kW, 480 V, 3-ph, 60 Hz	1	0

Case 3 – Account 12: Instrumentation and Control

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	DCS – Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W)	Operator stations/printers and engineering stations/printers	1	0
2	DCS – Processor	Microprocessor with redundant input/output	N/A	1	0
3	DCS – Data Highway	Fiber optic	Fully redundant, 25% spare	1	0

3.15 Case 3 – Cost Estimating

The cost estimating methodology was described previously in Section 2.11. Exhibit 3-46 shows a detailed breakdown of the capital costs; Exhibit 3-47 shows the owner's costs, TOC, and TASC; Exhibit 3-48 shows the initial and annual O&M costs; and Exhibit 3-49 shows the LCOH breakdown.

The estimated TOC of the autothermal reforming plant with CO_2 capture is $1,372/[kg H_2/day]. Process contingency represents less than 1 percent of the TOC and project contingency represents 8.3 percent. The LCOH is $1.59/kg H_2 with CO_2 T&S.

Exhibit 3-46. Case 3 total plant cost details

	Case:	3	– Autothermal Reforming w/ CCS						Estimate Type:	Conceptual	
	Plant Size (kg H_2/day):	659,999							Cost Base:	Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/[kg H_2/day]
3						Feedwater & Miscellaneous BOP Systems					
3.1	Feedwater System	$82	$311	$476	$372	$1,241	$248	$0	$223	$1,713	$3
3.2	Water Makeup & Pretreating	$4,458	$446	$2,526	$0	$7,429	$1,486	$0	$1,783	$10,698	$16
3.3	Other Feedwater Subsystems	$1,108	$363	$345	$0	$1,816	$363	$0	$327	$2,507	$4
3.4	Service Water Systems	$1,353	$2,582	$8,362	$0	$12,297	$2,459	$0	$2,951	$17,708	$27
3.5	Other Boiler Plant Systems	$230	$84	$209	$0	$523	$105	$0	$94	$722	$1
3.6	Natural Gas Pipeline and Start-Up System	$9,304	$400	$300	$0	$10,005	$2,001	$0	$1,801	$13,807	$21
3.7	Wastewater Treatment Equipment	$5,620	$0	$3,444	$0	$9,064	$1,813	$0	$2,175	$13,052	$20
3.9	Miscellaneous Plant Equipment	$14,217	$1,865	$7,225	$0	$23,306	$4,661	$0	$5,594	$33,561	$51
	Subtotal	**$36,372**	**$6,050**	**$22,888**	**$372**	**$65,682**	**$13,136**	**$0**	**$14,948**	**$93,766**	**$142**
4						Reformer & Accessories					
4.1	Primary Reformers	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.2	Air Blower	$162	$115	$160	$125	$562	$112	$0	$101	$776	$1
4.3	Sulfur Guard Bed	$271	$317	$460	$359	$1,407	$281	$0	$253	$1,941	$3
4.4	Prereformer	$1,511	$359	$549	$428	$2,847	$569	$0	$854	$4,271	$6
4.5	Syngas Coolers	$10,541	$586	$766	$597	$12,491	$2,498	$0	$3,747	$18,736	$28
4.6	Air Separation Unit/Oxidant Compression	$153,064	$255	$39,871	$31,097	$224,287	$44,857	$0	$0	$269,144	$408
4.7	Steam Drum(s)	$953	$428	$696	$543	$2,619	$524	$0	$471	$3,614	$5
4.11	Autothermal Reformer	$4,777	$500	$1,017	$793	$7,086	$1,417	$0	$2,126	$10,629	$16
	Subtotal	**$171,278**	**$2,560**	**$43,519**	**$33,942**	**$251,299**	**$50,260**	**$0**	**$7,553**	**$309,112**	**$468**
5						Flue Gas Cleanup					
5.1	Cansolv Carbon Dioxide (CO_2) Removal System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.4	Carbon Dioxide (CO_2) Compression & Drying	$43,131	$1,481	$3,681	$2,871	$51,165	$10,233	$0	$15,349	$76,747	$116
5.5	Carbon Dioxide (CO_2) Compressor Aftercooler	$827	$571	$1,397	$1,090	$3,884	$777	$0	$699	$5,360	$8
5.12	Gas Cleanup Foundations	w/ 5.4 & 5	w/ 5.4 & 5	w/ 5.4 & 5	w/ 5.4 & 5	$0	$0	$0	$0	$0	$0
	Subtotal	**$43,958**	**$2,052**	**$5,078**	**$3,961**	**$55,049**	**$11,010**	**$0**	**$16,049**	**$82,108**	**$124**

	Case:	3	– Autothermal Reforming w/ CCS					Estimate Type:		Conceptual	
	Plant Size (kg H_2/day):	659,999						Cost Base:		Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/[kg H_2/day]
6						Syngas Cleanup					
6.1	Recycle Compressor	$7,936	$189	$450	$351	$8,925	$1,785	$0	$2,678	$13,388	$20
6.2	Methyl Diethanolamine (MDEA) – Low Temperature Acid Gas Removal	$23,443	$0	$674	$526	$24,642	$4,928	$0	$5,914	$35,485	$54
6.7	Water Gas Shift (WGS) Reactors	$2,187	$2,070	$3,155	$2,460	$9,872	$1,974	$0	$2,369	$14,216	$22
6.12	Gas Cleanup Foundations	w/ 6.1,2,&7	w/6.1,2,&7	w/ 6.1,2,&7	w/ 6.1,2,&7	$0	$0	$0	$0	$0	$0
	Subtotal	**$33,566**	**$2,259**	**$4,278**	**$3,337**	**$43,440**	**$8,688**	**$0**	**$10,961**	**$63,088**	**$96**
7						Hydrogen Production					
7.1	Pressure Swing Adsorber	$20,140	$23	$898	$700	$21,762	$4,352	$0	$5,223	$31,337	$47
7.2	Hydrogen Compressor	$6,915	$618	$718	$560	$8,812	$1,762	$0	$2,115	$12,689	$19
	Subtotal	**$27,055**	**$641**	**$1,616**	**$1,261**	**$30,574**	**$6,115**	**$0**	**$7,338**	**$44,026**	**$67**
8						Fired Heater, Ductwork, & Stack					
8.1	Fired Heater	$6,675	$376	$1,899	$1,481	$10,431	$2,086	$0	$2,504	$15,021	$23
8.2	Fired Heater Accessories	w/8.1	w/8.1	w/8.1	w/8.1	$0	$0	$0	$0	$0	$0
8.3	Ductwork	w/8.1	w/8.1	w/8.1	w/8.1	$0	$0	$0	$0	$0	$0
8.4	Stack	w/8.1	w/8.1	w/8.1	w/8.1	$0	$0	$0	$0	$0	$0
8.5	Fired Heater, Ductwork & Stack Foundations	w/8.1	w/8.1	w/8.1	w/8.1	$0	$0	$0	$0	$0	$0
	Subtotal	**$6,675**	**$376**	**$1,899**	**$1,481**	**$10,431**	**$2,086**	**$0**	**$2,504**	**$15,021**	**$23**
9						Cooling Water System					
9.1	Cooling Towers	$9,138	$0	$2,767	$0	$11,906	$2,381	$0	$2,143	$16,430	$25
9.2	Circulating Water Pumps	$1,207	$0	$74	$0	$1,281	$256	$0	$231	$1,768	$3
9.3	Circulating Water System Auxiliaries	$8,860	$0	$1,169	$0	$10,029	$2,006	$0	$1,805	$13,840	$21
9.4	Circulating Water Piping	$0	$2,550	$2,309	$0	$4,858	$972	$0	$875	$6,705	$10
9.5	Make-up Water System	$333	$0	$428	$0	$760	$152	$0	$137	$1,049	$2
9.6	Component Cooling Water System	$376	$0	$288	$0	$664	$133	$0	$120	$917	$1
9.7	Circulating Water System Foundations	$0	$590	$979	$0	$1,569	$314	$0	$377	$2,260	$3
	Subtotal	**$19,914**	**$3,139**	**$8,014**	**$0**	**$31,068**	**$6,214**	**$0**	**$5,686**	**$42,968**	**$65**
11						Accessory Electric Plant					
11.2	Station Service Equipment	$0	$1,128	$109	$85	$1,321	$264	$0	$238	$1,823	$3
11.3	Switchgear & Motor Control	$0	$602	$188	$147	$937	$187	$0	$169	$1,293	$2
11.4	Conduit & Cable Tray	$0	$281	$508	$397	$1,186	$237	$0	$213	$1,637	$2
11.5	Wire & Cable	$0	$662	$467	$364	$1,493	$299	$0	$269	$2,060	$3
11.6	Protective Equipment	$0	$443	$785	$612	$1,840	$368	$0	$331	$2,539	$4
11.7	Standby Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0

	Case:	3	– Autothermal Reforming w/ CCS				Estimate Type:			Conceptual	
	Plant Size (kg H_2/day):	659,999					Cost Base:			Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/[kg H_2/day]
11.8	Main Power Transformers	$0	$151	$71	$56	$278	$56	$0	$50	$384	$1
11.9	Electrical Foundations	$0	$7	$22	$17	$46	$9	$0	$11	$66	$0
	Subtotal	**$0**	**$3,275**	**$2,149**	**$1,676**	**$7,101**	**$1,420**	**$0**	**$1,281**	**$9,802**	**$15**
12	**Instrumentation & Control**										
12.4	Other Major Component Control Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
12.5	Signal Processing Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
12.6	Control Boards, Panels & Racks	$0	$340	$40	$31	$411	$82	$21	$77	$591	$1
12.7	Distributed Control System Equipment	$9,148	$0	$280	$0	$9,428	$1,886	$471	$1,768	$13,552	$21
12.8	Instrument Wiring & Tubing	$0	$238	$408	$318	$965	$193	$0	$174	$1,332	$2
12.9	Other Instrumentation & Controls Equipment	$0	$1,474	$386	$301	$2,160	$432	$0	$389	$2,981	$5
	Subtotal	**$9,148**	**$2,052**	**$1,114**	**$650**	**$12,964**	**$2,593**	**$492**	**$2,407**	**$18,456**	**$28**
13	**Improvements to Site**										
13.1	Site Preparation	$0	$533	$11,312	$0	$11,845	$2,369	$0	$2,843	$17,056	$26
13.2	Site Improvements	$0	$290	$1,163	$907	$2,361	$472	$0	$567	$3,399	$5
13.3	Site Facilities	$1,645	$0	$1,725	$0	$3,370	$674	$0	$809	$4,853	$7
	Subtotal	**$1,645**	**$823**	**$14,201**	**$907**	**$17,575**	**$3,515**	**$0**	**$4,218**	**$25,308**	**$38**
14	**Buildings & Structures**										
14.2	Interconnecting Pipe	$0	$736	$1,744	$1,360	$3,840	$768	$0	$691	$5,299	$8
14.4	Administration Building	$0	$364	$246	$0	$610	$122	$0	$110	$842	$1
14.5	Circulation Water Pumphouse	$0	$64	$31	$0	$95	$19	$0	$17	$131	$0
14.6	Water Treatment Buildings	$0	$365	$333	$0	$698	$140	$0	$126	$963	$1
14.7	Machine Shop	$0	$544	$348	$0	$892	$178	$0	$161	$1,232	$2
14.8	Warehouse	$0	$424	$255	$0	$679	$136	$0	$122	$937	$1
14.9	Structural Steel Pipe Racks	$0	$1,726	$982	$766	$3,474	$695	$0	$625	$4,795	$7
14.10	Waste Treating Building & Structures	$0	$714	$1,274	$0	$1,988	$398	$0	$358	$2,743	$4
	Subtotal	**$0**	**$4,937**	**$5,214**	**$2,126**	**$12,277**	**$2,455**	**$0**	**$2,210**	**$16,942**	**$26**
	Total	**$349,614**	**$28,166**	**$109,998**	**$49,712**	**$537,490**	**$107,498**	**$492**	**$75,162**	**$720,642**	**$1,092**

Exhibit 3-47. Case 3 owner's costs

Description	$/1,000	$/[kg H_2/day]
Pre-Production Costs		
6 Months All Labor	$5,158	$8
1 Month Maintenance Materials	$761	$1
1 Month Non-Fuel Consumables	$6,430	$10
1 Month Waste Disposal	$9	$0
25% of 1 Months Fuel Cost at 100% CF	$3,876	$6
2% of TPC	$14,413	$22
Total	**$30,647**	**$46**
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	$1,017	$2
0.5% of TPC (spare parts)	$3,603	$5
Total	**$4,620**	**$7**
Other Costs		
Initial Cost for Catalyst and Chemicals	$21,445	$32
Land	$300	$0
Other Owner's Costs	$108,096	$164
Financing Costs	$19,457	$29
Total Overnight Costs (TOC)	**$905,208**	**$1,372**
TASC Multiplier (IOU, 35 year)	1.070	
Total As-Spent Cost (TASC)	**$968,393**	**$1,467**

Exhibit 3-48. Case 3 initial and annual operating and maintenance costs

Case:	**3**	**– Autothermal Reforming w/ CCS**			**Cost Base:**	**Dec 2018**
Plant Size (kg H_2/day):	**659,999**				**Capacity Factor (%):**	**90**
Operating & Maintenance Labor						
Operating Labor				**Operating Labor Requirements per Shift**		
Operating Labor Rate (base):		38.50	**$/hour**	Skilled Operator:		1.0
Operating Labor Burden:		30.00	**% of base**	Operator:		3.3
Labor O-H Charge Rate:		25.00	**% of labor**	Foreman:		1.0
				Lab Techs, etc.:		1.0
				Total:		**6.3**
Fixed Operating Costs						
					Annual Cost	
					($)	**($/[kg H_2/day])**
Annual Operating Labor:					$2,776,628	$4.207
Maintenance Labor:					$5,476,882	$8.298
Administrative & Support Labor:					$2,063,377	$3.126
Property Taxes and Insurance:					$14,412,847	$21.838
Total:					**$24,729,734**	**$37.469**
Variable Operating Costs						
					($)	**($/kg H_2)**
Maintenance Material:					$8,215,323	$0.04
Consumables						
	Initial Fill	**Per Day**	**Per Unit**	**Initial Fill**		
Water (gal/1000):	0	2,678	$1.90	$0	$1,671,599	$0.00771
Makeup and Wastewater Treatment Chemicals (ton):	0	8.0	$550	$0	$1,441,422	$0.00665
Zinc Oxide Sulfur Guard Catalyst (ft^3):	1120	0.8	$600	$672,000	$151,200	$0.00070
Prereformer Catalyst (ft^3):	1330	0.9	$1,250	$1,662,500	$374,063	$0.00173
Primary Reformer Catalyst (ft^3):	3600	2.0	$450	$1,620,000	$291,600	$0.00134
Water Gas Shift Catalyst (ft^3):	18809	12.9	$480	$9,028,174	$2,031,339	$0.00937
Methyl Diethanolamine Solution (gal):	847096	319	$2.80	$2,371,868	$293,432	$0.00135
Post-Combustion CO_2 Capture System[A]:	Proprietary				$0	$0.00000
PSA Unit Adsorbent (ft^3):	40600	5.6	$150	$6,090,000	$274,050	$0.00126
Triethylene Glycol (gal):	w/equip.	285	$6.80	$0	$635,685	$0.00293
Electricity (MWh):	0	2644	$71.70	$0	$62,282,859	$0.28727
Subtotal:				**$21,444,542**	**$69,447,248**	**$0.32031**
Waste Disposal						
Zinc Oxide Sulfur Guard Catalyst (ft^3):	0	0.8	$40.00	$0	$10,080	$0.00005
Prereformer Catalyst (ft^3):	0	0.9	$0.00	$0	$0	$0.00000
Primary Reformer Catalyst (ft^3):	0	2.0	$0.00	$0	$0	$0.00000
Water Gas Shift Catalyst (ft^3):	0	12.9	$2.50	$0	$10,580	$0.00005
Methyl Diethanolamine Solution (gal):	0	319	$0.35	$0	$36,679	$0.00017
PSA Unit Adsorbent (ft^3):	0	5.6	$1.50	$0	$2,741	$0.00001
Triethylene Glycol (gal):	0	285	$0.35	$0	$32,719	$0.00015
Amine Purification Unit Waste (ton):	0	0	$38.0	$0	$0	$0.00000
Thermal Reclaimer Unit Waste (ton):	0	0	$38.0	$0	$0	$0.00000
Subtotal:				**$0**	**$92,798**	**$0.00043**

Case:	**3**	**– Autothermal Reforming w/ CCS**			**Cost Base:**	**Dec 2018**
Plant Size (kg H_2/day):	**659,999**				**Capacity Factor (%):**	**90**
By-Products						
					($)	**(\$/kg H_2)**
Steam (ton):	0	0	$0.00	$0	$0	$0.00000
Electricity (MWh):	0	0	$71.70	$0	$0	$0.00000
Subtotal:				**$0**	**$0**	**$0.00000**
Variable Operating Costs Total:				**$21,444,542**	**$77,532,003**	**$0.35760**
Fuel Cost						
Natural Gas (MMBtu):	0	115,306	$4.42	$0	$167,428,981	$0.77224
Total:				**$0**	**$167,428,981**	**$0.77224**

[A]Post-combustion CO_2 capture system chemicals includes ion exchange resin, NaOH, and Cansolv solvent.

Exhibit 3-49. Case 3 LCOH breakdown

Component	Value, \$/kg H_2	Percentage
Capital	0.26	16%
Fixed	0.11	7%
Variable	0.36	22%
Fuel	0.77	48%
Total (Excluding T&S)	**1.51**	**N/A**
CO_2 T&S	0.09	6%
Total (Including T&S)	**1.59**	**N/A**

3.16 Reforming Case Summary

The performance and cost results of the three reforming plant configurations modeled in this report are summarized in Exhibit 3-50. The following section provides a discussion on the reforming performance, environmental, and cost results.

Exhibit 3-50. Estimated performance results for all reforming cases

	Reforming		
	SMR		**ATR**
Case Name	**1**	**2**	**3**
PERFORMANCE			
Nominal CO_2 Capture	0%	96.2%	94.5%
Capacity Factor	90%	90%	90%
Hydrogen Production Rate (lb/hr)	44,369	44,369	60,627
Gross Power Output (MWe)	0	0	0
Auxiliary Power Requirement (MWe)	13	41	110
Net Power Output (MWe)	-13	-41	-110
Coal Flowrate (lb/hr)	N/A	N/A	N/A

	Reforming		
	SMR		ATR
Case Name	**1**	**2**	**3**
Biomass Flowrate (lb/hr)	N/A	N/A	N/A
Natural Gas Flowrate (lb/hr)	156,482	166,387	213,694
HHV Thermal Input (kW_t)	1,031,068	1,096,328	1,408,040
HHV Effective Thermal Efficiency (%)	75.4%	68.4%	67.9%
HHV Cold Gas Efficiency (%)	76.7%	72.1%	75.7%
Raw Water Withdrawal (gpm)	1,457	2,727	3,720
Process Water Discharge (gpm)	63	586	775
Raw Water Consumption (gpm)	1,395	2,140	2,945
CO_2 Emissions (lb/MMBtu)	118	5	7
SO_2 Emissions[A] (lb/MMBtu)	0.000	0.000	0.000
NOx Emissions (lb/MMBtu)	0.003	0.002	0.000
PM Emissions (lb/MMBtu)	0.000	0.000	0.000
Hg Emissions (lb/TBtu)	0.000	0.000	0.000
COST			
Total Plant Cost (2018$/[kg H_2/day])	554	1,394	1,092
Total Overnight Cost (2018$/[kg H_2/day])	713	1,735	1,372
Owner's Costs	159	341	280
Process Contingency	0	52	1
Project Contingency	94	252	114
Home Office Expenses	77	182	163
Bare Erected Cost	384	908	814
Total As-Spent Cost (2018$/[kg H_2/day])	763	1,856	1,467
LCOH ($/kg H_2) (excluding T&S)	1.06	1.54	1.51
Capital Costs	0.14	0.33	0.26
Fixed Costs	0.07	0.15	0.11
Variable Costs	0.09	0.24	0.36
Fuel Costs	0.77	0.82	0.77
LCOH ($/kg H_2) (including T&S)	1.06	1.64	1.59
CO_2 T&S Costs	0.00	0.10	0.09

[A]Trace amounts of sulfur emissions may exist in the flue gas stream to the stack in capture cases

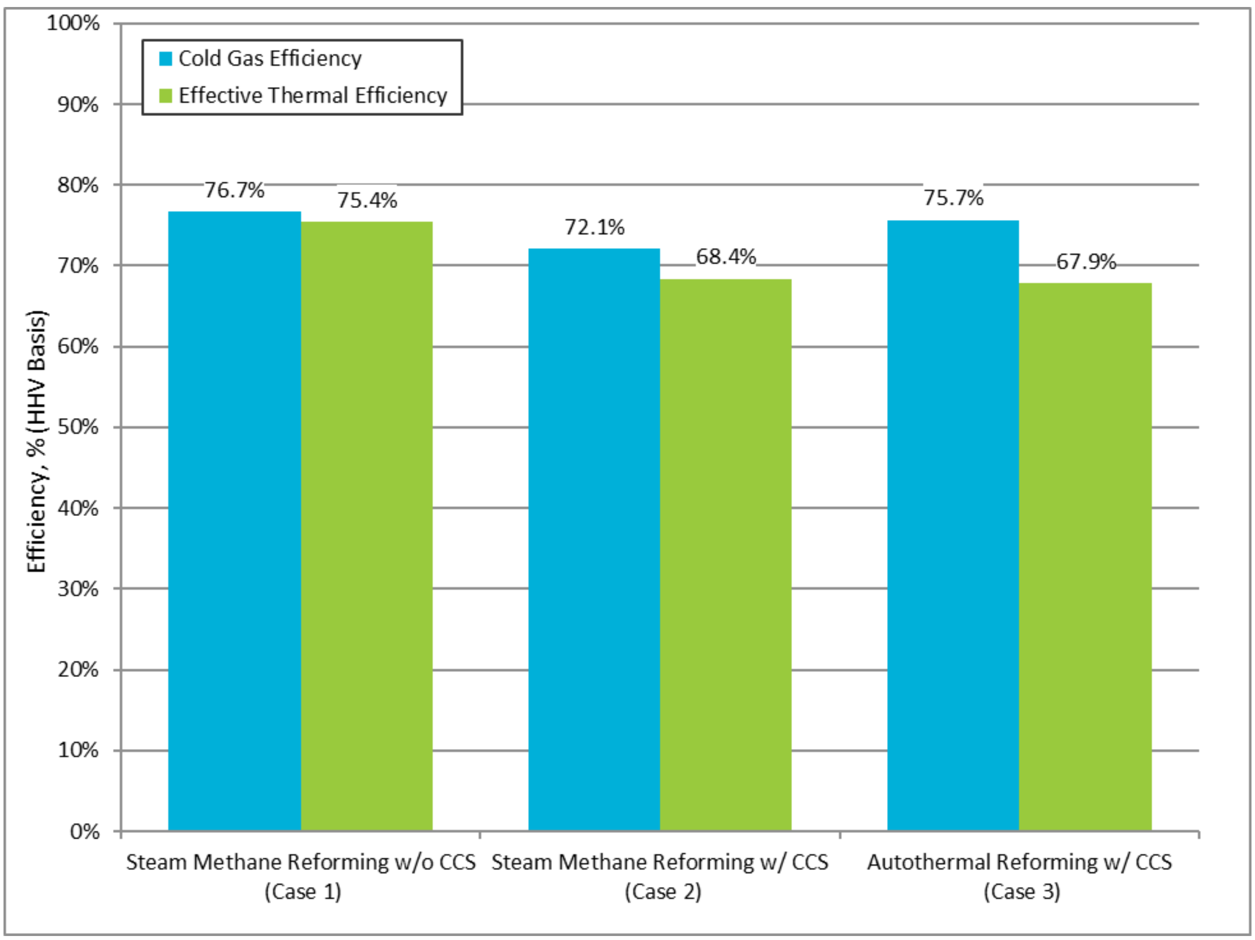
3.16.1 Performance Results Summary

A graph of the net plant efficiency (HHV basis) is provided in Exhibit 3-51.

The following observations can be made regarding plant performance:

- The SMR plant without capture (Case 1) performs the best, with a cold gas efficiency of 76.7 percent on an HHV-basis and an effective thermal efficiency of 75.4 percent on an HHV-basis. The worst performing configuration on a cold gas efficiency basis is the SMR plant with capture (Case 2), having a 72.1 percent efficiency on an HHV-basis. The ATR plant with capture (Case 3) configuration has the lowest effective thermal efficiency metric of 67.9 percent on an HHV-basis.
- The two cases that employ carbon capture, Case 2 and Case 3, have the lowest efficiency values. This is due to the energy penalty associated with adding CO_2 capture as additional energy is needed for the thermal demand and auxiliary load for the CO_2 separation and compression equipment. The reduction in effective thermal efficiency is about 7 percentage points (about 9.3 percent relative to non-capture) between the SMR cases, Case 1 and Case 2.
- The configuration with the highest auxiliary power requirement (110 MWe) is the ATR plant with capture (Case 3) due to the carbon capture energy penalty as well as the addition of the ASU in the plant configuration. The SMR plant without capture (Case 1) has the lowest auxiliary power requirement (13 MWe).

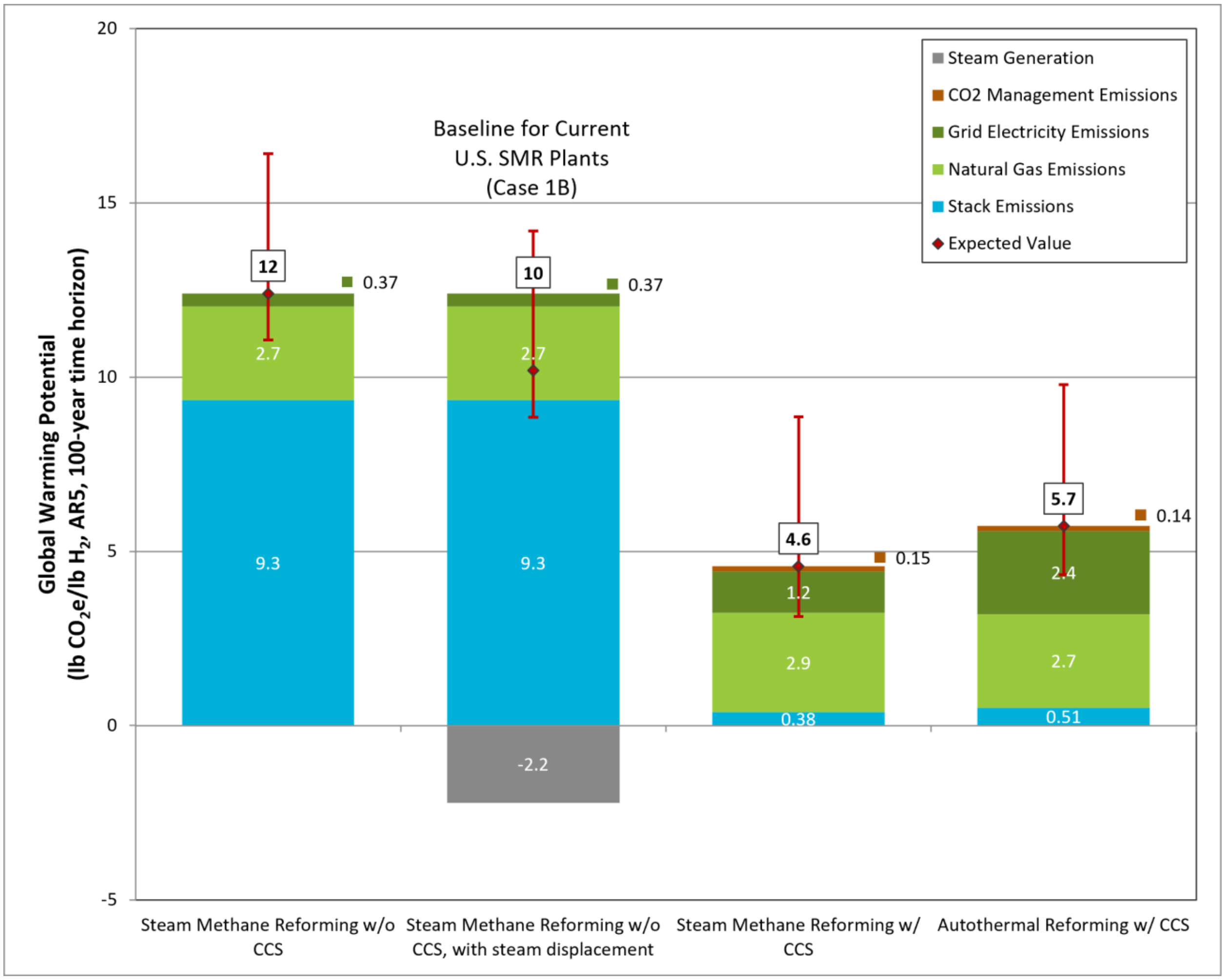
Exhibit 3-51. Net plant efficiency (HHV basis) for all reforming cases



3.16.2 Environmental Results Summary

A graph of the life cycle equivalent CO_2 emissions breakdown is shown in Exhibit 3-52. To generate these results, a Monte Carlo simulation with 5,000 iterations was used. The bars represent median values, and the error bars show results from the 5th and 95th percentiles. For the upstream natural gas contribution, the error bars also represent variability across basins. The case of SMR without CCS includes parallel models without and with steam displacement, labeled Case 1A and Case 1B, respectively. Further discussion of this steam credit is below.

Exhibit 3-52. CO_2e life cycle emissions for all reforming cases



- The highest total equivalent CO_2 life cycle emissions (12 lb CO_2e/lb H_2) comes from the SMR plant without capture and without steam displacement (Case 1A), which is expected. Seventy five percent of the total CO_2e emissions are from CO_2 stack emissions (9.3 lb CO_2e/lb H_2). Twenty two percent comes from natural gas procurement (3.6 lb CO_2e/lb H_2) and 3 percent comes from grid electricity usage (0.37 lb CO_2e/lb H_2). With the inclusion of steam displacement in the SMR plant without capture (Case 1B), the total equivalent CO_2 life cycle emissions reduce to 10 lb CO_2e/lb H_2.

- For the SMR case with CO_2 capture (Case 2), the total life cycle CO_2e emissions are more than 10 times that of the stack emissions alone. CO_2 stack emissions are eight percent of the total CO_2e emissions (0.38 lb CO_2e/lb H_2). Sixty two percent comes from natural gas procurement (2.9 lb CO_2e/lb H_2), 26 percent comes from grid electricity usage (1.2 lb CO_2e/lb H_2), and 3 percent comes from CO_2 management after it has been captured (0.15 lb CO_2e/lb H_2).
- Adding capture to the SMR plant cuts the total CO_2e emissions by more than half, about a 62 percent reduction. Compared to Case 1A, the SMR plant with capture has higher life cycle emissions sourced from grid electricity due to the higher power demand of this case. Similarly, life cycle GHG emissions from carbon management activities are included in the system boundary to account for the transport and geologic sequestration of the captured carbon.
- The ATR with capture case (Case 3) has a total equivalent CO_2 life cycle emissions value of 5.7 lb CO_2e/lb H_2. Nine percent of the total CO_2e emissions are sourced from the CO_2 stack emissions (0.51 lb CO_2e/lb H_2). Fourty seven percent comes from natural gas procurement (2.7 lb CO_2e/lb H_2), 42 percent comes from grid electricity usage (2.4 lb CO_2e/lb H_2), and 2 percent comes from carbon management (0.14 lb CO_2e/lb H_2).

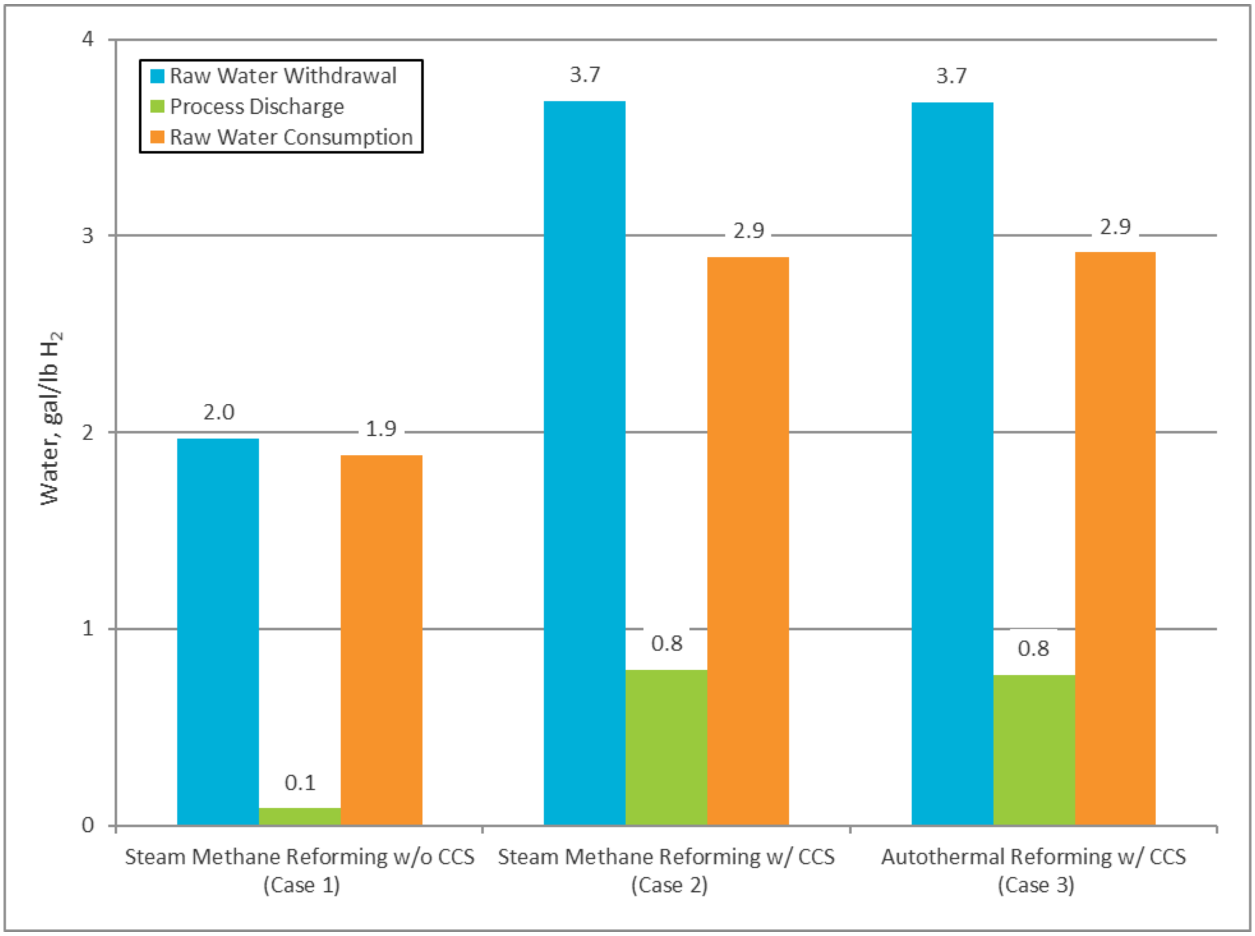
For these cases, the variability in the GWP results is fairly high (about 25%), and is primarily driven by differences in the upstream life cycle emissions from extraction through delivery of the natural gas to the facility, and to a lesser extent, the upstream grid electricity. More details on the sources of variability are in Section 5.1.2.

Hydrogen production today is commonly integrated or co-located with industrial operations that utilize the hydrogen directly within the industrial processes. Steam methane reforming of natural gas to produce hydrogen, the predominant commercial technology for hydrogen production today, often produces excess steam that is commonly used by integrated or co-located industrial operations. Future deployments of hydrogen production facilities may or may not be integrated or co-located with industrial operations with adequate demand for the excess steam produced from various hydrogen production methods. We model two types of SMR configurations without CCS. In Case 1A, the SMR plant configuration is assumed to be a stand-alone hydrogen production facility with no viable use for excess steam produced with the hydrogen product. In Case 1B, SMR hydrogen production is integrated or co-located with an industrial operational that can utilize the excess steam, thus a commensurate emissions credit based on system expansion using a natural gas boiler to produce steam of equivalent quality is used. The emissions credit for excess steam is calculated using the Case 1 modeled steam output of 514 GJ/hr, or 26 MJ/kg H_2, and an assumed 85 percent efficient (HHV basis) natural gas boiler, requiring the combustion of approximately 0.57 kg natural gas/kg H_2 [63]. Overall, the estimated GWP results for Case 1A and 1B are 12 and 10 kg CO_2e/kg H_2, respectively.

Each of the cases satisfies the appropriate hydrogen production facility air emissions targets for SO_2, NOx, particulate matter, and Hg. All three configurations have negligible SO_2, Hg, and PM emissions, on a lb of pollutant per MMBtu thermal input basis. The NOx emissions are highest in the SMR plant without capture (Case 1) configuration (0.003 lb/MMBtu) and lowest in the ATR with capture (Case 3) configuration (0.000 lb/MMBtu).

Finally, water use for the three reforming cases has been summarized in Exhibit 3-53.

Exhibit 3-53. Water use for all reforming cases

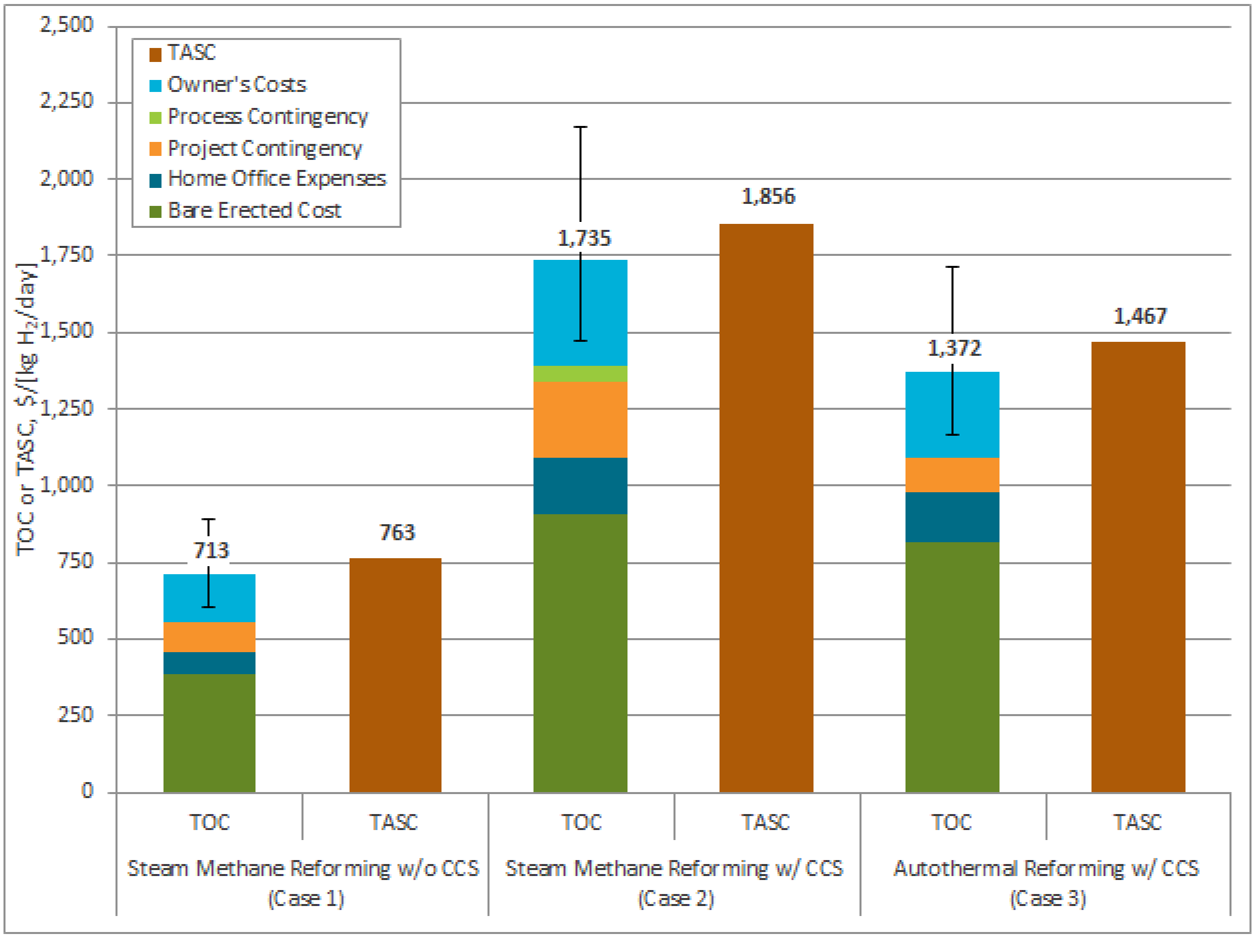


- The results are as expected for the three reforming cases. The SMR plant without capture (Case 1) has the lowest water demands of all three cases, with a raw water consumption of 1.9 gal/lb H_2. In this case, 2.0 gal/lb H_2 is needed for raw water withdrawal and 0.1 gal/lb H_2.
- The SMR (Case 2) and ATR (Case 3) plants with capture have approximately equal water demands, with raw water consumptions for both cases of 2.9 gal/lb H_2, about a 50 percent increase from the non-capture configuration. This is due to the increased cooling demands required by the CO_2 capture system. The raw water withdrawal for this case is 3.7 gal/lb H_2 and the process discharge is 0.8 gal/lb H_2.

3.16.3 Cost Results Summary

The components of TOC and the overall TASC of the three reforming cases are shown in Exhibit 3-54. The error bars included represent the potential TOC range relative to the maximum and minimum of the capital cost uncertainty range which is -15 percent/+25 percent for reforming cases.

Exhibit 3-54. Plant capital cost for all reforming cases



The following TOC observations are made with the caveat that the differences between cases are less than the estimated accuracy. However, all cases are evaluated using a common set of technical and economic assumptions allowing meaningful comparison among the cases:

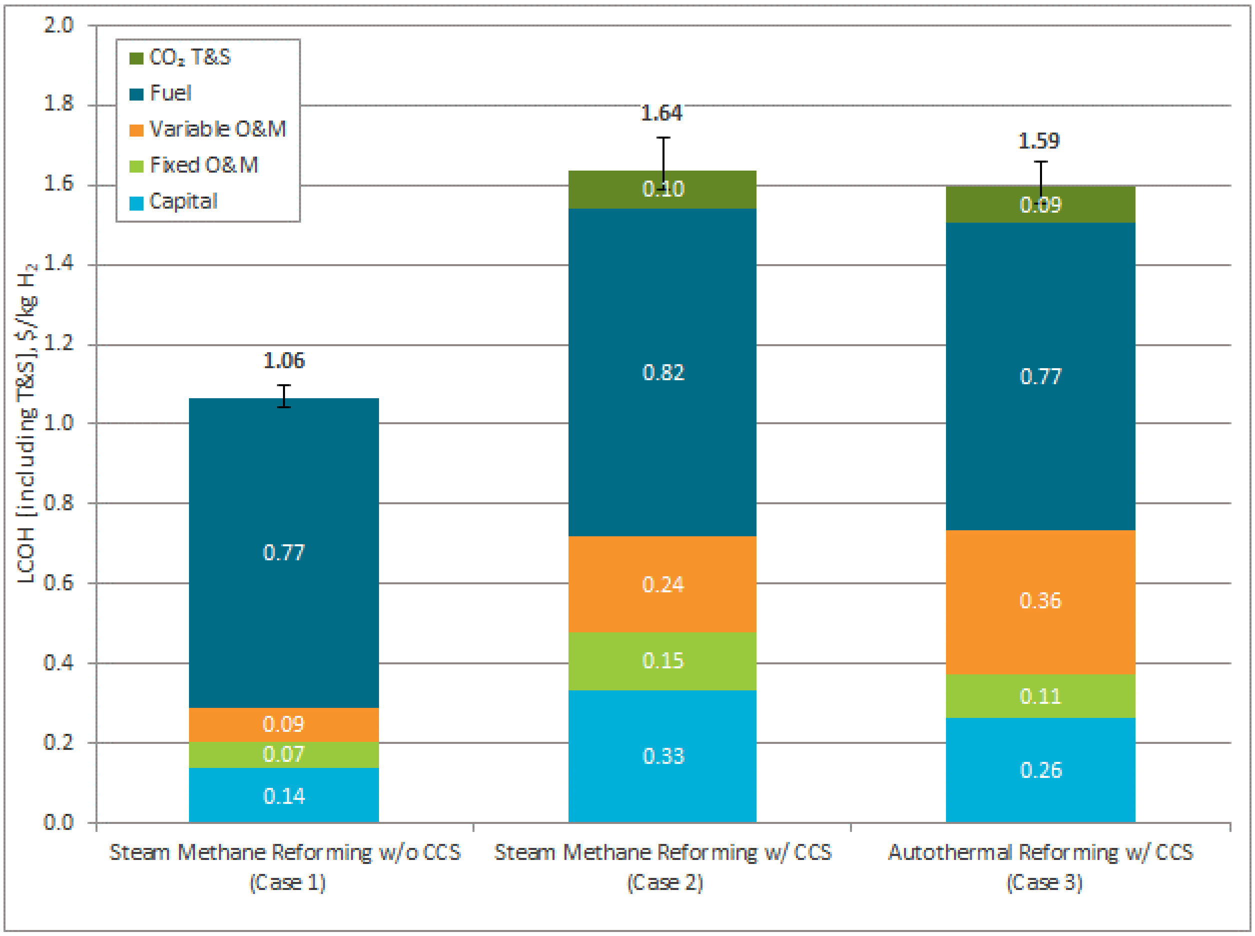
- The SMR plant without capture (Case 1) has the lowest TOC (\$713/[kg H_2/day]) and TASC (\$763/[kg H_2/day]) values of all reforming cases. The primary contributor to the TOC is the bare erected cost, accounting for 54 percent of the TOC for Case 1. The next largest contributor is the owner's costs, accounting for 22 percent. There are no process contingencies in this case.
- The SMR plant with capture (Case 2) has the highest TOC (\$1,735/[kg H_2/day]) and TASC (\$1,856/[kg H_2/day]) values of all reforming cases. A small fraction of the TOC, about

three percent, goes towards process contingency, unlike Case 1. Like Case 1, the bare erected cost accounts for more than half of the TOC, about 52 percent.

- The ATR plant with capture (Case 3) has a TOC of \$1,372/[kg H_2/day] and TASC of \$1,467/[kg H_2/day]. The bare erected cost accounts for 59 percent of the TOC.

The LCOH is shown for all three reforming cases in Exhibit 3-55. The error bars included represent the potential LCOH range relative to the maximum and minimum capital cost uncertainty ranges. The LCOH ranges presented are not reflective of other changes, such as variation in fuel price, labor price, CF, or other factors.

Exhibit 3-55. LCOH for all reforming cases



The following observations can be made:

- The lowest LCOH is from the SMR plant without CCS (Case 1, \$1.06/kg H_2) and the highest LCOH is from the SMR plant with CCS (Case 2, \$1.64/kg H_2). The ATR plant with CCS has a slightly lower LCOH than the SMR plant with CCS (Case 3, \$1.59/kg H_2).
- The largest contributor to the LCOH is the fuel, making up between 48 and 73 percent of the LCOH for the reforming cases.

- In the cases with capture (Case 2 and Case 3), the variable costs make up a larger portion, while the fuel costs make up a smaller portion of the LCOH in comparison to the non-capture case (Case 1). The capital costs make up a larger portion in the SMR plant with capture (Case 2) than the other two reforming cases.

As presented in Section 2.11.4, the breakeven CO_2 sales price and emissions penalty were calculated by using the steam methane reforming plant with capture (Case 2) and the autothermal reforming plant with capture (Case 3). The breakeven CO_2 sales price represents the minimum CO_2 plant gate sales price that will incentivize carbon capture in lieu of a defined reference non-capture plant. The breakeven CO_2 emissions penalty represents the minimum CO_2 emissions price, when applied to both the capture and non-capture plant that will incentivize carbon capture in lieu of a defined reference non-capture plant. Both the breakeven CO_2 sales price and emissions penalty were calculated using the steam methane reforming plant without capture (Case 1) as a non-capture reference.

The breakeven CO_2 sales price (excluding T&S) for the reforming cases with capture are $50.1/tonne ($45.4/ton) for the SMR plant with capture (Case 2) and $50.4/tonne ($45.7/ton) for the ATR plant with capture (Case 3). The breakeven CO_2 sales prices is roughly equivalent between the SMR and ATR plant configurations with CCS. The breakeven CO_2 emissions penalty for the reforming cases with capture are $64.1/tonne ($58.1/ton) for the SMR plant with capture (Case 2) and $60.3/tonne ($54.7/ton) for the ATR plant with capture (Case 3). A lower carbon tax is required to encourage the use of an ATR plant with capture as opposed to an SMR plant with capture, for which the CO_2 emissions penalty needs to be about $4/tonne greater.

GASIFICATION PLANTS

4 Gasification Plants

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

The first two gasification cases, Case 4 and Case 5, use a Shell gasifier and differ in that Case 4 does not include CO_2 capture while Case 5 does. The cases were sized based on the as-received coal feed rate for the Shell IGCC without CO_2 capture (Case B1B) in the NETL Fossil Energy Baseline study. [31] The original size selection was based on the constraints imposed by the fixed combustion turbine (CT) output of a comparable IGCC plant.

Since a coal/biomass co-gasification case was of interest in this study, the design total feedstock flowrate of the Buggenum facility was chosen as a reference facility for Case 6. Owned and operated by Nuon Power, the Buggenum facility was an IGCC plant of commercial size located in the Netherlands. In an effort to reduce CO_2 emissions, up to 30 percent by weight of biomass, including chicken litter, sewage sludge, and milled wood, was co-fed into the system with coal at a total gasifier feed rate of 1,400 tpd. [30] This Buggenum total feed rate was used as the capacity basis for Case 6; however, the percent of biomass was adjusted to achieve net zero life cycle CO_2 emissions.

The evaluation scope included developing energy and mass balances and estimating plant performance. Equipment lists were developed for each design. Section 4.1 covers general information that is common to all of the cases; case-specific information is subsequently presented in sections 4.5, 4.9, and 4.13.

4.1 Gasifier Background

The three gasification cases are based on performance data for an oxygen-blown, entrained-flow gasifier that is intended to be similar to the Shell gasifier, by which it is referred to from here on out. This gasifier has been published in numerous prior NETL studies, including most-recently, the Fossil Energy Baseline. The estimates were derived from public literature and relied heavily on conference presentations given by Shell at the Gasification Technologies Conference in 2009. It is not intended to faithfully reproduce the operating point of any specific gasifier either currently operating, previously operating, or planned. It is a conceptual design that extends some of the features of the gasifier island, such as integration with other plant components. It was intended to keep the design basis comparable to how other gasifiers were represented in previous studies like the Fossil Energy Baseline. In 2018, the Shell gasifier technology was purchased and is now licensed by Air Products. It is recognized that the technology is now owned by Air Products, not Shell, but the gasifier will continue to be called the “Shell” gasifier to be consistent with the Fossil Energy Baseline and avoid confusion.

The “Coal Gasification Guidebook: Status, Application, and Technologies” report published by EPRI provides a detailed history of the development of several types of gasifier technology, including the Shell gasifier, as well as gasifier capacity, distinguishing characteristics, and important coal characteristics. [64] As of 2009, Shell reported ten gasifiers in operation, producing 100,000–150,000 Nm^3/hr of syngas each and three of the same size in construction. Another three ranging from 150,000 to 250,000 Nm^3/hr are also in construction. [65] The large gasifier operating in the Netherlands has a bituminous coal-handling capacity of 1,633 tonnes/day (1,800 tpd) and produces dry gas at a rate of 158,575 Nm^3/hr (5.6 million scf/hr) with an energy content of about 1,792 GJ/hr (1,700 MMBtu/hr) (HHV). This gasifier was sized to match the fuel gas requirements for the Siemens/Kraftwerk Union V-94.2 CT and could easily be scaled up to match state-of-the-art 2008 F-class turbine requirements. [65]

Shell gasifiers are capable of utilizing a wide variety of coal types, and compared to slurry fed gasifiers, the dry-fed, cooled-refractory lined, Shell gasifier has a lower O_2 requirement and produces a gas with a higher H_2S/COS ratio, which improves sulfur recovery. [64]

While the use of dry feed allows for lower O_2 consumption, the feed system—which includes the coal drying system—is more complicated. [64]

The Shell gasifier has demonstrated the ability to handle coal types with an ash content up to 37%. The slag viscosity is a critical parameter, and this is tuned by the addition of fluxants such as limestone. The negative impact that high ash coals have on the operation of gasifiers are reduced in dry feed systems in comparison to slurry fed gasifiers. [64]

4.2 Key System Assumptions

System assumptions for cases 4, 5, and 6 (coal gasification without and with CO_2 capture and coal/biomass co-gasification with CO_2 capture) are compiled in Exhibit 4-1.

Exhibit 4-1. Gasification plant study configuration matrix

<table>
<tr><th>Case</th><th>4</th><th>5</th><th>6</th></tr>
<tr><td>Gasifier Pressure, MPa (psia)</td><td colspan="3">4.2 (615)</td></tr>
<tr><td>O_2:Coal Ratio, kg O_2/kg As-Received coal</td><td colspan="3">0.720</td></tr>
<tr><td>Carbon Conversion, %</td><td colspan="3">99.5</td></tr>
<tr><td>Steam Cycle, MPa/°C/°C (psig/°F/°F)</td><td>11.7/482/482
(1,700/900/900)</td><td>12.4/510/510
(1,800/950/950)</td><td>12.4/532/532
(1,800/990/990)</td></tr>
<tr><td>Condenser Pressure, mm Hg (in. Hg)</td><td colspan="3">51 (2.0)</td></tr>
<tr><td>Gasifier Technology</td><td colspan="3">Shell</td></tr>
<tr><td>Oxidant</td><td colspan="3">95 vol% O_2</td></tr>
<tr><td>Feedstock</td><td colspan="2">Illinois No. 6 Coal</td><td>Illinois No. 6 Coal/torrefied, woody biomass</td></tr>
<tr><td>Coal Feed Moisture Content, %</td><td colspan="3">5</td></tr>
<tr><td>Plant Capacity</td><td colspan="2">304 kNm^3/hr
(274 MMSCFD)
Match Case 3</td><td>61 kNm^3/hr (55 MMSCFD)</td></tr>
</table>

Case	4	5	6
			~1,400 tpd AR feedstock
Water Gas Shift	2x3 train configuration, high-temperature, 97.2% conversion		
H_2 Product Purity	99.90 vol%		
H_2 Product Pressure	939.7 psia		
H_2 Purification	Pressure Swing Adsorption		
COS Hydrolysis	Occurs in WGS		
H_2S Separation	MDEA	Selexol 1st Stage	
Sulfur Removal, %	99.6%	~100.0	
Sulfur Recovery	Claus Plant with Tail Gas Treatment/Elemental Sulfur		
Particulate Control	Cyclone, Candle Filter, Scrubber, Quench, and AGR Absorber		
Chloride Control	Venturi Scrubber, Vacuum Flash, Brine Concentrator, Crystallizer		
Mercury Control	Carbon Bed		
NOx Control	Low-NOx Burners (LNB)		
CO_2 Separation	N/A	Selexol 2nd Stage	
CO_2 Storage	N/A	Off-site Saline Formation	

4.2.1 Balance of Plant

The balance of plant assumptions are common to all three cases and presented in Exhibit 4-2.

Exhibit 4-2. Balance of plant assumptions

Parameters	Values
Cooling System	Recirculating Wet Cooling Tower
Fuel and Other Storage	
Coal/Biomass	30 days
Slag	30 days
Sulfur	30 days
Sorbent	30 days
Plant Distribution Voltage	
Motors below 1 hp	110/220 V
Motors between 1 hp and 250 hp	480 V
Motors between 250 hp and 5,000 hp	4,160 V
Motors above 5,000 hp	13,800 V
Steam Turbine Generators	24,000 V

Parameters	Values
Grid Interconnection Voltage	345 kV
Water and Wastewater	
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
Process Wastewater	Storm water that contacts equipment surfaces is collected and treated for discharge
Sanitary Waste Disposal	Design includes a packaged domestic sewage treatment plant with effluent discharged to the industrial wastewater treatment system. Sludge is hauled off site. Packaged plant is sized for 5.68 cubic meters per day (1,500 gallons per day)
Water Discharge	Blowdown from the cooling tower is softened and passed through a two-stage RO with pre-treatment and demineralizer before being discharged

4.3 Sparing Philosophy

The sparing philosophy for cases 4, 5, and 6 is provided below. Dual trains are used to accommodate the scale of the hydrogen production. There is no redundancy other than normal sparing of rotating equipment. The plant design consists of the following major subsystems:

- Two ASUs in Cases 4 & 5 (2 x 50 percent) and one ASU in Case 6 (1 x 100 percent)
- Two trains of gasification, including feedstock drying & feed, gasifier, synthesis gas cooler (SGC), and slag handling (2 x 50 percent)
- Two trains of syngas clean-up process (2 x 50 percent)
- Two trains of refrigerated MDEA in Case 4 (1 x 50) and one train of two-stage Selexol in Cases 5 & 6 (1 x 100 percent)
- One train of CO_2 compression (1 x 100 percent) in Case 5 and Case 6
- Two trains of process water treatment systems (2 x 50 percent)
- One train of Claus-based sulfur recovery (1 x 100 percent)
- One train of H_2 production and compression systems (1 x 100 percent)
- One steam turbine (1 x 100 percent)

4.4 Gasification Common Process Areas

The cases have process areas, which are common to each plant configuration, such as coal receiving and storage, O_2 supply, gas cleanup, hydrogen production, etc. As detailed descriptions of these process areas for each case would be repetitious, they are presented in this section for general background information. Where there is case-specific performance information, the performance features are presented in the relevant case sections.

4.4.1 Coal Receiving and Storage

The function of the coal receiving and storage system is to unload, convey, prepare, and store the coal delivered to the plant. The scope of the system is from the trestle bottom dumper and coal receiving hoppers up to and including the slide gate valves at the outlet of the coal storage silos. Coal receiving and storage is identical in design for all three cases.

The coal is delivered to the site by rail cars. The unloading is done by a trestle bottom dumper, which unloads the coal into two receiving hoppers. Coal from each hopper is fed directly into a vibratory feeder. The 8-centimeter (cm) x 0 (3-in. x 0) coal from the feeder is discharged onto a belt conveyor. Two conveyors with an intermediate transfer tower are assumed to convey the coal to the coal stacker, which transfer the coal to either the long-term storage pile or to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron and then to the reclaim pile.

The reclaimer loads the coal into two vibratory feeders located in the reclaim hopper under the pile. The feeders transfer the coal onto a belt conveyor that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3 cm x 0 (1.25 in. x 0) by the crusher. A conveyor then transfers the coal to a transfer tower. In the transfer tower, the coal is routed to the tripper, which loads the coal into one of three silos. Two sampling systems are supplied: the as-received sampling system and the as-fired sampling system. Data from the analyses are used to support the reliable and efficient operation of the plant.

4.4.2 Coal Preparation and Feed System

The receiving and handling subsystem ends at the coal silo. The Shell process uses a dry feed system, which is sensitive to the coal moisture content. Coal moisture consists of two parts, surface moisture and inherent moisture. For coal to flow smoothly through the lock hoppers, the surface moisture must be removed. The Illinois No. 6 coal used in this report contains 11.12 percent total moisture on an as-received basis. It was assumed that the coal must be dried to 5 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 with an approximate two-hour capacity. A slipstream of clean syngas is combusted with air in an incinerator, resulting in a flue gas with an O_2 content of 6 vol%, which is determined by the amount of air added to the drying loop. The incinerator’s flue gas is used to dry the coal in the mill before being vented to the atmosphere.

The coal is drawn from the surge hoppers and fed through a pressurization lock hopper system to a dense phase pneumatic conveyor, which uses N_2 from the ASU to convey the coal to the gasifiers.

4.4.3 Air Separation Unit

A high-pressure ASU, nearly identical in design to the ATR with CO_2 capture case (Case 3) described in Section 3.12.1, is considered in each of the three gasification cases with the following modifications:

- Nominal ASU capacities for the gasification cases are 4,400 tpd for Cases 4 and 5, and 1,100 tpd for Case 6 of 95 vol% O_2
- The 95 vol% O_2 is compressed to 740 psia to meet the design pressure of the Shell gasifier
- HP N_2 is also produced from the cold box and is further compressed to 815 psia for feedstock transport

The ASU process description can be found in Section 3.12.1 and is not repeated here.

4.4.4 Gasifier

There are two Shell dry feed, pressurized, up-flow, entrained, slagging gasifiers, operating at 4.2 MPa (615 psia). The air separation plant supplies 95 percent O_2 to the gasifiers. Coal reacts with O_2 and steam at a temperature of 1,427 °C (2,600 °F) in the gasifier to produce principally H_2 and CO with little CO_2 formed. The gasifier's refractory-lined water wall is protected by molten slag that partly solidifies on the cooled walls.

Note that the presented gasifier design is conceptual, not based on vendor quote, and may differ from operating plants.

4.4.5 Raw Gas Cooling and Particulate Removal

High-temperature heat recovery in each gasifier train is accomplished in three steps, including the gasifier jacket, which cools and solidifies slag touching the gasifier walls and maintains the syngas temperature at 1,427 °C (2,600 °F). The product gas from the gasifier is cooled below 1,093 °C (2,000 °F) by adding cooled recycled syngas to lower the temperature below the ash melting point. The mixed gas then goes through a duct cooler, which lowers the gas temperature to 899 °C (1,650 °F). Following the duct cooler is a direct contact water quench, which cools the raw syngas to a temperature of 399 °C (750 °F) while providing a portion of the water required for WGS and removing a significant portion of the PM from the syngas. The duct cooler produces HP steam for use in the steam cycle.

The majority of the fine particulates in the cooled gas from the syngas cooler are removed by passing through a cyclone collector, followed by an array of raw gas metallic or ceramic candle filter elements in a pressure vessel. Recycled syngas is used as the pulse gas to clean the candle filters. A carbon conversion of 99.5 percent is achieved by recycling the recovered fines, which are returned to the gasifier with the fuel.

The ash that is not carried out with the gas forms slag and runs down the interior walls, exiting the gasifier in liquid form. The slag is solidified in a quench tank for disposal. Lockhoppers are used to reduce the pressure of the solids from 4.2 to 0.1 MPa (615 to 15 psia). The syngas scrubber removes additional PM further downstream.

The syngas from the candle filter is further cooled to 232 °C (450 °F) by producing intermediate pressure (IP) steam at 5.1 MPa (740 psia) for use in the gasifier and producing LP steam at 0.4 MPa (65 psia).

4.4.6 Quench Gas Compressor

A portion of the cooled syngas is recycled back to the gasifier exit as quench gas. A single-stage compressor is utilized to boost the pressure of the cooled syngas stream to 4.3 MPa (625 psia).

4.4.7 Syngas Scrubber

The ejector-type venturi scrubber is common to all cases. The raw syngas exiting the final raw gas cooler at 232 °C (450 °F) enters the scrubber for removal of HCl and remaining PM. The treated syngas leaves the scrubber at saturation. Due to the wet nature of the syngas, a recycle of the scrubber effluent is required; however, a minimum blowdown is maintained to limit the chloride concentration buildup in the scrubber effluent to 5,000 ppmv. The recycled scrubber effluent is cooled to 44 °C (112 °F) by exchanging heating with feedwater to the WGS steam generator before being further cooled through the addition of process water. Process water and zero liquid discharge (ZLD) condensate are cooled to 32 °C (90 °F) by preheating fuel gas prior to the gas-fired boiler (GFB) and subsequently cooled further to 21 °C (70 °F) with cooling water before the water stream is injected into the scrubber. Since the chloride concentration is maintained by adjusting the rate of effluent recycle, the rate of process water addition is used to maintain the HCl removal rate at 96 percent, along with removing essentially all traces of entrained particles (principally unconverted carbon, slag, and metals).

A 50 percent by weight (wt%) solution of NaOH is added to the scrubber to maintain pH and form the HSS sodium chloride NaCl.

The blowdown from the syngas scrubber is sent to the process water treatment system for chloride removal and recycle.

4.4.8 Water Gas Shift Reactors

In all cases, the gasifier product must be converted to a CO_2 and H_2-rich syngas in order to maximize H_2 product yield. The water gas shift technology, which is also used in the reforming cases, has been previously described in Section 3.3.4.

The CO shift converter is located upstream of the AGR step (sour gas shift) in order to promote COS hydrolysis without a separate catalyst bed. The COS hydrolysis reaction is equimolar with a slightly exothermic heat of reaction, as shown in the following reaction:

$$\mathrm{COS} + \mathrm{H_2O} \leftrightarrow \mathrm{CO_2} + \mathrm{H_2S};\ \Delta H°_{\mathrm{rxn}} = -33.9\ \mathrm{kJ/mol}$$

Since the reaction is exothermic, higher conversion is achieved at lower temperatures. However, at lower temperatures the reaction kinetics are slower.

Steam injection upstream of the shift reactors is extracted from the steam cycle and is used to drive the reaction and control the outlet steam to dry gas ratio.

Cooling is provided between the series of reactors to control the exothermic temperature rise. In all cases, the HXs are used to raise steam for injection or otherwise integrated into the plant, such as for syngas reheating.

The heat generated from the first reactor is used to produce more steam than is required (3.9 MPa (563 psia) steam is exported for use in the steam cycle) to maintain the desired steam to dry gas ratio while cooling the syngas to 253 °C (487 °F) prior to entering the second stage. After the second WGS reactor, the syngas is again cooled to a temperature of 253 °C (487 °F) by exchanging heat with the feed water of the WGS steam generator. Prior to the syngas being sent to the low temperature heat recovery (LTHR) system, the warm syngas from the third stage of WGS is cooled to 202 °C (395 °F) by preheating the feed water of the WGS steam generator.

4.4.9 Low Temperature Heat Recovery

The raw syngas from the WGS unit is cooled through a series of four shell and tube HXs. The first stage cools the syngas from 202 °C (395 °F) to 162 °C (323 °F) by raising 0.4 MPa (65 psia) process steam. The second stage further cools the syngas to a temperature of 134 °C (274 °F) by preheating the cold feedwater for the WGS steam generator. The third stage cools the syngas from 59 °C (138 °F) by preheating FW to the GFB. Finally, the syngas is cooled to a temperature of 29 °C (85 °F) in the fourth stage by cooling water. During cooling, part of the water vapor condenses, along with significant amounts of NH_3, and is combined with the effluent of the NH_3 wash.

4.4.10 Mercury Removal

A conceptual design for a sulfur-impregnated, activated carbon bed adsorption system was developed for mercury control, where mercury is captured in the reduced state. Data on the performance of carbon bed systems were obtained from the Eastman Chemical Company, which uses carbon beds at its syngas facility in Kingsport, Tennessee. [66] The coal mercury content (0.15 ppmvd) and carbon bed removal efficiency (95 percent) were discussed previously in Section 2.

4.4.10.1 Carbon Bed Location

The packed carbon bed vessels are located after the ammonia wash section, upstream of the AGR process. Syngas is preheated from temperatures between 27 °C (80 °F) and 29 °C (85 °F), to temperatures of 37 °C (98 °F) and 38 °C (100 °F) before entering the bed. The preheating is necessary to avoid condensation within the bed. Eastman Chemical operates their beds ahead of their sulfur recovery unit (SRU) at a temperature of 30 °C (86 °F). [66]

4.4.10.2 Process Parameters

An empty vessel basis gas residence time of approximately 20 seconds was used based on Eastman Chemical's experience. [66] Allowable gas velocities are limited by considerations of particle entrainment, bed agitation, and pressure drop. One-foot-per-second superficial velocity is in the middle of the range normally encountered [67] and was selected for this application.

The bed density of 480 kg/m^3 (30 lb/ft^3) was based on the Calgon Carbon Corporation HGR®-P sulfur-impregnated pelleted activated carbon offering as of 2002. [68] These parameters determined the size of the vessels and the amount of carbon required. Each gasifier train has two sequential mercury removal beds; the first bed achieves 90 percent of the necessary removal, with the second bed removing the balance of the mercury necessary to meet the emissions limit of 95 percent removal. Since there are two gasifier trains per case, each case has four total carbon beds.

4.4.10.3 Carbon Replacement Time

Eastman Chemicals replaces its bed every 18 to 24 months. [66] However, bed replacement is not due to mercury loading, but rather from:

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

For this report, a 24-month carbon replacement cycle was assumed. Under these assumptions, the mercury loading in the bed would build up to 0.6–1.1 percent by weight (wt%). Mercury capacity of sulfur-impregnated carbon can be as high as 30 wt%. [69] The mercury laden carbon is considered to be a hazardous waste.

4.4.11 Sulfur Recovery/Tail Gas Cleanup

Sulfur compounds in syngas need to be removed in most gasification applications due to environmental regulations or to avoid catalyst poisoning. The Claus process is still the industry standard for sulfur recovery. Conventional three-stage Claus plants, with indirect reheat and feeds with a high H_2S content, can approach 98 percent sulfur recovery efficiency. However, since environmental regulations have become more stringent, sulfur recovery plants are required to recover sulfur with over 99.8 percent efficiency. To meet these stricter regulations, the Claus process underwent various modifications and add-ons. [70]

The add-on modification to the Claus plant selected for this report can be considered a separate option from the Claus process. In this context, it is often called a tail gas treating unit (TGTU) process. The Claus process converts H_2S to elemental sulfur via the following reactions:

$$\mathrm{H_2S} + \frac{3}{2}\mathrm{O_2} \leftrightarrow \mathrm{H_2O} + \mathrm{SO_2}$$

$$2\mathrm{H_2S} + \mathrm{SO_2} \leftrightarrow 2\mathrm{H_2O} + 3\mathrm{S}$$

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

$$3\mathrm{H_2S} + \frac{3}{2}\mathrm{O_2} \leftrightarrow 3\mathrm{H_2O} + 3\mathrm{S}$$

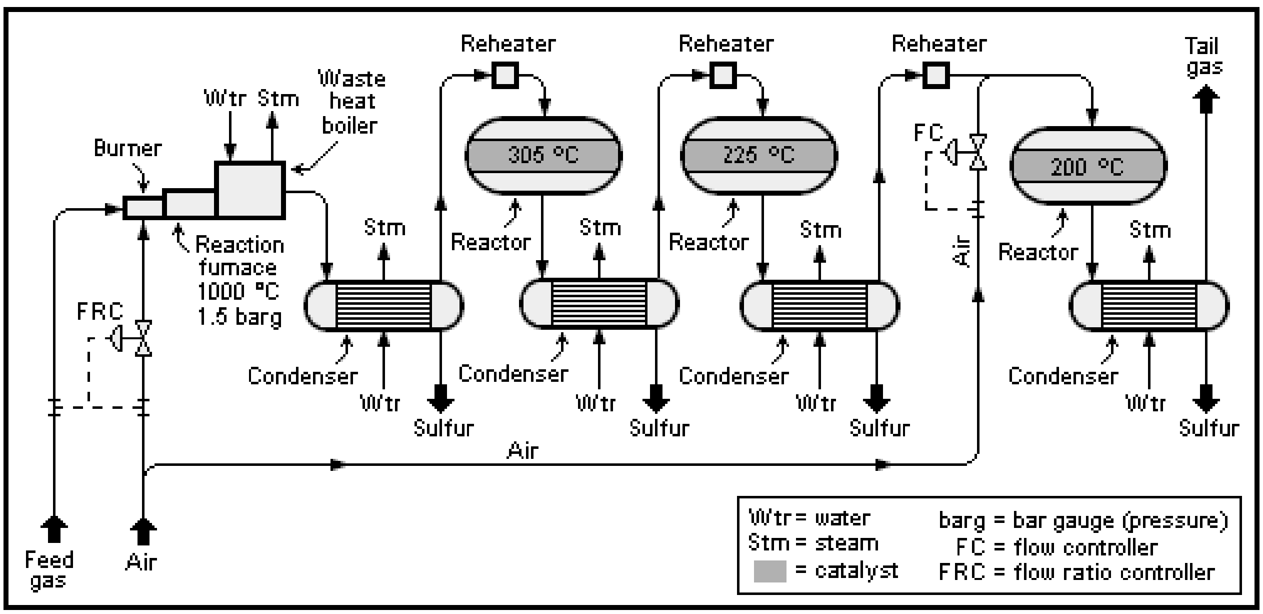
The sulfur in the vapor phase exists as S_2, S_6, and S_8, with the S_2 predominant at higher temperatures, and S_8 predominant at lower temperatures.

A simplified process flow diagram of a typical three-stage Claus plant is shown in Exhibit 4-3. [70] One-third of the H_2S is burned in the furnace with O_2 from the air to give sufficient SO_2 to react with the remaining H_2S. Since these reactions are highly exothermic, a boiler that recovers this heat to generate HP steam usually follows the furnace. Sulfur is condensed in a condenser that follows the HP steam recovery section. LP steam is raised in the condenser. The tail gas from the first condenser then goes to several catalytic conversion stages, usually two to three, where the remaining sulfur is recovered via the Claus reaction. Each catalytic stage consists of gas preheat, a catalytic reactor, and a sulfur condenser. The liquid sulfur goes to the sulfur pit, while the tail gas proceeds to the incinerator or for further processing in a TGTU.

4.4.11.1 Claus Plant Sulfur Recovery Efficiency

The Claus reaction is equilibrium limited, and sulfur conversion is sensitive to the reaction temperature. The highest sulfur conversion in the thermal zone is limited to about 75 percent. Typical furnace temperatures are in the range of 1,093–1,427 °C (2,000–2,600 °F), and as the temperature decreases, conversion increases dramatically.

Exhibit 4-3. Typical three-stage Claus sulfur plant



Used with permission from Beychok [71]

Claus plant sulfur recovery efficiency depends on many factors:

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

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

4.4.11.2 Oxygen-Blown Claus

One way to reduce diluent flows through the Claus plant and to obtain stable temperatures in the furnace for dilute H_2S streams is the O_2-blown Claus process.

The O_2-blown Claus process was originally developed to increase capacity at existing conventional Claus plants and to increase flame temperatures of low H_2S content gases. The process has also been used to provide the capacity and operating flexibility for sulfur plants where the feed gas is variable in flow and composition such as often found in refineries. The application of the process has now been extended to greenfield installations, even for rich H_2S feed streams, to provide operating flexibility compared to conventional Claus units. At least four of the gasification plants in Europe use O_2-enriched Claus units.

O_2 enrichment results in higher temperatures in the front-end furnace, potentially reaching temperatures as high as 1,593–1,649 °C (2,900–3,000 °F) as the enrichment moves beyond 40–70 vol% O_2 in the oxidant feed stream. Although O_2 enrichment has many benefits, its primary benefit for lean H_2S feeds is a stable furnace temperature. O_2 enrichment also allows for tail gas recycle without N_2 diluent, which would be needed for an air-blown Claus system. Sulfur recovery is not significantly enhanced by O_2 enrichment. Because the gasification process already requires an ASU, the O_2-blown Claus plant was chosen for all cases.

4.4.11.3 Tail Gas Treating

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

In the case of gasification applications, the tail gas from the Claus plant can be catalytically hydrogenated and then recycled back into the system with the choice of location being technology dependent, or it can be treated with a SCOT-type process. In each of the three gasification cases the Claus plant tail gas is hydrogenated, water is separated out, the tail gas is compressed, and the gas is returned to the AGR process for further treatment.

4.4.11.4 Flare Stack

In most Claus plants, a self-supporting, refractory-lined, carbon steel (CS) flare stack is typically provided to combust and dispose of unreacted gas during startup, shutdown, and upset conditions. However, in all three gasification cases, a flare stack was included for syngas dumping during startup, shutdown, etc. Hence, a separate dedicated Claus plant flare was not required.

4.4.12 Slag Handling

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

The general aspects of slag handling are the same for all three gasification cases. The slag is dewatered, the water is clarified and recycled, and the dried slag is transferred to a storage area for disposal.

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

4.4.13 Steam Generation

4.4.13.1 Gas-Fired Boiler

The gas-fired boiler (GFB) is an air-fired, multi-pressure, drum-type piece of equipment within the steam generation island. The H_2-rich fuel gas exiting the PSA is mixed with air and combusted in the GFB to recover the large quantity of thermal energy that remains.

In addition to generating and superheating steam, the GFB performs reheat duty for the cold/hot reheat steam for the steam turbine, provides condensate and feedwater (FW) heating, and provides deaeration of the condensate.

Natural circulation of BFW and steam is accomplished in the GFB by utilizing differences in densities due to temperature differences of the steam. The GFB drums include moisture separators, internal baffles, and piping for FW/steam. All tubes, including economizers, superheaters, and headers and drums, are equipped with drains.

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

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

4.4.13.2 Steam Turbine Generator and Auxiliaries

The steam turbine consists of an HP section, an IP section, and one double-flow LP section, all connected to the generator by a common shaft. The HP and IP sections are contained in a single-span, opposed-flow casing, with the double-flow LP section in a separate casing.

Main steam from the GFB and gasifier island is combined in a header, and then passes through the stop valves and control valves and enters the turbine at stream conditions shown in Exhibit 4-4. The steam initially enters the turbine near the middle of the HP span, flows through the turbine, and returns to the GFB for reheating. The reheat steam flows through the reheat stop valves and intercept valves and enters the IP section at stream conditions shown in Exhibit 4-4.

Exhibit 4-4. Gasification steam conditions

Steam Conditions		
Steam Parameter	**Capture**	**Non-Capture**
Main Pressure, MPa (psig)	12.4 (1,800)	11.7 (1,700)
Main Temperature, °C (°F)	510-532 (950-990)	482 (900)
Reheat Pressure, MPa (psig)	3.4 (491)	3.4 (491)
Reheat Temperature, °C (°F)	510-532 (950-990)	482 (900)

After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the LP section. The steam divides into two paths and flows through the LP sections, exhausting downward into the condenser.

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

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

The generator is a H_2-cooled synchronous type, generating power at 24 kV. A static, transformer type exciter is provided. The generator is cooled with a H_2 gas recirculation system using fans mounted on the generator rotor shaft. The heat absorbed by the gas is removed as it passes over finned tube gas coolers mounted in the stator frame. Gas is prevented from escaping at the rotor shafts by a CL oil seal system. The oil seal system consists of storage tank, pumps, filters, and pressure controls, all skid-mounted.

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

4.4.13.3 Condensate System

The condensate system transfers condensate from the condenser hotwell, through a series of economizers, to the deaerator. The economizers may exchange heat with either the tail-gas recycle coolers, LTHR system, and/or the low-temperature economizer section in the GFB, depending on the case. The system consists of one main condenser; two 50 percent capacity, motor-driven, vertical condensate pumps; one gland steam condenser; and a low-temperature tube bundle in the GFB. Condensate is delivered to a common discharge header through separate pump discharge lines, each with a check valve and a gate valve. A common minimum flow recirculation line discharging to the condenser is provided to maintain minimum flow requirements for the gland steam condenser and the condensate pumps.

4.4.13.4 Feedwater System

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

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

4.4.13.5 Main and Reheat Steam Systems

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

Main steam at conditions shown previously in Exhibit 4-4 exits the GFB superheater through a motor-operated stop/check valve and a motor-operated gate valve and then is routed to the HP turbine. Cold reheat steam exits the HP turbine, flows through a motor-operated isolation gate valve, to the GFB reheater. Hot reheat steam at the conditions shown previously in Exhibit 4-4 exits the GFB RH through a motor-operated gate valve and is routed to the IP turbines.

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

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

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

4.4.13.6 Circulating Water System

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

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

The condenser is a single-pass, horizontal type with divided water boxes. There are two separate circulating water circuits in each box. One-half of the condenser can be removed from service for cleaning or for plugging tubes. This can be done during normal operation at reduced load.

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

4.4.13.7 Raw Water, Fire Protection, and Cycle Makeup Water Systems

The raw water system supplies cooling tower makeup, steam cycle makeup, service water and potable water requirements. The water source is 50 percent from a POTW and 50 percent from groundwater (makeup water quality is provided in Section 2.1). Booster pumps within the plant boundary provide the necessary pressure.

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

The cycle makeup water system provides high-quality demineralized water for makeup to the GFB cycle and for steam injection ahead of the WGS reactors.

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

4.4.14 Process Water Systems

4.4.14.1 Process Water Sources

This section provides brief technology descriptions of equipment that produces process wastewater from gasification plants, including the syngas scrubber, LTHR, sour water stripper (SWS), NH_3 wash, and process water drum.

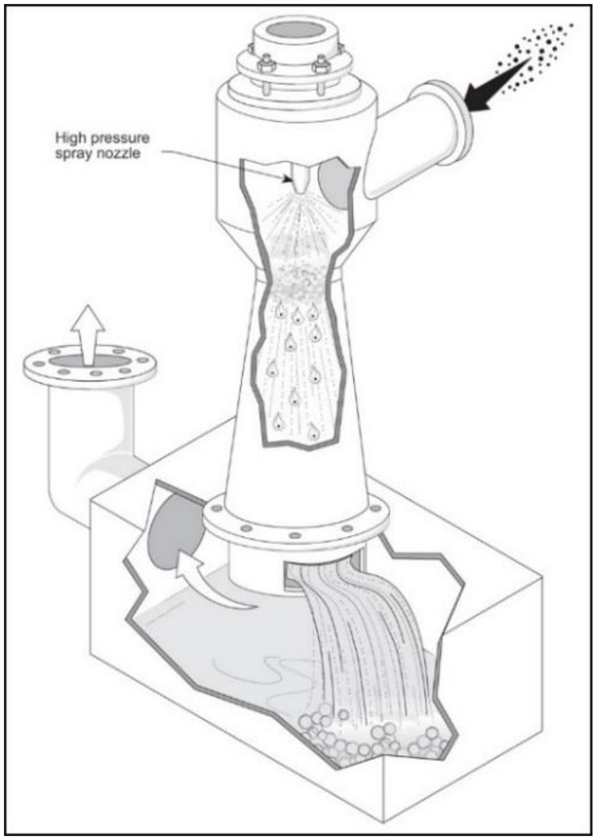
4.4.14.1.1 Syngas Scrubber

The majority of PM is removed from the syngas by upstream equipment, leaving only the finest particulates remaining prior to the syngas scrubber. Most of the remaining particulate is removed in the scrubber, although the primary concern of the syngas scrubber is to facilitate high-efficiency gas cleaning by maximizing the contact surface area between liquid and gas, as gases such as HCl are eliminated through absorption into the scrubbing liquid, which in this case is an alkali solution of water and NaOH.

A HP ejector venturi scrubber is particularly suitable for high-efficiency gas cleaning in HP operations and is frequently selected to facilitate this process, with expected HCl removal efficiencies in excess of 95 percent. [72] [73]

Exhibit 4-5 provides a diagram of an ejector type venturi scrubber. The gas enters the top section of the system where it comes into contact with a spray of fine water droplets, the spray is directed into a chamber that is shaped to conduct the gas through the atomized droplets, [74] where the HCl and other soluble gases are absorbed. The majority of the remaining particulates are removed from the gas stream by impingement against the high-velocity droplets. The liquid is collected in a reservoir, and the gas exits the side of the reservoir opposite the entrance. The liquid is pumped from the reservoir through a settling tank for particulate removal before being recycled.

Exhibit 4-5. Example diagram of an ejector type venturi scrubber



Source: EPA [75]

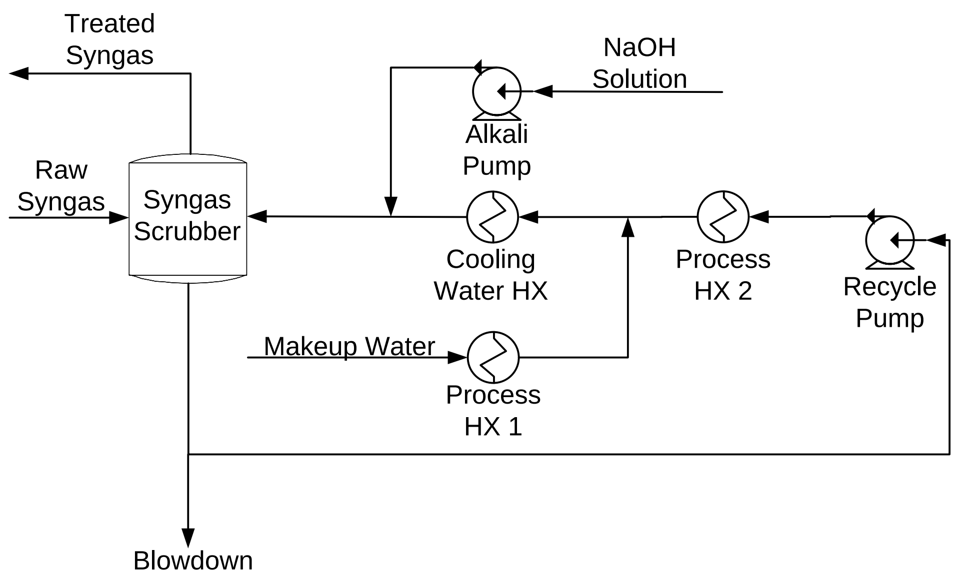
The ejector type venturi scrubbers are typically constructed of 316L stainless steel, operate with a relatively low pressure drop (approximately 2–3 percent of the inlet pressure) [76], have a liquid injection pressure of around 0.83 MPa (120 psi) above that of the inlet gas stream [74], and require between 30 and 100 gallons per 1,000 ft^3 of inlet gas. [72]

While 316L stainless steel has a very high tolerance to alkali solutions without concern for corrosion (concentrations of NaOH of up to 50 wt% can be used with negligible corrosion rates [77]), it can only withstand up to 2,000 ppmw of chloride ions. Considering the cost of the downstream ZLD equipment, priority was given to maximizing the chloride concentration and minimizing the process water discharge flowrate. Because 317L stainless steel can withstand up to 5,000 ppmw Cl^-, it was selected as the material of choice for the scrubber system. [78] The blowdown from the syngas scrubber is adjusted to maintain a Cl^- concentration of 5,000 ppmw, or lower.

While the tolerance of both 316L and 317L to alkali solutions is very high, the pitting rate of both steels rapidly increases with decreasing pH. At a pH of 5, severe pitting occurs with chloride concentrations as low as 500 ppmw. [79] In order to prevent excessive pitting, it is recommended that sufficient NaOH be added in the makeup water to maintain an alkali solution in the effluent. [72] [80]

Exhibit 4-6 provides a simplified block flow diagram (BFD) of the syngas scrubber system, including the alkali injection, makeup water, blowdown, and recycle. The alkali solution is assumed to be 50 wt% NaOH in water. The final cooling water HX cools the injection water to 21.1 °C (70 °F). The two process HXs are integrated with the syngas preheating system prior to the GFB and the WGS FW preheating system. The makeup water to the syngas scrubber is sourced from either ZLD condensate or process wastewater, depending on the selection of technologies utilized in each case.

Exhibit 4-6. Simplified syngas scrubber block flow diagram

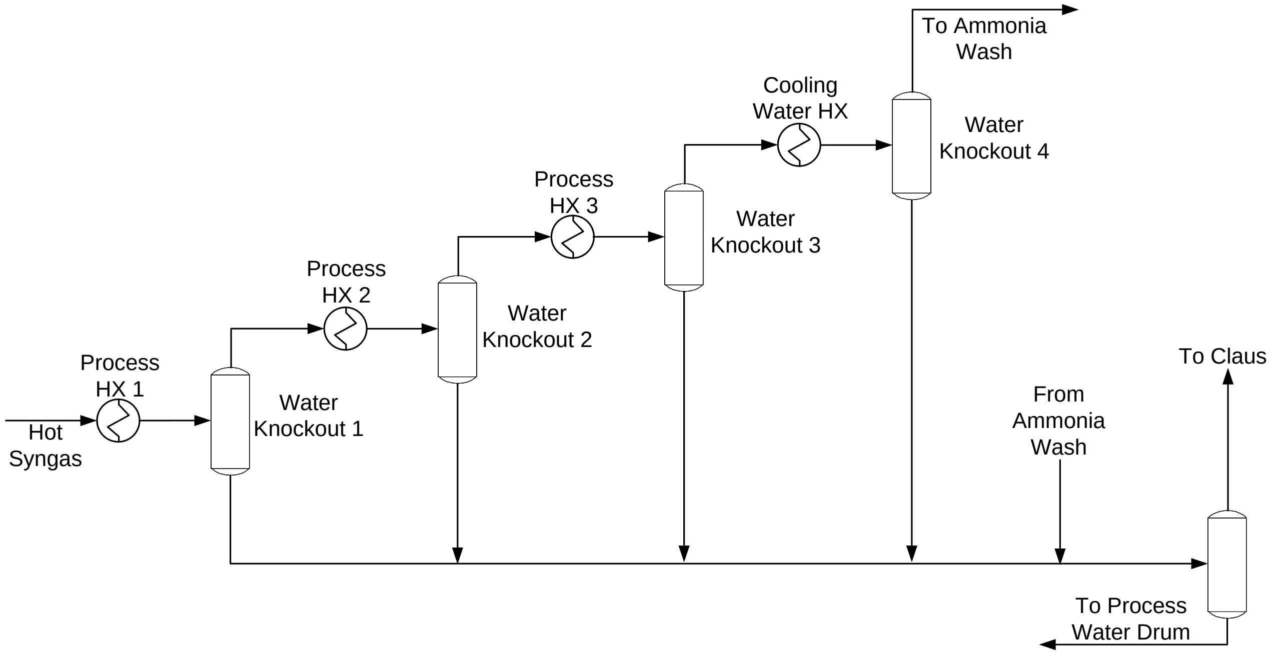


4.4.14.1.2 Low Temperature Heat Recovery

The gas exiting the WGS reactors enters the LTHR system before entering the NH_3 wash. The purpose of the LTHR system is to cool the syngas to the required operating temperature of the mercury control system, the NH_3 wash, and the AGR system, while recovering low grade heat.

As shown in Exhibit 4-7, LTHR consists of a series of shell-and-tube HXs; [81] depending on the inlet temperature, the LTHR system consists of one to three process integrated HXs, followed by a final heat exchange with cooling water. [82]

Exhibit 4-7. Simplified low temperature heat recovery system



During cooling, a significant amount of water is condensed, along with 55 percent of the NH_3 present at the inlet and nearly all the remaining HCl, leaving only trace amounts. Small amounts of CO_2, H_2, and H_2S are also removed. Knockout drums are located after each HX to remove condensed water.

The effluent from the LTHR system is combined with that of the NH_3 wash and is flashed to 0.5 MPa (65 psia). The vapor product is combined with the compressed sour gas from the SWS and sent to the Claus plant for incineration. The effluent from the flash drum is sent to the process water collection drum for use as process water recycle.

The first of the LTHR HXs is used to produce low pressure steam, which is used in the steam cycle, and the second of the LTHR HXs is used for fuel gas preheat prior and/or steam cycle FW preheat, depending on the plant configuration. The third LTHR HX is used to preheat steam cycle FW.

To withstand the NH_3 concentration in the effluent, which can approach 60,000 ppmw, 316L stainless steel was selected as the material of construction for its high tolerance to NH_3, as discussed in Section 4.4.14.1.5.

4.4.14.1.3 Primary Sour Water Stripper

As NH_3 is a highly soluble gas, it has a tendency to build-up in the plant process water. The solution is to utilize a combination of technologies that result in the presence of excess water from the process water collection drum that necessitates the utilization of a SWS to remove NH_3, H_2S, and other dissolved gases, so that the process water can be utilized as makeup water to downstream systems such as the NH_3 wash.

The presence of acids, chlorides, sulfates, and formates suppress NH_3 stripping. [83] Since small quantities of HCl are present in the sour water as a result of the LTHR effluent (discussed in

Section 4.4.14.1.2), which reacts with NH_3 to form ammonium chloride (NH_4Cl), small amounts of NaOH may be added to react with the NH_4Cl, releasing the NH_3 and producing the heat stable salt (HSS) sodium chloride (NaCl). [83] The caustic would be fed onto a tray far enough down the column that most of the H_2S has already been stripped out (tray 35 of 40). No more NaOH should be injected than is necessary to maintain pH, as it will chemically bind H_2S into the solution. [83] Other alkaline contaminants that can trap H_2S include sodium, potassium, and magnesium; however, they were assumed to be present in negligible quantities in this report.

HSS can cause the protonation of NH_3 and, therefore, cause a residual amount of NH_3 that cannot be removed (as little as 300 ppmw of HSS can prevent the treated water from reaching NH_3 concentrations of below 100 ppmw), which limits the usefulness of the effluent as recycle water. However, HSS can significantly improve H_2S removal rates, as they are weak acids. [83]

Despite the reasons presented above for adding caustic to the SWS, there are also potential adverse consequences, including a negative impact on the removal of H_2S, the fact that NaOH reacts with HCl to form NaCl, and the beneficial role that the presence of HCl can have on H_2S removal. Based on the net impacts, it was determined that the addition of NaOH would not be necessary and would not be utilized in this report.

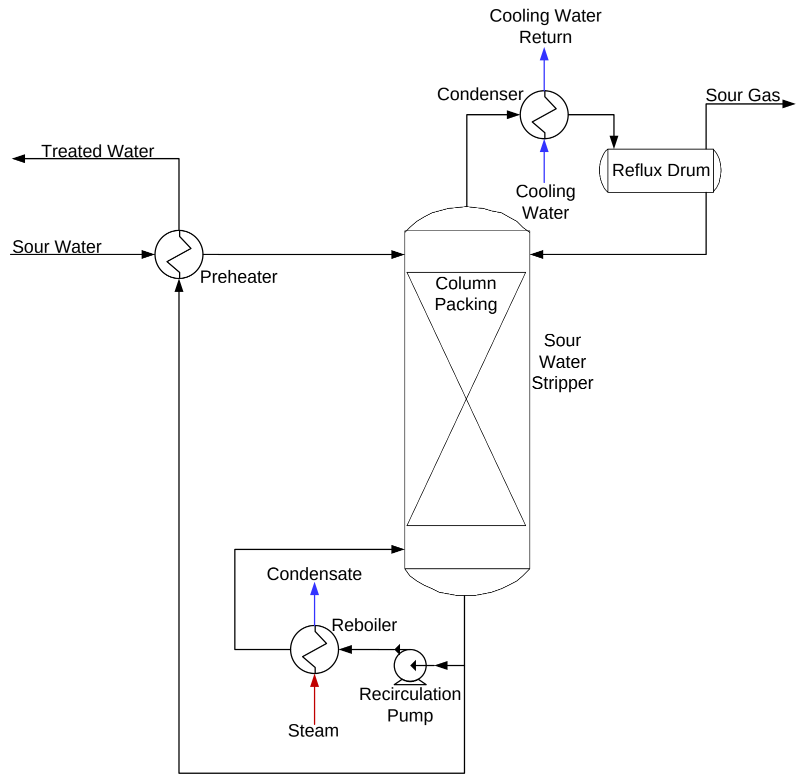
A selection of typical values for key operating parameters for a SWS are provided in Exhibit 4-8.

Exhibit 4-8. Typical operating parameters for sour water strippers

Parameter	Typical
Column Stages [83]	35-45
NH_3 in Effluent, ppmw [83]	30-80
H_2S in Effluent, ppmw [84]	<<1
Steam/Sour Water Feed, kg/m^3 (lb/ft^3) [84]	60-300 (3.7-18.7)
Column Operating Pressure, MPa (psia) [84]	0.1-0.5 (16-65)
pH of Sour Water Feed [83]	~9

Exhibit 4-9 provides a diagram of an SWS column. The excess water from the process water collection drum (described in Section 4.4.14.1.6) constitutes the sour water feed stream to the SWS. The sour water feed is preheated against the treated water product effluent prior to being injected at the top of the column. The sour water flows downward through the column packing [84] against an up-flow of sour gases and steam. A portion of the bottom water is recycled through the reboiler back to the column to increase the rate of recovery of the sour gases. The vapor product is passed through a partial-vapor condenser with the sour gases being separated from the condensate in the reflux drum. The condensate is returned to the column to increase the retention rate of water in the column.

Exhibit 4-9. Simplified sour water stripper diagram



To withstand the NH_3 concentration in the sour water feed stream, which can exceed 20,000 ppmw, 316L stainless steel was selected as the material of construction for its high tolerance to NH_3, as discussed in Section 4.4.14.1.5.

The sour water is fed to the SWS above the first stage at 0.4 MPa (65 psia), with the sour gas produced at 0.11 MPa (16 psia) and the effluent produced at stage 40 at 0.15 MPa (22 psia). The SWS utilizes an external kettle-type reboiler with 0.4 MPa (65 psia) process steam used as its heat source. [83] The partial-vapor condenser receives cold water from the cooling tower at 16 °C (60 °F) and the water is returned to the cooling tower at 27 °C (80 °F).

The SWS is designed to recover 99.5 percent by mole (mol%) of NH_3 while retaining 99.6 mol% of water. Only trace amounts of other dissolved gases remain, including H_2S. The sour gas is sent to the Claus plant for incineration, and the clean effluent is used as wash water in the NH_3 wash system.

4.4.14.1.4 Secondary Sour Water Stripper

In each gasification case, the selected combination of plant technologies resulted in excess process wastewater that required disposal. However, to qualify for ZLD, this wastewater needed to be utilized as process makeup water. Since the Shell gasifier requires steam addition at elevated pressure (~5.1 MPa [~740 psia]), it was determined that the production of mid-grade steam from process water would be beneficial, rather than extracting high-grade steam and letting it down to the required pressure.

The use of raw process water presents multiple design and operation issues, as these streams contain appreciable quantities of dissolved solids, which would cause fouling on the surface of the boiler tubes. Of particular concern is the presence of NaCl, which is typically limited to 10 ppmw in the boiler water feed. [85] In order to avoid this, the feed to the secondary SWS was sourced from the condensate of the ZLD processes (discussed in Section 4.4.14.2). This effluent is free from suspended solids, and contains less than 20 ppm of dissolved solids, consisting primarily of sodium- and calcium-based constituents. It is assumed that the sodium-based constituents comprise less than half of the total dissolved solids. Therefore, the only contaminant of concern is NH_3, which is limited to 200 ppmw at the desired operating pressure. [85]

The design of the secondary SWS is nearly identical to that of the primary SWS. The only two differences being that the secondary SWS operates with a feed pressure of 0.13 MPa (19.3 psia) and the column is designed to achieve an effluent with an NH_3 concentration of 200 ppmw. The sour gas from the secondary SWS is compressed and sent to the Claus plant for incineration.

4.4.14.1.5 NH_3 Wash

The operation of solvent-based AGR systems is sensitive to the presence of NH_3, which has a tendency to concentrate in the CO_2 reflux loop. [86] NH_3 forms ammonium sulfide with H_2S that is a contaminant in the product CO_2 and the off gas, and it can form ammonium carbamate with CO_2 that can cause plugging of equipment. [87] If concentrations are allowed to build-up to the point that performance starts to be impacted, the conventional recovery measure would consist of blowing down a portion of the reflux stream, which would then be disposed. [86]

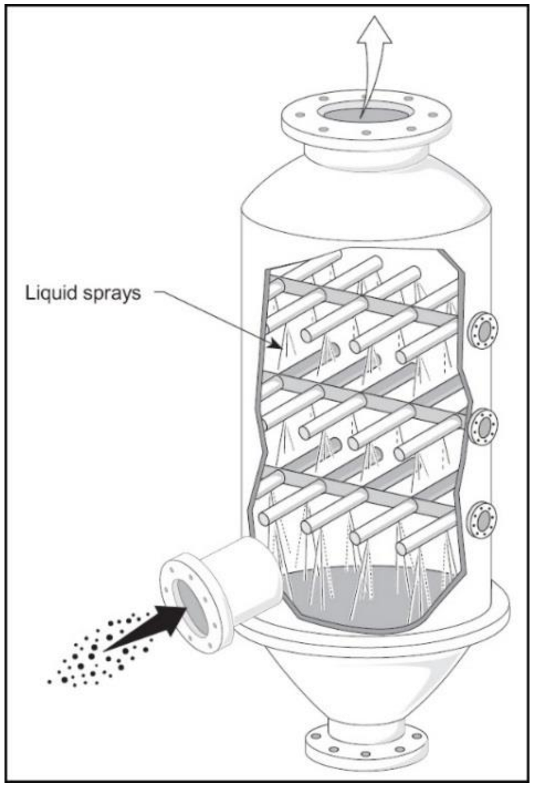
In order to minimize the loss of solvent and to maximize performance of the AGR, a pre-scrubber for NH_3 control is typically utilized to maintain NH_3 concentrations at or below 10 ppmv in the syngas. [86] [88] [89] A spray tower absorption column is often the scrubber of choice for highly soluble gases, such as NH_3, in systems that operate at low temperatures. [90] [86] [82]

NH_3, like HCl, is eliminated through absorption into the scrubbing liquid, [73] and as such, the efficiency of scrubbers increases with decreasing temperature; typical scrubber operating temperatures are around 26.7–29.4 °C (80–85 °F). [91] In order to achieve this, a cooling stage is included in the low temperature HX design (discussed in Section 4.4.14.1.2) that lowers the temperature to 29.4 °C (85 °F) and the wash water is cooled to 21.1 °C (70 °F) in a HX prior to injection.

A single-stage system can expect to achieve between 70 [86] and 85 [91] percent NH_3 reduction, with a typical configuration consisting of approximately 5 stages, [91] which can achieve total NH_3 reduction of over 99 percent. [91] To achieve the target concentration of NH_3 in the clean syngas of 10 ppm, 4 stages were required.

Exhibit 4-10 provides a diagram of an example spray tower absorption column. The spray tower operates by having a counter-current raw gas flow upward against a downward fine-mist spray of water. The water functions to both cool the syngas and absorb the NH_3, along with other soluble gases. The cleaned gas exits the top of the column and the water exits the bottom. To maximize NH_3 removal, the water is not recycled but is instead sent to a process water collection drum (discussed in Section 4.4.14.1.6) to be utilized as process water makeup. The water demand for the spray tower is primarily sourced from the effluent of the SWS, [92] [91] with raw water making up the balance, as necessary.

Exhibit 4-10. Example diagram of a spray tower absorption column



Source: EPA [75]

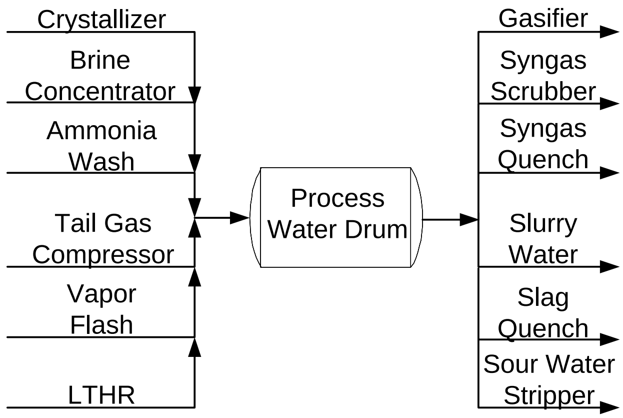
The NH_3 water wash column has the same relative pressure drop and water to gas ratio as the syngas scrubber, described in Section 4.4.14.1.1.

While spray scrubbers are generally constructed of carbon steel, [92] 316L stainless steel is required for high NH_3 concentration applications. [93] Considering the concentration of NH_3 can exceed 20,000 ppmw in the scrubber effluent and dissolved CO_2 is present in the system, 316L stainless steel was the material of choice for this study since no significant deterioration occurs in the presence of NH_3 at concentrations as high as 6 percent [60,000 ppmw]. [94]

4.4.14.1.6 Process Water Drum

The process water drum, depicted in Exhibit 4-11, is a collection tank and distribution point for process wastewater. The process water from various sources such as the Claus plant, ZLD process water treatment system, and NH_3 wash are collected together before being distributed to processes with a water demand, such as the syngas scrubber and gasifier.

Exhibit 4-11. Example diagram of a process water drum



Note: Slurry water is not necessary for dry-fed gasifiers like Shell

As is the case with the process wastewater sources, the process water drum would be subjected to a high concentration of NH_3 in the process water (greater than 20,000 ppmw). Therefore, 316L stainless steel is utilized as the material of construction, due to its high tolerance for NH_3 and CO_2, as discussed in Section 4.4.14.1.5.

4.4.14.1.7 Gasification Wastewater Quality

Gasification wastewater quality, summarized in Exhibit 4-12, represents the assumed quality of the water exiting the syngas scrubber that must be treated under the ELG rule, as stated earlier in Section 2.7.2. The gasification water quality is based on internal information from Black & Veatch IGCC projects utilizing a GE gasifier and discussions with GEP regarding their experience with gasification wastewater. The water quality is assumed to not be significantly different with a Shell gasifier. Exhibit 4-12 includes a range of values, an average, and the final selected gasification wastewater quality.

The wastewater composition reported in the following table is based on water qualities from actual operations and adjusted to account for chloride. The design concentration of each constituent is individually representative of a plant configuration comparable to those in this study. However, due to the interaction and interdependencies of each constituent and the multitude of potential species, the wastewater quality cannot be considered representative as a whole. The wastewater quality is intended to inform users of the contaminants likely present, and at what concentrations they may be expected at, to facilitate appropriate equipment selection and design.

Exhibit 4-12. Gasification wastewater quality

Parameter	Gasification Wastewater (Range)	Gasification Wastewater (Average)	Gasification Wastewater (Final)
pH	5.5–7.0		7.0
Chemical O_2 demand, ppm		1,500	1,500
Biological O_2 demand, ppm		1,000	1,000
Specific Conductance, µS/cm			14,000
Ammonia as N, ppm		<80	<80
Suspended Solids, ppm		<50	5
Total Dissolved Solids, ppm			14,995
Chloride as Cl, ppm		5,000	5,000
Sodium as Na, ppm		3,250	3,250
Formate, ppm		3,200	5,333
M-Alkalinity as $CaCO_3$, ppm[A]	600–2,000		700
Calcium as Ca, ppm	20–270		270
Sulfate as SO_4, ppm	25–100		100
Silica as SiO_2, ppm	25–50		50
Barium (total), ppm	0.20–40	20	40
Magnesium as Mg, ppm	4–20		20
Aluminum, ppm			20
Boron (total), ppm	2.5–10	5	10
Iron (total), ppm	2.5–10	5	10
Selenium (total), ppm	2.5–10	5	10
Sulfide, ppm		<10	<10
Cyanide, ppm		<5	<5
Chromium (total), ppm	0.5–2.0	1	2
Phosphorus, ppm	0.5–2.0	1	2
Potassium, ppm			2
Fluorine, ppm			2
Nickel (total), ppm	0.1–1.0	0.5	1
Molybdenum (total), ppm	0.2–0.8	0.4	0.8
Titanium (total), ppm	0.2–0.8	0.4	0.8
Lithium, ppm			0.3
Antimony (total), ppm	0.005–0.200	0.1	0.2
Arsenic (total), ppm	0.005–0.200	0.1	0.2
Lead (total), ppm	0.05–0.2	0.1	0.2
Thallium (total), ppm	0.05–0.2	0.1	0.2

Parameter	Gasification Wastewater (Range)	Gasification Wastewater (Average)	Gasification Wastewater (Final)
Vanadium (total), ppm	0.025–0.1	0.05	0.1
Uranium, ppm			0.1
Cobalt, ppm			0.1
Manganese (total), ppm	0.015–0.06	0.03	0.06
Beryllium (total), ppm	0.01–0.04	0.02	0.04
Copper (total), ppm	0.01–0.04	0.02	0.04
Zinc (total), ppm	0.01–0.04	0.02	0.04
Thorium, ppm			0.04
Cadmium (total), ppm	0.005–0.02	0.01	0.02
Tin, ppm			0.02
Mercury (total), ppm	0.002–0.008	0	0.01

[A]Alkalinity is reported as $CaCO_3$ equivalent, rather than the concentration of HCO_3. The concentration of HCO_3 can be obtained by dividing the alkalinity by 0.82

The gasification wastewater composition will be dependent on several factors, including composition of the coal, makeup water quality, syngas treatment systems that recycle water to the syngas scrubber, and other factors. The wastewater quality defined above will form the basis for discussion of the process water treatment systems, discussed in the following section.

4.4.14.2 Process Water Treatment

The updated ELG rule established gasification wastewater as a new category, with discharge limits that must be met. The gasification wastewater from the balance of plant is recycled within the gasification and syngas cleanup process, ultimately being utilized as makeup to the syngas scrubber. Therefore, all streams detailed in the updated ELG rule are included in the syngas scrubber blowdown (primary sources are described in Section 4.4.14.1) and can be treated by a single system with a composition described in Section 4.4.14.1.7.

It was a goal of this study to eliminate process water discharge in all the gasification cases presented in this report. Process water discharge is defined as any water discharged from systems that provide direct contact with contaminants foreign to the source-water (syngas scrubber, NH_3 wash, LTHR) to local waterways. Under these boundaries, blowdown from both the steam cycle and cooling tower are exempt from ZLD classification, provided that no process wastewater is utilized as makeup to either of these systems.

The equipment utilized to achieve ZLD is varied and dependent on several factors, such as contaminants being treated (e.g., selenium or chlorine), general water quality (e.g., pH), end use of product (e.g., process makeup or drinking water), land availability (evaporative ponds), geology characteristics (e.g., deep-well injection), and site characteristics (e.g., wetlands).

Wabash River, Kemper County, and Duke Edwardsport were designed with, and use, vapor-compression evaporation systems to treat their gasification wastewater as part of a ZLD operating practice. [39]

While multiple process configurations were assessed for feasibility of complying with the ELG rule, given this study's intention of maintaining general applicability of the cases presented, and the prevalence of utilizing ZLD operating practices in existing gasification plants, systems that would achieve ZLD were selected in all cases.

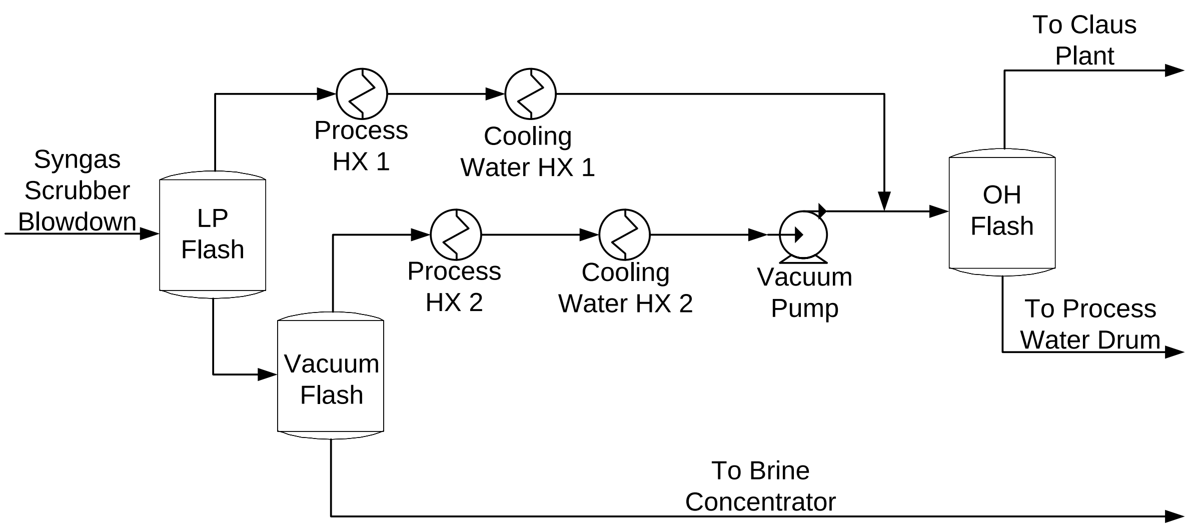
The process water treatment system for gasification cases includes a vacuum flash, brine concentrator, and crystallizer.

4.4.14.2.1 Vacuum Flash

The primary purpose of the vacuum flash system is to remove dissolved gases, such as NH_3, CO, H_2, and H_2S. The separation of dissolved gases prior to the brine concentrator aids in maintaining stable operation and avoiding upsets. Considering the desire to utilize the ZLD system condensate as FW to the auxiliary boiler, and the sensitivity of the auxiliary boiler to NH_3 (discussed in 4.4.14.1.3), it is particularly desirable to remove as much NH_3 as possible prior to the brine concentrator.

Exhibit 4-13 provides a diagram of a vacuum flash system. The blowdown from the syngas scrubber enters the LP drum where it is flashed to 0.5 MPa (70 psia). [95] The effluent of the LP drum enters the vacuum drum, where it is flashed to 0.05 MPa (7.5 psia). The effluent from the vacuum drum is then sent to the brine concentrator. The vapor overhead from both the LP and vacuum drums is first cooled to 46 °C (115 °F), by preheating fuel gas prior to the GFB, before being cooled further to 29 °C (85 °F) using cooling water. The vacuum flash is compressed in a vacuum pump before both overhead (OH) streams are sent to the overhead drum, where they are flashed to 0.2 MPa (35 psia). [95] The sour gas is compressed to 0.4 MPa (65 psia) and sent to the Claus plant for incineration and the effluent is sent to the process water drum.

Exhibit 4-13. Simplified diagram of vacuum flash system



While the syngas scrubber limits the chloride concentration of the blowdown stream to 5,000 ppmw, a portion of the water present in the blowdown will be lost with the NH_3 in the LP and vacuum flash drums, resulting in an increased chloride concentration in the effluent of the LP and vacuum flash drums.

The effluent from the LP flash drum exceeds the chloride tolerance of 317L stainless steel (5,000 ppmw). Therefore, a more advanced stainless steel will be required in both the LP and vacuum flash drum, such as 317LM or 317LMN. [96] The OH flash drum is required to be constructed of 316L stainless steel due to the high concentration of NH_3 (greater than 40,000 ppmw), as discussed in Section 4.4.14.1.2.

4.4.14.2.2 Brine Concentrator

A brine concentrator is a thermal evaporation process that is often selected as a component of a wastewater treatment system and is utilized as the first step in this study. There are two primary categories of evaporators used in the wastewater treatment industry: thin film and forced circulation. Most brine concentrators in operation are thin film evaporators configured to use a mechanical vapor compression vertical tube evaporation process, which partially evaporates water from the incoming waste stream and leaves behind a concentrated salt solution.

As NaOH is used for HCl scrubbing (discussed in Section 4.4.14.1.1), the wastewater contains primarily sodium-based salts, which are easy to crystallize compared to calcium- and magnesium-based salts, which are highly soluble and difficult to evaporate to a solid product. Therefore, the system can achieve full ZLD without a purge from the crystallizer.

Due to the nature of the salts and the low total suspended solids[f] (5 ppmw) in gasification wastewater, pretreatment upstream of the brine concentrator is not required. However, gasification wastewater typically contains constituents that will precipitate from solution within the brine concentrator when heated, adhering to the evaporator surface. While a seeded slurry process can be used to reduce precipitation on tube walls (calcium sulfate is added as a precipitation surface for low solubility salts), scaling cannot be completely eliminated through this method. Therefore, antiscalant addition was selected over seeding, which is fed at multiple points in the system.

To minimize carryover into downstream equipment, an antifoam is added to prevent foaming caused by biological O_2 demand, chemical O_2 demand, and other organics.

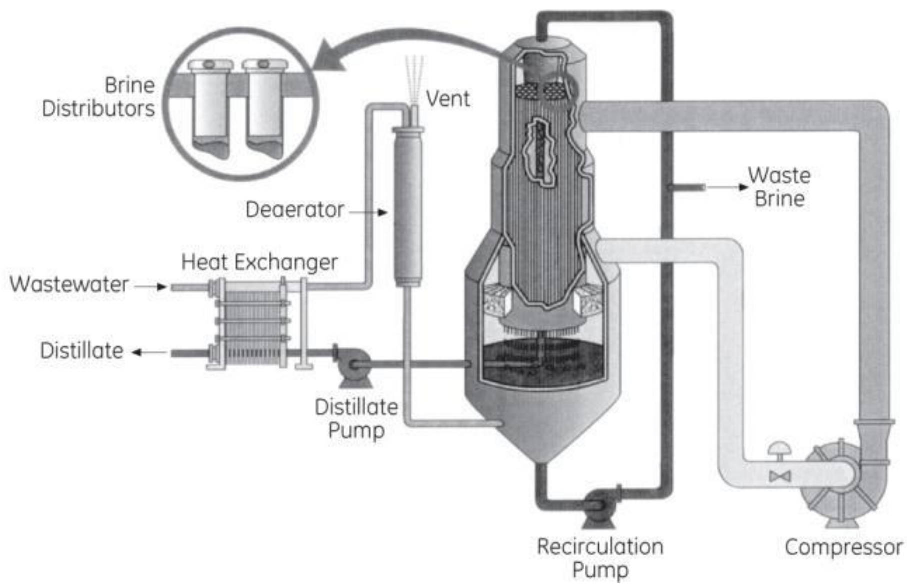
Lastly, sulfuric acid is added to prevent fouling and corrosion, as well as maintain consistent brine properties.

Exhibit 4-14 provides a depiction of a brine concentrator [97] and consists of a HX, heat transfer tubes, sump, sump pump, and compressor. The process water (depicted as "Wastewater") from the vacuum flash is pumped (not shown) through a HX, where its temperature is raised to the boiling point by the distillate product of the brine concentrator. While the depiction shows a deaerator following the HX, it is not included in the design utilized in this study as the vacuum flash serves the same purpose of removing dissolved gases. From the HX, the process water

[f] The feedwater to the brine concentrator is limited to 50 ppmw TSS to prevent plugging the inlet plate and frame heat exchanger.

enters the brine concentrator and is combined with the brine slurry in the sump. The brine slurry is recirculated from the sump to a floodbox at the top of a bundle of heat transfer tubes. [98] As the brine falls through the heat transfer tubes to the sump, a portion of the water evaporates and passes through a mist eliminator (not shown) before entering the vapor compressor. [98] The compressed vapor is sent to the outside of the heat transfer tubes where it is condensed against the brine falling inside the tubes. The condensed distillate is pumped back through the HX, where it heats the incoming process water. A small amount of waste brine is blown down from the sump to control the total dissolved solids (TDS). [98]

Exhibit 4-14. Example diagram of a brine concentrator



Copyright General Electric Company; used with permission [97]

The brine concentrator is expected to produce an effluent with up to 500,000 TDS [99], with typical performance achieving between 200,000 and 300,000 TDS. [100] Given the elevated operating temperature and the use of NaOH in the upstream syngas scrubber for salt conversion, along with the high solubility of NaCl, it was assumed that 250,000 TDS would be achievable.

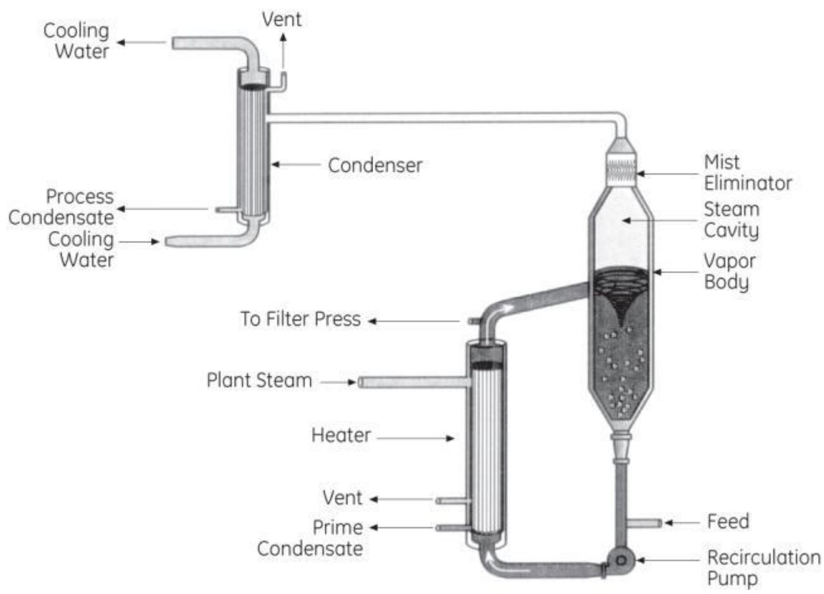
The brine concentrator operates at ambient pressure, and the vapor compressor varies the outlet pressure to ensure that sufficient heat is available to preheat the incoming process wastewater (0.14 MPa [20.6 psia]).

The brine concentrator must be constructed out of exotic materials to withstand the corrosive nature of the highly concentrated product stream. Either Inconel 625 or Hastelloy C276 can be used for piping and equipment construction; however, titanium would be required for the HX tubes.

4.4.14.2.3 Crystallizer

A crystallizer is often selected as a component of a wastewater treatment system and is utilized as the second step in this study. Exhibit 4-15 provides a depiction of a forced-circulation, steam-driven crystallizer [97] and consists of two HXs and a pump.

Exhibit 4-15. Example diagram of a forced-circulation, steam-driven crystallizer



Copyright General Electric Company; used with permission [97]

The process water (depicted as feed) from the brine concentrator is sent to the crystallizer sump, where the sump pump circulates the brine through a shell and tube HX. Because the tubes are flooded, the brine is under pressure and will not boil, which prevents scaling in the tubes. [97] The brine enters the crystallizer vapor body at an angle, where it swirls in a vortex. [97] As the water in the brine evaporates, crystals begin to form. The majority of the brine is recirculated back to the heater; however, approximately 20 percent [97] is blown down to the centrifuge/filter press for dewatering (depicted as "To Filter Press", but not shown). The water from the centrifuge and filter press is returned and mixed with the process water feed from the brine concentrator (not shown). The vapor from evaporation passes through a mist eliminator to remove entrained particles. [98] The product vapor is sent to a HX to be condensed against cooling water. The hot cooling water is returned to the cooling tower and the resulting condensate is utilized as process water makeup. An antifoam chemical feed is required to control foaming within the crystallizer.

An alternative configuration is to use a vapor compressor, rather than plant process steam, to provide heat to the system. In this alternative configuration, the compressed vapor heats the recirculating brine as it condenses. [98] The advantages of this process are the elimination of the condensing HX and process steam extraction in exchange for an electrical auxiliary load. Considering the energy penalties of using an electrical compressor versus process steam, along with the concern of sufficient vapor product to provide heat to the system, it was determined that a steam-driven crystallizer was most suitable for the cases in this study.

As with the brine concentrator, the crystallizer operates at ambient pressures and requires exotic materials of construction to withstand the corrosive concentrations of brine. Either Inconel 625 or Hastelloy C276 can be used for piping and equipment construction; however, titanium would be required for HX tubes.

The use of sulfuric acid for pH control in the brine concentrator results in the elimination of NaOH and the production of Na_2SO_4, which is removed as a dissolved solid in the moisture of the salt cake. It is assumed that the impurities present in the gasification wastewater result in a waste product, rather than a salable product, which is transported off-site to a waste-disposal site.

The combined distillate stream from the brine concentrator and crystallizer has a TDS level of less than 20 ppmw, generally consisting primarily of sodium and calcium as carryovers. For the purposes of this study, it is assumed that the combined distillate consists of less than 10 ppmw NaCl.

4.4.15 Accessory Electric Plant

The accessory electric plant is the same as in the reforming plants, described in Section 3.3.6.

4.4.16 Instrumentation and Control

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

This critical controller system is used to control the gasification process. The partial oxidation of the fuel feed and O_2 feed streams to form a syngas product is a stoichiometric, temperature- and pressure-dependent reaction. The critical controller utilizes a redundant microprocessor executing calculations and dynamic controls at 100- to 200-millisecond intervals. The enhanced execution speeds as well as evolved predictive controls allow the critical controller to mitigate process upsets and maintain the reactor operation within a stable set of operating parameters.

4.4.17 Miscellaneous Systems

An onsite water treatment facility treats all runoff, cleaning wastes, blowdown, and backwash. It is anticipated that the treated water will be suitable for discharge into existing systems and be within EPA standards for suspended solids, oil and grease, pH, and miscellaneous metals.

The waste treatment system is minimal and consists, primarily, of neutralization and oil/water separators (along with the associated pumps, piping, etc.).

A natural gas supply line has been included in all cases for start-up or emergency fuel.

Miscellaneous systems consisting of natural gas, service air, instrument air, and service water are provided. All truck roadways and unloading stations inside the fence area are provided.

4.4.18 Performance Summary Metrics

This section details the methodologies of several metrics reported in the performance summaries. The effective thermal efficiency is calculated in the same manner as in the reforming cases, shown in Section 3.3.10.1. Net power is a negative value in Case 4 and Case 5 and is a positive value in Case 6. The cold gas efficiency is also calculated using the equation described in Section 3.3.10.2.

4.4.18.1 Steam Turbine Efficiency

The steam turbine efficiency is calculated by taking the steam turbine power produced and dividing it by the difference between the thermal input and thermal output. This calculation is represented by the following equation:

$$STE = \frac{STP}{(TI - TO)} * 100\%$$

Where:

STE – steam turbine efficiency

STP – steam turbine power

TI – thermal input

TO – thermal output

Depending on the case, the thermal input is considered to include the main steam, makeup water, energy added during reheat, LP steam from the GFB, and/or IP steam from the WGS.

The IP blowdown, HP blowdown, and superheater losses are also credited to the thermal input as they are extracted from the cycle prior to the main steam but after the condensate boiler FW.

The thermal output is considered to be the BFW from the condenser and any extractions, such as the gasifier steam and the 1.7 MPa (250 psia) header.

4.4.18.2 Steam Turbine Heat Rate

The steam turbine heat rate is calculated by taking the inverse of the steam turbine efficiency. This calculation is represented by the following equation:

$$STHR = \frac{1}{STE} * 3{,}412 \text{ Btu/kWh}$$

Where:

STHR – steam turbine heat rate, Btu/kWh

STE – steam turbine efficiency, fraction

4.5 Case 4 – Shell Gasification Plant Without CO_2 Capture Process Description

This section contains an evaluation of a plant design for Case 4, which is based on a Shell coal gasification plant without CO_2 capture. The system descriptions follow the BFD provided in Exhibit 4-16 with the associated stream tables that show process data provided in Exhibit 4-17.

The balance of this section is organized as follows:

- Process Description – provides an overview of the technology operation as applied to Case 4. The systems that are common to all gasification cases were covered in Section 4.4 and only features that are unique to Case 4 are discussed further in this section
- Performance Results – provides the main modeling results from Case 4, including the performance summary, environmental performance, carbon balance, sulfur balance, water balance, energy balance, mass and energy balance diagrams. Note – some rounding error may be present in results reporting
- Equipment List – provides an itemized list of major equipment for Case 4
- Cost Estimates – provides a summary of capital and operating costs for Case 4

Exhibit 4-16. Case 4 block flow diagram, Shell gasification plant without CO_2 capture

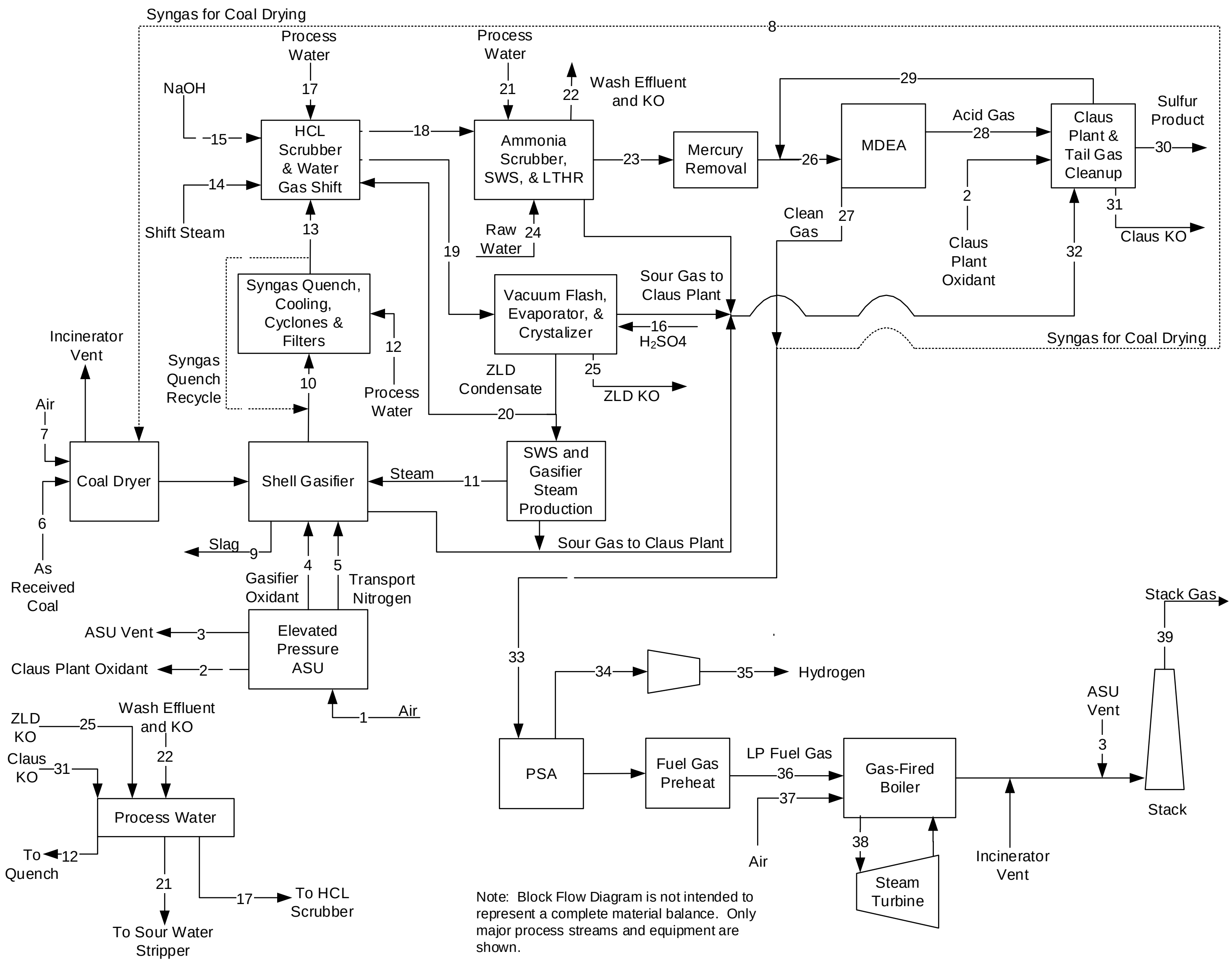


Exhibit 4-17. Case 4 stream table, Shell gasification plant without capture

	1	2	3	4	5	6	7	8	9	10
V-L Mole Fraction										
Ar	0.0092	0.0343	0.0025	0.0343	0.0000	0.0000	0.0092	0.0062	0.0000	0.0082
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	0.0000	0.0004
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0105	0.0000	0.5163
CO_2	0.0003	0.0000	0.0004	0.0000	0.0000	0.0000	0.0003	0.3805	0.0000	0.0170
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0006
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.5603	0.0000	0.2637
H_2O	0.0099	0.0000	0.0091	0.0000	0.0000	0.0000	0.0099	0.0017	0.0000	0.1293
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0073
H_2SO_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.7732	0.0157	0.9844	0.0157	0.9964	0.0000	0.7732	0.0391	0.0000	0.0516
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0051
O_2	0.2074	0.9501	0.0035	0.9501	0.0036	0.0000	0.2074	0.0014	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaOH	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	1.0000	0.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	24,168	188	17,942	5,017	945	0	1,089	488	0	26,522
V-L Flowrate (kg/hr)	697,414	6,061	501,902	161,609	26,484	0	31,415	9,560	0	536,840
Solids Flowrate (kg/hr)	0	0	0	0	0	211,967	0	0	21,230	0
Temperature (°C)	15	27	21	27	130	15	15	45	1,427	1,079
Pressure (MPa, abs)	0.10	0.86	0.45	5.10	5.62	0.10	0.10	3.04	4.24	4.24
Steam Table Enthalpy (kJ/kg)[A]	30.23	21.53	28.11	9.82	129.22	---	30.23	64.45	---	2,034.61
AspenPlus Enthalpy (kJ/kg)[B]	-97.58	-0.97	-97.36	-12.68	103.25	-2,119.02	-97.58	-7,702.04	2,165.43	-3,011.85
Density (kg/m^3)	1.2	11.2	5.2	68.6	46.4	---	1.2	22.9	---	7.6
V-L Molecular Weight	28.857	32.209	27.974	32.209	28.028	---	28.857	19.594	---	20.241
V-L Flowrate (lb_{mol}/hr)	53,281	415	39,555	11,062	2,083	0	2,400	1,076	0	58,471
V-L Flowrate (lb/hr)	1,537,535	13,363	1,106,504	356,286	58,388	0	69,258	21,076	0	1,183,529
Solids Flowrate (lb/hr)	0	0	0	0	0	467,308	0	0	46,803	0
Temperature (°F)	59	80	70	80	267	59	59	112	2,600	1,974
Pressure (psia)	14.7	125.0	65.0	740.0	815.0	14.7	14.7	441.1	615.0	615.0
Steam Table Enthalpy (Btu/lb)[A]	13.0	9.3	12.1	4.2	55.6	---	13.0	27.7	---	874.7
AspenPlus Enthalpy (Btu/lb)[B]	-42.0	-0.4	-41.9	-5.5	44.4	-911.0	-42.0	-3,311.3	931.0	-1,294.9
Density (lb/ft^3)	0.076	0.700	0.322	4.283	2.894	---	0.076	1.427	---	0.473

[A]Steam table reference conditions are 32.02 °F & 0.089 psia

[B]AspenPlus thermodynamic reference state is the component's constituent elements in an ideal gas state at 25 °C and 1 atm

Exhibit 4-17. Case 4 stream table, Shell gasification plant without capture (continued)

	11	12	13	14	15	16	17	18	19	20
V-L Mole Fraction										
Ar	0.0000	0.0000	0.0059	0.0000	0.0000	0.0000	0.0000	0.0048	0.0000	0.0000
CH_4	0.0000	0.0000	0.0003	0.0000	0.0000	0.0000	0.0000	0.0002	0.0000	0.0000
CO	0.0000	0.0000	0.3728	0.0000	0.0000	0.0000	0.0000	0.0083	0.0002	0.0000
CO_2	0.0000	0.0010	0.0125	0.0000	0.0000	0.0000	0.0007	0.3013	0.0001	0.0000
COS	0.0000	0.0000	0.0005	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.1904	0.0000	0.0000	0.0000	0.0000	0.4437	0.0001	0.0000
H_2O	0.9998	0.9798	0.3658	1.0000	0.6895	0.1000	0.9802	0.1999	0.9932	0.9993
HCl	0.0000	0.0000	0.0003	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2S	0.0000	0.0001	0.0053	0.0000	0.0000	0.0000	0.0001	0.0046	0.0001	0.0000
H_2SO_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.9000	0.0000	0.0000	0.0000	0.0000
N_2	0.0000	0.0000	0.0373	0.0000	0.0000	0.0000	0.0000	0.0299	0.0000	0.0000
NH_3	0.0002	0.0190	0.0090	0.0000	0.0000	0.0000	0.0190	0.0073	0.0038	0.0007
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0026	0.0000
NaOH	0.0000	0.0000	0.0000	0.0000	0.3105	0.0000	0.0000	0.0000	0.0001	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	1,130	10,213	29,021	8,247	29	0	711	36,155	3,480	1,627
V-L Flowrate (kg/hr)	20,363	184,094	569,529	148,564	731	13	12,810	697,887	63,058	29,312
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	343	188	232	288	16	15	66	202	190	88
Pressure (MPa, abs)	5.10	4.98	3.98	3.88	4.81	0.13	0.47	3.49	3.88	0.13
Steam Table Enthalpy (kJ/kg)[A]	3,083.36	768.16	1,189.25	2,971.61	-338.78	-8,206.86	237.61	803.97	792.35	368.05
AspenPlus Enthalpy (kJ/kg)[B]	-12,884.30	-14,942.19	-6,580.56	-13,008.68	-13,665.00	-8,526.27	-15,478.76	-8,422.21	-15,049.89	-15,593.15
Density (kg/m^3)	19.9	836.4	19.0	16.9	1,531.7	1,791.5	965.1	17.2	873.9	965.6
V-L Molecular Weight	18.015	18.025	19.625	18.015	24.842	90.073	18.017	19.303	18.120	18.015
V-L Flowrate (lb_{mol}/hr)	2,492	22,516	63,979	18,181	65	0	1,567	79,708	7,672	3,587
V-L Flowrate (lb/hr)	44,893	405,859	1,255,596	327,527	1,612	29	28,241	1,538,578	139,020	64,622
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	650	371	450	550	61	59	151	395	373	191
Pressure (psia)	740.0	722.7	577.7	562.7	697.7	18.2	67.7	506.1	562.7	19.3
Steam Table Enthalpy (Btu/lb)[A]	1,325.6	330.2	511.3	1,277.6	-145.7	-3,528.3	102.2	345.6	340.7	158.2
AspenPlus Enthalpy (Btu/lb)[B]	-5,539.3	-6,424.0	-2,829.1	-5,592.7	-5,874.9	-3,665.6	-6,654.7	-3,620.9	-6,470.3	-6,703.8
Density (lb/ft^3)	1.240	52.218	1.183	1.055	95.623	111.841	60.248	1.075	54.557	60.280

[A]Steam table reference conditions are 32.02 °F & 0.089 psia

[B]AspenPlus thermodynamic reference state is the component’s constituent elements in an ideal gas state at 25 °C and 1 atm

Exhibit 4-17. Case 4 stream table, Shell gasification plant without capture (continued)

	21	22	23	24	25	26	27	28	29	30
V-L Mole Fraction										
Ar	0.0000	0.0000	0.0060	0.0000	0.0000	0.0059	0.0062	0.0000	0.0035	0.0000
CH_4	0.0000	0.0000	0.0003	0.0000	0.0000	0.0003	0.0003	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0105	0.0000	0.0000	0.0099	0.0105	0.0001	0.0000	0.0000
CO_2	0.0007	0.0010	0.3784	0.0000	0.0002	0.4133	0.3805	0.9074	0.9512	0.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.5598	0.0000	0.0000	0.5256	0.5603	0.0042	0.0000	0.0000
H_2O	0.9802	0.9792	0.0016	0.9999	0.9833	0.0016	0.0017	0.0010	0.0025	0.0000
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2S	0.0001	0.0001	0.0057	0.0000	0.0003	0.0054	0.0000	0.0856	0.0000	0.0000
H_2SO_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0000	0.0000	0.0378	0.0000	0.0000	0.0367	0.0391	0.0002	0.0196	0.0000
NH_3	0.0190	0.0196	0.0000	0.0000	0.0162	0.0000	0.0000	0.0002	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0013	0.0014	0.0000	0.0218	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0000	0.0014	0.0014	0.0000
NaCl	0.0000	0.0000	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaOH	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000
V-L Flowrate (kg_{mol}/hr)	3,367	13,314	28,649	2,571	684	30,511	28,608	1,903	1,862	0
V-L Flowrate (kg/hr)	60,667	240,009	561,548	46,318	12,326	642,318	560,536	81,782	80,770	0
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	5,280
Temperature (°C)	66	69	29	15	30	37	45	45	38	182
Pressure (MPa, abs)	0.47	0.45	3.18	0.10	0.24	3.07	3.04	3.04	3.07	0.29
Steam Table Enthalpy (kJ/kg)[A]	237.61	247.91	39.43	62.75	91.05	45.70	64.45	9.96	5.20	---
AspenPlus Enthalpy (kJ/kg)[B]	-15,478.76	-15,454.40	-7,686.77	-15,905.25	-15,665.35	-7,799.78	-7,702.04	-8,376.94	-8,669.58	145.98
Density (kg/m^3)	965.1	962.4	25.2	999.4	985.7	25.6	22.9	58.0	60.7	5,270.5
V-L Molecular Weight	18.017	18.026	19.601	18.019	18.008	21.052	19.594	42.976	43.382	---
V-L Flowrate (lb_{mol}/hr)	7,423	29,353	63,161	5,667	1,509	67,266	63,070	4,195	4,105	0
V-L Flowrate (lb/hr)	133,748	529,130	1,238,001	102,113	27,175	1,416,069	1,235,771	180,298	178,068	0
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	11,640
Temperature (°F)	151	156	84	59	85	98	112	112	100	360
Pressure (psia)	67.7	65.0	460.7	14.7	35.0	445.2	441.1	441.1	445.2	41.7
Steam Table Enthalpy (Btu/lb)[A]	102.2	106.6	17.0	27.0	39.1	19.6	27.7	4.3	2.2	---
AspenPlus Enthalpy (Btu/lb)[B]	-6,654.7	-6,644.2	-3,304.7	-6,838.0	-6,734.9	-3,353.3	-3,311.3	-3,601.4	-3,727.2	62.8
Density (lb/ft^3)	60.248	60.083	1.572	62.391	61.535	1.599	1.427	3.622	3.787	329.025

[A]Steam table reference conditions are 32.02 °F & 0.089 psia

[B]AspenPlus thermodynamic reference state is the component's constituent elements in an ideal gas state at 25 °C and 1 atm

Exhibit 4-17. Case 4 stream table, Shell gasification plant without capture (continued)

	31	32	33	34	35	36	37	38	39
V-L Mole Fraction									
Ar	0.0000	0.0004	0.0062	0.0008	0.0008	0.0112	0.0092	0.0000	0.0074
CH_4	0.0000	0.0000	0.0003	0.0000	0.0000	0.0005	0.0000	0.0000	0.0000
CO	0.0000	0.0042	0.0105	0.0000	0.0000	0.0201	0.0000	0.0000	0.0000
CO_2	0.0001	0.3168	0.3805	0.0000	0.0000	0.7271	0.0003	0.0000	0.2330
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0282	0.5603	0.9990	0.9990	0.1606	0.0000	0.0000	0.0000
H_2O	0.9999	0.1576	0.0017	0.0000	0.0000	0.0032	0.0099	1.0000	0.0786
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2S	0.0000	0.0123	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2SO_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0000	0.0014	0.0391	0.0002	0.0002	0.0745	0.7732	0.0000	0.6422
NH_3	0.0000	0.4791	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.0014	0.0000	0.0000	0.0027	0.2074	0.0000	0.0388
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaOH	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	294	137	28,120	13,406	13,406	14,714	14,578	13,684	48,104
V-L Flowrate (kg/hr)	5,303	3,510	550,976	27,498	27,498	523,479	420,681	246,512	1,500,775
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0
Temperature (°C)	49	160	45	45	30	193	15	482	68
Pressure (MPa, abs)	0.27	0.45	3.04	2.97	6.48	0.50	0.10	11.82	0.11
Steam Table Enthalpy (kJ/kg)[A]	105.73	505.34	64.45	634.68	439.90	204.42	30.23	3,302.61	181.25
AspenPlus Enthalpy (kJ/kg)[B]	-15,863.29	-7,054.88	-7,702.04	284.27	89.50	-7,951.62	-97.58	-12,677.68	-3,504.91
Density (kg/m^3)	963.8	3.2	22.9	2.3	5.1	4.6	1.2	38.0	1.2
V-L Molecular Weight	18.017	25.588	19.594	2.051	2.051	35.576	28.857	18.015	31.198
V-L Flowrate (lb_{mol}/hr)	649	302	61,995	29,555	29,555	32,439	32,139	30,167	106,052
V-L Flowrate (lb/hr)	11,692	7,739	1,214,695	60,622	60,622	1,154,073	927,443	543,466	3,308,642
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0
Temperature (°F)	120	321	112	112	86	380	59	900	154
Pressure (psia)	39.5	65.0	441.1	431.1	940.0	72.5	14.7	1,714.7	15.4
Steam Table Enthalpy (Btu/lb)[A]	45.5	217.3	27.7	272.9	189.1	87.9	13.0	1,419.9	77.9
AspenPlus Enthalpy (Btu/lb)[B]	-6,820.0	-3,033.1	-3,311.3	122.2	38.5	-3,418.6	-42.0	-5,450.4	-1,506.8
Density (lb/ft^3)	60.168	0.201	1.427	0.142	0.319	0.287	0.076	2.372	0.073

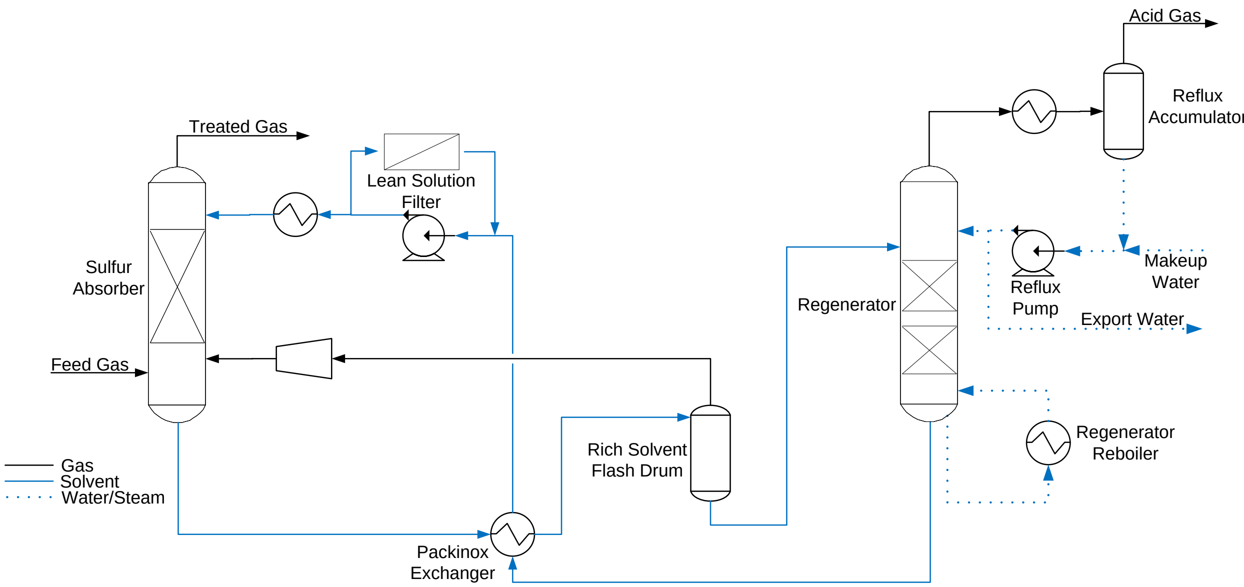
[A]Steam table reference conditions are 32.02 °F & 0.089 psia

[B]AspenPlus thermodynamic reference state is the component's constituent elements in an ideal gas state at 25 °C and 1 atm

4.5.1 Acid Gas Removal

H_2S removal generally consists of absorption by a regenerable solvent. The most commonly used technique is based on counter current contact with the solvent. Acid-gas-rich solution from the absorber is stripped of its acid gas in a regenerator, usually by application of heat. The regenerated lean solution is then cooled and recirculated to the top of the absorber, completing the cycle. Exhibit 4-18 is a simplified diagram of the AGR process. [101]

Exhibit 4-18. Flow diagram for a conventional single-stage AGR unit



There are well over 30 AGR processes in common commercial use throughout the oil, chemical, and natural gas industries. However, in a 2002 report by Sfa Pacific, Inc. a list of 42 operating and planned gasifiers shows that only six AGR processes are represented: Rectisol, Sulfinol-M, MDEA, Selexol, aqueous di-isopropanol (ADIP) amine, and FLEXSORB. [102] These processes can be separated into three general types: chemical reagents, physical solvents, and hybrid solvents.

There are numerous commercial AGR processes that could meet the sulfur environmental target of this report. The most frequently used AGR systems (Selexol, Sulfinol-M, MDEA, and Rectisol) have all been used with the Shell gasifiers in various applications.

For the non-capture case, Case 4, the MDEA system is chosen due to its low CO_2 selectivity which is preferred when treating a shifted syngas without the separation of CO_2. AGR technologies with higher CO_2 selectivities result in a buildup of CO_2 in the acid gas stream and the sulfur recovery unit due to the recycling of Claus tail gas to the AGR. Additionally, refrigerated MDEA is chosen over conventional MDEA because the sulfur emissions environmental target chosen is just below the range of the conventional, higher temperature MDEA system.

Cool, particulate-free syngas (stream 26) enters the absorber unit at approximately 3.1 MPa (445 psia) and 37 °C (98 °F). In the absorber, H_2S is preferentially removed from the syngas stream by contact with MDEA. The absorber column is operated at 45 °C (112 °F) by refrigerating the lean MDEA solvent. An outlet H_2S concentration of less than 40 ppmv in the sweet syngas is achieved. The stripper acid gas stream (stream 28), consisting of 9 vol% H_2S and 91 vol% CO_2, is sent to the Claus unit.

4.5.2 Hydrogen Production

The remaining clean syngas exiting the MDEA unit (stream 27) that is not used for coal drying (stream 33) is sent to a pressure swing adsorption (PSA) unit. The PSA produces a highly pure hydrogen product stream (stream 34) as well as a low pressure and low temperature off-gas stream that is used as fuel gas in the GFB. The PSA unit produces 99.9 vol% purity hydrogen at a H_2 recovery efficiency of 85 percent. The hydrogen stream is compressed using an integrally geared, multi-stage, intercooled centrifugal compressor to a pipeline-ready pressure of 6.48 MPa (925 psig).

4.5.3 Steam Generation

The steam conditions for the coal gasification plant without capture (Case 4) are 11.7 MPa/482 °C/482 °C (1,700 psig/900 °F/900 °F).

4.6 CASE 4 – PERFORMANCE RESULTS

Overall performance for the entire plant is summarized in Exhibit 4-19. The plant produces 27,498 kg/hr (60,622 lb/hr) of hydrogen at an effective thermal efficiency of 65.0 percent (HHV basis). The steam turbine generates 89 MWe and the total auxiliary load for the plant is 114 MWe, requiring 25 MW of electricity to be sourced from the grid.

Exhibit 4-20 provides a detailed breakdown of the auxiliary power requirements. The ASU accounts for approximately 67 percent of the total auxiliary load distributed between the MAC, booster compressor, N_2 compressor, O_2 pump, and ASU auxiliaries. The hydrogen compressor accounts for about 9 percent of the total auxiliary load. The BFW pumps and cooling water system, including the circulating water pumps and cooling tower fan, accounts for approximately 5 percent of the auxiliary load. All other systems together constitute the remaining 19 percent of the auxiliary load.

Exhibit 4-19. Case 4 plant performance summary

Performance Summary	
Steam Turbine Power, MWe	89
Total Gross Power, MWe	**89**
Air Separation Unit Main Air Compressor, kWe	66,970
Air Separation Unit Booster Compressor, kWe	5,270
N_2 Compressors, kWe	2,390
CO_2 Compression, kWe	0
Acid Gas Removal, kWe	5,610
Balance of Plant, kWe	33,350
Total Auxiliaries, MWe	**114**
Net Power, MWe	**-25**
Hydrogen Production, kg/hr (lb/hr)	**27,498 (60,622)**
CO_2 Capture, %	**–**
HHV Effective Thermal Efficiency[A], %	65.0%
HHV Cold Gas Efficiency[B], %	66.6%
LHV Effective Thermal Efficiency[A], %	56.8%
LHV Cold Gas Efficiency[B], %	58.4%
Steam Turbine Cycle Efficiency, %	52.9%
Steam Turbine Heat Rate, kJ/kWh (Btu/kWh)	6,799 (6,445)
Condenser Duty, GJ/hr (MMBtu/hr)	366 (347)
AGR Cooling Duty, GJ/hr (MMBtu/hr)	351 (333)
As-Received Coal Feed, kg/hr (lb/hr)	211,967 (467,308)
HHV Thermal Input, kWt	1,597,711
LHV Thermal Input, kWt	1,541,012
Raw Water Withdrawal, m^3/min (gpm)	12.1 (3,188)
Raw Water Consumption, m^3/min (gpm)	9.5 (2,516)

[A]ETE = (Hydrogen Heating Value + Net Power) / Fuel Heating Value
[B]CGE = Hydrogen Heating Value / Fuel Heating Value

Exhibit 4-20. Case 4 plant power summary

Power Summary	
Steam Turbine Power, MWe	89
Total Gross Power, MWe	**89**
Auxiliary Load Summary	
Acid Gas Removal, kWe	5,610
Air Blower, kWe	1,080
Air Separation Unit Auxiliaries, kWe	1,000
Air Separation Unit Main Air Compressor, kWe	66,970
Air Separation Unit Booster Compressor, kWe	5,270
Ammonia Wash Pumps, kWe	120
Circulating Water Pumps, kWe	2,910
Claus Plant TG Recycle Compressor, kWe	4,230
Claus Plant/TGTU Auxiliaries, kWe	680
CO_2 Compression, kWe	0
Coal Handling, kWe	460
Coal Milling, kWe	2,180
Condensate Pumps, kWe	110
Cooling Tower Fans, kWe	1,500
Dryer Air Compressor, kWe	80
Feedwater Pumps, kWe	1,540
Gasifier Water Pump, kWe	40
Ground Water Pumps, kWe	290
Hydrogen Compressor, kWe	10,430
Miscellaneous Balance of Plant[A], kWe	1,930
N_2 Compressors, kWe	2,390
O_2 Pump, kWe	330
Quench Water Pump, kWe	330
Shift Steam Pump, kWe	310
Slag Handling, kWe	550
Slag Reclaim Water Recycle Pump, kWe	0
Slurry Water Pump, kWe	0
Sour Gas Compressors, kWe	140
Sour Water Recycle Pumps, kWe	0
Steam Turbine Auxiliaries, kWe	80
Syngas Recycle Compressor, kWe	880
Syngas Scrubber Pumps, kWe	120
Process Water Treatment Auxiliaries, kWe	1,310
Transformer Losses, kWe	720
Total Auxiliaries, MWe	**114**
Net Power, MWe	**-25**

[A]Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

4.6.1 Environmental Performance

The environmental targets for emissions of Hg, HCl, NOx, SO_2, and PM were presented in Section 2.7.1. A summary of the plant air emissions for Case 4 is presented in Exhibit 4-21.

Exhibit 4-21. Case 4 air emissions

	kg/GJ (lb/MMBtu)	Tonne/year (ton/year)[A]	lb/lb H_2
SO_2	0.011 (0.027)	462 (509)	0.002
NOx	0.007 (0.017)	291 (321)	0.002
Particulate	0.003 (0.007)	123 (136)	0.001
Hg	2.46E-7 (5.71E-7)	0.010 (0.011)	5.14E-8
HCl	0.000 (0.000)	0.00 (0.00)	0.000
CO_2[B]	86 (199)	3,456,164 (3,809,769)	17.9
CO_2e[C]	95 (222)	3,845,486 (4,238,922)	20.

[A]Calculations based on an 80 percent capacity factor
[B]Plant stack emissions
[C]Life cycle emissions

The low level of SO_2 emissions is achieved by capturing the sulfur in the syngas by the MDEA AGR process. The AGR process removes over 99 percent of the sulfur compounds in the fuel gas down to a level of less than 36 ppmv. The H_2S-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur. The Claus plant tail gas is compressed and recycled back to the AGR to capture most of the remaining sulfur. The SO_2 emissions in Exhibit 4-21 include both the stack emissions and the coal dryer emissions.

NOx emissions are limited by the use of low NOx burners to 15 ppmvd (as NO at 15 percent O_2). NH_3 in the syngas is removed with process condensate prior to the low-temperature AGR process and destroyed in the Claus plant burner. This helps lower NOx levels as well.

Particulate discharge to the atmosphere is limited to extremely low values by the use of a cyclone and a barrier filter in addition to the syngas scrubber and the gas washing effect of the AGR absorber. The particulate emissions represent filterable particulate only.

Approximately 97 percent of the mercury is captured from the syngas by dual activated carbon beds.

Plant stack CO_2 emissions represent the uncontrolled discharge from the process. Life cycle CO_2 emissions represent the total GWP, in CO_2e emissions, of the process over the plant life cycle, as detailed in Section 2.8.

The carbon balance for the plant is shown in Exhibit 4-22. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal. Carbon in the air is not neglected here since the Aspen model accounts for air components throughout. Carbon leaves the plant as unburned carbon in the slag, a small portion of the hydrogen product, and the CO_2 in the stack gas, including the coal dryer vent gas and ASU vent gas.

Exhibit 4-22. Case 4 carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	135,118 (297,884)	Stack Gas	134,595 (296,732)
Air (CO_2)	156 (345)	CO_2 Product	–
		Slag	676 (1,489)
		H_2 Product	3 (7)
Total	**135,274 (298,229)**	**Total**	**135,274 (298,229)**

Exhibit 4-23 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant as elemental sulfur and sulfur emitted in the stack gas, including the coal drying vent. Sulfur in the slag is considered negligible.

Exhibit 4-23. Case 4 sulfur balance

Sulfur In		Sulfur Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	5,313 (11,713)	Stack Gas	33 (73)
		CO_2 Product	–
		Elemental Sulfur	5,280 (11,640)
Total	**5,313 (11,713)**	**Total**	**5,313 (11,713)**

Exhibit 4-24 shows the overall water balance for the plant. The primary water demand comes from the cooling tower (11.32 m^3/min), which makes up about 60 percent of the total water demand (18.94 m^3/min). The largest contributor to the internal recycle is the quench, making up about 45 percent of the total internal recycle (6.87 m^3/min). Process water is discharged from the cooling tower at a rate of 2.55 m^3/min. The raw water consumption, which can be described as the water demand minus the internal recycle and process water discharge, is 9.52 m^3/min in total for the coal gasification plant without capture (Case 4).

Exhibit 4-24. Case 4 water balance

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)
Slag Handling	0.46 (122)	0.46 (122)	–	–	–
Slurry Water	–	–	–	–	–
Gasifier Water	–	–	–	–	–
Quench	3.07 (812)	3.07 (812)	–	–	–

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)
HCl Scrubber	1.88 (496)	1.88 (496)	–	–	–
NH_3 Scrubber	1.76 (465)	0.99 (261)	0.77 (204)	–	0.77 (204)
Gasifier Steam	0.34 (90)	0.34 (90)	–	–	–
Condenser Makeup	**0.11 (29)**	**–**	**0.11 (29)**	**–**	**0.11 (29)**
BFW Makeup	0.11 (29)	–	0.11 (29)	–	0.11 (29)
Gasifier Steam	–	–	–	–	–
Shift Steam	–	–	–	–	–
Cooling Tower	**11.32 (2,990)**	**0.13 (35)**	**11.19 (2,955)**	**2.55 (672)**	**8.64 (2,283)**
BFW Blowdown	–	0.11 (29)	-0.11 (-29)	–	-0.11 (-29)
ASU Knockout	–	0.02 (6)	-0.02 (-6)	–	-0.02 (-6)
Total	**18.94 (5,004)**	**6.87 (1,815)**	**12.07 (3,188)**	**2.55 (672)**	**9.52 (2,516)**

An overall plant energy balance is provided in tabular form in Exhibit 4-25. The power out is the steam turbine power prior to generator losses.

Exhibit 4-25. Case 4 overall energy balance (0 °C [32 °F] reference)

	HHV	Sensible + Latent	Power	Total
Heat In, GJ/hr (MMBtu/hr)				
Coal	5,752 (5,452)	4.8 (4.6)	–	5,757 (5,456)
Air	–	34.7 (32.9)	–	34.7 (32.9)
Raw Water Makeup	–	45.4 (43.0)	–	45.4 (43.0)
Auxiliary Power	–	–	408.9 (387.6)	408.9 (387.6)
Total	**5,752 (5,452)**	**84.9 (80.5)**	**408.9 (387.6)**	**6,246 (5,920)**
Heat Out, GJ/hr (MMBtu/hr)				
Misc. Process Steam	–	5.0 (4.7)	–	5.0 (4.7)
Slag	22.1 (21.0)	35.8 (34.0)	–	58.0 (55.0)
Stack Gas	–	272 (258)	–	272 (258)
Sulfur	48.9 (46.4)	0.6 (0.6)	–	49.5 (47.0)
Motor Losses and Design Allowances	–	–	14.7 (13.9)	14.7 (13.9)
Hydrogen Product	3,828 (3,628)	12.1 (11.5)	–	3,840 (3,640)
Cooling Tower Load[A]	–	1,478 (1,401)	–	1,478 (1,401)
CO_2 Product Stream	–	–	–	–

	HHV	Sensible + Latent	Power	Total
Blowdown Streams	–	22.9 (21.7)	–	22.9 (21.7)
Ambient Losses[B]	–	89.8 (85.1)	–	89.8 (85.1)
Power	–	–	319 (302)	319 (302)
Total	**3,899 (3,696)**	**1,916 (1,816)**	**334 (316)**	**6,149 (5,828)**
Unaccounted Energy[C]	–	96.4 (91.4)	–	96.4 (91.4)

[A]Includes condenser, AGR, and miscellaneous cooling loads
[B]Ambient losses include all losses to the environment through radiation, convection, etc. Sources of these losses include the combustor, reheater, superheater, and transformers
[C]By difference

4.6.2 Energy and Mass Balance Diagrams

Energy and mass balance diagrams are shown for the following subsystems in Exhibit 4-26 through Exhibit 4-28:

- ASU, gasification, and gas cooling
- Gas cleanup system
- Hydrogen purification and power block

Exhibit 4-26. Case 4 coal gasification and ASU energy and mass balance

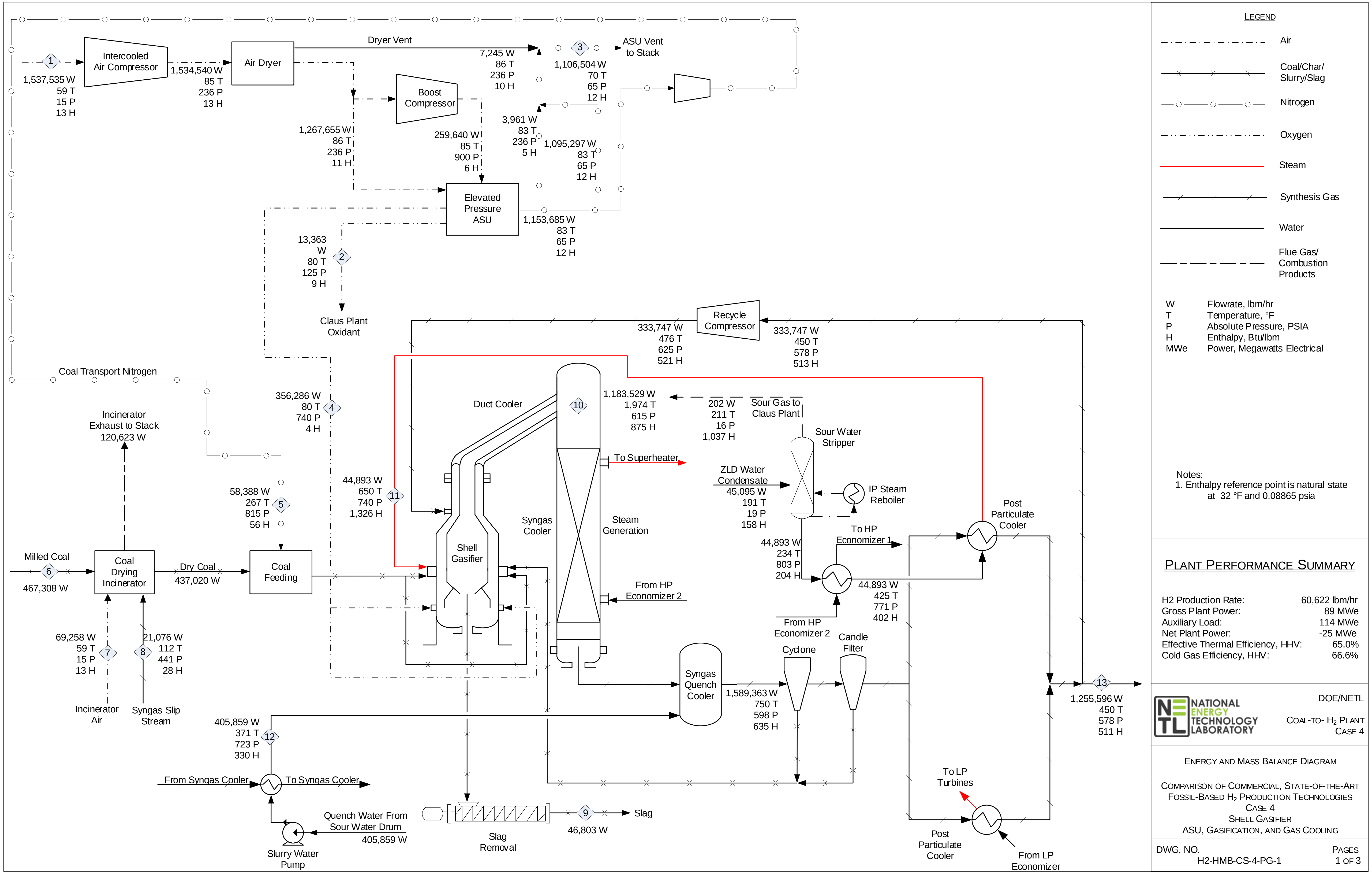


Exhibit 4-27. Case 4 syngas cleanup energy and mass balance

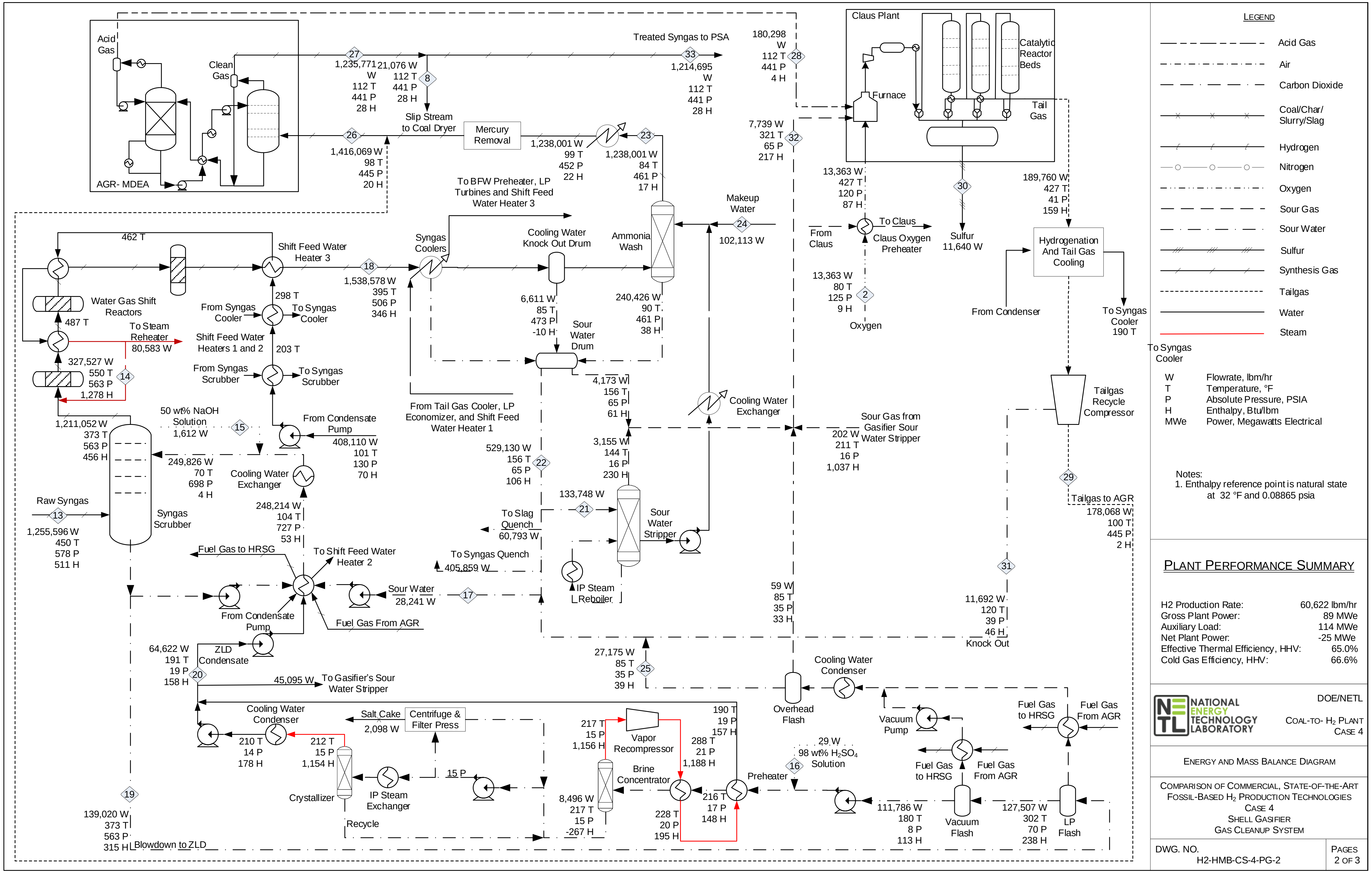
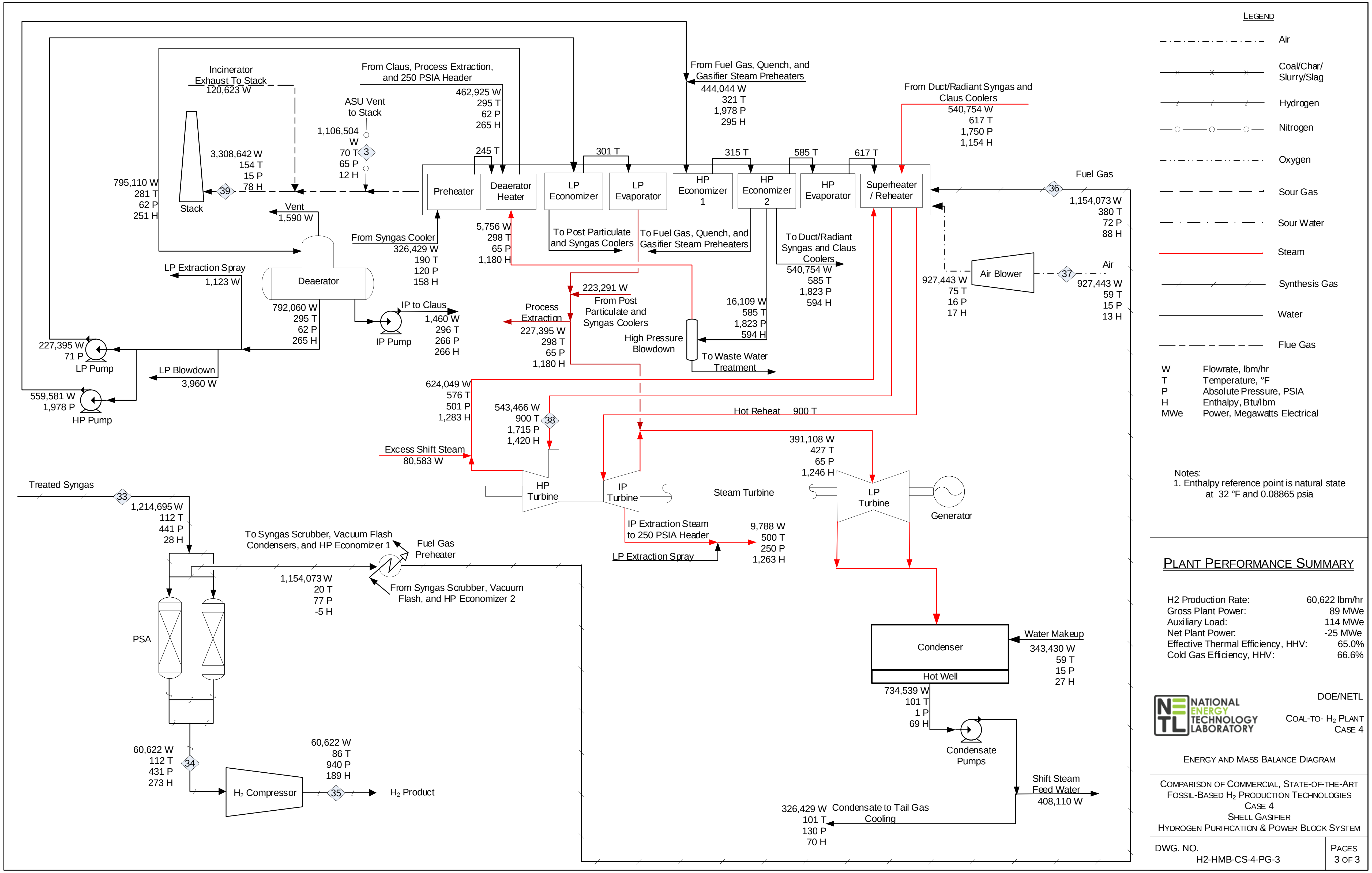


Exhibit 4-28. Case 4 hydrogen purification and power block energy and mass balance



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4.7 Case 4 – Major Equipment List

Major equipment items for the Shell gasifier with no CO_2 capture are shown in the following tables. In general, the design conditions include a 10 percent design allowance for flows and heat duties and a 21 percent design allowance for heads on pumps and fans.

Case 4 – Account 1: Coal Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	181 tonne (200 ton)	2	0
2	Feeder	Belt	570 tonne/hr (630 tph)	2	0
3	Conveyor No. 1	Belt	1,130 tonne/hr (1,250 tph)	1	0
4	Conveyor No. 2	Belt	1,130 tonne/hr (1,250 tph)	1	0
5	Stacker/Reclaimer	Traveling, linear	1,130 tonne/hr (1,250 tph)	1	0
6	Reclaim Hopper	N/A	40 tonne (50 ton)	2	1
7	Feeder	Vibratory	170 tonne/hr (190 tph)	2	1
8	Conveyor No. 3	Belt w/ tripper	350 tonne/hr (390 tph)	1	0
9	Coal Surge Bin w/ Vent Filter	Dual outlet	170 tonne (190 ton)	2	0
10	Crusher	Impactor reduction	8 cm x 0 – 3 cm x 0 (3" x 0 - 1-1/4" x 0)	2	0
11	Conveyor No. 4	Belt w/tripper	350 tonne/hr (390 tph)	1	0
12	Conveyor No. 5	Belt w/ tripper	350 tonne/hr (390 tph)	1	0
13	Coal Silo w/ Vent Filter and Slide Gates	Field erected	780 tonne (860 ton)	3	0

Case 4 – Account 2: Coal Preparation and Feed

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Feeder	Vibratory	80 tonne/hr (90 tph)	3	0
2	Conveyor No. 6	Belt w/tripper	230 tonne/hr (260 tph)	1	0
3	Roller Mill Feed Hopper	Dual Outlet	470 tonne (510 ton)	1	0
4	Weigh Feeder	Belt	120 tonne/hr (130 tph)	2	0
5	Coal Dryer and Pulverizer	Rotary	120 tonne/hr (130 tph)	2	0
6	Coal Dryer Feed Hopper	Vertical Hopper	230 tonne (260 ton)	2	0

Case 4 – Account 3: Feedwater and Miscellaneous Balance of Plant Systems

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Demineralized Water Storage Tank	Vertical, cylindrical, outdoor	10,288,000 liters (2,718,000 gal)	2	0
2	Condensate Pumps	Vertical canned	3,080 lpm @ 90 m H_2O (810 gpm @ 300 ft H_2O)	2	1
3	Deaerator (integral w/ GFB)	Horizontal spray type	198,000 kg/hr (437,000 lb/hr)	2	0
4	Intermediate Pressure Feedwater Pump	Horizontal centrifugal, single stage	980 lpm @ 10 m H_2O (260 gpm @ 20 ft H_2O)	2	1
5	High Pressure Feedwater Pump No. 1	Barrel type, multi-stage, centrifugal	HP water: 2,420 lpm @ 1,600 m H_2O (640 gpm @ 5,300 ft H_2O)	2	1
6	High Pressure Feedwater Pump No. 2	Barrel type, multi-stage, centrifugal	IP water: 1,920 lpm @ 190 m H_2O (510 gpm @ 640 ft H_2O)	2	1
7	Auxiliary Boiler	Shop fabricated, water tube	18,000 kg/hr, 2.8 MPa, 343 °C (40,000 lb/hr, 400 psig, 650 °F)	1	0
8	Service Air Compressors	Flooded Screw	28 m^3/min @ 0.7 MPa (1,000 scfm @ 100 psig)	2	1

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
9	Instrument Air Dryers	Duplex, regenerative	28 m^3/min (1,000 scfm)	2	1
10	Closed Cycle Cooling Heat Exchangers	Plate and frame	418 GJ/hr (397 MMBtu/hr) each	2	0
11	Closed Cycle Cooling Water Pumps	Horizontal centrifugal	150,200 lpm @ 20 m H_2O (39,700 gpm @ 70 ft H_2O)	2	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	3,785 lpm @ 110 m H_2O (1,000 gpm @ 350 ft H_2O)	1	1
13	Fire Service Booster Pump	Two-stage horizontal centrifugal	2,650 lpm @ 80 m H_2O (700 gpm @ 250 ft H_2O)	1	1
14	Municipal Water Pumps	Stainless steel, single suction	1,470 lpm @ 20 m H_2O (390 gpm @ 60 ft H_2O)	2	1
15	Ground Water Pumps	Stainless steel, single suction	2,930 lpm @ 270 m H_2O (770 gpm @ 880 ft H_2O)	1	1
16	Filtered Water Pumps	Stainless steel, single suction	1,850 lpm @ 50 m H_2O (490 gpm @ 160 ft H_2O)	2	1
17	Filtered Water Tank	Vertical, cylindrical	890,000 liter (235,000 gal)	2	0
18	Makeup Water Demineralizer	Anion, cation, and mixed bed	1,430 lpm (380 gpm)	2	0
19	Liquid Waste Treatment System	N/A	10 years, 24-hour storm	1	0
20	Process Water Treatment	Vacuum flash, brine concentrator, and crystallizer	Vacuum Flash Inlet: 35,000 kg/hr (76,000 lb/hr) Vacuum Flash Outlet: 6,218 ppmw Cl- Brine Concentrator Inlet: 28,000 kg/hr (61,000 lb/hr) Crystallizer Inlet: 2,000 kg/hr (5,000 lb/hr)	2	0

Case 4 – Account 4: Gasifier, ASU, and Accessories

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Gasifier	Pressurized dry-feed, entrained bed	2,800 tonne/day, 4.2 MPa (3,100 tpd, 615 psia)	2	0
2	Synthesis Gas Cooler	Convective spiral-wound tube boiler	295,000 kg/hr (651,000 lb/hr)	2	0
3	Synthesis Gas Cyclone	High efficiency	397,000 kg/hr (874,000 lb/hr) Design efficiency 90%	2	0
4	HCl Scrubber	Ejector Venturi	313,000 kg/hr (691,000 lb/hr)	2	0
5	Ammonia Wash	Counter-flow spray tower	311,000 kg/hr (685,000 lb/hr) @ 3.3 MPa (473 psia)	2	0
6	Primary Sour Water Stripper	Counter-flow with external reboiler	33,000 kg/hr (74,000 lb/hr)	2	0
7	Secondary Sour Water Stripper	Counter-flow with external reboiler	11,000 kg/hr (25,000 lb/hr)	2	0
8	Low Temperature Heat Recovery Coolers	Shell and tube with condensate drain	256,000 kg/hr (564,000 lb/hr)	9	0
9	Low Temperature Heat Recovery Knockout Drum	Vertical with mist eliminator	312,000 kg/hr, 59 °C, 3.3 MPa (689,000 lb/hr, 138 °F, 476 psia)	2	0
10	Synthesis Gas Reheaters	Shell and tube	Reheater 1: 0 kg/hr (0 lb/hr) Reheater 2: 54,000 kg/hr (119,000 lb/hr) Reheater 3: 61,000 kg/hr (135,000 lb/hr) Reheater 4: 173,000 kg/hr (382,000 lb/hr) Reheater 5: 288,000 kg/hr (635,000 lb/hr) Reheater 6: 288,000 kg/hr (635,000 lb/hr)	2	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
11	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	302,000 kg/hr (666,000 lb/hr) syngas	2	0
12	ASU Main Air Compressor	Centrifugal, multi-stage	5,000 m^3/min @ 1.6 MPa (185,000 scfm @ 236 psia)	2	0
13	Cold Box	Vendor design	2,200 tonne/day (2,400 tpd) of 95% purity oxygen	2	0
14	Gasifier O_2 Pump	Centrifugal, multi-stage	1,000 m^3/min (38,000 scfm) Suction – 1.0 MPa (130 psia) Discharge – 5.1 MPa (740 psia)	2	0
15	Gasifier Nitrogen Boost Compressor	Centrifugal, single-stage	210 m^3/min (7,000 scfm) Suction – 0.4 MPa (70 psia) Discharge – 5.6 MPa (820 psia)	2	0

Case 4 – Account 5: Syngas Cleanup

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Mercury Adsorber 1	Sulfated carbon bed	309,000 kg/hr (681,000 lb/hr) 29 °C (84 °F) 3.2 MPa (461 psia)	2	0
2	Mercury Adsorber 2	Sulfated carbon bed	309,000 kg/hr (681,000 lb/hr) 37 °C (99 °F) 3.1 MPa (448 psia)	2	0
3	Sulfur Plant	Claus type	139 tonne/day (154 tpd)	1	0
4	Water Gas Shift Reactors	Fixed bed, catalytic	256,000 kg/hr (282,000 lb/hr) 216 °C (420 °F) 3.9 MPa (560 psia)	6	0
5	Shift Reactor Heat Recovery Exchangers	Shell and Tube	Exchanger 1: 223 GJ/hr (211 MMBtu/hr) Exchanger 2: 91 GJ/hr (86 MMBtu/hr) Exchanger 3: 66 GJ/hr (63 MMBtu/hr) Exchanger 4: 83 GJ/hr (78 MMBtu/hr)	8	0
6	Acid Gas Removal Plant	MDEA	353,000 kg/hr (779,000 lb/hr) 37 °C (98 °F) 3.1 MPa (445 psia)	2	0
7	Hydrogenation Reactor	Fixed bed, catalytic	95,000 kg/hr (209,000 lb/hr) 219 °C (427 °F) 0.3 MPa (40.8 psia)	1	0
8	Tail Gas Recycle Compressor	Centrifugal	89,000 kg/hr (196,000 lb/hr) each	1	0
9	Candle Filter	Pressurized filter with pulse-jet cleaning	metallic filters	2	0

Case 4 – Account 6: Hydrogen Production

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Pressure Swing Adsorber	Polybed Proprietary	Syngas: 550,976 kg/hr (1,214,695 lb/hr) 45 °C (112 °F) 3.0 MPa (441.1 psia) Hydrogen: 27,498 kg/hr (60,622 lb/hr) 45 °C (112 °F) 3.0 MPa (431.1 psia)	1	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
			Off Gas: 523,479 kg/hr (1,154,073 lb/hr) -7 °C (20 °F) 0.5 MPa (77.0 psia)		
2	Hydrogen Compressor	Integrally geared, multi-stage centrifugal	99 m^3/min @ 6.5 MPa (3,482 acfm @ 940 psia)	1	1

Case 4 – Account 7: Stack, Ducting, & Off-Gas Fired Boiler

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Stack	CS plate, type 409SS liner	76 m (250 ft) high x 4.5 m (15 ft) diameter	1	0
2	Field-Erected Gas-Fired Boiler	Drum, multi-pressure with economizer section and integral deaerator Air-Fired	Off Gas: 523,479 kg/hr (1,154,073 lb/hr) 0.5 MPa (72.5 psia) 193 °C (380 °F) Air: 420,681 kg/hr (927,443 lb/hr) 0.1 MPa (16.0 psia) 24 °C (75 °F) 645 GJ/hr (611 MMBtu/hr)	1	0
3	Air Blower	Centrifugal	462,749 kg/hr, 6,315 m^3/min @ 111 cm WG (1,020,187 lb/hr, 223,000 acfm @ 44 in. WG)	2	1

Case 4 – Account 8: Steam Turbine and Accessories

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Steam Turbine	Commercially available advanced steam turbine	93 MW 11.7 MPa/482 °C/482 °C (1,700 psig/ 900 °F/900 °F)	1	0
2	Steam Turbine Generator	Hydrogen cooled, static excitation	100 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	1	0
3	Surface Condenser	Single pass, divided waterbox including vacuum pumps	400GJ/hr (380 MMBtu/hr), Inlet water temperature 16 °C (60 °F), Water temperature rise 11 °C (20 °F)	1	0
4	Steam Bypass	One per boiler	50% steam flow @ design steam conditions	1	0

Case 4 – Account 9: Cooling Water System

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Circulating Water Pumps	Vertical, wet pit	292,000 lpm @ 30 m (77,000 gpm @ 100 ft)	2	1
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	11 °C (52 °F) wet bulb/ 16 °C (60 °F) CWT/27 °C (80 °F) HWT/ 1,630 GJ/hr (1,540 MMBtu/hr) heat duty	1	0

Case 4 – Account 10: Slag Recovery and Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Slag Quench Tank	Water bath	223,000 liters (59,000 gal)	2	0
2	Slag Crusher	Roll	12 tonne/hr (13 tph)	2	0
3	Slag Depressurizer	Lock Hopper	12 tonne/hr (13 tph)	2	0
4	Slag Receiving Tank	Horizontal, weir	134,000 liters (35,000 gal)	2	0
5	Black Water Overflow Tank	Shop fabricated	60,000 liters (16,000 gal)	2	0
6	Slag Conveyor	Drag chain	12 tonne/hr (13 tph)	2	0
7	Slag Separation Screen	Vibrating	12 tonne/hr (13 tph)	2	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
8	Coarse Slag Conveyor	Belt/bucket	12 tonne/hr (13 tph)	2	0
9	Fine Ash Settling Tank	Vertical, gravity	190,000 liters (50,000 gal)	2	0
10	Fine Ash Recycle Pumps	Horizontal centrifugal	50 lpm @ 14 m H_2O (10 gpm @ 46 ft H_2O)	2	2
11	Grey Water Storage Tank	Field erected	61,000 liters (16,000 gal)	2	0
12	Grey Water Pumps	Centrifugal	210 lpm @ 430 m H_2O (60 gpm @ 1,420 ft H_2O)	2	2
13	Slag Storage Bin	Vertical, field erected	800 tonne (900 tons)	2	0
14	Unloading Equipment	Telescoping chute	100 tonne/hr (110 tph)	1	0

Case 4 – Account 11: Accessory Electric Plant

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	STG Transformer	Oil-filled	24 kV/345 kV, 60 MVA, 3-ph, 60 Hz	1	0
2	High Voltage Auxiliary Transformer	Oil-filled	345 kV/13.8 kV, 41 MVA, 3-ph, 60 Hz	2	0
3	Medium Voltage Transformer	Oil-filled	24 kV/4.16 kV, 42 MVA, 3-ph, 60 Hz	1	1
4	Low Voltage Transformer	Dry ventilated	4.16 kV/480 V, 6 MVA, 3-ph, 60 Hz	1	1
5	STG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	1	0
6	Medium Voltage Switchgear	Metal clad	4.16 kV, 3-ph, 60 Hz	1	1
7	Low Voltage Switchgear	Metal enclosed	480 V, 3-ph, 60 Hz	1	1
8	Emergency Diesel Generator	Sized for emergency shutdown	750 kW, 480 V, 3-ph, 60 Hz	1	0

Case 4 – Account 12: Instrumentation and Control

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	DCS – Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W)	Operator stations/printers and engineering stations/printers	1	0
2	DCS – Processor	Microprocessor with redundant input/output	N/A	1	0
3	DCS – Data Highway	Fiber optic	Fully redundant, 25% spare	1	0

4.8 Case 4 – Cost Estimating

The cost estimating methodology was described previously in Section 2.11. Exhibit 4-29 shows a detailed breakdown of the capital costs; Exhibit 4-30 shows the owner's costs, TOC, and TASC; Exhibit 4-31 shows the initial and annual O&M costs; and Exhibit 4-32 shows the LCOH breakdown.

The estimated TOC of the Shell gasification plant without CO_2 capture is $5,243/[kg H_2/day]. Process contingency represents 4.4 percent of the TOC and project contingency represents 12.0 percent. The LCOH is $2.58/kg H_2.

Exhibit 4-29. Case 4 total plant cost details

	Case:	4	– Coal Gasification w/o CCS				Estimate Type:			Conceptual	
	Plant Size (kg H_2/day):	659,947					Cost Base:			Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/[kg H_2/day]
1	Coal Handling										
1.1	Coal Receive & Unload	$971	$0	$468	$0	$1,438	$216	$0	$331	$1,985	$3
1.2	Coal Stackout & Reclaim	$3,173	$0	$758	$0	$3,931	$590	$0	$904	$5,425	$8
1.3	Coal Conveyors & Yard Crush	$30,270	$0	$7,704	$0	$37,974	$5,696	$0	$8,734	$52,404	$79
1.4	Other Coal Handling	$4,715	$0	$1,061	$0	$5,776	$866	$0	$1,328	$7,970	$12
1.5	Biomass Receive & Unload	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.6	Biomass Handling	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.7	Biomass Conveyors	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.8	Biomass Hnd. Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.9	Coal & Sorbent Handling Foundations	$0	$85	$222	$0	$307	$46	$0	$71	$424	$1
	Subtotal	**$39,128**	**$85**	**$10,213**	**$0**	**$49,426**	**$7,414**	**$0**	**$11,368**	**$68,208**	**$103**
2	Coal Preparation & Feed										
2.1	Coal Crushing & Drying	$2,349	$142	$338	$0	$2,829	$424	$0	$651	$3,904	$6
2.2	Prepared Coal Storage & Feed	$7,217	$1,734	$1,115	$0	$10,066	$1,510	$0	$2,315	$13,891	$21
2.3	Dry Coal Injection System	$9,212	$106	$844	$0	$10,162	$1,524	$0	$2,337	$14,023	$21
2.4	Miscellaneous Coal Preparation & Feed	$713	$521	$1,534	$0	$2,768	$415	$0	$637	$3,819	$6
2.5	Biomass Shredding & Drying	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.6	Prepared Biomass Storage & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.7	Dry Biomass Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.9	Coal, Biomass & Sorbent Feed Foundation	$0	$1,734	$1,488	$0	$3,221	$483	$0	$741	$4,446	$7
	Subtotal	**$19,491**	**$4,237**	**$5,318**	**$0**	**$29,045**	**$4,357**	**$0**	**$6,680**	**$40,083**	**$61**
3	Feedwater & Miscellaneous BOP Systems										
3.1	Feedwater System	$1,051	$1,802	$901	$0	$3,754	$563	$0	$864	$5,181	$8
3.2	Water Makeup & Pretreating	$3,738	$374	$2,118	$0	$6,230	$935	$0	$2,149	$9,314	$14
3.3	Other Feedwater Subsystems	$543	$178	$169	$0	$891	$134	$0	$205	$1,229	$2
3.4	Service Water Systems	$1,117	$2,133	$6,905	$0	$10,155	$1,523	$0	$3,503	$15,181	$23
3.5	Other Boiler Plant Systems	$138	$50	$126	$0	$314	$47	$0	$72	$433	$1
3.6	Natural Gas Pipeline and Start-Up System	$7,197	$310	$232	$0	$7,739	$1,161	$0	$1,780	$10,679	$16
3.7	Wastewater Treatment Equipment	$5,256	$0	$3,222	$0	$8,478	$1,272	$0	$2,925	$12,674	$19
3.8	Vacuum Flash, Brine Concentrator, & Crystallizer	$25,027	$0	$15,504	$0	$40,532	$6,080	$0	$13,983	$60,595	$92
3.9	Miscellaneous Plant Equipment	$15,325	$2,010	$7,788	$0	$25,124	$3,769	$0	$8,668	$37,560	$57
	Subtotal	**$59,393**	**$6,856**	**$36,966**	**$0**	**$103,216**	**$15,482**	**$0**	**$34,149**	**$152,847**	**$232**

	Case:	4	– Coal Gasification w/o CCS				Estimate Type:			Conceptual	
	Plant Size (kg H_2/day):	659,947					Cost Base:			Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/[kg H_2/day]
4	Gasifier, ASU, & Accessories										
4.1	Gasifier & Auxiliaries (Shell)	$604,151	$0	$259,074	$0	$863,224	$129,484	$120,851	$167,034	$1,280,593	$1,940
4.2	Syngas Cooler	$61,105	$0	$26,203	$0	$87,308	$13,096	$12,223	$16,894	$129,521	$196
4.3	Air Separation Unit/Oxidant Compression	$126,735	$29,005	$30,163	$0	$185,903	$37,181	$18,590	$36,251	$277,925	$421
4.5	Miscellaneous Gasification Equipment	$4,180	$0	$1,792	$0	$5,972	$896	$0	$1,030	$7,898	$12
4.6	Low Temperature Heat Recovery & Fuel Gas Saturation	$44,687	$0	$16,977	$0	$61,664	$9,250	$0	$14,183	$85,096	$129
4.7	Flare Stack System	$1,901	$0	$335	$0	$2,236	$335	$0	$514	$3,086	$5
4.15	Major Component Rigging	$233	$0	$100	$0	$332	$50	$0	$57	$439	$1
4.16	Gasification Foundations	$0	$470	$280	$0	$751	$113	$0	$216	$1,079	$2
	Subtotal	$842,990	$29,476	$334,925	$0	$1,207,390	$190,404	$151,665	$236,179	$1,785,638	$2,706
5	Syngas Cleanup										
5.2	Methyl Diethanolamine (MDEA) – Low Temperature Acid Gas Removal	$10,598	$0	$8,932	$0	$19,530	$2,929	$0	$4,492	$26,951	$41
5.3	Elemental Sulfur Plant	$47,453	$9,250	$60,802	$0	$117,505	$17,626	$0	$27,026	$162,157	$246
5.4	Carbon Dioxide (CO_2) Compression & Drying	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.5	Carbon Dioxide (CO_2) Compressor Aftercooler	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.6	Mercury Removal (Carbon Bed)	$480	$0	$363	$0	$843	$126	$42	$202	$1,214	$2
5.7	Water Gas Shift (WGS) Reactors	$133,061	$0	$53,194	$0	$186,255	$27,938	$0	$42,839	$257,032	$389
5.9	Particulate Removal	$1,841	$0	$790	$0	$2,631	$395	$0	$454	$3,480	$5
5.10	Blowback Gas Systems	$838	$471	$263	$0	$1,571	$236	$0	$361	$2,168	$3
5.11	Fuel Gas Piping	$0	$3,802	$2,490	$0	$6,292	$944	$0	$1,447	$8,683	$13
5.12	Gas Cleanup Foundations	$0	$220	$148	$0	$369	$55	$0	$127	$551	$1
	Subtotal	**$194,272**	**$13,743**	**$126,980**	**$0**	**$334,996**	**$50,249**	**$42**	**$76,949**	**$462,236**	**$700**
6	Hydrogen Production										
6.1	Pressure Swing Adsorber	$32,224	$23	$1,403	$1,095	$34,745	$5,212	$0	$7,991	$47,948	$73
6.2	Hydrogen Compressor	$9,097	$147	$526	$410	$10,179	$1,527	$0	$2,341	$14,048	$21
	Subtotal	**$41,321**	**$170**	**$1,929**	**$1,505**	**$44,925**	**$6,739**	**$0**	**$10,333**	**$61,996**	**$94**
7	Off Gas Fired Boiler, Ductwork, & Stack										
7.1	Off Gas Fired Boiler	$3,727	$367	$416	$324	$4,833	$725	$0	$1,112	$6,670	$10
7.2	Off Gas Fired Boiler Accessories	$523	$465	$741	$578	$2,307	$346	$0	$531	$3,184	$5
7.3	Ductwork	w/7.1	w/7.1	w/7.1	w/7.1	$0	$0	$0	$0	$0	$0
7.4	Stack	$1,157	$77	$211	$164	$1,610	$242	$0	$463	$2,314	$4
7.5	Off Gas Fired Boiler, Ductwork & Stack Foundations	w/7.1	w/7.1	w/7.1	w/7.1	$0	$0	$0	$0	$0	$0
	Subtotal	**$5,407**	**$910**	**$1,368**	**$1,067**	**$8,751**	**$1,313**	**$0**	**$2,105**	**$12,169**	**$18**

Item No.	Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	Total Plant Cost $/1,000	Total Plant Cost $/[kg H₂/day]
	Case:	4	– Coal Gasification w/o CCS				Estimate Type:			Conceptual	
	Plant Size (kg H_2/day):	659,947					Cost Base:			Dec 2018	
8						Steam Turbine & Accessories					
8.1	Steam Turbine Generator & Accessories	$28,196	$1,952	$3,109	$2,425	$35,682	$5,352	$0	$8,207	$49,242	$75
8.2	Steam Turbine Plant Auxiliaries	w/8.1	w/8.1	w/8.1	w/8.1	$0	$0	$0	$0	$0	$0
8.3	Condenser & Auxiliaries	$761	$620	$504	$393	$2,278	$342	$0	$524	$3,143	$5
8.4	Steam Piping	w/8.1	w/8.1	w/8.1	w/8.1	$0	$0	$0	$0	$0	$0
8.5	Turbine Generator Foundations	w/8.1	w/8.1	w/8.1	w/8.1	$0	$0	$0	$0	$0	$0
	Subtotal	**$28,957**	**$2,572**	**$3,613**	**$2,818**	**$37,960**	**$5,694**	**$0**	**$8,731**	**$52,385**	**$79**
9						Cooling Water System					
9.1	Cooling Towers	$8,329	$0	$2,520	$0	$10,849	$1,627	$0	$1,872	$14,348	$22
9.2	Circulating Water Pumps	$1,084	$0	$80	$0	$1,164	$175	$0	$201	$1,539	$2
9.3	Circulating Water System Auxiliaries	$7,666	$0	$1,089	$0	$8,755	$1,313	$0	$1,510	$11,578	$18
9.4	Circulating Water Piping	$0	$4,417	$4,000	$0	$8,418	$1,263	$0	$1,936	$11,616	$18
9.5	Make-up Water System	$464	$0	$638	$0	$1,102	$165	$0	$253	$1,520	$2
9.6	Component Cooling Water System	$158	$189	$130	$0	$476	$71	$0	$110	$657	$1
9.7	Circulating Water System Foundations	$0	$368	$654	$0	$1,022	$153	$0	$353	$1,529	$2
	Subtotal	**$17,701**	**$4,974**	**$9,111**	**$0**	**$31,786**	**$4,768**	**$0**	**$6,234**	**$42,788**	**$65**
10						Slag Recovery & Handling					
10.1	Slag Dewatering & Cooling	$1,934	$0	$947	$0	$2,881	$432	$0	$497	$3,810	$6
10.2	Gasifier Ash Depressurization	$1,096	$0	$537	$0	$1,632	$245	$0	$282	$2,158	$3
10.3	Cleanup Ash Depressurization	$492	$0	$241	$0	$734	$110	$0	$127	$970	$1
10.6	Ash Storage Silos	$1,104	$0	$1,193	$0	$2,297	$345	$0	$396	$3,038	$5
10.7	Ash Transport & Feed Equipment	$425	$0	$99	$0	$524	$79	$0	$90	$693	$1
10.8	Miscellaneous Ash Handling Equipment	$61	$75	$22	$0	$158	$24	$0	$27	$209	$0
10.9	Ash/Spent Sorbent Foundation	$0	$431	$573	$0	$1,004	$151	$0	$346	$1,501	$2
	Subtotal	**$5,112**	**$506**	**$3,612**	**$0**	**$9,230**	**$1,384**	**$0**	**$1,765**	**$12,380**	**$19**
11						Accessory Electric Plant					
11.1	Generator Equipment	$1,468	$0	$1,108	$0	$2,576	$386	$0	$444	$3,406	$5
11.2	Station Service Equipment	$3,409	$0	$292	$0	$3,702	$555	$0	$639	$4,896	$7
11.3	Switchgear & Motor Control	$20,574	$0	$3,569	$0	$24,143	$3,621	$0	$4,165	$31,929	$48
11.4	Conduit & Cable Tray	$0	$91	$263	$0	$354	$53	$0	$102	$509	$1
11.5	Wire & Cable	$0	$1,248	$2,231	$0	$3,479	$522	$0	$1,000	$5,002	$8
11.6	Protective Equipment	$241	$0	$837	$0	$1,078	$162	$0	$186	$1,426	$2
11.7	Standby Equipment	$307	$0	$284	$0	$591	$89	$0	$102	$781	$1
11.8	Main Power Transformers	$1,421	$0	$29	$0	$1,450	$218	$0	$250	$1,918	$3
11.9	Electrical Foundations	$0	$17	$43	$0	$59	$9	$0	$20	$89	$0

Case:		4	– Coal Gasification w/o CCS					Estimate Type:		Conceptual	
Plant Size (kg H_2/day):		659,947						Cost Base:		Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/[kg H_2/day]
	Subtotal	**$27,421**	**$1,356**	**$8,656**	**$0**	**$37,433**	**$5,615**	**$0**	**$6,908**	**$49,956**	**$76**
12	Instrumentation & Control										
12.1	Gasification Control Equipment	$654	$0	$280	$0	$934	$140	$0	$161	$1,235	$2
12.3	Steam Turbine Control Equipment	$604	$0	$82	$0	$686	$103	$0	$118	$907	$1
12.4	Other Major Component Control Equipment	$1,150	$0	$783	$0	$1,933	$290	$97	$348	$2,667	$4
12.5	Signal Processing Equipment	$892	$0	$29	$0	$921	$138	$0	$159	$1,219	$2
12.6	Control Boards, Panels & Racks	$259	$0	$170	$0	$428	$64	$21	$103	$617	$1
12.7	Distributed Control System Equipment	$9,364	$0	$306	$0	$9,669	$1,450	$483	$1,740	$13,344	$20
12.8	Instrument Wiring & Tubing	$466	$373	$1,491	$0	$2,329	$349	$116	$699	$3,494	$5
12.9	Other Instrumentation & Controls Equipment	$1,045	$0	$518	$0	$1,563	$234	$78	$281	$2,157	$3
	Subtotal	**$14,433**	**$373**	**$3,659**	**$0**	**$18,464**	**$2,770**	**$796**	**$3,610**	**$25,639**	**$39**
13	Improvements to Site										
13.1	Site Preparation	$0	$422	$9,603	$0	$10,025	$1,504	$0	$3,459	$14,987	$23
13.2	Site Improvements	$0	$1,908	$2,697	$0	$4,604	$691	$0	$1,589	$6,884	$10
13.3	Site Facilities	$2,978	$0	$3,344	$0	$6,322	$948	$0	$2,181	$9,451	$14
	Subtotal	**$2,978**	**$2,330**	**$15,643**	**$0**	**$20,951**	**$3,143**	**$0**	**$7,228**	**$31,322**	**$47**
14	Buildings & Structures										
14.3	Steam Turbine Building	$0	$2,585	$3,681	$0	$6,266	$940	$0	$1,081	$8,287	$13
14.4	Administration Building	$0	$844	$613	$0	$1,457	$219	$0	$251	$1,927	$3
14.5	Circulation Water Pumphouse	$0	$116	$61	$0	$178	$27	$0	$31	$235	$0
14.6	Water Treatment Buildings	$0	$250	$244	$0	$493	$74	$0	$85	$653	$1
14.7	Machine Shop	$0	$477	$326	$0	$803	$120	$0	$138	$1,062	$2
14.8	Warehouse	$0	$371	$240	$0	$611	$92	$0	$105	$808	$1
14.9	Other Buildings & Structures	$0	$272	$212	$0	$484	$73	$0	$83	$640	$1
14.10	Waste Treating Building & Structures	$0	$730	$1,393	$0	$2,123	$318	$0	$366	$2,808	$4
	Subtotal	**$0**	**$5,645**	**$6,769**	**$0**	**$12,415**	**$1,862**	**$0**	**$2,142**	**$16,418**	**$25**
	Total	**$1,298,603**	**$73,232**	**$568,762**	**$5,389**	**$1,945,987**	**$301,193**	**$152,503**	**$414,381**	**$2,814,065**	**$4,264**

Exhibit 4-30. Case 4 owner's costs

Description	$/1,000	$/[kg H_2/day]
Pre-Production Costs		
6 Months All Labor	$22,852	$35
1 Month Maintenance Materials	$5,716	$9
1 Month Non-Fuel Consumables	$2,275	$3
1 Month Waste Disposal	$695	$1
25% of 1 Months Fuel Cost at 100% CF	$2,216	$3
2% of TPC	$56,281	$85
Total	**$90,034**	**$136**
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	$19,393	$29
0.5% of TPC (spare parts)	$14,070	$21
Total	**$33,463**	**$51**
Other Costs		
Initial Cost for Catalyst and Chemicals	$23,857	$36
Land	$900	$1
Other Owner's Costs	$422,110	$640
Financing Costs	$75,980	$115
Total Overnight Costs (TOC)	**$3,460,409**	**$5,243**
TASC Multiplier (IOU, 35 year)	1.116	
Total As-Spent Cost (TASC)	**$3,862,980**	**$5,853**

Exhibit 4-31. Case 4 initial and annual operating and maintenance costs

Case:	**4**	**– Coal Gasification w/o CCS**			**Cost Base:**	**Dec 2018**
Plant Size (kg H_2/day):	**659,947**				**Capacity Factor (%):**	**80**
Operating & Maintenance Labor						
Operating Labor				**Operating Labor Requirements per Shift**		
Operating Labor Rate (base):		38.50	**$/hour**	Skilled Operator:		2.0
Operating Labor Burden:		30.00	**% of base**	Operator:		10.0
Labor O-H Charge Rate:		25.00	**% of labor**	Foreman:		1.0
				Lab Techs, etc.:		3.0
				Total:		**16.0**
Fixed Operating Costs						
					Annual Cost	
					($)	**($/[kg H_2/day])**
Annual Operating Labor:					$7,015,008	$10.63

Case:	**4**	**– Coal Gasification w/o CCS**			**Cost Base:**	**Dec 2018**
Plant Size (kg H_2/day):	**659,947**				**Capacity Factor (%):**	**80**
Maintenance Labor:					$29,547,681	$44.77
Administrative & Support Labor:					$9,140,672	$13.85
Property Taxes and Insurance:					$56,281,297	$85.28
Total:					**$101,984,658**	**$154.53**
Variable Operating Costs						
					($)	**($/kg H_2)**
Maintenance Material:					$54,874,264	$0.28
Consumables						
	Initial Fill	**Per Day**	**Per Unit**	**Initial Fill**		
Water (gal/1000):	0	2,296	$1.90	$0	$1,273,566	$0.00661
Makeup and Wastewater Treatment Chemicals (ton):	0	6.8	$550.00	$0	$1,098,198	$0.00570
Sulfur-Impregnated Activated Carbon (ton):	107	0.146	$12,000.00	$1,282,315	$512,926	$0.00266
Water Gas Shift Catalyst (ft^3):	25,993	17.8	$480.00	$12,476,654	$2,495,331	$0.01295
Methyl Diethanolamine Solution (gal):	231,310	107.3	$2.80	$647,667	$87,750	$0.00046
Sodium Hydroxide (50 wt%, ton):	0	19.3	$600.00	$0	$3,388,268	$0.01758
Sulfuric Acid (98 wt%, ton):	0	0.342	$210.00	$0	$20,992	$0.00011
Claus Catalyst (ft^3):	w/equip.	1.93	$48.00	$0	$27,085	$0.00014
PSA Unit Adsorbent (ft^3):	63,000	8.6	$150.00	$9,450,000	$378,000	$0.00196
Triethylene Glycol (gal):	w/equip.	0	$6.80	$0	$0	$0.00000
Electricity (MWh):	0	600	$71.70	$0	$12,556,815	$0.06516
Subtotal:				**$23,856,637**	**$21,838,931**	**$0.11333**
Waste Disposal						
Sulfur-Impregnated Activated Carbon (ton):	0	0.146	$80.00	$0	$3,420	$0.00002
Water Gas Shift Catalyst (ft^3):	0	17.8	$2.50	$0	$12,997	$0.00007
Methyl Diethanolamine Solution (gal):	0	107.3	$0.35	$0	$10,969	$0.00006
Claus Catalyst (ft^3):	0	1.93	$2.50	$0	$1,411	$0.00001
Crystallizer Solids (ton):	0	36.4	$38.00	$0	$404,177	$0.00210
Slag (ton):	0	562	$38.00	$0	$6,231,964	$0.03234
PSA Unit Adsorbent (ft^3):	0	8.6	$1.50	$0	$3,780	$0.00002
Triethylene Glycol (gal):	0	0	$0.35	$0	$0	$0.00000
Subtotal:				**$0**	**$6,668,716**	**$0.03461**
By-Products						
Sulfur (tons):	0	140	$0.00	$0	$0	$0.00000
Electricity (MWh):	0	0	$71.70	$0	$0	$0.00000
Subtotal:				**$0**	**$0**	**$0.00000**
Variable Operating Costs Total:				**$23,856,637**	**$83,381,911**	**$0.43269**
Fuel Cost						
Illinois Number 6 (ton):	0	5,608	$51.96	$0	$85,078,534	$0.44150
Woody Torrefied Biomass (ton):	0	0	$105.84	$0	$0	$0.00000
Total:				**$0**	**$85,078,534**	**$0.44150**

Exhibit 4-32. Case 4 LCOH breakdown

Component	Value, $/kg H_2	Percentage
Capital	1.17	46%
Fixed	0.53	21%
Variable	0.43	17%
Fuel	0.44	17%
Total (Excluding T&S)	**2.58**	**N/A**
CO_2 T&S	0.00	0%
Total (Including T&S)	**2.58**	**N/A**

4.9 Case 5 – Shell Gasification Plant with CO_2 Capture Process Description

This section contains an evaluation of a plant design for Case 5, which is based on a Shell coal gasification plant with CO_2 capture. The plant configuration is nearly identical to that of Case 4, with the exception that this case is configured to produce hydrogen with CO_2 capture. The hydrogen production rate of Case 5 is equal to the Case 4 capacity. The system descriptions follow the BFD provided in Exhibit 4-33 with the associated stream tables that show process data provided in Exhibit 4-34. Rather than repeating the entire process description, only differences from Case 4 are reported here.

Exhibit 4-33. Case 5 block flow diagram, Shell gasification plant with CO_2 capture

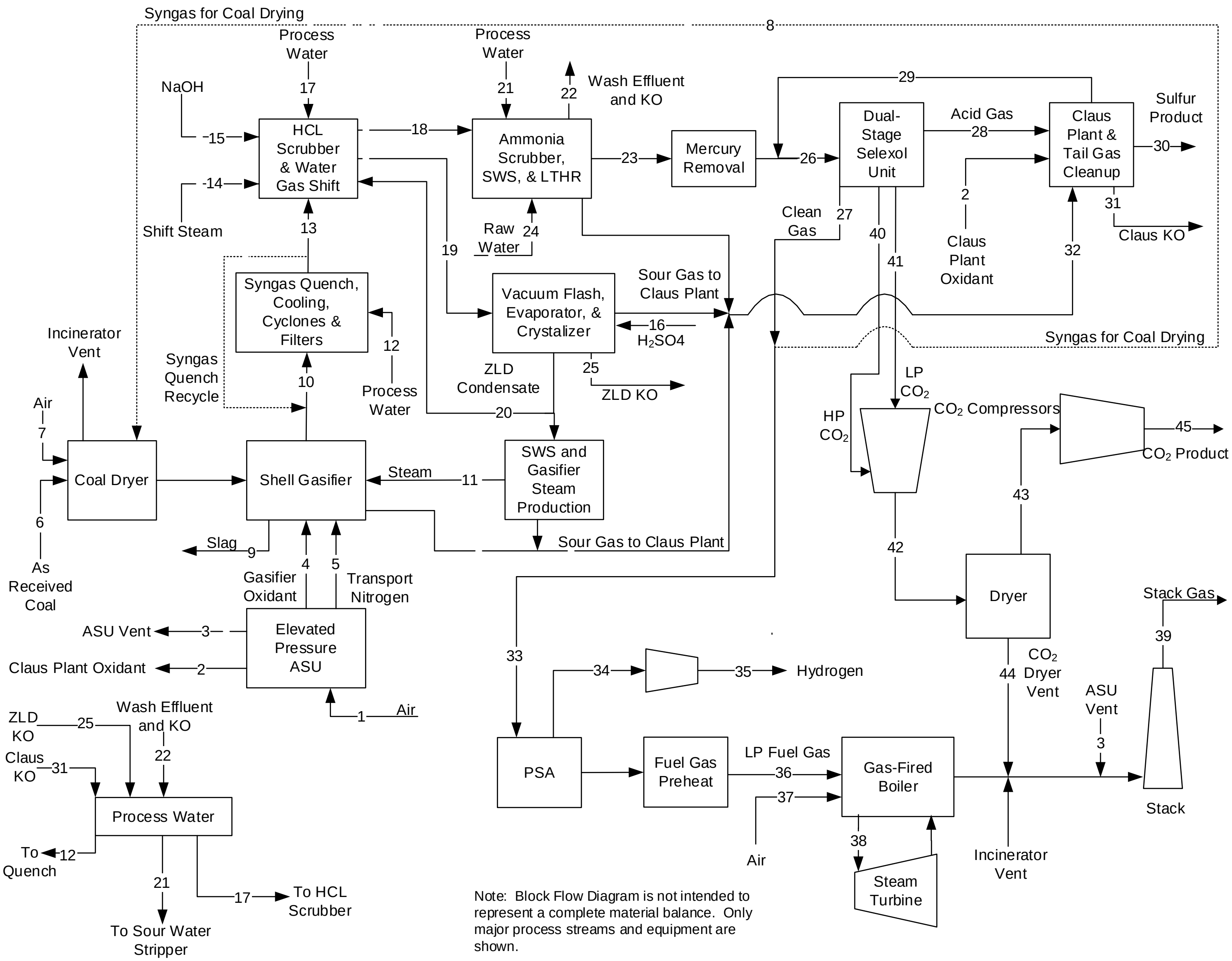


Exhibit 4-34. Case 5 stream table, Shell gasification plant with capture

	1	2	3	4	5	6	7	8	9	10	11	12
V-L Mole Fraction												
Ar	0.0092	0.0343	0.0025	0.0343	0.0000	0.0000	0.0092	0.0096	0.0000	0.0082	0.0000	0.0000
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0000	0.0004	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0167	0.0000	0.5163	0.0000	0.0000
CO_2	0.0003	0.0000	0.0004	0.0000	0.0000	0.0000	0.0003	0.0286	0.0000	0.0170	0.0000	0.0010
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0006	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.8832	0.0000	0.2637	0.0000	0.0000
H_2O	0.0099	0.0000	0.0091	0.0000	0.0000	0.0000	0.0099	0.0001	0.0000	0.1293	0.9998	0.9796
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0073	0.0000	0.0001
H_2SO_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.7732	0.0157	0.9844	0.0157	0.9964	0.0000	0.7732	0.0614	0.0000	0.0516	0.0000	0.0000
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0051	0.0002	0.0193
O_2	0.2074	0.9501	0.0035	0.9501	0.0036	0.0000	0.2074	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaOH	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	1.0000	0.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	23,745	97	17,612	5,017	945	0	1,003	305	0	26,522	1,130	10,214
V-L Flowrate (kg/hr)	685,216	3,129	492,660	161,608	26,484	0	28,929	1,716	0	536,844	20,363	184,111
Solids Flowrate (kg/hr)	0	0	0	0	0	211,967	0	0	21,230	0	0	0
Temperature (°C)	15	27	21	27	130	15	15	18	1,427	1,079	343	188
Pressure (MPa, abs)	0.10	0.86	0.45	5.10	5.62	0.10	0.10	2.89	4.24	4.24	5.10	4.98
Steam Table Enthalpy (kJ/kg)[A]	30.23	21.53	28.11	9.82	129.22	---	30.23	94.16	---	2,034.56	3,083.36	767.78
AspenPlus Enthalpy (kJ/kg)[B]	-97.58	-0.97	-97.45	-12.68	103.25	-2,119.02	-97.58	-2,374.94	2,165.43	-3,011.70	-12,884.30	-14,939.49
Density (kg/m³)	1.2	11.2	5.2	68.6	46.4	---	1.2	6.6	---	7.6	19.9	835.9
V-L Molecular Weight	28.857	32.209	27.974	32.209	28.028	---	28.857	5.617	---	20.241	18.015	18.025
V-L Flowrate (lb_{mol}/hr)	52,349	214	38,827	11,062	2,083	0	2,210	673	0	58,472	2,492	22,518
V-L Flowrate (lb/hr)	1,510,643	6,897	1,086,129	356,286	58,388	0	63,778	3,783	0	1,183,539	44,892	405,896
Solids Flowrate (lb/hr)	0	0	0	0	0	467,308	0	0	46,803	0	0	0
Temperature (°F)	59	80	70	80	267	59	59	65	2,600	1,974	650	371
Pressure (psia)	14.7	125.0	65.0	740.0	815.0	14.7	14.7	418.7	615.0	615.0	740.0	722.7
Steam Table Enthalpy (Btu/lb)[A]	13.0	9.3	12.1	4.2	55.6	---	13.0	40.5	---	874.7	1,325.6	330.1
AspenPlus Enthalpy (Btu/lb)[B]	-42.0	-0.4	-41.9	-5.5	44.4	-911.0	-42.0	-1,021.0	931.0	-1,294.8	-5,539.3	-6,422.8
Density (lb/ft³)	0.076	0.700	0.322	4.283	2.894	---	0.076	0.412	---	0.473	1.240	52.186

[A]Steam table reference conditions are 32.02 °F & 0.089 psia

[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25 °C and 1 atm

Exhibit 4-34. Case 5 stream table, Shell gasification plant with capture (continued)

	13	14	15	16	17	18	19	20	21	22	23
V-L Mole Fraction											
Ar	0.0059	0.0000	0.0000	0.0000	0.0000	0.0048	0.0000	0.0000	0.0000	0.0000	0.0060
CH_4	0.0003	0.0000	0.0000	0.0000	0.0000	0.0002	0.0000	0.0000	0.0000	0.0000	0.0003
CO	0.3728	0.0000	0.0000	0.0000	0.0000	0.0084	0.0002	0.0000	0.0000	0.0000	0.0106
CO_2	0.0125	0.0000	0.0000	0.0000	0.0007	0.3010	0.0001	0.0000	0.0007	0.0010	0.3783
COS	0.0005	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.1904	0.0000	0.0000	0.0000	0.0000	0.4433	0.0001	0.0000	0.0000	0.0000	0.5597
H_2O	0.3657	1.0000	0.6895	0.1000	0.9799	0.2004	0.9931	0.9993	0.9799	0.9791	0.0016
HCl	0.0003	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2S	0.0053	0.0000	0.0000	0.0000	0.0001	0.0046	0.0001	0.0000	0.0001	0.0001	0.0057
H_2SO_4	0.0000	0.0000	0.0000	0.9000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0373	0.0000	0.0000	0.0000	0.0000	0.0299	0.0000	0.0000	0.0000	0.0000	0.0378
NH_3	0.0091	0.0000	0.0000	0.0000	0.0192	0.0073	0.0038	0.0007	0.0192	0.0197	0.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0026	0.0000	0.0000	0.0000	0.0000
NaOH	0.0000	0.0000	0.3105	0.0000	0.0000	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	29,022	8,265	29	0	709	36,173	3,480	1,627	3,331	13,357	28,645
V-L Flowrate (kg/hr)	569,546	148,903	731	13	12,771	698,204	63,063	29,316	60,014	240,775	561,467
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	232	288	16	15	67	202	190	88	67	69	29
Pressure (MPa, abs)	3.98	3.88	4.81	0.13	0.47	3.49	3.88	0.13	0.47	0.45	3.18
Steam Table Enthalpy (kJ/kg)[A]	1,189.11	2,971.61	-338.78	-8,206.86	238.31	804.50	792.23	368.02	238.31	247.89	39.45
AspenPlus Enthalpy (kJ/kg)[B]	-6,579.94	-13,008.68	-13,665.00	-8,526.27	-15,475.10	-8,423.95	-15,049.56	-15,593.09	-15,475.10	-15,452.90	-7,685.67
Density (kg/m^3)	19.0	16.9	1,531.7	1,791.5	964.7	17.2	873.9	965.6	964.7	962.3	25.2
V-L Molecular Weight	19.625	18.015	24.842	90.073	18.017	19.302	18.120	18.015	18.017	18.026	19.601
V-L Flowrate (lb_{mol}/hr)	63,982	18,222	65	0	1,563	79,747	7,673	3,588	7,343	29,447	63,152
V-L Flowrate (lb/hr)	1,255,634	328,274	1,612	29	28,156	1,539,277	139,029	64,631	132,308	530,818	1,237,823
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	450	550	61	59	152	395	373	191	152	157	84
Pressure (psia)	577.7	562.7	697.7	18.2	67.7	506.1	562.7	19.3	67.7	65.0	460.7
Steam Table Enthalpy (Btu/lb)[A]	511.2	1,277.6	-145.7	-3,528.3	102.5	345.9	340.6	158.2	102.5	106.6	17.0
AspenPlus Enthalpy (Btu/lb)[B]	-2,828.9	-5,592.7	-5,874.9	-3,665.6	-6,653.1	-3,621.6	-6,470.1	-6,703.8	-6,653.1	-6,643.6	-3,304.2
Density (lb/ft^3)	1.183	1.055	95.623	111.841	60.225	1.075	54.553	60.280	60.225	60.075	1.572

[A]Steam table reference conditions are 32.02 °F & 0.089 psia

[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25 °C and 1 atm

Exhibit 4-34. Case 5 stream table, Shell gasification plant with capture (continued)

	24	25	26	27	28	29	30	31	32	33	34
V-L Mole Fraction											
Ar	0.0000	0.0000	0.0060	0.0096	0.0000	0.0093	0.0000	0.0000	0.0004	0.0096	0.0008
CH_4	0.0000	0.0000	0.0003	0.0004	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0000
CO	0.0000	0.0000	0.0106	0.0167	0.0004	0.0053	0.0000	0.0000	0.0042	0.0167	0.0000
CO_2	0.0000	0.0002	0.3816	0.0286	0.5187	0.6440	0.0000	0.0000	0.3172	0.0286	0.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0009	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.5557	0.8832	0.0122	0.2388	0.0000	0.0000	0.0283	0.8832	0.9990
H_2O	0.9999	0.9832	0.0016	0.0001	0.0226	0.0024	0.0000	1.0000	0.1566	0.0001	0.0000
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2S	0.0000	0.0003	0.0057	0.0000	0.4441	0.0050	0.0000	0.0000	0.0123	0.0000	0.0000
H_2SO_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0000	0.0000	0.0385	0.0614	0.0003	0.0952	0.0000	0.0000	0.0015	0.0614	0.0002
NH_3	0.0000	0.0164	0.0000	0.0000	0.0007	0.0000	0.0000	0.0000	0.4796	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaCl	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaOH	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	2,627	685	29,011	18,152	373	366	0	214	137	17,847	13,412
V-L Flowrate (kg/hr)	47,343	12,327	573,275	101,962	14,340	11,808	0	3,864	3,515	100,246	27,500
Solids Flowrate (kg/hr)	0	0	0	0	0	0	5,312	0	0	0	0
Temperature (°C)	15	30	37	18	27	38	184	50	160	18	18
Pressure (MPa, abs)	0.10	0.24	3.07	2.89	0.18	3.07	0.12	0.11	0.45	2.89	2.82
Steam Table Enthalpy (kJ/kg)[A]	62.75	90.75	50.96	94.16	40.84	23.14	---	109.79	503.09	94.16	263.41
AspenPlus Enthalpy (kJ/kg)[B]	-15,905.25	-15,663.68	-7,678.43	-2,374.94	-5,697.83	-7,909.59	147.51	-15,860.13	-7,050.59	-2,374.94	-86.91
Density (kg/m^3)	999.4	985.6	23.9	6.6	2.9	41.0	5,266.6	968.5	3.2	6.6	2.4
V-L Molecular Weight	18.019	18.008	19.760	5.617	38.486	32.224	---	18.016	25.597	5.617	2.050
V-L Flowrate (lb_{mol}/hr)	5,793	1,509	63,959	40,019	821	808	0	473	303	39,345	29,568
V-L Flowrate (lb/hr)	104,373	27,175	1,263,856	224,788	31,615	26,032	0	8,518	7,749	221,005	60,627
Solids Flowrate (lb/hr)	0	0	0	0	0	0	11,711	0	0	0	0
Temperature (°F)	59	85	99	65	80	100	364	121	320	65	65
Pressure (psia)	14.7	35.0	445.2	418.7	26.7	445.2	16.8	15.9	65.0	418.7	408.7
Steam Table Enthalpy (Btu/lb)[A]	27.0	39.0	21.9	40.5	17.6	9.9	---	47.2	216.3	40.5	113.2
AspenPlus Enthalpy (Btu/lb)[B]	-6,838.0	-6,734.2	-3,301.1	-1,021.0	-2,449.6	-3,400.5	63.4	-6,818.6	-3,031.2	-1,021.0	-37.4
Density (lb/ft^3)	62.391	61.529	1.491	0.412	0.181	2.561	328.780	60.459	0.201	0.412	0.147

[A]Steam table reference conditions are 32.02 °F & 0.089 psia

[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25 °C and 1 atm

Exhibit 4-34. Case 5 stream table, Shell gasification plant with capture (continued)

	35	36	37	38	39	40	41	42	43	44	45
V-L Mole Fraction											
Ar	0.0008	0.0364	0.0092	0.0000	0.0095	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.0016	0.0000	0.0000	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0671	0.0000	0.0000	0.0000	0.0005	0.0001	0.0004	0.0004	0.0000	0.0004
CO_2	0.0000	0.1150	0.0003	0.0000	0.0254	0.9841	0.9984	0.9879	0.9908	0.0500	0.9908
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.9990	0.5331	0.0000	0.0000	0.0000	0.0106	0.0008	0.0080	0.0080	0.0000	0.0080
H_2O	0.0000	0.0005	0.0099	1.0000	0.1124	0.0043	0.0007	0.0034	0.0005	0.9500	0.0005
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2SO_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0002	0.2464	0.7732	0.0000	0.8242	0.0004	0.0000	0.0003	0.0003	0.0000	0.0003
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.2074	0.0000	0.0285	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaOH	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	13,412	4,435	10,388	13,270	33,068	7,648	2,839	10,487	10,455	32	10,455
V-L Flowrate (kg/hr)	27,500	72,746	299,768	239,062	910,167	332,195	124,778	456,973	456,364	609	456,364
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	30	193	15	510	50	-3	-11	29	29	29	30
Pressure (MPa, abs)	6.48	0.50	0.10	12.51	0.11	0.55	0.12	2.50	2.39	2.50	15.27
Steam Table Enthalpy (kJ/kg)[A]	440.05	356.19	30.23	3,369.21	216.61	-8.36	-9.60	2.30	0.54	138.13	-226.95
AspenPlus Enthalpy (kJ/kg)[B]	89.72	-2,913.87	-97.58	-12,611.09	-1,345.41	-8,974.80	-8,972.84	-8,963.44	-8,956.66	-15,225.03	-9,184.16
Density (kg/m^3)	5.1	2.1	1.2	38.4	1.1	11.2	2.3	49.8	47.2	319.0	838.1
V-L Molecular Weight	2.050	16.403	28.857	18.015	27.524	43.435	43.956	43.576	43.650	19.315	43.650
V-L Flowrate (lb_{mol}/hr)	29,568	9,777	22,902	29,255	72,903	16,861	6,258	23,119	23,050	70	23,050
V-L Flowrate (lb/hr)	60,627	160,378	660,875	527,042	2,006,574	732,364	275,089	1,007,453	1,006,110	1,343	1,006,110
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	86	380	59	950	122	26	12	85	85	85	86
Pressure (psia)	939.7	72.5	14.7	1,814.7	15.4	80.0	16.7	363.0	346.5	363.0	2,214.7
Steam Table Enthalpy (Btu/lb)[A]	189.2	153.1	13.0	1,448.5	93.1	-3.6	-4.1	1.0	0.2	59.4	-97.6
AspenPlus Enthalpy (Btu/lb)[B]	38.6	-1,252.7	-42.0	-5,421.8	-578.4	-3,858.5	-3,857.6	-3,853.6	-3,850.7	-6,545.6	-3,948.5
Density (lb/ft^3)	0.319	0.132	0.076	2.395	0.069	0.698	0.146	3.108	2.947	19.917	52.320

[A]Steam table reference conditions are 32.02 °F & 0.089 psia

[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25 °C and 1 atm

4.9.1 Acid Gas Removal

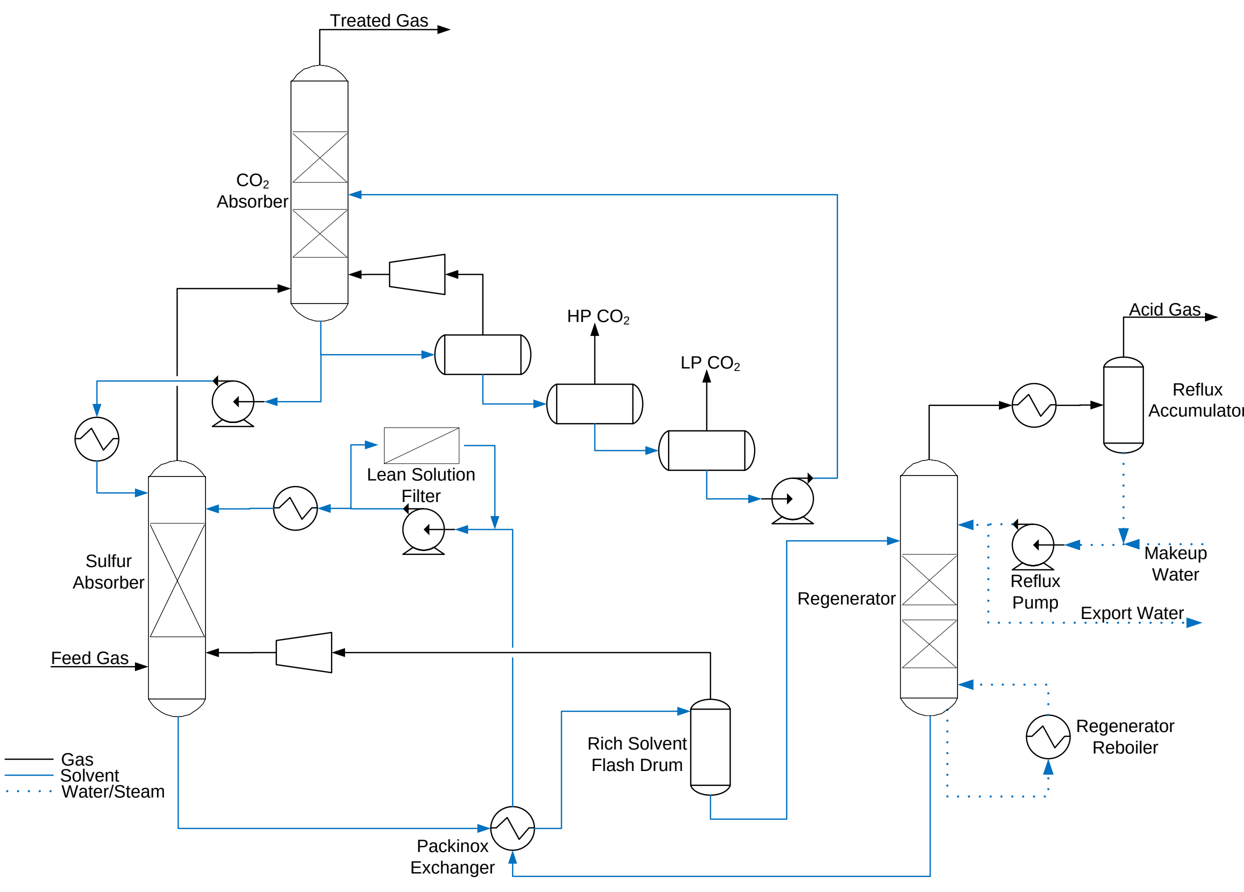
The AGR process in Case 5 is a two-stage Selexol process where H_2S is removed in the first stage and CO_2 in the second stage of absorption. The process results in four product streams, the clean syngas (stream 27), two CO_2-rich streams (streams 40 and 41), and an acid gas feed to the Claus plant (stream 28). The acid gas contains 44 vol% H_2S and 52 vol% CO_2 with the balance primarily water and H_2. The raw CO_2 stream from the Selexol process contains over 98 vol% CO_2.

The two-stage Selexol process is used in the two gasification cases that require CO_2 capture, Case 5 and Case 6. According to the previously referenced Sfa Pacific, Inc. report, "For future IGCC with CO_2 removal for storage, a two-stage Selexol process presently appears to be the preferred AGR process – as indicated by ongoing engineering studies at EPRI and by various engineering firms with IGCC interests." [70]

As several vendors have indicated that Selexol is sensitive to NH_3 and have specified that NH_3 should be reduced to a concentration below 10 ppmv in the syngas feed, a water wash column is included upstream of the AGR for NH_3 control, as detailed in Section 4.4.14.1.5.

As shown in Exhibit 4-35, syngas enters the bottom of the first of two absorbers and flows upward through packed beds where it contacts chilled solvent—loaded with CO_2—entering at the top of the column, which preferentially removes H_2S from the gas phase.

Exhibit 4-35. Flow diagram for a conventional two-stage AGR unit



The gas from the H_2S absorber flows upward through the packed beds of the second absorber. CO_2 is removed from the gas phase first by semi-lean, flash-regenerated solvent entering near the middle of the tower, followed by chilled, lean solvent entering at the top of the tower. The treated gas passes through de-entrainment devices at the top of the tower before exiting the absorber and being sent directly to the PSA. A portion of the gas can also be used for coal drying, when required.

The CO_2 loaded solvent exits the CO_2 absorber. A portion is sent to the H_2S absorber, which is pumped up to pressure and then chilled prior to reaching the H_2S absorber, and the remainder is sent to a series of flash drums for regeneration.

The portion sent to the flash regeneration is expanded in the HP CO_2 recycle flash drum at 2.0 MPa (289.7 psia) where H_2, CH_4, CO_2, and other dissolved gases are transferred to the gas phase. The flashed off gas is compressed and returned to the CO_2 absorber, minimizing product losses to the CO_2 stream.

The amount of H_2 recovered from the syngas stream is dependent on the Selexol process design conditions. In this report, H_2 recovery is 99.5 percent per pass. The minimal H_2 slip to the CO_2 storage stream maximizes the overall plant efficiency.

The semi-rich solvent from the HP CO_2 recycle flash drum is routed to two sequential CO_2 flash drums. The HP CO_2 stream is flashed at 0.55 MPa (80 psia) and the LP CO_2 stream is flashed at 0.1 MPa (16.7 psia). The flashed CO_2 gas is sent to the CO_2 compressors and the semi-lean solvent is pumped back to the CO_2 absorber.

The rich solvent exiting the H_2S absorber is heated using the lean solvent from the stripper. The hot, rich solvent enters the H_2S concentrator and partially flashes. The gas exiting the concentrator is compressed and recycled back to the H_2S absorber. The H_2S-rich solvent from the concentrator is sent to the regenerator for thermal regeneration.

The regenerator is composed of a lower section containing packed beds and an upper section containing several reflux trays used to contact the overhead vapor with the reflux water. The solvent from the concentrator enters the regenerator above the packed bed and flows downward, releasing H_2S, CO_2, and other components as it passes an upflow of hot vapor generated in the reboiler.

The combined gases and hot vapor flow upward through a demister and the trayed section, where it is contacted with downflowing reflux water, which cools and condenses the hot vapor and reduces solvent entrainment. The overhead stream passes through a de-entrainment device and exits the top of the column. The overhead gas then passes through the reflux condenser in order to recover the overhead liquid. The cooled liquid and vapor phases are separated in the reflux drum. The reflux liquid is pumped to the trayed section of the regenerator and the acid gas stream is sent to the Claus plant for further processing, as discussed in Section 4.4.11. The lean solvent exiting the stripper is first cooled by providing heat to the rich solvent, then further cooled by exchange with the product gas and finally chilled in the lean chiller before returning to the top of the CO_2 absorber.

The Selexol process unit can be constructed primarily out of killed carbon steel, which is a deoxidized steel that provides limited or no aging, and a harder material. High severity areas of the Selexol process require stainless steel.

High severity areas are defined as:

- High temperature
- CO_2/H_2S evolution
- High turbulence
- Areas not normally wetted by Selexol solvent

The reboiler, absorber, and regenerator packing and internals, regenerator column and reflux circuit, water wash section, and reflux pump impellers and casings are considered high severity areas.

4.9.2 CO_2 Compression and Drying System

The CO_2 discharge pressures from the AGR for each capture case are identical, as are the suction pressures at the CO_2 compressor inlets. Therefore, the CO_2 compressor specifications, stage pressure ratios and outlet stage pressures are identical for each of the CO_2 capture cases with the major differentiator being the inlet CO_2 flowrates.

The compression system was modeled based on vendor supplied data and using elements of the compressor design presented in the Carbon Capture Simulation Initiative's paper "Centrifugal Compressor Simulation User Manual." [58] The design was assumed to be an eight-stage front-loaded integrally geared centrifugal compressor with feed streams at stage one and stage three. The stage discharge pressures are presented in Exhibit 4-36.

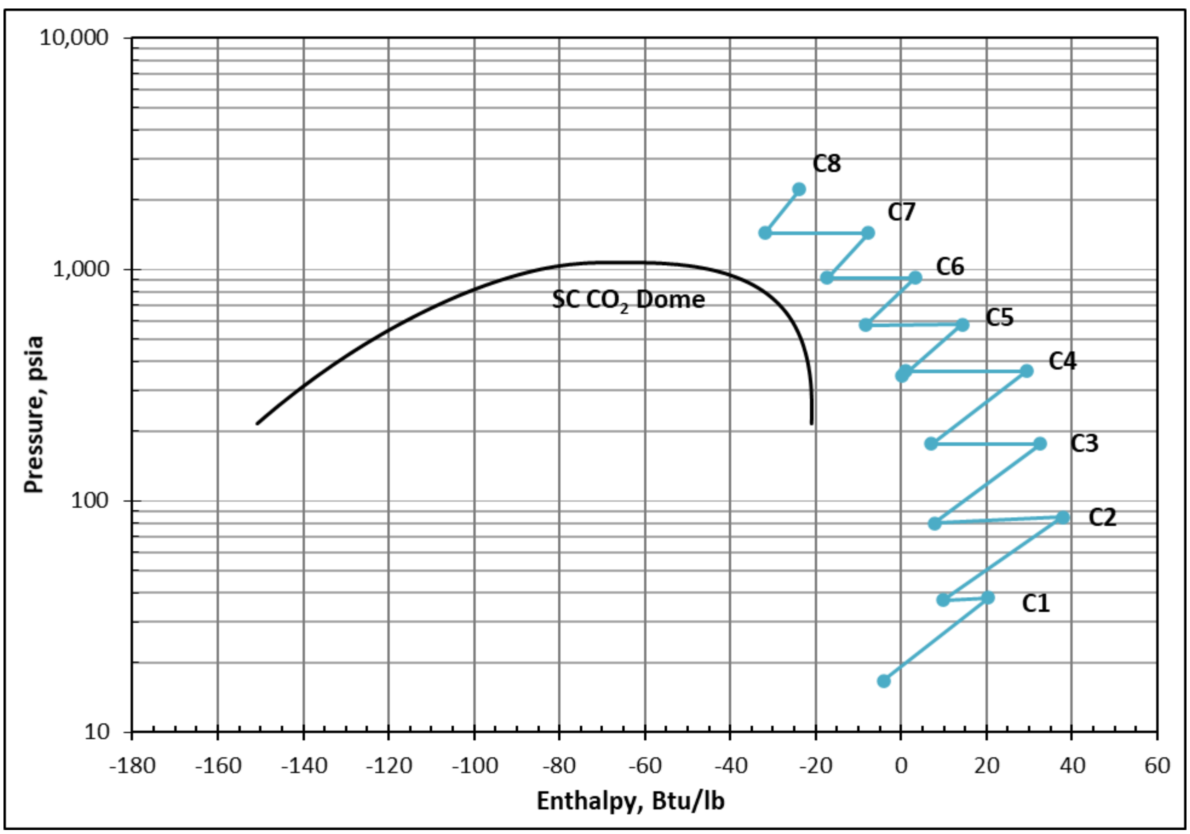
Exhibit 4-36. CO_2 compressor interstage pressures

Stage	Outlet Pressure, MPa (psia)	Stage Pressure Ratio
1	0.26 (38)	2.28
2	0.59 (85)	2.28
3	1.22 (177)	2.21
4	2.51 (364)	2.07
5	3.97 (576)	1.66
6	6.34 (920)	1.60
7	9.88 (1,433)	1.56
8	15.28 (2,217)	1.55

The AGR produces CO_2 at two pressure levels and contains approximately 99 percent CO_2. The LP CO_2 stream enters the first stage of the CO_2 compressor at 0.1 MPa (16.7 psia) and is compressed to 0.59 MPa (85 psia) in the first two compression stages with intercooling. The LP CO_2 stream exiting stage two of compression is flashed to 0.55 MPa (80 psia) and is combined with the MP CO_2 stream prior to stage three of compression. The combined stream is compressed in the following two stages to 2.51 MPa (364 psia) with intercooling, after which the combined stream is dehydrated using a triethylene glycol (TEG) dryer. The dried CO_2 stream is then further compressed in the final four stages, with intercooling, to the target product pressure of 15.27 MPa (2,215 psia).

Due to the temperature rise with compression, intercooling is included for each stage with the first three stages including water knockout. The first five intercoolers cool the CO_2 to 29 °C (85 °F), the sixth intercooler cools the CO_2 to 40 °C (104 °F), and the final intercooler cools the CO_2 to 55 °C (131 °F). The increased temperature is utilized in the final two stages of intercooling to provide a suitable buffer between the compressor operating profile and SC CO_2 dome. A CO_2 product aftercooler is also included to cool the CO_2 30 °C (86 °F). CO_2 transportation and storage costs assume that the CO_2 enters the transport pipeline as a dense phase liquid; thus, a pipeline inlet temperature of 30 °C (86 °F) is considered. Exhibit 4-37 shows the enthalpy versus pressure plot for the CO_2 compressor modeled in Case 5. Reference conditions for the data are 0.01 °C (32.02 °F) and 0.0006 MPa (0.089 psia), the same as those used for stream table results. Data points representing compression stage discharge pressures are labeled with the compression stage number (e.g., C1). Given that the two gasification cases with CO_2 capture utilize the same AGR, and thus, have the same CO_2 discharge pressures, the operating profile is identical across the two gasification cases with CO_2 capture. The CO_2 aftercooler is not represented in the compressor operating profile plot.

Exhibit 4-37. Case 5 CO_2 compressor enthalpy versus pressure operating profile



A TEG dehydration unit is included between stages 4 and 5, operating at 2.39 MPa (346 psia), to reduce the moisture concentration of the CO_2 stream to 500 ppmv. The dryer was designed based on a paper published by the Norwegian University of Science and Technology. [60]

In an absorption process, such as in a TEG dehydration unit, the gas containing water flows up through a column while the TEG flows downward. The solvent preferentially binds the water by physical absorption. The dried gas exits at the top of the column, while the solvent rich in water exits at the bottom. After depressurization to around atmospheric pressure, the solvent is regenerated by heating it and passing it through a regeneration column where the water is boiled off. A TEG unit is capable of reducing water concentrations to meet the QGESS design point of 500 ppmv. [59]

4.9.3 Hydrogen Production

No differences from Case 4.

4.9.4 Steam Generation

The steam conditions for the coal gasification plant with capture (Case 5) are 12.4 MPa/510 °C/510 °C (1,800 psig/950 °F/950 °F).

4.10 Case 5 – Performance Results

Overall performance for the plant is summarized in Exhibit 4-38. The plant produces 27,500 kg/hr (60,627 lb/hr) of hydrogen at an effective thermal efficiency of 64.1 percent (HHV basis). The removal of CO_2 from the syngas results in a higher quality PSA off-gas than in Case 4 and leads to higher grade heat available for steam generation. The steam turbine generates 109 MWe and the total auxiliary load for the plant is 148 MWe, requiring 39 MW of electricity to be sourced from the grid.

Exhibit 4-39 provides a detailed breakdown of the auxiliary power requirements. The ASU accounts for approximately 50 percent of the auxiliary load between the MAC, booster compressor, N_2 compressors, O_2 pump, and ASU auxiliaries. The two-stage Selexol process and CO_2 compression account for an additional 29 percent of the auxiliary power load. The hydrogen compressor accounts for about 7 percent. The BFW and circulating water system, including the circulating water pumps and cooling tower fan, composes approximately 5 percent of the load, with all other systems together constituting the remaining 9 percent of the auxiliary load.

Exhibit 4-38. Case 5 plant performance summary

Performance Summary	
Steam Turbine Power, MWe	109
Total Gross Power, MWe	**109**
Air Separation Unit Main Air Compressor, kWe	65,800
Air Separation Unit Booster Compressor, kWe	5,170

Performance Summary	
N_2 Compressors, kWe	2,390
CO_2 Compression, kWe	31,880
Acid Gas Removal, kWe	11,640
Balance of Plant, kWe	31,550
Total Auxiliaries, MWe	**148**
Net Power, MWe	**-39**
Hydrogen Production, kg/hr (lb/hr)	**27,500 (60,627)**
CO2 Capture, %	**92.5%**
HHV Effective Thermal Efficiency[A], %	64.1%
HHV Cold Gas Efficiency[B], %	66.6%
LHV Effective Thermal Efficiency[A], %	55.9%
LHV Cold Gas Efficiency[B], %	58.4%
Steam Turbine Cycle Efficiency, %	42.7%
Steam Turbine Heat Rate, kJ/kWh (Btu/kWh)	8,430 (7,991)
Condenser Duty, GJ/hr (MMBtu/hr)	615 (583)
AGR Cooling Duty, GJ/hr (MMBtu/hr)	148 (141)
As-Received Coal Feed, kg/hr (lb/hr)	211,967 (467,308)
HHV Thermal Input, kWt	1,597,711
LHV Thermal Input, kWt	1,541,012
Raw Water Withdrawal, m^3/min (gpm)	13.8 (3,638)
Raw Water Consumption, m^3/min (gpm)	10.8 (2,866)

[A]ETE = (Hydrogen Heating Value + Net Power) / Fuel Heating Value
[B]CGE = Hydrogen Heating Value / Fuel Heating Value

Exhibit 4-39. Case 5 plant power summary

Power Summary	
Steam Turbine Power, MWe	109
Total Gross Power, MWe	**109**
Auxiliary Load Summary	
Acid Gas Removal, kWe	11,640
Air Blower, kWe	780
Air Separation Unit Auxiliaries, kWe	1,000
Air Separation Unit Main Air Compressor, kWe	65,800
Air Separation Unit Booster Compressor, kWe	5,170
Ammonia Wash Pumps, kWe	120
Circulating Water Pumps, kWe	3,340

Power Summary	
Claus Plant TG Recycle Compressor, kWe	1,240
Claus Plant/TGTU Auxiliaries, kWe	250
CO_2 Compression, kWe	31,880
Coal Handling, kWe	460
Coal Milling, kWe	2,180
Condensate Pumps, kWe	150
Cooling Tower Fans, kWe	1,730
Dryer Air Compressor, kWe	80
Feedwater Pumps, kWe	1,600
Gasifier Water Pump, kWe	40
Ground Water Pumps, kWe	330
Hydrogen Compressor, kWe	10,720
Miscellaneous Balance of Plant[A], kWe	2,520
N_2 Compressors, kWe	2,390
O_2 Pump, kWe	330
Quench Water Pump, kWe	330
Shift Steam Pump, kWe	310
Slag Handling, kWe	550
Slag Reclaim Water Recycle Pump, kWe	0
Slurry Water Pump, kWe	0
Sour Gas Compressors, kWe	140
Sour Water Recycle Pumps, kWe	0
Steam Turbine Auxiliaries, kWe	90
Syngas Recycle Compressor, kWe	880
Syngas Scrubber Pumps, kWe	120
Process Water Treatment Auxiliaries, kWe	1,310
Transformer Losses, kWe	950
Total Auxiliaries, MWe	**148**
Net Power, MWe	**-39**

[A]Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

4.10.1 Environmental Performance

The environmental targets for emissions of Hg, HCl, NOx, SO_2, CO_2, and PM were presented in Section 2.7.1. A summary of the plant air emissions for Case 5 is presented in Exhibit 4-40.

Exhibit 4-40. Case 5 air emissions

	kg/GJ (lb/MMBtu)	Tonne/year (ton/year)[A]	lb/lb H_2
SO_2	0.000 (0.000)	0 (0)	0.000
NOx	0.003 (0.007)	121 (133)	0.001
Particulate	0.003 (0.007)	123 (136)	0.001
Hg	2.46E-7 (5.71E-7)	0.010 (0.011)	5.14E-8
HCl	0.000 (0.000)	0.00 (0.00)	0.000
CO_2[B]	6 (15)	259,485 (286,033)	1.3
CO_2e[C]	19 (44)	757,406 (834,897)	4.1

[A]Calculations based on an 80 percent capacity factor
[B]Plant stack emissions
[C]Life cycle emissions

The low level of SO_2 emissions is achieved by capturing the sulfur in the gas by the two-stage Selexol AGR process. The CO_2 capture target results in more sulfur compounds being removed than required in the environmental targets of Section 2.7.1. The clean syngas exiting the AGR process has a sulfur concentration of approximately 0 ppmv. The H_2S-rich regeneration gas from the AGR system is fed to a Claus plant, producing elemental sulfur. The Claus plant tail gas is compressed and recycled back to the AGR where most of the remaining sulfur is removed.

NOx, PM, and Hg emissions are the same as described in Case 4.

Ninety five percent of the CO_2 from the syngas is captured in the AGR system and compressed for storage. Because not all CO is converted to CO_2 in the shift reactors, the overall carbon removal is 92.5 percent. Additionally, an LCA was conducted to determine the total GWP of the plant.

The carbon balance for the plant is shown in Exhibit 4-41. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal. Carbon leaves the plant as unburned carbon in the slag, a small portion of the hydrogen product, the captured CO_2 product, and the CO_2 in the stack gas, including the coal dryer vent gas and ASU vent gas. The carbon capture efficiency is defined as one minus the mass of carbon in the stack gas relative to the total carbon in less carbon contained in the slag, represented by the following equation:

$$Carbon\ Capture\ Efficiency\ (\%) = \left(1 - \left(\frac{Carbon\ in\ Stack}{(Total\ Carbon\ In) - (Carbon\ in\ Slag)}\right)\right) * 100\%$$

Exhibit 4-41. Case 5 carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	135,118 (297,884)	Stack Gas	10,105 (22,278)
Air (CO_2)	138 (304)	CO_2 Product	124,472 (274,414)
		Slag	676 (1,489)
		H_2 Product	3 (6)
Total	**135,256 (298,188)**	**Total**	**135,256 (298,188)**

Exhibit 4-42 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant and sulfur in the CO_2 product. Sulfur in the slag is considered negligible.

Exhibit 4-42. Case 5 sulfur balance

Sulfur In		Sulfur Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	5,313 (11,713)	Stack Gas	–
		CO_2 Product	1 (2)
		Elemental Sulfur	5,312 (11,711)
Total	**5,313 (11,713)**	**Total**	**5,313 (11,713)**

Exhibit 4-43 shows the overall water balance for the plant. The exhibit is presented in an identical manner as for Case 4.

Exhibit 4-43. Case 5 water balance

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)
Slag Handling	0.46 (122)	0.46 (122)	–	–	–
Slurry Water	–	–	–	–	–
Gasifier Water	–	–	–	–	–
Quench	3.07 (812)	3.07 (812)	–	–	–
HCl Scrubber	1.88 (497)	1.88 (497)	–	–	–
NH_3 Scrubber	1.77 (467)	0.98 (258)	0.79 (209)	–	0.79 (209)
Gasifier Steam	0.34 (90)	0.34 (90)	–	–	–
Condenser Makeup	**0.10 (26)**	**–**	**0.10 (26)**	**–**	**0.10 (26)**
BFW Makeup	0.10 (26)	–	0.10 (26)	–	0.10 (26)
Gasifier Steam	–	–	–	–	–
Shift Steam	–	–	–	–	–
Cooling Tower	**13.00 (3,435)**	**0.12 (32)**	**12.88 (3,403)**	**2.92 (773)**	**9.96 (2,630)**
BFW Blowdown	–	0.10 (26)	-0.10 (-26)	–	-0.10 (-26)
ASU Knockout	–	0.02 (6)	-0.02 (-6)	–	-0.02 (-6)
Total	**20.63 (5,449)**	**6.86 (1,811)**	**13.77 (3,638)**	**2.92 (773)**	**10.85 (2,866)**

An overall plant energy balance is provided in tabular form in Exhibit 4-44. The power out is the steam turbine power prior to generator losses.

Exhibit 4-44. Case 5 overall energy balance (0 °C [32 °F] reference)

	HHV	Sensible + Latent	Power	Total
Heat In, GJ/hr (MMBtu/hr)				
Coal	5,752 (5,452)	4.8 (4.6)	–	5,757 (5,456)
Air	–	30.6 (29.0)	–	30.6 (29.0)
Raw Water Makeup	–	51.8 (49.1)	–	51.8 (49.1)
Auxiliary Power	–	–	534.3 (506.5)	534.3 (506.5)
Total	**5,752 (5,452)**	**87.2 (82.7)**	**534.3 (506.5)**	**6,373 (6,041)**
Heat Out, GJ/hr (MMBtu/hr)				
Misc. Process Steam	–	4.8 (4.6)	–	4.8 (4.6)
Slag	22.1 (21.0)	35.8 (34.0)	–	58.0 (55.0)
Stack Gas	–	197 (187)	–	197 (187)
Sulfur	49.2 (46.7)	0.6 (0.6)	–	49.9 (47.3)
Motor Losses and Design Allowances	–	–	18.8 (17.8)	18.8 (17.8)
Hydrogen Product	3,830 (3,630)	12.1 (11.5)	–	3,842 (3,641)
Cooling Tower Load[A]	–	1,698 (1,609)	–	1,698 (1,609)
CO_2 Product Stream	–	-103.6 (-98.2)	–	-103.6 (-98.2)
Blowdown Streams	–	25.3 (24.0)	–	25.3 (24.0)
Ambient Losses[B]	–	88.9 (84.3)	–	88.9 (84.3)
Power	–	–	393 (373)	393 (373)
Total	**3,901 (3,698)**	**1,959 (1,857)**	**412 (391)**	**6,272 (5,945)**
Unaccounted Energy[C]	–	101.1 (95.8)	–	101.1 (95.8)

[A]Includes condenser, AGR, and miscellaneous cooling loads
[B]Ambient losses include all losses to the environment through radiation, convection, etc. Sources of these losses include the combustor, reheater, superheater, and transformers
[C]By difference

4.10.2 Energy and Mass Balance Diagrams

Energy and mass balance diagrams are shown for the following subsystems in Exhibit 4-45 through Exhibit 4-47:

- ASU, gasification, and gas cooling
- Gas cleanup system
- Hydrogen purification and power block

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Exhibit 4-45. Case 5 coal gasification and ASU energy and mass balance

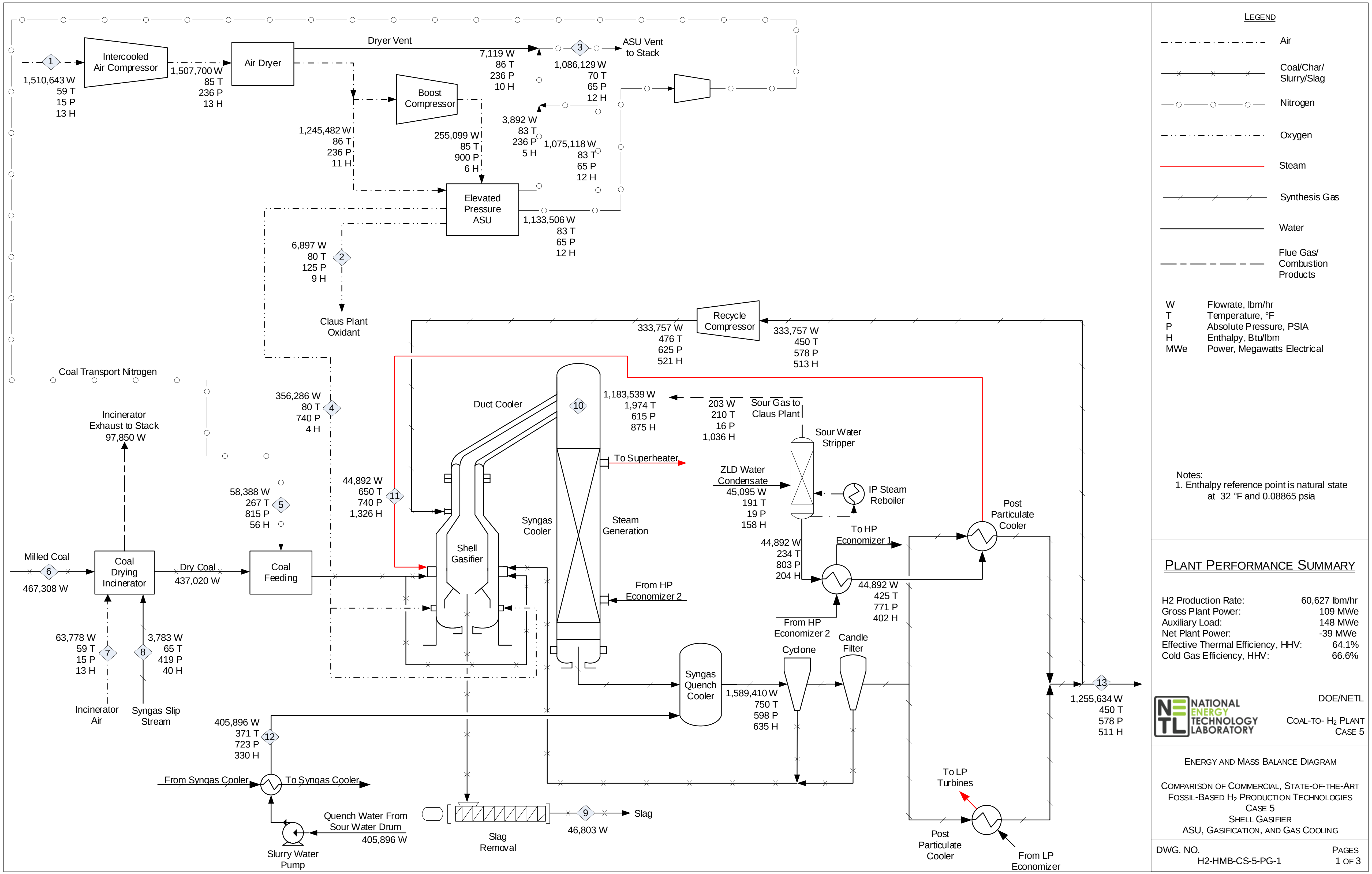


Exhibit 4-46. Case 5 syngas cleanup energy and mass balance

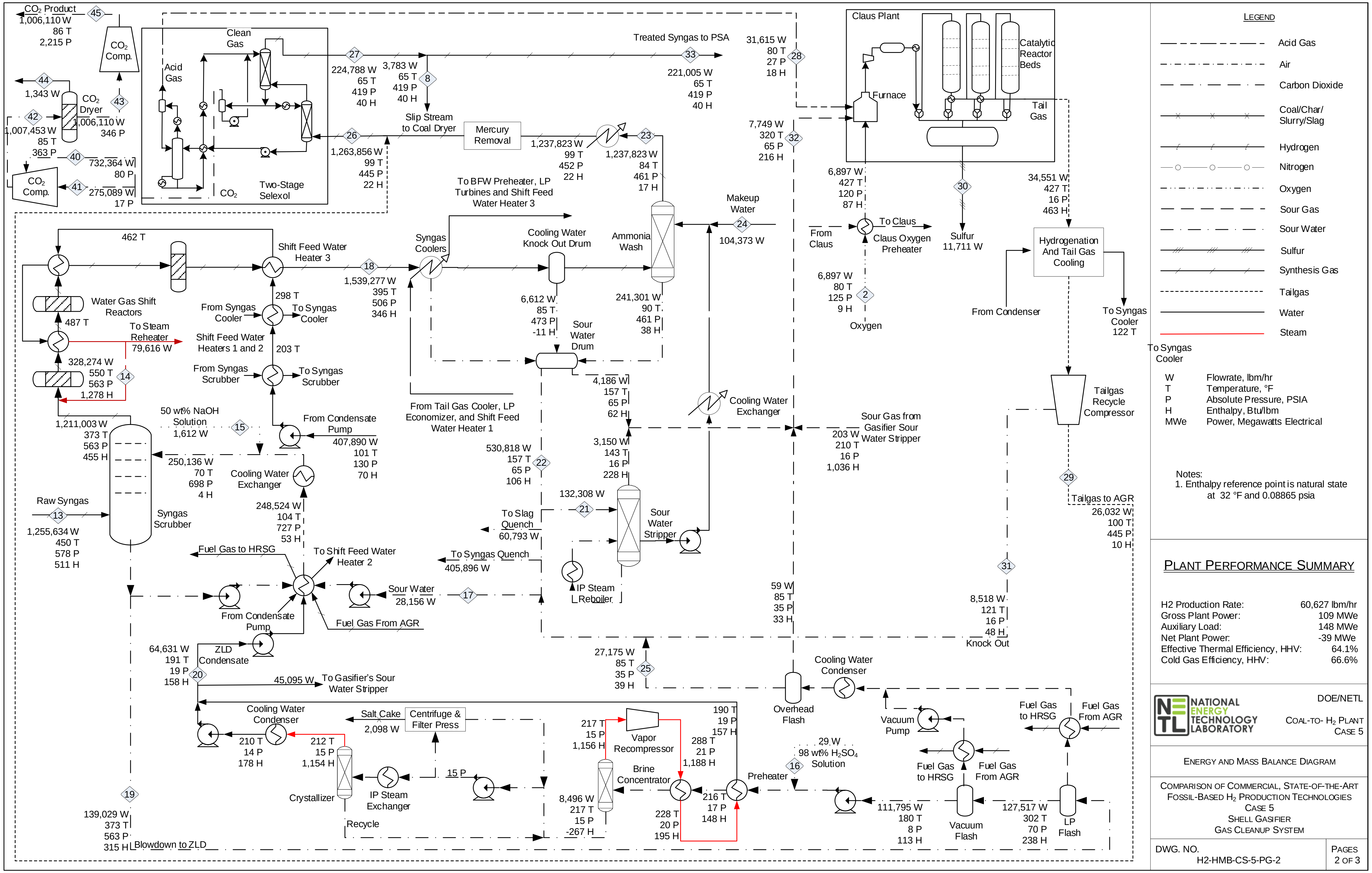
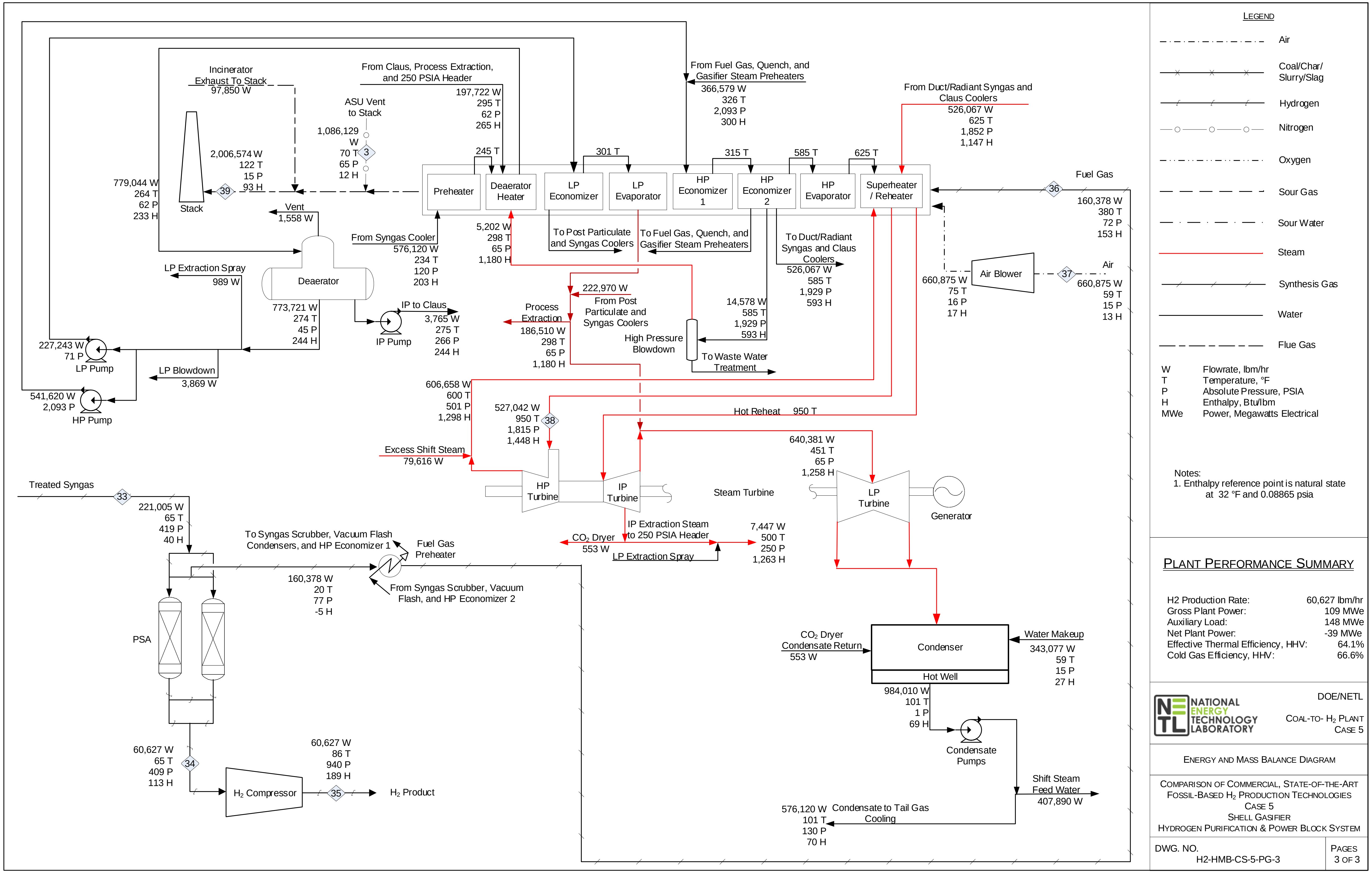


Exhibit 4-47. Case 5 hydrogen purification and power block energy and mass balance



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4.11 Case 5 – Major Equipment List

Major equipment items for the Shell gasifier with CO_2 capture are shown in the following tables. In general, the design conditions include a 10 percent design allowance for flows and heat duties and a 21 percent design allowance for heads on pumps and fans.

Case 5 – Account 1: Coal Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	181 tonne (200 ton)	2	0
2	Feeder	Belt	570 tonne/hr (630 tph)	2	0
3	Conveyor No. 1	Belt	1,130 tonne/hr (1,250 tph)	1	0
4	Conveyor No. 2	Belt	1,130 tonne/hr (1,250 tph)	1	0
5	Stacker/Reclaimer	Traveling, linear	1,130 tonne/hr (1,250 tph)	1	0
6	Reclaim Hopper	N/A	40 tonne (50 ton)	2	1
7	Feeder	Vibratory	170 tonne/hr (190 tph)	2	1
8	Conveyor No. 3	Belt w/ tripper	350 tonne/hr (390 tph)	1	0
9	Coal Surge Bin w/ Vent Filter	Dual outlet	170 tonne (190 ton)	2	0
10	Crusher	Impactor reduction	8 cm x 0 – 3 cm x 0 (3" x 0 – 1-1/4" x 0)	2	0
11	Conveyor No. 4	Belt w/tripper	350 tonne/hr (390 tph)	1	0
12	Conveyor No. 5	Belt w/ tripper	350 tonne/hr (390 tph)	1	0
13	Coal Silo w/ Vent Filter and Slide Gates	Field erected	780 tonne (860 ton)	3	0

Case 5 – Account 2: Coal Preparation and Feed

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Feeder	Vibratory	80 tonne/hr (90 tph)	3	0
2	Conveyor No. 6	Belt w/tripper	230 tonne/hr (260 tph)	1	0
3	Roller Mill Feed Hopper	Dual Outlet	470 tonne (510 ton)	1	0
4	Weigh Feeder	Belt	120 tonne/hr (130 tph)	2	0
5	Coal Dryer and Pulverizer	Rotary	120 tonne/hr (130 tph)	2	0
6	Coal Dryer Feed Hopper	Vertical Hopper	230 tonne (260 ton)	2	0

Case 5 – Account 3: Feedwater and Miscellaneous Balance of Plant Systems

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Demineralized Water Storage Tank	Vertical, cylindrical, outdoor	10,277,000 liters (2,715,000 gal)	2	0
2	Condensate Pumps	Vertical canned	4,120 lpm @ 90 m H_2O (1,090 gpm @ 300 ft H_2O)	2	1
3	Deaerator (integral w/ GFB)	Horizontal spray type	194,000 kg/hr (428,000 lb/hr)	2	0
4	Intermediate Pressure Feedwater Pump	Horizontal centrifugal, single stage	980 lpm @ 20 m H_2O (260 gpm @ 70 ft H_2O)	2	1
5	High Pressure Feedwater Pump No. 1	Barrel type, multi-stage, centrifugal	HP water: 2,340 lpm @ 1,700 m H_2O (620 gpm @ 5,700 ft H_2O)	2	1
6	High Pressure Feedwater Pump No. 2	Barrel type, multi-stage, centrifugal	IP water: 1,590 lpm @ 210 m H_2O (420 gpm @ 670 ft H_2O)	2	1

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
7	Auxiliary Boiler	Shop fabricated, water tube	18,000 kg/hr, 2.8 MPa, 343 °C (40,000 lb/hr, 400 psig, 650 °F)	1	0
8	Service Air Compressors	Flooded Screw	28 m^3/min @ 0.7 MPa (1,000 scfm @ 100 psig)	2	1
9	Instrument Air Dryers	Duplex, regenerative	28 m^3/min (1,000 scfm)	2	1
10	Closed Cycle Cooling Heat Exchangers	Plate and frame	514 GJ/hr (487 MMBtu/hr) each	2	0
11	Closed Cycle Cooling Water Pumps	Horizontal centrifugal	184,400 lpm @ 20 m H_2O (48,700 gpm @ 70 ft H_2O)	2	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	3,785 lpm @ 110 m H_2O (1,000 gpm @ 350 ft H_2O)	1	1
13	Fire Service Booster Pump	Two-stage horizontal centrifugal	2,650 lpm @ 80 m H_2O (700 gpm @ 250 ft H_2O)	1	1
14	Municipal Water Pumps	Stainless steel, single suction	1,940 lpm @ 20 m H_2O (510 gpm @ 60 ft H_2O)	2	1
15	Ground Water Pumps	Stainless steel, single suction	3,870 lpm @ 270 m H_2O (1,020 gpm @ 880 ft H_2O)	1	1
16	Filtered Water Pumps	Stainless steel, single suction	1,860 lpm @ 50 m H_2O (490 gpm @ 160 ft H_2O)	2	1
17	Filtered Water Tank	Vertical, cylindrical	893,000 liter (236,000 gal)	2	0
18	Makeup Water Demineralizer	Anion, cation, and mixed bed	1,430 lpm (380 gpm)	2	0
19	Liquid Waste Treatment System	N/A	10 years, 24-hour storm	1	0
20	Process Water Treatment	Vacuum flash, brine concentrator, and crystallizer	Vacuum Flash Inlet: 35,000 kg/hr (76,000 lb/hr) Vacuum Flash Outlet: 6,218 ppmw Cl- Brine Concentrator Inlet: 28,000 kg/hr (61,000 lb/hr) Crystallizer Inlet: 2,000 kg/hr (5,000 lb/hr)	2	0

Case 5 – Account 4: Gasifier, ASU, and Accessories

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Gasifier	Pressurized dry-feed, entrained bed	2,800 tonne/day, 4.2 MPa (3,100 tpd, 615 psia)	2	0
2	Synthesis Gas Cooler	Convective spiral-wound tube boiler	295,000 kg/hr (651,000 lb/hr)	2	0
3	Synthesis Gas Cyclone	High efficiency	397,000 kg/hr (874,000 lb/hr) Design efficiency 90%	2	0
4	HCl Scrubber	Ejector Venturi	313,000 kg/hr (691,000 lb/hr)	2	0
5	Ammonia Wash	Counter-flow spray tower	311,000 kg/hr (685,000 lb/hr) @ 3.3 MPa (473 psia)	2	0
6	Primary Sour Water Stripper	Counter-flow with external reboiler	33,000 kg/hr (73,000 lb/hr)	2	0
7	Secondary Sour Water Stripper	Counter-flow with external reboiler	11,000 kg/hr (25,000 lb/hr)	2	0
8	Low Temperature Heat Recovery Coolers	Shell and tube with condensate drain	256,000 kg/hr (564,000 lb/hr)	9	0
9	Low Temperature Heat Recovery Knockout Drum	Vertical with mist eliminator	312,000 kg/hr, 59 °C, 3.3 MPa (689,000 lb/hr, 138 °F, 476 psia)	2	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
10	Synthesis Gas Reheaters	Shell and tube	Reheater 1: N/A Reheater 2: 31,000 kg/hr (69,000 lb/hr) Reheater 3: 9,000 kg/hr (20,000 lb/hr) Reheater 4: N/A Reheater 5: 40,000 kg/hr (88,000 lb/hr) Reheater 6: 40,000 kg/hr (88,000 lb/hr)	2	0
11	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	302,000 kg/hr (666,000 lb/hr) syngas	2	0
12	ASU Main Air Compressor	Centrifugal, multi-stage	5,000 m^3/min @ 1.6 MPa (182,000 scfm @ 236 psia)	2	0
13	Cold Box	Vendor design	2,200 tonne/day (2,400 tpd) of 95% purity oxygen	2	0
14	Gasifier O_2 Pump	Centrifugal, multi-stage	1,000 m^3/min (38,000 scfm) Suction – 1.0 MPa (130 psia) Discharge – 5.1 MPa (740 psia)	2	0
15	Gasifier Nitrogen Boost Compressor	Centrifugal, single-stage	210 m^3/min (7,000 scfm) Suction – 0.4 MPa (70 psia) Discharge – 5.6 MPa (820 psia)	2	0

Case 5 – Account 5: Syngas Cleanup

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Mercury Adsorber 1	Sulfated carbon bed	309,000 kg/hr (681,000 lb/hr) 29 °C (84 °F) 3.2 MPa (461 psia)	2	0
2	Mercury Adsorber 2	Sulfated carbon bed	309,000 kg/hr (681,000 lb/hr) 37 °C (99 °F) 3.1 MPa (448 psia)	2	0
3	Sulfur Plant	Claus type	140 tonne/day (155 tpd)	1	0
4	Water Gas Shift Reactors	Fixed bed, catalytic	128,000 kg/hr (282,000 lb/hr) 216 °C (420 °F) 3.9 MPa (560 psia)	6	0
5	Shift Reactor Heat Recovery Exchangers	Shell and Tube	Exchanger 1: 223 GJ/hr (211 MMBtu/hr) Exchanger 2: 91 GJ/hr (86 MMBtu/hr) Exchanger 3: 66 GJ/hr (63 MMBtu/hr) Exchanger 4: 83 GJ/hr (78 MMBtu/hr)	8	0
6	Acid Gas Removal Plant	Two-stage Selexol	631,000 kg/hr (1,390,000 lb/hr) 37 °C (99 °F) 3.1 MPa (445 psia)	1	0
7	Hydrogenation Reactor	Fixed bed, catalytic	17,000 kg/hr (38,000 lb/hr) 219 °C (427 °F) 0.1 MPa (16.4 psia)	1	0
8	Tail Gas Recycle Compressor	Centrifugal	13,000 kg/hr (29,000 lb/hr) each	1	0
9	Candle Filter	Pressurized filter with pulse-jet cleaning	metallic filters	2	0
10	CO_2 Dryer	Triethylene glycol	Inlet: 153 m^3/min @ 2.5 MPa (5,403 acfm @ 363 psia) Outlet: 2.4 MPa (346 psia)	1	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
			Water Recovered: 609 kg/hr (1,343 lb/hr)		
11	CO_2 Compressor	Integrally geared, multi-stage centrifugal	10 m^3/min @ 15.3 MPa (353 acfm @ 2,217 psia)	1	0
12	CO_2 Aftercooler	Shell and tube heat exchanger	Outlet: 15.3 MPa, 30 °C (2,215 psia, 86 °F) Duty: 78 GJ/hr (74 MMBtu/hr)	1	0

Case 5 – Account 6: Hydrogen Production

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Pressure Swing Adsorber	Polybed Proprietary	Syngas: 100,246 kg/hr (221,005 lb/hr) 18 °C (65 °F) 2.9 MPa (418.7 psia) Hydrogen: 27,500 kg/hr (60,627 lb/hr) 18 °C (65 °F) 2.8 MPa (408.7 psia) Off Gas: 72,746 kg/hr (160,378 lb/hr) -7 °C (20 °F) 0.5 MPa (77.0 psia)	1	0
2	Hydrogen Compressor	Integrally geared, multi-stage centrifugal	99 m^3/min @ 6.5 MPa (3,484 acfm @ 940 psia)	1	1

Case 5 – Account 7: Stack, Ducting, & Off-Gas Fired Boiler

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Stack	CS plate, type 409SS liner	76 m (250 ft) high x 3.6 m (12 ft) diameter	1	0
2	Field-Erected Gas-Fired Boiler	Drum, multi-pressure with economizer section and integral deaerator Air-Fired	Off Gas: 72,746 kg/hr (160,378 lb/hr) 0.5 MPa (72.5 psia) 193 °C (380 °F) Air: 299,768 kg/hr (660,875 lb/hr) 0.1 MPa (16.0 psia) 24 °C (75 °F) 632 GJ/hr (599 MMBtu/hr)	1	0
3	Air Blower	Centrifugal	329,744 kg/hr, 4,500 m^3/min @ 111 cm WG (726,962 lb/hr, 158,905 acfm @ 44 in. WG)	2	1

Case 5 – Account 8: Steam Turbine and Accessories

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Steam Turbine	Commercially available advanced steam turbine	115 MW 12.4 MPa/510 °C/510 °C (1,800 psig/ 950 °F/950 °F)	1	0
2	Steam Turbine Generator	Hydrogen cooled, static excitation	130 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	1	0
3	Surface Condenser	Single pass, divided waterbox including vacuum pumps	680 GJ/hr (640 MMBtu/hr), Inlet water temperature 16 °C (60 °F), Water temperature rise 11 °C (20 °F)	1	0
4	Steam Bypass	One per boiler	50% steam flow @ design steam conditions	1	0

Case 5 – Account 9: Cooling Water System

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Circulating Water Pumps	Vertical, wet pit	335,000 lpm @ 30 m (89,000 gpm @ 100 ft)	2	1
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	11 °C (52 °F) wet bulb/ 16 °C (60 °F) CWT/27 °C (80 °F) HWT/ 1,870 GJ/hr (1,770 MMBtu/hr) heat duty	1	0

Case 5 – Account 10: Slag Recovery and Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Slag Quench Tank	Water bath	223,000 liters (59,000 gal)	2	0
2	Slag Crusher	Roll	12 tonne/hr (13 tph)	2	0
3	Slag Depressurizer	Lock Hopper	12 tonne/hr (13 tph)	2	0
4	Slag Receiving Tank	Horizontal, weir	134,000 liters (35,000 gal)	2	0
5	Black Water Overflow Tank	Shop fabricated	60,000 liters (16,000 gal)	2	0
6	Slag Conveyor	Drag chain	12 tonne/hr (13 tph)	2	0
7	Slag Separation Screen	Vibrating	12 tonne/hr (13 tph)	2	0
8	Coarse Slag Conveyor	Belt/bucket	12 tonne/hr (13 tph)	2	0
9	Fine Ash Settling Tank	Vertical, gravity	190,000 liters (50,000 gal)	2	0
10	Fine Ash Recycle Pumps	Horizontal centrifugal	50 lpm @ 14 m H_2O (10 gpm @ 46 ft H_2O)	2	2
11	Grey Water Storage Tank	Field erected	61,000 liters (16,000 gal)	2	0
12	Grey Water Pumps	Centrifugal	210 lpm @ 430 m H_2O (60 gpm @ 1,420 ft H_2O)	2	2
13	Slag Storage Bin	Vertical, field erected	800 tonne (900 tons)	2	0
14	Unloading Equipment	Telescoping chute	100 tonne/hr (110 tph)	1	0

Case 5 – Account 11: Accessory Electric Plant

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	STG Transformer	Oil-filled	24 kV/345 kV, 80 MVA, 3-ph, 60 Hz	1	0
2	High Voltage Auxiliary Transformer	Oil-filled	345 kV/13.8 kV, 58 MVA, 3-ph, 60 Hz	2	0
3	Medium Voltage Transformer	Oil-filled	24 kV/4.16 kV, 47 MVA, 3-ph, 60 Hz	1	1
4	Low Voltage Transformer	Dry ventilated	4.16 kV/480 V, 7 MVA, 3-ph, 60 Hz	1	1
5	STG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	1	0
6	Medium Voltage Switchgear	Metal clad	4.16 kV, 3-ph, 60 Hz	1	1
7	Low Voltage Switchgear	Metal enclosed	480 V, 3-ph, 60 Hz	1	1
8	Emergency Diesel Generator	Sized for emergency shutdown	750 kW, 480 V, 3-ph, 60 Hz	1	0

Case 5 – Account 12: Instrumentation and Control

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	DCS – Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W)	Operator stations/printers and engineering stations/printers	1	0
2	DCS – Processor	Microprocessor with redundant input/output	N/A	1	0
3	DCS – Data Highway	Fiber optic	Fully redundant, 25% spare	1	0

4.12 Case 5 – Cost Estimating

The cost estimating methodology was described previously in Section 2.11. Exhibit 4-48 shows a detailed breakdown of the capital costs; Exhibit 4-49 shows the owner's costs, TOC, and TASC; Exhibit 4-50 shows the initial and annual O&M costs; and Exhibit 4-51 shows the LCOH breakdown.

The estimated TOC of the Shell gasification plant with CO_2 capture is \$6,044/[kg H_2/day]. Process contingency represents 5.0 percent of the TOC and project contingency represents 12.2 percent. The LCOH is \$3.09/kg H_2 with CO_2 T&S.

Exhibit 4-48. Case 5 total plant cost details

	Case:	5	– Coal Gasification w/ CCS				Estimate Type:			Conceptual	
	Plant Size (kg H_2/day):	660,004					Cost Base:			Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/[kg H_2/day]
	1					Coal Handling					
1.1	Coal Receive & Unload	$971	$0	$468	$0	$1,438	$216	$0	$331	$1,985	$3
1.2	Coal Stackout & Reclaim	$3,173	$0	$758	$0	$3,931	$590	$0	$904	$5,425	$8
1.3	Coal Conveyors & Yard Crush	$30,270	$0	$7,704	$0	$37,974	$5,696	$0	$8,734	$52,404	$79
1.4	Other Coal Handling	$4,715	$0	$1,061	$0	$5,776	$866	$0	$1,328	$7,970	$12
1.5	Biomass Receive & Unload	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.6	Biomass Handling	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.7	Biomass Conveyors	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.8	Biomass Hnd. Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.9	Coal & Sorbent Handling Foundations	$0	$85	$222	$0	$307	$46	$0	$71	$424	$1
	Subtotal	**$39,128**	**$85**	**$10,213**	**$0**	**$49,426**	**$7,414**	**$0**	**$11,368**	**$68,208**	**$103**
	2					Coal Preparation & Feed					
2.1	Coal Crushing & Drying	$2,349	$142	$338	$0	$2,829	$424	$0	$651	$3,904	$6
2.2	Prepared Coal Storage & Feed	$7,217	$1,734	$1,115	$0	$10,066	$1,510	$0	$2,315	$13,891	$21
2.3	Dry Coal Injection System	$9,212	$106	$844	$0	$10,162	$1,524	$0	$2,337	$14,023	$21
2.4	Miscellaneous Coal Preparation & Feed	$713	$521	$1,534	$0	$2,768	$415	$0	$637	$3,819	$6
2.5	Biomass Shredding & Drying	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.6	Prepared Biomass Storage & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.7	Dry Biomass Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.8	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.9	Coal, Biomass & Sorbent Feed Foundation	$0	$1,734	$1,488	$0	$3,221	$483	$0	$741	$4,446	$7
	Subtotal	**$19,491**	**$4,237**	**$5,318**	**$0**	**$29,045**	**$4,357**	**$0**	**$6,680**	**$40,083**	**$61**
	3					Feedwater & Miscellaneous BOP Systems					
3.1	Feedwater System	$1,027	$1,761	$880	$0	$3,669	$550	$0	$844	$5,063	$8
3.2	Water Makeup & Pretreating	$4,105	$411	$2,326	$0	$6,842	$1,026	$0	$2,361	$10,229	$15
3.3	Other Feedwater Subsystems	$531	$174	$165	$0	$870	$131	$0	$200	$1,201	$2
3.4	Service Water Systems	$1,227	$2,342	$7,584	$0	$11,153	$1,673	$0	$3,848	$16,674	$25
3.5	Other Boiler Plant Systems	$135	$49	$123	$0	$307	$46	$0	$71	$423	$1
3.6	Natural Gas Pipeline and Start-Up System	$7,197	$310	$232	$0	$7,739	$1,161	$0	$1,780	$10,679	$16
3.7	Wastewater Treatment Equipment	$5,801	$0	$3,555	$0	$9,356	$1,403	$0	$3,228	$13,987	$21
3.8	Vacuum Flash, Brine Concentrator, & Crystallizer	$25,030	$0	$15,506	$0	$40,536	$6,080	$0	$13,985	$60,602	$92
3.9	Miscellaneous Plant Equipment	$15,325	$2,010	$7,788	$0	$25,124	$3,769	$0	$8,668	$37,560	$57
	Subtotal	**$60,378**	**$7,056**	**$38,161**	**$0**	**$105,595**	**$15,839**	**$0**	**$34,983**	**$156,417**	**$237**

	Case:	5	– Coal Gasification w/ CCS				Estimate Type:			Conceptual	
	Plant Size (kg H_2/day):	660,004					Cost Base:			Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/[kg H_2/day]
4	Gasifier, ASU, & Accessories										
4.1	Gasifier & Auxiliaries (Shell)	$604,151	$0	$259,074	$0	$863,224	$129,484	$120,851	$167,034	$1,280,593	$1,940
4.2	Syngas Cooler	$61,106	$0	$26,204	$0	$87,310	$13,096	$12,223	$16,894	$129,524	$196
4.3	Air Separation Unit/Oxidant Compression	$125,401	$28,700	$29,845	$0	$183,946	$36,789	$18,395	$35,870	$275,000	$417
4.5	Miscellaneous Gasification Equipment	$4,180	$0	$1,792	$0	$5,972	$896	$0	$1,030	$7,898	$12
4.6	Low Temperature Heat Recovery & Fuel Gas Saturation	$44,687	$0	$16,977	$0	$61,664	$9,250	$0	$14,183	$85,096	$129
4.7	Flare Stack System	$1,901	$0	$335	$0	$2,236	$335	$0	$514	$3,086	$5
4.15	Major Component Rigging	$233	$0	$100	$0	$332	$50	$0	$57	$439	$1
4.16	Gasification Foundations	$0	$470	$280	$0	$751	$113	$0	$216	$1,079	$2
	Subtotal	**$841,657**	**$29,170**	**$334,608**	**$0**	**$1,205,435**	**$190,013**	**$151,469**	**$235,798**	**$1,782,715**	**$2,701**
5	Syngas Cleanup										
5.1	Double Stage Selexol	$170,726	$0	$69,733	$0	$240,459	$36,069	$48,092	$64,924	$389,544	$590
5.2	Sulfur Removal	w/5.1	w/5.1	w/5.1	$0	$0	$0	$0	$0	$0	$0
5.3	Elemental Sulfur Plant	$47,647	$9,288	$61,050	$0	$117,985	$17,698	$0	$27,137	$162,819	$247
5.4	Carbon Dioxide (CO_2) Compression & Drying	$32,713	$4,907	$13,735	$0	$51,354	$7,703	$0	$11,812	$70,869	$107
5.5	Carbon Dioxide (CO_2) Compressor Aftercooler	$483	$77	$207	$0	$767	$115	$0	$176	$1,058	$2
5.6	Mercury Removal (Carbon Bed)	$480	$0	$363	$0	$843	$126	$42	$202	$1,214	$2
5.7	Water Gas Shift (WGS) Reactors	$131,699	$0	$52,649	$0	$184,348	$27,652	$0	$42,400	$254,401	$385
5.9	Particulate Removal	$1,841	$0	$790	$0	$2,631	$395	$0	$454	$3,480	$5
5.10	Blowback Gas Systems	$838	$471	$263	$0	$1,571	$236	$0	$361	$2,168	$3
5.11	Fuel Gas Piping	$0	$1,115	$730	$0	$1,844	$277	$0	$424	$2,545	$4
5.12	Gas Cleanup Foundations	$0	$221	$149	$0	$371	$56	$0	$128	$554	$1
	Subtotal	**$386,428**	**$16,078**	**$199,668**	**$0**	**$602,174**	**$90,326**	**$48,134**	**$148,018**	**$888,653**	**$1,346**
6	Hydrogen Production										
6.1	Pressure Swing Adsorber	$20,140	$23	$932	$727	$21,822	$3,273	$0	$5,019	$30,114	$46
6.2	Hydrogen Compressor	$5,166	$273	$524	$408	$6,371	$956	$0	$1,465	$8,792	$13
	Subtotal	**$25,306**	**$296**	**$1,455**	**$1,135**	**$28,193**	**$4,229**	**$0**	**$6,484**	**$38,906**	**$59**
7	Off Gas Fired Boiler, Ductwork, & Stack										
7.1	Off Gas Fired Boiler	$3,880	$375	$426	$332	$5,013	$752	$0	$1,153	$6,918	$10
7.2	Off Gas Fired Boiler Accessories	$398	$320	$496	$387	$1,601	$240	$0	$368	$2,210	$3
7.3	Ductwork	w/7.1	w/7.1	w/7.1	w/7.1	$0	$0	$0	$0	$0	$0
7.4	Stack	$579	$77	$211	$164	$1,032	$155	$0	$297	$1,483	$2
7.5	Off Gas Fired Boiler, Ductwork & Stack Foundations	w/7.1	w/7.1	w/7.1	w/7.1	$0	$0	$0	$0	$0	$0
	Subtotal	**$4,857**	**$773**	**$1,133**	**$883**	**$7,646**	**$1,147**	**$0**	**$1,818**	**$10,611**	**$16**

	Case:	5	– Coal Gasification w/ CCS				Estimate Type:		Conceptual	
	Plant Size (kg H_2/day):	660,004					Cost Base:		Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	Total Plant Cost $/1,000	Total Plant Cost $/[kg H_2/day]
8	Steam Turbine & Accessories										
8.1	Steam Turbine Generator & Accessories	$32,224	$1,952	$3,109	$2,425	$39,710	$5,957	$0	$9,133	$54,800	$83
8.2	Steam Turbine Plant Auxiliaries	w/8.1	w/8.1	w/8.1	w/8.1	$0	$0	$0	$0	$0	$0
8.3	Condenser & Auxiliaries	$1,216	$647	$609	$475	$2,948	$442	$0	$848	$4,238	$6
8.4	Steam Piping	w/8.1	w/8.1	w/8.1	w/8.1	$0	$0	$0	$0	$0	$0
8.5	Turbine Generator Foundations	w/8.1	w/8.1	w/8.1	w/8.1	$0	$0	$0	$0	$0	$0
	Subtotal	**$33,440**	**$2,599**	**$3,719**	**$2,900**	**$42,658**	**$6,399**	**$0**	**$9,981**	**$59,038**	**$89**
9	Cooling Water System										
9.1	Cooling Towers	$9,205	$0	$2,785	$0	$11,990	$1,799	$0	$2,068	$15,857	$24
9.2	Circulating Water Pumps	$1,198	$0	$88	$0	$1,286	$193	$0	$222	$1,701	$3
9.3	Circulating Water System Auxiliaries	$8,413	$0	$1,195	$0	$9,608	$1,441	$0	$1,657	$12,707	$19
9.4	Circulating Water Piping	$0	$4,808	$4,354	$0	$9,162	$1,374	$0	$2,107	$12,643	$19
9.5	Make-up Water System	$504	$0	$693	$0	$1,197	$180	$0	$275	$1,652	$3
9.6	Component Cooling Water System	$173	$206	$142	$0	$520	$78	$0	$120	$718	$1
9.7	Circulating Water System Foundations	$0	$400	$710	$0	$1,110	$166	$0	$383	$1,659	$3
	Subtotal	**$19,493**	**$5,414**	**$9,967**	**$0**	**$34,874**	**$5,231**	**$0**	**$6,833**	**$46,938**	**$71**
10	Slag Recovery & Handling										
10.1	Slag Dewatering & Cooling	$1,934	$0	$947	$0	$2,881	$432	$0	$497	$3,810	$6
10.2	Gasifier Ash Depressurization	$1,096	$0	$537	$0	$1,632	$245	$0	$282	$2,158	$3
10.3	Cleanup Ash Depressurization	$492	$0	$241	$0	$734	$110	$0	$127	$970	$1
10.6	Ash Storage Silos	$1,104	$0	$1,193	$0	$2,297	$345	$0	$396	$3,038	$5
10.7	Ash Transport & Feed Equipment	$425	$0	$99	$0	$524	$79	$0	$90	$693	$1
10.8	Miscellaneous Ash Handling Equipment	$61	$75	$22	$0	$158	$24	$0	$27	$209	$0
10.9	Ash/Spent Sorbent Foundation	$0	$431	$573	$0	$1,004	$151	$0	$346	$1,501	$2
	Subtotal	**$5,112**	**$506**	**$3,612**	**$0**	**$9,230**	**$1,384**	**$0**	**$1,765**	**$12,380**	**$19**
11	Accessory Electric Plant										
11.1	Generator Equipment	$1,644	$0	$1,241	$0	$2,885	$433	$0	$498	$3,815	$6
11.2	Station Service Equipment	$3,846	$0	$330	$0	$4,176	$626	$0	$720	$5,522	$8
11.3	Switchgear & Motor Control	$23,206	$0	$4,026	$0	$27,232	$4,085	$0	$4,697	$36,014	$55
11.4	Conduit & Cable Tray	$0	$103	$296	$0	$399	$60	$0	$115	$574	$1
11.5	Wire & Cable	$0	$1,408	$2,517	$0	$3,925	$589	$0	$1,128	$5,642	$9
11.6	Protective Equipment	$241	$0	$837	$0	$1,078	$162	$0	$186	$1,426	$2
11.7	Standby Equipment	$340	$0	$314	$0	$653	$98	$0	$113	$864	$1
11.8	Main Power Transformers	$1,650	$0	$34	$0	$1,684	$253	$0	$290	$2,227	$3
11.9	Electrical Foundations	$0	$19	$49	$0	$69	$10	$0	$24	$103	$0
	Subtotal	**$30,926**	**$1,530**	**$9,643**	**$0**	**$42,100**	**$6,315**	**$0**	**$7,771**	**$56,186**	**$85**

	Case:	5	– Coal Gasification w/ CCS					Estimate Type:		Conceptual	
	Plant Size (kg H_2/day):	660,004						Cost Base:		Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/[kg H_2/day]
12						Instrumentation & Control					
12.1	Gasification Control Equipment	$677	$0	$290	$0	$967	$145	$0	$167	$1,279	$2
12.3	Steam Turbine Control Equipment	$625	$0	$85	$0	$710	$107	$0	$123	$939	$1
12.4	Other Major Component Control Equipment	$1,190	$0	$811	$0	$2,001	$300	$100	$360	$2,762	$4
12.5	Signal Processing Equipment	$924	$0	$30	$0	$954	$143	$0	$165	$1,262	$2
12.6	Control Boards, Panels & Racks	$268	$0	$176	$0	$443	$67	$22	$106	$638	$1
12.7	Distributed Control System Equipment	$9,695	$0	$317	$0	$10,011	$1,502	$501	$1,802	$13,816	$21
12.8	Instrument Wiring & Tubing	$482	$386	$1,544	$0	$2,412	$362	$121	$724	$3,618	$5
12.9	Other Instrumentation & Controls Equipment	$1,082	$0	$536	$0	$1,618	$243	$81	$291	$2,233	$3
	Subtotal	**$14,943**	**$386**	**$3,788**	**$0**	**$19,117**	**$2,868**	**$824**	**$3,737**	**$26,547**	**$40**
13						Improvements to Site					
13.1	Site Preparation	$0	$426	$9,702	$0	$10,129	$1,519	$0	$3,494	$15,142	$23
13.2	Site Improvements	$0	$1,927	$2,725	$0	$4,652	$698	$0	$1,605	$6,955	$11
13.3	Site Facilities	$3,009	$0	$3,378	$0	$6,388	$958	$0	$2,204	$9,549	$14
	Subtotal	**$3,009**	**$2,354**	**$15,805**	**$0**	**$21,168**	**$3,175**	**$0**	**$7,303**	**$31,647**	**$48**
14						Buildings & Structures					
14.3	Steam Turbine Building	$0	$2,618	$3,727	$0	$6,346	$952	$0	$1,095	$8,392	$13
14.4	Administration Building	$0	$852	$618	$0	$1,469	$220	$0	$253	$1,943	$3
14.5	Circulation Water Pumphouse	$0	$124	$66	$0	$189	$28	$0	$33	$251	$0
14.6	Water Treatment Buildings	$0	$274	$267	$0	$542	$81	$0	$93	$717	$1
14.7	Machine Shop	$0	$479	$328	$0	$806	$121	$0	$139	$1,066	$2
14.8	Warehouse	$0	$373	$241	$0	$613	$92	$0	$106	$811	$1
14.9	Other Buildings & Structures	$0	$273	$213	$0	$486	$73	$0	$84	$642	$1
14.10	Waste Treating Building & Structures	$0	$739	$1,410	$0	$2,149	$322	$0	$371	$2,842	$4
	Subtotal	**$0**	**$5,731**	**$6,869**	**$0**	**$12,600**	**$1,890**	**$0**	**$2,174**	**$16,664**	**$25**
	Total	**$1,484,168**	**$76,215**	**$643,960**	**$4,919**	**$2,209,263**	**$340,587**	**$200,428**	**$484,715**	**$3,234,992**	**$4,901**

Exhibit 4-49. Case 5 owner's costs

Description	$/1,000	$/[kg H_2/day]
Pre-Production Costs		
6 Months All Labor	$25,888	$39
1 Month Maintenance Materials	$6,571	$10
1 Month Non-Fuel Consumables	$3,229	$5
1 Month Waste Disposal	$701	$1
25% of 1 Months Fuel Cost at 100% CF	$2,216	$3
2% of TPC	$64,700	$98
Total	**$103,305**	**$157**
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	$19,784	$30
0.5% of TPC (spare parts)	$16,175	$25
Total	**$35,959**	**$54**
Other Costs		
Initial Cost for Catalyst and Chemicals	$41,169	$62
Land	$900	$1
Other Owner's Costs	$485,249	$735
Financing Costs	$87,345	$132
Total Overnight Costs (TOC)	**$3,988,919**	**$6,044**
TASC Multiplier (IOU, 35 year)	1.116	
Total As-Spent Cost (TASC)	**$4,452,975**	**$6,747**

Exhibit 4-50. Case 5 initial and annual operating and maintenance costs

Case:	5	– Coal Gasification w/ CCS			Cost Base:	Dec 2018
Plant Size (kg H_2/day):	660,004				Capacity Factor (%):	80
Operating & Maintenance Labor						
Operating Labor				**Operating Labor Requirements per Shift**		
Operating Labor Rate (base):		38.50	$/hour	Skilled Operator:		2.0
Operating Labor Burden:		30.00	% of base	Operator:		11.0
Labor O-H Charge Rate:		25.00	% of labor	Foreman:		1.0
				Lab Techs, etc.:		3.0
				Total:		**17.0**
Fixed Operating Costs						
					Annual Cost	
					($)	**($/[kg H_2/day])**
Annual Operating Labor:					$7,453,446	$11.29
Maintenance Labor:					$33,967,412	$51.47

Case:	5	– Coal Gasification w/ CCS			Cost Base:	Dec 2018
Plant Size (kg H_2/day):	**660,004**				**Capacity Factor (%):**	**80**
Administrative & Support Labor:					$10,355,215	$15.69
Property Taxes and Insurance:					$64.699.832	$98.03
Total:					**$116,475,905**	**$176.48**
Variable Operating Costs						
					($)	**($/kg H_2)**
Maintenance Material:					$63,082,337	$0.33
Consumables						
	Initial Fill	**Per Day**	**Per Unit**	**Initial Fill**		
Water (gal/1000):	0	2,620	$1.90	$0	$1,453,327	$0.00754
Makeup and Wastewater Treatment Chemicals (ton):	0	7.8	$550.00	$0	$1,253,206	$0.00650
Sulfur-Impregnated Activated Carbon (ton):	107	0.146	$12,000.00	$1,282,161	$512,864	$0.00266
Water Gas Shift Catalyst (ft^3):	25,661	17.6	$480.00	$12,317,212	$2,463,442	$0.01278
Selexol Solution (gal):	555,790	55.1	$38.00	$21,120,035	$611,538	$0.00317
Sodium Hydroxide (50 wt%, ton):	0	19.3	$600.00	$0	$3,388,165	$0.01758
Sulfuric Acid (98 wt%, ton):	0	0.342	$210.00	$0	$20,983	$0.00011
Claus Catalyst (ft^3):	w/equip.	1.94	$48.00	$0	$27,237	$0.00014
PSA Unit Adsorbent (ft^3):	43,000	5.9	$150.00	$6,450,000	$258,000	$0.00134
Triethylene Glycol (gal):	w/equip.	680	$6.80	$0	$1,350,417	$0.00701
Electricity (MWh):	0	939	$71.70	$0	$19,661,792	$0.10202
Subtotal:				**$41,169,408**	**$31,000,972**	**$0.16086**
Waste Disposal						
Sulfur-Impregnated Activated Carbon (ton):	0	0.146	$80.00	$0	$3,419	$0.00002
Water Gas Shift Catalyst (ft^3):	0	17.6	$2.50	$0	$12,830	$0.00007
Selexol Solution (gal):	0	55.1	$0.35	$0	$5,633	$0.00003
Claus Catalyst (ft^3):	0	1.94	$2.50	$0	$1,419	$0.00001
Crystallizer Solids (ton):	0	36.4	$38.00	$0	$404,178	$0.00210
Slag (ton):	0	562	$38.00	$0	$6,231,964	$0.03234
PSA Unit Adsorbent (ft^3):	0	5.9	$1.50	$0	$2,580	$0.00001
Triethylene Glycol (gal):	0	680	$0.35	$0	$69,507	$0.00036
Subtotal:				**$0**	**$6,731,529**	**$0.03493**
By-Products						
Sulfur (tons):	0	141	$0.00	$0	$0	$0.00000
Electricity (MWh):	0	0	$71.70	$0	$0	$0.00000
Subtotal:				**$0**	**$0**	**$0.00000**
Variable Operating Costs Total:				**$41,169,408**	**$100,814,837**	**$0.03493**
Fuel Cost						
Illinois Number 6 (ton):	0	5,608	$51.96	$0	$85,078,534	$0.44146
Woody Torrefied Biomass (ton):	0	0	$105.84	$0	$0	$0.00000
Total:				**$0**	**$85,078,534**	**$0.44146**

Exhibit 4-51. Case 5 LCOH breakdown

Component	Value, $/kg H_2	Percentage
Capital	1.35	44%
Fixed	0.60	20%
Variable	0.52	17%
Fuel	0.44	14%
Total (Excluding T&S)	**2.92**	**N/A**
CO_2 T&S	0.17	5%
Total (Including T&S)	**3.09**	**N/A**

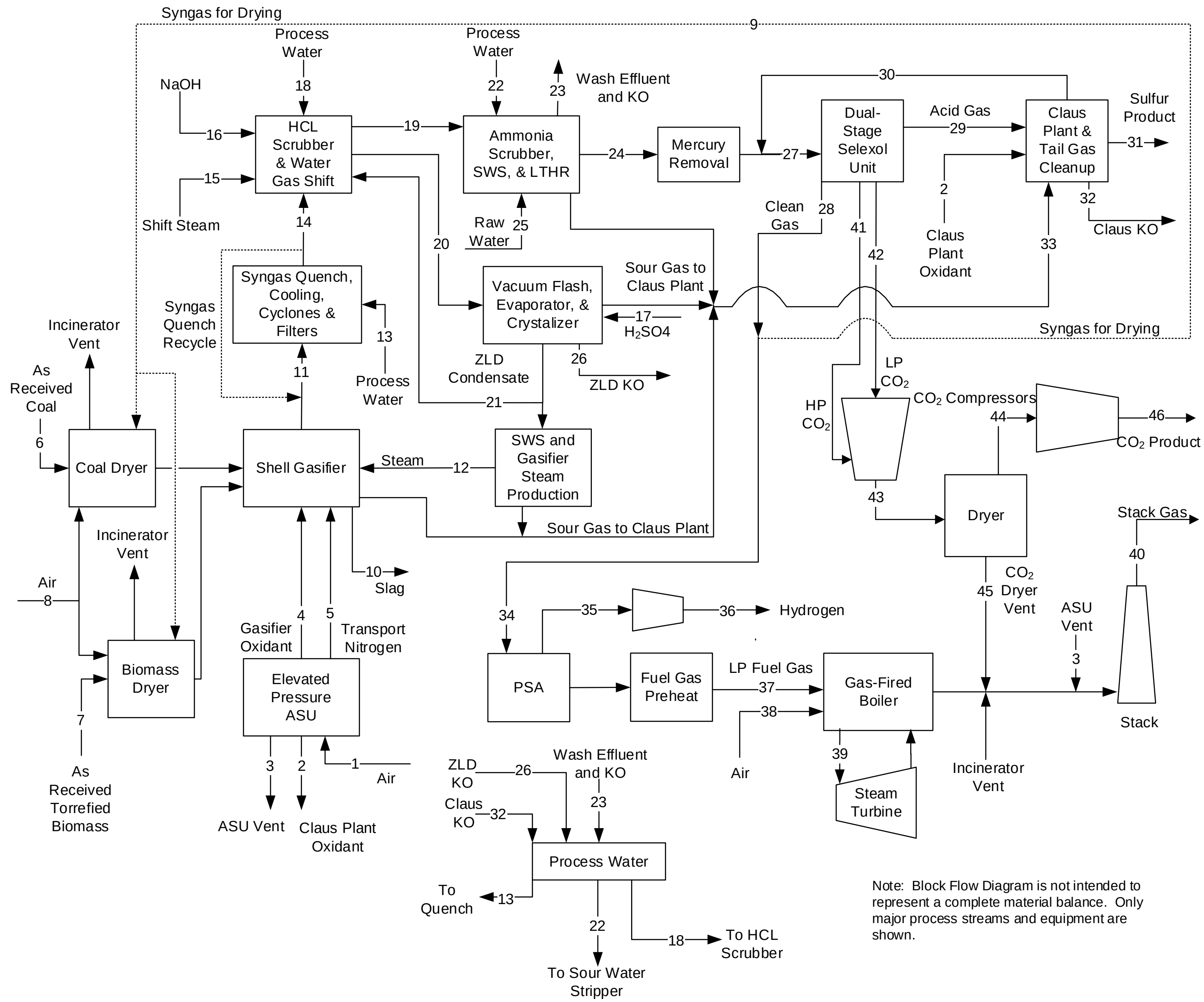
4.13 Case 6 – Shell Co-Gasification Plant with CO_2 Capture Process Description

In this section, the Shell gasification process for Case 6 is described. The plant configuration is nearly identical to that of Case 5, with the exception that this case is configured to co-feed biomass and is at a reduced capacity. Furthermore, some adjustments have been made to the PSA to ensure the plant auxiliary loads can be met entirely by the steam turbine output. The capacity for Case 6 has been scaled to match the Buggenum power plant as-received total gasifier input, as discussed in Section 4, while the percentage of biomass fed to the system was adjusted to achieve net-zero life cycle emissions. As shown in Exhibit 2-7 the alkali content of the torrefied, woody biomass was not modeled.

The process descriptions for Case 6 are similar to Case 5 with several notable exceptions to accommodate the biomass co-feed. The system descriptions follow the BFD provided in

Exhibit 4-52 with the associated stream tables that show process data provided in Exhibit 4-53. Rather than repeating the entire process description, only differences from Case 5 are reported here.

Exhibit 4-52. Case 6 block flow diagram, Shell co-gasification plant with CO_2 capture



Note: Block Flow Diagram is not intended to represent a complete material balance. Only major process streams and equipment are shown.

Exhibit 4-53. Case 6 stream table, Shell co-gasification plant with capture

	1	2	3	4	5	6	7	8	9	10	11	12
V-L Mole Fraction												
Ar	0.0092	0.0343	0.0025	0.0343	0.0000	0.0000	0.0000	0.0092	0.0104	0.0000	0.0082	0.0000
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0000	0.0001	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0168	0.0000	0.4738	0.0000
CO_2	0.0003	0.0000	0.0004	0.0000	0.0000	0.0000	0.0000	0.0003	0.0304	0.0000	0.0444	0.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.8770	0.0000	0.2361	0.0000
H_2O	0.0099	0.0000	0.0091	0.0000	0.0000	0.0000	0.0000	0.0099	0.0001	0.0000	0.1776	0.9998
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0041	0.0000
H_2SO_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.7732	0.0157	0.9844	0.0157	0.9964	0.0000	0.0000	0.7732	0.0652	0.0000	0.0504	0.0000
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0047	0.0002
O_2	0.2074	0.9501	0.0035	0.9501	0.0036	0.0000	0.0000	0.2074	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaOH	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	1.0000	1.0000	0.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	5,890	16	4,367	1,252	236	0	0	210	64	0	6,642	282
V-L Flowrate (kg/hr)	169,965	526	122,159	40,337	6,612	0	0	6,048	374	0	138,770	5,084
Solids Flowrate (kg/hr)	0	0	0	0	0	29,899	23,020	0	0	3,181	0	0
Temperature (°C)	15	27	21	27	130	15	15	15	18	1,427	1,080	343
Pressure (MPa, abs)	0.10	0.86	0.45	5.10	5.62	0.10	0.10	0.10	2.89	4.24	4.24	5.10
Steam Table Enthalpy (kJ/kg)[A]	30.23	21.53	28.12	9.82	129.22	---	---	30.23	90.68	---	2,120.07	3,083.36
AspenPlus Enthalpy (kJ/kg)[B]	-97.58	-0.97	-97.48	-12.68	103.25	-2,119.02	-5,119.15	-97.58	-2,412.27	1,526.12	-3,716.57	-12,884.30
Density (kg/m³)	1.2	11.2	5.2	68.6	46.4	---	---	1.2	6.8	---	7.8	19.9
V-L Molecular Weight	28.857	32.209	27.974	32.209	28.028	---	---	28.857	5.821	---	20.892	18.015
V-L Flowrate (lb_{mol}/hr)	12,985	36	9,627	2,761	520	0	0	462	142	0	14,644	622
V-L Flowrate (lb/hr)	374,708	1,159	269,315	88,927	14,577	0	0	13,334	825	0	305,936	11,207
Solids Flowrate (lb/hr)	0	0	0	0	0	65,917	50,750	0	0	7,012	0	0
Temperature (°F)	59	80	70	80	267	59	59	59	65	2,600	1,975	650
Pressure (psia)	14.7	125.0	65.0	740.0	815.0	14.7	14.7	14.7	418.7	615.0	615.0	740.0
Steam Table Enthalpy (Btu/lb)[A]	13.0	9.3	12.1	4.2	55.6	---	---	13.0	39.0	---	911.5	1,325.6
AspenPlus Enthalpy (Btu/lb)[B]	-42.0	-0.4	-41.9	-5.5	44.4	-911.0	-2,200.8	-42.0	-1,037.1	656.1	-1,597.8	-5,539.3
Density (lb/ft³)	0.076	0.700	0.322	4.283	2.894	---	---	0.076	0.427	---	0.488	1.240

[A]Steam table reference conditions are 32.02 °F & 0.089 psia

[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25 °C and 1 atm

Exhibit 4-53. Case 6 stream table, Shell co-gasification plant with capture (continued)

	13	14	15	16	17	18	19	20	21	22	23	24
V-L Mole Fraction												
Ar	0.0000	0.0059	0.0000	0.0000	0.0000	0.0000	0.0051	0.0000	0.0000	0.0000	0.0000	0.0064
CH_4	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0001
CO	0.0000	0.3391	0.0000	0.0000	0.0000	0.0000	0.0083	0.0001	0.0000	0.0000	0.0000	0.0105
CO_2	0.0010	0.0321	0.0000	0.0000	0.0000	0.0007	0.3134	0.0002	0.0000	0.0007	0.0010	0.3936
COS	0.0000	0.0003	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.1690	0.0000	0.0000	0.0000	0.0000	0.4317	0.0001	0.0000	0.0000	0.0000	0.5449
H_2O	0.9794	0.4055	1.0000	0.6895	0.1000	0.9797	0.1999	0.9932	0.9993	0.9797	0.9786	0.0016
HCl	0.0000	0.0002	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2S	0.0001	0.0029	0.0000	0.0000	0.0000	0.0000	0.0028	0.0000	0.0000	0.0000	0.0001	0.0034
H_2SO_4	0.0000	0.0000	0.0000	0.0000	0.9000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0000	0.0361	0.0000	0.0000	0.0000	0.0000	0.0312	0.0000	0.0000	0.0000	0.0000	0.0394
NH_3	0.0195	0.0089	0.0000	0.0000	0.0000	0.0195	0.0075	0.0036	0.0007	0.0195	0.0203	0.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0026	0.0000	0.0000	0.0000	0.0000
NaOH	0.0000	0.0000	0.0000	0.3105	0.0000	0.0000	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	2,637	7,331	1,575	4	0	0	8,465	490	44	688	3,130	6,705
V-L Flowrate (kg/hr)	47,534	147,178	28,382	103	2	0	167,579	8,882	799	12,388	56,411	135,607
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	188	232	288	16	15	67	203	193	88	67	69	29
Pressure (MPa, abs)	4.98	3.98	3.88	4.81	0.13	0.47	3.49	3.88	0.13	0.47	0.45	3.18
Steam Table Enthalpy (kJ/kg)[A]	767.69	1,252.51	2,971.61	-338.78	-8,206.86	241.60	787.57	806.98	368.22	241.60	246.32	37.86
AspenPlus Enthalpy (kJ/kg)[B]	-14,938.00	-7,102.42	-13,008.68	-13,665.00	-8,526.27	-15,469.43	-8,445.47	-15,036.25	-15,593.65	-15,469.43	-15,449.42	-7,743.98
Density (kg/m³)	835.5	19.5	16.9	1,531.7	1,791.5	964.0	17.6	870.5	965.6	964.0	962.1	26.0
V-L Molecular Weight	18.024	20.077	18.015	24.842	90.073	18.017	19.798	18.122	18.015	18.017	18.025	20.225
V-L Flowrate (lb_{mol}/hr)	5,814	16,161	3,473	9	0	0	18,661	1,081	98	1,516	6,900	14,782
V-L Flowrate (lb/hr)	104,795	324,471	62,572	227	4	0	369,448	19,582	1,761	27,310	124,365	298,962
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	371	450	550	61	59	153	398	379	191	153	156	84
Pressure (psia)	722.7	577.7	562.7	697.7	18.2	67.7	506.1	562.7	19.3	67.7	65.0	460.7
Steam Table Enthalpy (Btu/lb)[A]	330.0	538.5	1,277.6	-145.7	-3,528.3	103.9	338.6	346.9	158.3	103.9	105.9	16.3
AspenPlus Enthalpy (Btu/lb)[B]	-6,422.2	-3,053.5	-5,592.7	-5,874.9	-3,665.6	-6,650.7	-3,630.9	-6,464.4	-6,704.1	-6,650.7	-6,642.1	-3,329.3
Density (lb/ft³)	52.157	1.218	1.055	95.623	111.841	60.178	1.099	54.347	60.283	60.178	60.060	1.625

[A]Steam table reference conditions are 32.02 °F & 0.089 psia

[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25 °C and 1 atm

Exhibit 4-53. Case 6 stream table, Shell co-gasification plant with capture (continued)

	25	26	27	28	29	30	31	32	33	34	35
V-L Mole Fraction											
Ar	0.0000	0.0000	0.0064	0.0104	0.0001	0.0071	0.0000	0.0000	0.0004	0.0104	0.0008
CH_4	0.0000	0.0000	0.0001	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0000
CO	0.0000	0.0000	0.0104	0.0168	0.0005	0.0056	0.0000	0.0000	0.0027	0.0168	0.0000
CO_2	0.0000	0.0004	0.3974	0.0304	0.6390	0.7092	0.0000	0.0000	0.3482	0.0304	0.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0007	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.5406	0.8770	0.0140	0.1820	0.0000	0.0000	0.0280	0.8770	0.9990
H_2O	0.9999	0.9842	0.0016	0.0001	0.0269	0.0024	0.0000	1.0000	0.1558	0.0001	0.0000
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2S	0.0000	0.0002	0.0035	0.0000	0.3177	0.0039	0.0000	0.0000	0.0074	0.0000	0.0000
H_2SO_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0000	0.0000	0.0400	0.0652	0.0004	0.0898	0.0000	0.0000	0.0015	0.0652	0.0002
NH_3	0.0000	0.0152	0.0000	0.0000	0.0008	0.0000	0.0000	0.0000	0.4560	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaCl	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaOH	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	0.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	711	99	6,786	4,160	74	81	0	38	30	4,095	2,697
V-L Flowrate (kg/hr)	12,809	1,789	138,414	24,212	2,913	2,807	0	681	798	23,838	5,531
Solids Flowrate (kg/hr)	0	0	0	0	0	0	749	0	0	0	0
Temperature (°C)	15	30	37	18	27	38	183	49	156	18	18
Pressure (MPa, abs)	0.10	0.24	3.07	2.89	0.18	3.07	0.12	0.11	0.45	2.89	2.82
Steam Table Enthalpy (kJ/kg)[A]	62.75	92.90	48.94	90.68	39.82	18.89	---	107.20	481.26	90.68	263.32
AspenPlus Enthalpy (kJ/kg)[B]	-15,905.25	-15,675.87	-7,739.44	-2,412.27	-6,706.91	-8,085.18	146.79	-15,862.69	-7,262.49	-2,412.27	-86.95
Density (kg/m^3)	999.4	986.4	24.7	6.8	3.0	44.9	5,268.4	967.0	3.4	6.8	2.4
V-L Molecular Weight	18.019	18.013	20.398	5.821	39.543	34.712	---	18.016	26.339	5.821	2.051
V-L Flowrate (lb_{mol}/hr)	1,567	219	14,960	9,171	162	178	0	83	67	9,029	5,945
V-L Flowrate (lb/hr)	28,239	3,943	305,150	53,379	6,423	6,188	0	1,501	1,759	52,554	12,194
Solids Flowrate (lb/hr)	0	0	0	0	0	0	1,652	0	0	0	0
Temperature (°F)	59	85	99	65	80	100	362	120	312	65	65
Pressure (psia)	14.7	35.0	445.2	418.7	26.7	445.2	17.1	16.2	65.0	418.7	408.7
Steam Table Enthalpy (Btu/lb)[A]	27.0	39.9	21.0	39.0	17.1	8.1	---	46.1	206.9	39.0	113.2
AspenPlus Enthalpy (Btu/lb)[B]	-6,838.0	-6,739.4	-3,327.4	-1,037.1	-2,883.5	-3,476.0	63.1	-6,819.7	-3,122.3	-1,037.1	-37.4
Density (lb/ft^3)	62.391	61.576	1.542	0.427	0.186	2.803	328.893	60.369	0.209	0.427	0.147

[A]Steam table reference conditions are 32.02 °F & 0.089 psia

[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25 °C and 1 atm

Exhibit 4-53. Case 6 stream table, Shell co-gasification plant with capture (continued)

	36	37	38	39	40	41	42	43	44	45	46
V-L Mole Fraction											
Ar	0.0008	0.0290	0.0092	0.0000	0.0094	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.0003	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0492	0.0000	0.0000	0.0000	0.0005	0.0001	0.0004	0.0004	0.0000	0.0004
CO_2	0.0000	0.0889	0.0003	0.0000	0.0215	0.9850	0.9985	0.9886	0.9913	0.0500	0.9913
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.9990	0.6419	0.0000	0.0000	0.0000	0.0099	0.0007	0.0074	0.0075	0.0000	0.0075
H_2O	0.0000	0.0004	0.0099	1.0000	0.1244	0.0042	0.0007	0.0032	0.0005	0.9500	0.0005
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2SO_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0002	0.1904	0.7732	0.0000	0.8119	0.0004	0.0000	0.0003	0.0003	0.0000	0.0003
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.2074	0.0000	0.0327	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NaOH	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	2,697	1,399	3,658	4,270	9,310	1,861	691	2,552	2,545	7	2,545
V-L Flowrate (kg/hr)	5,531	18,307	105,547	76,932	254,701	80,898	30,390	111,288	111,145	143	111,145
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0	0	0
Temperature (°C)	30	193	15	532	56	-3	-11	29	29	29	30
Pressure (MPa, abs)	6.48	0.51	0.10	12.51	0.11	0.55	0.12	2.50	2.39	2.50	15.27
Steam Table Enthalpy (kJ/kg)[A]	439.89	442.46	30.23	3,428.46	265.12	-8.36	-9.63	2.14	0.50	138.13	-227.22
AspenPlus Enthalpy (kJ/kg)[B]	89.62	-2,710.88	-97.58	-12,551.84	-1,376.14	-8,974.80	-8,972.87	-8,963.60	-8,957.03	-15,225.03	-9,184.75
Density (kg/m^3)	5.1	1.7	1.2	36.8	1.1	11.2	2.3	49.8	47.2	319.0	839.2
V-L Molecular Weight	2.051	13.087	28.857	18.015	27.356	43.470	43.959	43.602	43.673	19.315	43.673
V-L Flowrate (lb_{mol}/hr)	5,945	3,084	8,064	9,415	20,526	4,103	1,524	5,627	5,611	16	5,611
V-L Flowrate (lb/hr)	12,194	40,360	232,692	169,606	561,519	178,350	66,998	245,347	245,033	315	245,033
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0	0	0
Temperature (°F)	86	380	59	990	132	26	12	85	85	85	86
Pressure (psia)	939.7	74.0	14.7	1,814.7	15.4	80.0	16.7	363.0	346.5	363.0	2,214.7
Steam Table Enthalpy (Btu/lb)[A]	189.1	190.2	13.0	1,474.0	114.0	-3.6	-4.1	0.9	0.2	59.4	-97.7
AspenPlus Enthalpy (Btu/lb)[B]	38.5	-1,165.5	-42.0	-5,396.3	-591.6	-3,858.5	-3,857.6	-3,853.7	-3,850.8	-6,545.6	-3,948.7
Density (lb/ft^3)	0.319	0.107	0.076	2.300	0.066	0.698	0.146	3.110	2.949	19.917	52.393

[A]Steam table reference conditions are 32.02 °F & 0.089 psia

[B]Aspen thermodynamic reference state is the component’s constituent elements in an ideal gas state at 25 °C and 1 atm

4.13.1 Biomass Receiving and Storage

Torrefied, woody biomass has similar physical characteristics as coal. As such, the biomass receiving and storage systems mimic but are separate from the coal preparation and feed systems. The system is comprised of a biomass bottom trestle dumper and receiving hoppers, belt feeder, conveyers, stacker/reclaimer, reclaim hopper, vibratory feeder, surge bin with vent filter, crusher, and silo with vent filter and slide gates. Details on the coal receiving and storage systems can be found in Section 4.4.1.

4.13.2 Biomass Preparation and Feed Systems

The biomass preparation and feed systems mimic but are separate from the coal preparation and feed systems. The system is comprised of a biomass feeder, conveyor, roller mill feed hopper, weigh feeder, dryer and pulverizer, and dryer feed hopper. Although the as-received moisture content of the biomass is near the dried coal moisture content of 5 percent by weight, a biomass dryer is included in the plant configuration to match the dried coal specification. Additionally, although fluctuations in feedstock moisture content are not considered, the dryer is assumed to be necessary to handle expected fluctuations.

4.13.3 Acid Gas Removal

The dual-stage Selexol system is employed for acid gas removal in this case. The process is the same as in Case 5 except the acid gas contains 32 vol% H_2S and 64 vol% CO_2 with the balance primarily water and H_2.

4.13.4 CO_2 Compression and Drying System

No differences from Case 5.

4.13.5 Hydrogen Production

There are no differences from Case 5, except the PSA operates with a hydrogen recovery rate of 75 percent. The recovery efficiency was lowered from the 85 percent assumed for the other cases in this study in order to produce a PSA off-gas having sufficient heating value to generate enough power in the steam turbine to fully offset the auxiliary loads in the plant. As a result, life cycle GHG emissions associated with grid electricity are avoided and the required quantity of biomass feed is minimized.

4.13.6 Steam Generation

The steam conditions for the biomass/coal co-gasification plant with capture (Case 6) are 12.4 MPa/532 °C/532 °C (1,800 psig/990 °F/990 °F).

4.14 CASE 6 – PERFORMANCE RESULTS

Overall performance for the plant is summarized in Exhibit 4-54. The plant produces 5,531 kg/hr (12,194 lb/hr) of hydrogen at an effective thermal efficiency of 57.9 percent (HHV basis). The steam turbine generates 37 MWe and the total auxiliary load for the plant is about 37 MWe, allowing about 1 MW of electricity to be sold back to the grid.

Exhibit 4-55 provides a detailed breakdown of the auxiliary power requirements. The ASU accounts for approximately 51 percent of the auxiliary load between the MAC, booster compressor, N_2 compressors, O_2 pump, and ASU auxiliaries. The two-stage Selexol process and CO_2 compression account for an additional 29 percent of the auxiliary power load. The hydrogen compressor accounts for about 6 percent. The BFW and circulating water system, including the circulating water pumps and cooling tower fan, composes approximately 5 percent of the load, with all other systems together constituting the remaining 9 percent of the auxiliary load.

Exhibit 4-54. Case 6 plant performance summary

Performance Summary	
Steam Turbine Power, MWe	37
Total Gross Power, MWe	**37**
Air Separation Unit Main Air Compressor, kWe	16,320
Air Separation Unit Booster Compressor, kWe	1,280
N_2 Compressors, kWe	610
CO_2 Compression, kWe	7,800
Acid Gas Removal, kWe	2,820
Balance of Plant, kWe	7,770
Total Auxiliaries, MWe	**37**
Net Power, MWe	**1**
Hydrogen Production, kg/hr (lb/hr)	**5,531 (12,194)**
CO_2 Capture, %	**92.7%**
HHV Effective Thermal Efficiency[A], %	57.9%
HHV Cold Gas Efficiency[B], %	57.7%
LHV Effective Thermal Efficiency[A], %	51.2%
LHV Cold Gas Efficiency[B], %	51.1%
Steam Turbine Cycle Efficiency, %	42.9%
Steam Turbine Heat Rate, kJ/kWh (Btu/kWh)	8,393 (7,955)
Condenser Duty, GJ/hr (MMBtu/hr)	211 (200)
AGR Cooling Duty, GJ/hr (MMBtu/hr)	36 (34)
As-Received Coal Feed, kg/hr (lb/hr)	29,899 (65,917)
As-Received Biomass Feed, kg/hr (lb/hr)	23,020 (50,750)
HHV Thermal Input, kWt	370,367
LHV Thermal Input, kWt	354,249
Raw Water Withdrawal, m^3/min (gpm)	3.9 (1,035)
Raw Water Consumption, m^3/min (gpm)	3.1 (814)

[A]ETE = (Hydrogen Heating Value + Net Power) / Fuel Heating Value
[B]CGE = Hydrogen Heating Value / Fuel Heating Value

Exhibit 4-55. Case 6 plant power summary

Power Summary	
Steam Turbine Power, MWe	37
Total Gross Power, MWe	**37**
Auxiliary Load Summary	
Acid Gas Removal, kWe	2,820
Air Blower, kWe	280
Air Separation Unit Auxiliaries, kWe	250
Air Separation Unit Main Air Compressor, kWe	16,320
Air Separation Unit Booster Compressor, kWe	1,280
Ammonia Wash Pumps, kWe	30
Biomass Handling, kWe	90
Biomass Milling, kWe	240
Circulating Water Pumps, kWe	970
Claus Plant TG Recycle Compressor, kWe	280
Claus Plant/TGTU Auxiliaries, kWe	50
CO_2 Compression, kWe	7,800
Coal Handling, kWe	270
Coal Milling, kWe	310
Condensate Pumps, kWe	40
Cooling Tower Fans, kWe	500
Dryer Air Compressor, kWe	20
Feedwater Pumps, kWe	510
Gasifier Water Pump, kWe	10
Ground Water Pumps, kWe	90
Hydrogen Compressor, kWe	2,170
Miscellaneous Balance of Plant[A], kWe	620
N_2 Compressors, kWe	610
O_2 Pump, kWe	80
Quench Water Pump, kWe	90
Shift Steam Pump, kWe	70
Slag Handling, kWe	80
Slag Reclaim Water Recycle Pump, kWe	0
Slurry Water Pump, kWe	0
Sour Gas Compressors, kWe	30
Sour Water Recycle Pumps, kWe	0

Power Summary	
Steam Turbine Auxiliaries, kWe	30
Syngas Recycle Compressor, kWe	220
Syngas Scrubber Pumps, kWe	20
Process Water Treatment Auxiliaries, kWe	180
Transformer Losses, kWe	240
Total Auxiliaries, MWe	**37**
Net Power, MWe	**1**

[A]Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

4.14.1 Environmental Performance

The environmental targets for emissions of Hg, HCl, NOx, SO_2, CO_2, and PM were presented in Section 2.5. A summary of the plant air emissions for Case 6 is presented in Exhibit 4-56.

Exhibit 4-56. Case 6 air emissions

	kg/GJ (lb/MMBtu)	Tonne/year (ton/year)[A]	lb/lb H_2
SO_2	0.000 (0.000)	0 (0)	0.000
NOx	0.004 (0.010)	39 (43)	0.001
Particulate	0.003 (0.007)	29 (31)	0.001
Hg	1.49E-7 (3.48E-7)	0.001 (0.002)	3.60E-8
HCl	0.000 (0.000)	0.00 (0.00)	0.000
CO_2[B]	7 (15)	61,731 (68,047)	1.6
CO_2e[C]	0 (0)	-1,352 (-1,490)	-1.0

[A]Calculations based on an 80 percent capacity factor
[B]Plant stack emissions
[C]Life cycle emissions

SO_2, NOx, PM, and Hg emissions are the same as described in Case 4 and Case 5.

Ninety five percent of the CO_2 from the syngas is captured in the AGR system and compressed for storage. Because not all CO is converted to CO_2 in the shift reactors, the overall carbon removal is 92.7 percent. Additionally, an LCA was conducted to determine the total GWP of the plant.

The carbon balance for the plant is shown in Exhibit 4-57. The carbon input to the plant consists of carbon in the air in addition to carbon in the coal and biomass. Carbon leaves the plant as unburned carbon in the slag, a small portion of the hydrogen product, the captured CO_2 product, and the CO_2 in the stack gas, including the coal dryer vent gas and ASU vent gas. The carbon capture efficiency is defined as one minus the amount of carbon in the stack gas relative to the total carbon in less carbon contained in the slag, represented by the following fraction:

$$Carbon\ Capture\ Efficiency\ (\%) = \left(1 - \left(\frac{Carbon\ in\ Stack}{(Total\ Carbon\ In) - (Carbon\ in\ Slag)}\right)\right) * 100\%$$

Exhibit 4-57. Case 6 carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	19,059 (42,018)	Stack Gas	2,404 (5,300)
Biomass	13,786 (30,392)	CO_2 Product	30,314 (66,832)
Air (CO_2)	38 (84)	Slag	164 (362)
		H_2 Product	1 (1)
Total	**32,883 (72,495)**	**Total**	**32,883 (72,495)**

Exhibit 4-58 shows the sulfur balance for the plant. Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the sulfur recovered in the Claus plant and sulfur in the CO_2 product. Sulfur in the slag is considered negligible.

Exhibit 4-58. Case 6 sulfur balance

Sulfur In		Sulfur Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	749 (1,652)	Stack Gas	–
Biomass	0 (0)	CO_2 Product	0 (0)
		Elemental Sulfur	749 (1,652)
Total	**749 (1,652)**	**Total**	**749 (1,652)**

Exhibit 4-59 shows the overall water balance for the plant. The exhibit is presented in an identical manner as for Case 5.

Exhibit 4-59. Case 6 water balance

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)
Slag Handling	0.07 (18)	0.07 (18)	–	–	–
Slurry Water	–	–	–	–	–
Gasifier Water	–	–	–	–	–
Quench	0.79 (210)	0.79 (210)	–	–	–
HCl Scrubber	0.58 (153)	0.58 (153)	–	–	–
NH_3 Scrubber	0.42 (110)	0.20 (53)	0.21 (56)	–	0.21 (56)
Gasifier Steam	0.08 (22)	0.08 (22)	–	–	–
Condenser Makeup	**0.03 (8)**	**–**	**0.03 (8)**	**–**	**0.03 (8)**
BFW Makeup	0.03 (8)	–	0.03 (8)	–	0.03 (8)
Gasifier Steam	–	–	–	–	–
Shift Steam	–	–	–	–	–

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)
Cooling Tower	**3.71 (980)**	**0.04 (9)**	**3.67 (970)**	**0.83 (220)**	**2.84 (750)**
BFW Blowdown	–	0.03 (8)	-0.03 (-8)	–	-0.03 (-8)
ASU Knockout	–	0.01 (1)	-0.01 (-1)	–	-0.01 (-1)
Total	**5.68 (1,500)**	**1.76 (465)**	**3.92 (1,035)**	**0.83 (220)**	**3.08 (814)**

An overall plant energy balance is provided in tabular form in Exhibit 4-60. The power out is the steam turbine power prior to generator losses. The power at the generator is calculated by multiplying the power out by a combined generator efficiency of 98.5 percent.

Exhibit 4-60. Case 6 overall energy balance (0 °C [32 °F] reference)

	HHV	Sensible + Latent	Power	Total
Heat In, GJ/hr (MMBtu/hr)				
Coal	811 (769)	0.7 (0.6)	–	812 (770)
Biomass	522 (495)	0.5 (0.5)	–	523 (495)
Air	–	8.5 (8.1)	–	8.5 (8.1)
Raw Water Makeup	–	14.7 (14.0)	–	14.7 (14.0)
Auxiliary Power	–	–	131.8 (124.9)	131.8 (124.9)
Total	**1,333 (1,264)**	**24.4 (23.2)**	**131.8 (124.9)**	**1,490 (1,412)**
Heat Out, GJ/hr (MMBtu/hr)				
Misc. Process Steam	–	4.8 (4.6)	–	4.8 (4.6)
Slag	5.4 (5.1)	5.4 (5.1)	–	10.8 (10.2)
Stack Gas	–	68 (64)	–	68 (64)
Sulfur	6.9 (6.6)	0.1 (0.1)	–	7.0 (6.7)
Motor Losses and Design Allowances	–	–	5.5 (5.2)	5.5 (5.2)
Hydrogen Product	770 (730)	2.4 (2.3)	–	772 (732)
Cooling Tower Load[A]	–	484 (459)	–	484 (459)
CO_2 Product Stream	–	-25.3 (-23.9)	–	-25.3 (-23.9)
Blowdown Streams	–	7.3 (6.9)	–	7.3 (6.9)
Ambient Losses[B]	–	21.8 (20.7)	–	21.8 (20.7)
Power	–	–	134 (127)	134 (127)
Total	**782 (741)**	**568 (538)**	**139 (132)**	**1,490 (1,412)**
Unaccounted Energy[C]	–	0.0 (0.0)	–	0.0 (0.0)

[A]Includes condenser, AGR, and miscellaneous cooling loads
[B]Ambient losses include all losses to the environment through radiation, convection, etc. Sources of these losses include the combustor, reheater, superheater, and transformers
[C]By difference

4.14.2 Energy and Mass Balance Diagrams

Energy and mass balance diagrams are shown for the following subsystems in Exhibit 4-61 through Exhibit 4-63:

- ASU, gasification, and gas cooling
- Gas cleanup system
- Hydrogen purification and power block

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Exhibit 4-61. Case 6 coal gasification and ASU energy and mass balance

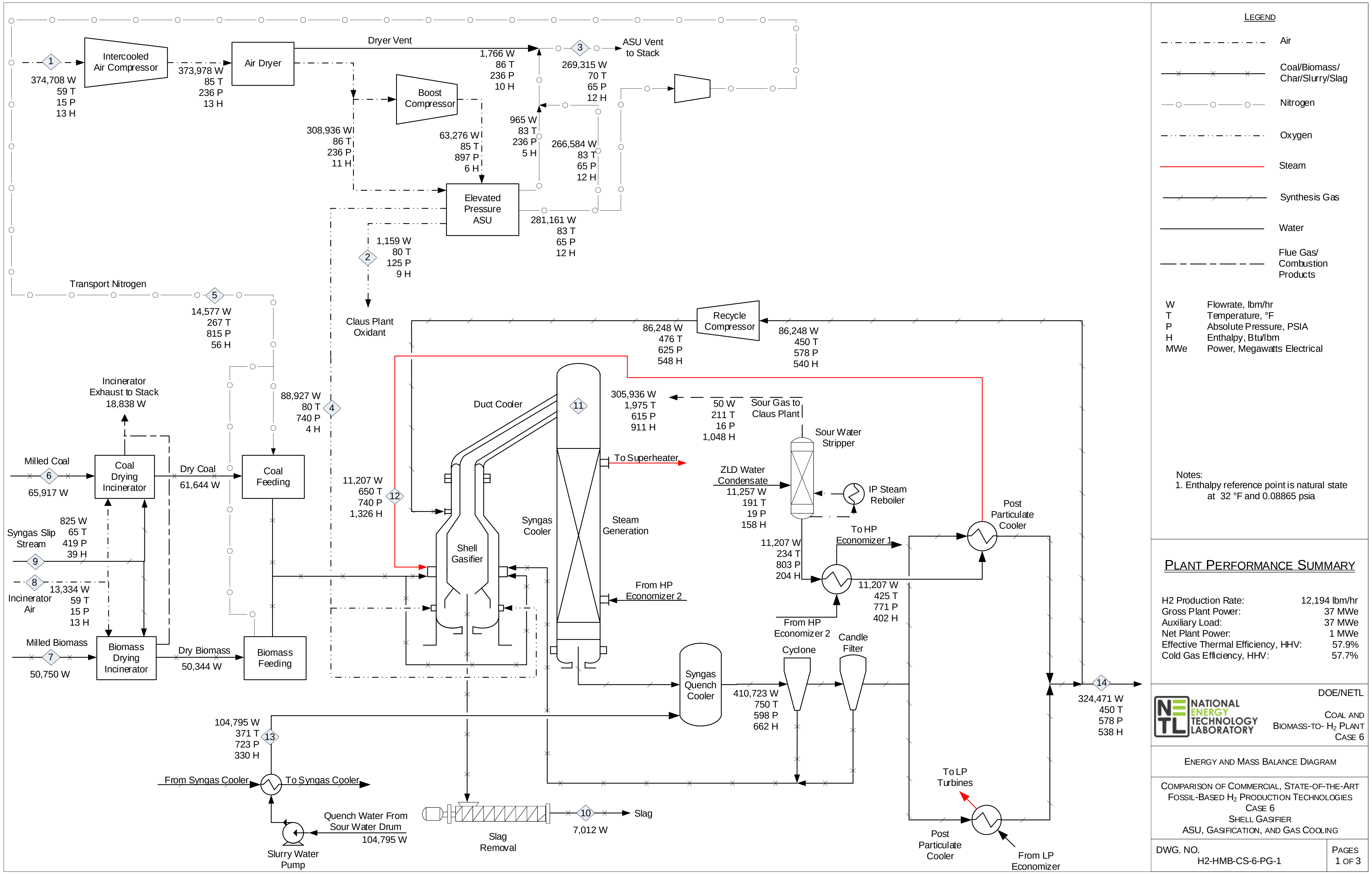


Exhibit 4-62. Case 6 syngas cleanup energy and mass balance

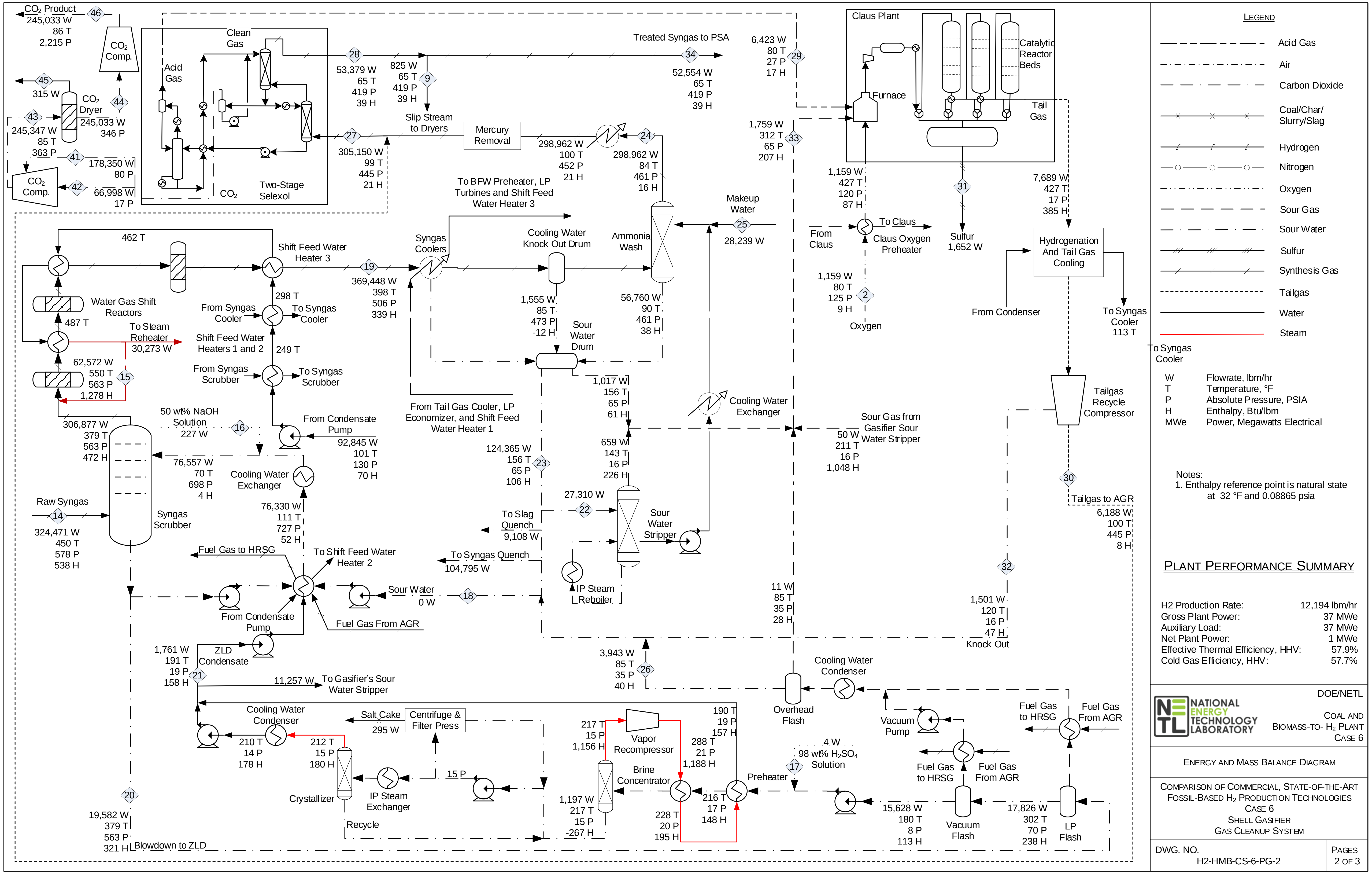
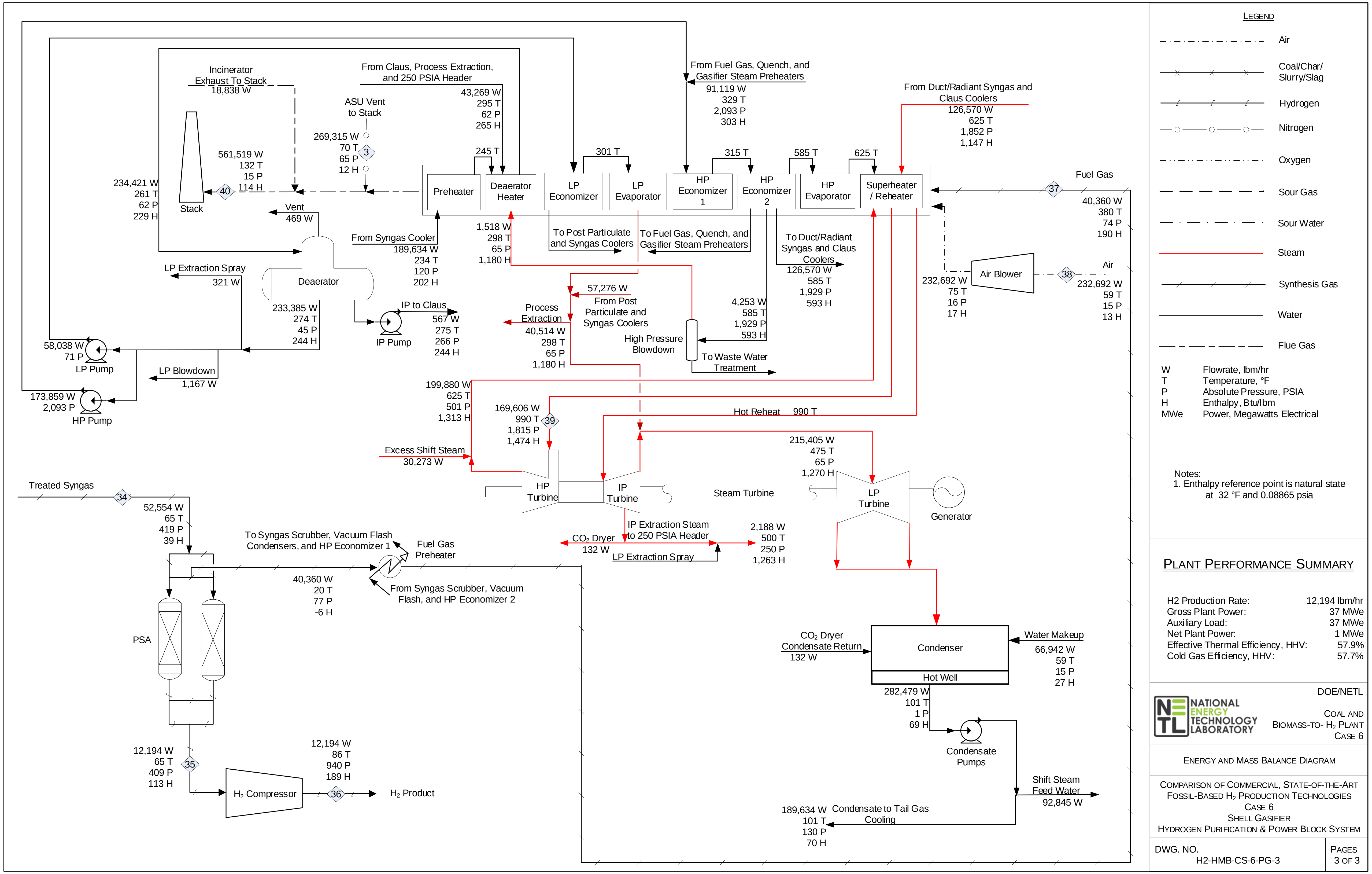


Exhibit 4-63. Case 6 hydrogen purification and power block energy and mass balance



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4.15 Case 6 – Major Equipment List

Major equipment items for the Shell gasifier with CO_2 capture are shown in the following tables. In general, the design conditions include a 10 percent design allowance for flows and heat duties and a 21 percent design allowance for heads on pumps and fans.

Case 6 – Account 1: Coal & Biomass Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Coal Bottom Trestle Dumper and Receiving Hoppers	N/A	181 tonne (200 ton)	2	0
2	Coal Feeder	Belt	570 tonne/hr (630 tph)	2	0
3	Coal Conveyor No. 1	Belt	1,130 tonne/hr (1,250 tph)	1	0
4	Coal Conveyor No. 2	Belt	1,130 tonne/hr (1,250 tph)	1	0
5	Coal Stacker/Reclaimer	Traveling, linear	1,130 tonne/hr (1,250 tph)	1	0
6	Coal Reclaim Hopper	N/A	10 tonne (10 ton)	2	0
7	Coal Feeder	Vibratory	20 tonne/hr (30 tph)	2	0
8	Coal Conveyor No. 3	Belt w/ tripper	50 tonne/hr (50 tph)	1	0
9	Coal Surge Bin w/ Vent Filter	Dual outlet	20 tonne (30 ton)	2	0
10	Coal Crusher	Impactor reduction	8 cm x 0 – 3 cm x 0 (3" x 0 – 1-1/4" x 0)	2	0
11	Coal Conveyor No. 4	Belt w/tripper	50 tonne/hr (50 tph)	1	0
12	Coal Conveyor No. 5	Belt w/ tripper	50 tonne/hr (50 tph)	1	0
13	Coal Silo w/ Vent Filter and Slide Gates	Field erected	110 tonne (120 ton)	3	0
14	Biomass Bottom Trestle Dumper and Receiving Hoppers	N/A	32 tonne (35 ton)	2	0
15	Biomass Feeder	Belt	80 tonne/hr (90 tph)	2	0
16	Biomass Conveyor No. 1	Belt	160 tonne/hr (180 tph)	1	0
17	Biomass Conveyor No. 2	Belt	160 tonne/hr (180 tph)	1	0
18	Biomass Stacker/Reclaimer	Traveling, linear	160 tonne/hr (180 tph)	1	0
19	Biomass Reclaim Hopper	N/A	5 tonne (10 ton)	2	0
20	Biomass Feeder	Vibratory	20 tonne/hr (20 tph)	2	0
21	Biomass Conveyor No. 3	Belt w/ tripper	40 tonne/hr (40 tph)	1	0
22	Biomass Surge Bin w/ Vent Filter	Dual outlet	20 tonne (20 ton)	2	0
23	Biomass Crusher	Impactor reduction	8 cm x 0 – 3 cm x 0 (3" x 0 – 1-1/4" x 0)	2	0
24	Biomass Conveyor No. 4	Belt w/tripper	40 tonne/hr (40 tph)	1	0
25	Biomass Conveyor No. 5	Belt w/ tripper	40 tonne/hr (40 tph)	1	0
26	Biomass Silo w/ Vent Filter and Slide Gates	Field erected	80 tonne (90 ton)	3	0

Case 6 – Account 2: Coal & Biomass Preparation and Feed

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Coal Feeder	Vibratory	10 tonne/hr (10 tph)	3	0
2	Coal Conveyor No. 6	Belt w/tripper	30 tonne/hr (40 tph)	1	0
3	Coal Roller Mill Feed Hopper	Dual Outlet	70 tonne (70 ton)	1	0
4	Coal Weigh Feeder	Belt	20 tonne/hr (20 tph)	2	0
5	Coal Dryer and Pulverizer	Rotary	20 tonne/hr (20 tph)	2	0
6	Coal Dryer Feed Hopper	Vertical Hopper	30 tonne (40 ton)	2	0
7	Biomass Feeder	Vibratory	10 tonne/hr (10 tph)	3	0
8	Biomass Conveyor No. 6	Belt w/tripper	30 tonne/hr (30 tph)	1	0
9	Biomass Roller Mill Feed Hopper	Dual Outlet	50 tonne (60 ton)	1	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
10	Biomass Weigh Feeder	Belt	10 tonne/hr (10 tph)	2	0
11	Biomass Dryer and Pulverizer	Rotary	10 tonne/hr (10 tph)	2	0
12	Biomass Dryer Feed Hopper	Vertical Hopper	30 tonne (30 ton)	2	0

Case 6 – Account 3: Feedwater and Miscellaneous Balance of Plant Systems

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Demineralized Water Storage Tank	Vertical, cylindrical, outdoor	2,005,000 liters (530,000 gal)	2	0
2	Condensate Pumps	Vertical canned	1,180 lpm @ 90 m H_2O (310 gpm @ 300 ft H_2O)	2	1
3	Deaerator (integral w/ GFB)	Horizontal spray type	58,000 kg/hr (129,000 lb/hr)	2	0
4	Intermediate Pressure Feedwater Pump	Horizontal centrifugal, single stage	250 lpm @ 20 m H_2O (70 gpm @ 70 ft H_2O)	2	1
5	High Pressure Feedwater Pump No. 1	Barrel type, multi-stage, centrifugal	HP water: 750 lpm @ 1,700 m H_2O (200 gpm @ 5,700 ft H_2O)	2	1
6	High Pressure Feedwater Pump No. 2	Barrel type, multi-stage, centrifugal	IP water: 390 lpm @ 210 m H_2O (100 gpm @ 670 ft H_2O)	2	1
7	Auxiliary Boiler	Shop fabricated, water tube	18,000 kg/hr, 2.8 MPa, 343 °C (40,000 lb/hr, 400 psig, 650 °F)	1	0
8	Service Air Compressors	Flooded Screw	28 m^3/min @ 0.7 MPa (1,000 scfm @ 100 psig)	2	1
9	Instrument Air Dryers	Duplex, regenerative	28 m^3/min (1,000 scfm)	2	1
10	Closed Cycle Cooling Heat Exchangers	Plate and frame	131 GJ/hr (124 MMBtu/hr) each	2	0
11	Closed Cycle Cooling Water Pumps	Horizontal centrifugal	46,900 lpm @ 20 m H_2O (12,400 gpm @ 70 ft H_2O)	2	1
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	3,785 lpm @ 110 m H_2O (1,000 gpm @ 350 ft H_2O)	1	1
13	Fire Service Booster Pump	Two-stage horizontal centrifugal	2,650 lpm @ 80 m H_2O (700 gpm @ 250 ft H_2O)	1	1
14	Municipal Water Pumps	Stainless steel, single suction	600 lpm @ 20 m H_2O (160 gpm @ 60 ft H_2O)	2	1
15	Ground Water Pumps	Stainless steel, single suction	1,200 lpm @ 270 m H_2O (320 gpm @ 880 ft H_2O)	1	1
16	Filtered Water Pumps	Stainless steel, single suction	400 lpm @ 50 m H_2O (100 gpm @ 160 ft H_2O)	2	1
17	Filtered Water Tank	Vertical, cylindrical	190,000 liter (50,000 gal)	2	0
18	Makeup Water Demineralizer	Anion, cation, and mixed bed	280 lpm (70 gpm)	2	0
19	Liquid Waste Treatment System	N/A	10 years, 24-hour storm	1	0
20	Process Water Treatment	Vacuum flash, brine concentrator, and crystallizer	Vacuum Flash Inlet: 5,000 kg/hr (11,000 lb/hr) Vacuum Flash Outlet: 6,264 ppmw Cl- Brine Concentrator Inlet: 4,000 kg/hr (9,000 lb/hr) Crystallizer Inlet: 300 kg/hr (1,000 lb/hr)	2	0

Case 6 – Account 4: Gasifier, ASU, and Accessories

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Gasifier	Pressurized dry-feed, entrained bed	400 tonne/day, 4.2 MPa (400 tpd, 615 psia)	2	0
2	Synthesis Gas Cooler	Convective spiral-wound tube boiler	76,000 kg/hr (168,000 lb/hr)	2	0
3	Synthesis Gas Cyclone	High efficiency	102,000 kg/hr (226,000 lb/hr) Design efficiency 90%	2	0
4	HCl Scrubber	Ejector Venturi	81,000 kg/hr (178,000 lb/hr)	2	0
5	Ammonia Wash	Counter-flow spray tower	75,000 kg/hr (165,000 lb/hr) @ 3.3 MPa (473 psia)	2	0
6	Primary Sour Water Stripper	Counter-flow with external reboiler	7,000 kg/hr (15,000 lb/hr)	2	0
7	Secondary Sour Water Stripper	Counter-flow with external reboiler	3,000 kg/hr (6,000 lb/hr)	2	0
8	Low Temperature Heat Recovery Coolers	Shell and tube with condensate drain	92,000 kg/hr (203,000 lb/hr)	6	0
9	Low Temperature Heat Recovery Knockout Drum	Vertical with mist eliminator	75,000 kg/hr, 59 °C, 3.3 MPa (166,000 lb/hr, 138 °F, 476 psia)	2	0
10	Synthesis Gas Reheaters	Shell and tube	Reheater 1: 9,000 kg/hr (19,000 lb/hr) Reheater 2: 1,200 kg/hr (3,000 lb/hr) Reheater 3: N/A Reheater 4: N/A Reheater 5: 10,000 kg/hr (22,000 lb/hr) Reheater 6: 10,000 kg/hr (22,000 lb/hr)	2	0
11	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	77,000 kg/hr (169,000 lb/hr) syngas	2	0
12	ASU Main Air Compressor	Centrifugal, multi-stage	1,000 m^3/min @ 1.6 MPa (45,000 scfm @ 236 psia)	1	0
13	Cold Box	Vendor design	500 tonne/day (600 tpd) of 95% purity oxygen	1	0
14	Gasifier Oxygen Pump	Centrifugal, multi-stage	300 m^3/min (10,000 scfm) Suction – 1.0 MPa (130 psia) Discharge – 5.1 MPa (740 psia)	1	0
15	AGR Nitrogen Boost Compressor	Centrifugal, multi-stage	N/A	1	0
16	Gasifier Nitrogen Boost Compressor	Centrifugal, single-stage	50 m^3/min (2,000 scfm) Suction – 0.4 MPa (70 psia) Discharge – 5.6 MPa (820 psia)	1	0

Case 6 – Account 5: Syngas Cleanup

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Mercury Adsorber 1	Sulfated carbon bed	75,000 kg/hr (164,000 lb/hr) 29 °C (84 °F) 3.2 MPa (461 psia)	2	0
2	Mercury Adsorber 2	Sulfated carbon bed	75,000 kg/hr (164,000 lb/hr) 38 °C (100 °F) 3.1 MPa (448 psia)	2	0
3	Sulfur Plant	Claus type	20 tonne/day (22 tpd)	1	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
4	Water Gas Shift Reactors	Fixed bed, catalytic	31,000 kg/hr (68,000 lb/hr) 210 °C (410 °F) 3.9 MPa (560 psia)	6	0
5	Shift Reactor Heat Recovery Exchangers	Shell and Tube	Exchanger 1: 50 GJ/hr (47 MMBtu/hr) Exchanger 2: 21 GJ/hr (20 MMBtu/hr) Exchanger 3: 15 GJ/hr (14 MMBtu/hr) Exchanger 4: 19 GJ/hr (18 MMBtu/hr)	8	0
6	Acid Gas Removal Plant	Two-stage Selexol	152,000 kg/hr (336,000 lb/hr) 37 °C (99 °F) 3.1 MPa (445 psia)	1	0
7	Hydrogenation Reactor	Fixed bed, catalytic	4,000 kg/hr (8,000 lb/hr) 219 °C (427 °F) 0.1 MPa (17 psia)	1	0
8	Tail Gas Recycle Compressor	Centrifugal	3,000 kg/hr (7,000 lb/hr) each	1	0
9	Candle Filter	Pressurized filter with pulse-jet cleaning	metallic filters	2	0
10	CO_2 Dryer	Triethylene glycol	Inlet: 37 m^3/min @ 2.5 MPa (1,315 acfm @ 363 psia) Outlet: 2.4 MPa (346 psia) Water Recovered: 143 kg/hr (315 lb/hr)	1	0
11	CO_2 Compressor	Integrally geared, multi-stage centrifugal	2 m^3/min @ 15.3 MPa (86 acfm @ 2,217 psia)	1	0
12	CO_2 Aftercooler	Shell and tube heat exchanger	Outlet: 15.3 MPa, 30 °C (2,215 psia, 86 °F) Duty: 19 GJ/hr (18 MMBtu/hr)	1	0

Case 6 – Account 6: Hydrogen Production

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Pressure Swing Adsorber	Polybed Proprietary	Syngas 23,838 kg/hr (52,554 lb/hr) 18 °C (65 °F) 2.9 MPa (418.7 psia) Hydrogen 5,531 kg/hr (12,194 lb/hr) 18 °C (65 °F) 2.8 MPa (408.7 psia) Off Gas 18,307 kg/hr (40,360 lb/hr) -7 °C (20 °F) 0.5 MPa (77.0 psia)	1	0
2	Hydrogen Compressor	Integrally geared, multi-stage centrifugal	20 m^3/min @ 6.5 MPa (701 acfm @ 940 psia)	1	1

Case 6 – Account 7: Stack, Ducting, & Off-Gas Fired Boiler

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Stack	CS plate, type 409SS liner	76 m (250 ft) high x 2.0 m (6 ft) diameter	1	0
2	Field-Erected Gas-Fired Boiler	Drum, multi-pressure with economizer section and integral deaerator Air-Fired	Off Gas: 18,307 kg/hr (40,360 lb/hr) 0.5 MPa (74.0 psia) 193 °C (380 °F) Air: 105,547 kg/hr (232,692 lb/hr) 0.1 MPa (16.0 psia) 24 °C (75 °F) 226 GJ/hr (214 MMBtu/hr)	1	0
3	Air Blower	Centrifugal	116,102 kg/hr, 1,584 m^3/min @ 111 cm WG (255,961 lb/hr, 55,950 acfm @ 44 in. WG)	2	1

Case 6 – Account 8: Steam Turbine and Accessories

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Steam Turbine	Commercially available advanced steam turbine	39 MW 12.4 MPa/532 °C/532 °C (1,800 psig/ 990 °F/990 °F)	1	0
2	Steam Turbine Generator	Hydrogen cooled, static excitation	40 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	1	0
3	Surface Condenser	Single pass, divided waterbox including vacuum pumps	230 GJ/hr (220 MMBtu/hr), Inlet water temperature 16 °C (60 °F), Water temperature rise 11 °C (20 °F)	1	0
4	Steam Bypass	One per boiler	50% steam flow @ design steam conditions	1	0

Case 6 – Account 9: Cooling Water System

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Circulating Water Pumps	Vertical, wet pit	96,000 lpm @ 30 m (25,000 gpm @ 100 ft)	2	1
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	11 °C (52 °F) wet bulb/ 16 °C (60 °F) CWT/27 °C (80 °F) HWT/ 530 GJ/hr (500 MMBtu/hr) heat duty	1	0

Case 6 – Account 10: Slag Recovery and Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Slag Quench Tank	Water bath	33,000 liters (9,000 gal)	2	0
2	Slag Crusher	Roll	2 tonne/hr (2 tph)	2	0
3	Slag Depressurizer	Lock Hopper	2 tonne/hr (2 tph)	2	0
4	Slag Receiving Tank	Horizontal, weir	20,000 liters (5,000 gal)	2	0
5	Black Water Overflow Tank	Shop fabricated	9,000 liters (2,000 gal)	2	0
6	Slag Conveyor	Drag chain	2 tonne/hr (2 tph)	2	0
7	Slag Separation Screen	Vibrating	2 tonne/hr (2 tph)	2	0
8	Coarse Slag Conveyor	Belt/bucket	2 tonne/hr (2 tph)	2	0
9	Fine Ash Settling Tank	Vertical, gravity	29,000 liters (8,000 gal)	2	0
10	Fine Ash Recycle Pumps	Horizontal centrifugal	10 lpm @ 14 m H_2O (2 gpm @ 46 ft H_2O)	2	2
11	Grey Water Storage Tank	Field erected	9,000 liters (2,000 gal)	2	0
12	Grey Water Pumps	Centrifugal	30 lpm @ 430 m H_2O (10 gpm @ 1,420 ft H_2O)	2	2
13	Slag Storage Bin	Vertical, field erected	100 tonne (100 tons)	2	0
14	Unloading Equipment	Telescoping chute	10 tonne/hr (20 tph)	1	0

Case 6 – Account 11: Accessory Electric Plant

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	STG Transformer	Oil-filled	24 kV/345 kV, 30 MVA, 3-ph, 60 Hz	1	0
2	High Voltage Auxiliary Transformer	Oil-filled	345 kV/13.8 kV, 14 MVA, 3-ph, 60 Hz	2	0
3	Medium Voltage Transformer	Oil-filled	24 kV/4.16 kV, 12 MVA, 3-ph, 60 Hz	1	1
4	Low Voltage Transformer	Dry ventilated	4.16 kV/480 V, 2 MVA, 3-ph, 60 Hz	1	1

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
5	STG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	1	0
6	Medium Voltage Switchgear	Metal clad	4.16 kV, 3-ph, 60 Hz	1	1
7	Low Voltage Switchgear	Metal enclosed	480 V, 3-ph, 60 Hz	1	1
8	Emergency Diesel Generator	Sized for emergency shutdown	750 kW, 480 V, 3-ph, 60 Hz	1	0

Case 6 – Account 12: Instrumentation and Control

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	DCS – Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W)	Operator stations/printers and engineering stations/printers	1	0
2	DCS – Processor	Microprocessor with redundant input/output	N/A	1	0
3	DCS – Data Highway	Fiber optic	Fully redundant, 25% spare	1	0

4.16 Case 6 – Cost Estimating

The cost estimating methodology was described previously in Section 2.11. Exhibit 4-64 shows a detailed breakdown of the capital costs; Exhibit 4-65 shows the owner's costs, TOC, and TASC; Exhibit 4-66 shows the initial and annual O&M costs; and Exhibit 4-67 shows the LCOH breakdown.

The estimated TOC of the Shell co-gasification plant with CO_2 capture is \$6,515/[kg H_2/day]. Process contingency represents 3.2 percent of the TOC and project contingency represents 11.7 percent. The LCOH is \$3.64/kg H_2 with CO_2 T&S.

Exhibit 4-64. Case 6 total plant cost details

	Case:	6	– Coal + Bio Gasification w/ CCS					Estimate Type:	Conceptual	
	Plant Size (kg H_2/day):	132,746						Cost Base:	Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/[kg H_2/day]
1	Coal Handling										
1.1	Coal Receive & Unload	$288	$0	$139	$0	$427	$64	$0	$98	$589	$4
1.2	Coal Stackout & Reclaim	$942	$0	$225	$0	$1,167	$175	$0	$268	$1,611	$12
1.3	Coal Conveyors & Yard Crush	$8,987	$0	$2,287	$0	$11,275	$1,691	$0	$2,593	$15,559	$117
1.4	Other Coal Handling	$1,400	$0	$315	$0	$1,715	$257	$0	$394	$2,367	$18
1.5	Biomass Receive & Unload	$245	$0	$118	$0	$363	$0	$0	$0	$363	$3
1.6	Biomass Handling	$801	$0	$191	$0	$993	$0	$0	$0	$993	$7
1.7	Biomass Conveyors	$7,642	$0	$1,945	$0	$9,587	$0	$0	$0	$9,587	$72
1.8	Biomass Hnd. Foundations	$1,190	$0	$268	$0	$1,458	$0	$0	$0	$1,458	$11
1.9	Coal & Sorbent Handling Foundations	$0	$25	$66	$0	$91	$14	$0	$21	$126	$1
	Subtotal	**$21,496**	**$25**	**$5,555**	**$0**	**$27,076**	**$2,201**	**$0**	**$3,375**	**$32,653**	**$246**
2	Coal & Biomass Preparation & Feed										
2.1	Coal Crushing & Drying	$645	$39	$93	$0	$777	$116	$0	$179	$1,072	$8
2.2	Prepared Coal Storage & Feed	$1,981	$476	$306	$0	$2,763	$415	$0	$636	$3,814	$29
2.3	Dry Coal Injection System	$2,529	$29	$232	$0	$2,790	$418	$0	$642	$3,850	$29
2.4	Miscellaneous Coal Preparation & Feed	$196	$143	$421	$0	$760	$114	$0	$175	$1,049	$8
2.5	Biomass Crushing & Drying	$543	$33	$78	$0	$654	$0	$0	$0	$654	$5
2.6	Prepared Biomass Storage & Feed	$1,667	$401	$258	$0	$2,325	$0	$0	$0	$2,325	$18
2.7	Dry Biomass Injection System	$2,128	$25	$195	$0	$2,348	$0	$0	$0	$2,348	$18
2.8	Miscellaneous Biomass Preparation & Feed	$165	$120	$354	$0	$639	$0	$0	$0	$639	$5
2.9	Coal, Biomass & Sorbent Feed Foundation	$0	$37	$32	$0	$70	$10	$0	$16	$96	$1
	Subtotal	**$9,854**	**$1,303**	**$1,969**	**$0**	**$13,125**	**$1,074**	**$0**	**$1,647**	**$15,845**	**$119**
3	Feedwater & Miscellaneous BOP Systems										
3.1	Feedwater System	$458	$786	$393	$0	$1,637	$246	$0	$377	$2,259	$17
3.2	Water Makeup & Pretreating	$1,681	$168	$953	$0	$2,802	$420	$0	$967	$4,189	$32
3.3	Other Feedwater Subsystems	$237	$78	$74	$0	$388	$58	$0	$89	$536	$4
3.4	Service Water Systems	$502	$959	$3,105	$0	$4,567	$685	$0	$1,576	$6,827	$51
3.5	Other Boiler Plant Systems	$61	$22	$56	$0	$139	$21	$0	$32	$192	$1
3.6	Natural Gas Pipeline and Start-Up System	$4,498	$193	$145	$0	$4,836	$725	$0	$1,112	$6,674	$50
3.7	Wastewater Treatment Equipment	$2,380	$0	$1,459	$0	$3,839	$576	$0	$1,324	$5,739	$43
3.8	Vacuum Flash, Brine Concentrator, & Crystallizer	$5,659	$0	$3,506	$0	$9,165	$1,375	$0	$3,162	$13,702	$103
3.9	Miscellaneous Plant Equipment	$10,984	$1,441	$5,582	$0	$18,007	$2,701	$0	$6,212	$26,921	$203
	Subtotal	**$26,461**	**$3,647**	**$15,272**	**$0**	**$45,380**	**$6,807**	**$0**	**$14,851**	**$67,039**	**$505**

Case:		6	– Coal + Bio Gasification w/ CCS						Estimate Type:	Conceptual	
Plant Size (kg H_2/day):		132,746							Cost Base:	Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/[kg H_2/day]
4	Gasifier, ASU, & Accessories										
4.1	Gasifier & Auxiliaries (Shell)	$171,150	$489	$809	$631	$173,079	$25,962	$24,231	$33,491	$256,763	$1,934
4.2	Syngas Cooler	$5,438	$390	$403	$314	$6,545	$982	$916	$1,267	$9,710	$73
4.3	Air Separation Unit/Oxidant Compression	$52,364	$108	$13,983	$10,906	$77,362	$11,604	$0	$13,345	$102,311	$771
4.5	Miscellaneous Gasification Equipment	$2,234	$59	$182	$142	$2,617	$393	$0	$451	$3,461	$26
4.6	Low Temperature Heat Recovery & Fuel Gas Saturation	$670	$1,017	$1,476	$1,151	$4,314	$647	$0	$992	$5,953	$45
4.7	Flare Stack System	$202	$48	$128	$99	$477	$72	$0	$110	$658	$5
4.15	Major Component Rigging	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.16	Gasification Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	Subtotal	**$232,057**	**$2,113**	**$16,981**	**$13,244**	**$264,394**	**$39,659**	**$25,147**	**$49,656**	**$378,856**	**$2,854**
5	Syngas Cleanup										
5.1	Double Stage Selexol	$9,669	$81	$469	$365	$10,584	$1,588	$2,117	$2,858	$17,146	$129
5.2	Sulfur Removal	w/5.1	w/5.1	w/5.1	w/5.1	$0	$0	$0	$0	$0	$0
5.3	Elemental Sulfur Plant	$1,007	$18	$111	$87	$1,223	$183	$0	$281	$1,688	$13
5.4	Carbon Dioxide (CO2) Compression & Drying	$4,683	$395	$1,021	$797	$6,897	$1,035	$0	$1,586	$9,517	$72
5.5	Carbon Dioxide (CO2) Compressor Aftercooler	$201	$175	$177	$138	$692	$104	$0	$159	$955	$7
5.6	Mercury Removal (Carbon Bed)	$201	$209	$439	$342	$1,191	$179	$60	$286	$1,716	$13
5.7	Water Gas Shift (WGS) Reactors	$1,357	$1,750	$2,630	$2,051	$7,788	$1,168	$0	$1,791	$10,747	$81
5.8	Hydrogen Chloride Scrubber	$904	$529	$1,027	$801	$3,261	$489	$0	$750	$4,500	$34
5.9	Particulate Removal	$162	$42	$65	$51	$320	$48	$0	$55	$423	$3
5.10	Blowback Gas Systems	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.11	Fuel Gas Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.12	Gas Cleanup Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	Subtotal	**$18,184**	**$3,200**	**$5,940**	**$4,633**	**$31,956**	**$4,793**	**$2,176**	**$7,767**	**$46,693**	**$352**
6	Hydrogen Production										
6.1	Pressure Swing Adsorber	$7,553	$23	$359	$280	$8,215	$1,232	$0	$1,889	$11,336	$85
6.2	Hydrogen Compressor	$3,978	$98	$314	$245	$4,636	$695	$0	$1,066	$6,397	$48
	Subtotal	**$11,531**	**$122**	**$673**	**$525**	**$12,850**	**$1,928**	**$0**	**$2,956**	**$17,734**	**$134**
7	Off Gas Fired Boiler, Ductwork, & Stack										
7.1	Off Gas Fired Boiler	$391	$135	$239	$187	$952	$143	$0	$219	$1,313	$10
7.2	Off Gas Fired Boiler Accessories	$108	$95	$133	$104	$440	$66	$0	$101	$608	$5
7.3	Ductwork	w/7.1	w/7.1	w/7.1	w/7.1	$0	$0	$0	$0	$0	$0
7.4	Stack	$458	$70	$194	$151	$874	$131	$0	$251	$1,256	$9
7.5	Off Gas Fired Boiler, Ductwork & Stack Foundations	w/7.1	w/7.1	w/7.1	w/7.1	$0	$0	$0	$0	$0	$0
	Subtotal	**$957**	**$300**	**$567**	**$442**	**$2,266**	**$340**	**$0**	**$571**	**$3,177**	**$24**

	Case:	6	– Coal + Bio Gasification w/ CCS					Estimate Type:		Conceptual	
	Plant Size (kg H_2/day):	132,746						Cost Base:		Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/[kg H_2/day]
8	Steam Turbine & Accessories										
8.1	Steam Turbine Generator & Accessories	$13,091	$1,159	$1,861	$1,452	$17,563	$2,634	$0	$4,040	$24,237	$183
8.2	Steam Turbine Plant Auxiliaries	w/8.1	w/8.1	w/8.1	w/8.1	$0	$0	$0	$0	$0	$0
8.3	Condenser & Auxiliaries	$510	$333	$336	$262	$1,442	$216	$0	$332	$1,990	$15
8.4	Steam Piping	w/8.1	w/8.1	w/8.1	w/8.1	$0	$0	$0	$0	$0	$0
8.5	Turbine Generator Foundations	w/8.1	w/8.1	w/8.1	w/8.1	$0	$0	$0	$0	$0	$0
	Subtotal	**$13,601**	**$1,492**	**$2,198**	**$1,714**	**$19,005**	**$2,851**	**$0**	**$4,371**	**$26,227**	**$198**
9	Cooling Water System										
9.1	Cooling Towers	$3,730	$0	$1,128	$0	$4,858	$729	$0	$838	$6,425	$48
9.2	Circulating Water Pumps	$485	$0	$36	$0	$521	$78	$0	$90	$689	$5
9.3	Circulating Water System Auxiliaries	$3,629	$0	$516	$0	$4,145	$622	$0	$715	$5,482	$41
9.4	Circulating Water Piping	$0	$2,236	$2,025	$0	$4,261	$639	$0	$980	$5,881	$44
9.5	Make-up Water System	$228	$0	$314	$0	$542	$81	$0	$125	$748	$6
9.6	Component Cooling Water System	$77	$92	$63	$0	$233	$35	$0	$54	$322	$2
9.7	Circulating Water System Foundations	$0	$191	$339	$0	$529	$79	$0	$183	$791	$6
	Subtotal	**$8,150**	**$2,519**	**$4,421**	**$0**	**$15,090**	**$2,263**	**$0**	**$2,984**	**$20,337**	**$153**
10	Slag Recovery & Handling										
10.1	Slag Dewatering & Cooling	$574	$0	$281	$0	$855	$128	$0	$147	$1,131	$9
10.2	Gasifier Ash Depressurization	$325	$0	$159	$0	$484	$73	$0	$84	$640	$5
10.3	Cleanup Ash Depressurization	$146	$0	$72	$0	$218	$33	$0	$38	$288	$2
10.6	Ash Storage Silos	$389	$0	$420	$0	$809	$121	$0	$139	$1,069	$8
10.7	Ash Transport & Feed Equipment	$150	$0	$35	$0	$184	$28	$0	$32	$244	$2
10.8	Miscellaneous Ash Handling Equipment	$21	$26	$8	$0	$56	$8	$0	$10	$73	$1
10.9	Ash/Spent Sorbent Foundation	$0	$152	$202	$0	$353	$53	$0	$122	$528	$4
	Subtotal	**$1,605**	**$178**	**$1,176**	**$0**	**$2,959**	**$444**	**$0**	**$571**	**$3,974**	**$30**
11	Accessory Electric Plant										
11.1	Generator Equipment	$918	$0	$692	$0	$1,610	$241	$0	$278	$2,129	$16
11.2	Station Service Equipment	$2,048	$0	$176	$0	$2,224	$334	$0	$384	$2,941	$22
11.3	Switchgear & Motor Control	$12,359	$0	$2,144	$0	$14,503	$2,175	$0	$2,502	$19,180	$144
11.4	Conduit & Cable Tray	$0	$55	$158	$0	$213	$32	$0	$61	$306	$2
11.5	Wire & Cable	$0	$750	$1,340	$0	$2,090	$314	$0	$601	$3,005	$23
11.6	Protective Equipment	$241	$0	$837	$0	$1,078	$162	$0	$186	$1,426	$11
11.7	Standby Equipment	$202	$0	$187	$0	$389	$58	$0	$67	$514	$4
11.8	Main Power Transformers	$766	$0	$16	$0	$782	$117	$0	$135	$1,034	$8
11.9	Electrical Foundations	$0	$9	$23	$0	$32	$5	$0	$11	$48	$0
	Subtotal	**$16,534**	**$814**	**$5,573**	**$0**	**$22,920**	**$3,438**	**$0**	**$4,224**	**$30,583**	**$230**

Case:	6	– Coal + Bio Gasification w/ CCS				Estimate Type:			Conceptual	
Plant Size (kg H_2/day):	132,746					Cost Base:			Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	Total Plant Cost $/1,000	Total Plant Cost $/[kg H_2/day]
12	Instrumentation & Control										
12.1	Gasification Control Equipment	$564	$0	$242	$0	$806	$121	$0	$139	$1,066	$8
12.3	Steam Turbine Control Equipment	$521	$0	$71	$0	$592	$89	$0	$102	$783	$6
12.4	Other Major Component Control Equipment	$992	$0	$676	$0	$1,668	$250	$83	$300	$2,302	$17
12.5	Signal Processing Equipment	$770	$0	$25	$0	$795	$119	$0	$137	$1,052	$8
12.6	Control Boards, Panels & Racks	$223	$0	$146	$0	$370	$55	$18	$89	$532	$4
12.7	Distributed Control System Equipment	$8,082	$0	$264	$0	$8,346	$1,252	$417	$1,502	$11,517	$87
12.8	Instrument Wiring & Tubing	$402	$322	$1,287	$0	$2,011	$302	$101	$603	$3,016	$23
12.9	Other Instrumentation & Controls Equipment	$902	$0	$447	$0	$1,349	$202	$67	$243	$1,861	$14
	Subtotal	**$12,457**	**$322**	**$3,158**	**$0**	**$15,936**	**$2,390**	**$687**	**$3,115**	**$22,129**	**$167**
13	Improvements to Site										
13.1	Site Preparation	$0	$377	$8,587	$0	$8,965	$1,345	$0	$3,093	$13,402	$101
13.2	Site Improvements	$0	$369	$521	$0	$890	$134	$0	$307	$1,331	$10
13.3	Site Facilities	$125	$0	$140	$0	$264	$40	$0	$91	$395	$3
	Subtotal	**$125**	**$746**	**$9,249**	**$0**	**$10,119**	**$1,518**	**$0**	**$3,491**	**$15,128**	**$114**
14	Buildings & Structures										
14.3	Steam Turbine Building	$0	$2,454	$3,493	$0	$5,947	$892	$0	$1,026	$7,865	$59
14.4	Administration Building	$0	$816	$592	$0	$1,407	$211	$0	$243	$1,861	$14
14.5	Circulation Water Pumphouse	$0	$70	$37	$0	$106	$16	$0	$18	$141	$1
14.6	Water Treatment Buildings	$0	$167	$163	$0	$330	$50	$0	$57	$437	$3
14.7	Machine Shop	$0	$468	$321	$0	$789	$118	$0	$136	$1,043	$8
14.8	Warehouse	$0	$365	$235	$0	$600	$90	$0	$104	$794	$6
14.9	Other Buildings & Structures	$0	$267	$208	$0	$475	$71	$0	$82	$629	$5
14.10	Waste Treating Building & Structures	$0	$659	$1,259	$0	$1,919	$288	$0	$331	$2,538	$19
	Subtotal	**$0**	**$5,266**	**$6,308**	**$0**	**$11,574**	**$1,736**	**$0**	**$1,997**	**$15,307**	**$115**
	Total	**$373,010**	**$22,046**	**$79,038**	**$20,557**	**$494,652**	**$71,443**	**$28,011**	**$101,576**	**$695,682**	**$5,241**

Exhibit 4-65. Case 6 owner's costs

Description	$/1,000	$/[kg H_2/day]
Pre-Production Costs		
6 Months All Labor	$7,580	$57
1 Month Maintenance Materials	$1,413	$11
1 Month Non-Fuel Consumables	$262	$2
1 Month Waste Disposal	$106	$1
25% of 1 Months Fuel Cost at 100% CF	$803	$6
2% of TPC	$13,914	$105
Total	**$24,077**	**$181**
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	$6,859	$52
0.5% of TPC (spare parts)	$3,478	$26
Total	**$10,338**	**$78**
Other Costs		
Initial Cost for Catalyst and Chemicals	$10,726	$81
Land	$900	$7
Other Owner's Costs	$104,352	$786
Financing Costs	$18,783	$141
Total Overnight Costs (TOC)	**$864,858**	**$6,515**
TASC Multiplier (IOU, 35 year)	1.116	
Total As-Spent Cost (TASC)	**$965,472**	**$7,273**

Exhibit 4-66. Case 6 initial and annual operating and maintenance costs

Case:	6	– Coal + Bio Gasification w/ CCS			Cost Base:	Dec 2018
Plant Size (kg H_2/day):	132,746				Capacity Factor (%):	80
Operating & Maintenance Labor						
Operating Labor				**Operating Labor Requirements per Shift**		
Operating Labor Rate (base):		38.50	**$/hour**	Skilled Operator:		2.0
Operating Labor Burden:		30.00	**% of base**	Operator:		6.0
Labor O-H Charge Rate:		25.00	**% of labor**	Foreman:		1.0
				Lab Techs, etc.:		2.0
				Total:		**11.0**
Fixed Operating Costs						
					Annual Cost	
					($)	**($/[kg H_2/day])**
Annual Operating Labor:					$4,822,818	$36.331
Maintenance Labor:					$7,304,660	$55.027
Administrative & Support Labor:					$3,031,869	$22.840
Property Taxes and Insurance:					$13,913,637	$104.814

Case:	6	– Coal + Bio Gasification w/ CCS			Cost Base:	Dec 2018
Plant Size (kg H_2/day):	132,746				Capacity Factor (%):	80
Total:					$29,072,984	$219.012
Variable Operating Costs						
					($)	($/kg H_2)
Maintenance Material:					$13,565,796	$0.35
Consumables						
	Initial Fill	Per Day	Per Unit	Initial Fill		
Water (gal/1000):	0	745	$1.90	$0	$413,249	$0.01066
Makeup and Wastewater Treatment Chemicals (ton):	0	2.2	$550.00	$0	$356,345	$0.00919
Sulfur-Impregnated Activated Carbon (ton):	25	0.034	$12,000.00	$299,608	$119,843	$0.00309
Water Gas Shift Catalyst (ft^3):	6,012	4.1	$480.00	$2,885,997	$577,199	$0.01489
Selexol Solution (gal):	134,490	13.3	$38.00	$5,110,603	$147,979	$0.00382
Sodium Hydroxide (50 wt%, ton):	0	2.7	$600.00	$0	$477,915	$0.01233
Sulfuric Acid (98 wt%, ton):	0	0.048	$210.00	$0	$2,957	$0.00008
Claus Catalyst (ft^3):	w/equip.	0.41	$48.00	$0	$5,777	$0.00015
PSA Unit Adsorbent (ft^3):	16,200	2.2	$150.00	$2,430,000	$97,200	$0.00251
Triethylene Glycol (gal):	w/equip.	159	$6.80	$0	$316,438	$0.00816
Electricity (MWh):	0	0	$71.70	$0	$0	$0.00000
Subtotal:				**$10,726,209**	**$2,514,904**	**$0.06488**
Waste Disposal						
Sulfur-Impregnated Activated Carbon (ton):	0	0.034	$80.00	$0	$799	$0.00002
Water Gas Shift Catalyst (ft^3):	0	4.1	$2.50	$0	$3,006	$0.00008
Selexol Solution (gal):	0	13.3	$0.35	$0	$1,363	$0.00004
Claus Catalyst (ft^3):	0	0.41	$2.50	$0	$301	$0.00001
Crystallizer Solids (ton):	0	5.1	$38.00	$0	$56,926	$0.00147
Slag (ton):	0	84	$38.00	$0	$933,694	$0.02409
PSA Unit Adsorbent (ft^3):	0	2.2	$1.50	$0	$972	$0.00003
Triethylene Glycol (gal):	0	159	$0.35	$0	$16,287	$0.00042
Subtotal:				**$0**	**$1,013,348**	**$0.02614**
By-Products						
Sulfur (tons):	0	20	$0.00	$0	$0	$0.00000
Electricity (MWh):	0	12	$71.70	$0	$251,237	$0.00648
Subtotal:				**$0**	**$251,237**	**$0.00648**
Variable Operating Costs Total:				**$10,726,209**	**$16,842,812**	**$0.43452**
Fuel Cost						
Illinois Number 6 (ton):	0	791	$51.96	$0	$12,000,850	$0.30960
Woody Torrefied Biomass (ton):	0	609	$105.84	$0	$18,821,492	$0.48557
Total:				**$0**	**$30,822,342**	**$0.79517**

Exhibit 4-67. Case 6 LCOH breakdown

Component	Value, $/kg H_2	Percentage
Capital	1.46	40%
Fixed	0.75	21%
Variable	0.43	12%
Fuel	0.80	22%
Total (Excluding T&S)	**3.44**	**N/A**
CO_2 T&S	0.20	6%
Total (Including T&S)	**3.64**	**N/A**

4.17 Gasification Case Summary

The performance and cost results of the three gasification plant configurations modeled in this report are summarized in Exhibit 4-68. The following section provides a discussion on the gasification performance, environmental, and cost results.

Exhibit 4-68. Estimated performance results for all gasification cases

	Gasification		
	Shell		
Case Name	4	5	6
PERFORMANCE			
Nominal CO_2 Capture	0%	92.5%	92.7%
Capacity Factor	80%	80%	80%
Hydrogen Production Rate (lb/hr)	60,622	60,627	12,194
Gross Power Output (MWe)	89	109	37
Auxiliary Power Requirement (MWe)	114	148	37
Net Power Output (MWe)	-25	-39	1
Coal Flowrate (lb/hr)	467,308	467,308	65,917
Biomass Flowrate (lb/hr)	N/A	N/A	50,750
Natural Gas Flowrate (lb/hr)	N/A	N/A	N/A
HHV Thermal Input (kW_t)	1,597,711	1,597,711	370,367
HHV Effective Thermal Efficiency (%)	65.0%	64.1%	57.9%
HHV Cold Gas Efficiency (%)	66.6%	66.6%	57.7%
Raw Water Withdrawal (gpm)	3,188	3,638	1,035
Process Water Discharge (gpm)	672	773	220
Raw Water Consumption (gpm)	2,516	2,866	814
CO_2 Emissions (lb/MMBtu)	199	15	15
SO_2 Emissions[A] (lb/MMBtu)	0.027	0.000	0.000
NOx Emissions (lb/MMBtu)	0.017	0.007	0.010
PM Emissions (lb/MMBtu)	0.007	0.007	0.007
Hg Emissions (lb/TBtu)	0.571	0.571	0.348
COST			
Total Plant Cost (2018$/[kg H_2/day])	4,264	4,901	5,241
Total Overnight Cost (2018$/[kg H_2/day])	5,243	6,044	6,515
Owner's Costs	979	1,142	1,274
Process Contingency	231	304	211
Project Contingency	628	734	765
Home Office Expenses	456	516	538
Bare Erected Cost	2,949	3,347	3,726
Total As-Spent Cost (2018$/[kg H_2/day])	5,853	6,747	7,273
LCOH ($/kg H_2) (excluding T&S)	2.58	2.92	3.44
Capital Costs	1.17	1.35	1.46
Fixed Costs	0.53	0.60	0.75
Variable Costs	0.43	0.52	0.43
Fuel Costs	0.44	0.44	0.80
LCOH ($/kg H_2) (including T&S)	2.58	3.09	3.64
CO_2 T&S Costs	0.00	0.17	0.20

[A]Trace amounts of sulfur emissions may exist in the flue gas stream to the stack in capture cases

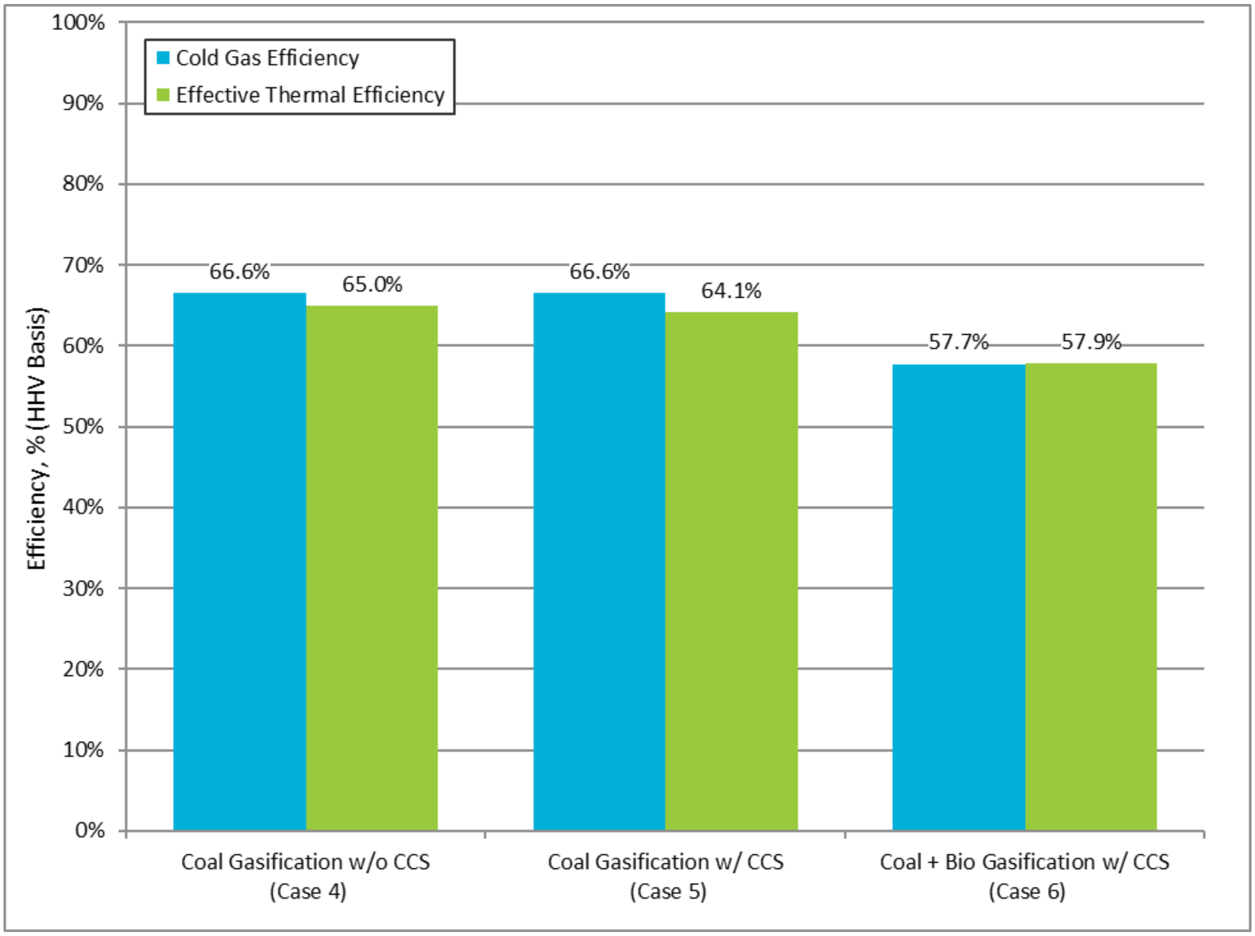
4.17.1 Performance Results Summary

A graph of the net plant efficiency (HHV basis) is provided in Exhibit 4-69.

The following observations can be made regarding plant performance:

- In terms of net hydrogen cold gas efficiency, the gasification plant without capture (Case 4) and with capture (Case 5) performed approximately the same (66.6% HHV-basis) indicating the addition of CO_2 capture does not decrease the yield of H_2 per pound of coal basis. This is a consequence of the two cases having the same configuration and performance for the syngas generation and hydrogen purification systems.
- The gasification plant without capture (Case 4) had the highest effective thermal efficiency (65.0% HHV-basis) compared to the other two cases due to being coal-fed only and having a lower net power demand than Case 5. The biomass co-gasification plant with capture (Case 6) performed the worst overall with respect to efficiency due to the lower heating value of the biomass and reduced H_2 recovery in the PSA.
- Despite performing the same as Case 4 on a cold gas efficiency basis, Case 5 performed slightly worse on an effective thermal efficiency basis. This is due to the energy penalty associated with adding CO_2 capture as additional energy is needed for steam extraction for use in the WGS reaction and the auxiliary load for the CO_2 separation and compression equipment. The reduction in effective thermal efficiency is about 0.9 percentage points (about 1.3 percent relative to non-capture).
- Consequently, the configuration with the highest auxiliary power requirement, normalized on the hydrogen production rate, (0.28 kWe/[kg H_2/day]) MWe) is the biomass/coal co-gasification plant with capture (Case 6) due to the carbon capture and smaller scale energy penalty. The coal gasification plant without capture (Case 4) has the lowest auxiliary power requirement (0.17 kWe/[kg H_2/day]).
- Case 6 is also the only case that has a positive net power output of 1 MWe compared to -25 MWe for Case 4 and -39 MWe for Case 5. This is due to intentional differences in the PSA design for Case 6, which has a lower hydrogen recovery that produces an off-gas that is more energetic and therefore generates more electricity in the steam cycle. The positive net power becomes important for the life cycle equivalent CO_2 emissions as sourcing electricity from the grid is a contributor to total CO_2e emissions.

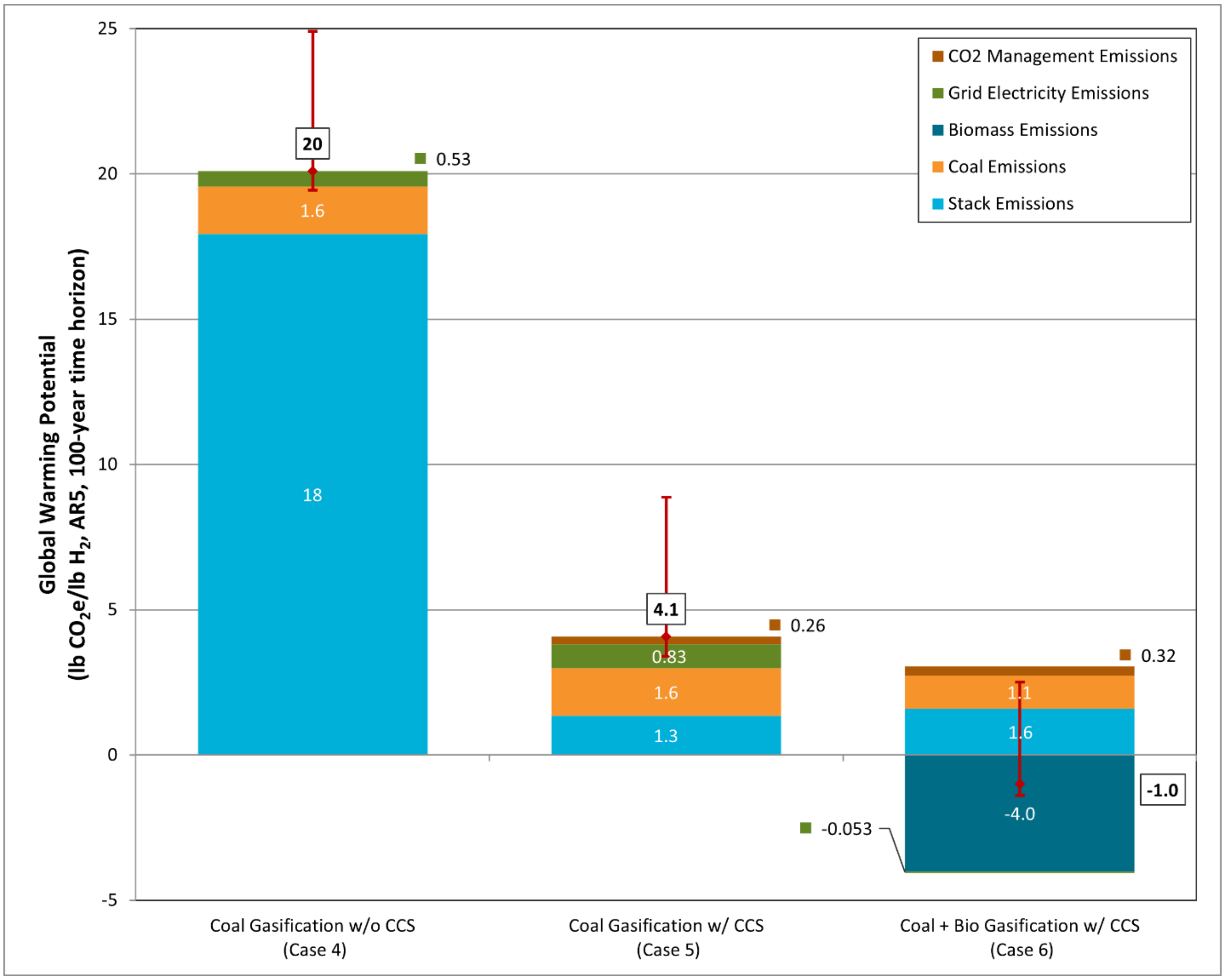
Exhibit 4-69. Net plant efficiency (HHV basis) for all gasification cases



4.17.2 Environmental Results Summary

A graph of the life cycle equivalent CO_2 emissions breakdown is shown in Exhibit 4-70. The results for each case are generated from a Monte Carlo simulation with 5,000 iterations done in openLCA. The bar height is the median value, and the "error bars" represent the 5th and 95th percentile values. The relative variability of results of these cases is generally similar to that of Cases 1-3. Further discussion of sources of variability is provided in Section 5.1.2.

Exhibit 4-70. CO_2e life cycle emissions for all gasification cases



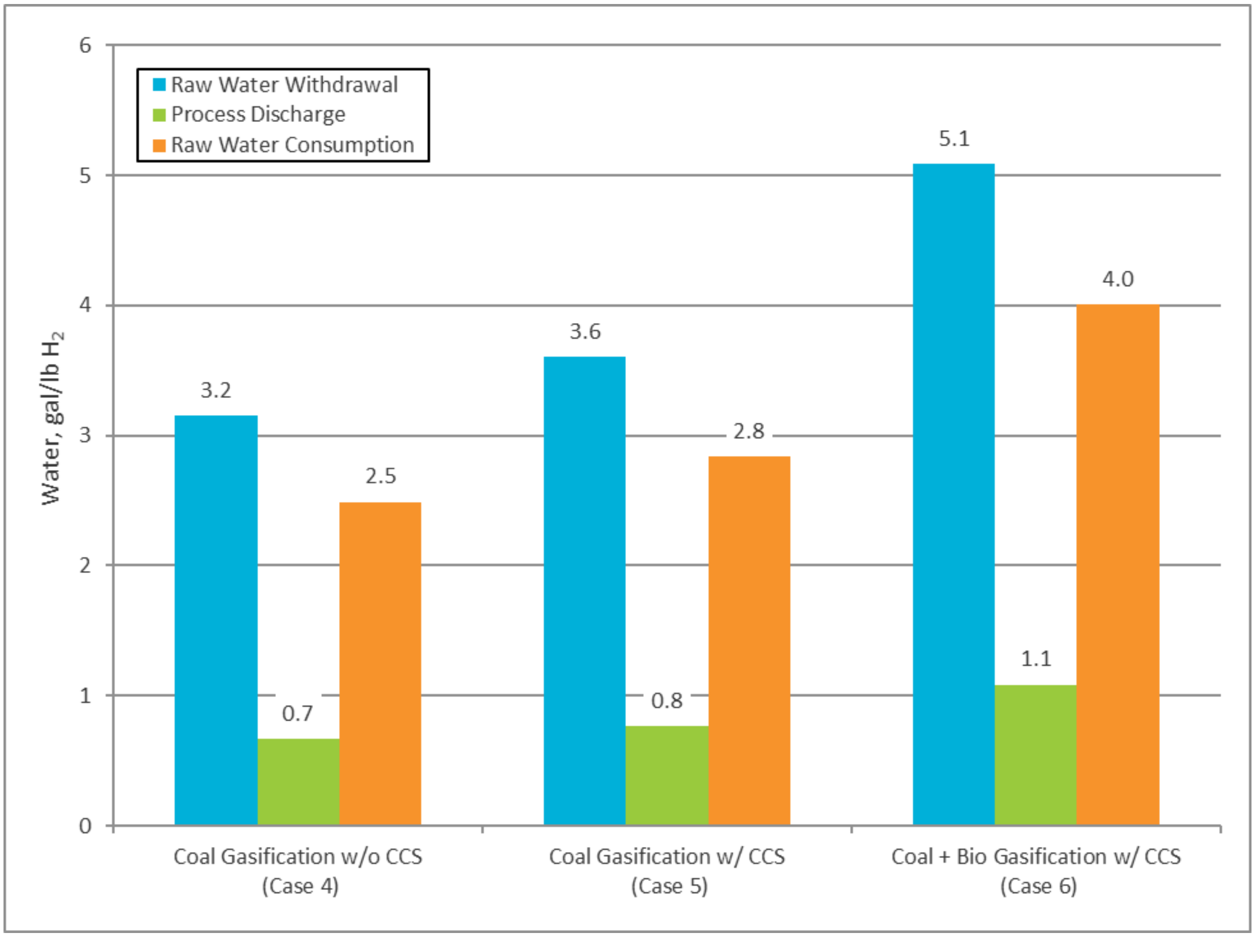
- The highest total equivalent CO_2 life cycle emissions (20 lb CO_2e/lb H_2) comes from the coal gasification plant without capture (Case 4), which is expected. Eighty nine percent of total CO_2e emissions comes from CO_2 released via the plant stack (18 lb CO_2e/lb H_2). 8 percent is from coal (1.6 lb CO_2e/lb H_2) and 2 percent is from grid electricity (0.53 lb CO_2e/lb H_2).
- The coal gasification plant with capture produces over two-thirds of its CO_2e emissions outside the plant boundary, as only 32 percent of the total life cycle emissions are from the plant stack (1.3 lb CO_2e/lb H_2). Forty percent is sourced from the coal life cycle emissions (1.6 lb CO_2e/lb H_2), 20 percent is from grid electricity (0.83 lb CO_2e/lb H_2), and 8 percent is from CO_2 management after it has been captured (0.26 lb CO_2e/lb H_2).
- Although overall CO_2e emissions are down 80 percent when CO_2 capture is employed, the coal gasification plant with capture (Case 5) has a higher life cycle emissions contribution from grid electricity due to the higher power demand of this case and additional life cycle emissions from management of the captured carbon.

- The coal and biomass co-gasification plant with capture has a net life cycle emission value of -1.0 lb CO_2e/lb H_2. The 3.0 lb CO_2e/lb H_2 emitted from the stack, coal life cycle, and CO_2 management is more than counteracted by the 4.0 lb CO_2e/lb H_2 life cycle emissions that are offset by using biomass as a feedstock. If this plant configuration needed to source electricity from the grid, like in the other two cases, a feedstock with more than 50 percent biomass would be needed to account for the life cycle emissions of the grid electricity. Instead, this plant produces excess electricity which results in a credit, assuming that this electricity will displace average electricity production. This explains the motivation for reducing the H_2 recovery in the PSA in order to increase internal power production, as explained previously in Section 4.13.5.

Furthermore, all of the cases satisfy the appropriate hydrogen production facility air emissions targets for SO_2, NOx, particulate matter, and Hg. The coal gasification plant without carbon capture has the highest SO_2 (0.027) and NOx (0.017) emissions, on a lb of pollutant per MMBtu thermal input basis while the gasification plant with capture has the lowest, 0.000 and 0.007 respectively. Each case has approximately the same PM emissions (0.007 lb/MMBtu). The coal-only plants have the highest Hg emissions (0.571 lb/TBtu) on a thermal input basis while the coal/biomass co-gasification plant with capture has the lowest (0.348 lb/TBtu).

Finally, water use for the three gasification cases has been summarized in Exhibit 4-71.

Exhibit 4-71. Water use for all gasification cases

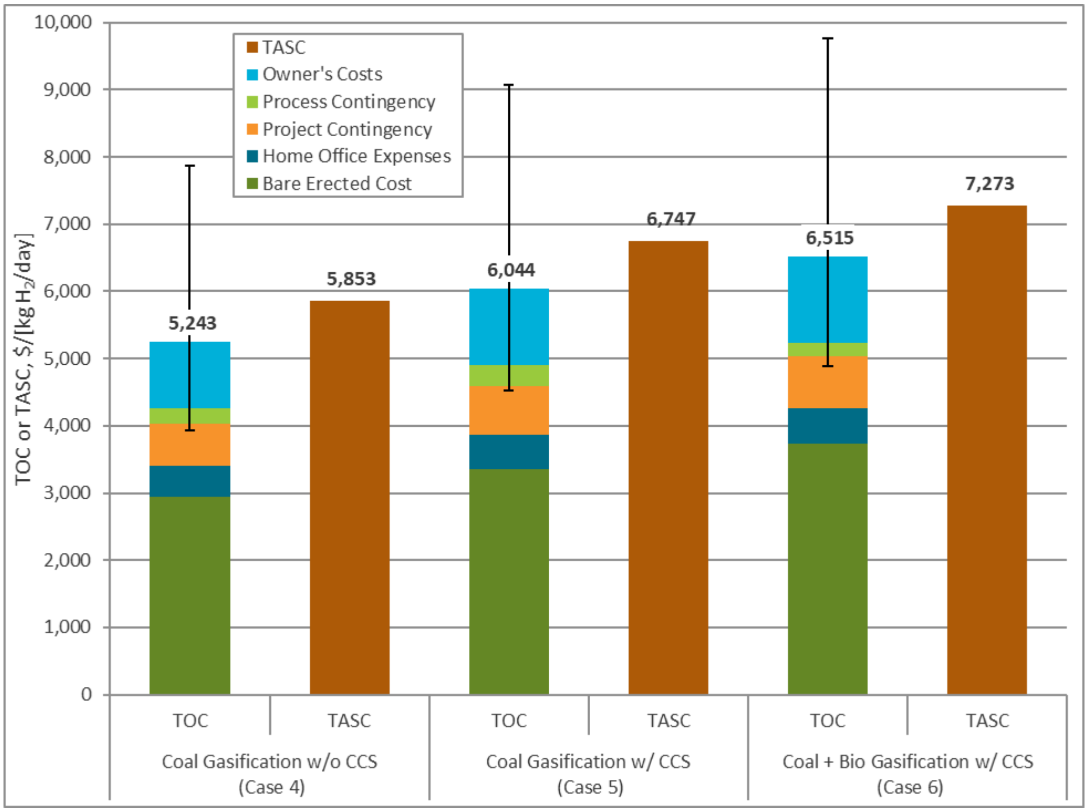


- The results are as expected for the three gasification cases. The coal gasification plant without capture (Case 4) has the lowest water demands of all three cases, with a raw water consumption of 2.5 gal/lb H_2. The coal and biomass co-gasification plant with carbon capture has the highest, with a raw water consumption of 4.0 gal/lb H_2.
- The coal gasification plant with capture (Case 5) has a higher raw water withdrawal rate (3.6 gal/lb H_2) with an approximately equivalent process discharge to the non-capture plant (Case 4). As a result, there is about a 14 percent increase in raw water consumption demand for Case 5 compared to Case 4. This is due to the increased cooling demand required by the CO_2 capture system.
- The coal and biomass gasification plant with capture (Case 6) has a slightly higher process discharge rate (1.1 gal/lb H_2) and approximately a 50 percent higher raw water withdrawal (5.1 gal/lb H_2) than Case 4 and Case 5. This is due to the combined effects of the lower PSA H_2 recovery and additional chemical energy made available in the PSA offgas, which increases evaporative cooling of the steam cycle condenser in Case 6.

4.17.3 Cost Results Summary

The components of TOC and the overall TASC of the three gasification cases are shown in Exhibit 4-72. The error bars included represent the potential TOC range relative to the maximum and minimum of the capital cost uncertainty range, which is -25 percent/+50 percent for gasification cases.

Exhibit 4-72. Plant capital cost for all gasification cases

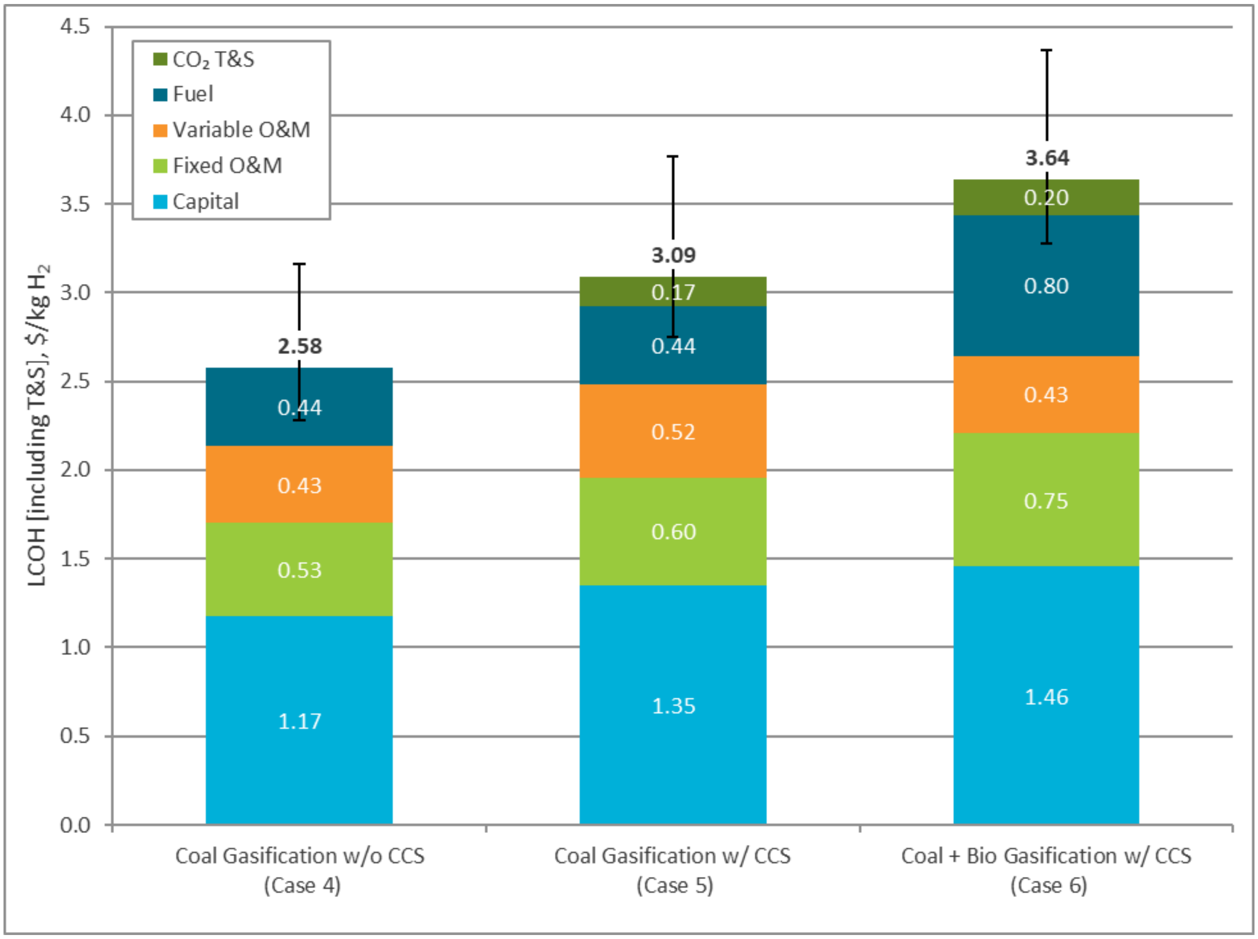


The following TOC observations are made:

- The coal gasification plant without capture (Case 4) has the lowest TOC ($5,243/[kg H_2/day]) and TASC ($5,853/[kg H_2/day]) values of all gasification cases. The coal/biomass co-gasification plant with CCS (Case 6) has the highest TOC ($6,515/[kg H_2/day]) and TASC ($7,273/[kg H_2/day]) values of all gasification cases.
- The primary contributor to the TOC is the bare erected cost, accounting for 55 to 57 percent of the TOC for all cases. The smallest contributor to the TOC is the process contingency, accounting for only three to five percent.

The LCOH is shown for the three gasification cases in Exhibit 4-73. The error bars included represent the potential LCOH range relative to the maximum and minimum capital cost uncertainty ranges. The LCOH ranges presented are not reflective of other changes, such as variation in fuel price, labor price, CF, or other factors.

Exhibit 4-73. LCOH for all gasification cases



The following observations can be made:

- The lowest LCOH is from the coal gasification plant without capture (Case 4, \$2.58/kg H_2) and the highest LCOH is from the coal/biomass co-gasification plant with capture (Case 6, \$3.64/kg H_2). The coal gasification plant with capture (Case 5) has a LCOH in between the other two plants, about \$3.09/kg H_2.
- The largest contributor to the LCOH is the capital cost, making up between about 40 and 46 percent of the LCOH for the gasification cases.
- The capital cost portion of the LCOH is the largest in the coal/biomass co-gasification plant with capture (Case 6), accounting for \$1.46/kg H_2 and the lowest in the coal gasification plant without capture (Case 4, \$1.17/kg H_2).
- The fuel cost portion of the LCOH in the coal/biomass co-gasification plant (Case 6) is about double that of the coal-only plants due to the higher cost of biomass compared to coal. Fuel is \$0.80/kg H_2 for Case 6 and is only \$0.44/kg H_2 for Case 4 and Case 5.
- Variable costs are highest in the coal gasification plant with capture (Case 5, \$0.52/kg H_2) and lowest in the coal gasification plant without capture (Case 4, \$0.43/kg H_2). This is

primarily due to the higher amount of electricity needed to be purchased in Case 5 and lower maintenance material costs in Case 4.

As presented in Section 2.11.4, the breakeven CO_2 sales price and emissions penalty were calculated by using the coal gasification plant with capture (Case 5) and the biomass/coal co-gasification plant with capture (Case 6). The breakeven CO_2 sales price represents the minimum CO_2 plant gate sales price that will incentivize carbon capture. The breakeven CO_2 emissions penalty represents the minimum CO_2 emissions price, when applied to both the capture and non-capture plant that will incentivize carbon capture in lieu of a defined reference non-capture plant. Both the breakeven CO_2 sales price and emissions penalty were calculated using the gasification plant without capture (Case 4) as a non-capture reference.

The breakeven CO_2 sales price (excluding T&S) for the gasification cases with capture are $20.8/tonne ($18.8/ton) for the gasification plant with capture (Case 5) and $42.9/tonne ($38.9/ton) for the co-gasification plant with capture (Case 6). The breakeven CO_2 emissions penalty for the gasification cases with capture are $30.8/tonne ($27.9/ton) for the gasification plant with capture (Case 5) and $65.0/tonne ($59.0/ton) for the co-gasification plant with capture (Case 6). A carbon penalty amount about 111 percent more than the penalty needed for the coal-only case with capture is needed to encourage the use of the biomass/coal co-gasification plant.

5 Results Analysis

Summaries of the individual technologies were provided in Section 3 and Section 4. This section provides the results of all technologies for cross-comparison.

5.1 Performance

Exhibit 5-1 provides a summary of the performance and environmental profile for all cases.

Exhibit 5-1. Estimated performance results for all cases

Case Name	Reforming			Gasification		
	SMR		ATR	Shell Gasifier		
	1	2	3	4	5	6
CO_2 Capture	0%	96.2%	94.5%	0%	92.5%	92.7%
Capacity Factor	90%	90%	90%	80%	80%	80%
Hydrogen Production Rate (lb/hr)	44,369	44,369	60,627	60,622	60,627	12,194
Gross Power Output (MWe)	0	0	0	89	109	37
Auxiliary Power Requirement (MWe)	13	41	110	114	148	37
Net Power Output (MWe)	-13	-41	-110	-25	-39	1
Coal Flowrate (lb/hr)	N/A	N/A	N/A	467,308	467,308	65,917
Biomass Flowrate (lb/hr)	N/A	N/A	N/A	N/A	N/A	50,750
Natural Gas Flowrate (lb/hr)	156,482	166,387	213,694	N/A	N/A	N/A
HHV Thermal Input (kW_t)	1,031,068	1,096,328	1,408,040	1,597,711	1,597,711	370,367
HHV Effective Thermal Efficiency (%)	75.4%	68.4%	67.9%	65.0%	64.1%	57.9%
HHV Cold Gas Efficiency (%)	76.7%	72.1%	75.7%	66.6%	66.6%	57.7%
Raw Water Withdrawal (gpm)	1,457	2,727	3,720	3,188	3,638	1,035
Process Water Discharge (gpm)	63	586	775	672	773	220
Raw Water Consumption (gpm)	1,395	2,140	2,945	2,516	2,866	814
CO_2 Emissions (lb/MMBtu)	118	5	7	199	15	15
SO_2 Emissions[A] (lb/MMBtu)	0.000	0.000	0.000	0.027	0.000	0.000
NOx Emissions (lb/MMBtu)	0.003	0.002	0.000	0.017	0.007	0.010
PM Emissions (lb/MMBtu)	0.000	0.000	0.000	0.007	0.007	0.007
Hg Emissions (lb/TBtu)	0.000	0.000	0.000	0.571	0.571	0.348

[A]Trace amounts of sulfur emissions may exist in the flue gas stream to the stack in capture cases

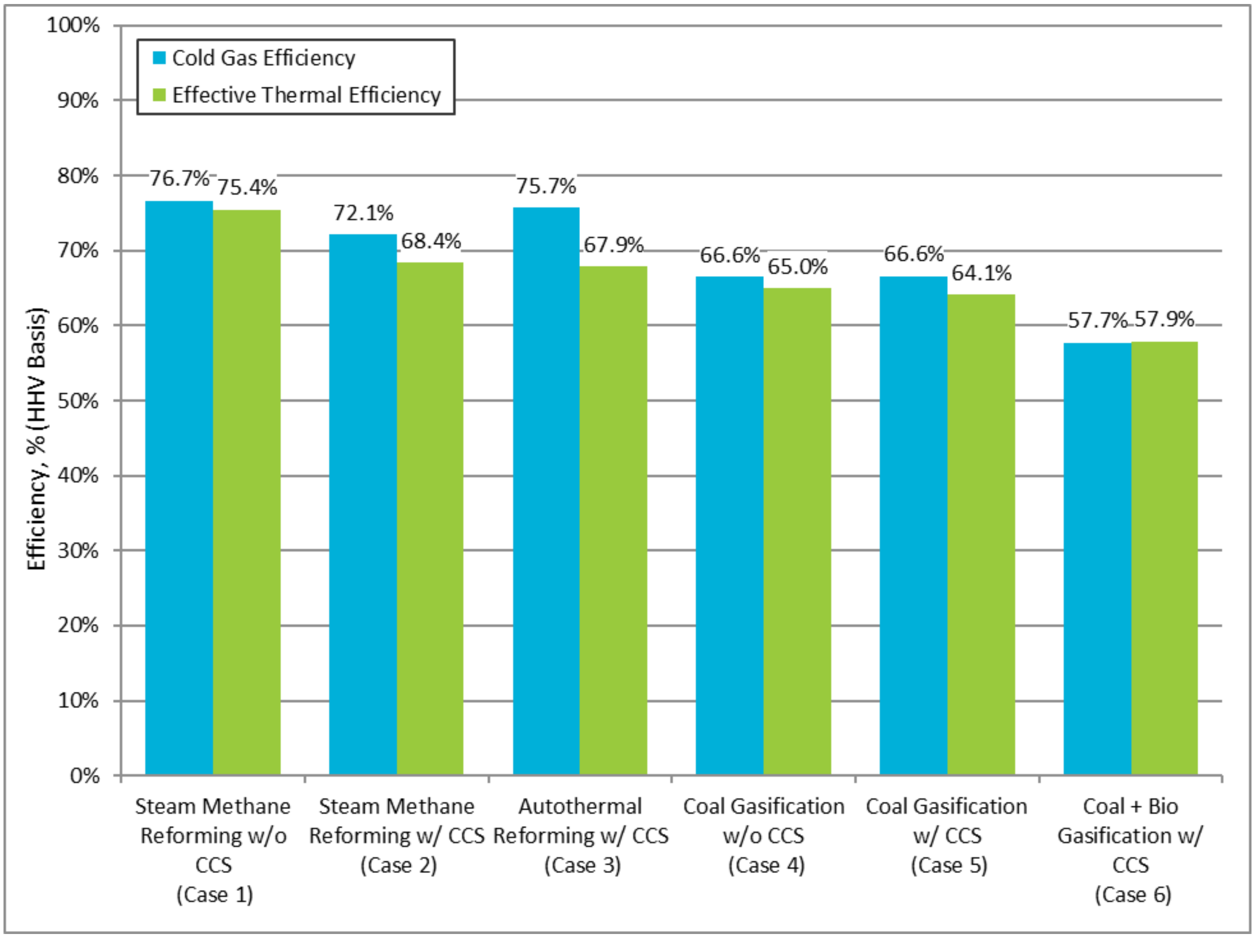
5.1.1 Energy Efficiency

A graph of the net plant efficiency (HHV basis) is provided in Exhibit 5-2. The primary conclusions that can be drawn are:

- In general, the reforming cases have higher efficiencies than the gasification cases, on average about 10 percentage points higher. The highest effective thermal efficiency (75.4%) is from the SMR plant without capture (Case 1) and the lowest effective thermal efficiency (57.9%) is from the coal/biomass co-gasification case with capture (Case 6).

- Adding capture to the SMR plant decreases the ETE by about seven percentage points due to a greater percentage of the NG feedstock being combusted to satisfy the thermal demands of the CO_2 removal processes; while adding capture to the coal gasification plant decreases the CGE by less than one percentage point since additional feedstock is not required to support CO_2 removal.
- The coal/biomass co-gasification plant with CCS (Case 6) is the only case in which the ETE is higher than the cold gas efficiency, due to the positive net power production of the system. This reflects the design assumption made for this case to eliminate power imports so life cycle GHG emissions associated with power imports are eliminated.

Exhibit 5-2. Plant efficiencies for all cases (HHV basis)



5.1.2 Environmental Emissions

Estimated emissions of Hg, HCl, PM, NOx, and SO_2 are all at or below the applicable regulatory limits for all cases. The SO_2, NOx, and PM emissions for each case are shown in Exhibit 5-3, reported in units of pounds of pollutant emitted per 1,000 pounds of H_2 produced. The Hg emissions for each case are shown in Exhibit 5-4.

Natural gas does not contain Hg, PM, or HCl, which makes its environmental profile more attractive compared to the gasification cases. In this report, it was assumed that the only sulfur present in natural gas is from the addition of the odorant, mercaptan. The addition of a sulfur polishing system for the purpose of preventing catalyst poisoning results in an SO_2 emission rate well below the regulatory limits.

The following systems for the gasification plants are required to bring the cases considered into compliance with the ELG rule: vacuum flash, brine concentrator, and crystallizer. These are applied to the syngas scrubber blowdown to ensure that ZLD is satisfied.

Exhibit 5-3. SO_2, NOx, and PM emissions for all cases

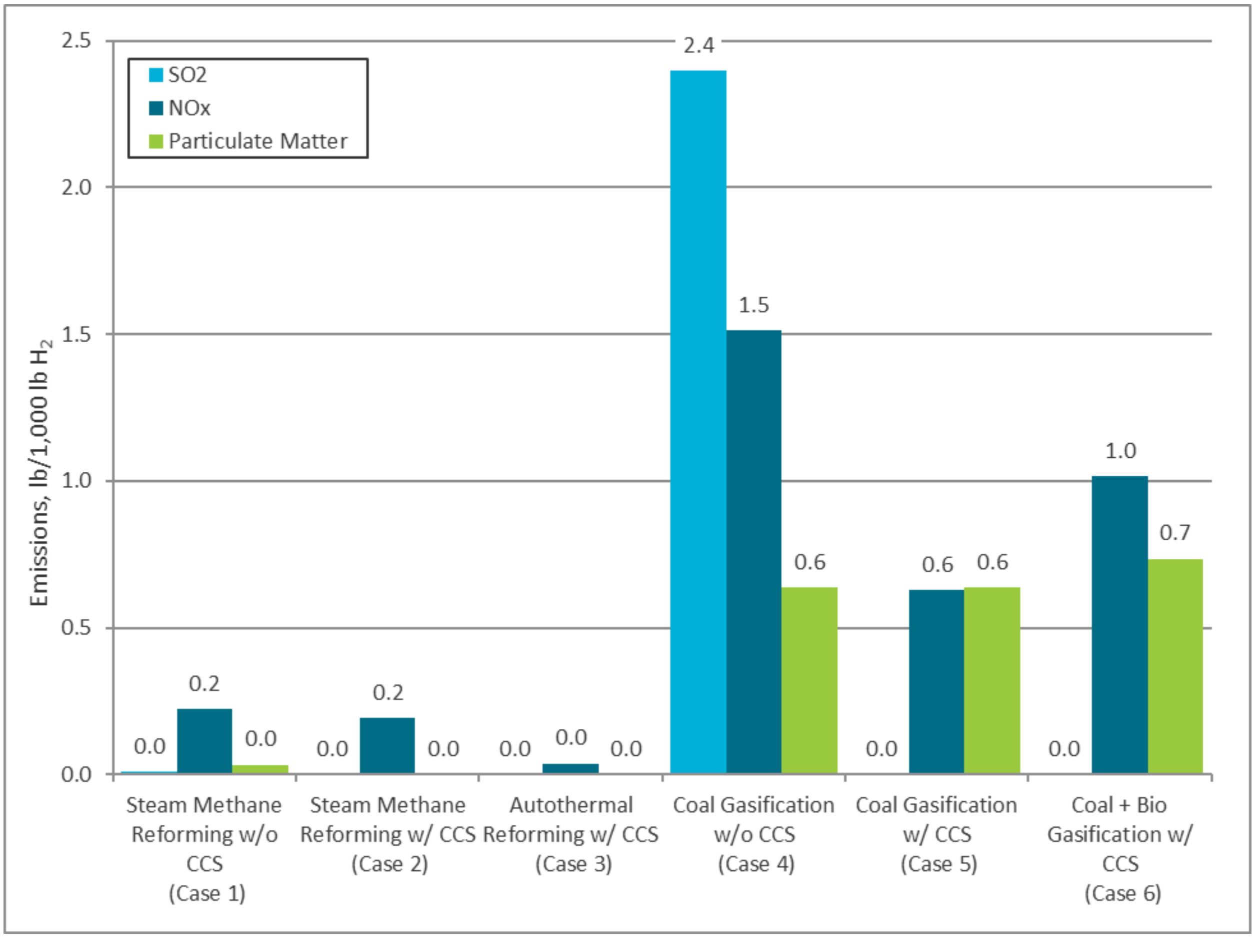
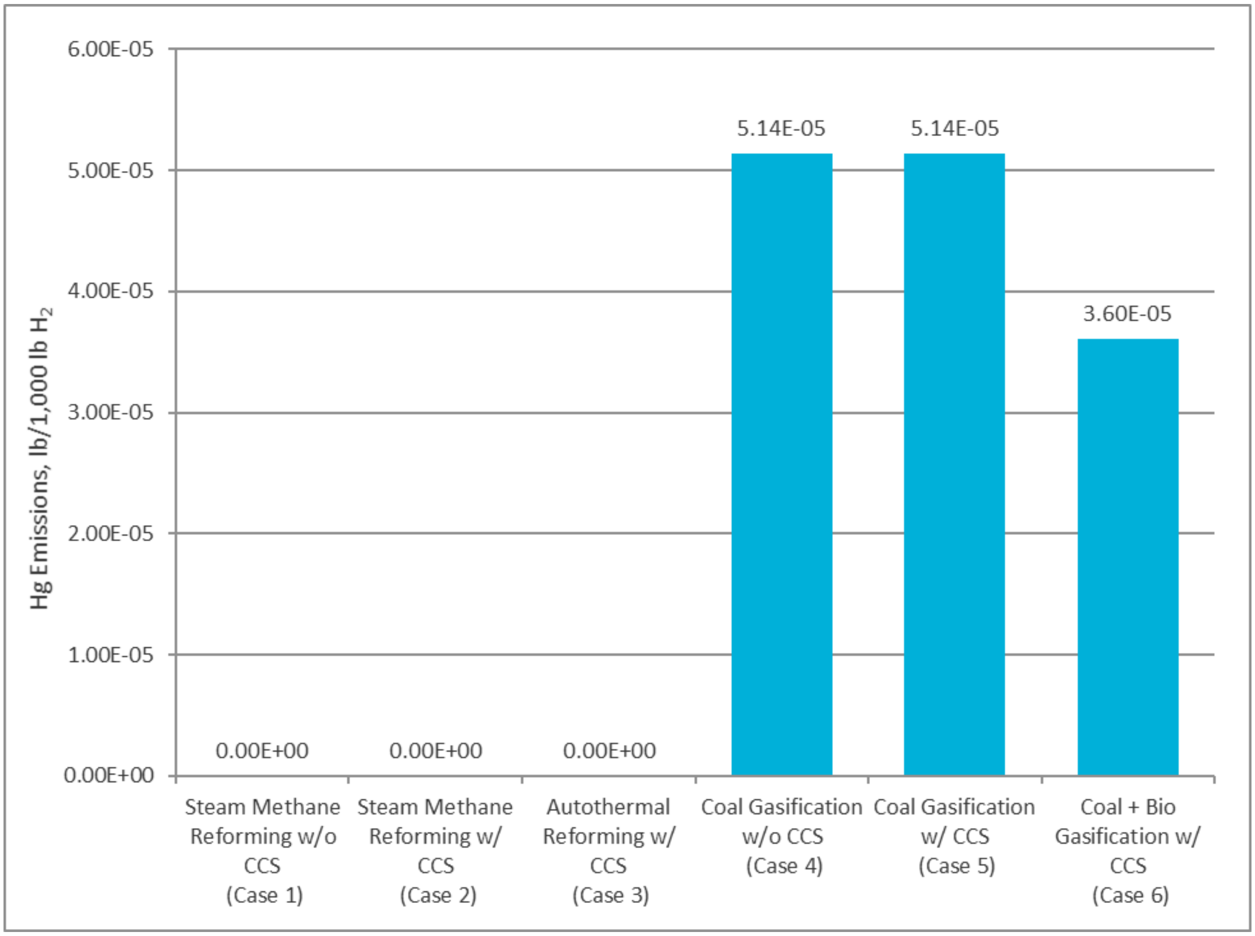


Exhibit 5-4. Hg emissions for all cases



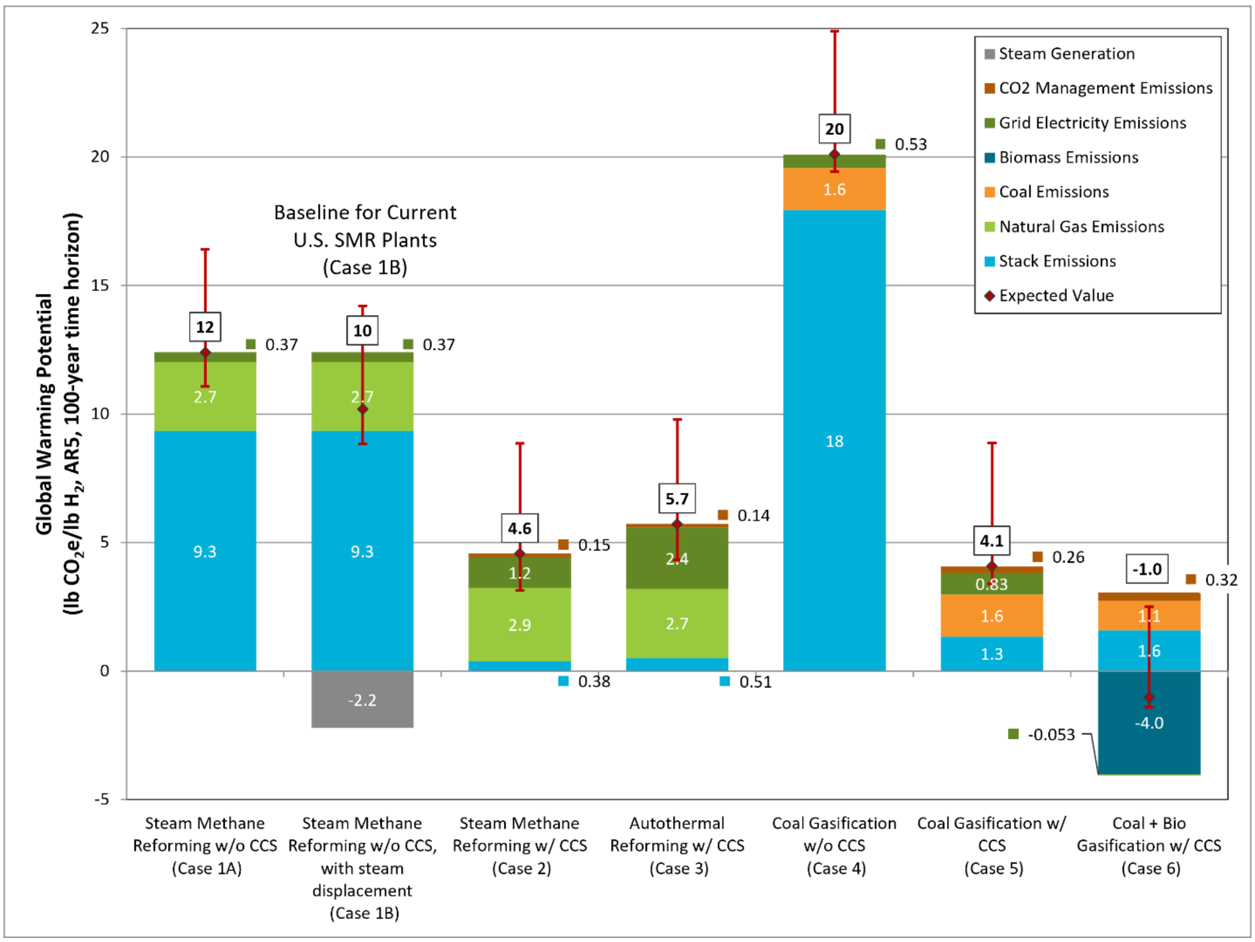
The following observations can be made about the SO_2, NOx, PM, and Hg emissions:

- The coal gasification plant without CCS (Case 4) releases the most SO_2 pollutant (2.4 lb SO_2/1,000 lb H_2) on a normalized basis. All other cases release a negligible amount of SO_2. This is due to the minimal amount of sulfur present in the natural gas feedstock and the extra AGR systems present in the gasification cases with CO_2 capture.
- The largest emitter of NOx on a normalized basis is the coal gasification plant without capture (Case 4), emitting 1.5 lb NOx/1,000 lb H_2 and the smallest emitter is the ATR plant with capture (Case 3). The emission of NOx is kept to a minimum by the use of low-NOx burners in the reforming and gasification plants.
- Each of the gasification cases emits approximately the same normalized amount of PM, about 0.6 – 0.7 lb PM/1,000 lb H_2. The PM emitted is limited by the use of the direct contact water quench, cyclone collector, candle filter, and syngas scrubber. The reforming cases emit a negligible quantity of PM.
- The largest normalized mercury emissions are released in the coal-only gasification cases (Case 4 and Case 5, 5.14×10^{-5} lb Hg/1,000 lb H_2), which is about 43 percent more than the emissions of the coal/biomass co-gasification plant with CCS (Case 6). This is due to

the large proportion of woody biomass in Case 6 which does not contain mercury. None of the reforming cases emit any mercury.

Of particular interest in this study is the global warming potential of each plant, throughout the life cycle of the system. This analysis has been discussed previously in Section 2.8. To summarize, the global warming potential, on a CO_2e unit-basis, is estimated over the life cycle of the plant by considering the GHG emissions from the following sources: grid electricity usage, coal, natural gas, and biomass feedstock curation, CO_2 management, and CO_2 stack emissions. Exhibit 5-5 shows the life cycle global warming potential, in units of lb CO_2e per lb H_2, of each of the cases.

Exhibit 5-5. CO_2e life cycle emissions for all cases



The "error bars" represent results of the 5th and 95th percentile after running a Monte Carlo simulation with 5,000 iterations for each technology. Heights of bars represent median values. Simulations were performed in openLCA, and the results distribution relies on underlying parameterized data in upstream models. The upstream natural gas model has many stochastically defined parameters, such as the triangular distribution for the number of natural gas transmission stations [9]. For upstream natural gas, regional variability is also accounted for in the error bars. The median value represents the U.S. average, while the low case assumes

Alaskan Offshore gas and the high case assumes San Juan Conventional gas. The addition of regional variability to the natural gas error was included to represent the variability in GWP values in the domestic natural gas sector, which is obscured when applying a Monte Carlo simulation to the U.S. average value. The upstream electricity model includes a lognormal uncertainty distribution for select emission flows, such as ammonia and carbon dioxide [11, 103]. This derives from variation in emissions factors measured and reported in facility data. The upstream biomass model includes triangular distributions for key parameters, like biomass yield and fertilization rates [13], and uncertainty in the upstream coal model is driven by the coal mine methane parameter [14]. Both the coal and the biomass cases exhibit wider emissions bands because of known variability in the sourcing and physical properties. For example, coal is modeled using a median value for coal mine methane emissions for all the underground bituminous coal mines in the Illinois basin. There is a wide variability in coal mine methane emissions within that region, as shown by the uncertainty bars. This is appropriate because it is unknown which coal producer might be providing coal to these notional gasification facilities. In contrast, natural gas and electricity exhibit much tighter uncertainty because they are commoditized products with gas or electricity coming from many producers in known amounts.

The following observations can be made about the life cycle emissions:

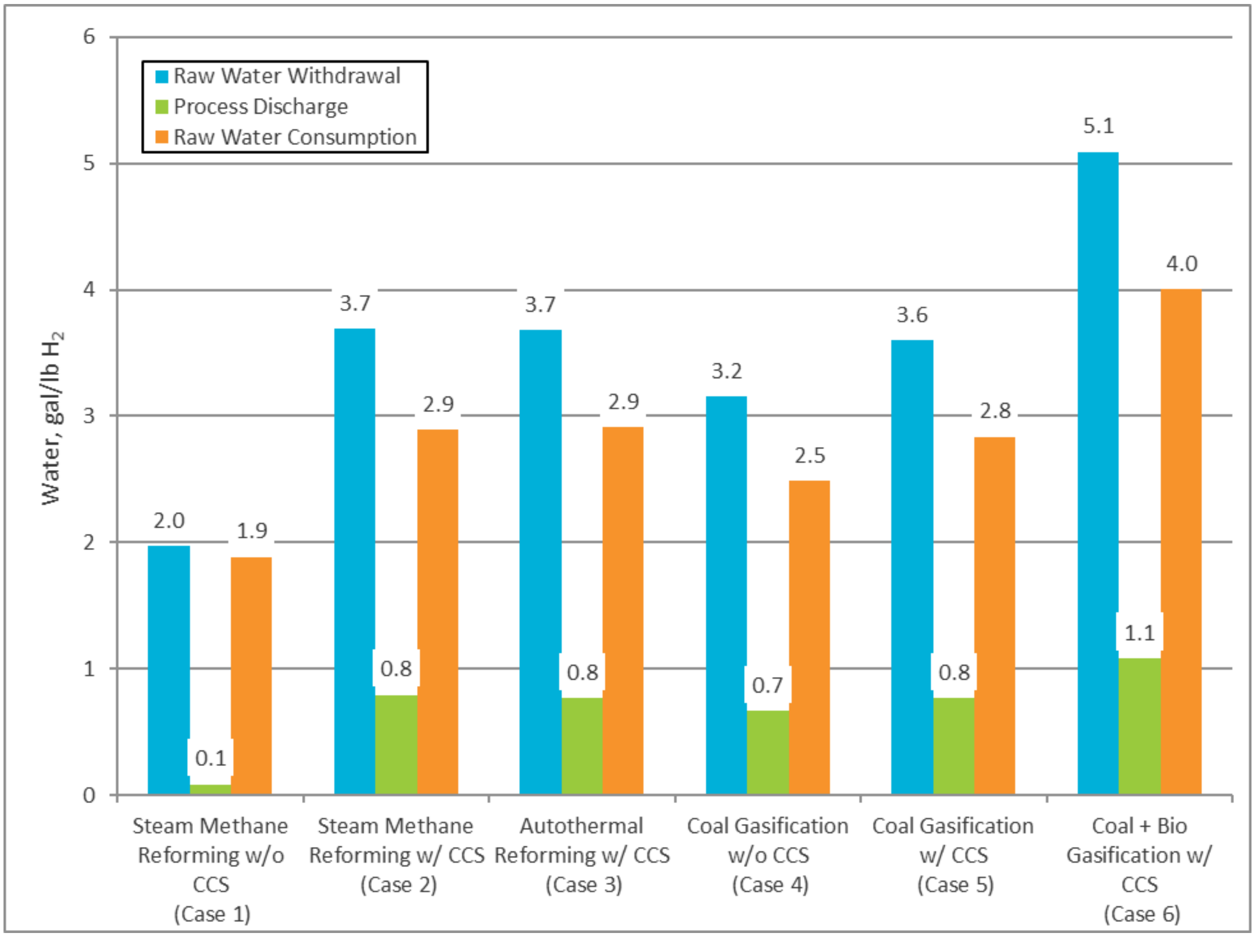
- The configuration with the highest global warming potential on a life cycle basis is the coal gasification plant without CCS (Case 4, 20 lb CO_2e/lb H_2). The configuration with the lowest GWP is the coal/biomass co-gasification plant with CCS (Case 6, -1.0 lb CO_2e/lb H_2). The life cycle emissions of Case 6 include a small credit (-0.053 lb CO_2e/lb H_2) for electricity sold back to the grid.
- The average life cycle GHG emissions of the reforming cases with capture (Case 2, 4.6 lb CO_2e/lb H_2 and Case 3, 5.7 lb CO_2e/lb H_2) is approximately 10 to 40 percent higher than life cycle GHG emissions of the coal gasification case with capture (Case 5, 4.1 lb CO_2e/lb H_2). This is primarily due to the life cycle GHG emissions of the grid electricity and natural gas feedstock procurement, which make up approximately 90 percent of the life cycle emissions of the reforming cases with capture.
- The smallest contributor to the life cycle emissions is the CO_2 management emissions, accounting from between zero, for non-capture cases, and two and eight, for capture cases, percent of the life cycle GHG emissions.
- Monte Carlo simulation results show a strong positive skew for gasification cases. For Case 6, this variation pushes the GWP results to positive 2.6 lb CO_2e/lb H_2. This would be the result of a plant getting coal from a mine with high coal mine methane emissions and its biomass from a site with a low yield rate.

Appendix B shows results with additional detail, as well as from using GWP factors other than the IPCC AR5 100-year time horizon to provide additional perspective on the results from the various cases.

5.1.3 Water Use

Three water values are presented for each technology in Exhibit 5-6: raw water withdrawal, process discharge, and raw water consumption. Each value is normalized by the hydrogen production rate.

Exhibit 5-6. Water use for all cases



The following observations can be made:

- The SMR plant without capture (Case 1) has the lowest water demand of all cases (1.9 gal/lb H_2 raw water consumption). The coal/biomass co-gasification plant with capture (Case 6) has the highest water demands of all cases (4.0 gal/lb H_2 raw water consumption), about two times more than Case 1.
- CO_2 capture imposes a significant water demand on all technologies. The capture technologies have significant cooling water demands that result in increased raw water consumption because of increased cooling tower blowdown and cooling tower evaporative losses. Raw water consumption increases by 53 percent for both SMR and ATR cases with capture, 12 percent for the coal-only gasification plant with capture, and 60 percent for the coal/biomass co-gasification plant with capture.

5.2 Cost Results

Exhibit 5-7 provides a summary of the costs for all cases.

Exhibit 5-7. Cost summary for all cases

Case Name	Reforming[A]			Gasification[A]		
	SMR		ATR	Shell Gasifier		
	1	2	3	4	5	6
Total Plant Cost (2018\$/[kg H_2/day])	554	1,394	1,092	4,264	4,901	5,241
Total Overnight Cost (2018\$/[kg H_2/day])	713	1,735	1,372	5,243	6,044	6,515
Owner's Costs	159	341	280	979	1,142	1,274
Process Contingency	0	52	1	231	304	211
Project Contingency	94	252	114	628	734	765
Home Office Expenses	77	182	163	456	516	538
Bare Erected Cost	384	908	814	2,949	3,347	3,726
Total As-Spent Cost (2018\$/[kg H_2/day])	763	1,856	1,467	5,853	6,747	7,273
LCOH (\$/kg H_2) (excluding T&S)	1.06	1.54	1.51	2.58	2.92	3.44
Capital Costs	0.14	0.33	0.26	1.17	1.35	1.46
Fixed Costs	0.07	0.15	0.11	0.53	0.60	0.75
Variable Costs	0.09	0.24	0.36	0.43	0.52	0.43
Fuel Costs	0.77	0.82	0.77	0.44	0.44	0.80
LCOH (\$/kg H_2) (including T&S)	1.06	1.64	1.59	2.58	3.09	3.64
CO_2 T&S Costs	0.00	0.10	0.09	0.00	0.17	0.20
Breakeven CO_2 Sales Price (ex. T&S), \$/tonne[B]	N/A	50.1	50.4	N/A	20.8	42.9
Breakeven CO_2 Emissions Penalty (incl. T&S), \$/tonne[B]	N/A	64.1	60.3	N/A	30.8	65.0

[A]Financing structures are presented in NETL's "QGESS: Cost Estimation Methodology for NETL Assessments of Power Plant Performance" [6]

[B]Both the breakeven CO_2 sales price and emissions penalty were calculated based on the respective non-capture cases within the reforming cases (Case 1) and gasification cases (Case 4)

5.2.1 TOC and TASC

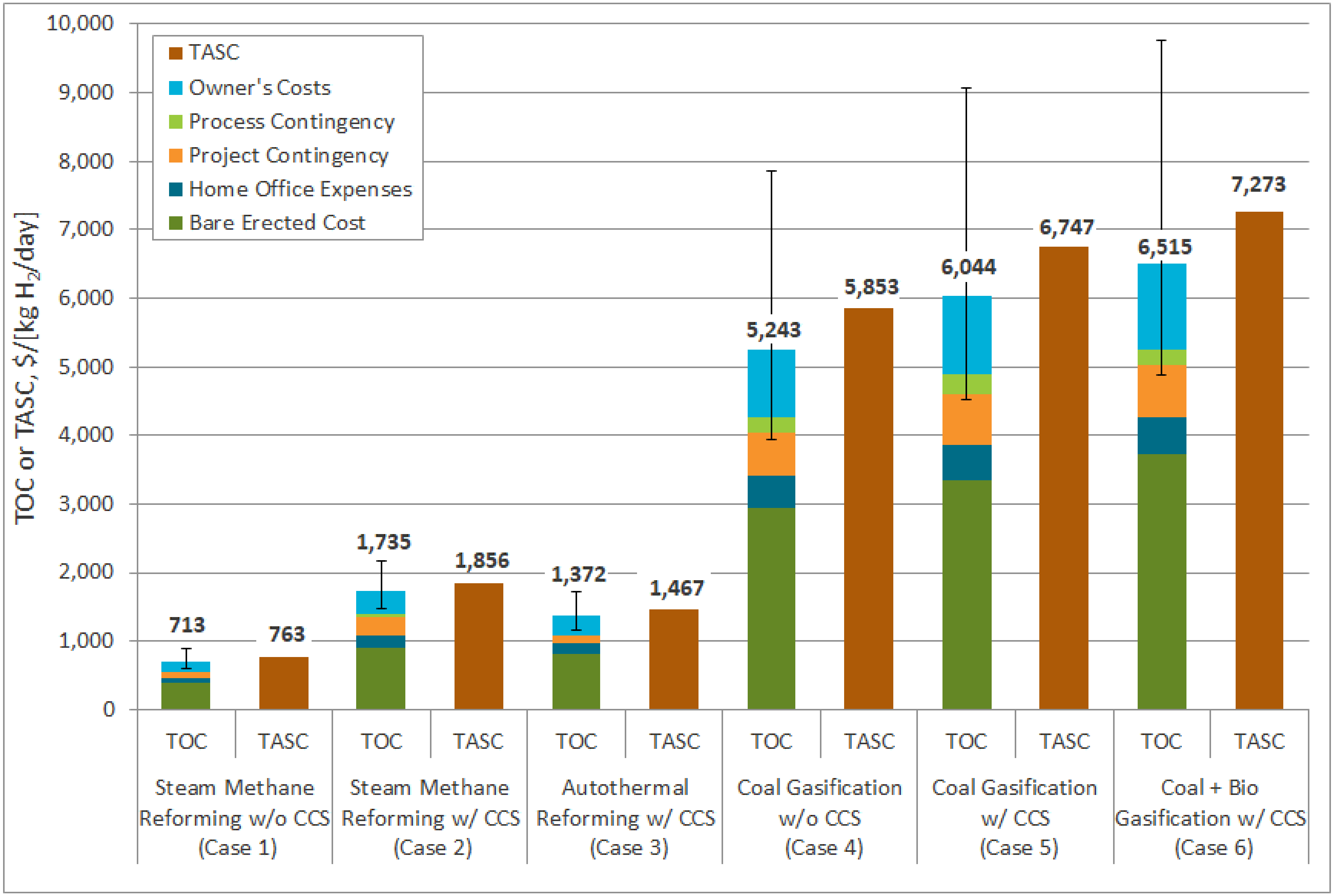
In Exhibit 5-8, the normalized components of TOC and overall TASC are shown for each technology. The error bars included represent the potential TOC range relative to the maximum and minimum of the capital cost uncertainty range, as described previously in the technology summary sections. This range is -25 percent/+50 percent for gasification cases and -15 percent/+25 percent for reforming cases.

The following observations can be made:

- The configuration with the lowest TOC is the SMR plant without capture (Case 1, \$713/[kg H_2/day]) and the configuration with the highest TOC is the coal/biomass co-gasification plant with capture (Case 6, \$6,515/[kg H_2/day]). There is almost an order of magnitude difference between the low and high TOC values.
- The average TOC for the gasification cases is more than four times greater than the average TOC for the reforming cases.

- The addition of CO_2 capture technology impacts all technologies in different magnitudes. The TOC increase for the addition of CO_2 capture technology in reforming cases is 143 percent for the SMR plant (Case 2). The TOC increases by 15 percent for the coal gasification plant (Case 5).

Exhibit 5-8. Plant capital costs

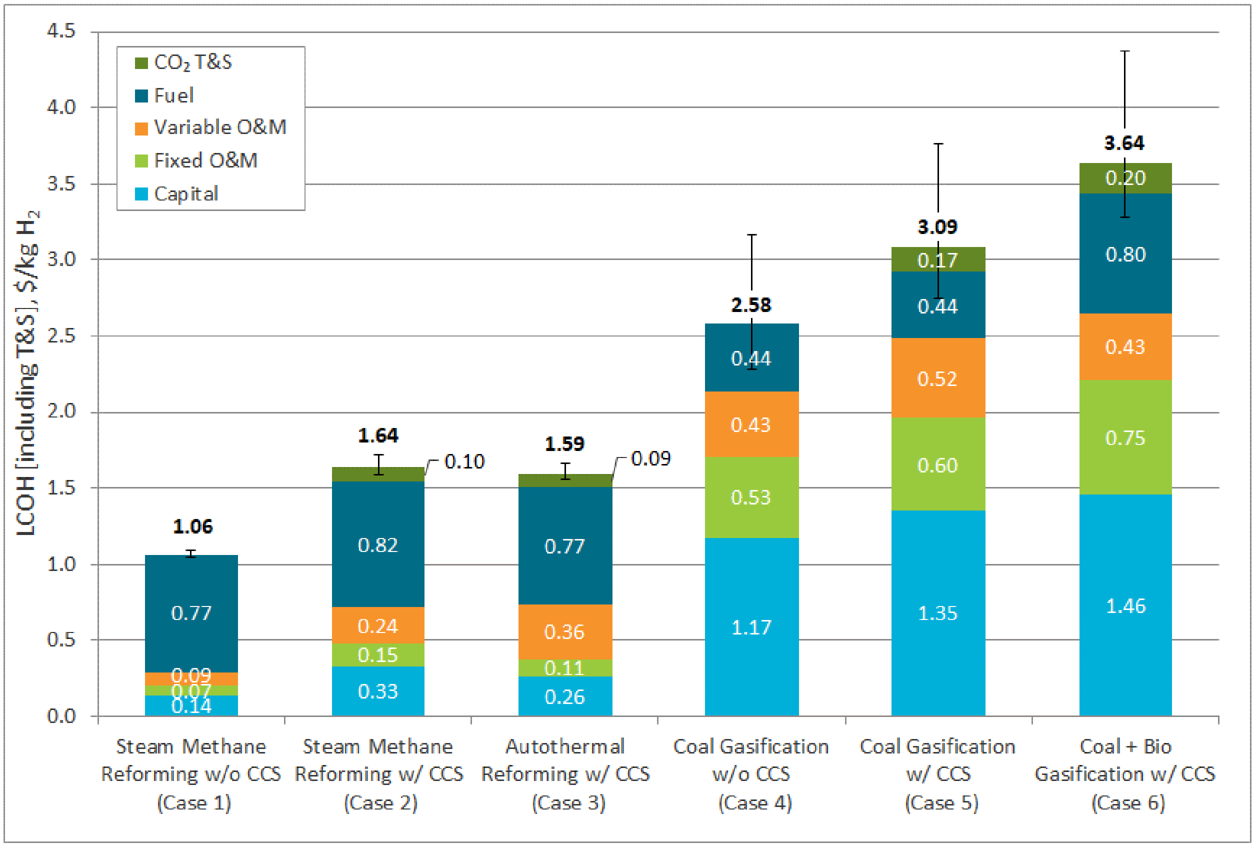


5.2.2 LCOH

A graph of the LCOH by cost component is provided in Exhibit 5-9. The error bars included represent the potential LCOH range relative to the maximum and minimum capital cost uncertainty ranges. The LCOH ranges presented are not reflective of other changes, such as variation in fuel price, labor price, CF, or other factors. The primary observations that can be made are:

- The lowest LCOH is \$1.06/kg H_2 from the SMR plant without capture (Case 1). The highest LCOH is \$3.64/kg H_2 from the coal/biomass co-gasification case with capture (Case 6).
- The average LCOH for the gasification cases is about two times greater than the average LCOH for the reforming cases. This is mainly due to the higher capital and fixed costs needed in the gasification cases compared to the reforming cases.
- The largest contributor to the LCOH for the reforming cases is the fuel cost, accounting for between 48 and 73 percent of the total LCOH. The largest contributor for the gasification cases is the capital cost, accounting for between 40 and 46 percent of the total LCOH. The cost of the feedstock, on a per thermal content unit, is about double for natural gas compared to coal. On the other hand, the greater complexity of the gasification plants explains the large contribution of capital costs to the LCOH.
- Fixed O&M costs account for 20 percent of the gasification case's average LCOH, which is 13 percentage points higher than the reforming cases. This is attributed to the relatively higher operating, maintenance, and administrative labor burden in the gasification cases as well as higher property taxes and insurance.
- The addition of CO_2 capture technology impacts the reforming plant's LCOH more than the gasification plant's LCOH. Adding CCS to the reforming cases increases the LCOH by 54 percent for the SMR plant (Case 2). Adding capture to the coal gasification plant (Case 5) increases the LCOH by 20 percent.
- The CO_2 T&S LCOH component represents about 5 or 6 percent of the total LCOH across the cases with CO_2 capture considered in this study.

Exhibit 5-9. LCOH by cost component



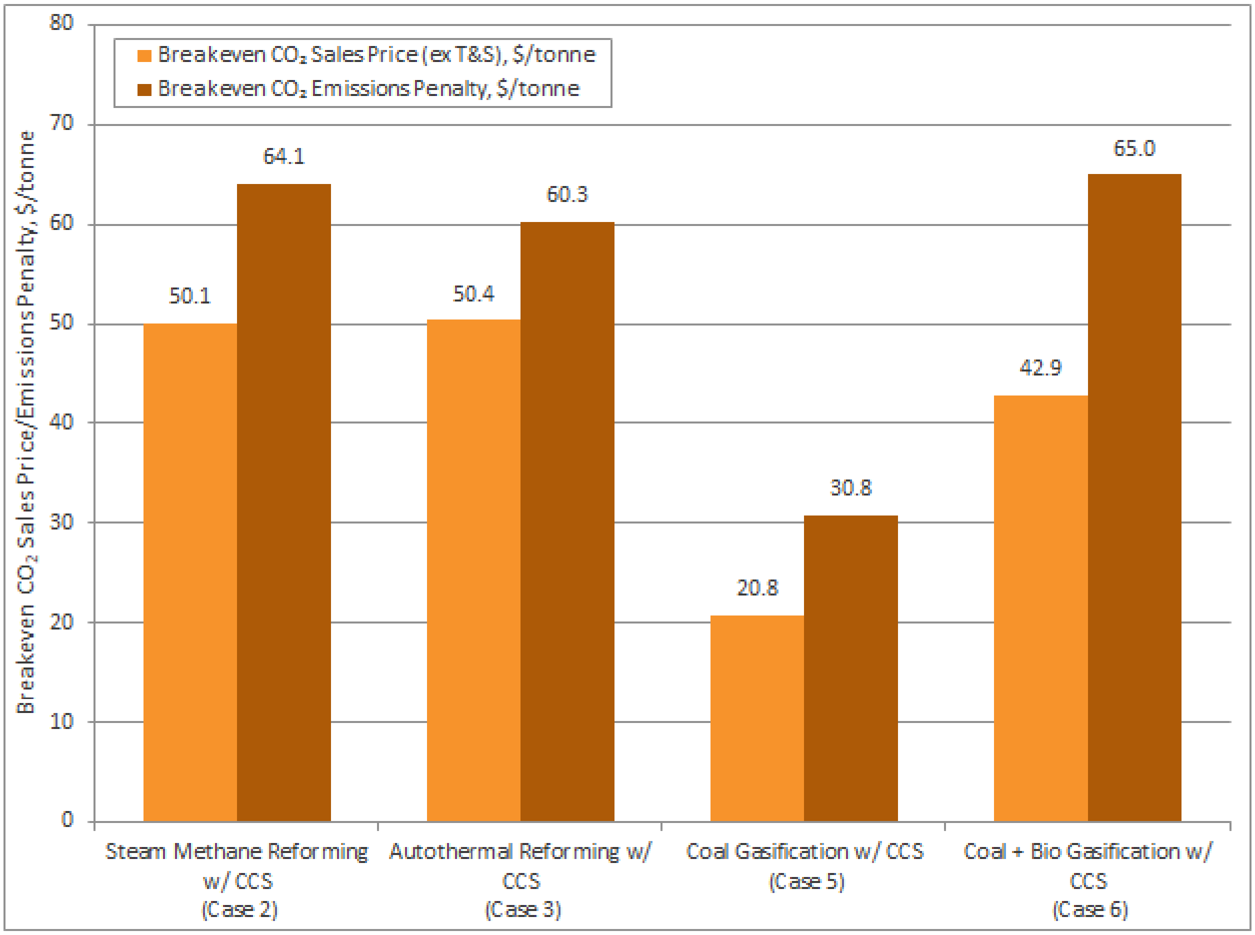
5.2.3 Stack CO_2 Emissions Penalty/Sales Price

If future legislation assigns a cost to carbon emissions, all the technologies examined in this report will become more expensive. The technologies without carbon capture will be impacted to a larger extent than those with carbon capture, and coal-based technologies will be impacted more than natural gas-based technologies. The breakeven CO_2 emissions penalty is described in Section 2.11.4 as the carbon tax on stack CO_2 emissions necessary to increase the LCOH of the non-capture cases, Case 1 and Case 4, to match the LCOH of the comparable capture cases.

Another valuable analysis considers the sale of the captured CO_2 for utilization and storage in CO_2 enhanced oil recovery (EOR), which has the potential to provide a revenue stream for capture plant configurations. The plant gate CO_2 sales price will ultimately depend on a number of factors, including plant location and crude oil prices. The breakeven CO_2 sales price represents the minimum CO_2 plant gate sales price that will incentivize carbon capture relative to a defined reference non-capture plant.

Exhibit 5-10 shows a comparison of the breakeven CO_2 sales price and emissions penalty using the equivalent non-capture cases for the reforming (Case 1) and gasification (Case 4) cases.

Exhibit 5-10. Breakeven stack CO_2 emissions penalty/sales price using equivalent non-capture cases



The following observations can be made about the CO_2 emissions penalty and sales price:

- The lowest CO_2 sales price that would create parity in the LCOH is in the coal gasification plant with capture (Case 5, $20.8/tonne CO_2). The lowest CO_2 emissions penalty that would create this parity is also in the coal gasification plant with capture (Case 5, $30.8/tonne).
- The reforming plants require a higher cost of CO_2 for their LCOH to reach parity with the non-capture case. This is primarily due to the smaller percentage difference between the non-capture and capture LCOH in gasification plants as well as the larger amount of CO_2 that is present in the gasification systems.

5.3 Sensitivity Analysis

Sensitivity analysis is a valuable tool for determining how much the results of the study will change when there are significant variations in key process variable assumptions. One of the most volatile of these process assumptions is the price of the feedstocks. In the following three plots, the LCOH is shown as a function of changes to the price of one of the feedstocks. In each sensitivity analysis, only a single feedstock price is varied with the others held constant. While it is recognized that isolated variations in these prices is unlikely, the single variable sensitivities (slopes) provide a simple approximation to estimate the impact of multi-variable changes.

Exhibit 5-11 shows the LCOH sensitivity to coal fuel costs for all cases, even though the coal fuel cost has no effect on the reforming cases. The coal price has been varied from $1 to $10/MMBtu for the sensitivity, with the study coal price being $2.23/MMBtu. This graph highlights that even at coal prices that approach zero, the gasification cases never become economically favorable compared to the reforming cases unless CO_2 is monetized. Furthermore, outside of the range of this graph, the coal/biomass co-gasification plant (Case 6) becomes more favorable than the coal only gasification plant with capture (Case 5). This occurs at high coal prices above $11.5/MMBtu.

Exhibit 5-11. LCOH sensitivity to coal fuel costs

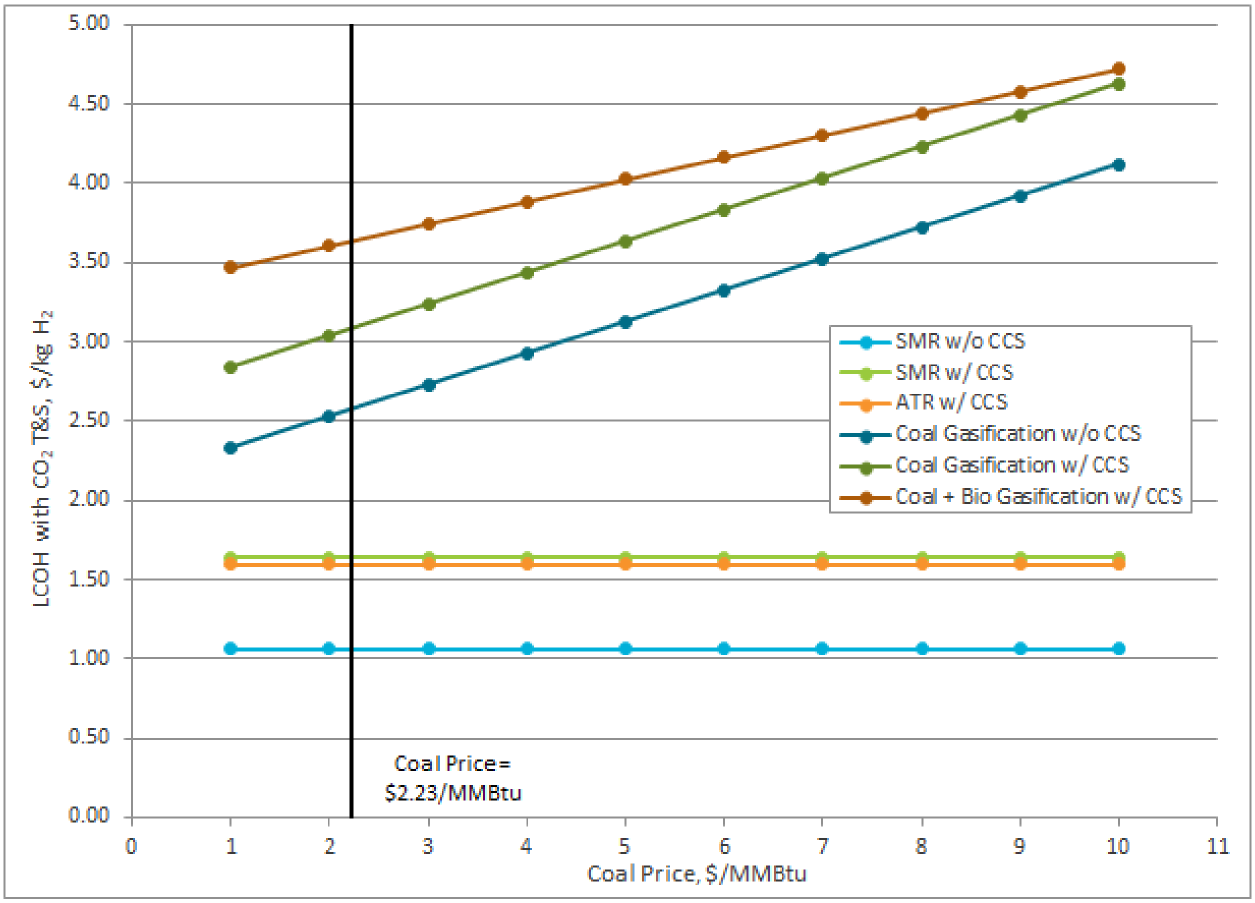


Exhibit 5-12 shows the LCOH sensitivity to the biomass fuel costs, which only affects Case 6. The biomass price has been varied from $1 to $17/MMBtu. The baseline biomass price assumed in this study was $5.43/MMBtu. The sensitivity plot shows that the coal/biomass co-gasification case (Case 6) will only become more favorable than the coal only gasification case with capture (Case 5) at a negative biomass fuel cost.

Exhibit 5-12. LCOH sensitivity to biomass fuel costs

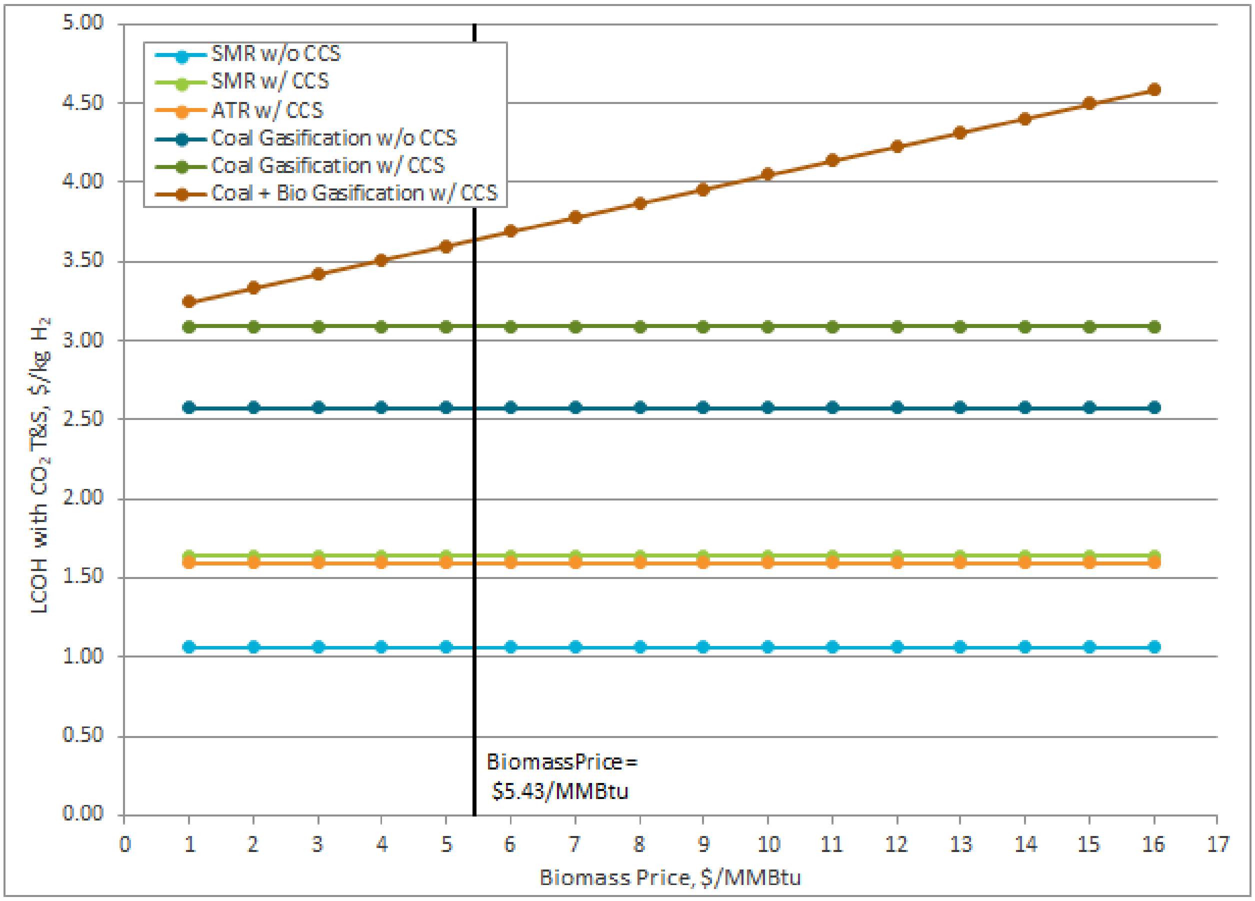


Exhibit 5-13 shows the sensitivity plot in which only the natural gas fuel costs are varied and the LCOH is examined; therefore, the gasification cases have no change. The natural gas price was varied from $1 to $11/MMBtu, with the study price being $4.42/MMBtu. The results show that at a low natural gas price of $1/MMBtu, the reforming cases with capture (Case 2 and Case 3), become equivalent, economically, at a LCOH of about $1/kg H_2. Furthermore, at high natural gas prices above $13/MMBtu, the gasification case with capture becomes more favorable than the reforming cases with capture.

Exhibit 5-13. LCOH sensitivity to natural gas fuel costs

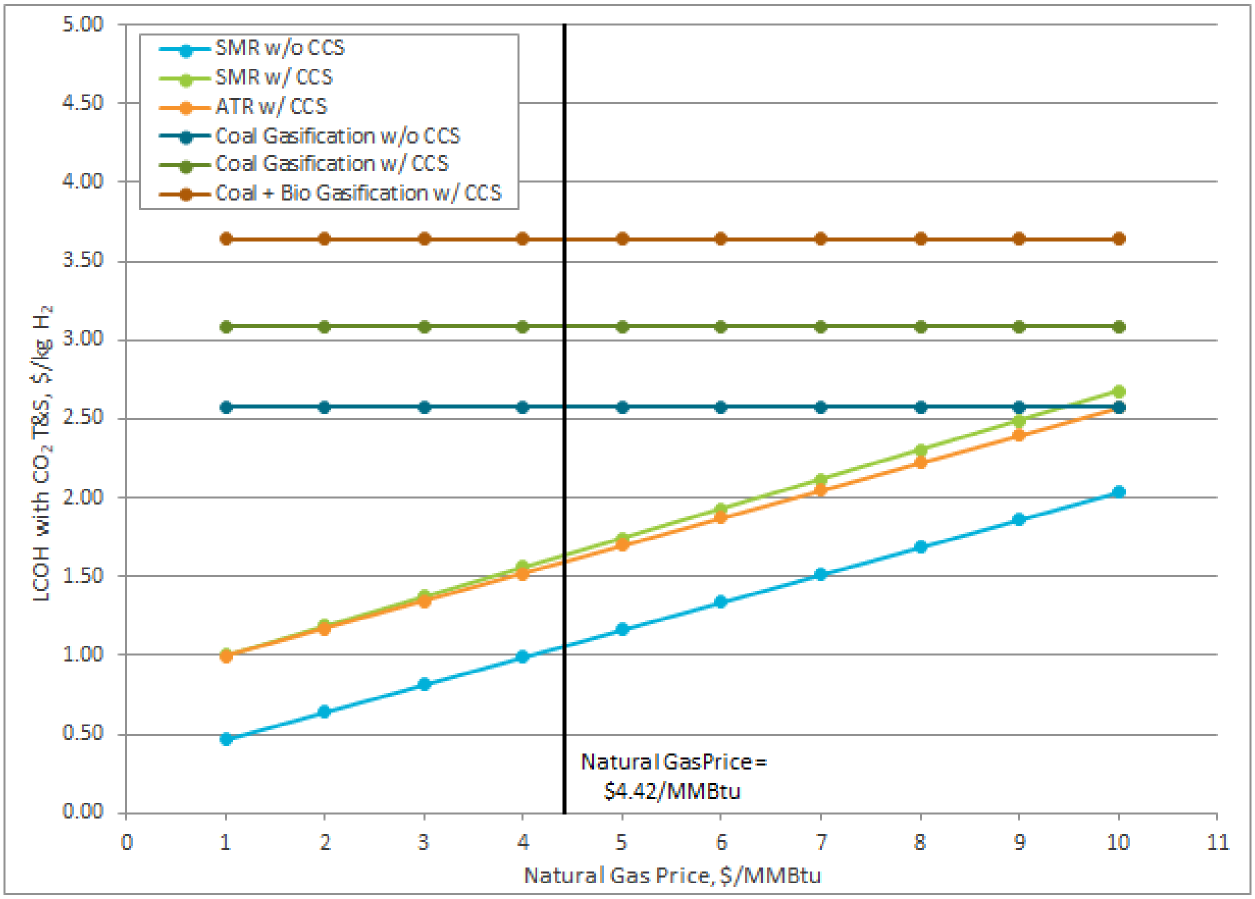
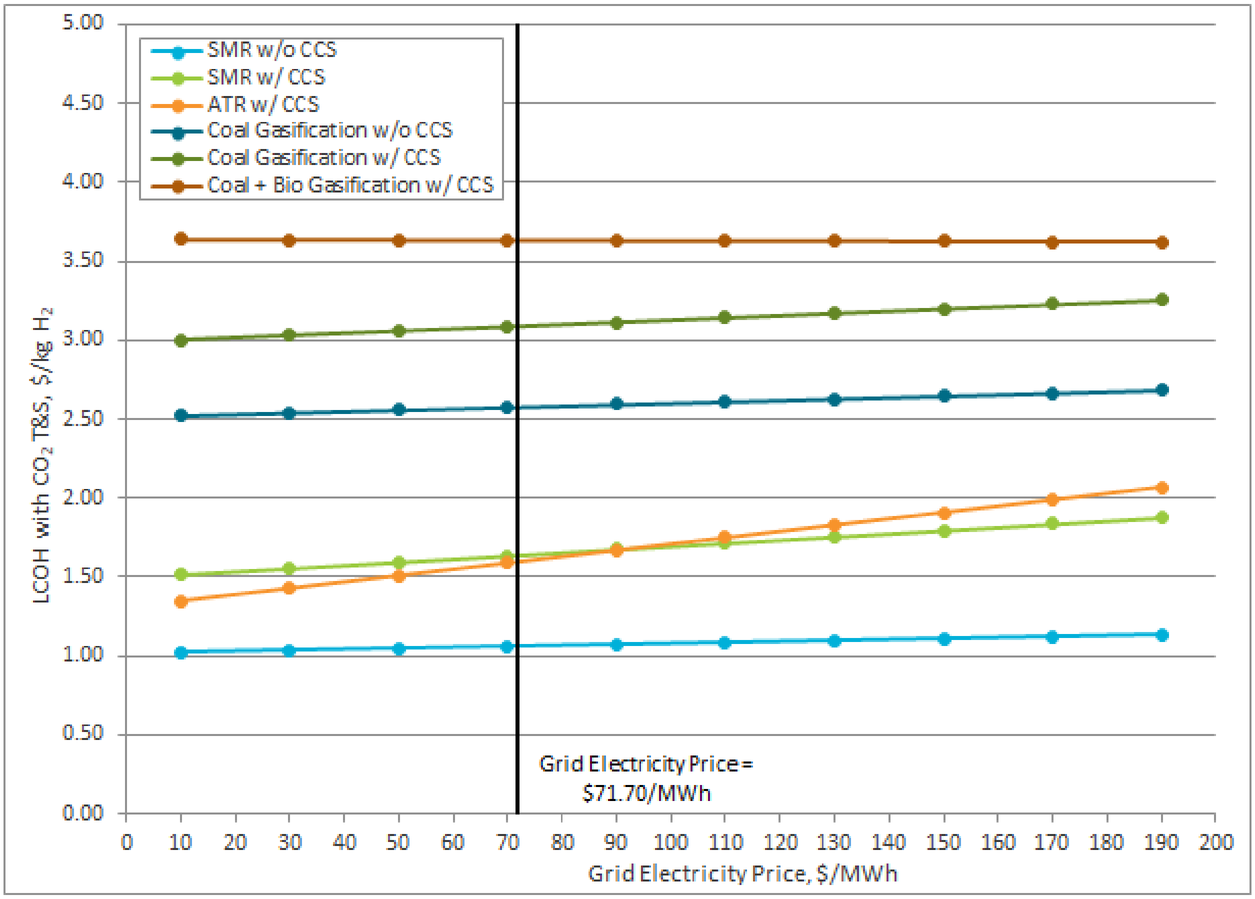


Exhibit 5-14 shows the LCOH sensitivity to the grid electricity price over a range of prices from $10 to $190/MWh. The price point used in this study is $71.70/MWh. The LCOH is relatively insensitive to changes in grid electricity price, with Case 3 and Case 5 being the most sensitive due to their large grid power demand. At grid electricity prices approaching $100/MWh, the LCOH of the ATR plant with capture (Case 3) becomes approximately equal to that of the SMR plant with capture (Case 2). This is due to the higher net power and therefore higher grid electricity demands of Case 3.

Exhibit 5-14. LCOH sensitivity to grid electricity price



In Exhibit 5-15, the sensitivity of LCOH to CF is shown for all technologies. The curves are relatively grouped by technology and whether CO_2 capture is included or not. The CF is plotted from 30 to 95 percent. The baseline CF is 90 percent for reforming cases and 80 percent for gasification cases. The curves plotted in Exhibit 5-15 assume that the CF could be extended to 95 percent with no additional capital equipment, and that lower capacity factors do not result in capital cost savings. An analysis examining the LCOH sensitivity to additional capital expenditures needed to achieve a capacity factor greater than the 80 percent baseline assumption in the gasification cases is presented in Section 5.4.3.

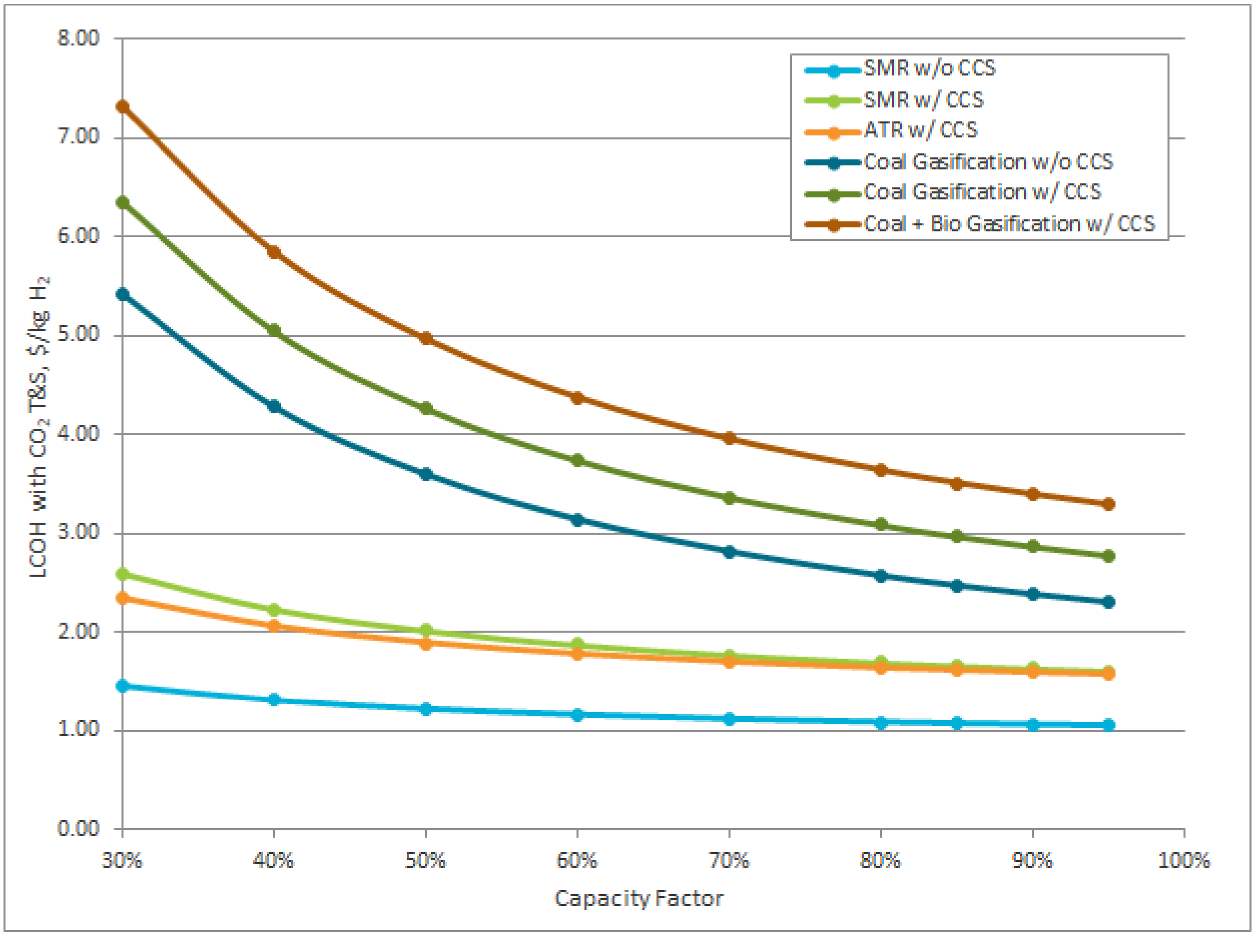
Technologies with high capital cost, such as all the gasification cases, show a greater increase in LCOH with decreased CF. Conversely, the SMR plant with no CO_2 capture is relatively flat because the LCOH is dominated by fuel charges, which decrease as the CF decreases.

Conclusions that can be drawn from Exhibit 5-15 include:

- In general, the gasification cases are more sensitive to changes in the CF than the reforming cases. The cases employing CO_2 capture are slightly more sensitive to changes in the CF than cases without CO_2 capture.

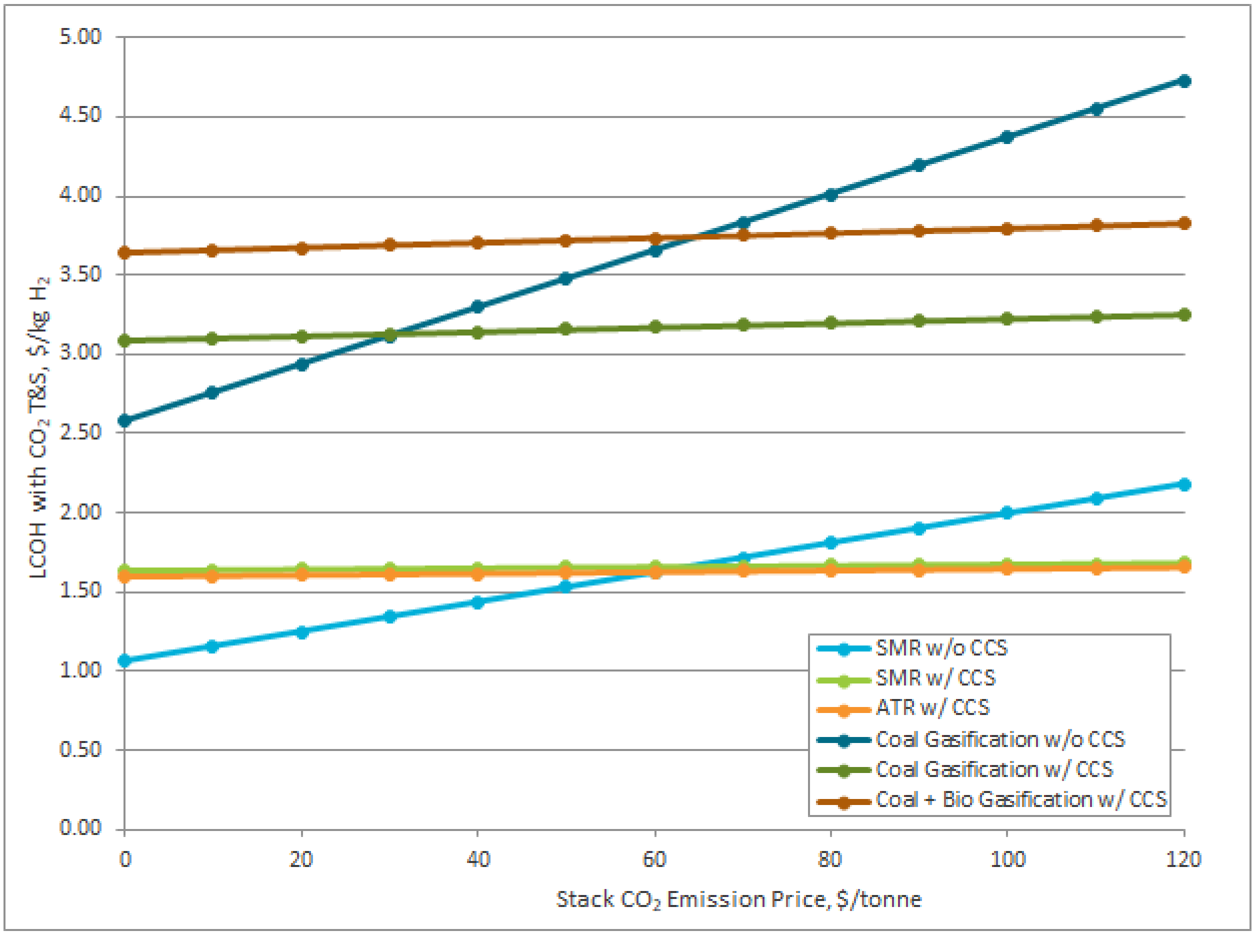
- At high capacity factors, the LCOH difference between the SMR and ATR plants with capture (Case 2 and Case 3, respectively) become negligible. None of the cases obtain a LCOH below \$1/kg H_2, even at high capacity factors.

Exhibit 5-15. LCOH sensitivity to capacity factor



The final sensitivity examines how the LCOH is impacted by a range of stack CO_2 emissions penalties, from \$0, which is used as a baseline assumption in this study, to \$120/tonne CO_2. All of the lines have a positive slope due to greater-than-zero CO_2 stack emissions; however, the slopes of the non-capture lines are steeper due to the larger amount of CO_2 released. The intersection points between the non-capture and capture cases' lines are the breakeven CO_2 emission prices, previously discussed in Section 5.2.3. At high CO_2 emission penalties, above \$100/tonne CO_2, the non-capture reforming case produces hydrogen that costs more than \$2.00/kg H_2 and the non-capture gasification case produces hydrogen that costs more than \$4.00/kg H_2.

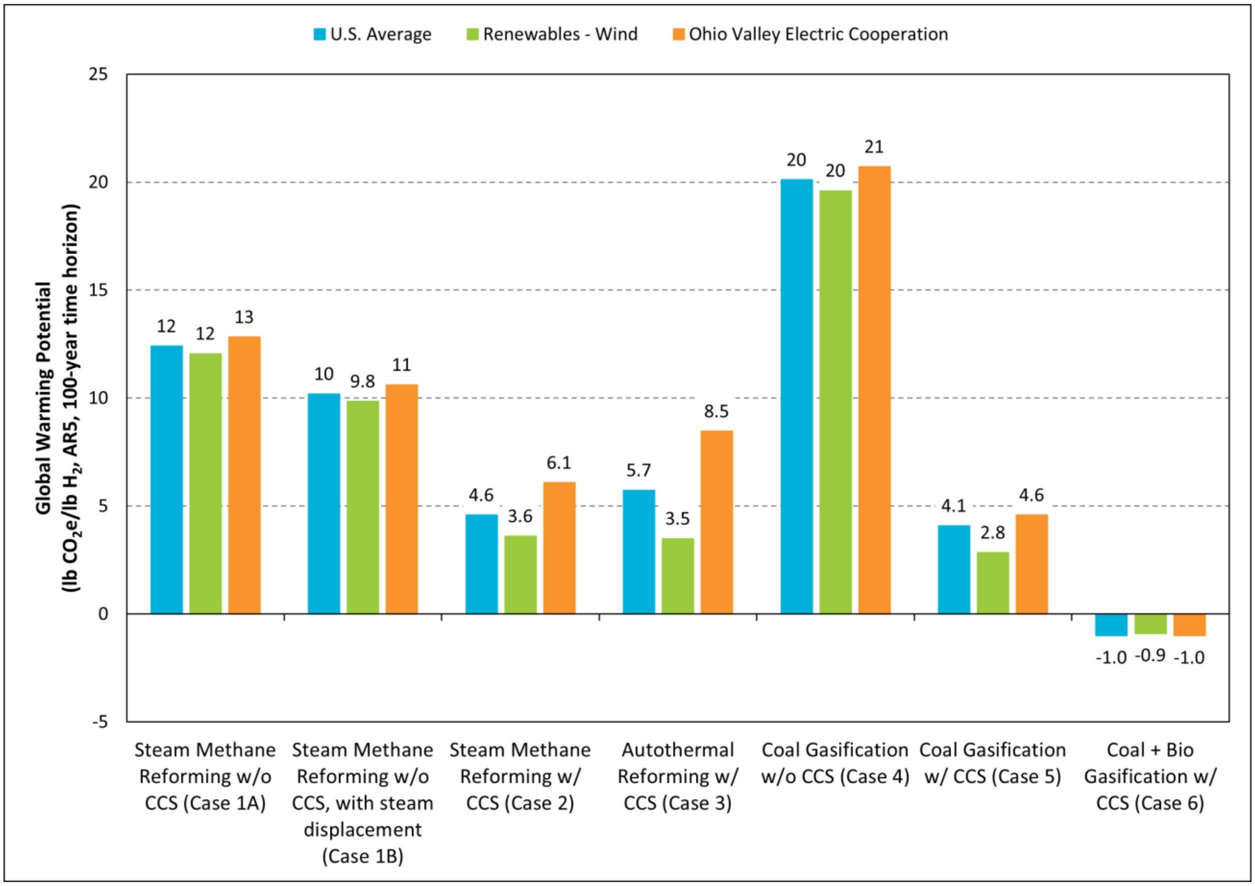
Exhibit 5-16. LCOH sensitivity to stack CO_2 emissions price



In addition to LCOH, a sensitivity analysis was performed on the life cycle greenhouse gas emissions results, reported as global warming potential (GWP). The inputs assessed for this sensitivity analysis include the effects of electricity source, natural gas source, and methane leakage rates.

Exhibit 5-17 shows the effects of electricity source on GWP for each baseline technology (where Case 1 represents Case 1A). The low-emissions electricity source is 100% wind energy, modeled based on the U.S. average, and the high-emissions electricity source is modeled as the Ohio Valley Electric Cooperation grid mix. [103] [11] [12] Indirect emissions changes in the other upstream supply chains caused by varying electricity sources are not modeled.

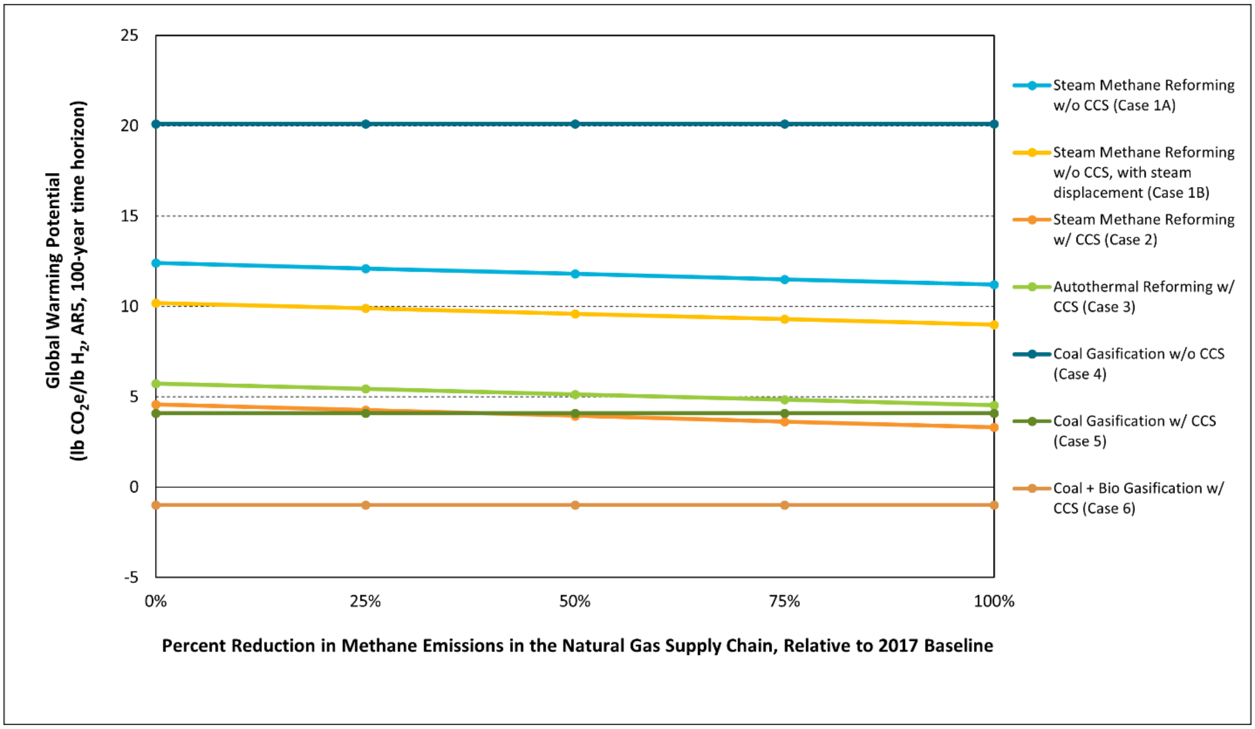
Exhibit 5-17. GWP sensitivity to electricity source



Case 3 imports the most electricity per unit of H_2 produced and therefore is the most sensitive to the electricity source. As a net electricity exporter, Case 6 receives credit for displacing emissions that would have been caused from electricity generation, and thus shows a lower GWP for the high-emissions case. No matter the electricity source, Case 4 shows the highest GWP and Case 6 shows the lowest GWP. Overall, electricity source is not a large contributor to the overall GHG emissions but is generally much more significant in the cases with CCS than those without.

Exhibit 5-18 shows the sensitivity of total GWP results to levels of methane emissions in the upstream gas supply chain. Methane emissions contribute nearly 40% of the GWP in the natural gas supply chain and are currently a focus of EPA and DOE GHG identification and mitigation efforts. [9] [104] Similar to the regional natural gas sensitivity analysis, indirect emission reductions in the electricity supply chain caused by lower methane leakage are not modeled. The natural gas supply chain is modeled for U.S. average natural gas production through transmission, resulting in a GWP of 0.76 kg CO_2e per kg natural gas. [9]

Exhibit 5-18. GWP sensitivity to natural gas supply chain methane emissions



As the upstream methane emissions are reduced, there is a linear reduction in GWP for Cases 1-3. At 100% reduction in natural gas supply chain methane emissions, Case 1A has a 9.6% reduction in GWP, Case 1B has a 12% reduction in GWP, Case 2 has a 26% reduction in GWP, and Case 3 has a 21% reduction in GWP. This suggests that methane reduction has a higher impact on technologies with CCS. Also, at 100% reduction, Case 2 reaches a lower GWP value than Case 5. At all methane emissions reduction rates, Case 4 shows the highest GWP and Case 6 shows the lowest GWP.

The following statements are a few high-level takeaways from the sensitivity analysis. First, no scenarios resulted in a noticeable reduction in GWP for Case 4, which had a GWP 1.5 times greater than the second highest-emitting technology in all scenarios. Next, Case 6 continued to have the lowest GWP in all scenarios. Meanwhile, Cases 2, 3, and 5 were relatively close in GWP and their ranking changed with each scenario. This suggests that geographic region will have a meaningful impact on GHG emissions for these cases in real-world implementation.

5.4 Other Analyses

5.4.1 Hydrogen Compression Credit

The DOE Hydrogen and Fuel Cells Program develops and maintains a suite of Hydrogen Analysis Production Models (H2A) to allow users to estimate the cost of hydrogen production through various fossil and non-fossil routes. [105] Case studies for the production routes are publicly available with documentation detailing underlying assumptions. Hydrogen production costs from projected current (2019) and projected future (2035) distributed Polymer Electrolyte

Membrane (PEM) Electrolysis are available for production capacities representative of distributed (1,500 kg H_2/day) and centralized (50,000 kg H_2/day) facilities. [106]

Within the H2A models, a compression credit is applied to the projected future PEM electrolysis cases that inherently produce a hydrogen product above the 300 psig comparative standard. The credit is calculated by estimating the capital and operating cost of a notional hydrogen compressor, since there is no hydrogen compressor within the projected future PEM electrolysis configuration. After being normalized to a $/kg H_2 basis, the credit is subtracted from the LCOH. The calculated credits were $0.09/kg H_2 and $0.05/kg H_2 for the projected future distributed and centralized cases, respectively. [106]

The following analysis was conducted to determine a compression credit for the six study cases in a similar manner to the H2A models. Each of the study cases has an inherent hydrogen pressure greater than the NREL comparative standard of 300 psig and includes additional compression to a final pipeline-ready pressure of 925 psig by a hydrogen compressor. Crediting the LCOH of each study case makes the results comparable to lower-pressure hydrogen production routes and analyses, like in the H2A model.

The hydrogen compressor in each study case, which has known capital and operating costs, is used to determine the appropriate compression credit. A compression credit factor is applied to the known cost of compression to account for any pressurization greater than 300 psig that is accomplished before the hydrogen compressor. This factor, as calculated by the following equation, is greater than one for the six study cases because each case’s inherent hydrogen pressure prior to the compressor is greater than the NREL comparative standard of 300 psig:

$$Compression\ Credit\ Factor = \frac{(P_{final} - P_{ref,NREL})}{(P_{final} - P_{initial})}$$

Where:

P_{final} – compressed H_2 product pressure, psig

$P_{ref,NREL}$ – NREL H2A H_2 product reference pressure, 300 psig

$P_{initial}$ – uncompressed inherent H_2 pressure, psig

The hydrogen compression credit is the factored cost of compression that is normalized on an annual H_2 production basis. The credit is subtracted from the baseline LCOH to produce an adjusted LCOH which approximates a 300 psig H_2 delivery price. The adjusted LCOH is shown for all cases in Exhibit 5-19. The result of the analysis is that the LCOH for each case is reduced by about $0.04/kg H_2.

Exhibit 5-19. Hydrogen compression credit applied to LCOH for all cases

Case Name	Reforming			Gasification		
	SMR		ATR	Shell Gasifier		
	1	2	3	4	5	6
Inherent H_2 Pressure, psig (uncompressed)	331	316	316	416	394	394
H_2 Product Pressure, psig (compressed)	925					
Compression Credit Factor	1.053	1.027	1.027	1.229	1.177	1.177
Cost of Compression (MM$)	6.37	6.61	8.75	6.16	5.96	1.51
Baseline LCOH ($/kg H_2) (including T&S)	1.06	1.64	1.59	2.58	3.09	3.64
Hydrogen Compression Credit	-0.042	-0.043	-0.041	-0.039	-0.036	-0.046
Adjusted LCOH ($/kg H_2) (including T&S)	1.02	1.59	1.55	2.54	3.05	3.59

5.4.2 Export Steam Sales Price

Superheated steam at 399 °C (750 °F) and 3.1 MPa (450 psia) is produced in the SMR plant without CO_2 capture (Case 1) at quantities that exceed the internal process demand for the steam. The plant water balance is presented in Exhibit 3-10 and shows that of the steam generated within the plant 53 percent is consumed in the SMR reactor for hydrogen production while the balance is exported from the plant. This is a common occurrence in operating SMR facilities without the additional thermal demands of solvent regeneration from CO_2 removal processes.

Effectively, export steam generation serves as a process cooling mechanism resulting in a potentially saleable by-product. However, challenges exist to produce steam at conditions and at purity levels meeting potential customer requirements in situations where potential customers are co-located. Flowsheet solutions also exist to eliminate the generation of export steam in situations where a customer is not co-located. [107]

It was not an objective of the SMR without CO_2 capture plant (Case 1) in this study to produce a saleable export steam byproduct or eliminate the production of export steam. It was assumed that export steam would be exported off-site as a zero-revenue byproduct. However, a simple sensitivity to steam price was performed in which a value of the steam was $6/ton. The steam value was calculated by using a simplified approach to estimate the amount of process heat available for steam generation and the cost of the natural gas consumed to provide only the sensible heating of the superheated steam. Applying the $6/ton value as a variable O&M revenue stream reduced the LCOH from $1.06 to $1.01/kg H_2.

5.4.3 Gasifier Capacity Factor and Sparing

Section 2.9.3 describes the 80 percent capacity factor assumption used for the gasification cases. This assumption differs from that of the reforming cases which are assumed to have a capacity factor of 90 percent. A sensitivity analysis was performed to quantify the impact to LCOH by adding a spare gasifier train and associated auxiliary capital equipment in order to achieve a 90 percent capacity factor. Exhibit 5-20 shows the results of the analysis.

In each of the three gasification study cases the LCOH is increased by adding the spare gasifier and increasing the capacity factor to 90 percent. This indicates that the additional revenue produced from operating at a higher capacity factor throughout the year is not enough to offset the additional capital required to achieve the higher capacity factor.

Adding a spare gasifier had the most impact on the coal gasification case without CO_2 capture, increasing the LCOH by 9 percent. The gasification cases with CO_2 capture were impacted to lesser extent by 7 and 4 percent for coal (Case 5) and coal/biomass (Case 6), respectively. This is primarily due to the additional costs associated with the dual stage Selexol and biomass co-feedstock diluting the overall impact of adding the spare gasifier costs.

Exhibit 5-20. LCOH impact from gasifier train sparing

Cost	Case (*=gasifier sparing)					
	4	4*	5	5*	6	6*
LCOH ($/kg H_2) (excluding T&S)	2.58	2.82	2.92	3.13	3.44	3.58
Capital Costs	1.17	1.32	1.35	1.48	1.46	1.55
Fixed Costs	0.53	0.58	0.6	0.66	0.75	0.77
Variable Costs	0.43	0.44	0.52	0.55	0.43	0.46
Fuel Costs	0.44	0.44	0.44	0.44	0.80	0.80
LCOH ($/kg H_2) (including T&S)	**2.58**	**2.82**	**3.09**	**3.30**	**3.64**	**3.78**
CO_2 T&S Costs	0.00	0.00	0.17	0.17	0.20	0.20

6 FUTURE WORK

This report has presented the results of an independent assessment of the cost and performance of select hydrogen production plants utilizing fossil fuel resources as the primary feedstocks. As stated previously, the cost and performance results presented are important inputs to assessments and determinations of technology combinations to be utilized to meet the projected demands of the nascent hydrogen economy. Future iterations of this report may leverage the included cases to assess R&D goals and metrics, expand the number of commercially representative case configurations included, modify the existing representations, or modify design basis assumptions. As such, this section identifies several potential areas of interest for future work.

6.1 REDUCING LIFE CYCLE GHG EMISSIONS

While these hydrogen baseline technologies were modeled with the best available data, there are GHG emissions that were not accounted for. Technologies with water as a feedstock must meet specific water quality requirements to be used in the process. In many cases, this requires upstream water treatment. However, feedstock water treatment was not modeled for these baseline technologies due to an assumed negligible impact. ISO 14040 specifies that all inputs with impacts greater than 1% of the total should be included in the model [1] [2]; thus, this exclusion is insignificant to the results. Construction of plants and components for hydrogen production technologies were also not modeled. Previous LCA models of energy-related technologies show that construction impacts rarely surpass the 1% threshold [43] [14], which was the basis for excluding them from the scope.

As explained in Section 2.7.3, this study examined plant arrangements capable of achieving capture rates in excess of 90 percent of the feedstock carbon entering the facility. The resulting life cycle GHG emission intensities for gasification were -1.1 and 4.1 lb CO_2e/lb H_2 for coal/biomass and coal with CCS, respectively. The reforming cases were 5.5 and 6.6 lb CO_2e/lb H_2 for SMR with CCS and ATR with CCS, respectively. Future study to reduce life cycle GHG emissions are identified in the sections following.

6.1.1 Renewable Natural Gas Blending

RNG is a NG pipeline quality gas that has been processed from a renewable, methane-containing source gas. RNG can be blended into an existing NG transmission or distribution network to reduce the GHG intensity of the pipeline gas. Typical source gases for RNG production are biogases from landfills, dairies, and municipal wastewater treatment plants. Impurities such as CO_2, water vapor, H_2S, and dioxins are removed from the source gas to upgrade the heat content and composition to meet NG pipeline specifications.

By feeding a blended NG/RNG feedstock to a reforming plant, the life cycle GHG intensity of the plant is reduced. Further study into establishing the magnitude of life cycle GHG reductions and whether RNG-blending is a viable pathway to achieving net-zero, or net-negative, life cycle GHG emissions is needed.

6.1.2 Direct Air Capture Pairing

Direct air capture (DAC) technology is the use of forced draft circulation to adsorb atmospheric CO_2 using solvents or fixed-bed sorbents. As a standalone technology DAC provides an atmospheric CO_2 sink much like photosynthesis during the growth cycle of biomass. However, the sorbent (or sorbent) must be regenerated, and separated CO_2 must be compressed and pipelined for storage or utilization.

Integrated system concepts where DAC technology is paired with a hydrogen production facility may offer a pathway to further reduce uncaptured CO_2 emissions emitted directly from the plant stack and/or through atmospheric CO_2 removal offsets. If sufficient atmospheric CO_2, or a combination of both CO_2 sources is treated, net-negative life cycle GHG emissions attributable to the facility is possible. Potential synergies exist where the CO_2 compression train of the host hydrogen production plant can be leveraged by an on-site DAC facility.

6.1.3 Low Carbon Auxiliary Power

GHG emissions associated with imported grid electricity are a significant contributor to the life cycle GHG profile of the reforming cases. Importing or co-producing no- or low-carbon electricity from low-carbon syngas, fossil-based sources with carbon capture, hydrogen, or renewables can provide a pathway to reducing the overall GHG profile of the hydrogen production facility. Analyses examining the LCOH impacts of several alternative auxiliary powering options under various grid power cost and emission profile assumptions would be beneficial in assessing the most cost competitive strategies for lowering this portion of the GHG profile.

6.1.4 Higher CO_2 Capture Rates

As described in Section 2.7.3, the capture cases in this report maximized CO_2 capture using a defined approach. No adjustments were made to the vendor-provided CO_2 separation efficiencies of the AGR technologies modeled. Obtaining cost and performance information from technology vendors for AGR systems designed for greater CO_2 separation efficiency would enable quantification of the magnitude of plant stack CO_2 emission reduction, as well as the LCOH impact, from incorporating such designs.

6.2 ADVANCED TECHNOLOGY ASSESSMENTS

Plant configurations containing advanced technologies may offer potential cost and performance improvements to the systems examined in this study after overcoming fundamental R&D barriers before market adoption. Part of the utility of this study is as a performance and cost benchmark against which advanced hydrogen production plant configurations can be measured. The following are brief descriptions of advanced technology concepts that may be compared in future analyses.

6.2.1 Advanced Reforming Concepts

Chemical looping reforming (CLR), gas switching reforming (GSR), and membrane-assisted ATR (MA-ATR) are emerging advanced reforming concepts that potentially offer improvements to current NG reforming technologies by eliminating the need for certain conventional processes. [108]

Like the approach in other chemical looping concepts, CLR uses a circulating oxygen carrier to provide oxidant to partially oxidize a NG feed to generate a syngas product. R&D is ongoing to develop oxygen carriers also capable of catalyzing the reforming reaction to generate additional H_2 in the syngas product.

GSR aims to avoid the fundamental issues associated with circulating the solid oxygen carrier at the pressures and temperatures required for CLR. This concept involves a series of fixed reactor beds containing the oxygen carrier material. The fixed beds are periodically exposed to air and NG/steam streams by switching between streams, thereby allowing these reactants to flow over the fixed beds without the need for solids circulation. Demonstrating chemical and mechanical stability and durability of a suitable oxygen carrier as well as high-temperature switching valves is a challenge for this concept. [108]

The MA-ATR concept utilizes a fluidized, catalytic reactor which is fed with NG and oxygen from a conventional ASU. Hydrogen-selective membranes continuously separate hydrogen from the reactor reforming zone while a CO_2-rich stream is separated at the opposite end of the reactor. The in-situ hydrogen separation is a process intensification concept that is aimed at eliminating the WGS and PSA steps needed in conventional reforming systems. Challenges with this concept include significant electricity consumption, membrane scale-up difficulties, and low hydrogen permeate pressure. [108]

6.2.2 NG Pyrolysis

The pyrolysis of natural gas is an endothermic process that occurs in the absence of oxygen to form hydrogen and a solid carbon product. Although currently thought of as a production route for carbon black, the technology has potential as a form of hydrogen production with low CO_2 emissions. Future work is needed to better understand the current market applications and limitations for commercial, state-of-the-art NG pyrolysis. Additionally, integrated process system concepts may be developed to begin to understand the relative performance and cost when compared to conventional NG reforming processes.

6.2.3 Supercritical Water Gasification

Supercritical water gasification (SCWG) is an alternative gasification approach to conventional thermal gasification which reacts coal, biomass, or a mixture with water above its supercritical point (374 °C and 22.1 MPa). SCWG differs from thermal gasification by utilizing the enhanced solubility of organics in supercritical water and relying on supercritical water itself as the oxidant. Due to the use of supercritical water as the reaction media, a high-hydrogen syngas results with one study reporting a hydrogen fraction of nearly 73 mol% in the syngas. [109] Potential advantages of this technology include the elimination of the ASU as well as a reduction in gas cleanup requirements. Challenges include scale-up, reactor materials, and overcoming the large heat demand needed for water to reach supercritical conditions.

Future analysis in this area is needed to understand the potential cost and performance benefits of this technology as an advanced option for co-gasifying biomass with coal as a net-zero hydrogen production technology.

6.2.4 Advanced H_2/CO_2 Separation Concepts

Advanced concepts for H_2 and CO_2 separation continue to be an active area of R&D. Various separation methods including absorption, adsorption, and membranes seek to improve overall system cost and performance beyond current pre-combustion separation methods. Historically having focused on IGCC applications, these technologies can likely be applied to gasification-based hydrogen production plants as well. Exhibit 6-1. provides a summary of current, NETL-funded projects for pre-combustion H_2/CO_2 separation technologies.

Future work in this area is planned in order to develop a TEA to assess the potential performance and cost benefits of select technologies within a coal and biomass "net-zero" plant arrangement.

Exhibit 6-1. NETL-funded advanced H_2/CO_2 R&D projects

Project Title	Performer	Agreement Number	Date
Technology Gap in Coal FIRST Gasification Based Poly-Generation: Advanced Ceramic Membranes	Media and Process Technology, Inc.	FE0031930	10/1/2020 – 9/30/2023
Bench-Scale Development of a Transformative Membrane Process for Pre-Combustion CO_2 Capture	Membrane Technology and Research, Inc.	FE0031632	10/1/2018 – 9/30/2021
Development and Testing of a High Temperature PBI Hollow-Fiber Membrane Technology for Pre-Combustion CO_2 Capture	SRI International	FE0031633	10/1/2018 – 3/31/2022
Transformational Membranes for Pre-Combustion Carbon Capture	Ohio State University	FE0031635	10/1/2018 – 9/30/2021
Development of Carbon Molecular Sieves Hollow Fiber Membranes Based on Polybenzimidazole Doped with Polyprotic Acids with Superior H_2/CO_2 Separation Properties	State University of New York	FE0031636	10/1/2018 – 9/30/2021
Bench Scale Testing of a High Efficiency, Ultra-Compact Process for Pre-Combustion CO_2 Capture	University of Southern California	FE0031737	6/1/2019 – 5/31/2022
A High Efficiency, Modular Pre-Combustion Capture System for Coal FIRST Poly-Generation Process	TDA Research, Inc.	FE0031926	10/1/2020 – 9/30/2023
Alkanolamines for Acid Gas Removal in Gasification Processes	Pacific Northwest National Laboratory	FWP-72564	10/1/2018 – 3/31/2021

Project Title	Performer	Agreement Number	Date
High-Temperature Ceramic-Carbonate Dual-Phase Membrane Reactor for Pre-Combustion Carbon Dioxide Capture	Arizona State University	FE0031634	10/1/2018 – 9/30/2021
Integrated Multichannel Water Gas Shift Catalytic Membrane Reactor for Pre-Combustion Carbon Capture	Bettergy Corporation	SC0018853	7/2/2018 – 8/18/2021
Integrated Water-Gas-Shift (WGS) / Pre-Combustion Carbon Capture Process	TDA Research, Inc.	FE0023684	10/1/2014 – 9/30/2021

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APPENDIX A: HYDROGEN UNIT CONVERSIONS

This appendix provides select hydrogen unit conversions, which may help put the data presented in this report into various contexts.

First, there are multiple ways that hydrogen capacities can be reported, including on mass and volumetric bases. Exhibit A shows various unit conversions of an example capacity of 100 MMSCFD of hydrogen.

Exhibit A-1. Equivalent hydrogen capacities in various units

Value	Unit	Notes
100	MMSCFD	"standard conditions" are 60 °F and 14.7 psia [110]
111	kNm^3/hr	"normal conditions" are 32 °F and 14.7 psia [110]
22,167	lb/hr	
266	ton/day	
87,955	metric ton/year	assumes 100% capacity factor

For turbine firing applications, 100 MMSCFD of H_2 can fuel a 212 MW NGCC plant, assuming a NGCC HHV efficiency of 53.6% as reported in the Fossil Energy Baseline. For ammonia (NH_3) production applications, 100 MMSCFD of H_2 supports a 1,496 tpd NH_3 plant using the Haber-Bosch process.

Next, the cost of hydrogen, which is typically reported in units of \$/kg H_2, can be reported on an energy content basis. \$1/kg H_2 is equivalent to \$7.56/MMBtu on a HHV basis and \$8.94/MMBtu on a LHV basis. This assumes a H_2 HHV of 59,970 Btu/lb and a LHV of 50,740 Btu/lb. Finally, the barrel of oil equivalent (BOE), which provides a comparison to the energy content of a barrel of crude oil, of \$1/kg H_2 is \$47/BOE on a LHV basis.

Appendix B: Detailed Life Cycle Results

Detailed Results Using Only IPCC AR5 100-year GWP Values

The results shown in Exhibit 5-5 are detailed in tabular form in Exhibit B-1. Note that more significant figures are shown in this Appendix than elsewhere in the report in order to help highlight the often small differences in the total, 5th, and 95th percentile values. Inclusion of these additional digits should not otherwise imply more faith in the underlying results. The 5th and 95th percentile values were derived from a Monte Carlo simulation with 5,000 iterations, and as stated previously, cases including upstream natural gas also reflect regional variability across natural gas basins. The low case is represented by Alaskan Offshore gas and included in 5th percentile results, and the high case is represented by San Juan Conventional gas and included in 95th percentile results.

Exhibit B-1. Life cycle GHG emissions for all hydrogen scenarios (IPCC AR5, 100-yr) (lb CO_2e/lb H_2)

	Reforming				Gasification		
	SMR			ATR	CG		CG/BG
Case Name	1A	1B	2	3	4	5	6
Stack Emissions	9.35	9.35	0.380	0.514	17.9	1.35	1.59
Natural Gas Emissions	2.69	2.69	2.86	2.69	-	-	-
Coal Emissions	-	-	-	-	1.64	1.63	1.15
Biomass Emissions	-	-	-	-	-	-	-4.01
Grid Electricity Emissions	0.368	0.368	1.19	2.39	0.533	0.834	-0.0530
CO_2 Management Emissions	-	-	0.151	0.140	-	0.263	0.318
Steam Generation	-	-2.21	-	-	-	-	-
Total	12.4	10.2	4.57	5.72	20.1	4.08	-1.00
Low (5th percentile)	11.1	8.85	3.13	4.33	19.4	3.40	-1.40
High (95th percentile)	16.4	14.2	8.86	9.79	24.9	8.87	2.51

Given hydrogen's HHV of 59,970 Btu/lb, GWP results can also be presented in terms of energy. as shown in Exhibit B-2 and Exhibit B-3.

Exhibit B-2. Life cycle GHG emissions for all hydrogen scenarios (IPCC AR5, 100-yr) (lb CO_2e/MMBtu H_2)

	Reforming				Gasification		
	SMR			ATR	CG		CG/BG
Case Name	1A	1B	2	3	4	5	6
Stack Emissions	156	156	6.33	8.57	299	22.5	26.6
Natural Gas Emissions	44.8	44.8	47.6	44.8	-	-	-
Coal Emissions	-	-	-	-	27.7	27.3	19.1
Biomass Emissions	-	-	-	-	-	-	-66.8

	Reforming				Gasification		
	SMR			ATR	CG		CG/BG
Case Name	1A	1B	2	3	4	5	6
Grid Electricity Emissions	6.14	6.14	19.8	39.8	8.88	13.9	-0.884
CO_2 Management Emissions	-	-	2.52	2.33	-	4.38	5.30
Steam Generation	-	-36.9	-	-	-	-	-
Total	207	170	76.3	95.5	335	68.0	-16.4
Low (5th percentile)	184	148	52.2	72.3	324	56.6	-23.3
High (95th percentile)	274	237	148	163	415	148	41.8

Exhibit B-3. Life cycle GHG emissions for all hydrogen scenarios (IPCC AR5, 100-yr) (kg CO_2e/MJ H_2)

	Reforming				Gasification		
	SMR			ATR	CG		CG/BG
Case Name	1A	1B	2	3	4	5	6
Stack Emissions	0.0670	0.0670	0.00272	0.00368	0.129	0.00966	0.0114
Natural Gas Emissions	0.0192	0.0192	0.0205	0.0192	-	-	-
Coal Emissions	-	-	-	-	0.0117	0.0117	0.00822
Biomass Emissions	-	-	-	-	-	-	-0.0287
Grid Electricity Emissions	0.00264	0.00264	0.00851	0.0171	0.00382	0.00598	-0.000380
CO_2 Management Emissions	-	-	0.00108	0.00100	-	0.00188	0.00228
Steam Generation	-	-0.0159	-	-	-	-	-
Total	0.0889	0.0730	0.0328	0.0410	0.144	0.0292	-0.00720
Low (5th percentile)	0.0793	0.0627	0.0224	0.0311	0.139	0.0243	-0.0100
High (95th percentile)	0.118	0.102	0.0635	0.0702	0.178	0.0636	0.0180

Detailed Results Using Various IPCC GWP Values

The results shown in the main study all use the IPCC AR5 100-year GWP values. Impact results for all cases were also calculated with five other GWP methods produced by IPCC to help demonstrate the sensitivity of results to different impact assessment methods. These results are displayed in Exhibit B-4.

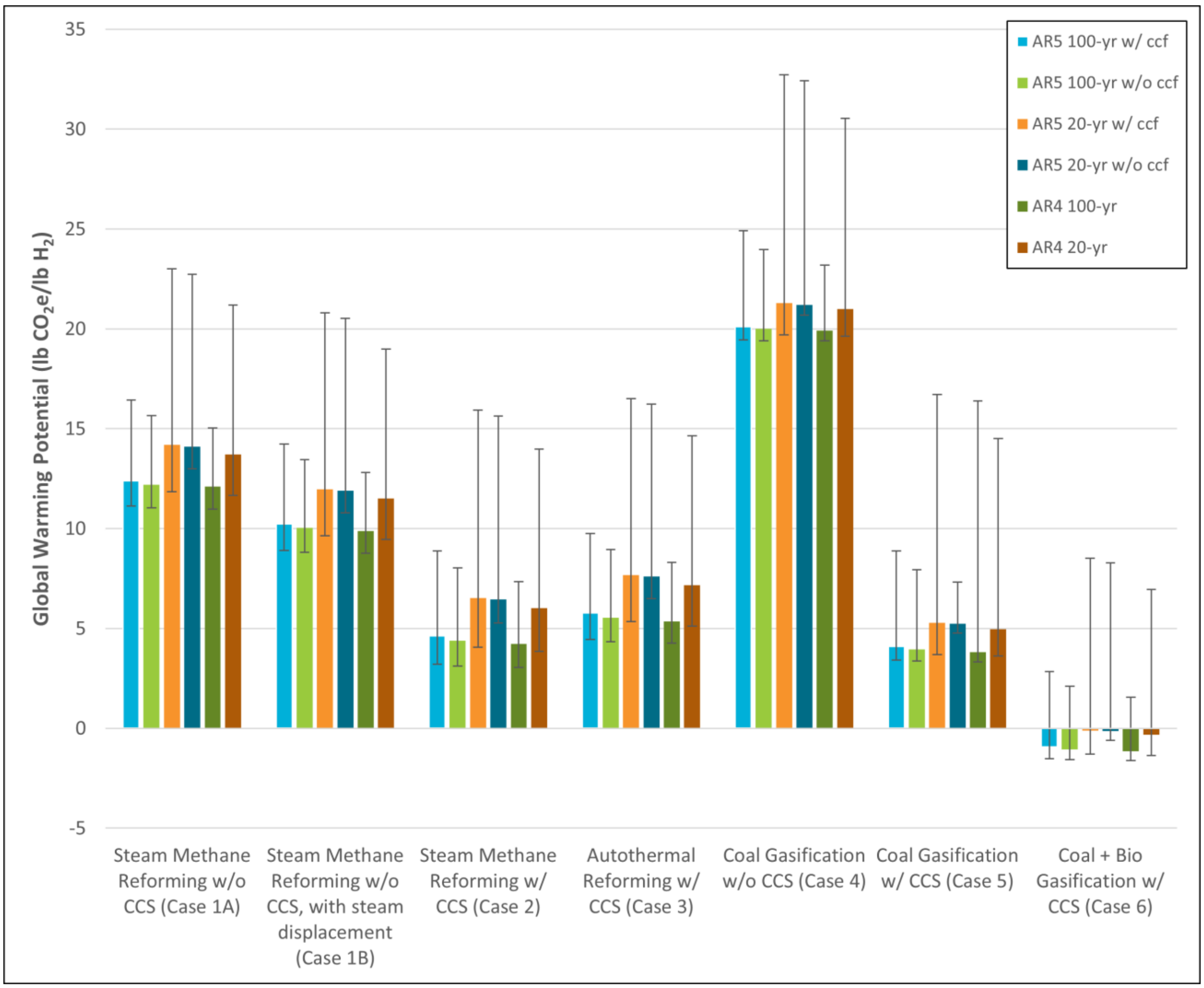
Exhibit B-4. GWP results for all cases by IPCC scenario (lb CO_2e/lb H_2)

	Reforming				Gasification		
	SMR			ATR	CG		CG/BG
Case Name	1A	1B	2	3	4	5	6
AR5, 100-yr time horizon, with climate carbon feedback	12	10	4.6	5.8	20	4.1	-0.94
AR5, 100-yr time horizon, without climate carbon feedback	12	10	4.4	5.5	20	4.0	-1.1

	Reforming				Gasification		
	SMR			ATR	CG		CG/BG
Case Name	1A	1B	2	3	4	5	6
AR5, 20-yr time horizon, with climate carbon feedback	14	12	6.5	7.7	21	5.3	-0.11
AR5, 20-yr time horizon, without climate carbon feedback	14	12	6.5	7.6	21	5.2	-0.14
AR4, 100-yr time horizon	12	9.9	4.2	5.4	20	3.8	-1.1
AR4, 20-yr time horizon	14	11	6.0	7.2	21	5.0	-0.33

These results are also shown graphically in Exhibit B-5, including uncertainty. While not shown explicitly in the graph, the major source of variability in the results are different values used for methane, and to a lesser extent, nitrous oxide. And, as expected, the use of the 20-year GWP values (which are higher than the 100-year values) leads to a boost in GWP results.

Exhibit B-5. GWP results for all cases and across six GWP methods (lb CO_2e/lb H_2)



All GWP factors used for the IPCC scenarios in this section of the Appendix are recorded in Exhibit B-6.

Exhibit B-6. Summary of GWP factors used in all IPCC scenarios [41] [111]

Common Name	Formula	AR5 100-yr w/ ccf	AR5 100-yr w/o ccf	AR5 20-yr w/ ccf	AR5 20-yr w/o ccf	AR4 100-yr	AR4 20-yr
Carbon dioxide	CO_2	1	1	1	1	1	1
Methane	CH_4	36	30	86	84	25	72
Nitrous oxide	N_2O	298	265	268	264	298	289
CFC-11	CCl_3F	5,350	4,660	7,020	6,900	4,750	4,750
CFC-12	CCl_2F_2	10,200	10,200	10,200	10,200	10,900	10,900
CFC-13	$CClF_3$	13,900	13,900	13,900	13,900	14,400	14,400
CFC-113	CCl_2FCClF_2	5,820	5,820	5,820	5,820	6,130	6,130
CFC-114	$CClF_2CClF_2$	8,590	8,590	8,590	8,590	10,000	10,000
CFC-115	$CClF_2CF_3$	7,670	7,670	7,670	7,670	7,370	7,370
Halon-1301	$CBrF_3$	6,290	6,290	6,290	6,290	7,140	7,140
Halon-1211	$CBrClF_2$	1,750	1,750	1,750	1,750	1,890	1,890
Carbon tetrachloride	CCl_4	1,730	1,730	1,730	1,730	1,400	1,400
Methyl bromide	CH_3Br	2	2	2	2	5	5
Methyl chloroform	CH_3CCl_3	160	160	160	160	146	146
HCFC-21	$CHCl_2F$	148	148	148	148	-	-
R-22 (HCFC-22)	$CHClF_2$	1,760	1,760	1,760	1,760	1,810	1,810
HCFC-123	$CHCl_2CF_3$	79	79	79	79	77	77
HCFC-124	$CHClFCF_3$	527	527	527	527	609	609
HCFC-141b	CH_3CCl_2F	782	782	782	782	725	725
HCFC-142b	CH_3CClF_2	1,980	1,980	1,980	1,980	2,310	2,310
HCFC-225ca	$CHCl_2CF_2CF_3$	127	127	127	127	122	122
HCFC-225cb	$CHClFCF_2CClF_2$	525	525	525	525	595	595
Methylene chloride	CH_2Cl_2	9	9	9	9	9	9
Trichloromethane (Chloroform)	$CHCl_3$	16	16	16	16	-	-
Ethane	C_2H_6	-	-	-	-	6	6
Propane	C_3H_8	-	-	-	-	3	3
Butane	C_4H_{10}	-	-	-	-	4	4
Ethylene	C_2H_4	-	-	-	-	4	4

Common Name	Formula	AR5 100-yr w/ ccf	AR5 100-yr w/o ccf	AR5 20-yr w/ ccf	AR5 20-yr w/o ccf	AR4 100-yr	AR4 20-yr
Propylene	C_3H_6	-	-	-	-	2	2
Toluene	C_7H_8	-	-	-	-	3	3
Isoprene	C_5H_8	-	-	-	-	3	3
Methanol	CH_3OH	-	-	-	-	3	3
Acetaldehyde	CH_3CHO	-	-	-	-	1	1
Acetone	CH_3COCH_3	-	-	-	-	1	1
HFC-23	CHF_3	12,400	12,400	12,400	12,400	14,800	12,000
difluoromethane (HFC-32)	CH_2F_2	675	675	675	675	675	675
HFC-125	C_2HF_5	3,500	3,500	3,500	3,500	3,500	3,500
1,1,1,2-tetrafluoroethane (HFC-134a)	$C_2H_2F_4$ (CH_2FCF_3)	1,550	1,300	3,790	3,710	1,430	1,430
HFC-143	$C_2H_3F_3$ (CHF_2CH_2F)	353	353	353	353	353	353
1,1,1-Trifluoroethane (HFC-143a)	$C_2H_3F_3$ (CF_3CH_3)	4,470	4,470	4,470	4,470	4,470	4,470
HFC-236fa	$C_3H_2F_6$	9,810	9,810	9,810	9,810	9,810	9,810
HFC-245fa	$CHF_2CH_2CF_3$	1,030	1,030	1,030	1,030	1,030	1,030
Carbon tetrafluoride (PFC-14)	CF_4	7,390	6,630	4,950	4,880	7,390	7390
Hexafluoroethane (PFC-116)	C_2F_6	12,200	12,200	12,200	12,200	12,200	12,200
Sulphur hexafluoride	SF_6	23,500	23,500	17,500	17,500	22,800	16,300
Nitrogen trifluoride	NF_3	17,200	17,200	17,200	17,200	17,200	17,200
Trifluoromethyl sulfur pentafluoride	SF_5CF_3	17,400	17,400	17,400	17,400	17,700	17,700

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