NATIONAL ENERGY TECHNOLOGY LABORATORY

Assessment of Hydrogen Production with CO_2 Capture Volume 1: Baseline State-of-the-Art Plants

Revision 1, November 14, 2011

DOE/NETL-2011/1434

U.S. DEPARTMENT OF ENERGY

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ASSESSMENT OF HYDROGEN PRODUCTION WITH CO_2 CAPTURE VOLUME 1: BASELINE STATE-OF-THE-ART PLANTS

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FINAL REPORT

Revision 1, November 14, 2011

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LIST OF ACRONYMS AND ABBREVIATIONS

acfm	Actual cubic feet per minute
AEO	Annual Energy Outlook
AGR	Acid gas removal
ASU	Air separation unit
atm	Atmosphere
BACT	Best available control technology
BEC	Bare erected cost
BFW	Boiler feed water
BLS	U.S. Bureau of Labor Statistics
Btu	British thermal unit
Btu/hr	British thermal unit per hour
CCF	Capital charge rate factor
CF	Capacity factor
CGE	Cold gas efficiency
cm	Centimeter
CLP	Calcium looping process
CM	Construction management
CO_2	Carbon dioxide
COE	Cost of electricity
COH	Cost of hydrogen
COS	Carbonyl sulfide
CWT	Cold water temperature
DB	Dry basis
DCS	Distributed control system
DI	De-ionized
DOE	Department of Energy
EIA	Energy Information Administration
EPC	Engineering, procurement, construction
EPRI	Electric Power Research Institute
ERC	Emission reduction credits
ETE	Effective thermal efficiency
FD	Forced draft
FG	Flue gas
ft	Foot, Feet
FW	Feedwater
FO&M	Fixed operations and maintenance
gal	Gallon
gal/MWh	Gallon per megawatt hour
gpm	Gallons per minute
GEE	General Electric Energy
GJ	Gigajoule
GRI	Gas Research Institute
GT	Gas turbine
H_2	Hydrogen
H2A	DOE's Hydrogen Advisory Group

HHV	Higher heating value
HCl	Hydrogen chloride
HO	Home office
hp	Horsepower
HP	High pressure
hr	Hour
HRSG	Heat recovery steam generator
HSS	Heat stable salts
HVAC	Heating, ventilating, and air conditioning
HWT	Hot water temperature
Hz	Hertz
ID	Induced draft
IEA	International Energy Agency
IGCC	Integrated coal gasification combined cycle
in. H_2O	Inches water
in. Hga	Inches mercury (absolute pressure)
in. W.C.	Inches water column
in	Inch, Inches
kg/hr	Kilogram per hour
kJ	Kilojoules
kJ/hr	Kilojoules per hour
KO	Knockout
kPa	Kilopascal absolute
kV	Kilovolt
kW	Kilowatt
kWe	Kilowatts electric
kWh	Kilowatt-hour
kWt	Kilowatts thermal
LAER	Lowest achievable emission rate
lb	Pound
lb/hr	Pounds per hour
lb/ft^2	Pounds per square foot
LHV	Lower heating value
LIBOR	London Interbank Offered Rate
LP	Low pressure
lpm	Liters per minute
LV	Low voltage
m	Meters
m/min	Meters per minute
m^3/min	Cubic meter per minute
MAF	Moisture and Ash Free
MCR	Maximum continuous rate
md	millidarcy, is a measure of permeability defined as roughly 10^{-12} Darcy
MDEA	Methyldiethanolamine
MEA	Monoethanolamine
MHz	Megahertz

million $	Millions of dollars
MMBtu	Million British thermal units (also shown as 10^6 Btu)
MMBtu/hr	Million British thermal units (also shown as 10^6 Btu) per hour
MMSCFD	Million standard cubic feet per day
mole%	Mole percent (percent by mole)
MPa	Megapascals absolute
MVA	Megavolt (average)
MWe	Megawatts electric
MWh	Megawatt-hour
MWt	Megawatts thermal
Neg.	Negligible
NETL	National Energy Technology Laboratory
N/A	Not applicable
nm^3	Normal cubic meter
NOx	Oxides of nitrogen
NSR	New Source Review
O&M	Operations and maintenance
OC	Operating cost
OSU	Ohio State University
PM	Particulate matter
POTW	Publicly owned treatment works
ppm	Parts per million
ppmd	Parts per million, dry basis
ppmv	Parts per million volume
ppmvd	Parts per million volume, dry
ppmw	Parts per million weight
PSA	Pressure swing adsorption
PSFM	Power Systems Financial Model
psia	Pounds per square inch absolute
psid	Pounds per square inch differential
psig	Pounds per square inch gage
QGESS	Quality Guidelines for Energy System Studies
Qty	Quantity
Ref.	Reference
RDS	Research and Development Solutions, LLC
scfd	Standard cubic feet per day
scfm	Standard cubic feet per minute
Sch.	Schedule
SCR	Selective catalytic reduction
SG	Specific gravity
SMR	Steam methane reforming
SNCR	Selective non-catalytic reduction
SNG	Synthetic natural gas
SO_2	Sulfur dioxide
SOx	Oxides of sulfur
SS	Stainless steel

ST	Steam turbine
STG	Steam turbine generator
SWS	Sour water scrubber
TASC	Total as-spent cost
TGTU	Tail gas treatment unit
TOC	Total overnight cost
TPC	Total plant cost
tpd	Tons per day
tph	Tons per hour
TPI	Total plant investment
tonne	Metric ton (1000 kg)
TS&M	Transport, storage, and monitoring
UOP	UOP LLC, a Honeywell company
Vert.	Vertical
V-L	Vapor and liquid portion of stream (excluding solids)
VO&M	Variable Operations and maintenance
vol%	Volume percent (percent by volume)
WB	Wet bulb
WG	Water gauge
WGS	Water gas shift
WPI	Worcester Polytechnic Institute
wt%	Weight percent (percent by weight)
$M	Millions of dollars

EXECUTIVE SUMMARY

The objective of this study was to analyze potential plant configurations to determine their baseline performance and cost of producing hydrogen from natural gas and coal. The plants were assumed to be designed and constructed in the near future based on technologies as they exist today, with a planned startup year of 2015. This report covers the following base cases:

- **Case 1-1** – Baseline Steam Methane Reforming (SMR) Hydrogen Plant with CO_2 Capture and Sequestration matching the hydrogen generation rate of case 2-1
- **Case 1-2** – Baseline Steam Methane Reforming (SMR) Hydrogen Plant with CO_2 Capture and Sequestration matching the hydrogen generation rate of case 2-2
- **Case 2-1** – Baseline Coal Gasification Hydrogen Plant using GE Energy Radiant-Only Gasifier with CO_2 Capture and Sequestration and Hydrogen Separation by Pressure Swing Adsorption
- **Case 2-2** – Baseline Coal Gasification Hydrogen Plant using GE Energy Quench Gasifier with CO_2 Capture and Sequestration and Hydrogen Separation by Pressure Swing Adsorption

This report is part of a larger study that seeks to evaluate and compare a relatively large number of potential design configurations in a relatively short period of time. As such, the level of engineering effort expended in the production of these reports is commensurate with a conceptual design and is not sufficient for producing a preliminary design. The results should be viewed in this context. Performance and process limits are best estimates based upon published reports, information obtained from vendors, scaling of vendor information, and best engineering judgment.

Hydrogen cost is first determined by preparing plant designs for hydrogen production based on currently available process technology and meeting current permitting regulations for environmental compliance. To arrive at an estimated cost for producing hydrogen, the designs include commercially available process technology obtained from verifiable sources. While input was sought from various technology vendors, the final assessment of performance and cost was determined independently and may not represent the views of the technology vendors. Plants in this study were designed and estimated for 90 percent availability.

PERFORMANCE SUMMARY

Simplified process flow diagrams are presented in Exhibit ES-1 through Exhibit ES-3. More details including stream data are presented in Sections 3 through 5 in this report. Overall performance for each case is summarized in Exhibit ES-4.

Exhibit ES-1 Case 1-1 & 1-2 Block Flow Diagram: SMR with PSA & CO_2 Capture

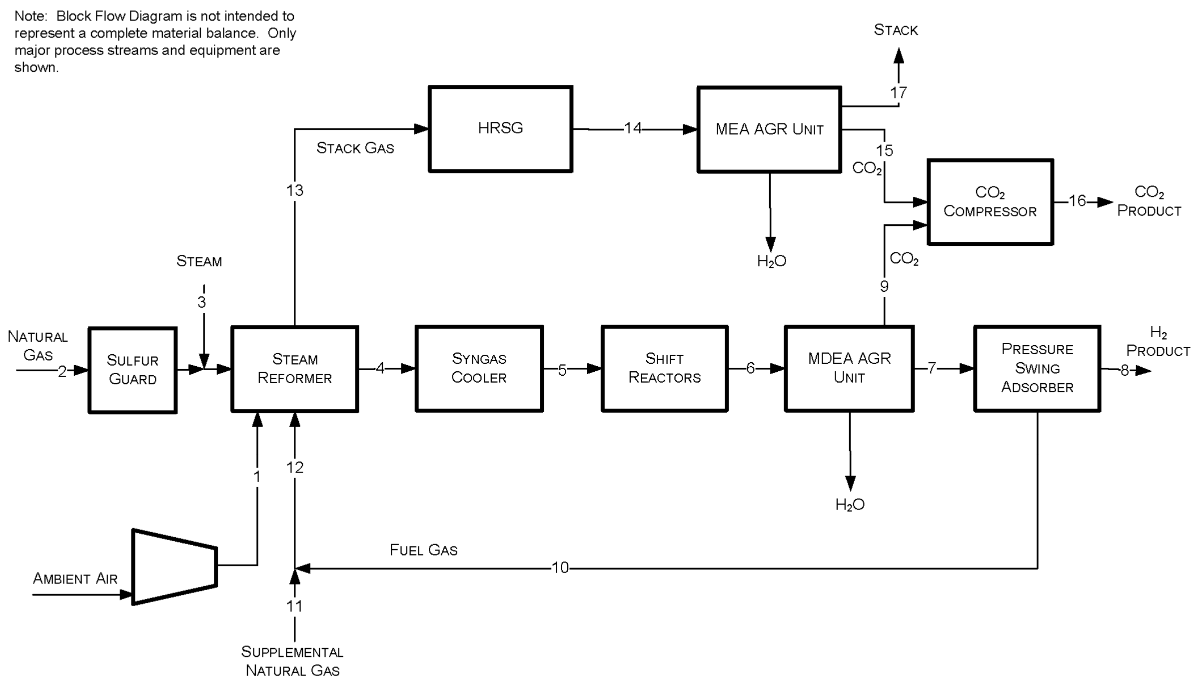


Exhibit ES-2 Case 2-1 Block Flow Diagram: Coal to Hydrogen with PSA & CO_2 Capture

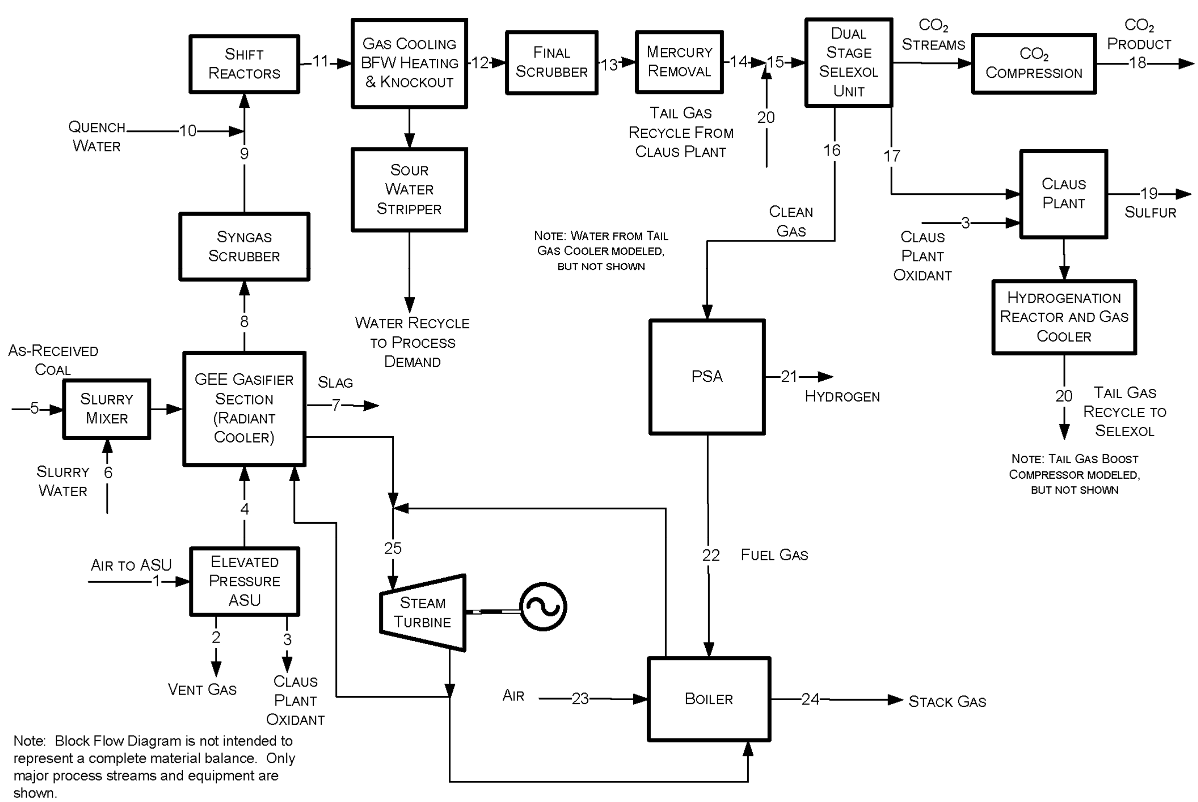


Exhibit ES-3 Case 2-2 Block Flow Diagram: Coal to Hydrogen with PSA & CO_2 Capture

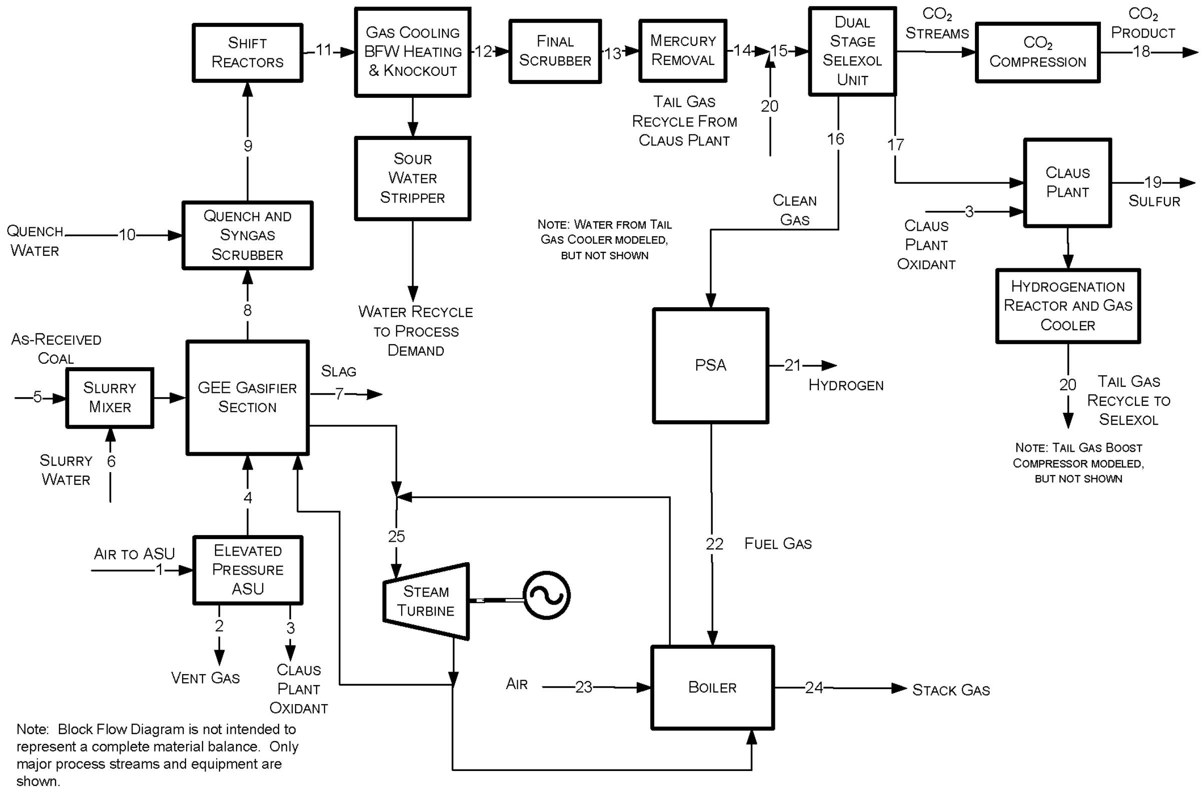


Exhibit ES-4 Overall Performance

Case	1-1	1-2	2-1	2-2
Plant Size, MMSCFD (kg/day) Hydrogen	242 (616,528)	243 (618,936)	242 (616,527)	243 (618,940)
Fuel	Natural Gas	Natural Gas	Illinois #6 Coal	Illinois #6 Coal
Natural Gas Feed to SMR, GJ/hr (MMBtu/hr)	4,500 (4,270)	4,520 (4,290)	N/A	N/A
Supplemental Natural Gas, GJ/hr (MMBtu/hr)	550 (520)	550 (520)	N/A	N/A
Coal Feed to Gasification, GJ/hr (MMBtu/hr)	N/A	N/A	5,994 (5,681)	5,994 (5,681)
Coal Feed to Gasification, tonne/day (ton/day)	N/A	N/A	5,302 (5,844)	5,301 (5,844)
Plant Availability	90%	90%	90%	90%
Cold Gas Efficiency[1], HHV	72.18%	72.17%	60.81%	61.05%
Effective Thermal Efficiency[2], HHV	69.74%	69.73%	61.24%	58.94%
CO_2 Recovered, tonne/day (ton/day)	5,456 (6,014)	5,478 (6,038)	10,954 (12,075)	10,951 (12,071)
CO_2 Emissions, tonne/day (ton/day)	606 (668)	609 (671)	1,183 (1,304)	1,192 (1,313)
Gross Power Generated, kWe	N/A	N/A	155,600	112,700
Auxiliary Power Consumed, kWe	34,200	34,330	148,440	147,830
Net Power, kWe	-34,200	-34,330	7,160	-35,130

[1] CGE = (Hydrogen Product Heating Value)/ Fuel Heating Value, HHV
[2] ETE = (Hydrogen + Power Heating Value)/ Fuel Heating Value, HHV

COST ESTIMATING SUMMARY

The cost estimates carry an accuracy of ±30 percent, consistent with the screening study level of engineering effort expended in the design. The results of the capital estimation calculations are shown in Exhibit ES-5. The value of this study lies not in the absolute accuracy of the individual case results but in the fact that all cases were evaluated under the same set of technical and economic assumptions. This consistency of approach allows meaningful comparisons among the cases evaluated. All capital and operations and maintenance (O&M) costs are presented as "overnight costs" expressed in June 2007 dollars. The cost estimation methodology is explained in more detail in Section 2.7 of "Cost and Performance Baseline for Fossil Energy Power Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity" [Ref. 1].

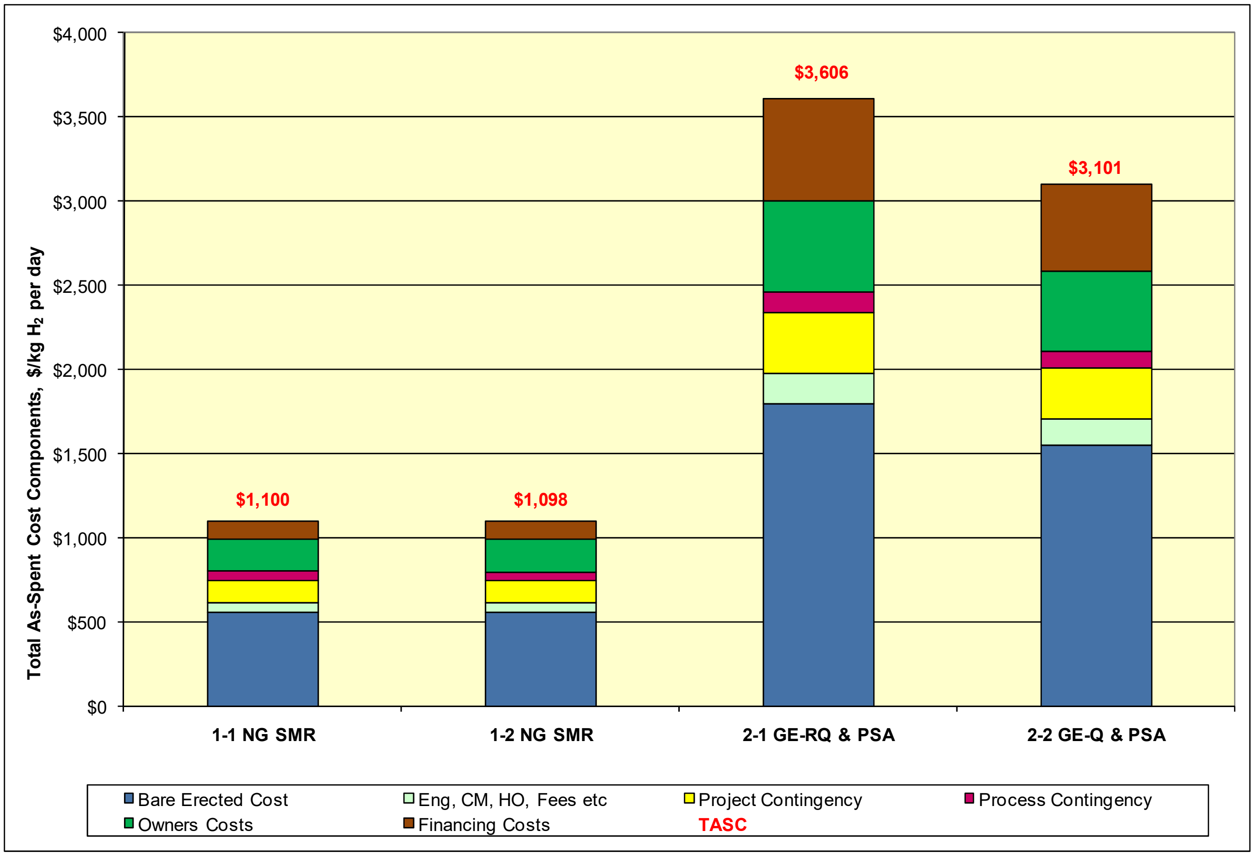
The bare erected costs (BEC) for the equipment were factored from cases 2 and 2A of the bituminous baseline study [Ref. 1] for all equipment that was included in those original cases. Additional equipment costs were obtained from other similar studies. The estimates were prepared by factoring the capital estimate on the basis of coal, gas, and stream flows and conditions.

The total plant cost (TPC) includes all equipment (complete with initial chemical and catalyst loadings), materials, labor (direct and indirect), engineering and construction management, and contingencies (process and project). The total overnight cost (TOC) for each plant was calculated by adding owner's costs to the TPC. Additional financing costs including escalation during construction were estimated and added to the TOC to provide the total as-spent cost (TASC). The TASC normalized on net hydrogen output is shown for each plant configuration in Exhibit ES-6. The coal to hydrogen cases are substantially more capital intensive than the SMR cases.

Exhibit ES-5 Capital Cost Estimation Results (June 2007 dollars)

Case	1-1	1-2	2-1	2-2
H_2 Production (kg H_2/day)	616,528	618,936	616,527	618,940
Bare Erected Cost, 1000$	$342,553	$343,355	$1,107,930	$958,576
Eng, CM, HO, Fees, etc., 1000$	$34,255	$34,335	$110,793	$95,858
Project Contingency, 1000$	$82,913	$83,107	$221,901	$190,503
Process Contingency, 1000$	$32,832	$32,909	$77,533	$61,446
Total Plant Cost, 1000$	**$492,553**	**$493,706**	**$1,518,158**	**$1,306,383**
Total Plant Cost, $/(kg H_2/day)	**$799**	**$798**	**$2,462**	**$2,111**
Owner's Cost, 1000$	$118,642	$118,926	$332,624	$291,118
Total Overnight Cost, 1000$	**$611,195**	**$612,632**	**$1,850,782**	**$1,597,501**
Total Overnight Cost, $/(kg H_2/day)	**$991**	**$990**	**$3,002**	**$2,581**
Financing Cost, 1000$	$66,905	$67,062	$372,543	$321,561
Total As-Spent Cost 1000$	**$678,100**	**$679,694**	**$2,223,325**	**$1,919,061**
Total As-Spent Cost, $/(kg H_2/day)	**$1,100**	**$1,098**	**$3,606**	**$3,101**

Exhibit ES-6 Total As-Spent Cost Components (June 2007 dollars)



Fixed and variable operating costs were estimated for each case. Baseline fuel costs for this analysis were determined using data from the Energy Information Administration's (EIA) Annual Energy Outlook (AEO) 2008. The costs used are $1.55/GJ ($1.64/MMBtu) for coal (Illinois No. 6) and $6.21/GJ ($6.55/MMBtu) for natural gas, both on a HHV basis and in 2007 United States (U.S.) dollars. All other consumable unit costs were also assumed to match those used in the baseline reference report [Ref. 1]. A value of $30/tonne of CO_2 emitted was also applied to reflect potential environmental regulations. A value of $105/MWh was applied for any excess power generated or additional power required for each case. These values are consistent with electricity generated in an environment where coal-based power plants are built with carbon capture and sequestration systems.

The first year costs of hydrogen (COH) were derived using the NETL Power Systems Financial Model (PSFM). COH is assumed to escalate at three percent per year for the thirty-year economic life of the plant. The project financial structure is representative of a high-risk fuels project with no loan guarantees or other government subsidies. The annual operating costs and COH values are shown in Exhibit ES-7.

Exhibit ES-7 First Year Cost of Hydrogen Estimation Results (June 2007 dollars)

Case	1-1	1-2	2-1	2-2
H_2 Production (kg H_2/day)	616,528	618,936	616,527	618,940
Fuel	Natural Gas	Natural Gas	Illinois #6	Illinois #6
Natural Gas Price ($/MMBtu)	$6.55	$6.55	N/A	N/A
Natural Gas Price ($/ton)	$298.47	$298.47	N/A	N/A
Natural Gas Consumption, tpd	2,520	2,530	N/A	N/A
Coal Price ($/MMBtu)	N/A	N/A	$1.64	$1.64
Coal Price ($/ton)	N/A	N/A	$38.19	$38.19
Coal Consumption, tpd	N/A	N/A	5,844	5,844
Capacity Factor, %	90%	90%	90%	90%
First Year Fuel Cost, $/yr	$247,114,305	$248,102,221	$73,307,753	$73,302,484
First Year Fixed O&M Cost, $/yr	$22,668,479	$22,703,075	$53,436,617	$47,083,365
First Year Variable O&M Cost, $/yr	$14,937,300	$14,978,807	$40,717,934	$36,216,846
First Year Electricity Cost (Revenue), $/yr	$28,311,444	$28,419,061	($5,927,191)	$29,081,317
First Year Carbon Emissions Value, $/yr	$5,974,186	$5,998,070	$11,661,501	$11,742,952
First Year Capital, $/yr	$124,920,703	$125,214,495	$459,987,842	$397,140,235
First Year COH, $/kg H_2	**2.19**	**2.19**	**3.13**	**2.92**
First year COH, $/1000scf H_2	**5.59**	**5.58**	**7.97**	**7.45**
First Year CO_2 TS&M, $/yr	$23,520,371	$23,675,800	$41,344,396	$41,349,212
First year COH including CO_2 TS&M, $/kg H_2	**2.31**	**2.31**	**3.33**	**3.13**
First year COH including CO_2 TS&M, $/1000scf H_2	**5.88**	**5.88**	**8.49**	**7.97**

The first year COH results are shown graphically in Exhibit ES-8 with the capital cost, fixed operating cost, variable operating cost, and fuel cost components shown separately. CO_2 transport, storage, and monitoring (TS&M) costs are also shown as a separate bar segment. The following conclusions can be drawn:

- The COH is dominated by capital charges in both of the coal cases. The capital cost component of COH comprises 62-68 percent in the coal cases but only 27 percent in the natural gas SMR cases.
- The fuel cost component is relatively minor in the coal cases, representing 11-12 percent of the COH, but it dominates the natural gas SMR cases at 53 percent.
- The excess power generated in case 2-1 reduces the variable O&M for that case by 10 percent.
- The TS&M component of COH in all cases is 5-7 percent.

Exhibit ES-8 First Year COH by Cost Component (June 2007 dollars)

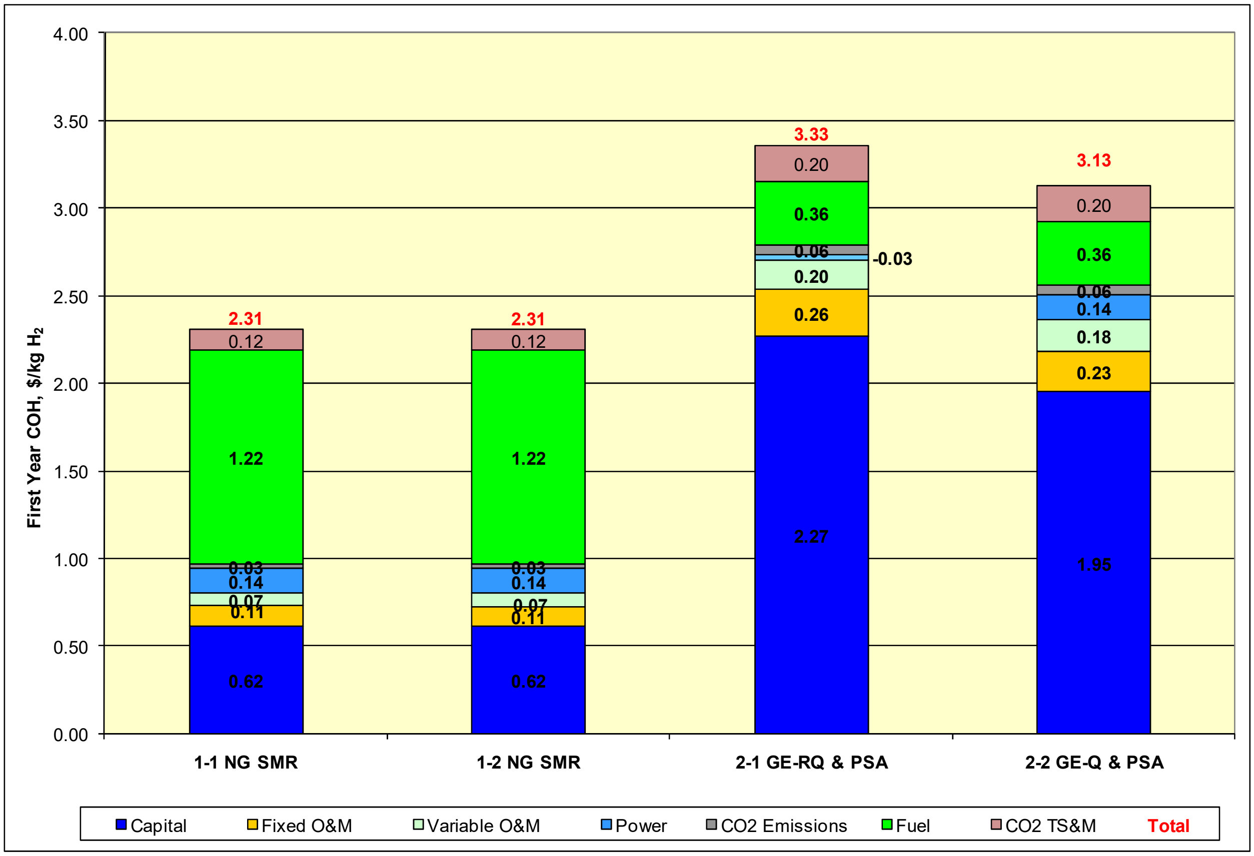
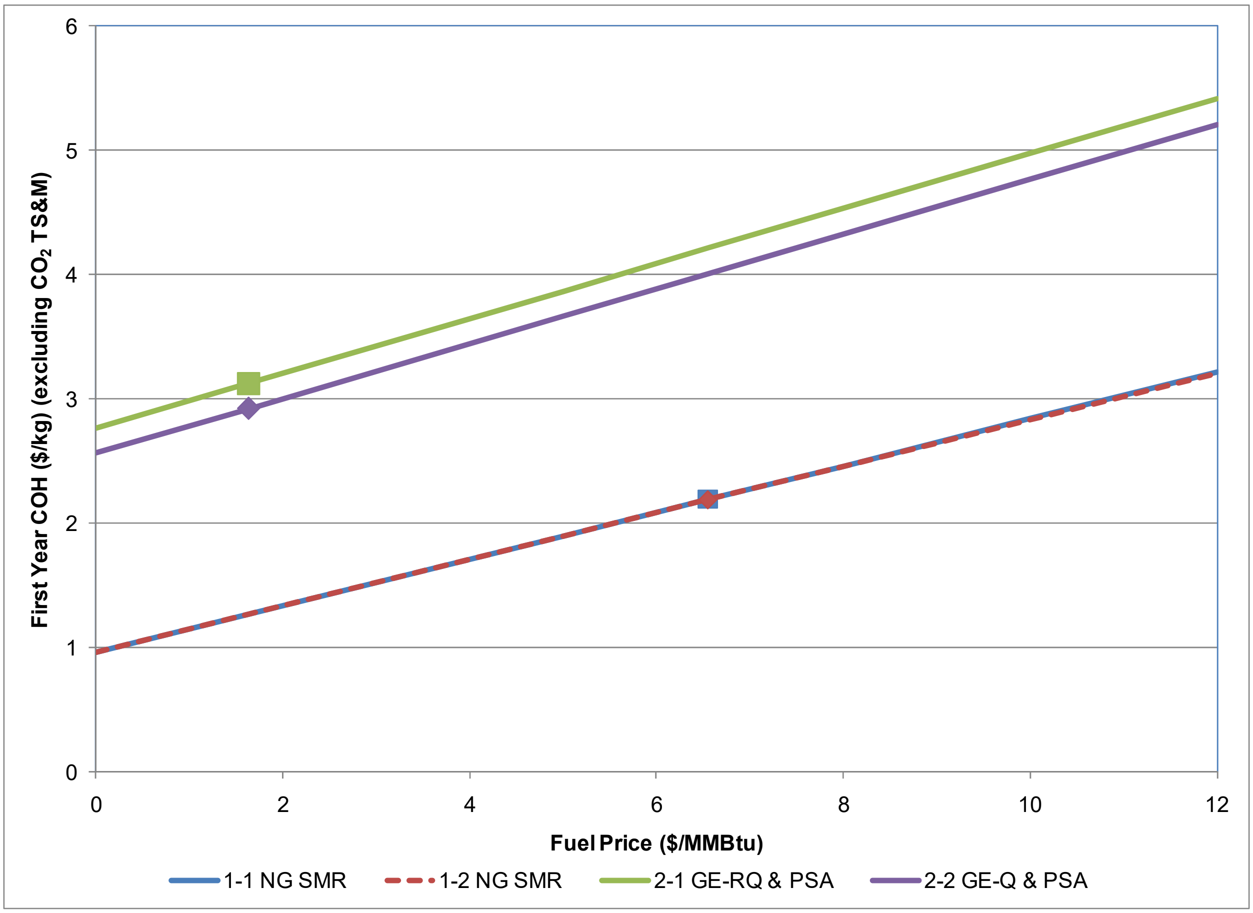


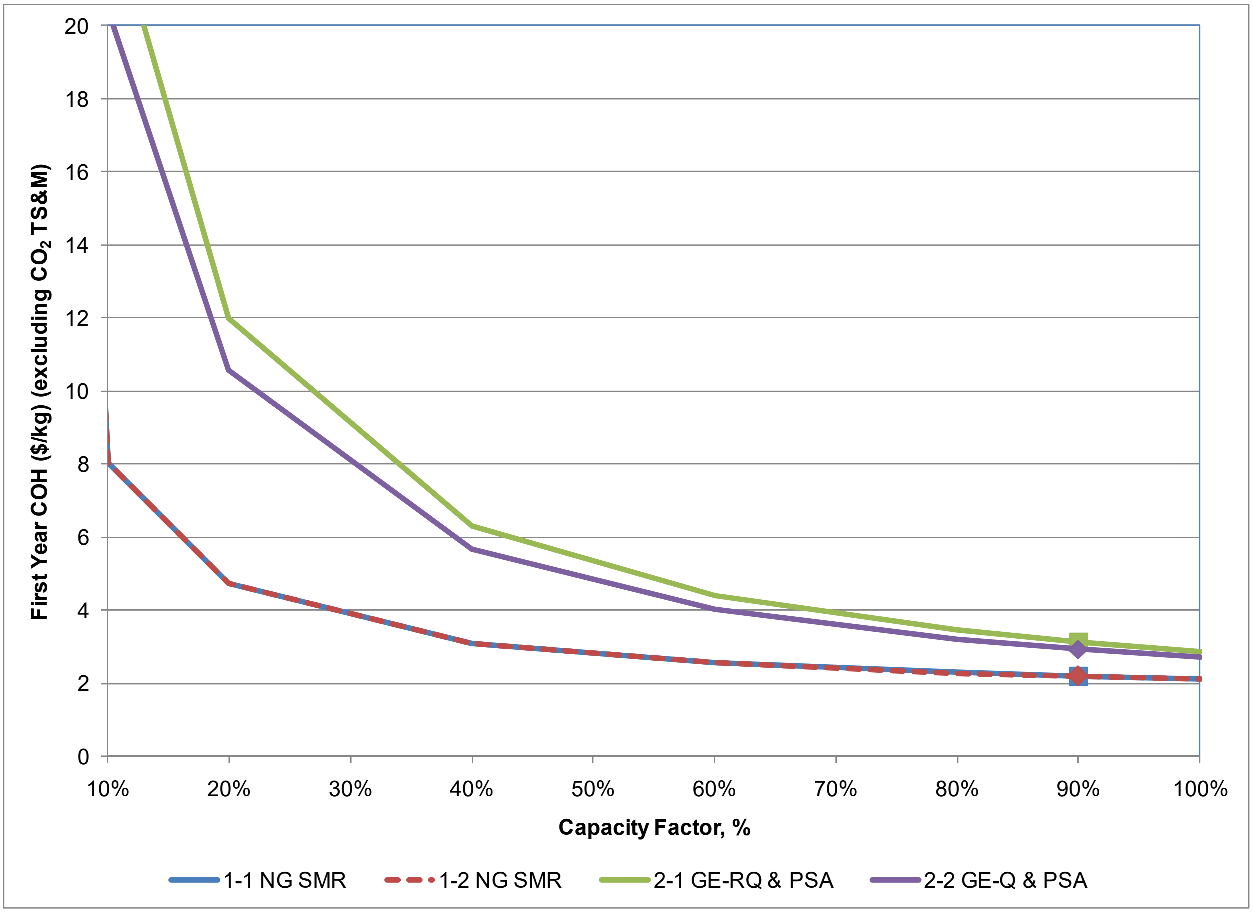
Exhibit ES-9 shows the first year COH sensitivity to fuel costs. Again the COH values calculated for the prices of natural gas and coal assumed in this study are shown on each line. As expected, all cases show a linear increase in COH with an increase in fuel prices. The COHs for the natural gas SMR cases increase by approximately 19 cents for each \$/MMBtu increase in natural gas price. The COHs for the coal cases increase by approximately 22 cents for each \$/MMBtu increase in coal prices. In general, the values for the SMR cases approach the values for the coal to hydrogen cases as natural gas prices increase and coal prices decrease.

Exhibit ES-9 First Year COH Sensitivity to Fuel Costs (June 2007 dollars)



The sensitivity of first year COH to capacity factor is shown in Exhibit ES-10. Again the COH values calculated for the capacity factor assumed in this study are shown on each line. At high capacity factors, the COH value for the coal cases approaches the COH for the natural gas cases. All cases show a substantial decrease in COH as the capacity factor increases. At very lower capacity factors (10 to 20 percent), the COH for the natural gas cases are less than one-half that of the two coal cases.

Exhibit ES-10 First Year COH Sensitivity to Capacity Factor (June 2007 dollars)



The first year COH for coal to hydrogen with CO_2 capture cases is estimated to be approximately \$1.08 to \$1.35 per kg of hydrogen greater than the COH for hydrogen generated from natural gas by the current commercial SMR technology with the addition of CO_2 capture. As gasification and CO_2 capture technologies become more commercially available, this differential should decrease and the coal to hydrogen cases would likely become more economically viable. Natural gas prices above \$9.60/MMBtu would also make the coal to hydrogen cases more competitively attractive.

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1. INTRODUCTION

This report presents the design configuration and performance summaries for four baseline cases in the Advanced Hydrogen Production Plant study. This report is part of a larger study that seeks to evaluate and compare a relatively large number of potential design configurations in a relatively short period of time. As such, the level of engineering effort expended in the production of these reports is commensurate with a conceptual design and is not sufficient for producing a preliminary design. The results should be viewed in this context. Performance and process limits are best estimates based upon published reports, information obtained from vendors, scaling of vendor information, and best engineering judgment.

Objective: The objective of this study is to estimate the performance and cost of fossil-based hydrogen production from both baseline and advanced systems. Baseline systems will utilize state-of-the-art technology available in 2015 and advanced systems will feature hydrogen separation membranes and the calcium looping process, technologies currently under development for future plants.

1.1 SITE DESCRIPTION

All plants in this study are assumed to be located at a generic plant site in Midwestern USA, with ambient conditions and site characteristics as presented in Exhibit 1-1 and Exhibit 1-2.

Exhibit 1-1
Site Ambient Conditions

Elevation, ft	0
Barometric Pressure, psia	14.696
Design Ambient Temperature, Dry Bulb, °F	59
Design Ambient Temperature, Wet Bulb, °F	51.5
Design Ambient Relative Humidity, %	60

Exhibit 1-2
Site Characteristics

Location	Green-field, Midwestern USA
Topography	Level
Size, acres	300
Transportation	Rail
Ash Disposal	Off Site
Water	Municipal (50%) / Groundwater (50%)
Access	Land locked, having access by railway and highway

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

- Flood plain considerations
- Existing soil/site conditions
- Water discharges and reuse
- Rainfall/snowfall criteria
- Seismic design
- Buildings/enclosures
- Fire protection
- Local code height requirements
- Noise regulations – Impact on site and surrounding area

1.2 PRODUCT SPECIFICATIONS

Elemental Sulfur – 99 percent sulfur

Hydrogen – 99.9 percent hydrogen

The baseline plant designs produced hydrogen with a purity of 99.9 percent (by volume), minimum delivery pressure of 300 psig utilizing pressure swing adsorption (PSA) technology. The PSA performance is based on POLYBED ten-bed unit by UOP, LLC (a Honeywell company).

The hydrogen separation performance for all cases is presented in Exhibit 1-3.

Exhibit 1-3 Hydrogen Product Specification

Hydrogen Recovery	80%
Hydrogen purity	$\geq$99.9%
Max. CO	5 ppm
Max. H_2S	5 ppb
Max H_2O	1 ppb

Carbon Dioxide Sequestration

The study assumed that sequestration-ready CO_2 is transported to the plant boundary as a supercritical fluid. The CO_2 and pipeline requirements are presented in Exhibit 1-4.

Exhibit 1-4 Carbon Dioxide Pipeline Specification

Inlet Pressure	~15.3 MPa (~2,200 psig)
Water Content	-40°C (-40°F) dew point
N_2	< 300 ppmv
O_2	< 40 ppmv
Ar	< 10 ppmv

1.3 DESIGN NATURAL GAS

DOE's Hydrogen Advisory Group (H2A) previously utilized a GRI survey (U.S. Natural Gas Composition Based on 26-City Survey: Gas Research Institute Report GRI-92/0123) to determine the average composition of natural gas as a feedstock for SMR hydrogen plants. The composition that was used in the current analysis is the National Average, shown in Exhibit 1-5.

Exhibit 1-5
Design Natural Gas Analysis

Volume share	National Average
Methane	93.90%
Ethane	3.20%
Propane	0.70%
C4+	0.40%
CO_2+N_2	2.60%
Water	85-105ppmv
Sulfur	6ppmv

1.4 DESIGN COAL

The design coal assumed for this study is presented in Exhibit 1-6. All coal-fired cases will be modeled with Illinois No. 6 coal.

Exhibit 1-6
Design Coal Analysis – Illinois No. 6

Rank	**Bituminous**	
Seam	**Illinois No. 6 (Herrin)**	
Source	**Old Ben Mine**	
Proximate Analysis (weight %) (Note 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	27,113	30,506
HHV, Btu/lb	11,666	13,126
LHV, kJ/kg	26,151	29,544
LHV, Btu/lb	11,252	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	0.29	0.33
Sulfur	2.51	2.82
Ash	9.70	10.91
Oxygen (Note B)	6.88	7.75
Total	100.00	100.00

Notes: A. The proximate analysis assumes sulfur as volatile matter
B. By difference

1.5 PLANT CAPACITY FACTOR AND AVAILABILITY

The overall availability of the natural gas case operating plant was assumed to be 90 percent, which is consistent with commercial SMR plants. The balance of plant will be single train, operating at 100 percent capacity, based on commercial process operating experience as verified by equipment vendors.

The goal of the designs for the two coal cases was to achieve an overall availability for the operating plant of 90 percent. This is a high factor for single train gasification and will result in the requirement for two gasifier trains operating at full coal throughput along with a hot standby gasifier train with the capability to ramp up and maintain 100 percent hydrogen production. The balance of plant will be single train, operating at 100 percent capacity, based on commercial process operating experience as verified by equipment vendors.

1.6 ENVIRONMENTAL REQUIREMENTS

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 probably 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 will be used for the plant modeled in this evaluation. The production of hydrogen from steam methane reforming is inherently emissions free, with the exception of NOx from the SMR burner. Low NOx burners will be included to minimize the NOx emissions and 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 which 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 auxiliary boilers will be utilized. BACT control technologies and emission limits are summarized in Exhibit 1-7 and Exhibit 1-8.

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 for this screening.
- 90 percent removal of carbon in design fuel.

Exhibit 1-7 BACT Environmental Design Basis for Natural Gas Cases

Pollutant	Environmental Design Basis	
	Control Technology	Limit
Sulfur Oxides (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 1-8 BACT Environmental Design Basis for Coal Cases

Pollutant	Environmental Design Basis	
	Control Technology	Limit
Sulfur Oxides (SO_2)	Selexol + 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

1.7 PROCESS DESIGN CRITERIA

The original design basis for case 1-1 was to size the plant as was done previously in the H2A model to produce nominally 150 million SCFD hydrogen (381,000 kg/day). This is considered to be a world-class plant scale and can be achieved with a single train steam-methane reformer. The design basis, shown in Exhibit 1-9, was modified to increase the capacity of the plant to approximately match the hydrogen output of cases 2-1 and 2-2 (coal to hydrogen plants) at 25,740 kg/hr (56,750 lb/hr) or 240 million SCFD.

The design basis for the coal gasification cases is similar to case 2 and 2A in the NETL report entitled "Cost and Performance Baseline for Fossil Energy Plants," [Ref. 1] and is shown in Exhibit 1-10. The baseline design produces the maximum amount of hydrogen with CO_2 capture for sequestration from 5,845 tons of coal per day at 100 percent capacity. The plant is based on the General Electric Energy (GEE) gasification technology operating at approximately 965 psia.

Exhibit 1-9 Design Criteria for Conventional SMR Hydrogen Production Plant

Parameter	Design Basis
Plant Size	~240 MMSCFD (~56,750 lb/hr) 99.9 % purity hydrogen
Hydrogen Pressure	>300 psig at plant gate
Plant Capacity Factor	90 %
Ambient Conditions	14.7 psia, 60°F
Natural Gas Feed	Pipeline, 450 psia
Desulfurization	Zinc oxide guard bed for natural gas feed to reformer
Reformer	Vertical tube steam methane reformer, externally heated
Water Gas Shift	High-temperature, 98 % conversion
Syngas CO_2 Recovery	Coastal, proprietary MDEA, 95 % removal
Stack gas CO_2 Recovery	Fluor Econamine, proprietary MEA, achieve 90 % total
Hydrogen Purification	Pressure Swing Adsorption
PSA Retentate Gas	Recycled to reformer as fuel
CO_2 Product Pressure	2,215 psia

Exhibit 1-10 Design Criteria for Coal to Hydrogen Production Plants

Parameter	Design Basis
Ambient Conditions	14.7 psia, 60°F
Coal Feed	Illinois No. 6
Gasifier	Oxygen-blown GE Energy
Plant Size	Maximum hydrogen production from ~5,845 tpd coal feed
Hot Gas Temperature	~2,500°F
Gasifier Outlet Pressure	~ 965 psia
Gas Quench/Cooling	~450°F
Water Gas Shift	High-temperature, sulfur-tolerant
Mercury Removal	Carbon Bed
Desulfurization	Selexol
Sulfur Recovery	Elemental sulfur
CO_2 Recovery	Selexol
Hydrogen Purification	Pressure Swing Adsorption
PSA or Membrane Retentate Gas	Fired in auxiliary boiler
CO_2 Product Pressure	2,200 psia
Hydrogen Production	770 psia at plant gate
Auxiliary Power Block	Steam turbine generator
Plant Capacity Factor	90 %

1.8 BALANCE OF PLANT

Assumed balance of plant requirements are as follows:

Cooling system	Recirculating, Evaporative Cooling Tower
Fuel and Other storage	
Natural Gas	On-site pipeline
Coal	30 days
Slag	30 days
Sulfur	30 days
Plant Distribution Voltage	
Motors below 1 hp	110/220 volt
Motors 250 hp and below	480 volt
Motors above 250 hp	4,160 volt
Motors above 5,000 hp	13,800 volt
Steam and Gas Turbine generators	24,000 volt
Grid Interconnection voltage	345 kV
Water and Waste	
Makeup Water	The water supply is assumed to be 50 % from a local Publicly Owned Treatment Works (POTW) and 50 % from groundwater and is assumed to be in sufficient quantities to meet plant makeup requirements. Makeup for potable, process, and de-ionized (DI) water will be drawn from municipal sources.
Feed water	The quality of feedwater (i.e., water treatment systems) required is assumed to be similar regardless of the technology.
Process Wastewater	Water associated with process activity and storm water that contacts equipment surfaces will be collected and treated for discharge through a permitted discharge permit.
Sanitary Waste Disposal	Design will include a packaged domestic sewage treatment plant with effluent discharged to the industrial wastewater treatment system. Sludge will be hauled off site.
Water Discharge	Most of the wastewater is to be recycled for plant needs. Blowdown will be treated for chloride and metals, and discharged.
Solid Waste	Fly ash, bottom ash, scrubber sludge, and gasifier slag are assumed to be solid wastes that are classified as non-hazardous wastes. Offsite waste disposal sites are assumed to have the capacity to accept waste generated throughout the life of the facility. Solid wastes sent to disposal are at an assumed nominal fee per ton, even if the waste is hauled back to the mine. Solid waste generated that can be recycled or reused is assumed at a zero cost to the technology.

2. COST ESTIMATING METHODOLOGY

The estimates carry an accuracy of ±30 percent, consistent with the screening study level of engineering effort expended in the design. Capital and O&M costs are presented as "overnight costs" expressed in June 2007 dollars. The cost estimation methodology is explained in more detail in Section 2.7 of "Cost and Performance Baseline for Fossil Energy Power Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity" [Ref. 1].

The capital costs for the equipment were factored from cases 2 and 2A of the bituminous baseline study [Ref. 1] for all equipment that was included in that original case. Additional equipment costs were obtained from other similar studies. The estimates were prepared by factoring the capital estimate on the basis of coal, gas, and steam flows and conditions.

Bare erected capital costs (BEC) include:

- Equipment (complete with initial chemical and catalyst loadings)
- Materials
- Labor (direct and indirect)

Project contingencies were added to the BEC and Engineering, Construction Management, Home Office & Fees (Eng'g CM, H.O. & Fee) costs to cover project uncertainty and the cost of additional equipment that could result from a more detailed design. The project contingencies represent costs that are expected to occur. Each capital account was evaluated against the level of estimate detail, field experience, and the basis for the equipment pricing to define project contingency. Process contingencies were added to compensate for uncertainty in cost estimates caused by performance uncertainties associated with the development status of technologies for the gasification and CO_2 removal systems. Contingency values were applied based on recommendations in "Quality Guidelines for Energy System Studies (QGESS)" [Ref. 2] and values used in the baseline cost and performance report [Ref. 1]. The percentages assessed for both types of contingency vary between accounts and cases.

All the capital costs are then summed to calculate the total plant cost (TPC). Owner's costs were subsequently calculated and added to the TPC, the result of which is total overnight cost (TOC). Additionally, financing costs were estimated by applying a factor to the TOC value to calculate total as-spent costs (TASC). The first year cost of hydrogen production (COH) was calculated using TOC.

Fixed and variable operating costs were estimated for each case. Baseline fuel costs for this analysis were determined using data from the Energy Information Administration's (EIA) Annual Energy Outlook (AEO) 2008. The costs used are $1.55/GJ ($1.64/MMBtu) for coal (Illinois No. 6) and $6.21/GJ ($6.55/MMBtu) for natural gas, both on a HHV basis and in 2007 United States (U.S.) dollars. All other consumable unit costs were assumed to match those used in the baseline reference report [Ref. 1]. A value of $30/tonne of CO_2 emitted was also applied to reflect potential environmental regulations. A value of $105/MWh was applied for any excess power generated or required for each case. These values are consistent with electricity generated in an environment where coal-based power plants are built with carbon capture and sequestration systems.

The revenue requirement method of performing an economic analysis of a prospective energy plant has been widely used in the electric utility industry. This method permits the incorporation of the various dissimilar components for a potential new plant into a single value that can be compared to various alternatives. The revenue requirement figure-of-merit in this report is the cost of hydrogen (COH) expressed in $/kg. The first year COHs were calculated by the Power Systems Financial Model [Ref. 3] using the financial assumptions specified in Exhibit 2-1 and Exhibit 2-2 [Ref. 4]. The first year COH is estimated to be the value calculated when the required return on equity (ROE) equals the internal rate of return (IRR) for 30 years of operation based on the assumed financial structure and escalations. COH is assumed to escalate at three percent per year for the thirty-year economic life of the plant.

All costs are expressed in June 2007 year dollars. In this study the first year of plant construction is assumed to be 2010. A three year capital expenditure/construction period is assumed for the natural gas based case plants with startup in 2013. For the coal based case plants, a five year capital expenditure/construction period is assumed with startup in 2015. The five-year period is assumed to include at least one year prior to the start of construction, during which capital costs associated with items such as detailed design, permitting, and long-lead equipment orders might be incurred. A five-year capital expenditure period is appropriate for more complex projects, including many coal-based energy projects, while the shorter, three-year capital expenditure period may be representative for simpler projects such as the natural gas based plant.

The capital and operating costs for CO_2 transport, storage, and monitoring (TS&M) were independently estimated by NETL. Those costs were combined with the plant capital and operating costs to produce an overall COH. The TS&M cost estimation methodology is also explained in more detail in Section 2.7 of "Cost and Performance Baseline for Fossil Energy Power Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity" [Ref. 1].

Exhibit 2-1 Financial Structure for High-Risk Fuels Projects

Type of Security	% of Total	Current (Nominal) Dollar Cost	Weighted Current (Nominal) Cost	After Tax Weighted Cost of Capital
Debt	50	9.5% (LIBOR plus 6%)	4.75%	
Equity	50	20%	10.0%	
Total			14.75%	12.945%

Exhibit 2-2 Parameter Assumptions for Cost of Hydrogen Calculations

Parameter	Value
TAXES	
Income Tax Rate	38% (Effective 34% Federal, 6% State)
Capital Depreciation	20 years, 150% declining balance
Investment Tax Credit	0%
Tax Holiday	0 years
FINANCING TERMS	
Repayment Term of Debt	15 years
Grace Period on Debt Repayment	0 years
Debt Reserve Fund	None
TREATMENT OF CAPITAL COSTS	
Capital Cost Escalation During Construction (nominal annual rate)	3.6%[1]
Distribution of Total Overnight Capital over the Capital Expenditure Period (before escalation)	3-Year Period: 10%, 60%,30% 5-Year Period: 10%, 30%, 25%, 20%, 15%
Working Capital	zero for all parameters
% of Total Overnight Capital that is Depreciated	100% *(this assumption introduces a very small error even if a substantial amount of TOC is actually non-depreciable)*
INFLATION	
COH, O&M, Fuel Escalation (nominal annual rate)	3.0%[2] COH, O&M, Fuel

[1] A nominal average annual rate of 3.6% is assumed for escalation of capital costs during construction. This rate is equivalent to the nominal average annual escalation rate for process plant construction costs between 1941 and 2008 according to the *Chemical Engineering* Plant Cost Index.

[2] An average annual inflation rate of 3.0% is assumed. This rate is equivalent to the average annual escalation rate between 1947 and 2008 for the U.S. Department of Labor's Producer Price Index for Finished Goods, the so-called "headline" index of the various Producer Price Indices. (The Producer Price Index for the Electric Power Generation Industry may be more applicable, but that data does not provide a long-term historical perspective since it only dates back to December 2003.)

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3. PROCESS DESCRIPTIONS

The hydrogen production cases described here are based on a combination of commercially-proven and developmental processes. For cases 1-1 and 1-2, steam reforming of methane continues to be the most widely used process for the production of hydrogen and hydrogen/carbon monoxide mixtures. The process involves catalytic conversion of hydrocarbons and steam to produce hydrogen and carbon oxides. Since the process works only with light hydrocarbons which can be vaporized completely without carbon deposition, the feedstocks used can range from methane (natural gas) to naphtha to No. 2 fuel oil.

For the coal to hydrogen processes, cases 2-1 and 2-2 are configured with two gasifier trains and a hot standby spare train. The GEE single-stage coal gasification technology features an oxygen-blown, entrained flow, refractory lined gasifier with continuous slag removal. Coal/water slurry reacts with oxygen at about 2,500 ºF and 965 psia. A turnkey, dedicated air separation unit (ASU) supplies oxygen at 95 percent purity to the gasifier. The gasifier trains include processes to progressively cool and clean the gas, making it suitable for hydrogen production.

Individual process components for all the cases are described below.

3.1 STEAM METHANE REFORMING – CASES 1-1 AND 1-2

Natural Gas Conditioning

Natural gas is fed to the plant from the pipeline at a pressure of 450 psia. To protect the catalysts in the hydrogen plant, the natural gas has to be desulfurized before being fed to the reformer. The gas is generally sulfur-free but mercaptan-based odorizers must be cleaned from the gas to prevent contamination of the reformer catalyst. This is accomplished with a zinc oxide polishing bed also known as a sulfur guard.

Natural Gas Reformer/Boiler

The desulfurized natural gas feedstock is mixed with process steam and reacted over a nickel-based catalyst contained inside a system of high alloy steel tubes. The following reactions take place in the reformer:

$$CH_4 + H_2O = CO + 3H_2$$

$$CO + H_2O = CO_2 + H_2$$

$$CO + 3H_2 = CH_4 + H_2O$$

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 hydrogen purification process plus supplemental natural gas as needed. The metallurgy of the tubes usually limits the reaction temperature to 1,400-1,700ºF. The flue gas path of the fired reformer is integrated with additional boiler surfaces to produce about 1,428,000 lb/hr steam. About 655,000 lb/hr of this steam is superheated to 450 psia and 750°F, to be added to the incoming natural gas.

Additional steam from the boiler is used within the plant for regeneration of CO_2 solvent from the acid gas removal (AGR) processes. The reformer burner uses a low-NO_x design to limit NO_x emissions to 20 ppm, very low for a gas-fired boiler. This consists of burning predominantly PSA purge gas along with supplemental natural gas with air at ambient temperature. The use of SNCR or SCR for NO_x reduction is not required with this plant design.

The CO-shift and methanation reactions quickly reach 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 generally employs a steam to carbon ratio of 3 to 5 at a process temperature of about 1,500ºF and pressures up to 500 psig to convert more than 70 percent of hydrocarbons to oxides of carbon at the outlet of the reformer so as to ensure a minimum concentration of CH_4 in the product gas. The typical composition of the synthesis gas at 450 psia leaving a steam-methane reformer is shown in Exhibit 3-1.

Exhibit 3-1 Typical Composition of the Synthesis Gas Leaving a Steam-Methane Reformer

Component	Volume %
CH_4	2
CO	7
CO_2	6
H_2	44
H_2O	41
Total	100

Leaving the reformer, the process gas mixture of CO and H_2 passes through a heat recovery step and is fed into a water gas shift (WGS) reactor to produce additional H_2.

Water Gas Shift Reactor

For the conversion of the reformer gas to hydrogen, the first step is to convert most of the carbon monoxide (CO) to hydrogen and carbon dioxide (CO_2) by reacting the CO with steam over a bed containing iron-based catalysts which promote the WGS reaction. This increases the balance of the gross hydrogen product by converting approximately 98 percent of the carbon monoxide to hydrogen and CO_2. The product stream from the reformer contains sufficient amounts of water vapor to meet the necessary water to gas criteria at the shift reactor inlet. The CO shift converter consists of four fixed-bed reactors with two reactors in series and two in parallel. Two reactors in series with cooling between the two are required to control the exothermic temperature rise. The two reactors in parallel are required due to the high gas mass flow rate.

Effluent from the second stage is cooled by exchanging heat with incoming feed, by an air cooled exchanger, and finally by a water-cooled exchanger. The exit gas is predominantly hydrogen and CO_2 with some residual CO and methane.

Acid Gas Removal - Shifted Syngas

With conventional production of hydrogen from natural gas, CO_2 is normally not recovered from the syngas stream and the excess steam generated in the boiler is exported off site. However, the case 1-1 plant designed to capture CO_2 utilizes a proprietary amine-based process to remove and recover 95 percent of the CO_2 from the syngas stream. The CO_2 is removed by chemical absorption with a highly selective, hybrid amine. From the shift reactor, gas is passed through an amine tower where it is contacted counter-currently with a circulating stream of lean aqueous amine solution. CO_2 in the feed averages approximately 12 mole percent and 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 this plant. 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.

<u>Acid Gas Removal - Stack Gas</u>

If the CO_2 were only captured from the shifted syngas stream, the overall CO_2 recovery would be about 65 percent. To increase the overall carbon recovery to 90 percent, a second CO_2 removal process is utilized in the reformer heater stack to remove CO_2 resulting from reformer heater combustion.

The CO_2 recovery process for the stack is based on the Fluor Econamine FG Plus technology [Ref. 5]. The Econamine FG Plus process uses a formulation of monoethanolamine (MEA) and a proprietary oxidation inhibitor to recover CO_2 from the flue gas. This process is designed to recover high-purity CO_2 from low-pressure streams that contain oxygen, such as flue gas from coal-fired power plants, gas turbine exhaust gas, and other waste gases.

A fraction of the flue gas exiting the HRSG from the reformer heater enters the bottom of the CO_2 Absorber and flows up the tower countercurrent to a stream of lean monoethanolamine (MEA)-based solvent (Econamine FG Plus). This results in approximately 70 percent of the CO_2 in the stack gas being absorbed into the lean solvent, and the remaining gas leaves the top of the absorber section and flows into the water wash section of the tower. This extraction, combined with CO_2 removed from the shifted syngas stream, results in an overall CO_2 capture of 90 percent. The lean solvent enters the top of the absorber, absorbs the CO_2 from the flue gases, and leaves the bottom of the absorber with the absorbed CO_2. The purpose of the water wash section is to minimize solvent losses due to mechanical entrainment and evaporation.

A solvent stripper is used to separate the CO_2 from the rich solvent feed exiting the bottom of the CO_2 absorber. The rich solvent is collected on a chimney tray below the bottom packed section of the solvent stripper and routed to the solvent stripper reboiler where the rich solvent is heated by steam, stripping the CO_2 from the solution. The uncondensed CO_2-rich gas is then delivered to the CO_2 product compressor. The condensed liquid from the solvent stripper reflux drum is pumped via the solvent stripper reflux pumps where a portion of condensed overhead liquid is used as make-up water for the water wash section of the CO_2 absorber. The rest of the pumped liquid is routed back to the solvent stripper as reflux, which aids in limiting the amount of solvent vapors entering the stripper overhead system.

A small slipstream of the lean solvent from the solvent stripper bottoms is fed to the solvent stripper reclaimer for the removal of high-boiling nonvolatile impurities (heat stable salts - HSS), volatile acids and iron products from the circulating solvent solution. The solvent bound in the HSS is recovered by reaction with caustic and heating with steam. The solvent reclaimer system reduces corrosion, foaming, and fouling in the solvent system.

3.2 COAL GASIFICATION – CASES 2-1 AND 2-2

Coal Grinding and Slurry Preparation

Coal is fed onto a conveyor by vibratory feeders located below each coal silo. The conveyor feeds the coal to an inclined conveyor that delivers the coal to the rod mill feed hopper. The feed hopper provides a surge capacity of about two hours and contains two hopper outlets. A vibrating feeder on each hopper outlet supplies the weigh feeder, which in turn feeds a rod mill. Each rod mill is sized to process 60 percent of the coal feed requirements of the gasifier. The rod mill grinds the coal and wets it with treated slurry water transferred from the slurry water tank by the slurry water pumps. The coal slurry is discharged into the rod mill product tank, and then the slurry is pumped from the rod mill product tank to the slurry storage and slurry blending tanks.

The coal grinding system is equipped with a dust suppression system consisting of water sprays aided by a wetting agent. The degree of dust suppression required depends on local environmental regulations. All of the tanks are equipped with vertical agitators to keep the coal slurry solids suspended.

Air Separation Unit

The air separation unit (ASU) is designed to produce a nominal output of 6,000 tpd of 95 percent pure O_2 for use in the gasifier for both of the gasification cases. The designs also include the generation of additional O_2 for the Claus plants, thermal oxidizers, and calcium looping process as needed for specific cases. The ASU is designed with two production trains. The air compressors are powered by an electric motor.

Gasification

This plant utilizes two gasification trains to process a total of about 6,000 tpd of Illinois No. 6 coal. Each of the 2 x 50% gasifiers operates at nearly maximum capacity. To achieve 90 percent availability, each plant is configured with a third spare gasifier train (gasifier, radiant cooler, and/or quench) on hot standby. The slurry feed pump takes suction from the slurry run tank, and the discharge is sent to the feed injector of the GEE gasifier. Oxygen from the ASU is vented during preparation for startup and is sent to the feed injector during normal operation. The air separation plant supplies about 6,000 tpd of 95 percent purity oxygen to the gasifiers and the Claus plant.

The gasifier vessel is a refractory-lined, high-pressure combustion chamber. Coal slurry is transferred from the slurry storage tank to the gasifier with a high-pressure pump. At the top of the gasifier vessel a combination fuel injector is located through which coal slurry feedstock and oxidant (oxygen) are fed. The coal slurry and the oxygen feeds react in the gasifier at about 965 psia at a high temperature (in excess of 2,500 °F) to produce syngas.

The syngas consists primarily of hydrogen and carbon monoxide, with lesser amounts of water vapor and carbon dioxide, and small amounts of hydrogen sulfide, carbonyl sulfide, methane, argon, hydrogen chloride, and nitrogen. The heat in the gasifier liquefies coal ash. Hot syngas and molten solids from the reactor flow downward either into a radiant heat exchanger where the syngas is cooled to 1,250°F and the slag solidifies or to the syngas quench chamber for cooling and removal of entrained solids. The solids collect in the water sump at the bottom of the gasifier and are removed periodically using a lock hopper system.

Raw Gas Cooling (Cases 2-1)

Hot raw gas exits the gasifier at 965 psia and 2,500 ºF. This gas stream is cooled to approximately 1,250 ºF in a radiant exchange boiler. The waste heat from this cooling is used to generate high-pressure steam. Boiler feedwater in the tubes is saturated and then steam and water are separated in a steam drum. The raw syngas is saturated and cooled further in a water bath quench.

3.3 GAS CLEANUP – CASES 2-1 AND 2-2

Syngas Quench/Scrubber

Syngas enters the syngas quench area and is directed downwards by a dip tube into a water sump at the bottom of the gasifier vessel or radiant cooler. Most of the solids are separated from the syngas at the bottom of the dip tube as the syngas goes upwards through the water. From the overhead of the quench, the syngas enters the low-temperature gas cooling section for further cooling and particulate capture.

The water removed from the syngas scrubber contains all the solids that were not removed in the quench gasifier water sump. In order to limit the amount of solids recycled to the quench chamber and to limit HCl concentration to below 800 ppmw, a continuous blowdown stream is removed from the bottom of the syngas quench. The blowdown is sent to the sour water stripper. The circulating quench water is pumped by circulating pumps to the quench gasifier.

Solids collected in the water sump are removed by gravity and forced circulation of water from the lock hopper circulating pump. Fine material, which does not settle as easily, is removed in the gasification blowdown and goes to the vacuum flash drum by way of the syngas scrubber.

The slag handling system removes solids from the gasification process equipment. These solids consist of a small amount of unconverted carbon and essentially all of the ash contained in the feed coal. These solids are in the form of a glassy frit, which fully encapsulates any metals.

Sour Water Stripper

The sour water stripper removes chloride, NH_3, SO_2, and other impurities from the waste stream of the gasifier quench blowdown. The sour gas stripper consists of a sour drum that accumulates sour water from the gasifier quench blowdown, following quench blowdown treatment for chloride. The treated water flows to the sour stripper, which consists of a packed column with a steam-heated reboiler. Sour gas is stripped from the liquid and sent to the sulfur recovery unit. Remaining water is then recycled to the gasifier slurry makeup.

Water Gas Shift Reactors

The WGS reactor is the same as that described for the SMR cases above consisting of two sets of parallel fixed-bed reactors arranged in series. Steam is added to the syngas before it enters the WGS reactor. Cooling is provided between the series of two reactors to control the exothermic temperature rise. The parallel set of reactors is required due to the high gas mass flow rate. Feed to the shift converter is first preheated by hot effluent from the second converter, and finally fed to the top of the two parallel first-stage converters. Effluent from the first stage is cooled and fed to the top of the second-stage converters. A nominal 80 percent of the CO is converted to CO_2 and H_2. The effluent from the second stage is cooled through a series of gas coolers to about 100 ºF before mercury and H_2S removal.

Mercury Removal

Performance of carbon bed systems was based on information obtained from the Eastman Chemical Company, which uses carbon beds at its syngas facility in Kingsport, Tennessee. The packed carbon bed vessels are located upstream of the sulfur recovery unit and at a temperature of about 100°F.

Mercury removal is accomplished by packed beds of sulfur-impregnated carbon similar to that used at Eastman Chemical's gasification plant. Dual beds of sulfur-impregnated carbon with approximately a 20-second superficial gas residence time achieve 95 percent mercury reduction in addition to removal of other volatile heavy metals such as arsenic.

Acid Gas Removal

H_2S and CO_2 are removed within the same process system, the Selexol unit. The purpose of the Selexol unit is to preferentially remove H_2S as a product stream and then to remove CO_2 as a separate product stream. This is achieved in the double-stage Selexol unit.

Cool, dry, and particulate-free synthesis gas enters the first absorber unit at approximately 867 psia and 103°F. In this absorber, H_2S is preferentially removed from the fuel gas stream by "loading" the lean Selexol solvent with CO_2. The solvent, saturated with CO_2, preferentially removes H_2S. The rich solution leaving the bottom of the absorber is regenerated in a stripper through the indirect application of thermal energy via condensing low-pressure steam in a reboiler. The stripper acid gas stream, consisting of 35 percent H_2S and 52 percent CO_2, is then sent to the Claus unit.

Sweet fuel gas flowing from the first absorber is cooled and routed to the second absorber unit. In this absorber, the fuel gas is contacted with "unloaded" lean solvent. The solvent removes approximately 95 percent of the CO_2 in the fuel gas stream. A CO_2 balance is maintained by hydraulically expanding the CO_2-saturated rich solution and then flashing CO_2 vapor off the liquid at reduced pressure. Sweet fuel gas off the second absorber is sent to the hydrogen separation process.

Claus Unit

Acid gas from the first-stage stripper of the Selexol unit is routed to the Claus plant. The Claus plant partially oxidizes the H_2S in the acid gas to elemental sulfur. About 150 tpd of elemental sulfur are recovered from the fuel gas stream. This value represents an overall sulfur recovery efficiency of 99.8 percent.

Acid gas from the Selexol unit is preheated to 450°F. A portion of the acid gas along with all of the sour gas and oxidant are fed to the Claus furnace. In the furnace, H_2S is catalytically oxidized to SO_2 using 95 percent pure oxygen from the ASU. A furnace temperature greater than 2,450°F must be maintained in order to thermally decompose all of the NH_3 present in the sour gas stream.

Three preheaters and three sulfur converters are used to obtain a per-pass H_2S conversion of approximately 97.8 percent. In the furnace waste heat boiler, 650 psia steam is generated. This steam is used to satisfy all Claus process preheating and reheating requirements as well as steam to the medium-pressure steam header. The sulfur condensers produce 50 psig steam for the low-pressure steam header.

3.4 PRESSURE SWING ADSORPTION SEPARATION PROCESS

The pressure swing adsorption (PSA) process is used for hydrogen purification, based on the ability to produce high purity hydrogen, low amounts of CO and CO_2, and ease of operation. Treated gas from the amine unit is fed directly to the PSA unit where hydrogen is purified up to approximately 99.9 percent. The PSA process is based on the principle of adsorbent beds adsorbing more impurities at high gas-phase partial pressure than at low partial pressure. The PSA process operates on a cyclic basis and is controlled by automatic switching valves. Multiple beds are used in order to provide constant product and purge gas flows.

In case 1-1, the gas stream is passed through adsorbent beds at 375 to 400 psia; and the impurities are purged from the beds at 20 psia. Purified hydrogen is produced at 380 psia. The purified product hydrogen leaves the PSA unit at 380 psia, and the PSA off-gas (containing remaining hydrogen and CO) is sent as fuel to the gas-fired reformer heater which is primarily fired with supplemental natural gas. The exhaust from the reformer heater enters a heat recovery steam generator (HRSG) for steam generation from the reformer stack gas. The stack gas then passes through an MEA process to capture 70 percent of the CO_2.

In cases 2-1 and 2-2, the gas stream is passed through adsorbent beds at 770 psia and the impurities are purged from the beds at 77 psia. The purified product hydrogen leaves the PSA unit at 770 psia, and the PSA off-gas (containing remaining hydrogen and CO) is sent as fuel to the auxiliary boiler where it is fired with air in a low-NOx burner to produce a stack gas of nitrogen, water vapor, and a minor amount of CO_2.

3.5 CO_2 COMPRESSION AND DEHYDRATION – ALL CASES

The CO_2 streams from the MEA and MDEA units in case 1-1 and those from the Selexol units in cases 2-1 and 2-2 are compressed and dried. In the compression section, the CO_2 is compressed to approximately 2,200 psia by a multi-stage intercooled centrifugal compressor. The discharge pressures of the stages are balanced to give reasonable power distribution and discharge temperatures across the various stages as shown in Exhibit 3-2 for eight stages.

Exhibit 3-2 CO_2 Compressor Interstage Pressures

Stage	Outlet Pressure, MPa (psia)
1	0.24 (35)
2	0.50 (73)
3	1.0 (150)
4	2.1 (300)
5	2.8 (410)
6	5.2 (750)
7	8.5 (1,230)
8	15.3 (2,215)

Power consumption for this large compressor was estimated assuming a polytropic efficiency of 86 percent. During compression to approximately 2,200 psia in the multiple-stage, intercooled compressor, the CO_2 stream is dehydrated to a dew point of -40º with triethylene glycol. The virtually moisture-free supercritical CO_2 stream is then ready for pipeline transport.

The overall availability of the operating plant will be 90 percent, which is consistent with commercial SMR plants. The balance of plant will be single train, operating at 100 percent capacity, based on commercial process operating experience as verified by equipment vendors.

3.6 CO_2 TRANSPORT, STORAGE AND MONITORING – ALL CASES

CO_2 is supplied to the pipeline at the plant gate at a pressure of 15.3 MPa (2,215 psia). The CO_2 product gas composition varies in the cases presented but is expected to meet the specification described in Exhibit 3-3. The CO_2 is transported 80 km (50 miles) via pipeline to a geologic sequestration field for injection into a saline formation.

The CO_2 is transported and injected as a supercritical fluid in order to avoid two-phase flow and achieve maximum efficiency [Ref. 6]. The pipeline is assumed to have an outlet pressure above the supercritical pressure of 10.4 MPa (1,515 psia) with no recompression during transport. Accordingly, CO_2 flow in the pipeline was modeled to determine the pipe diameter that results in a pressure drop of 4.8 MPa (700 psi) over a fifty mile pipeline length [Ref. 7]. (Although not explored in this study, the use of boost compressors and a smaller pipeline diameter could possibly reduce capital costs for sufficiently long pipelines.) The diameter of the injection pipe will be of sufficient size that frictional losses during injection are minimal and no booster compression is required at the well-head in order to achieve an appropriate down-hole pressure.

The saline formation is at a depth of 1,236 meters (4,055 feet) and has a permeability of 22 md and formation pressure of 8.4 MPa (1,220 psig). This is considered an average storage site and requires roughly one injection well for each 9,360 metric tons (10,300 short tons) of CO_2 injected per day [Ref. 8]. The assumed aquifer characteristics are tabulated in Exhibit 3-4.

Exhibit 3-3 CO_2 Pipeline Specification

Parameter	Units	Molar Composition
Inlet Pressure	MPa (psia)	15.3 (2,215)
Outlet Pressure	MPa (psia)	10.4 (1,515)
Inlet Temperature	°C (°F)	35 (95)
N_2 Composition	ppmv	< 300
O_2 Composition	ppmv	< 40
Ar Composition	ppmv	< 10

Exhibit 3-4 Deep, Saline Aquifer Specification

Parameter	Units	Base Case
Pressure	MPa (psi)	8.4 (1,220)
Thickness	m (ft)	161 (530)
Depth	m (ft)	1,236 (4,055)
Permeability	md	22
Pipeline Distance	km (miles)	80 (50)
Injection Rate per Well	tonne (ton) CO_2/day	9,360 (10,320)

4. CASES 1-1 AND 1-2: SMR RESULTS

This case is an upgrade from the original H2A design basis. Changes to the scope of work as requested in technical direction from NETL include adjusting the hydrogen production rate to match that of the revised cases 2-1 and 2-2, hydrogen from coal gasification, and adjusting the amount of CO_2 recovered to 90 percent. This design modification affected the steam balance as discussed below.

- **Modified Steam Balance**

Because of the steam load required to regenerate CO_2 absorbents, there is no steam export from this plant. Prior SMR cases estimated the amount of steam required to regenerate the CO_2 from the MDEA and MEA systems. The estimate assumed that the steam generated within the plant would approximate the regeneration requirements and that the steam was balanced without additional fuel.

In the revised design, the MDEA and MEA regeneration steam exceeded that which is generated internally and therefore the model required additional steam generation. This additional generation was modeled as an integral component of the reformer heater and that, combined with the steam generated in the HRSG in the reformer flue gas train, achieves the required amount of steam for plant operation. The increased steam production results in an additional reformer heater fuel feed by supplementing the baseline PSA off-gas with natural gas (approximately 12 percent of the total plant natural gas consumption).

Steam reforming of hydrocarbons continues to be the most efficient, economical, and widely used process for production of hydrogen and hydrogen/carbon monoxide mixtures. The process involves catalytic conversion of hydrocarbons and steam to produce hydrogen and carbon oxides. Since the process works only with light hydrocarbons which can be vaporized completely without carbon deposition, the feedstocks used range from methane (natural gas) to naphtha to No. 2 fuel oil.

A block flow diagram of the process is shown in Exhibit 4-1 with the corresponding stream tables shown in Exhibit 4-2 and Exhibit 4-3 for cases 1-1 and 1-2 respectively.

Exhibit 4-1 Case 1-1 and 1-2 Block Flow Diagram: Baseline SMR with CO_2 Capture

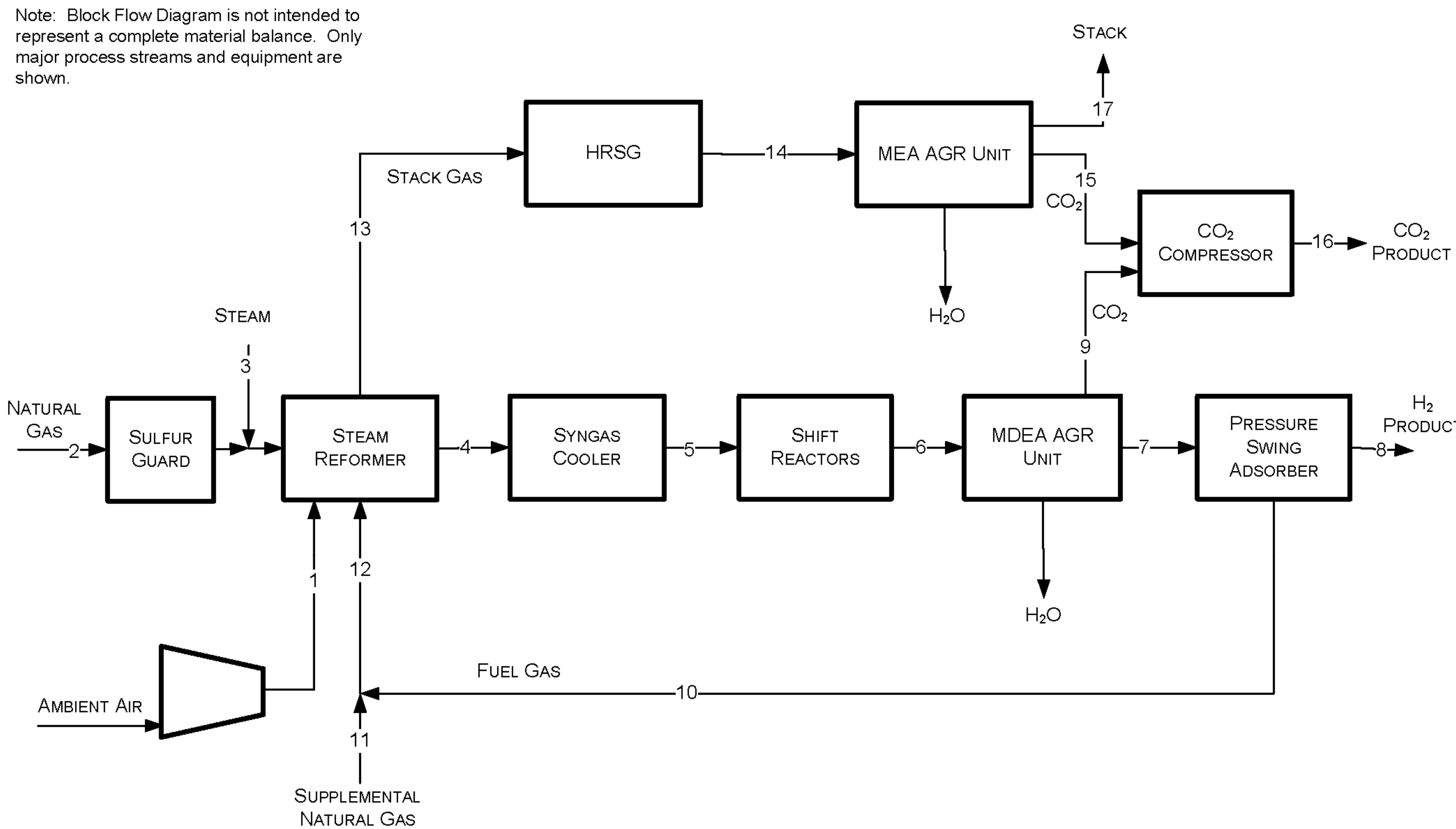


Exhibit 4-2 Case 1-1 Stream Table: Baseline SMR with CO_2 Capture

	1	2	3	4	5	6	7	8	9
V-L Mole Fraction									
AR	0.0094	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.9315	0.0000	0.0179	0.0179	0.0179	0.0333	0.0000	0.0000
C_2H_6	0.0000	0.0317	0.0000	0.0049	0.0049	0.0049	0.0091	0.0000	0.0000
C_3H_8	0.0000	0.0069	0.0000	0.0011	0.0011	0.0011	0.0020	0.0000	0.0000
C_4H_{10}	0.0000	0.0040	0.0000	0.0006	0.0006	0.0006	0.0011	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0699	0.0699	0.0010	0.0018	0.0000	0.0000
CO_2	0.0003	0.0129	0.0000	0.0579	0.0579	0.1268	0.0118	0.0000	0.9458
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.4331	0.4331	0.5020	0.9370	1.0000	0.0001
H_2O	0.0104	0.0000	1.0000	0.4128	0.4128	0.3439	0.0000	0.0000	0.0540
N_2	0.7722	0.0129	0.0000	0.0020	0.0020	0.0020	0.0037	0.0000	0.0000
O_2	0.2077	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)	25,890	4,892	18,861	31,733	31,733	31,733	17,000	12,743	4,040
V-L Flowrate (kg/hr)	747,048	84,947	339,788	424,735	424,735	424,735	59,986	25,689	172,101
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0
Temperature (°C)	16	15	399	866	204	39	38	38	49
Pressure (MPa, abs)	0.11	3.10	3.10	3.03	3.00	2.79	2.69	2.62	0.14
Enthalpy (kJ/kg)[A]	31.61	-6.89	3,228.10	3,665.98	1,857.72	92.17	316.08	546.87	98.41
Density (kg/m^3)	1.3	24.5	10.5	4.3	10.3	21.6	3.6	2.0	2.2
V-L Molecular Weight	28.854	17.365	18.015	13.385	13.385	13.385	3.529	**2.016**	42.599
V-L Flowrate (lb_{mol}/hr)	57,078	10,785	41,582	69,960	69,960	69,960	37,478	28,094	8,907
V-L Flowrate (lb/hr)	1,646,960	187,276	749,105	936,381	936,381	936,381	132,247	56,634	379,418
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0
Temperature (°F)	60	59	750	1,590	400	103	100	100	120
Pressure (psia)	16.0	450.0	450.0	440.0	435.0	405.0	390.0	380.0	19.7
Enthalpy (Btu/lb)[A]	13.6	-3.0	1,387.8	1,576.1	798.7	39.6	135.9	235.1	42.3
Density (lb/ft^3)	0.083	1.529	0.656	0.267	0.644	1.347	0.226	0.126	0.136

A - Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 4-2 Case 1-1 Stream Table: Baseline SMR with CO_2 Capture (continued)

	10	11	12	13	14	15	16	17
V-L Mole Fraction								
AR	0.0000	0.0000	0.0000	0.0083	0.0083	0.0000	0.0000	0.0106
CH_4	0.1332	0.9315	0.2310	0.0000	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0365	0.0317	0.0359	0.0000	0.0000	0.0000	0.0000	0.0000
C_3H_8	0.0080	0.0069	0.0079	0.0000	0.0000	0.0000	0.0000	0.0000
C_4H_{10}	0.0046	0.0040	0.0045	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0073	0.0000	0.0064	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.0472	0.0129	0.0430	0.0655	0.0655	0.9999	0.9967	0.0251
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.7485	0.0000	0.6568	0.0000	0.0000	0.0000	0.0001	0.0000
H_2O	0.0000	0.0000	0.0000	0.2212	0.2212	0.0000	0.0031	0.0626
N_2	0.0148	0.0129	0.0146	0.6850	0.6850	0.0001	0.0000	0.8761
O_2	0.0000	0.0000	0.0000	0.0200	0.0200	0.0000	0.0000	0.0256
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	4,256	594	4,851	29,291	29,291	1,345	5,182	22,900
V-L Flowrate (kg/hr)	34,298	10,319	44,617	791,665	791,665	59,172	227,631	641,590
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0
Temperature (°C)	38	15	32	643	121	21	124	138
Pressure (MPa, abs)	0.14	3.10	0.14	0.11	0.11	0.16	15.27	0.101
Enthalpy (kJ/kg)^A	147.33	-6.89	111.66	1,146.08	506.24	15.46	16.92	241.021
Density (kg/m^3)	0.4	24.5	0.5	0.4	0.9	2.9	276.5	0.8
V-L Molecular Weight	**8.058**	17.365	9.198	27.028	27.028	44.008	43.924	28.017
V-L Flowrate (lb_{mol}/hr)	9,384	1,310	10,694	64,575	64,575	2,964	11,425	50,486
V-L Flowrate (lb/hr)	75,614	22,750	98,364	1,745,323	1,745,323	130,453	501,839	1,414,465
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0
Temperature (°F)	100	59	90	1,190	250	69	255	281
Pressure (psia)	20.0	450.0	20.0	16.0	16.0	23.5	2,214.7	14.7
Enthalpy (Btu/lb)^A	63.3	-3.0	48.0	492.7	217.6	6.6	7.3	103.6
Density (lb/ft^3)	0.027	1.529	0.031	0.024	0.057	0.184	17.263	0.052

A - Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 4-3 Case 1-2 Stream Table: Baseline SMR with CO_2 Capture

	1	2	3	4	5	6	7	8	9
V-L Mole Fraction									
AR	0.0094	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.9315	0.0000	0.0179	0.0179	0.0179	0.0335	0.0000	0.0000
C_2H_6	0.0000	0.0317	0.0000	0.0049	0.0049	0.0049	0.0091	0.0000	0.0000
C_3H_8	0.0000	0.0069	0.0000	0.0011	0.0011	0.0011	0.0020	0.0000	0.0000
C_4H_{10}	0.0000	0.0040	0.0000	0.0006	0.0006	0.0006	0.0011	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0698	0.0698	0.0010	0.0018	0.0000	0.0000
CO_2	0.0003	0.0129	0.0000	0.0579	0.0579	0.1267	0.0118	0.0000	0.9458
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.4329	0.4329	0.5018	0.9369	1.0000	0.0001
H_2O	0.0104	0.0000	1.0000	0.4129	0.4129	0.3441	0.0000	0.0000	0.0541
N_2	0.7722	0.0129	0.0000	0.0020	0.0020	0.0020	0.0037	0.0000	0.0000
O_2	0.2077	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)	25,997	4,914	18,946	31,871	31,871	31,871	17,069	12,793	4,056
V-L Flowrate (kg/hr)	750,135	85,328	341,312	426,640	426,640	426,640	60,268	25,789	172,779
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0
Temperature (°C)	16	15	399	865	204	39	38	38	49
Pressure (MPa, abs)	0.11	3.10	3.10	3.03	3.00	2.79	2.69	2.62	0.14
Enthalpy (kJ/kg)[A]	31.61	-6.89	3,228.10	3,665.15	1,857.85	92.16	315.89	546.87	98.47
Density (kg/m^3)	1.3	24.5	10.5	4.3	10.3	21.6	3.6	2.0	2.2
V-L Molecular Weight	28.854	17.365	18.015	13.386	13.386	13.386	3.531	**2.016**	42.598
V-L Flowrate (lb_{mol}/hr)	57,314	10,833	41,768	70,263	70,263	70,263	37,630	28,204	8,942
V-L Flowrate (lb/hr)	1,653,764	188,116	752,463	940,579	940,579	940,579	132,868	56,855	380,912
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0
Temperature (°F)	60	59	750	1,589	400	103	100	100	120
Pressure (psia)	16.0	450.0	450.0	440.0	435.0	405.0	390.0	380.0	19.7
Enthalpy (Btu/lb)[A]	13.6	-3.0	1,387.8	1,575.7	798.7	39.6	135.8	235.1	42.3
Density (lb/ft^3)	0.083	1.529	0.656	0.267	0.644	1.347	0.227	0.126	0.136

A - Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 4-3 Case 1-2 Stream Table: Baseline SMR with CO_2 Capture (continued)

	10	11	12	13	14	15	16	17
V-L Mole Fraction								
AR	0.0000	0.0000	0.0000	0.0083	0.0083	0.0000	0.0000	0.0106
CH_4	0.1337	0.9315	0.2311	0.0000	0.0000	0.0000	0.0000	0.0000
C_2H_6	0.0365	0.0317	0.0359	0.0000	0.0000	0.0000	0.0000	0.0000
C_3H_8	0.0080	0.0069	0.0079	0.0000	0.0000	0.0000	0.0000	0.0000
C_4H_{10}	0.0046	0.0040	0.0045	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0073	0.0000	0.0064	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.0472	0.0129	0.0430	0.0655	0.0655	0.9999	0.9967	0.0251
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.7480	0.0000	0.6567	0.0000	0.0000	0.0000	0.0001	0.0000
H_2O	0.0000	0.0000	0.0000	0.2212	0.2212	0.0000	0.0031	0.0626
N_2	0.0148	0.0129	0.0146	0.6850	0.6850	0.0001	0.0000	0.8761
O_2	0.0000	0.0000	0.0000	0.0200	0.0200	0.0000	0.0000	0.0256
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	4,276	594	4,870	29,411	29,411	1,350	5,203	22,994
V-L Flowrate (kg/hr)	34,479	10,319	44,798	794,933	794,933	59,423	228,541	644,231
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0
Temperature (°C)	38	15	32	643	121	21	124	138
Pressure (MPa, abs)	0.14	3.10	0.14	0.11	0.11	0.16	15.27	0.101
Enthalpy (kJ/kg)[A]	147.23	-6.89	111.73	1,146.06	506.22	15.46	16.92	240.992
Density (kg/m^3)	0.4	24.5	0.5	0.4	0.9	2.9	276.5	0.8
V-L Molecular Weight	**8.064**	17.365	9.199	27.028	27.028	44.008	43.924	28.017
V-L Flowrate (lb_{mol}/hr)	9,427	1,310	10,737	64,841	64,841	2,977	11,471	50,694
V-L Flowrate (lb/hr)	76,013	22,750	98,763	1,752,527	1,752,527	131,004	503,846	1,420,287
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0
Temperature (°F)	100	59	90	1,190	250	69	255	281
Pressure (psia)	20.0	450.0	20.0	16.0	16.0	23.5	2,214.7	14.7
Enthalpy (Btu/lb)[A]	63.3	-3.0	48.0	492.7	217.6	6.6	7.3	103.6
Density (lb/ft^3)	0.027	1.529	0.031	0.024	0.057	0.184	17.263	0.052

A - Reference conditions are 32.02 F & 0.089 PSIA

4.1 COMPONENT TABLES

The component tables below contain general specifications and overall process data for the major process component systems for cases 1-1 and 1-2.

NATURAL GAS SULFUR GUARD

Case	**1-1**	**1-2**
Technology Type	Fixed bed zinc oxide pellets	Fixed bed zinc oxide pellets
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions Inlet – Gas	95,267 kg/hr (210,029 lb/hr) natural gas, as received 3.1MPa (450.0 psia), 15°C (59°F) (flow includes supplemental gas)	95,648 kg/hr (210,868 lb/hr) natural gas, as received 3.1MPa (450.0 psia), 15°C (59°F) (flow includes supplemental gas)
Operating Conditions Outlet – Gas	95,266kg/hr (210,026 lb/hr) natural gas, as received 3.1 MPa (450.0 psia), 15°C (59°F) (flow includes supplemental gas)	95,647kg/hr (210,866 lb/hr) natural gas, as received 3.1 MPa (450.0 psia), 15°C (59°F) (flow includes supplemental gas)
Assumed or Specified Performance Characteristics	Operating pressure, temperature, sulfur removal, sorbent capacity, all based on information published by vendor.	Operating pressure, temperature, sulfur removal, sorbent capacity, all based on information published by vendor.
Calculated Performance Characteristics	Calculated 99.9% sulfur removal	Calculated 99.9% sulfur removal
Contaminant Removed, %	99.9%	99.9%
Assumptions Regarding Anticipated Application Issues	Commercial, no issues	Commercial, no issues

STEAM METHANE REFORMER

Case	1-1	1-2
Technology Type	Single-stage, catalytic, externally heated	Single-stage, catalytic, externally heated
Basis for Design and Performance	Vendor data – commercial	Vendor data – commercial
Operating Conditions Inlet – Gas Inlet - Steam Fuel - PSA Off-Gas + Nat Gas	84,947 kg/hr (187,276 lb/hr) natural gas, sulfur free 3.1MPa (450.0 psia), 15°C (59°F) 339,788 kg/hr (749,105 lb/hr) steam 3.1 MPa (450.0 psia), 399°C (750°F) 44,617 kg/hr (98,364 lb/hr) fuel 0.1MPa (20.0 psia), 32°C (90°F)	85,328 kg/hr (188,116 lb/hr) natural gas, sulfur free 3.1MPa (450.0 psia), 15°C (59°F) 341,312 kg/hr (752,463 lb/hr) steam 3.1 MPa (450.0 psia), 399°C (750°F) 44,798 kg/hr (98,763 lb/hr) fuel 0.1MPa (20.0 psia), 32°C (90°F)
Operating Conditions Outlet – Syngas Outlet - Stack Gas	424,735 kg/hr (936,381 lb/hr) raw syngas 3.0MPa (440.0 psia), 866°C (1,590°F) 791,665 kg/hr (1,745,323 lb/hr) stack gas 0.1MPa (16.0 psia), 643°C (1,190°F)	426,640 kg/hr (940,579 lb/hr) raw syngas 3.0MPa (440.0 psia), 866°C (1,590°F) 794,933 kg/hr (1,752,527 lb/hr) stack gas 0.1MPa (16.0 psia), 643°C (1,190°F)
Assumed or Specified Performance Characteristics	Operating temperature, pressure; equilibrium syngas composition	Operating temperature, pressure; equilibrium syngas composition
Calculated Performance Characteristics	Modeled performance based on performance assumptions	Modeled performance based on performance assumptions
Contaminant Removed, %	N/A	N/A
Assumptions Regarding Anticipated Application Issues	Commercial, no issues	Commercial, no issues

SYNGAS COOLER

Case	1-1	1-2
Technology Type	Convective Boiler - Drum	Convective Boiler - Drum
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions Inlet - Syngas	424,735 kg/hr (936,381 lb/hr) raw syngas, 3.0 MPa (440.0 psia), 874°C (1,605°F)	426,640 kg/hr (940,579 lb/hr) raw syngas, 3.0 MPa (440.0 psia), 874°C (1,605°F)
Operating Conditions Outlet - Syngas	424,735 kg/hr (936,381 lb/hr) raw syngas, 3.0 MPa (435.0 psia), 204°C (400°F)	426,640 kg/hr (940,579 lb/hr) raw syngas, 3.0 MPa (435.0 psia), 204°C (400°F)
Assumed or Specified Performance Characteristics	Syngas temperature profile	Syngas temperature profile
Calculated Performance Characteristics	Modeled performance based on performance assumptions	Modeled performance based on performance assumptions
Contaminant Removed, %	N/A	N/A
Assumptions Regarding Anticipated Application Issues	Commercial design – no issues.	Commercial design – no issues.

WATER GAS SHIFT REACTOR

Case	1-1	1-2
Technology Type	Haldor Topsoe Two-Stage Shift Catalyst	Haldor Topsoe Two-Stage Shift Catalyst
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions Inlet – Gas (1)	424,735 kg/hr (936,381 lb/hr) syngas, 3.0 MPa (435.0 psia), 204°C (400°F)	424,640 kg/hr (940,579 lb/hr) syngas, 3.0 MPa (435.0 psia), 204°C (400°F)
Operating Conditions Cooled Outlet – Syngas (2)	424,735 kg/hr (936,381 lb/hr) syngas, 2.8 MPa (405.0 psia), 39°C (103°F)	424,640 kg/hr (940,579 lb/hr) syngas, 2.8 MPa (405.0 psia), 39°C (103°F)
Assumed or Specified Performance Characteristics	Specified shift catalyst which also promotes COS hydrolysis; Assume minimum H_2O/CO ratio of 2.0 at inlet; Interstage cooling; Assume two stages required to achieve >97% CO conversion	Specified shift catalyst which also promotes COS hydrolysis; Assume minimum H_2O/CO ratio of 2.0 at inlet; Interstage cooling; Assume two stages required to achieve >97% CO conversion
Calculated Performance Characteristics	Modeled WGS reaction based on equilibrium to >97% CO conversion	Modeled WGS reaction based on equilibrium to >97% CO conversion
Contaminant Removed, %	N/A	N/A
Assumptions Regarding Anticipated Application Issues	Commercial design; no issues	Commercial design; no issues

(1) The SMR outlet syngas contains sufficient water for the WGS reaction and no steam is added.

(2) The WGS Reactor component table includes the catalytic reactors, intercoolers, and the syngas aftercooler in preparation for entrance to the MDEA absorber.

SYNGAS MDEA CO_2 REMOVAL

Case	1-1	1-2
Technology Type	Proprietary MDEA Solvent	Proprietary MDEA Solvent
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions Inlet - Syngas	424,735 kg/hr (936,381 lb/hr) shifted syngas, 2.8 MPa (405.0 psia), 39°C (103°F)	426,640 kg/hr (940,579 lb/hr) shifted syngas, 2.8 MPa (405.0 psia), 39°C (103°F)
Operating Conditions Outlet - CO_2 Outlet – Syngas (1)	172,101 kg/hr (379,418 lb/hr) CO_2, 0.1 MPa (19.7 psia), 49°C (120°F) 59,986 kg/hr (132,247 lb/hr) Claus feed gas, 2.7 MPa (390.0 psia), 38°C (100°F)	172,779 kg/hr (380,912 lb/hr) CO_2, 0.1 MPa (19.7 psia), 49°C (120°F) 60,268 kg/hr (132,868 lb/hr) Claus feed gas, 2.7 MPa (390.0 psia), 38°C (100°F)
Assumed or Specified Performance Characteristics	Specified 95% specific to CO_2 removal	Specified 95% specific to CO_2 removal
Calculated Performance Characteristics	Model results reflect design assumptions	Model results reflect design assumptions
Contaminant Removed, %	95% CO_2	95% CO_2
Assumptions Regarding Anticipated Application Issues	Vendor design specifically applied to similar Hydrogen production conditions	Vendor design specifically applied to similar Hydrogen production conditions

(1) Condensed water separates from the cool syngas at the entrance to the absorber, reducing the net outlet syngas mass flow

PRESSURE SWING ADSORBER

Case	**1-1**	**1-2**
Technology Type	Pressure Swing Adsorption	Pressure Swing Adsorption
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions Inlet - Syngas	59,986 kg/hr (132,247 lb/hr) syngas, 2.7 MPa (390.0 psia), 38°C (100°F)	60,268 kg/hr (132,868 lb/hr) syngas, 2.7 MPa (390.0 psia), 38°C (100°F)
Operating Conditions Outlet – Hydrogen Outlet – To SMR burner	25,689 kg/hr (56,634 lb/hr) H_2 2.6 MPa (380.0 psia), 38°C (100°F) 34,298 kg/hr (75,614 lb/hr) off-gas, 0.1 MPa (20.0 psia), 38°C (100°F)	25,789 kg/hr (56,855 lb/hr) H_2 2.6 MPa (380.0 psia), 38°C (100°F) 34,479 kg/hr (76,013 lb/hr) off-gas, 0.1 MPa (20.0 psia), 38°C (100°F)
Assumed or Specified Performance Characteristics	PSA operates at 80% hydrogen removal efficiency. Off-gas is sent back to reformer heater/boiler as fuel	PSA operates at 80% hydrogen removal efficiency. Off-gas is sent back to reformer heater/boiler as fuel
Calculated Performance Characteristics	>99.9 % purity H_2	>99.9 % purity H_2
Contaminant Removed, %	N/A	N/A
Assumptions Regarding Anticipated Application Issues	Commercial, no issues	Commercial, no issues

STACK GAS HEAT RECOVERY STEAM GENERATOR (HRSG)

Case	1-1	1-2
Technology Type	Sub-Critical Drum	Sub-Critical Drum
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions Inlet - Flue Gas	791,665 kg/hr (1,745,323 lb/hr) syngas, 0.1 MPa (16.0 psia), 644°C (1,191°F)	794,933 kg/hr (1,752,527 lb/hr) syngas, 0.1 MPa (16.0 psia), 644°C (1,191°F)
Operating Conditions Outlet -Flue Gas	791,665 kg/hr (1,745,323 lb/hr) syngas, 0.1 MPa (16.0 psia), 121°C (250°F)	794,933 kg/hr (1,752,527 lb/hr) syngas, 0.1 MPa (16.0 psia), 121°C (250°F)
Assumed or Specified Performance Characteristics	Specified pressure and steam requirement	Specified pressure and steam requirement
Calculated Performance Characteristics	Calculated 507 GJ/hr (480 MMBtu/hr) Heat Recovery based on gas flow and temperatures	Calculated 509 GJ/hr (482 MMBtu/hr) Heat Recovery based on gas flow and temperatures
Contaminant Removed, %	N/A	N/A
Assumptions Regarding Anticipated Application Issues	Commercial, no issues	Commercial, no issues

STACK GAS HINDERED AMINE (MEA) CO_2 REMOVAL

Case	1-1	1-2
Technology Type	Fluor Econamine FG Plus	Fluor Econamine FG Plus
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions		
Operating Conditions Inlet - Syngas	791,665 kg/hr (1,745,323 lb/hr) flue gas, 0.1 MPa (16.0 psia), 121°C (250°F)	794,933 kg/hr (1,752,527 lb/hr) flue gas 0.1 MPa (16.0 psia), 121°C (250°F)
Operating Conditions Outlet - CO_2 Outlet - To Stack	59,172 kg/hr (130,453 lb/hr) CO_2, 0.2 MPa (23.5 psia), 21°C (69°F) 641,590 kg/hr (1,414,465 lb/hr) stack gas 0.1 MPa (14.7 psia), 138°C (281°F)	59,423 kg/hr (131,004 lb/hr) CO_2, 0.2 MPa (23.5 psia), 21°C (69°F) 644,231 kg/hr (1,420,287 lb/hr) stack gas 0.1 MPa (14.7 psia), 138°C (281°F)
Assumed or Specified Performance Characteristics	Specified 70% specific to CO_2 removal	Specified 70% specific to CO_2 removal
Calculated Performance Characteristics	Model results reflect design assumptions	Model results reflect design assumptions
Contaminant Removed, %	70% CO_2 from gas to stack	70% CO_2 from gas to stack
Assumptions Regarding Anticipated Application Issues	Vendor design specifically provided for low pressure syngas conditions	Vendor design specifically provided for low pressure syngas conditions

CO_2 COMPRESSION

Case	1-1	1-2
Technology Type	Multi-Stage Integral Gear Compressor	Multi-Stage Integral Gear Compressor
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions Inlet - CO_2	231,273 kg/hr (509,871 lb/hr) CO_2, 0.2 MPa (23.5 psia), 21°C (69°F)	232,202 kg/hr (511,917 lb/hr) CO_2, 0.2 MPa (23.5 psia), 21°C (69°F)
Operating Conditions Outlet - CO_2 Outlet – H_2O	227,631 kg/hr (501,839 lb/hr) CO_2, 15.3 MPa (2,214.7 psia), 124°C (255°F) 3,643 kg/hr (8,031 lb/hr) H_2O various streams	228,541 kg/hr (503,846 lb/hr) CO_2, 15.3 MPa (2,214.7 psia), 124°C (255°F) 3,661 kg/hr (8,071 lb/hr) H_2O various streams
Assumed or Specified Performance Characteristics	Replicated Great Plains Gasification installation; Assumed 80% isentropic efficiency with intercooled stages	Replicated Great Plains Gasification installation; Assumed 80% isentropic efficiency with intercooled stages
Calculated Performance Characteristics	Model results reflect design assumptions	Model results reflect design assumptions
Contaminant Removed, %	Dehydrated to –40º dew point	Dehydrated to –40º dew point
Assumptions Regarding Anticipated Application Issues	Commercial, no issues	Commercial, no issues

4.2 PERFORMANCE SUMMARY

The overall performance for the plant is summarized in Exhibit 4-4. The overall plant effective thermal efficiency (thermal value of hydrogen and power produced divided by the thermal value of natural gas) is 69.85 percent on an HHV basis. The total amount of CO_2 captured and sent off site is 5,350 tons/day and represents 90 percent of the carbon in the natural gas.

**Exhibit 4-4 Cases 1-1 and 1-2 Plant Performance Summaries
100 Percent Load**

Case	1-1	1-2	Units
Plant Output			
Steam Turbine Power	0	0	kW_e
Total	**0**	**0**	kW_e
Auxiliary Load			
Boiler Feedwater Pumps	450	460	kW_e
Primary Air Fans	2,350	2,360	kW_e
CO_2 Compressor	23,020	23,110	kW_e
Circulating Water Pump	2,560	2,570	kW_e
Ground Water Pumps	300	300	kW_e
Cooling Tower Fans	1,320	1,330	kW_e
CO_2 Removal	3,200	3,200	kW_e
Miscellaneous Balance-of-Plant[2]	1,000	1,000	kW_e
Transformer Losses	0	0	kW_e
Total	**34,200**	**34,330**	**kW_e**
Plant Performance			
Net Plant Power	**-34,200**	**-34,330**	**kW_e**
Plant Capacity Factor	90.0	90.0	
Net Plant Efficiency (HHV)	N/A	N/A	
Net Plant Heat Rate (HHV)	N/A	N/A	
Natural Gas SMR Feed Flow rate	84,947 (187,276)	85,328 (188,116)	kg/hr (lb/hr)
Supplemental NG Feed Flow rate	10,319 (22,750)	10,319 (22,750)	Kg/hr (lb/hr)
Hydrogen Production Flow rate	25,689 (56,634)	25,789 (56,855)	kg/hr (lb/hr)
Thermal Input[1]	1,402,907	1,408,515	kW_t
Effective Thermal Efficiency[3]	69.74%	69.73%	%
Cold Gas Efficiency[4]	72.18%	72.17%	%
Condenser Duty	N/A	N/A	
Raw Water Withdrawal	12.4 (3,265)	12.4 (3,278)	m^3/min (gpm)
Raw Water Consumption	10.1 (2,673)	10.2 (2,683)	m^3/min (gpm)

[1] HHV of Natural Gas 53,014 kJ/kg (22,792 Btu/lb)
[2] Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads
[3] ETE = (Hydrogen Heating Value + Net Power)/ Natural Gas Heating Value, HHV
[4] CGE = (Hydrogen Product Value) / Natural Gas Heating Value, HHV

4.3 MASS AND ENERGY BALANCES

The overall energy balances for the SMR plants are shown in Exhibit 4-5 and Exhibit 4-6.

Exhibit 4-5 Case 1-1 Overall Energy Balance

	HHV	Sensible + Latent	Power	Total
Energy In, GJ/hr (MMBtu/hr)				
SMR Natural Gas	4,503 (4,268)	0.3 (0.3)		4,504 (4,269)
SMR Reaction Endothermic		-858.2 (-813.4)		-858 (-813)
Supplemental Natural Gas	547 (519)	0.04 (0.04)		547 (519)
SMR Air		23.6 (22.4)		24 (22)
Raw Water Makeup		46.5 (44.0)		46 (44)
Auxiliary Power			123 (117)	123 (117)
TOTAL	**5,050 (4,787)**	**-787.7 (-746.6)**	**123 (117)**	**4,386 (4,157)**
Energy Out, GJ/hr (MMBtu/hr)				
H_2 Product	3,645 (3,455)	14.0 (13.3)		3,659 (3,468)
CO_2 Product		4 (4)		4 (4)
CO_2 intercoolers		94.16 (89.25)		94 (89)
Cooling Tower Blowdown		16.6 (15.8)		17 (16)
HRSG Flue Gas		154.6 (146.6)		155 (147)
MEA Reboiler		210.4 (199.4)		210 (199)
MDEA Reboiler		625.9 (593.2)		626 (593)
LP Water Recycle cooler		0.1 (0.1)		0.1 (0.1)
Process Losses*		*-379 (-359)*		*-379 (-359)*
Power		0.0 (0.0)		0 (0)
TOTAL	**3,645 (3,455)**	**741 (702)**	**0 (0)**	**4,386 (4,157)**

* Process losses are estimated to match the heat input to the plant.
Process losses include losses from: HRSG, SMR, combustion reactions, and gas cooling.
Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 4-6 Case 1-2 Overall Energy Balance

	HHV	Sensible + Latent	Power	Total
Energy In, GJ/hr (MMBtu/hr)				
SMR Natural Gas	4,524 (4,288)	0.3 (0.3)		4,524 (4,288)
SMR Reaction Endothermic		-864.1 (-819.0)		-864 (-819)
Supplemental Natural Gas	547 (519)	0.04 (0.04)		547 (519)
SMR Air		23.7 (22.5)		24 (22)
Raw Water Makeup		46.7 (44.2)		47 (44)
Auxiliary Power			124 (117)	124 (117)
TOTAL	**5,071 (4,806)**	**-793.4 (-752.0)**	**124 (117)**	**4,401 (4,171)**
Energy Out, GJ/hr (MMBtu/hr)				
H_2 Product	3,659 (3,468)	14.1 (13.4)		3,674 (3,482)
CO_2 Product		4 (4)		4 (4)
CO_2 intercoolers		94.55 (89.61)		95 (90)
Cooling Tower Blowdown		16.7 (15.8)		17 (16)
HRSG Flue Gas		155.3 (147.2)		155 (147)
MEA Reboiler		211.2 (200.2)		211 (200)
MDEA Reboiler		628.3 (595.5)		628 (596)
LP Water Recycle cooler		0.1 (0.1)		0.1 (0.1)
Process Losses*		*-383 (-363)*		*-383 (-363)*
Power		0.0 (0.0)		0 (0)
TOTAL	**3,659 (3,468)**	**741 (703)**	**0 (0)**	**4,401 (4,171)**

* Process losses are estimated to match the heat input to the plant.
Process losses include losses from: HRSG, SMR, combustion reactions, and gas cooling.
Reference conditions are 32.02 F & 0.089 PSIA

4.3.1 Water Balance

The overall water balances for the plants are shown in Exhibit 4-7 and Exhibit 4-8. Raw water is obtained from groundwater (50 percent) and from municipal sources (50 percent). Water demand represents the total amount of water required for a particular process. Some water is recovered within the process as syngas condensate and that water is re-used as internal recycle. Raw water makeup is the difference between water demand and internal recycle.

Exhibit 4-7 Case 1-1 Water Balance

Water Use	Water Demand, m^3/min (gpm)	Internal Recycle, m^3/min (gpm)	Raw Water Withdrawal, m^3/min (gpm)	Process Water Discharge, m^3/min (gpm)	Raw Water Usage, m^3/min (gpm)
MEA Steam	1.66 (438)	1.66 (438)			
MDEA Steam	4.9 (1,302)	4.9 (1,302)			
SMR Steam	5.67 (1,498)		5. 7 (1,498)		5.7 (1,498)
Cooling Tower	**10.0 (2,632)**	**3.3 (865)**	**6.7 (1,767)**	**2.2 (592)**	**4.4 (1,175)**
SWS Blowdown		3.2 (849)	3.2 (849)		
CO_2 Product Knockout		0.06 (16)	0.06 (16)		
Total	**22.2 (5,870)**	**9.9 (2,605)**	**12.4 (3,265)**	**2.2 (592)**	**10.1 (2,673)**

Exhibit 4-8 Case 1-2 Water Balance

Water Use	Water Demand, m^3/min (gpm)	Internal Recycle, m^3/min (gpm)	Raw Water Withdrawal, m^3/min (gpm)	Process Water Discharge, m^3/min (gpm)	Raw Water Usage, m^3/min (gpm)
MEA Steam	1.66 (439)	1.66 (439)			
MDEA Steam	4.9 (1,307)	4.9 (1,307)			
SMR Steam	5.70 (1,505)		5.7 (1,505)		5.7 (1,505)
Cooling Tower	**10.0 (2,642)**	**3.3 (870)**	**6.7 (1,773)**	**2.2 (594)**	**4.5 (1,178)**
SWS Blowdown		3.2 (854)	3.2 (854)		
CO_2 Product Knockout		0.06 (16)	0.06 (16)		
Total	**22.2 (5,894)**	**9.9 (2,616)**	**12.4 (3,278)**	**2.2 (594)**	**10.2 (2,683)**

4.3.2 Carbon Balance

The carbon balances for the plants are shown in Exhibit 4-9 and Exhibit 4-10. 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 CO_2 in the stack gas and the CO_2 product. The percent of total carbon sequestered is defined as the amount of carbon product produced (as sequestration-ready CO_2) divided by the carbon in the feedstock, expressed as a percentage:

(Carbon in Product for Sequestration)/ (Carbon in the Feed) or
136,783/(151,757)*100 or
90 percent

(Carbon in Product for Sequestration)/ (Carbon in the Feed) or
137,330/(152,363)*100 or
90 percent

Exhibit 4-9 Case 1-1 Carbon Balance

Carbon In, kg/hr (lb/hr)		Carbon Out, kg/hr (lb/hr)	
Natural Gas	61,379 (135,318)	Stack Gas	6,894 (15,198)
Supplemental Natural Gas	7,456 (16,438)	Hydrogen Product	0 (0)
Air (CO_2)	102 (225)	CO_2 Product	62,044 (136,783)
Water In	0 (0)	Convergence Tolerance*	0 (0)
Total	**68,938 (151,981)**	**Total**	**68,938 (151,981)**

*by difference

Exhibit 4-10 Case 1-2 Carbon Balance

Carbon In, kg/hr (lb/hr)		Carbon Out, kg/hr (lb/hr)	
Natural Gas	61,655 (135,925)	Stack Gas	6,921 (15,258)
Supplemental Natural Gas	7,456 (16,438)	Hydrogen Product	0 (0)
Air (CO_2)	102 (226)	CO_2 Product	62,292 (137,330)
Water In	0 (0)	Convergence Tolerance*	0 (0)
Total	**69,213 (152,589)**	**Total**	**69,213 (152,589)**

*by difference

4.3.3 Sulfur Balance

The sulfur in the natural gas is assumed to be a very low concentration (6 ppmv). Virtually all the sulfur is removed in the zinc oxide guard bed and the supplemental firing emissions are negligible.

4.3.4 Air Emissions

The environmental targets for emissions of NOx, SO_2, and particulate matter were presented in Section 1.6. A summary of the plant air emissions is presented in Exhibit 4-11.

Exhibit 4-11 Cases 1-1 and 1-2 Air Emissions

	Case 1-1 Kg/GJ (lb/10^6 Btu)	Case 1-1 Tonne/year (tons/year) 90% Capacity Factor	Case 1-2 Kg/GJ (lb/10^6 Btu)	Case 1-2 Tonne/year (tons/year) 90% Capacity Factor
SO_2	negligible	negligible	negligible	negligible
NO_X	0.014 (0.032)	60 (66)	0.014 (0.032)	60 (66)
Particulates	negligible	negligible	negligible	negligible
Hg	negligible	negligible	negligible	negligible
CO_2	46.2 (107.4)	199,140 (219,514)	46.4 (107.8)	199,936 (220,391)

4.4 MAJOR EQUIPMENT LIST

This section contains the equipment list corresponding to the SMR plant configuration for case 1-1. This list, along with the heat and material balance and supporting performance data, was used to generate plant costs and complete the financial analysis.

ACCOUNT 1 - FEEDWATER AND MISCELLANEOUS SYSTEMS AND EQUIPMENT

Equip-ment No.	Description	Type	Case 1-1 Design Condition	Case 1-2 Design Condition	Oper. Qty. (Spares)
1	Demineralized Water Storage Tank	Vertical, cylindrical, outdoor	2,150,114 liters (568,000 gal)	2,161,470 liters (571,000 gal)	2 (0)
2	Intermediate Pressure Feedwater Pump	Horizontal centrifugal, single stage	3,899 lpm @ 396 m H_2O (1,030 gpm @ 1300 ft H_2O)	3,899 lpm @ 396 m H_2O (1,030 gpm @ 1300 ft H_2O)	2 (1)
3	Primary Air Fan	Centrifugal	410,955 kg/hr @ 0 m^3/min (906,000 lb/hr @ 0 acfm)	412,769 kg/hr @ 0 m^3/min (910,000 lb/hr @ 0 acfm)	2 (0)
4	Service Air Compressors	Flooded Screw	28 m^3/min @ 0.7 MPa (1,000 scfm @ 100 psig)	28 m^3/min @ 0.7 MPa (1,000 scfm @ 100 psig)	2 (1)
5	Ground Water Pumps	Stainless steel, single suction	2,574 lpm @ 91 m H_2O (680 gpm @ 300 ft H_2O)	2,574 lpm @ 91 m H_2O (680 gpm @ 300 ft H_2O)	2 (1)

Equip-ment No.	Description	Type	Case 1-1 Design Condition	Case 1-2 Design Condition	Oper. Qty. (Spares)
6	Raw Water Pumps	Stainless steel, single suction	6,814 lpm @ 268 m H_2O (1,800 gpm @ 880 ft H_2O)	6,814 lpm @ 268 m H_2O (1,800 gpm @ 880 ft H_2O)	2 (2)

ACCOUNT 2 - REFORMER AND ACCESSORIES

Equip-ment No.	Description	Type	Case 1-1 Design Condition	Case 1-2 Design Condition	Oper. Qty. (Spares)
1	Natural Gas Sulfur Guard	Zinc Oxide Bed	95,270 kg/hr (210,130 lb/hr) natural gas, 3.1 MPa (450 psia)	95,650 kg/hr (210,870 lb/hr) natural gas, 3.1 MPa (450 psia)	1 (0)
2	Reformer	Single Stage Catalytic, Externally Heated	93,440 kg/hr natural gas (206,000 lb/hr) natural gas, 373,800 kg/hr (824,000 lb/hr) 3.1 MPa (450 psia)	93,870 kg/hr natural gas (206,930 lb/hr) natural gas, 375,400 kg/hr (827,700 lb/hr) 3.1 MPa (450 psia)	1 (0)
3	Synthesis Gas Cooler	Fire-tube boiler	233,600 kg/hr (515,000 lb/hr) syngas, 3.0 MPa (440 psia)	234,500 kg/hr (517,000 lb/hr) syngas, 3.0 MPa (440 psia)	1 (0)
4	Stack Gas Cooler	Fire-tube boiler	352,900 kg/hr (777,960 lb/hr) stack gas, 0.1 MPa (16 psia),	354,300 kg/hr (781,160 lb/hr) stack gas, 0.1 MPa (16 psia)	1 (0)
5	Stack	CS plate, type 409SS liner	15 m (50 ft) high x 2.6 m (9 ft) diameter	15 m (50 ft) high x 2.6 m (9 ft) diameter	1 (0)

ACCOUNT 3A - WATER GAS SHIFT, SYNGAS CLEANUP AND HYDROGEN PURIFICATION

Equip-ment No.	Description	Type	Case 1-1 Design Condition	Case 1-2 Design Condition	Oper Qty. (Spares)
1	WGS Reactors	Fixed bed, catalytic	233,600 kg/hr (515,000 lb/hr) 204°C (400°F) 3.0 MPa (440 psia)	234,500 kg/hr (517,000 lb/hr) 204°C (400°F) 3.0 MPa (440 psia)	4 (0)
2	Shift Reactor Heat Recovery Exchangers	Shell and Tube	Exchanger 1: 45 GJ/hr (43 MMBtu/hr) Exchanger 2: 363 GJ/hr (344 MMBtu/hr)	Exchanger 1: 43 GJ/hr (45 MMBtu/hr) Exchanger 2: 364 GJ/hr (345 MMBtu/hr)	4 (0)
3	Acid Gas Removal Plant-Syngas Stream	Proprietary MDEA Process	233,600 kg/hr (515,000 lb/hr) 39°C (103°F) 2.8 MPa (405 psia)	234,500 kg/hr (517,000 lb/hr) 39°C (103°F) 2.8 MPa (405 psia)	2 (0)
4	Pressure Swing Adsorber	Polybed Proprietary	32,990 kg/hr (72,735 lb/hr) Syngas 38°C (100°F) 2.7 MPa (390. psia) 14,130 kg/hr (31,150 lb/hr) Hydrogen 38°C (100°F) 2.6 MPa (380. psia) 18,860 kg/hr (41,585 lb/hr) Off Gas 38°C (100°F) 0.1 MPa (20. psia)	33,146 kg/hr (73,075 lb/hr) Syngas 38°C (100°F) 2.7 MPa (390. psia) 14,184 kg/hr (31,270 lb/hr) Hydrogen 38°C (100°F) 2.6 MPa (380. psia) 18,960 kg/hr (41,805 lb/hr) Off Gas 38°C (100°F) 0.1 MPa (20. psia)	2 (0)
5	Acid Gas Removal Plant-Stack Gas	Proprietary MEA Process	435,450 kg/hr (960,000 lb/hr) stack gas, 0.1 MPa (15 psia), 90% CO_2 removal	437,260 kg/hr (964,000 lb/hr) stack gas, 0.1 MPa (15 psia), 90% CO_2 removal	2 (0)

ACCOUNT 3B - CO_2 COMPRESSION

Equip-ment No.	Description	Type	Case 1-1 Design Condition	Case 1-2 Design Condition	Oper Qty. (Spares)
1	CO_2 Compressor	Integrally geared, multi-stage centrifugal	564 m^3/min @ 15.3 MPa (19,900 scfm @ 2,215 psia)	566 m^3/min @ 15.3 MPa (20,000 scfm @ 2,215 psia)	4 (1)

ACCOUNT 4 - COOLING WATER SYSTEM

Equip -ment No.	Description	Type	Case 1-1 Design Condition	Case 1-2 Design Condition	Oper Qty. (Spares)
1	Circulating Water Pumps	Vertical, wet pit	128,704 lpm @ 30 m (34,000 gpm @ 100 ft)	128,704 lpm @ 30 m (34,000 gpm @ 100 ft)	4 (1)
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	11°C (51.5°F) wet bulb / 16°C (60°F) CWT / 27°C (80°F) HWT / 1435 GJ/hr (1360 MMBtu/hr) heat duty	11°C (51.5°F) wet bulb / 16°C (60°F) CWT / 27°C (80°F) HWT / 1435 GJ/hr (1360 MMBtu/hr) heat duty	1 (0)

4.5 COST ESTIMATION

The total plant cost (TPC) for the plant was estimated from bare erected costs (BEC) for equipment in June 2007 dollars. The production costs consist of several broad categories of cost elements. These cost elements include operating labor, maintenance material and labor, administrative and support labor, consumables (water and water treating chemicals, solid waste disposal cost, and fuel costs). A surcharge is added for the imported electricity necessary for the auxiliary plant loads. Capital cost estimating methodology is explained in Section 2.

4.5.1 Equipment Costing

Reformer, Shift Reactor, and PSA

The capital cost of the steam methane reformer specified for production of H_2 was based on a budgetary quotation from Krupp-Uhde to RDS. The 1998 quotation encompassed the turnkey installation of an SMR plant to produce 70 MMSCFD of hydrogen. The sulfur polisher, reformer, shift reactor, and PSA were included in the quotation. The estimate was based on U.S. Gulf Coast labor rates. The case 1-1 and 1-2 estimates were prepared by upgrading the Krupp estimate to 2007 utilizing *Chemical Engineering* Plant Cost Indices [Ref. 9] and factoring the capital estimate on the basis of hydrogen production capacity (240 MMSCFD).

Acid Gas Removal

The AGR processes for removing CO_2 from the hydrogen production plants are both proprietary MDEA and MEA systems. The cost for the MDEA system was factored from the bituminous baseline study, case 3 [Ref. 1]. The cost for the MEA system was factored from a Fluor quotation to RDS for another application of the Econamine process.

CO_2 Compression

The cost for the CO_2 compressor and drier was factored from the bituminous baseline study, case 4 [Ref. 1].

Balance of Plant

The cost of the balance of plant that constitutes the complete hydrogen production plant was based on an in-house model that has been used to develop the capital costs and economic results for many process applications. The costs attributed to balance of plant components amount to 15 percent of the installed plant equipment cost.

Contingency

A 20 percent project contingency was added to all equipment capital cost accounts for this case. Project contingencies were added to cover project uncertainty and the cost of additional equipment that could result from a more detailed design. The project contingencies represent costs that are expected to occur. A 20 percent process contingency was added for the CO_2 removal systems for this case based on the relative unproven status of the technologies at commercial scale for power plant and hydrogen production applications.

4.5.2 O&M Costs

Fixed and variable operating costs were estimated for each case. The natural gas price used for this study was $6.21/GJ ($6.55/MMBtu) on a HHV basis. All other consumable costs were assumed to match those used in the baseline reference report [Ref. 1]. An emissions value of $30/tonne of CO_2 emitted was also applied to reflect potential environmental regulations. A value of $105/MWh was applied for the power requirements for each case. This value is consistent with the cost of electricity (COE) generated in an environment where coal-based power plants are built with carbon capture and sequestration systems.

4.5.3 Cost Estimation Results

The total overnight cost (TOC) for the case 1-1 plant producing 242 MMSCFD (617 metric tons) of hydrogen (99.6 percent H_2 by volume) per day from natural gas with CO_2 capture is estimated to be $611.2 million in June 2007 dollars resulting in a first year cost of hydrogen (COH) of $2.07/kg H_2 for this case of hydrogen production from natural gas with CO_2 capture. Exhibit 4-12 and Exhibit 4-13 show the capital and operating costs for this SMR plant.

The TOC for the case 1-2 plant producing 243 MMSCFD (619 metric tons) of hydrogen (99.6 percent H_2 by volume) per day from natural gas with CO_2 capture is estimated to be $612.6 million in June 2007 dollars resulting in a COH of $2.07/kg H_2 for this case also. Exhibit 4-14 and Exhibit 4-15 show the capital and operating costs for this SMR plant.

The additional cost of CO_2 TS&M is estimated to be $0.09/kg H_2 for both cases bringing the total COH with CO_2 capture to $2.16/kg H_2. The additional costs for CO_2 TS&M are shown in Exhibit 4-16.

The small differences in the scale of the two plants are reflected in the consistency of the final COH values between these two cases.

Exhibit 4-12 Case 1-1 Capital Cost Summary

		Hydrogen Production:	kg H_2/day	616,528			
Identifier	**Component**	**Description**	**June 2007 Bare Erected Cost**	**Eng'g CM, H.O.& Fee**	**Process Contingency**	**Project Contingency**	**Total Plant Cost**
				0.1	0.2	0.2	**$1,000**
SMR-1	Methanation Reactor #1	Externally Heated SMR Reactor	$57,841	$5,784	$0	$12,725	$76,350
ZnO-1	Sulfur polisher	Zinc Oxide Bed	$243	$24	$0	$53	$321
COMP-1	Primary Air Compressor	Supply Air to SMR Reactor Burner	$861	$86	$0	$189	$1,136
WGS-1	Water Gas Shift reactor	Convert CO to H2 and CO2	$12,918	$1,292	$0	$2,842	$17,052
AGR-1	MDEA CO2 Removal Process	Remove excess CO2 from reformer product stream	$95,985	$9,599	$19,197	$24,956	$149,737
COMP-2	CO2 Compressor/Drier	Increase CO2 stream to Pipeline Pressure	$16,494	$1,649	$0	$3,629	$21,772
AGR-2	MEA CO2 Removal Process	Remove excess CO2 from reformer stack gas	$68,176	$6,818	$13,635	$17,726	$106,355
PSA-1	Pressure Swing Adsorber	Separate and Purify Hydrogen	$38,047	$3,805	$0	$8,370	$50,222
SMR B-1	SMR Additional Boiler Surface	Produce supplemental steam	$7,306	$731	$0	$1,607	$9,644
	Total Installed Equipment Costs		**$297,872**	**$29,787**	**$32,832**	**$72,098**	**$432,590**
BOP-1	Balance of Plant		$44,681	$4,468	$0	$10,815	$59,964
	TOTALS		**$342,553**	**$34,255**	**$32,832**	**$82,913**	**$492,553**

Owner's Costs	
Preproduction Costs	
6 Months All Labor	$6,409
1 Month Maintenance Materials	$985
1 Month Non-fuel Consumables	$3,019
1 Month Waste Disposal	$0
25% of 1 Months Fuel Cost at 100% CF	$5,720
2% of TPC	$9,851
Total	**$25,985**
Inventory Capital	
60 day supply of consumables at 100% CF	$795
0.5% of TPC (spare parts)	$2,463
Total	**$3,258**
Initial Cost for Catalyst and Chemicals	$1,318
Land	$900
Other Owner's Costs	$73,883
Financing Costs	$13,299
Total Overnight Costs (TOC)	**$611,195**
TASC Multiplier	1.109
Total As-Spent Cost (TASC)	**$678,100**

Exhibit 4-13 Case 1-1 Operating Cost Summary

Annual Fixed O&M Labor and Material Costs

Operators per Shift	16	Operating Labor	$6,313,507
Operator Base Rate	$34.65	Maintenance Labor	$4,925,532
Operator Labor Burden (% of Base)	30.00%	Admin & Support	$1,578,377
Labor O-H Charge Rate (% of labor)	25.00%	Property Taxes and Insurance	$9,851,063
Total Overnight Cost	**$611,195,209**	**TOTAL FIXED O&M**	**$22,668,479**

Variable O&M Operating Costs

	Consumption		Unit Rate	Unit Cost	Unit	Initial Fill Cost	Annual Variable O&M Costs
Maintenance Material							$10,639,148
	Initial Fill	/Day					
Water	0	3,849	1,000 gals/day	$1.08	1000 gal	$0	$1,367,629
Water Treatment Chemicals	0	11,240	lb/day	$0.17	lb	$0	$639,026
Reforming Catalyst (ft^3)	0	9	ft^3/day	$440.00	ft^3	$0	$1,302,950
Water Gas Shift Catalyst (ft^3)	1,574	5	ft^3/day	$498.83	ft^3	$785,355	$738,581
MDEA Solution (gal)	28,187	40	gal/day	$8.70	gal	$245,168	$113,971
MEA Solution (lb)	255,152	358	lb/day	$1.12	lb	$287,031	$132,390
Electric Power Purchased (Generated)		34	MW		MWh		
Purchased Electric Power (Revenue)	0	821	MWh/day	$105.00	MWh	$0	$28,311,444
Solid Waste Disposal	0	1	ton/day	$16.23	ton	$0	$3,603
Carbon Dioxide Emissions Tax		668	ton/day	$ 27.22	ton CO_2	$0	$5,974,186
TOTAL VARIABLE O&M						**$1,317,555**	**$49,222,930**

Annual Fuel Costs

Natural Gas Feed	2,520	tons/day	$298	ton	**$247,114,305**
			$6.55	MMBtu	
			TOTAL ANNUAL O&M COSTS		**$319,005,714**
			FIRST YEAR CAPITAL CHARGE		**$124,920,703**
Hydrogen Production	**616,528**	**kg/day**			
Plant Capacity Factor:	**90%**		**First Year H_2 COST**	**$/kg**	**$2.19**

Exhibit 4-14 Case 1-2 Capital Cost Summary

Identifier	Component	Description	June 2007 Bare Erected Cost	Eng'g CM, H.O.& Fee	Process Contingency	Project Contingency	Total Plant Cost
	Hydrogen Production:	kg H_2/day	618,936				
				0.1	0.2	0.2	$1,000
SMR-1	Methanation Reactor #1	Externally Heated SMR Reactor	$57,976	$5,798	$0	$12,755	$76,528
ZnO-1	Sulfur polisher	Zinc Oxide Bed	$244	$24	$0	$54	$322
COMP-1	Primary Air Compressor	Supply Air to SMR Reactor Burner	$863	$86	$0	$190	$1,139
WGS-1	Water Gas Shift reactor	Convert CO to H2 and CO2	$12,949	$1,295	$0	$2,849	$17,092
AGR-1	MDEA CO2 Removal Process	Remove excess CO2 from reformer product stream	$96,210	$9,621	$19,242	$25,015	$150,088
COMP-2	CO2 Compressor/Drier	Increase CO2 stream to Pipeline Pressure	$16,533	$1,653	$0	$3,637	$21,823
AGR-2	MEA CO2 Removal Process	Remove excess CO2 from reformer stack gas	$68,336	$6,834	$13,667	$17,767	$106,604
PSA-1	Pressure Swing Adsorber	Separate and Purify Hydrogen	$38,136	$3,814	$0	$8,390	$50,340
SMR B-1	SMR Additional Boiler Surface	Produce supplemental steam	$7,323	$732	$0	$1,611	$9,667
	Total Installed Equipment Costs		**$298,569**	**$29,857**	**$32,909**	**$72,267**	**$433,602**
BOP-1	Balance of Plant		$44,785	$4,479	$0	$10,840	$60,104
	TOTALS		**$343,355**	**$34,335**	**$32,909**	**$83,107**	**$493,706**

Owner's Costs	
Preproduction Costs	
6 Months All Labor	$6,414
1 Month Maintenance Materials	$987
1 Month Non-fuel Consumables	$3,031
1 Month Waste Disposal	$0
25% of 1 Months Fuel Cost at 100% CF	$5,743
2% of TPC	$9,874
Total	**$26,050**
Inventory Capital	
60 day supply of consumables at 100% CF	$798
0.5% of TPC (spare parts)	$2,469
Total	**$3,267**
Initial Cost for Catalyst and Chemicals	$1,323
Land	$900
Other Owner's Costs	$74,056
Financing Costs	$13,330
Total Overnight Costs (TOC)	**$612,632**
TASC Multiplier	1.109
Total As-Spent Cost (TASC)	**$679,694**

Exhibit 4-15 Case 1-2 Operating Cost Summary

Annual Fixed O&M Labor and Material Costs

Operators per Shift	16	Operating Labor	$6,313,507
Operator Base Rate	$34.65	Maintenance Labor	$4,937,064
Operator Labor Burden (% of Base)	30.00%	Admin & Support	$1,578,377
Labor O-H Charge Rate (% of labor)	25.00%	Property Taxes and Insurance	$9,874,127
Total Overnight Cost	**$612,632,020**	**TOTAL FIXED O&M**	**$22,703,075**

Variable O&M Operating Costs

	Initial Fill	Consumption /Day	Unit Rate	Unit Cost	Unit	Initial Fill Cost	Annual Variable O&M Costs
Maintenance Material							$10,664,058
Water	0	3,864	1,000 gals/day	$1.08	1000 gal	$0	$1,373,004
Water Treatment Chemicals	0	11,280	lb/day	$0.17	lb	$0	$641,300
Reforming Catalyst (ft^3)	0	9	ft^3/day	$440.00	ft^3	$0	$1,308,038
Water Gas Shift Catalyst (ft^3)	1,581	5	ft^3/day	$498.83	ft^3	$788,422	$741,466
MDEA Solution (gal)	28,298	40	gal/day	$8.70	gal	$246,125	$114,416
MEA Solution (lb)	256,148	360	lb/day	$1.12	lb	$288,152	$132,907
Electric Power Purchased (Generated)		34	MW		MWh		
Purchased Electric Power (Revenue)	0	824	MWh/day	$105.00	MWh	$0	$28,419,061
Solid Waste Disposal	0	1	ton/day	$16.23	ton	$0	$3,618
Carbon Dioxide Emissions Tax		671	ton/day	$ 27.22	ton CO_2	$0	$5,998,070
TOTAL VARIABLE O&M						**$1,322,700**	**$49,395,938**

Annual Fuel Costs

Natural Gas Feed	2,530	tons/day	$298	ton	**$248,102,221**
			$6.55	MMBtu	
			TOTAL ANNUAL O&M COSTS		**$320,201,234**
			FIRST YEAR CAPITAL CHARGE		**$125,214,495**
Hydrogen Production	**618,936**	**kg/day**			
Plant Capacity Factor:	**90%**		**First Year H_2 COST**	**$/kg**	**$2.19**

Exhibit 4-16 Case 1-1 and Case 1-2 Cost Estimate CO_2 TS&M

Parameter	Value	Value
TPC of Transport, million $	64.25	64.25
TPC of Storage, million $	31.16	31.23
Capital Fund for Life-Cycle CO_2 Monitoring Costs, million $	16.78	16.78
Total Capital TS&M	**112.19**	**112.26**
First Year Annual Operating Costs at 100% Capacity Factor		
Transport - Fixed O&M, million $	0.43	0.43
Storage - Variable O&M, million $	0.02	0.02
Storage - Fixed O&M, million $	0.14	0.28
Total First Year Cost CO_2 TS&M, $/kg H_2	**0.12**	**0.12**
Total First Year Cost without CO_2 TS&M, $/kg H_2 (see Exhibit 4-13 and Exhibit 4-15**)**	**2.19**	**2.19**
TOTAL First Year COH, $/kg H_2	**2.31**	**2.31**

5. CASES 2-1 AND 2-2 RESULTS: COAL GASIFICATION WITH H_2 SEPARATION BY PSA

Certain design considerations that represent a change in the original scope of work for the baseline cases are presented in this section of the report. Changes to the scope of work from the original H2A basis as requested in technical direction from NETL include: changing the design coal to Illinois #6, expanding the plant capacity to two gasifiers, adjusting the radiant cooler outlet temperature to 1250°F, and increasing the scrubber/quench blowdown to appropriately limit chloride concentration. Also, the form of recovered sulfur was changed from sulfuric acid to elemental sulfur. As appropriate, these changes are discussed below.

GE Energy Gasifier

Case 2-1 Radiant-Only: The original H2A design was based on a plant utilizing the E-Gas gasifier which has a relatively low operating pressure. This approach was satisfactory for applications which do not require high pressure syngas applications downstream. Future advanced cases to be evaluated will utilize membrane separations for hydrogen recovery that operate more effectively at higher pressure. For this reason, the revised plant design features the GE Energy slurry fed gasifier in the radiant-only mode. In this mode, the hot gas from the gasifier outlet is cooled in a radiant syngas cooler to 1,250°F before flowing through a water-filled quench chamber. This gasifier is offered commercially to operate at pressures greater than 900 psia.

Case 2-2 Full Quench: Case 2-1 was based on the GEE radiant-only gasifier which produced high pressure steam for power production, albeit at a higher capital cost. Case 2-2 differs from case 2-1 by eliminating the radiant cooler and is based on the GEE gasification technology with the quench gasifier option operating at approximately 965 psia. In this mode, the hot gas from the gasifier outlet is cooled by flowing through a water-filled quench chamber without raising high pressure steam. The objective of this case is to test the assumption that this configuration will not produce enough power to supply auxiliary loads.

CO_2 Recovery

Previous H2A designs fired the off-gas from the PSA with oxygen in an oxy-combustion mode to recover maximum CO_2 from the plant. The design utilized shift reactors which had a CO conversion of about 75 percent and it was necessary to fire the PSA off-gas with oxygen to recover the CO_2 and achieve the 90 percent recovery target. The more recent designs utilize equilibrium shift reactors in series to achieve CO conversions in the 80 percent range. As such, the syngas contains sufficient CO_2 from both the raw gas and the shift reaction to achieve the 90 percent recovery target without firing the off-gas in an oxy-combustion mode.

Recovery of elemental sulfur, rather than H_2SO_4

The original H2A design produced sulfuric acid as a byproduct. Since sulfur is less expensive to produce and transport, the revised design for this plant and subsequent plants in this series will produce elemental sulfur as a byproduct.

Limit chloride concentration

Case 2-1 Radiant-Only: The high chlorine content of Illinois No. 6 coal (0.29 percent as received) raises concern over potential chloride corrosion in the quench chamber and downstream piping. The shift catalyst is not affected by the presence of chloride or sulfur. The raw syngas leaves the gasifier syngas cooler and enters the quench chamber at 1,250°F with a hydrogen chloride (HCl) concentration of 768 ppmv. 756,000 lb/hr of water enters the quench chamber and 378,000 lb/hr of water are evaporated to the gas while cooling the gas to 415°F. 95 percent of the chloride is removed from the syngas in the quench chamber, necessitating a high water flow rate to achieve an equilibrium HCl concentration of 794 ppmw to avoid quench chamber material issues. The remaining chloride in the syngas eventually drops out with condensed water downstream. A total of 472,000 lb/hr knockout from the downstream condensers is recycled back to the quench chamber to augment the quench inventory to control chloride concentration.

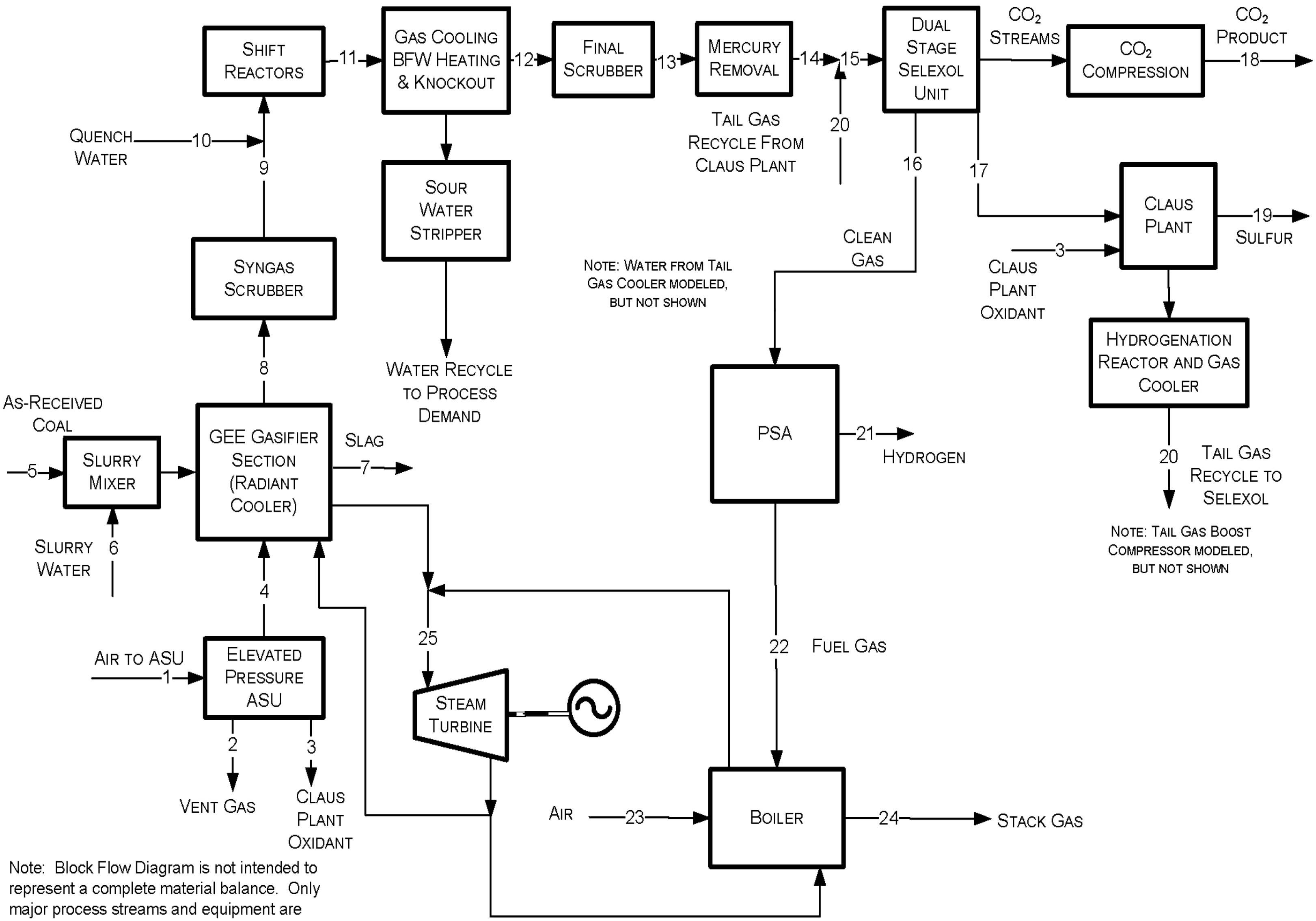
Case 2-2 Full Quench: The raw syngas leaves the gasifier and enters the quench chamber at 2,250°F with a hydrogen chloride (HCl) concentration of 768 ppmv. 1,665,000 lb/hr of water enters the quench chamber and 878,000 lb/hr of water are evaporated to the gas while cooling the gas to 520°F. 95 percent of the chloride is removed from the syngas in the quench chamber, necessitating a high water flow rate to achieve an equilibrium HCl concentration of 804 ppmw to avoid quench chamber material issues. The remaining chloride in the syngas eventually drops out with condensed water downstream. A total of 1,031,000 lb/hr knockout from the downstream condensers is recycled back to the quench chamber to augment the quench inventory to control chloride concentration.

The quench outlet streams are sent to a treatment plant where calcium hydroxide is used to precipitate calcium chloride as a byproduct. The stream then goes to a sour water stripper and the stripper bottoms are recycled to satisfy the water makeup for the quench.

A block flow diagram of the case 2-1 process is shown in Exhibit 5-1 with the corresponding stream table shown in Exhibit 5-2.

A block flow diagram of the case 2-2 process is shown in Exhibit 5-3 with the corresponding stream table shown in Exhibit 5-4.

Exhibit 5-1 Case 2-1 Block Flow Diagram: Baseline Coal to Hydrogen 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 5-2 Case 2-1 Stream Table: Baseline Coal to Hydrogen with CO_2 Capture

	1	2	3	4	5	6	7	8	9	10	11	12	13
V-L Mole Fraction													
Ar	0.0092	0.0045	0.0318	0.0318	0.0000	0.0000	0.0000	0.0086	0.0068	0.0000	0.0054	0.0071	0.0071
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0011	0.0009	0.0000	0.0007	0.0009	0.0009
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3576	0.2816	0.0000	0.0060	0.0077	0.0078
CO_2	0.0003	0.0007	0.0000	0.0000	0.0000	0.0000	0.0000	0.1380	0.1087	0.0000	0.3082	0.4011	0.4019
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0001	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3406	0.2682	0.0000	0.4364	0.5679	0.5691
H_2O	0.0099	0.0182	0.0000	0.0000	0.0000	0.9993	0.0000	0.1369	0.3202	1.0000	0.2324	0.0012	0.0012
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0073	0.0057	0.0000	0.0047	0.0061	0.0061
N_2	0.7732	0.9535	0.0178	0.0178	0.0000	0.0000	0.0000	0.0070	0.0055	0.0000	0.0044	0.0058	0.0058
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0007	0.0000	0.0019	0.0020	0.0000	0.0016	0.0019	0.0000
O_2	0.2074	0.0231	0.9504	0.9504	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	0.0000
S8	0.0000	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	0.0000	1.0000	0.0000	0.9992	0.9998	1.0000	0.9999	0.9998	1.0000
V-L Flowrate (kg_{mol}/hr)	27,379	12,598	99	5,526	0	5,037	0	23,122	29,355	7,134	36,489	28,036	27,978
V-L Flowrate (kg/hr)	790,073	352,613	3,198	177,828	0	90,748	0	465,243	577,297	128,519	705,816	553,532	552,441
Solids Flowrate (kg/hr)	0	0	0	0	220,904	0	24,237	0	0	0	0	0	0
Temperature (°C)	15	15	32	32	15	142	1,316	677	206	288	240	35	35
Pressure (MPa, abs)	0.10	0.11	0.86	0.86	0.10	5.79	5.62	5.55	5.52	5.52	5.41	5.17	5.14
Enthalpy (kJ/kg)^A	30.23	33.30	26.67	26.67	---	537.54	---	1,424.65	1,069.00	2,918.18	941.96	36.06	37.05
Density (kg/m^3)	1.2	1.3	11.0	11.0	---	872.0	---	14.0	27.2	25.6	24.8	41.0	40.8
V-L Molecular Weight	28.857	27.989	32.181	32.181	---	18.015	---	20.121	19.666	18.015	19.343	19.744	19.746
V-L Flowrate (lb_{mol}/hr)	60,360	27,774	219	12,183	0	11,106	0	50,976	64,717	15,728	80,444	61,809	61,681
V-L Flowrate (lb/hr)	1,741,813	777,378	7,050	392,044	0	200,064	0	1,025,686	1,272,722	283,335	1,556,057	1,220,328	1,217,923
Solids Flowrate (lb/hr)	0	0	0	0	487,011	0	53,433	0	0	0	0	0	0
Temperature (°F)	59	58	90	90	59	287	2,400	1,250	403	550	463	95	95
Pressure (psia)	14.7	16.4	125.0	125.0	14.7	840.0	815.0	805.0	800.0	800.0	785.0	750.0	745.0
Enthalpy (Btu/lb)^A	13.0	14.3	11.5	11.5	---	231.1	---	612.5	459.6	1,254.6	405.0	15.5	15.9
Density (lb/ft^3)	0.076	0.083	0.687	0.687	---	54.436	---	0.871	1.699	1.597	1.550	2.562	2.544

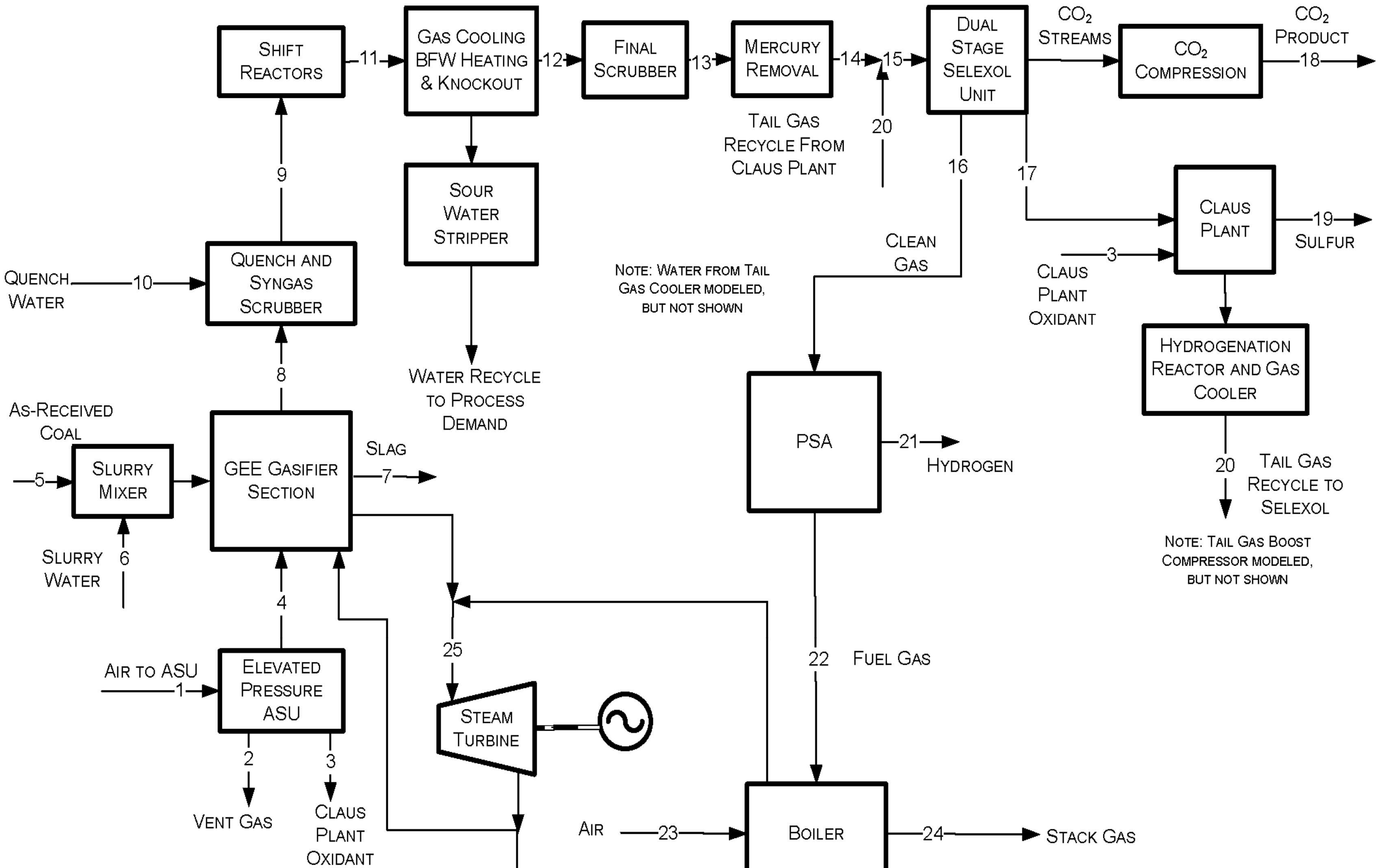
A - Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 5-2 Case 2-1 Stream Table: Baseline Coal to Hydrogen with CO_2 Capture (continued)

	14	15	16	17	18	19	20	21	22	23	24	25
V-L Mole Fraction												
Ar	0.0071	0.0071	0.0115	0.0018	0.0002	0.0000	0.0103	0.0000	0.0427	0.0094	0.0217	0.0000
CH_4	0.0009	0.0009	0.0015	0.0004	0.0000	0.0000	0.0000	0.0000	0.0055	0.0000	0.0000	0.0000
CO	0.0078	0.0077	0.0124	0.0022	0.0002	0.0000	0.0064	0.0000	0.0462	0.0000	0.0000	0.0000
CO_2	0.4019	0.4054	0.0502	0.5220	0.9948	0.0000	0.6471	0.0000	0.1864	0.0003	0.0801	0.0000
COS	0.0000	0.0000	0.0000	0.0002	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.5691	0.5648	0.9136	0.1030	0.0048	0.0000	0.2666	1.0000	0.6791	0.0000	0.0000	0.0000
H_2O	0.0012	0.0012	0.0001	0.0225	0.0000	0.0000	0.0017	0.0000	0.0003	0.0104	0.2396	1.0000
H_2S	0.0061	0.0061	0.0000	0.3470	0.0000	0.0000	0.0035	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0058	0.0066	0.0107	0.0008	0.0000	0.0000	0.0645	0.0000	0.0399	0.7722	0.6206	0.0000
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2077	0.0381	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
S8	0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	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	1.0000
V-L Flowrate (kg_{mol}/hr)	27,978	28,379	17,434	496	10,426	0	401	12,743	4,691	11,004	13,994	24,144
V-L Flowrate (kg/hr)	552,441	565,102	90,337	17,666	456,694	0	12,661	25,689	64,649	317,515	382,163	434,960
Solids Flowrate (kg/hr)	0	0	0	0	0	5,525	0	0	0	0	0	0
Temperature (°C)	35	35	35	48	51	178	38	35	-7	15	138	538
Pressure (MPa, abs)	5.10	5.10	5.10	0.16	15.27	0.12	5.5	5.102	0.531	0.101	0.105	12.512
Enthalpy (kJ/kg)[A]	37.05	36.38	195.26	74.69	-162.30	---	7.0	512.283	-15.178	31.074	551.044	3,441.857
Density (kg/m^3)	40.5	40.9	10.1	2.2	641.8	5,280.0	75.8	3.9	3.3	1.2	0.8	36.5
V-L Molecular Weight	19.746	19.913	5.182	35.591	43.804	---	32	2.016	13.781	28.854	27.309	18.015
V-L Flowrate (lb_{mol}/hr)	61,681	62,565	38,436	1,094	22,985	0	884	28,094	10,342	24,260	30,852	53,228
V-L Flowrate (lb/hr)	1,217,923	1,245,837	199,159	38,947	1,006,837	0	27,913	56,634	142,526	700,000	842,526	958,923
Solids Flowrate (lb/hr)	0	0	0	0	0	12,181	0	0	0	0	0	0
Temperature (°F)	95	95	95	119	124	352	100	95	20	59	280	1,000
Pressure (psia)	740.0	740.0	740.0	23.7	2,214.7	17.3	799.5	740.0	77.0	14.7	15.2	1,814.7
Enthalpy (Btu/lb)[A]	15.9	15.6	83.9	32.1	-69.8	---	3.0	220.2	-6.5	13.4	236.9	1,479.7
Density (lb/ft^3)	2.528	2.552	0.631	0.137	40.068	329.622	5	0.245	0.206	0.076	0.052	2.280

A - Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 5-3 Case 2-2 Block Flow Diagram: Baseline Coal to Hydrogen 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 5-4 Case 2-2 Stream Table: Baseline Coal to Hydrogen with CO_2 Capture

	1	2	3	4	5	6	7	8	9	10	11	12	13
V-L Mole Fraction													
Ar	0.0092	0.0045	0.0318	0.0318	0.0000	0.0000	0.0000	0.0085	0.0052	0.0000	0.0052	0.0070	0.0070
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0012	0.0007	0.0000	0.0007	0.0010	0.0010
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3590	0.2192	0.0000	0.0059	0.0080	0.0080
CO_2	0.0003	0.0007	0.0000	0.0000	0.0000	0.0000	0.0000	0.1366	0.0836	0.0000	0.2970	0.3999	0.4010
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0001	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.3420	0.2088	0.0000	0.4221	0.5684	0.5699
H_2O	0.0099	0.0183	0.0000	0.0000	0.0000	0.9995	0.0000	0.1355	0.4714	0.9995	0.2580	0.0012	0.0012
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0073	0.0045	0.0000	0.0046	0.0061	0.0061
N_2	0.7732	0.9533	0.0178	0.0178	0.0000	0.0000	0.0000	0.0070	0.0043	0.0000	0.0043	0.0057	0.0058
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0005	0.0000	0.0019	0.0020	0.0005	0.0020	0.0025	0.0000
O_2	0.2074	0.0232	0.9504	0.9504	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	0.0000
S8	0.0000	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	0.0000	1.0000	0.0000	0.9992	0.9999	1.0000	0.9999	0.9999	1.0000
V-L Flowrate (kg_{mol}/hr)	27,248	12,467	107	5,491	0	5,037	0	23,117	37,845	22,116	37,845	28,106	28,032
V-L Flowrate (kg/hr)	786,301	348,938	3,451	176,711	0	90,741	0	464,105	729,291	398,414	729,291	553,834	552,497
Solids Flowrate (kg/hr)	0	0	0	0	220,889	0	24,235	0	0	0	0	0	0
Temperature (°C)	15	15	32	32	15	146	1,316	1,316	227	188	249	35	35
Pressure (MPa, abs)	0.10	0.11	0.86	0.86	0.10	5.79	5.62	5.62	5.58	8.27	5.45	5.21	5.17
Enthalpy (kJ/kg)[A]	30.23	33.30	26.67	26.67	---	558.85	---	2,632.17	1,485.52	764.86	1,022.09	36.21	36.96
Density (kg/m^3)	1.2	1.3	11.0	11.0	---	867.0	---	8.5	25.9	816.7	24.5	41.2	41.0
V-L Molecular Weight	28.857	27.989	32.181	32.181	---	18.015	---	20.076	19.270	18.015	19.270	19.705	19.710
V-L Flowrate (lb_{mol}/hr)	60,072	27,485	236	12,106	0	11,105	0	50,965	83,435	48,757	83,435	61,964	61,800
V-L Flowrate (lb/hr)	1,733,496	769,276	7,607	389,581	0	200,050	0	1,023,177	1,607,812	878,352	1,607,812	1,220,994	1,218,046
Solids Flowrate (lb/hr)	0	0	0	0	486,976	0	53,430	0	0	0	0	0	0
Temperature (°F)	59	58	90	90	59	295	2,400	2,400	440	371	481	95	95
Pressure (psia)	14.7	16.4	125.0	125.0	14.7	840.0	815.0	815.0	810.0	1,200.0	790.0	755.0	750.0
Enthalpy (Btu/lb)[A]	13.0	14.3	11.5	11.5	---	240.3	---	1,131.6	638.7	328.8	439.4	15.6	15.9
Density (lb/ft^3)	0.076	0.083	0.687	0.687	---	54.126	---	0.529	1.616	50.988	1.528	2.574	2.557

A - Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 5-4 Case 2-2 Stream Table: Baseline Coal to Hydrogen with CO_2 Capture (continued)

	14	15	16	17	18	19	20	21	22	23	24	25
V-L Mole Fraction												
Ar	0.0070	0.0071	0.0114	0.0018	0.0002	0.0000	0.0104	0.0000	0.0422	0.0094	0.0216	0.0000
CH_4	0.0010	0.0010	0.0015	0.0004	0.0001	0.0000	0.0000	0.0000	0.0056	0.0000	0.0000	0.0000
CO	0.0080	0.0080	0.0128	0.0023	0.0002	0.0000	0.0061	0.0000	0.0475	0.0000	0.0000	0.0000
CO_2	0.4010	0.4042	0.0499	0.5191	0.9947	0.0000	0.6162	0.0000	0.1853	0.0003	0.0805	0.0000
COS	0.0000	0.0000	0.0000	0.0002	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.5699	0.5655	0.9132	0.1028	0.0048	0.0000	0.2768	1.0000	0.6779	0.0000	0.0000	0.0000
H_2O	0.0012	0.0012	0.0001	0.0220	0.0000	0.0000	0.0016	0.0000	0.0003	0.0104	0.2403	1.0000
H_2S	0.0061	0.0062	0.0000	0.3505	0.0000	0.0000	0.0091	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0058	0.0069	0.0111	0.0008	0.0000	0.0000	0.0799	0.0000	0.0412	0.7722	0.6204	0.0000
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2077	0.0372	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
S8	0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	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	1.0000
V-L Flowrate (kg_{mol}/hr)	28,032	28,455	17,511	499	10,423	0	423	12,793	4,718	11,004	14,011	12,980
V-L Flowrate (kg/hr)	552,497	565,529	90,833	17,752	456,545	0	13,033	25,789	65,044	317,515	382,558	233,831
Solids Flowrate (kg/hr)	0	0	0	0	0	5,540	0	0	0	0	0	0
Temperature (°C)	35	35	35	48	51	179	38	35	-7	15	138	538
Pressure (MPa, abs)	5.14	5.14	5.14	0.16	15.27	0.12	5.5	5.137	0.531	0.101	0.105	12.512
Enthalpy (kJ/kg)^A	36.96	36.32	194.98	74.17	-162.26	---	9.0	512.243	-15.188	31.074	552.341	3,441.857
Density (kg/m^3)	40.7	41.1	10.2	2.2	641.7	5,277.8	73.4	3.9	3.3	1.2	0.8	36.5
V-L Molecular Weight	19.710	19.875	5.187	35.573	43.803	---	31	2.016	13.787	28.854	27.305	18.015
V-L Flowrate (lb_{mol}/hr)	61,800	62,732	38,605	1,100	22,978	0	932	28,204	10,401	24,260	30,888	28,615
V-L Flowrate (lb/hr)	1,218,046	1,246,778	200,252	39,136	1,006,510	0	28,732	56,855	143,397	700,000	843,397	515,508
Solids Flowrate (lb/hr)	0	0	0	0	0	12,214	0	0	0	0	0	0
Temperature (°F)	95	95	95	119	124	354	100	95	20	59	280	1,000
Pressure (psia)	745.0	745.0	745.0	23.7	2,214.7	17.3	799.5	745.0	77.0	14.7	15.2	1,814.7
Enthalpy (Btu/lb)^A	15.9	15.6	83.8	31.9	-69.8	---	3.9	220.2	-6.5	13.4	237.5	1,479.7
Density (lb/ft^3)	2.540	2.564	0.636	0.137	40.059	329.483	5	0.246	0.206	0.076	0.052	2.280

A - Reference conditions are 32.02 F & 0.089 PSIA

5.1 COMPONENT TABLES

The component tables below contain general specifications and overall process data for the major process component systems for cases 2-1 and 2-2.

GASIFIER

Case	2-1	2-2
Technology Type	GE Energy Slurry Feed, Oxygen Blown Radiant-Only Gasifier	GE Energy Slurry Feed, Oxygen Blown Quench Gasifier
Basis for Design and Performance	Vendor furnished data – upgrade from Polk IGCC	Vendor furnished data – upgrade from Polk IGCC
Operating Conditions Inlet – Coal Inlet – Slurry Water	220,904 kg/hr (487,011 lb/hr) Illinois No. 6 90,748 kg/hr (200,064 lb/hr) 5.8 MPa (840.0 psia), 142°C (287°F) Slurry Water	220,889 kg/hr (486,976 lb/hr) Illinois No. 6 90,741 kg/hr (200,050 lb/hr) 5.8 MPa (840.0 psia), 146°C (295°F) Slurry Water
Operating Conditions Outlet – Gas	465,243 kg/hr (1,025,686 lb/hr) raw syngas 5.6 MPa (815.0 psia), 1,316°C (2,400°F)	464,105 kg/hr (1,023,177 lb/hr) raw syngas 5.6 MPa (815.0 psia), 1,316°C (2,400°F)
Assumed or Specified Performance Characteristics	Operating pressure, slurry concentration, oxygen/coal ratio, cold gas efficiency, carbon conversion, outlet temperature and syngas composition, all based on proprietary information provided by vendor	Operating pressure, slurry concentration, oxygen/coal ratio, cold gas efficiency, carbon conversion, outlet temperature and syngas composition, all based on proprietary information provided by vendor
Calculated Performance Characteristics	Overall heat and material balance modeled from vendor information	Overall heat and material balance modeled from vendor information
Contaminant Removed, %	N/A	N/A
Assumptions Regarding Anticipated Application Issues	Gasifier performance is based on projected technology improvements beyond current Polk IGCC operation.	Gasifier performance is based on projected technology improvements beyond current Polk IGCC operation.

SYNGAS COOLER

Case	2-1	2-2
Technology Type	High pressure radiant heat exchanger followed with water quench	N/A
Basis for Design and Performance	Vendor data – future design	
Operating Conditions Inlet – Gas	465,243 kg/hr (1,025,686 lb/hr) raw syngas 5.6 MPa (815.0 psia), 1,316°C (2,400°F)	
Operating Conditions Outlet – Gas	465,243 kg/hr (1,025,686 lb/hr) raw syngas 5.6 MPa (805.0 psia), 677°C (1,250°F)	
Assumed or Specified Performance Characteristics	BFW pressure based on design IGCC	
Calculated Performance Characteristics	Steam generation 562 GJ/hr (532 MMBtu/hr) calculated from syngas flow and temperatures	
Contaminant Removed, %	N/A	
Assumptions Regarding Anticipated Application Issues	Commercial gasifier-based	

GASIFIER QUENCH AND SYNGAS SCRUBBER

Case	2-1	2-2
Technology Type	Gasifier Quench Chamber and Counter-Current Spray Tower	Gasifier Quench Chamber and Counter-Current Spray Tower
Basis for Design and Performance	Vendor data – future design	Vendor data – future design
Operating Conditions Inlet – Gas	465,243 kg/hr (1,025,686 lb/hr) raw syngas 5.6 MPa (805.0 psia), 677°C (1,250°F)	464,105 kg/hr (1,023,177 lb/hr) raw syngas 5.6 MPa (815.0 psia), 1,316°C (2,400°F)
Operating Conditions Outlet – Gas	636,837 kg/hr (1,403,986 lb/hr) raw syngas 5.6 MPa (805.0 psia), 232°C (450°F)	862,519 kg/hr (1,901,529 lb/hr) raw syngas 5.6 MPa (815.0 psia), 288°C (550°F)
Assumed or Specified Performance Characteristics	Quench and spray tower equilibrium with exiting gas stream. Particulate, chloride, and alkali removal	Quench and spray tower equilibrium with exiting gas stream. Particulate, chloride, and alkali removal
Calculated Performance Characteristics	Calculated 99.9% particulate removal, 90% HCl and complete alkali removal	Calculated 99.9% particulate removal, 90% HCl and complete alkali removal
Contaminant Removed, %	Particulate, 99.9%; chloride, 90%	Particulate, 99.9%; chloride, 90%
Assumptions Regarding Anticipated Application Issues	Commercial gasifier-based	Commercial gasifier-based

ASU

Case	2-1	2-2
Technology Type	Cryogenic Distillation	Cryogenic Distillation
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions Inlet – Gas	789,368 kg/hr (1,740,258 lb/hr) Air 1.3 MPa (189.5 psia), 38°C (100°F)	785,599 kg/hr (1,731,949 lb/hr) Air 1.3 MPa (189.5 psia), 38°C (100°F)
Operating Conditions Outlet – Gas	181,026 kg/hr (399,094 lb/hr) 95% Oxygen 0.9 MPa (125.0 psia), 32°C (90°F) 255,729 kg/hr (563,786 lb/hr) 100% Nitrogen 1.3 MPa (182.0 psia), 10°C (50°F)	180,161 kg/hr (397,188 lb/hr) 95% Oxygen 0.9 MPa (125.0 psia), 32°C (90°F) 256,499 kg/hr (565,484 lb/hr) 100% Nitrogen 1.3 MPa (182.0 psia), 10°C (50°F)
Assumed or Specified Performance Characteristics	Specified plant production to produce 95% purity oxygen, and air compressor and oxygen compressor performance. There is no gas turbine integration.	Specified plant production to produce 95% purity oxygen, and air compressor and oxygen compressor performance. There is no gas turbine integration.
Calculated Performance Characteristics	Modeled performance based on gasifier and sulfur plant requirements.	Modeled performance based on gasifier and sulfur plant requirements.
Contaminant Removed, %	100% inlet CO_2 with adsorption	100% inlet CO_2 with adsorption
Assumptions Regarding Anticipated Application Issues	Commercial plant – no issues.	Commercial plant – no issues.

WATER GAS SHIFT REACTOR

Case	2-1	2-2
Technology Type	Haldor Topsoe Two Stage Shift Catalyst	Haldor Topsoe Two Stage Shift Catalyst
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions Inlet – Gas Inlet Steam	705,816 kg/hr (1,556,057 lb/hr) syngas 5.5 MPa (795.0 psia), 225°C (436°F) 128,519 kg/hr (283,335 lb/hr) steam 5.5 MPa (800.0 psia), 288°C (550°F)	729,291 kg/hr (1,607,812 lb/hr) syngas 5.5 MPa (800.0 psia), 232°C (450°F) 0 kg/hr (0 lb/hr) steam 5.5 MPa (800.0 psia), 288°C (550°F)
Operating Conditions Outlet - Syngas	705,816 kg/hr (1,556,057 lb/hr) syngas 5.4 MPa (780.0 psia), 235°C (456°F)	729,291 kg/hr (1,607,812 lb/hr) syngas 5.4 MPa (785.0 psia), 244°C (470°F)
Assumed or Specified Performance Characteristics	Specified shift catalyst which also promotes COS hydrolysis; Assume minimum H_2O/CO ratio of 2.0 at inlet; Interstage cooling; Assume two stages required to achieve ~80 percent CO conversion	Specified shift catalyst which also promotes COS hydrolysis; Assume minimum H_2O/CO ratio of 2.0 at inlet; Interstage cooling; Assume two stages required to achieve ~80 percent CO conversion
Calculated Performance Characteristics	Modeled WGS reaction based on equilibrium to ~80% CO conversion	Modeled WGS reaction based on equilibrium to ~80% CO conversion
Contaminant Removed, %	N/A	N/A
Assumptions Regarding Anticipated Application Issues	Commercial design; no issues	Commercial design; no issues

PRE-MERCURY REMOVAL SYNGAS SCRUBBER

Case	2-1	2-2
Technology Type	Venturi Water Spray Scrubber	Venturi Water Spray Scrubber
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions Inlet - Syngas	553,532 kg/hr (1,220,328 lb/hr) syngas 5.2 MPa (750.0 psia), 35°C (95°F)	553,834 kg/hr (1,220,994 lb/hr) syngas 5.2 MPa (755.0 psia), 35°C (95°F)
Operating Conditions Outlet - Syngas	552,441 kg/hr (1,217,923 lb/hr) syngas 5.1 MPa (745.0 psia), 35°C (95°F)	552,497 kg/hr (1,218,046 lb/hr) syngas 5.2 MPa (750.0 psia), 35°C (95°F)
Assumed or Specified Performance Characteristics	Spray tower equilibrium with exiting gas stream. Chloride and ammonia removal	Spray tower equilibrium with exiting gas stream. Chloride and ammonia removal
Calculated Performance Characteristics	Complete chloride and ammonia removal	Complete chloride and ammonia removal
Contaminant Removed, %	Chloride and ammonia, 99.9%	Chloride and ammonia, 99.9%
Assumptions Regarding Anticipated Application Issues	Commercial, no issues	Commercial, no issues

MERCURY REMOVAL

Case	2-1	2-2
Technology Type	Low temperature Activated Carbon	Low temperature Activated Carbon
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions Inlet - Syngas	552,441 kg/hr (1,217,923 lb/hr) syngas 5.1 MPa (745.0 psia), 35°C (95°F)	552,497 kg/hr (1,218,046 lb/hr) syngas 5.2 MPa (750.0 psia), 35°C (95°F)
Operating Conditions Outlet - Syngas	565,102 kg/hr (1,245,837 lb/hr) syngas 5.1 MPa (740.0 psia), 35°C (95°F)	565,529 kg/hr (1,246,778 lb/hr) syngas 5.1 MPa (745.0 psia), 35°C (95°F)
Assumed or Specified Performance Characteristics	95% performance based on Eastman Chemical experience; Specified gas velocity, 20 sec. retention time, operating temperature, 69 kPa drop (10 psi), pressure drop	95% performance based on Eastman Chemical experience; Specified gas velocity, 20 sec. retention time, operating temperature, 69 kPa drop (10 psi), pressure drop
Calculated Performance Characteristics	Model results reflect design assumptions	Model results reflect design assumptions
Contaminant Removed, %	Mercury, 95% removal	Mercury, 95% removal
Assumptions Regarding Anticipated Application Issues	Commercial, no issues	Commercial, no issues

SELEXOL H_2S REMOVAL

Case	2-1	2-2
Technology Type	Refrigerated Selexol Physical solvent	Refrigerated Selexol Physical solvent
Basis for Design and Performance	Vendor Data – Commercial Design	Vendor Data – Commercial Design
Operating Conditions Inlet - Syngas	565,102 kg/hr (1,245,837 lb/hr) syngas 5.1 MPa (740.0 psia), 35°C (95°F)	565,529 kg/hr (1,246,778 lb/hr) syngas 5.1 MPa (745.0 psia), 35°C (95°F)
Operating Conditions Regenerated Sulfur Stream	17,666 kg/hr (38,947 lb/hr) • 52% CO_2 • 2% H_2O • 35% H_2S • 9% H_2 0.2 MPa (23.7 psia), 49°C (119°F)	17,752 kg/hr (39,136 lb/hr) • 52% CO_2 • 2% H_2O • 35% H_2S • 9% H_2 0.2 MPa (23.7 psia), 49°C (119°F)
Assumed or Specified Performance Characteristics	Specified 99.9% specific to H_2S	Specified 99.9% specific to H_2S
Calculated Performance Characteristics	Model results reflect design assumptions	Model results reflect design assumptions
Contaminant Removed, %	H_2S, 99.9%	H_2S, 99.9%
Assumptions Regarding Anticipated Application Issues	Vendor design specifically provided for GEE syngas conditions	Vendor design specifically provided for GEE syngas conditions

SELEXOL CO_2 REMOVAL

Case	2-1	2-2
Technology Type	Refrigerated Selexol Physical solvent	Refrigerated Selexol Physical solvent
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions Inlet - Syngas	1,022,201 kg/hr (2,253,567 lb/hr) syngas 5.1 MPa (740.0 psia), 35°C (95°F)	1,022,474 kg/hr (2,254,169 lb/hr) syngas 5.1 MPa (745.0 psia), 35°C (95°F)
Operating Conditions Outlet - CO_2	457,099 kg/hr (1,007,730 lb/hr) CO_2 0.9 MPa (135.0 psia), 16°C (60°F) 17,666 kg/hr (38,947 lb/hr) Claus Feed Gas 0.2 MPa (23.7 psia), 48°C (119°F)	456,945 kg/hr (1,007,390 lb/hr) CO_2 0.9 MPa (135.0 psia), 16°C (60°F) 17,752 kg/hr (39,136 lb/hr) Claus Feed Gas 0.2 MPa (23.7 psia), 48°C (119°F)
Assumed or Specified Performance Characteristics	Specified 91.7% CO_2 removal to achieve overall system reduction of 90%	Specified 91.5% CO_2 removal to achieve overall system reduction of 90%
Calculated Performance Characteristics	Model results reflect design assumptions	Model results reflect design assumptions
Contaminant Removed, %	91.7% CO_2 from syngas	91.5% CO_2 from syngas
Assumptions Regarding Anticipated Application Issues	Vendor design specifically provided for GEE syngas conditions	Vendor design specifically provided for GEE syngas conditions

CO_2 COMPRESSION

Case	2-1	2-2
Technology Type	Multi-Stage Integral Gear Compressor	Multi-Stage Integral Gear Compressor
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions Inlet - CO_2	457,099 kg/hr (1,007,730 lb/hr) CO_2 0.9 MPa (135.0 psia), 16°C (60°F)	456,945 kg/hr (1,007,390 lb/hr) CO_2 0.9 MPa (135.0 psia), 16°C (60°F)
Operating Conditions Outlet - CO_2	456,694 kg/hr (1,006,837 lb/hr) CO_2 15.3 MPa (2,214.7 psia), 51°C (124°F)	456,545 kg/hr (1,006,510 lb/hr) CO_2 15.3 MPa (2,214.7 psia), 51°C (124°F)
Assumed or Specified Performance Characteristics	Assumed 86% polytropic efficiency with intercooled stage	Assumed 86% polytropic efficiency with intercooled stage
Calculated Performance Characteristics	Model results reflect design assumptions	Model results reflect design assumptions
Contaminant Removed, %	Dehydrated to –40º dew point	Dehydrated to –40º dew point
Assumptions Regarding Anticipated Application Issues	Commercial, no issues	Commercial, no issues

CLAUS SULFUR RECOVERY

Case	2-1	2-2
Technology Type	Oxygen Enriched Claus Plant with Recycled Tail Gas to Gasifier	Oxygen Enriched Claus Plant with Recycled Tail Gas to Gasifier
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions Inlet - Gas Inlet - Oxygen	18,955 kg/hr (41,789 lb/hr) • 52% CO_2 • 2% H_2O • 35% H_2S 0.2 MPa (23.7 psia), 48°C (119°F)	19,498 kg/hr (42,986 lb/hr) • 52% CO_2 • 2% H_2O • 35% H_2S 0.2 MPa (23.7 psia), 48°C (119°F)
Operating Conditions Outlet -Sulfur	5,525 kg/hr (12,181 lb/hr) sulfur 0.1 MPa (17.3 psia), 178°C (352°F)	5,540 kg/hr (12,214 lb/hr) sulfur 0.1 MPa (17.3 psia), 179°C (354°F)
Assumed or Specified Performance Characteristics	Assumed >97% conversion; 3 trains	Assumed >97% conversion; 3 trains
Calculated Performance Characteristics	Model calculated sulfur production and tail gas composition based on design assumptions	Model calculated sulfur production and tail gas composition based on design assumptions
Contaminant Removed, %	N/A	N/A
Assumptions Regarding Anticipated Application Issues	Commercial, no issues	Commercial, no issues

PRESSURE SWING ADSORBER

Case	**2-1**	**2-2**
Technology Type	Pressure Swing Adsorption	Pressure Swing Adsorption
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions Inlet - Syngas	90,337 kg/hr (199,159 lb/hr) Syngas 5.1 MPa (740.0 psia), 35°C (95°F)	90,833 kg/hr (200,252 lb/hr) Syngas 5.1 MPa (745.0 psia), 35°C (95°F)
Operating Conditions Outlet - H_2 Outlet – Off-Gas	25,689 kg/hr (56,634 lb/hr) H_2 5.1 MPa (740.0 psia), 35°C (95°F) 64,649 kg/hr (142,526 lb/hr) Off Gas 0.5 MPa (77.0 psia), -7°C (20°F)	25,789 kg/hr (56,855 lb/hr) H_2 5.1 MPa (745.0 psia), 35°C (95°F) 65,044 kg/hr (143,397 lb/hr) Off Gas 0.5 MPa (77.0 psia), -7°C (20°F)
Assumed or Specified Performance Characteristics	PSA operates at 80% hydrogen removal efficiency. Off-gas is sent to auxiliary boiler as fuel	PSA operates at 80% hydrogen removal efficiency. Off-gas is sent to auxiliary boiler as fuel
Calculated Performance Characteristics	>99.9 % Purity H_2	>99.9 % Purity H_2
Contaminant Removed, %	N/A	N/A
Assumptions Regarding Anticipated Application Issues	Commercial, no issues	Commercial, no issues

OFF-GAS FIRED BOILER

Case	2-1	2-2
Technology Type	Sub-Critical Drum	Sub-Critical Drum
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions Inlet - Flue Gas Inlet Air	64,649 kg/hr (142,526 lb/hr) waste gas 0.5 MPa (77.0 psia), -7°C (20°F) 317,515 kg/hr (700,000 lb/hr) air 0.1 MPa (14.7 psia), 15°C (59°F)	65,044 kg/hr (143,397 lb/hr) waste gas 0.5 MPa (77.0 psia), -7°C (20°F) 317,515 kg/hr (700,000 lb/hr) air 0.1 MPa (14.7 psia), 15°C (59°F)
Operating Conditions Outlet -Flue Gas	382,163 kg/hr (842,526 lb/hr) stack gas 0.1 MPa (15.2 psia), 138°C (280°F)	382,558 kg/hr (843,397 lb/hr) stack gas 0.1 MPa (15.2 psia), 138°C (280°F)
Assumed or Specified Performance Characteristics	Specified BFW pressure and steam from syngas cooler	Specified BFW pressure and steam from syngas cooler
Calculated Performance Characteristics	796 GJ/hr (754 MMBtu/hr)	801 GJ/hr (759 MMBtu/hr)
Contaminant Removed, %	N/A	N/A
Assumptions Regarding Anticipated Application Issues	Commercial, no issues	Commercial, no issues

STEAM TURBINE GENERATOR AND AUXILIARIES

Case	2-1	2-2
Technology Type	GE Steam Turbine	GE Steam Turbine
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions - Inlet	Multi-pressure steam	Multi-pressure steam
Operating Conditions - Outlet	Condensate - 0.007 MPa (0.98 psia), 38°C (101°F)	Condensate - 0.007 MPa (0.98 psia), 38°C (101°F)
Assumed or Specified Performance Characteristics	Specified	Specified
Calculated Performance Characteristics	155,600 kWe (at generator terminals)	112,700 kWe (at generator terminals)
Contaminant Removed, %	N/A	N/A
Assumptions Regarding Anticipated Application Issues	Commercial, no issues	Commercial, no issues

SLAG RECOVERY AND HANDLING

Case	2-1	2-2
Technology Type	Proprietary General Electric	Proprietary General Electric
Basis for Design and Performance	Vendor data – commercial design	Vendor data – commercial design
Operating Conditions Inlet – Slag	24,237 kg/hr (53,433 lb/hr) slag 5.6 MPa (815.0 psia), 1,316°C (2,400°F)	24,235 kg/hr (53,430 lb/hr) slag 5.6 MPa (815.0 psia), 1,316°C (2,400°F)
Operating Conditions Outlet -Slag	24,237 kg/hr (53,433 lb/hr) slag 0.1 MPa (14.7 psia), 15°C (59°F)	24,235 kg/hr (53,430 lb/hr) slag 0.1 MPa (14.7 psia), 15°C (59°F)
Assumed or Specified Performance Characteristics	Specified as commercial equipment	Specified as commercial equipment
Calculated Performance Characteristics	Model based on percentage of coal flow	Model based on percentage of coal flow
Contaminant Removed, %	N/A	N/A
Assumptions Regarding Anticipated Application Issues	Commercial slag removal as utilized at commercial IGCC. No issues	Commercial slag removal as utilized at commercial IGCC. No issues

5.2 PERFORMANCE SUMMARY

The overall performance for cases 2-1 and 2-2 is summarized in Exhibit 5-5.

Exhibit 5-5 Cases 2-1 & 2-2 Plant Performance Summaries
100 Percent Load

Case	2-1	2-2	Units
Plant Output			
Steam Turbine Power	155,600	112,700	kW_e
Total	**155,600**	**112,700**	kW_e
Auxiliary Load			
Coal Handling	470	470	kW_e
Coal Milling	2,270	2,270	kW_e
Coal Slurry Pumps	190	200	kW_e
Slag Handling	1,160	1,160	kW_e
Air Separation Unit Auxiliaries	1,000	1,000	kW_e
ASU Main Air Compressor	67,370	67,050	kW_e
Oxygen Compressor	10,640	10,580	kW_e
CO_2 Compressor	31,160	31,150	kW_e
Feedwater Pumps	2,850	1,690	kW_e
Condensate Pump	150	80	kW_e
Quench Water Pump	540	1,270	kW_e
Circulating Water Pump	3,110	3,080	kW_e
Ground Water Pumps	380	390	kW_e
Cooling Tower Fans	1,600	1,590	kW_e
Scrubber Pumps	230	470	kW_e
Acid Gas Removal	19,230	19,220	kW_e
Steam Turbine Auxiliaries	100	100	kW_e
Claus Plant/TGTU Auxiliaries	250	250	kW_e
Claus Plant TG Recycle Compressor	1,840	1,940	kW_e
Miscellaneous Balance of Plant[2]	3,000	3,000	kW_e
Transformer Losses	900	870	kW_e
Total	**148,440**	**147,830**	**kW_e**
Plant Performance			
Net Plant Power	**7,160**	**-35,130**	**kW_e**
Plant Capacity Factor	90.0	90.0	
Coal Feed Flow rate	220,904 (487,011)	220,889 (486,976)	kg/hr (lb/hr)
Hydrogen Production	25,689 (56,634)	25,789 (56,855)	kg/hr (lb/hr)
Thermal Input[1]	1,665,075	1,664,955	kW_t
Effective Thermal Efficiency[3]	61.24%	58.94%	
Cold GasEfficiency[4]	60.81%	61.05%	
Condenser Duty	717 (680)	549 (520)	GJ/hr (MMBtu/hr)
Raw Water Withdrawal	16.1 (4,253)	16.4 (4,324)	m^3/min (gpm)
Raw Water Consumption	13.4 (3,529)	13.6 (3,604)	m^3/min (gpm)

[1] HHV Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb)
[2] Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads
[3] ETE = (Hydrogen HHV + Net Power) / Fuel HHV
[4] CGE = (Hydrogen Product Value) / Fuel Heating Value, HHV

5.3 MASS AND ENERGY BALANCES

Overall energy balances for case 2-1 and case 2-2 are shown in Exhibit 5-6 and Exhibit 5-7 respectively.

Exhibit 5-6 Case 2-1 Overall Energy Balance

	HHV	Sensible + Latent	Power	Total
Energy In, GJ/hr (MMBtu/hr)				
Coal	5,994 (5,681)	5.0 (4.7)		5,999 (5,686)
ASU Air		23.9 (22.6)		24 (23)
Boiler Air		9.9 (9.4)		10 (9)
Raw Water Makeup		60.5 (57.4)		61 (57)
Auxiliary Power			534 (506)	534 (506)
TOTAL	**5,994 (5,681)**	**99.3 (94.1)**	**534 (506)**	**6,628 (6,282)**
Energy Out, GJ/hr (MMBtu/hr)				
ASU Vent		11.7 (11.1)		12 (11)
Slag	92 (88)	38 (36)		130 (123)
Sulfur	51 (49)	0.63 (0.60)		52 (49)
CO_2		-74.1 (-70.3)		-74 (-70)
Cooling Tower Blowdown		20.2 (19.1)		20 (19)
Gasifier Heat Loss		44.6 (42.3)		45 (42)
Hydrogen	3,645 (3,455)	13.2 (12.5)		3,658 (3,467)
Boiler Flue Gas		211 (200)		211 (200)
Condenser		716 (679)		716 (679)
Non-Condenser Cooling Tower Loads*		658 (624)		658 (624)
*Process Losses***		*640 (607)*		*640 (607)*
Power			560 (531)	560 (531)
TOTAL	**3,789 (3,591)**	**2,275 (2,157)**	**560 (531)**	**6,628 (6,282)**

* Includes ASU compressor intercoolers, CO_2 compressor intercoolers, sour water stripper condenser, syngas cooler (low level heat rejection) and extraction air cooler.

** Calculated by difference to close the energy balance

Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 5-7 Case 2-2 Overall Energy Balance

	HHV	Sensible + Latent	Power	Total
Energy In, GJ/hr (MMBtu/hr)				
Coal	5,994 (5,681)	5.0 (4.7)		5,999 (5,686)
ASU Air		23.8 (22.5)		24 (23)
Boiler Air		9.9 (9.4)		10 (9)
Raw Water Makeup		61.6 (58.3)		62 (58)
Auxiliary Power			532 (504)	532 (504)
TOTAL	**5,994 (5,681)**	**100.2 (95.0)**	**532 (504)**	**6,626 (6,280)**
Energy Out, GJ/hr (MMBtu/hr)				
ASU Vent		11.6 (11.0)		12 (11)
Slag	92 (88)	38 (36)		130 (123)
Sulfur	51 (49)	0.64 (0.61)		52 (49)
CO_2		-74.1 (-70.2)		-74 (-70)
Cooling Tower Blowdown		20.0 (18.9)		20 (19)
Gasifier Heat Loss		30.8 (29.2)		31 (29)
Hydrogen	3,659 (3,469)	13.2 (12.5)		3,673 (3,481)
Boiler Flue Gas		211 (200)		211 (200)
Condenser		554 (525)		554 (525)
Non-Condenser Cooling Tower Loads*		805 (763)		805 (763)
*Process Losses***		*808 (765)*		*808 (765)*
Power			406 (385)	406 (385)
TOTAL	**3,803 (3,605)**	**2,417 (2,291)**	**406 (385)**	**6,626 (6,280)**

* Includes ASU compressor intercoolers, CO_2 compressor intercoolers, sour water stripper condenser, syngas cooler (low level heat rejection) and extraction air cooler.

** Calculated by difference to close the energy balance

Reference conditions are 32.02 F & 0.089 PSIA

5.3.1 Water Balance

Overall water balances for case 2-1 and case 2-2 are shown in Exhibit 5-8 and Exhibit 5-9 respectively. Raw water is obtained from groundwater (50 percent) and from municipal sources (50 percent). Water demand represents the total amount of water required for a particular process. Some water is recovered within the process as syngas condensate and that water is re-used as internal recycle. Raw water makeup is the difference between water demand and internal recycle.

Exhibit 5-8 Case 2-1 Water Balance

Water Use	Water Demand, m^3/min (gpm)	Internal Recycle, m^3/min (gpm)	Raw Water Withdrawal, m^3/min (gpm)	Process Water Discharge, m^3/min (gpm)	Raw Water Usage, m^3/min (gpm)
Slag Handling	0.5 (139)	0.5 (139)	0 (0)	0 (0)	0 (0)
Quench/Wash	5.7 (1,514)	3.6 (945)	2.2 (569)	0 (0)	2.2 (569)
Slurry Water	1.5 (400)	1.5 (400)	0 (0)	0 (0)	0 (0)
Venturi Scrubber Water	0.6 (150)	0.6 (150)	0 (0)	0.03 (7)	-0.03 (-7)
Condenser Makeup	**2.4 (622)**	**0.0 (0)**	**2.4 (622)**	**0.0 (0)**	**2.4 (622)**
Shift Steam	2.1 (567)	0 (0)	2.1 (567)	0 (0)	0 (0)
Cooling Tower	**12.1 (3,190)**	**0.5 (128)**	**11.6 (3,062)**	**2.7 (717)**	**8.9 (2,345)**
SWS Blowdown	0 (0)	0.3 (73)	-0.3 (-73)	0 (0)	0 (0)
Total	**22.8 (6,016)**	**6.7 (1,762)**	**16.1 (4,253)**	**2.7 (725)**	**13.4 (3,529)**

Exhibit 5-9 Case 2-2 Water Balance

Water Use	Water Demand, m^3/min (gpm)	Internal Recycle, m^3/min (gpm)	Raw Water Withdrawal, m^3/min (gpm)	Process Water Discharge, m^3/min (gpm)	Raw Water Usage, m^3/min (gpm)
Slag Handling	0.5 (139)	0.5 (139)	0 (0)	0 (0)	0 (0)
Quench/Wash	12.6 (3,333)	7.8 (2,065)	4.8 (1,269)	0 (0)	4.8 (1,269)
Slurry Water	1.5 (400)	1.5 (400)	0 (0)	0 (0)	0 (0)
Venturi Scrubber Water	0.6 (150)	0.6 (150)	0 (0)	0.04 (10)	-0.04 (-10)
Condenser Makeup	**0.0 (0)**	**0.0 (0)**	**0.0 (0)**	**0.0 (0)**	**0.0 (0)**
Shift Steam	0 (0)	0 (0)	0 (0)	0 (0)	0 (0)
Cooling Tower	**12.0 (3,158)**	**0.4 (103)**	**11.6 (3,055)**	**2.7 (710)**	**8.9 (2,345)**
SWS Blowdown	0 (0)	0.4 (103)	-0.4 (-103)	0 (0)	0 (0)
Total	**27.2 (7,180)**	**10.8 (2,856)**	**16.4 (4,324)**	**2.7 (720)**	**13.6 (3,604)**

5.3.2 Carbon Balance

Carbon balances for case 2-1 and case 2-2 are shown in Exhibit 5-10 and Exhibit 5-11 respectively. 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, CO_2 in the boiler stack gas, and CO_2 product. Gray wastewater is recycled within the plant as slurry water. The percent of total carbon sequestered is defined as the amount of carbon product produced (as sequestration-ready CO_2) divided by the carbon in the coal feedstock, less carbon contained in solid byproducts (slag), expressed as a percentage:

(Carbon in Product for Sequestration) / (Carbon in Feed – Carbon in Slag) * 100 or
Case 2-1: 274,699/(310,444-6,209)*100 = 90%
Case 2-2: 274,610/(310,422-6,208)*100 = 90%

Exhibit 5-10 Case 2-1 Carbon Balance

Carbon In, kg/hr (lb/hr)		Carbon Out, kg/hr (lb/hr)	
Coal	140,815 (310,444)	Slag	2,816 (6,209)
Air (CO_2)	107 (237)	Boiler Stack Gas	13,456 (29,665)
		ASU Vent	107 (237)
		CO_2 Product	124,601 (274,699)
		Convergence Tolerance*	-59 (-129)
Total	**140,922 (310,681)**	**Total**	**140,922 (310,681)**

*by difference

Exhibit 5-11 Case 2-2 Carbon Balance

Carbon In, kg/hr (lb/hr)		Carbon Out, kg/hr (lb/hr)	
Coal	140,805 (310,422)	Slag	2,816 (6,208)
Air (CO_2)	107 (236)	Boiler Stack Gas	13,550 (29,873)
		ASU Vent	107 (236)
		CO_2 Product	124,561 (274,610)
		Convergence Tolerance*	-122 (-269)
Total	**140,912 (310,657)**	**Total**	**140,912 (310,657)**

*by difference

5.3.3 Sulfur Balance

Exhibit 5-12 and Exhibit 5-13 show the sulfur balances for case 2-1 and case 2-2 respectively. Sulfur input is the sulfur in the coal. Sulfur output is the sulfur recovered in the Claus plant and SO_2 in the stack gas. Sulfur in the slag and sulfur stripped from the wastewater streams are considered negligible. The convergence tolerance is split between sulfur product and stack gas in proportion to the amounts shown for those two categories to calculate sulfur capture. The total sulfur capture is represented by the following fraction:

Sulfur byproduct/Sulfur in the coal or
Case 2-1: 12,181/12,207 = 99.8%
Case 2-2: 12,485/12,492 = 99.9%

Exhibit 5-12 Case 2-1 Sulfur Balance

Sulfur In, kg/hr (lb/hr)		Sulfur Out, kg/hr (lb/hr)	
Coal	5,537(12,207)	Elemental Sulfur	5,525 (12,181)
		Stack Gas	3 (6)
		CO_2 Product	10 (22)
		Convergence Tolerance*	-2 (-4)
Total	**5,537 (12,207)**	**Total**	**5,537 (12,207)**

*by difference

Exhibit 5-13 Case 2-2 Sulfur Balance

Sulfur In, kg/hr (lb/hr)		Sulfur Out, kg/hr (lb/hr)	
Coal	5,536 (12,206)	Elemental Sulfur	5,540 (12,214)
		Stack Gas	3 (6)
		CO_2 Product	10 (23)
		Convergence Tolerance*	-17 (-38)
Total	**5,536 (12,206)**	**Total**	**5,536 (12,206)**

*by difference

5.3.4 Air Emissions

The environmental targets for emissions were presented in Section 1.6. A summary of the plant air emissions for cases 2-1 and 2-2 are presented in Exhibit 5-14.

Exhibit 5-14 Cases 2-1 and 2-2 Air Emissions

	kg/GJ (lb/10^6 Btu)	Tonne/year (tons/year) 90% Capacity Factor	kg/GJ (lb/10^6 Btu)	Tonne/year (tons/year) 90% Capacity Factor
SO_2	0.001 (0.002)	44 (49)	0.001 (0.002)	44 (49)
NO_X	0.003 (0.008)	154 (170)	0.003 (0.008)	155 (171)
Particulates	0.003 (0.007)	144 (159)	0.003 (0.007)	144 (159)
Hg	2.46E-7 (5.71E-7)	0.012 (0.013)	2.46E-7 (5.71E-7)	0.012 (0.013)
CO_2	8.2 (19.1)	388,717 (428,487)	8.3 (19.3)	391,432 (431,480)

5.4 MAJOR EQUIPMENT LISTS

This section contains the equipment lists corresponding to the plant configurations for cases 2-1 and 2-2. These lists, along with the heat and material balances and supporting performance data, were used to generate plant costs and complete the financial analysis.

Account 1 – Coal Handling

Equip-ment No.	Description	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	181 tonne (200 ton)	181 tonne (200 ton)	2 (0)
2	Feeder	Belt	572 tonne/hr (630 tph)	572 tonne/hr (630 tph)	2 (0)
3	Conveyor No. 1	Belt	1,134 tonne/hr (1,250 tph)	1,134 tonne/hr (1,250 tph)	1 (0)
4	Transfer Tower No. 1	Enclosed	N/A	N/A	1 (0)
5	Conveyor No. 2	Belt	1,134 tonne/hr (1,250 tph)	1,134 tonne/hr (1,250 tph)	1 (0)
6	As-Received Coal Sampling System	Two-stage	N/A	N/A	1 (0)
7	Stacker/Reclaimer	Traveling, linear	1,134 tonne/hr (1,250 tph)	1,134 tonne/hr (1,250 tph)	1 (0)
8	Reclaim Hopper	N/A	45 tonne (50 ton)	45 tonne (50 ton)	2 (1)
9	Feeder	Vibratory	181 tonne/hr (200 tph)	181 tonne/hr (200 tph)	2 (1)
10	Conveyor No. 3	Belt w/ tripper	363 tonne/hr (400 tph)	363 tonne/hr (400 tph)	1 (0)
11	Crusher Tower	N/A	N/A	N/A	1 (0)
12	Coal Surge Bin w/ Vent Filter	Dual outlet	181 tonne (200 ton)	181 tonne (200 ton)	2 (0)
13	Crusher	Impactor reduction	8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0)	8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0)	2 (0)
14	As-Fired Coal Sampling System	Swing hammer	N/A	N/A	1 (1)
15	Conveyor No. 4	Belt w/tripper	363 tonne/hr (400 tph)	363 tonne/hr (400 tph)	1 (0)
16	Transfer Tower No. 2	Enclosed	N/A	N/A	1 (0)
17	Conveyor No. 5	Belt w/ tripper	363 tonne/hr (400 tph)	363 tonne/hr (400 tph)	1 (0)
18	Coal Silo w/ Vent Filter and Slide Gates	Field erected	816 tonne (900 ton)	816 tonne (900 ton)	3 (0)

Account 2 – Coal Preparation and Feed

Equip-ment No.	Description	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
1	Coal Feeder	Gravimetric	82 tonne/hr (90 tph)	82 tonne/hr (90 tph)	3 (0)
2	Conveyor No. 6	Belt w/tripper	245 tonne/hr (270 tph)	245 tonne/hr (270 tph)	1 (0)
3	Rod Mill Feed Hopper	Dual Outlet	490 tonne (540 ton)	490 tonne (540 ton)	1 (0)
4	Weigh Feeder	Belt	118 tonne/hr (130 tph)	118 tonne/hr (130 tph)	2 (0)
5	Rod Mill	Rotary	118 tonne/hr (130 tph)	118 tonne/hr (130 tph)	2 (0)
6	Slurry Water Storage Tank with Agitator	Field erected	299,921 liters (79,230 gal)	299,883 liters (79,220 gal)	2 (0)
7	Slurry Water Pumps	Centrifugal	833 lpm (220 gpm)	833 lpm (220 gpm)	2 (2)
8	Trommel Screen	Coarse	172 tonne/hr (190 tph)	172 tonne/hr 190 tph)	2 (0)
9	Rod Mill Discharge Tank with Agitator	Field erected	343,302 liters (90,690 gal)	343,302 liters (90,690 gal)	2 (0)
10	Rod Mill Product Pumps	Centrifugal	3,028 lpm (800 gpm)	3,028 lpm (800 gpm)	2 (2)
11	Slurry Storage Tank with Agitator	Field erected	1,030,019 liters (272,100 gal)	1,030,019 liters (272,100 gal)	2 (0)
12	Slurry Recycle Pumps	Centrifugal	5,678 lpm (1,500 gpm)	5,678 lpm (1,500 gpm)	2 (2)
13	Slurry Product Pumps	Positive displacement	3,028 lpm (800 gpm)	3,028 lpm (800 gpm)	2 (2)

Account 3 – Feedwater and Miscellaneous Systems and Equipment

Equip-ment No.	Description	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
1	Demineralized Water Storage Tank	Vertical, cylindrical, outdoor	386,112 liters (102,000 gal)	257,408 liters (68,000 gal)	2 (0)
2	Condensate Pumps	Vertical canned	4,240 lpm @ 91 m H_2O (1,120 gpm @ 300 ft H_2O)	2,347 lpm @ 91 m H_2O (620 gpm @ 300 ft H_2O)	2 (1)

Equip-ment No.	Description	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
3	Deaerator (integral w/ HRSG)	Horizontal spray type	322,504 kg/hr (711,000 lb/hr)	215,456 kg/hr (475,000 lb/hr)	2 (0)
4	Intermediate Pressure Feedwater Pump	Horizontal centrifugal, single stage	1,287 lpm @ 658 m H_2O (340 gpm @ 2160 ft H_2O)	1,287 lpm @ 658 m H_2O (340 gpm @ 2160 ft H_2O)	2 (1)
5	High Pressure Feedwater Pump No. 1	Barrel type, multi-stage, centrifugal	HP water: 4,164 lpm @ 1,890 m H_2O (1,100 gpm @ 6,200 ft H_2O)	HP water: 2,271 lpm @ 1,890 m H_2O (600 gpm @ 6,200 ft H_2O)	2 (1)
6	High Pressure Feedwater Pump No. 2	Barrel type, multi-stage, centrifugal	IP water: 189 lpm @ 485 m H_2O (50 gpm @ 1,590 ft H_2O)	IP water: 189 lpm @ 485 m H_2O (50 gpm @ 1,590 ft H_2O)	2 (1)
7	Auxiliary Boiler	Shop fabricated, water tube	18,144 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F)	18,144 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)	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)	28 m^3/min (1,000 scfm)	2 (1)
10	Closed Cycle Cooling Heat Exchangers	Plate and frame	377 GJ/hr (356.95042955 MMBtu/hr) each	457 GJ/hr (433.2617091 MMBtu/hr) each	2 (0)
11	Closed Cycle Cooling Water Pumps	Horizontal centrifugal	135,139 lpm @ 21 m H_2O (35,700 gpm @ 70 ft H_2O)	163,908 lpm @ 21 m H_2O (43,300 gpm @ 70 ft H_2O)	2 (1)
12	Engine-Driven Fire Pump	Vertical turbine, diesel engine	3,785 lpm @ 107 m H_2O (1,000 gpm @ 350 ft H_2O)	3,785 lpm @ 107 m H_2O (1,000 gpm @ 350 ft H_2O)	1 (1)
13	Fire Service Booster Pump	Two-stage horizontal centrifugal	2,650 lpm @ 76 m H_2O (700 gpm @ 250 ft H_2O)	2,650 lpm @ 76 m H_2O (700 gpm @ 250 ft H_2O)	1 (1)
14	Raw Water Pumps	Stainless steel, single suction	4,430 lpm @ 18 m H_2O (1,170 gpm @ 60 ft H_2O)	4,505 lpm @ 18 m H_2O (1,190 gpm @ 60 ft H_2O)	2 (1)
15	Ground Water Pumps	Stainless steel, single suction	2,953 lpm @ 268 m H_2O (780 gpm @ 880 ft H_2O)	2,990 lpm @ 268 m H_2O (790 gpm @ 880 ft H_2O)	2 (1)

Equip-ment No.	Description	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
16	Filtered Water Pumps	Stainless steel, single suction	6,246 lpm @ 49 m H_2O (1,650 gpm @ 160 ft H_2O)	7,684 lpm @ 49 m H_2O (2,030 gpm @ 160 ft H_2O)	2 (1)
17	Filtered Water Tank	Vertical, cylindrical	2,998,046 liter (792,000 gal)	3,694,562 liter (976,000 gal)	2 (0)
18	Makeup Water Demineralizer	Anion, cation, and mixed bed	1,363 lpm (360 gpm)	151 lpm (40 gpm)	2 (0)
19	Liquid Waste Treatment System		10 years, 24-hour storm	10 years, 24-hour storm	1 (0)

Account 4 – Gasifier, Syngas Cooler, Scrubber, ASU, and Accessories

Equip-ment No.	Description	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
1	Gasifier	Pressurized slurry-feed, entrained bed	2,903 tonne/day, 5.6 MPa (3,200 tpd, 814.96 psia)	2,903 tonne/day, 5.6 MPa (3,200 tpd, 814.96 psia)	2 (1)
2	Synthesis Gas Cooler	Vertical downflow radiant heat exchanger	255,826 kg/hr (564,000 lb/hr)	N/A	2 (1)
3	Syngas Quench-Scrubber Including Sour Water Stripper	Quench Chamber and Vertical upflow	350,173 kg/hr (772,000 lb/hr)	474,458 kg/hr (1,046,000 lb/hr)	2 (0)
4	Flare Stack	Self-supporting, carbon steel, stainless steel top, pilot ignition	350,173 kg/hr (772,000 lb/hr) syngas	474,458 kg/hr (1,046,000 lb/hr) syngas	2 (0)
5	ASU Main Air Compressor	Centrifugal, multi-stage	5,947 m^3/min @ 1.3 MPa (210,000 scfm @ 190 psia)	5,918 m^3/min @ 1.3 MPa (209,000 scfm @ 190 psia)	2 (0)
6	Cold Box	Vendor design	2,359 tonne/day (2,600 tpd) of 95% purity oxygen	2,359 tonne/day (2,600 tpd) of 95% purity oxygen	2 (0)

Equip-ment No.	Description	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
7	Oxygen Compressor	Centrifugal, multi-stage	1,189 m^3/min (42,000 scfm) Suction - 0.9 MPa (130 psia) Discharge - 6.5 MPa (940 psia)	1,189 m^3/min (42,000 scfm) Suction - 0.9 MPa (130 psia) Discharge - 6.5 MPa (940 psia)	2 (0)

Account 5A – Sour Gas Shift, Raw Gas Coolers, and Syngas Cleanup

Equip-ment No.	Description	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
1	Water Gas Shift Reactors	Fixed bed, catalytic	388,275 kg/hr (856,000 lb/hr) 227°C (440°F) 5.4 MPa (790 psia)	400,976 kg/hr (884,000 lb/hr) 232°C (450°F) 5.5 MPa (800 psia)	4 (0)
2	Shift Reactor Heat Recovery Exchangers	Shell and Tube	Exchanger 1: 157 GJ/hr (149 MMBtu/hr) Exchanger 2: -3 GJ/hr (-3 MMBtu/hr)	Exchanger 1: 155 GJ/hr (147 MMBtu/hr) Exchanger 2: -4 GJ/hr (-4 MMBtu/hr)	4 (0)
3	Raw Gas Coolers	Shell and tube with condensate drain	388,275 kg/hr (856,000 lb/hr)	400,976 kg/hr (884,000 lb/hr)	8 (0)
4	Raw Gas Knockout Drum	Vertical with mist eliminator	304,360 kg/hr (671,000 lb/hr)	304,814 kg/hr (672,000 lb/hr)	2 (0)
5	Venturi Water Spray Scrubber	Spray tower equilibrium with exiting gas stream	607,687 kg/hr (1,339,720 lb/hr) 35°C (95°F) 5.1 MPa (745 psia)	607,746 kg/hr (1,339,850 lb/hr) 35°C (95°F) 5.2 MPa (750 psia)	1 (0)
6	Mercury Adsorber	Sulfated carbon bed	303,907 kg/hr (670,000 lb/hr) 35°C (95°F) 5.1 MPa (745 psia)	303,907 kg/hr (670,000 lb/hr) 35°C (95°F) 5.2 MPa (750 psia)	2 (0)
7	Sulfur Plant	Claus type	146 tonne/day (161 tpd)	146 tonne/day (161 tpd)	1 (0)
8	Acid Gas Removal Plant	Two-stage Selexol	310,711 kg/hr (685,000 lb/hr) 35°C (95°F) 5.1 MPa (740 psia)	311,164 kg/hr (686,000 lb/hr) 35°C (95°F) 5.1 MPa (745 psia)	2 (0)
9	Hydrogen-ation Reactor	Fixed bed, catalytic	18,290 kg/hr (40,323 lb/hr) 232°C (450°F) 0.1 MPa (12.3 psia)	19,149 kg/hr (42,216 lb/hr) 232°C (450°F) 0.1 MPa (12.3 psia)	1 (0)

Equip-ment No.	Description	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
10	Tail Gas Recycle Compressor	Centrifugal	13,942 kg/hr (30,737 lb/hr)	14,351 kg/hr (31,639 lb/hr)	1 (0)

Account 5B – CO_2 Compressor

Equip-ment No.	Description	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
1	CO_2 Compressor	Integrally geared, multi-stage centrifugal	1,133 m^3/min @ 15.3 MPa (40,000 scfm @ 2,200 psia)	1,133 m^3/min @ 15.3 MPa (40,000 scfm @ 2,200 psia)	4 (1)

Account 6 – Hydrogen Purification

Equip-ment No.	Description	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
1	Pressure Swing Adsorber	Polybed Proprietary	99,373 kg/hr (219,080 lb/hr) Syngas 35°C (94.619767°F) 5.1 MPa (740. psia) 28,259 kg/hr (62,300 lb/hr) Hydrogen 35°C (94.619767°F) 5.1 MPa (740. psia) 71,114 kg/hr (156,780 lb/hr) Off Gas -7°C (20°F) 0.5 MPa (77. psia)	99,917 kg/hr (220,280 lb/hr) Syngas 35°C (94.5996075°F) 5.1 MPa (745. psia) 28,368 kg/hr (62,540 lb/hr) Hydrogen 35°C (94.5996075°F) 5.1 MPa (745. psia) 71,550 kg/hr (157,740 lb/hr) Off Gas -7°C (20°F) 0.5 MPa (77. psia)	1 (0)

Account 7 – Stack, Ducting, and Off-Gas Fired Boiler

Equip-ment No.	Descrip tion	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
1	Stack	CS plate, type 409SS liner	76 m (250 ft) high x 2.7 m (9 ft) diameter	76 m (250 ft) high x 2.7 m (9 ft) diameter	1 (0)
2	Field-Erected Gas-Fired Boiler	Drum, multi-pressure with economizer section and integral deaerator Air-Fired	71,114 kg/hr (156,780 lb/hr) Off gas 1MPa (77psig) -6.7°C (20°F) 349,266 kg/hr (770,000 lb/hr) Air 3MPa (452.30396psig) 537.8°C (1000°F) 876 GJ/hr (830 MMBtu/hr)	71,550 kg/hr (157,740 lb/hr) Off gas 1MPa (77psig) -6.7°C (20°F) 349,266 kg/hr (770,000 lb/hr) Air 3MPa (452.30396psig) 537.8°C (1000°F) 886 GJ/hr (840 MMBtu/hr)	1 (0)

Account 8 – Steam turbine Generator and Auxiliaries

Equip-ment No.	Description	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
1	Steam Turbine	Commercially available advanced steam turbine	164 MW 12.4 MPa/538°C/538°C (1800 psig/ 1,000°F/1,000°F)	119 MW 12.4 MPa/538°C/538°C (1800 psig/ 1,000°F/1,000°F)	1 (0)
2	Steam Turbine Generator	Hydrogen cooled, static excitation	180 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	130 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3-phase	1 (0)
3	Steam Bypass	One per HRSG	50% steam flow @ design steam conditions	50% steam flow @ design steam conditions	2 (0)
4	Surface Condenser	Single pass, divided waterbox including vacuum pumps	791 GJ/hr (750 MMBtu/hr), Inlet water temperature 16°C (60°F), Water temperature rise 11°C (20°F)	612 GJ/hr (580 MMBtu/hr), Inlet water temperature 16°C (60°F), Water temperature rise 11°C (20°F)	1 (0)

Account 9 – Cooling water System

Equip-ment No.	Description	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
1	Circulating Water Pumps	Vertical, wet pit	310,404 lpm @ 30 m (82,000 gpm @ 100 ft)	306,618 lpm @ 30 m (81,000 gpm @ 100 ft)	2 (0)
8	Cooling Tower	Evaporative, mechanical draft, multi-cell	11°C (51.5°F) wet bulb / 16°C (60°F) CWT / 27°C (80°F) HWT / 1,730 GJ/hr (1,640 MMBtu/hr) heat duty	11°C (51.5°F) wet bulb / 16°C (60°F) CWT / 27°C (80°F) HWT / 1,720 GJ/hr (1,630 MMBtu/hr) heat duty	1 (0)

Account 10 – Slag Recovery and Handling

Equip-ment No.	Description	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
1	Slag Quench Tank	Water bath	253,623 liters (67,000 gal)	253,623 liters (67,000 gal)	2 (1)
2	Slag Crusher	Roll	14 tonne/hr (15 tph)	14 tonne/hr (15 tph)	2 (1)
3	Slag Depressurizer	Lock Hopper	14 tonne/hr (15 tph)	14 tonne/hr (15 tph)	2 (1)

Equip-ment No.	Description	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
4	Slag Receiving Tank	Horizontal, weir	151,416 liters (40,000 gal)	151,416 liters (40,000 gal)	2 (1)
5	Black Water Overflow Tank	Shop fabricated	68,137 liters (18,000 gal)	68,137 liters (18,000 gal)	2 (0)
6	Slag Conveyor	Drag chain	14 tonne/hr (15 tph)	14 tonne/hr (15 tph)	2 (0)
7	Slag Separation Screen	Vibrating	14 tonne/hr (15 tph)	14 tonne/hr (15 tph)	2 (0)
8	Coarse Slag Conveyor	Belt/bucket	14 tonne/hr (15 tph)	14 tonne/hr (15 tph)	2 (0)
9	Fine Ash Settling Tank	Vertical, gravity	215,768 liters (57,000 gal)	215,768 liters (57,000 gal)	2 (0)
10	Fine Ash Recycle Pumps	Horizontal centrifugal	38 lpm @ 14 m H_2O (10 gpm @ 46 ft H_2O)	38 lpm @ 14 m H_2O (10 gpm @ 46 ft H_2O)	2 (2)
11	Grey Water Storage Tank	Field erected	68,137 liters (18,000 gal)	68,137 liters (18,000 gal)	2 (0)
12	Grey Water Pumps	Centrifugal	227 lpm @ 564 m H_2O (60 gpm @ 1,850 ft H_2O)	227 lpm @ 564 m H_2O (60 gpm @ 1,850 ft H_2O)	2 (2)
13	Slag Storage Bin	Vertical, field erected	998 tonne (1,100 tons)	998 tonne (1,100 tons)	2 (0)
14	Unloading Equipment	Telescoping chute	109 tonne/hr (120 tph)	109 tonne/hr (120 tph)	1 (0)

Account 11 – Accessory Electric Plant

Equip-ment No.	Description	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
1	STG Step-up Transformer	Oil-filled	24 kV/345 kV, 180 MVA, 3-ph, 60 Hz	24 kV/345 kV, 130 MVA, 3-ph, 60 Hz	1 (0)
2	High Voltage Auxiliary Transformer	Oil-filled	345 kV/13.8 kV, 61 MVA, 3-ph, 60 Hz	345 kV/13.8 kV, 60 MVA, 3-ph, 60 Hz	2 (0)
3	Medium Voltage Auxiliary Transformer	Oil-filled	24 kV/4.16 kV, 43 MVA, 3-ph, 60 Hz	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	4.16 kV/480 V, 6 MVA, 3-ph, 60 Hz	1 (1)

Equip-ment No.	Description	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
5	STG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	24 kV, 3-ph, 60 Hz	1 (0)
6	Medium Voltage Switchgear	Metal clad	4.16 kV, 3-ph, 60 Hz	4.16 kV, 3-ph, 60 Hz	1 (1)
7	Low Voltage Switchgear	Metal enclosed	480 V, 3-ph, 60 Hz	480 V, 3-ph, 60 Hz	1 (1)
8	Emergency Diesel Generator	Sized for emergency shutdown	750 kW, 480 V, 3-ph, 60 Hz	750 kW, 480 V, 3-ph, 60 Hz	1 (0)

Account 12 – Instrumentation and Control

Equip-ment No.	Description	Type	Case 2-1 Design Condition	Case 2-2 Design Condition	Oper. Qty. (Spares)
1	DCS - Main Control	Monitor/keyboard; Operator printer; Engineering printer	Operator stations/printers and engineering stations/printers	Operator stations/printers and engineering stations/printers	1 (0)
2	DCS - Processor	Microprocessor with redundant input/output	N/A	N/A	1 (0)
3	DCS - Data Highway	Fiber optic	Fully redundant, 25% spare	Fully redundant, 25% spare	1 (0)

5.5 COST ESTIMATION

5.5.1 Equipment Costing

The capital costs for the equipment in this case were factored from the bituminous baseline study, case 2 [Ref. 1] for all equipment that was included in that original case. The case 2-1 estimate was prepared by factoring the capital estimate on the basis of coal, gas, and steam flows and conditions.

GEE Radiant-Only Gasifier

The capital cost for the GEE radiant-only gasifier was taken from the bituminous baseline study. However, that study was based on utilization of two gasifier trains, each operating at 50 percent to achieve full plant capacity. To achieve 90 percent availability, case 2-1 is configured with two gasifier trains operating at 50 percent and a third spare gasifier train (gasifier, radiant cooler, and quench) on hot standby. This has resulted in a significant increase in the capital cost.

GEE Quench Gasifier

The capital cost for the GEE quench gasifier is adapted from the bituminous baseline study and factored utilizing the relative costs for radiant-only and quench from a detailed Texaco gasifier cost estimate by Parsons in 2002. Case 2-2 is also configured with two gasifier trains operating at 50 percent and a third spare gasifier train (gasifier, radiant cooler, and quench) on hot standby.

PSA

The capital cost of the PSA is based on a budgetary quotation from Krupp-Uhde to RDS. The 1998 quotation encompassed the turnkey installation of an SMR plant to produce 70 MMSCFD of hydrogen. The case 2 estimates were prepared by extracting the PSA cost and upgrading the Krupp estimate to June 2007 utilizing *Chemical Engineering* Plant Cost Indices [Ref. 9] and factoring the capital estimate on the basis of hydrogen production capacity.

Off-Gas Fired Boiler

The cost of the PSA off-gas air fired boiler was based on published estimates for field-erected boilers.

Balance of Plant

The cost of the balance of plant that constitutes the complete hydrogen production plant was based on an in-house model that has been used to develop the capital costs and economic results for many process applications. Cost attributed to balance of plant amounts to 15 percent of the installed plant equipment cost.

Contingency

Project and process contingencies were added to the estimates for the coal to hydrogen cases based on values includes in the original reference study [Ref. 1]. Project contingencies were added to cover project uncertainty and the cost of additional equipment that could result from a more detailed design. The project contingencies represent costs that are expected to occur. Each capital account was evaluated against the level of estimate detail, field experience, and the basis for the equipment pricing to define project contingency. Process contingencies were added for the gasification and CO_2 removal system elements of the technology that are not considered commercially proven based on the level of detail available and commercially proven status for the system elements.

5.5.2 O&M Costs

Fixed and variable operating costs were estimated for each case. The coal price used for this study was $1.55/GJ ($1.64/MMBtu) for Illinois No. 6 on a HHV basis. All other consumable costs were assumed to match those used in the baseline reference report [Ref. 1]. An emissions value of $30/tonne of CO_2 emitted was also applied to reflect potential environmental regulations. A credit of $105/MWh for the net electricity generated in case 2-1 and a debit of $105/MWh for the net electricity consumed in case 2-2 are included in the costs. This value is consistent with the cost of electricity (COE) generated in an environment where coal-based power plants are built with carbon capture and sequestration systems.

5.5.3 Cost Estimating Results

The total overnight cost (TOC) for the case 2-1 plant producing 242 MMSCFD (617 metric tons) of hydrogen (99.6 percent H_2 by volume) per day from coal with CO_2 capture is estimated to be $1,851 million in June 2007 dollars resulting in a first year cost of hydrogen(COH) of $3.09/kg H_2. Exhibit 5-15 and Exhibit 5-16 show the capital and operating costs for this plant.

The TOC for the case 2-2 plant producing 243 MMSCFD of hydrogen (619 metric tons) of hydrogen (99.6 percent H_2 by volume) per day from coal with CO_2 capture is estimated to be $1,597 million in June 2007 dollars resulting in a first year COH of $2.89/kg H_2. Exhibit 5-17 and Exhibit 5-18 show the capital and operating costs for this plant. Capital cost estimating methodology is explained in Section 2.

The additional costs of CO_2 TS&M for cases 2-1 and 2-2 are shown in Exhibit 5-19. Both sets of values are estimated to total $0.20/kg H_2, bringing the total cost of hydrogen production with CO_2 capture to $3.29/kg H_2 for case 2-1 and $3.09/kg H_2 for case 2-2.

Exhibit 5-15 Case 2-1 Capital Cost Summary

Client: USDOE/NETL Report Date 21-Aug-10
Project: 401.01.04 Activity 5 Assessment of Baseline and Advanced Hydrogen Production Plants
Case: Case 2-1 Baseline Coal-to-Hydrogen with CC&S
Plant Size: 616,527 kg H_2/day Cost Base Jun-07

Acct No.	Item/Description	Bare Erected Cost $1,000	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	TOTAL PLANT COST $1,000
1	COAL HANDLING	**$28,008**	**$2,801**	**$0**	**$6,162**	**$36,970**
2	COAL PREP & FEED	**$43,489**	**$4,349**	**$1,578**	**$9,883**	**$59,300**
3	FEEDWATER & MISC. BOP SYSTEMS	**$18,574**	**$1,857**	**$0**	**$4,840**	**$25,272**
4	GASIFIER ISLAND					
4.1	GEE Syngas Cooler Gasifier System	$341,298	$34,130	$47,365	**$65,956**	$488,748
4.2	Syngas Cooler(w/ Gasifier - 4.1)		$0	$0	$0	
4.3	ASU/Oxidant Compression	$193,147	$19,315	$0	$21,246	$233,707
4.4	Scrubber & Low Temperature Cooling	$16,102	$1,610	$0	$3,542	$21,255
4.4-4.9	Other Gasification Equipment	$16,352	$1,635	$0	$4,497	$22,484
	SUBTOTAL 4.	**$566,898**	**$56,690**	**$47,365**	**$95,241**	**$766,194**
5A	GAS CLEANUP & PIPING					
5A.1	Double Stage Selexol	$137,069	$13,707	$27,414	$35,638	$213,828
5A.2	Elemental Sulfur Plant	$25,719	$2,572	$0	$5,658	$33,949
5A.3	Mercury Removal	$2,436	$244	$122	$560	$3,362
5A.4	Shift Reactors	$13,487	$1,349	$0	$2,967	$17,802
5A.7	Fuel Gas Piping	$1,080	$108	$0	$238	$1,426
5A.9	HGCU Foundations	$1,058	$106	$0	$233	$1,397
	SUBTOTAL 5A.	**$180,849**	**$18,085**	**$27,536**	**$45,294**	**$271,764**
5B	CO2 COMPRESSION					
5B.1	CO2 Compressor & Drying	$29,448	$2,945	$0	$6,479	$38,872
	SUBTOTAL 5B.	**$29,448**	**$2,945**	**$0**	**$6,479**	**$38,872**
6	HYDROGEN PRODUCTION					
6.1	Pressure Swing Adsorber	$40,412	$4,041	$0	$8,891	$53,344
6.2	Hydrogen Compressor	$0	$0	$0	$0	$0
	SUBTOTAL 7	**$40,412**	**$4,041**	**$0**	**$8,891**	**$53,344**
7	HRSG, DUCTING, STACK					
7.1	Off Gas Fired Boiler and Stack	$20,309	$2,031	$0	$6,702	$29,042
7.2	Hot Gas Expander	$0	$0	$0	$0	$0
	SUBTOTAL 6	**$20,309**	**$2,031**	**$0**	**$6,702**	**$29,042**
8	STEAM TURBINE GENERATOR					
8.1	Steam TG & Accessories	$21,606	$2,161	$0	$4,753	$28,521
8.2	Turbine Plant Auxiliaries and Steam Piping	$11,976	$1,198	$0	$2,635	$15,808
	SUBTOTAL 8	**$33,582**	**$3,358**	**$0**	**$7,388**	**$44,329**
9	COOLING WATER SYSTEM	**$15,715**	**$1,572**	**$0**	$3,457	**$20,744**
10	ASH/SPENT SORBENT HANDLING SYS	**$56,865**	**$5,687**	**$0**	$9,383	**$71,935**
11	ACCESSORY ELECTRIC PLANT	**$20,017**	**$2,002**	**$0**	$4,404	**$26,422**
12	INSTRUMENTATION & CONTROL	**$21,083**	**$2,108**	**$1,054**	$4,849	**$29,095**
13	IMPROVEMENTS TO SITE	**$15,813**	**$1,581**	**$0**	$5,218	**$22,612**
14	BUILDINGS & STRUCTURES	**$16,867**	**$1,687**	**$0**	$3,711	**$22,264**
	TOTAL COST	**$1,107,930**	**$110,793**	**$77,533**	**$221,901**	**$1,518,158**

Exhibit 5-15 Case 2-1 Capital Cost Summary (continued)

Client: USDOE/NETL — Report Date: 21-Aug-10
Project: 401.01.04 Activity 5 Assessment of Baseline and Advanced Hydrogen Production Plants
Case: Case 2-1 Baseline Coal-to-Hydrogen with CC&S
Plant Size: 616,527 kg H_2/day — Cost Base: Jun-07

Acct No.	Item/Description	Bare Erected Cost $1,000	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	TOTAL PLANT COST $1,000
	TOTAL COST	$1,107,930	$110,793	$77,533	$221,901	$1,518,158
	Owner's Costs					
	Preproduction Costs					
	6 Months All Labor					$11,537
	1 Month Maintenance Materials					$3,036
	1 Month Non-fuel Consumables					$417
	1 Month Waste Disposal					$316
	25% of 1 Months Fuel Cost at 100% CF					$1,697
	2% of TPC					$30,363
	Total					**$47,367**
	Inventory Capital					
	60 day supply of consumables at 100% CF					$835
	0.5% of TPC (spare parts)					$7,591
	Total					**$8,426**
	Initial Cost for Catalyst and Chemicals					$7,218
	Land					$900
	Other Owner's Costs					$227,724
	Financing Costs					$40,990
	Total Overnight Costs (TOC)					**$1,850,782**
	TASC Multiplier					1.201
	Total As-Spent Cost (TASC)					**$2,223,325**

Exhibit 5-16 Case 2-1 Operating Cost Summary

Annual Fixed O&M Labor and Material Costs

Skilled Operator	2		
Operator	10		
Foreman	1		
Lab Tech's, etc.	3		
TOTAL-O.J.'s	16	Operating Labor	$6,313,507
Operator Base Rate	$34.65	Maintenance Labor	$15,181,578
Operator Labor Burden (% of Ba:	30.00%	Admin & Support	$1,578,377
Labor O-H Charge Rate (% of lal	25.00%	Property Taxes and Insurance	$30,363,155
Total Overnight Cost	**$1,850,781,835**	**TOTAL FIXED O&M**	**$53,436,617**

Variable O&M Operating Costs

	Initial Fill	Consumption /Day	Unit Rate	Unit Cost	Unit	Initial Fill Cost	Annual Variable O&M Costs
Maintenance Material							$32,792,208
Water		5,081	1,000 gals/day	$1.08	1000 gal	$0	$1,805,574
Water Treatment Chemicals	0	25,410	lb/day	$0.17	lb	$0	$1,444,631
Carbon (Mercury Removal) (lb)	76,126	104	lb/day	$1.05	lb	$79,945	$35,878
Water Gas Shift Catalyst (ft^3)	6,290	4	ft^3/day	$498.83	ft^3	$3,137,696	$705,659
Selexol Solution (gal)	298,541	95	gal/day	$13.40	gal	$3,999,928	$418,126
Claus Catalyst(ft^3)	0	2	ft^3/day	$131.27	ft^3	$0	$98,008
Electric Power Bought (Generated)		(7)	MW		MWh		
Purchased Electric Power (Revenue)		(172)	MWh/day	$105.00	MWh	$0	($5,927,191)
Solid Waste Disposal		641	ton/day	$16.23	ton	$0	$3,417,849
Carbon Dioxide Emissions Tax		1,304	ton/day	$ 27.22	ton CO_2	$0	$11,661,501
					TOTAL VARIABLE O&M	**$7,217,569**	**$46,452,244**

Annual Fuel Costs

Coal (Illinois #6)	5,844	tons/day	$38.19	ton	**$73,307,753**
			$1.64	MMBtu	
			TOTAL ANNUAL O&M COSTS		**$173,196,613**
			FIRST YEAR CAPITAL CHARGE		**$459,987,842**
Hydrogen Production	**616,527**	**kg/day**			
Plant Capacity Factor:	**90%**		**First Year H_2 COST**	**$/kg**	**$3.13**

Exhibit 5-17 Case 2-2 Capital Cost Summary

Client: USDOE/NETL — **Report Date** 21-Aug-10
Project: 401.01.04 Activity 5 Assessment of Baseline and Advanced Hydrogen Production Plants

Case: Case 2-2 Baseline Coal-to-Hydrogen with CC&S & Full Quench
Plant Size: 618,940 kg H_2/day — **Cost Base** Jun-07

Acct No.	Item/Description	Bare Erected Cost $1,000	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	TOTAL PLANT COST $1,000
1	COAL HANDLING	**$28,007**	**$2,801**	**$0**	**$6,161**	**$36,969**
2	COAL PREP & FEED	**$43,459**	**$4,346**	**$1,577**	**$9,876**	**$59,259**
3	FEEDWATER & MISC. BOP SYSTEMS	**$14,703**	**$1,470**	**$0**	**$3,831**	**$20,004**
4	GASIFIER ISLAND					
4.1	GEE Quench Gasifier System	$239,472	$23,947	$31,372	**$45,987**	$340,779
4.2	Syngas Cooler(w/ Gasifier - 4.1)		$0	$0	$0	
4.3	ASU/Oxidant Compression	$192,375	$19,237	$0	$21,161	$232,774
4.4	Scrubber & Low Temperature Cooling	$7,121	$712	$0	$1,567	$9,400
4.4-4.9	Other Gasification Equipment	$4,902	$490	$0	$1,348	$6,740
	SUBTOTAL 4.	**$443,870**	**$44,387**	**$31,372**	**$70,063**	**$589,692**
5A	GAS CLEANUP & PIPING					
5A.1	Double Stage Selexol	$137,403	$13,740	$27,481	$35,725	$214,349
5A.2	Elemental Sulfur Plant	$25,775	$2,577	$0	$5,670	$34,022
5A.3	Mercury Removal	$2,434	$243	$122	$560	$3,359
5A.4	Shift Reactors	$14,158	$1,416	$0	$3,115	$18,689
5A.7	Fuel Gas Piping	$1,080	$108	$0	$238	$1,426
5A.9	HGCU Foundations	$1,058	$106	$0	$233	$1,397
	SUBTOTAL 5A.	**$181,909**	**$18,191**	**$27,602**	**$45,540**	**$273,242**
5B	CO2 COMPRESSION					
5B.1	CO2 Compressor & Drying	$29,442	$2,944	$0	$6,477	$38,863
	SUBTOTAL 5B.	**$29,442**	**$2,944**	**$0**	**$6,477**	**$38,863**
6	HYDROGEN PRODUCTION					
6.1	Pressure Swing Adsorber	$40,538	$4,054	$0	$8,918	$53,511
6.2	Hydrogen Compressor	$0	$0	$0	$0	$0
	SUBTOTAL 7	**$40,538**	**$4,054**	**$0**	**$8,918**	**$53,511**
7	HRSG, DUCTING, STACK					
7.1	Off Gas Fired Boiler and Stack	$20,409	$2,041	$0	$6,735	$29,184
7.2	Hot Gas Expander	$0	$0	$0	$0	$0
	SUBTOTAL 6	**$20,409**	**$2,041**	**$0**	**$6,735**	**$29,184**
8	STEAM TURBINE GENERATOR					
8.1	Steam TG & Accessories	$17,074	$1,707	$0	$3,756	$22,537
8.2	Turbine Plant Auxiliaries and Steam Piping	$8,933	$893	$0	$1,965	$11,792
	SUBTOTAL 8	**$26,007**	**$2,601**	**$0**	**$5,722**	**$34,329**
9	COOLING WATER SYSTEM	**$13,416**	**$1,342**	**$0**	$2,952	**$17,709**
10	ASH/SPENT SORBENT HANDLING SYS	**$56,864**	**$5,686**	**$0**	$9,382	**$71,932**
11	ACCESSORY ELECTRIC PLANT	**$16,131**	**$1,613**	**$0**	$3,549	**$21,293**
12	INSTRUMENTATION & CONTROL	**$17,887**	**$1,789**	**$894**	$4,114	**$24,684**
13	IMPROVEMENTS TO SITE	**$13,415**	**$1,342**	**$0**	$4,427	**$19,184**
14	BUILDINGS & STRUCTURES	**$12,521**	**$1,252**	**$0**	$2,755	**$16,527**
	TOTAL COST	**$958,576**	**$95,858**	**$61,446**	**$190,503**	**$1,306,383**

Exhibit 5-17 Case 2-2 Capital Cost Summary (continued)

Client: USDOE/NETL — **Report Date** 21-Aug-10
Project: 401.01.04 Activity 5 Assessment of Baseline and Advanced Hydrogen Production Plants
Case: Case 2-2 Baseline Coal-to-Hydrogen with CC&S & Full Quench
Plant Size: 618,940 kg H_2/day — **Cost Base** Jun-07

Acct No.	Item/Description	Bare Erected Cost $1,000	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	TOTAL PLANT COST $1,000
	Owner's Costs					
	Preproduction Costs					
	6 Months All Labor					$10,478
	1 Month Maintenance Materials					$2,613
	1 Month Non-fuel Consumables					$3,117
	1 Month Waste Disposal					$316
	25% of 1 Months Fuel Cost at 100% CF					$1,697
	2% of TPC					$26,128
	Total					**$44,348**
	Inventory Capital					
	60 day supply of consumables at 100% CF					$848
	0.5% of TPC (spare parts)					$6,532
	Total					**$7,380**
	Initial Cost for Catalyst and Chemicals					$7,260
	Land					$900
	Other Owner's Costs					$195,957
	Financing Costs					$35,272
	Total Overnight Costs (TOC)					**$1,597,501**
	TASC Multiplier					1.201
	Total As-Spent Cost (TASC)					**$1,919,061**

Exhibit 5-18 Case 2-2 Operating Cost Summary

Annual Fixed O&M Labor and Material Costs

Skilled Operator	2		
Operator	10		
Foreman	1		
Lab Tech's, etc.	3		
TOTAL-O.J.'s	16	Operating Labor	$6,313,507
Operator Base Rate	$34.65	Maintenance Labor	$13,063,827
Operator Labor Burden (% of Base)	30.00%	Admin & Support	$1,578,377
Labor O-H Charge Rate (% of labor)	25.00%	Property Taxes and Insurance	$26,127,654
Total Overnight Cost	**$1,597,500,891**	**TOTAL FIXED O&M**	**$47,083,365**

Variable O&M Operating Costs

	Initial Fill	Consumption /Day	Unit Rate	Unit Cost	Unit	Initial Fill Cost	Annual Variable O&M Costs
Maintenance Material							$28,217,867
Water		5,189	1,000 gals/day	$1.08	1000 gal	$0	$1,843,971
Water Treatment Chemicals	0	25,950	lb/day	$0.17	lb	$0	$1,475,332
Carbon (Mercury Removal) (lb)	76,126	104	lb/day	$1.05	lb	$79,945	$35,878
Water Gas Shift Catalyst (ft^3)	6,290	4	ft^3/day	$498.83	ft^3	$3,137,696	$705,659
Selexol Solution (gal)	301,684	96	gal/day	$13.40	gal	$4,042,032	$422,527
Claus Catalyst(ft^3)	0	2	ft^3/day	$131.27	ft^3	$0	$98,008
Electric Power Bought (Generated)		35	MW				
Purchased Electric Power (Revenue)		843	MWh/day	$105.00	MWh	$0	$29,081,317
Solid Waste Disposal		641	ton/day	$16.23	ton	$0	$3,417,604
Carbon Dioxide Emissions Tax		1,313	ton/day	$ 27.22	ton CO_2	$0	$11,742,952
				TOTAL VARIABLE O&M		**$7,259,673**	**$77,041,114**

Annual Fuel Costs

Coal (Illinois #6)	5,844	tons/day	$38.19	ton	**$73,302,484**
			$1.64	MMBtu	
			TOTAL ANNUAL O&M COSTS		**$197,426,964**
			FIRST YEAR CAPITAL CHARGE		**$397,140,235**
Hydrogen Production	**618,940**	**kg/day**			
Plant Capacity Factor:	**90%**		**First Year H_2 COST**	**$/kg**	**$2.92**

Exhibit 5-19 Cases 2-1 & 2-2 Cost Estimate CO_2 TS&M

Parameter	Case 2-1	Case 2-2
TPC of Transport, million $	80.43	80.43
TPC of Storage, million $	50.17	50.16
Capital Fund for Life-Cycle CO_2 Monitoring Costs, million $	33.01	33.00
Total Capital TS&M	**163.62**	**163.60**
First Year Annual Operating Costs at 100% Capacity Factor		
Transport - Fixed O&M, million $	0.43	0.43
Storage - Variable O&M, million $	0.04	0.04
Storage - Fixed O&M, million $	0.21	0.21
Total First Year Cost CO_2 TS&M, $/kg H_2	**0.20**	**0.20**
Total First Year Cost without CO_2 TS&M, $/kg H_2 (see Exhibit 5-16 **and** Exhibit 5-18**)**	**3.13**	**2.92**
TOTAL First Year COH, $/kg H_2	**3.33**	**3.13**

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6. SUMMARY

The objective of this study was to analyze potential plant configurations to determine the baseline performance and cost of producing hydrogen from natural gas and coal. The plants were assumed to be designed and constructed in the near future based on technologies as they exist today, with a planned startup year of 2015. This report covers the following cases:

- **Case 1-1** – Baseline Steam Methane Reforming (SMR) Hydrogen Plant with CO_2 Capture and Sequestration matching the hydrogen generation rate of case 2-1
- **Case 1-2** – Baseline Steam Methane Reforming (SMR) Hydrogen Plant with CO_2 Capture and Sequestration matching the hydrogen generation rate of case 2-2
- **Case 2-1** – Baseline Coal Gasification Hydrogen Plant using GE Energy Radiant-Only Gasifier with CO_2 Capture and Sequestration and Hydrogen Separation by Pressure Swing Adsorption
- **Case 2-2** – Baseline Coal Gasification Hydrogen Plant using GE Energy Quench Gasifier with CO_2 Capture and Sequestration and Hydrogen Separation by Pressure Swing Adsorption

The overall performance results for all four cases are summarized in Exhibit 6-1.

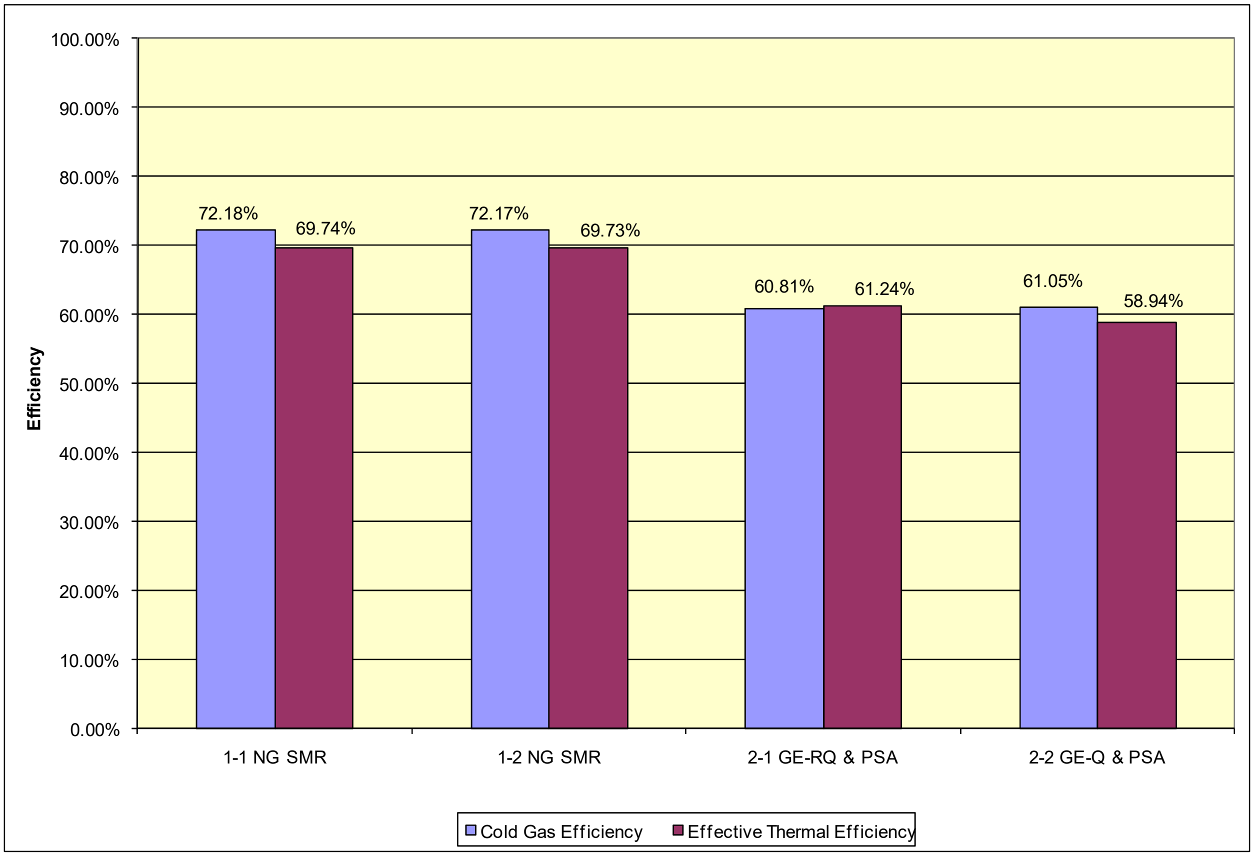
The effective thermal efficiencies (based on HHV) are shown graphically in Exhibit 6-2. The SMR cases have the highest ETE.

Exhibit 6-1 Overall Performance

Case	1-1	1-2	2-1	2-2
Steam Turbine Power, kWe	0	0	155,600	112,700
Net Auxiliary Load, kWe	34,200	34,330	148,440	147,830
Net Plant Power, kWe	-34,200	-34,330	7,160	-35,130
Natural Gas SMR Feed Flow rate, kg/hr (lb/hr)	84,947 (187,276)	85,328 (188,116)	N/A	N/A
Supplemental Natural Gas Feed Flow rate, kg/hr (lb/hr)	10,319 (22,750)	10,319 (22,750)	N/A	N/A
Coal Feed Flow rate, kg/hr (lb/hr)	N/A	N/A	220,904 (487,011)	220,889 (486,976)
Thermal Input[1], GJ/hr (MMBtu/hr)	5,051 (4,787)	5,071 (4,806)	5,994 (5,681)	5,994 (5,681)
Hydrogen Production, kg/day	616,528	618,936	616,527	618,940
Hydrogen Production, lb/hr	56,634	56,855	56,634	56,855
Hydrogen Production, million nm^3/day (MMSCFD)	6.9 (242)	6.9 (243)	6.9 (242)	6.9 (243)
Cold Gas Efficiency[2]	72.18%	72.17%	60.81%	61.05%
Effective Thermal Efficiency[3]	69.74%	69.73%	61.24%	58.94%
Plant Availability	90.00%	90.00%	90.00%	90.00%
Condenser Duty, GJ/hr (MMBtu/hr)	N/A	N/A	680	520
CO_2 Captured, tonne/day (tpd)	5,456 (6,014)	5,478 (6,038)	10,954 (12,075)	10,951 (12,071)
CO_2 Emissions, tonne/day (tpd)	606 (668)	609 (671)	1,183 (1,304)	1,191 (1,313)
Raw Water Withdrawal, m^3/min (gpm)	12.4 (3,265)	12.4 (3,278)	16.1 (4,253)	16.4 (4,324)
Raw Water Consumption, m^3/min (gpm)	10.1 (2,673)	10.2 (2,683)	13.4 (3,529)	13.6 (3,604)

[1] HHV of Natural Gas is 53,014 kJ/kg (22,792 Btu/lb)
& HHV of Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb)
[2] CGE = (Hydrogen Product Heating Value)/ Fuel Heating Value, HHV
[3] ETE = (Hydrogen + - Power Heating Value)/ Fuel Heating Value, HHV

Exhibit 6-2 Effective Thermal Efficiencies



The environmental targets were described in Section 1.6. The projected annual emissions of SO_2, NOx, and particulate matter are shown in Exhibit 6-3 and the projected annual mercury emissions are shown in Exhibit 6-4. Projected annual CO_2 emissions are shown in Exhibit 6-5 . The following observations can be made:

- NOx emissions are lowest for the SMR cases.
- CO_2 emissions are lowest for the SMR case.

Exhibit 6-3 Annual Air Emissions

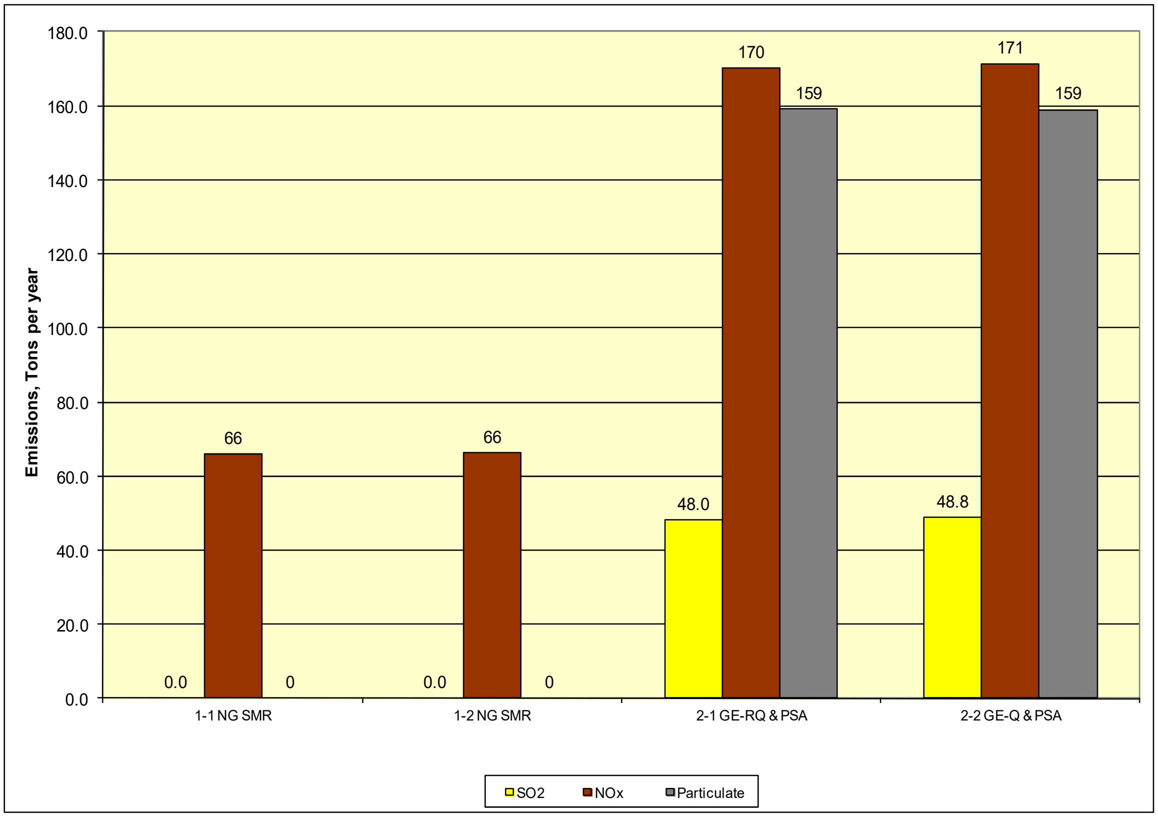


Exhibit 6-4 Annual Mercury Emissions

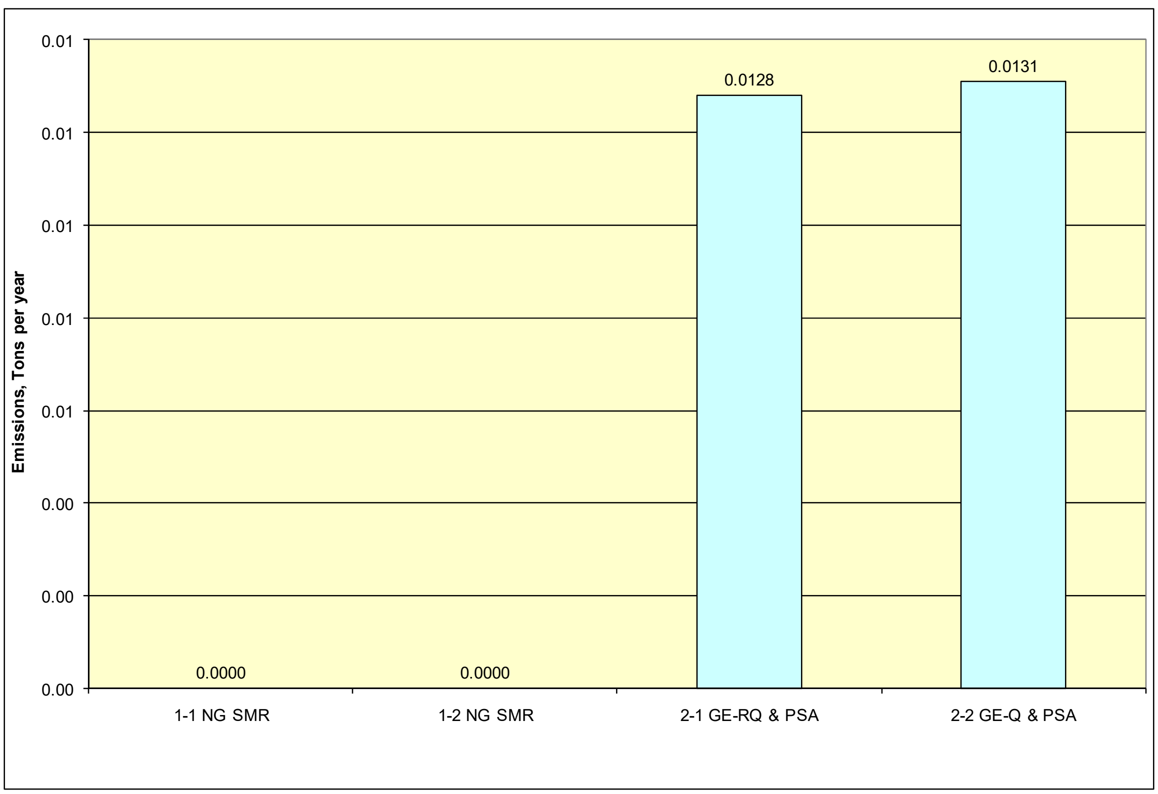
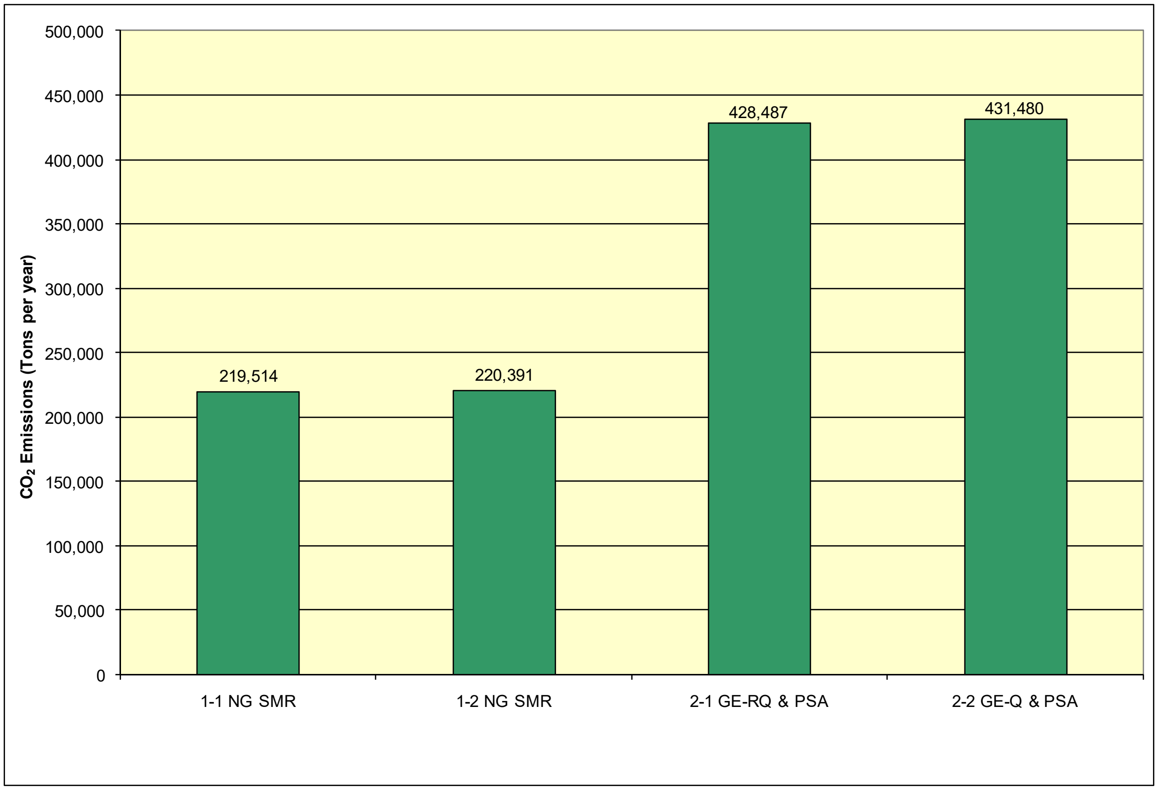
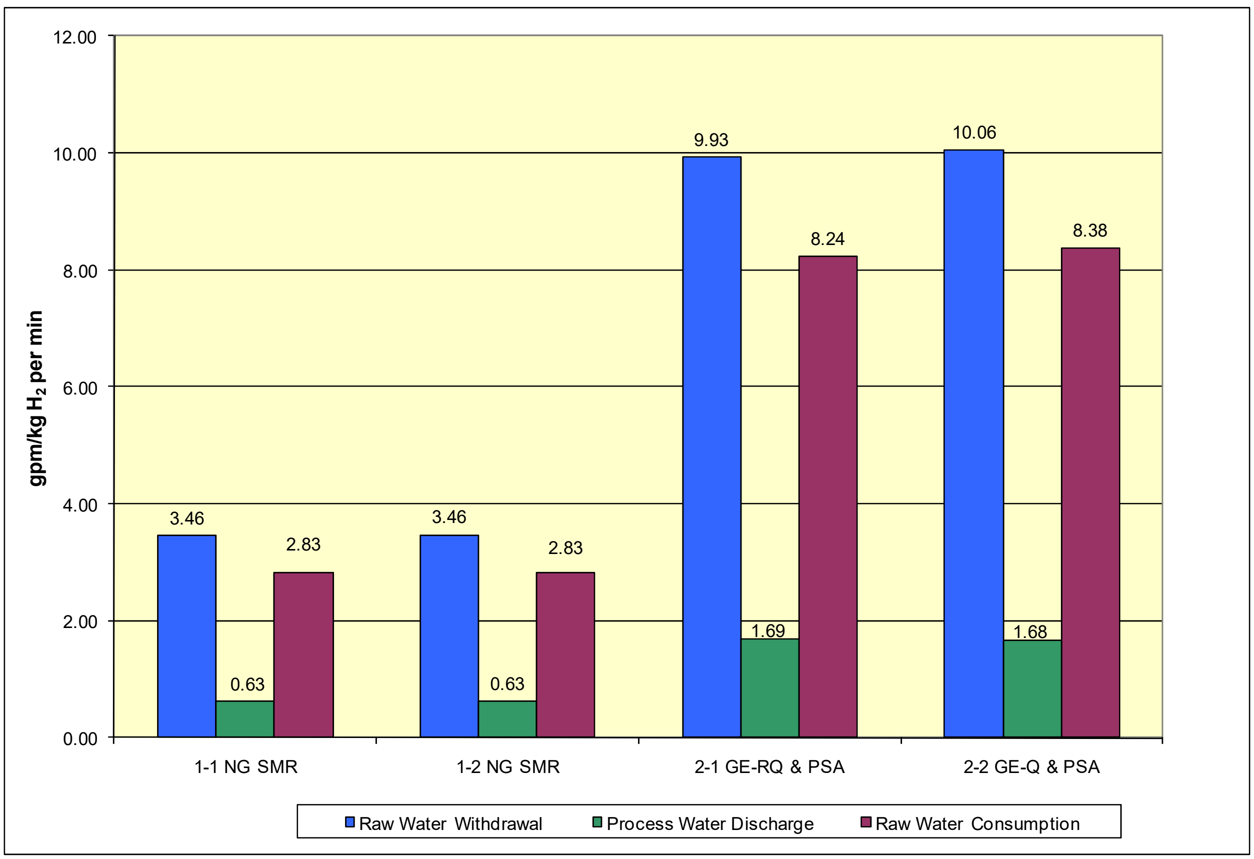


Exhibit 6-5 Annual CO_2 Emissions



Raw water withdrawal, process discharge, and raw water consumption – all normalized by net output – are presented in Exhibit 6-6. Raw water withdrawal is the difference between demand and internal recycle. Demand is the amount of water required to satisfy a particular process (slurry, quench, flue gas desulfurization makeup, etc.) and internal recycle is water available within the process (boiler feedwater blowdown, condensate, etc.). Raw water withdrawal is the water removed from the ground or diverted from a surface-water source for use in the plant. Raw water consumption is the portion of the raw water withdrawn that is evaporated, transpired, incorporated into products, or otherwise not returned to the water source it was withdrawn from. Raw water consumption is the difference between withdrawal and process discharge, and it represents the overall impact of the process on the water source, which in this study is considered to be 50 percent from groundwater (wells) and 50 percent from a municipal source. All plants are equipped with evaporative cooling towers, and all process blowdown streams are assumed to be treated and recycled to the cooling tower. The raw water usage is significantly lower for the SMR cases.

Exhibit 6-6 Raw Water Withdrawal, Discharge, and Consumption



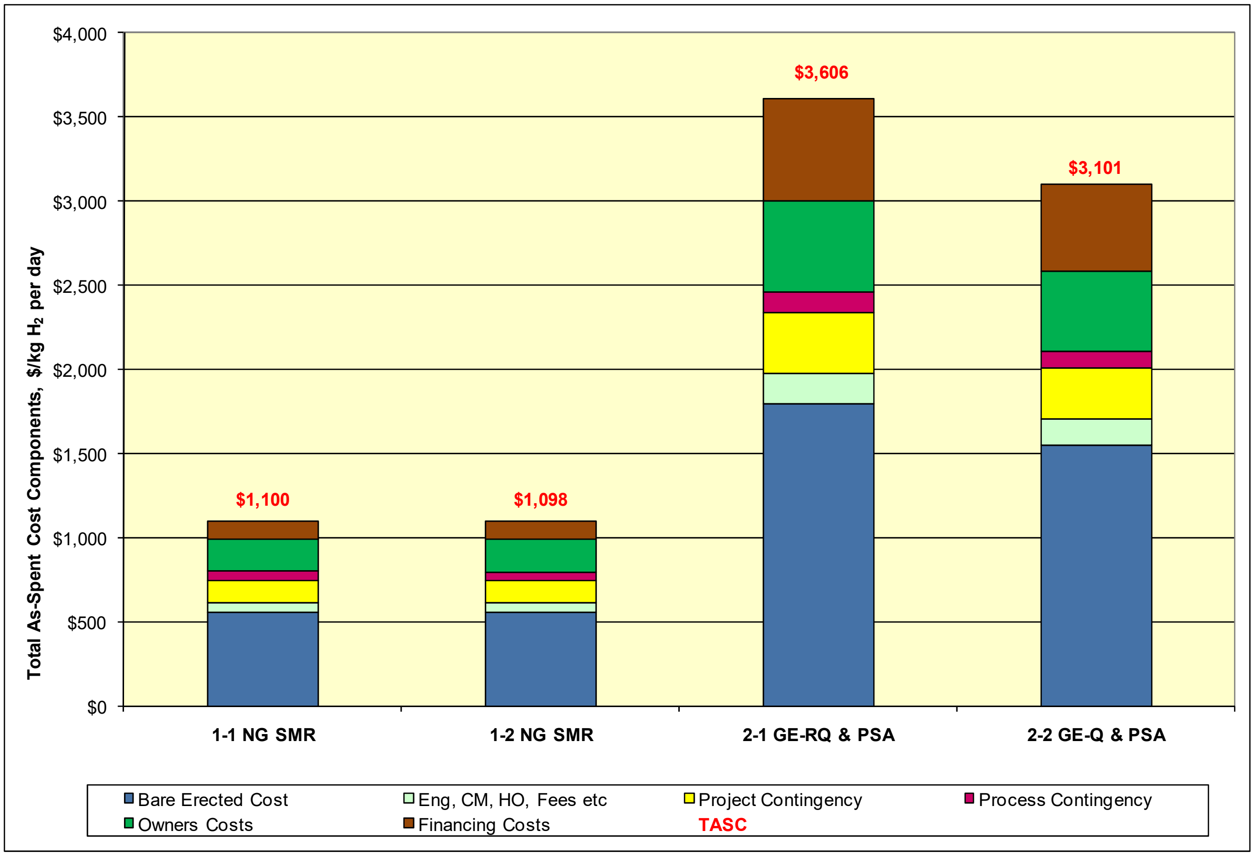
The cost estimates carry an accuracy of ±30 percent, consistent with the screening study level of engineering effort expended in the design. The results of the capital estimation calculations are shown in Exhibit 6-7. All capital and operations and maintenance (O&M) costs are presented as "overnight costs" expressed in June 2007 dollars. The estimating methodology and calculations are presented in Section 2.

The total plant cost (TPC) includes all equipment (complete with initial chemical and catalyst loadings), materials, labor (direct and indirect), engineering and construction management, and contingencies (process and project). The total overnight cost (TOC) for each plant was calculated by adding owner's costs to the TPC. Additional financing costs including escalation during construction were estimated and added to the TOC to provide the total as-spent cost (TASC). The TASC normalized on net hydrogen output is shown for each plant configuration in Exhibit 6-8. The coal to hydrogen cases are substantially more capital intensive than the SMR cases.

Exhibit 6-7 Capital Cost Estimation Results

Case	1-1	1-2	2-1	2-2
H_2 Production (kg H_2/day)	616,528	618,936	616,527	618,940
Bare Erected Cost, 1000$	$342,553	$343,355	$1,107,930	$958,576
Eng, CM, HO, Fees, etc., 1000$	$34,255	$34,335	$110,793	$95,858
Project Contingency, 1000$	$82,913	$83,107	$221,901	$190,503
Process Contingency, 1000$	$32,832	$32,909	$77,553	$61,446
Total Plant Cost, 1000$	**$492,553**	**$493,706**	**$1,518,158**	**$1,306,383**
Total Plant Cost, $/(kg H_2/day)	**$799**	**$798**	**$2,462**	**$2,111**
Owner's Cost, 1000$	$118,642	$118,926	$332,624	$291,118
Total Overnight Cost, 1000$	**$611,195**	**$612,632**	**$1,850,782**	**$1,597,501**
Total Overnight Cost, $/(kg H_2/day)	**$991**	**$990**	**$3,002**	**$2,581**
Financing Cost, 1000$	$66,905	$67,062	$372,543	$321,561
Total As-Spent Cost 1000$	**$678,100**	**$679,694**	**$2,223,325**	**$1,918,061**
Total As-Spent Cost, $/(kg H_2/day)	**$1,100**	**$1,098**	**$3,606**	**$3,101**

Exhibit 6-8 Total As-Spent Cost Components



Fixed and variable operating costs were estimated for each case. An emissions value of \$30/tonne of CO_2 emitted was also applied to reflect potential environmental regulations. A value of \$105/MWh was applied for any excess power generated or required for each case. This value is consistent with the cost of electricity (COE) generated in an environment where coal-based power plants are built with carbon capture and sequestration systems.

The first year costs of hydrogen (COH) were derived using the NETL Power Systems Financial Model (PSFM). COH is assumed to escalate at three percent per year for the thirty-year economic life of the plant. The project financial structure is representative of a high-risk fuels project with no loan guarantees or other government subsidies. The annual operating costs and resulting first year COH values are shown in Exhibit 6-9.

Exhibit 6-9 Cost of Hydrogen Estimation Results

Case	1-1	1-2	2-1	2-2
H_2 Production (kg H_2/day)	616,528	618,936	616,527	618,940
Fuel	Natural Gas	Natural Gas	Illinois #6	Illinois #6
Natural Gas Price ($/MMBtu)	$6.55	$6.55	N/A	N/A
Natural Gas Price ($/Ton)	$298.47	$298.47	N/A	N/A
Natural Gas Consumption, tpd	2,520	2,530	N/A	N/A
Coal Price ($/MMBtu)	N/A	N/A	$1.64	$1.64
Coal Price ($/Ton)	N/A	N/A	$38.19	$38.19
Coal Consumption, tpd	N/A	N/A	5,844	5,844
Capacity Factor, %	90%	90%	90%	90%
First Year Fuel Cost, $/yr	$247,114,305	$248,102,221	$73,307,753	$73,302,484
First Year Fixed O&M Cost, $/yr	$22,668,479	$22,703,075	$53,436,617	$47,083,365
First Year Variable O&M Cost, $/yr	$14,937,300	$14,978,807	$40,717,934	$36,216,846
First Year Electricity Cost (Revenue), $/yr	$28,311,444	$28,419,061	($5,927,191)	$29,081,317
First Year Carbon Emissions Value, $/yr	$5,974,186	$5,998,070	$11,661,501	$11,742,952
First Year Capital Charges, $/yr	$124,920,703	$125,214,495	$459,987,842	$397,140,235
Capital, $/kg H_2	0.62	0.62	2.27	1.95
Fixed O&M, $/kg H_2	0.11	0.11	0.26	0.23
Variable O&M, $/kg H_2	0.07	0.07	0.17	0.18
Power purchased (sold,) $/kg H_2	0.14	0.14	0.03	0.14
CO_2 Emissions Value, $/kg H_2	0.03	0.03	0.06	0.06
Fuel, $/kg H_2	1.22	1.22	0.36	0.36
Total First Year COH, $/kg H_2	**2.19**	**2.19**	**3.13**	**2.92**
First year COH, $/1000scf H_2	**5.59**	**5.58**	**7.96**	**7.45**
First Year CO_2 TS&M, $/yr	$23,520,371	$23,675,800	$41,344,396	$41,349,212
CO_2 TS&M, $/kg H_2	**0.12**	**0.12**	**0.20**	**0.20**
First year COH including CO_2 TS&M, $/kg H_2	**2.31**	**2.31**	**3.33**	**3.13**
First year COH including CO_2 TS&M, $/1000scf H_2	**5.88**	**5.88**	**8.49**	**7.97**

The first year COH results are shown graphically in Exhibit 6-10 with the capital cost, fixed operating cost, variable operating cost, and fuel cost components shown separately. CO_2 Transport, Storage and Monitoring (TS&M) costs are also shown as a separate bar segment. The following conclusions can be drawn:

- The COH is dominated by capital charges in both of the coal cases. The capital cost component of COH comprises 62-68 percent in the coal cases but only 23 percent in the natural gas SMR cases.
- The fuel cost component is relatively minor in the coal cases, representing 11-12 percent of the COH, but it dominates the natural gas SMR cases at 53 percent.

- The excess power generated in case 2-1 reduces the variable O&M for that case by 10 percent.
- The TS&M component of COH in all cases is 5-7 percent.

Exhibit 6-10 First Year COH by Cost Component (June 2007 dollars)

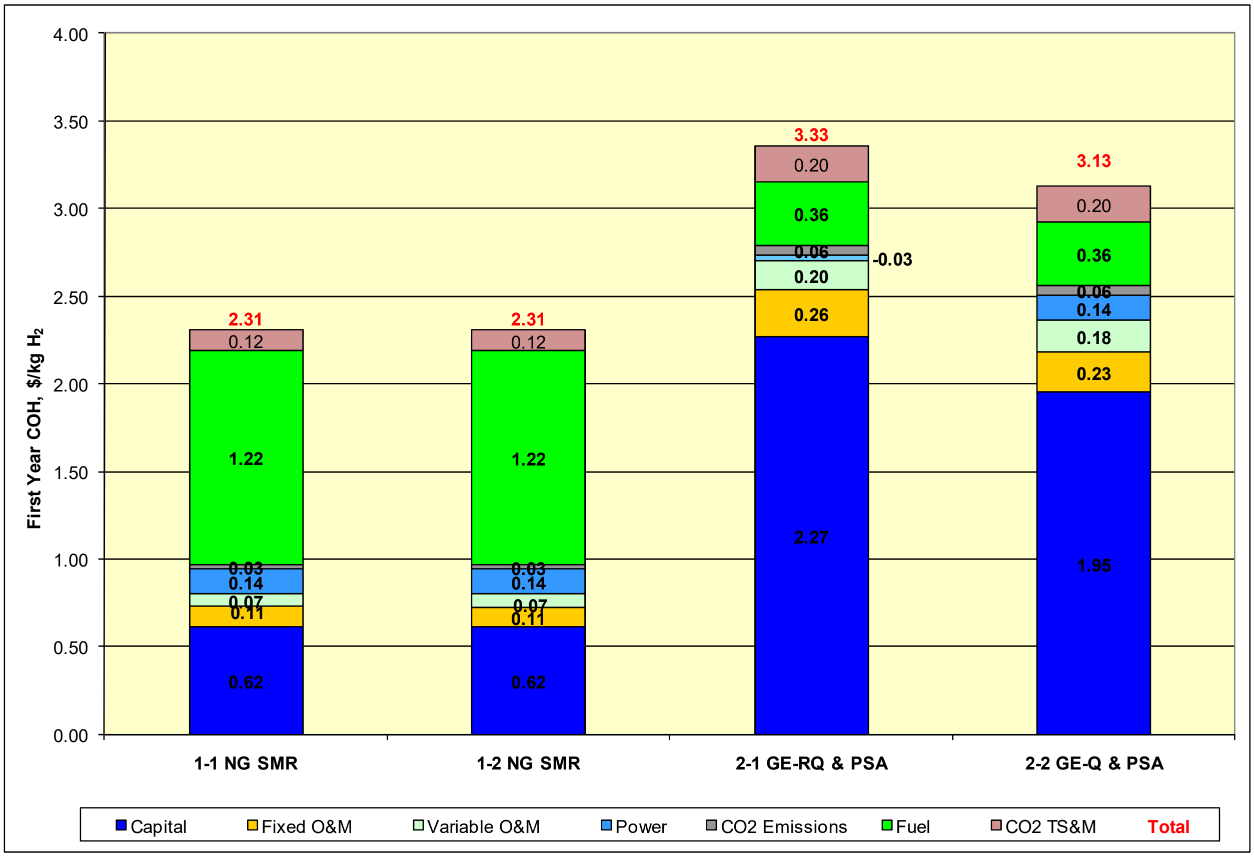


Exhibit 6-11 shows the first year COH sensitivity to the carbon dioxide emissions value. The COH values calculated for the $30/tonne assumed in this study are shown on each line. As expected, all cases show a slight linear increase in COH with an increase in the CO_2 emissions value. The natural gas based SMR cases increase by approximately four percent for a $100/tonne increase in value. The coal to hydrogen cases increase by approximately six percent for the same $100/tonne increase.

Exhibit 6-11 First Year COH Sensitivity to CO_2 Emissions Value (June 2007 dollars)

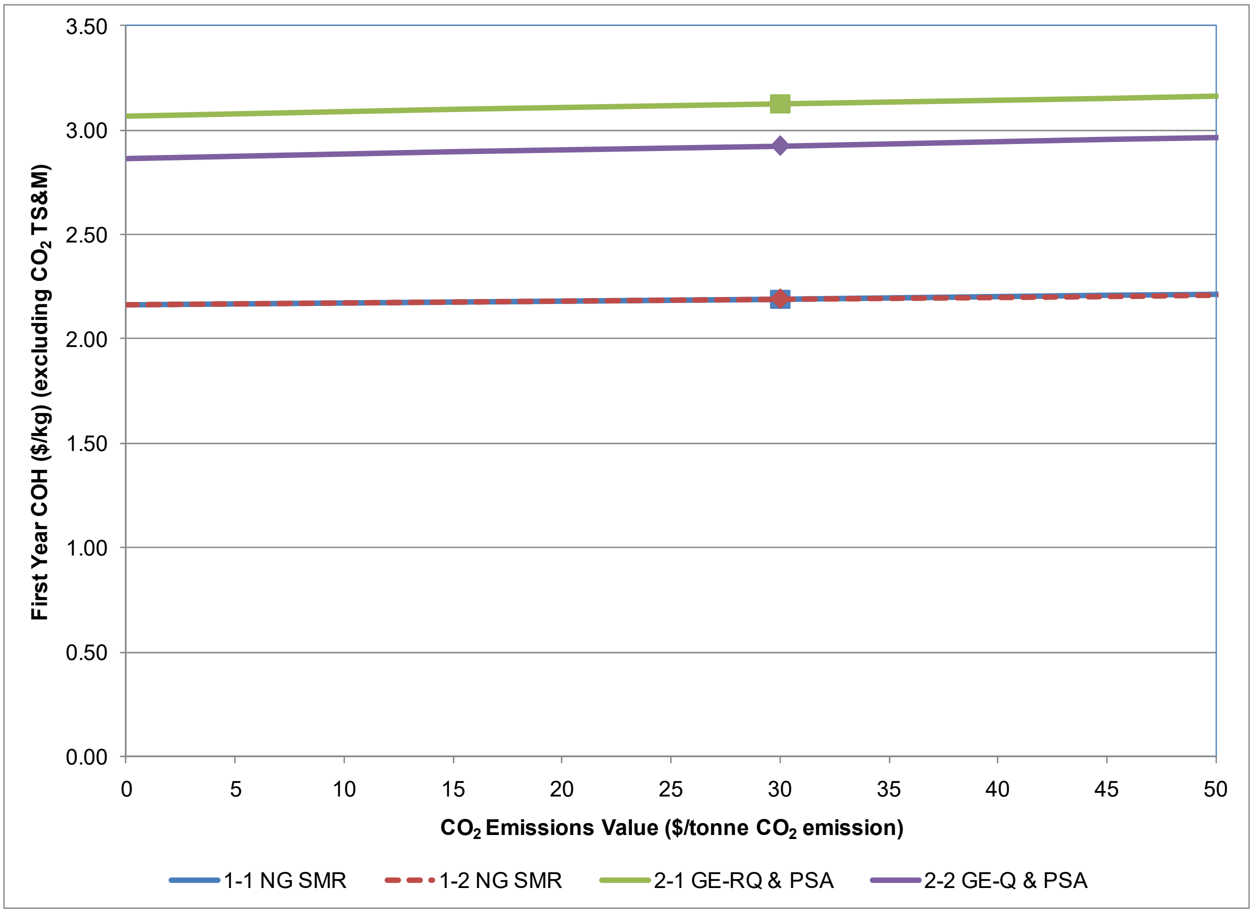


Exhibit 6-12 shows the first year COH sensitivity to the cost of electricity applied for excess and required power. The COH values calculated for the $105/MWh assumed in this study are shown on each line. As expected, the values for cases 1-1, 1-2, and 2-2 which require additional power show a slight linear increase in COH with an increase in the COE. Case 2-1 which generates some excess power shows a slight linear decrease in COH with an increase in the COE. The increase for the SMR cases which contain no power generation is approximately six percent for a $100 increase in COE. The increase for case 2-2 which requires some additional power is approximately five percent for a $100 increase in COE. The decrease for case 2-1 case which generates excess power is approximately one percent for a $100 increase in COE. The coal to hydrogen cases can potentially be modified to generate more electricity and thus reduce the COH, but the modifications must be balanced against the additional costs of equipment and consumables for the increased generation.

Exhibit 6-12 First Year COH Sensitivity to Cost of Electricity (June 2007 dollars)

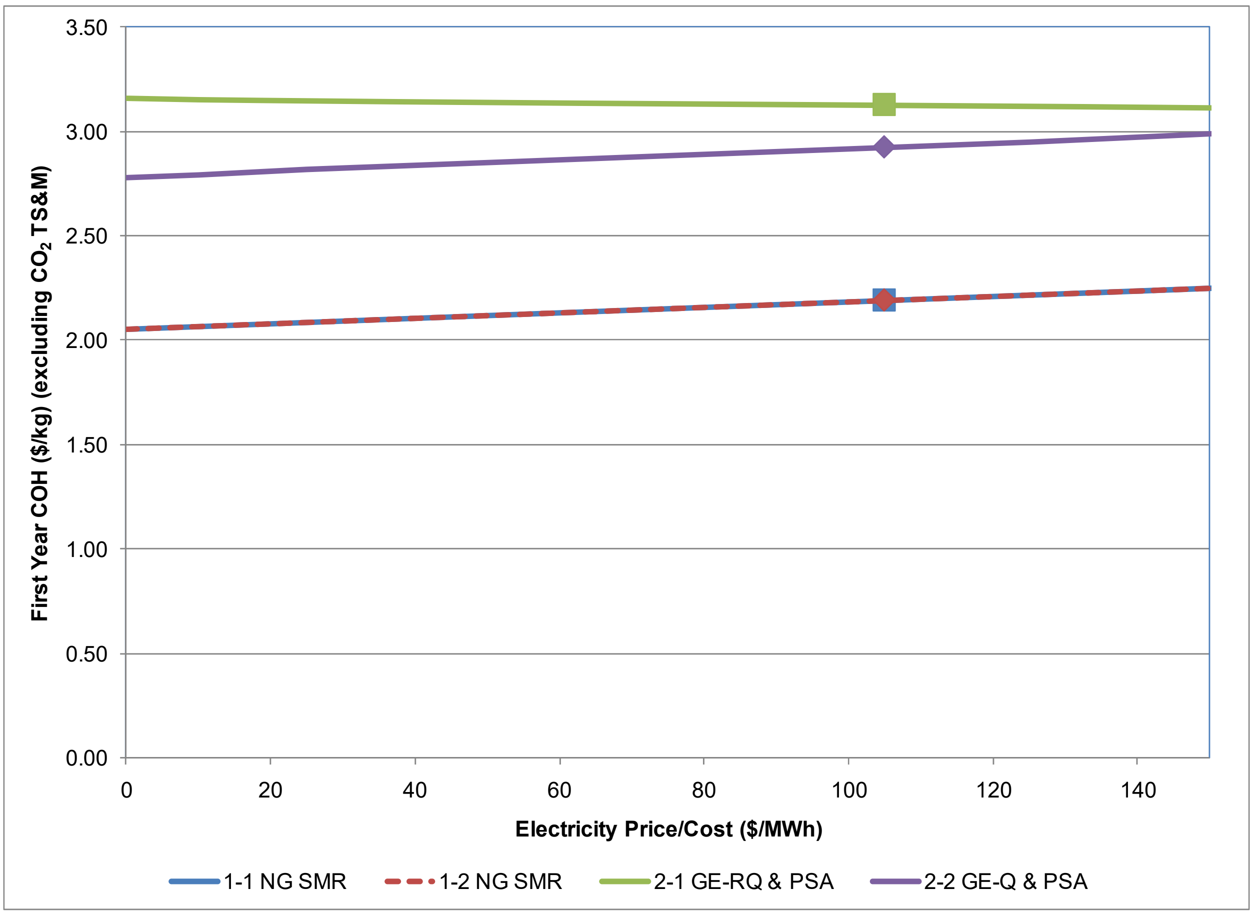
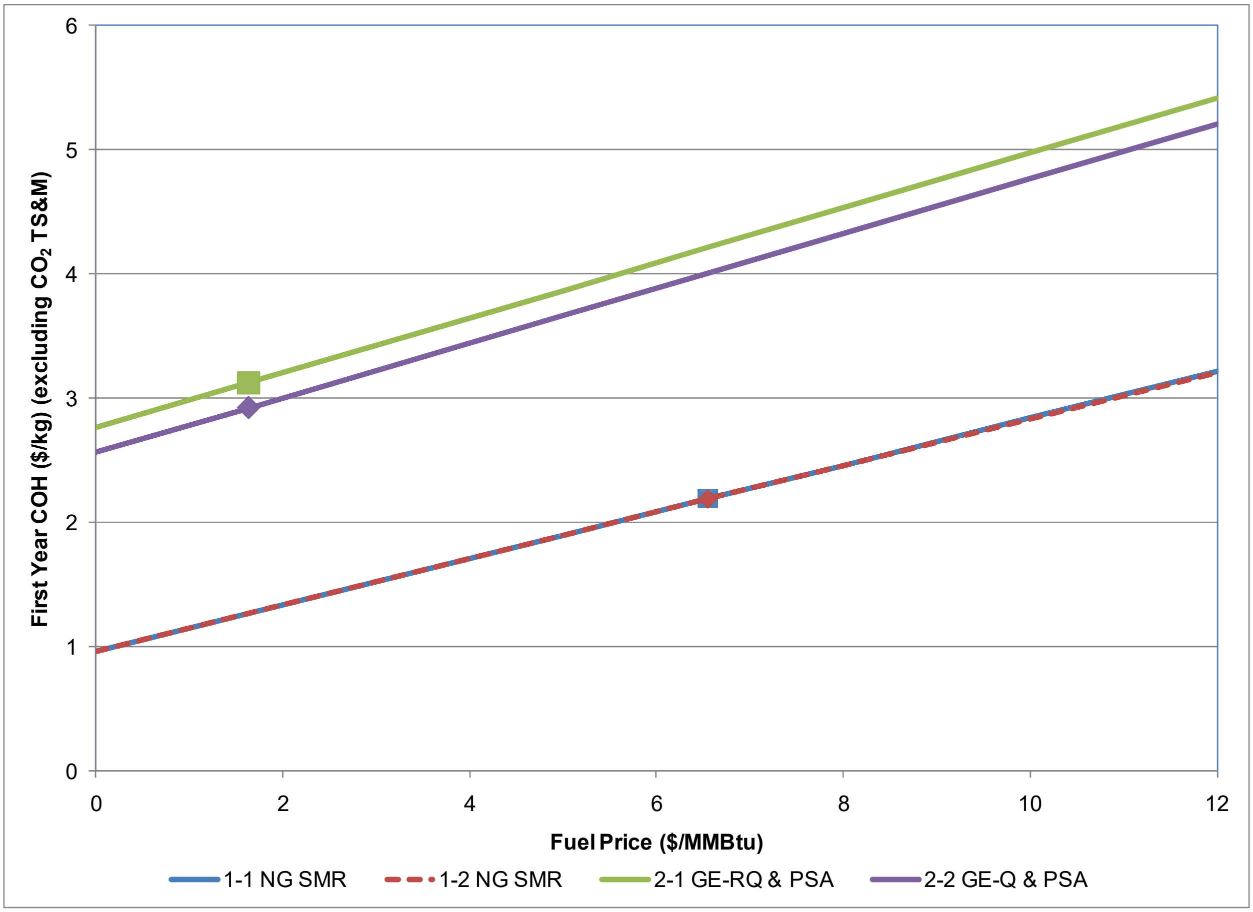


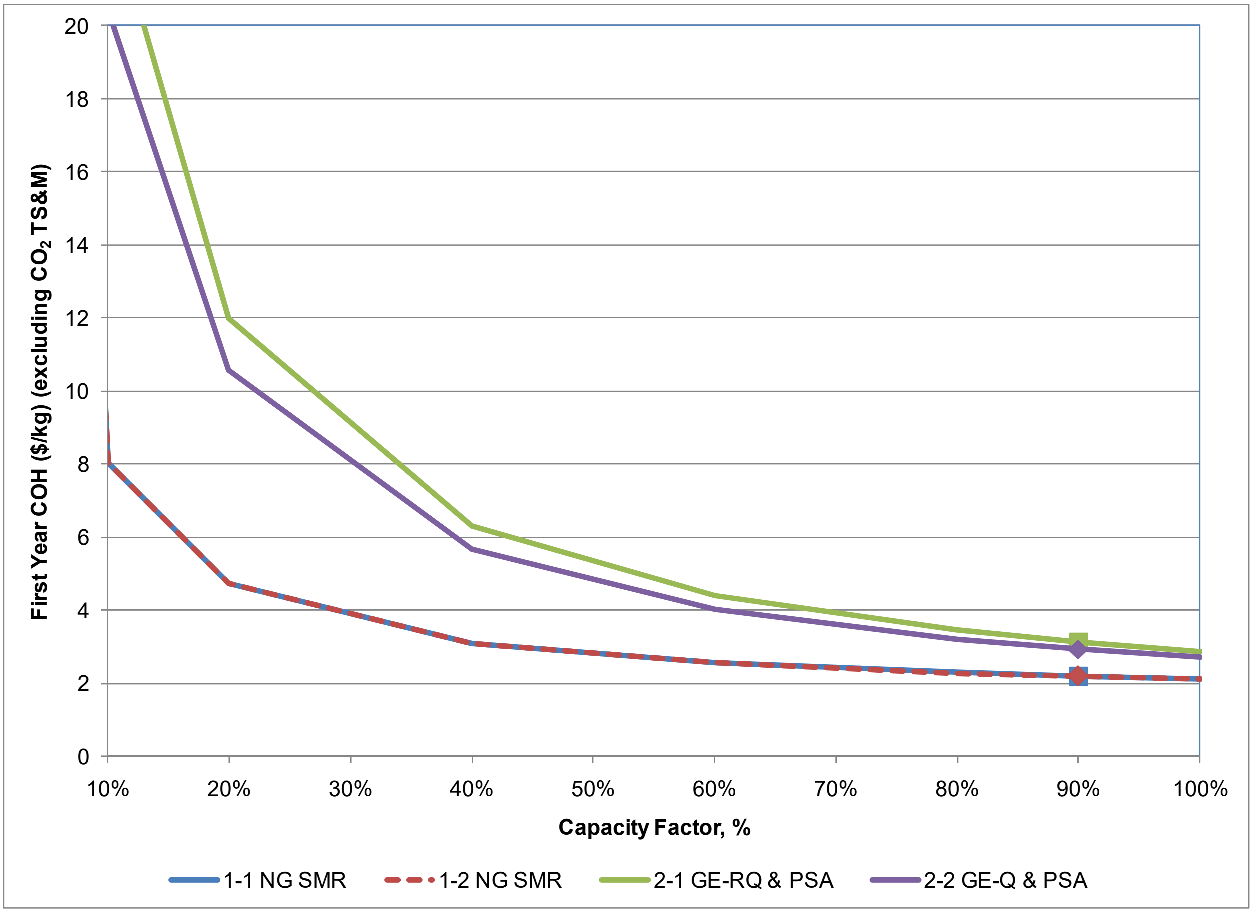
Exhibit 6-13 shows the first year COH sensitivity to fuel costs. Again the COH values calculated for the prices of natural gas and coal assumed in this study are shown on each line. As expected, all cases show a linear increase in COH with an increase in fuel prices. The COHs for the natural gas SMR cases increase by approximately 19 cents for each $/MMBtu increase in natural gas price. The COHs for the coal cases increase by approximately 22 cents for each $/MMBtu increase in coal prices. In general, the values for the SMR cases approach the values for the coal to hydrogen cases as natural gas prices increase and coal prices decrease.

Exhibit 6-13 First Year COH Sensitivity to Fuel Costs (June 2007 dollars)



The sensitivity of first year COH to capacity factor is shown in Exhibit 6-14. Again the COH values calculated for the capacity factor assumed in this study are shown on each line. At high capacity factors, the COH value for the coal cases approaches the COH for the natural gas cases. All cases show a substantial decrease in COH as the capacity factor increases. At very lower capacity factors (10 to 20 percent), the COH for the natural gas cases are less than one-half that of the two coal cases.

Exhibit 6-14 First Year COH Sensitivity to Capacity Factor (June 2007 dollars)



The first year COH for coal to hydrogen with CO_2 capture cases is estimated to be approximately \$1.08 to \$1.35 per kg of hydrogen greater than the COH for hydrogen generated from natural gas by the current commercial SMR technology with the addition of CO_2 capture. As gasification and CO_2 capture technologies become more commercially available, this differential should decrease and the coal to hydrogen cases would likely become more economically viable. Natural gas prices above \$9.60/MMBtu would also make the coal to hydrogen cases more competitively attractive.

7. REFERENCES

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