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Analysis of Integrated Gasification Fuel Cell Plant Configurations

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Exhibits

Acronyms and Abbreviations

AACE	Association for the Advancement of Cost Engineering
AC	Alternating current
AEO	Annual Energy Outlook
AGR	Acid gas removal
ASU	Air separation unit
BACT	Best available control technology
BEC	Bare erected cost
BFD	Block flow diagram
Btu	British thermal unit
Btu/h	British thermal unit per hour
Btu/kWh	British thermal unit per kilowatt hour
Btu/lb	British thermal unit per pound
Btu/scf	British thermal unit per standard cubic foot
C	Cost of equipment in study plant area of section
Carb	Number of atoms of carbon in the syngas
CCS	Carbon capture and storage
CF	Capacity factor
cm	Centimeter
CO_2	Carbon dioxide
COE	Cost of electricity
CoP	ConocoPhillips
COS	Carbonyl sulfide
Cref	Cost of equipment in reference plant area or section
CRT	Cathode ray tube
DCS	Distributed control system
DI	De-ionized
DOE	Department of Energy
EAF	Equivalent availability factor
E-Gas™	ConocoPhillips gasifier technology
EIA	Energy Information Administration
EPA	Environmental Protection Agency
EPC	Engineer/Procure/Construct
EPRI	Electric Power Research Institute
EPCM	Engineering/Procurement/Construction Management
F	Capacity of study plant area or section
Fref section	Capacity of reference plant area or
ft	Foot, Feet
gpm	Gallons per minute
h	Hour
H_2	Hydrogen
HCl	Hydrogen chloride
Hg	Mercury
HHV	Higher heating value
hp	Horsepower
HP	High pressure
HRSG	Heat recovery steam generator
IGCC	Integrated gasification combined cycle
IGFC	Integrated gasification fuel cell
IOU	Investor-owned utility
IP	Intermediate pressure
ISO	International Standards Organization
kg/GJ	Kilogram per gigajoule
kg/h	Kilogram per hour
kJ	Kilojoules
kJ/h	Kilojoules per hour
kJ/kg	Kilojoules per kilogram
kV	Kilovolt
kW	Kilowatt
kWe	Kilowatts electric
kWh	Kilowatt-hour
lb	Pound
lb/h	Pounds per hour
lb/MMBtu	Pounds per million British thermal units
lb/MWh	Pounds per megawatt hour
LCOE	Levelized cost of electricity
LGTI	Louisiana Gasification Technology, Inc.
LHV	Lower heating value
LNB	Low NOx burner
LP	Low pressure
m	Meters
m^3/min	Cubic meter per minute
MAF	Moisture and Ash Free
MDEA	Methyldiethanolamine
MMBtu	Million British thermal units (also shown as 10^6 Btu)
MMBtu/h	Million British thermal units (also shown as 10^6 Btu) per hour
MMkJ kJ)	Million kilojoules (also shown as 10^6
MMkJ/h	Million kilojoules (also shown as 10^6 kJ) per hour
MPa	Megapascals
MWh	Megawatt-hour
N in parallel	Number of study plant areas or sections
Nref	Number of reference plant areas or sections in parallel
N/A	Not applicable
NERC Council	North American Electric Reliability
NETL Laboratory	National Energy Technology
NGCC	Natural gas combined cycle
Nm^3	Normal cubic meter

NOx	Oxides of nitrogen
NSPS	New Source Performance Standards
O&M	Operation and maintenance
OC_{Fn}	Category n fixed operating cost for the initial year of operation
OC_{Vnq}	Category n variable operating cost for the initial year of operation
OSHA	Occupational Safety and Health Administration
Oxy syngas	Number of atoms of oxygen in the
PC	Pulverized coal
POTW	Publicly Owned Treatment Works
ppm	Parts per million
ppmv	Parts per million volume
ppmvd	Parts per million volume, dry
PSA	Pressure Swing Adsorption
psia	Pounds per square inch absolute
psid	Pounds per square inch differential
psig	Pounds per square inch gage
R&D	Research and development
S	Scaling factor for plant areas or section cost
SCOT	Shell Claus Off-gas Treating
SCR	Selective catalytic reduction
SG	Specific gravity
SGC	Synthesis gas cooler
SGS	Sour gas shift
SO_2	Sulfur dioxide
SOFC	Solid oxide fuel cell
SRU	Sulfur recovery unit
T	Temperature
TASC	Total as-spent cost
TOC	Total overnight cost
TPC	Total plant cost
TGTU	Tail gas treating unit
Tonne	Metric Ton (1000 kg)
TPC	Total plant cost
TPD	Tons per day
TS&M	Transport, storage and monitoring
V-L	Vapor Liquid portion of stream (excluding solids)
vol%	Volume percent
wt%	Weight percent
$/MMBtu	Dollars per million British thermal units
$/MMkJ	Dollars per million kilojoul

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Executive Summary

This report presents the results of a Pathway Study for coal-based, integrated gasification fuel cell (IGFC) power systems with carbon capture and sequestration (CCS). The results quantify the performance and cost benefits for a series of projected gains made through the development of advanced technologies or improvements in plant operation and maintenance. The results represent the potential future benefits of IGFC technology development. They also provide DOE with a basis to select the most appropriate development path for IGFC, and to measure and prioritize the contribution of its R&D program to future power systems technology.

The IGFC plants in this study apply advanced, planar, solid oxide fuel cell (SOFC) technology with separate anode and cathode off-gas steams, and incorporate anode off-gas oxy-combustion for nearly complete carbon capture. The SOFC simulations utilize the expected operating conditions and performance capabilities of this solid oxide fuel cell technology, operating initially at atmospheric-pressure. The power plant cost and performance estimates reflect performance projections based on the current state of SOFC development, as well as projecting a pathway of SOFC technology development advances. The following fuel cell system advances are incorporated in a cumulative manner:

- Reduced SOFC stack performance degradation
- Reduced stack overpotential
- SOFC stack cost reduction
- Improved inverter efficiency
- Pressurized SOFC.

Advances in IGFC plant operation are also included in the pathway, being represented as improved plant availability and capacity factor achieved through advanced component monitoring, improved maintenance practices, and plant operation experience.

This document characterizes two parallel pathways of IGFC development, both incorporating CCS, and estimates overall plant performance and cost along these pathways in a consistent technical and economic manner. The first pathway applies conventional coal gasification technology, the ConocoPhillips E-Gas™ gasifier (CoP). This gasification technology produces syngas having limited methane content, roughly 6 mole percent. Increased syngas methane content is projected to benefit the performance of the IGFC plant. The first pathway consists of two scenarios. Scenario 1 applies atmospheric-pressure SOFC and follows both SOFC technology advances and a near-term enhancement in the conventional gasifier technology to generate syngas having slightly higher methane content. The potential benefit of an additional, near-term technology enhancement step with conventional gasifier technology and atmospheric-pressure SOFC has also been explored in Case 1-6 as a branch-point to Scenario 1, considering the use of natural gas injection into the coal syngas as a means to achieve significantly higher syngas methane content. Scenario 2 considers the incorporation of pressurized-SOFC technology as a longer term enhancement, and represents an additional branch-point to Scenario 1.

The second pathway applies an advanced, catalytic coal gasification technology projected to produce syngas having very high methane content of roughly 30 mole percent, greatly improving the IGFC performance. This pathway follows similar advances in SOFC technology development as used for the pathway with the conventional gasifier.

Summaries of the pathway parameters considered in this study are presented in Exhibit ES-1and Exhibit ES-2. The Baseline plant utilizes SOFC operating conditions and performance capabilities based on the current status of sub-scale SOFC testing.

Exhibit ES-1 Conventional Gasifier IGFC Pathway Parameters (Scenarios 1, and 2)

Case	Pathway Parameter	Gasifier (methane %[1]	SOFC Pressure & Overpotential	Capacity Factor %	Cell Degradation %/1000 h	SOFC Stack Cost $/kW SOFC[2]	Inverter Efficiency (%)
1-1	**Baseline Atm-pressure SOFC**	CoP (6%)	15.6 psia 140 mV	80	1.5	296	97
1-2	Degradation	CoP (6%)	15.6 psia 140 mV	80	**0.2**	296	97
1-3	Overpotential	CoP (6%)	15.6 psia **70 mV**	80	0.2	296	97
1-4	Capacity Factor	CoP (6%)	15.6 psia 70 mV	**85**	0.2	296	97
1-5	Gasifier	**Enhanced (10%)**	15.6 psia 70mV	85	0.2	296	97
1-6 Branch	Natural Gas Injection	Enhanced (24.6%)	15.6 psia 70mV	85	0.2	296	97
1-7	Capacity Factor	Enhanced (10%)	15.6 psia 70 mV	**90**	0.2	296	97
1-8	SOFC cost reduction)	Enhanced (10%)	15.6 psia 70 mV	90	0.2	**268**	97
1-9	Inverter Efficiency	Enhanced (10%)	15.6 psia 70 mV	90	0.2	268	**98**
2-1	**Pressurized SOFC**	Enhanced (11%)	**285 psia** 70 mV	85	0.2	442	97
2-2	Capacity Factor	Enhanced (11%)	285 psia 70 mV	**90**	0.2	442	97
2-3	SOFC cost reduction)	Enhanced (11%)	285 psia 70 mV	90	0.2	**414**	97
2-4	Inverter Efficiency	Enhanced (11%)	285 psia 70 mV	90	0.2	414	**98**

1 – Methane content (mole percent) of clean, dry syngas
2 – Cost (TPC) of the SOFC stack unit (stacks, enclosures, inverters) in $ per kW of SOFC AC output
3 – Natural gas injected in the syngas as percent of the total fuel energy input

Exhibit ES-2 Catalytic Gasifier IGFC Pathway Parameters (scenarios 3 and 4)

Case	Pathway Parameter	SOFC Pressure & Overpotential	Capacity Factor %	Cell Degradation %/1000 h	SOFC Stack Cost $/kW	Inverter Efficiency %
3-1	**Baseline Atm-pressure SOFC**	15.6 140 mV	80	1.5	296	97
3-2	Degradation	15.6 140 mV	80	**0.2**	296	97
3-3	Overpotential	15.6 **70 mV**	80	0.2	296	97
3-4	Capacity Factor	15.6 70 mV	**85**	0.2	296	97
3-5	Capacity Factor	15.6 70 mV	**90**	0.2	296	97
3-6	SOFC cost reduction	15.6 70 mV	90	0.2	**268**	97
3-7	Inverter Efficiency	15.6 70 mV	90	0.2	296	**98**
4-1	**Pressurized SOFC**	**285 psia** 70 mV	85	0.2	442	97
4-2	Increased Capacity Factor	285 psia 70mV	**90**	0.2	442	97
4-3	SOFC cost reduction	285 psia 70 mV	90	0.2	**414**	97
4-4	Inverter Efficiency	285 psia 70 mV	90	0.2	414	**98**

1 – Cost (TPC) of the SOFC stack unit (stacks, enclosures, inverters) in $ per kW of SOFC AC output

Scenario1 represents the pathway for a conventional gasifier plant generating syngas for use in an atmospheric-pressure SOFC power island. The Case 1-6 branch-point simulates an alternative approach for generating a high methane syngas without the need to develop advanced gasification technology, accomplishes this by injecting sufficient quantity of natural gas into the conventional gasifier syngas. It branches from Scenario 1 after Case 1-5. Scenario 2 represents the transition of Scenario 1 to a pressurized-SOFC power island configuration after several pathway enhancements in the Scenario 1 plant, branching from Scenario 1 after Case 1-5.

Scenario 3 applies an advanced catalytic gasifier for the production of a high methane syngas for use in an atmospheric-pressure SOFC power island. Scenario 4 transitions to a pressurized-SOFC power island. It branches from Scenario 3 after Case 3-5. Components for each plant configuration are described in more detail in the corresponding report sections for each scenario.

The design and cost bases for this evaluation have been largely extracted from "Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity," (herein referred to as the Bituminous Baseline report) published by NETL in 2010 [1], so these IGFC plant results can be directly compared to the baseline results for other fossil fuel power generation technologies. The basis for the design of the SOFC power island components and their cost estimates are described in Section 2 of the report. All of the IGFC plants are designed for baseload operation with the following key basis specifications:

- Illinois No.6 coal
- ISO ambient conditions
- conventional cryogenic air separation technology
- conventional dry syngas cleaning and polishing technology
- net plant capacity of 550,000 kW.

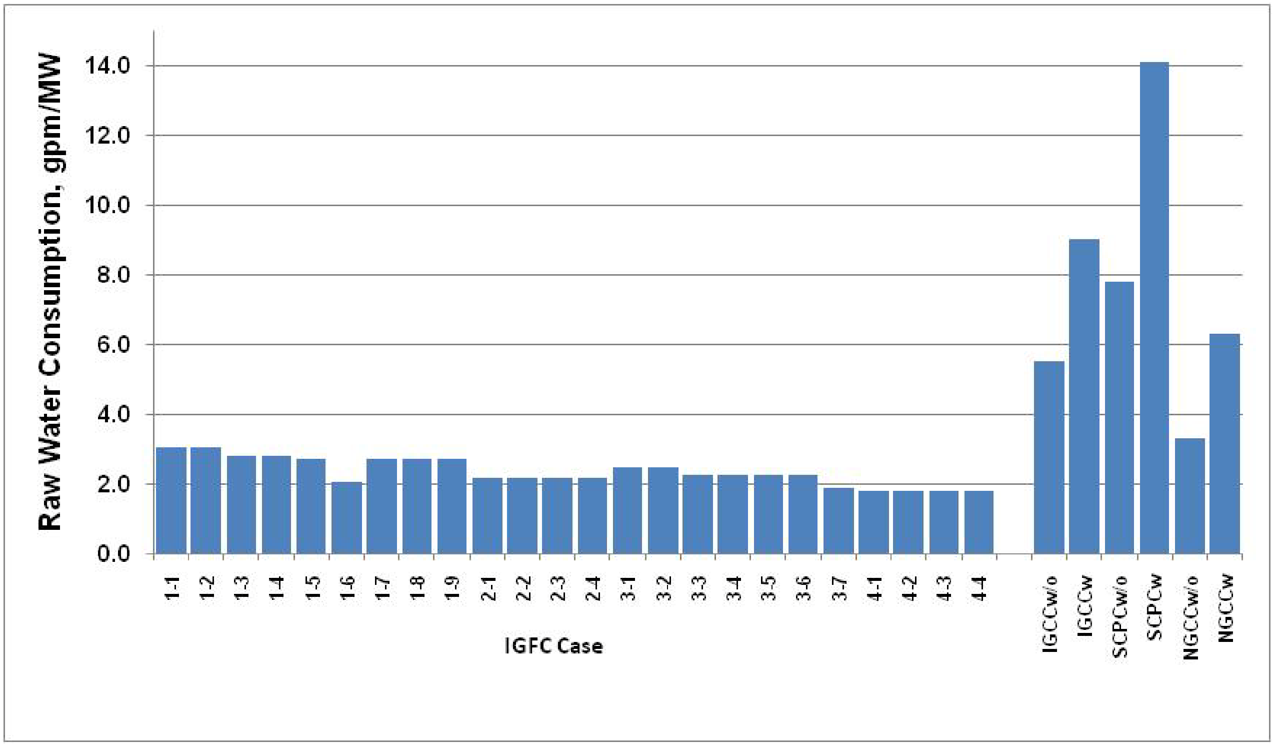
Summary listings of the performance and cost results for both of the pathways and all of the study cases are provided in Exhibit ES-8, Exhibit ES-9, Exhibit ES-10, Exhibit ES-11.Shown are some plant sizing factors (coal feed rate and number of parallel processing trains in the plant), performance factors (cell voltage, net plant efficiency, raw water consumption, and CO_2 emission), and cost factors (plant total overnight cost, and first-year cost of electricity). The exhibits show the increased performance and cost reduction that result as the IGFC technologies mature and innovations are developed. Similar performance and cost factors are listed in Exhibit ES-12 for conventional fossil fuel power generation technologies.

Conventional fossil fuel power plants such as those characterized in the Bituminous Baseline report apply a design basis of 90 percent CO_2 removal. With the higher power conversion efficiencies in the IGFC plants than in conventional fossil fuel power plants, and CO_2 removal efficiencies that ranges from >98 percent for pressurized-SOFC plants, and >99 percent for atmospheric-pressure SOFC plants, the IGFC plant emissions of CO_2 are lower than in conventional fossil fuel power plants by a factor of 20 to 50. The emissions other gas phase contaminants are also very limited in the IGFC power plants because the syngas has very stringent cleaning standards for sulfur species, halides, and trace metals. Finally, because of the use of an oxy-combustor for CO_2 capture, all of the IGFC plant's remaining syngas environmental contaminants are sequestered with the CO_2 product stream. The only issue relating to the IGFC plant's CO_2 product is its oxygen content, at about 2 to 3 mole percent being greater that typical CO_2 composition specifications for sequestration. It is expected that low-temperature, high-pressure processing of the CO_2 product stream can reduce the oxygen content to acceptable levels with little cost or performance impact if this is required.

The raw water consumption rate in the IGFC plant cases is compared to raw water emissions from conventional fossil fuel power plants in Exhibit ES-3. The IGFC raw water consumption is less than 50 percent of the raw water consumption from the conventional plants with CCS due to the IGFC plant high thermal efficiency and the oxy-combustion process characteristic of recovering and recycling all of the anode off-gas water vapor content.

The first-year cost of electricity (COE) for the conventional gasifier IGFC pathway and the catalytic gasifier pathway are shown in Exhibit ES-4 and Exhibit ES-5. The exhibits also display the net plant efficiency for each pathway. The conventional gasifier pathway with atmospheric-pressure SOFC (Scenario 1) shows a significant reduction in first-year COE with each of the first five pathway steps, followed by smaller impacts from SOFC cost reduction and inverter efficiency improvement. Cell degradation rate reduction and overpotential reduction represent important SOFC technology gains, and improved plant availability represents a key integrated power plant gain. The enhanced conventional gasifier, with increased syngas methane content, is a near-term enhancement having significant impact. Also, the injection of natural gas, with a natural gas price of $6.55/MMBtu, reduces the COE about $6/MWh by efficiently increasing the syngas methane content and represents another near-term enhancement possibility.

Exhibit ES-3 IGFC Raw Water Consumption Compared with Conventional Fossil Fuel Plants



The conventional gasifier pathway (Scenario 1) net plant efficiency shows gains of 3.7 and 2.3 percentage-points from the reduction in SOFC cell overpotential and the enhanced gasifier, respectively. The injection of natural gas into the IGFC syngas (Case 1-6) results in an almost 5 percentage-point gain in net plant efficiency.

The introduction of pressurized-SOFC into the conventional gasifier pathway (Scenario 2) results in a substantial increase of about 4 percentage-points in the net plant efficiency. Due to the large increase in equipment cost with pressurization, though, the COE is only reduced about $3/MWh by pressurization, reaching a COE about the same as achieved by natural gas injection.

Exhibit ES-5 focuses on the advanced catalytic gasifier pathway. As with the conventional gasifier pathway, the catalytic gasifier pathway with atmospheric-pressure SOFC (Scenario 3) shows a significant reduction in COE with each of the first four pathway steps, followed by smaller impacts from SOFC cost reduction and inverter efficiency improvement. Again, cell degradation rate reduction and overpotential reduction represent important SOFC technology gains, and improved plant availability represents a key integrated power plant gain. The catalytic gasifier pathway (Scenario 3) net plant efficiency shows a gain of 4.6 percentage-points from the reduction in SOFC cell overpotential, being the only pathway parameter having significant influence on efficiency.

The introduction of pressurized-SOFC into the catalytic gasifier pathway (Scenario 4) results in a substantial increase of almost 5 percentage-points in the net plant efficiency, reaching a level of 60 percent (HHV). Due to the large increase in equipment cost with pressurization, though, the COE in this case is slightly increased by pressurization, and no benefit is found to result with pressurization.

Exhibit ES-6 plots the COE against the CO_2 emissions price, a parameter representing the cost debit assessed for emitting CO_2 to the atmosphere, and assumed to range up to 100 $/tonne of CO_2 emitted. Lines are plotted for conventional IGCC, SCPC, and NGCC COE, both with and without CCS. Lines representing the COE for the baseline and the most advanced IGFC cases with atmospheric-pressure SOFC are also plotted. These lines are nearly horizontal because of the very small CO_2 emissions from the IGFC plants. Conventional and catalytic gasifier pathways are considered, as well as a line for natural gas injection. For very low CO_2 emission prices the competing conventional technologies are SCPC and NGCC, both without CCS. At the low CO_2 emissions prices, the COE for SCPC and NGCC without CCS is about the same as the COE for the most advanced catalytic gasifier IGFC case (Case 3-7). As the CO_2 emissions price increases, other IGFC pathway cases become competitive. Thus, at a CO_2 emissions price of about 35 $/tonne the conventional gasifier natural gas injection case (Case 1-6) and the most advanced conventional gasifier case (Case 1-9) both become competitive. Over a broad natural gas price range of 4 to 12 $/MMBtu, the Exhibit shows that the natural gas injection options maintains significant merit. As the CO_2 emissions price approaches 100 $/tonne, all of the IGFC pathway cases become competitive with all of the conventional fossil fuel power plant technologies, with most IGFC cases having very large COE advantage.

Exhibit ES-7 compares the cost of CO_2 avoided for all of the cases in the IGFC pathways. The avoided cost is relative to the supercritical PC plant without CCS. The plot strongly illustrates the cost advantage of the catalytic gasifier technology over the conventional gasifier technology, and shows the lack of a significant advantage of pressurized-SOFC over atmospheric-pressure SOFC. The cost of CO_2 avoided of the most advanced cases in the catalytic gasifier pathway approach very small values.

These results indicate that:

- The IGFC power plant technologies evaluated have significant environmental advantages over all other fossil fuel power plants, being near-zero emission power plants.
- IGFC using a catalytic coal gasifier and atmospheric-pressure SOFC will provide the greatest benefits, with the cost of electricity projected to be significantly lower than IGCC, PC, and NGCC all with CCS. IGFC with a catalytic coal gasifier has the potentially for costs comparable to IGCC, PC, and NGCC without CCS. This IGFC system requires development of the catalytic gasifier, development of the SOFC stack unit capable of reliable operation on high-methane syngas, and the development of the oxy-combustor technology.
- IGFC using a catalytic coal gasifier and pressurized-SOFC provides no cost benefit over IGFC with a catalytic gasifier and atmospheric-pressure SOFC. The conventional

gasifier applied with pressurized-SOFC shows a moderate cost improvement over the conventional gasifier with atmospheric-pressure SOFC. The IGFC plant configuration and operating conditions selected for the pressurized SOFC evaluation in this study have not been optimized and, thus, there are opportunities for further benefit. This IGFC configuration requires development of the pressurized-SOFC technology.

- IGFC using conventional gasifier technology or an enhanced-conventional gasifier technology that results in a moderate increase in methane content in the syngas will have significant performance and cost advantages over IGCC and PC with CCS, although not as great as those provided by the catalytic gasifier. In this case, development is limited to the SOFC technology and the oxy-combustor technology.
- Natural gas injection at rates up to 43 percent of the total plant fuel energy input can greatly increase the performance and cost potential of the IGFC plant using conventional or enhanced-conventional coal gasification. The COE of IGFC with natural gas injection is significantly lower than that of NGCC with CCS, and reaches parity with NGCC without CCS at a natural gas price of about 10 $/MMBtu. IGFC with natural gas injection can have COE lower than IGFC with conventional gasification or catalytic gasification under baseline SOFC conditions. The use of natural gas injection into the coal-syngas stream provides an opportunity to achieve significant IGFC plant performance and cost enhancements with limited need for advanced technology development.

There are other technological innovations that might also benefit the IGFC power plant performance and cost, such as humid gas cleaning (HGC), the ion transport membrane (ITM) technology for oxygen separation incorporating integration with the pressurized SOFC cathode air compressor, and shock wave CO_2 compression. It is recommended that these technology advances be included in future IGFC pathway evaluations.

Exhibit ES-4 Conventional Gasifier IGFC Pathway (Scenarios 1, and 2)

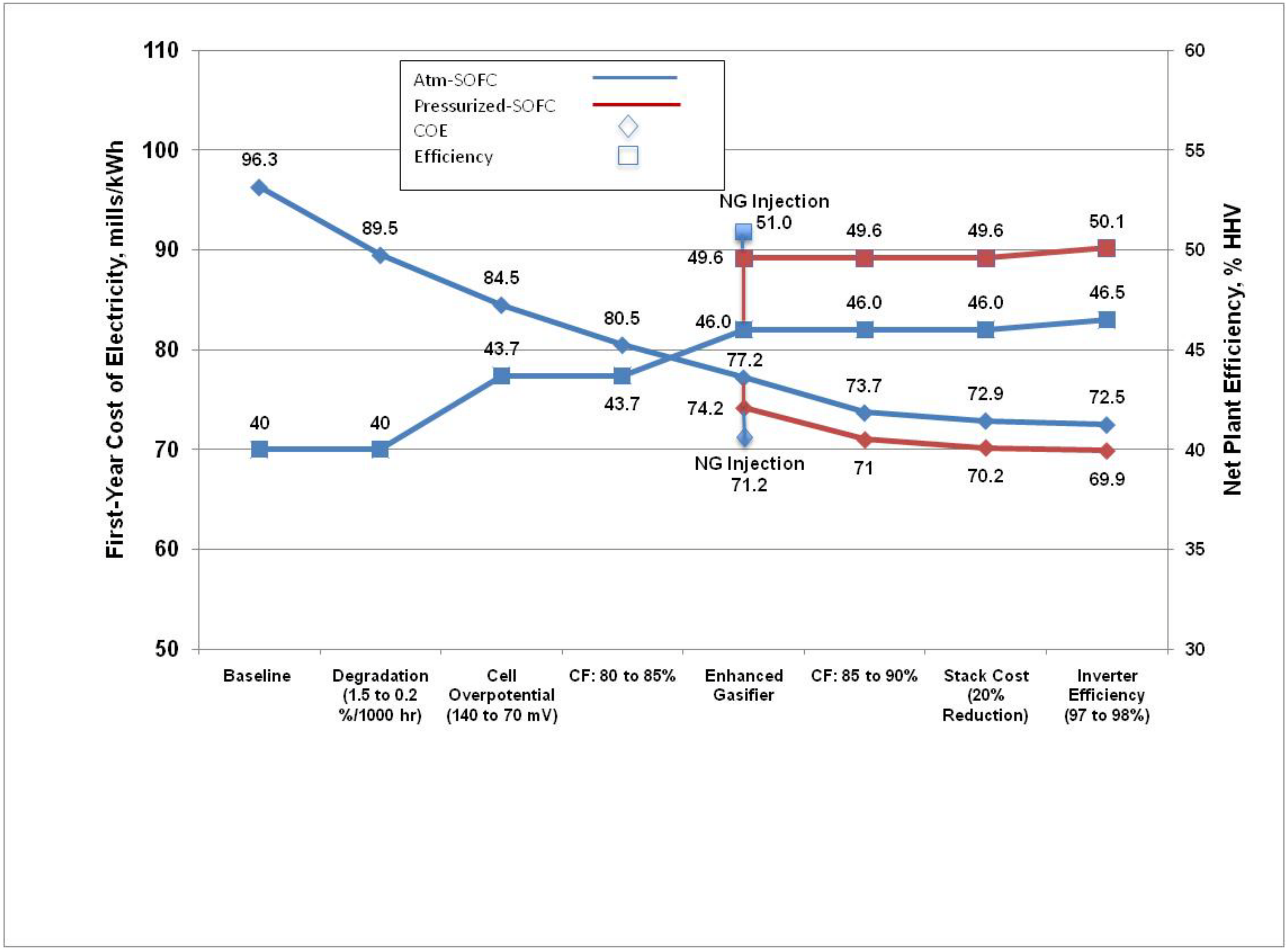


Exhibit ES-5 Catalytic Gasifier IGFC Pathway (Scenarios 3 and 4)

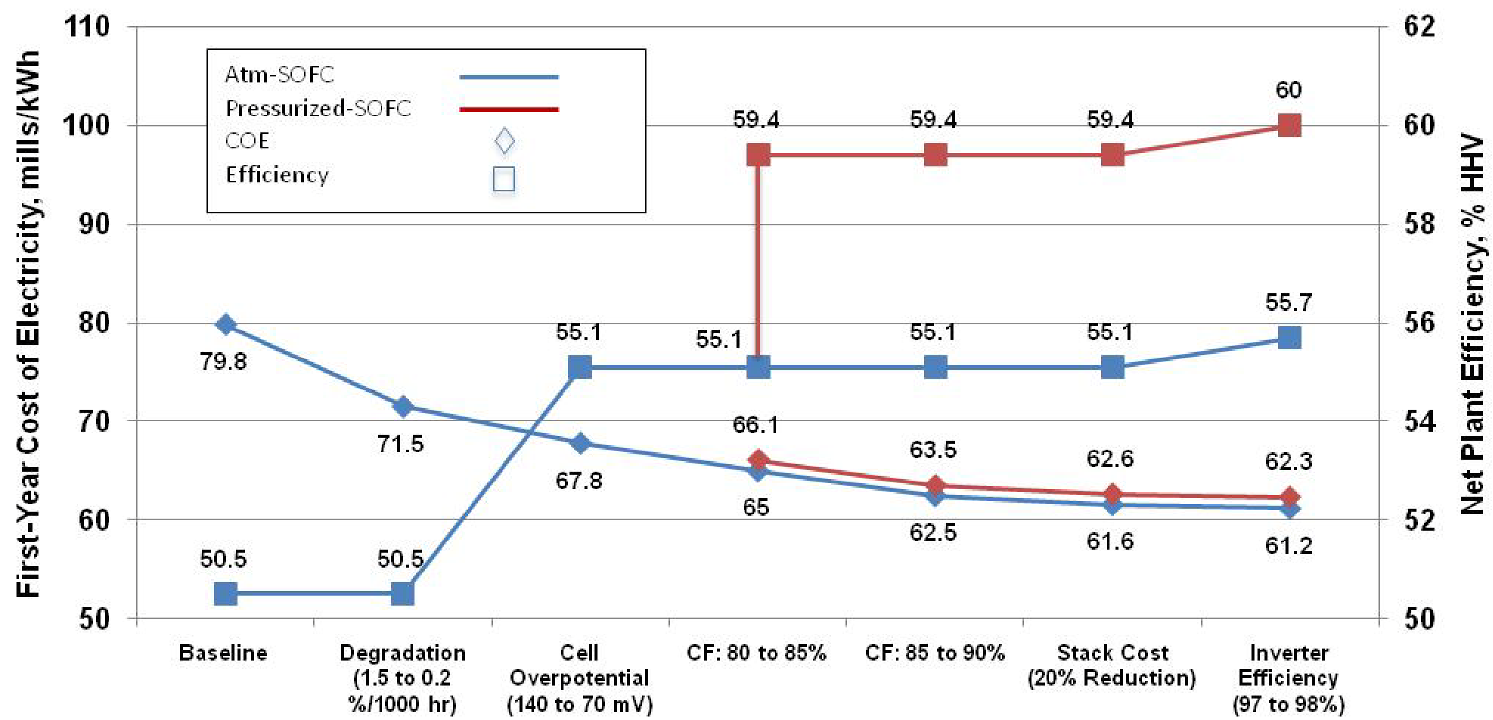


Exhibit ES-6 IGFC COE Comparison with Conventional Fossil Fuel Plants

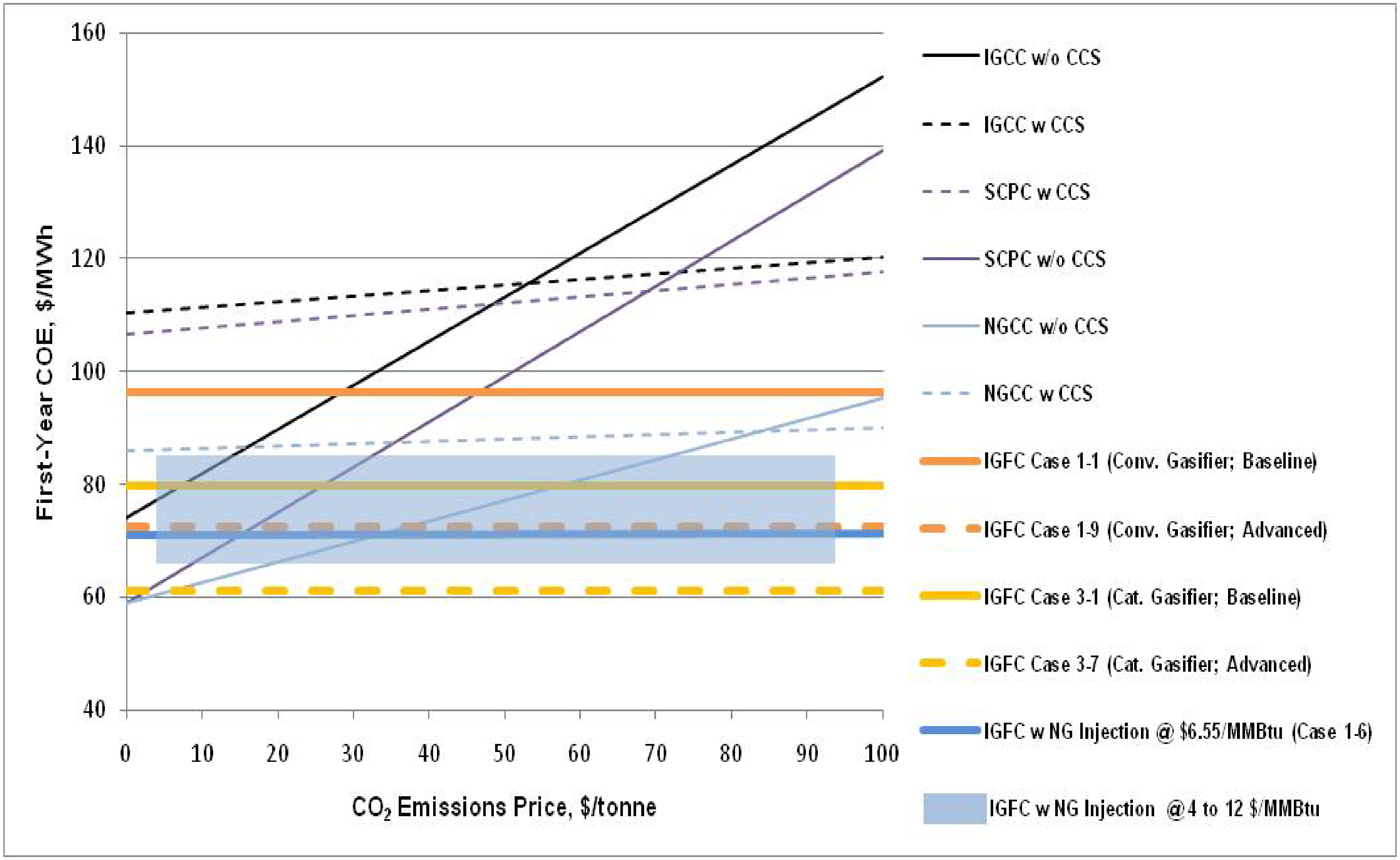


Exhibit ES-7 Cost of CO_2 Avoided for IGFC Pathways

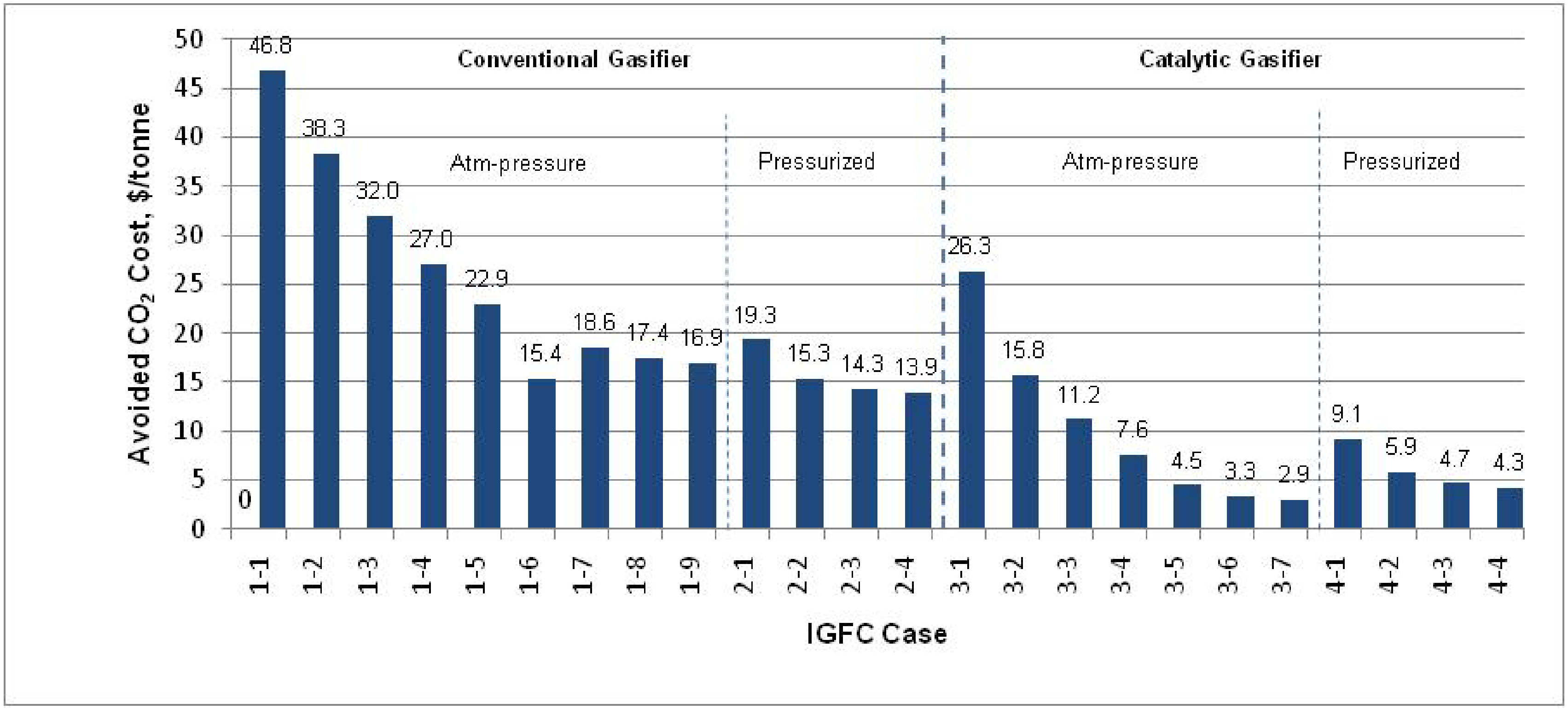


Exhibit ES-8 Conventional Gasifier, Atm-Pressure SOFC Pathway Results (Scenario 1)

Case	Pathway Parameter	Change Made	Coal Feed Rate, kg/h (lb/h)	Number Parallel Trains	Cell Voltage V	Plant Efficiency %, HHV	Raw Water Consumed gpm/MW	CO_2 Emission kg/MWh	Capital Cost, TOC $/kW	COE mills/kWh	Cost of CO_2 Avoided $/tonne
1-1	**Baseline Atm-pressure**	Baseline	182,264 (401,823)	2	0.816	40.0	3.07	2.5	3,001	96.3	46.8
1-2	**Degradation**	1.5 to 0.2 %/1000 hours	182,264 (401,823)	2	0.816	40.0	3.07	2.5	2,844	89.5	38.3
1-3	**Cell Over-potential**	140 to 70 mV	166,990 (368,151)	2	0.885	43.7	2.82	2.3	2,666	84.5	32.0
1-4	**Capacity Factor**	80 to 85 %	166,990 (368,151)	2	0.885	43.7	2.82	2.3	2,666	80.5	27.0
1-5	**Enhanced Gasifier**	5.9 to 10.2 mole% CH_4	158,481 (349.390)	2	0.878	46.0	2.74	2.2	2,552	77.2	22.9
1-6	**Natural Gas Injection**	38.5% injection	87,954 (193,905)	1	0.86	51.0	2.05	1.3	1,794	71.2 @ $6.55/MM Btu natural gas	15.4
1-7	**Capacity Factor**	85 to 90 %	158,481 (349.390)	2	0.878	46.0	2.74	2.2	2,552	73.7	18.6
1-8	**SOFC Stack Cost**	296 to 268 $/kW	158,481 (349.390)	2	0.878	46.0	2.74	2.2	2,512	72.9	17.4
1-9	**Inverter Efficiency**	97 to 98 %	156,885 (345,873)	2	0.878	46.0	2.71	2.2	2,497	72.5	16.9

Exhibit ES-9 Conventional Gasifier, Pressurized SOFC Pathway Results (Scenario 2)

Case	Pathway Parameter	Change Made	Coal Feed Rate kg/h (lb/h)	Number Parallel Trains	Cell Voltage V	Plant Efficiency %, HHV	Raw Water Consumed gpm/MW	CO_2 Emission kg/MWh	Capital Cost TOC $/kW	COE mills/kWh	Cost of CO_2 Avoided $/tonne
2-1	**Pressurized SOFC**	15.6 to 285 psia	146,735 (324,386)	1	0.937	49.6	2.20	5.7	2,436	74.2	19.3
2-2	**Capacity Factor**	85 to 90 %	146,735 (324,386)	1	0.937	49.6	2.20	5.7	2,436	71.0	15.3
2-3	**SOFC Stack Cost**	442 to 414 $/kW	146,735 (324,386)	1	0.937	49.6	2.20	5.7	2,397	70.2	14.3
2-4	**Inverter Efficiency**	97 to 98 5	145,671 (321,149)	1	0.937	50.1	2.18	5.7	2,384	69.9	13.9

Exhibit ES-10 Catalytic Gasifier, Atm-Pressure SOFC Pathway Results (Scenario 3)

Case	Pathway Description	Change Made	Coal Feed Rate kg/h (lb/h)	Number Parallel Trains	Cell Voltage V	Plant Efficiency %, HHV	Raw Water Consumed gpm/MW	CO_2 Emission kg/MWh	Capital Cost, TOC $/kW	COE mills/kWh	Cost of CO_2 Avoided $/tonne
3-1	**Baseline Atm-pressure**	Baseline	135,961 (299,744)	1	0.787	50.5	2.49	1.8	2,194	79.8	26.3
3-2	**Degradation**	1.5 to 0.2 %/1000 hours	135,961 (299,744)	1	0.787	50.5	2.49	1.8	2,043	71.5	15.8
3-3	**Cell Over-potential**	140 to 70 mV	124,495 (274,465)	1	0.852	55.1	2.26	1.6	1,918	67.8	11.2
3-4	**Capacity Factor**	80 to 85 %	124,495 (274,465)	1	0.852	55.1	2.26	1.6	1,918	65.0	7.6
3-5	**Capacity Factor**	85 to 90 %	124,495 (274,465)	1	0.852	55.1	2.26	1.6	1,918	62.5	4.5
3-6	**SOFC Stack Cost**	296 to 268 $/kW	124,495 (274,465)	1	0.852	55.1	2.26	1.6	1,877	61.6	3.3
3-7	**Inverter Efficiency**	97 to 98 %	123,199 (271,608)	1	0.852	55.7	1.88	1.6	1,866	61.2	2.9

Exhibit ES-11 Catalytic Gasifier, Pressurized SOFC Pathway Results (Scenario 4)

Case	Pathway Description	Change Made	Coal Feed Rate kg/h (lb/h)	Number Parallel Trains	Cell Voltage V	Plant Efficiency %, HHV	Raw Water Consumed gpm/MW	CO_2 Emission g/MWh	Capital Cost, TOC $/kW	COE mills/kWh	Cost of CO_2 Avoided $/tonne
4-1	**Pressurized SOFC**	15.6 to 285 psia	115,524 (254,687)	1	0.912	59.4	1.81	5.7	2,026	66.1	9.1
4-2	**Capacity Factor**	85 to 90 %	115,524 (254,687)	1	0.912	59.4	1.81	5.7	2,026	63.5	5.9
4-3	**SOFC Stack Cost**	442 to 414 $/kW	115,524 (254,687)	1	0.912	59.4	1.81	5.7	1,986	62.6	4.7
4-4	**Inverter Efficiency**	97 to 98 %	114,330 (252,055)	1	0.912	60.0	1.79	5.7	1,976	62.3	4.3

Exhibit ES-12 Conventional Fossil Fuel Power Generation Technology Performance and Cost [1]

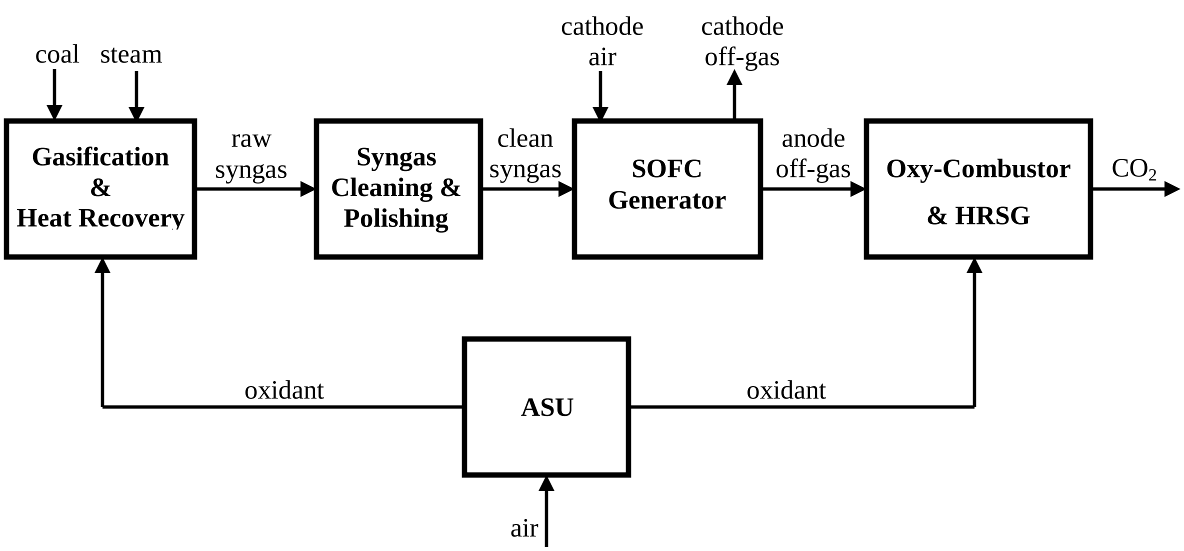
Technology	IGCC (CoP Gasifier)		PC (Supercritical)		NGCC	
CCS	No	Yes	No	Yes	No	Yes
Capacity Factor (%)	80	80	85	85	85	85
Capacity (kW)	625,060	513,610	550,020	550,000	550,080	473,570
Efficiency (%, HHV)	39.7	31.0	39.3	28.4	50.2	42.8
Raw Water Consumption (gpm/MW)	5.5	9.0	7.8	14.1	3.3	6.3
CO_2 Emission (kg/MWh)	776	98	803	111	365	43
TOC ($/kW)	2,351	3,952	2,024	3,570	718	1,497
COE (mills/kWh)	74.0	110.4	58.9	106.6	58.9	85.0
Cost of CO_2 avoided, $/tonne	NA	73.0	NA	68.9	NA	34.3

1. Introduction

This report presents the results of a Pathway Study for coal-based, integrated gasification fuel cell (IGFC) power systems with carbon capture and sequestration (CCS). The results quantify the performance and cost benefits for a series of projected gains made through the development of advanced technologies or improvements in plant operation and maintenance. The results represent the potential future benefits of IGFC technology development. They also provide DOE with a basis to select the most appropriate development path for IGFC, and to measure and prioritize the contribution of its R&D program to future power systems technology.

The IGFC power plant is analogous to an IGCC power plant, but with the gas turbine power island replaced with an SOFC power island (Exhibit 1-1). An inherent characteristic of SOFC is that the anode fuel gas is oxidized by the passage of cathode oxygen ions across the cell interface, resulting in a partially oxy-combusted SOFC anode off-gas. If this anode off-gas combustion is completed with a separate oxygen stream, the plant exhaust gas stream is essentially a CO_2 product ready for compression, dehydration, and sequestration. The only other exhaust gas stream in the plant is the cathode off-gas which is uncontaminated, vitiated air.

Exhibit 1-1 IGFC Plant Configuration with CCS



The IGFC plants in this study apply advanced, planar, solid oxide fuel cell (SOFC) technology with separate anode and cathode off-gas steams, and incorporate anode off-gas oxy-combustion for nearly complete carbon capture. The SOFC simulations utilize the expected operating conditions and performance capabilities of this solid oxide fuel cell technology, operating initially at atmospheric-pressure. The power plant cost and performance estimates reflect performance projections based on the current state of SOFC development, as well as projecting a pathway of SOFC technology development advances. The following fuel cell system advances are incorporated in a cumulative manner:

- Reduced SOFC stack performance degradation
- Reduced stack overpotential
- SOFC stack cost reduction
- Improved inverter efficiency
- Pressurized SOFC.

Advances in IGFC plant operation are also included in the pathway, being represented as improved plant availability and capacity factor achieved through advanced component monitoring, improved maintenance practices, and plant operation experience.

This document characterizes two parallel pathways of IGFC development, both incorporating CCS, and estimates overall plant performance and cost along these pathways in a consistent technical and economic manner. The first pathway applies conventional coal gasification technology, the ConocoPhillips E-Gas™ gasifier (CoP). This gasification technology produces syngas having limited methane content, roughly 6 mole percent. Increased syngas methane content is projected to benefit the performance of the IGFC plant. The first pathway consists of two scenarios. Scenario 1 looks at atmospheric-pressure SOFC and follows both SOFC technology advances and a near-term enhancement in the conventional gasifier technology to generate syngas having slightly higher methane content. The potential benefit of an additional, near-term technology enhancement step with conventional gasifier technology and atmospheric-pressure SOFC has also been explored in Case 1-6 as a branch-point to Scenario 1, considering the use of natural gas injection into the coal syngas as a means to achieve significantly higher syngas methane content. Scenario 2 considers the incorporation of pressurized-SOFC technology as a longer term enhancement, and represents an additional branch-point to Scenario 1.

The second pathway applies an advanced, catalytic coal gasification technology projected to produce syngas having very high methane content of roughly 30 mole percent, greatly improving the IGFC performance. This pathway follows similar advances in SOFC technology development as used for the pathway with the conventional gasifier.

Summaries of plant configurations and pathway parameters considered in this study are presented in Exhibit 1-2 and Exhibit 1-3. The Baseline plant utilizes SOFC operating conditions and performance capabilities based on the current status of sub-scale testing. Components for each plant configuration are described in more detail in the corresponding report sections for each pathway.

Exhibit 1-2 Conventional Gasifier IGFC Pathway Parameters (Scenarios 1 and 2)

Case	Pathway Parameter	Gasifier (methane %[1]	SOFC Pressure & Overpotential	Capacity Factor %	Cell Degradation %/1000 h	SOFC Stack Cost $/kW SOFC[2]	Inverter Efficiency (%)
1-1	**Baseline Atm-pressure SOFC**	CoP (6%)	15.6 psia 140 mV	80	1.5	296	97
1-2	Degradation	CoP (6%)	15.6 psia 140 mV	80	**0.2**	296	97
1-3	Overpotential	CoP (6%)	15.6 psia **70 mV**	80	0.2	296	97
1-4	Capacity Factor	CoP (6%)	15.6 psia 70 mV	**85**	0.2	296	97
1-5	Gasifier	**Enhanced (10%)**	15.6 psia 70mV	85	0.2	296	97
1-6	Natural Gas Injection	Enhanced (24.6%)	15.6 psia 70mV	85	0.2	296	97
1-7	Capacity Factor	Enhanced (10%)	15.6 psia 70 mV	**90**	0.2	296	97
1-8	SOFC cost reduction)	Enhanced (10%)	15.6 psia 70 mV	90	0.2	**268**	97
1-9	Inverter Efficiency	Enhanced (10%)	15.6 psia 70 mV	90	0.2	268	**98**
2-1	**Pressurized SOFC**	Enhanced (11%)	**285 psia** 70 mV	85	0.2	442	97
2-2	Capacity Factor	Enhanced (11%)	285 psia 70 mV	**90**	0.2	442	97
2-3	SOFC cost reduction)	Enhanced (11%)	285 psia 70 mV	90	0.2	**414**	97
2-4	Inverter Efficiency	Enhanced (11%)	285 psia 70 mV	90	0.2	414	**98**

1 – Methane content (mole percent) of clean, dry syngas
2 – Cost (TPC) of the SOFC stack unit (stacks, enclosures, inverters) in $ per kW of SOFC AC output
3 – Natural gas injected in the syngas as percent of the total fuel energy input

Exhibit 1-3 Catalytic Gasifier IGFC Pathway Parameters (Scenarios 3 and 4)

Case	Pathway Parameter	SOFC Pressure & Overpotential	Capacity Factor %	Cell Degradation %/1000 h	SOFC Stack Cost $/kW	Inverter Efficiency %
3-1	**Baseline Atm-pressure SOFC**	15.6 140 mV	80	1.5	296	97
3-2	Degradation	15.6 140 mV	80	**0.2**	296	97
3-3	Overpotential	15.6 **70 mV**	80	0.2	296	97
3-4	Capacity Factor	15.6 70 mV	**85**	0.2	296	97
3-5	Capacity Factor	15.6 70 mV	**90**	0.2	296	97
3-6	SOFC cost reduction	15.6 70 mV	90	0.2	**268**	97
3-7	Inverter Efficiency	15.6 70 mV	90	0.2	296	**98**
4-1	**Pressurized SOFC**	**285 psia** 70 mV	85	0.2	442	97
4-2	Increased Capacity Factor	285 psia 70mV	**90**	0.2	442	97
4-3	SOFC cost reduction	285 psia 70 mV	90	0.2	**414**	97
4-4	Inverter Efficiency	285 psia 70 mV	90	0.2	414	**98**

1 – Cost (TPC) of the SOFC stack unit (stacks, enclosures, inverters) in $ per kW of SOFC AC output

Scenario1 represents the pathway for a conventional gasifier plant generating syngas for use in an atmospheric-pressure SOFC power island. The Case 1-6 branch-point simulates an alternative approach for generating a high methane syngas without the need to develop advanced gasification technology, accomplishes this by injecting sufficient quantity of natural gas into the conventional gasifier syngas. It branches from Scenario 1 after Case 1-5. Scenario 2 represents the transition of Scenario 1 to a pressurized-SOFC power island configuration after several pathway enhancements in the Scenario 1 plant, branching from Scenario 1 after Case 1-5.

Scenario 3 applies an advanced catalytic gasifier for the production of a high methane syngas for use in an atmospheric-pressure SOFC power island. Scenario 4 transitions to a pressurized-

SOFC power island. It branches from Scenario 3 after Case 3-5. Components for each plant configuration are described in more detail in the corresponding report sections for each scenario.

The balance of this report is organized as follows:

- Section 2 provides the basis for the technical and cost evaluations.
- Section 3 reports the pathway results for IGFC using conventional coal gasification technology. It first describes the baseline, conventional coal gasifier-based IGFC plant with atmospheric-pressure SOFC, and presents the baseline plant performance and cost results, followed by a summary of its pathway performance and cost results (Scenario1). Section 3 then addresses the IGFC pathway branch that applies pressurized SOFC (Scenario 2), describing the plant and reporting its pathway performance and cost results.
- Section 4 is analogous to Section 3, but it describes the catalytic coal gasifier-based IGFC plant simulations and presents the results for the atmospheric-pressure SOFC (Scenario 3) and pressurized SOFC (Scenario 4) pathways.
- Section 5 provides process description, performance results, and cost results for Case 1-6, the conventional gasifier-based, atmospheric-pressure SOFC plant with natural gas injection into the clean syngas since this is a unique case.
- Section 6 provides the reference list.

2. Pathway Study Basis

For each of the plant configurations in this study a ChemCad process simulator (commercial process simulator by ChemStations, Houston, TX) model was developed and used to generate material and energy balances, which in turn were used as the design basis for the major equipment items. The major equipment characterizations were used to generate capital and operating cost estimates for the IGFC plants. Performance and process limits were based upon published reports, information obtained from vendors and users of the technology, performance data from design/build utility projects, and/or best engineering judgment as described in the Bituminous Baseline report.

Capital and operating costs for most of the conventional equipment items were scaled from estimates made in the Bituminous Baseline report. A current-dollar, first-year cost of electricity (COE) was calculated for each of the cases and is reported as the revenue requirement figure-of-merit.

The balance of this chapter documents the design basis for the pathway study, as well as environmental targets and cost assumptions used in the study. This basis was largely a duplicate of the design basis applied in the Bituminous Baseline report, and any changes in that basis are noted in this section.

2.1 Site Characteristics

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 2-1 and Exhibit 2-2. The ambient conditions are the same as ISO conditions.

Exhibit 2-1 Site Ambient Conditions

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

Exhibit 2-2 Site Characteristics

Location	Greenfield, Midwestern USA
Topography	Level
Size, acres	150 (IGFC)
Transportation	Rail
Ash/Slag Disposal	Off Site
Water	Municipal (50%) / Groundwater (50%)
Access	Land locked, having access by train and highway
CO_2 Storage	Compressed to 15.3 MPa (2,215 psia), transported 80 kilometers (50 miles) and sequestered in a saline formation at a depth of 1,239 meters (4,055 feet)

The land area for IGFC cases assumes 15 acres are required for the plant proper and the balance provides a buffer of approximately 0.25 miles to the fence line. The extra land could also provide for a rail loop if required. In all cases it was assumed that the steam turbine is enclosed in a turbine building. The gasifier and the SOFC stack units are not enclosed in buildings.

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

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

2.2 Coal Characteristics

The design coal is Illinois No. 6 with characteristics presented in Exhibit 2-3. The coal properties are from NETL's Coal Quality Guidelines [2].

The first year cost of coal used in this study is $1.55/MMkJ ($1.64/MMBtu). This cost was determined and applied in the Bituminous Baseline report.

Exhibit 2-3 Design Coal

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

2.3 Natural Gas Characteristics

Natural gas is utilized as secondary fuel in one case (Section 5), and its composition is presented in Exhibit 2-4 [3].

Exhibit 2-4 Natural Gas Composition

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

Note: Fuel composition is normalized and heating values are calculated

The first year cost of natural gas used in this study is \$ 6.21/MMkJ (\$6.55/MMBtu). This cost was determined and applied in the Bituminous Baseline report.

2.4 Environmental Targets

Typical environmental targets for fossil-based power plants were identified in the Bituminous Baseline report, and it is projected that the emissions from IGFC plants will be inherently much lower than these typical targets. In setting the environmental targets a number of factors were considered, including current emission regulations, regulation trends, results from recent permitting activities and the status of current best available control technology (BACT).

The IGFC plant emissions are projected to be very limited because (1) the total sulfur content in the clean syngas must be maintained at less than 100 ppbv to protect the critical fuel cell materials; (2) 95 percent of the mercury and other trace components are removed from the syngas; (3) the oxy-combustor is a low NO_x producing combustor; and (4) all of the coal syngas contaminant species remaining after syngas cleaning are sequestered with the CO_2 product. Thus, the IGFC plant has nearly 100 percent removal of all environmental contaminants, including CO_2. Only the plant solid waste streams, the coal ash, and spent sorbents and waste catalysts might be of environmental concern.

The coal mercury content for this study was assumed to be 0.15 ppm (dry), and this is consistent with the mercury content estimated and applied in the Bituminous Baseline report. It was further assumed that all of the coal Hg enters the gas phase and none leaves with the bottom ash or slag.

The IGCC environmental targets were chosen for IGFC plants to match the Electric Power Research Institute's (EPRI) design basis for their CoalFleet for Tomorrow Initiative and are shown in Exhibit 2-5[4]. EPRI notes that these are design targets and are not to be used for permitting values.

Exhibit 2-5 Environmental Targets for IGFC Cases

Pollutant	Environmental Target	NSPS Limit	Control Technology
NOx	15 ppmv (dry) @ 15% O_2	1.0 lb/MWh	Low NOx oxy-combustors
SO_2	0.0128 lb/MMBtu	1.4 lb/MWh	Selexol and ZnO-polishing
Particulate Matter (Filterable)	0.0071 lb/MMBtu	0.015 lb/MMBtu	Cyclones and candle filters
Mercury	> 90% capture	20×10^{-6} lb/MWh	Carbon bed

To achieve an environmental target for SO_2 of 0.0128 lb/MMBtu requires approximately 28 ppmv sulfur in the sweet syngas. The acid gas removal (AGR) process must have a sulfur capture efficiency of about 99.7 percent to reach the environmental target. Since the syngas sulfur content for SOFC application is estimated to be no more than 100 ppbv, additional sulfur polishing is requires.

Most of the coal ash is removed from the gasifier as slag or bottom ash. The ash that remains entrained in the syngas is captured in the downstream equipment, including the syngas scrubber and a cyclone and either ceramic or metallic candle filters. The environmental target of 0.0071 lb/MMBtu filterable particulates can be achieved with these particulate control devices.

The environmental target for mercury capture is greater than 90 percent removal. Based on experience at the Eastman Chemical plant, where syngas from a GEE gasifier is treated, the actual mercury removal efficiency used is 95 percent. Sulfur-impregnated activated carbon is used by Eastman as the adsorbent in the packed beds operated at 30°C (86°F) and 6.2 MPa (900 psig). Mercury removal between 90 and 95 percent has been reported with a bed life of 18 to 24 months. Removal efficiencies may be even higher, but at 95 percent the measurement precision limit was reached. Eastman has yet to experience any mercury contamination in its product [5]. Mercury removals of greater than 99 percent can be achieved by the use of dual beds, i.e., two beds in series. However, this study assumes that the use of sulfur-impregnated carbon in a single carbon bed achieves 95 percent reduction of mercury emissions which meets the environmental target and NSPS limits in all cases. In addition, the carbon beds are assumed to effectively remove other trace metals that are of concern to the SOFC.

Carbon dioxide (CO_2) is not currently regulated nationally. However, the possibility exists that federal carbon limits will be imposed in the future and this study examines cases that include nearly complete elimination of CO_2 emissions. The highest level of CO_2 achievable using the IGFG technology is the goal of this evaluation, and this level is near 100 percent.

2.5 Balance of Plant

The balance of plant assumptions are common to all cases and are presented in Exhibit 2-6.

Exhibit 2-6 Balance of Plant Assumptions

Cooling system	Recirculating Wet Cooling Tower
Fuel and Other storage Coal	30 days
Slag/ash	30 days
Sulfur	30 days
Sorbent/catalyst	30 days
Plant Distribution Voltage Motors below 1 hp	110/220 volt
Motors between 1 hp and 250 hp	480 volt
Motors between 250 hp and 5,000 hp	4,160 volt
Motors above 5,000 hp	13,800 volt
Steam and Gas Turbine Generators	24,000 volt
Grid Interconnection Voltage	345 kV
Water and Waste Water Makeup Water	The water supply is 50 percent from a local Publicly Owned Treatment Works and 50 percent from groundwater, and is assumed to be in sufficient quantities to meet plant makeup requirements. Makeup for potable, process, and de-ionized (DI) water is drawn from municipal sources
Process Wastewater	Water associated with gasification activity and storm water that contacts equipment surfaces is collected and treated for discharge through a permitted discharge.
Sanitary Waste Disposal	Design includes a packaged domestic sewage treatment plant with effluent discharged to the industrial wastewater treatment system. Sludge is hauled off site. Packaged plant was sized for 5.68 cubic meters per day (1,500 gallons per day)
Water Discharge	Most of the process wastewater is recycled to the cooling tower basin. Blowdown is treated for chloride and metals, and discharged.

2.6 Plant Capacity

The IGFC plant's net generating capacity is fixed at 550 MW in this pathway study. The coal feed rate varies over a broad range from 114,330 to 182,264 kg/h (252,055 to 401,823 lb/h, as-received) over all of the IGFC cases evaluated.

The study case which injects natural gas into the coal-derived syngas (Case 1-6), maintains a coal feed rate at only 87,954 kg/h (193,906 lb/h, as-received), injecting natural gas at 38.5 percent of the total plant fuel energy input, and resulting in a plant net generating capacity of 550 MW.

2.7 Sparing Philosophy and Number of Parallel Process Trains

There is no redundancy provided in the case evaluations, other than normal sparing of rotating equipment. Spare SOFC cells are provided with on-line switching capability to control cell degradation effects and maintain nearly constant SOFC power output [6].

The number of parallel processing trains utilized in the IGFC plant depends on the flow capacities for each case. The number of parallel trains used in the pathway study are taken to be comparable to the design basis applied for IGCC in the Bituminous Baseline report: Single ASU maximum oxidant rate of 113,400 kg/h (250,000 lb/h), single gasification and syngas cooling train maximum coal feed rate of 249,500 kg/h (325,000 lb/h), single conventional syngas cleaning train maximum syngas flow rate of 147,400 kg/h (550,000 lb/h), and single CO_2 compression train maximum CO_2 stream rate of 136,100 kg/h (300,000 lb/h).

With this basis, the Scenario 1 plants consist of the following major subsystems:

- Two parallel air separation units (2 x 100 percent)
- Two train gasification section, including gasifier, synthesis gas cooler, quench and scrubber (2 x 100 percent).
- Two parallel train syngas clean-up section (2 x 100 percent).
- Two parallel trains Selexol acid gas removal (2 x 100 percent), and two Claus-based sulfur recovery units (1 x 100 percent).
- Two oxy-combustor/HRSG trains (2 x 100 percent).
- One steam turbine system (1 x 100 percent).
- Four parallel CO_2 compression trains (4 x 100 percent)

The other cases in Scenarios 2, 3 and 4, use single processing trains, these having sufficiently small coal, oxidant, syngas, and CO_2 product flow capacities to operate with single processing trains and two CO_2 compression trains.

2.8 SOFC Power Island Characterization

Several assumptions were applied to estimate the performance of the IGFC power island components.

Estimation of SOFC Operating Voltage

The SOFC operating voltage has a large impact on the total plant performance and cost. An experimental basis or detailed modeling basis for estimating the operating voltage has not yet been established. For the pathway study cases, the SOFC cell operating potential has been estimated based on the evaluation of representative stack test data, using the difference between the anode inlet Nernst potential and a calibration over-potential to determine the operating potential. Thus, the operating voltage, V, is estimated as

$$V = E - OP$$

where E is the stack anode-inlet Nernst potential as calculated from the anode gas composition, and OP is the calibration overpotential value. The Nernst potential is a function of the anode gas molar ratio of hydrogen to water vapor, the cathode gas oxygen mole fraction, the temperature, and the pressure [7]. This procedure provides operating voltages that are comparable to SOFC vendor test results with comparable conditions and fuel gas composition.

SOFC Carbon Deposition Control

The cell stack inlet anode gas composition can induce the formation of solid carbon deposits, which can disrupt the normal performance of the stack. A criterion is applied in all of the cases to represent anode gas inlet conditions where carbon deposition should not occur, and it is demanded in the simulations that this criterion be satisfied. The criterion for carbon deposit-free behavior is

$$Oxy / Carb > 2.0$$

where Oxy is the inlet anode gas total molar atomic oxygen content (with the main species being CO, CO_2, and H_2O), and Carb is the inlet anode gas total molar atomic carbon content (with the main species being CH_4, CO, and CO_2). Anode gas recirculation using hot gas blowers or syngas jet pumps maintains the inlet anode gas composition in a safe range by recirculating water vapor.

Estimation of Steam Bottoming Cycle Performance

The anode off-gas stream is combusted with oxygen, providing a hot gas stream that passes through a heat recovery steam generation system that produces high-pressure process steam, low-pressure process steam, and high-pressure steam for power generation in a steam bottoming cycle. The steam bottoming cycle is a subcritical steam cycle that varies greatly in its steam conditions and capacity in the study cases, providing a relatively small proportion of the total plant generation output. In some cases the heat recovery temperature available is relatively low and results in poor steam superheat conditions. Rather than perform detailed simulation each of these unique steam bottoming cycles, a correlation method was applied that relates the steam bottoming cycle efficiency to the flue gas temperature available for steam generation [8].

For steam cycles limited to subcritical conditions, the correlation for the bottoming cycle efficiency is

$$\text{Efficiency (percent of heat absorbed)} = -0.000048223\ T^2 + 0.100981\ T - 5.747913$$

where T is the heat recovery inlet gas temperature (°C). For inlet temperatures greater than 648 °C, the efficiency is limited to 39.45 percent of the heat absorbed.

2.9 Capacity Factor

The capacity factor for the IGFC baseline plant is assumed to be 80 percent, identical to that of baseline IGCC used in the Bituminous Baseline report. The plant processing sections are designed for 100 percent capacity, with no excess capacity provided for any component other than the SOFC stack. The SOFC stack is designed with excess cell capacity that can be activated during operation to maintain the fuel cell output nearly constant in response to cell performance degradation.

This study assumes that each new plant would be dispatched any time it is available and would be capable of generating maximum capacity when online. Therefore the capacity factor and plant availability are equal.

NERC defines an equivalent availability factor (EAF), which is essentially a measure of the plant capacity factor assuming there is always a demand for the output. The EAF accounts for planned and scheduled derated hours as well as seasonal derated hours. As such, the EAF matches this study's definition of capacity factor.

EPRI examined the historical forced and scheduled outage times for IGCCs and concluded that the reliability factor (which looks at forced or unscheduled outage time only) for a single train IGCC (no spares) would be about 90 percent [9]. To get the availability factor, one has to deduct the scheduled outage time. In reality the scheduled outage time differs from gasifier technology-to-gasifier technology, but the differences are relatively small and would have minimal impact on the capacity factor, so for this study it was assumed to be constant at a 30-day planned outage per year (or two 15-day outages). The planned outage would amount to 8.2 percent of the year, so the availability factor would be (90 percent - 8.2 percent), or 81.2 percent.

There are four operating IGCC's worldwide that use a solid feedstock and are primarily power producers (Polk, Wabash, Buggenum and Puertollano). A 2006 report by Higman et al. examined the reliability of these IGCC power generation units and concluded that typical annual on-stream times are around 80 percent.[10] The capacity factor would be somewhat less than the on-stream time since most plants operate at less than full load for some portion of the operating year. Given the results of the EPRI study and the Higman paper, a capacity factor of 80 percent was chosen for IGFC with no spare gasifier required.

2.10 Raw Water Withdrawal and Consumption

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

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

Boiler feedwater blowdown and a portion of the sour water stripper blowdown were assumed to be treated and recycled to the cooling tower. The cooling tower blowdown and the balance of the SWS blowdown streams were assumed to be treated and 90 percent returned to the water source with the balance sent to the ash ponds for evaporation.

The largest consumer of raw water in all cases is cooling tower makeup. It was assumed that all cases utilized a mechanical draft, evaporative cooling tower, and all process blowdown streams were assumed to be treated and recycled to the cooling tower.

Exhibit 2-1 was used to achieve a cooling water temperature of 16°C (60°F) using an approach of 5°C (8.5°F). The cooling water range was assumed to be 11°C (20°F). The cooling tower makeup rate was determined using the following [11]:

- Evaporative losses of 0.8 percent of the circulating water flow rate per 10°F of range
- Drift losses of 0.001 percent of the circulating water flow rate
- Blowdown losses were calculated as follows:
    - Blowdown Losses = Evaporative Losses / (Cycles of Concentration - 1)

      Where cycles of concentration is a measure of water quality, and a mid-range value of 4 was chosen for this study.

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

2.11 Cost Estimating Methodology

Following the basis used in the Bituminous Baseline report, the capital costs at the Total Overnight Cost (TOC) level include equipment, materials, labor, indirect construction costs, engineering, owner's costs, and contingencies. Where applicable, the cost of major conventional plant sections in the study case plants are based on scaled estimates from costs presented in the Bituminous Baseline report, applying the general cost-scaling equation:

$$C = N * (C_{ref} / N_{ref}) * [(F / N) / (F_{ref} / N_{ref})]^S$$

where C is the cost of the study case plant section, N is the number of parallel sections in the study case plant,

C_{ref} is the cost of the reference plant section,

N_{ref} is the number of parallel sections in the reference plant,

F is the capacity of the study case plant section,

F_{ref} is the capacity for the reference plant section, and

S is the scaling factor characteristic of the plant section equipment (a fraction usually between 0.5 and 0.8).

The Total Plant Cost (TPC) and Operation and Maintenance (O&M) costs for each of the cases in the study were estimated using data generated by WorleyParsons Group Inc. (WorleyParsons) in the Bituminous Baseline report. The Bituminous Baseline report estimates carry an accuracy of ±30 percent, consistent with the screening study level of information available for the various study power technologies.

All capital costs are presented as "overnight costs" expressed in June 2007 dollars. A first year of operation of 2015 is assumed for all cases.

Capital costs at the Total Plant Cost (TPC) level includes:

- Equipment (complete with initial chemical and catalyst loadings),
- Materials,
- Labor (direct and indirect),
- Engineering and construction management, and
- Contingencies (process and project).

Owner's costs are subsequently calculated and added to the TPC, the result of which is Total Overnight Cost (TOC). Additionally, financing costs are estimated and added to TOC to provide Total As-Spent Cost (TASC). The current-dollar, first-year cost of electricity is calculated using TOC.

Plant Maturity

The case estimates provided include technologies at different commercial maturity levels, and the overall IGFC plants represent very advanced, immature technologies. The commercial components in the IGFC plants are based on data from commercial IGCC offerings, however, there have been very limited sales of these units so far.

The SOFC and oxy-combustion technologies for the IGFC cases are very immature. This technology is unproven at commercial scale in power generation applications. The developing SOFC technology performance and cost has been estimated through scaling to commercial levels by the SOFC developers. While commercial pre-combustion CO_2 removal technology could be applied in place of the oxy-combustion based CO_2 removal, the advantages of oxy-combustion approach over pre-combustion CO_2 removal are so large that the oxy-combustion technology merits additional development.

The catalytic gasification technology is based on prior extensive development work conducted for a similar coal gasification technology by Exxon in the 1970s for the purpose of SNG production. The specific catalytic gasifier simulated for application to IGFC has not been tested and represents a conceptual processing step in the pathway evaluation.

Estimate Scope

The estimates represent a complete power plant facility on a generic site. Site-specific considerations such as unusual soil conditions, special seismic zone requirements, or unique local conditions such as accessibility, local regulatory requirements, etc. are not considered in the estimates.

The estimate boundary limit is defined as the total plant facility within the “fence line” including coal receiving and water supply system, but terminating at the high voltage side of the main power transformers. The single exception to the fence line limit is in the CO_2 capture cases where costs are included for TS&M of the CO_2.

Capital Costs

WorleyParsons developed the capital cost estimates for IGCC plants in the Bituminous Baseline report using the company’s in-house database and conceptual estimating models for each of the specific technologies. A reference bottoms-up estimate for each major component provides the basis for the estimating models. This provides a basis for subsequent comparisons and easy modification when comparing between specific case-by-case variations.

Some equipment costs for the cases were calibrated to reflect recent quotations and/or purchase orders for other ongoing in-house power or process projects. These include, but are not limited to the following equipment:

- Steam Turbine Generators
- Circulating Water Pumps and Drivers
- Cooling Towers
- Condensers
- Air Separation Units (partial)
- Main Transformers

Other key estimate considerations include the following:

- Labor costs are based on Midwest, Merit Shop. Costs would need to be re-evaluated for projects at different locations or for projects employing union labor.
- The estimates are based on a competitive bidding environment, with adequate skilled craft labor available locally.
- Labor is based on a 50-hour work-week (5-10s). No additional incentives such as per-diems or bonuses have been included to attract craft labor.
- While not included at this time, labor incentives may ultimately be required to attract and retain skilled labor depending on the amount of competing work in the region, and the availability of skilled craft in the area at the time the projects proceed to construction. Current indications are that regional craft shortages are likely over the next several years. The types and amounts of incentives will vary based on project location and timing

relative to other work. The cost impact resulting from an inadequate local work force can be significant.

- The estimates are based on a greenfield site.
- The site is considered to be Seismic Zone 1, relatively level, and free from hazardous materials, archeological artifacts, or excessive rock. Soil conditions are considered adequate for spread footing foundations. The soil bearing capability is assumed adequate such that piling is not needed to support the foundation loads.
- Costs are limited to within the “fence line,” terminating at the high voltage side of the main power transformers with the exception of costs included for TS&M of CO_2 in all capture cases.
- Engineering and Construction Management were estimated as a percent of bare erected cost; 10 percent for IGCC and PC technologies, and 9 percent for NGCC technologies. These costs consist of all home office engineering and procurement services as well as field construction management costs. Site staffing generally includes a construction manager, resident engineer, scheduler, and personnel for project controls, document control, materials management, site safety and field inspection.
- All capital costs are presented as “Overnight Costs” in June 2007 dollars. Escalation to period-of-performance is specifically excluded.

The current-dollar, first-year COE was calculated for each case using economic parameters for high-risk technologies resulting in a capital charge factor of 0.1773. The capital component of COE was calculated using TOC.

SOFC Stack Unit Cost Estimation

The rationale used to estimate the cost of the SOFC power island for both atmospheric-pressure SOFC and pressurized SOFC applications is described here. The cost basis for the key SOFC stack unit (the cell blocks arranged as stack modules, their enclosures, and the DC-AC inverters) is proposed. The major basis for the estimates made here are a DOE 2010 SOFC cost goal, and cost estimates generated by Fuel Cell Energy, Inc. (FCE) [12].

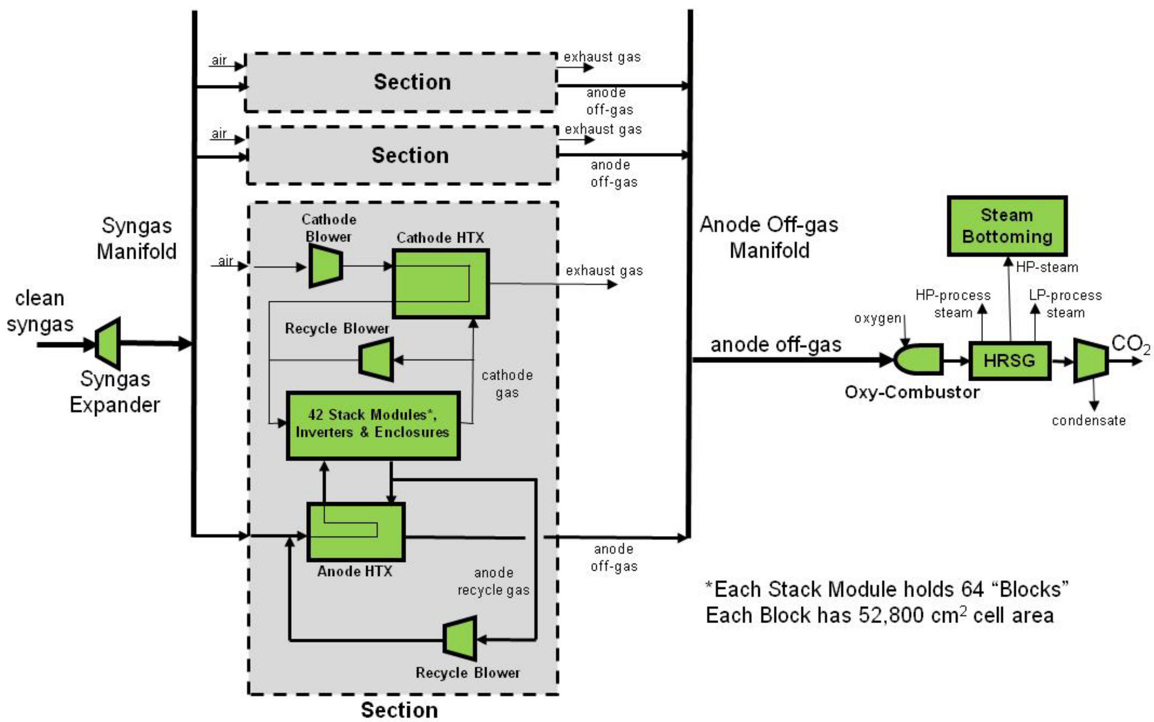
Exhibit 2-7 illustrates a generic, planar technology IGFC power island configuration. The IGFC power island consists of an array of factory assembled SOFC Sections, a syngas expander, an oxy-combustor, and steam bottoming components that are separately shipped and installed with the SOFC Sections at the plant site. Each SOFC Section consist of an array of stack modules, with the anode and cathode blowers and heat exchangers being factory assembled and shipped as complete, integrated units to the power plant. Each SOFC stack unit contains (1) SOFC “blocks” arranged as stack modules, (2) an enclosure for each stack module, and (3) a DC-AC inverter for each stack module.

A basic “cell” has an area of 550 cm^2, and a block contains 96 cells, or 52,800 cm^2 of active cell area. Each stack module holds 64 blocks, and each section holds 42 stack modules.

The DOE 2010 SOFC cost goal for the factory–assembled, atmospheric-pressure SOFC blocks and enclosures is 175 $ per kW of plant net power, in June 2007 dollars. This cost is interpreted as the factory-assembled cost, not including transportation to the site, and labor and materials for the site foundation, and for placing the equipment at the site. The other section components (blowers and heat exchangers) are separately estimated as factory-installed items. The other power island components (syngas expander, oxy-combustor, and steam bottoming components) are estimated as separately shipped components installed at the plant site.

It is assumed that the atmospheric-pressure SOFC stack unit has a power density of 400 mW/cm^2 for all of the plant conditions simulated. This is valid since the temperature, fuel utilization, and syngas composition vary over a limited range generally selected for high levels of performance. The SOFC blocks and enclosure cost must be converted to units of dollars per kW of SOFC power, rather than dollars per kW of net plant power, in order to be able to use the cost for general plant cost estimation. The ratio of the net plant power to the SOFC power ranges from approximately 0.94 to 1.0 for prior plant simulations performed, and a value of 0.945 from a base plant configuration is applied here to produce a cost for the integrated blocks and enclosures of 165 $/SOFC kW, in June 2007 dollars.

Exhibit 2-7 SOFC Power Island Configuration Showing Section Components



From FCE estimates, the separate enclosures cost is about 25 $/SOFC kW, and thus the integrated blocks cost about 140 $/SOFC kW. With the power density being 400 mW/cm^2, and assuming an inverter efficiency of 97 percent, the integrated blocks cost per cm^2 of active surface area is 140/1,000 * 400/1,000 * 0.97 = 0.054 $/cm^2 active surface area. This value is used to estimate the pressurized SOFC stack unit cost.

The inverters' cost is estimated from FCE information as 82 $/SOFC kW using NIST SiC inverter technology [13]. This advanced technology is considerably cheaper than the more conventional Satcon technology.

The total cost of the atmospheric-pressure, integrated SOFC stack unit (blocks, enclosures, inverters) is 165 + 82 = 247 $/SOFC kW. To this is also added the rough estimate for the cost of transport and placement of the sections (12 $/SOFC kW) and the cost for the section foundations at the site (37 $/SOFC kW), for a total installed cost of 296 $/kW of SOFC AC generation. This represents the SOFC stack unit cost on a total plant cost basis, and no contingencies have been applied to this since the cost estimate is taken from a DOE cost goal and vendor estimates having their own contingencies applied.

A similar configuration is assumed to apply for pressurized SOFC, where the enclosures now require pressure capability to a 300 psia design pressure. It is assumed that the pressurized cells have a fixed power density of 500 mW/cm^2, increased from 400 mW/cm^2 by the enhanced performance resulting from pressurization.

The integrated block cost will then be 0.054 / (500 * 0.97) * 1x10^6 = 111 $/SOFC kW, based on the atmospheric blocks cost of 0.054 $/cm^2. The enclosure cost is estimated to be a factor of 10 higher than the atmospheric-pressure enclosure cost to house the modules having dimensions of roughly 10-ft width by 15-ft length by 10-ft height. This makes the enclosure cost 25 * 400/500 * 10 = 200 $/SOFC kW.

With the inverter cost being the same as in the atmospheric-pressure application, the total cost of the pressurized, integrated SOFC stack unit (blocks, enclosures, inverters) is 111 + 200 + 82 = 392 $/SOFC kW. With transportation, placement, and foundations, the total cost is 442 $/kW of SOFC AC generation.

Exclusions

The capital cost estimate includes all anticipated costs for equipment and materials, installation labor, professional services (Engineering and Construction Management), contingency, and owner's costs. The following items are excluded from the capital costs:

- Site specific considerations – including but not limited to seismic zone, accessibility, local regulatory requirements, excessive rock, piles, laydown space, etc.
- Labor incentives in excess of a 5-day/10-hour work week
- Additional premiums associated with an EPC contracting approach

Contingency

Both the project contingency and process contingency costs represent costs that are expected to be spent in the development and execution of the project that are not yet fully reflected in the design. It is industry practice to include project contingency in the total plant cost (TPC) to cover project uncertainty and the cost of any additional equipment that would result during detailed design. Likewise, the estimates include process contingency to cover the cost of any additional equipment that would be required as a result of continued technology development, and the project and process contingencies applied are consistent with those used in the Bituminous Baseline study.

Based on the AACE International contingency guidelines as presented in NETL's "Quality Guidelines for Energy System Studies" it would appear that the overall project contingencies for the subject cases should be in the range of 30 to 40 percent.[3] However, such contingencies are believed to be too high when the basis for the cost numbers is considered. The costs have been extrapolated from an extensive data base of project costs (estimated, quoted, and actual), based on both conceptual and detailed designs for the various technologies. This information has been used to calibrate the costs in the current studies, thus improving the quality of the overall estimates. As such, the overall project contingencies should be more in the range of 15 to 20 percent based on the specific technology.

No project contingency has been applied to the SOFC stack unit cost, these contingencies already being incorporated by vendor estimates for the SOFC stack unit. A 15 percent project contingency has been applied to the ancillary components in the SOFC power island.

Process contingency is intended to compensate for uncertainties arising as a result of the state of technology development. No process contingency was placed on the SOFC stack unit cost, with the IGFC plant cost sensitivity to variations in the SOFC stack unit cost to be separately examined. Process contingencies have been applied to the estimates as follows:

- Slurry Prep and Feed – 5 percent on CoP IGFC cases - systems are operating at a high as 800 psia as compared to 600 psia in IGCC experience
- Gasifiers and Syngas Coolers – 15 percent on all IGFC cases – next-generation commercial offering and integration with the power island
- Trace Element Removal – 5 percent – minimal commercial scale experience in IGCC applications
- SOFC power island ancillary components – 15 percent.

Operations and Maintenance (O&M)

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

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

There are two components of O&M costs; fixed O&M, which is independent of power generation, and variable O&M, which is proportional to power generation. The approach followed in estimating these costs are consistent with that applied in the Bituminous Baseline report.

Operating Labor

Operating labor cost was determined based on of the number of operators required for each specific case. The average base labor rate used to determine annual cost is $34.65/h. The associated labor burden is estimated at 30 percent of the base labor rate.

Maintenance Material and Labor

Maintenance cost was evaluated on the basis of relationships of maintenance cost to initial capital cost. This represents a weighted analysis in which the individual cost relationships were considered for each major plant component or section. The exception to this is the maintenance cost for the combustion turbines, which is calculated as a function of operating hours.

The gasifier maintenance factors used for this study are as follows: CoP and Catalytic – 7.5 percent on the gasifier and related components, and 4.5 percent on the syngas cooling.

Another significant production cost is associated with cell performance degradation. Test data indicate that the cell performance degrades at less than 1 percent per 1,000 hours and levels as low as 0.05 percent per 1,000 hours can be considered [13]. The SOFC cells will operate with constant cell voltage and with decreasing cell current, resulting in degraded plant power generation with time. Spare cell capacity in the form of blocks and enclosures must be incorporated into the SOFC system design to be “switched on” at regular periods (1,000-hour intervals assumed) to increase the operating cell surface. This will maintain a near-constant plant power output from the SOFC cells to avoid total power plant performance degradation.

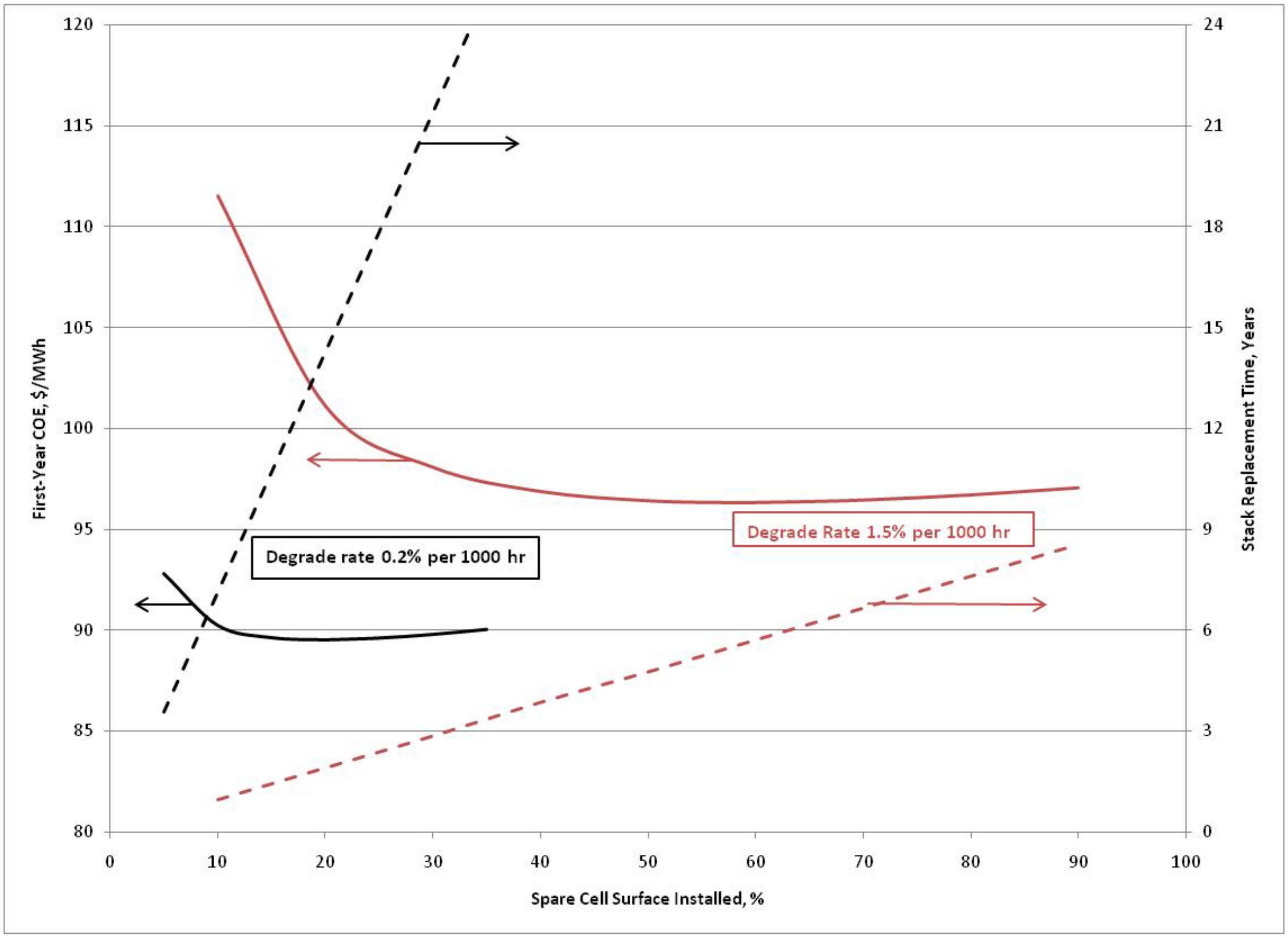
It is assumed in this evaluation that spare SOFC cell surface (blocks and enclosures) are provided at a cost of 165 $/SOFC kW based on the cost considerations in this section, and with the spare surface based on the cell degradation rate and the selected cell replacement period. The entire cell surface would be replaced (the blocks only) at a cost of 140 $/SOFC kW, with an assumed 10 percent discount rate after the cell has degraded a selected extent.

Exhibit 2-8 shows an illustration of the impact of the cell degradation rate and the spare cell surface initially installed in the plant on the plant first-year cost of electricity (COE) for plants having cell degradation rates of 1.5 percent per 1000 hours and 0.2 percent per 1000 hours. If too little space cell surface is installed the COE will be high due to the need to frequently replace the stacks. Increased spare cell surface installation leads to a relatively flat COE region where the COE is little influenced by the amount of spare surface installed and the stack replacement period can be selected for best plant maintenance schedule. An optimum spare surface installed exists and this is applied in the pathway study. For 1.5 percent per 1000 hour degradation, the optimum spare surface is 58.4 percent with 5.5 year stack replacement time. For 0.2 percent per 1000 hour degradation, the optimum spare surface is 19.7 percent with 14.1 year stack replacement time.

Administrative and Support Labor

Labor administration and overhead charges are assessed at rate of 25 percent of the burdened operation and maintenance labor.

Exhibit 2-8 Impact of Cell Degradation and Cell Stack Replacement Period



Consumables

The cost of consumables, including fuel, was determined on the basis of individual rates of consumption, the unit cost of each specific consumable commodity, and the plant annual operating hours.

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

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

Initial fills of the consumables, fuels and chemicals, are different from the initial chemical loadings, which are included with the equipment pricing in the capital cost.

Waste Disposal

Waste quantities and disposal costs were estimated similarly to the consumables. In this study slag/ash and sorbents from the IGFC cases are considered a waste with a disposal cost of

$17.89/tonne ($16.23/ton). The carbon used for trace element control in the IGFC cases is considered a hazardous waste with disposal cost of $926/tonne ($840/ton).

Co-Products and By-Products

By-product quantities were also determined similarly to the consumables. However, due to the variable marketability of these by-products, specifically sulfur, no credit was taken for their potential salable value. Nor were any of the technologies penalized for their potential disposal cost. That is, for this evaluation, it is assumed that the by-product or co-product value simply offset disposal costs, for a net zero in operating costs.

Owner's Costs

The owner's costs included in the TOC cost estimate are shown in Exhibit 2-9.

Exhibit 2-9 Owner's Costs Included in TOC

Owner's Cost	Comprised of
Preproduction Costs	• 6 months operating, maintenance, and administrative & support labor • 1 month maintenance materials • 1 month non-fuel consumables • 1 month of waste disposal costs • 25% of one month's fuel cost @ 100% capacity factor • 2% of TPC
Inventory Capital	• 60 day supply of fuel and consumables @100% capacity factor • 0.5% of TPC (spare parts)
Land	• $3,000/acre (300 acres for greenfield IGCC and PC)
Financing Costs	• 2.7% of TPC
Other Owner's Costs	• 15% of TPC
Initial Cost for Catalyst and Chemicals	• All initial fills not included in BEC
Prepaid Royalties	• Not included in owner's costs (included with BEC)
Property Taxes & Insurance	• 2% of TPC (Fixed O&M cost)
AFUDC and Escalation	• Varies based on levelization period and financing scenario • 33-yr IOU high risk: TASC = TOC *1.078 • 33-yr IOU low risk: TASC = TOC * 1.075 • 35-yr IOU high risk: TASC = TOC * 1.140 • 35-yr IOU low risk: TASC = TOC * 1.134

The category labeled "Other Owner's Costs" includes the following:

- Preliminary feasibility studies, including a Front-End Engineering Design (FEED) study
- Economic development (costs for incentivizing local collaboration and support)
- Construction and/or improvement of roads and/or railroad spurs outside of site boundary.

- Legal fees
- Permitting costs
- Owner's engineering (staff paid by owner to give third-party advice and to help the owner oversee/evaluate the work of the EPC contractor and other contractors)
- Owner's contingency: sometimes called "management reserve", these are funds to cover costs relating to delayed startup, fluctuations in equipment costs, unplanned labor incentives in excess of a five-day/ten-hour-per-day work week

Cost items excluded from "Other Owner's Costs" include:

- EPC Risk Premiums: Costs estimates are based on an Engineering Procurement Construction Management (EPCM) approach utilizing multiple subcontracts, in which the owner assumes project risks for performance, schedule and cost. This approach provides the owner with greater control of the project, while minimizing, if not eliminating most of the risk premiums typically included in a lump-sum, "turnkey" Engineer/Procure/Construct (EPC) contract, under which the EPC contractor assumes some or all of the project risks. The EPCM approach used as the basis for the estimates here is anticipated to be the most cost effective approach for the owner.
- Transmission interconnection: the cost of interconnecting with power transmission infrastructure beyond the plant busbar.
- Taxes on capital costs: all capital costs are assumed to be exempt from state and local taxes.
- Unusual site improvements: normal costs associated with improvements to the plant site are included in the bare erected cost, assuming that the site is level and requires no environmental remediation. Unusual costs associated with the following design parameters are excluded: 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.

CO_2 Transport, Storage and Monitoring

An approach for estimating capital and operating costs for CO_2 transport, storage and monitoring (TS&M), as independently estimated by the National Energy Technology Laboratory (NETL) [1], was applied in this pathway study. Those costs were converted to a current-dollar, cost of electricity (COE) and combined with the plant capital and operating costs to produce an overall COE.

CO_2 TS&M costs were estimated based on the following assumptions:

- CO_2 is supplied to the pipeline at the plant fence line at a pressure of 15.3 MPa (2,215 psia).
- The dried CO_2 sequestration stream will contain about 2 mole percent oxygen and it is assumed that this will be acceptable for the CO_2 piping system and the storage formation even though it does not meet the specification described in Exhibit 2-10 [14].

Exhibit 2-10 CO_2 Pipeline Specification

Parameter	Units	Parameter Value
Inlet Pressure	MPa (psia)	15.3 (2,215)
Outlet Pressure	MPa (psia)	10.4 (1,515)
Inlet Temperature	°C (°F)	26 (79)
N_2 Concentration	ppmv	< 300
O_2 Concentration	ppmv	< 40
Ar Concentration	ppmv	< 10

- The CO_2 is transported 80 kilometers (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.[14] The pipeline is assumed to have an outlet pressure (above the supercritical pressure) of 10.4 MPa (1,515 psia) with no recompression along the way. 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 an 80 kilometer (50 mile) pipeline length [16]. (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 1,239 meters (4,055 ft) and has a permeability of 22 millidarcy (a measure of permeability defined as roughly 10^{-12} Darcy) 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 tonnes (10,320 short tons) of CO_2 injected per day [15]. The assumed aquifer characteristics are tabulated in Exhibit 2-11.

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Exhibit 2-11 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)

First-Year, Current-Dollar Cost of Electricity

The revenue requirement method of performing an economic analysis of a prospective power 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 a current-dollar, first-year cost of electricity (LCOE). The COE is expressed in mills/kWh (numerically equivalent to $/MWh). The current-dollar, first-year COE was calculated using a simplified equation derived from the NETL Power Systems Financial Model [17].

The equation used to calculate COE is as follows:

$$COE = \frac{(CCF_P)(TOC) + [(OC_{F1}) + (OC_{F2}) + \ldots] + (CF)[(OC_{V1}) + (OC_{V2}) + \ldots]}{(CF)(MWh)}$$

where

$LCOE_P$ =	levelized cost of electricity over P years, $/MWh
CCF_P =	capital charge factor for a levelization period of P years
TOC =	total overnight cost, $
OC_{Fn} =	category n fixed operating cost for the initial year of operation
CF =	plant capacity factor
OC_{Vn} =	category n variable operating cost at 100 percent capacity factor for the initial year of operation
MWh =	annual net megawatt-hours of power generated at 100 percent capacity factor

All costs are expressed in June, 2007 year dollars, and the resulting COE is also expressed in June, 2007 year dollars. The COE for TS&M costs are added to the COE calculated using the above equation to generate a total cost including CO_2 capture, sequestration and subsequent monitoring.

The economic assumptions used to derive the capital charge factors are shown in Exhibit 2-12. The difference between the high risk and low risk categories is manifested in the debt-to-equity ratio and the weighted cost of capital. The values used to generate the capital charge factors and levelization factors in this study are shown in Exhibit 2-13.

Exhibit 2-12 Parameter Assumptions for Capital Charge Factors

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%
INFLATION	
LCOE, O&M, Fuel Escalation (nominal annual rate) Escalation rates must be the same for LCOE approximation to be valid	3.0%[2] COE, 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 1947 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.)

Exhibit 2-13 Financial Structure for Investor Owned Utility High and Low Risk Projects

Type of Security	% of Total	Current (Nominal) Dollar Cost	Weighted Current (Nominal) Cost	After Tax Weighted Cost of Capital
Low Risk				
Debt	50	4.5%	2.25%	
Equity	50	12%	6%	
Total			8.25%	7.39%
High Risk				
Debt	45	5.5%	2.475%	
Equity	55	12%	6.6%	
Total			9.075%	8.13%

The cost of CO_2 avoided for each pathway case is defined as:

$$\{COE_{\text{with CCS}} - COE_{\text{reference}}\} / \{CO_2 \text{ emission}_{\text{reference}} - CO_2 \text{ emission}_{\text{with CCS}}\}, \$/\text{tonne}$$

where the reference plant is taken to be the superciritical PC plant without CCS, having a first-year COE of 58.9 $/MWh, and a CO_2 emission of 802 tonne/hr [1].

3. IGFC Pathway with Conventional Gasification Technology

Two IGFC power plant scenarios with a series of pathway parameters, all using conventional coal gasification technology, are evaluated in this section. The cases utilize the commercial CoP E-Gas™ gasifier technology. The Scenario 1 plant configuration uses the SOFC operated at atmospheric-pressure. A branch of this pathway (Case 1-6) applies natural gas injection into the clean syngas to raise the methane content in the syngas and promote improve SOFC power island performance. The Scenario 2 configuration uses SOFC operated at elevated pressure. The performance of the steam bottoming cycle in these IGFC plants vary based on the oxy-combustor exhaust conditions, and the steam bottoming cycle represents a much smaller portion of the overall plant power generation than is true for other types of fossil power plants, such as PC, IGCC, and NGCC power plants.

3.1 Descriptions of Process Areas

The IGFC plant, like the IGCC plant, consists of several integrated process areas, the primary ones being the coal receiving and storage area, the air separation unit, the gasification area, the gas cleaning area, the power island, and the CO_2 dehydration and compression area. Descriptions of these areas and their selected technologies are presented in this report section, many of these plant areas having descriptions analogous to those used for IGCC in the Bituminous Baseline report. Additional case-specific performance information, and the performance features for these areas are presented in the relevant case sections.

3.1.1 Coal Receiving and Storage Area

The function of the Coal Receiving and Storage system is to unload, convey, prepare, and store the coal delivered to the plant. The scope of the system is from the trestle bottom dumper and coal receiving hoppers up to and including the slide gate valves at the outlet of the coal storage silos. Coal receiving and storage is identical for all of the IGFC cases; however, coal preparation and feed are gasifier-specific.

The coal is delivered to the site by 100-car unit trains comprised of 91 tonne (100 ton) rail cars. The unloading is done by a trestle bottom dumper, which unloads the coal into two receiving hoppers. Coal from each hopper is fed directly into a vibratory feeder. The 8 cm x 0 (3" x 0) coal from the feeder is discharged onto a belt conveyor. Two conveyors with an intermediate transfer tower are assumed to convey the coal to the coal stacker, which transfer the coal to either the long-term storage pile or to the reclaim area. The conveyor passes under a magnetic plate separator to remove tramp iron and then to the reclaim pile.

The reclaimer loads the coal into two vibratory feeders located in the reclaim hopper under the pile. The feeders transfer the coal onto a belt conveyor that transfers the coal to the coal surge bin located in the crusher tower. The coal is reduced in size to 3 cm x 0 (1¼" x 0) by the crusher. A conveyor then transfers the coal to a transfer tower. In the transfer tower the coal is routed to the tripper, which loads the coal into one of three silos. Two sampling systems are supplied: the as-received sampling system and the as-fired sampling system.

3.1.2 Air Separation Unit

The air separation unit (ASU) generates oxidant for use in three sections of the IGFC plant: the coal gasifier, the Claus sulfur recovery process, and the anode gas oxy-combustor. In this study, the ASU main air compressor discharge pressure was set at 0.5 MPa (79 psia), providing oxygen product at sufficient pressure, 0.16 MPa (23 psia), to operate the oxy-combustor for the atm-pressure SOFC applications. The ASU is designed to generate 99.5 percent pure oxygen for IGFC applications to maintain the sequestered CO_2 stream with low nitrogen and argon content. There is no opportunity for ASU air-side integration in the IGFC plant like there are in IGCC plants, and there is no need or benefit from syngas nitrogen dilution in the IGFC. In this study, the ASU nitrogen product was used only for plant inerting needs and solids transport needs, with the remainder vented.

An air compressor providing air to the ASU is powered by an electric motor. Air to this stand-alone compressor is first filtered in a suction filter upstream of the compressor. This air filter removes particulate, which may tend to cause compressor wheel erosion and foul intercoolers. The filtered air is then compressed in the centrifugal compressor, with intercooling between each stage.

Air from the compressor is cooled and fed to an adsorbent-based pre-purifier system. The adsorbent removes water, carbon dioxide, and C_4+ saturated hydrocarbons in the air. After passing through the adsorption beds, the air is filtered with a dust filter to remove any adsorbent fines that may be present. Downstream of the dust filter a small stream of air is withdrawn to supply the instrument air requirements of the ASU.

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

The air from the pre-purifier is then split into three streams. About 70 percent of the air is fed directly to the cold box. About 25 percent of the air is compressed in an air booster compressor. This boosted air is then cooled in an aftercooler against cooling water in the first stage and against chilled water in the second stage before it is fed to the cold box. The chiller utilizes low pressure process steam at 0.3 MPa (50 psia) to drive the absorption refrigeration cycle. The remaining 5 percent of the air is fed to a turbine-driven, single-stage, centrifugal booster compressor. This stream is cooled in a shell and tube aftercooler against cooling water before it is fed to the cold box.

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

Inside the cold box the air is separated into oxygen and nitrogen products. The oxygen product is withdrawn from the distillation columns as a liquid and is pressurized by a cryogenic pump. The pressurized liquid oxygen is then vaporized against the high-pressure air feed before being warmed to ambient temperature. The gaseous oxygen exits the cold box and a portion is fed to the power island's oxy-combustor and Claus plant. The remainder of the oxygen is fed to the centrifugal compressor with intercooling between each stage of compression. This compressed oxygen is then fed to the gasification unit.

3.1.3 Conventional Coal Gasification Area

Two gasification technologies were selected for this pathway: a conventional, ConocoPhillips (CoP), entrained coal gasification technology, and a conceptual, near-term, enhanced coal gasifier. The conventional CoP coal gasifier technology was selected for use in the IGFC plant because it can produce a syngas having a moderate methane content of about 5.9 mole percent. The syngas produced by the E-Gas™ gasifier is higher in methane content than either the GEE or Shell gasifier. The two stage design allows for improved cold gas efficiency and lower oxygen consumption, but the quenched second stage allows some CH_4 to remain. The syngas CH_4 concentration exiting the gasifier is 5.9 vol percent (dry gas), compared to 0.10 vol percent for the GEE and 0.001 vol percent for the Shell gasifier.

Methane is expected to be beneficial to the IGFC plant performance because it can provide cooling of the SOFC stack when methane reforms in parallel with the syngas oxidation. This reforming, thus, reduces the excess cathode air flow needed for SOFC stack temperature control.

A conceptual enhanced coal gasifier having design features similar to the commercial CoP gasifier, but operated to achieve a higher syngas methane content of about 10 mole percent was also considered to determine the potential benefits of developing and applying such a gasifier as part of the IGFC pathway.

ConocoPhillips E-Gas™ Gasifier

The conventional, entrained, CoP E-Gas™ gasification technology can be operated to generate a syngas having a moderate methane content of approximately 6 mole percent, and it represents one of the best conventional coal gasifier technologies for use in IGFC. The design basis and performance estimates for the CoP gasifier were taken from the Bituminous Baseline report. Its cold gas efficiency is estimated to be 81 percent (HHV). This design basis is described in Exhibit 3-1.

The Scenario 1 plant cases in this study utilize two parallel gasification trains to process Illinois No. 6 coal. The gasifiers operate at maximum capacity. The E-Gas™ two-stage coal gasification technology features an oxygen-blown, entrained-flow, refractory-lined gasifier with continuous slag removal.

Coal from the coal silo is fed onto a conveyor by vibratory feeders located below each 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. Each hopper outlet discharges onto a weigh feeder, which in turn feeds a rod mill. Each rod mill is sized to process 55 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 through a trommel screen into the rod mill discharge tank, and then the slurry is pumped to the slurry storage tanks. The dry solids concentration of the final slurry is 63 percent.

Exhibit 3-1 Coal Gasification Section Assumptions with CoP E-Gas™ Gasifier

	Specification/Assumptions
Gasifier	
Technology	CoP 2-stage coal-water slurry
Number in parallel	2
Dried coal moisture, wt%	11.0 (as-received)
Coal feed type	coal-water slurry pumps
Oxygen-to-coal feed ratio	0.68
Slurry coal content, wt%	71
Steam-to-coal ratio	0.33
Steam temperature, ºC (ºF)	288 (550) saturated
Recycle gas-to-coal ratio	0.31
Recycle gas compressor eff., % (adiabatic)	85
Exit temperature, ºC (ºF)	999 (1830)
Exit pressure, MPa (psia)	3.10 (450)
Carbon loss with ash, % of coal carbon	0.8
Raw syngas composition basis	Equilibrium approach
Syngas methane content, vol% (dry)	5.9
Raw Syngas Cooler	
Technology	Fire-tube boiler
Number in parallel	2
Outlet temperature, ºC (ºF)	316 (600)

About 78 percent of the total slurry feed is fed to the first (or bottom) stage of the gasifier. All oxygen for gasification is fed to this stage of the gasifier at a pressure of 4.2 MPa (615 psia). This stage is best described as a horizontal cylinder with two horizontally opposed burners. The highly exothermic gasification/oxidation reactions take place rapidly at temperatures of 1,316 to 1,427°C (2,400 to 2,600°F). The hot raw gas from the first stage enters the second (top) stage, which is a vertical cylinder perpendicular to the first stage. The remaining 22 percent of coal slurry is injected into this hot raw gas. The endothermic gasification and devolatilization reactions in this stage reduce the final gas temperature to about 999°C (1,830°F).

The coal ash in the first-stage is converted to molten slag, which flows down through a tap hole. The molten slag is quenched in water and removed through a proprietary continuous-pressure letdown/dewatering system. Char is produced in the second gasifier stage and is captured and recycled to the hotter first stage to be gasified.

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

The slag is dewatered, the water is clarified and recycled and the dried slag is transferred to a storage area for disposal. The specifics of slag handling vary among the gasification technologies regarding how the water is separated and the end uses of the water recycle streams.

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

Enhanced, Conventional Gasifier

The enhanced conventional gasifier represents a conceptual extrapolation of the CoP gasifier. Conventional gasifier enhancement activities are currently being conducted [27]. Its estimated operating parameters are listed in Exhibit 3-2. The cold gas efficiency of this gasifier is estimated to be 82.5 % (HHV). The other features of the gasifier, including its stage-one characteristics, are expected to be very similar to the CoP E-Gas™ gasifier.

Exhibit 3-2 Coal Gasification Section Assumptions with Enhanced Gasifier

	Specification/Assumptions
Gasifier	
Technology	2-stage coal-water slurry
Number in parallel	2
Dried coal moisture, wt%	11.0 (as-received)
Coal feed type	coal-water slurry pumps
Oxygen-to-coal feed ratio	0.61
Slurry coal content, wt%	71
Steam-to-coal ratio	0.33
Steam temperature, ºC (ºF)	288 (550) saturated
Recycle gas-to-coal ratio	0.31
Recycle gas compressor eff., % (adiabatic)	85
Exit temperature, ºC (ºF)	935 (1715)
Exit pressure, MPa (psia)	4.82 (700)
Carbon loss with ash, % of coal carbon	0.8
Raw syngas composition basis	Equilibrium approach
Syngas methane content, vol% (dry)	10.2
Raw Syngas Cooler	
Technology	Fire-tube boiler
Number in parallel	2
Outlet temperature, ºC (ºF)	316 (600)

3.1.4 Syngas Cleaning Area

The Gas Cleaning Area's function is to remove contaminants from the gasifier raw syngas to protect the downstream equipment from damage and to satisfy environmental emission requirements. In IGFC the dominant requirement is the protection of the SOFC from syngas contaminants, with these specifications being much more stringent than the environmental requirements.

All of the IGFC plant configuration cases apply conventional, dry gas cleaning technology. Single-stage Selexol acid gas removal is applied in all of the IGFC cases, with this technology expected to generate a clean syngas more acceptable to the fuel cell application than alternatives such as amine-based acid gas removal. The Selexol acid gas removal is preceded by COS hydrolysis and by low-temperature, activated-carbon beds to remove mercury and other trace elements. The Selexol acid gas removal step is followed by a syngas reheat step and conventional zinc oxide (ZnO) beds to polish the syngas to acceptable sulfur levels (less than 100 ppbv total sulfur). This clean syngas is expanded to the required pressure and fed to the fuel cell as its anode feed gas.

The Gas Cleaning Area uses conventional dry gas cleaning technology based on single-stage Selexol acid gas removal, illustrated in the Exhibit 3-3 block flow diagram. The Area components are a high-temperature barrier filter, a water scrubbing system, a COS hydrolysis unit, a low-temperature syngas cooling system, a trace element removal system, a Selexol single-stage acid gas removal process, a syngas reheat unit, and a ZnO fixed-bed sulfur-polishing unit. Its conditions and configuration are nearly identical for both the catalytic gasifier and the conventional gasifier pathways, with only a few differences in operating conditions. The configuration is slightly different, with the conventional gasifier pathway cases performing syngas reheat using HP-steam indirect heating, and the catalytic gasifier pathway cases using recuperative gas-to-gas heating, as shown in Exhibit 3-3.

Exhibit 3-4 summarizes the major syngas cleaning section assumptions and specifications. The inherent assumption in this evaluation is that the coal syngas subjected to the listed cleaning steps will be acceptable to the SOFC stack unit for long term operation. This long-term success has not yet been demonstrated.

Exhibit 3-3 Syngas Cleaning for IGFC

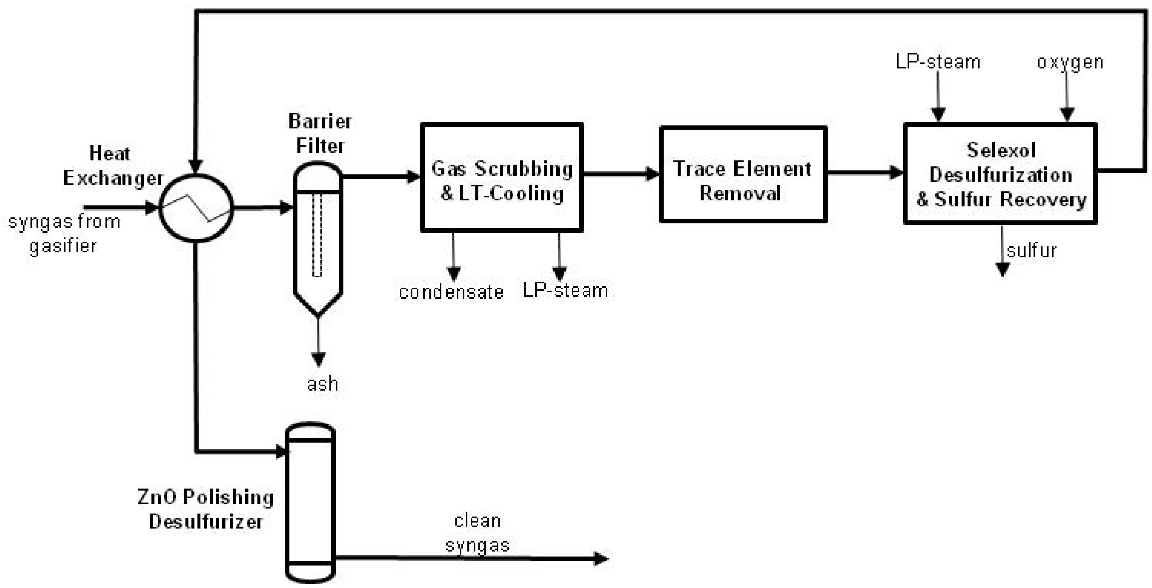


Exhibit 3-4 Gas Cleaning Area Assumptions Conventional Gasifier Cases

	Specification/Assumptions
Gas Cleaning technology	
Technology	Conventional dry gas cleaning
Number parallel trains	2
Particulate removal	Barrier filter at 316ºC (600ºF)
HCl removal	Water scrubber
Ammonia removal	Low-temperature gas cooling to 35 ºC (95 ºF)
Hg, As, Se, Cd, P	Activated-Carbon fixed beds at 35 ºC (95 ºF)
Bulk desulfurization	Selexol at 35 ºC (95 ºF)
Sulfur recovery	Conventional Claus plant with tail gas recycle
Polishing Desulfurization	ZnO fixed beds at 316ºC (600ºF)
Syngas Preheating Source	HP-steam heating for CoP gasifier

Raw Gas Cooling/Particulate Removal

The raw syngas from the gasifier is cooled to its desired temperature in the syngas cooler unit, which consists of a fire-tube boiler and convective superheating and economizing sections. Fire-tube boilers cost markedly less than comparable duty, water-tube boilers. This is because of the large savings in high-grade steel associated with containing the hot, high-pressure synthesis gas in relatively small tubes.

The cooled gas from the syngas cooler is cleaned of remaining particulate via a cyclone collector followed by a ceramic candle filter. Recycled syngas is used as the pulse gas to clean the candle filters. In the cases using the conventional gasifier, the recovered fines are pneumatically returned to the first stage of the gasifier. The recycled char and recycled particulate results in high overall carbon conversion.

With the conventional gasifier, following particulate removal, additional heat is removed from the syngas to raise saturated IP steam at 0.4 MPa (65 psia). In this manner the syngas is cooled to 232°C (450°F) prior to the syngas scrubber.

Syngas Scrubber and Low-Temperature Cooling Section

The cooled syngas passes to a syngas scrubber where a water wash is used to remove primarily chlorides, and any particulate that might have penetrated the barrier filter. The syngas exits the scrubber saturated at about 169°C (337°F). This is followed by low-temperature cooling to 35°C (95°F), removing primarily NH_3 and generating condensate streams.

The sour water stripper removes NH_3, H_2S, and other impurities from the scrubber and other condensate streams. The stripper consists of a sour drum that accumulates sour water from the gas scrubber and condensate from synthesis gas coolers. Sour water from the drum 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 sent to wastewater treatment.

COS Hydrolysis

The use of COS hydrolysis pretreatment in the feed to the acid gas removal process provides a means to reduce the COS concentration. Several catalyst manufacturers including Haldor Topsoe and Porocel offer a catalyst that promotes the COS hydrolysis reaction.

Syngas exiting the scrubber is reheated to about 186°C (367°F) by using HP steam from the HRSG evaporator prior to entering a COS hydrolysis reactor. About 99.5 percent of the COS is converted to CO_2 and H_2O.

The COS hydrolysis reaction is equimolar with a slightly exothermic heat of reaction. The reaction is represented as follows.

$$COS + H_2O \leftrightarrow CO_2 + H_2S$$

Since the reaction is exothermic, higher conversion is achieved at lower temperatures. However, at lower temperatures the reaction kinetics are slower. Based on the feed gas for this evaluation, Porocel recommended a temperature of 177 to 204°C (350 to 400°F). Since the exit gas COS concentration is critical to the amount of H_2S that must be removed with the AGR process, a retention time of 50-75 seconds was used to achieve 99.5 percent conversion of the COS. The

Porocel activated alumina-based catalyst, designated as Hydrocel 640 catalyst, promotes the COS hydrolysis reaction without promoting reaction of H_2S and CO to form COS and H_2.

Although the reaction is exothermic, the heat of reaction is dissipated among the large amount of non-reacting components. Therefore, the reaction is essentially isothermal. The product gas, now containing less than 4 ppmv of COS, is cooled prior to entering the mercury removal process and the AGR.

Trace Removal

The gas exiting the COS reactor passes through a series of heat exchangers and knockout drums to lower the syngas temperature to 35°C (95°F) and to separate entrained water. The cooled syngas then passes through a carbon bed to remove 95 percent of the Hg and other trace metals.

A conceptual design for an activated, sulfur-impregnated, carbon bed adsorption system was developed for mercury control in the IGCC plants being studied. Data on the performance of carbon bed systems were obtained from the Eastman Chemical Company, which uses carbon beds at its syngas facility in Kingsport, Tennessee [19]. IGFC-specific design considerations are discussed below.

The packed carbon bed vessels are located upstream of the Selexol acid gas removal unit and syngas enters at a temperature near 38°C (100°F). Eastman Chemical also operates their beds ahead of their acid gas removal unit at a temperature of 30°C (86°F) [18].

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

The bed density of 30 lb/ft^3 was based on the Calgon Carbon Corporation HGR-P sulfur-impregnated pelletized activated carbon [19]. These parameters determined the size of the vessels and the amount of carbon required. The gasifier train has one mercury removal.

Eastman Chemicals replaces its bed every 18 to 24 months. However, bed replacement is not because of mercury loading, but for other reasons including:

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

For this study a 24 month carbon replacement cycle was assumed. Under these assumptions, the mercury loading in the bed would build up to 0.6 - 1.1 weight percent (wt percent). Mercury capacity of sulfur-impregnated carbon can be as high as 20 wt percent [19]. The mercury laden carbon is considered to be a hazardous waste, and the disposal cost estimate reflects this categorization.

It is assumed that other trace species, such as arsenic, selenium, cadmium, and phosphorus will also be effectively removed by this unit.

Acid gas Removal Process

A key function of syngas cleaning is acid gas removal (AGR) with sulfur recovery. The total sulfur content of the syngas is reduced to less than 30 ppmv. This includes all sulfur species, but in particular the total of COS and H_2S. Selexol was chosen for AGR in all of the pathways based on the gasifier operating at high pressure which favors the physical solvent used in the Selexol process. The Selexol process uses dimethyl ether of polyethylene glycol as a solvent [20].

Cool, particulate-free synthesis gas enters the Selexol absorber unit at approximately 34°C (94°F). In this absorber, H_2S is preferentially removed from the fuel gas stream along with smaller amounts of CO_2, COS and other gases such as hydrogen. The rich solution leaving the bottom of the absorber is heated against the lean solvent returning from the regenerator before entering the H_2S concentrator. A portion of the non-sulfur bearing absorbed gases is driven from the solvent in the H_2S concentrator using N_2 from the ASU as the stripping medium. The temperature of the H_2S concentrator overhead stream is reduced prior to entering the reabsorber where a second stage of H_2S absorption occurs. The rich solvent from the reabsorber is combined with the rich solvent from the absorber and sent to the stripper where it is regenerated through flash pressure reduction in a series of flash vessels. The stripper acid gas stream, consisting of H_2S and CO_2, with some N_2, is then sent to the Claus unit.

Sulfur Recovery/Tail Gas Cleanup Process

The conventional three-stage Claus plant, with indirect reheat and feeds with a high H_2S content, exceeds 98 percent sulfur recovery efficiency [20].

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

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

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

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

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

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

One-third of the H_2S is burned in the furnace with oxygen to give sufficient SO_2 to react with the remaining H_2S. Since these reactions are highly exothermic, a waste heat boiler that recovers this heat to generate high-pressure steam following the furnace. Sulfur is condensed in a condenser that follows the high-pressure steam recovery section. Low-pressure steam is raised in the condenser. The tail gas from the first condenser then goes to several catalytic conversion stages, usually 2 to 3, where the remaining sulfur is recovered via the Claus reaction. Each catalytic stage consists of gas preheat, a catalytic reactor, and a sulfur condenser. The liquid sulfur goes to the sulfur pit, while the tail gas proceeds to the incinerator or for further processing in a TGTU.

The Claus reaction is equilibrium limited, and sulfur conversion is sensitive to the reaction temperature. The highest sulfur conversion in the thermal zone is limited to about 75 percent. Typical furnace temperatures are in the range from 1093 to 1427°C (2000 to 2600°F), and as the temperature decreases, conversion increases dramatically. Claus plant sulfur recovery efficiency depends on many factors such as H_2S concentration of the feed gas, number of catalytic stages, and the gas reheat method. In many refinery and other conventional Claus applications, tail gas treating involves the removal of the remaining sulfur compounds from gases exiting the sulfur recovery unit. Tail gas from a typical Claus process contains small, but varying quantities of COS, CS_2, H_2S, SO_2, and elemental sulfur vapors. In addition, there is some H_2, CO, and CO_2 in the tail gas. In order to remove the rest of the sulfur compounds from the tail gas, all of the sulfur-bearing species must first be converted to H_2S. Then, the resulting H_2S is absorbed into a solvent and the clean gas vented or recycled for further processing. In all of the IGFC cases, the Claus plant tail gas is hydrogenated, water is separated, and tail gas is compressed and is then returned to the AGR process for further treatment.

3.1.5 Sulfur Polishing

Several commercial sorbents are available for syngas sulfur polishing. Zinc oxide-based sorbents, having one of the highest affinities for hydrogen sulfide removal, are applicable for desulfurization to levels less than 100 ppbv and are offered by several catalyst vendors.

They operate at relatively high temperatures, 260-427°C (500-800°F) and are typical applied in batch operated, packed bed vessels. These vessels are normally operated with syngas downflow through the packed bed, and the packed bed is supported on a ceramic or metal syngas distribution device that promotes uniform syngas flow through the bed, and maintains gas velocities at the distributor low enough to prevent sorbent particle attrition. The sorbents are manufactured with sizes that allow reasonable gas velocities through the beds with acceptable pressure drops. The sorbent particles have pore structures that provide rapid reaction conditions so that a distinct reaction front moved through the bed. When sulfur breakthrough is approached in the bed, or when the bed pressure drop becomes excessive, the vessel is taken out of service, is drained and is refilled with fresh sorbent. The bulk desulfurized syngas from the Selexol unit must be preheated by gas-to-gas heat exchange with the warm syngas from the barrier filter, or by indirect steam heating with high-pressure steam.

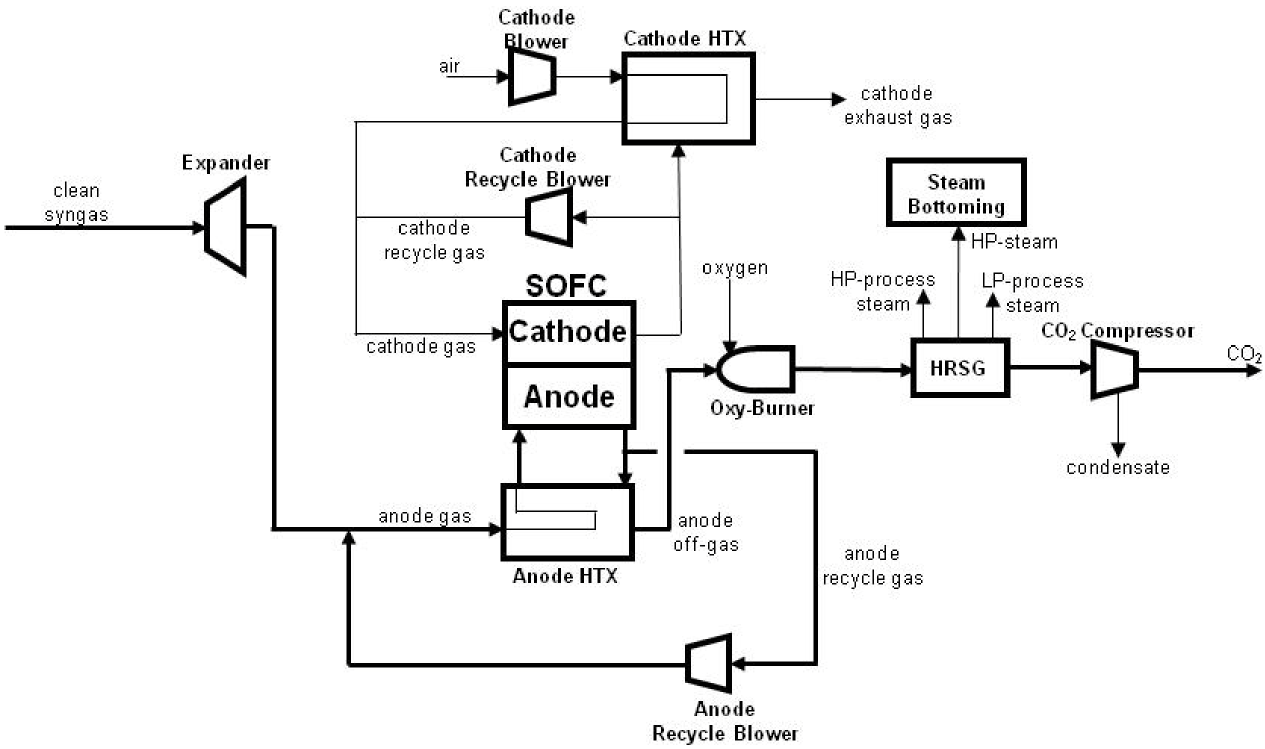
3.1.6 SOFC Power Island

The SOFC power island components are shown in Exhibit 3-5 block flow diagram. They consist of a syngas expander that expands the syngas from its high-pressure condition down to the operating pressure of the fuel cell unit, the SOFC fuel cell unit with DC-AC inverters, an anode off-gas oxy-combustor, a heat recovery steam generator that captures heat from the combusted anode off-gas, and a steam bottoming cycle. The SOFC fuel cell unit ancillary components consist of cathode air blowers, cathode heat exchangers that recuperatively heat the cathode air up to the fuel cell inlet temperature, cathode hot gas recycle blowers, anode heat exchangers that recuperatively heat the anode gas up to the fuel cell inlet temperature, and anode hot gas recycle blowers. Hot gas blowers capable of operation at the required conditions of the anode and cathode recycle gas streams are currently under development [21].

The heat recovery steam generator produces low-pressure and high-pressure process steam, and high-pressure power steam for the subcritical steam bottoming cycle. The cooling water system uses a mechanical draft, wet cooling tower arrangement.

In Scenario 2, in which pressurized fuel cell operation is used, the cathode air is compressed to the pressurized fuel gas inlet pressure, and no cathode gas recycle is used. The cathode off-gas is expanded to atmospheric pressure to generate power to drive the cathode gas compressor. Anode gas recycle is accomplished using a syngas-driven jet pump in this pressurized case.

Exhibit 3-5 IGFC Power Island



The major assumptions and information sources for the atmospheric-pressure SOFC power island are listed in Exhibit 3-6. In all of the study cases, it is assumed that the anode inlet gas to the fuel cell must have a total oxygen-to-carbon atomic ratio of at least 2.0 to avoid carbon deposition in the fuel cell. This constraint is satisfied by maintaining sufficiently high anode gas recycle, with the hot anode gas recycle increasing the water vapor content, and the associated oxygen-to-carbon atomic ratio, in the anode inlet gas.

The anode off-gas is combusted using oxygen in an advanced oxy-combustor with excess oxygen limited to 1 mole percent. It is assumed that an anode off-gas oxy-combustor can be developed that can operate stably with 1 mole percent excess oxygen.

The combusted anode gas consists of CO_2, one mole-percent excess oxygen, water vapor, and minor traces of syngas contaminants (i.e., sulfur species, HCl, NO_x, trace elements). This combusted gas is dehydrated and compressed to the sequestration pressure of 2,200 psig. In its dry state it will contain about two mole percent oxygen. It is assumed that this will be acceptable, although it far exceeds the currently adopted criteria for CO_2 sequestration gas.

Exhibit 3-6 Atmospheric-Pressure Power Island Base Assumptions

	Specification/Assumptions
Syngas Expander	
Outlet pressure, MPa (psia)	0.21 (30)
Efficiency, adiabatic %	90
Generator efficiency (%)	98.5
Fuel Cell System	
Cell stack inlet temperature, ºC (ºF)	650 (1202)
Cell stack outlet temperature, ºC (ºF)	750 (1382)
Cell stack outlet pressure, MPa (psia)	0.12 (15.6)
Fuel single-step utilization, %	75
Fuel overall utilization, %	90
Stack anode-side pressure drop, MPa (psi)	0.0014 (0.2)
Stack cathode-side pressure drop, MPa (psi)	0.0014 (0.2)
Power density, mW/cm^2	400
Stack over-potential, mV	140
Operating voltage estimation method	Section 8.1.4
Cell degradation rate (% per 1000 hours)	1.5
Cell replacement period (% degraded)	20
Fuel Cell System Ancillary Components	
Anode gas recycle method	Hot gas fan
Anode recycle gas fan efficiency, adiabatic %	80
Anode heat exchanger pressure drop, MPa (psi)	0.0014 (0.2)
Cathode gas recycle method	Hot gas fan
Cathode recycle gas rate, %	50
Cathode recycle gas fan eff., adiabatic %	80
Cathode heat exchanger pressure drop, MPa	0.0014 (0.2)
Cathode blower efficiency, adiabatic %	90
Rectifier DC-to-AC efficiency, %	97.0
Recycle blower motor drives eff., %	87.6
Other electric motor drives efficiency, %	95
Transformer efficiency, %	99.65
Oxy-Combustor	
Technology	Atm-pressure diffusion flame
Outlet excess O_2, mole%	1
Steam Bottoming Cycle	
Technology level	subcritical
Modeling approach	Empirical approximation
Other steam generation duties	HP and LP process steam

Heat Recovery Steam Generator

The heat recovery steam generator (HRSG) is a horizontal gas flow, drum-type, multi-pressure design that is matched to the characteristics of the oxy-combustor exhaust gas. High-temperature flue gas exiting the oxy-combustor is conveyed through the HRSG to recover the quantity of thermal energy that remains. High-pressure steam for power generation, and high-pressure and low-pressure process steam are generated in the HRSG. Flue gas travels through the HRSG gas path and exits at about 132°C (270°F).

Steam Turbine Generator and Auxiliaries

The steam turbine consists of an HP section, an IP section, and one double-flow low pressure (LP) section, all connected to the generator by a common shaft. The HP and IP sections are contained in a single-span, opposed-flow casing, with the double-flow LP section in a separate casing. The LP turbine has a last stage bucket length of 76 cm (30 in).

Main steam from the HRSG and gasifier island is combined in a header, and then passes through the stop valves and control valves and enters the turbine at either 12.4 MPa/559°C to 562°C (1800 psig/1038°F to 1043°F) for the non-carbon capture cases, or 12.4 MPa/534°C (1800 psig/993°F to 994°F) for the carbon capture cases. The steam initially enters the turbine near the middle of the high-pressure span, flows through the turbine, and returns to the HRSG for reheating. The reheat steam flows through the reheat stop valves and intercept valves and enters the IP section at 3.1 MPa/558°C to 561°C (443 psig/1036°F to 1041°F) for the non-carbon capture cases or 3.1 MPa/532°C to 533°C (443 psig/990°F to 992°F) for the carbon capture cases. After passing through the IP section, the steam enters a crossover pipe, which transports the steam to the LP section. The steam divides into two paths and flows through the LP sections, exhausting downward into the condenser.

The generator is a hydrogen-cooled synchronous type, generating power at 24 kV. A static, transformer type exciter is provided. The generator is cooled with a hydrogen gas recirculation system using fans mounted on the generator rotor shaft. The heat absorbed by the gas is removed as it passes over finned tube gas coolers mounted in the stator frame.

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

Condensate System

The condensate system transfers condensate from the condenser hotwell to the deaerator, through the gland steam condenser, gasifier, and the low-temperature economizer section in the HRSG. The system consists of one main condenser; two 50 percent capacity, motor-driven, vertical condensate pumps; one gland steam condenser; and a low-temperature tube bundle in the HRSG. Condensate is delivered to a common discharge header through separate pump discharge lines, each with a check valve and a gate valve. A common minimum flow recirculation line discharging to the condenser is provided to maintain minimum flow requirements for the gland steam condenser and the condensate pumps.

Feedwater System

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

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

Main and Reheat Steam Systems

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

Circulating Water System

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

Raw Water, Fire Protection, and Cycle Makeup Water Systems

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

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

The cycle makeup water system provides high quality demineralized water for makeup to the HRSG cycle, for steam injection ahead of the water gas shift reactors in CO_2 capture cases, and for injection steam to the auxiliary boiler for control of NOx emissions, if required.

3.1.7 CO_2 Dehydration and Compression Area

The oxy-combustion off-gas stream, after all heat recovery in completed, is compressed from its delivery pressure to a supercritical condition at 15.3 MPa (2215 psia) using four parallel multiple-stage, intercooled compressors. During compression, the CO_2 stream is dehydrated before each compression stage by water cooling and water knockout, and ultimately to a dewpoint of -40ºC (-40°F) with a triethylene glycol system. The CO_2 is transported to the plant fence line and is sequestration ready.

3.1.8 Accessory Electric Plant

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

3.1.9 Instrumentation and Control

An integrated plant-wide distributed control system (DCS) is provided. The DCS is a redundant microprocessor-based, functionally distributed control system. The control room houses an array of multiple video monitor (CRT) and keyboard units. The CRT/keyboard units are the primary interface between the generating process and operations personnel. The DCS incorporates plant monitoring and control functions for all the major plant equipment. The DCS is designed to be operational and accessible 99.5 percent of the time it is required (99.5 percent availability). The plant equipment and the DCS are designed for automatic response to load changes from minimum load to 100 percent. Startup and shutdown routines are manually implemented, with operator selection of modular automation routines available. The exception to this, and an important facet of the control system for gasification, is the critical controller system, which is a part of the license package from the gasifier supplier and is a dedicated and distinct hardware segment of the DCS.

This critical controller system is used to control the gasification process. The partial oxidation of the fuel feed and oxygen feed streams to form a syngas product is a stoichiometric, temperature- and pressure-dependent reaction.

3.2 Scenario 1 – IGFC with Atmospheric-Pressure SOFC

The Case 1-1 baseline configuration uses the conventional ConocoPhillips (CoP) E-Gas™ gasifier combined with atmospheric-pressure SOFC. The Coal Gasification Area contains the coal preparation system, the slag handling system, the coal water-slurry feeding system, the coal gasification system, the air separation system, and the raw syngas cooling system.

The Gas Cleaning Area uses conventional dry gas cleaning technology based on single-stage Selexol acid gas removal. The area components are a high-temperature barrier filter, a water scrubbing system, a COS hydrolysis unit, a low-temperature syngas cooling system, a trace element removal system, a Selexol single-stage acid gas removal system, a syngas reheat unit, and a ZnO fixed-bed sulfur-polishing unit. The baseline, atmospheric-pressure Power Island assumptions and specifications are listing in Exhibit 3-7.

3.2.1 Case 1-1 Baseline Plant Performance Results

The following information is presented in tabular form for Case 1-1:

- Block Flow Diagram and Stream Table
- Performance Summary
- Mass and Energy Flow Diagrams
- Steam Balance
- Water Balance
- Carbon Balance
- Sulfur Balance
- Air Emissions.

The system description follows the BFD in Exhibit 3-8, and stream numbers reference the same Exhibit. The table in Exhibit 3-9 provides process data for the numbered streams in the BFD. Note that 66.7 percent of the anode off-gas is recycled to the anode inlet stream, reducing the syngas methane content of 5.87 mole percent to 1.79 mole percent in the actual anode inlet stream. Exhibit 3-10 provides the power plant breakdown and overall thermal performance. Note that the steam turbine power represents about 16 percent of the total plant power generated, with the SOFC system being the overwhelmingly dominate power generator. The baseline plant efficiency of 40 percent (HHV) is very high for a power plant with carbon removal compared to other fossil fuel power plant technologies.

Exhibit 3-7 Case 1-1 Atmospheric-Pressure Power Island Base Assumptions

	Specification/Assumptions
Syngas Expander	
Outlet pressure, MPa (psia)	0.21 (30)
Efficiency, adiabatic %	90
Generator efficiency (%)	98.5
Fuel Cell System	
Cell stack inlet temperature, ºC (ºF)	650 (1202)
Cell stack outlet temperature, ºC (ºF)	750 (1382)
Cell stack outlet pressure, MPa (psia)	0.12 (15.6)
Fuel single-step utilization, %	75
Fuel overall utilization, %	90
Stack anode-side pressure drop, MPa (psi)	0.0014 (0.2)
Stack cathode-side pressure drop, MPa (psi)	0.0014 (0.2)
Power density, mW/cm^2	400
Stack over-potential, mV	140
Operating voltage estimation method	Section 8.1.4
Cell degradation rate (% per 1000 hours)	1.5
Cell replacement period (% degraded)	20
Fuel Cell System Ancillary Components	
Anode gas recycle method	Hot gas fan
Anode recycle gas fan efficiency, adiabatic %	80
Anode heat exchanger pressure drop, MPa (psi)	0.0014 (0.2)
Cathode gas recycle method	Hot gas fan
Cathode recycle gas rate, %	50
Cathode recycle gas fan eff., adiabatic %	80
Cathode heat exchanger pressure drop, MPa	0.0014 (0.2)
Cathode blower efficiency, adiabatic %	90
Rectifier DC-to-AC efficiency, %	97.0
Recycle blower motor drives eff., %	87.6
Other electric motor drives efficiency, %	95
Transformer efficiency, %	99.65
Oxy-Combustor	
Technology	Atm-pressure diffusion flame
Outlet excess O_2, mole%	1
Steam Bottoming Cycle	
Technology level	subcritical
Modeling approach	Empirical approximation
Other steam generation duties	HP and LP process steam

To enhance the understanding of the IGFC power plant, mass flow and energy flows diagrams are presented in Exhibit 3-11 and Exhibit 3-12 on a basis relative to the coal as-received mass feed rate, and relative to the coal feed energy (HHV), respectively. The mass flow diagram indicates that the mass of the CO_2 product stream is 2.4 times the mass of the coal feed stream, and the largest mass flows in the plant are associated with the cathode gas streams, these being as large as almost 24 times the coal feed flow. In this case, the oxidant flow to the oxy-combustor is about 24 percent of the oxidant flow to the coal gasifier.

The energy flow diagram indicates that the CoP gasifier cold gas efficiency is about 81 percent (HHV) and that 78.6 percent of the coal feed energy is contained in the syngas feed stream to the SOFC power island, and 8.3 percent of the coal feed energy is contained on the anode off-gas stream going to the heat recovery section of the power island. This diagram lists the key stream energy flows and temperatures, and lists the heat loads for major heat exchangers, the auxiliary power consumption and power generation outputs of major plant components. The SOFC operating voltage is 0.82 V. The cathode air preheat heat exchanger is very large with a heat load of about 48 percent of the coal feed energy input. The dominant auxiliary powers in the plant are the ASU at 3.9 percent of the coal energy, the CO_2 compression section at 3.7 percent, and the cathode air blower at 0.8 percent.

Steam balances for high-pressure and low-pressure steam are shown in Exhibit 3-13 and Exhibit 3-14. The high-pressure process steam feed for the CoP gasifier is generated at the oxy-combustor heat recovery step, and the high-pressure steam for power generation is generated from the raw syngas cooler and the oxy-combustor heat recovery section.

The IGFC power plant water balance is shown on Exhibit 3-15. The nearly complete recovery of water from the oxy-combustion CO_2 product stream results in water consumption in the IGFC plant being significantly lower than with other fossil fuel power plant technologies.

Carbon and sulfur balances are displayed in Exhibit 3-16 and Exhibit 3-17. Nearly complete carbon capture is achieved, with a 99.7 percent carbon removal from the raw syngas. Note that the CO_2 product stream contains about 1.7 mole% oxygen and this exceeds expected CO_2 product transportation and sequestration specifications for oxygen. If this is the case, low-temperature processing of the CO_2 stream can be conducted to separate oxygen, resulting in little plant performance or cost impact, but reducing the plant carbon capture by about 5 percentage-points. Likewise, nearly complete sulfur removal is achieved, with 99.99 percent sulfur removal from the coal.

The air emissions are listed in Exhibit 3-18. The IGFC plant acts as a nearly zero emission power plant, with the only significant emission being the small release of CO_2. This emissions performance is dictated, in part, by the need to protect the SOFC stack components from contamination.

Exhibit 3-8 Case 1-1 Block Flow Diagram

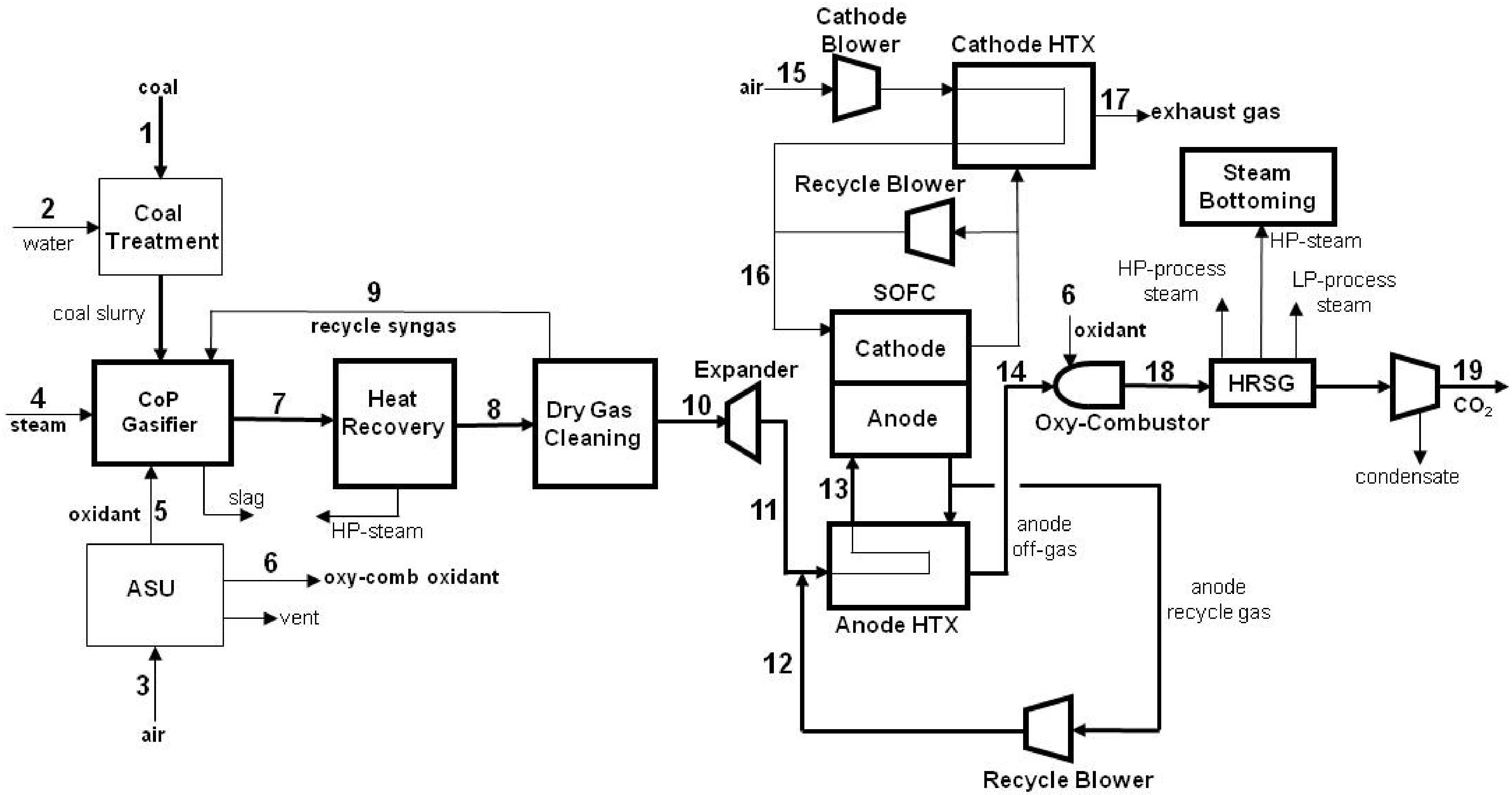


Exhibit 3-9 Case 1-1 Stream Table

	1	2	3	4	5	6	7	8	9	10	11
V-L Mole Fraction											
Ar	0.0000	0.0000	0.0094	0.0000	0.0031	0.0031	0.0006	0.0006	0.0008	0.0008	0.0008
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0443	0.0443	0.0578	0.0587	0.0587
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2889	0.2889	0.3768	0.3824	0.3824
CO_2	0.0000	0.0000	0.0003	0.0000	0.0000	0.0000	0.1535	0.1535	0.1997	0.2023	0.2023
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	0.0003	0.0003	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2667	0.2667	0.3479	0.3530	0.3530
H_2O	0.0000	1.0000	0.0104	1.0000	0.0000	0.0000	0.2315	0.2315	0.0026	0.0002	0.0002
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0008	0.0009	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0067	0.0067	0.0067	0.0000	0.0000
N_2	0.0000	0.0000	0.7722	0.0000	0.0019	0.0019	0.0021	0.0021	0.0027	0.0027	0.0027
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0045	0.0045	0.0038	0.0000	0.0000
O_2	0.0000	0.0000	0.2077	0.0000	0.9950	0.9950	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	0.0000	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)	0	4,156	24,013	3,290	3,891	895	23,156	23,156	2,657	14,875	14,875
V-L Flowrate (kg/hr)	0	74,876	692,881	59,272	124,579	28,668	479,421	479,421	57,035	317,962	317,962
Solids Flowrate (kg/hr)	182,263	0	0	0	0	0	3,535	177	0	0	0
Temperature (°C)	15	149	15	288	125	27	999	316	78	316	44
Pressure (MPa, abs)	0.10	3.45	0.10	3.45	3.45	0.16	3.10	3.00	4.14	2.32	0.14
Specific Enthalpy (kJ/kg)[A]	---	-15,353.3	-101.7	-13,035.8	88.9	1.1	-5,604.4	-6,884.6	-5,783.0	-5,463.0	-5,881.4
Density (kg/m^3)	---	917.6	1.2	14.7	33.2	2.0	6.0	12.7	30.4	10.1	1.1
V-L Molecular Weight	---	18.015	28.855	18.015	32.016	32.016	20.704	20.704	21.469	21.376	21.376
V-L Flowrate (lb_{mol}/hr)	0	9,163	52,939	7,254	8,578	1,974	51,051	51,051	5,857	32,793	32,793
V-L Flowrate (lb/hr)	0	165,073	1,527,542	130,673	274,651	63,201	1,056,942	1,056,942	125,740	700,987	700,987
Solids Flowrate (lb/hr)	401,822	0	0	0	0	0	7,793	390	0	0	0
Temperature (°F)	59	300	59	550	257	80	1,830	600	173	600	112
Pressure (psia)	14.7	500	14.7	500	500	23	450	435	600	337	20
Specific Enthalpy (Btu/lb)[A]	---	-6600.7	-43.7	-5604.4	38.2	0.5	-2409.4	-2959.8	-2486.3	-2348.7	-2528.6
Density (lb/ft^3)	---	57.281	0.076	0.916	2.075	0.127	0.377	0.790	1.897	0.628	0.070

A - Standard Reference State is the ideal vapor heat of formation at 298.15°K

Exhibit 3-11 Case 1-1 Stream Table (continue)

	12	13	14	15	16	17	18	19
V-L Mole Fraction								
Ar	0.0007	0.0008	0.0007	0.0094	0.0094	0.0098	0.0009	0.0015
CH_4	0.0000	0.0181	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0455	0.1495	0.0455	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.5303	0.4290	0.5303	0.0003	0.0003	0.0003	0.5698	0.9771
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0415	0.1377	0.0415	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.3796	0.2625	0.3796	0.0104	0.0104	0.0109	0.4168	0.0000
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0024	0.0025	0.0024	0.7722	0.7722	0.8069	0.0025	0.0042
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.2077	0.2077	0.1721	0.0100	0.0172
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	33,290	48,165	16,620	150,873	150,873	144,388	16,793	9,789
V-L Flowrate (kg/hr)	1,052,958	1,370,922	525,691	4,353,404	4,353,404	4,145,857	554,358	428,085
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0
Temperature (°C)	758	650	560	15	545	224	972	38
Pressure (MPa, abs)	0.11	0.11	0.11	0.10	0.11	0.11	0.10	15.27
Specific Enthalpy (kJ/kg)[A]	-8,686.3	-7,929.3	-8,976.6	-101.7	455.3	108.2	-8,569.7	-9,009.3
Density (kg/m^3)	0.4	0.4	0.5	1.2	0.5	0.7	0.3	676.0
V-L Molecular Weight	31.630	28.463	31.630	28.855	28.855	28.713	33.012	43.730
V-L Flowrate (lb_{mol}/hr)	73,392	106,185	36,641	332,619	332,619	318,321	37,021	21,582
V-L Flowrate (lb/hr)	2,321,377	3,022,367	1,158,950	9,597,622	9,597,622	9,140,057	1,222,151	943,767
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0
Temperature (°F)	1,397	1,203	1,040	59	1,013	435	1,782	100
Pressure (psia)	16.2	16.2	15.4	14.7	15.8	15.4	14.8	2215
Specific Enthalpy (Btu/lb)[A]	-3734.4	-3409.0	-3859.2	-43.7	195.8	46.5	-3684.3	-3873.3
Density (lb/ft^3)	0.026	0.026	0.030	0.076	0.029	0.046	0.020	42.202

Exhibit 3-10 Case 1-1 Plant Performance Summary (100 Percent Load)

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	
SOFC Power	551,342
Syngas Expander Power	36,197
Steam Turbine Power	112,866
TOTAL POWER, kWe	**700,405**
AUXILIARY LOAD SUMMARY, kWe	
Coal handling	396
Coal size reduction	1,876
Sour water recycle slurry pumps	157
Ash handling	964
ASU Auxiliary power	824
ASU air compressor	39,525
Oxygen compressor	12,877
Nitrogen compression	641
Anode recycle compressor	4,252
Claus Tail Gas Recycle compressor	1,008
CO_2 compressor	50,994
BFW pump	1,790
Condensate pump	120
Syngas recycle compressor	429
Quench water pump	446
Circulating water pump	1,979
Ground water pump	437
Cooling tower fans	1,381
Scrubber pumps	204
Selexol auxiliary power	2,803
ST auxiliaries	38
Cathode air blower	10,328
Cathode recycle blower	11,411
Claus / TGTU auxiliaries	164
BOP	2,727
Transformer losses	2,634
TOTAL AUXILIARIES, kWe	**150,405**
NET POWER, kWe	**550,000**
Net Plant Efficiency, % (HHV)	**40.0**
Net Plant Heat Rate, kJ/kWh (Btu/kWh)	**8,993 (8,523)**
CONDENSER COOLING DUTY 10^6 kJ/h (10^6 Btu/h)	**313 (197)**
CONSUMABLES	
As-Received Coal Feed, kg/h (lb/h)	182,263 (401,822)
Thermal Input[1], kWt	1,373,923
Raw Water Consumption, m^3/min (gpm)	6.4 (1,690)

Exhibit 3-11 Case 1-1 Mass Flow Diagram

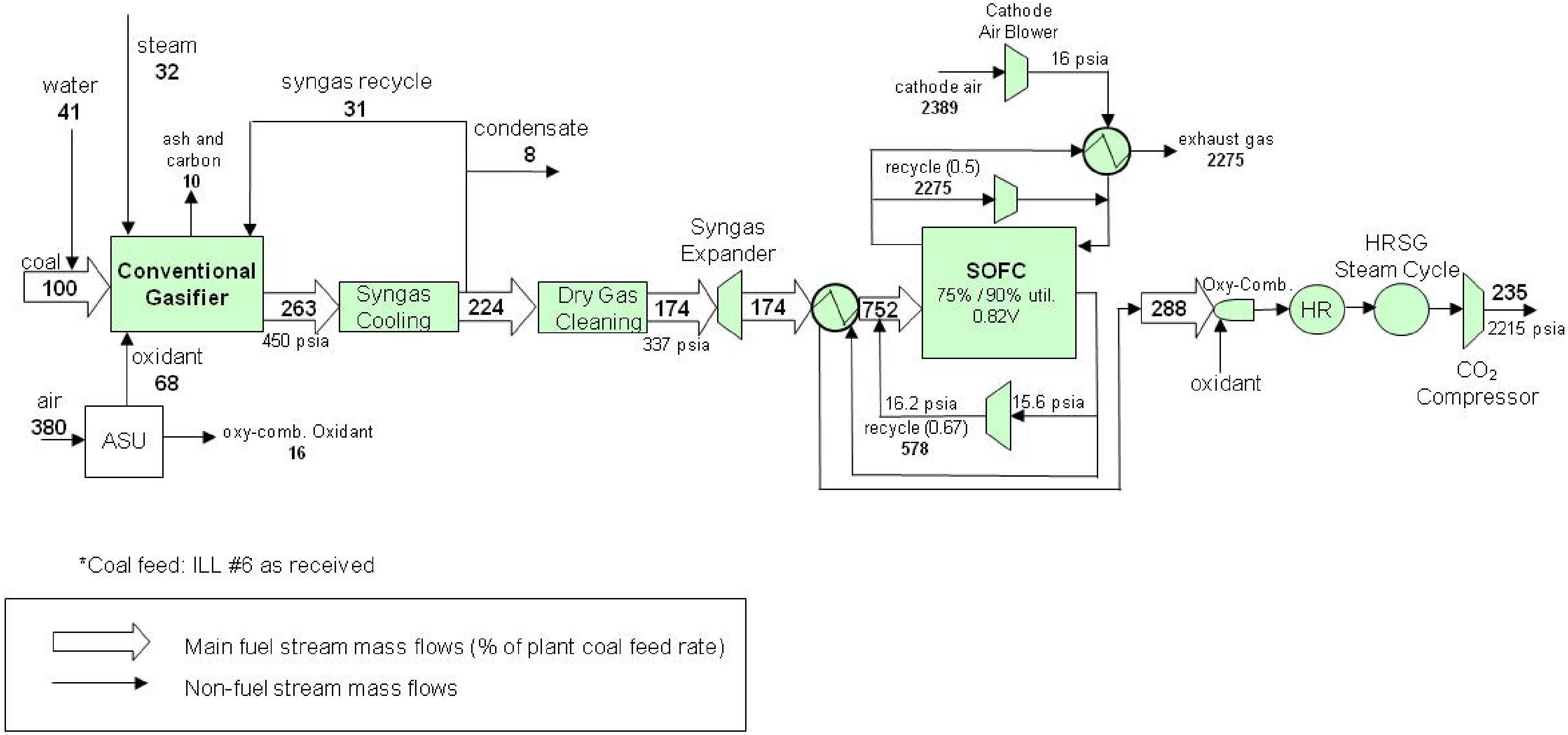


Exhibit 3-12 Case 1-1 Energy Flow Diagram

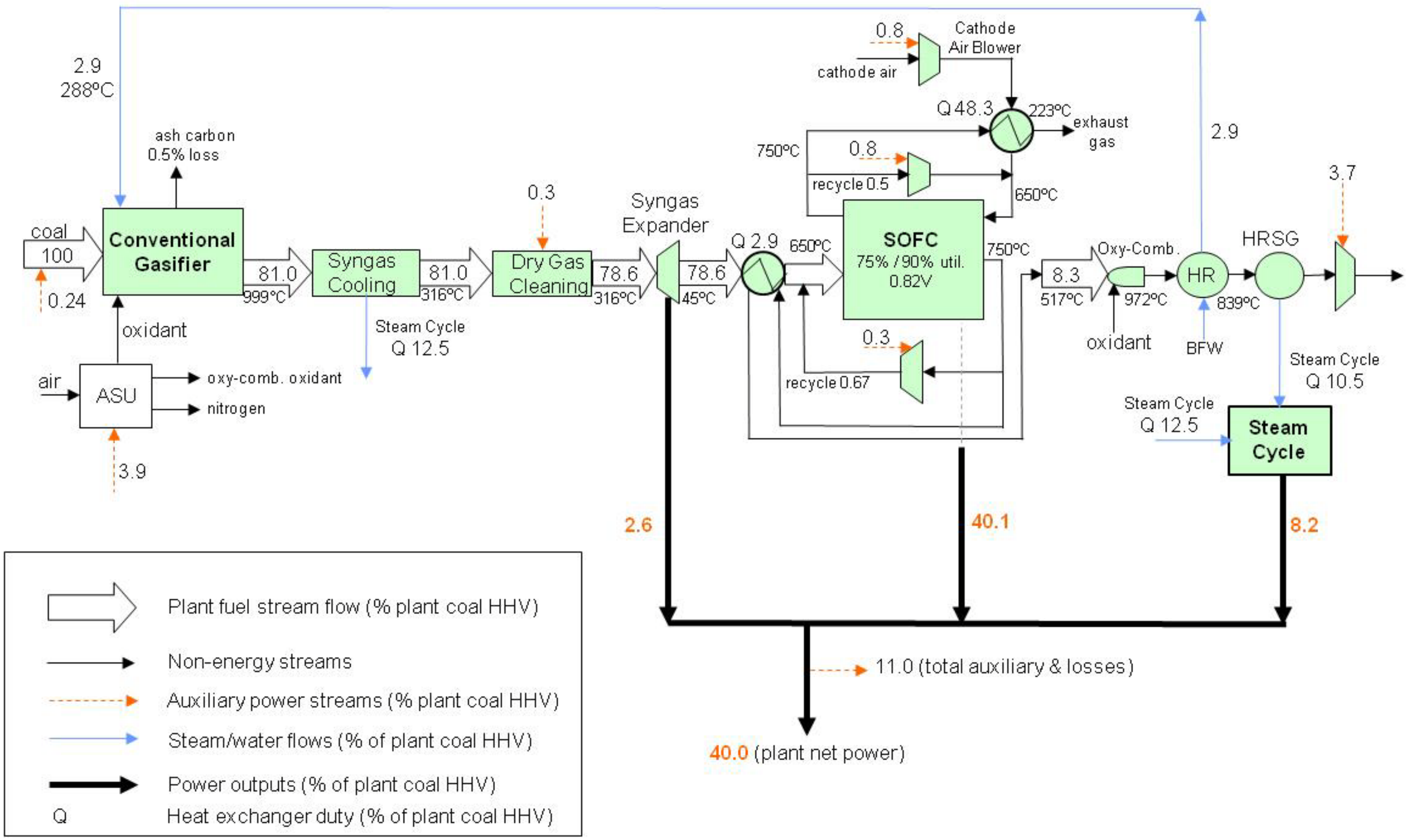


Exhibit 3-13 Case 1-1 High-Pressure Steam Balance

HP Process Steam Use, kg/h (lb/h)		HP Process Steam Generation, kg/h (lb/h)	
Gasifier feed	59,272 (130,673)	**Raw syngas cooling**	0 (0)
		Oxy-combustor heat recovery	59,272 (130,673)
Total	59,272 (130,673)	**Total**	59,272 (130,673)
HP Power-Steam generation, GJ/h (MMBtu/h)			
Slag cooling	30 (28)		
Raw syngas cooling	618(586)		
Oxy-combustor heat recovery	517 (490)		
Syngas reheat for polishing	-134 (-127)		
Total	1,030 (976)		

Exhibit 3-14 Case 1-1 Low-Pressure Steam Balance

LP Process Steam Use, GJ/h (MMBtu/h)		LP Process Steam Generation, GJ/h (MMBtu/h)	
Selexol stripping	114 (108)	**LT syngas cooling**	218 (207)
ASU	61 (58)	**Recycle syngas cooler**	101 (96)
Sour water stripping	132 (125)		
Syngas hydrolysis preheat	12 (12)		
Total	319 (303)	**Total**	319 (303)

Exhibit 3-15 Case 1-1 Water Balance

	m^3/min (gpm)
Water Demand	**7.40 (1,954)**
Slag Handling	0.40 (106)
Slurry Water	1.25 (330)
Condenser Makeup *Gasifier Steam* *BFW Makeup*	1.10 (290) *0.99 (261)* *0.11 (29)*
Cooling Tower Makeup	9.48 (2,505)
Water Recovery for Reuse	**3.73 (987)**
Low-temperature Cooling	1.63 (432)
CO_2 Dehydration	2.10 (555)
Process Discharge Water	**2.47 (652)**
Cooling Tower Water Blowdown	2.13 (564)
Low-temperature Cooling	0.15 (39)
CO_2 Dehydration	0.19 (50)
Raw Water Consumed	**6.40 (1,690)**

Exhibit 3-16 Case 1-1 Carbon Balance

Carbon In, kg/h (lb/h)		Carbon Out, kg/h (lb/h)	
Coal	116,182 (256,140)	**Slag**	929 (2,049)
		Exhaust Gas	368 (812)
		CO_2 Product	114,885 (253,279)
Total	116,182 (256,140)	**Total**	116,182 (256,140)

Exhibit 3-17 Case 1-1 Sulfur Balance

Sulfur In, kg/h (lb/h)		Sulfur Out, kg/h (lb/h)	
Coal	4,568 (10,071)	**Elemental Sulfur**	4,563 (10,059)
		Polishing Sorbent	5 (10)
		CO_2 Product	1 (2)
Total	4,568 (10,071)	**Total**	4,568 (10,071)

Exhibit 3-18 Case 1-1 Air Emissions

	kg/GJ ($lb/10^6$ Btu)	Tonne/year (tons/year) 80% capacity factor	kg/MWh (lb/MWh)
SO_2	0 (0)	0 (0)	0 (0)
NO_X	0 (0)	0 (0)	0 (0)
Particulate	0 (0)	0 (0)	0 (0)
Hg	0 (0)	0 (0)	0 (0)
CO_2	0.075 (0.17)	2,582 (2,846)	0.67 (1.48)

3.2.2 Case 1-1 Baseline Plant Cost Results

The capital cost estimate for Case 1-1 is broken down in Exhibit 3-19. Owner's costs are included in Exhibit -3-20. The dominant area costs are the gasification area and the SOFC power island. The first-year cost-of-electricity for Case 1-1 is displayed in Exhibit 3-21. The dominant contributor to the COE is capital recovery, with fuel cost being relatively small because of the high plant efficiency.

Exhibit 3-19 Case 1-1 Capital Cost Breakdown

	TOTAL PLANT COST	
Item/Description	**$ x 1000**	**$/kW**
COAL & SORBENT HANDLING	**32,152**	**58**
COAL & SORBENT PREP & FEED	**49,817**	**91**
FEEDWATER & MISC. BOP SYSTEMS	**14,372**	**26**
GASIFIER & ACCESSORIES	**433,714**	**789**
Gasifier & Syngas Cooler	236,855	431
ASU & Oxidant Compressor	179,515	326
Other Gasification Equipment	17,344	32
GAS CLEANUP & PIPING	**145,922**	**265**
Scrubber & Low Temperature Cooling	24,798	45
Single-Stage Selexol/MDEA	71,042	129
Claus Plant	29,604	54
Trace removal	2,058	4
COS Hydrolysis	7,739	14
Blowback, Piping, Foundations	3,891	7
Sulfur polishing	6,790	12
CO_2 DRYING & COMPRESSION	**54,688**	**99**
SOFC POWER ISLAND	**365,073**	**664**
Syngas expander	6,641	12
SOFC Stack Units (stack modules, enclosures, inverters)	258,523	470
Cathode Air Blower	4,550	8
Cathode Gas Recycle Blower	10,788	20
Cathode Heat Exchanger	60,756	110
Anode Heat Exchanger	10,845	20
Anode Recycle Blower	902	2
Oxy-Combustor	12,068	22
HRSG, DUCTING & STACK	**21,403**	**39**
STEAM POWER SYSTEM	**28,520**	**52**
COOLING WATER SYSTEM	**14,588**	**27**
ASH/SPENT SORBENT HANDLING SYS	**33,069**	**60**
ACCESSORY ELECTRIC PLANT	**75,743**	**138**
INSTRUMENTATION & CONTROL	**27,743**	**50**
IMPROVEMENTS TO SITE	**17,653**	**32**
BUILDING & STRUCTURES	**16,331**	**30**
TOTAL PLANT COST ($1000)	**1,345,703**	**2,447**

Note the cost of the gasifier and syngas cooler at $236,855,000. This represents the cost of two, parallel CoP gasifiers which are mechanically complex, two-stage, high-temperature, slagging pressure vessels having multiple coal and oxidant feed points and slag removal nozzles. Included in this cost are two, large tar cracking pressure vessels that directly follow the gasifiers, and a pair of convective heat exchangers for cooling the 999°C (1900°F) syngas to 316°C (600°F) under highly fouling conditions. The Gasifier & Accessories area has the greatest total cost at 789 $/kW.

The SOFC stack units (stacks, enclosures, and inverters) have a total cost of 470 $/kW, while the entire SOFC power island total cost is 664 $/kW. The SOFC stack units represent the single most expensive component in the baseline IGFC plant.

Exhibit -3-20 Case 1-1 Owner's Costs

Owner's Costs		
Preproduction Costs		
6 Months All Labor	10,846	20
1 Month Maintenance Materials	2,301	4
1 Month Non-fuel Consumables	305	1
1 Month Waste Disposal	275	1
25% of 1 Months Fuel Cost at 100% CF	1,400	3
2% of TPC	26,914	49
Total	**42,042**	**76**
Inventory Capital		
60 day supply of fuel and consumables at 100% CF	11,561	21
0.5% of TPC (spare parts)	6,729	12
Total	**18,289**	**33**
Initial Cost for Catalyst and Chemicals	5,378	10
Land	900	2
Other Owner's Costs	201,855	367
Financing Costs	36,334	66
Total Overnight Costs (TOC)	**1,650,502**	**3,001**
Total As-Spent Cost (TASC)	**1,881,572**	**3,421**

Exhibit 3-21 Case 1-1 Cost-of-Electricity Breakdown

			Annual Cost	Annual Unit Cost
OPERATING & MAINTENANCE LABOR			$	mills/kWh
Annual Operating Labor Cost			5,918,913	
Maintenance Labor Cost			11,435,340	
Administrative & Support Labor			4,338,563	
Property Taxes and Insurance			26,812,403	
TOTAL FIXED OPERATING COSTS			48,505,220	**12.6**
VARIABLE OPERATING COSTS				
Maintenance Material Cost			22,093,077	
Stack Replacement Cost			17,511,337	
Subtotal			**39,604,414**	
Consumables	Unit Cost	Initial Cost		
Water (/1000 gallons)	1.08	0	**1,200,317**	
Chemicals				
MU & WT Chem. (lbs)	0.17	0	332,321	
Carbon (Trace Removal) (lb)	1.05	503,902	201,561	
COS Catalyst (m3)	2,397	870,582	174,116	
Selexol Solution (gal)	13.40	3,290,593	151,435	
Claus / DSRP Catalyst (ft3)	131.27	64,049	475,112	
ZnO polishing sorbent (lb)	1.50	712,668	803,951	
Subtotal Chemicals		**5,377,745**	**1,727,433**	
Waste Disposal				
Spent Trace Catalyst (lb.)	0.42	0	87,060	
Ash (ton)	16.23	0	2,332,461	
Spent sorbents (lb)	0.42	0	225,106	
Subtotal-Waste Disposal ($)			**2,644,628**	
TOTAL VARIABLE OPERATING COSTS		**5,377,745**	**45,176,792**	**11.7**
Fuel Coal (ton)	38.18	0	**53,763,318**	**13.9**
Capital Recovery (mills/kWh)				**53.2**
TS&M (mills/kWh)				**5.1**
COE First Year (mills/kWh)				**96.5**

3.2.3 Scenario 1 Pathway Results

The Scenario 1 pathway performance and cost estimates were performed for progressions in 1) an IGFC plant with baseline SOFC conditions, using conventional coal gasifier technology (Case 1-1); 2) the cell performance degradation rate, improved from 1.5 percent /1000 hours to 0.2 percent /1000 hours; 3) the cell overpotential, reduced from 140 mV to 70 mV; 4) the plant capacity factor, increased from 80 percent to 85 percent; 5) the gasifier technology, improved from the E-Gas to an Enhanced technology with methane increased to 10.2 mole percent (dry); 6) the plant capacity factor, increased from 85 percent to 90 percent; 7) a branch point for natural gas injection into the clean syngas, this not being a cumulative enhancement, 8) the cost of the SOFC stack blocks reduced 20 percent, with the total SOFC cost reduced from 296 to 268 $ per kW of SOFC output; and 9) the SOFC DC-to-AC inverter efficiency increased from 97 to 98 percent.

The performance and cost results are tabulated in Exhibit 3-24. The total progression increases the plant efficiency to 46.0 percent (HHV), with the Case 1-6 branch point using natural gas injection increasing the plant efficiency to 51.0 percent (HHV). The progression reduces the COE to 72.5 mills/kWh, with the Case 1-6 branch point achieving 71.2 mills/kWh COE, for the assumed natural gas price of 6.55 $/MMBtu. There are corresponding reductions in the plant capital investment and the raw water consumption rate along the progression.

It is also of interest to observe some of the characteristics of the most expensive component systems in the plant, the gasifier, the SOFC stack units, and the SOFC power island. Exhibit 3-22 shows some key characteristics of the coal gasifier in Scenario 1. The exit volumetric flow and cost do not change to a large extent, except in Case 1-6 where natural gas injection is used.

Exhibit 3-22 Scenario 1 Conventional Coal Gasifier Characteristics

Case	Gasifier Coal Feed Rate, kg/hr (lb/hr)	Gasifier Exit Pressure, MPa (psia)	Gasifier Exit Temperature, °C (°F)	Gasifier Exit Syngas Rate, 1000 m^3/h (1000 ft^3/h)	Gasifier & Heat Recovery Cost ($1000)
1-1	182,264 (401,823)	3.10 (450)	999 (1830)	79.4 (2,805)	236,855
1-2	182,264 (401,823)	3.10 (450)	999 (1830)	79.4 (2,805)	236,855
1-3	166,990 (368,151)	3.10 (450)	999 (1830)	72.7 (2,570)	222,780
1-4	166,990 (368,151)	3.10 (450)	999 (1830)	72.7 (2,570)	222,780
1-5	158,481 (349.390)	3.10 (450)	945 (1733)	69.0 (2,439)	214,771
1-6	87,954 (193,905)	3.10 (450)	945 (1733)	38.3 (1,354)	115,521
1-7	158,481 (349.390)	3.10 (450)	945 (1733)	69.0 (2,439)	214,771
1-8	158,481 (349.390)	3.10 (450)	945 (1733)	69.0 (2,439)	214,771
1-9	156,885 (345,873)	3.10 (450)	945 (1733)	68.3 (2,415)	213,255

Exhibit 3-23 lists some of the SOFC characteristic along the Scenario 1 pathway. The SOFC current density generally decreases along the pathway, as the cell voltage remains near 0.88 V. The spare cell surface installed and the stack replacement times are the optimum values estimated. The SOFC stack unit cost and power island cost generally decrease along the pathway.

Exhibit 3-23 Scenario 1 SOFC Characteristics

Case	Cell Voltage V	Power Density mW DC/cm^2	Current Density mA/cm^2	Spare Cell Surface Installed %	Stack Replacement Time years	SOFC Stack Unit Cost $/kW	Power Island Cost $/kW
1-1	0.816	400	490	58.4	5.6	470	664
1-2	0.816	400	490	19.7	14.5	355	549
1-3	0.885	400	452	19.7	14.5	353	508
1-4	0.885	400	452	19.7	14.5	353	508
1-5	0.878	400	455	19.7	13.2	352	493
1-6	0.86	400	464	19.7	13.2	346	454
1-7	0.878	400	455	19.7	12.5	352	493
1-8	0.878	400	455	19.7	12.5	319	460
1-9	0.878	400	455	19.7	12.5	319	458

Exhibit 3-24 Scenario 1 Pathway Results

Case	Pathway Parameter	Change Made	Coal Feed Rate, kg/h (lb/h)	Number Parallel Trains	Cell Voltage V	Plant Efficiency %, HHV	Raw Water Consumed gpm/MW	CO_2 Emission kg/MWh	Capital Cost, TOC $/kW	COE mills/kWh	Cost of CO_2 Avoided $/tonne
1-1	**Baseline Atm-pressure**	Baseline	182,264 (401,823)	2	0.816	40.0	3.07	2.5	3,001	96.3	46.8
1-2	**Degradation**	1.5 to 0.2 %/1000 hours	182,264 (401,823)	2	0.816	40.0	3.07	2.5	2,844	89.5	38.3
1-3	**Cell Over-potential**	140 to 70 mV	166,990 (368,151)	2	0.885	43.7	2.82	2.3	2,666	84.5	32.0
1-4	**Capacity Factor**	80 to 85 %	166,990 (368,151)	2	0.885	43.7	2.82	2.3	2,666	80.5	27.0
1-5	**Enhanced Gasifier**	5.9 to 10.2 mole% CH_4	158,481 (349.390)	2	0.878	46.0	2.74	2.2	2,552	77.2	22.9
1-6	**Natural Gas Injection**	38.5% injection	87,954 (193,905)	1	0.86	51.0	2.05	1.3	1,794	71.2 @ $6.55/MM Btu natural gas	15.4
1-7	**Capacity Factor**	85 to 90 %	158,481 (349.390)	2	0.878	46.0	2.74	2.2	2,552	73.7	18.6
1-8	**SOFC Stack Cost**	296 to 268 $/kW	158,481 (349.390)	2	0.878	46.0	2.74	2.2	2,512	72.9	17.4
1-9	**Inverter Efficiency**	97 to 98 %	156,885 (345,873)	2	0.878	46.0	2.71	2.2	2,497	72.5	16.9

3.3 Scenario 2 - IGFC with Pressurized-SOFC

Scenario 2 applies the enhanced conventional coal gasifier, but considers a configuration for an IGFC plant using pressurized SOFC. Pressurized SOFC can be configured in two general, alternative arrangements:

1. The anode off-gas oxy-combustor is followed by hot gas expander power generation (expansion ratio about 18). A heat recovery steam generator (HRSG) produces steam for power generation, and the remaining, low-pressure, wet CO_2 stream is dried and compressed (compression ratio about 149).
2. The anode off-gas oxy-combustor is followed directly by a HRSG for steam bottoming power generation. The remaining, high-pressure, wet CO_2 stream is dried and compressed (compression ratio about 8.4).

Configuration 2 is expected to be the least complex and most effective approach and is utilized for this evaluation. Note that further optimization of the pressurized configuration and its operating conditions are recommended and could produce superior results to those presented here. All areas of the plant are identical to the Case 1 plant areas except for the power island and the CO_2 dehydration and compression area.

The Scenario 2 pressurized-SOFC Power Block assumptions and specifications are listing in Exhibit 3-25. Exhibit 3-26 lists the assumptions and basis for the CO_2 dehydration and compression section when applying pressurized SOFC.

Exhibit 3-25 Scenario 2 Pressurized Power Island Assumptions

	Specification/Assumptions
Syngas Expander	
Outlet pressure, MPa (psia)	2.0 (290)
Efficiency, adiabatic %	90
Generator efficiency, %	98.5
Fuel Cell System	
Cell stack inlet temperature, ºC (ºF)	650 (1202)
Cell stack outlet temperature, ºC (ºF)	750 (1382)
Cell stack outlet pressure, MPa (psia)	1.97 (285)
Fuel single-step utilization, %	75
Fuel overall utilization, %	90
Stack anode-side pressure drop, MPa (psi)	0.014 (2)
Stack cathode-side pressure drop, MPa (psi)	0.014 (2)
Power density, mW/cm^2	500
Stack over-potential, mV	70
Operating voltage estimation method	Section 8.1.4
Cell degradation rate (% per 1000 hours)	0.2
Cell replacement period (% degraded)	20
Fuel Cell Ancillary Components	
Anode gas recycle method	Syngas jet pump [22]
Syngas motive gas rate	3% of circulation rate
Anode heat exchanger pressure drop, MPa	0.02 (3)
Cathode recycle gas rate, %	0
Cathode heat exchanger pressure drop, MPa	0.02 (3)
Cathode compressor efficiency, adiabatic %	90
Rectifier DC-to-AC efficiency, %	97.0
Other electric motor drives efficiency, %	95
Transformer efficiency, %	99.65

Exhibit 3-26 Scenario 2 CO_2 Dehydration and Compression Section Assumptions

	Specification/Assumptions
CO_2 Dehydration	
technology	Water cooling & Glycol column
CO_2 Compressor	
number of compression stages	2
efficiency, adiabatic %	80
electric motor drive efficiency, %	95

3.3.1 Case 2-1 IGFC Plant Performance Results

The following information is presented in tabular form for Case 2-1:

- Block Flow Diagrams and Stream Table
- Performance Summary
- Mass and Energy Flow Diagrams
- Steam Balance
- Water Balance
- Carbon Balance
- Sulfur Balance
- Air Emissions.

The system description follows the BFD in Exhibit 3-27 and stream numbers reference the same Exhibit. Exhibit 3-28 provides process data for the numbered streams in the BFD. Note that 66.7 percent of the anode off-gas is recycled to the anode inlet stream, reducing the syngas methane content of 10.91 mole percent to 3.18 mole percent in the actual anode inlet stream.

Exhibit 3-29 provides the power plant breakdown and overall thermal performance. Note that the steam turbine power represents only about 13 percent of the total plant power generated, with the SOFC system being the overwhelmingly dominate power generator. The Scenario 2, Case 2-1 plant efficiency of 50.1 percent (HHV) is extremely high for a power plant with carbon removal compared to other fossil fuel power plant technologies.

Mass flow and energy flows diagrams are presented in Exhibit 3-30 and Exhibit 3-31 on a basis relative to the coal as-received mass feed rate, and relative to the coal feed energy (HHV), respectively. The mass flow diagram indicates that the mass of the CO_2 product stream is 2.35 times the mass of the coal feed stream, and the largest mass flows in the plant are associated with the cathode gas streams, these being as large as almost 27 times the coal feed flow. In this case, the oxidant flow to the oxy-combustor is about 26 percent of the oxidant flow to the coal gasifier.

The energy flow diagram indicates that the enhanced conventional gasifier cold gas efficiency is about 82.5 percent (HHV) and that 80.3 percent of the coal feed energy is contained in the syngas feed stream to the SOFC power island, and 8.8 percent of the coal feed energy is contained on the anode off-gas stream going to the heat recovery section of the power island. This diagram lists the key stream energy flows and temperatures, and lists the heat loads for major heat exchangers, the auxiliary power consumption and power generation outputs of major plant components.

The SOFC operating voltage is 0.94 V, much of the increase in this voltage being due to the increased pressure of the SOFC system. The cathode air preheat heat exchanger in Case 2-1 is not as large as in Case 1-1, with a heat load of about 24 percent of the coal feed energy input, because the compression of the cathode air partially preheats the stream. The dominant auxiliary powers in the plant are the ASU at 5.2 percent of the coal energy, the cathode air compressor-expander at 3.4 percent, and the CO_2 compression area at 1.4 percent. The ASU auxiliary power is increased relative to Case 1-1 because the oxy-combustion oxidant stream must be compressed to the pressurized condition of the anode off-gas. The CO_2 compression area auxiliary power is relatively small because the oxy-combustor off-gas is at high pressure.

Steam balances for high-pressure and low-pressure steam are shown in Exhibit 3-32 and Exhibit 3-33. The high-pressure process steam feed for the enhanced gasifier is generated at the oxy-combustor heat recovery step, and the high-pressure steam for power generation is generated from the raw syngas cooler and the oxy-combustor heat recovery section.

The IGFC power plant water balance is shown on Exhibit 3-34. As in Case 1-1, the nearly complete recovery of water from the oxy-combustion CO_2 product stream in Case 2-1 results in water consumption in the IGFC plant being significantly lower than with other fossil fuel power plant technologies.

Carbon and sulfur balances are displayed in Exhibit 3-35 and Exhibit 3-36. Again, nearly complete carbon capture is achieved, with a 99.7 percent carbon removal from the raw syngas. Note that the CO_2 product stream contains about 1.7 mole percent oxygen and this may exceed CO_2 product transportation and sequestration specifications for oxygen once these are established. If this is the case, low-temperature processing of the CO_2 stream can be conducted to separate oxygen, resulting in little plant performance of cost impact, but reducing the plant carbon capture by about 5 percentage-points. Likewise, nearly complete sulfur removal is achieved, with 99.985 percent sulfur removal from the coal.

The air emissions are listed in Exhibit 3-37. The IGFC plant acts as a nearly zero emission power plant, with the only significant emission being the small release of CO_2. This emissions performance is dictated, in part, by the need to protect the SOFC stack components from contamination.

Exhibit 3-27 Case 2-1 Block Flow Diagram

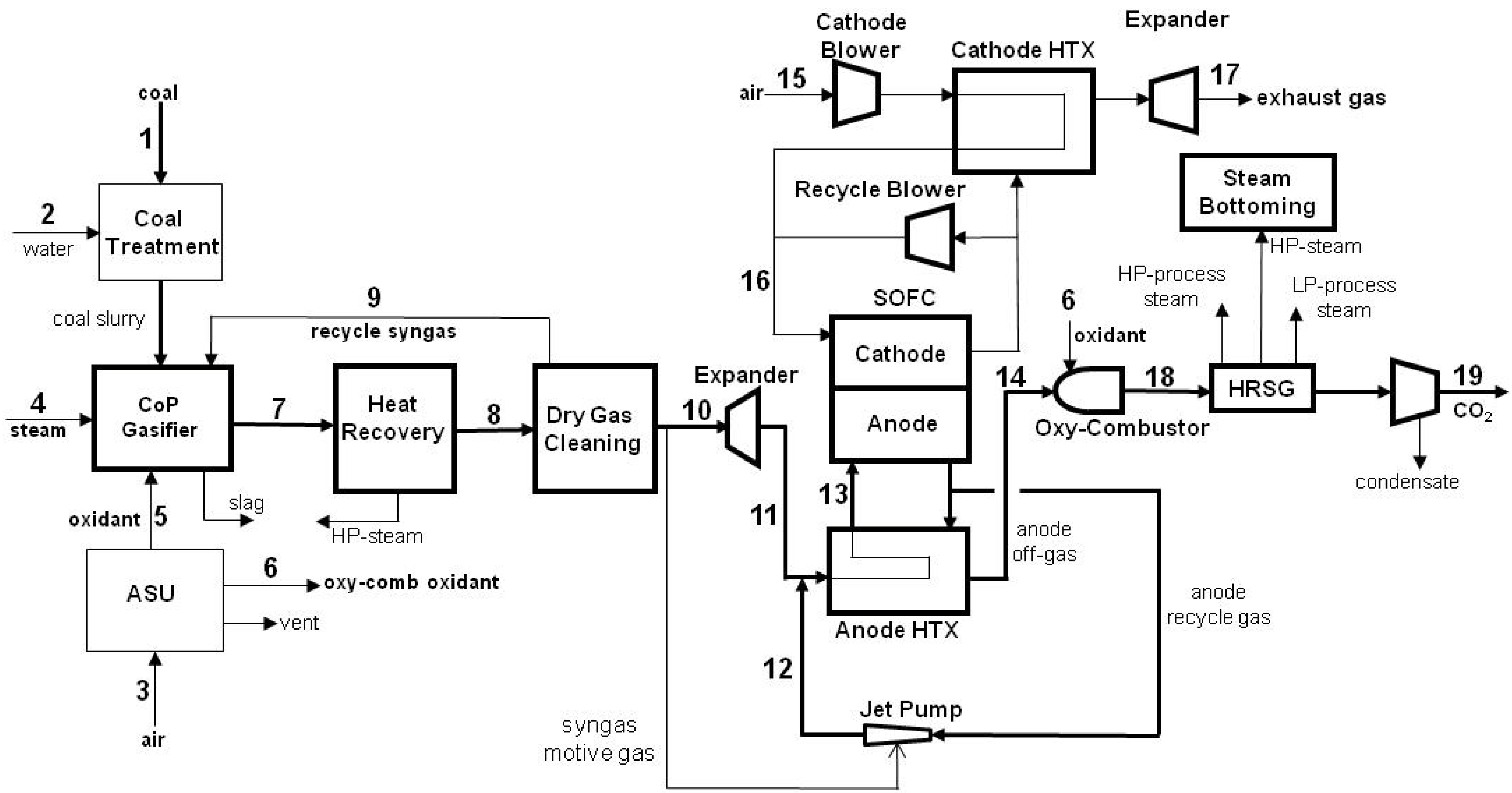


Exhibit 3-28 Case 2-1 Stream Table

	1	2	3	4	5	6	7	8	9	10	11
V-L Mole Fraction											
Ar	0.0000	0.0000	0.0094	0.0000	0.0031	0.0031	0.0006	0.0006	0.0008	0.0008	0.0008
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0827	0.0827	0.1078	0.1091	0.1091
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2627	0.2627	0.3425	0.3468	0.3468
CO_2	0.0000	0.0000	0.0003	0.0000	0.0000	0.0000	0.1744	0.1744	0.2263	0.2285	0.2285
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	0.0003	0.0002	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2359	0.2359	0.3075	0.3114	0.3114
H_2O	0.0000	1.0000	0.0104	1.0000	0.0000	0.0000	0.2287	0.2287	0.0016	0.0001	0.0001
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0008	0.0009	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0070	0.0070	0.0062	0.0000	0.0000
N_2	0.0000	0.0000	0.7722	0.0000	0.0019	0.0019	0.0025	0.0025	0.0033	0.0033	0.0033
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0040	0.0040	0.0028	0.0000	0.0000
O_2	0.0000	0.0000	0.2077	0.0000	0.9950	0.9950	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	0.0000	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)	0	3,322	17,696	2,630	2,773	750	17,391	17,391	2,040	9,478	9,478
V-L Flowrate (kg/hr)	0	59,843	510,614	47,372	88,771	24,021	372,396	372,396	45,608	211,122	211,122
Solids Flowrate (kg/hr)	145,671	0	0	0	0	0	2,825	141	0	0	0
Temperature (°C)	15	149	15	271	157	136	945	316	69	316	239
Pressure (MPa, abs)	0.10	5.52	0.10	5.52	5.31	1.97	4.83	4.70	6.00	3.87	2.00
Specific Enthalpy (kJ/kg)[A]	---	-15,353.3	-101.7	-13,189.6	117.7	100.9	-5,808.4	-6,999.4	-6,025.0	-5,685.4	-5,807.9
Density (kg/m^3)	---	917.6	1.2	26.7	47.2	18.5	10.1	20.5	47.5	17.4	10.4
V-L Molecular Weight	---	18.015	28.855	18.015	32.016	32.016	21.413	21.413	22.361	22.274	22.274
V-L Flowrate (lb_{mol}/hr)	0	7,323	39,013	5,797	6,113	1,654	38,341	38,341	4,497	20,896	20,896
V-L Flowrate (lb/hr)	0	131,932	1,125,713	104,438	195,707	52,957	820,994	820,994	100,549	465,444	465,444
Solids Flowrate (lb/hr)	321,149	0	0	0	0	0	6,228	311	0	0	0
Temperature (°F)	59	300	59	520	314	276	1,733	600	156	600	462
Pressure (psia)	14.7	800	14.7	800	770	285	700	681	870	562	290
Specific Enthalpy (Btu/lb)[A]	---	-6600.7	-43.7	-5670.5	50.6	43.4	-2497.2	-3009.2	-2590.3	-2444.3	-2496.9
Density (lb/ft^3)	---	57.281	0.076	1.664	2.944	1.153	0.630	1.279	2.966	1.086	0.649

A - Standard Reference State is the ideal vapor heat of formation at 298.15°K

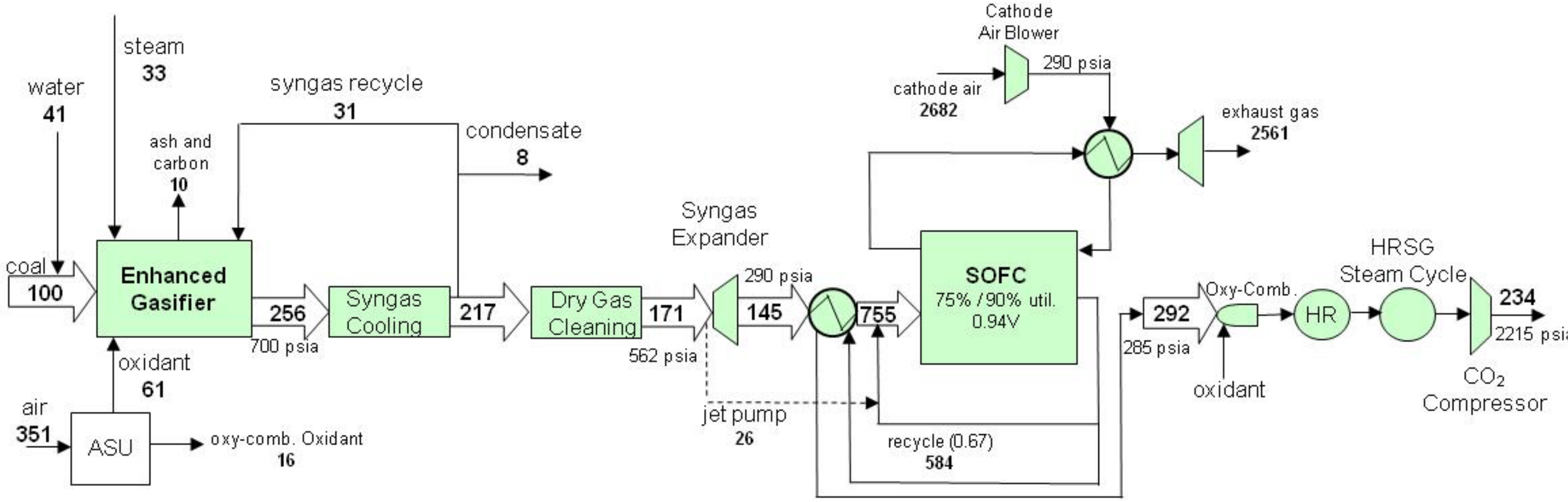
Exhibit 3-29 Case 2-1 Stream Table (continued)

	12	13	14	15	16	17	18	19
V-L Mole Fraction								
Ar	0.0006	0.0007	0.0006	0.0094	0.0094	0.0098	0.0008	0.0014
CH_4	0.0064	0.0318	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0631	0.1331	0.0455	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.4995	0.4326	0.5163	0.0003	0.0003	0.0003	0.5560	0.9761
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0593	0.1216	0.0438	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.3682	0.2774	0.3910	0.0104	0.0104	0.0108	0.4304	0.0000
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0027	0.0029	0.0027	0.7722	0.7722	0.8049	0.0028	0.0049
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.2077	0.2077	0.1741	0.0100	0.0176
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	28,917	38,395	13,596	135,382	135,382	129,882	13,739	7,785
V-L Flowrate (kg/hr)	888,040	1,099,162	424,629	3,906,400	3,906,400	3,730,398	448,650	340,315
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0
Temperature (°C)	729	650	682	15	650	132	1,131	38
Pressure (MPa, abs)	1.99	1.98	1.94	0.10	1.98	0.11	1.89	15.27
Specific Enthalpy (kJ/kg)^A	-8,598.3	-8,023.0	-8,828.3	-101.7	574.1	12.9	-8,350.4	-9,003.1
Density (kg/m^3)	7.3	7.4	7.6	1.2	7.4	0.9	5.3	674.8
V-L Molecular Weight	30.710	28.628	31.232	28.855	28.855	28.722	32.654	43.714
V-L Flowrate (lb_{mol}/hr)	63,751	84,646	29,974	298,466	298,466	286,340	30,290	17,163
V-L Flowrate (lb/hr)	1,957,795	2,423,239	936,148	8,612,144	8,612,144	8,224,127	989,106	750,266
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0
Temperature (°F)	1,344	1,202	1,260	59	1,202	269	2,067	100
Pressure (psia)	289	287	282	14.7	287	15.5	274	2215
Specific Enthalpy (Btu/lb)^A	-3696.6	-3449.3	-3795.5	-43.7	246.8	5.6	-3590.0	-3870.6
Density (lb/ft^3)	0.457	0.459	0.476	0.076	0.461	0.057	0.329	42.124

Exhibit 3-29 Case 2-1 Plant Performance Summary (100 Percent Load)

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	
SOFC Power	541,858
Syngas Expander Power	7,037
Steam Turbine Power	129,805
TOTAL POWER, kWe	**678,699**
AUXILIARY LOAD SUMMARY, kWe	
Coal handling	295
Coal size reduction	1,499
Sour water recycle slurry pumps	125
Ash handling	770
ASU Auxiliary power	607
ASU air compressor	43,744
Oxygen compressor	12,558
Nitrogen compression	512
Claus Tail Gas Recycle compressor	806
CO2 compressor	15,462
BFW pump	2,059
Condensate pump	138
Syngas recycle compressor	343
Quench water pump	356
Circulating water pump	2,276
Ground water pump	349
Cooling tower fans	1,058
Scrubber pumps	158
Selexol auxiliary power	2,240
ST auxiliaries	43
Cathode air compressor	37,058
Claus / TGTU auxiliaries	131
Miscellaneous Balance of Plant	2,643
Transformer losses	2,552
TOTAL AUXILIARIES, kWe	**128,699**
NET POWER, kWe	**550,000**
Net Plant Efficiency, % (HHV)	**50.1**
Net Plant Heat Rate, kJ/kWh (Btu/kWh)	**7,187 (6,812)**
CONDENSER COOLING DUTY 10^6 kJ/h (10^6 Btu/h)	**404 (3830**
CONSUMABLES	
As-Received Coal Feed, kg/h (lb/h)	145,671 [illegible]
Thermal Input[1], kWt	1,098,082
Raw Water Consumption, m^3/min (gpm)	4.5 (1,197)

Exhibit 3-30 Case 2-1 Mass Flow Diagram



*Coal feed: ILL #6 as received

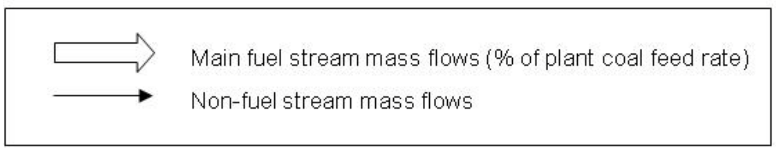


Exhibit 3-31 Cases 2-1 Energy Flow Diagram

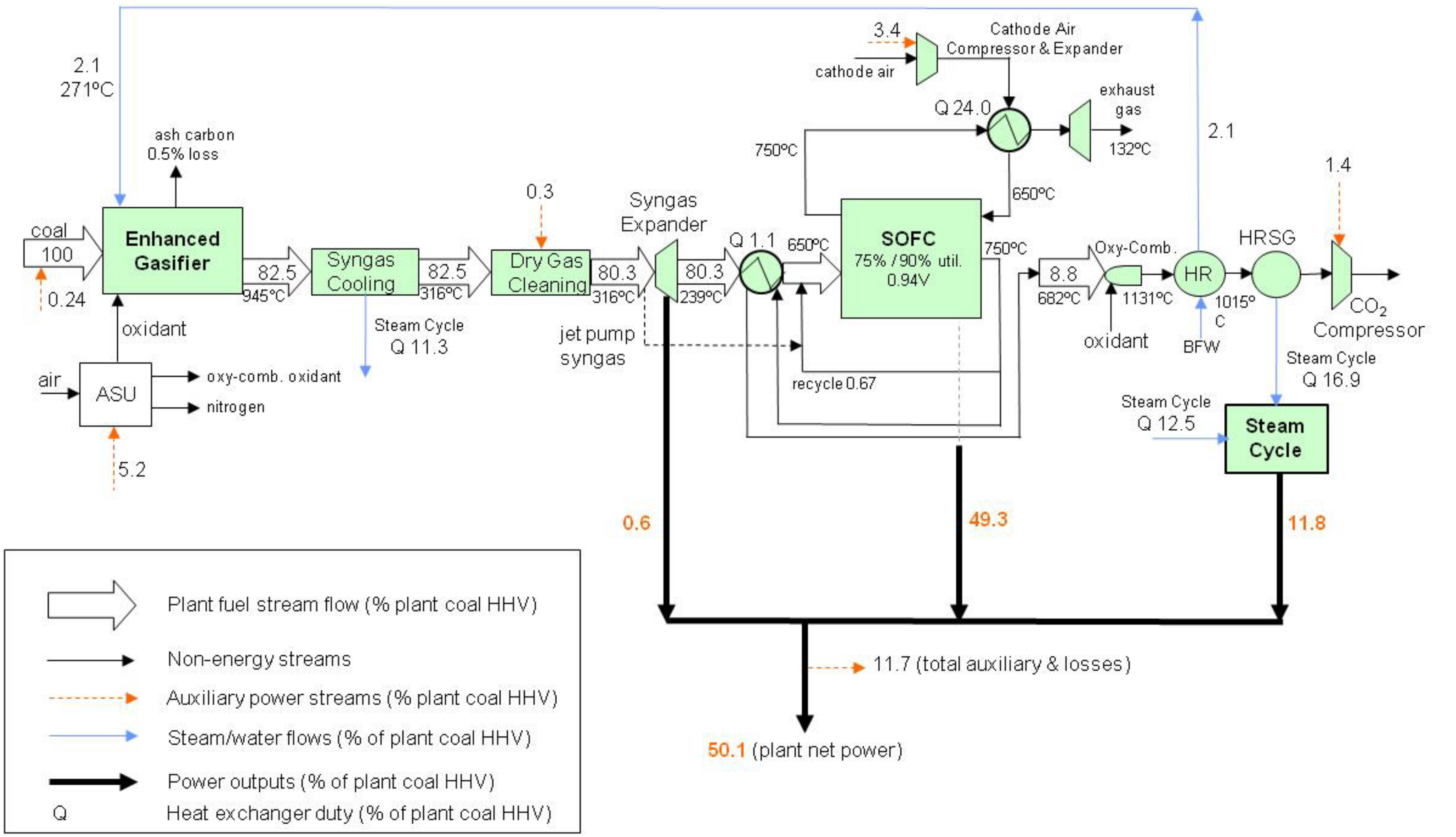


Exhibit 3-32 Case 2-1 High-Pressure Steam Balance

HP Process Steam Use, kg/h (lb/h)		HP Process Steam Generation, kg/h (lb/h)	
Gasifier feed	35,824 (104,438)	**Raw syngas cooling**	0 (0)
		Oxy-combustor heat recovery	35,824 (104,438)
Total	35,824 (104,438)	**Total**	35,824 (104,438)
HP Power-Steam generation, GJ/h (MMBtu/h)			
Slag cooling	24 (22)		
Raw syngas cooling	494 (468)		
Oxy-combustor heat recovery	667 (632)		
Syngas reheat for polishing	-107 (-102)		
Total	1,127 (1,068)		

Exhibit 3-33 Case 2-1 Low-Pressure Steam Balance

LP Process Steam Use, GJ/h (MMBtu/h)		LP Process Steam Generation, GJ/h (MMBtu/h)	
Selexol stripping	91 (86)	**LT syngas cooling**	160 (152)
ASU	45 (43)	**Recycle syngas cooler**	88 (83)
Sour water stripping	102 (97)		
Syngas hydrolysis preheat	10 (9)		
Total	248 (235)	**Total**	248 (235)

Exhibit 3-34 Case 2-1 Water Balance

	m^3/min (gpm)
Water Demand	**9.49 (2,507)**
Slag Handling	0.32 (85)
Slurry Water	1.00 (264)
Condenser Makeup *Gasifier Steam* *BFW Makeup*	0.91 (241) *0.79 (209)* *0.12 (32)*
Cooling Tower Makeup	7.26 (1,918)
Water Recovery for Reuse	**2.73 (721)**
Low-temperature Cooling	1.11 (292)
CO_2 Dehydration	1.62 (429)
Process Discharge Water	**1.91 (504)**
Cooling Tower Water Blowdown	1.63 (432)
Low-temperature Cooling	0.11 (29)
CO_2 Dehydration	0.16 (43)
Raw Water Consumed	**4.85 (1,282)**

Exhibit 3-35 Case 2-1 Carbon Balance

Carbon In, kg/h (lb/h)		Carbon Out, kg/h (lb/h)	
Coal	92,857 (204,715)	**Slag**	743 (1,638)
		Exhaust Gas	847 (1,869)
		CO_2 Product	91,267 (201,208)
Total	92,857 (204,715)	**Total**	92,857 (204,715)

Exhibit 3-36 Case 2-1 Sulfur Balance

Sulfur In, kg/h (lb/h)		Sulfur Out, kg/h (lb/h)	
Coal	3,652 (8,049)	**Elemental Sulfur**	3,647 (8,040)
		Polishing Sorbent	4 (8)
		CO_2 Product	1 (1)
Total	3,652 (8,049)	**Total**	3,652 (8,049)

Exhibit 3-37 Case 2-1 Air Emissions

	kg/GJ ($lb/10^6$ Btu)	Tonne/year (tons/year) 80% capacity factor	kg/MWh (lb/MWh)
SO_2	0 (0)	0 (0)	0 (0)
NO_X	0 (0)	0 (0)	0 (0)
Particulate	0 (0)	0 (0)	0 (0)
Hg	0 (0)	0 (0)	0 (0)
CO_2	0.214 (0.499)	5,942 (6,550)	1.54 (3.40)

3.3.2 Case 2-1 IGFC Plant Cost Results

The capital cost estimate for Case 2-1 is broken down in Exhibit 3-38. Owner's costs are included in Exhibit 3-39. The dominant area costs are the gasification area and the SOFC power island. The single highest cost component in the plant is the SOFC stacks and inverters. The first-year cost-of-electricity for Case 2-1 is displayed in Exhibit 3-40. The dominant contributor to the COE is capital recovery, with fuel cost being relatively small because of the high plant efficiency.

Exhibit 3-38 Case 2-1 Capital Cost Breakdown

	TOTAL PLANT COST	
Item/Description	$ x 1000	$/kW
COAL & SORBENT HANDLING	**27,484**	**50**
COAL & SORBENT PREP & FEED	**34,589**	**63**
FEEDWATER & MISC. BOP SYSTEMS	**10,746**	**20**
GASIFIER & ACCESSORIES	**294,256**	**535**
Gasifier & Syngas Cooler	164,454	299
ASU & Oxidant Compressor	117,759	214
Other Gasification Equipment	12,043	22
GAS CLEANUP & PIPING	**99,033**	**180**
Scrubber & Low Temperature Cooling	16,833	31
Single-Stage Selexol/MDEA	48,577	88
Claus Plant	20,554	37
Trace removal	1,407	3
COS Hydrolysis	5,253	10
Blowback, Piping, Foundations	2,647	5
Sulfur polishing	3,762	7
CO_2 DRYING & COMPRESSION	**21,758**	**40**
SOFC POWER ISLAND	**380,711**	**692**
Syngas expander	1,872	3
SOFC Stack Units (stack modules, enclosures, inverters)	279,837	509
Cathode Air Compressor	51,406	93
Cathode Heat Exchanger	15,999	29
Cathode Gas Expander	22,135	40
Anode Heat Exchanger	48	0
Anode Recycle Jet Pump	185	0
Oxy-Combustor	9,229	17
HRSG, DUCTING & STACK	**23,349**	**42**
STEAM POWER SYSTEM	**31,453**	**57**
COOLING WATER SYSTEM	**14,588**	**27**
ASH/SPENT SORBENT HANDLING SYS	**19,634**	**36**
ACCESSORY ELECTRIC PLANT	**67,915**	**123**
INSTRUMENTATION & CONTROL	**27,743**	**50**
IMPROVEMENTS TO SITE	**15,090**	**27**
BUILDING & STRUCTURES	**13,960**	**25**
TOTAL PLANT COST ($1000)	**1,084,926**	**1,973**

Exhibit 3-39 Case 2-1 Owner's Costs

Owner's Costs		
Preproduction Costs		
6 Months All Labor	10,846	20
1 Month Maintenance Materials	2,166	4
1 Month Non-fuel Consumables	228	0
1 Month Waste Disposal	220	0
25% of 1 Months Fuel Cost at 100% CF	1,119	2
2% of TPC	21,699	39
Total	**36,278**	**66**
Inventory Capital		
60 day supply of fuel and consumables at 100% CF	9,235	17
0.5% of TPC (spare parts)	5,425	10
Total	**14,660**	**27**
Initial Cost for Catalyst and Chemicals	4,055	7
Land	900	2
Other Owner's Costs	162,739	296
Financing Costs	29,293	53
Total Overnight Costs (TOC)	**1,332,851**	**2,423**
Total As-Spent Cost (TASC)	**1,519,450**	**2,763**

Exhibit 3-40 Case 2-1 Cost-of-Electricity Breakdown

					Annual Cost	Annual Unit Cost
OPERATING & MAINTENANCE LABOR					$	mills/kWh
Annual Operating Labor Cost					5,918,913	
Maintenance Labor Cost					11,435,340	
Administrative & Support Labor					4,338,563	
Property Taxes and Insurance					21,652,161	
TOTAL FIXED OPERATING COSTS					43,344,977	**10.6**
VARIABLE OPERATING COSTS						
Maintenance Material Cost					22,093,077	
Stack Replacement Cost					4,572,498	
Subtotal					**26,665,575**	
Consumables	Consumption		Unit	Initial		
	Initial Fill	/Day	Cost	Cost		
Water (/1000 gallons)	0	2,629	1.08	0	**882,577**	
Chemicals						
MU & WT Chem. (lbs)	0	4,835	0.17	0	259,623	
Carbon (Trace Removal) (lb)	383,494	525	1.05	402,734	171,162	
COS Catalyst (m3)	290	0.20	2,397	695,796	147,857	
Selexol Solution (gal)	196,290	30.94	13.40	2,629,944	128,596	
Claus / DSRP Catalyst (ft3)	0	1.34	131.27	54,389	217,550	
ZnO polishing sorbent (lb)	217,550	1,467	1.50	326,325	682,702	
Subtotal Chemicals				**4,054,799**	**1,444,329**	
Waste Disposal						
Spent Trace Catalyst (lb.)	0	567	0.42	0	73,930	
Ash (ton)	0	393	16.23	0	1,980,686	
Spent sorbents (lb)	0	1,467	0.42	0	191,156	
Subtotal-Waste Disposal ($)					**2,245,773**	
TOTAL VARIABLE OPERATING COSTS				**4,054,799**	**31,238,253**	**7.6**
Fuel Coal (ton)	0	2,914	38.18	0	**45,654,894**	**11.1**
Capital Recovery (mills/kWh)						**40.4**
TS&M (mills/kWh)						**4.1**
COE First Year (mills/kWh)						**73.9**

3.3.3 Scenario 2 Pathway Results

Scenario 2 pathway performance and cost estimates were performed for progressions in 1) the SOFC power island converted from the Case 1-5 condition to a pressurized-SOFC condition, 2) the plant capacity factor, increased from 85 percent to 90 percent; 3) the cell stack cost reduced by 20 percent, with the total cell cost (stacks and enclosures) dropping from 442 to 414 $/kW of SOFC power; and 4) the DC to AC inverter efficiency increased from 97 percent to 98 percent . The results are tabulated in Exhibit 3-43.

The progression increases the plant efficiency to 50.1 percent (HHV), with the COE reduced to 69.9 mills/kWh. There are corresponding reductions in the plant capital investment, and the raw water consumption rate is maintained at about 2.2 gpm/MW. Compared to the Scenario 1 pathway results, the benefits of pressurized SOFC in the selected configuration of Scenario 2 does provide some performance benefit, but the cost benefit is limited because of the high capital investment associated with pressurizing the SOFC system.

It is also of interest to observe some of the characteristics of the most expensive component systems in the plant, the gasifier, the SOFC stack units, and the SOFC power island. Exhibit 3-41 shows some key characteristics of the coal gasifier in Pathway 2. The exit volumetric flow and cost do not change significantly along the pathway.

Exhibit 3-41 Scenario 2 Conventional Coal Gasifier Characteristics

Case	Gasifier Coal Feed Rate kg/hr (lb/hr)	Gasifier Exit Pressure MPa (psia)	Gasifier Exit Temperature °C (°F)	Gasifier Exit Syngas Rate 1000 m^3/h (1000 ft^3/h)	Gasifier & Heat Recovery Cost $1000
2-1	146,735 (324,386)	4.83 (700)	945 (1733)	37.2 (1,316)	165,612
2-2	146,735 (324,386)	4.83 (700)	945 (1733)	37.2 (1,316)	165,612
2-3	146,735 (324,386)	4.83 (700)	945 (1733)	37.2 (1,316)	165,612
2-4	145,671 (321,149)	4.83 (700)	945 (1733)	37.2 (1,316)	164,454

Exhibit 3-42 lists some of the SOFC characteristic along Pathway 2. The SOFC current density remains fixed along the pathway, as the cell voltage remains at 0.937 V. The spare cell surface installed and the stack replacement times are the optimum values estimated. The SOFC stack unit cost and power island cost decrease for the Case 2-3 SOFC stack cost reduction.

Exhibit 3-42 Scenario 2 Pressurized SOFC Characteristics

Case	Cell Voltage V	Power Density mW DC/cm^2	Current Density mA/cm^2	Spare Cell Surface Installed %	Stack Replacement Time years	SOFC Stack Unit Cost $/kW	Power Island Cost $/kW
2-1	0.937	500	534	16.8	11.3	509	694
2-2	0.937	500	534	16.8	10.7	509	694
2-3	0.937	500	534	16.8	10.7	477	662
2-4	0.937	500	534	16.8	10.7	477	662

Exhibit 3-43 Scenario 2 Pathway Results

Case	Pathway Parameter	Change Made	Coal Feed Rate kg/h (lb/h)	Number Parallel Trains	Cell Voltage V	Plant Efficiency %, HHV	Raw Water Consumed gpm/MW	CO_2 Emission kg/MWh	Capital Cost TOC $/kW	COE mills/kWh	Cost of CO_2 Avoided $/tonne
2-1	**Pressurized SOFC**	15.6 to 285 psia	146,735 (324,386)	1	0.937	49.6	2.20	5.7	2,436	74.2	19.3
2-2	**Capacity Factor**	85 to 90 %	146,735 (324,386)	1	0.937	49.6	2.20	5.7	2,436	71.0	15.3
2-3	**SOFC Stack Cost**	442 to 414 $/kW	146,735 (324,386)	1	0.937	49.6	2.20	5.7	2,397	70.2	14.3
2-4	**Inverter Efficiency**	97 to 98 5	145,671 (321,149)	1	0.937	50.1	2.18	5.7	2,384	69.9	13.9

4. IGFC Pathway with Catalytic Gasification Technology

The performance of the IGFC plant is expected to increase with increased syngas methane content, up to some limiting value. This expected increase results from enhanced cell cooling due to methane in-situ reforming. An effective route to generating syngas with high methane content is the use of a catalytic, low-temperature coal gasifier.

4.1 Description of Process Areas

All of the IGFC plant areas with catalytic gasification are similar in their technologies and configurations, except for the gasification area. Some modifications to equipment and operating conditions used in the gas cleaning area are also made with this catalytic gasifier-based IGFC plant.

4.1.1 Catalytic Gasifier Area

The coal gasifier converts coal into a syngas to be applied as SOFC anode gas, and in this conversion process, losses of coal energy through partial-combustion of the coal and from carbon lost with the ash will occur. Catalytic coal gasification promotes the efficient gasification of coal at relatively low temperature where oxygen consumption is minimized, carbon conversion remains acceptably high, and the gasifier cold gas efficiency is high compared to conventional gasifiers. Under these conditions, especially if operated at high pressure, the methane content of the product syngas is also high, making it desirable for use with SOFC.

Catalytic coal gasification has not been tested beyond early development stages. It is assumed that the catalytic gasifier can be successfully developed for operation at the operating conditions selected and with the performance estimated in this evaluation should the benefits justify the development investment. There is currently no ongoing development effort for this type of coal gasifier.

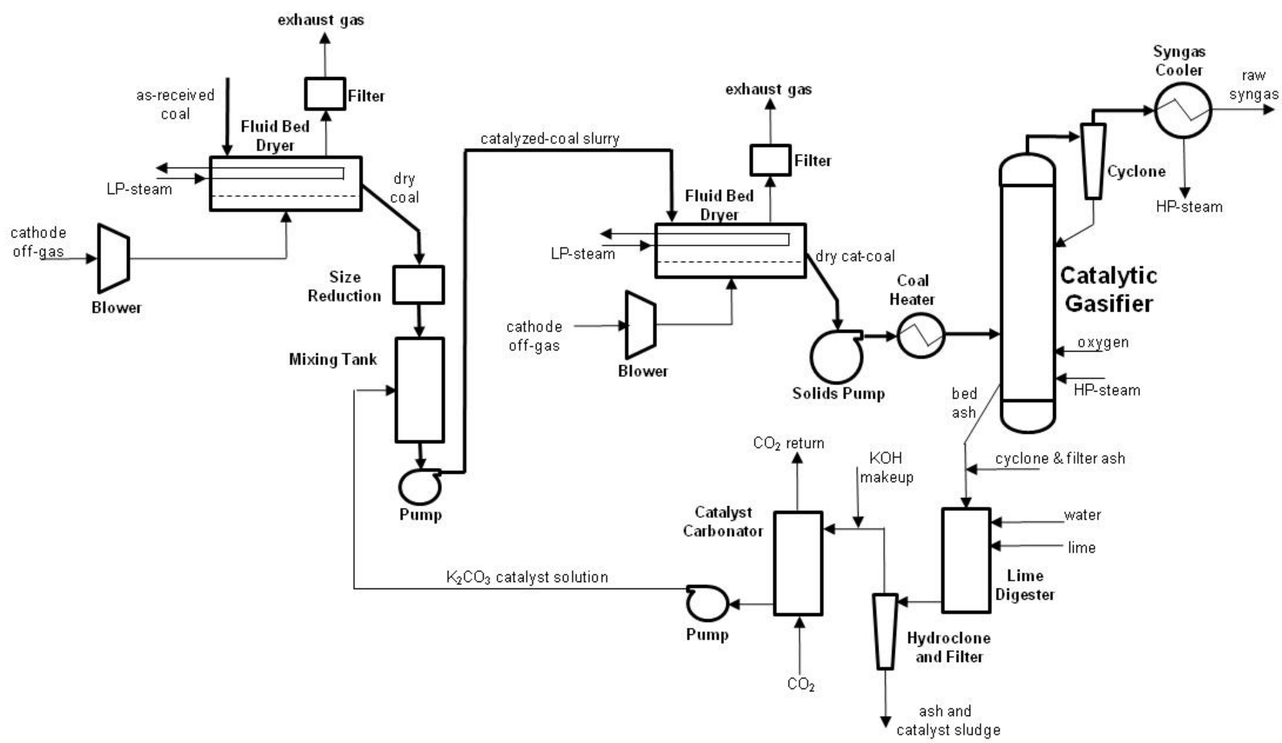
While a number of gasifier catalysts have been tested in laboratory studies, it has been found that the catalyst applied by Exxon (K_2CO_3 with KOH makeup) in their prior development program is very effective, but relatively expensive compared to other, less effective catalysts [23]. The catalyst material, K_2CO_3, is used as the primary catalyst in this evaluation, with KOH being the catalyst makeup form because of its lower cost.

The catalytic coal gasifier, assumed to use fluid bed contacting with steam and oxygen injection, was selected for the IGFC application because of its theoretical capability to efficiently generate a syngas having high methane content (approximately 30 mole percent). High-methane syngas is expected to promote more effective fuel cell cooling performance through internal SOFC methane reforming, leading to enhanced total plant efficiency due to a reduction in the needed cathode air rate that results. The demonstration of this enhancement capability using high-methane fuel in SOFC has not yet been completed.

Prior catalytic coal gasifier development by the Exxon Corporation was applied to a different fluid bed concept that used steam injection and recycle of a high-temperature stream of hydrogen-rich syngas, with the industrial application being synthetic natural gas production [24]. The design basis for the steam-oxygen catalytic gasifier applied in this evaluation was generated from thermodynamic equilibrium estimates for the gasifier operating at high pressure (975 psia exit pressure) and moderate temperature 704 °C (1,300 °F), as well as from Exxon catalytic gasifier design assumptions for the coal-catalyst treatment and catalyst recovery processes. The estimated performance for this gasifier is supported by Exxon catalytic gasifier data, assuming a carbon loss of 5 weight percent of the coal feed carbon.

A general process diagram for the catalytic coal gasifier and its associated coal-catalyst treatment and catalyst recovery equipment is shown in Exhibit 4-1.

Exhibit 4-1 Catalytic Gasifier Coal/Catalyst Processing



As-received coal is first dried in a fluid bed dryer using warm cathode off-gas from the power island for fluidization, and using low-pressure steam for additional in-bed heating [26]. The dried coal is reduced in size and is then mixed with a K_2CO_3 catalyst solution. This slurry is then dried in a second fluid bed dryer similar to the first, again using warm cathode off-gas for fluidization, and LP-steam for in-bed heating. The processed coal is preheated to 149°C (300°F) using low-pressure steam or cathode off-gas indirect heating, and is pressurized in a dry coal pump to the catalytic gasifier coal feed nozzles.

The gasifier ash and overhead fines are collected and are treated in a lime digester to release the catalyst from the ash constituents. The ash and catalyst sludge is separated from the slurry, and the catalyst solution is mixed with makeup catalyst (KOH). The catalyst solution is carbonated using a small portion of the plant CO_2 product. This step completes the recovery of the K_2CO_3 catalyst solution.

Details of the coal-catalyst processing steps assumed are as follows:

Coal Catalyst Treatment:

- Coal is crushed to -8 mesh (-2,380 microns or 0.0937 inch).
- Coal is mixed with recycled catalyst solution (37 wt percent K_2CO_3).
- The coal-catalyst solution is dried in fluid bed dryer at 54 ºC (130 ºF) using cathode off-gas and LP-steam heat source.
- The process results in a coal catalyst loading of 15 wt percent K_2CO_3 (dry coal).

Catalyst Recovery Factors:

- First step is $Ca(OH)_2$ digestion plus water washing, operated at149 ºC (300 ºF) with a mass ratio for Ca/K of 0.7 lb/lb.
- Soluble K recovery is 90 percent of the solids content to the digester.
- Solid/liquid separation is conducted using hydroclones.
- Overall catalyst recovery is 87 percent of the total loading.
- Catalyst makeup rate is 13 percent of the total catalyst feed rate.
- The makeup catalyst form is KOH.
- The recovered catalyst solution has 37 wt percent K_2CO_3 equivalent.

Gasifier Catalyst Reactions:

- It is estimated that some of the K_2CO_3 catalyst decomposes in the gasifier, releasing CO_2.
- K_2O reacts with the char and ash, producing water-soluble and insoluble forms.
- An equivalent stream of CO_2 is recycled from the plant CO_2 product stream to the makeup catalyst carbonator vessel.

The assumptions for the coal gasifier and the raw syngas cooler are listed in Exhibit 4-2. Note that it has been assumed in this study that the ash and catalyst mixture from the catalytic gasifier cases can be landfilled at the same per ton cost as the slag from the conventional coal gasifier.

Exhibit 4-2 Coal Gasification Section Assumptions with Catalytic Gasifier

	Specification/Assumptions
Gasifier	
Technology	Advanced steam-O_2 catalytic
Number in parallel	1
Dried coal-catalyst moisture, wt%	5.5
Coal feed technology	Advanced dry feed pump
Coal-catalyst preheat temperature, ºC (ºF)	149 (300)
Oxygen-to-coal feed ratio	0.19
Steam-to-coal ratio	1.445
Steam temperature, ºC (ºF)	538 (1000)
Recycle gas-to-coal ratio	0
Exit temperature, ºC (ºF)	704 (1300)
Exit pressure, MPa (psia)	6.72 (975)
Carbon loss with ash, wt% of coal carbon	5
Raw syngas composition basis	Equilibrium
Syngas methane content, vol% (dry)	31.3
Raw Syngas Cooler	
Technology	Fire-tube boiler
Number in parallel	1
Outlet temperature, ºC (ºF)	427 (800)

For the catalytic gasifier, a recuperative heat exchanger cools the syngas to 232°C (450°F) by preheating the clean syngas to 371°C (700°F).

The catalytic coal gasifier is a fluidized bed reactor contained within a cylindrical, refractory-lined, pressure vessel. It is assumed to operate with a superficial velocity of 1.2 ft/s. The gas residence time is very long at about 100 seconds, resulting in a very deep bed. Coal, oxygen, and steam are introduced into the vessel with mixing conditions to avoid the creation of hot spots within the fluidized bed.

4.1.2 Syngas Cleaning Area

The gas cleaning area is modified slightly in its configuration used with the conventional gasifier technology, as is indicated in Exhibit 4-3. The particulate removal temperature has been increased to 427°C (800°F), and zinc oxidize syngas polishing temperature has been increased to 371°C (700°F). Syngas reheat for sulfur polishing is accomplished by gas-gas recuperation heat exchange rather than by steam heating.

Exhibit 4-3 Gas Cleaning Section Assumptions with Catalytic Gasifier

	Specification/Assumptions
Gas Cleaning Technology	
Technology	Conventional dry gas cleaning
Number parallel trains	1
Particulate removal	Barrier filter at 371ºC (700ºF)
HCl removal	Water scrubber
Ammonia removal	Low-temperature gas cooling to 35 ºC (95 ºF)
Hg, As, Se, Cd, P	Activated-Carbon fixed beds at 35 ºC (95 ºF)
Bulk desulfurization	Selexol at 35 ºC (95 ºF)
Sulfur recovery	Conventional Claus plant with tail gas recycle
Polishing Desulfurization	ZnO fixed beds at 371ºC (700ºF) to 100 ppbv total sulfur
Syngas Preheating Source	Syngas recuperation

4.2 Scenario 3 – IGFC with Atmospheric-Pressure SOFC

The Scenario 3 baseline configuration uses the advanced, catalytic gasifier technology combined with atmospheric-pressure SOFC. The Coal Gasification Section contains the coal-catalyst preparation system, the ash handling system, the coal feeding system, the coal gasification system, the air separation system, and the raw syngas cooling system. The Gas Cleaning Section uses conventional dry gas cleaning technology based on single-stage Selexol acid gas removal. The section components are a high-temperature barrier filter, a water scrubbing system, a COS hydrolysis unit, a low-temperature syngas cooling system, a trace element removal system, a Selexol single-stage acid gas removal, a syngas reheat unit, and a ZnO fixed-bed sulfur-polishing unit.

The Scenario 3 baseline atmospheric-pressure Power Block assumptions and specifications are listing in Exhibit 4-4 and are identical to those applied for Case 1-1.

4.2.1 Case 3-1 Baseline Plant Performance Results

The following information is presented in tabular form for Case 3-1:

- Block Flow Diagrams and Stream Tables
- Performance Summaries
- Energy Flow Diagrams
- Steam Balances
- Water Balances
- Carbon Balances
- Sulfur Balances
- Air Emissions.

The system description follows the BFD in Exhibit 4-5, and stream numbers reference the same Exhibit. The table in Exhibit 4-6 provides process data for the numbered streams in the BFD. Note that 66.7 percent of the anode off-gas is recycled to the anode inlet stream, reducing the syngas methane content of 31.3 mole percent to 7.28 mole percent in the actual anode inlet stream. Exhibit 4-7 provides the power plant breakdown and overall thermal performance. Note that the steam turbine power represents only about 6 percent of the total plant power generated, with the SOFC system being the overwhelmingly dominate power generator. The baseline plant efficiency of 50.5 percent (HHV) is extremely high for a power plant with carbon removal compared to other fossil fuel power plant technologies.

Exhibit 4-4 Case 3-1 Atmospheric-Pressure Power Island Base Assumptions

	Specification/Assumptions
Syngas Expander	
Outlet pressure, MPa (psia)	0.21 (30)
Efficiency, adiabatic %	90
Generator efficiency (%)	98.5
Fuel Cell System	
Cell stack inlet temperature, ºC (ºF)	650 (1202)
Cell stack outlet temperature, ºC (ºF)	750 (1382)
Cell stack outlet pressure, MPa (psia)	0.12 (15.6)
Fuel single-step utilization, %	75
Fuel overall utilization, %	90
Stack anode-side pressure drop, MPa (psi)	0.0014 (0.2)
Stack cathode-side pressure drop, MPa (psi)	0.0014 (0.2)
Power density, mW/cm^2	400
Stack over-potential, mV	140
Operating voltage estimation method	Section 8.1.4
Cell degradation rate (% per 1000 hours)	1.5
Cell replacement period (% degraded)	20
Fuel Cell System Ancillary Components	
Anode gas recycle method	Hot gas fan
Anode recycle gas fan efficiency, adiabatic %	80
Anode heat exchanger pressure drop, MPa (psi)	0.0014 (0.2)
Cathode gas recycle method	Hot gas fan
Cathode recycle gas rate, %	50
Cathode recycle gas fan eff., adiabatic %	80
Cathode heat exchanger pressure drop, MPa	0.0014 (0.2)
Cathode blower efficiency, adiabatic %	90
Rectifier DC-to-AC efficiency, %	97.0 – 98.0
Recycle blower motor drives eff., %	87.6
Other electric motor drives efficiency, %	95
Transformer efficiency, %	99.65
Oxy-Combustor	
Technology	Atm-pressure diffusion flame
Outlet excess O_2, mole%	1
Steam Bottoming Cycle	
Technology level	subcritical
Modeling approach	Empirical approximation
Other steam generation duties	HP and LP process steam

To enhance the understanding of the IGFC power plant, mass flow and energy flows diagrams are presented in Exhibit 4-8 and Exhibit 4-9 on a basis relative to the coal as-received mass feed rate, and relative to the coal feed energy (HHV), respectively. The mass flow diagram indicates that the mass of the CO_2 product stream is 2.3 times the mass of the coal feed stream, and the largest mass flows in the plant are associated with the cathode gas streams, these being as large as almost 18 times the coal feed flow. In this case, the oxidant flow to the oxy-combustor is about 124 percent of the oxidant flow to the coal gasifier because the catalytic gasifier has relatively small oxygen consumption.

The energy flow diagram indicates that the CoP gasifier cold gas efficiency is about 95 percent (HHV) and that 93.1 percent of the coal feed energy is contained in the syngas feed stream to the SOFC power island, and 11.2 percent of the coal feed energy is contained on the anode off-gas stream going to the heat recovery section of the power island. This diagram lists the key stream energy flows and temperatures, and lists the heat loads for major heat exchangers, the auxiliary power consumption and power generation outputs of major plant components. The SOFC operating voltage is 0.79 V, lower than in Case 1 because the inlet anode gas is diluted by water vapor and methane. The cathode air preheat heat exchanger is large, but is smaller than in Case 1, with a heat load of about 37 percent of the coal feed energy input. The dominant auxiliary powers in the plant are the coal-catalyst treatment at 2.0 percent of the coal energy, the ASU at 1.7 percent, the CO_2 compression section at 3.6 percent, and the cathode air blower and recycle blower, each at 0.6 percent.

Steam balances for high-pressure and low-pressure steam are shown in Exhibit 4-10 and Exhibit 4-11. The high-pressure process steam feed for the catalytic gasifier is very large and is generated at the raw syngas cooling and at the oxy-combustor heat recovery step. The high-pressure steam for power generation is generated from the oxy-combustor heat recovery section and is significantly smaller than in Case 1-1.

The IGFC power plant water balance is shown on Exhibit 4-12. As in Case 1-1, the nearly complete recovery of water from the oxy-combustion CO_2 product stream results in water consumption in the IGFC plant being significantly lower than with other fossil fuel power plant technologies.

Carbon and sulfur balances are displayed in Exhibit 4-13 and Exhibit 4-14. The carbon inputs to the Case 3-1 plant syngas consist of carbon in the coal and carbon in the gasifier catalyst (potassium carbonate). It is assumed that all of the catalyst carbon is released to the syngas product in the gasifier. The recovered gasifier catalyst and the makeup catalyst, in the form of potassium hydroxide, are recarbonated to potassium carbonate using a portion of the plant CO_2 product. It is assumed that a 25 percent excess of recycled CO_2 is needed to perform the catalyst recarbonation. Nearly complete carbon capture is achieved, with a 99.1 percent carbon removal from the raw syngas. Note that the CO_2 product stream contains about 2.1 mole percent oxygen.

Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the elemental sulfur captured in the Claus plant, the trace levels of sulfur captured by the sulfur polishing sorbent, and the very small sulfur dioxide component that is part of the CO_2 product. Sulfur in the ash is considered to be negligible. Nearly complete sulfur removal is achieved, with 99.999 percent sulfur removal from the coal.

The air emissions are listed in Exhibit 4-15. Air emissions are nearly zero for Case 3-1 because all of the controlled species remaining in the very clean syngas are sequestered with the CO_2 product. The only CO_2 emission is from vented exhaust streams from condensate processing, with the total carbon removal exceeding 99 percent. The NOx emission estimate assumes that the SOFC off-gas air-combustor can operate with a NOx content of 5 ppmv. The Hg, and other trace element emission results from an assumed 95 percent removal performance.

Exhibit 4-5 Case 3-1 Block Flow Diagram

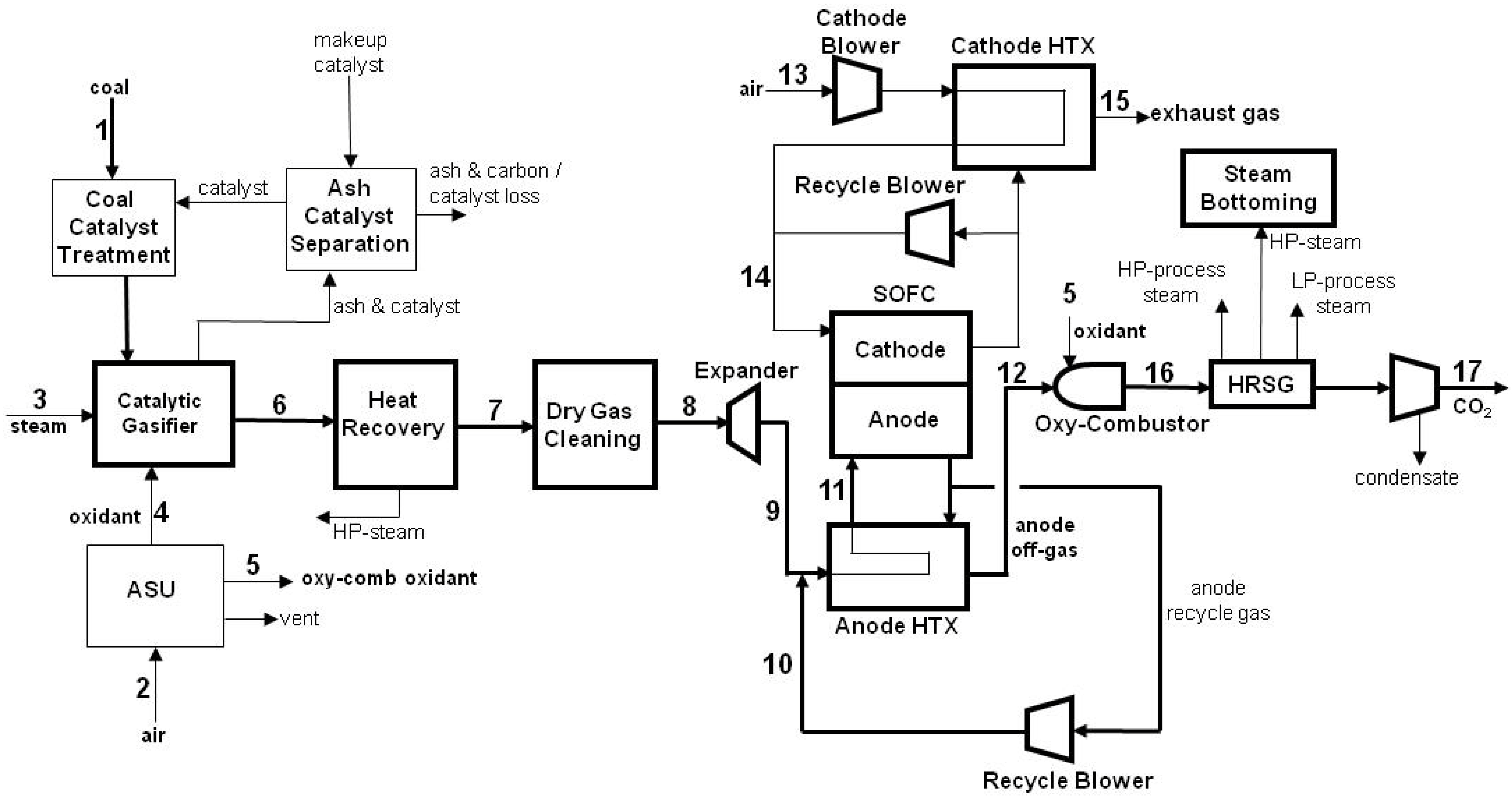


Exhibit 4-6 Case 3-1 Stream Table

	1	2	3	4	5	6	7	8	9	10	11
V-L Mole Fraction											
Ar	0.0000	0.0094	0	0.0031	0.0031	0.0002	0.0002	0.0003	0.0003	0.0002	0.0002
CH_4	0.0000	0	0	0	0	0.193	0.193	0.3135	0.3135	0	0.0736
CO	0.0000	0	0	0	0	0.0563	0.0563	0.0915	0.0915	0.0397	0.0519
CO_2	0.0000	0.0003	0	0	0	0.2116	0.2116	0.3424	0.3424	0.4196	0.4015
COS	0.0000	0	0	0	0	0.0001	0.0001	0	0	0	0
H_2	0.0000	0	0	0	0	0.1509	0.1509	0.2451	0.2451	0.0578	0.1018
H_2O	0.0000	0.0104	1	0	0	0.3748	0.3748	0.0009	0.0009	0.4788	0.3666
HCl	0.0000	0	0	0	0	0.0008	0.0008	0	0	0	0
H_2S	0.0000	0	0	0	0	0.007	0.007	0	0	0	0
N_2	0.0000	0.7722	0	0.0019	0.0019	0.0038	0.0038	0.0062	0.0062	0.0038	0.0044
NH_3	0.0000	0	0	0	0	0.0007	0.0007	0	0	0	0
O_2	0.0000	0.2077	0	0.995	0.995	0	0	0	0	0	0
SO_2	0.0000	0	0	0	0	0	0	0	0	0	0
Total	0.0000	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)	0	8,797	10,903	786	937	15,794	15,794	9,727	9,727	31,701	41,429
V-L Flowrate (kg/hr)	0	253,836	196,424	25,178	29,989	338,941	338,942	227,212	227,212	901,501	1,128,734
Solids Flowrate (kg/hr)	144,558	0	0	0	0	25,477	1,274	0	0	0	0
Temperature (°C)	15	15	538	133	27	705	427	371	53	759	650
Pressure (MPa, abs)	0.10	0.10	7.58	7.24	0.16	6.72	6.45	5.64	0.14	0.11	0.11
Specific Enthalpy (kJ/kg)^A	---	-101.7	-12,488.5	92.7	1.1	-7,796.2	-8,389.3	-6,641.2	-7,175.6	-8,998.1	-8,536.3
Density (kg/m^3)	---	1.2	21.3	67.9	2.0	17.5	23.9	24.2	1.2	0.4	0.4
V-L Molecular Weight	---	28.855	18.015	32.016	32.016	21.460	21.460	23.358	23.358	28.438	27.245
V-L Flowrate (lb_{mol}/hr)	0	19,394	24,038	1,734	2,065	34,820	34,820	21,445	21,445	69,889	91,335
V-L Flowrate (lb/hr)	0	559,612	433,042	55,508	66,114	747,238	747,239	500,916	500,916	1,987,471	2,488,434
Solids Flowrate (lb/hr)	318,697	0	0	0	0	56,167	2,808	0	0	0	0
Temperature (°F)	59	59	1,000	272	80	1,301	800	700	127	1,398	1,202
Pressure (psia)	14.7	14.7	1100	1050	23	975	935	818	20	16.2	16.2
Specific Enthalpy (Btu/lb)^A	---	-43.7	-5,369.1	39.8	0.5	-3,351.8	-3,606.7	-2,855.2	-3,085.0	-3,868.5	-3,670.0
Density (lb/ft^3)	---	0.076	1.330	4.241	0.127	1.095	1.490	1.510	0.074	0.023	0.025

A - Standard Reference State is the ideal vapor heat of formation at 298.15°K

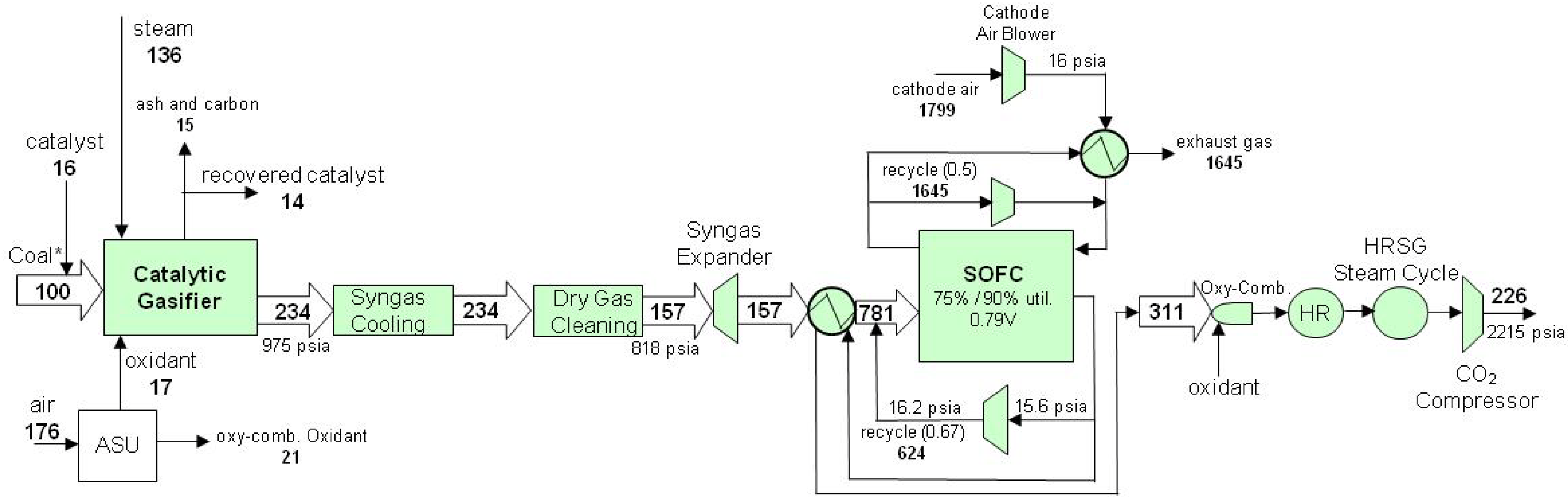
Exhibit 4-6 Case 3-1 Stream Table (continued)

	12	13	14	15	16	17
V-L Mole Fraction						
Ar	0.0002	0.0094	0.0098	0.0102	0.0003	0.0007
CH_4	0	0	0	0	0	0
CO	0.0397	0	0	0	0	0
CO_2	0.4196	0.0003	0.0003	0.0003	0.4547	0.9696
COS	0	0	0	0	0	0
H_2	0.0578	0	0	0	0	0
H_2O	0.4788	0.0104	0.0108	0.0113	0.5311	0
HCl	0	0	0	0	0	0
H_2S	0	0	0	0	0	0
N_2	0.0038	0.7722	0.7722	0.8369	0.0039	0.0084
NH_3	0	0	0	0	0	0
O_2	0	0.2077	0.1759	0.1414	0.01	0.0213
SO_2	0	0	0	0	0	0
Total	1.0000	1.0000	0.9690	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	15,827	90,134	90,134	83,172	15,992	7,494
V-L Flowrate (kg/hr)	450,085	2,600,795	2,600,795	2,378,005	480,074	326,886
Solids Flowrate (kg/hr)	0	0	0	0	0	0
Temperature (°C)	617	15	650	199	820	38
Pressure (MPa, abs)	0.11	0.10	0.11	0.11	0.10	15.27
Specific Enthalpy (kJ/kg)[A]	-9,207.9	-101.7	458.4	79.2	-9,123.6	-8,961.7
Density (kg/m^3)	0.4	1.2	0.4	0.8	0.3	667.3
V-L Molecular Weight	28.438	28.855	28.728	28.591	30.020	43.617
V-L Flowrate (lb_{mol}/hr)	34,893	198,712	198,712	183,363	35,256	16,522
V-L Flowrate (lb/hr)	992,269	5,733,776	5,733,776	5,242,608	1,058,384	720,661
Solids Flowrate (lb/hr)	0	0	0	0	0	0
Temperature (°F)	1,143	59	1,202	391	1,508	100
Pressure (psia)	15.4	14.7	15.8	15.4	14.8	2215
Specific Enthalpy (Btu/lb)[A]	-3,958.7	-43.7	197.1	34.0	-3,922.4	-3,852.8
Density (lb/ft^3)	0.026	0.076	0.025	0.048	0.021	41.660

Exhibit 4-7 Case 3-1 Plant Performance Summary (100 Percent Load)

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	
SOFC Power	570,305
Syngas Expander Power	33,070
Steam Turbine Power	35,976
TOTAL POWER, kWe	**639,350**
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling	275
Coal Size Reduction	465
Catalyst-Coal Processing	1,807
Coal Feeding	994
Ash Handling	716
Air Separation Unit Auxiliaries	297
Air Separation Unit Main Air Compressor	14,481
Oxygen Compressor	3,256
Nitrogen Compression	508
Anode Recycle Blower	4,049
Claus Tail Gas Recycle Compressor	799
CO_2 Compressor	39,086
BFW Pump	571
Condensate Pump	38
Circulating Water Pump	631
Ground Water Pumps	326
Cooling Tower Fans	1,047
Scrubber Pumps	170
Selexol Auxiliary Power	2,223
Steam Turbine Auxiliaries	12
Cathode Air Blower	6,169
Cathode Gas Recycle Blower	6,574
Claus / TGTU Auxiliaries	130
Miscellaneous Balance of Plant	2,490
Transformer Losses	2,238
TOTAL AUXILIARIES, kWe	**89,350**
NET POWER, kWe	**550,000**
Net Plant Efficiency, % (HHV)	**50.5**
Net Plant Heat Rate, kJ/kWh (Btu/kWh)	**7,132 (6,760)**
CONDENSER COOLING DUTY 10^6 kJ/h (10^6 Btu/h)	**204 (193)**
CONSUMABLES	
As-Received Coal Feed, kg/h (lb/h)	144,558 (318,697)
Thermal Input[1], kWt	1,089,696
Raw Water Consumption, m^3/min (gpm)	5.2 (1,368)

Exhibit 4-8 Cases 3-1 Mass Flow Diagram



*Coal feed: ILL #6 as received

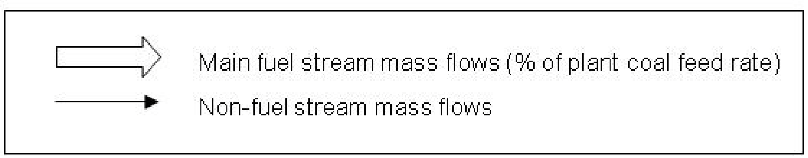


Exhibit 4-9 Cases 3-1 Energy Flow Diagram

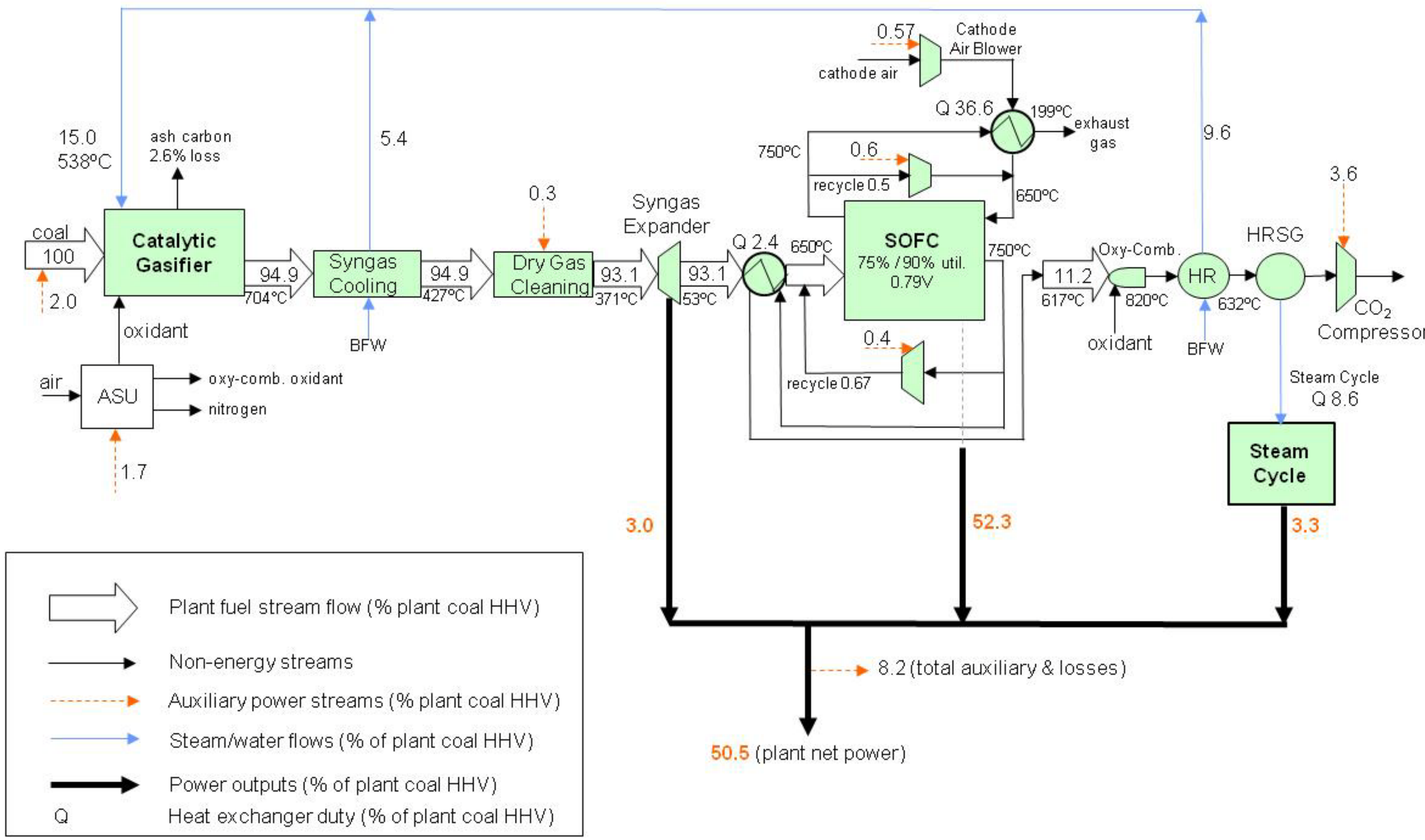


Exhibit 4-10 Case 3-1 High-Pressure Steam Balance

HP Process Steam Use, kg/h (lb/h)		HP Process Steam Generation, kg/h (lb/h)	
Gasifier feed	196,424 (433,042)	**Raw syngas cooling**	70,396 (155,196)
		Oxy-combustor heat recovery	126,029 (277,846)
Total	196,424 (433,042)	**Total**	196,424 (433,042)

HP Power-Steam generation, GJ/h (MMBtu/h)		
	Case 1	
Raw syngas cooling	0 (0)	
Oxy-combustor heat recovery	334 (316)	
Total	334 (316)	

Exhibit 4-11 Case 3-1 Low-Pressure Steam Balance

LP Process Steam Use, GJ/h (MMBtu/h)		LP Process Steam Generation, GJ/h (MMBtu/h)	
Coal-Catalyst drying	30 (28)	**LT syngas cooling**	270 (256)
Selexol stripping	91 (86)	**Cathode exhaust cooling**	14 (13)
ASU	22 (21)		
Sour water stripping	110 (104)		
Coal preheat	32 (30)		
Total	284 (269)	**Total**	284 (269)

Exhibit 4-12 Case 3-1 Water Balances

	m^3/min (gpm)
Water Demand	**11.08 (2,927)**
Catalyst Treatment	0.56 (147)
Condenser Makeup *Gasifier Steam* *BFW Makeup*	3.34 (882) *3.28 (865)* *0.06 (17)*
Cooling Tower Makeup	7.18 (1,898)
Water Recovery for Reuse	**3.90 (1,029)**
Low-temperature Cooling	1.60 (423)
CO_2 Dehydration	2.30 (607)
Process Discharge Water	**2.01 (530)**
Cooling Tower Water Blowdown	1.62 (427)
Low-temperature Cooling	0.16 (42)
CO_2 Dehydration	0.23 (61)
Raw Water Consumed	**5.18 (1,368)**

Exhibit 4-13 Cases 3-1 Carbon Balance

Carbon In, kg/h (lb/h)		Carbon Out, kg/h (lb/h)	
Coal	92,148 (203,152)	**Ash**	4,067 (10,158)
Gasifier Catalyst	1,970 (4,344)	**Catalyst CO_2 Recycle**	3,232 (7,126)
		Exhaust Gas	663 (1,461)
		CO_2 Product	85,616 (188,752)
Total	94,119 (207,496)	**Total**	94,119 (207,496)

Exhibit 4-14 Cases 3-1 Sulfur Balance

Sulfur In, kg/h (lb/h)		Sulfur Out, kg/h (lb/h)	
Coal	3,623 (7,988)	**Elemental Sulfur**	3,617 (7,974)
		Polishing Sorbent	6 (13)
		CO_2 Product	0 (0)
Total	3,623 (7,988)	**Total**	3,623 (7,988)

Exhibit 4-15 Cases 3-1 Air Emissions

	kg/GJ ($lb/10^6$ Btu)	Tonne/year (tons/year) 80% capacity factor	kg/MWh (lb/MWh)
SO_2	0 (0)	0 (0)	0 (0)
NO_X	0 (0)	0 (0)	0 (0)
Particulates	0 (0)	0 (0)	0 (0)
Hg	0 (0)	0 (0)	0 (0)
CO_2	0.07 (0.16)	1,883 (2,075)	0.49 (1.08)

4.2.2 Case 3-1 Baseline Plant Cost Results

The capital cost estimate for Case 3-1 is broken down in Exhibit 4-16. Owner's costs are included in Exhibit 4-17. The dominant area costs are the gasification area and the SOFC power island. The single highest cost component in the plant is the SOFC stacks and inverters. The first-year cost-of-electricity for Cases 3-1 is displayed in Exhibit 4-18. The dominant contributor to the COE is capital recovery, with fuel cost being relatively small because of the high plant efficiency. These are the same type of cost characteristics found for Case 1-1.

Exhibit 4-16 Case 3-1 Capital Cost Breakdowns

	TOTAL PLANT COST	
Item/Description	**$ x 1000**	**$/kW**
COAL & SORBENT HANDLING	**26,188**	**48**
COAL & SORBENT PREP & FEED	**62,420**	**113**
Coal prep & feed system	52,292	95
Catalyst Treatment	10,129	18
FEEDWATER & MISC. BOP SYSTEMS	**14,263**	**26**
GASIFIER & ACCESSORIES	**188,828**	**343**
Gasifier & Syngas Cooler	103,135	188
ASU & Oxidant Compressor	73,131	133
Other Gasification Equipment	12,561	23
GAS CLEANUP & PIPING	**103,747**	**189**
Scrubber & Low Temperature Cooling	11,859	22
Single-Stage Selexol/MDEA	56,745	103
Claus Plant	20,469	37
Trace removal	1,424	3
COS Hydrolysis	6,261	11
Blowback, Piping, Foundations	3,543	6
Sulfur polishing	2,895	5
Heat Interchanger	553	1
CO_2 DRYING & COMPRESSION	**41,645**	**76**
SOFC POWER ISLAND	**338,966**	**616**
Syngas expander	6,332	12
SOFC Stack Units (stack modules, enclosures, inverters)	267,414	486
Cathode Air Blower	2,005	4
Cathode Heat Exchanger	5,639	10
Anode Heat Exchanger	6,425	12
Anode Recycle Blower	875	2
Oxy-Combustor	11,068	20
HRSG, DUCTING & STACK	**22,493**	**41**
STEAM POWER SYSTEM	**12,810**	**23**
COOLING WATER SYSTEM	**17,078**	**31**
ASH/SPENT SORBENT HANDLING SYS	**38,868**	**71**
Ash handling	25,499	46
Catalyst Recovery	13,369	24
ACCESSORY ELECTRIC PLANT	**52,605**	**96**
INSTRUMENTATION & CONTROL	**27,743**	**50**
IMPROVEMENTS TO SITE	**14,378**	**26**
BUILDING & STRUCTURES	**13,302**	**24**
TOTAL PLANT COST ($1000)	**975,335**	**1,773**

Exhibit 4-17 Case 3-1 Owner's Costs

Owner's Costs		
Preproduction Costs		
6 Months All Labor	10,846	20
1 Month Maintenance Materials	2,301	4
1 Month Non-fuel Consumables	1,407	3
1 Month Waste Disposal	461	1
25% of 1 Months Fuel Cost at 100% CF	1,109	2
2% of TPC	19,507	35
Total	**35,632**	**65**
Inventory Capital		
60 day supply of fuel and consumables at 100% CF	11,410	21
0.5% of TPC (spare parts)	4,877	9
Total	16,286	30
Initial Cost for Catalyst and Chemicals	5,908	11
Land	900	2
Other Owner's Costs	146,300	266
Financing Costs	26,334	48
Total Overnight Costs (TOC)	**1,206,695**	**2,194**
Total As-Spent Cost (TASC)	**1,375,632**	**2,501**

Exhibit 4-18 Case 3-1 Cost-of-Electricity Breakdown

OPERATING & MAINTENANCE LABOR					Annual Cost $	Annual Unit Cost mills/kWh
Annual Operating Labor Cost					5,918,913	
Maintenance Labor Cost					11,435,340	
Administrative & Support Labor					4,338,563	
Property Taxes and Insurance					19,602,753	
TOTAL FIXED OPERATING COSTS					41,295,570	**10.7**
VARIABLE OPERATING COSTS						
Maintenance Material Cost					22,093,077	
Stack Replacement Cost					18,113,664	
Subtotal					**40,206,741**	
Consumables	Consumption Initial Fill	/Day	Unit Cost	Initial Cost		
Water (/1000 gallons)	0	4,186	1.08	0	1,322,577	
Chemicals						
MU & WT Chem. (lbs)	0	7,246	0.17	0	366,170	
Carbon (Trace Removal) (lb)	357,933	490	1.05	375,891	150,356	
COS Catalyst (m3)	271	0.19	2,397	649,419	129,884	
Selexol Solution (gal)	183,207	28.87	13.40	2,454,651	112,964	
Claus / DSRP Catalyst (ft3)	0	1.25	131	0	47,778	
ZnO polishing sorbent (lb)	147,471	1,456	1.50	221,207	637,636	
KOH Coal Catalyst makeup (lb)	5,920,051	98,668	0.16	947,208	4,609,746	
Lime for catalyst recovery	31,479,635	524,661	0.04	1,259,185	6,128,036	
Subtotal Chemicals				**5,907,561**	**12,182,570**	
Waste Disposal						
Spent Trace Catalyst (lb.)	0	530	0.42	0	64,943	
Ash (ton)	0	883	16.23	0	4,185,034	
Spent sorbents (lb)	0	1,456	0.42	0	178,538	
Subtotal-Waste Disposal ($)					**4,428,515**	
TOTAL VARIABLE OPERATING COSTS				**5,907,561**	**58,140,403**	**15.1**
Fuel Coal (ton)	0	2,914	38.18	0	42,583,697	**11.0**
Capital Recovery (mills/kWh)						**38.9**
TS&M (mills/kWh)						**4.3**
COE First Year (mills/kWh)						**80.1**

4.2.3 Scenario 3 Pathway Results

Scenario 3 pathway performance and cost estimates were performed for progressions in 1) a IGFC plant configuration with baseline SOFC conditions, using the advanced, catalytic coal gasifier technology; 2) the cell performance degradation rate, improved from 1.5 percent /1000 hours to 0.2 percent /1000 hours; 3) the cell overpotential, reduced from 140 mV to 70 mV; 4) the plant capacity factor, increased from 80percent to 85percent; 5) the plant capacity factor, increased from 85 percent to 90 percent; 6) the cost of the SOFC stack blocks reduced 20 percent, with the total SOFC cost reduced from 296 to 268 $ per kW of SOFC output; and 7) the SOFC DC-to-AC inverter efficiency increased from 97 to 98 percent. The results are tabulated in Exhibit 4-21.

The progression increases the plant efficiency to 55.7 percent (HHV), with the COE reduced to 61.2 mills/kWh. These performance and cost results are improved significantly over the Scenario 1 results and are far superior to performance and cost of other fossil fuel power plant technologies. There are corresponding reductions in the plant capital investment and the raw water consumption rate over the pathway.

It is also of interest to observe some of the characteristics of the most expensive component systems in the plant, the gasifier, the SOFC stack units, and the SOFC power island. Exhibit 4-19 shows some key characteristics of the coal gasifier in Pathway 3. The coal feed rates and syngas exit volumetric flows are much smaller than those for the conventional gasifier in pathway 1, resulting in a much lower catalytic gasifier cost.

Exhibit 4-19 Scenario 3 Catalytic Coal Gasifier Characteristics

Case	Gasifier Coal Feed Rate kg/hr (lb/hr)	Gasifier Exit Pressure MPa (psia)	Gasifier Exit Temperature °C (°F)	Gasifier Exit Syngas Rate, 1000 m^3/h (1000 ft^3/h)	Gasifier & Heat Recovery Cost $1000
3-1	135,961 (299,744)	6.72 (975)	705 (1300)	19.3 (683)	103,135
3-2	135,961 (299,744)	6.72 (975)	705 (1300)	19.3 (683)	103,135
3-3	124,495 (274,465)	6.72 (975)	705 (1300)	17.7 (625)	96,966
3-4	124,495 (274,465)	6.72 (975)	705 (1300)	17.7 (625)	96,966
3-5	124,495 (274,465)	6.72 (975)	705 (1300)	17.7 (625)	96,966
3-6	124,495 (274,465)	6.72 (975)	705 (1300)	17.7 (625)	96,966
3-7	123,199 (271,608)	6.72 (975)	705 (1300)	17.5 (619)	96,259

Exhibit 4-20 lists some of the SOFC characteristic along Pathway 3. The SOFC current density generally decreases along the pathway, as the cell voltage remains near 0.88 V. The spare cell surface installed and the stack replacement times are the optimum values estimated. The SOFC stack unit cost and power island cost generally decrease along the pathway.

Exhibit 4-20 Scenario 3 SOFC Characteristics

Case	Cell Voltage V	Power Density mW DC/cm^2	Current Density mA/cm^2	Spare Cell Surface Installed %	Stack Replacement Time years	SOFC Stack Unit Cost $/kW	Power Island Cost $/kW
3-1	0.787	400	508	58.4	5.6	486	616
3-2	0.787	400	508	19.7	14.0	367	493
3-3	0.852	400	470	19.7	14.0	364	461
3-4	0.852	400	470	19.7	13.2	364	461
3-5	0.852	400	470	19.7	12.5	364	461
3-6	0.852	400	470	19.7	12.5	330	427
3-7	0.852	400	470	19.7	12.5	330	427

Exhibit 4-21 Scenario 3 Pathway Results

Case	Pathway Description	Change Made	Coal Feed Rate kg/h (lb/h)	Number Parallel Trains	Cell Voltage V	Plant Efficiency %, HHV	Raw Water Consumed gpm/MW	CO_2 Emission kg/MWh	Capital Cost, TOC $/kW	COE mills/kWh	Cost of CO_2 Avoided $/tonne
3-1	**Baseline Atm-pressure**	Baseline	135,961 (299,744)	1	0.787	50.5	2.49	1.8	2,194	79.8	26.3
3-2	**Degradation**	1.5 to 0.2 %/1000 hours	135,961 (299,744)	1	0.787	50.5	2.49	1.8	2,043	71.5	15.8
3-3	**Cell Over-potential**	140 to 70 mV	124,495 (274,465)	1	0.852	55.1	2.26	1.6	1,918	67.8	11.2
3-4	**Capacity Factor**	80 to 85 %	124,495 (274,465)	1	0.852	55.1	2.26	1.6	1,918	65.0	7.6
3-5	**Capacity Factor**	85 to 90 %	124,495 (274,465)	1	0.852	55.1	2.26	1.6	1,918	62.5	4.5
3-6	**SOFC Stack Cost**	296 to 268 $/kW	124,495 (274,465)	1	0.852	55.1	2.26	1.6	1,877	61.6	3.3
3-7	**Inverter Efficiency**	97 to 98 %	123,199 (271,608)	1	0.852	55.7	1.88	1.6	1,866	61.2	2.9

4.3 Scenario 4 - IGFC with Pressurized-SOFC

Scenario 4 applies the catalytic coal gasifier with a configuration using pressurized SOFC. Pressurized SOFC can be configured in two general, alternative arrangements:

1. The anode off-gas oxy-combustor is followed by hot gas expander power generation (expansion ratio about 18). A heat recovery steam generator (HRSG) produces steam for power generation, and the remaining, low-pressure, wet CO_2 stream is dried and compressed (compression ratio about 149).
2. The anode off-gas oxy-combustor is followed directly by a HRSG for steam bottoming power generation. The remaining, high-pressure, wet CO_2 stream is dried and compressed (compression ratio about 8.4).

Configuration 2 is expected to be the least complex and most effective approach and is utilized for this evaluation. Note that the pressurize configuration and its operating conditions have not been optimized. All areas of the plant are identical to the Case 3 plant areas. The assumptions and specifications for the power island and the CO_2 dehydration and compression area are identical to those used in Case 2-1.

Case 4-1 Plant Performance Results

The following information is presented in tabular form for Case 4-1:

- Block Flow Diagrams and Stream Table
- Performance Summary
- Mass and Energy Flow Diagrams
- Steam Balance
- Water Balance
- Carbon Balance
- Sulfur Balance
- Air Emissions

The system description follows the BFD in Exhibit 4-22 and stream numbers reference the same Exhibit. Exhibit 4-23 provides process data for the numbered streams in the BFD. Exhibit 4-24 provides the power plant breakdown and overall thermal performance. Note that the steam turbine power represents only about 9 percent of the total plant power generated, with the SOFC system being the overwhelmingly dominate power generator. The Case 2 Baseline plant efficiency of 60.0 percent (HHV) is extremely high for a power plant with carbon removal compared to other fossil fuel power plant technologies.

Mass flow and energy flows diagrams are presented in Exhibit 4-25 and Exhibit 4-26 on a basis relative to the coal as-received mass feed rate, and relative to the coal feed energy (HHV), respectively. The mass flow diagram indicates that the mass of the CO_2 product stream is 2.24 times the mass of the coal feed stream, and the largest mass flows in the plant are associated with the cathode gas streams, these being as large as almost 14 times the coal feed flow. In this case, the oxidant flow to the oxy-combustor is about 124 percent of the oxidant flow to the coal gasifier.

The energy flow diagram indicates that the enhanced conventional gasifier cold gas efficiency is about 94.9 percent (HHV) and that 93.1 percent of the coal feed energy is contained in the syngas feed stream to the SOFC power island, and 11.2 percent of the coal feed energy is contained on the anode off-gas stream going to the heat recovery section of the power island. This diagram lists the key stream energy flows and temperatures, and lists the heat loads for major heat exchangers, auxiliary power consumption and power generation outputs of major plant components. The SOFC operating voltage is 0.91 V, much of the increase in this voltage being due to the increased pressure of the SOFC system. The cathode air preheat heat exchanger in Case 4-1 is not as large as in Case 2-1, with a heat load of about 13 percent of the coal feed energy input, because the compression of the cathode air partially preheats the stream. The dominant auxiliary powers in the plant are the ASU at 1.9 percent of the coal energy, the cathode air compressor-expander at 3.6 percent, and the CO_2 compression area at 1.4 percent. The ASU auxiliary power is increased relative to Case 3 because the oxy-combustion oxidant stream must be compressed to the pressurized condition of the anode off-gas. The CO_2 compression area auxiliary power is relatively small because the oxy-combustor off-gas is at high pressure.

Steam balances for high-pressure and low-pressure steam are shown in Exhibit 4-27 and Exhibit 4-28. The high-pressure process steam feed for the catalytic gasifier is very large and is generated at the raw syngas cooling and at the oxy-combustor heat recovery step. The high-pressure steam for power generation is generated from the oxy-combustor heat recovery section and is significantly smaller than in Case 3-1.

The IGFC power plant water balance is shown on Exhibit 4-29. As in Case 3-1, the nearly complete recovery of water from the oxy-combustion CO_2 product stream results in water consumption in the IGFC plant being significantly lower than with other fossil fuel power plant technologies.

Carbon and sulfur balances are displayed in Exhibit 4-30 and Exhibit 4-31. The carbon inputs to the Case 4-1 plant syngas consist of carbon in the coal and carbon in the gasifier catalyst (potassium carbonate). It is assumed that all of the catalyst carbon is released to the syngas product in the gasifier. The recovered gasifier catalyst and the makeup catalyst, in the form of potassium hydroxide, are recarbonated to potassium carbonate using a portion of the plant CO_2 product. It is assumed that a 25 percent excess of recycled CO_2 is needed to perform the catalyst recarbonation. Nearly complete carbon capture is achieved, with a 98.4 percent carbon removal from the raw syngas. Note that the CO_2 product stream contains about 2.2 mole percent oxygen.

Sulfur input comes solely from the sulfur in the coal. Sulfur output includes the elemental sulfur captured in the Claus plant, the trace levels of sulfur captured by the sulfur polishing sorbent, and the very small sulfur dioxide component that is part of the CO_2 product. Sulfur in the ash is considered to be negligible. Nearly complete sulfur removal is achieved, with 99.999 percent sulfur removal from the coal.

The air emissions are listed in Exhibit 4-32. Air emissions are nearly zero for Case 4-1 because all of the controlled species remaining in the very clean syngas are sequestered with the CO_2 product. The only CO_2 emission is from vented exhaust streams from condensate processing, with the total carbon removal exceeding 98 percent. The NOx emission estimate assumes that the SOFC off-gas air-combustor can operate with a NOx content of 5 ppmv. The Hg and other trace element emissions result from an assumed 95 percent removal performance.

Exhibit 4-22 Case 4-1 Block Flow Diagram

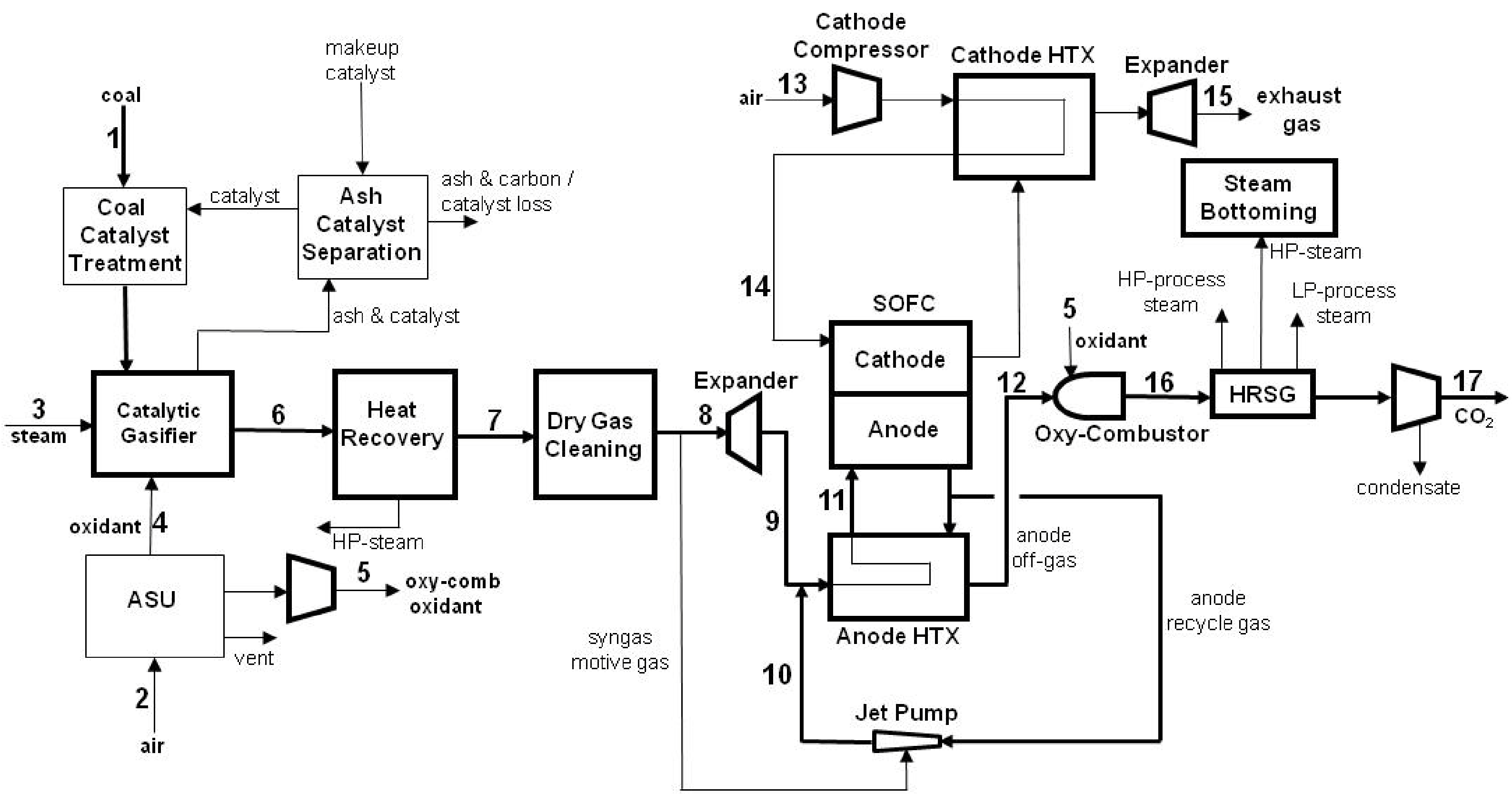


Exhibit 4-23 Case 4-1 Stream Table

	1	2	3	4	5	6	7	8	9	10	11
V-L Mole Fraction											
Ar	0.0000	0.0094	0.0000	0.0031	0.0031	0.0002	0.0002	0.0003	0.0003	0.0002	0.0002
CH_4	0.0000	0	0.0000	0.0000	0.0000	0.1929	0.1929	0.3132	0.3132	0.0113	0.0737
CO	0.0000	0	0.0000	0.0000	0.0000	0.0564	0.0564	0.0915	0.0915	0.0416	0.0519
CO_2	0.0000	0.0003	0.0000	0.0000	0.0000	0.2117	0.2117	0.3423	0.3423	0.4168	0.4014
COS	0.0000	0	0.0000	0.0000	0.0000	0.0001	0.0001	0	0	0	0
H_2	0.0000	0	0.0000	0.0000	0.0000	0.1512	0.1512	0.2455	0.2455	0.0643	0.1018
H_2O	0.0000	0.0104	1.0000	0.0000	0.0000	0.3746	0.3746	0.0009	0.0009	0.4619	0.3666
HCl	0.0000	0	0.0000	0.0000	0.0000	0.0008	0.0008	0	0	0	0
H_2S	0.0000	0	0.0000	0.0000	0.0000	0.007	0.007	0	0	0	0
N_2	0.0000	0.7722	0.0000	0.0019	0.0019	0.0038	0.0038	0.0062	0.0062	0.0039	0.0044
NH_3	0.0000	0	0.0000	0.0000	0.0000	0.0007	0.0007	0	0	0	0
O_2	0.0000	0.2077	0.0000	0.9950	0.9950	0	0	0	0	0	0
SO_2	0.0000	0	0.0000	0.0000	0.0000	0	0	0	0	0	0
Total	0.0000	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)	0	7,410	9,169	661	790	13,284	13,284	7,202	7,202	27,635	34,837
V-L Flowrate (kg/hr)	0	213,800	165,173	21,172	25,296	285,025	285,025	168,174	168,174	780,906	860,085
Solids Flowrate (kg/hr)	121,559	0	0	0	0	19,407	1,071	0	0	0	0
Temperature (°C)	15	15	538	133	136	704	427	371	263	736	649
Pressure (MPa, abs)	0.10	0.10	7.58	7.24	1.97	6.72	6.45	5.64	2.00	1.99	1.98
Specific Enthalpy (kJ/kg)[A]	---	-101.7	-12,488.5	92.7	100.9	-7,796.3	-8,388.2	-6,641.0	-6,836.5	-8,999.1	-8,537.9
Density (kg/m^3)	---	1.2	21.3	67.9	18.5	17.6	23.9	24.2	10.4	6.7	7.0
V-L Molecular Weight	---	28.850	18.010	32.020	32.020	21.460	21.460	23.351	23.351	28.258	27.244
V-L Flowrate (lb_{mol}/hr)	0	16,335	20,213	1,458	1,742	29,285	29,285	15,877	15,877	60,925	76,802
V-L Flowrate (lb/hr)	0	471,349	364,145	46,677	55,768	628,373	628,373	370,760	370,760	1,721,603	1,896,165
Solids Flowrate (lb/hr)	267,992	0	0	0	0	42,785	2,361	0	0	0	0
Temperature (°F)	59	59	1000	272	276	1300	800	700	505	1356	1201
Pressure (psia)	14.7	14.7	1100	1050	285	975	935	818	290	289	287
Specific Enthalpy (Btu/lb)[A]	---	-43.7	-5,369.1	39.8	43.4	-3,351.8	-3,606.3	-2,855.1	-2,939.2	-3,868.9	-3,670.6
Density (lb/ft^3)	---	0.076	1.330	4.240	1.153	1.100	1.490	1.510	0.651	0.418	0.438

A - Standard Reference State is the ideal vapor heat of formation at 298.15°K

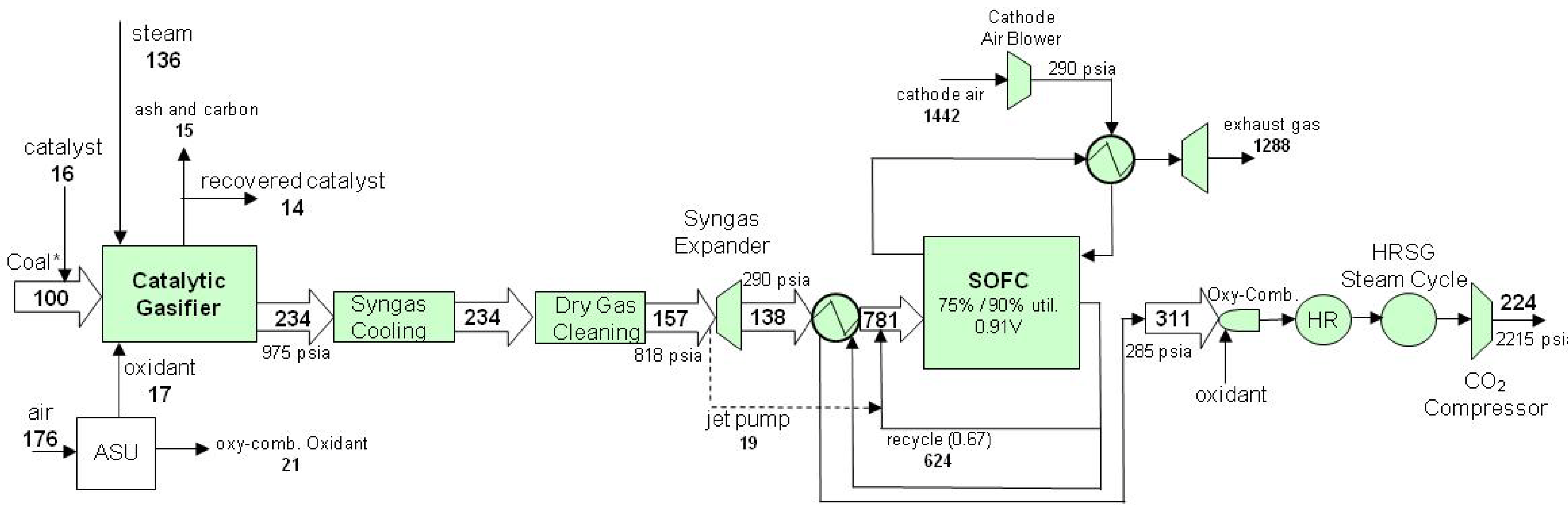
Exhibit 4-21 Case 4-1 Stream Table (continued)

	12	13	14	15	16	17
V-L Mole Fraction						
Ar	0.0002	0.0094	0.0094	0.0104	0.0003	0.0007
CH_4	0.0001	0.0000	0	0	0	0
CO	0.0398	0.0000	0	0	0	0
CO_2	0.4196	0.0003	0.0003	0.0003	0.4546	0.9694
COS	0	0.0000	0	0	0	0
H_2	0.0576	0.0000	0	0	0	0
H_2O	0.4789	0.0104	0.0104	0.0115	0.5311	0
HCl	0	0.0000	0	0	0	0
H_2S	0	0.0000	0	0	0	0
N_2	0.0038	0.7722	0.7722	0.8545	0.0039	0.0084
NH_3	0	0.0000	0	0	0	0
O_2	0	0.2077	0.2077	0.1232	0.01	0.0215
SO_2	0	0.0000	0	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	13,306	60,760	60,760	54,907	13,449	6,256
V-L Flowrate (kg/hr)	378,417	1,753,215	1,753,214	1,565,904	403,713	272,866
Solids Flowrate (kg/hr)	0	0	0	0	0	0
Temperature (°C)	717	15	650	122	1,205	38
Pressure (MPa, abs)	1.94	0.10	1.98	0.11	1.89	15.27
Specific Enthalpy (kJ/kg)[A]	-9,051.9	-101.7	574.1	-2.8	-8,479.2	-8,960.4
Density (kg/m^3)	6.7	1.2	7.4	0.9	4.6	667.1
V-L Molecular Weight	28.439	28.850	28.855	28.520	30.019	41.645
V-L Flowrate (lb_{mol}/hr)	29,336	133,953	133,953	121,048	29,649	13,793
V-L Flowrate (lb/hr)	834,267	3,865,180	3,865,179	3,452,230	890,037	601,568
Solids Flowrate (lb/hr)	0	0	0	0	0	0
Temperature (°F)	1323	59	1202	252	2202	100
Pressure (psia)	282	14.7	287	15.5	274	2215
Specific Enthalpy (Btu/lb)[A]	-3,891.6	-43.7	246.8	-1.2	-3,645.4	-3,852.3
Density (lb/ft^3)	0.418	0.076	0.461	0.058	0.287	41.645

Exhibit 4-24 Case 4-1 Plant Performance Summary (100 Percent Load)

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	
SOFC Power	561,268
Syngas Expander Power	8,951
Steam Turbine Power	57,286
TOTAL POWER, kWe	**627,505**
AUXILIARY LOAD SUMMARY, kWe	
Coal Handling	232
Coal Size Reduction	391
Catalyst-Coal Processing	1,519
Coal Feeding	836
Ash Handling	602
Air Separation Unit Auxiliaries	250
Air Separation Unit Main Air Compressor	12,196
Oxygen Compressor	4,770
Nitrogen Compression	427
Claus Tail Gas Recycle Compressor	672
CO_2 Compressor	12,453
BFW Pump	909
Condensate Pump	61
Circulating Water Pump	1,004
Ground Water Pumps	274
Cooling Tower Fans	763
Scrubber Pumps	143
Selexol Auxiliary Power	1,869
Steam Turbine Auxiliaries	19
Cathode Air Compressor-Expander	32,822
Claus / TGTU Auxiliaries	109
Miscellaneous Balance of Plant	2,444
Transformer Losses	2,359
TOTAL AUXILIARIES, kWe	**77,505**
NET POWER, kWe	**550,000**
Net Plant Efficiency, % (HHV)	**60.0**
Net Plant Heat Rate, kJ/kWh (Btu/kWh)	**5,998 (5,685)**
CONDENSER COOLING DUTY 10^6 kJ/h (10^6 Btu/h)	**317 (300)**
CONSUMABLES	
As-Received Coal Feed, kg/h (lb/h)	121,559 ([illegible])
Thermal Input[1], kWt	916,327
Raw Water Consumption, m^3/min (gpm)	1.5 (398)

Exhibit 4-25 Cases 4-1 Mass Flow Diagram



*Coal feed: ILL #6 as received

Main fuel stream mass flows (% of plant coal feed rate)
Non-fuel stream mass flows

Exhibit 4-26 Cases 4-1 Energy Flow Diagram

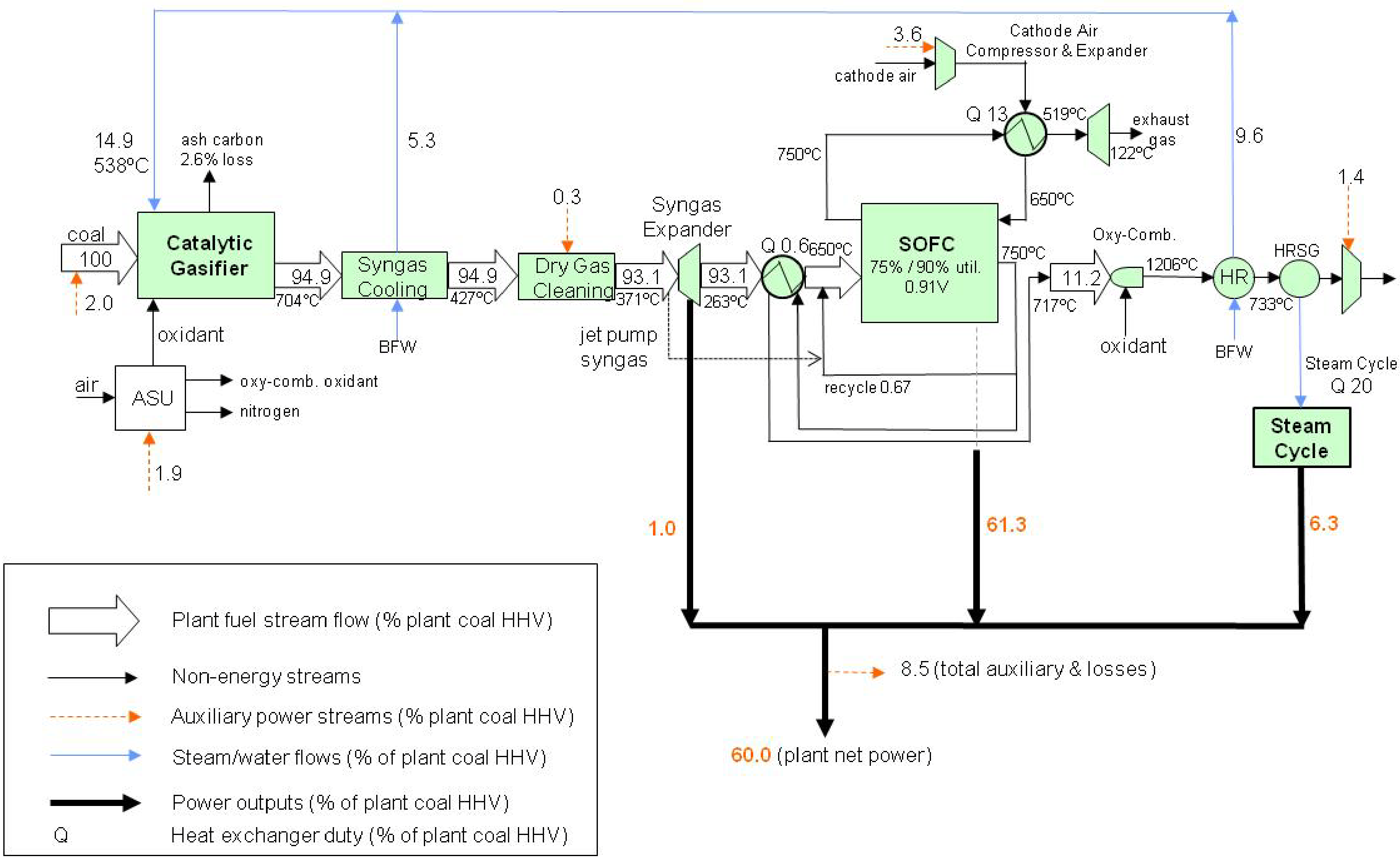


Exhibit 4-27 Case 4-1 High-Pressure Steam Balance

HP Process Steam Use, kg/h (lb/h)		HP Process Steam Generation, kg/h (lb/h)	
Gasifier feed	165,173 (364,146)	**Raw syngas cooling**	59,100 (130,293)
		Oxy-combustor heat recovery	106,073 (233,852)
Total	165,173 (364,146)	**Total**	165,173 (364,146)
HP Power-Steam generation, GJ/h (MMBtu/h)			
Raw syngas cooling	0 (0)		
Oxy-combustor heat recovery	523 (495)		
Total	523 (495)		

Exhibit 4-28 Case 4-1 Low-Pressure Steam Balance

LP Process Steam Use, GJ/h (MMBtu/h)		LP Process Steam Generation, GJ/h (MMBtu/h)	
Coal-Catalyst drying	25 (24)	**LT syngas cooling**	227 (215)
Selexol stripping	76 (72)	**Cathode exhaust cooling**	12 (11)
ASU	19 (18)		
Sour water stripping	92 (87)		
Coal preheat	27 (25)		
Total	239 (226)	**Total**	239 (226)

Exhibit 4-29 Case 4-1 Water Balance

	m^3/min (gpm)
Water Demand	**8.53 (2,255)**
Catalyst Treatment	0.47 (124)
Condenser Makeup *Gasifier Steam* *BFW Makeup*	2.83 (728) *2.75 (659)* *0.08 (20)*
Cooling Tower Makeup	5.23 (1,383)
Water Recovery for Reuse	**3.31 (874)**
Low-temperature Cooling	1.35 (355)
CO_2 Dehydration	1.96 (518)
Process Discharge Water	**1.51 (398)**
Cooling Tower Water Blowdown	1.18 (311)
Low-temperature Cooling	0.13 (36)
CO_2 Dehydration	0.20 (52)
Raw Water Consumed	**3.72 (983)**

Exhibit 4-30 Case 4-1 Carbon Balance

Carbon In, kg/h (lb/h)		Carbon Out, kg/h (lb/h)	
Coal	77,487 (170,831)	**Ash**	3,874 (8,542)
Gasifier Catalyst	1,657 (3,653)	**Catalyst CO_2 Recycle**	2,286 (5,039)
		Exhaust Gas	770 (1,698)
		CO_2 Product	72,214 (159,205)
Total	79,144 (174,484)	**Total**	79,144 (174,484)

Exhibit 4-31 Case 4-1 Sulfur Balance

Sulfur In, kg/h (lb/h)		Sulfur Out, kg/h (lb/h)	
Coal	3,047 (6,717)	**Elemental Sulfur**	3,042 (6,706)
		Polishing Sorbent	5 (11)
		CO_2 Product	0 (0)
Total	3,047 (6,717)	**Total**	3,047 (6,717)

Exhibit 4-32 Case 4-1 Air Emissions

	kg/GJ ($lb/10^6$ Btu)	Tonne/year (tons/year) 80% capacity factor	kg/MWh (lb/MWh)
SO_2	0 (0)	0 (0)	0 (0)
NO_X	0 (0)	0 (0)	0 (0)
Particulate	0 (0)	0 (0)	0 (0)
Hg	0 (0)	0 (0)	0 (0)
CO_2	0.23 (0.54)	5,397 (5,949)	1.40 (3.09)

4.3.1 Case 4-1 IGFC Plant Cost Results

The capital cost estimate for Case 4-1 is broken down in Exhibit 4-33. The dominant area costs are the gasification area and the SOFC power island. The single highest cost component in the plant is the SOFC stacks and inverters. Owner's costs presenting in Exhibit 4-34.

Exhibit 4-33 Case 4-1 Capital Cost Breakdown

	TOTAL PLANT COST	
Item/Description	**\$ x 1000**	**\$/kW**
COAL & SORBENT HANDLING	**23,197**	**42**
COAL & SORBENT PREP & FEED	**55,290**	**101**
Coal prep & feed system	46,318	84
Catalyst Treatment	8,972	16
FEEDWATER & MISC. BOP SYSTEMS	**11,235**	**20**
GASIFIER & ACCESSORIES	**167,332**	**304**
Gasifier & Syngas Cooler	91,354	166
ASU & Oxidant Compressor	64,852	118
Other Gasification Equipment	11,126	20
GAS CLEANUP & PIPING	**92,098**	**167**
Scrubber & Low Temperature Cooling	10,504	19
Single-Stage Selexol/MDEA	50,380	92
Claus Plant	18,131	33
Trace removal	1,261	2
COS Hydrolysis	5,546	10
Blowback, Piping, Foundations	3,139	6
Sulfur polishing	2,595	5
Heat Interchanger	543	1
CO_2 DRYING & COMPRESSION	**16,558**	**30**
SOFC POWER ISLAND	**340,754**	**620**
Syngas expander	2,240	4
SOFC Stack Units (stack modules, enclosures, inverters)	289,862	527
Cathode Air Compressor	23,071	42
Cathode Heat Exchanger	6,592	12
Cathode Gas Expander	9,151	17
Anode Heat Exchanger	21	0
Anode Syngas Jet Pump	189	0
Oxy-Combustor	9,629	18
HRSG, DUCTING & STACK	**25,284**	**46**
STEAM POWER SYSTEM	**17,741**	**32**
COOLING WATER SYSTEM	**13,683**	**25**
ASH/SPENT SORBENT HANDLING SYS	**34,451**	**63**
Ash handling	22,609	41
Catalyst Recovery	11,842	22
ACCESSORY ELECTRIC PLANT	**47,620**	**87**
INSTRUMENTATION & CONTROL	**27,743**	**50**
IMPROVEMENTS TO SITE	**12,736**	**23**
BUILDING & STRUCTURES	**11,783**	**21**
TOTAL PLANT COST (\$1000)	**897,505**	**1,632**

Exhibit 4-34 Case 4-1 Owner's Costs

Owner's Costs		
Preproduction Costs		
6 Months All Labor	10,123	18
1 Month Maintenance Materials	2,022	4
1 Month Non-fuel Consumables	1,154	2
1 Month Waste Disposal	388	1
25% of 1 Months Fuel Cost at 100% CF	933	2
2% of TPC	17,950	33
Total	**32,570**	**59**
Inventory Capital		
60 day supply of fuel and consumables at 100% CF	9,584	17
0.5% of TPC (spare parts)	4,488	8
Total	**14,072**	**26**
Initial Cost for Catalyst and Chemicals	4,968	9
Land	900	2
Other Owner's Costs	134,626	245
Financing Costs	24,233	44
Total Overnight Costs (TOC)	**1,108,873**	**2,016**
Total As-Spent Cost (TASC)	**1,264,115**	**2,298**

The first-year cost-of-electricity for Cases 4-1 is displayed in Exhibit 4-35. The dominant contributor to the COE is capital recovery, with fuel cost being relatively small because of the high plant efficiency. These are the same type of cost characteristics found in the other IGFC cases.

Exhibit 4-35 Case 4-1 Cost-of-Electricity Breakdown

					Annual Cost	Annual Unit Cost
OPERATING & MAINTENANCE LABOR					$	mills/kWh
Annual Operating Labor Cost					5,524,319	
Maintenance Labor Cost					10,672,984	
Administrative & Support Labor					4,049,326	
Property Taxes and Insurance					18,013,640	
TOTAL FIXED OPERATING COSTS					38,260,268	**9.3**
VARIABLE OPERATING COSTS						
Maintenance Material Cost					20,620,205	
Stack Replacement Cost					4,736,293	
Subtotal					**25,356,498**	
Consumables	Consumption Initial Fill	/Day	Unit Cost	Initial Cost		
Water (/1000 gallons)	0	2,802	1.08	0	940,454	
Chemicals						
MU & WT Chem. (lbs)	0	5,152	0.17	0	276,648	
Carbon (Trace Removal) (lb)	300,986	412	1.05	316,087	134,337	
COS Catalyst (m3)	228	0.16	2,397	546,097	116,046	
Selexol Solution (gal)	154,059	24.28	13.40	2,064,118	100,929	
Claus / DSRP Catalyst (ft3)	0	1.05	131	0	42,688	
ZnO polishing sorbent (lb)	124,079	1,224	1.50	186,119	569,701	
KOH Coal Catalyst makeup (lb)	4,978,176	82,970	0.16	796,508	4,118,611	
Lime for catalyst recovery	26,471,252	441,188	0.04	1,058,850	5,475,137	
Subtotal Chemicals				**4,967,779**	**10,834,096**	
Waste Disposal						
Spent Trace Catalyst (lb.)	0	445	0.42	0	58,024	
Ash (ton)	0	743	16.23	0	3,742,549	
Spent sorbents (lb)	0	1,224	0.42	0	159,516	
Subtotal-Waste Disposal ($)					**3,960,089**	
TOTAL VARIABLE OPERATING COSTS				**4,967,779**	**41,091,137**	**10.0**
Fuel Coal (ton)	0	2,914	38.18	0	38,046,708	**9.3**
Capital Recovery (mills/kWh)						**33.6**
TS&M (mills/kWh)						**3.6**
COE First Year (mills/kWh)						**65.9**

4.3.2 Scenario 4 Pathway Results

Scenario 4 pathway performance and cost estimates were performed for progressions in 1) the IGFC plant converted to a pressurized-SOFC power island following the Case 3-5 SOFC conditions; 2) the plant capacity factor, increased from 85 percent to 90 percent; and 3) the cell stack cost reduced by 20 percent, with the total cell cost (stacks and enclosures) dropping from 442 to 414 $/kW of SOFC power; and 4) the DC to AC inverter efficiency increased from 97 percent to 98 percent. The results are tabulated in

Exhibit 4-38.

The progression maintains the plant efficiency at about 60.0 percent (HHV), with the COE reduced to 62.3 mills/kWh. There are corresponding reductions in the plant capital investment and the raw water consumption rate is maintained at 1.8 gpm/MW. Compared to the Scenario 3 pathway results, the benefits of pressurized SOFC in the selected configuration of Scenario 4 does provide some performance benefit, but there is no cost benefit because of the high capital investment associated with pressurizing the SOFC system.

It is also of interest to observe some of the characteristics of the most expensive component systems in the plant, the gasifier, the SOFC stack units, and the SOFC power island. Exhibit 4-36 shows some key characteristics of the coal gasifier in the Scenario 4 pathway. The exit volumetric flow and cost do not change significantly along the pathway.

Exhibit 4-36 Scenario 4 Conventional Coal Gasifier Characteristics

Case	Gasifier Coal Feed Rate kg/hr (lb/hr)	Gasifier Exit Pressure MPa (psia)	Gasifier Exit Temperature °C (°F)	Gasifier Exit Syngas Rate 1000 m^3/h (1000 ft^3/h)	Gasifier & Heat Recovery Cost $1000
4-1	115,524 (254,687)	6.72 (975)	704 (1300)	16.4 (580)	92,021
4-2	114,330 (252,055)	6.72 (975)	704 (1300)	16.4 (580)	92,021
4-3	114,330 (252,055)	6.72 (975)	704 (1300)	16.4 (580)	92,021
4-4	114,330 (252,055)	6.72 (975)	704 (1300)	16.2 (574)	91,354

Exhibit 4-37 lists some of the SOFC characteristic along the Scenario 4 pathway. The SOFC current density remains fixed along the pathway, as the cell voltage remains at 0.937 V. The spare cell surface installed and the stack replacement times are the optimum values estimated. The SOFC stack unit cost and power island cost decrease for the Case 1-3 SOFC stack cost reduction.

Exhibit 4-37 Scenario 4 Pressurized SOFC Characteristics

Case	Cell Voltage, V	Power Density mW DC/cm^2	Current Density mA/cm^2	Spare Cell Surface Installed %	Stack Replacement Time years	SOFC Stack Unit Cost $/kW	Power Island Cost $/kW
4-1	0.912	500	548	16.8	11.3	527	621
4-2	0.912	500	548	16.8	10.7	527	621
4-3	0.912	500	548	16.8	10.7	494	587
4-4	0.912	500	548	16.8	10.7	494	587

Exhibit 4-38 Scenario 4 Pathway Results

Case	Pathway Description	Change Made	Coal Feed Rate kg/h (lb/h)	Number Parallel Trains	Cell Voltage V	Plant Efficiency %, HHV	Raw Water Consumed gpm/MW	CO_2 Emission g/MWh	Capital Cost, TOC $/kW	COE mills/kWh	Cost of CO_2 Avoided $/tonne
4-1	**Pressurized SOFC**	15.6 to 285 psia	115,524 (254,687)	1	0.912	59.4	1.81	5.7	2,026	66.1	9.1
4-2	**Capacity Factor**	85 to 90 %	115,524 (254,687)	1	0.912	59.4	1.81	5.7	2,026	63.5	5.9
4-3	**SOFC Stack Cost**	442 to 414 $/kW	115,524 (254,687)	1	0.912	59.4	1.81	5.7	1,986	62.6	4.7
4-4	**Inverter Efficiency**	97 to 98 %	114,330 (252,055)	1	0.912	60.0	1.79	5.7	1,976	62.3	4.3

5. IGFC with Natural Gas Injection (Case 1-6)

With high methane content in the syngas expected to be beneficial to the IGFC plant performance, another approach to achieving high methane content is to inject a portion of natural gas into the cleaned syngas stream, and this plant configuration was simulated in Scenario 1 as Case 1-6. Its detailed results are described here because it represents a unique configuration.

The Case 1-6 IGFC plant was simulated using the plant assumptions associated with the use of the enhanced-conventional gasifier with an atmospheric-pressure SOFC system. Natural gas, provided at 500 psia, was injected into the clean syngas before it was expanded in this plant. The major assumptions for the plant are listing in Exhibit 5-1. With the natural gas injection flow representing 38.5 percent of the total plant energy input, the syngas methane content is 24.6 mole percent (dry). The SOFC performance assumptions represent advanced technology assumptions.

Exhibit 5-1 IGFC with Natural Gas Injection (Case 1-6) Plant Assumptions

Gasifier (methane %)	Natural Gas % of total plant energy	SOFC Pressure & Overpotential	Capacity Factor %	Cell Degradation %/1000 h	SOFC Stack Cost $/kW	Inverter Efficiency %
Enhanced (24.6)	38.5	15.6 psia 70 mV	85	0.2	296	97

For the purpose of this study, natural gas is assumed to contain 5 ppmv of total sulfur constituents, and therefore must be desulfurized prior to use. The pipeline natural gas is assumed to contain no particulate matter or trace elements, resulting in no control requirements being needed. The natural gas is desulfurized from its assumed 5 ppmv total sulfur content to 100 ppbv total sulfur using low-temperature sorbent beds of the TDA Research Inc. SulfaTrap^{TM} sorbent [25]. The supply pressure of the natural gas is assumed to be 500 psia, and the natural gas is preheated after desulfurization, combined with clean coal syngas, and expanded across the plant expander to generate additional power.

The overall performance results are summarized in Exhibit 5-2. The plant power breakdown is listed in Exhibit 5-3.

Exhibit 5-2 IGFC with Natural Gas Injection (Case 1-6) Overall Performance Results

Anode Fuel	Ill #6 syngas mixed with Natural Gas
Net Power (kW)	550,000
Nernst potential (V)	0.93
Operating Potential (V)	0.86
Raw Water Consumption (gpm/MW)	2.05
Carbon Removal (% in syngas)	99.7
CO_2 Emission (kg/MWh)	1.33
Time to Replace Stack (years)	13.2
Net Plant Efficiency (%, HHV)	51.0

The plant capital cost breakdown is presented in Exhibit 5-4, and the owner's cost are shown in Exhibit 5-5. The cost-of-electricity breakdown is presented in Exhibit 5-6. Because of the use of natural gas injection, the Case 1-6 IGFC plant has very low capital investment, with a total overnight cost of only 1,794 $/kW. The COE is only 71.2 mills/kWh.

The use of natural gas injection into a coal gasifier syngas has the potential to provide an IGFC power plant with high performance and low cost using conventional power plant component technologies (coal gasification technology, gas cleaning technology, CO_2 dehydration and compression technology) other than the SOFC system and the oxy-combustion system. The combined use of coal and natural gas with other power plant technologies that incorporate CCS (PC or IGCC) will not result in similar benefits.

Exhibit 5-3 IGFC with Natural Gas Injection (Case 1-6) Power Summary

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	
SOFC Power	537,312
Syngas Expander Power	20,707
Steam Turbine Power	78,836
TOTAL POWER, kWe	**636,854**
AUXILIARY LOAD SUMMARY, kWe	
Coal handling	191
Coal size reduction	905
Sour water recycle slurry pumps	76
Ash handling	465
ASU Auxiliary power	426
ASU air compressor	20,397
Oxygen compressor	6,214
Nitrogen compression	309
Anode recycle compressor	3,253
Claus Tail Gas Recycle compressor	486
CO_2 compressor	32,797
BFW pump	1,251
Condensate pump	84
Syngas recycle compressor	207
Quench water pump	215
Circulating water pump	1,382
Ground water pump	211
Cooling tower fans	1,050
Scrubber pumps	98
Selexol auxiliary power	1,352
ST auxiliaries	26
Cathode air blower	5,095
Cathode recycle blower	5,409
Claus / TGTU auxiliaries	79
BOP	2,480
Transformer losses	2,395
TOTAL AUXILIARIES, kWe	**86,854**
NET POWER, kWe	**550,000**
Net Plant Efficiency, % (HHV)	**51.0**
Net Plant Heat Rate, kJ/kWh (Btu/kWh)	**7,061 (6,693)**
CONDENSER COOLING DUTY 10^6 kJ/h (10^6 Btu/h)	**306 (290)**
CONSUMABLES	
As-Received Coal Feed, kg/h (lb/h)	87,954 (193,905)
Natural Gas Feed, kg/h (lb/h)	28,610 (63,075)
Thermal Input[1], kWt	1,078,870
Raw Water Consumption, m^3/min (gpm)	4.3 (1,130)

Exhibit 5-4 IGFC with Natural Gas Injection (Case 1-6) Capital Investment

Item/Description	$ x 1000	$/kW
COAL & SORBENT HANDLING	**19,306**	**35**
COAL & SORBENT PREP & FEED	**24,297**	**44**
FEEDWATER & MISC. BOP SYSTEMS	**11,582**	**21**
GASIFIER & ACCESSORIES	**215,744**	**392**
Gasifier & Syngas Cooler	115,521	210
ASU & Oxidant Compressor	91,763	167
Other Gasification Equipment	8,459	15
GAS CLEANUP & PIPING	**75,892**	**138**
Scrubber & Low Temperature Cooling	12,095	22
Single-Stage Selexol/MDEA	38,151	69
Claus Plant	14,439	26
Trace removal	1,105	2
COS Hydrolysis	3,774	7
Blowback, Piping, Foundations	1,951	4
Sulfur polishing	4,378	8
CO_2 DRYING & COMPRESSION	**32,614**	**59**
SOFC POWER ISLAND	**249,590**	**454**
Syngas expander	4,576	8
SOFC Stack Units (stack modules, enclosures, inverters)	190,330	346
Cathode Air Blower	2,004	4
Cathode Gas Recycle Blower	4,630	8
Cathode Heat Exchanger	32,879	60
Anode Heat Exchanger	4,530	8
Anode Recycle Blower	805	1
Oxy-Combustor	9,838	18
HRSG, DUCTING & STACK	**19,864**	**36**
STEAM POWER SYSTEM	**22,185**	**40**
COOLING WATER SYSTEM	**14,588**	**27**
ASH/SPENT SORBENT HANDLING SYS	**16,129**	**29**
ACCESSORY ELECTRIC PLANT	**51,572**	**94**
INSTRUMENTATION & CONTROL	**27,743**	**50**
IMPROVEMENTS TO SITE	**10,600**	**19**
BUILDING & STRUCTURES	**9,806**	**18**
TOTAL PLANT COST ($1000)	**804,042**	**1,462**

Exhibit 5-5 IGFC with Natural Gas Injection (Case 1-6) Owner's Costs

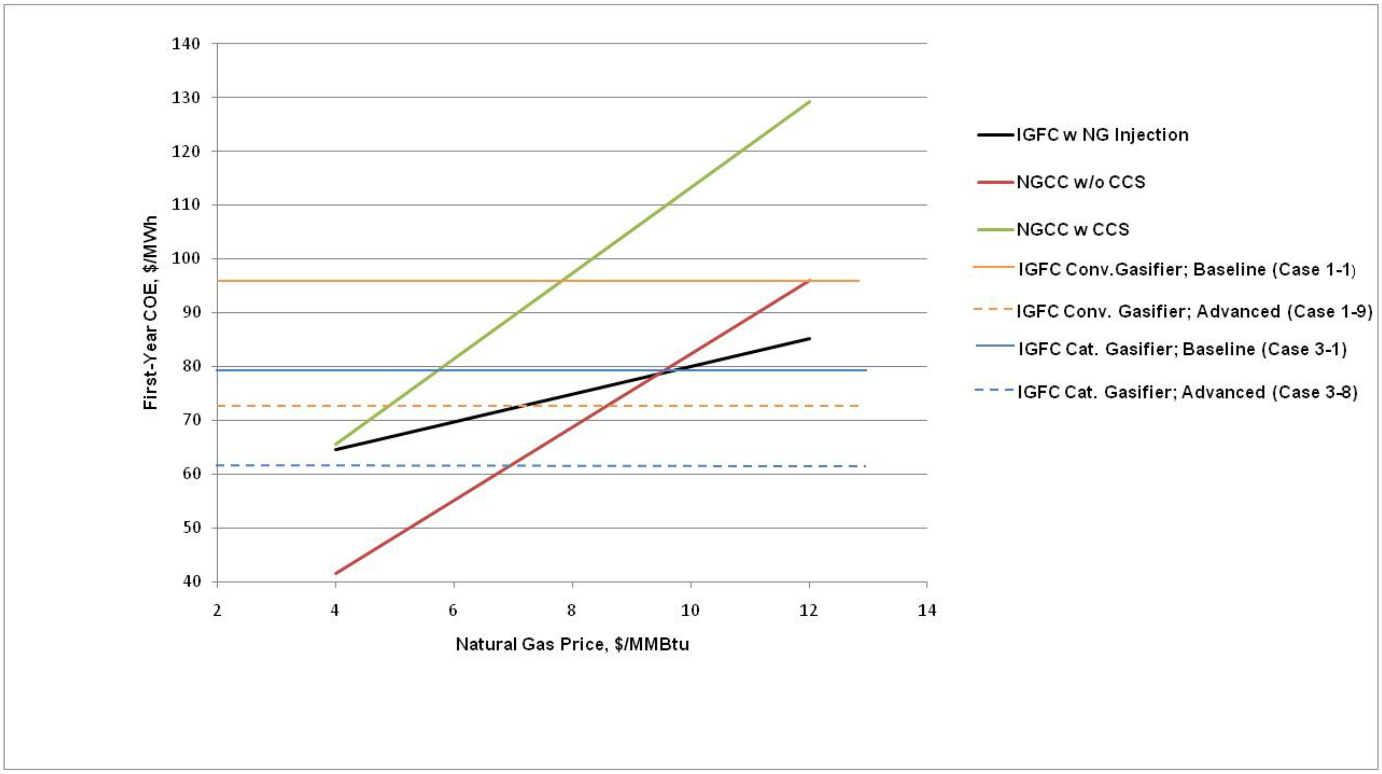
Owner's Costs		
Preproduction Costs		
6 Months All Labor	8,677	16
1 Month Maintenance Materials	1,733	3
1 Month Non-fuel Consumables	182	0
1 Month Waste Disposal	133	0
25% of 1 Months Fuel Cost at 100% CF	676	1
2% of TPC	16,081	29
Total	**27,481**	**50**
Inventory Capital		
60 day supply of fuel and consumables at 100% CF	5,596	10
0.5% of TPC (spare parts)	4,020	7
Total	**9,616**	**17**
Initial Cost for Catalyst and Chemicals	2,595	5
Land	900	2
Other Owner's Costs	120,606	219
Financing Costs	21,709	39
Total Overnight Costs (TOC)	**986,950**	**1,794**
Total As-Spent Cost (TASC)	**1,125,123**	**2,046**

Exhibit 5-6 IGFC with Natural Gas Injection (Case 1-6) COE

					Annual Cost	Annual Unit Cost
OPERATING & MAINTENANCE LABOR					$	mills/kWh
Annual Operating Labor Cost					4,735,130	
Maintenance Labor Cost					9,148,272	
Administrative & Support Labor					3,470,851	
Property Taxes and Insurance					16,033,001	
TOTAL FIXED OPERATING COSTS					33,387,254	**8.2**
VARIABLE OPERATING COSTS						
Maintenance Material Cost					17,674,461	
Stack Replacement Cost					3,569,801	
Subtotal					**21,244,262**	
Consumables	nsumption		Unit	Initial		
	Initial Fill	/Day	Cost	Cost		
Water (/1000 gallons)	0	2,627	1.08	0	**881,831**	
Chemicals						
MU & WT Chem. (lbs)	0	4,831	0.17	0	259,404	
Carbon (Trace Removal) (lb)	231,549	317	1.05	243,165	103,345	
COS Catalyst (m3)	175	0.12	2,397	420,112	89,274	
Selexol Solution (gal)	118,517	18.68	13.40	1,587,923	77,644	
Claus / DSRP Catalyst (ft3)	0	0.81	131.27	32,840	229,272	
ZnO polishing sorbent (lb)	229,272	886	1.50	343,908	412,206	
Subtotal Chemicals				**2,595,109**	**974,712**	
Waste Disposal						
Spent Trace Catalyst (lb.)	0	343	0.42	0	44,638	
Ash (ton)	0	238	16.23	0	1,195,910	
Spent sorbents (lb)	0	886	0.42	0	115,418	
Subtotal-Waste Disposal ($)					**1,355,966**	
TOTAL VARIABLE OPERATING COSTS				**2,595,109**	**24,456,771**	**6.0**
Fuel Coal (ton)	0	2,327	38.18	0	**27,565,781**	**6.7**
Fuel Natural Gas (MMBtu)	0	1,419	6.55		**69,202,979**	**16.9**
Capital Recovery (mills/kWh)						**29.9**
TS&M (mills/kWh)						**3.6**
COE First Year (mills/kWh)						**71.3**

The COE for the IGFC with natural gas injection is compared to the COE for NGCC with and without CCS as a function of the natural gas price in Exhibit 5-7. Also plotted are the COEs for IGFC representing four cases: IGFC with a conventional gasifier, and atmospheric-pressure SOFC at the baseline SOFC conditions (Case 1-1); IGFC with a conventional gasifier, and atmospheric-pressure SOFC at advanced conditions (Case 1-9); IGFC with a catalytic gasifier, and atmospheric-pressure SOFC at the baseline SOFC conditions (Case 3-1); and IGFC with a catalytic gasifier, and atmospheric-pressure SOFC at advanced SOFC conditions (Case 3-8).

Exhibit 5-7 IGFC with Natural Gas Injection COE Comparison with NGCC



Because the plant efficiency for IGFC with natural gas injection, using the assumed advanced SOFC conditions, is greater than that of NGCC with and without CCS, and the IGFC plant is only partially fueled with natural gas, the COE of IGFC with natural gas injection does not increase as fast with increasing natural gas price. The COE of IGFC with natural gas injection is significantly lower than that of NGCC with CCS, and reaches parity with NGCC without CCS at a natural gas price of about 10 $/MMBtu. IGFC with natural gas injection can have COE lower than IGFC with conventional gasification or catalytic gasification under baseline SOFC conditions. IGFC with natural gas injection can also achieve COE that is comparable to the IGFC COE with advanced SOFC performance and cost assumptions. The use of natural gas injection into the coal-syngas stream provides an opportunity to achieve significant IGFC plant performance and cost enhancements with limited need for advanced technology development.

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