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Techno-Economic Analysis of Integrated Gasification Fuel Cell Systems

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Arun K.S. Iyengar, Richard A. Newby, Dale L. Keairns
Booz Allen Hamilton, Inc.

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Table of Contents

Exhibits

Acronyms and Abbreviations

AACE	Association for the Advancement of Cost Engineering
AC	Alternating current
AFUDC	Allowance for funds used during construction
AGR	Acid gas removal
Aspen	Aspen Plus®
ASR	Area specific resistance
ASU	Air separation unit
atm	Atmospheric
AUSC	Advanced ultrasupercritical
BACT	Best available control technology
BB	Bituminous Baseline
BEC	Bare erected cost
BFD	Block flow diagram
BFW	Boiler feedwater
BOP	Balance of plant
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
CEPCI	Chemical Engineering Plant Cost Index
CF	Capacity factor
CH_4	Methane
cm	Centimeter
CO	Carbon monoxide
CO_2	Carbon dioxide
COE	Cost of electricity
CoP	ConocoPhillips
COS	Carbonyl sulfide
CPU	CO_2 purification unit
Cref	Cost of equipment in reference plant area or section
CRT	Cathode ray tube
DC	Direct current
DCS	Distributed control system
DI	De-ionized
DOE	Department of Energy
DSRP	Direct sulfur recovery process
EAF	Equivalent availability factor
E-Gas™	ConocoPhillips gasifier technology
EIA	Energy Information Administration
EOR	Enhanced oil recovery
EPA	Environmental Protection Agency
EPC	Engineer/Procure/Construct
EPCM	Engineering/Procurement/Construction Management
EPRI	Electric Power Research Institute
ESPA	Energy Sector Planning and Analysis
F	Capacity of study plant area or section
FEED	Front-End Engineering Design
Fref	Capacity of reference plant area or section
ft	Foot, Feet
ft^3	Cubic feet
gal	Gallon
GEE	General Electric Energy
GHG	Greenhouse gas
gpm	Gallons per minute
h, hr	Hour
H_2	Hydrogen
H_2S	Hydrogen sulfide
HCl	Hydrogen chloride
Hg	Mercury
HGC	Humid gas cleaning
HHV	Higher heating value
hp	Horsepower
HP	High pressure
HRSG	Heat recovery steam generator
HTX	Heat Exchanger
HVAC	Heating, ventilation, and air conditioning
I&C	Instrumentation and control
IGCC	Integrated gasification combined cycle
IGFC	Integrated gasification fuel cell
IOU	Investor-owned utility
IP	Intermediate pressure
ISO	International Standards Organization
ITM	Ion transport membrane
kg/GJ	Kilogram per gigajoule
kg/h	Kilogram per hour
kJ	Kilojoules
kJ/h	Kilojoules per hour
kJ/kg	Kilojoules per kilogram
KOH	Potassium hydroxide
kV	Kilovolt
kW	Kilowatt
kWe	Kilowatts electric
kWh	Kilowatt-hour
kWt	Kilowatts thermal
lb	Pound
lb/hr	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
mA/cm^2	Milliamps per square cm
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	Million kilojoules (also shown as 10^6 kJ)
MMkJ/h	Million kilojoules (also shown as 10^6 kJ) per hour
MPa	Megapascals
MU	Make up
mV	millivolt
MWe	Megawatt electric
MWh	Megawatt-hour
N	Number of study plant areas or sections in parallel
Nref	Number of reference plant areas or sections in parallel
N/A	Not applicable
NERC	North American Electric Reliability Council
NETL	National Energy Technology Laboratory
NG	Natural gas
NGCC	Natural gas combined cycle
NGFC	Natural gas fuel cell
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	Number of atoms of oxygen in the syngas
PC	Pulverized coal
PFD	Process flow diagrams
POTW	Publicly owned treatment works
ppbv	Parts per billion volume
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
QGESS	Quality Guidelines for Energy System Studies
R&D	Research and development
S	Scaling factor for plant areas or section cost
SC	Supercritical
SCOT	Shell Claus Off-gas Treating
SCR	Selective catalytic reduction
SG	Specific gravity
SGC	Synthesis gas cooler
SGS	Sour gas shift
SNG	Synthetic natural gas
SO_2	Sulfur dioxide
SOFC	Solid oxide fuel cell
SRU	Sulfur recovery unit
Syngas	Synthesis gas
SWS	Sour Water Stripper
T	Temperature
T&S	Transport and storage
TASC	Total as-spent cost
TG	Tail gas
TGTU	Tail gas treating unit
TOC	Total overnight cost
ton	Short ton (2000 lbs)
Tonne	Metric Ton (1000 kg)
TPC	Total plant cost
TPD	Tons per day
T&S	?Transport and storage
TS&M	Transport, storage, and monitoring
U.S.	United States
V	Volts
V-L	Vapor Liquid portion of stream (excluding solids)
vol%	Volume percent
WT	Water treatment
wt%	Weight percent
ZnO	Zinc oxide
$/MMBtu	Dollars per million British thermal units
$/MMkJ	Dollars per million kilojoule
°C	Degrees Celsius
°F	Degrees Fahrenheit

Executive Summary

The results of a pathway study for coal-based, integrated gasification fuel cell (IGFC) power systems with carbon capture and storage (CCS) are presented in this report. They represent the potential future benefits of IGFC technology development by quantifying the performance and cost benefits for a series of projected gains made through the development of advanced technologies for improvements in plant operation and maintenance. In addition, the effectiveness of an IGFC system without CCS in meeting the proposed greenhouse gas (GHG) emission limit is discussed using results from select cases. The objective of the study is to provide guidance to the research and development (R&D) program of the Department of Energy (DOE).

The IGFC power plant is analogous to an integrated gasification combined cycle (IGCC) power plant, but with the gas turbine power island replaced with a solid oxide fuel cell (SOFC) power island. The SOFC provides for high electric efficiencies associated with the nearly reversible electrochemical conversion of syngas chemical potential to electric power, as compared to a conventional Carnot-driven heat engine. In addition, the clean oxy-reaction of syngas in the sealed fuel cell system renders itself readily available for CCS with the requirement of only a small oxy-combustor downstream of the fuel cell to react the fuel that is not utilized electrochemically. The heat rejected by the fuel cell system can be recovered further in a combination of Brayton and Rankine cycles, depending on whether the fuel cell system is operating at elevated or atmospheric pressures. The only other exhaust gas stream in the plant is the cathode off-gas which is uncontaminated, vitiated air.

The present study assumes the use of advanced, planar, SOFC technology featuring separated anode and cathode off-gas steams with anode off-gas oxy-combustion for cases with carbon capture. The SOFC simulations utilize the expected operating conditions and performance capabilities of the technology, operating initially at atmospheric pressure. The power plant cost and performance are estimated based not only on the current state of SOFC development but also on a projected pathway of SOFC technology development advances. In particular, the following SOFC system advances are incorporated in a cumulative manner:

- Reduction of SOFC stack performance degradation
- Reduction of stack overpotential (under normal operating conditions)
- Reduction of SOFC stack cost
- Improvement of inverter efficiency
- Operation of SOFC under pressurized conditions

General advances in IGFC plant operation are also included in the pathway in the form of improved plant availability and increased capacity factor, which are assumed to be achieved through advanced component monitoring, improved maintenance practices, and plant operation experience.

The overall plant performance and costs estimates of two parallel pathways of IGFC development are considered here. The first pathway utilizes conventional coal gasification technology, and features the ConocoPhillips E-Gas™ gasifier (CoP) with syngas methane (CH_4) content of ~ 6 mole percent. Two variants of this pathway system are considered based on the SOFC operating pressure. Scenario 1 investigates SOFC systems operating at atmospheric pressure and progressively includes pertinent SOFC technology advances. Both the near-term enhancement in the conventional gasifier technology and the injection of natural gas into the syngas stream were also considered in this scenario to boost the CH_4 content of the syngas, which has been projected to benefit the performance of the IGFC plant. The incorporation of pressurized-SOFC technology as a longer term enhancement is considered in Scenario 2.

The second pathway utilizes an advanced, catalytic coal gasification technology projected to produce syngas having a high CH_4 content of ~ 30 mole percent, which considerably improves the IGFC plant performance. Scenarios 3 and 4 represent, respectively, the atmospheric and pressurized SOFC variants under this pathway, which follows similar advances in SOFC technology development as in the first pathway with the conventional gasifier.

Summaries of plant configurations and pathway parameters considered in this study are presented in Exhibit ES-1 and Exhibit ES-2. 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.

The design and cost bases for this evaluation are largely based on National Energy Technology Laboratory's (NETL) Bituminous Baseline (BB) report, (1) to facilitate a direct comparison 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. The IGFC plants are designed for baseload operation with the following key design basis specifications:

- Illinois No. 6 coal
- International Standards Organization (ISO) ambient conditions
- Conventional cryogenic air separation technology
- Conventional dry syngas cleaning and polishing technology
- Cryogenic distillation-based carbon dioxide (CO_2) purification process to meet enhanced oil recovery (EOR) specifications for the CO_2 product stream
- Net plant capacity of 550,000 kW
- Coal price is assumed to be \$68/ton and the NG price was assumed to be \$6.13/MMBtu

In practice, degradation of SOFC stack performance, a major contributor to production costs, is mitigated by providing additional capacity in the form of extra stack surface. A constant power output is maintained by operating the stack at a voltage above the design voltage (and the current below the design value) initially, and increasing the current subsequently as the stack voltage declines; the system efficiency, however, varies from a value that is higher than the nominal value to a value below it over the stack lifetime. The corresponding stack operational scenarios, which effectively compensate for stack degradation, were modeled to optimize the extra area installed and evaluate the corresponding stack replacement period. Both linear and first-order stack degradation models at various degradation rates were investigated in the present study. It was found that the first order degradation assumption results in stack life values generally 25 percent higher than the values computed assuming a linear degradation. For the projected SOFC stack degradation rate of 0.2 percent per 1000 h, installation of 10 percent extra area was found to be an optimum. The corresponding stack life was predicted to be about ~ 6.4 years, and 8.1 years for the linear and the first order models, respectively. Accordingly, a stack with additional 10 percent area with an average stack replacement period of 7.3 years was assumed for the Nth of a kind IGFC unit in the cost of electricity calculations.

Exhibit ES-1 Conventional gasifier IGFC pathway parameters (Scenarios 1 and 2)

<table>
<tr><th>Case</th><th>1-1</th><th>1-2</th><th>1-3</th><th>1-4</th><th>1-5</th><th>1-6</th><th>1-7</th><th>1-8</th><th>1-9</th><th>2-1</th><th>2-2</th><th>2-3</th><th>2-4</th></tr>
<tr><th>Pathway Parameter</th><td>Baseline</td><td>Degradation</td><td>Overpotential</td><td>Capacity Factor</td><td>Gasifier</td><td>NG Injection (34%)[2]</td><td>Capacity Factor</td><td>SOFC Cost Reduction</td><td>Inverter Efficiency</td><td>Pressurized SOFC</td><td>Capacity Factor</td><td>SOFC Cost Reduction</td><td>Inverter Efficiency</td></tr>
<tr><th>SOFC Pressure (psia)</th><td colspan="9">Atmospheric (15.6 psia)</td><td colspan="4">Pressurized (285 psia)</td></tr>
<tr><th>SOFC Degradation Rate (%/1000h)</th><td>1.5</td><td colspan="12">0.2</td></tr>
<tr><th>SOFC Overpotential (mV)</th><td colspan="2">140</td><td colspan="11">70</td></tr>
<tr><th>Capacity Factor (%)</th><td colspan="3">80</td><td colspan="3">85</td><td colspan="3">90</td><td>85</td><td colspan="3">90</td></tr>
<tr><th>Dry Syngas CH_4 Content (%)</th><td colspan="4">5.8</td><td>10.8</td><td>24.6</td><td colspan="7">10.8</td></tr>
<tr><th>SOFC Stack Cost ($/kW)[1]</th><td colspan="7">225</td><td colspan="2">200</td><td colspan="2">225</td><td colspan="2">200</td></tr>
<tr><th>Inverter Efficiency (%)</th><td colspan="8">97</td><td>98</td><td colspan="3">97</td><td>98</td></tr>
</table>

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

[2]*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)

<table>
<tr><th>Case</th><th>3-1</th><th>3-2</th><th>3-3</th><th>3-4</th><th>3-5</th><th>3-6</th><th>3-7</th><th>4-1</th><th>4-2</th><th>4-3</th><th>4-4</th></tr>
<tr><td>Pathway Parameter</td><td>Baseline</td><td>Degradation</td><td>Overpotential</td><td>Capacity Factor</td><td>Capacity Factor</td><td>SOFC Cost Reduction</td><td>Inverter Efficiency</td><td>Pressurized SOFC</td><td>Capacity Factor</td><td>SOFC Cost Reduction</td><td>Inverter Efficiency</td></tr>
<tr><td>SOFC Pressure (psia)</td><td colspan="7">Atmospheric (15.6 psia)</td><td colspan="4">Pressurized (285 psia)</td></tr>
<tr><td>Dry Syngas CH_4 Content %</td><td colspan="11">31.6</td></tr>
<tr><td>SOFC Degradation Rate (%/1000h)</td><td>1.5</td><td colspan="10">0.2</td></tr>
<tr><td>SOFC Overpotential (mV)</td><td colspan="2">140</td><td colspan="9">70</td></tr>
<tr><td>Capacity Factor (%)</td><td colspan="3">80</td><td>85</td><td colspan="3">90</td><td>85</td><td colspan="3">90</td></tr>
<tr><td>SOFC Stack Cost ($/kW)[1]</td><td colspan="5">225</td><td colspan="2">200</td><td colspan="2">225</td><td colspan="2">200</td></tr>
<tr><td>Inverter Efficiency (%)</td><td colspan="6">97</td><td>98</td><td colspan="3">97</td><td>98</td></tr>
</table>

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

The cost and performance of the cases are summarized in Exhibit ES-3, Exhibit ES-4, Exhibit ES-5, and Exhibit ES-6 for Scenarios 1 to 4. The costs are reported in 2011$ and are based on the NETL goal of $225/kW for the SOFC stack. The cost of electricity (COE) is computed using high-risk financial assumptions[1]. The exhibits show the increased performance and cost reduction that result from the maturity of IGFC plants supplemented with technological advancements. The total plant cost (TPC), the total overnight cost (TOC), and the total as-spent cost (TASC) are compared in Exhibit ES-7. The corresponding cost of COE including the charges estimated for CO_2 transport and storage (T&S) are plotted in Exhibit ES-8 for all the cases.

Although the TOCs for the baseline plants, Cases 1-1 and 3-1 for the CoP and the catalytic gasifier, respectively, are similar to the other cases within the corresponding pathway cases, their COE is significantly higher due to the 1.5 percent per 1000 h stack degradation rate imposed in this plant. The sensitivity of the COE and the cost of captured CO_2 to the stack degradation rate and stack cost, shown in Exhibit ES-9 and Exhibit ES-10, clearly illustrate the drawback of the high degradation rates associated with the current SOFC technology. They underline the current focus of NETL research program in fuel cells to reduce degradation rates, in addition to enhancement of SOFC performance, which is also required to meet specific NETL goals (20 percent reduction COE over a reference IGCC system with CCS, $40/tonne Cost of CO_2 relative to an appropriate reference plant without CCS).

The conventional gasifier pathway (Scenario 1) net plant efficiency shows gains of 7.0 percentage points from the reduction in SOFC cell overpotential. The enhanced gasifier is observed to limit this gain to 4 percentage points; the cooling benefit of the increased syngas CH_4 content in this case is overshadowed by the increased O_2 and H_2O demand from the gasifier (as modeled) and by the decrease in the SOFC inlet syngas Nernst voltage. The injection of natural gas corresponding to ~34 percent of the total thermal input into the IGFC syngas (Case 1-6) results in an additional 5 percentage-point gain in net plant efficiency.

The introduction of pressurized-SOFC into the conventional gasifier pathway (Scenario 2) results in an increase of ~2.2 percentage points in the net plant efficiency. However, due to the large increase in equipment cost with pressurization, the COE (without CO_2 T&S charges) increases slightly by ~$1/MWh (Case 1-5 versus Case 2-1, which uses similar SOFC performance and capacity factors).

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 (12 percent) 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 4.4 percentage points from the reduction in SOFC cell overpotential, being the only pathway parameter having significant influence on efficiency.

[1] The COE (excluding T&S) for case 1-9, which represents the pathway end point for a conventional gasifier-based IGFC, decreases from $96.4/MWh to $92.9/MWh using conventional financing assumptions.

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 59 percent (higher heating value [HHV]). As in the CoP gasifier case, pressurization is not found to be beneficial, the overall cost increases with pressurization (Case 3-4 versus Case 4-1, which use similar SOFC performance and capacity factors) due to the large increase in equipment cost with pressurization, which is not completely offset by the fuel cost savings associated with the increased pressurized SOFC system efficiencies.

The catalytic gasifier-based IGFC systems perform better and cost less with greater than 10 percentage-point gain in system efficiency, and higher than $13/MWh reduction in costs, when compared to conventional gasifier based systems.

Exhibit ES-3 Summary of cost and performance for conventional gasifier with atmospheric SOFC cases (Scenario 1)

CASE	1-1	1-2	1-3	1-4	1-5	1-6	1-7	1-8	1-9
SOFC Degradation Rate (%/1000 h)	1.5	0.2							
SOFC Overpotential (mV)	140		70						
Capacity Factor (%)	80			85			90		
Dry Syngas CH_4 Content (%)	5.8				10.8	24.6	10.8		
SOFC Stack Cost ($/kW)	225							200	
Inverter Efficiency (%)	97								98
PERFORMANCE									
SOFC Cell Voltage	0.816	0.816	0.885	0.885	0.877	0.866	0.877	0.877	0.877
Gross Power (MWe)	729.9	729.9	713.8	713.8	698.2	658.1	698.2	698.2	696.7
Auxiliary Loads (MWe)	179.9	179.9	163.7	163.7	148.0	107.9	148.0	148.0	146.6
Net Power (MWe)	550.0	550.0	550.1	550.1	550.2	550.2	550.2	550.2	550.1
Coal Flowrate (lb/hr)	406,900	406,900	378,000	378,000	352,250	204,600	352,250	352,250	348,675
NG Flowrate (lb/hr)	-	-	-	-	-	54,326	-	-	-
Net Electric Efficiency, HHV (%)	**39.5**	**39.5**	**42.6**	**42.6**	**45.7**	**51.9**	**45.7**	**45.7**	**46.1**
Net Plant Heat Rate, HHV (Btu/kWh)	8,630	8,630	8,017	8,017	7,468	6,569	7,468	7,468	7,394
CO_2 Capture rate (%)	98.6	98.6	98.6	98.6	98.6	98.5	98.6	98.6	98.6
CO_2 Emissions (lb/MWhgross)	23.6	23.6	21.5	21.5	20.1	17.8	20.1	20.1	19.9
CO_2 Emissions (lb/MWhnet)	31.3	31.3	27.9	27.9	25.5	21.2	25.5	25.5	25.3
Raw Water Consumption (gpm/MWnet)	2.3	2.3	2.0	2.0	2.7	2.0	2.7	2.7	2.0
COST									
Total Plant Cost (TPC) (1000$)	1,620,361	1,620,361	1,531,544	1,531,544	1,450,569	1,166,265	1,450,569	1,435,438	1,426,949
Total Overnight Cost (TOC) (1000$)	1,997,000	1,997,000	1,887,644	1,887,524	1,787,666	1,432,881	1,787,561	1,769,373	1,758,854
Total As-Spent Cost (TASC) (1000$)	2,276,579	2,276,579	2,151,914	2,151,777	2,037,939	1,633,485	2,037,820	2,017,085	2,005,093
Cost of Electricity ($/MWh)									
Variable COE	83.8	37.4	35.5	34.9	33.2	37.0	32.7	32.3	32.0
Fuel	25.4	25.4	23.6	23.6	22.0	26.4	22.0	22.0	21.7
Variable O&M	58.4	12.0	11.9	11.3	11.2	10.6	10.7	10.3	10.3
Fixed O&M	13.6	13.6	13.1	12.3	11.9	10.5	11.3	11.2	11.2
Capital Charges	64.4	64.4	60.9	57.3	54.2	43.5	51.2	50.7	50.4
Total First Year COE (excluding T&S)	**161.7**	**115.3**	**109.4**	**104.5**	**99.3**	**91.0**	**95.2**	**94.2**	**93.6**
T&S	8.4	8.4	7.8	7.8	7.3	5.5	7.3	7.3	7.2
Total First Year COE (including T&S)	**170.2**	**123.8**	**117.3**	**112.3**	**106.6**	**96.5**	**102.4**	**101.5**	**100.8**
NETL Metrics									
% COE reduction (COE_{IGFC} - $COE_{IGCC\ with\ CCS}$)/ $COE_{IGCC\ with\ CCS}$	-21.6	13.3	17.7	21.4	25.3	31.6	28.5	29.2	29.6
Cost of Captured CO_2 (\$/tonne CO_2) (COE_{IGFC} - $COE_{AUSC\ PC}$)/ CO_2 Captured (tonnes/MWh)	113.0	52.5	48.2	41.2	36.5	31.7	30.2	28.8	28.2

Exhibit ES-4 Summary of cost and performance for conventional gasifier with pressurized SOFC cases (Scenario 2)

CASE	2-1	2-2	2-3	2-4
SOFC Degradation Rate (%/1000 h)	0.2			
SOFC Overpotential (mV)	70			
Dry Syngas CH_4 Content (%)	10.8			
Capacity Factor (%)	85	90		
SOFC Stack Cost ($/kW)	225		200	
Inverter Efficiency (%)	97			98
PERFORMANCE				
SOFC Cell Voltage	0.949	0.949	0.949	0.949
Gross Power (MWe)	712.5	712.5	712.5	710.5
Auxiliary Loads (MWe)	162.2	162.2	162.2	160.4
Net Power (MWe)	550.3	550.3	550.3	550.1
Coal Flowrate (lb/hr)	349,900	349,900	349,900	346,000
NG Flowrate (lb/hr)	-	-	-	-
Net Electric Efficiency, HHV (%)	**46.0**	**46.0**	**46.0**	**46.5**
Net Plant Heat Rate, HHV (Btu/kWh)	7,418	7,418	7,418	7,337
CO_2 Capture rate (%)	98.6	98.6	98.6	98.6
CO_2 Emissions (lb/MWhgross)	20.9	20.9	20.9	20.7
CO_2 Emissions (lb/MWhnet)	27.1	27.1	27.1	26.8
Raw Water Consumption (gpm/MWnet)	2.6	2.6	2.6	2.5
COST				
Total Plant Cost (TPC) (1000$)	1,450,479	1,450,479	1,435,346	1,426,677
Total Overnight Cost (TOC) (1000$)	1,787,333	1,787,228	1,769,038	1,758,272
Total As-Spent Cost (TASC) (1000$)	2,037,560	2,037,440	2,016,704	2,004,430
Cost of Electricity ($/MWh)				
Variable COE	36.5	35.8	35.4	35.1
Fuel	21.8	21.8	21.8	21.6
Variable O&M	14.7	14.0	13.6	13.5
Fixed O&M	11.9	11.3	11.2	11.2
Capital Charges	54.2	51.2	50.7	50.4
Total First Year COE (excluding T&S)	**102.7**	**98.3**	**97.3**	**96.7**
T&S	7.2	7.2	7.2	7.2
Total First Year COE (including T&S)	**109.9**	**105.5**	**104.6**	**103.8**
NETL Metrics				
% COE reduction (COE_{IGFC} - $COE_{IGCC\ with\ CCS}$)/ $COE_{IGCC\ with\ CCS}$	22.8	26.1	26.8	27.3
Cost of Captured CO_2 ($/tonne CO_2) (COE_{IGFC} - $COE_{AUSC\ PC}$)/ CO_2 Captured (tonnes/MWh)	41.9	35.2	33.7	33.1

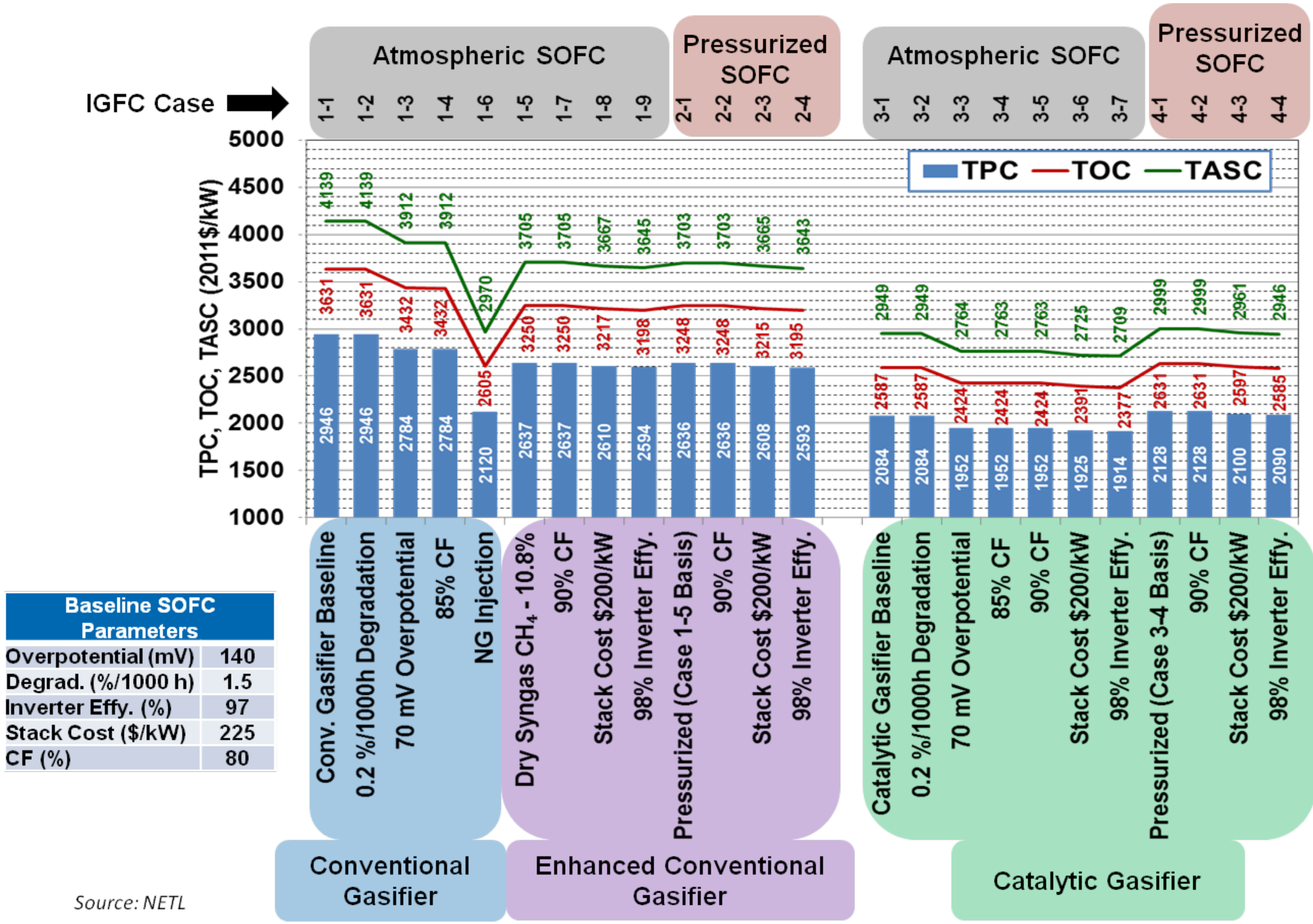
Exhibit ES-5 Summary of cost and performance for catalytic gasifier with atmospheric SOFC cases (Scenario 3)

CASE	3-1	3-2	3-3	3-4	3-5	3-6	3-7
Dry Syngas CH_4 Content (%)	31.6						
SOFC Degradation Rate (%/1000 h)	1.5	0.2					
SOFC Overpotential (mV)	140		70				
Capacity Factor (%)	80			85	90		
SOFC Stack Cost ($/kW)	225					200	
Inverter Efficiency (%)	97						98
PERFORMANCE							
SOFC Cell Voltage	0.796	0.796	0.862	0.862	0.862	0.862	0.862
Gross Power (MWe)	668.9	668.9	654.8	654.8	654.8	654.8	653.7
Auxiliary Loads (MWe)	118.6	118.6	104.7	104.7	104.7	104.7	103.7
Net Power (MWe)	550.3	550.3	550.1	550.1	550.1	550.1	550.0
Coal Flowrate (lb/hr)	329,238	329,238	300,800	300,800	300,800	300,800	297,610
NG Flowrate (lb/hr)	-	-	-	-	-	-	-
Net Electric Efficiency, HHV (%)	**48.9**	**48.9**	**53.5**	**53.5**	**53.5**	**53.5**	**54.1**
Net Plant Heat Rate, HHV (Btu/kWh)	6,980	6,980	6,380	6,380	6,380	6,380	6,312
CO_2 Capture rate (%)	98.5	98.5	98.5	98.5	98.5	98.5	98.5
CO_2 Emissions (lb/MWhgross)	20.5	20.5	18.1	18.1	18.1	18.1	17.9
CO_2 Emissions (lb/MWhnet)	25.0	25.0	21.5	21.5	21.5	21.5	21.3
Raw Water Consumption (gpm/MWnet)	2.4	2.4	2.2	2.2	2.2	2.2	2.1
COST							
Total Plant Cost (TPC) (1000$)	1,146,491	1,146,491	1,074,226	1,074,226	1,074,226	1,059,099	1,052,954
Total Overnight Cost (TOC) (1000$)	1,423,360	1,423,360	1,333,965	1,333,814	1,333,680	1,315,498	1,307,827
Total As-Spent Cost (TASC) (1000$)	1,622,631	1,622,631	1,520,720	1,520,548	1,520,395	1,499,668	1,490,922
Cost of Electricity ($/MWh)							
Variable COE	83.9	34.6	32.6	31.9	31.3	30.9	30.7
Fuel	20.4	20.4	18.6	18.6	18.6	18.6	18.4
Variable O&M	63.6	14.2	14.0	13.3	12.7	12.3	12.2
Fixed O&M	12.4	12.4	12.1	11.4	10.7	10.7	10.6
Capital Charges	45.9	45.9	43.0	40.5	38.2	37.7	37.5
Total First Year COE (excluding T&S)	**142.3**	**92.9**	**87.7**	**83.8**	**80.3**	**79.3**	**78.8**
CO_2 T&S	6.5	6.5	6.0	6.0	6.0	6.0	5.9
Total First Year COE (including T&S)	**148.8**	**99.5**	**93.7**	**89.7**	**86.2**	**85.3**	**84.7**
NETL Metrics							
% COE reduction (COE_{IGFC} - $COE_{IGCC\ with\ CCS}$)/ $COE_{IGCC\ with\ CCS}$	-7.0	30.1	34.1	37.0	39.6	40.4	40.8
Cost of Captured CO_2 ($/tonne CO_2) (COE_{IGFC} - $COE_{AUSC\ PC}$)/ CO_2 Captured (tonnes/MWh)	113.2	30.1	23.2	16.0	9.5	7.8	6.9

Exhibit ES-6 Summary of cost and performance for catalytic gasifier with pressurized SOFC cases (Scenario 4)

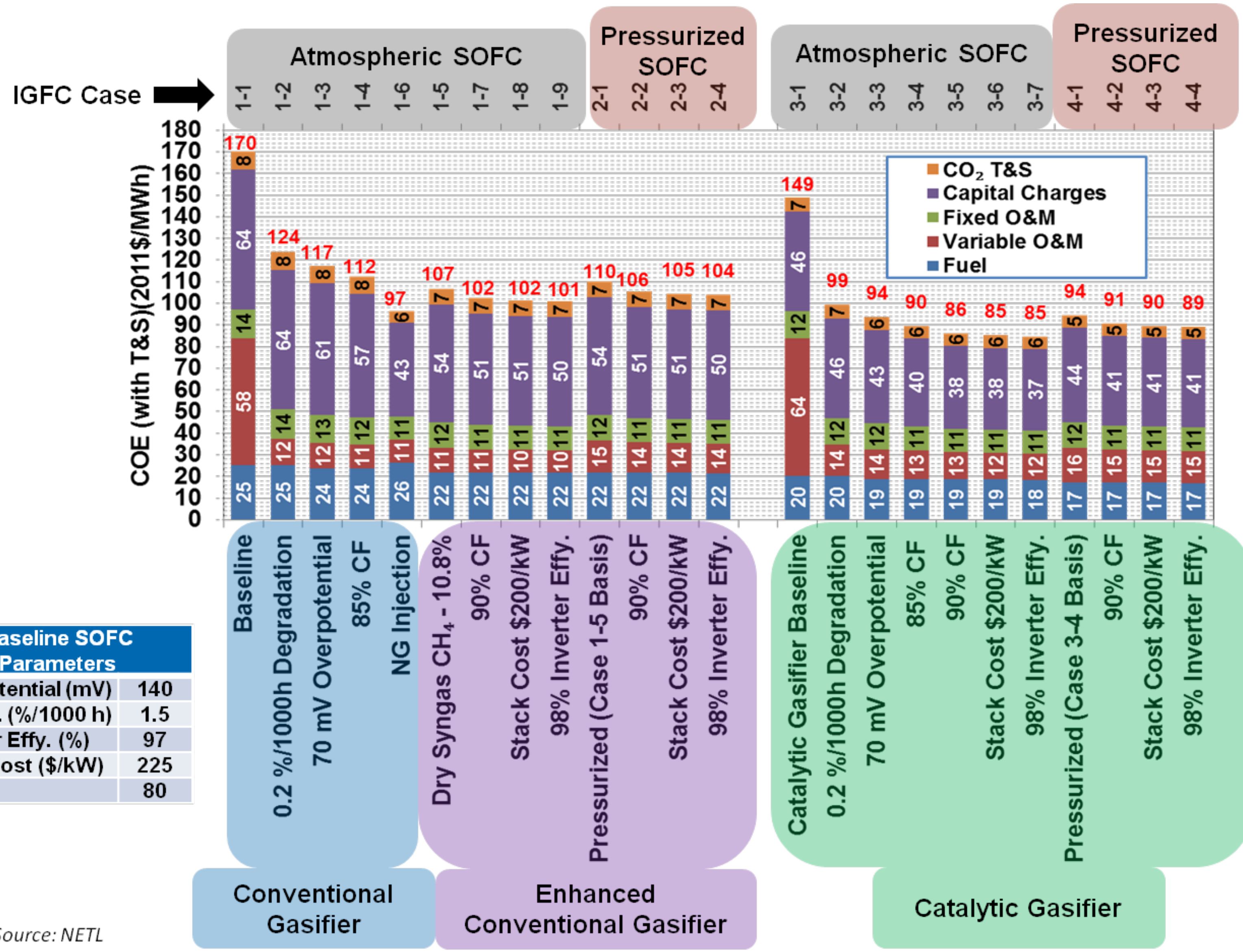
CASE	4-1	4-2	4-3	4-4
Dry Syngas CH_4 Content (%)	31.6			
SOFC Degradation Rate (%/1000 h)	0.2			
SOFC Overpotential (mV)	70			
Capacity Factor (%)	85	90		
SOFC Stack Cost ($/kW)	225		200	
Inverter Efficiency (%)	97			98
PERFORMANCE				
SOFC Cell Voltage	0.922	0.922	0.922	0.922
Gross Power (MWe)	651.0	651.0	651.0	650.0
Auxiliary Loads (MWe)	100.8	100.8	100.8	99.8
Net Power (MWe)	550.2	550.2	550.2	550.2
Coal Flowrate (lb/hr)	275,650	275,650	275,650	272,800
NG Flowrate (lb/hr)	-	-	-	-
Net Electric Efficiency, HHV (%)	**58.4**	**58.4**	**58.4**	**59.0**
Net Plant Heat Rate, HHV (Btu/kWh)	5,845	5,845	5,845	5,784
CO_2 Capture rate (%)	98.4	98.4	98.4	98.4
CO_2 Emissions (lb/MWhgross)	17.8	17.8	17.8	22.0
CO_2 Emissions (lb/MWhnet)	21.0	21.0	21.0	26.0
Raw Water Consumption (gpm/MWnet)	1.4	1.4	1.4	1.4
COST				
Total Plant Cost (TPC) (1000$)	1,170,703	1,170,703	1,155,573	1,149,842
Total Overnight Cost (TOC) (1000$)	1,447,479	1,447,344	1,429,157	1,422,015
Total As-Spent Cost (TASC) (1000$)	1,650,126	1,649,972	1,629,238	1,621,098
Cost of Electricity ($/MWh)				
Variable COE	33.2	32.4	32.1	31.8
Fuel	17.1	17.1	17.1	16.9
Variable O&M	16.1	15.4	15.0	14.9
Fixed O&M	11.8	11.1	11.1	11.1
Capital Charges	43.9	41.5	41.0	40.7
Total First Year COE (excluding T&S)	**88.9**	**85.1**	**84.1**	**83.6**
T&S	5.5	5.5	5.5	5.4
Total First Year COE (including T&S)	**94.4**	**90.5**	**89.6**	**89.0**
NETL Metrics				
% COE reduction (COE_{IGFC} - $COE_{IGCC\ with\ CCS}$)/ $COE_{IGCC\ with\ CCS}$	33.1	36.0	36.8	37.1
Cost of Captured CO_2 ($/tonne CO_2) (COE_{IGFC} - $COE_{AUSC\ PC}$)/ CO_2 Captured (tonnes/MWh)	27.9	20.1	18.2	17.3

Exhibit ES-7 Comparison of IGFC TPC, TOC, and TASC



Source: NETL

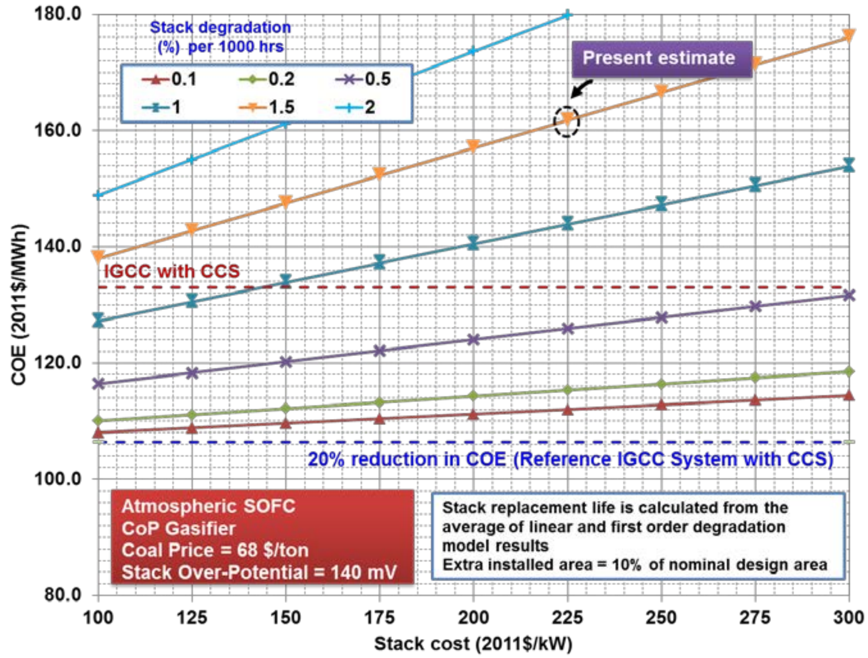
Exhibit ES-8 Comparison of IGFC COE (with CO_2 T&S)



Baseline SOFC Parameters	
Overpotential (mV)	140
Degrad. (%/1000 h)	1.5
Inverter Effy. (%)	97
Stack Cost ($/kW)	225
CF (%)	80

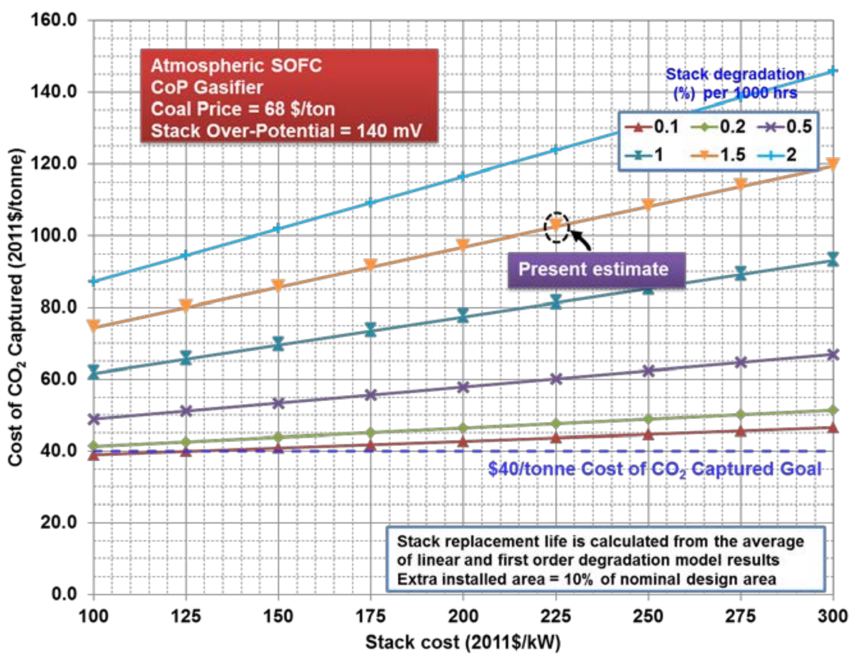
Source: NETL

Exhibit ES-9 Sensitivity of Case 1-1 COE to SOFC degradation rate and cost



Source: DOE/NETL

Exhibit ES-10 Sensitivity of Case 1-1 cost of captured CO_2 to SOFC degradation rate and cost



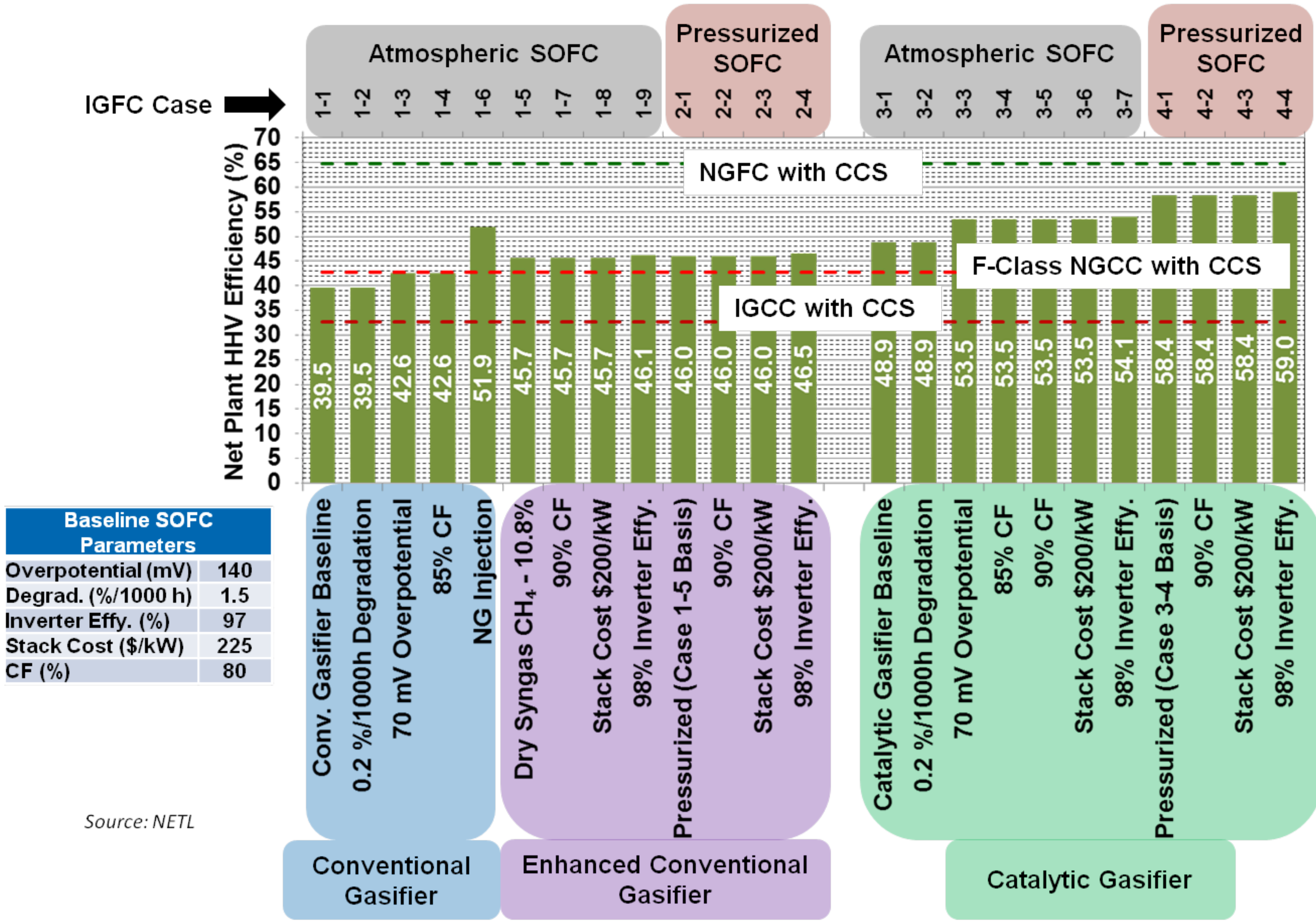
Source: DOE/NETL

It is clear that the performances of most of the IGFC plants evaluated in the study are superior to conventional fossil fuel technologies (with CCS) (1) as shown by the graphical comparison in Exhibit ES-11. While efficiency gains of ~7 percentage points over a comparable IGCC plant are possible even with current technology, combinations of advanced gasification technology and enhanced SOFC performance consistent with NETL's fuel cell program goals have the potential to yield efficiencies close to 60 percent. This corresponds to efficiency gains of ~26 and 16 percentage points over current IGCC and F-Class natural gas combined cycle (NGCC) systems. (1) However, as mentioned earlier, high performance degradation is a major drawback of current SOFC technology, which results in higher production costs and, consequently, higher COE for the baseline plants (conventional gasifier Case 1-1 and catalytic gasifier Case 3-1). As the SOFC stack degradation improves to the desired value of 0.2 percent per 1000 h, coupled with enhanced SOFC performance, the COE decreases dramatically, as shown in Exhibit ES-12, with most of the IGFC plants meeting the NETL goal of 20 percent reduction in COE over the reference IGCC system with CCS (COE = \$133/MWh) goal with a significant margin. The corresponding cost of CO_2 captured, shown in Exhibit ES-13, with reference to an advanced ultrasupercritical (AUSC) pulverized coal (PC) plant (COE = \$75.1/MWh), also follows a similar trend with most of the plants meeting the NETL \$40/tonne CO_2 goal, and supporting the use of EOR revenue stream to aid in the development of carbon capture systems.

Conventional fossil fuel power plants such as those characterized in the BB report (1) apply a design basis of 90 percent CO_2 removal. With the higher power conversion efficiencies in IGFC plants than conventional fossil fuel power plants, and CO_2 removal efficiencies that are greater than 98 percent, IGFC plant emissions of CO_2 are lower than in conventional fossil fuel power plants by a factor of 4 to 10 on a per MWh_{NET} basis (see Exhibit ES-3, Exhibit ES-4, Exhibit ES-5, and Exhibit ES-6). The emissions and other gas phase contaminants are also very limited in the IGFC power plants, because the SOFC imposes cleaning standards on the syngas that are considerably more stringent than current emission requirements for sulfur species, halides, and trace metals.

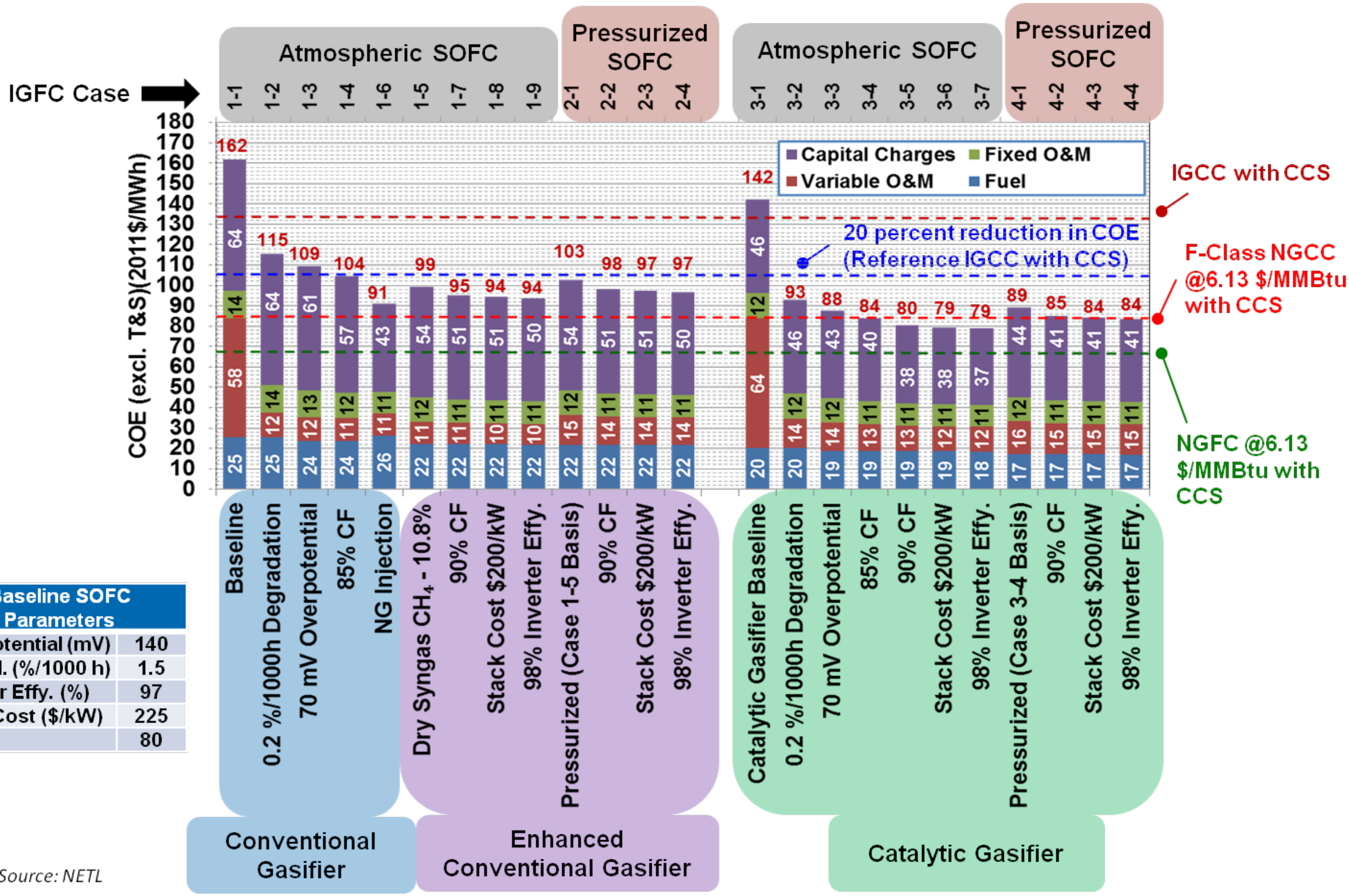
Another attractive feature of the IGFC systems is the considerably lower water consumption, compared to conventional fossil fuel power plants, as shown in Exhibit ES-14. The IGFC consumes 50 percent less raw water than even the water-economical NGCC system with CCS due to the higher IGFC plant efficiency, the recycle of the anode off-gas water vapor content, and the recovery of water through the carbon dioxide capture process.

Exhibit ES-11 Comparison IGFC HHV efficiencies



Source: NETL

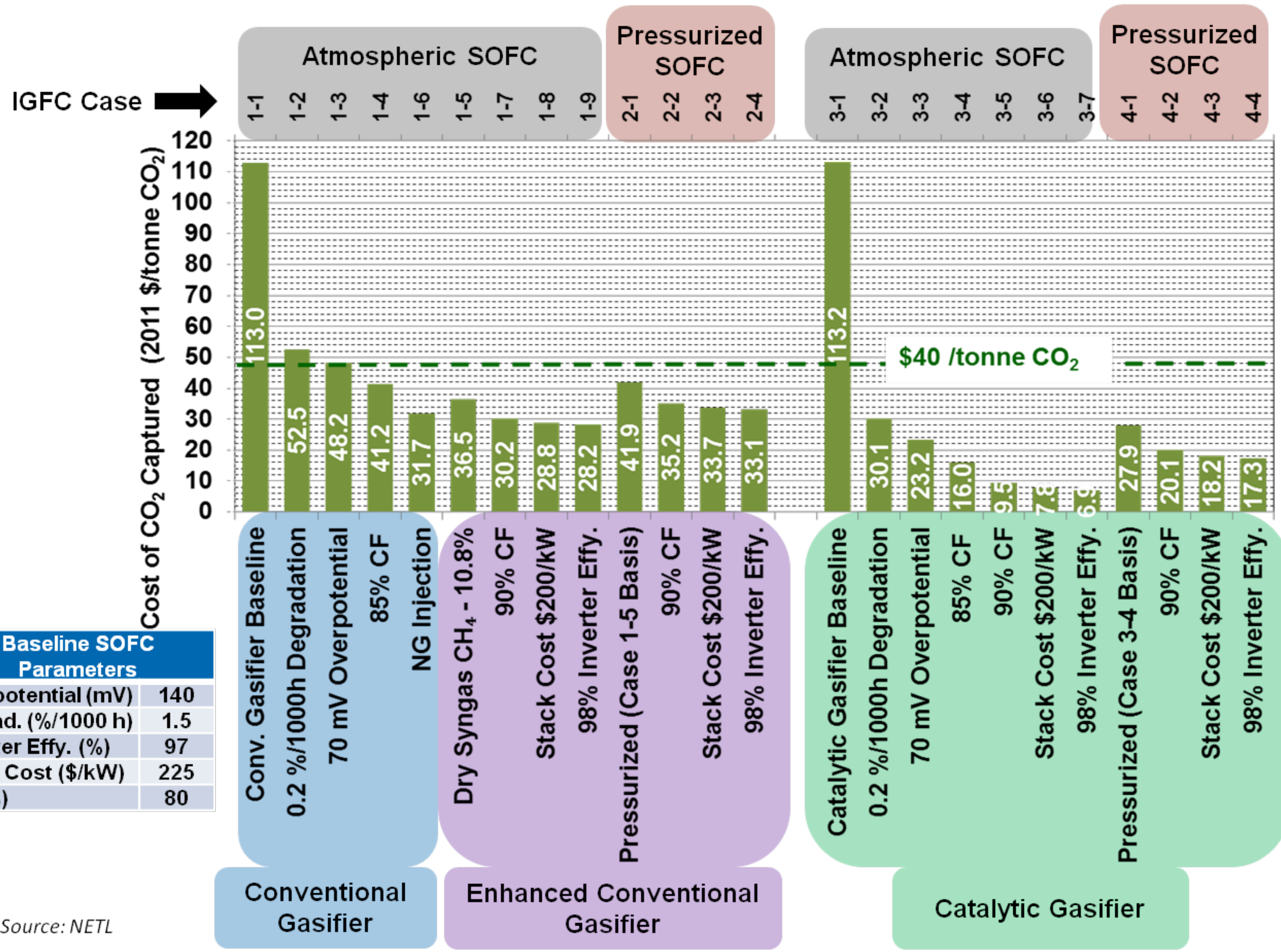
Exhibit ES-12 Comparison of IGFC COE (without CO_2 T&S)



Baseline SOFC Parameters	
Overpotential (mV)	140
Degrad. (%/1000 h)	1.5
Inverter Effy. (%)	97
Stack Cost ($/kW)	225
CF (%)	80

Source: NETL

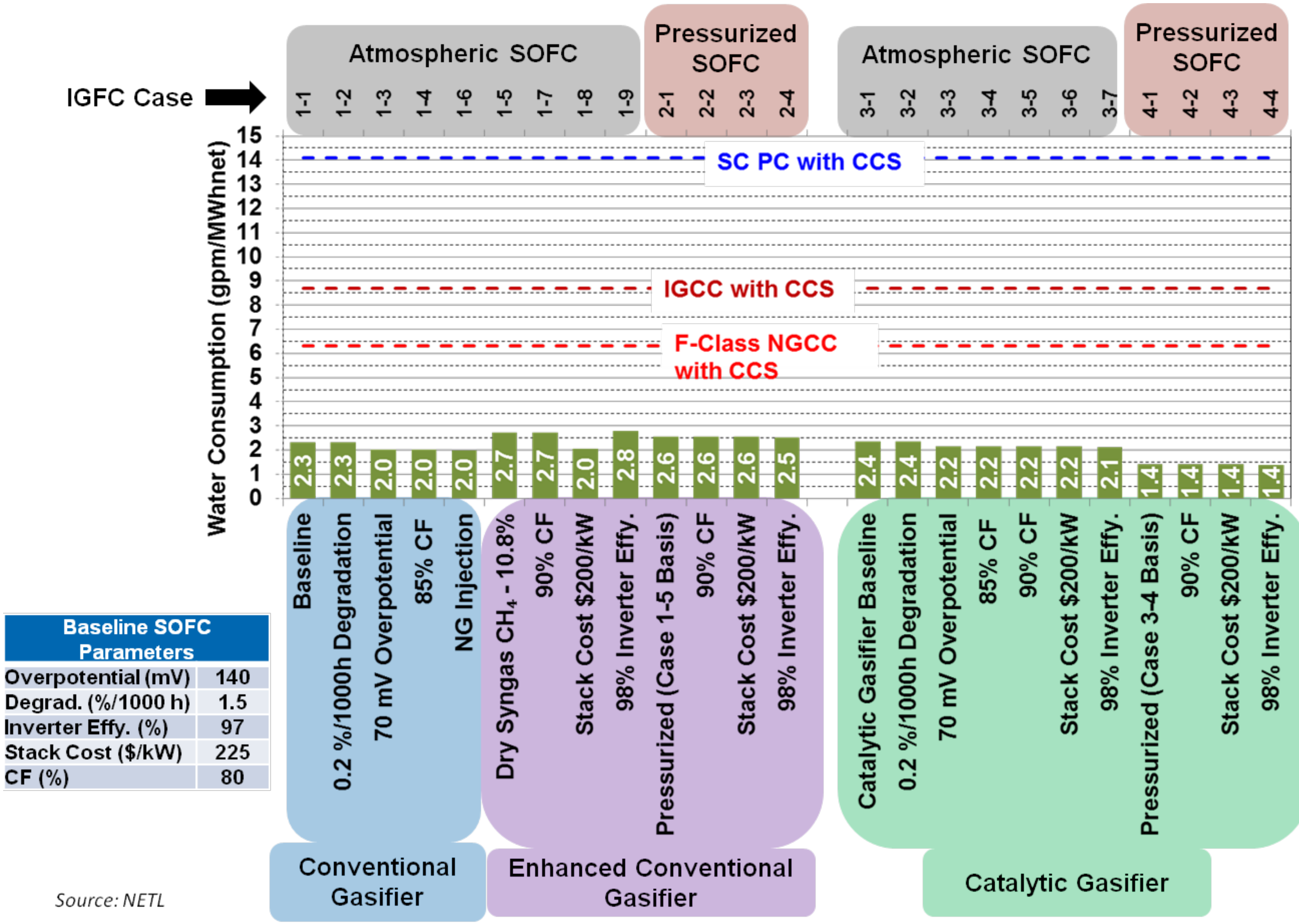
Exhibit ES-13 Comparison IGFC cost of captured CO_2



Baseline SOFC Parameters	
Overpotential (mV)	140
Degrad. (%/1000 h)	1.5
Inverter Effy. (%)	97
Stack Cost ($/kW)	225
CF (%)	80

Source: NETL

Exhibit ES-14 Comparison IGFC raw water consumption



Source: NETL

Currently, the Environmental Protection Agency (EPA) has proposed a GHG emissions limit of 1100 lb/MWh$_{GROSS}$ for coal-fueled plants. Salient IGFC cases were analyzed to evaluate their potential in meeting the GHG limits sans the performance and cost burden associated with carbon capture.

The performance and cost of salient IGFC systems without carbon capture are summarized in Exhibit ES-15. An IGFC system with the CoP gasifier technology along with an atmospheric SOFC, both of which are projected to become commercially competitive in the GHG regulation time-frame, can be designed to meet the proposed 1100 lb/MWh$_{GROSS}$ GHG limit with 14 percent natural gas (NG) injection. Although the CoP gasifier cases with the pressurized SOFC, and the catalytic gasifier cases, with both atmospheric and pressurized SOFC, clearly exceed the proposed GHG limit, they reflect advanced futuristic systems in the NETL transformational time frame, and may be subjected to a more stringent GHG limit. In particular, the emissions of the pressurized catalytic gasifier case are still higher, albeit slightly, than the emissions of present NGCC systems.

Exhibit ES-15 Performance and cost of salient IGFC cases without CCS

Gasifier	CoP			Catalytic	
SOFC Operating Condition Operating Pressure (psia)	Atm. 15.6	Atm. 15.6	Press. 285	Atm. 15.6	Press. 285
Case	1-4	1-6	2-1	3-4	4-1
NG Injection (% of total input thermal energy)	-	14.2	-	-	-
PERFORMANCE					
Net Power (MWe)	551.0	550.1	550.2	550.0	550.1
Net Electric Efficiency, HHV (%)	**47.5**	**54.2**	**49.6**	**60.2**	**64.8**
COST					
Total Plant Cost (TPC) (1000$)	1,280,642	1,090,041	1,281,806	854,820	972,238
Total Overnight Cost (TOC) (1000$)	1,582,392	1,346,221	1,582,203	1,067,048	1,206,471
Total As-Spent Cost (TASC) (1000$)	1,803,927	1,534,692	1,803,712	1,216,434	1,375,377
Cost of Electricity ($/MWh)					
Variable COE	31.6	31.1	34.0	29.0	30.5
Fuel	21.1	20.9	20.2	16.6	15.4
Variable O&M	10.5	10.2	13.8	12.4	15.1
Fixed O&M	11.1	10.2	11.1	10.3	10.9
Capital Charges	47.9	40.9	48.0	32.4	36.6
Total First Year COE	**90.7**	**82.2**	**93.2**	**71.7**	**78.0**
CO_2 Emissions					
lb/MWhgross	**1252**	**1083**	**1060**	**1018**	**880**
lb/MWhnet	1432	1193	1373	1078	1002

In essence, the 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.

- Significant reduction of SOFC performance degradation rate, in addition to enhancement of SOFC electrical performance, is required for the IGFC system to be economically competitive with other technologies.
- The cost of electricity with the IGFC system is projected to be significantly lower than IGCC and PC systems with CCS, while being competitive with NGCC systems with CCS (@ $6.13/MMBtu) and exceed the NETL goals. The IGFC systems using a catalytic coal gasifier and atmospheric-pressure SOFC result in the greatest benefit with a COE that is less than even the NGCC system with CCS (@$6.13/MMBtu). This IGFC system requires the 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.
- A pressurized-SOFC operating at 285 psia provides no cost benefit over systems operating with an atmospheric-pressure SOFC. However, 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. In particular, the cost of the pressure vessel is an important parameter that needs further refinement. This IGFC configuration requires development of the pressurized-SOFC 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 comparable to that of an NGCC system with CCS (@ $6.13/MMBtu). IGFC with natural gas injection can have a 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.
- The COE of all the IGFC plants considered herein are still higher than a natural gas fuel cell (NGFC) system with capture (COE ~$68/MMBtu @ $6.13/MMBtu).
- The natural gas injection case also represents an IGFC configuration that can meet the proposed EPA 1100 lb/MWh$_{GROSS}$ limit on GHG emissions for a coal power plant without any need for the CCS equipment. It is particularly attractive as it utilizes conventional gasification and SOFC technologies, which are likely to be developed within the regulation time-frame, unlike the pressurized SOFC and catalytic gasifier plants. However, the COE of this IGFC plant is still higher than an NGCC system without CCS (@ $6.13/MMBtu).

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.

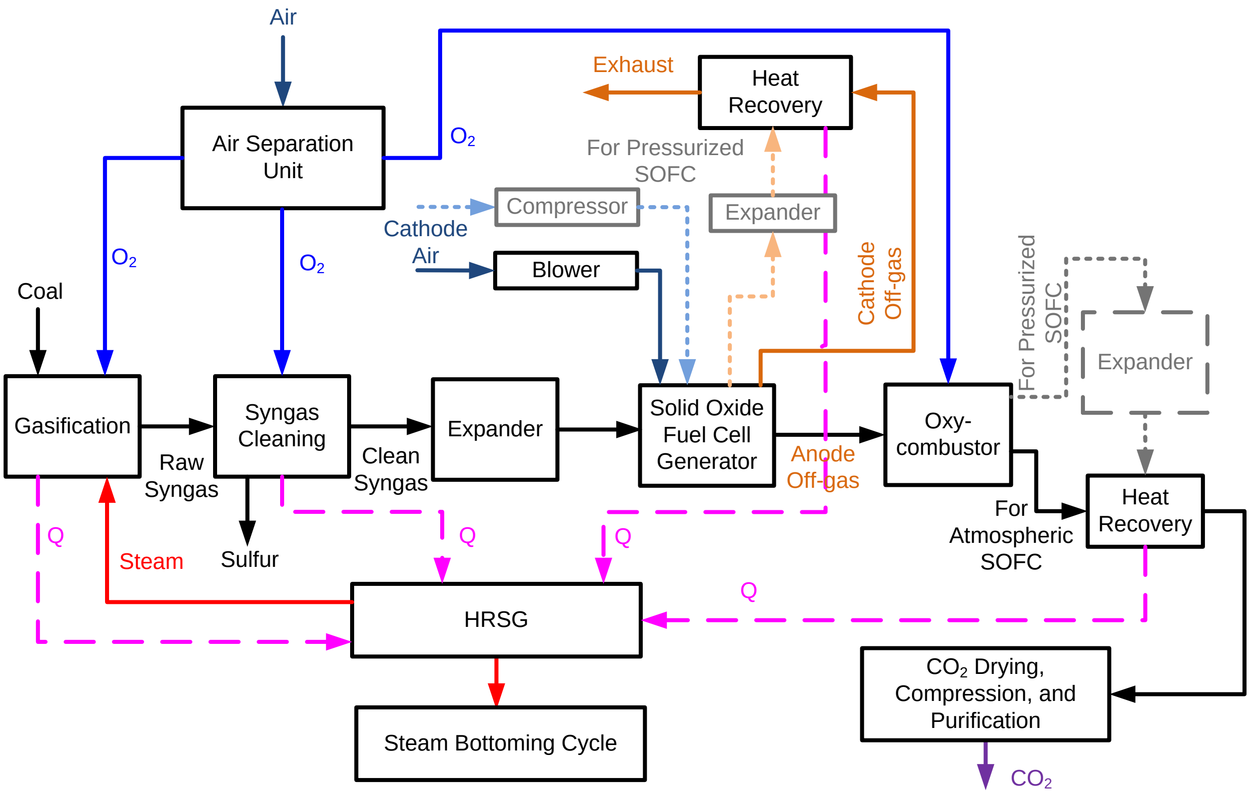
1 Introduction

The results of a pathway study for coal-based, integrated gasification fuel cell (IGFC) power systems with carbon capture and storage (CCS) are presented in this report. They represent the potential future benefits of IGFC technology development by quantifying the performance and cost benefits for a series of projected gains made through the development of advanced technologies for improvements in plant operation and maintenance. In addition, the effectiveness of an IGFC system without CCS in meeting the proposed greenhouse gas (GHG) emissions are discussed using results from select cases. The objective of the study is to provide guidance to the research and development (R&D) program of the Department of Energy (DOE).

The IGFC power plant is analogous to an integrated gasification combined cycle (IGCC) power plant, but with the gas turbine power island replaced with a solid oxide fuel cell (SOFC) power island. The SOFC provides for high electric efficiencies associated with the nearly reversible electrochemical conversion of syngas chemical potential to electric power, as compared to a conventional Carnot-driven heat engine. In addition, the clean oxy-reaction of syngas in the sealed fuel cell system renders itself readily available for CCS with the requirement of only a small oxy-combustor downstream of the fuel cell to react the fuel that is not utilized electrochemically (electrochemical utilization of fuels typically varies between 75-90 percent for current fuel cell technology, due to practical considerations). The heat rejected by the fuel cell system can be recovered further in a combination of Brayton and Rankine cycles, depending on whether the fuel cell system is operating at elevated or atmospheric pressures. The only other exhaust gas stream in the plant is the cathode off-gas which is uncontaminated, vitiated air.

A general schematic block flow diagram of the IGFC system with carbon capture is shown in Exhibit 1-1. The syngas exiting the gasifier goes through a cleaning step before expansion to SOFC operating pressure. Heat recovered from cooling the raw syngas, the hot oxy-combustor exhaust gas, and the hot vitiated air-exhaust gas from the SOFC system along with any process heat generated during syngas cleaning and sulfur extraction is sent to a heat recovery steam generator (HRSG), which, in addition to meeting all the process steam requirements of the system, supplies steam to the bottoming cycle. In systems featuring a pressurized SOFC generator, a major portion of the compression work needed to supply air at pressure to the generator is recovered by expanding the SOFC cathode air exhaust gas back to atmospheric pressure (as shown by the dotted lines in Exhibit 1-1). An air separation unit (ASU) supplies oxygen to the gasifier/reformer, to the sulfur extraction process (IGFC systems), and to the oxy-combustor to enable efficient CO_2 capture.

Exhibit 1-1 IGFC system configuration with CCS



Source: NETL/DOE

The present study assumes the use of advanced, planar, SOFC technology featuring separated anode and cathode off-gas steams with anode off-gas oxy-combustion for cases with carbon capture. The SOFC is assumed to operate initially at atmospheric pressure, and the simulations utilize the expected operating conditions and performance capabilities of the technology. The power plant cost and performance are estimated based not only on the current state of SOFC development, but also a projected pathway of SOFC technology development advances. In particular, the following SOFC system advances are incorporated in a cumulative manner:

- Reduction of SOFC stack performance degradation
- Reduction of stack overpotential (at normal operating conditions)
- Reduction of SOFC stack cost
- Improvement of inverter efficiency
- Operation of SOFC under pressurized conditions

General advances in IGFC plant operation are also included in the pathway in the form of improved plant availability and increased capacity factor, which are assumed to be achieved through advanced component monitoring, improved maintenance practices, and plant operation experience.

The overall plant performance and costs estimates of two parallel pathways of IGFC development are considered here. The first pathway utilizes conventional coal gasification technology, and features the gasifier (CoP) with syngas methane (CH_4) content of roughly 6 mole percent. Two variants of this pathway system are considered based on the SOFC operating pressure. Scenario 1 investigates SOFC systems operating at atmospheric pressure and progressively includes pertinent SOFC technology advances. Both the near-term enhancement in the conventional gasifier technology and the injection of natural gas into the syngas stream were also considered in this scenario to boost the CH_4 content of the syngas, which has been projected to benefit the performance of the IGFC plant. The incorporation of pressurized-SOFC technology as a longer term enhancement is considered in Scenario 2.

The second pathway utilizes an advanced, catalytic coal gasification technology projected to produce syngas having a high CH_4 content of roughly 30 mole percent, which considerably improves the IGFC performance. Scenarios 3 and 4 represent, respectively, the atmospheric and pressurized SOFC variants under this pathway, which follow similar advances in SOFC technology development as in the first 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.

The basis for the technical and cost evaluations is provided in Section 2. The results for IGFC pathway cases (Scenarios 1 and 2) using conventional coal gasification technology are presented in Section 3, while the corresponding results for the catalytic coal gasifier-based IGFC plant simulations (Scenarios 3 and 4) are discussed in Section 4 along with a description of the catalytic gasification technology.

Section 5 summarizes the results for salient IGFC cases without carbon capture to evaluate their potential in meeting the GHG limits currently being proposed by the United States (U.S.) Environmental Protection Agency (EPA).

Exhibit 1-2 Conventional gasifier IGFC pathway parameters (Scenarios 1 and 2)

<table>
<tr><th>Case</th><th>1-1</th><th>1-2</th><th>1-3</th><th>1-4</th><th>1-5</th><th>1-6</th><th>1-7</th><th>1-8</th><th>1-9</th><th>2-1</th><th>2-2</th><th>2-3</th><th>2-4</th></tr>
<tr><th>Pathway Parameter</th><td>Baseline</td><td>Degradation</td><td>Overpotential</td><td>Capacity Factor</td><td>Gasifier</td><td>NG Injection (34%)[2]</td><td>Capacity Factor</td><td>SOFC Cost Reduction</td><td>Inverter Efficiency</td><td>Pressurized SOFC</td><td>Capacity Factor</td><td>SOFC Cost Reduction</td><td>Inverter Efficiency</td></tr>
<tr><th>SOFC Pressure (psia)</th><td colspan="9">Atmospheric (15.6 psia)</td><td colspan="4">Pressurized (285 psia)</td></tr>
<tr><th>SOFC Degradation Rate (%/1000h)</th><td>1.5</td><td colspan="12">0.2</td></tr>
<tr><th>SOFC Overpotential (mV)</th><td colspan="2">140</td><td colspan="11">70</td></tr>
<tr><th>Capacity Factor (%)</th><td colspan="3">80</td><td colspan="3">85</td><td colspan="3">90</td><td>85</td><td colspan="3">90</td></tr>
<tr><th>Dry Syngas CH_4 Content (%)</th><td colspan="4">5.8</td><td>10.8</td><td>24.6</td><td colspan="7">10.8</td></tr>
<tr><th>SOFC Stack Cost ($/kW)[1]</th><td colspan="7">225</td><td colspan="2">200</td><td colspan="2">225</td><td colspan="2">200</td></tr>
<tr><th>Inverter Efficiency (%)</th><td colspan="8">97</td><td>98</td><td colspan="3">97</td><td>98</td></tr>
</table>

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

[2]*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)

<table>
<tr><th>Case</th><th>3-1</th><th>3-2</th><th>3-3</th><th>3-4</th><th>3-5</th><th>3-6</th><th>3-7</th><th>4-1</th><th>4-2</th><th>4-3</th><th>4-4</th></tr>
<tr><th>Pathway Parameter</th><td>Baseline</td><td>Degradation</td><td>Overpotential</td><td>Capacity Factor</td><td>Capacity Factor</td><td>SOFC Cost Reduction</td><td>Inverter Efficiency</td><td>Pressurized SOFC</td><td>Capacity Factor</td><td>SOFC Cost Reduction</td><td>Inverter Efficiency</td></tr>
<tr><th>SOFC Pressure (psia)</th><td colspan="7">Atmospheric (15.6 psia)</td><td colspan="4">Pressurized (285 psia)</td></tr>
<tr><th>Dry Syngas CH_4 Content (%)</th><td colspan="11">31.6</td></tr>
<tr><th>SOFC Degradation Rate (%/1000h)</th><td>1.5</td><td colspan="10">0.2</td></tr>
<tr><th>SOFC Overpotential (mV)</th><td colspan="2">140</td><td colspan="9">70</td></tr>
<tr><th>Capacity Factor (%)</th><td colspan="3">80</td><td>85</td><td colspan="3">90</td><td>85</td><td colspan="3">90</td></tr>
<tr><th>SOFC Stack Cost ($/kW)[1]</th><td colspan="5">225</td><td colspan="2">200</td><td colspan="2">225</td><td colspan="2">200</td></tr>
<tr><th>Inverter Efficiency (%)</th><td colspan="6">97</td><td>98</td><td colspan="3">97</td><td>98</td></tr>
</table>

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

2 Pathway Study Basis

Systems models were developed under the Aspen Plus® (Aspen) platform to simulate the IGFC process configurations. 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 (BB) report. (1)

Capital and operating costs for most of the conventional equipment items were scaled based on the updated BB cost estimates. (2) All the costs are reported in 2011 dollars, and the first-year cost of electricity (COE) is presented as the revenue requirement figure-of-merit for each of the cases.

The design basis for the pathway study, which is largely based on the BB study, (1) is reported in this section along with the environmental targets and cost assumptions.

2.1 Site Characteristics

All plants in this study are assumed to be located at a generic plant site in Midwestern U.S., with ambient conditions and site characteristics as presented in Exhibit 2-1 and Exhibit 2-2. The ambient conditions are the same as International Standards Organization (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 U.S.
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 IGFC cases assume that 15 acres of land area are required for the main plant along with an extra area amount that provides a buffer of approximately 0.25 miles to the fence line, which could also provide for a rail loop, if required. The steam turbine, unlike the gasifier and the SOFC stack units, is assumed to be enclosed in a building.

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 as per National Energy Technology Laboratory (NETL) Quality Guidelines for Energy System Studies (QGESS). (3) The first-year cost of coal used in this study is $2.76/MMkJ ($2.91/MMBtu) in 2011$ in accordance with the values in the updated BB report. (2) The coal mercury content for this study was assumed to be 0.15 ppm (dry), which is consistent with the mercury content estimated and applied in the BB report. It was also assumed that all of the coal Hg enters the gas phase, and none leaves with the bottom ash or slag.

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 (Case 1-6), 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
Unit	**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 $5.81/MMkJ ($6.13/MMBtu) as per the updated BB report. (2)

2.4 Environmental Targets

The IGFC plant emissions are projected to be highly 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 combustor; and
4. Most of the coal syngas contaminant species remaining after syngas cleaning are sequestered with the CO_2 product.

Only the plant solid waste streams, coal ash, spent sorbents, and waste catalysts might be of environmental concern. Accordingly, the emissions from the IGFC plants are expected to be much lower than the typical environmental targets identified in the BB report (1) for the conventional fossil-fuel based power plants.

The environmental targets for the IGFC plants, shown in Exhibit 2-5, were chosen to match the Electric Power Research Institute's (EPRI) design basis for their CoalFleet for Tomorrow Initiative (4) as in the BB IGCC studies. (1) EPRI notes that these are only 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 x 10^{-6} lb/MWh	Carbon bed

An acid gas removal (AGR) process with sulfur capture efficiency of ~99.7 percent is required to meet the environmental target for sulfur dioxide (SO_2) emissions of 0.0128 lb/MMBtu, which corresponds to ~28 ppmv sulfur in the sweet syngas. However, an additional sulfur polishing step is required for the IGFC systems to reduce the sulfur concentration in the syngas to less than 100 ppbv to prevent poisoning of the SOFCs.

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 along with either ceramic or metallic candle filters, to meet the environmental target of 0.0071 lb/MMBtu for filterable particulates.

Based on experience at the Eastman Chemical plant, where syngas from a General Electric Energy (GEE) gasifier is treated, the actual mercury removal efficiency used is 95 percent, which exceeds the environmental target for mercury capture of at least 90 percent removal. 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 the New Source Performance Standards (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) capture along with purification to enhanced oil recovery (EOR) levels is investigated in the present study. In addition, CO_2 emissions of salient IGFC systems without carbon capture are evaluated against the proposed EPA GHG limit of 1100 lb/MWh$_{GROSS}$[2] for coal-fueled power plants.

[2] Proposed value pending review at the time of writing the manuscript; final codified regulation may be different.

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 123,740 to 184,567 kg/h (272,800 to 406,900 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 93,395 kg/h (205,900 lb/h, as-received), injecting natural gas at 35 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. In practice, degradation of SOFC performance is mitigated by providing additional capacity in the form of extra SOFC surface area, coupled with operational strategies to maintain a constant power output. Stack operational scenarios, which effectively compensate for stack degradation, were modeled to optimize the extra area installed and evaluate the corresponding stack replacement period. Description and results of the model are discussed in detail later in this section.

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

2.8.1 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 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 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 (H_2) to water vapor (H_2O), the cathode gas oxygen mole fraction, and the SOFC operating temperature and the pressure. (6) This procedure provides operating voltages that are comparable to SOFC vendor test results with comparable conditions and fuel gas composition. The SOFC is assumed to operate at a nominal current density of 400 mA/cm^2.

2.8.2 SOFC Carbon Deposition Control

The cell stack inlet anode gas composition can induce the formation of solid carbon deposits, which can degrade and disrupt the normal performance of the SOFC unit. Anode gas recirculation is used to control the anode gas inlet conditions to maintain an atomic oxygen-to-carbon ratio greater than 2.0, which is a generally used criterion to prevent carbon deposition anywhere in the SOFC fuel flow domain. Anode gas recirculation is accomplished using hot gas blowers or syngas jet pumps.

2.8.3 Estimation of Steam Bottoming Cycle Performance

The anode off-gas stream is combusted with oxygen for the cases with carbon capture, while a portion of the SOFC cathode exhaust is utilized for combustion in cases without carbon capture. In both cases, the hot stream from the combustor exchanges heat in an HRSG system to produce steam, which generates power in a subcritical steam bottoming cycle after satisfying process steam requirements. The steam cycle provides a relatively small proportion of the total plant generation output with steam conditions and capacity varying greatly across study cases. In some cases, the heat recovery temperature available is relatively low, and results in poor steam superheat conditions. A simple subcritical steam cycle was included in the conventional coal gasifier case simulations. The steam cycle performance for the catalytic gasifier cases, on the other hand, was simply estimated using a nominal steam cycle efficiency of 38.1 percent, based on the conventional coal gasifier cases. This approach has been shown to result for the conventional IGFC cases in efficiencies that could be ±1.5 percentage points lower or higher than fully simulated steam cycle model values depending on whether the SOFC is operating at atmospheric or elevated pressures, respectively. These uncertainties have to be kept in mind while interpreting the catalytic gasifier case results.

2.8.4 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 BB report. (1) The plant processing sections are designed for 100 percent capacity, with no excess capacity provided for any component other than the SOFC stack. 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.

The North American Electric Reliability Council (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. (7) To get the availability factor, one has to deduct the scheduled outage time. In reality, the scheduled outage time varies with gasifier technology. However, the variations are relatively small and would have minimal impact on the capacity factor, and a total of 30-day planned outage per year (or two 15-day outages) was assumed in this study. 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. The sensitivity of the costs to capacity factor variations was explored to discern the effect.

There are four operating IGCC's worldwide that use a solid feedstock and are primarily power producers (Polk, Wabash, Buggenum and Puertollano). Higman et al. (8) examined the reliability of these IGCC power generation units and concluded that typical annual on-stream times are around 80 percent. 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.

2.9 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. 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 sour water stripper (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.

A cooling water temperature of 16°C (60°F) was assumed based on Exhibit 2-1 along with 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 (9) assumptions:

- 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.

2.10 Cost Estimating Methodology

Following the basis used in the BB report, (1) 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 BB 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 typically 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) for the BB report. (1) The 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 2011 dollars. A first year of operation of 2015 is assumed for all cases.

Capital costs at the 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).

The total as-spent cost (TASC) is the calculated by adding the estimated financing costs to the TOC, which is a sum of the owner's costs and the TPC. The current-dollar, first-year cost of electricity is calculated using TOC.

2.10.1 Plant Maturity

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

The SOFC and oxy-combustion technologies for the IGFC cases are also immature and are 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 oxy-combustion technology merits additional development due to its large advantage over pre-combustion CO_2 removal. 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 synthetic natural gas (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.

2.10.2 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, or local regulatory requirements 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 transport and storage (T&S) of the sequestered CO_2.

2.10.3 Capital Costs

WorleyParsons developed the capital cost estimates for IGCC plants in the BB report (1) using the company's in-house database and conceptual estimating models for each of the specific technologies. The estimating models are based on a reference bottoms-up estimate for each major component. 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 (5-10 h days). No additional incentives such as per-diem 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 T&S of the sequestered CO_2 in all capture cases.
- Engineering and Construction Management were estimated as a percent of bare erected cost; 10 percent for IGCC and pulverized coal (PC) technologies, and 9 percent for natural gas combined cycle (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.

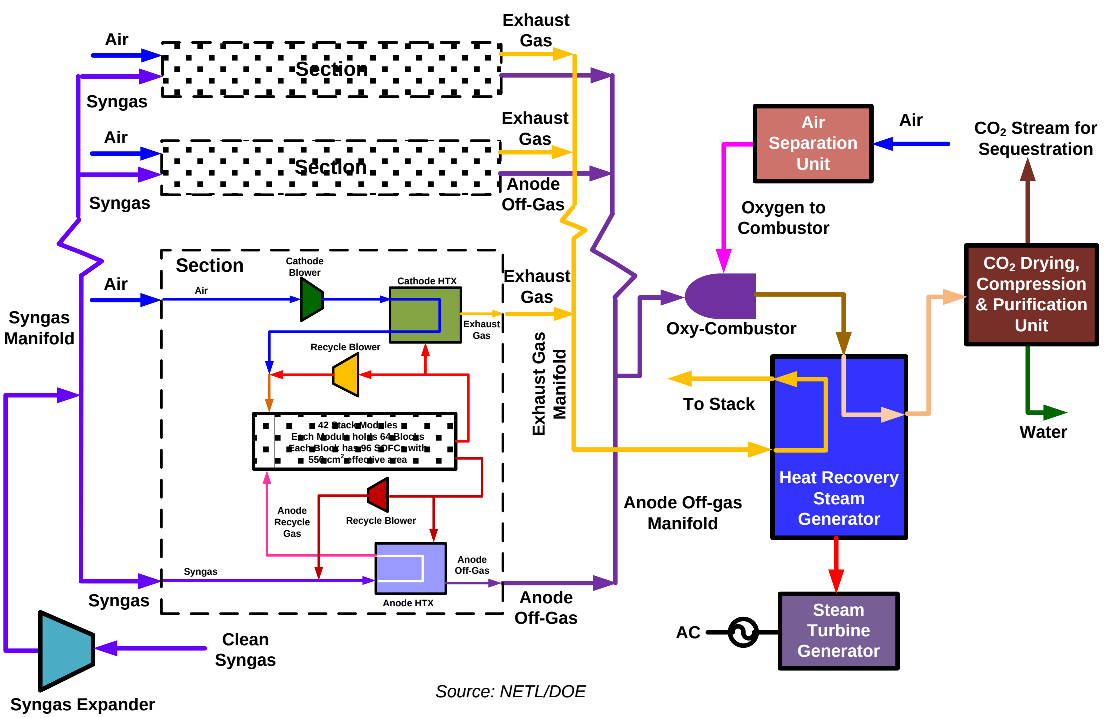
- All capital costs are presented as “Overnight Costs” in June 2011 dollars. Escalation to period-of-performance is specifically excluded.

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 0.124. The capital component of COE was calculated using TOC.

2.10.4 SOFC Module Costs

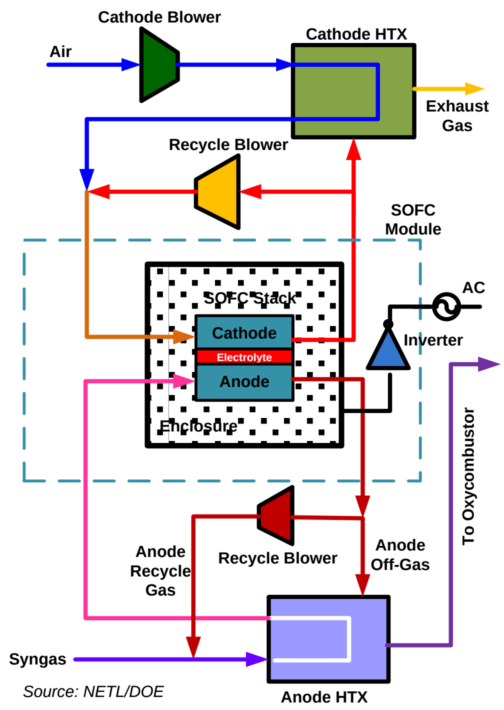
The atmospheric SOFC power island configuration, shown in Exhibit 2-7, is based on a generic planar technology power island. It consists of several parallel trains of modularized SOFC sections each consisting of 42 planar SOFC modules. A block is defined to be a stack of 96 SOFCs with 550 cm^2 effective area and 64 blocks comprise a single SOFC module. The module envelope, as shown in Exhibit 2-8, is defined to include, in addition to the SOFC stacks, the enclosure, and the direct current (DC) – alternating current (AC) inverter. The incoming syngas fuel is distributed into each section, which also houses an individual SOFC balance of plant (BOP) including an air blower, recycle blowers, and heat exchangers. A single ASU is assumed to drive an oxy-combustor, which is fed with to the anode off-gas collected from all the sections. An air compressor replaces the air blower in the pressurized SOFC cases as shown in Exhibit 2-9, which also features a cathode expander to extract power from the pressurized cathode exhaust before exiting to the stack.

Exhibit 2-7 Atmospheric SOFC power island configuration showing section components



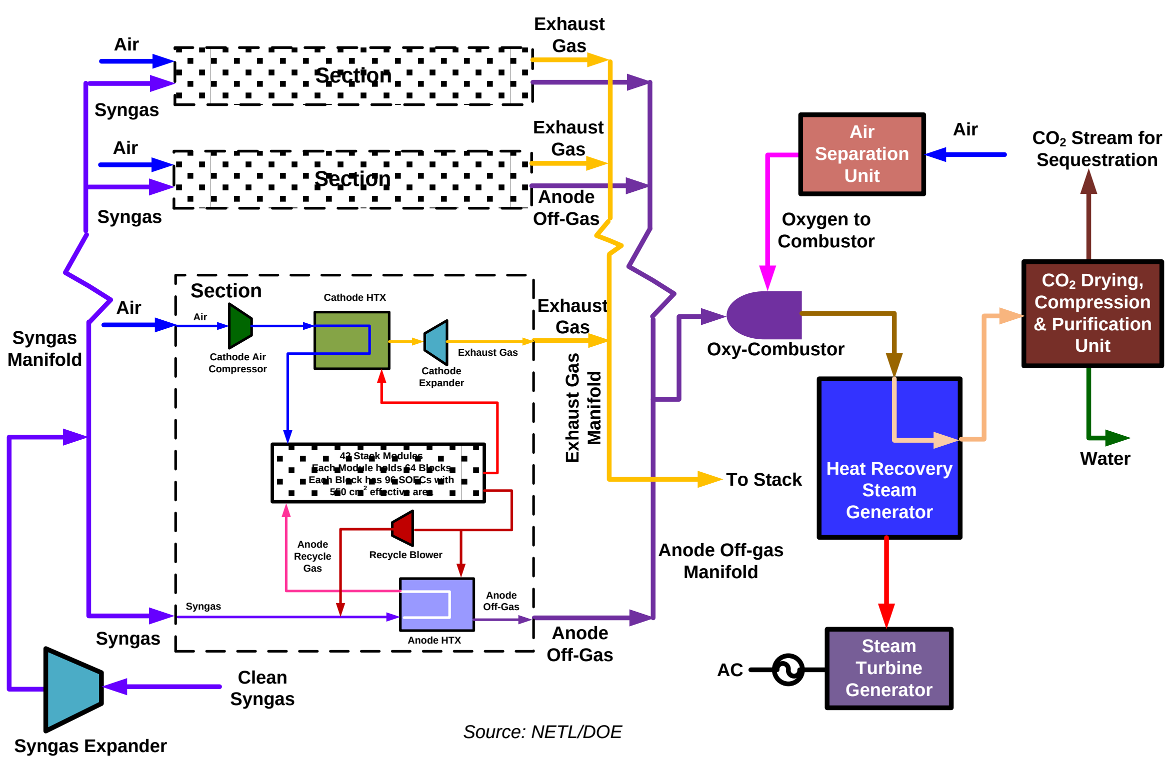
Source: NETL/DOE

Exhibit 2-8 SOFC module



Source: NETL/DOE

Exhibit 2-9 Pressurized SOFC power island configuration showing section components



Source: NETL/DOE

The NETL goal of $225/kWe in 2011 dollars for the stack cost forms the basis for the SOFC stack cost calculations. The atmospheric SOFC module costs, summarized in Exhibit 2-10, include the module transport and placement costs and the site foundation costs, which represent the costs associated with the installation of each module. These costs were generally estimated by escalating the earlier IGFC study (10) costs in 2007$ to 2011$ based on the BB cost update, (1) and the Chemical Engineering Plant Cost Index (CEPCI). (11) The corresponding costs for the pressurized SOFC module are shown in Exhibit 2-11 and differ only by the enclosure cost, which reflects the significantly higher cost for the pressure vessels required for pressurized SOFC operation.

Exhibit 2-10 Atmospheric SOFC module costs

Cost Component	Specific Cost ($/kWe AC)
	2011$
SOFC Module	
SOFC Stack	225
Enclosure	30
Transport and Placement	14
Site Foundations	44
Inverter	68
Total SOFC Module	**382**

Exhibit 2-11 Pressurized SOFC module costs

Cost Component	Specific Cost ($/kWe AC)
	2011$
SOFC Module	
SOFC Stack	225
Enclosure	240
Transport and Placement	14
Site Foundations	44
Inverter	68
Total SOFC Module	**592**
Total SOFC Module with 10% Extra Installed Area	**651**

2.11 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 engineer/procure/construct (EPC) contracting approach

2.12 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 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 Association for the Advancement of Cost Engineering (AACE) international contingency guidelines, as presented in NETL's QGESS, it would appear that the overall project contingencies for the subject cases should be in the range of 30 to 40 percent. (12) 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

2.13 Operation and Maintenance

The production costs or operating costs and related maintenance expenses 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 is consistent with that applied in the Bituminous Baseline report.

2.13.1 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 $39.70/h. The associated labor burden is estimated at 30 percent of the base labor rate.

2.13.2 Maintenance Material and Labor

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

2.13.3 Stack Degradation and Production Costs

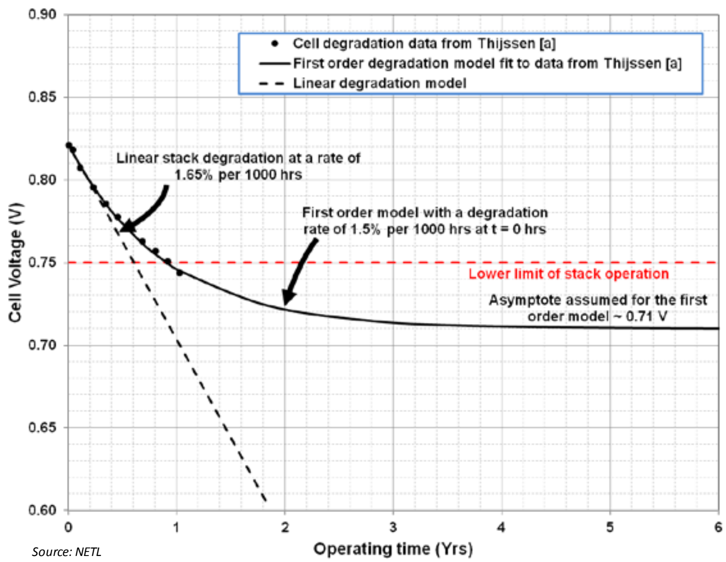
Solid oxide fuel cells have the potential to operate over a long period of time, which has been demonstrated in laboratory scale tests, where operation for over five years has been demonstrated without appreciable loss of performance. (13) However, with current planar stack technologies, stack performance has been observed to decline over its lifetime, generally due to an increase in the apparent electrical resistance of the stack associated with a variety of material and design related factors, which limits the permissible current at the same voltage (in a constant voltage operation mode). Performance degradation limits the operating lifetime of the capital intensive SOFC stack and forms a significant component of the production costs. Apart from long-term material developments, practical considerations to mitigate the adverse effects of stack degradation are investigated here to enable an estimation of the production costs that is consistent with industry practice.

2.13.3.1 Stack Degradation

Stack performance degradation rates are generally between 1-2 percent per 1000 hours with current stack technology. (14) Reducing stack degradation to values below the 0.2 percent per 1000 hours generally observed with conventional heat engine-based power generation systems (15) forms the focus of current SOFC research and development. Published experimental data on the long-term (over 20,000 h) cell performance degradation data is generally limited to tubular cell designs. Exhibit 2-12 shows intermediate duration (~1 yr, 8000 h) performance degradation data for planar cell taken from the SOFC system study of Thijssen. (15) While the cell performance degradation rate, usually expressed as a loss in cell voltage (mV) per 1000 h, suggests a constant degradation rate with a linear decay of cell voltage with time, a first-order degradation model appears to fit the experimental data better (Exhibit 2-12). (15) The linear degradation model is determined by setting the initial voltage and a degradation rate. On the other hand, the first order model also needs the prescription of an asymptotic value. An initial degradation rate of 1.65 percent per 1000 hours along with an asymptotic value of ~0.71 V was seen to result in a good fit to the data.

With a nickel-cermet anode, the nickel oxidation potential, which is ~0.70 V for the range of temperatures of interest, sets a lower limit for cell voltage. In practice, the stack is generally operated at a voltage with sufficient margin over the nickel oxidation potential to minimize potential operational risks[3] associated with temperature and cell resistance variations within a stack. The linear degradation model results in a stack lifetime of ~7 months (0.6 yr), whereas the first order model predicts a 50 percent higher stack lifetime of ~11 months (0.9 yr), assuming that the lower limit of stack voltage is 0.75 V (50 mV greater than the nickel-oxidation potential) from an operational perspective. It is clear that the reality lies somewhere in between these two extreme limits.

Exhibit 2-12 Cell degradation data and model fits (15)



Source: NETL

2.13.3.2 Degradation and Operation

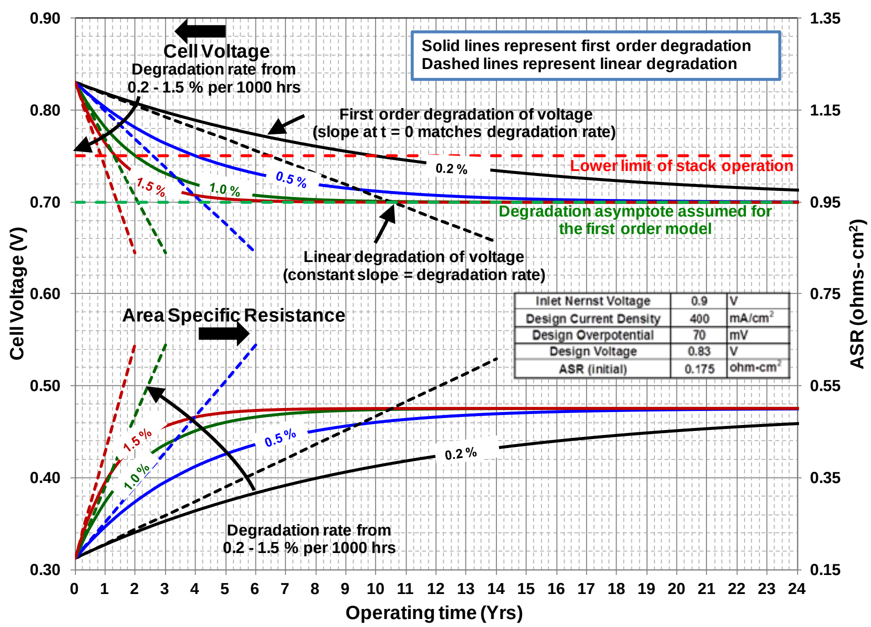
If the stack is operated at constant current as in the experimental data depicted in Exhibit 2-12, the stack power decreases quickly with time, which is commercially unattractive due to its detrimental effects on warranty costs. One way of prolonging the stack lifetime and still maintain the rated power is to provide additional capacity in the form extra stack surface. In theory, the system can be designed to periodically enable additional SOFC area online, to maintain a near-constant plant power output from the SOFC system. However, this approach is often economically prohibitive, and in practice, constant power output is maintained by operating

[3]The nickel oxidation potential represents the voltage at which the nickel in the anode is getting oxidized, which often results in delamination of the anode leading to cell failure that could cascade into a catastrophic failure of the stack.

the stack at a voltage above the design voltage (and the current below the design value) initially, and subsequently increasing the current as the stack voltage declines. While this approach is effective in maintaining a constant power output from the SOFC stack, the system efficiency varies from a value that is higher than the nominal value to a value below it over the stack lifetime. The corresponding stack operational scenarios, which effectively compensates for stack degradation, were modeled to optimize the extra area installed and evaluate the corresponding stack replacement period.

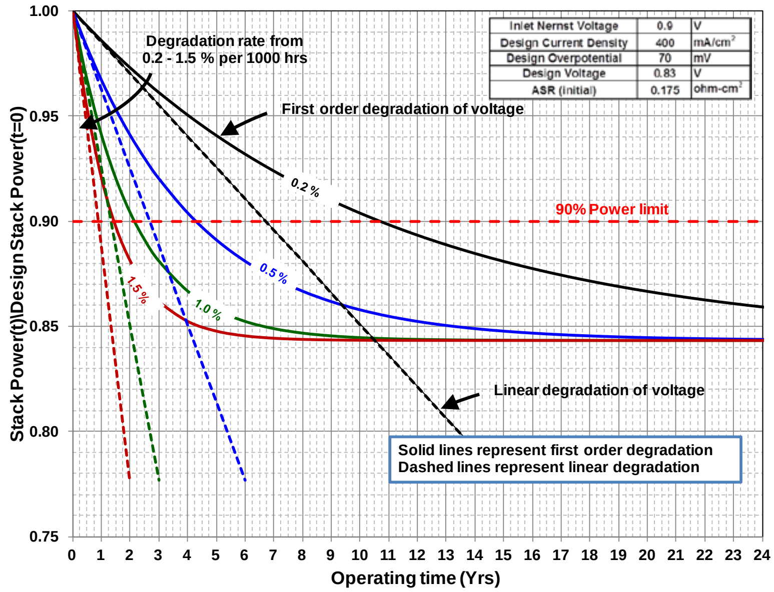
Both linear and first-order degradation models were investigated in the present study. Performance degradation rates were varied between 0.1 to 2 percent per 1000 hours along with a consideration of extra installed areas up to 500 percent of the nominal design area. The assumed stack voltage variations and the corresponding variations in the area specific resistance (ASR) of the stack (at a current density of J = 400 mA/cm^2) are shown in Exhibit 2-13 for different degradation rates. The degradation rate is the slope of the voltage decay for the linear model while it represents the gradient of the voltage curve at t = 0 h for the first order model. The initial cell voltage (at t = 0) and the Nernst voltage match the system model calculations. An asymptote of 0.7 V is used for the first order model. The corresponding power output is shown in Exhibit 2-14. With no additional area installed, stack power decreases rapidly with time at the 1.5 percent per 1000 hours degradation rate of the current technologies and a stack replacement is warranted almost within a year of operation. It also shows that a degradation rate of 0.2 percent per 1000 hours or below is required for an attractive commercial proposition.

Exhibit 2-13 Cell voltage and ASR variations for the degradation models



Source: NETL

Exhibit 2-14 Power variation with time for the two degradation models



Source: NETL

The cell voltage, V_t, at time, t, is related to its instantaneous current density, J_t, and the ASR, Ω_t, by the equation,

$$V_t = E - J_t \Omega_t \quad (1)$$

where E is the inlet Nernst potential. Multiplying Equation 1 by the instantaneous current, $I_t = J_t A$, we get,

$$P = V_t I_t = E I_t - \frac{I_t^2 \Omega_t}{A} \quad (2)$$

where P is the power generated and A is the total installed area.

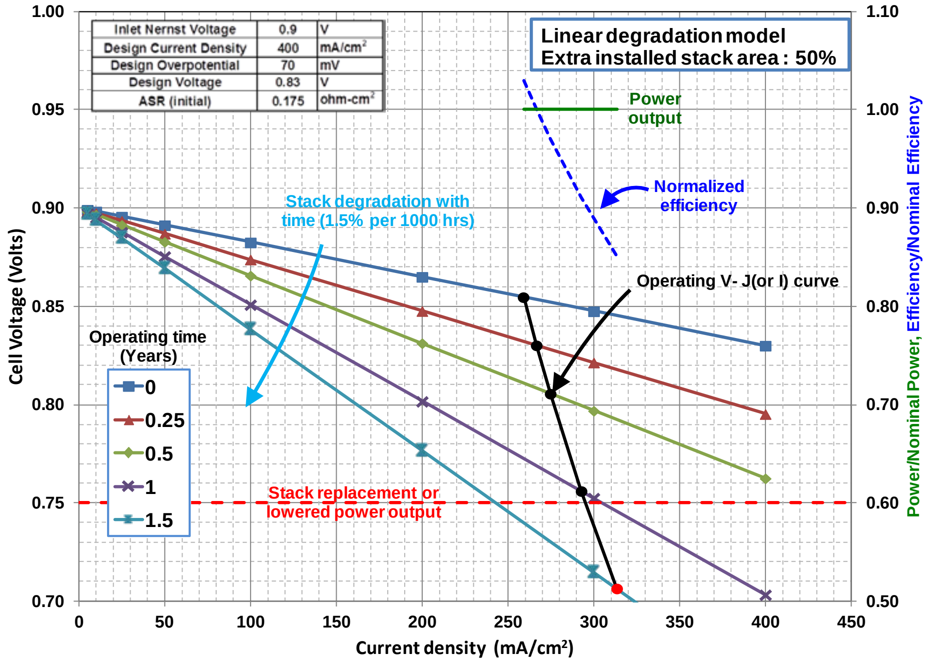
Assuming that the ASR is independent of current density, variation of $\Omega_t = f(t)$ can be evaluated using (1), which enables the computation of the instantaneous current and current density to maintain constant power, P, using Equation 2. The corresponding instantaneous voltage can be obtained by simply dividing the power value by the current. This process can be carried out for different values of the installed area.

Exhibit 2-15 shows the results for a degradation rate of 1.5 percent per thousand hours for 50 percent extra installed area (total area 1.5 times nominal area) using a linear voltage degradation model. The voltage versus current density (computed using the total area including the extra installed area) decay with time (stack performance degradation) is plotted along with the operating points, the V-J pairs, at which the power output is the same as the design power output (also shown in the figure). The efficiency normalized to its design efficiency, however, decreases with time due to the increasing current (or fuel flow rate) at the V-J pairs for the same power output. Assuming a stack voltage limit of 0.75 V, the addition of the 50 percent extra area along with the modified operating curve extends the life of the stack from ~9 months to ~12 months (~35 percent increase in stack life) while maintaining the power output at its design levels. The corresponding result assuming a first order degradation model with the same 1.5 percent per 1000 hour degradation rate initially, shown in Exhibit 2-16, indicates a longer stack life of 16 months for the same extra installed area. The stack life increases, but not in direct proportions as the amount of extra installed area is increased, as shown in Exhibit 2-17, which indicates that installation of 100 percent additional area (over the nominal design) extends the stack life to ~18 months. At an initial degradation rate of 0.2 percent per 1000 hours, the stack life is ~10 yr, as shown in Exhibit 2-18, assuming a first order degradation process.

Installation of additional area instantly reduces the current density at which the stack operates. Even with the operational scenarios described, which require operation at progressively increasing current densities, the operating current density is still below the design current density due to the additional area. The tendency of operating at the reduced current density to mitigate stack degradation rates presents a secondary benefit that is not taken into account in the present calculations.

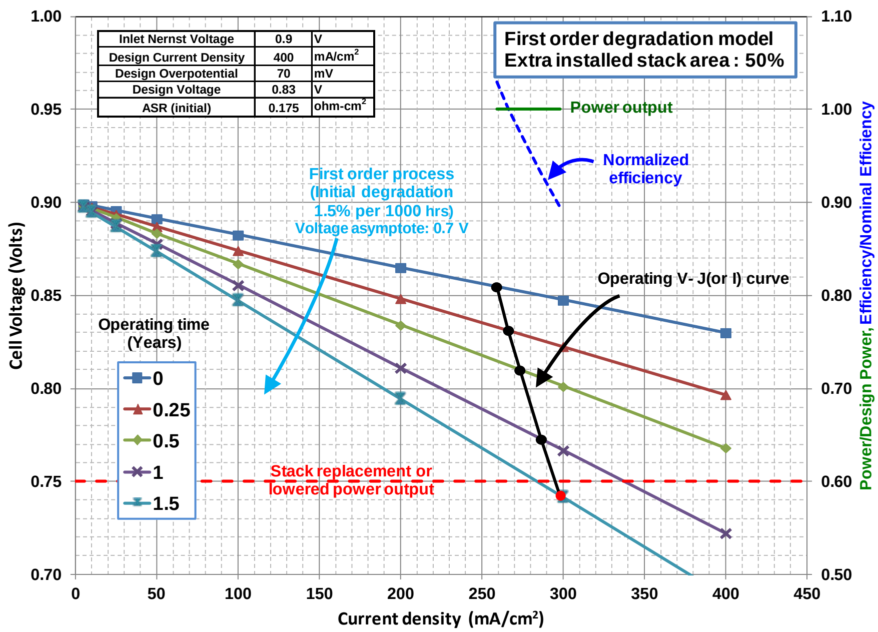
The variation of stack replacement period (or stack life), defined as the time before the stack voltages reaches the operating limit of 0.75 V, with the degradation rate and the amount of extra area installed is shown in Exhibit 2-19 and Exhibit 2-20 for the linear and first order degradation models, respectively. An average of the results of the two degradation models is shown in Exhibit 2-21. The first order degradation assumption results in stack life values that are generally 25 percent higher than the values computed assuming a linear degradation. In most of the cases, adding extra area beyond 100 percent yields diminishing returns. The average yearly cost of stack over stack life is shown for different amounts if extra installed area for a 0.2 percent per 1000 hours degradation rate is shown in Exhibit 2-22. A 10 percent extra installed area appears to be an optimum value for both the models at this degradation rate as it results in the lowest normalized yearly stack cost. At this point the stack life is predicted to be ~6.4 years and ~8.1 years for the linear and the first order models respectively. Accordingly, a stack with additional 10 percent area with an average stack replacement period of 7.3 years was assumed for the Nth of a kind natural gas fuel cell (NGFC) unit in the cost of electricity calculations.

Exhibit 2-15 Constant power operation scenario at 1.5% per 1000 h linear degradation rate with 50% extra installed area



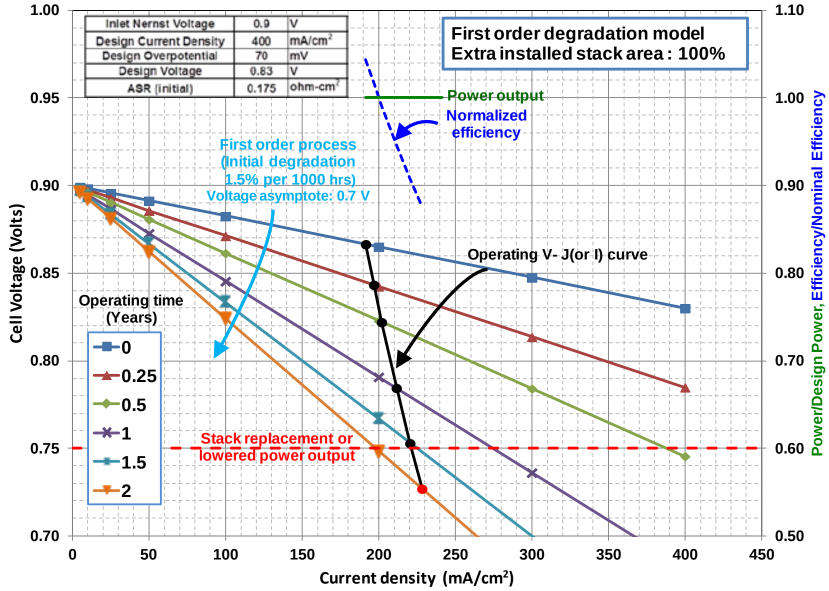
Source: NETL

Exhibit 2-16 Constant power operation scenario with first order degradation for an initial rate of 1.5% per 1000 h with 50% extra installed area



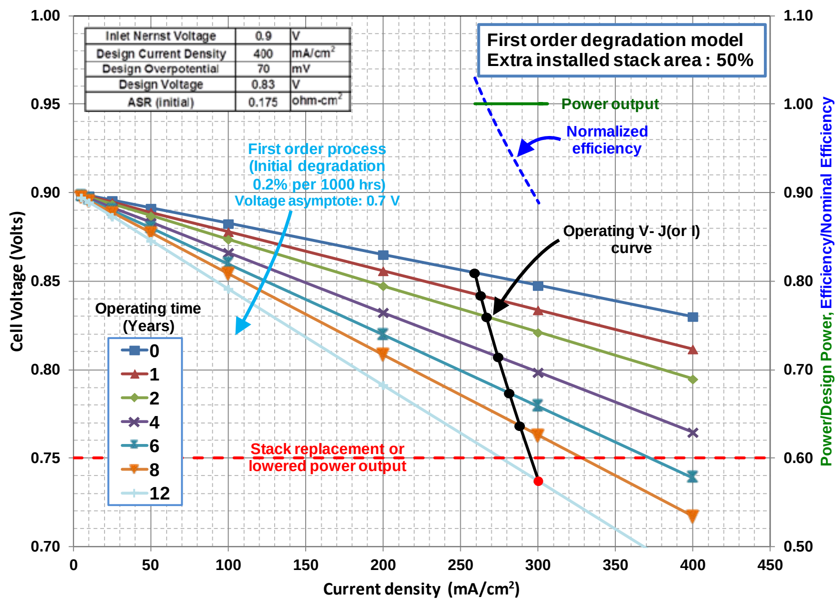
Source: NETL

Exhibit 2-17 Constant power operation scenario with first order degradation for an initial rate of 1.5% per 1000 h with 100% extra installed area



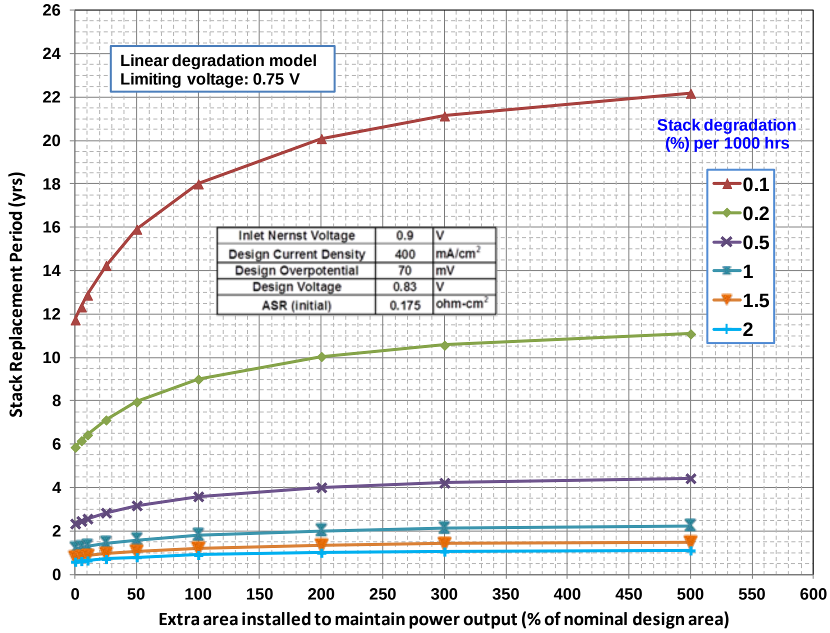
Source: NETL

Exhibit 2-18 Constant power operation scenario with first order degradation for an initial rate of 0.2% per 1000 h with 50% extra installed area



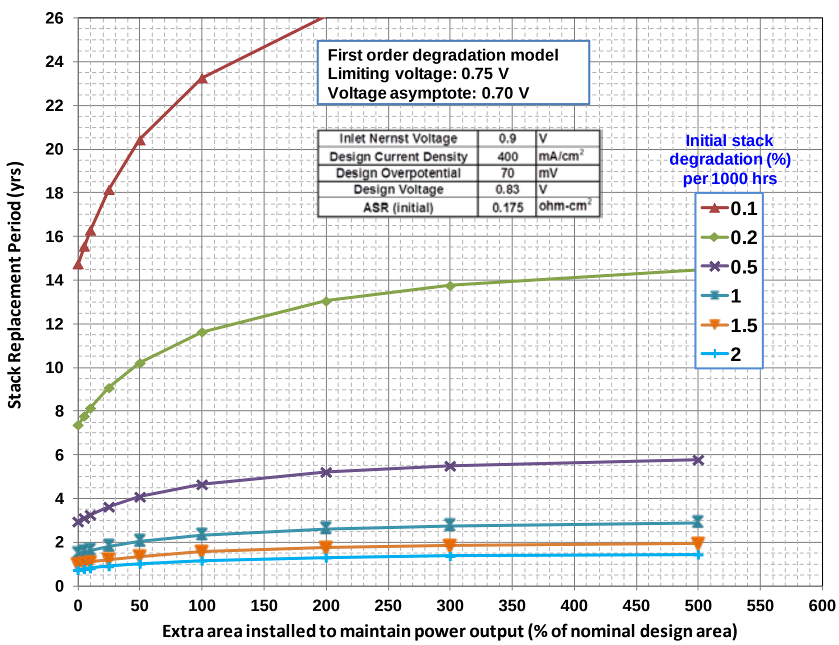
Source: NETL

Exhibit 2-19 Variation of stack replacement period with degradation rate and extra installed area assuming linear degradation



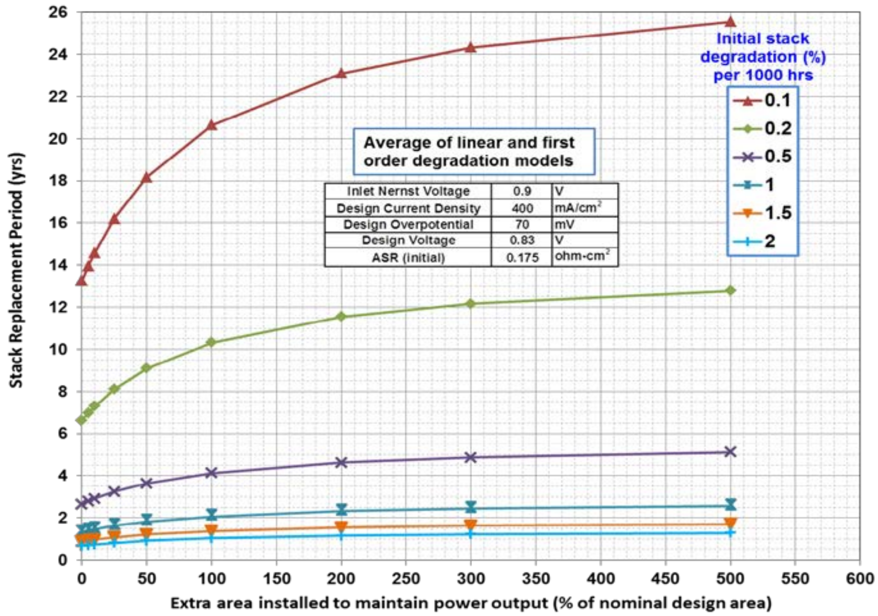
Source: NETL

Exhibit 2-20 Variation of stack replacement period with degradation rate and extra installed area assuming first order degradation



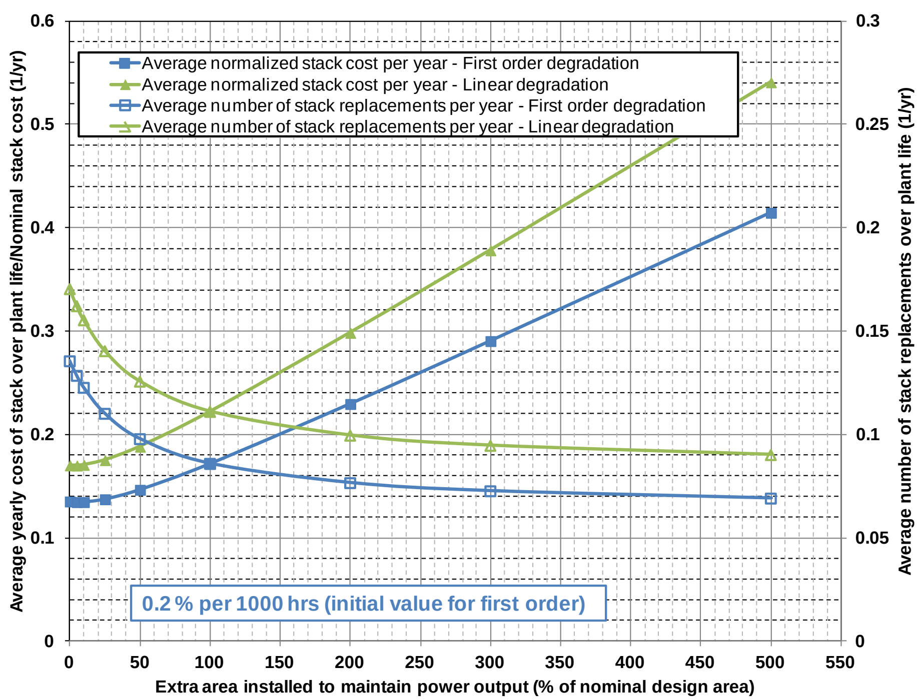
Source: NETL

Exhibit 2-21 Variation of the average stack replacement period with degradation rate and extra installed area – average of the values form the two models



Source: NETL

Exhibit 2-22 Average yearly cost of stack and the number of stack replacements per year for a 0.2% per 1000 h degradation rate for the two models



Source: NETL

2.13.4 Administrative and Support Labor

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

2.13.5 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.

2.13.6 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).

2.13.7 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.

The different components of the O&M costs used in the present study are summarized in Exhibit 2-23 for convenience.

Exhibit 2-23 Summary of O&M costs

Component	Specific Cost 2011$
Operating Labor Rate ($/h)	39.7
Stack Replacement O&M ($/h)	18.3
Water ($/1000 gal)	1.67
Chemicals	
MU & WT Chemicals ($/lb)	0.27
Carbon (Trace Removal) ($/lb)	1.6
COS Catalyst ($/m^3)	3,752
Selexol Solution ($/gal)	36.79
Claus / DSRP Catalyst ($ft^3)	36.79
ZnO polishing sorbent ($/lb)	1.8
KOH Coal Catalyst makeup ($/lb)	0.19
Lime ($/lb)	0.048
Waste Disposal	
Spent Trace Catalyst ($/lb)	0.65
Ash + HCl Sorbent ($/ton)	25.11
Spent sorbents ($/lb)	0.65
Fuel	
Coal ($/tons)	68.60
Natural Gas ($/MMBtu)	6.13

2.14 Owner's Costs

The owner's costs included in the TOC estimate are shown in Exhibit 2-24.

Exhibit 2-24 Owner's costs included in TOC

Owner's Cost	Comprising
Preproduction Costs	• 6 months operating, maintenance, and administrative and 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 and 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” 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.

2.15 CO_2 Transport and Storage

The CO_2 T&S costs were calculated as $11/tonne based on the Four-Basin study update. (16) Those costs were converted to a current-dollar, COE and combined with the plant capital and operating costs to produce an overall COE.

2.16 Cost of CO_2 Captured

The cost of CO_2 captured was computed using the equation,

$$Cost\ of\ CO_2\ Captured\ (\$/tonne\ CO_2) = \frac{COE_{IGFC\ System} - COE_{AUSC-PC\ without\ Capture}}{CO_2\ captured\ (tonnes/MWh)}$$

where,

$COE_{IGFC\ System}$ is the COE of the IGFC system under consideration, and $COE_{AUSC\text{-}PC\ without\ Capture}$ is the COE of the reference advanced ultrasupercritical (AUSC) PC without capture = $75.1/MWh.

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 utilizes advantages of SOFC operation at elevated pressures. The steam bottoming cycle represents a much smaller portion of the overall plant power generation relative to conventional 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 BB report. (1) Additional case-specific information is 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

A cryogenic ASU is assumed to generate the total 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 to 0.5 MPa (79 psia), providing oxygen product at sufficient pressure, 0.16 MPa (23 psia), to operate the oxy-combustor for the atmospheric pressure SOFC applications. The ASU is designed to generate 99.5 percent pure oxygen for IGFC applications to keep the nitrogen and argon content in the sequestered CO_2 stream low. Unlike IGCC plants, there is no opportunity for ASU air-side integration, as there is neither a requirement nor a benefit from syngas nitrogen dilution in the IGFC plant. In this study, the ASU nitrogen product was used only to provide an inert atmosphere, wherever applicable in the plant, and for transporting solids, 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 considered for this pathway: a conventional, CoP, entrained coal gasification technology, and a conceptual, near-term, enhanced coal gasifier. The conventional CoP E-Gas™ coal gasifier technology was selected for use in the IGFC plant because it can produce a syngas having a moderate CH_4 content of about 5.9 mole percent, which is higher than the CH_4 content of the syngas produced by 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 retains some CH_4. 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. The presence of CH_4 in the syngas is expected to be beneficial to the IGFC plant performance, because it reduces the excess cathode air flow needed for SOFC stack temperature control by providing local cooling of the SOFC stack through the endothermic reformation reaction.

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.

3.1.3.1 ConocoPhillips E-Gas™ Gasifier

The conventional, entrained, CoP E-Gas™ gasification technology represents one of the best conventional coal gasifier technologies for use in IGFC with its ability to generate a syngas having a moderate methane content of approximately 6 mole percent. The design basis and performance estimates for the CoP gasifier were based on the BB study (1) and the assumptions are shown in Exhibit 3-1. The cold gas efficiency of the gasifier was estimated to be 81 percent (higher heating value [HHV]).

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

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 transport 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.

3.1.3.2 Enhanced, Conventional Gasifier

The estimated operating parameters enhanced conventional gasifier, which represents a conceptual extrapolation of the CoP gasifier based on gasifier enhancement activities (17), are listed in Exhibit 3-2. The cold gas efficiency is estimated to be 82.5 percent (HHV) for this gasifier whose other features, including its stage-one characteristics, 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)	10.8
Raw Syngas Cooler	
Technology	Fire-tube boiler
Number in parallel	2
Outlet temperature, ºC (ºF)	316 (600)

3.1.4 Syngas Cleaning Area

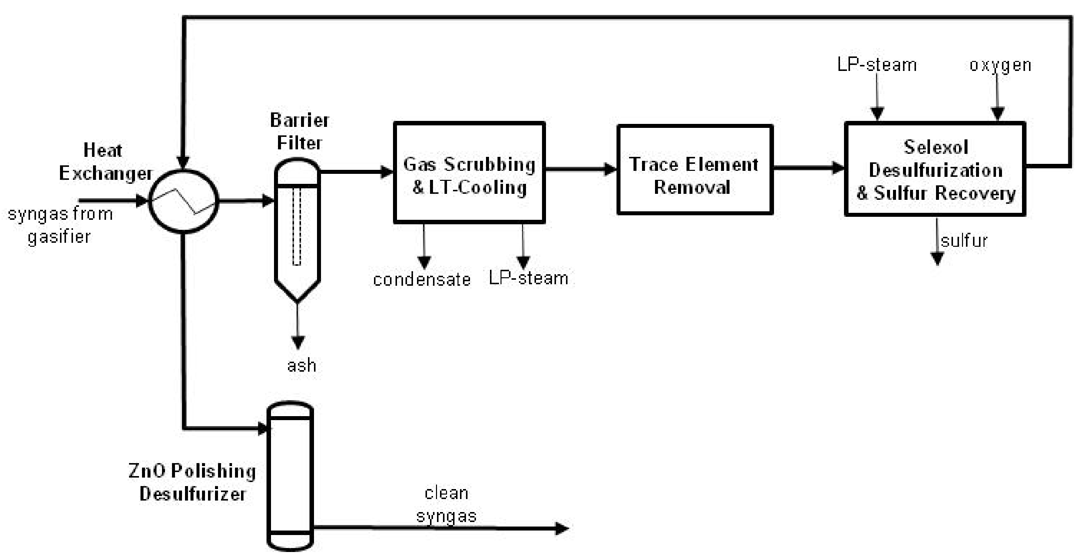
The function of the Gas Cleaning Area is to remove contaminants from the gasifier raw syngas to protect downstream equipment from damage in addition to satisfying the environmental emission requirements. In IGFC systems, the acceptable levels of contaminants in the syngas for SOFC operation are more stringent than the environmental requirements.

All of the IGFC plant configuration cases utilize conventional, dry gas cleaning technology. A single-stage Selexol AGR technology, which is expected to generate a cleaner syngas than alternatives such as amine-based AGR, is employed in all of the IGFC cases. The Selexol AGR is preceded by carbonyl sulfide (COS) hydrolysis and by low-temperature, activated-carbon beds to remove mercury and other trace elements. The syngas from the Selexol AGR step is reheated to support a conventional zinc oxide (ZnO) polishing step to further remove the sulfur to levels (less than 100 ppbv of total sulfur) that are acceptable to the SOFC. The clean syngas is expanded to the required pressure and fed to the fuel cell as its anode feed gas.

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 comprises the Gas Cleaning Area illustrated schematically in Exhibit 3-3. The configuration of the Gas Cleaning Area is nearly identical for both the conventional and catalytic gasifier pathways with minor differences in operating conditions. The reheating of the syngas for the subsequent ZnO polishing step is accomplished through high pressure (HP) steam indirect heating in the conventional gasifier cases unlike the recuperative gas-to-gas heat exchange employed for the catalytic gasifier cases 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 subsequent to the listed cleaning steps is acceptable to the SOFC unit for long term operation. This long-term success has not yet been demonstrated.

Exhibit 3-3 Syngas cleaning for IGFC



Source: NETL/DOE

Exhibit 3-4 Gas cleaning area assumptions conventional gasifier cases

Specification/Assumptions	
Gas cleaning technology	Conventional dry gas cleaning
Number of parallel trains	2
Particulate removal	Barrier filter at 316 ºC (600 ºF)
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 316 ºC (600 ºF)
Syngas Preheating Source	HP-steam heating for CoP gasifier

3.1.4.1 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.

The syngas, after particulate removal, is cooled to 232°C (450°F) through heat exchange with the steam cycle before entering the syngas scrubber in conventional gasifier cases.

3.1.4.2 Syngas Scrubber and Low-temperature Cooling Section

The cooled syngas passes to a syngas scrubber where a water wash is primarily used to remove 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), primarily removing 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.

3.1.4.3 COS Hydrolysis

The COS hydrolysis pretreatment provides a means to reduce the COS concentration in the feed to the acid gas removal process. 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 prior to entering a COS hydrolysis reactor. About 99.5 percent of the COS is converted to CO_2 and H_2S.

The equimolar COS hydrolysis reaction is represented as follows.

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

Although the slightly exothermic reaction favors higher conversion at lower temperatures, the reaction kinetics are slower at lower temperatures. 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 hydrogen sulfide (H_2S) and carbon monoxide (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, and 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.

3.1.4.4 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. (1) 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). (5)

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 was assumed to be 30 lb/ft^3 based on the Calgon Carbon Corporation HGR-P sulfur-impregnated pelletized activated carbon. (1) These parameters determined the size of the vessels and the amount of carbon required. The gasifier train has one mercury removal step.

Eastman Chemicals replaces its bed every 18 to 24 months. However, bed replacement is not because of mercury loading, but for other reasons including buildup of pressure drop, water, and other contaminants in the bed.

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. (18) 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.

3.1.4.5 Acid gas Removal Process

A key function of syngas cleaning is acid gas removal with sulfur recovery. The total sulfur content of the syngas is reduced to less than 30 ppmv including all sulfur species, but in particular the total of COS and H_2S. The Selexol process was chosen for AGR in all of the pathways based on the gasifier operation at high pressure which favors the physical solvent dimethyl ether of polyethylene glycol, used in the Selexol process. (19)

Cool, particulate-free syngas 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 H_2. 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.

3.1.4.6 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. (19)

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 two to three, where the remaining sulfur is recovered via the Claus reaction. Each catalytic stage consists of gas preheat, a catalytic reactor, and a sulfur condenser. The liquid sulfur goes to the sulfur pit, while the tail gas proceeds to the incinerator, or for further processing in a tail gas treating unit (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 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.

3.1.4.7 Sulfur Polishing

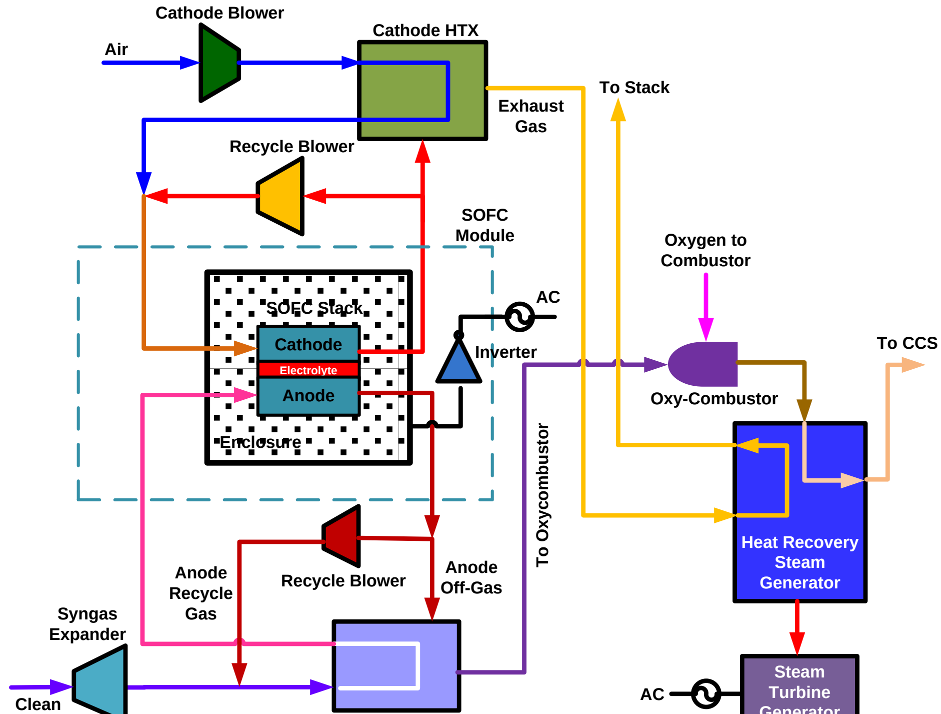
Several commercial sorbents are available for syngas sulfur polishing. Zinc oxide-based sorbents, having one of the highest affinities for hydrogen sulfide removal, are applicable for desulfurization to levels less than 100 ppbv and are offered by several catalyst vendors.

They operate at relatively high temperatures, 260-427°C (500-800°F) and are typically 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 refilled with fresh sorbent. The bulk desulfurized syngas from the Selexol unit is preheated by gas-to-gas heat exchange with the warm syngas from the barrier filter, or by indirect steam heating with high-pressure steam.

3.1.5 IGFC Power Island

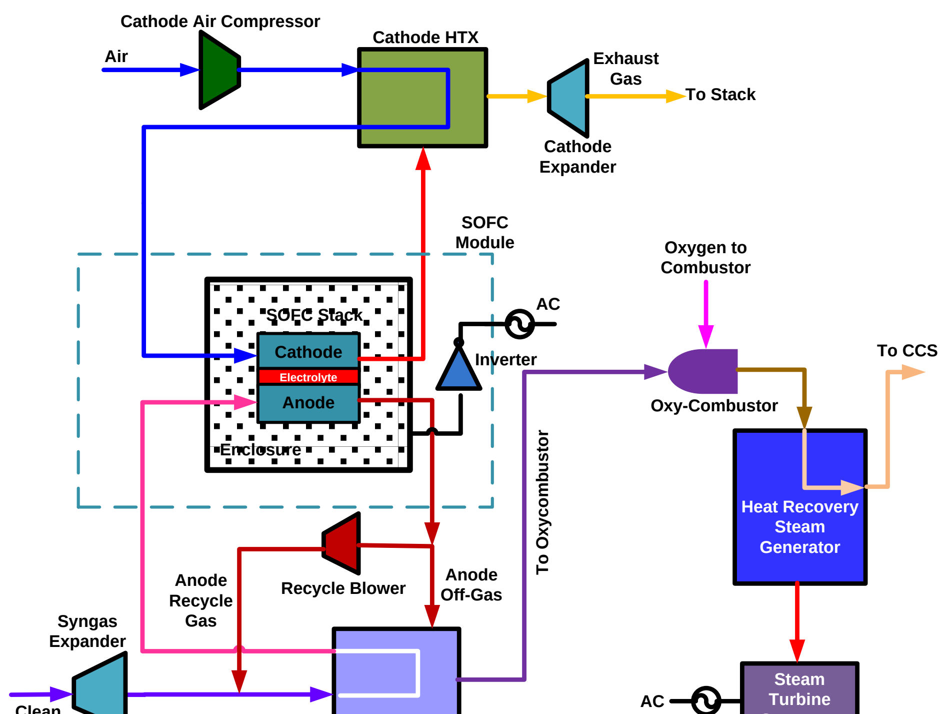
The IGFC power island for Scenario 1, shown in Exhibit 3-5, consists of a syngas expander that expands the syngas from its high-pressure condition down to the operating pressure of the fuel cell unit, the SOFC fuel cell unit with DC-AC inverters, an anode off-gas oxy-combustor, a heat recovery steam generator that captures heat from the combusted anode off-gas, and a steam bottoming cycle. The corresponding power island configuration for Scenario 2, which features pressurized fuel cell operation, is shown in Exhibit 3-6. In this case, the cathode off-gas is expanded to atmospheric pressure to generate power to drive the compressor that pressurizes the cathode air to the SOFC operating pressure. No cathode gas recycle is used while the anode gas recycle is accomplished using a syngas-driven jet pump, in this case.

Exhibit 3-5 IGFC power island – atmospheric SOFC



Source: NETL/DOE

Exhibit 3-6 IGFC power island – pressurized SOFC



Source: NETL/DOE

3.1.5.1 SOFC Unit

The SOFC 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. (20)

The major assumptions for the base case atmospheric-pressure SOFC power island are listed in Exhibit 3-7. 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 (or anywhere in the module). This constraint is satisfied by maintaining sufficiently high anode gas recycle, which increases 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 (e.g., sulfur species, HCl, NO_x, trace elements) is directed towards the CO_2 drying, purification, and compression unit discussed next.

Exhibit 3-7 Atmospheric-pressure power island base case 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)
Current density, mA/cm^2	400
Stack over-potential, mV	140
Operating voltage estimation method	Section 2.8.1
Cell degradation rate (% per 1000 hours)	1.5
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 (psi)	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	Atmospheric 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

3.1.5.2 Heat Recovery Steam Generator

The heat recovery steam generator produces steam to drive the subcritical steam bottoming cycle after meeting the low pressure (LP) and HP process steam needs. The 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).

3.1.5.3 Natural Gas Injection

An approach to achieve high methane content in the syngas, which is desirable from an IGFC performance perspective, is to inject natural gas into the cleaned syngas stream as in Case 1-6. Natural gas (NG), provided at 500 psia, was injected into the clean syngas, before it was expanded in this plant. An NG gas injection flow representing 38.5 percent of the total plant energy input, which resulted in dry syngas methane content ~24.6 mole percent (dry) was considered in this evaluation. The SOFC unit was assumed to accommodate the resulting cooling without any additional design modifications. A NG price of $6.13/MMBtu was used in estimating the costs.

3.1.5.4 Steam Bottoming Cycle

3.1.5.4.1 Steam Turbine Generator and Auxiliaries

The steam turbine consists of an HP section, an intermediate pressure (IP) section, and one double-flow LP section, all connected to the generator by a common shaft. The HP and IP sections are contained in a single-span, opposed-flow casing, with the double-flow LP section in a separate casing. 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 monitor operator interfacing, and datalink interfaces to the balance-of-plant distributed control system (DCS), and incorporates on-line repair capability.

3.1.5.4.2 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.

3.1.5.4.3 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.

3.1.5.4.4 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.

3.1.5.5 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.

3.1.6 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.

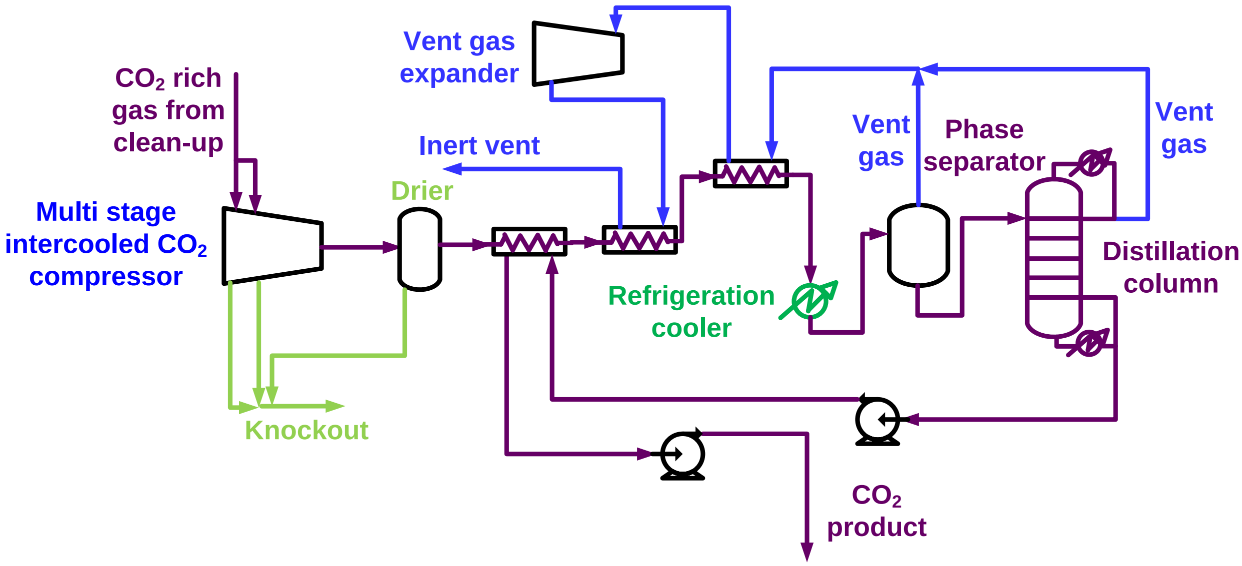
3.1.7 CO_2 Compression, Drying, and Purification

After completion of heat recovery, the oxy-combustion off-gas stream is sent to the CO_2 purification unit (CPU) for compression, drying, and purification to EOR specifications. (21) The CO_2 purification process, shown schematically in Exhibit 3-8 is based on an available quote, which was used in estimating the CPU costs. The CO_2 stream is dried and liquefied by cooling to -60 °F (to avoid the triple point to avert solid CO_2 formation), immediately following the initial compression to 30 bar (~ 450 psia). A distillation process subsequent to a phase separation stage subsequent to liquefaction is used to purify the CO_2 stream to EOR levels[4]. The cooling for the liquefaction is accomplished by a combination of external refrigeration and recuperative heat exchange with vent gases from the downstream distillation steps. The Aspen model of the corresponding process utilizes a RadFrac distiller with twenty stages to model the distillation, which is essentially a stripping column. A design spec that varied the boil-up ratio was used to control the O_2 purity at the exit of the distiller to a value of 10 ppm[5]. The reboiler heat was not assumed to be integrated with the external refrigeration cycle, which was not modeled.

[4] The QGESS reference (21) recommends a purity of the CO_2 at least 95 percent, as a conceptual design basis for EOR purposes; however, distillation methods used to meet the more stringent O_2 concentration requirement of 100 ppmv or less generally result in 99.9 percent + CO_2 purity.

[5] An O_2 concentration of 10ppmv in the CO_2 product stream was selected as the basis for conceptual design since it represents the lower limit of the range of values recommended in literature (21) for EOR applications. The number of distillation stages can be reduced slightly to design to the upper limit of 100 ppmv for O_2 concentration. (21) However, the impact of the associated small decrement in distillation cost on the overall cost is expected to be insignificant.

Exhibit 3-8 CO_2 purification process



Source: NETL/DOE

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 DCS is provided. The DCS is a redundant microprocessor-based, functionally distributed control system. The control room houses an array of multiple video monitors and keyboard units. The monitors/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. The 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, which is highly dependent on stoichiometry, temperature, and pressure.

3.2 Scenario 1 – IGFC with Atmospheric-Pressure SOFC

The Case 1-1 baseline configuration uses the conventional CoP E-Gas™ gasifier combined with an SOFC unit operating at atmospheric pressure. 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 together with a ZnO fixed-bed sulfur-polishing unit to supply clean syngas to the SOFC power island. The baseline, atmospheric pressure SOFC power island assumptions, and specifications are listed in Exhibit 3-7.

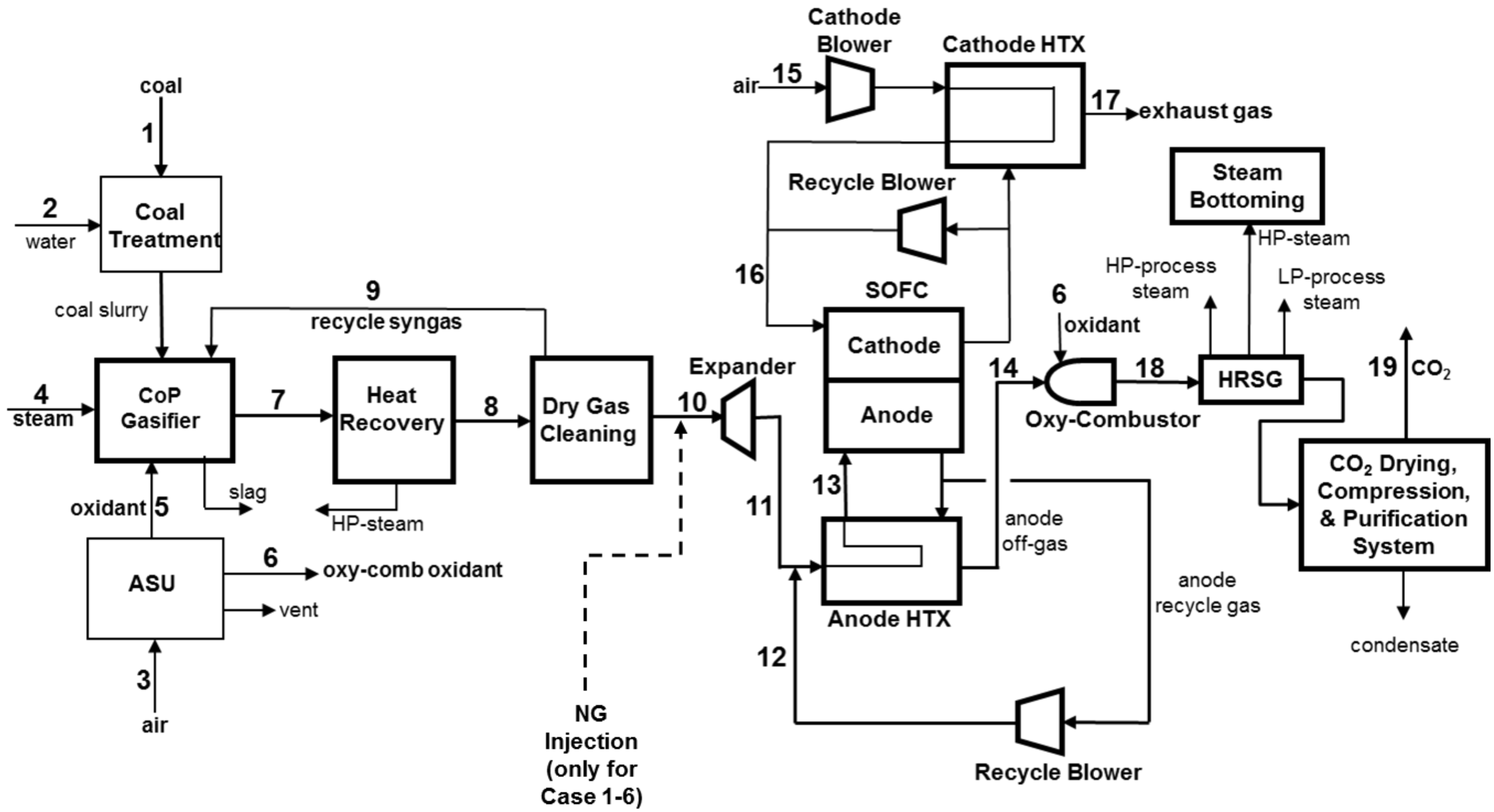
3.2.1 Case 1-1 Baseline Plant Performance Results

The relevant process data for the numbered streams in the block flow diagram (BFD) for the baseline plant shown in Exhibit 3-9 are tabulated in Exhibit 3-10. The syngas methane content of ~5.8 mole percent reduces to 1.8 mole percent in the anode inlet gas stream because of the 66 percent anode off-gas recirculation. The performance summary listed in Exhibit 3-11 shows that the baseline plant results in a net plant HHV efficiency of ~ 39.5 percent, which is much higher than the values achievable with conventional fossil fuel power plant technologies. The carbon capture rate for the process is 98.6 percent. Exhibit 3-12 shows that the steam cycle accounts for ~ 19 percent of the gross power generated, which is dominated by the SOFC power island contribution. The ASU and the CPU parasitics make up ~73 percent of the auxiliary load as shown in Exhibit 3-13. The SOFC power island losses[6] are not insignificant and comprise ~ 15 percent of the parasitic loads. The heat and mass balance diagram for the gasifier, ASU, and syngas clean-up is shown in Exhibit 3-14 while the corresponding process flow diagrams (PFD) for the power island and the CPU are shown in Exhibit 3-15 and Exhibit 3-16, respectively. Salient material and energy balances are shown in Exhibit 3-17. The nearly complete recovery of water from the oxy-combustion CO_2 product stream results in an IGFC plant water consumption, also shown in Exhibit 3-17, which is significantly lower than the corresponding value for conventional fossil fuel power plant technologies.

The IGFC plant acts as a nearly zero emission power plant, with the only significant emission being the small release of CO_2 shown in the emissions listed in Exhibit 3-17. This emissions performance is dictated, in part, by the need to protect the SOFC stack components from contamination.

[6] The DC-AC inverter losses and the SOFC polarization losses are included in the SOFC gross power estimates.

Exhibit 3-9 Case 1-1 block flow diagram



Source: DOE/NETL

Exhibit 3-10 Case 1-1 stream table

	1	2	3	4	5	6	7	8	9	10
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
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0443	0.0443	0.0577	0.0583
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2867	0.2867	0.3736	0.3774
CO_2	0.0000	0.0000	0.0003	0.0000	0.0000	0.0000	0.1549	0.1549	0.2023	0.2044
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	0.0003	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2675	0.2675	0.3486	0.3519
H_2O	0.0000	1.0000	0.0104	1.0000	0.0000	0.0000	0.2315	0.2315	0.0023	0.0013
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0008	0.0010	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0069	0.0069	0.0094	0.0000
N_2	0.0000	0.0000	0.7722	0.0000	0.0019	0.0019	0.0020	0.0020	0.0027	0.0060
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0044	0.0044	0.0016	0.0000
O_2	0.0000	0.0000	0.2077	0.0000	0.9950	0.9950	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	0.0000	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	4,209	24,698	3,377	3,940	907	23,506	23,506	2,706	15,182
V-L Flowrate (kg/hr)	0	75,820	712,634	60,835	126,159	29,047	486,792	486,792	58,249	325,385
Solids Flowrate (kg/hr)	184,567	0	0	0	0	0	172	172	0	0
Temperature (°C)	15	148	15	288	130	27	999	232	78	316
Pressure (MPa, abs)	0.10	0.69	0.10	3.45	3.45	0.16	3.10	2.96	4.14	2.32
Enthalpy (kJ/kg)[A]	---	566.85	31.06	2,973.90	114.27	23.91	2,286.76	859.84	103.53	484.63
Density (kg/m^3)	---	864.7	1.2	14.8	33.1	2.0	6.0	14.7	30.6	10.1
V-L Molecular Weight	---	18.015	28.854	18.015	32.016	32.016	20.710	20.710	21.528	21.433
V-L Flowrate (lb_{mol}/hr)	0	9,278	54,449	7,445	8,687	2,000	51,821	51,821	5,965	33,470
V-L Flowrate (lb/hr)	0	167,155	1,571,088	134,119	278,133	64,038	1,073,193	1,073,193	128,418	717,350
Solids Flowrate (lb/hr)	406,900	0	0	0	0	0	379	379	0	0
Temperature (°F)	59	299	59	550	265	80	1,830	450	173	600
Pressure (psia)	14.7	100.0	14.7	500.0	500.0	23.0	450.0	430.0	600.0	337.0
Enthalpy (Btu/lb)[A]	---	243.7	13.4	1,278.5	49.1	10.3	983.1	369.7	44.5	208.4
Density (lb/ft^3)	---	53.983	0.076	0.923	2.067	0.127	0.377	0.918	1.910	0.631

A - Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 3-10 Case 1-1 stream table (continued)

	11	12	13	14	15	16	17	18	19
V-L Mole Fraction									
Ar	0.0008	0.0008	0.0008	0.0008	0.0094	0.0096	0.0098	0.0009	0.0000
CH_4	0.0583	0.0000	0.0181	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.3774	0.0477	0.1502	0.0477	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.2044	0.5255	0.4256	0.5255	0.0003	0.0003	0.0003	0.5673	1.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.3519	0.0385	0.1359	0.0385	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0013	0.3823	0.2638	0.3823	0.0104	0.0106	0.0109	0.4164	0.0000
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0060	0.0053	0.0055	0.0053	0.7722	0.7892	0.8069	0.0054	0.0000
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.2077	0.1903	0.1721	0.0100	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	15,182	33,627	48,808	16,952	152,754	298,933	146,179	17,128	9,585
V-L Flowrate (kg/hr)	325,385	1,062,824	1,388,206	535,780	4,407,606	8,604,813	4,197,207	564,818	421,836
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0
Temperature (°C)	45	758	649	562	15	644	167	1,007	24
Pressure (MPa, abs)	0.14	0.11	0.11	0.1	0.101	0.109	0.106	0.102	15.272
Enthalpy (kJ/kg)[A]	67.37	1,550.40	1,309.12	1,262.5	31.057	700.982	187.381	1,934.728	-242.130
Density (kg/m^3)	1.1	0.4	0.4	0.5	1.2	0.4	0.8	0.3	740.6
V-L Molecular Weight	21.433	31.606	28.442	32	28.854	28.785	28.713	32.976	44.010
V-L Flowrate (lb_{mol}/hr)	33,470	74,135	107,603	37,372	336,765	659,035	322,270	37,761	21,131
V-L Flowrate (lb/hr)	717,350	2,343,126	3,060,470	1,181,194	9,717,108	18,970,366	9,253,258	1,245,210	929,989
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0
Temperature (°F)	113	1,397	1,200	1,043	59	1,192	333	1,845	74
Pressure (psia)	20.0	16.2	16.2	15.4	14.7	15.8	15.4	14.8	2,215.0
Enthalpy (Btu/lb)[A]	29.0	666.6	562.8	542.8	13.4	301.4	80.6	831.8	-104.1
Density (lb/ft^3)	0.070	0.026	0.026	0	0.076	0.026	0.052	0.020	46.235

A - Reference conditions are 32.02 F & 0.089 PSIA

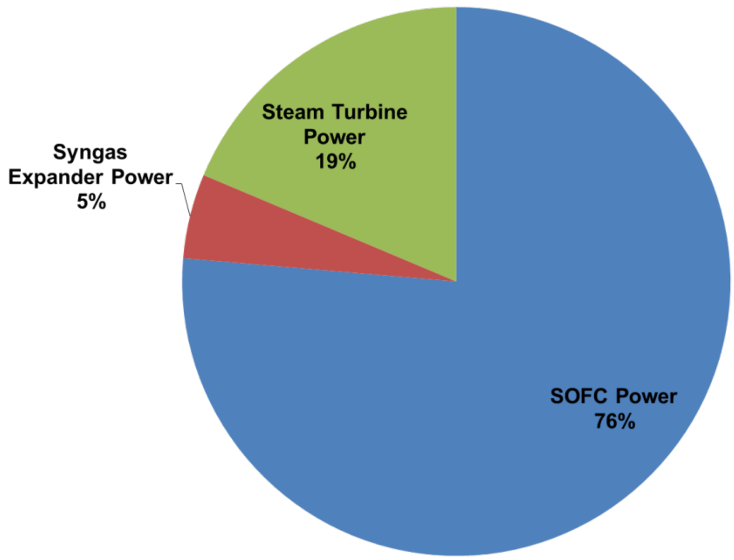
Exhibit 3-11 Case 1-1 plant performance summary (100 percent load)

POWER SUMMARY (Gross Power at Generator Terminals, kW_e)	
SOFC Power	557,300
Syngas Expander Power	36,400
Steam Turbine Power	136,200
TOTAL GROSS POWER (kW_e)	**729,900**
AUXILIARY LOAD SUMMARY, kW_e	
Coal Handling	430
Coal Milling	1,900
Sour Water Recycle Slurry Pump	168
Ash Handling	790
Air Separation Unit Auxiliaries	907
Air Separation Unit Main Air Compressor	41,910
Oxygen Compressor	12,760
Claus Plant TG Recycle Compressor	1,450
CO_2 Compression	39,760
CO_2 Purification	34,956
Boiler Feedwater Pumps	2,364
Condensate Pump	176
Syngas Recycle Compressor	1,050
Circulating Water Pump	2,830
Ground Water Pumps	437
Cooling Tower Fans	1,460
Scrubber Pumps	317
Quench Water Pump	58
Selexol Auxiliary Power	2,901
Steam Turbine Auxiliaries	57
Claus Plant/TGTU Auxiliaries	210
Gas Turbine Auxiliaries	78
Cathode Air Blower	11,530
Cathode Recycle Blower	11,540
Anode Recycle Blower	4,000
Miscellaneous Balance of Plant[2]	3,112
Transformer Losses	2,700
TOTAL AUXILIARIES, kW_e	**179,850**
NET POWER, kW_e	**550,050**
NET PLANT EFFICIENCY, % (HHV)	**39.5**
NET PLANT HEAT RATE, kJ/kWh (Btu/kWh)	**9,105 (8,630)**
CO_2 Capture Rate (%)	**98.6**
CONDENSER COOLING DUTY 10^6 kJ/h (10^6 Btu/h)	**717 (680)**
As-Received Coal Feed, kg/h (lb/h)	184,567 (406,900)
Thermal Input[1], kWt	1,391,178
Raw Water Consumption, m^3/min (gpm)	4.8 (1,278)

[1] HHV of as received Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb)

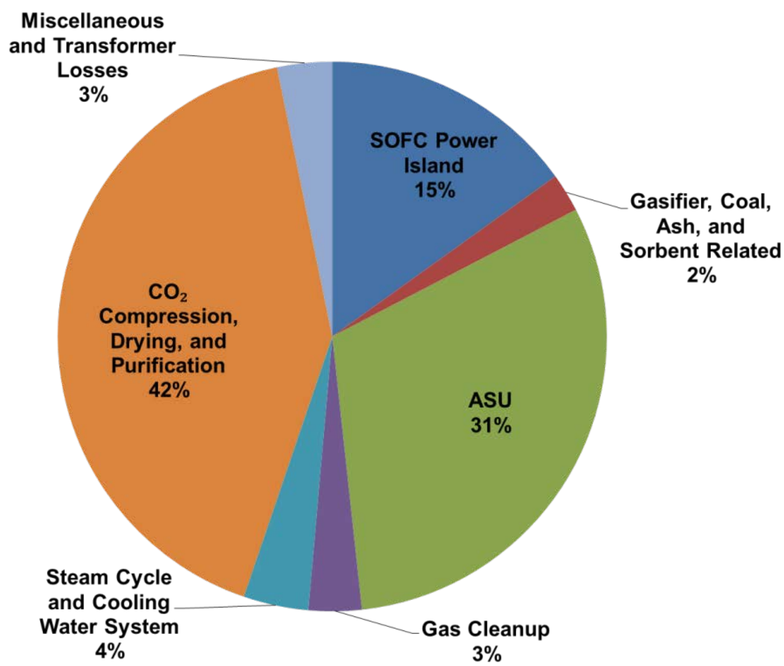
[2] Includes plant control systems, lighting, HVAC, and miscellaneous low-voltage loads

Exhibit 3-12 Case 1-1 gross power generation



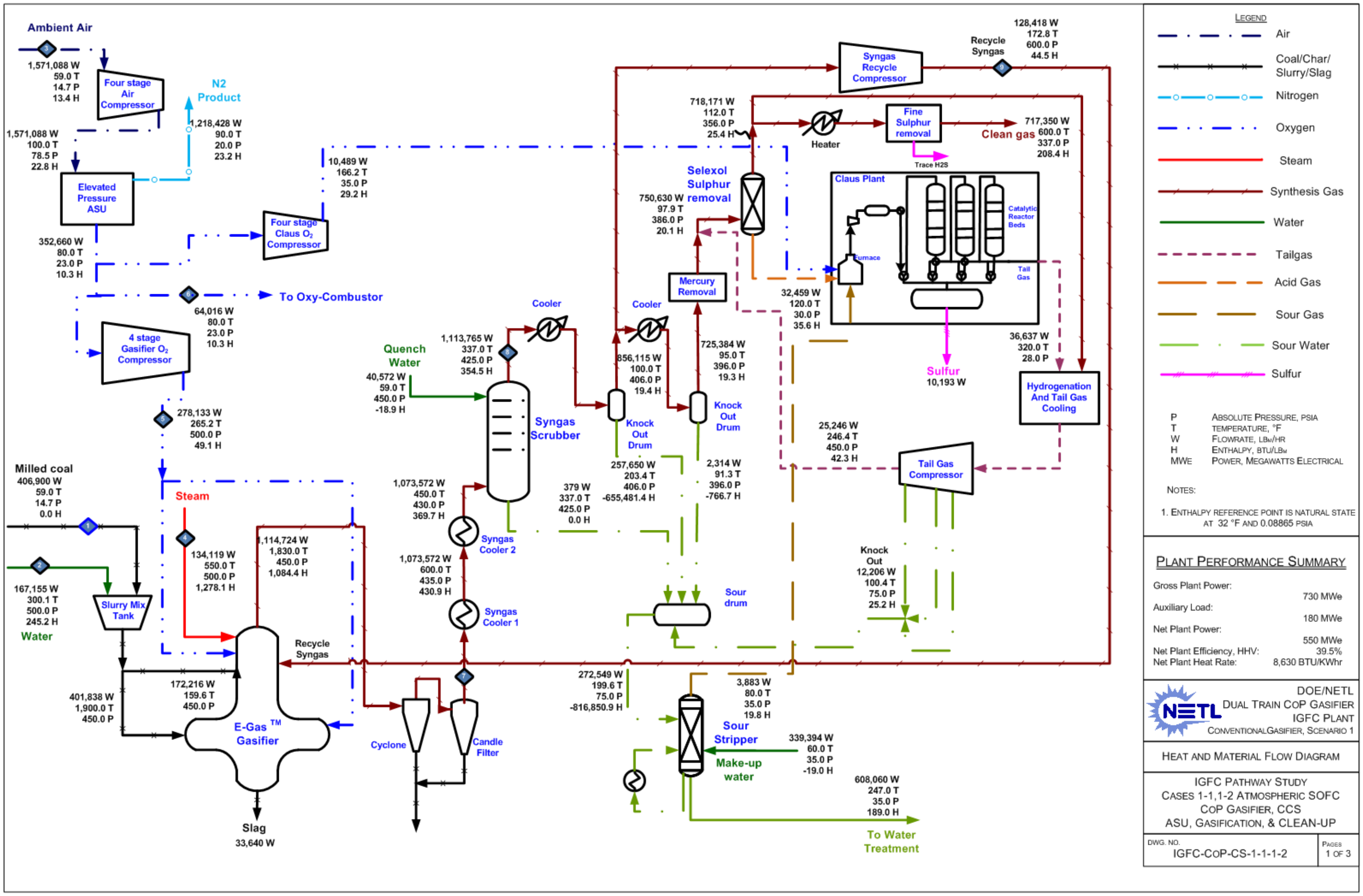
Source: DOE/NETL

Exhibit 3-13 Components of Case 1-1 auxiliary load



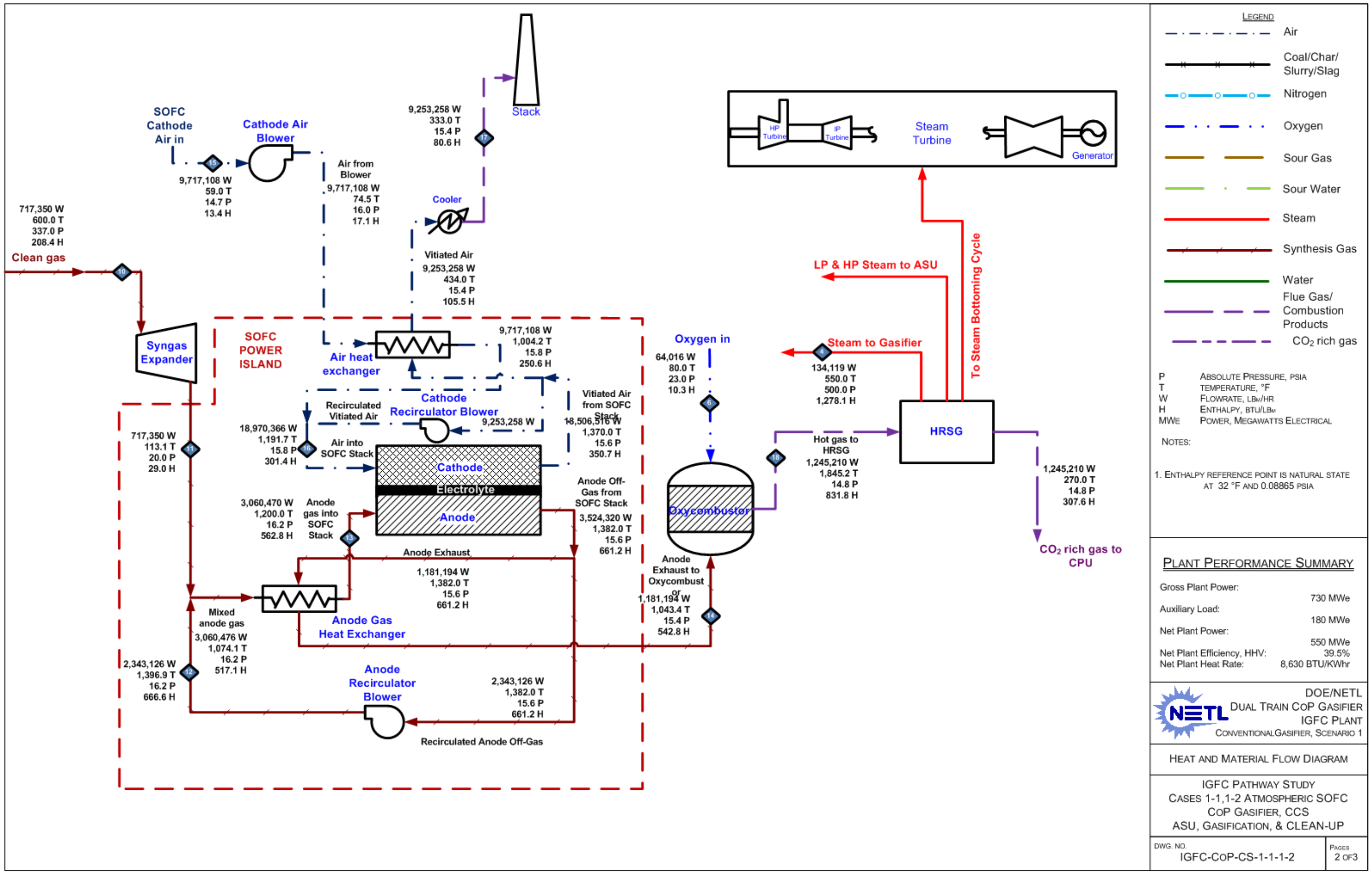
Source: DOE/NETL

Exhibit 3-14 Case 1-1 process flow diagram – gasifier, ASU, and syngas clean-up



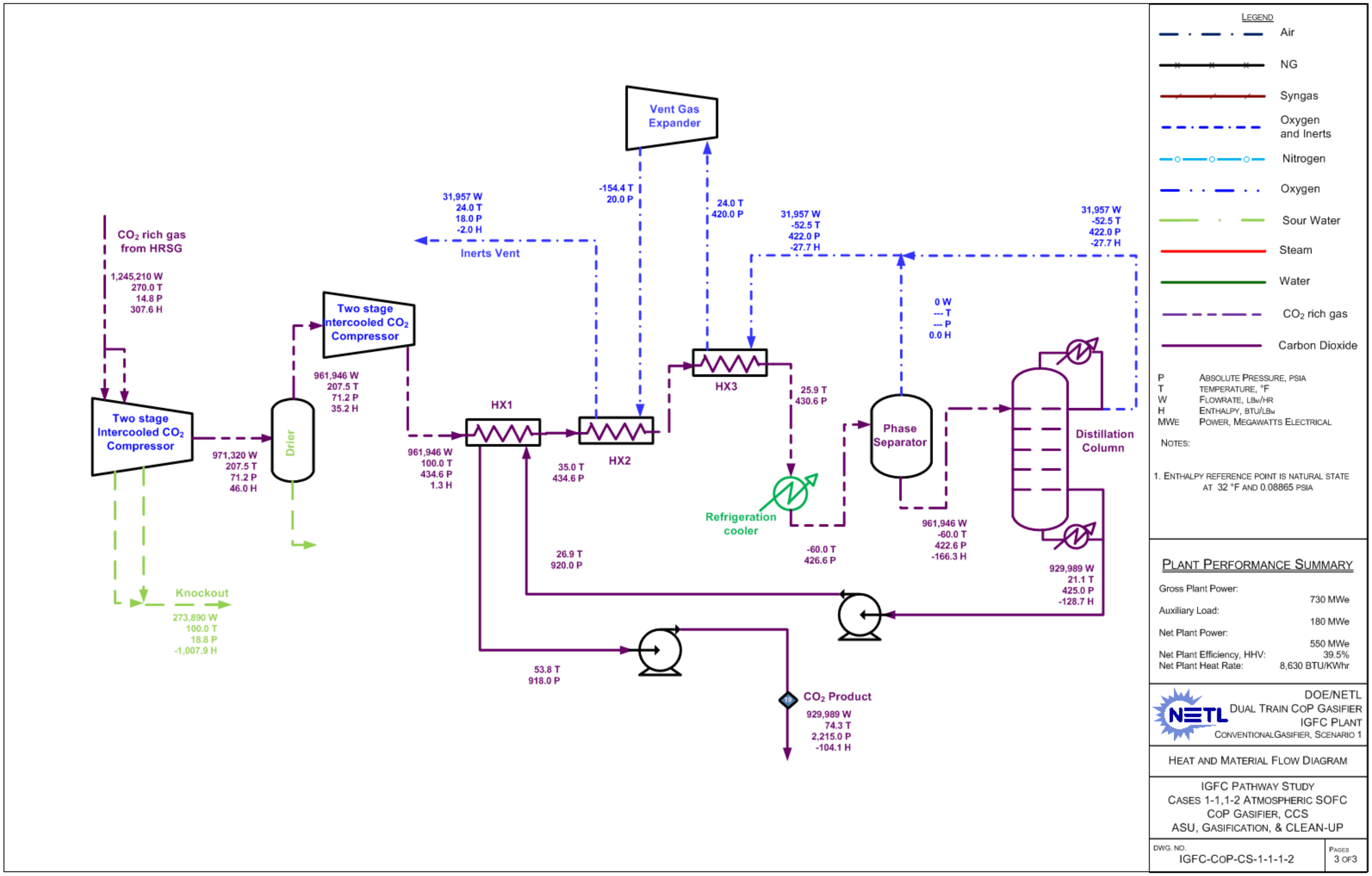
Source: DOE/NETL

Exhibit 3-15 Case 1-1 process flow diagram – IGFC power island



Source: DOE/NETL

Exhibit 3-16 Case 1-1 process flow diagram – CO_2 separation and purification



Source: DOE/NETL

Exhibit 3-17 Case 1-1 mass and energy balances

Carbon balance

Carbon In		Carbon Out	
kg/hr(lb/hr)		kg/hr(lb/hr)	
Coal	117,652 (259,377)	Slag	941 (2,075)
Air (CO2)	639 (1,410)	Stack Gas	550 (1,213)
		CO_2 Product	115,125 (253,807)
		N_2 Product	89 (196)
		Vent Gas	1,581 (3,486)
		Convergence Tolerance	4 (9)
Total	118,291 (260,787)	Total	118,291 (260,787)

Sulfur balance

Sulfur In		Sulfur Out	
kg/hr(lb/hr)		kg/hr(lb/hr)	
Coal	4,626 (10,199)	Elemental Sulfur	4,624 (10,193)
		Polishing Sorbent	2 (5)
		Convergence Tolerance	0 (0)
Total	4,626 (10,199)	Total	4,626 (10,199)

Energy balance

	HHV	Sensible + Latent	Power	Total
Heat In GJ/hr (MMBtu/hr)				
Coal	5,008 (4,747)	4.2 (4.0)		5,012 (4,751)
ASU Air		22.1 (21.0)		22 (21)
Fuel cell Air		136.9 (129.7)		137 (130)
Raw Water Makeup		62.1 (58.8)		62 (59)
Auxiliary Power			647 (614)	647 (614)
TOTAL	**5,008 (4,747)**	**225.3 (213.5)**	**647 (614)**	**5,881 (5,574)**
Heat Out GJ/hr (MMBtu/hr)				
Slag	25 (24)	15.7 (14.9)		41 (39)
Sulfur	43 (41)	0.5 (0.5)		43 (41)
CO_2		-102.1 (-96.8)		-102 (-97)
CO_2 Refrigeration		191.9 (181.9)		192 (182)
Cooling Tower Blowdown		18.4 (17.4)		18 (17)
Flue gas, Process Steam		786.5 (745.4)		786 (745)
Condenser		722 (684)		722 (684)
Non-Condenser Cooling Tower Loads		716 (678)		716 (678)
Process losses*		800 (759)		800 (759)
Difference**		16 (15)		16 (15)
Power			2,648 (2,510)	2,648 (2,510)
TOTAL	**68 (64)**	**3,165 (3,000)**	**2,648 (2,510)**	**5,881 (5,574)**

Water balance

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m3/min (gpm)	m3/min (gpm)	m3/min (gpm)	m3/min (gpm)	m3/min (gpm)
Slag Handling	0.33 (87)	0.33 (87)			
Slurry Water	1.27 (334)	1.27 (334)			
Quench/Wash	0.3 (81)	0.31 (81)			
Condenser Makeup	**1.1 (285)**	**0.0 (0)**	**1.1 (285)**	**0.0 (0)**	**1.1 (285)**
Gasifier Steam	1.0 (268)	0.0 (0)	1.0 (268)		
BFW Makeup	0.1 (17)	0.0 (0)	0.1 (17)		
Cooling Tower	**11.0 (2,909)**	**4.77 (1261)**	**6.2 (1,648)**	**2.5 (654)**	**3.8 (993)**
CO2 Dehydration	0.0 (0)	2.07 (548)	-2.07 (-548)		
SWS Excess Water	0.0 (0)	2.70 (713)	-2.70 (-713)		
Total	**14.0 (3,696)**	**6.68 (1764)**	**7.3 (1,933)**	**2.5 (654)**	**4.8 (1,278)**

Emissions

	kg/GJ ($lb/10^6 Btu$)	Tonne/year (tons/year)	kg/MWhgross (lb/MWhgross)
SO2	0 (0)	0 (0)	0 (0)
NOx	0 (0)	0 (0)	0 (0)
Particulate	0 (0)	0 (0)	0 (0)
Hg	0 (0)	0 (0)	0 (0)
CO2	2 (4)	54,779 (60,384)	11 (24)

** Process losses include losses from steam turbines, expanders, inverter, and blowers.*

***Value needed to match heat input to the plant and includes minor process losses.*

3.2.2 Case 1-1 Baseline Plant Cost Results

The SOFC power island capital costs are shown in Exhibit 3-18. The SOFC module costs account for a major portion, ~63 percent, of the SOFC power island costs as shown by the categorized cost distribution in Exhibit 3-19. The cathode side heat exchanger is the next significant expense contributing to ~19 percent of the SOFC power island costs.

Exhibit 3-18 Case 1-1 SOFC power island capital costs

Cost Component	Cost ($1000)	Specific Cost ($/kWe AC)
	2011$	
SOFC Module		
SOFC Stack	123,761	225
Enclosure	16,501	30
Transport and Placement	7,921	14
Site Foundations	24,422	44
Inverter	37,512	68
Total SOFC Module	**210,118**	**382**
Total SOFC Module with 10% Extra Installed Area	**231,130**	**420**
SYNGAS EXPANDER	**8,014**	**15**
SOFC BOP		
Cathode Air Blower	4,451	8
Cathode Gas Recycle Blower	9,994	18
Cathode Heat Exchanger	70,093	127
Anode Recycle Blower	1,030	2
Anode Heat Exchanger	27,642	50
Oxy-Combustor	13,960	25
Total SOFC BOP	**127,170**	**231**
TOTAL SOFC POWER ISLAND	**366,313**	**666**

Exhibit 3-19 Distribution of Case 1-1 SOFC power island capital costs

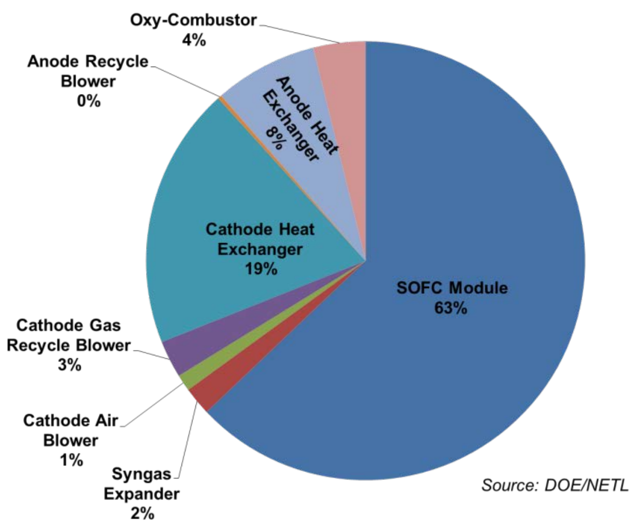
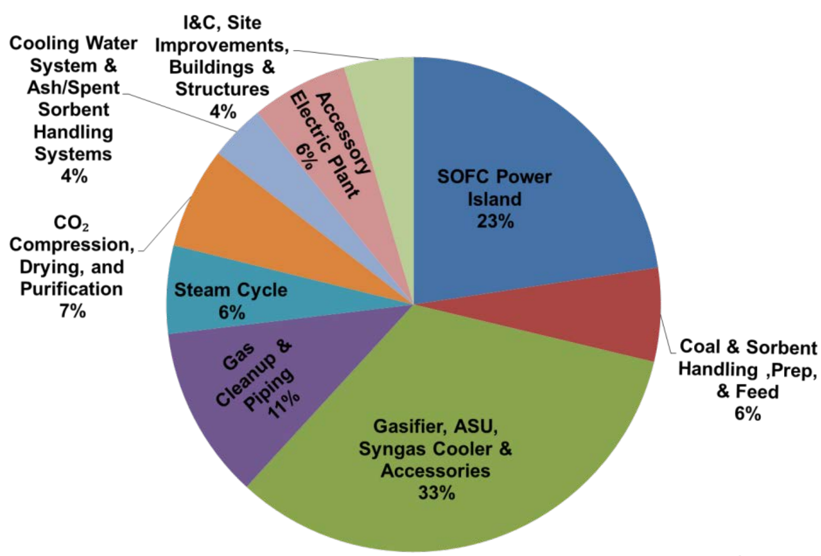


Exhibit 3-20, which depicts graphically the IGFC capital costs listed in Exhibit 3-21, indicates that the SOFC power island capital costs, the gasifier, and the ASU costs form ~ 53 percent of the total IGFC plant capital costs. The gasifier and syngas cooler costs are significant, ~ 295 MM$, and reflect the costs of two mechanically complex parallel two-stage CoP gasifiers that have high-temperature slagging pressure vessels, with 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 component cost at $975/kW.

The TOC, calculated as in Exhibit 3-22, results in COEs of ~$162/MWh and ~ $170/MWh with and without CO_2 T&S charges, respectively, as shown in Exhibit 3-23, which includes the O&M costs tabulated in Exhibit 3-24. The variable O&M is the dominant contributor to the COE, reflecting frequent stack replacement expenses associated with the stack degradation rate of 1.5 percent per 1000 h assumed in this baseline case. The importance of stack degradation rate becomes apparent from the results of the Scenario 1 pathway cases, which are discussed next.

Exhibit 3-20 Case 1-1 IGFC plant cost distribution



Source: DOE/NETL

Exhibit 3-21 Case 1-1 IGFC plant capital costs

Cost Component	Cost ($1000)	Specific Cost ($/kWe AC)
	2011$	
SOFC POWER ISLAND	**366,313**	**666**
COAL & SORBENT HANDLING	**38,858**	**71**
COAL & SORBENT PREP & FEED	**60,308**	**110**
GASIFIER, ASU, SYNGAS COOLER & ACCESSORIES		
Gasifier & Syngas Cooler	294,620	536
ASU & oxidant compressor	221,024	402
Other gasification equip & foundations	20,455	37
TOTAL GASIFIER, ASU, SYNGAS COOLER & ACCESSORIES	**536,099**	**975**
GAS CLEANUP & PIPING		
Scrubber & Low Temperature Cooling	33,691	61
Single-Stage Selexol/MDEA	87,317	159
Claus Plant	35,748	65
Trace removal	2,528	5
COS Hydrolysis	10,514	19
Blowback, Piping, Foundations	5,147	9
Sulfur polishing/NG desulfurization	7,873	14
TOTAL GAS CLEANUP & PIPING	**182,818**	**332**
STEAM CYCLE		
HRSG, Ducting, and Stack	28,809	52
Steam Power System	46,517	85
Feedwater and Misc BOP systems	17,899	33
TOTAL STEAM CYCLE	**93,225**	**169**
CO_2 COMPRESSION & PURIFICATION		
CO_2 Drying and Compression	-	-
CO_2 Purification	107,497	195
TOTAL CO_2 COMPRESSION & PURIFICATION	**107,497**	**195**
COOLING WATER SYSTEM	**24,494**	**45**
ASH/SPENT SORBENT HANDLING SYSTEM	**34,548**	**63**
ACCESSORY ELECTRIC PLANT	**102,581**	**186**
INSTRUMENTATION & CONTROL	**33,097**	**60**
IMPROVEMENTS TO SITE	**20,818**	**38**
BUILDING & STRUCTURES	**19,704**	**36**
TOTAL PLANT COST (TPC)	**1,620,361**	**2946**

Exhibit 3-22 Case 1-1 TOC and TASC

Cost Component	Cost ($1000)	Specific Cost ($/kWe AC)
	2011$	2011$
OWNER'S COSTS		
Preproduction Costs		
6 Months All Labor	9,942	
1 Month Maintenance Materials	2,109	
1 Month Non-fuel Consumables	349	
1 Month Waste Disposal	359	
25% of 1 Months Fuel Cost at 100% CF	2,547	
2% of TPC	32,407	
Total Preproduction Costs	**47,714**	
Inventory Capital		
60 day supply of fuel and consumables at 100% CF	20,926	
0.5% of TPC (spare parts)	8,102	
Total Inventory Capital	**29,028**	
Initial Cost for Catalyst and Chemicals	**12,193**	
Land	**900**	
Other Owner's Costs	**243,054**	
Financing Costs	**43,750**	
TOTAL OWNER'S COSTS	**376,639**	
TOTAL OVERNIGHT COST (TOC)	**1,997,000**	**3631**
TASC Multiplier	1.14	
TOTAL AS-SPENT COST (TASC)	**2,276,579**	**4139**

Exhibit 3-23 Case 1-1 cost of electricity

Cost Of Electricity (2011$/MWh) with CPU	
	$/MWh
Variable COE	**83.8**
Fuel	**25.4**
Variable O&M	**58.4**
Fixed O&M	**13.6**
Capital Charges	**64.4**
First Year COE (excluding T&S)	**161.7**
CO_2 T&S	**8.4**
First Year COE (with T&S)	**170.2**

Exhibit 3-24 Case 1-1 O&M costs

Case 1-1 Atm-SOFC CoP Gasifier						
Net Power: 550.0 Mwe, Capacity Factor: 80%, Heat Rate: 8,630 Btu/kWh						
Cost Component					**Cost ($)**	**Cost ($/MWh)**
					2011$	
OPERATING & MAINTENANCE LABOR						
Operating Labor Rate (base)					39.70	
Operating Labor Burden					30%	
Labor O-H Charge Rate					25%	
Total Operators per shift					12	
Maintenance labor/Operating labor					2.302	
Maintenance materials/Maintenance labor					1.932	
Stack replacement O&M, $/h per stack kW					18.33	
Annual Operating Labor Cost					5,425,243	
Maintenance Labor Cost					10,481,570	
Administrative & Support Labor					3,976,703	
Property Taxes and Insurance					32,441,258	
TOTAL FIXED OPERATING COSTS					**52,324,774**	**13.57**
VARIABLE OPERATING COSTS						
Maintenance Material Cost					**20,250,393**	**5.25**
Stack replacement						
SOFC stack life (years)					0.973	
Discount rate for stack replacement (%)					10.0%	
SOFC stack replacement cost, $/kW AC					$314	
SOFC Stack replacement O&M, $/yr per kW					$323	
Stack Replacement Cost					**198,081,070**	**51.39**
CONSUMABLES	**Consumption**		**Cost ($)**			
	Initial Fill	**/ Day**	**/ Unit**	**Initial Fill**		
Water (/1000 gallons)		1,472	1.670		**717,600**	
Chemicals						
MU & WT Chem. (lbs)		6,910	0.27		544,753	
Carbon (Trace Removal) (lb)	485,892	666	1.6	792,004	316,801	
COS Catalyst (m^3)	368	0.25	3,752	1,379,613	275,923	
Selexol Solution (gal)	248,702	39	36.79	9,149,747	421,076	
Claus / DSRP Catalyst (ft^3)		1.69	203		100,372	
ZnO polishing sorbent (lb)	484,147	1,859	1.8	871,465	976,931	
Sub Total Chemicals				**12,192,828**	**2,635,855**	
Waste Disposal						
Spent Trace Catalyst (lb)		719	0.65		136,438	
Ash + HCl Sorbent (ton)		404	25.11		2,959,863	
Spent sorbents (lb)		1,859	0.65		352,781	
Subtotal Waste Disposal					**3,449,082**	
TOTAL VARIABLE OPERATING COSTS				**12,192,828**	**225,134,001**	**58.40**
Fuel Coal (ton)		68.60			**97,808,343**	**25.37**

3.2.3 Scenario 1 Pathway Results

The Scenario 1 pathway estimated performance and cost for various SOFC system advances in a cumulative manner:

Case 1-2: The stack degradation rate was assumed to improve from the 1.5 percent /1000 hours in the baseline case 1-1 to 0.2 percent /1000 hours.

Case 1-3: An enhancement in cell performance was assumed in this case and the cell overpotential in Case 1-2 was reduced from 140 mV to 70 mV.

Case 1-4: Case 1-3 plant capacity factor was increased to 85 percent from 80 percent.

Case 1-5: The improvement in gasifier technology from the E-Gas to an Enhanced technology with methane increased to ~ 10.8 mole percent (dry) was explored in this case.

Case 1-6: NG injection to boost the dry CH_4 content of the syngas was analyzed in this case.

Case 1-7: The capacity factor of Case 1-5 was further increased to 90 percent.

Case 1-8: The stack cost was reduced from $225/kW in Case 1-7 to $200/kW.

Case 1-9: An improvement in the inverter efficiency from the 97 percent in Case 1-8 to 98 percent was analyzed in this case.

The performances and costs of the Scenario 1 pathway cases are summarized in Exhibit 3-25 and Exhibit 3-26, respectively. The net plant HHV efficiency varies from a value of 39.5 percent for the baseline case to a value of 50.7 percent for the NG injection case attributable mainly to the increased CH_4 methane content in the dry syngas. The effect of stack performance degradation rate is immediately visible in Exhibit 3-26 where the COE (without T&S) of Case 1-2 is lower than the baseline Case 1-1 COE (without T&S) by ~$36/MWh, a nearly 27 percent reduction, attributable directly to the decrease in degradation rate from 1.5 percent per 1000 h to 0.2 percent per 1000 h. The COE decreases progressively, albeit at a slower rate, for the other cases due to increases in performance and plant availability coupled with a reduction in stack cost. The combined effects of stack cost and stack degradation rate on the COE and the cost of captured CO_2 are presented in Exhibit 3-27 and Exhibit 3-28. It is clear from these exhibits that a stack degradation rate below 0.2 percent per 1000 h is necessary for the IGFC system to be competitive with conventional technologies.

Exhibit 3-25 Comparison of performance of Scenario 1 pathway cases

CASE	1-1 1-2	1-3 1-4	1-5 1-7 1-8	1-9	1-6
Dry Syngas CH_4 Content (%)	5.8		10.8		24.6
SOFC Overpotential (mV)	140	70			
Inverter Efficiency (%)	97			98	97
POWER SUMMARY (Gross Power at Generator Terminals, kW_e)					
SOFC Power	**557,300**	**562,200**	**552,000**	**551,900**	**551,100**
Syngas Expander Power	**36,400**	**33,800**	**33,100**	**32,800**	**19,300**
Steam Turbine Power	**136,200**	**117,800**	**113,100**	**112,000**	**87,700**
TOTAL GROSS POWER (kW_e)	**729,900**	**713,800**	**698,200**	**696,700**	**658,100**
AUXILIARY LOAD SUMMARY (kW_e)					
Coal Handling	430	420	400	400	340
Coal Milling	1,900	1,760	1,640	1,630	950
Sour Water Recycle Slurry Pump	168	156	145	144	85
Ash Handling	790	730	680	680	400
ASU Auxiliaries	907	842	719	712	479
ASU Main Air Compressor	41,910	38,940	33,260	32,940	22,140
Oxygen Compressor	12,760	11,850	10,750	10,650	6,280
Claus Plant TG Recycle Compressor	1,450	1,350	1,400	1,390	830
CO_2 Compression	39,760	36,940	34,550	34,200	26,820
CO_2 Purification	34,956	32,475	30,261	29,953	22,874
Boiler Feedwater Pumps	2,364	2,044	1,963	1,944	1,522
Condensate Pump	176	152	146	145	113
Syngas Recycle Compressor	1,050	970	600	590	350
Circulating Water Pump	2,830	2,540	2,430	2,400	2,000
Ground Water Pumps	437	406	378	374	220
Cooling Tower Fans	1,460	1,310	1,250	1,240	1,050
Scrubber Pumps	317	294	266	263	154
Quench Water Pump	58	54	360	359	210
Selexol Auxiliary Power	2,901	2,695	2,448	2,424	1,422
Steam Turbine Auxiliaries	57	49	47	47	37
Claus Plant/TGTU Auxiliaries	210	195	182	180	106
Gas Turbine Auxiliaries	78	73	71	71	42
Cathode Air Blower	11,530	9,090	7,570	7,490	5,980
Cathode Recycle Blower	11,540	9,020	7,440	7,360	5,760
Anode Recycle Blower	4,000	3,720	3,520	3,490	2,620
Miscellaneous Balance of Plant[2]	3,112	3,043	2,977	2,970	2,806
Transformer Losses	2,700	2,610	2,510	2,510	2,280
TOTAL AUXILIARIES (kW_e)	**179,850**	**163,729**	**147,964**	**146,556**	**107,868**
NET POWER (kW_e)	**550,050**	**550,071**	**550,236**	**550,144**	**550,232**
NET PLANT EFFICIENCY, % (HHV)	**39.5**	**42.6**	**45.7**	**46.1**	**51.9**
NET PLANT HEAT RATE, kJ/kWh (Btu/kWh)	**9,105 (8,630)**	**8,458 (8,017)**	**7,880 (7,468)**	**7,801 (7,394)**	**6,931 (6,569)**
CO_2 Capture Rate (%)	**98.6**	**98.6**	**98.6**	**98.6**	**98.5**
CONDENSER COOLING DUTY 10^6 kJ/h (10^6 Btu/h)	**717 (680)**	**622 (590)**	**601 (570)**	**591 (560)**	**464 (440)**
As-Received Coal Feed, kg/h (lb/h)	184,567 (406,900)	157,306 (346,800)	166,775 (367,675)	165,092 (363,965)	93,395 (205,900)
NG Feed Rate, kg/h (lb/h)	-	-		-	26,088 (57,515)
Thermal Input[1], kWt	1,391,178	1,185,698	1,257,069	1,244,384	1,084,910
Raw Water Consumption, m^3/min	4.8	4.1	5.8	5.8	4.2

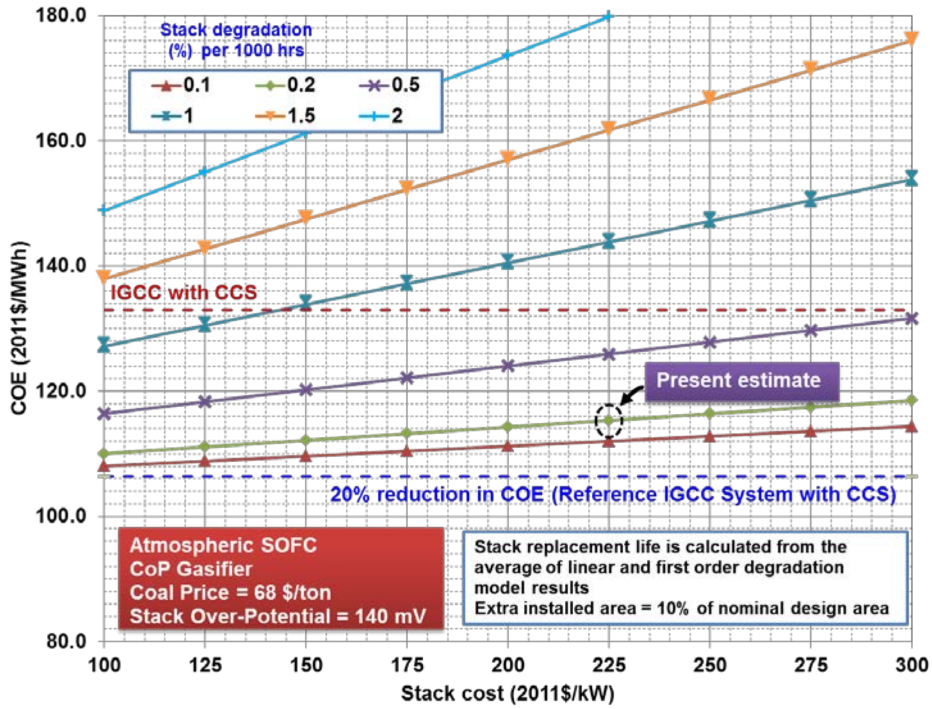
[1] *HHV of as received Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb)*

[2] *Includes plant control systems, lighting, HVAC, and miscellaneous low-voltage loads*

Exhibit 3-26 Comparison of cost of Scenario 1 pathway cases

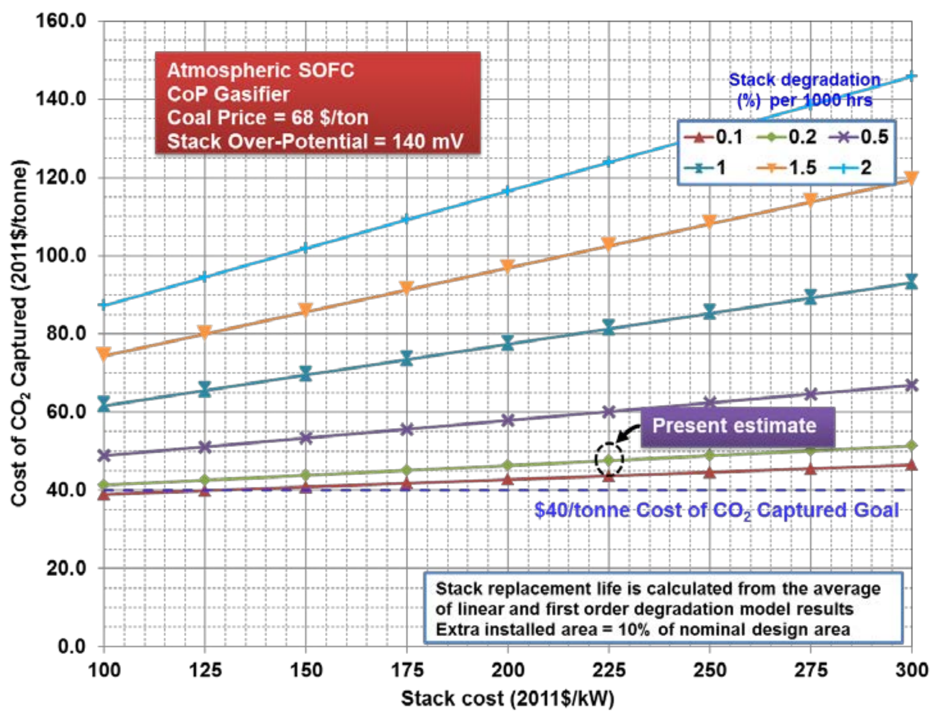
CASE	1-1	1-2	1-3	1-4	1-5	1-6	1-7	1-8	1-9
SOFC Degradation Rate (%/1000 h)	1.5	0.2							
SOFC Overpotential (mV)	140		70						
Capacity Factor	80			85			90		
Dry Syngas CH_4 Content (%)	5.8				10.8	24.6	10.8		
SOFC Stack Cost ($/kW)	225							200	
Inverter Efficiency (%)	97								98
PERFORMANCE									
Gross Power (MWe)	729.9	729.9	713.8	713.8	698.2	658.1	698.2	698.2	696.7
Auxiliary Loads (MWe)	179.9	179.9	163.7	163.7	148.0	107.9	148.0	148.0	146.6
Net Power (MWe)	550.0	550.0	550.1	550.1	550.1	550.3	550.1	550.1	550.0
Net Electric Efficiency, HHV (%)	39.5	39.5	42.6	42.6	45.7	51.9	45.7	45.7	46.1
CO_2 Capture rate (%)	98.6	98.6	98.6	98.6	98.6	98.5	98.6	98.6	98.6
CO_2 Emissions (lb/MWhgross)	23.6	23.6	21.5	21.5	20.1	17.8	20.1	20.1	19.9
CO_2 Emissions (lb/MWhnet)	31.3	31.3	27.9	27.9	25.5	21.2	25.5	25.5	25.3
Raw Water Consumption (gpm/MWnet)	2.3	2.3	2.0	2.0	2.7	2.0	2.7	2.7	2.0
COST									
Total Plant Cost (TPC) (1000$)	1,620,361	1,620,361	1,531,544	1,531,544	1,450,569	1,166,265	1,450,569	1,435,438	1,426,949
Total Overnight Cost (TOC) (1000$)	1,997,000	1,997,000	1,887,644	1,887,524	1,787,666	1,432,881	1,787,561	1,769,373	1,758,854
Total As-Spent Cost (TASC) (1000$)	2,276,579	2,276,579	2,151,914	2,151,777	2,037,939	1,633,485	2,037,820	2,017,085	2,005,093
Cost of Electricity ($/MWh)									
Variable COE	83.8	37.4	35.5	34.9	33.2	37.0	32.7	32.3	32.0
Fuel	25.4	25.4	23.6	23.6	22.0	26.4	22.0	22.0	21.7
Variable O&M	58.4	12.0	11.9	11.3	11.2	10.6	10.7	10.3	10.3
Fixed O&M	13.6	13.6	13.1	12.3	11.9	10.5	11.3	11.2	11.2
Capital Charges	64.4	64.4	60.9	57.3	54.2	43.5	51.2	50.7	50.4
Total First Year COE (excluding T&S)	161.7	115.3	109.4	104.5	99.3	91.0	95.2	94.2	93.6
CO_2 T&S	8.4	8.4	7.8	7.8	7.3	5.5	7.3	7.3	7.2
Total First Year COE (including T&S)	170.2	123.8	117.3	112.3	106.6	96.5	102.4	101.5	100.8
NETL Metrics									
% COE reduction (COE_{IGFC} - $COE_{IGCC\ with\ CCS}$)/ $COE_{IGCC\ with\ CCS}$	-21.6	13.3	17.7	21.4	25.3	31.6	28.5	29.2	29.6
Cost of Captured CO_2 ($/tonne CO_2) (COE_{IGFC} - $COE_{AUSC\ PC}$)/ CO_2 Captured (tonnes/MWh)	113.0	52.5	48.2	41.2	36.5	31.7	30.2	28.8	28.2

Exhibit 3-27 Sensitivity of Case 1-2 COE to stack cost and degradation rate



Source: DOE/NETL

Exhibit 3-28 Sensitivity of Case 1-2 cost of CO_2 captured to stack cost and degradation rate



Source: DOE/NETL

3.3 Scenario 2 – IGFC with Pressurized-SOFC

Scenario 2 applies the enhanced conventional coal gasifier technology with a pressurized SOFC unit. 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 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. Further optimization of the pressurized configuration and its operating conditions are recommended and could produce superior results over 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 listed in Exhibit 3-29. The CPU remains essentially the same with the LP compressor working at a lower pressure ratio. However, no cost benefit was assumed in this study to reflect the reduced LP compressor requirements.

Exhibit 3-29 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

3.3.1 Case 2-1 IGFC Plant Performance Results

The results are presented in the same fashion as in Scenario 1 and include the following:

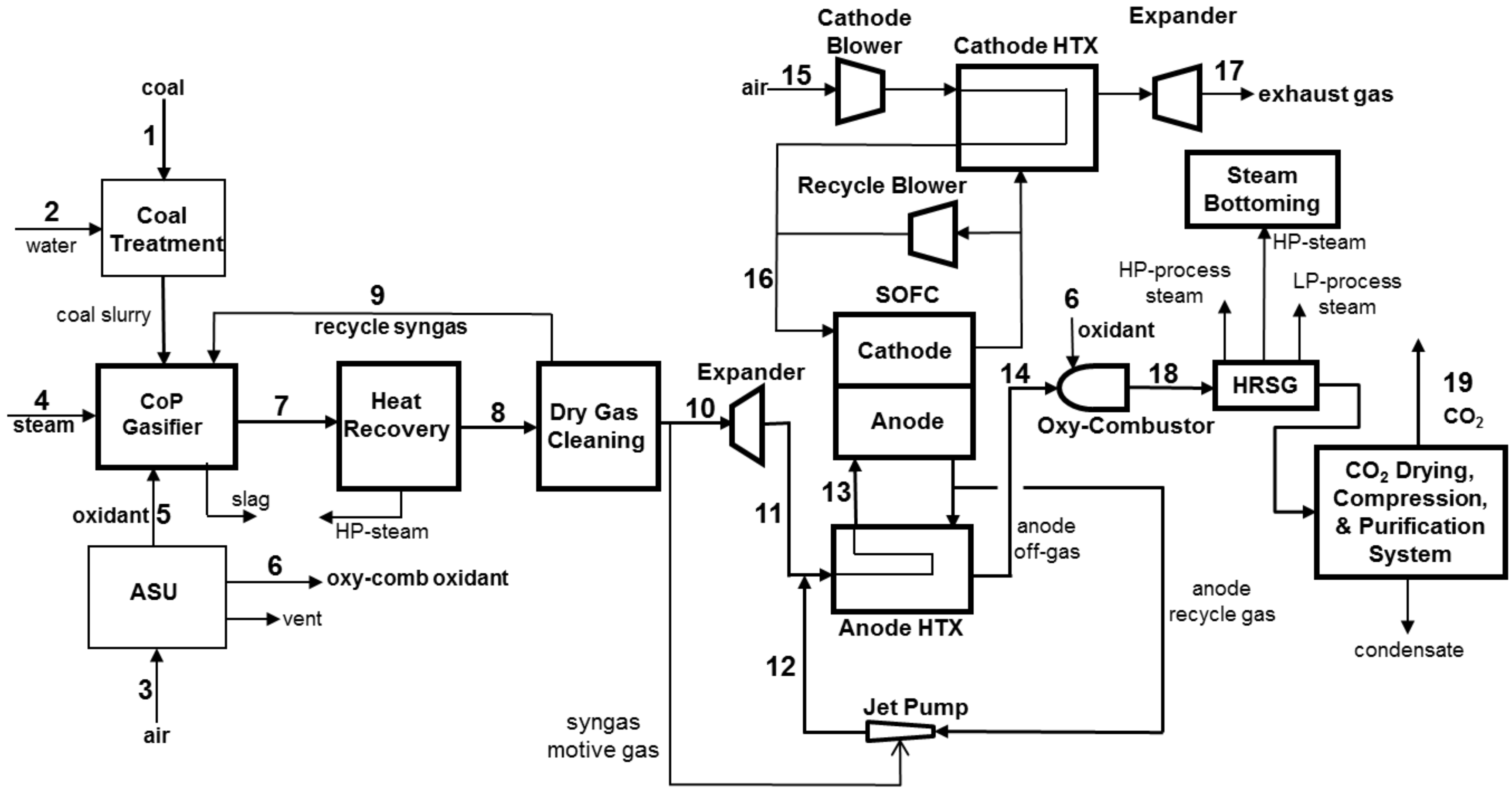
- BFD and stream table
- Performance summary
- Heat and mass balance diagrams
- Material and energy balances

The relevant process data for the numbered streams in the BFD for the baseline plant shown in Exhibit 3-30 are tabulated in Exhibit 3-31. The syngas methane content of ~10.8 mole percent reduces to 3.0 mole percent in the anode inlet gas stream, because of the 66 percent anode off-gas recirculation. The performance summary listed in Exhibit 3-32 shows that the baseline plant results in a net plant HHV efficiency of ~46.0 percent, which is much higher than the values achievable with conventional fossil fuel power plant technologies. The carbon capture rate for the process is 98.6 percent as in the atmospheric SOFC cases. Exhibit 3-32 shows that the steam cycle accounts for ~17 percent of the gross power generated, which is dominated by the SOFC power island contribution. The ASU and the CPU parasitics make up ~51 percent of the auxiliary load, as shown in Exhibit 3-34. The SOFC power island losses[7] are higher than in Case 1-1 and comprise ~37 percent of the parasitic loads mainly due to the compressor losses. The SOFC operating voltage is 0.94 V, a direct result of the SOFC pressurization. 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 20 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 4.0 percent of the coal energy, the cathode air compressor-expander at ~5.0 percent, and the CO_2 compression area at ~3.0 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. The heat and mass balance diagram for the gasifier, ASU, and syngas clean-up is shown in Exhibit 3-35 while the corresponding PFDs for the power island and the CPU are shown in Exhibit 3-36 and Exhibit 3-37, respectively. Salient material and energy balances are shown in Exhibit 3-38. The nearly complete recovery of water from the oxy-combustion CO_2 product stream results in an IGFC plant water consumption, also shown in Exhibit 3-38, which is significantly lower than the corresponding value for conventional fossil fuel power plant technologies.

The IGFC plant acts as a nearly zero emission power plant, with the only significant emission being the small release of CO_2 shown in the emissions listed in Exhibit 3-38. This emissions performance is dictated, in part, by the need to protect the SOFC stack components from contamination.

[7] The DC-AC inverter losses and the SOFC polarization losses are included in the SOFC gross power estimates.

Exhibit 3-30 Case 2-1 block flow diagram



Source: DOE/NETL

Exhibit 3-31 Case 2-1 stream table

	1	2	3	4	5	6	7	8	9	10
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
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0827	0.0827	0.1078	0.1089
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2581	0.2581	0.3366	0.3401
CO_2	0.0000	0.0000	0.0003	0.0000	0.0000	0.0000	0.1802	0.1802	0.2352	0.2383
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0004	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2315	0.2315	0.3019	0.3049
H_2O	0.0000	1.0000	0.0104	1.0000	0.0000	0.0000	0.2315	0.2315	0.0016	0.0006
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0008	0.0011	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0073	0.0073	0.0099	0.0000
N_2	0.0000	0.0000	0.7722	0.0000	0.0019	0.0019	0.0022	0.0022	0.0029	0.0063
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0047	0.0047	0.0024	0.0000
O_2	0.0000	0.0000	0.2077	0.0000	0.9950	0.9950	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	0.0000	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	3,619	19,971	2,829	3,100	811	18,852	18,852	2,168	10,333
V-L Flowrate (kg/hr)	0	65,199	576,247	50,956	99,254	25,960	407,079	407,079	49,157	233,512
Solids Flowrate (kg/hr)	158,712	0	0	0	0	0	148	148	0	0
Temperature (°C)	15	148	15	288	139	27	999	232	66	316
Pressure (MPa, abs)	0.10	0.69	0.10	5.52	4.59	0.16	4.83	4.69	6.00	3.80
Enthalpy (kJ/kg)[A]	---	566.85	31.06	2,918.18	122.01	23.91	2,269.81	824.88	72.27	472.49
Density (kg/m^3)	---	864.7	1.2	25.6	43.0	2.0	9.8	24.4	49.0	17.4
V-L Molecular Weight	---	18.015	28.854	18.015	32.016	32.016	21.594	21.594	22.673	22.599
V-L Flowrate (lb_{mol}/hr)	0	7,979	44,028	6,236	6,835	1,788	41,561	41,561	4,780	22,780
V-L Flowrate (lb/hr)	0	143,739	1,270,408	112,340	218,817	57,232	897,455	897,455	108,373	514,805
Solids Flowrate (lb/hr)	349,900	0	0	0	0	0	326	326	0	0
Temperature (°F)	59	299	59	550	283	80	1,830	450	151	600
Pressure (psia)	14.7	100.0	14.7	800.0	665.0	23.0	700.0	680.0	870.0	551.0
Enthalpy (Btu/lb)[A]	---	243.7	13.4	1,254.6	52.5	10.3	975.8	354.6	31.1	203.1
Density (lb/ft^3)	---	53.983	0.076	1.597	2.683	0.127	0.610	1.522	3.061	1.083

A - Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 3-32 Case 2-1 stream table (continued)

	11	12	13	14	15	16	17	18	19
V-L Mole Fraction									
Ar	0.0008	0.0007	0.0007	0.0007	0.0094	0.0094	0.0099	0.0008	0.0000
CH_4	0.1089	0.0070	0.0340	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.3401	0.0671	0.1396	0.0484	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.2383	0.4983	0.4292	0.5160	0.0003	0.0003	0.0003	0.5586	1.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.3049	0.0571	0.1229	0.0402	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0006	0.3646	0.2679	0.3894	0.0104	0.0104	0.0109	0.4252	0.0000
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0063	0.0053	0.0056	0.0052	0.7722	0.7722	0.8098	0.0053	0.0000
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.2077	0.2077	0.1691	0.0100	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	10,333	28,572	38,907	14,805	127,199	127,189	121,278	14,960	8,242
V-L Flowrate (kg/hr)	233,512	879,431	1,113,006	463,933	3,670,238	3,669,940	3,480,789	489,894	362,711
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0	0	0
Temperature (°C)	238	727	615	750	15	649	167	1,186	24
Pressure (MPa, abs)	2.00	2.00	1.98	1.9	0.101	1.979	0.100	1.889	15.272
Enthalpy (kJ/kg)^A	348.90	1,505.81	1,263.11	1,556.6	31.057	705.665	187.516	2,243.102	-242.125
Density (kg/m^3)	10.6	7.4	7.6	7.1	1.2	7.4	0.8	5.1	740.6
V-L Molecular Weight	22.599	30.779	28.607	31	28.854	28.854	28.701	32.748	44.010
V-L Flowrate (lb_{mol}/hr)	22,780	62,991	85,775	32,639	280,426	280,404	267,372	32,980	18,170
V-L Flowrate (lb/hr)	514,805	1,938,814	2,453,758	1,022,798	8,091,490	8,090,832	7,673,826	1,080,031	799,641
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0	0	0
Temperature (°F)	460	1,340	1,138	1,382	59	1,200	333	2,167	74
Pressure (psia)	290.0	290.0	287.0	282.0	14.7	287.0	14.5	274.0	2,215.0
Enthalpy (Btu/lb)^A	150.0	647.4	543.0	669.2	13.4	303.4	80.6	964.4	-104.1
Density (lb/ft^3)	0.661	0.461	0.478	0	0.076	0.462	0.049	0.317	46.233

A - Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 3-32 Case 2-1 plant performance summary (100 percent load)

POWER SUMMARY (Gross Power at Generator Terminals, kW_e)	
SOFC Power	583,200
Syngas Expander Power	7,700
Steam Turbine Power	121,600
TOTAL GROSS POWER (kW_e)	**712,500**
AUXILIARY LOAD SUMMARY (kW_e)	
Coal Handling	400
Coal Milling	1,630
Sour Water Recycle Slurry Pump	145
Ash Handling	680
Air Separation Unit Auxiliaries	733
Air Separation Unit Main Air Compressor	33,890
Oxygen Compressor	13,070
Claus Plant TG Recycle Compressor	1,400
CO_2 Compression	5,180
CO_2 Purification	30,057
Boiler Feedwater Pumps	2,110
Condensate Pump	157
Syngas Recycle Compressor	590
Circulating Water Pump	2,330
Ground Water Pumps	376
Cooling Tower Fans	1,210
Scrubber Pumps	265
Quench Water Pump	358
Selexol Auxiliary Power	2,447
Steam Turbine Auxiliaries	51
Claus Plant/TGTU Auxiliaries	181
Gas Turbine Auxiliaries	128
Cathode Air Compressor - Cathode Expander	59,200
Miscellaneous Balance of Plant[2]	3,038
Transformer Losses	2,590
TOTAL AUXILIARIES, kW_e	162,215
NET POWER (kW_e)	**550,285**
NET PLANT EFFICIENCY, % (HHV)	**46.0**
NET PLANT HEAT RATE, kJ/kWh (Btu/kWh)	**7,826 (7,418)**
CO_2 Capture Rate (%)	**98.6**
CONDENSER COOLING DUTY 10^6 kJ/h (10^6 Btu/h)	**644 (610)**
As-Received Coal Feed, kg/h (lb/h)	158,712 (349,900)
Thermal Input[1], kWt	1,196,297
Raw Water Consumption, m^3/min (gpm)	5.3 (1,412)

[1] *HHV of as received Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb)*

[2] *Includes plant control systems, lighting, HVAC, and miscellaneous low-voltage loads*

Exhibit 3-33 Case 2-1 power generation components

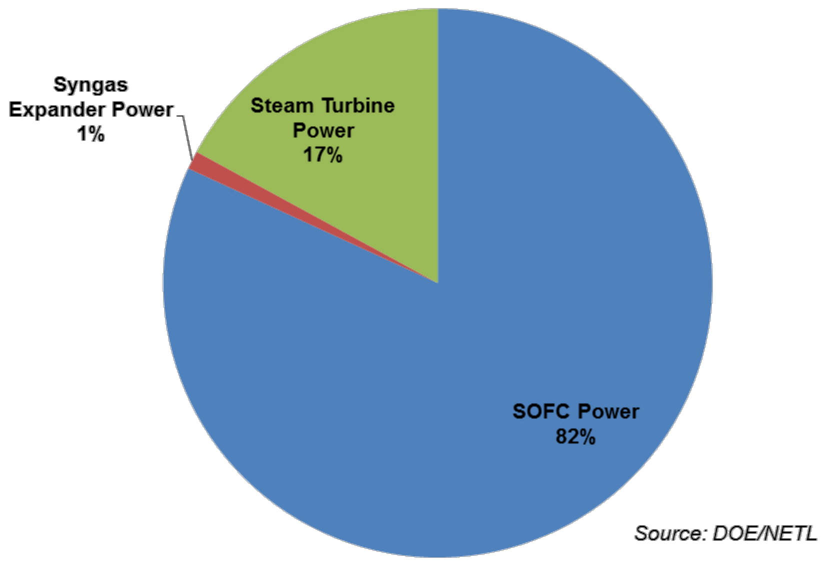


Exhibit 3-34 Components of Case 2-1 auxiliary load

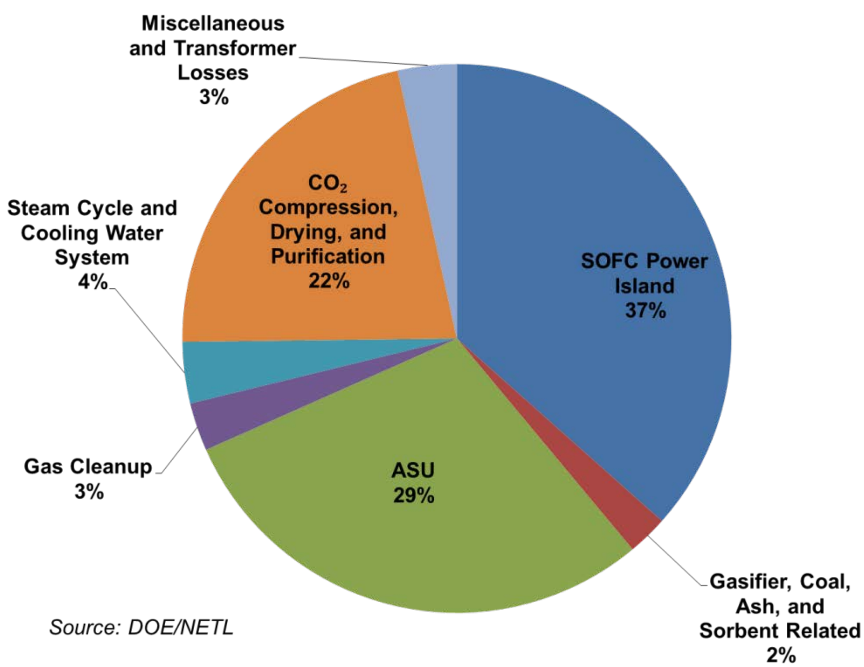
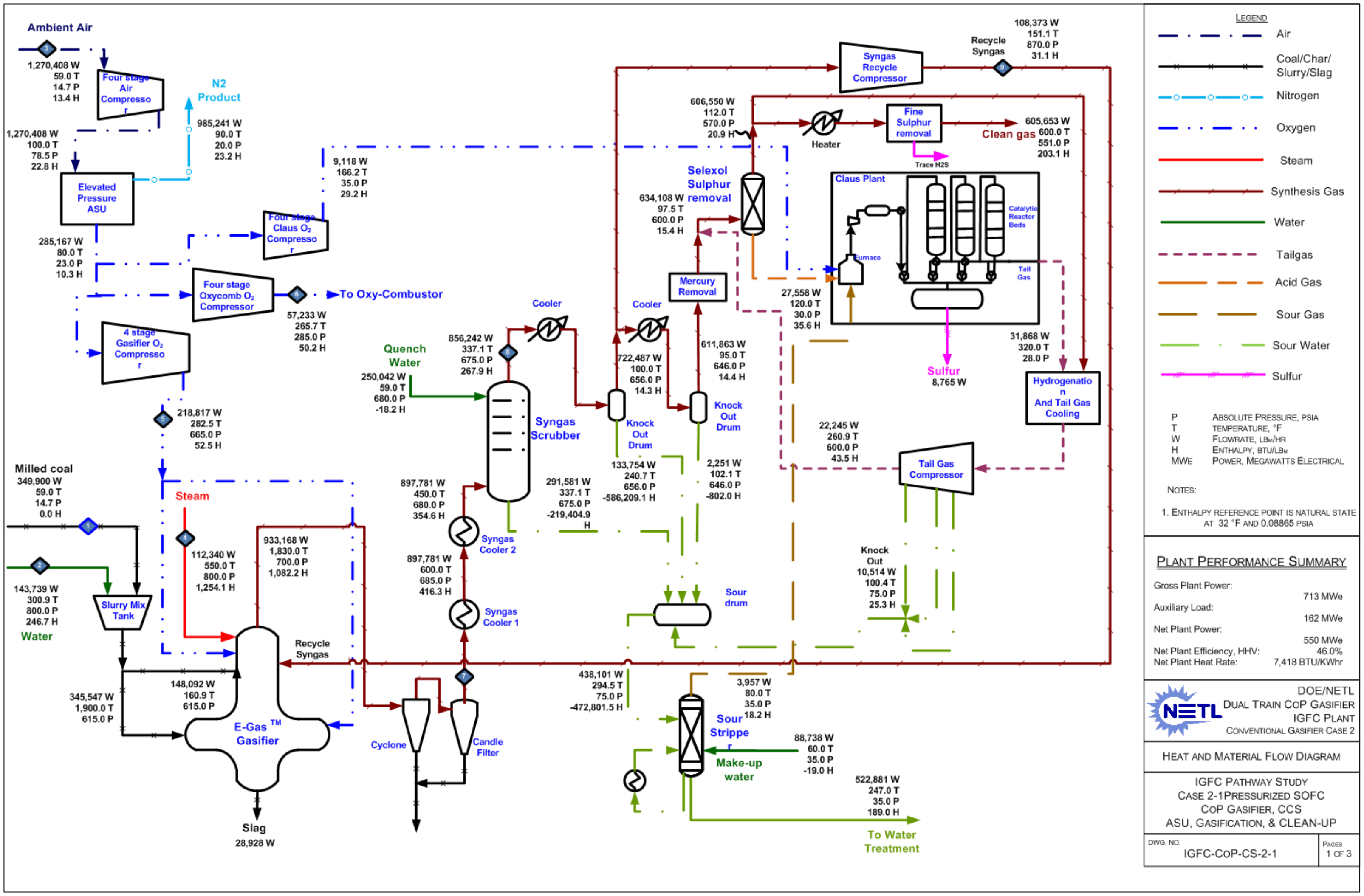
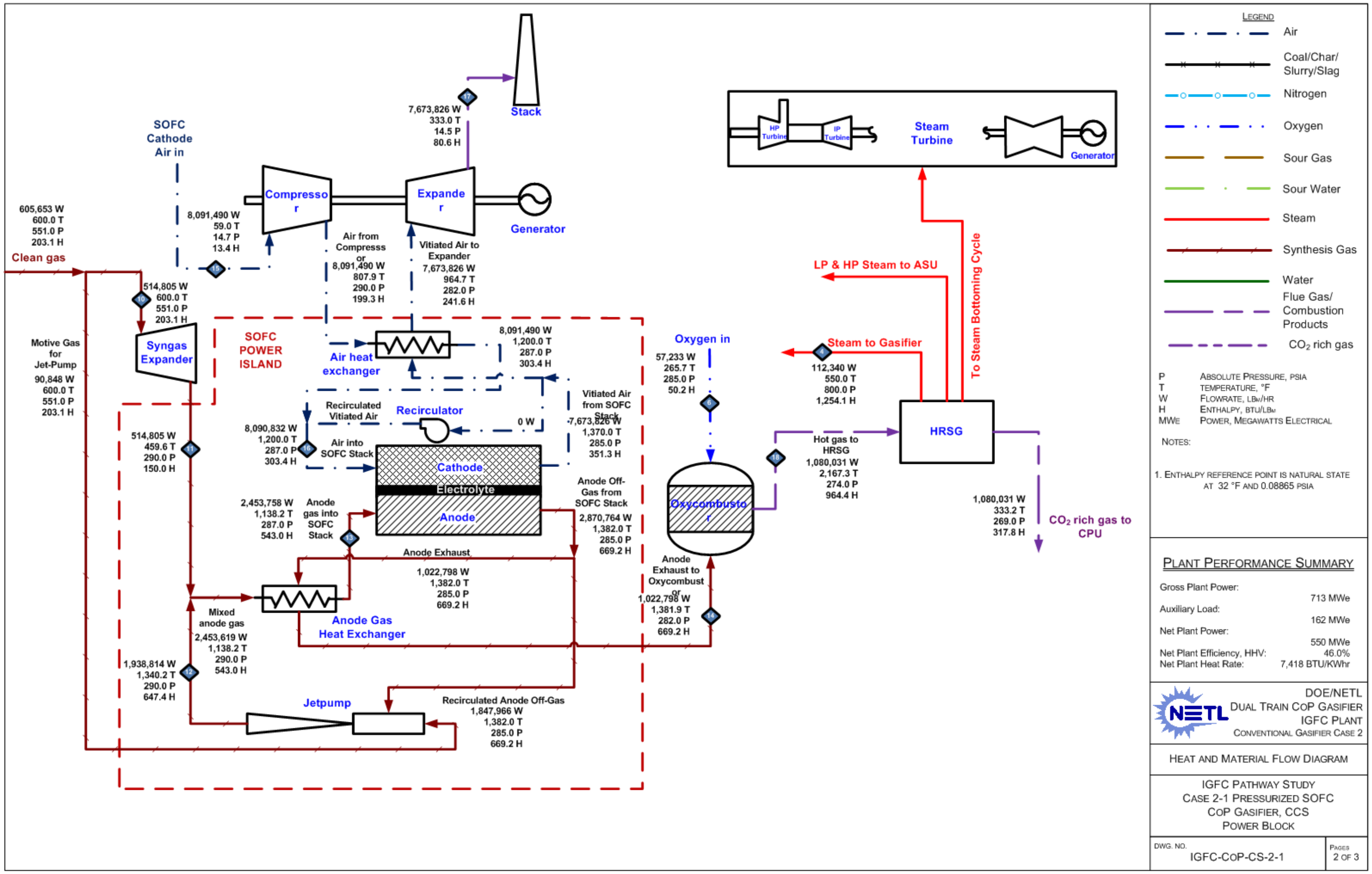


Exhibit 3-35 Case 2-1 process flow diagram – gasifier, ASU, and syngas clean-up



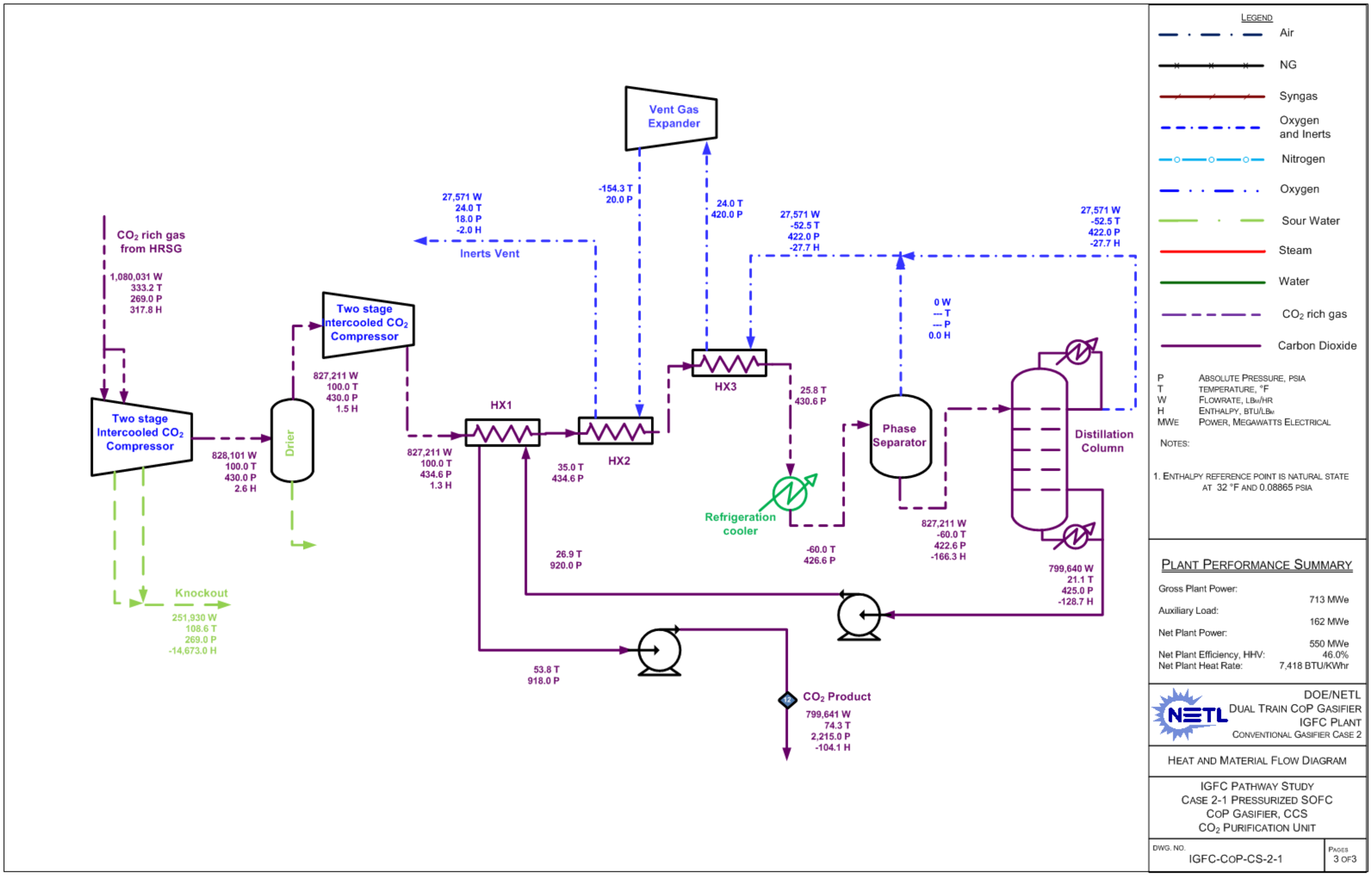
Source: DOE/NETL

Exhibit 3-36 Case 2-1 process flow diagram – IGFC power island



Source: DOE/NETL

Exhibit 3-37 Case 2-1 process flow diagram – CO_2 separation and purification



Source: DOE/NETL

Exhibit 3-38 Case 2-1 mass and energy balances

Carbon balance

Carbon In		Carbon Out	
kg/hr(lb/hr)		kg/hr(lb/hr)	
Coal	101,171 (223,043)	Slag	809 (1,784)
Air (CO2)	530 (1,169)	Stack Gas	458 (1,010)
		CO_2 Product	98,989 (218,234)
		N2 Product	72 (159)
		Vent Out	1,365 (3,009)
		CO_2 Dehydration	20 (44)
		Convergence Tolerance	-13 (-28)
Total	101,701 (224,212)	Total	101,701 (224,212)

Sulfur balance

Sulfur In		Sulfur Out	
kg/hr(lb/hr)		kg/hr(lb/hr)	
Coal	3,978 (8,770)	Elemental Sulfur	3,976 (8,765)
		Polishing Sorbent	2 (5)
		Convergence Tolerance	0 (0)
Total	3,978 (8,770)	Total	3,978 (8,770)

Energy balance

	HHV	Sensible + Latent	Power	Total
Heat In GJ/hr (MMBtu/hr)				
Coal	4,307 (4,082)	3.6 (3.4)		4,310 (4,085)
ASU Air		17.9 (17.0)		18 (17)
Fuel cell Air		114.0 (108.0)		114 (108)
Raw Water Makeup		78.0 (73.9)		78 (74)
Auxiliary Power			584 (553)	584 (553)
TOTAL	**4,307 (4,082)**	**213.5 (202.3)**	**584 (553)**	**5,104 (4,838)**
Heat Out GJ/hr (MMBtu/hr)				
Slag	21 (20)	13.2 (12.5)		35 (33)
Sulfur	37 (35)	0.4 (0.4)		37 (35)
CO_2		-87.8 (-83.2)		-88 (-83)
CO_2 Refrigeration		129.2 (122.4)		129 (122)
Cooling Tower Blowdown		15.2 (14.4)		15 (14)
HRSG Flue Gas		652.7 (618.6)		653 (619)
Condenser		645 (611)		645 (611)
Non-Condenser Cooling Tower Loads		515 (488)		515 (488)
Process losses*		642 (608)		642 (608)
Difference**		-43 (-41)		-43 (-41)
Power			2,565 (2,431)	2,565 (2,431)
TOTAL	**58 (55)**	**2,481 (2,351)**	**2,565 (2,431)**	**5,104 (4,838)**

Water balance

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m3/min (gpm)	m3/min (gpm)	m3/min (gpm)	m3/min (gpm)	m3/min (gpm)
Slag Handling	0.28 (75)	0.28 (75)			
Slurry Water	1.09 (287)	1.09 (287)			
Quench/Wash	1.9 (500)	1.89 (500)			
Condenser Makeup	**0.9 (240)**	**0.0 (0)**	**0.9 (240)**	**0.0 (0)**	**0.9 (240)**
Gasifier Steam	0.9 (225)	0.0 (0)	0.9 (225)		
BFW Makeup	0.1 (15)	0.0 (0)	0.1 (15)		
Cooling Tower	**9.1 (2,399)**	**2.60 (687)**	**6.5 (1,712)**	**2.0 (539)**	**4.4 (1,172)**
CO2 Dehydration	0.0 (0)	1.91 (504)	-1.91 (-504)		
SWS Excess Water	0.0 (0)	0.69 (183)	-0.69 (-183)		
Total	**13.3 (3,501)**	**5.87 (1550)**	**7.4 (1,951)**	**2.0 (539)**	**5.3 (1,412)**

Emissions

	kg/GJ ($lb/10^6Btu$)	Tonne/year (tons/year)	kg/MWh (lb/MWh)
SO2	0 (0)	0 (0)	0 (0)
NOx	0 (0)	0 (0)	0 (0)
Particulate	0 (0)	0 (0)	0 (0)
Hg	0 (0)	0 (0)	0 (0)
CO2	2 (4)	50,291 (55,436)	9 (21)

Process losses include losses from steam turbines, expanders, inverter, and blowers.

***Value needed to match heat input to the plant and includes minor process losses*

3.3.2 Case 2-1 IGFC Plant Cost Results

The SOFC power island capital costs for Case 2-1 are shown in Exhibit 3-39, and a distribution of cost amongst its major components are shown in Exhibit 3-40. The SOFC module costs account for a larger portion, ~77 percent, of the SOFC power island costs relative to the atmospheric case primarily due to the pressure vessel costs. The cathode air compressor and the cathode expander constitute the next significant expense. The cathode heat exchanger is not as large as in the atmospheric cases and does not influence the SOFC power island costs significantly in this case.

Exhibit 3-39 Case 2-1 SOFC power island capital costs

Cost Component	Cost ($1000)	Specific Cost ($/kWe AC)
	2011$	
SOFC Module		
SOFC Stack	123,814	225
Enclosure	132,068	240
Transport and Placement	7,924	14
Site Foundations	24,433	44
Inverter	37,528	68
Total SOFC Module	**325,768**	**592**
Total SOFC Module with 10% Extra Installed Area	**358,345**	**651**
SYNGAS EXPANDER	**2,702**	**5**
SOFC BOP		
Cathode Air Compressor	53,981	98
Cathode Gas Expander	23,476	43
Cathode Heat Exchanger	13,712	25
Anode Syngas Jet Pump	458	1
Anode Heat Exchanger	44	0.1
Oxy-Combustor	11,278	20
Total SOFC BOP	**102,948**	**187**
TOTAL SOFC POWER ISLAND	**463,994**	**843**

Case 1-1, Exhibit 3-41, which graphically depicts the IGFC capital costs listed in Exhibit 3-42, indicates that the SOFC power island capital costs, the gasifier, and the ASU costs form a major portion, ~58 percent, of the total IGFC plant capital costs. The TOC, calculated as in Exhibit 3-43 results in COEs of ~$103/MWh and ~ $110/MWh with and without CO_2 T&S charges, respectively as shown in Exhibit 3-44, which includes the O&M costs tabulated in Exhibit 3-45. Unlike Case 1-1, the capital charges have the highest value amongst the Case 2-1 COE components due to the assumption of lower (0.2 percent per 1000 h) stack degradation rate in the latter. However, the pressurized Case 2-1 does not seem to have any particular advantage while it costs slightly higher than Case 1-3, which features an atmospheric SOFC with a similar performance but with less operational complexities.

Exhibit 3-40 Distribution of Case 2-1 SOFC power island capital costs

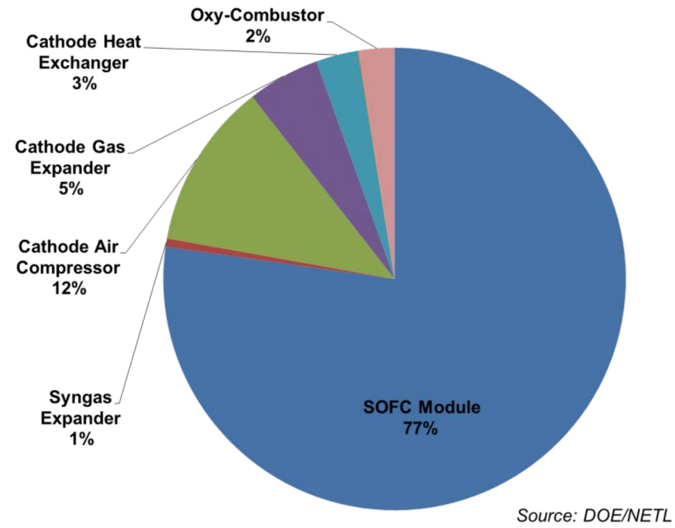


Exhibit 3-41 Case 2-1 IGFC plant cost distribution

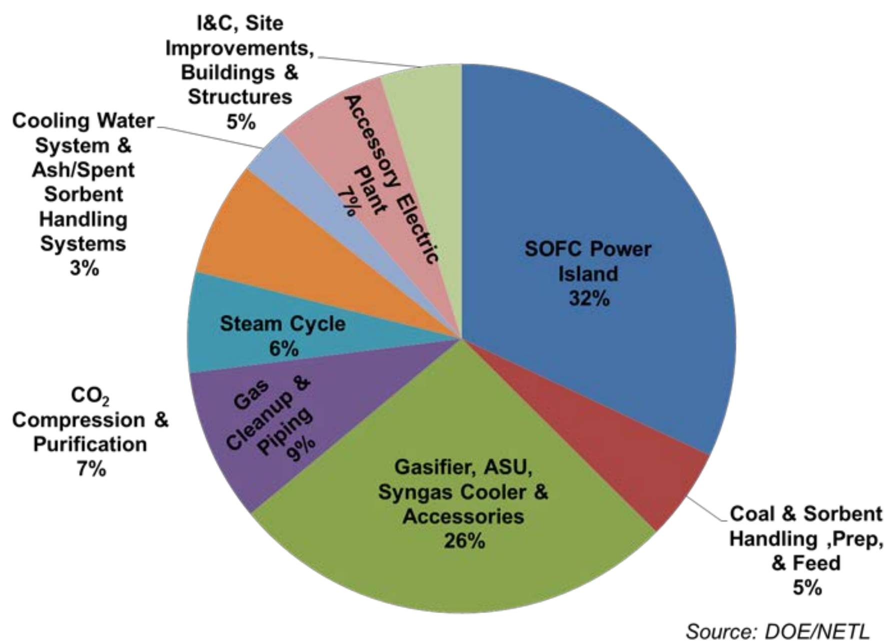


Exhibit 3-42 Case 2-1 IGFC plant capital costs

Cost Component	Cost ($1000)	Specific Cost ($/kWe AC)
	2011$	
SOFC POWER ISLAND	**463,994**	**843**
COAL & SORBENT HANDLING	**34,962**	**64**
COAL & SORBENT PREP & FEED	**44,074**	**80**
GASIFIER, ASU, SYNGAS COOLER & ACCESSORIES		
Gasifier & Syngas Cooler	215,314	391
ASU & oxidant compressor	154,722	281
Other gasification equip & foundations	14,949	27
TOTAL GASIFIER, ASU, SYNGAS COOLER & ACCESSORIES	**384,985**	**700**
GAS CLEANUP & PIPING		
Scrubber & Low Temperature Cooling	24,146	44
Single-Stage Selexol/MDEA	62,958	114
Claus Plant	26,126	47
Trace removal	1,823	3
COS Hydrolysis	7,535	14
Blowback, Piping, Foundations	3,695	7
Sulfur polishing/NG desulfurization	4,631	8
TOTAL GAS CLEANUP & PIPING	**130,913**	**238**
STEAM CYCLE		
HRSG, Ducting, and Stack	25,361	46
Steam Power System	42,968	78
Feedwater and Misc BOP systems	17,793	32
TOTAL STEAM CYCLE	**86,122**	**157**
CO_2 COMPRESSION & PURIFICATION		
CO_2 Drying and Compression	-	-
CO_2 Purification	97,447	177
TOTAL CO_2 COMPRESSION & PURIFICATION	**97,447**	**177**
COOLING WATER SYSTEM	**21,404**	**39**
ASH/SPENT SORBENT HANDLING SYSTEM	**21,590**	**39**
ACCESSORY ELECTRIC PLANT	**95,431**	**173**
INSTRUMENTATION & CONTROL	**33,097**	**60**
IMPROVEMENTS TO SITE	**18,731**	**34**
BUILDING & STRUCTURES	**17,729**	**32**
TOTAL PLANT COST (TPC)	**1,450,479**	**2636**

Exhibit 3-43 Case 2-1 TOC and TASC

Cost Component	Cost ($1000)	Specific Cost ($/kWe AC)
	2011$	2011$
OWNER'S COSTS		
Preproduction Costs		
6 Months All Labor	9,942	
1 Month Maintenance Materials	1,985	
1 Month Non-fuel Consumables	338	
1 Month Waste Disposal	309	
25% of 1 Months Fuel Cost at 100% CF	2,190	
2% of TPC	29,010	
Total Preproduction Costs	**43,774**	
Inventory Capital		
60 day supply of fuel and consumables at 100% CF	18,022	
0.5% of TPC (spare parts)	7,252	
Total Inventory Capital	**25,275**	
Initial Cost for Catalyst and Chemicals	**10,171**	
Land	**900**	
Other Owner's Costs	**217,572**	
Financing Costs	**39,163**	
TOTAL OWNER'S COSTS	**336,854**	
TOTAL OVERNIGHT COST (TOC)	**1,787,333**	**3248**
TASC Multiplier	1.14	
TOTAL AS-SPENT COST (TASC)	**2,037,560**	**3703**

Exhibit 3-44 Case 2-1 cost of electricity

COST OF ELECTRICITY (2011$/MWh) with CPU	
	$/MWh
Variable COE	36.5
Fuel	21.8
Variable O&M	14.7
Fixed O&M	11.9
Capital Charges	54.2
First Year COE (excluding T&S)	102.7
CO_2 T&S	7.2
First Year COE (with T&S)	109.9

Exhibit 3-45 Case 2-1 O&M costs

Case 2-1 Atm-SOFC CoP Gasifier						
Net Power: 550.3 Mwe, Capacity Factor: 85%, Heat Rate: 7,418 Btu/kWh						
Cost Component					**Cost ($)**	**Cost ($/MWh)**
					2011$	
OPERATING & MAINTENANCE LABOR						
Operating Labor Rate (base)					39.70	
Operating Labor Burden					30%	
Labor O-H Charge Rate					25%	
Total Operators per shift					12	
Maintenance labor/Operating labor					2.302	
Maintenance materials/Maintenance labor					1.932	
Stack replacement O&M, $/hr per stack kW					18.33	
Annual Operating Labor Cost					5,425,243	
Maintenance Labor Cost					10,481,570	
Administrative & Support Labor					3,976,703	
Property Taxes and Insurance					29,035,225	
TOTAL FIXED OPERATING COSTS					**48,918,742**	**11.94**
VARIABLE OPERATING COSTS						
Maintenance Material Cost					**20,250,393**	**4.94**
Stack replacement						
SOFC stack life (years)					7.294	
Discount rate for stack replacement (%)					10.0%	
SOFC stack replacement cost, $/kW AC					$524	
SOFC Stack replacement O&M, $/yr per kW					$52	
Stack Replacement Cost					**33,455,590**	**8.17**
CONSUMABLES	**Consumption**		**Cost ($)**			
	Initial Fill	**/Day**	**/Unit**	**Initial Fill**		
Water (/1000 gallons)		1,727	1.670	0	**894,721**	
Chemicals						
MU & WT Chem. (lb)		7,631	0.27	0	639,257	
Carbon (Trace Removal) (lb)	417,826	572	1.6	681,057	289,449	
COS Catalyst (m^3)	316	0.22	3,752	1,186,352	252,100	
Selexol Solution (gal)	213,863	34	36.79	7,868,018	384,721	
Claus / DSRP Catalyst (ft^3)		1.46	203	0	91,706	
ZnO polishing sorbent (lb)	241,979	1,598	1.8	435,562	892,584	
Sub Total Chemicals				**10,170,989**	**2,549,816**	
Waste Disposal						
Spent Trace Catalyst (lb)		618	0.65		124,658	
Ash + HCl Sorbent (ton)		347	25.11		2,704,312	
Spent sorbents (lb)		1,598	0.65		322,322	
Subtotal Waste Disposal					**3,151,293**	
TOTAL VARIABLE OPERATING COSTS				**10,170,989**	**60,301,813**	**14.72**
Fuel Coal (ton)		68.60			**89,363,690**	**21.81**

3.3.3 Scenario 2 Pathway Results

The Scenario 2 pathway estimated performance and cost for various SOFC system advances in a cumulative manner:

Case 2-2: The capacity factor of Case 2-1 was further increased to 90 percent.

Case 2-3: The stack cost was reduced from $225/kW in Case 2-2 to $200/kW.

Case 2-4: An improvement in the inverter efficiency from the 97 percent in Case 2-3 to 98 percent was analyzed in this case.

The performances and costs of the Scenario 1 pathway cases are summarized in Exhibit 3-46 and Exhibit 3-47, respectively. The 1 absolute percent improvement in the inverter efficiency in Case 2-4 results in a 0.5 percentage point increase in plant HHV efficiency over Case 2-1. The decrease observed in the COE from Case 2-1 to Case 2-4 is mainly attributable to the assumptions of increased capacity factor and lower stack cost. The combined effects of stack cost and stack degradation rate on the COE and the cost of captured CO_2 are presented in Exhibit 3-48 and Exhibit 3-49. These Exhibits further emphasize the need to achieve a stack degradation rate below 0.2 percent per 1000 h for the IGFC system to be competitive with conventional technologies.

Exhibit 3-46 Comparison of performance of Scenario 2 pathway cases

CASE	2-1	2-2	2-3	2-4
Dry Syngas CH_4 Content (%)	**10.8**			
SOFC Overpotential (mV)	**70**			
Inverter Efficiency (%)	**97**			**98**
POWER SUMMARY (Gross Power at Generator Terminals, kW_e)				
SOFC Power	583,200			582,600
Syngas Expander Power	7,700			7,700
Steam Turbine Power	121,600			120,200
TOTAL GROSS POWER (kW_e)	**712,500**			**710,500**
AUXILIARY LOAD SUMMARY (kW_e)				
Coal Handling	400			400
Coal Milling	1,630			1,610
Sour Water Recycle Slurry Pump	145			143
Ash Handling	680			670
Air Separation Unit Auxiliaries	733			725
Air Separation Unit Main Air Compressor	33,890			33,510
Oxygen Compressor	13,070			12,910
Claus Plant TG Recycle Compressor	1,400			1,380
CO_2 Compression	5,180			5,120
CO_2 Purification	30,057			29,719
Boiler Feedwater Pumps	2,110			2,086
Condensate Pump	157			155
Syngas Recycle Compressor	590			580
Circulating Water Pump	2,330			2,310
Ground Water Pumps	376			372
Cooling Tower Fans	1,210			1,190
Scrubber Pumps	265			262
Quench Water Pump	358			354
Selexol Auxiliary Power	2,447			2,419
Steam Turbine Auxiliaries	51			50
Claus Plant/TGTU Auxiliaries	181			179
Gas Turbine Auxiliaries	128			126
Cathode Air Compressor - Cathode Expander	59,200			58,490
Miscellaneous Balance of Plant[2]	3,038			3,029
Transformer Losses	2,590			2,580
TOTAL AUXILIARIES (kW_e)	**162,215**			**160,369**
NET POWER (kW_e)	**550,285**			**550,131**
NET PLANT EFFICIENCY, % (HHV)	**46.0**			**46.5**
NET PLANT HEAT RATE, kJ/kWh (Btu/kWh)	**7,826 (7,418)**			**7,741 (7,337)**
CO_2 Capture Rate (%)	**98.6**			**98.6**
CONDENSER COOLING DUTY 10^6 kJ/h (10^6 Btu/h)	**644 (610)**			**633 (600)**
As-Received Coal Feed, kg/h (lb/h)	158,712 (349,900)			156,943 (346,000)
Thermal Input[1], kWt	1,196,297			1,182,963
Raw Water Consumption, m^3/min (gpm)	5.3 (1,412)			5.3 (1,396)

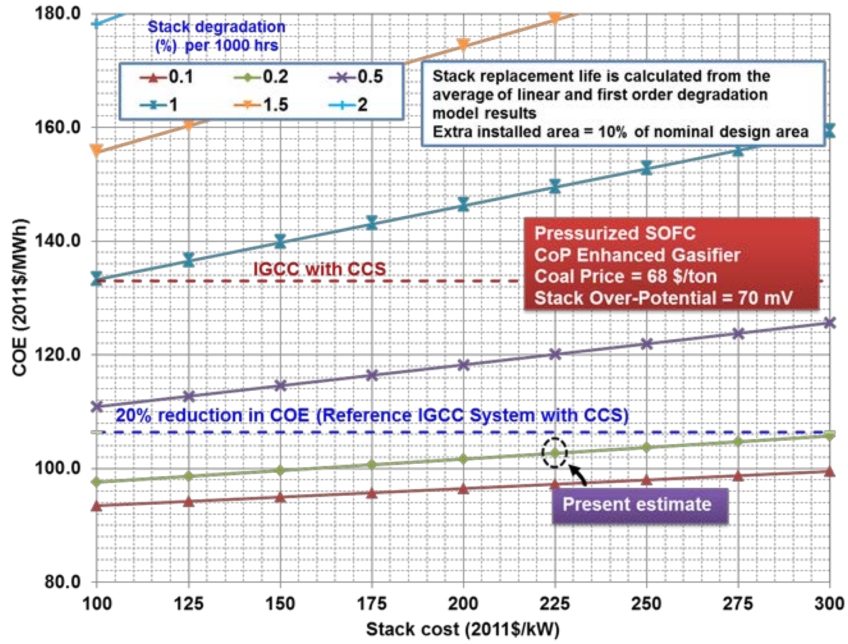
[1] HHV of as received Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb)

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

Exhibit 3-47 Comparison of Scenario 2 pathway cases costs

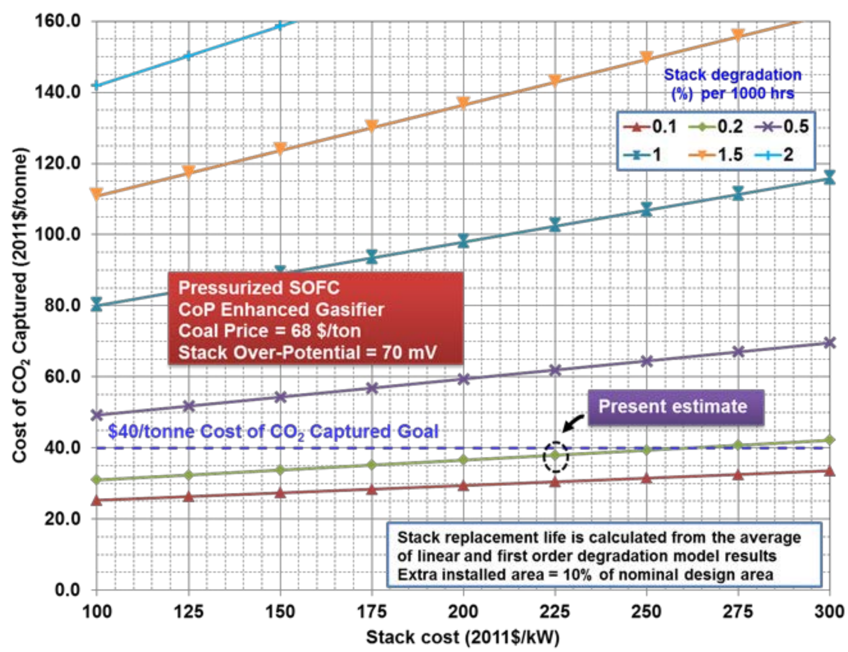
CASE	2-1	2-2	2-3	2-4
SOFC Degradation Rate (%/1000 h)	0.2			
SOFC Overpotential (mV)	70			
Dry Syngas CH_4 Content (%)	10.8			
Capacity Factor (%)	85	90		
SOFC Stack Cost ($/kW)	225		200	
Inverter Efficiency (%)	97			98
PERFORMANCE				
Gross Power (MWe)	712.5	712.5	712.5	710.5
Auxiliary Loads (MWe)	162.2	162.2	162.2	160.4
Net Power (MWe)	550.3	550.3	550.3	550.1
Net Electric Efficiency, HHV (%)	46.0	46.0	46.0	46.5
CO_2 Capture rate (%)	98.6	98.6	98.6	98.6
CO_2 Emissions (lb/MWhgross)	20.9	20.9	20.9	20.7
CO_2 Emissions (lb/MWhnet)	27.1	27.1	27.1	26.8
Raw Water Consumption (gpm/MWnet)	2.6	2.6	2.6	2.5
COST				
Total Plant Cost (TPC) (1000$)	1,450,479	1,450,479	1,435,346	1,426,677
Total Overnight Cost (TOC) (1000$)	1,787,333	1,787,228	1,769,038	1,758,272
Total As-Spent Cost (TASC) (1000$)	2,037,560	2,037,440	2,016,704	2,004,430
Cost of Electricity ($/MWh)				
Variable COE	36.5	35.8	35.4	35.1
Fuel	21.8	21.8	21.8	21.6
Variable O&M	14.7	14.0	13.6	13.5
Fixed O&M	11.9	11.3	11.2	11.2
Capital Charges	54.2	51.2	50.7	50.4
Total First Year COE (excluding T&S)	102.7	98.3	97.3	96.7
T&S	7.2	7.2	7.2	7.2
Total First Year COE (including T&S)	109.9	105.5	104.6	103.8
NETL Metrics				
% COE reduction (COE_{IGFC} - $COE_{IGCC\ with\ CCS}$)/ $COE_{IGCC\ with\ CCS}$	22.8	26.1	26.8	27.3
Cost of Captured CO_2 ($/tonne CO_2) (COE_{IGFC} - $COE_{AUSC\ PC}$)/ CO_2 Captured (tonnes/MWh)	41.9	35.2	33.7	33.1

Exhibit 3-48 Sensitivity of Case 2-1 COE to stack cost and degradation rate



Source: DOE/NETL

Exhibit 3-49 Sensitivity of Case 2-1 cost of CO_2 captured to stack cost and degradation rate



Source: DOE/NETL

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. However, some modifications to the equipment and operating conditions used in the Gas Cleaning Area are made to suit the catalytic gasifier-based IGFC plant. The steam cycle was not modeled directly as in the CoP gasifier cases. Instead, the steam cycle performance was computed using the net heat available in the system and applying an efficiency of 38.1 percent.

4.1.1 Catalytic Gasifier Area

The partial-combustion of the coal and the loss of carbon with the ash constitute significant coal energy losses in the gasification of coal into the syngas. Catalytic coal gasification promotes efficient gasification of coal at a relatively low temperature where oxygen consumption is minimized, carbon conversion remains acceptably high, resulting in a gasifier cold gas efficiency, which 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 developmental stages. It is assumed that the catalytic gasifier can be successfully developed for operation at the selected conditions, and with the performance estimated in this evaluation. While there is currently no ongoing development effort for this type of coal gasifier, an objective of this analysis is to assess the benefits of the catalytic gasification technology to justify future investments.

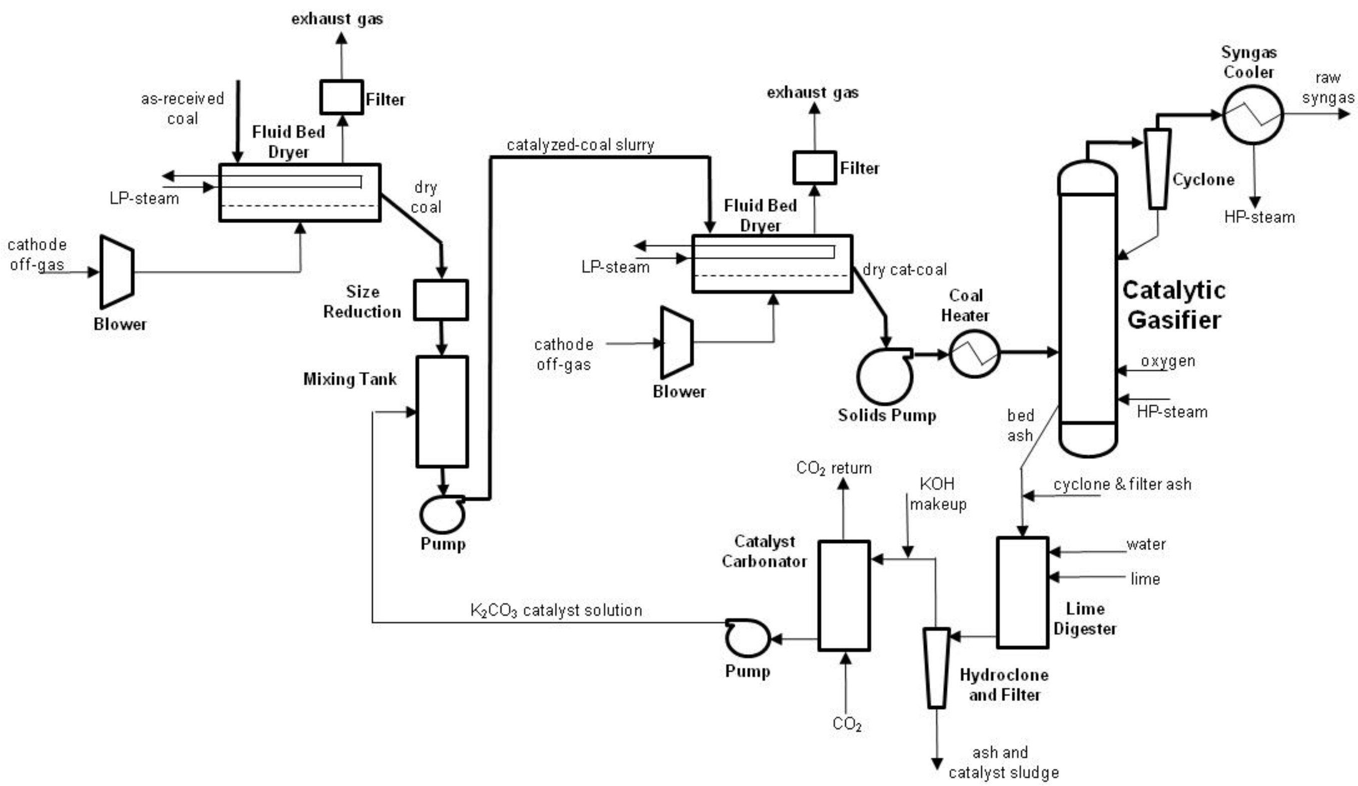
While a number of gasifier catalysts have been tested in laboratory studies, it has been found that the catalyst applied by Exxon (K_2CO_3 with KOH makeup) in their prior development program is very effective, but relatively expensive compared to other, less effective catalysts. (22) The catalyst material, 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. (23) 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 with warm cathode off-gas from the power island for fluidization, and with low-pressure steam for additional in-bed heating. (24) The dried coal is reduced in size and 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 weight percent K_2CO_3)
- The coal-catalyst solution is dried in fluid bed dryer at 54°C (130°F) using cathode off-gas and LP-steam heat source
- The process results in a coal catalyst loading of 15 weight percent K_2CO_3 (dry coal)

Catalyst Recovery Factors:

- First step is $Ca(OH)_2$ digestion plus water washing, operated at149°C (300°F) 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. 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)

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). Clean syngas is reheated to 371°C (700°F) for sulfur polishing by a gas-gas recuperative heat exchange, which cools the gasifier syngas to 232°C (450°F).

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 371 ºC (700 ºF)
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 371 ºC (700 ºF) 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, as in the CoP gasifier cases, uses conventional dry gas cleaning technology based on single-stage Selexol acid gas removal.

The Scenario 3 baseline atmospheric-pressure Power Block assumptions and specifications are listed again 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 BFD for the plant is shown in Exhibit 4-5. The process data corresponding to the numbered streams are tabulated in Exhibit 4-6, which shows that the syngas methane content of ~31.8 mole percent reduces to 10.4 mole percent in the SOFC inlet fuel stream because of the 56 percent anode off-gas recirculation. The performance summary listed in Exhibit 4-7 shows that the catalytic gasifier baseline IGFC plant results in a net plant HHV efficiency of ~49.1 percent, which is 16 percentage points higher than the efficiency achievable with current conventional IGCC technology with carbon capture; it is also significantly higher, ~7 percentage points, than conventional NGCC efficiencies with carbon capture. (1) The carbon capture rate for the process is 98.6 percent as in the other cases. Exhibit 4-8 shows that the SOFC power island contribution dominates the power generation with only 6 percent of power generated in the steam cycle. The SOFC operating voltage is 0.796 V, and is lower than in Case 1-1 due to the dilution of the inlet anode gas 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 38 percent of the coal feed energy input.

The ASU and the CPU parasitics make up ~70 percent of the auxiliary load as shown in Exhibit 4-9. The SOFC power island losses[8] comprise ~15 percent of the parasitic loads. The PFDs for the IGFC plant components are shown in Exhibit 4-10 , Exhibit 4-11, and Exhibit 4-12. Salient material and energy balances are shown in Exhibit 4-13. 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.

The nearly complete recovery of water from the oxy-combustion CO_2 product stream results in an IGFC plant water consumption, also shown in Exhibit 4-13, which is significantly lower than the corresponding value for conventional fossil fuel power plant technologies.

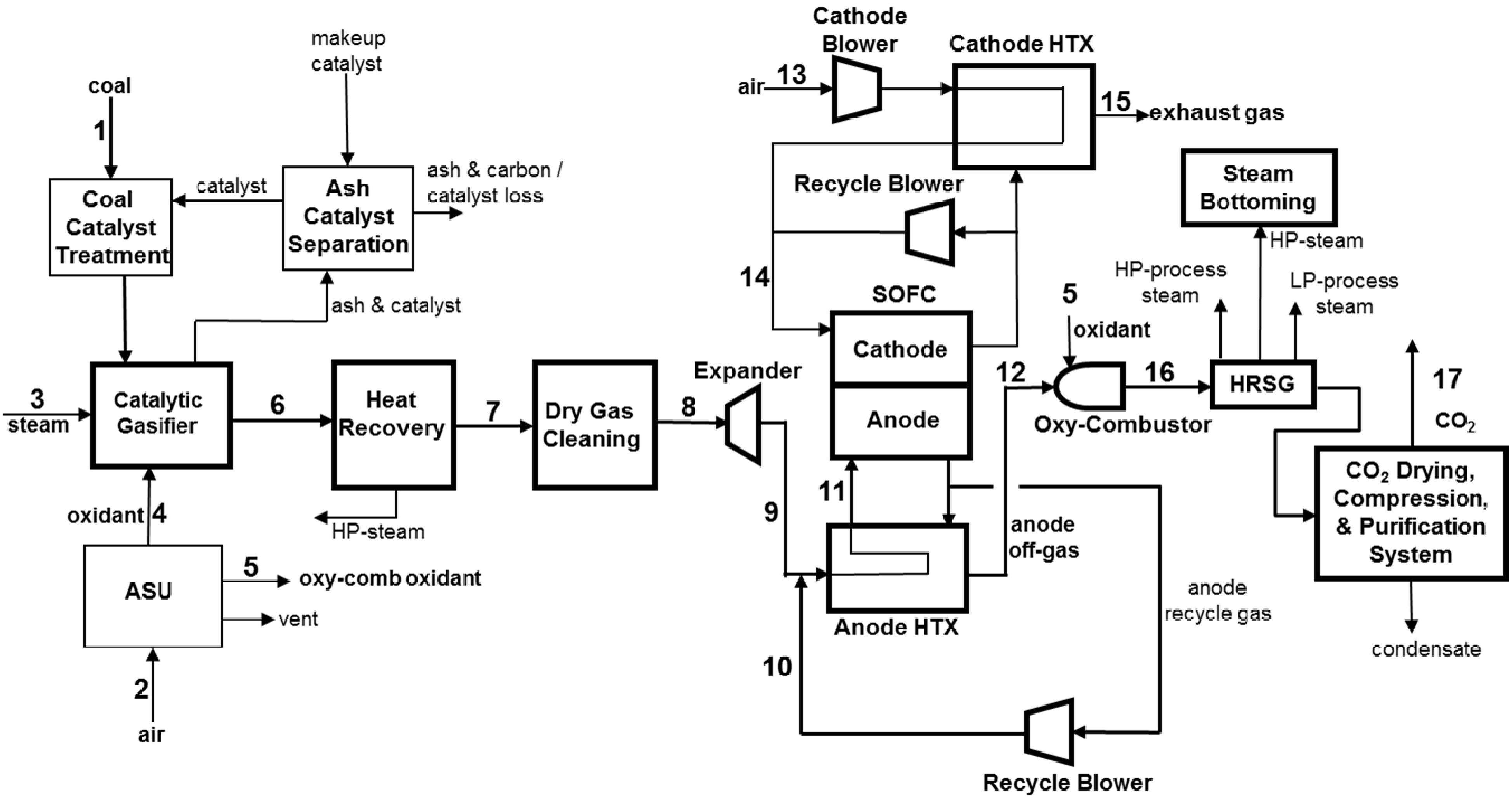
The IGFC plant acts as a nearly zero emission power plant, with the only significant emission being the small release of CO_2 shown in the emissions listed in Exhibit 4-13. This emissions performance is dictated, in part, by the need to protect the SOFC stack components from contamination.

[8] The DC-AC inverter losses and the SOFC polarization losses are included in the SOFC gross power estimates.

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 (psi)	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

Exhibit 4-5 Case 3-1 block flow diagram



Source: DOE/NETL

Exhibit 4-6 Case 3-1 stream table

	1	2	3	4	5	6	7	8	9	10
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
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.1950	0.1950	0.3159	0.3159	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0563	0.0563	0.0914	0.0914	0.0480
CO_2	0.0000	0.0003	0.0000	0.0000	0.0000	0.2128	0.2128	0.3465	0.3465	0.4140
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0001	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.1472	0.1472	0.2386	0.2386	0.0497
H_2O	0.0000	0.0104	1.0000	0.0000	0.0000	0.3760	0.3760	0.0004	0.0004	0.4840
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0008	0.0000	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0071	0.0071	0.0000	0.0000	0.0000
N_2	0.0000	0.7722	0.0000	0.0019	0.0019	0.0039	0.0039	0.0069	0.0069	0.0042
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0007	0.0007	0.0000	0.0000	0.0000
O_2	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
Total	0.0000	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	9,301	11,209	834	965	16,219	16,219	9,988	9,988	20,454
V-L Flowrate (kg/hr)	0	268,377	201,942	26,706	30,900	349,443	349,443	235,429	235,429	583,097
Solids Flowrate (kg/hr)	149,340	0	0	0	0	0	0	0	0	0
Temperature (°C)	15	15	291	130	27	704	204	371	54	759
Pressure (MPa, abs)	0.10	0.10	7.58	3.45	0.16	6.72	6.62	5.61	0.14	0.11
Enthalpy (kJ/kg)[A]	---	31.06	2,860.25	114.27	23.91	2,107.45	906.26	606.83	77.49	1,848.02
Density (kg/m^3)	---	1.2	38.4	33.1	2.0	17.7	42.2	24.4	1.2	0.4
V-L Molecular Weight	---	28.854	18.015	32.016	32.016	21.546	21.546	23.570	23.570	28.507
V-L Flowrate (lb_{mol}/hr)	0	20,505	24,713	1,839	2,128	35,756	35,756	22,021	22,021	45,094
V-L Flowrate (lb/hr)	0	591,669	445,206	58,876	68,122	770,390	770,390	519,032	519,032	1,285,508
Solids Flowrate (lb/hr)	329,238	0	0	0	0	0	0	0	0	0
Temperature (°F)	59	59	555	265	80	1,300	400	700	129	1,398
Pressure (psia)	14.7	14.7	1,100.0	500.0	23.0	975.0	960.0	814.0	20.0	16.2
Enthalpy (Btu/lb)[A]	---	13.4	1,229.7	49.1	10.3	906.0	389.6	260.9	33.3	794.5
Density (lb/ft^3)	---	0.076	2.398	2.067	0.127	1.104	2.635	1.523	0.075	0.023

A - Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 4-6 Case 3-1 stream table (continued)

	11	12	13	14	15	16	17
V-L Mole Fraction							
Ar	0.0002	0.0002	0.0094	0.0098	0.0102	0.0004	0.0000
CH_4	0.1036	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0622	0.0480	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.3918	0.4140	0.0003	0.0003	0.0003	0.4572	1.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.1117	0.0497	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.3253	0.4840	0.0104	0.0108	0.0112	0.5282	0.0000
HCl	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
N_2	0.0051	0.0042	0.7722	0.8021	0.8345	0.0043	0.0000
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.2077	0.1770	0.1438	0.0100	0.0000
SO_2	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
V-L Flowrate (kg_{mol}/hr)	30,443	16,299	95,994	184,822	88,830	16,468	7,417
V-L Flowrate (kg/hr)	818,524	464,641	2,769,837	5,310,378	2,540,582	495,541	326,416
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0
Temperature (°C)	649	536	15	650	122	1,051	24
Pressure (MPa, abs)	0.11	0.11	0.10	0.1	0.101	0.102	15.272
Enthalpy (kJ/kg)[A]	1,529.04	1,498.71	31.06	708.6	141.883	2,321.270	-242.014
Density (kg/m^3)	0.4	0.5	1.2	0.4	0.9	0.3	740.0
V-L Molecular Weight	26.887	28.507	28.854	29	28.601	30.091	44.010
V-L Flowrate (lb_{mol}/hr)	67,115	35,933	211,631	407,464	195,836	36,306	16,351
V-L Flowrate (lb/hr)	1,804,537	1,024,359	6,106,446	11,707,379	5,601,025	1,092,480	719,624
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0
Temperature (°F)	1,200	996	59	1,202	251	1,924	74
Pressure (psia)	16.2	15.4	14.7	15.8	14.7	14.8	2,215.0
Enthalpy (Btu/lb)[A]	657.4	644.3	13.4	304.6	61.0	998.0	-104.0
Density (lb/ft^3)	0.024	0.028	0.076	0	0.055	0.017	46.198

A - Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 4-7 Case 3-1 plant performance summary (100 percent load)

POWER SUMMARY (Gross Power at Generator Terminals, kW_e)	
SOFC Power	593,200
Syngas Expander Power	33,600
Steam Turbine Power	42,100
TOTAL GROSS POWER (kW_e)	**668,900**
AUXILIARY LOAD SUMMARY (kW_e)	
Coal Handling	390
Coal size reduction	480
Catalyst-coal processing	1,866
Catalyst coal feeding	1,027
Sour Water Recycle Slurry Pump	136
Ash Handling	1,090
Air Separation Unit Auxiliaries	341
Air Separation Unit Main Air Compressor	15,790
Oxygen Compressor	2,710
Claus Plant TG Recycle Compressor	1,260
CO_2 Compressor	37,640
CO_2 Purification	27,063
Boiler Feedwater Pumps	731
Condensate Pump	54
Circulating Water Pump	1,800
Ground Water Pumps	354
Cooling Tower Fans	940
Scrubber Pumps	227
Selexol Auxiliary Power	2,099
Steam Turbine Auxiliaries	18
Claus Plant/TGTU Auxiliaries	170
Gas Turbine Auxiliaries	14
Cathode Air Blower	6,640
Cathode Recycle Blower	7,890
Anode Recycle Blower	2,710
Miscellaneous Balance of Plant[2]	2,852
Transformer Losses	2,340
TOTAL AUXILIARIES (kW_e)	**118,632**
NET POWER (kW_e)	**550,268**
NET PLANT EFFICIENCY, % (HHV)	**48.9**
NET PLANT HEAT RATE, kJ/kWh (Btu/kWh)	**7,364 (6,980)**
CO_2 Capture Rate (%)	**98.5**
CONDENSER COOLING DUTY 10^6 kJ/h (10^6 Btu/h)	**243 (230**
As-Received Coal Feed, kg/h (lb/h)	149,340 (329,238)
Thermal Input[1], kWt	1,125,654
Raw Water Consumption, m^3/min (gpm)	4.9 (1,304)

[1] *HHV of as received Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb)*

[2] *Includes plant control systems, lighting, HVAC, and miscellaneous low-voltage loads*

Exhibit 4-8 Case 3-1 gross power generation components

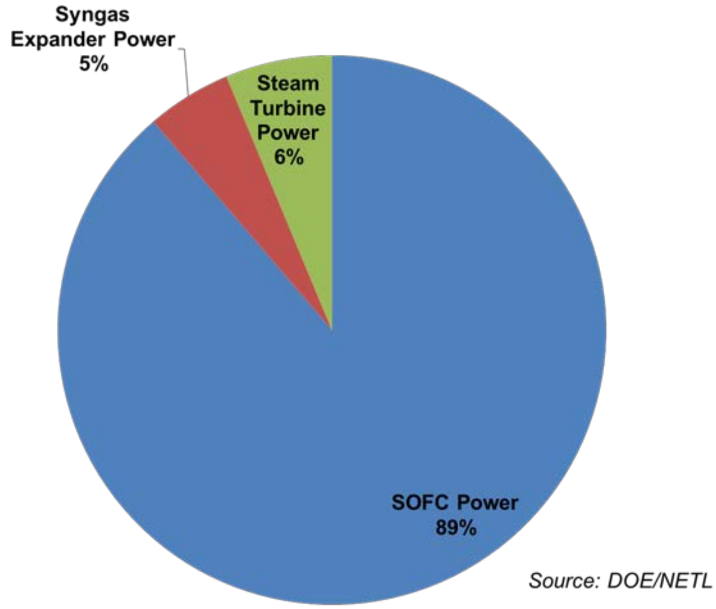


Exhibit 4-9 Components of Case 3-1 auxiliary load

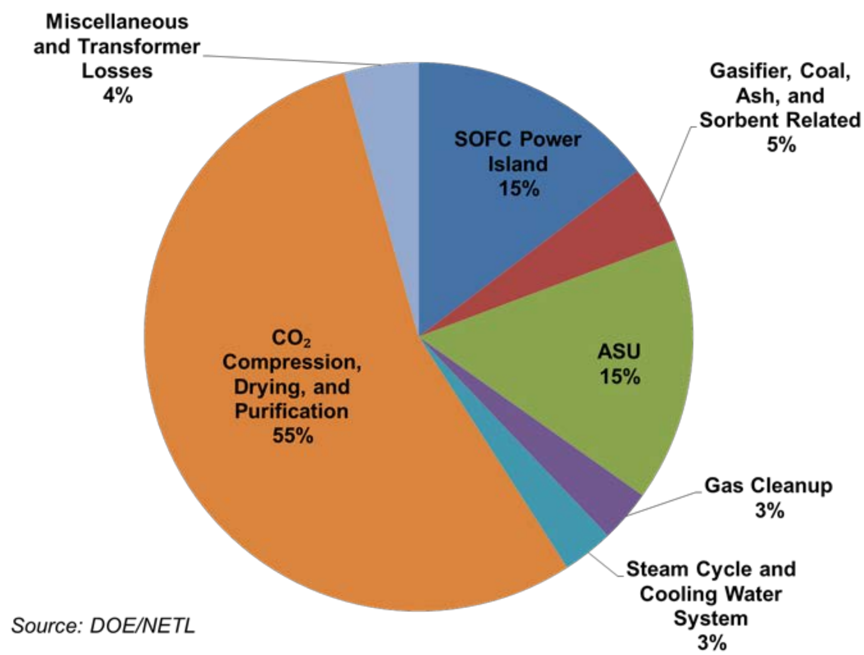
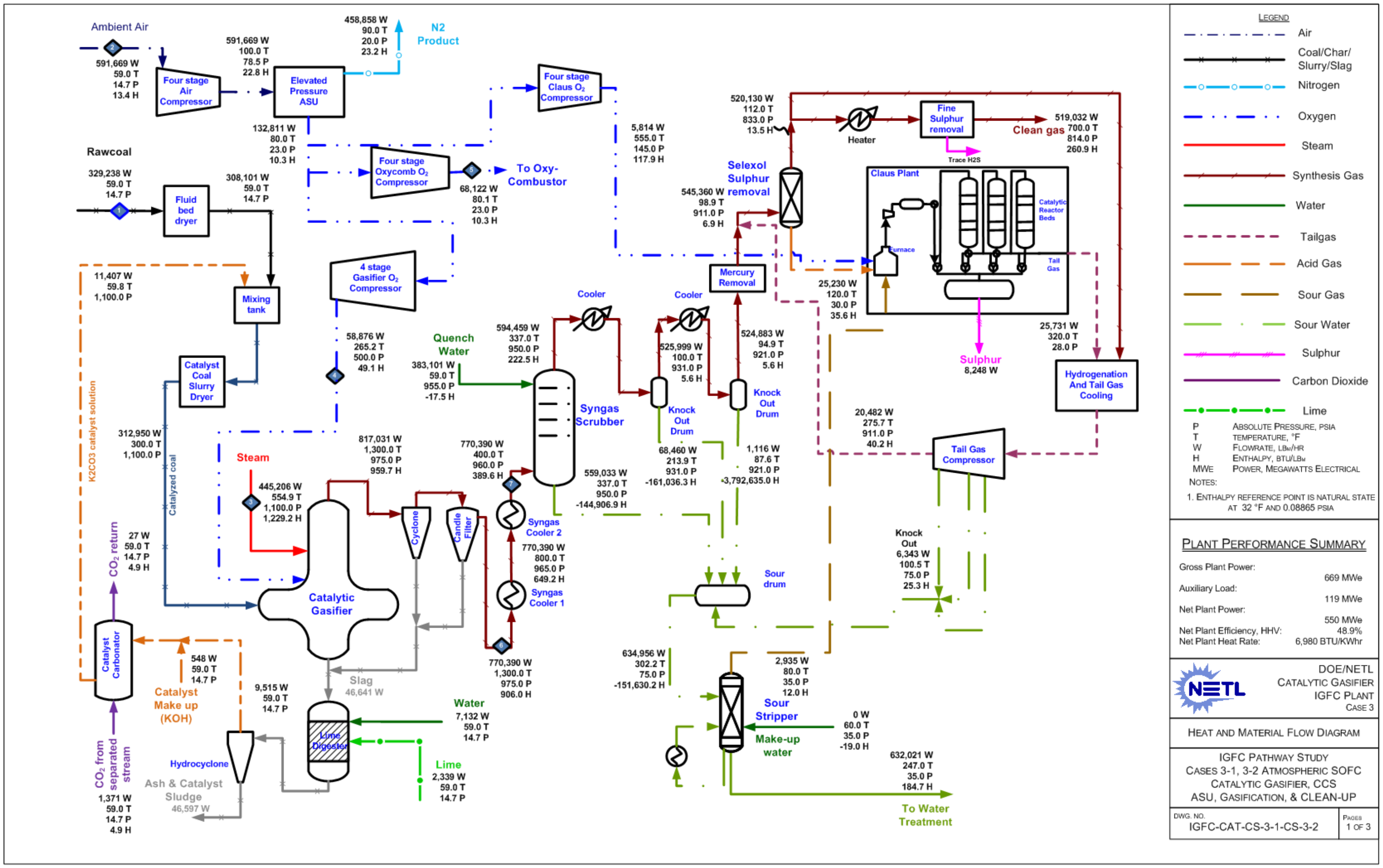
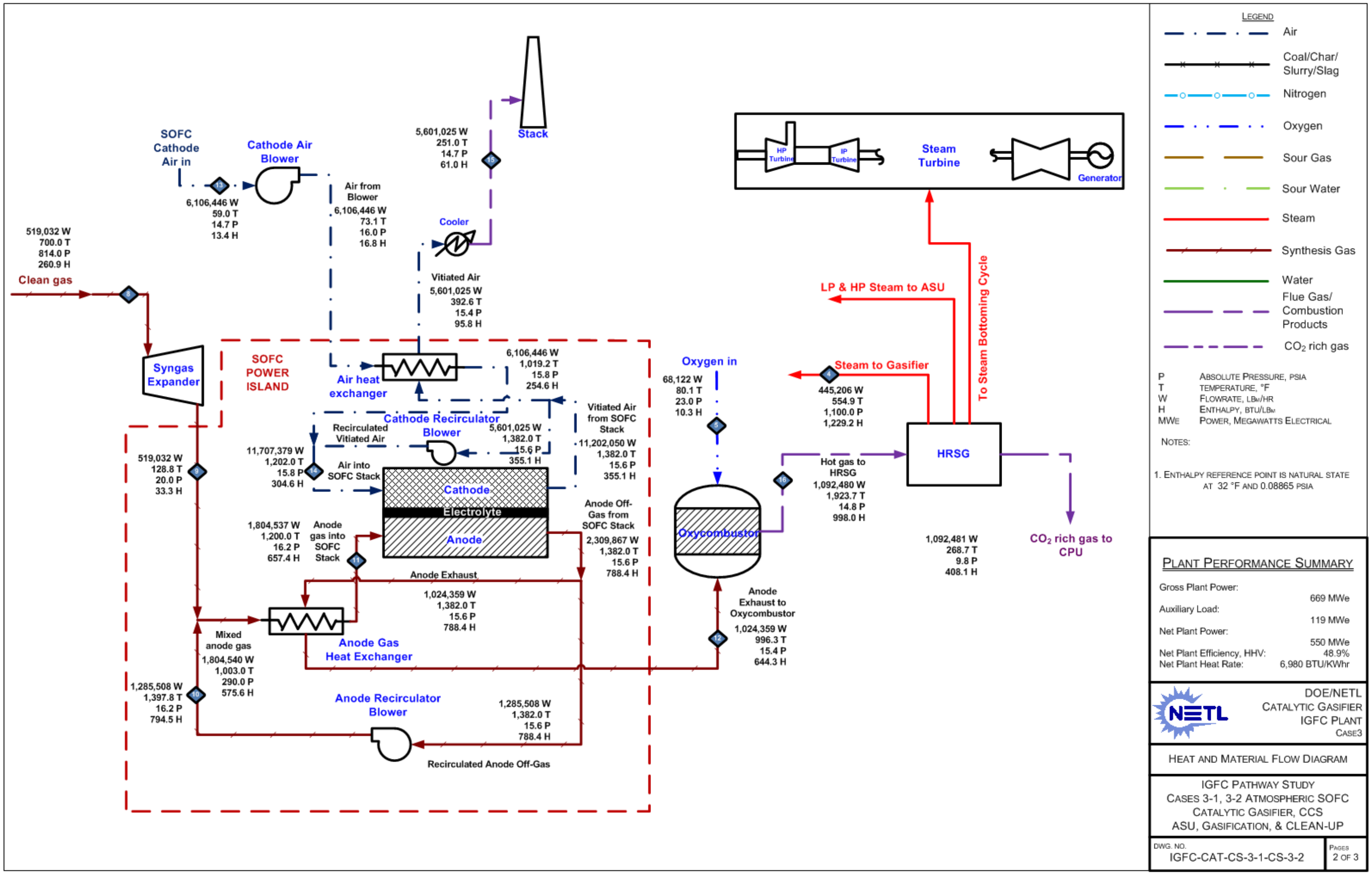


Exhibit 4-10 Case 3-1 process flow diagram – gasifier, ASU, and syngas clean-up



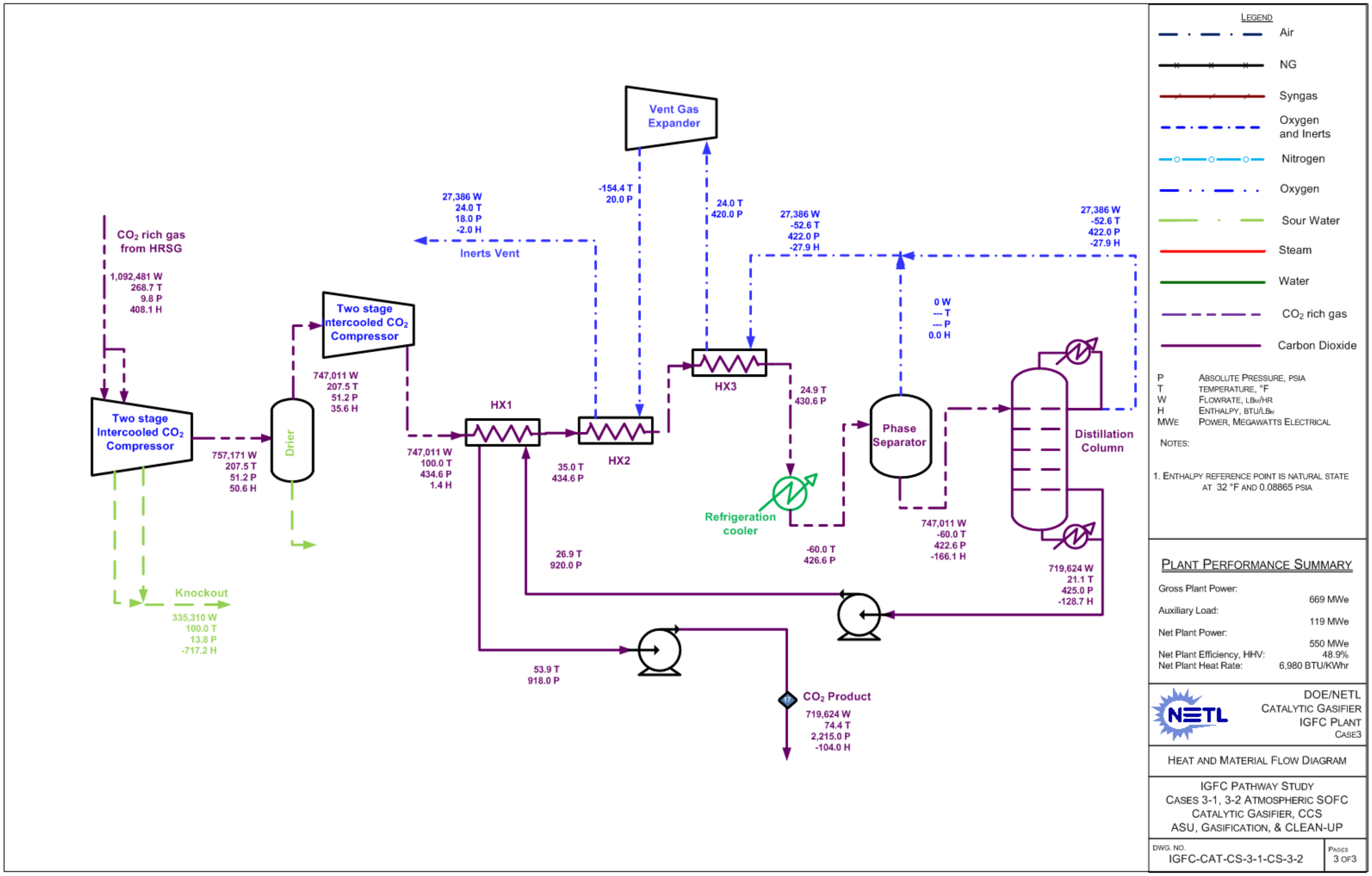
Source: DOE/NETL

Exhibit 4-11 Case 3-1 process flow diagram – IGFC power island



Source: DOE/NETL

Exhibit 4-12 Case 3-1 process flow diagram – CO_2 separation and purification



Source: DOE/NETL

Exhibit 4-13 Case 3-1 mass and energy balances

Carbon balance

Carbon In		Carbon Out	
kg/hr(lb/hr)		kg/hr(lb/hr)	
Coal	95,196 (209,872)	Slag	4,931 (10,872)
Air (CO2)	379 (836)	Stack Gas	346 (763)
Cat Carbonator	170 (374)	CO_2 Product	89,084 (196,396)
		N_2 Product	34 (74)
		Carbonator Vent	3 (7)
		Vent Gas	1,353 (2,984)
		Convergence Tolerance	-6 (-12)
Total	95,745 (211,082)	Total	95,745 (211,082)

Sulfur balance

Sulfur In		Sulfur Out	
kg/hr(lb/hr)		kg/hr(lb/hr)	
Coal	3,743 (8,252)	Elemental Sulfur	3,741 (8,248)
		Polishing Sorbent	2 (4)
		Convergence Tolerance	0 (0)
Total	3,743 (8,252)	Total	3,743 (8,252)

Energy balance

	HHV	Sensible + Latent	Power	Total
Heat In GJ/hr (MMBtu/hr)				
Coal	4,052 (3,841)	3.4 (3.2)		4,056 (3,844)
ASU Air		8.3 (7.9)		8 (8)
Fuel cell Air		86.0 (81.5)		86 (82)
Raw Water Makeup		24.5 (23.2)		24 (23)
Auxiliary Power			427 (405)	427 (405)
TOTAL	**4,052 (3,841)**	**122.2 (115.8)**	**427 (405)**	**4,602 (4,362)**
Heat Out GJ/hr (MMBtu/hr)				
Slag	162 (153)	15.7 (14.9)		177 (168)
Sulfur	35 (33)	0.4 (0.4)		35 (33)
CO_2		-79.0 (-74.9)		-79 (-75)
CO_2 Refrigeration		116.3 (110.2)		116 (110)
Cooling Tower Blowdown		11.7 (11.1)		12 (11)
HRSG Flue Gas		360.5 (341.7)		360 (342)
Condenser		246 (233)		246 (233)
Non-Condenser Cooling Tower Loads		637 (604)		637 (604)
Process losses*		665 (630)		665 (630)
Difference**		24 (23)		24 (23)
Power			2,408 (2,282)	2,408 (2,282)
TOTAL	**196 (186)**	**1,997 (1,893)**	**2,408 (2,282)**	**4,602 (4,362)**

Water balance

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m3/min (gpm)	m3/min (gpm)	m3/min (gpm)	m3/min (gpm)	m3/min (gpm)
Slag Handling	0.46 (121)	0.46 (121)			
Slurry Water	0.05 (14)	0.05 (14)			
Quench/Wash	2.9 (766)	2.90 (766)			
Condenser Makeup	**3.4 (907)**	**0.0 (0)**	**3.4 (907)**	**0.0 (0)**	**3.4 (907)**
Gasifier Steam	3.4 (890)	0.0 (0)	3.4 (890)		
BFW Makeup	0.1 (16)	0.0 (0)	0.1 (16)		
Cooling Tower	**7.0 (1,846)**	**3.91 (1033)**	**3.1 (813)**	**1.6 (415)**	**1.5 (398)**
CO2 Dehydration	0.0 (0)	2.54 (671)	-2.54 (-671)		
SWS Excess Water	0.0 (0)	1.37 (363)	-1.37 (-363)		
Total	**13.8 (3,654)**	**7.32 (1935)**	**6.5 (1,720)**	**1.6 (415)**	**4.9 (1,304)**

Emissions

	kg/GJ ($lb/10^6 Btu$)	Tonne/year (tons/year)	kg/MWh (lb/MWh)
SO2	0 (0)	0 (0)	0 (0)
NOx	0 (0)	0 (0)	0 (0)
Particulate	0 (0)	0 (0)	0 (0)
Hg	0 (0)	0 (0)	0 (0)
CO2	2 (4)	43,670 (48,138)	9 (21)

*

Process losses include losses from steam turbines, expanders, inverter, and blowers.

***Value needed to match heat input to the plant and includes minor process losses*

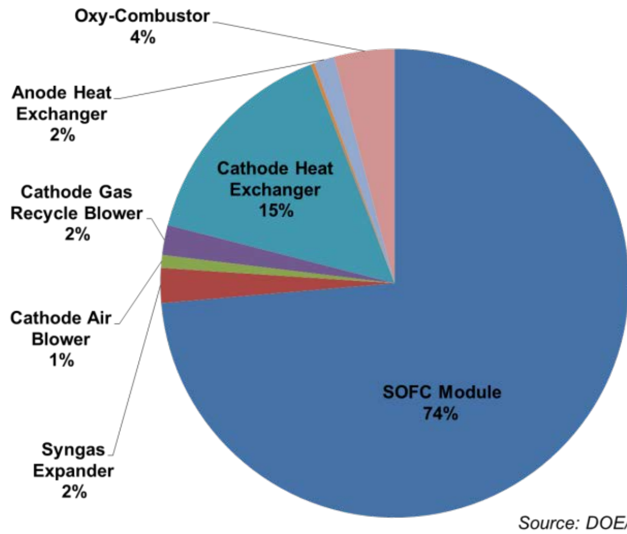
4.2.2 Case 3-1 Plant Cost Results

The SOFC power island capital costs are shown in Exhibit 4-14. The SOFC module costs account for a major portion, ~74 percent, of the SOFC power island costs as shown by the categorized cost distribution in Exhibit 4-15. The cooling due to internal (to the SOFC stack) reformation of the CH_4 in the syngas results in a decrease in cathode heat exchanger size (due to reduced airflow) and cost relative to Case 1-1. But it is still a significant expense contributing to ~15 percent of the SOFC power island costs.

Exhibit 4-14 Case 3-1 SOFC power island capital costs

Cost Component	Cost ($1000)	Specific Cost ($/kWe AC)
	2011$	
SOFC Module		
SOFC Stack	123,810	225
Enclosure	16,508	30
Transport and Placement	7,924	14
Site Foundations	24,432	44
Inverter	37,527	68
Total SOFC Module	**210,201**	**382**
Total SOFC Module with 10% Extra Installed Area	**231,221**	**420**
SYNGAS EXPANDER	**7,577**	**14**
SOFC BOP		
Cathode Air Blower	2,783	5
Cathode Gas Recycle Blower	6,470	12
Cathode Heat Exchanger	47,580	86
Anode Recycle Blower	790	1
Anode Heat Exchanger	4,428	8
Oxy-Combustor	13,116	24
Total SOFC BOP	**75,167**	**137**
TOTAL SOFC POWER ISLAND	**313,966**	**571**

Exhibit 4-15 Distribution of Case 3-1 SOFC power island capital costs

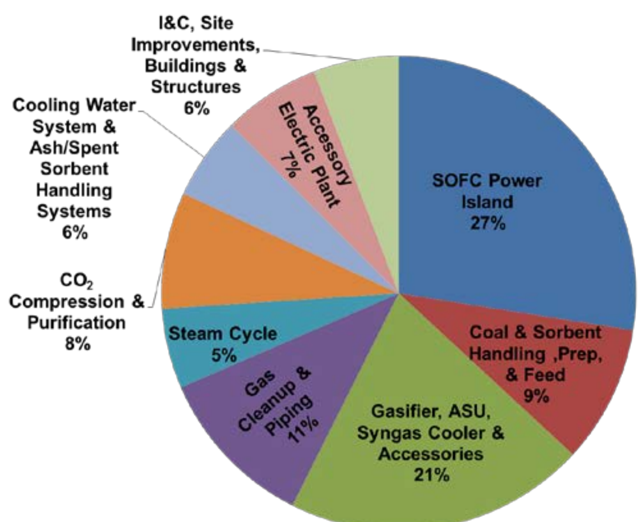


Source: DOE/NETL

Exhibit 4-16, which graphically depicts the IGFC capital costs listed in Exhibit 4-17, indicates that the SOFC power island capital costs, the gasifier, and the ASU costs form ~ 48 percent of the total IGFC plant capital costs.

The TOC in Exhibit 4-18, results in COEs of ~\$142/MWh and ~\$148/MWh with and without CO_2 T&S charges, respectively, as shown in Exhibit 4-19, which incorporates the O&M costs tabulated in Exhibit 4-20. These costs are high despite the high electrical efficiency due to the dominant O&M costs that, as in Case 1-1, are a consequence of frequent stack replacement expenses associated with the stack degradation rate of 1.5 percent per 1000 h assumed in this case. The importance of stack degradation rate becomes apparent from the results of the Scenario 3 pathway cases, which are discussed next.

Exhibit 4-16 Case 3-1 IGFC plant cost distribution



Source: DOE/NETL

Exhibit 4-17 Case 3-1 IGFC plant capital costs

Cost Component	Cost ($1000)	Specific Cost ($/kWe AC)
	2011$	
SOFC POWER ISLAND	**313,966**	**571**
COAL & SORBENT HANDLING	**31,983**	**58**
COAL & SORBENT PREP & FEED	**75,740**	**138**
GASIFIER, ASU, SYNGAS COOLER & ACCESSORIES		
Gasifier & Syngas Cooler	129,637	236
ASU & oxidant compressor	91,798	167
Other gasification equip & foundations	14,970	27
TOTAL GASIFIER, ASU, SYNGAS COOLER & ACCESSORIES	**236,404**	**430**
GAS CLEANUP & PIPING		
Scrubber & Low Temperature Cooling	14,537	26
Single-Stage Selexol/MDEA	70,358	128
Claus Plant	24,977	45
Trace removal	1,764	3
COS Hydrolysis	7,675	14
Blowback, Piping, Foundations	4,360	8
Sulfur polishing/NG desulfurization	3,325	6
Heat Interchanger	633	1
TOTAL GAS CLEANUP & PIPING	**127,629**	**232**
STEAM CYCLE		
HRSG, Ducting, and Stack	25,104	46
Steam Power System	20,450	37
Feedwater and Misc BOP systems	16,832	31
TOTAL STEAM CYCLE	**62,386**	**113**
CO_2 COMPRESSION & PURIFICATION		
CO_2 Drying and Compression	-	-
CO_2 Purification	90,992	165
TOTAL CO_2 COMPRESSION & PURIFICATION	**90,992**	**165**
COOLING WATER SYSTEM	**17,780**	**32**
ASH/SPENT SORBENT HANDLING SYSTEM	**46,500**	**85**
ACCESSORY ELECTRIC PLANT	**76,660**	**139**
INSTRUMENTATION & CONTROL	**33,097**	**60**
IMPROVEMENTS TO SITE	**17,135**	**31**
BUILDING & STRUCTURES	**16,218**	**29**
TOTAL PLANT COST (TPC)	**1,146,491**	**2084**

Exhibit 4-18 Case 3-1 TOC and TASC

Cost Component	Cost ($1000) 2011$	Specific Cost ($/kWe AC) 2011$
OWNER'S COSTS		
Preproduction Costs		
6 Months All Labor	12,427	
1 Month Maintenance Materials	2,637	
1 Month Non-fuel Consumables	463	
1 Month Waste Disposal	494	
25% of 1 Months Fuel Cost at 100% CF	2,048	
2% of TPC	22,930	
Total Preproduction Costs	**40,998**	
Inventory Capital		
60 day supply of fuel and consumables at 100% CF	17,156	
0.5% of TPC (spare parts)	5,732	
Total Inventory Capital	**22,888**	
Initial Cost for Catalyst and Chemicals	**9,154**	
Land	**900**	
Other Owner's Costs	**171,974**	
Financing Costs	**30,955**	
TOTAL OWNER'S COSTS	**276,869**	
TOTAL OVERNIGHT COST (TOC)	**1,423,360**	**2587**
TASC Multiplier	1.14	
TOTAL AS-SPENT COST (TASC)	**1,622,631**	**2949**

Exhibit 4-19 Case 3-1 cost of electricity

COST OF ELECTRICITY (2011$/MWh) with CPU	
	$/MWh
Variable COE	**83.9**
Fuel	**20.4**
Variable O&M	**63.6**
Fixed O&M	**12.4**
Capital Charges	**45.9**
First Year COE (excluding T&S)	**142.3**
CO_2 T&S	**6.5**
First Year COE (with T&S)	**148.8**

Exhibit 4-20 Case 3-1 O&M costs

Case 3-1 Atm-SOFC Catalytic Gasifier						
Net Power: 550.2 Mwe, Capacity Factor: 80%, Heat Rate: 6,942 Btu/kWh						
Cost Component					Cost ($)	Cost ($/MWh)
					2011$	
OPERATING & MAINTENANCE LABOR						
Operating Labor Rate (base)					39.70	
Operating Labor Burden					30%	
Labor O-H Charge Rate					25%	
Total Operators per shift					15	
Maintenance labor/Operating labor					2.302	
Maintenance materials/Maintenance labor					1.932	
Stack replacement O&M, $/hr per stack kW					18.33	
Annual Operating Labor Cost					6,781,554	
Maintenance Labor Cost					13,101,962	
Administrative & Support Labor					4,970,879	
Property Taxes and Insurance					23,122,486	
TOTAL FIXED OPERATING COSTS					**47,976,881**	**12.44**
VARIABLE OPERATING COSTS						
Maintenance Material Cost					**25,312,991**	**6.56**
Stack replacement						
SOFC stack life (years)					0.973	
Discount rate for stack replacement (%)					10.0%	
SOFC stack replacement cost, $/kW AC					$314	
SOFC Stack replacement O&M, $/yr per kW					$323	
Stack Replacement Cost					**210,580,189**	**54.61**
CONSUMABLES	**Consumption**		**Cost ($)**		**Annual ($)**	
	Initial Fill	**/ Day**	**/ Unit**	**Initial Fill**		
Water (/1000 gallons)		1,501	1.670		**732,136**	
Chemicals						
MU & WT Chem. (lb)		7,050	0.27		555,787	
Carbon (Trace Removal) (lb)	367,913	504	1.6	599,698	239,879	
COS Catalyst (m^3)	278	0.19	3,752	1,044,630	208,926	
Selexol Solution (gal)	188,315	30	36.79	6,928,104	318,835	
Claus / DSRP Catalyst (ft^3)		1.28	203		76,001	
ZnO polishing sorbent (lb)	148,774	1,496	1.8	267,792	786,497	
KOH Coal Catalyst makeup (lb)	789,616	13,160	0.192	151,606	737,817	
Lime for catalyst recovery	3,368,341	56,139	0.048	161,680	786,845	
Sub Total Chemicals				**9,153,511**	**3,710,586**	
Waste Disposal						
Spent Trace Catalyst (lb)		544	0.65		103,310	
Ash + HCl Sorbent (ton)		594	25.11		4,353,879	
Spent sorbents (lb)		1496	0.65		284,013	
Subtotal Waste Disposal					4,741,201	
TOTAL VARIABLE OPERATING COSTS				**9,153,511**	**245,077,103**	**63.55**
Fuel Coal (ton)		68.60			**78,636,289**	**20.39**

4.2.3 Scenario 3 Pathway Results

The Scenario 3 pathway estimated performance and cost for various SOFC system advances in a cumulative manner:

Case 3-2: The stack degradation rate was assumed to improve from the 1.5 percent /1000 hours in the baseline Case 3-1 to 0.2 percent /1000 hours.

Case 3-3: An enhancement in cell performance was assumed in this case and the cell overpotential in Case 3-2 was reduced from 140 mV to 70 mV.

Case 3-4: Case 3-3 plant capacity factor was increased to 85 percent from 80 percent.

Case 3-5: The capacity factor of Case 3-4 was further increased to 90 percent.

Case 3-6: The stack cost was reduced from $225/kW in Case 3-5 to $200/kW.

Case 3-7: An improvement in the inverter efficiency from the 97 percent in Case 3-6 to 98 percent was analyzed in this case.

The performances and costs of the Scenario 3 pathway cases are summarized in Exhibit 4-21 and Exhibit 4-22, respectively. The net plant HHV efficiency varies from a value of 49.1 percent for the baseline case to a value of 54.1 percent attributable mainly to the enhancement in SOFC electrical performance. The dramatic effect of stack performance degradation rate is immediately visible in Exhibit 4-22 where the COE (without T&S) of Case 3-2 is lower than the baseline Case 3-1 COE (without T&S) by nearly 34 percent, attributable directly to the decrease in degradation rate from 1.5 percent per 1000 h to 0.2 percent per 1000 h. The reduction in stack degradation by itself can propel the catalytic gasifier IGFC system to exceed significantly the NETL goals of 20 percent reduction in COE over an IGCC system with carbon capture and a cost of less than $40 per tonne of captured CO_2. Case 3-7 of this pathway results in the lowest value of COE ~$79/MWh of all the cases investigated in the present study.

As in the Pathway 1 cases, the COE decreases progressively, albeit at a slower rate, stepping through the other cases, due to increases in performance and plant availability coupled with a reduction in stack cost. The combined effects of stack cost and stack degradation rate on the COE and the cost of captured CO_2 are presented in Exhibit 4-23 and Exhibit 4-24. Unlike the CoP gasifier case a stack degradation rate of 0.5 percent per 1000 h seems to be acceptable for the catalytic gasifier to be competitive with respect to IGCC systems. However, in reality, a stack degradation rate that is lower than 0.2 percent per 1000 h is desirable to compete with other technologies such as the NGCC systems.

Exhibit 4-21 Comparison of performance of Scenario 3 pathway cases

CASE	3-1 3-2	3-3 3-4 3-5 3-6	3-7
Dry Syngas CH_4 Content (%)	**31.6**		
SOFC Overpotential (mV)	**140**	**70**	
Inverter Efficiency (%)	**97**		**98**
POWER SUMMARY (Gross Power at Generator Terminals, kW_e)			
SOFC Power	593,200	586,500	586,200
Syngas Expander Power	33,600	30,700	30,400
Steam Turbine Power	42,100	37,600	37,100
TOTAL GROSS POWER (kW_e)	**668,900**	**654,800**	**653,700**
AUXILIARY LOAD SUMMARY (kW_e)			
Coal Handling	390	380	380
Coal size reduction	480	439	434
Catalyst-coal processing	1,866	1,705	1,687
Catalyst coal feeding	1,027	938	928
Sour Water Recycle Slurry Pump	136	124	123
Ash Handling	1,090	1,000	990
Air Separation Unit Auxiliaries	341	312	309
Air Separation Unit Main Air Compressor	15,790	14,420	14,260
Oxygen Compressor	2,710	2,480	2,450
Claus Plant TG Recycle Compressor	1,260	1,150	1,140
CO_2 Compressor	37,640	34,390	34,030
CO_2 Purification	27,063	24,725	24,463
Boiler Feedwater Pumps	731	653	644
Condensate Pump	54	49	48
Circulating Water Pump	1,800	1,640	1,620
Ground Water Pumps	354	323	320
Cooling Tower Fans	940	860	850
Scrubber Pumps	227	208	206
Selexol Auxiliary Power	2,099	1,917	1,897
Steam Turbine Auxiliaries	18	16	15
Claus Plant/TGTU Auxiliaries	170	155	154
Gas Turbine Auxiliaries	14	9	9
Cathode Air Blower	6,640	4,340	4,290
Cathode Recycle Blower	7,890	4,990	4,930
Anode Recycle Blower	2,710	2,470	2,450
Miscellaneous Balance of Plant[2]	2,852	2,792	2,787
Transformer Losses	2,340	2,260	2,250
TOTAL AUXILIARIES (kW_e)	**118,632**	**104,745**	**103,663**
NET POWER (kW_e)	**550,268**	**550,055**	**550,037**
NET PLANT EFFICIENCY, % (HHV)	**48.9**	**53.5**	**54.1**
NET PLANT HEAT RATE, kJ/kWh (Btu/kWh)	**7,364 (6,980)**	**6,731 (6,380)**	**6,660 (6,312)**
CO_2 Capture Rate (%)	**98.5**	**98.5**	**98.5**
CONDENSER COOLING DUTY 10^6 kJ/h (10^6 Btu/h)	**243 (230)**	**222 (210)**	**222 (210)**
As-Received Coal Feed, kg/h (lb/h)	149,340 (329,238)	136,441 (300,800)	134,994 (297,610)
Thermal Input[1], kWt	1,125,654	1,028,425	1,017,519
Raw Water Consumption, m^3/min (gpm)	4.9 (1,304)	4.5 (1,188)	4.4 (1,174)

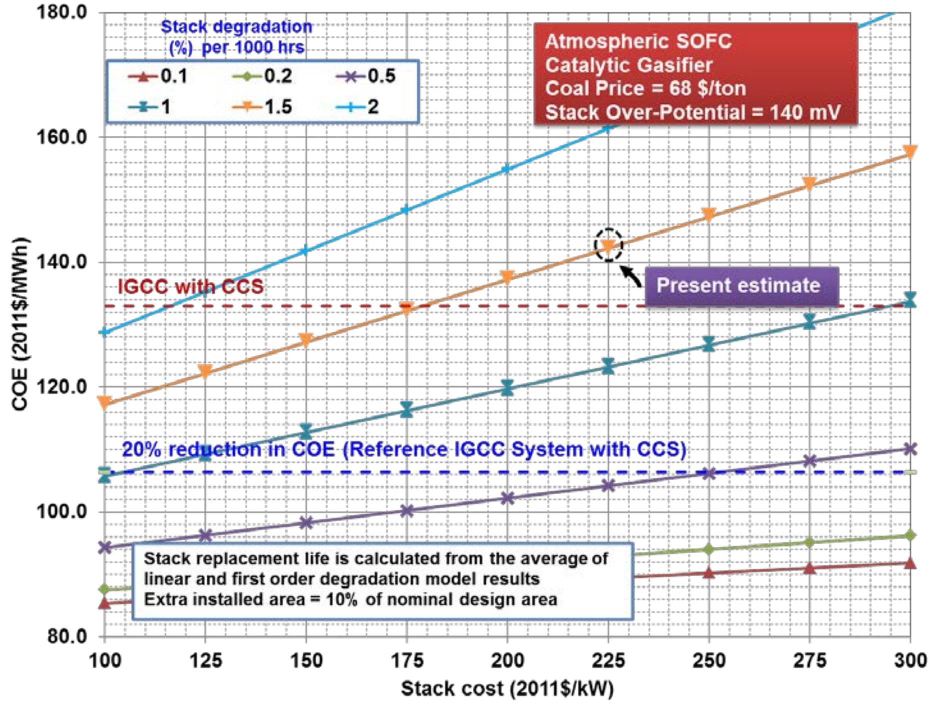
[1] HHV of as received Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb)

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

Exhibit 4-22 Comparison of Scenario 3 pathway cases costs

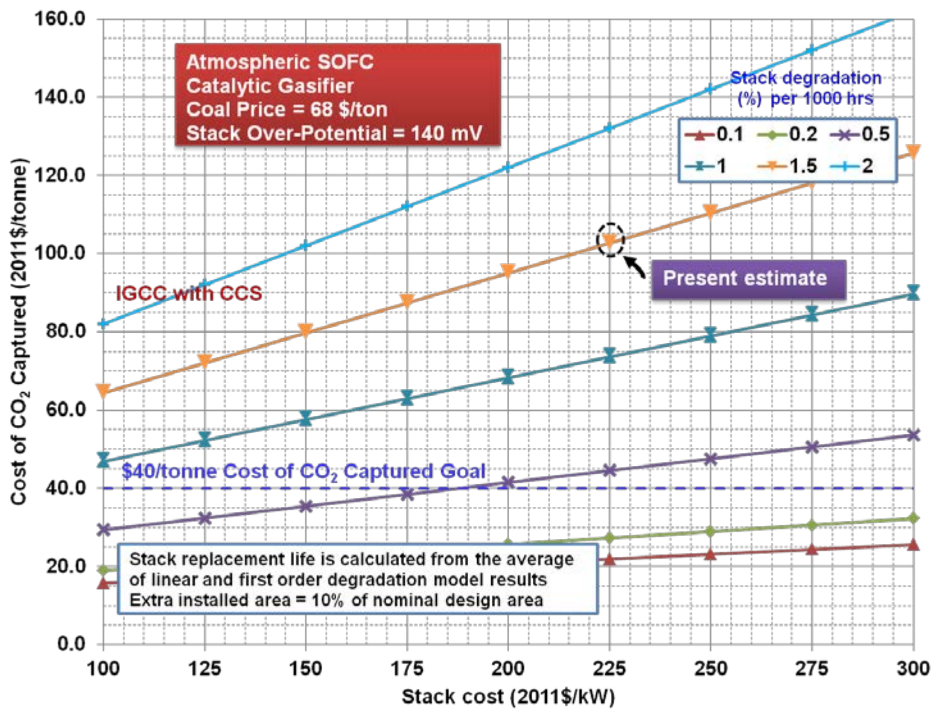
CASE	3-1	3-2	3-3	3-4	3-5	3-6	3-7
Dry Syngas CH_4 Content (%)	31.6						
SOFC Degradation Rate (%/1000 h)	1.5	0.2					
SOFC Overpotential (mV)	140		70				
Capacity Factor (%)	80			85	90		
SOFC Stack Cost ($/kW)	225					200	
Inverter Efficiency (%)	97						98
PERFORMANCE							
Gross Power (MWe)	668.9	668.9	654.8	654.8	654.8	654.8	653.7
Auxiliary Loads (MWe)	118.6	118.6	104.7	104.7	104.7	104.7	103.7
Net Power (MWe)	550.3	550.3	550.1	550.1	550.1	550.1	550.0
Net Electric Efficiency, HHV (%)	48.9	48.9	53.5	53.5	53.5	53.5	54.1
CO_2 Capture rate (%)	98.5	98.5	98.5	98.5	98.5	98.5	98.5
CO_2 Emissions (lb/MWhgross)	20.5	20.5	18.1	18.1	18.1	18.1	17.9
CO_2 Emissions (lb/MWhnet)	25.0	25.0	21.5	21.5	21.5	21.5	21.3
Raw Water Consumption (gpm/MWnet)	2.4	2.4	2.2	2.2	2.2	2.2	2.1
COST							
Total Plant Cost (TPC) (1000$)	1,146,491	1,146,491	1,074,226	1,074,226	1,074,226	1,059,099	1,052,954
Total Overnight Cost (TOC) (1000$)	1,423,360	1,423,360	1,333,965	1,333,814	1,333,680	1,315,498	1,307,827
Total As-Spent Cost (TASC) (1000$)	1,622,631	1,622,631	1,520,720	1,520,548	1,520,395	1,499,668	1,490,922
Cost of Electricity ($/MWh)							
Variable COE	83.9	34.6	32.6	31.9	31.3	30.9	30.7
Fuel	20.4	20.4	18.6	18.6	18.6	18.6	18.4
Variable O&M	63.6	14.2	14.0	13.3	12.7	12.3	12.2
Fixed O&M	12.4	12.4	12.1	11.4	10.7	10.7	10.6
Capital Charges	45.9	45.9	43.0	40.5	38.2	37.7	37.5
Total First Year COE (excluding T&S)	142.3	92.9	87.7	83.8	80.3	79.3	78.8
CO_2 T&S	6.5	6.5	6.0	6.0	6.0	6.0	5.9
Total First Year COE (including T&S)	148.8	99.5	93.7	89.7	86.2	85.3	84.7
NETL Metrics							
% COE reduction $(COE_{IGFC} - COE_{IGCC\ with\ CCS}) / COE_{IGCC\ with\ CCS}$	-7.0	30.1	34.1	37.0	39.6	40.4	40.8
Cost of Captured CO_2 ($/tonne CO_2) $(COE_{IGFC} - COE_{AUSC\ PC}) / CO_2$ Captured (tonnes/MWh)	113.2	30.1	23.2	16.0	9.5	7.8	6.9

Exhibit 4-23 Sensitivity of Case 3-1 COE to stack cost and degradation rate



Source: DOE/NETL

Exhibit 4-24 Sensitivity of Case 3-1 cost of CO_2 captured to stack cost and degradation rate



Source: DOE/NETL

4.3 Scenario 4 – IGFC with Pressurized-SOFC

Scenario 4 combines the catalytic coal gasifier with the advantages of a pressurized SOFC. Pressurized SOFC can be configured in two general, alternative arrangements:

The anode off-gas oxy-combustor is followed by hot gas expander power generation (expansion ratio about 18). An HRSG produces steam for power generation, and the remaining, low-pressure, wet CO_2 stream is dried and compressed (compression ratio about 149).

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. The pressurized 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. The Scenario 4 pressurized-SOFC Power Block assumptions and specifications are listed in Exhibit 4-25.The CPU remains essentially the same with the LP compressor working at a lower pressure ratio. However, no cost benefit was assumed in this study to reflect the reduced LP compressor requirements.

Exhibit 4-25 Scenario 4 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 (psi)	0.02 (3)
Cathode recycle gas rate, %	0
Cathode heat exchanger pressure drop, MPa (psi)	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

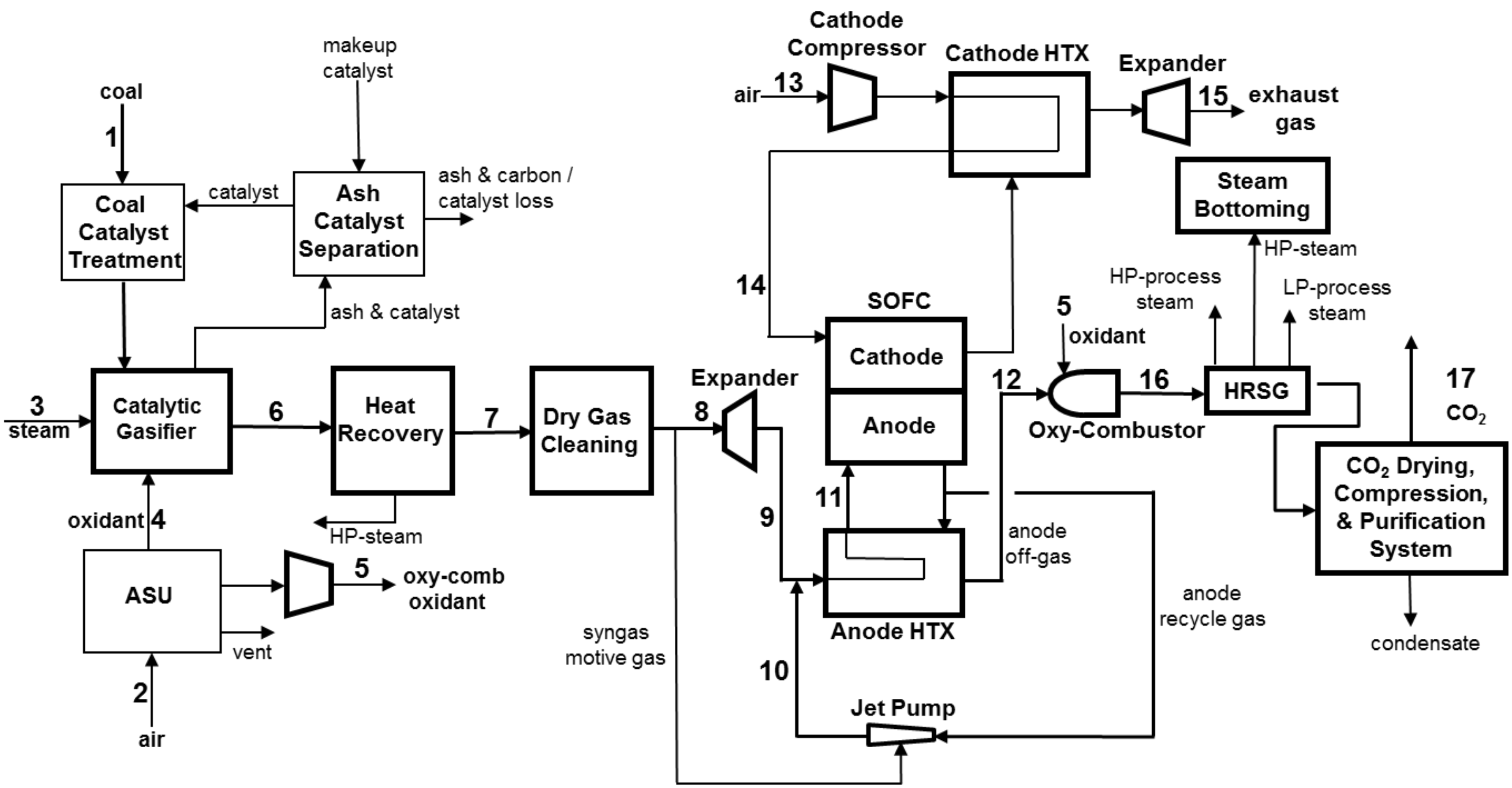
4.3.1 Case 4-1 IGFC Plant Performance Results

The relevant process data for the numbered streams in the BFD for the baseline plant shown in Exhibit 4-26 are tabulated in Exhibit 4-27. As in Case 3-1, the anode off-gas recirculation reduces the syngas methane content from ~31.6 mole percent to 10.4 mole percent in the anode. This case results in a plant HHV efficiency of ~58.4 percent, as shown in Exhibit 4-28, which is significantly higher than the efficiency of any of the earlier cases, and represents a transformational leap over other fossil-fuel based technologies. Exhibit 4-29 shows that the gross power generated is dominated by the SOFC power island contribution. The ASU and the CPU parasitics make up ~42 percent of the auxiliary load as shown in Exhibit 4-30. The SOFC power island losses[9] comprise ~43 percent of the parasitic loads, mainly due to the compressor losses. The SOFC operates at a voltage of 0.92 V, a direct result of the SOFC pressurization. The cathode air preheat heat exchanger in Case 4-1 is not as large as in Case 3-1, with a heat load of about 14 percent of the coal feed energy input, because the compression of the cathode air partially preheats the stream. The pressurized configuration increases the ASU auxiliary power relative to Case 3-1, due to the added compression of the oxy-combustion oxidant stream; however, the CO_2 compression power is reduced since the anode off-gas is already at an elevated pressure. The heat and mass balance diagram for the gasifier, ASU, and syngas clean-up is shown in Exhibit 4-31 while the corresponding PFDs for the power island and the CPU are shown in Exhibit 4-32 and Exhibit 4-33, respectively. Salient material and energy balances are shown in Exhibit 4-34. As in Case 3-1, the carbon inputs to plant 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.

The nearly complete recovery of water from the oxy-combustion CO_2 product stream results in an IGFC plant water consumption, also shown in Exhibit 4-34, which is significantly lower than the corresponding value for conventional fossil fuel power plant technologies. The IGFC plant acts as a nearly zero emission power plant, with the only significant emission being the small release of CO_2, shown in the emissions listed in Exhibit 4-34. This emissions performance is dictated, in part, by the need to protect the SOFC stack components from contamination.

[9] The DC-AC inverter losses and the SOFC polarization losses are included in the SOFC gross power estimates.

Exhibit 4-26 Case 4-1 block flow diagram



Source: DOE/NETL

Exhibit 4-27 Case 4-1 stream table

	1	2	3	4	5	6	7	8	9	10
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
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.1950	0.1950	0.3159	0.3159	0.0182
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0563	0.0563	0.0914	0.0914	0.0507
CO_2	0.0000	0.0003	0.0000	0.0000	0.0000	0.2128	0.2128	0.3465	0.3465	0.4099
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0001	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.1472	0.1472	0.2386	0.2386	0.0604
H_2O	0.0000	0.0104	1.0000	0.0000	0.0000	0.3760	0.3760	0.0004	0.0004	0.4563
HCl	0.0000	0.0000	0.0000	0.0000	0.0000	0.0008	0.0008	0.0000	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0071	0.0071	0.0000	0.0000	0.0000
N_2	0.0000	0.7722	0.0000	0.0019	0.0019	0.0039	0.0039	0.0069	0.0069	0.0044
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0007	0.0007	0.0000	0.0000	0.0000
O_2	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
Total	0.0000	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	7,787	9,385	698	808	13,579	13,579	7,317	7,317	18,168
V-L Flowrate (kg/hr)	0	224,686	169,073	22,358	25,869	292,566	292,566	172,471	172,471	512,755
Solids Flowrate (kg/hr)	125,033	0	0	0	0	0	0	0	0	0
Temperature (°C)	15	15	291	130	27	704	204	371	264	727
Pressure (MPa, abs)	0.10	0.10	7.58	3.45	0.16	6.72	6.62	5.61	2.00	2.00
Enthalpy (kJ/kg)[A]	---	31.06	2,860.25	114.27	23.91	2,107.45	906.26	606.83	414.52	1,772.26
Density (kg/m^3)	---	1.2	38.4	33.1	2.0	17.7	42.2	24.4	10.5	6.8
V-L Molecular Weight	---	28.854	18.015	32.016	32.016	21.546	21.546	23.570	23.570	28.224
V-L Flowrate (lb_{mol}/hr)	0	17,167	20,690	1,540	1,781	29,936	29,936	16,132	16,132	40,052
V-L Flowrate (lb/hr)	0	495,349	372,742	49,291	57,032	644,997	644,997	380,233	380,233	1,130,431
Solids Flowrate (lb/hr)	275,650	0	0	0	0	0	0	0	0	0
Temperature (°F)	59	59	555	265	80	1,300	400	700	507	1,341
Pressure (psia)	14.7	14.7	1,100.0	500.0	23.0	975.0	960.0	814.0	290.0	290.0
Enthalpy (Btu/lb)[A]	---	13.4	1,229.7	49.1	10.3	906.0	389.6	260.9	178.2	761.9
Density (lb/ft^3)	---	0.076	2.398	2.067	0.127	1.104	2.635	1.523	0.657	0.423

A - Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 4-27 Case 4-1 stream table (continued)

	11	12	13	14	15	16	17
V-L Mole Fraction							
Ar	0.0002	0.0002	0.0094	0.0094	0.0103	0.0004	0.0000
CH_4	0.1037	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0624	0.0482	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.3917	0.4138	0.0003	0.0003	0.0003	0.4572	1.0000
COS	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.1115	0.0495	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.3254	0.4842	0.0104	0.0104	0.0114	0.5282	0.0000
HCl	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
N_2	0.0051	0.0042	0.7722	0.7722	0.8478	0.0043	0.0000
NH_3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.2077	0.2077	0.1302	0.0100	0.0000
SO_2	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
V-L Flowrate (kg_{mol}/hr)	25,485	13,646	67,285	67,285	61,288	13,788	6,205
V-L Flowrate (kg/hr)	685,225	389,015	1,941,467	1,941,467	1,749,561	414,885	273,088
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0
Temperature (°C)	649	644	15	652	123	1,144	24
Pressure (MPa, abs)	1.98	1.94	0.10	2.0	0.107	1.889	15.272
Enthalpy (kJ/kg)^A	1,526.66	1,661.74	31.06	708.8	143.369	2,480.020	-242.014
Density (kg/m^3)	6.9	7.3	1.2	7.4	0.9	4.8	740.0
V-L Molecular Weight	26.888	28.508	28.854	29	28.547	30.091	44.010
V-L Flowrate (lb_{mol}/hr)	56,184	30,084	148,339	148,339	135,117	30,397	13,680
V-L Flowrate (lb/hr)	1,510,663	857,632	4,280,203	4,280,203	3,857,121	914,664	602,057
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0
Temperature (°F)	1,200	1,191	59	1,205	253	2,092	74
Pressure (psia)	287.0	282.0	14.7	287.0	15.5	274.0	2,215.0
Enthalpy (Btu/lb)^A	656.3	714.4	13.4	304.7	61.6	1,066.2	-104.0
Density (lb/ft^3)	0.432	0.454	0.076	0	0.058	0.301	46.197

A - Reference conditions are 32.02 F & 0.089 PSIA

Exhibit 4-28 Case 4-1 plant performance summary (100 percent load)

POWER SUMMARY (Gross Power at Generator Terminals, kW_e)	
SOFC Power	575,200
Syngas Expander Power	8,900
Steam Turbine Power	66,900
TOTAL GROSS POWER (kW_e)	**651,000**
AUXILIARY LOAD SUMMARY (kW_e)	
Coal Handling	370
Coal size reduction	402
Catalyst-coal processing	1,562
Catalyst coal feeding	860
Sour Water Recycle Slurry Pump	114
Ash Handling	920
Air Separation Unit Auxiliaries	286
Air Separation Unit Main Air Compressor	13,220
Oxygen Compressor	2,270
Claus Plant TG Recycle Compressor	1,060
CO_2 Compressor	3,790
CO_2 Purification	22,641
Boiler Feedwater Pumps	1,161
Condensate Pump	87
Circulating Water Pump	1,120
Ground Water Pumps	296
Cooling Tower Fans	590
Scrubber Pumps	190
Selexol Auxiliary Power	1,757
Steam Turbine Auxiliaries	92
Claus Plant/TGTU Auxiliaries	28
Gas Turbine Auxiliaries	142
Cathode Air Compressor - Cathode Expander	42,830
Miscellaneous Balance of Plant[2]	2,775
Transformer Losses	2,230
TOTAL AUXILIARIES, kW_e	100,795
NET POWER (kW_e)	**550,205**
NET PLANT EFFICIENCY, % (HHV)	**58.4**
NET PLANT HEAT RATE, kJ/kWh (Btu/kWh)	**6,166 (5,845)**
CO_2 Capture Rate (%)	**98.4**
CONDENSER COOLING DUTY 10^6 kJ/h (10^6 Btu/h)	**390 (370)**
As-Received Coal Feed, kg/h (lb/h)	125,033 (275,650)
Thermal Input[1], kWt	942,438
Raw Water Consumption, m^3/min (gpm)	2.9 (774)

[1] *HHV of as received Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb)*

[2] *Includes plant control systems, lighting, HVAC, and miscellaneous low-voltage loads*

Exhibit 4-29 Case 4-1 gross power generation components

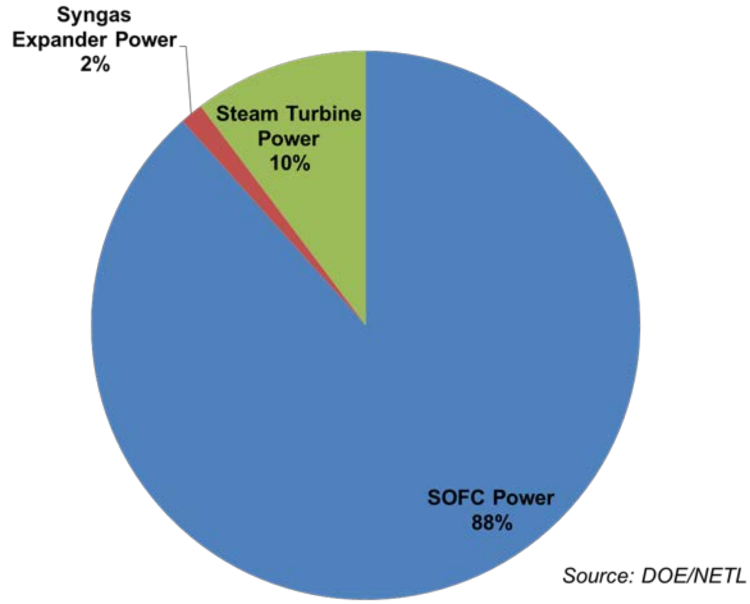


Exhibit 4-30 Components of Case 4-1 auxiliary load

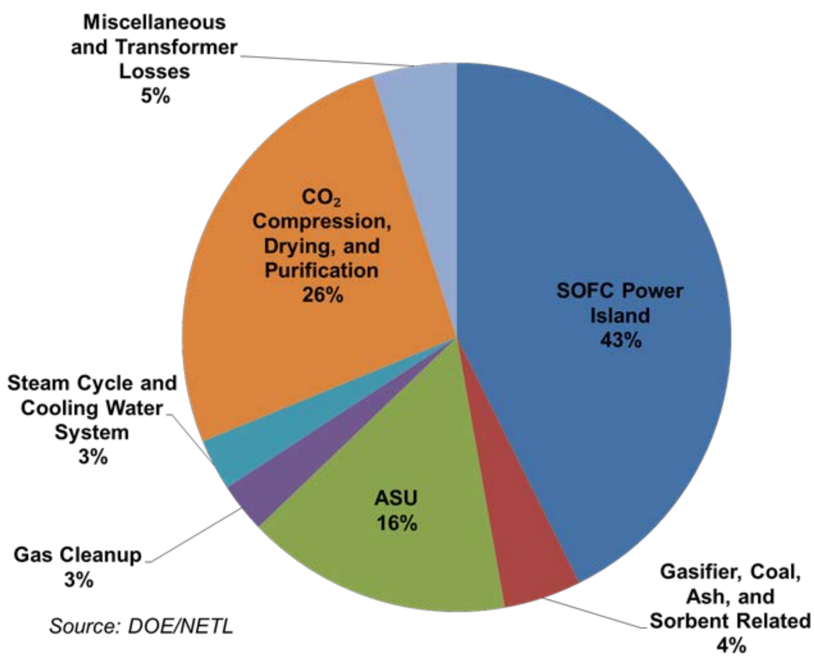
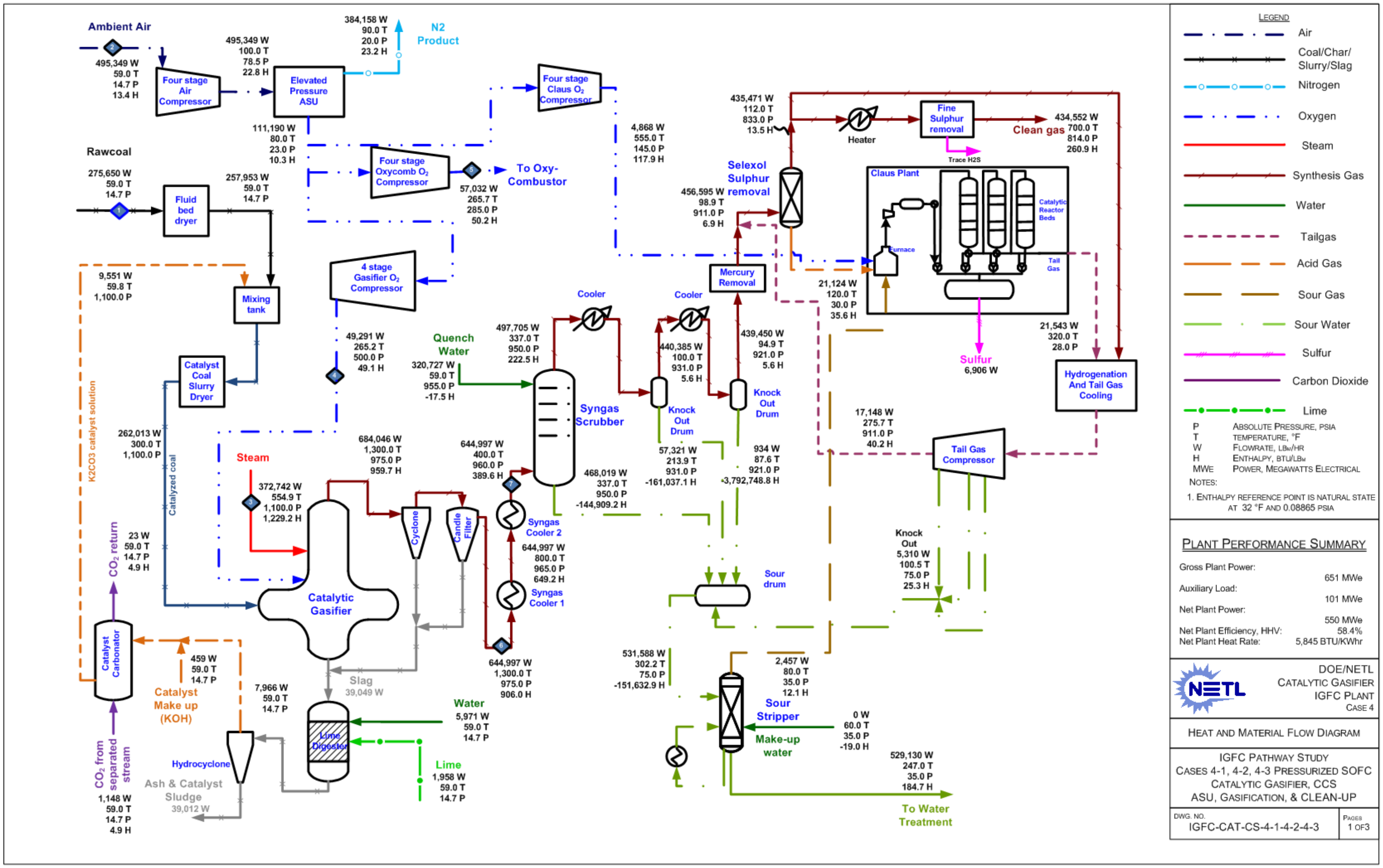
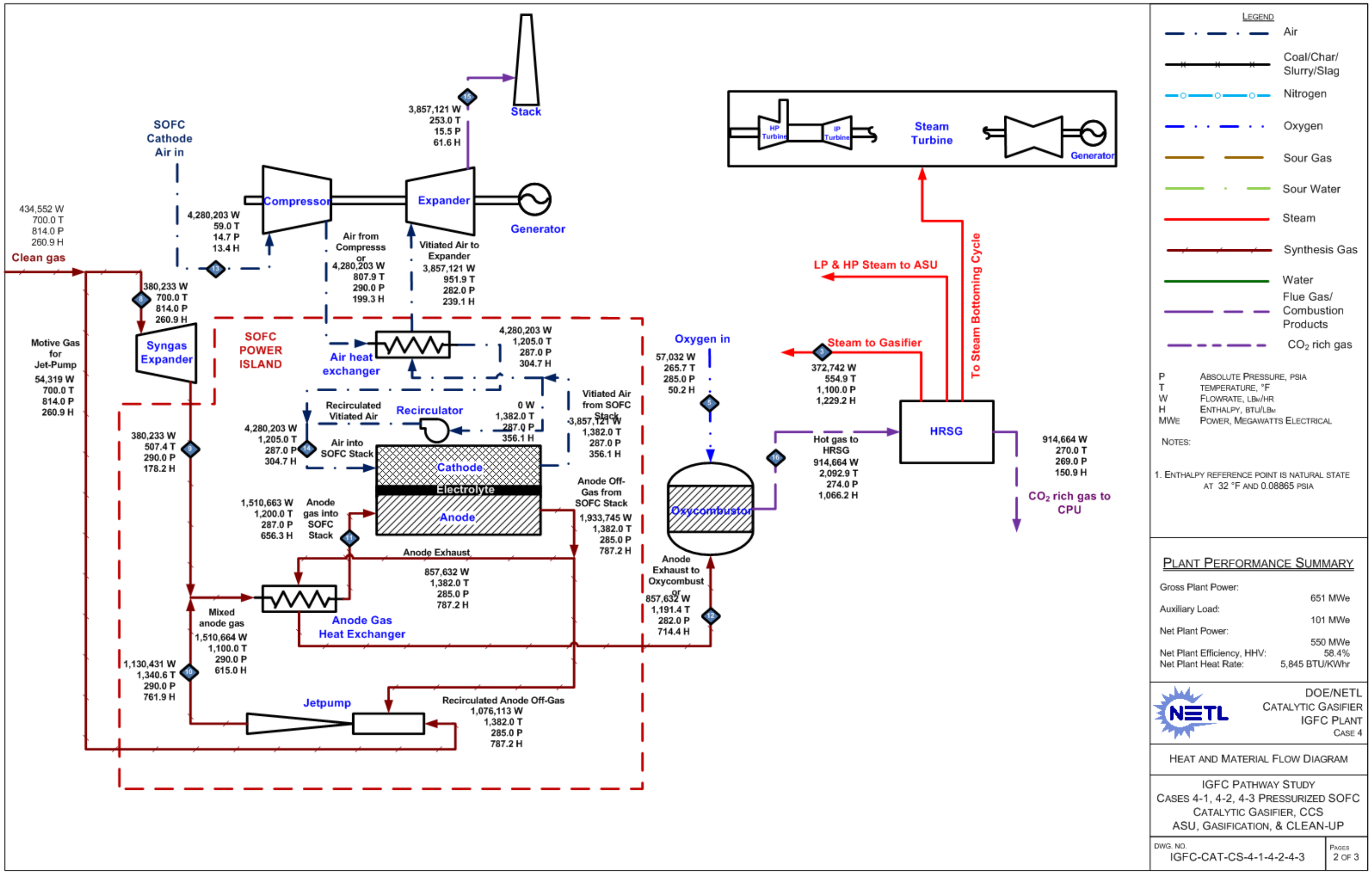


Exhibit 4-31 Case 4-1 process flow diagram – gasifier, ASU, and syngas clean-up



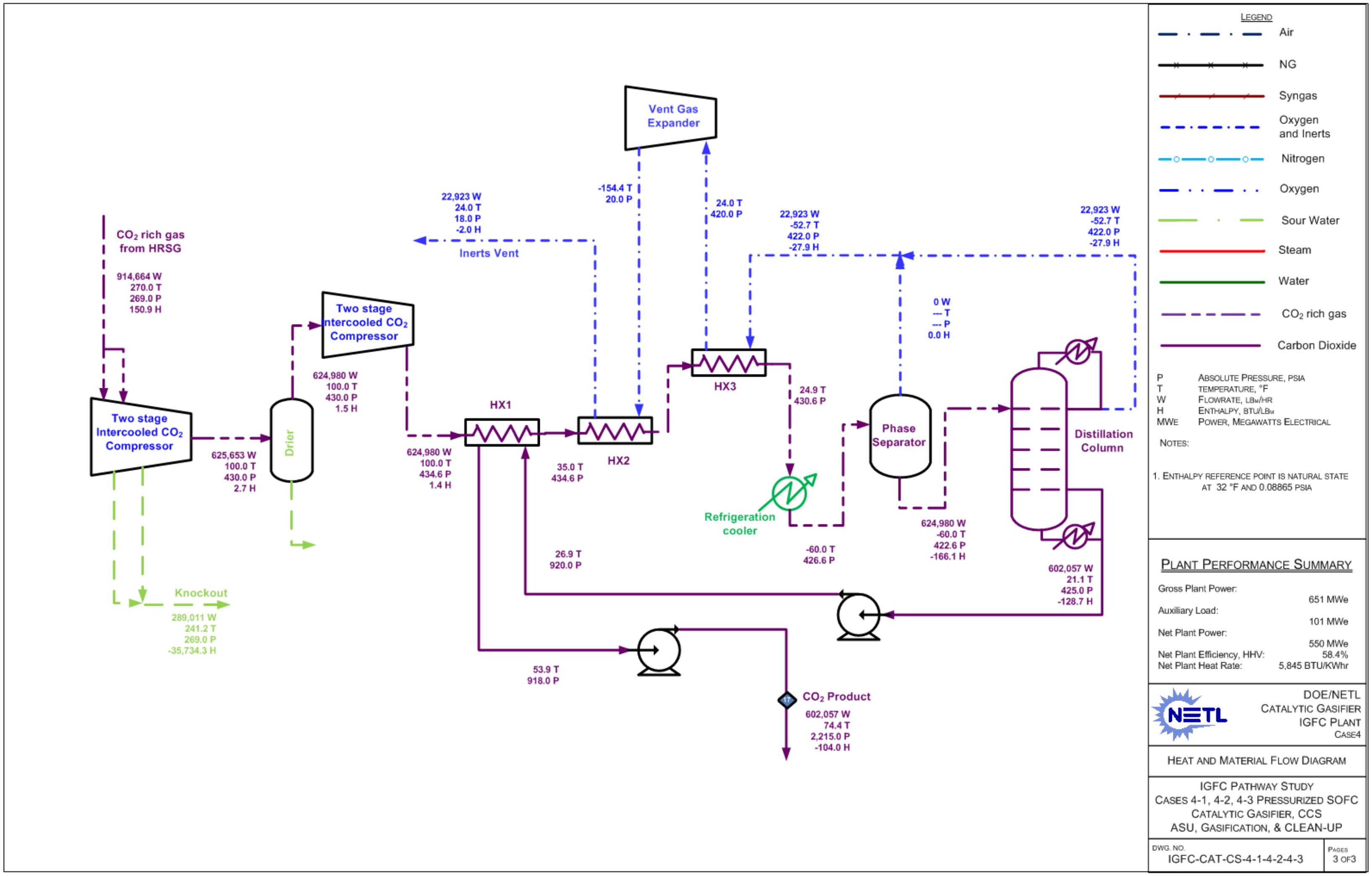
Source: DOE/NETL

Exhibit 4-32 Case 4-1 process flow diagram – IGFC power island



Source: DOE/NETL

Exhibit 4-33 Case 4-1 process flow diagram – CO_2 separation and purification



Source: DOE/NETL

Exhibit 4-34 Case 4-1 mass and energy balances

Carbon In		Carbon Out	
kg/hr(lb/hr)		kg/hr(lb/hr)	
Coal	79,702 (175,712)	Slag	4,129 (9,102)
Air (CO2)	271 (596)	Stack Gas	242 (535)
Cat Carbonator	142 (313)	CO_2 Product	74,530 (164,310)
		N_2 Product	28 (62)
		CO_2 Dehydration	56 (123)
		Vent Gas	1,133 (2,497)
		Convergence Tolerance	-3 (-7)
Total	80,114 (176,622)	Total	80,114 (176,622)

	HHV	Sensible + Latent	Power	Total
Heat In GJ/hr (MMBtu/hr)				
Coal	3,393 (3,216)	2.8 (2.7)		3,396 (3,218)
ASU Air		7.0 (6.6)		7 (7)
Fuel cell Air		60.3 (57.1)		60 (57)
Raw Water Makeup		14.7 (13.9)		15 (14)
Auxiliary Power			363 (344)	363 (344)
TOTAL	**3,393 (3,216)**	**84.8 (80.4)**	**363 (344)**	**3,840 (3,640)**
Heat Out GJ/hr (MMBtu/hr)				
Slag	135 (128)	13.2 (12.5)		149 (141)
Sulfur	29 (28)	0.3 (0.3)		29 (28)
CO_2		-66.1 (-62.6)		-66 (-63)
CO_2 Refrigeration		97.3 (92.2)		97 (92)
Cooling Tower Blowdown		7.3 (6.9)		7 (7)
HRSG Flue Gas		250.8 (237.7)		251 (238)
Condenser		391 (371)		391 (371)
Non-Condenser Cooling Tower Loads*		174 (165)		174 (165)
Process losses*		448 (425)		448 (425)
Difference**		16 (15)		16 (15)
Power			2,344 (2,221)	2,344 (2,221)
TOTAL	**164 (156)**	**1,332 (1,263)**	**2,344 (2,221)**	**3,840 (3,640)**

Sulfur balance

Sulfur In		Sulfur Out	
kg/hr(lb/hr)		kg/hr(lb/hr)	
Coal	3,134 (6,909)	Elemental Sulfur	3,132 (6,906)
		Polishing Sorbent	2 (3)
		Convergence Tolerance	0 (0)
Total	3,134 (6,909)	Total	3,134 (6,909)

Water balance

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m3/min (gpm)	m3/min (gpm)	m3/min (gpm)	m3/min (gpm)	m3/min (gpm)
Slag Handling	0.38 (101)	0.38 (101)			
Slurry Water	0.05 (12)	0.05 (12)			
Quench/Wash	2.4 (641)	2.43 (641)			
Condenser Makeup	**2.9 (764)**	**0.0 (0)**	**2.9 (764)**	**0.0 (0)**	**2.9 (764)**
Gasifier Steam	2.8 (745)	0.0 (0)	2.8 (745)		
BFW Makeup	0.1 (19)	0.0 (0)	0.1 (19)		
Cooling Tower	**4.3 (1,149)**	**3.34 (882)**	**1.0 (268)**	**1.0 (258)**	**0.0 (09)**
CO2 Dehydration	0.0 (0)	2.19 (578)	-2.19 (-578)		
SWS Excess Water	0.0 (0)	1.15 (304)	-1.15 (-304)		
Total	**10.1 (2,668)**	**6.19 (1636)**	**3.9 (1,032)**	**1.0 (258)**	**2.9 (774)**

Emissions

	kg/GJ ($lb/10^6 Btu$)	Tonne/year (tons/year)	kg/MWh (lb/MWh)
SO2	0 (0)	0 (0)	0 (0)
NOx	0 (0)	0 (0)	0 (0)
Particulate	0 (0)	0 (0)	0 (0)
Hg	0 (0)	0 (0)	0 (0)
CO2	2 (4)	39,048 (43,043)	8 (18)

** Process losses include losses from steam turbines, expanders, inverter, and blowers*

***Value needed to match heat input to the plant and includes minor process losses*

4.3.2 Case 4-1 IGFC Plant Cost Results

The SOFC power island capital costs for Case 2-1 are shown in Exhibit 4-35 and a distribution of cost among its major components are shown in Exhibit 4-36. The SOFC module costs account for nearly 85 percent of the SOFC power island costs relative to the atmospheric case, the SOFC pressure vessel costs being a significant addition to the costs over the atmospheric cases. The cathode air compressor and the cathode expander constitute the next significant expense. The cathode heat exchanger is not as large as in the atmospheric cases and does not influence the SOFC power island costs significantly in this case.

Exhibit 4-35 Case 4-1 SOFC power island capital costs

Cost Component	Cost ($1000)	Specific Cost ($/kWe AC)
	2011$	
SOFC Module		
SOFC Stack	123,796	225
Enclosure	132,049	240
Transport and Placement	7,923	14
Site Foundations	24,429	44
Inverter	37,523	68
Total SOFC Module	**325,720**	**592**
Total SOFC Module with 10% Extra Installed Area	**358,293**	**651**
SYNGAS EXPANDER	**2,990**	**5**
SOFC BOP		
Cathode Air Compressor	28,555	52
Cathode Gas Expander	11,735	21
Cathode Heat Exchanger	6,654	12
Anode Syngas Jet Pump	347	1
Anode Heat Exchanger	13	0
Oxy-Combustor	11,338	21
Total SOFC BOP	**58,642**	**107**
TOTAL SOFC POWER ISLAND	**419,924**	**763**

As in Case 1-1, Exhibit 4-37, which graphically depicts the IGFC capital costs listed in Exhibit 4-38, indicates that the SOFC power island capital costs, the gasifier, and the ASU costs form a major portion, ~54 percent, of the total IGFC plant capital costs. The TOC, calculated as in Exhibit 4-39 results in COEs of ~$89/MWh and ~$94/MWh with and without CO_2 T&S charges, respectively, as shown in Exhibit 4-40, with the O&M costs as tabulated in Exhibit 4-41. Unlike Case 1-1, the capital charges have the highest value amongst the Case 2-1 COE components due to the assumption of lower (0.2 percent per 1000 h) stack degradation rate in the latter. However, the pressurized Case 4-1 does not seem to have any particular advantage while it costs slightly higher than Case 3-3, which features an atmospheric SOFC with a similar performance, but with less operational complexities.

Exhibit 4-36 Distribution of Case 4-1 SOFC power island capital costs

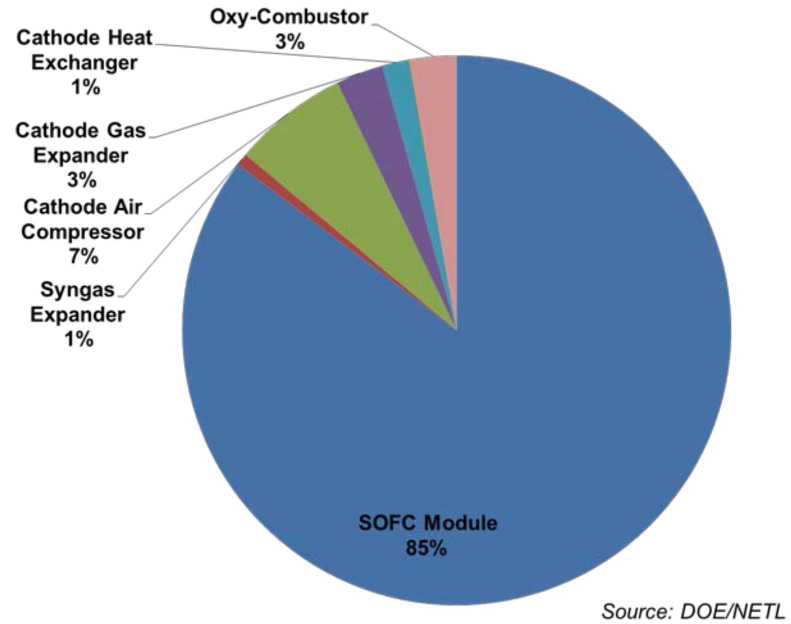
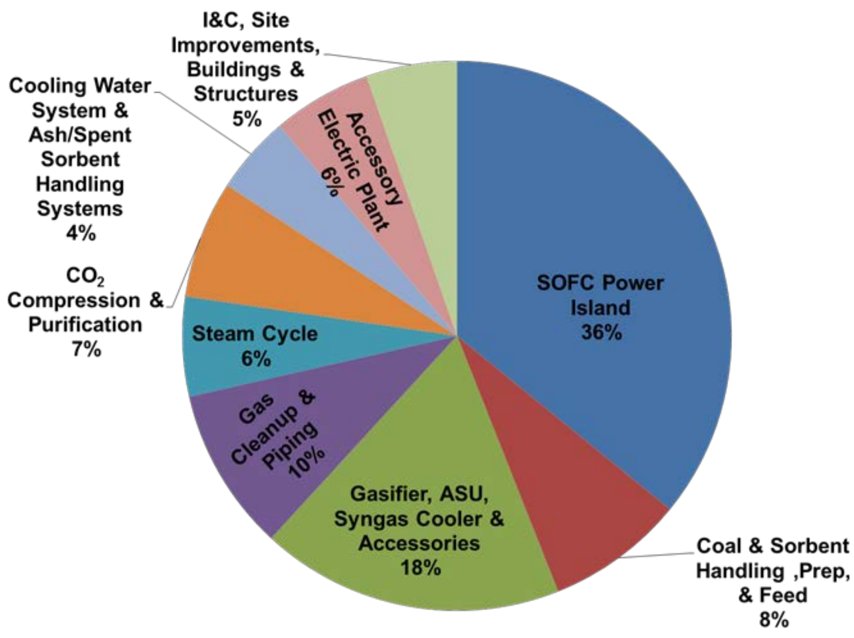


Exhibit 4-37 Case 4-1 IGFC plant cost distribution



Source: DOE/NETL

Exhibit 4-38 Case 4-1 IGFC plant capital costs

Cost Component	Cost ($1000)	Specific Cost ($/kWe AC)
	2011$	
SOFC POWER ISLAND	**419,924**	**763**
COAL & SORBENT HANDLING	**28,243**	**51**
COAL & SORBENT PREP & FEED	**66,884**	**122**
GASIFIER, ASU, SYNGAS COOLER & ACCESSORIES		
Gasifier & Syngas Cooler	114,478	208
ASU & oxidant compressor	81,062	147
Other gasification equip & foundations	13,219	24
TOTAL GASIFIER, ASU, SYNGAS COOLER & ACCESSORIES	**208,759**	**379**
GAS CLEANUP & PIPING		
Scrubber & Low Temperature Cooling	12,838	23
Single-Stage Selexol/MDEA	62,004	113
Claus Plant	22,057	40
Trace removal	1,557	3
COS Hydrolysis	6,778	12
Blowback, Piping, Foundations	3,850	7
Sulfur polishing/NG desulfurization	2,970	5
Heat Interchanger	569	1
TOTAL GAS CLEANUP & PIPING	**112,622**	**205**
STEAM CYCLE		
HRSG, Ducting, and Stack	28,524	52
Steam Power System	28,281	51
Feedwater and Misc BOP systems	11,677	21
TOTAL STEAM CYCLE	**68,482**	**124**
CO_2 COMPRESSION & PURIFICATION		
CO_2 Drying and Compression	-	-
CO_2 Purification	81,031	147
TOTAL CO_2 COMPRESSION & PURIFICATION	**81,031**	**147**
COOLING WATER SYSTEM	**12,749**	**23**
ASH/SPENT SORBENT HANDLING SYSTEM	**41,063**	**75**
ACCESSORY ELECTRIC PLANT	**68,397**	**124**
INSTRUMENTATION & CONTROL	**33,097**	**60**
IMPROVEMENTS TO SITE	**15,131**	**28**
BUILDING & STRUCTURES	**14,322**	**26**
TOTAL PLANT COST (TPC)	**1,170,703**	**2128**

Exhibit 4-39 Case 4-1 TOC and TASC

Cost Component	Cost ($1000)	Specific Cost ($/kWe AC)
	2011$	2011$
OWNER'S COSTS		
Preproduction Costs		
6 Months All Labor	12,427	
1 Month Maintenance Materials	2,482	
1 Month Non-fuel Consumables	358	
1 Month Waste Disposal	413	
25% of 1 Months Fuel Cost at 100% CF	1,715	
2% of TPC	23,414	
Total Preproduction Costs	**40,808**	
Inventory Capital		
60 day supply of fuel and consumables at 100% CF	14,335	
0.5% of TPC (spare parts)	5,854	
Total Inventory Capital	**20,189**	
Initial Cost for Catalyst and Chemicals	**7,664**	
Land	**900**	
Other Owner's Costs	**175,606**	
Financing Costs	**31,609**	
TOTAL OWNER'S COSTS	**276,775**	
TOTAL OVERNIGHT COST (TOC)	**1,447,479**	**2631**
TASC Multiplier	1.14	
TOTAL AS-SPENT COST (TASC)	**1,650,126**	**2999**

Exhibit 4-40 Case 4-1 cost of electricity

COST OF ELECTRICITY (2011$/MWh) with CPU	
	$/MWh
Variable COE	**33.2**
Fuel	**17.1**
Variable O&M	**16.1**
Fixed O&M	**11.8**
Capital Charges	**43.9**
First Year COE (excluding T&S)	**88.9**
CO_2 T&S	**5.5**
First Year COE (with T&S)	**94.4**

Exhibit 4-41 Case 4-1 O&M costs

Case 4-1 Pressurized SOFC Catalytic Gasifier						
Net Power: 550.2 Mwe, Capacity Factor: 80%, Heat Rate: 6,942 Btu/kWh						
Cost Component					**Cost ($)**	**Cost ($/MWh)**
					2011$	
OPERATING & MAINTENANCE LABOR						
Operating Labor Rate (base)					39.70	
Operating Labor Burden					30%	
Labor O-H Charge Rate					25%	
Total Operators per shift					15	
Maintenance labor/Operating labor					2.302	
Maintenance materials/Maintenance labor					1.932	
Stack replacement O&M, $/hr per stack kW					18.33	
Annual Operating Labor Cost					6,781,554	
Maintenance Labor Cost					13,101,962	
Administrative & Support Labor					4,970,879	
Property Taxes and Insurance					23,514,291	
TOTAL FIXED OPERATING COSTS					**48,368,686**	**11.81**
VARIABLE OPERATING COSTS						
Maintenance Material Cost					**25,312,991**	**6.18**
Stack replacement						
SOFC stack life (years)					7.294	
Discount rate for stack replacement (%)					10.0%	
SOFC stack replacement cost, $/kW AC					$524	
SOFC Stack replacement O&M, $/yr per kW					$52	
Stack Replacement Cost					**32,982,632**	**8.05**
CONSUMABLES	**Consumption**		**Cost ($)**		**Annual ($)**	
	Initial Fill	**/ Day**	**/ Unit**	**Initial Fill**		
Water (/1000 gallons)		946	1.670		**490,220**	
Chemicals						
MU & WT Chem. (lb)		4,181	0.27		350,251	
Carbon (Trace Removal) (lb)	308,030	422	1.6	502,089	213,388	
COS Catalyst (m^3)	233	0.16	3,752	874,602	185,853	
Selexol Solution (gal)	157,664	25	36.79	5,800,460	283,624	
Claus / DSRP Catalyst (ft^3)		1.07	203		67,608	
ZnO polishing sorbent (lb)	124,559	1,253	1.8	224,205	699,639	
KOH Coal Catalyst makeup (lb)	661,069	11,018	0.192	126,925	656,310	
Lime for catalyst recovery	2,820,098	47,002	0.048	135,365	699,948	
Sub Total Chemicals				**7,663,646**	**3,156,619**	
Waste Disposal						
Spent Trace Catalyst (lb)		456	0.65		91,901	
Ash + HCl Sorbent (ton)		497	25.11		3,873,045	
Spent sorbents (lb)		1253	0.65		252,647	
Subtotal Waste Disposal					**4,217,593**	
TOTAL VARIABLE OPERATING COSTS				**7,663,646**	**66,160,056**	**16.15**
Fuel Coal (ton)		68.60			**69,951,977**	**17.07**

4.3.3 Scenario 4 Pathway Results

The performance and cost for various SOFC system advances were evaluated in a cumulative manner for Scenario 4:

Case 4-2: The capacity factor of Case 4-1 was further increased to 90 percent.

Case 4-3: The stack cost was reduced from $225/kW in Case 4-2 to $200/kW.

Case 4-4: An improvement in the inverter efficiency from the 97 percent in Case 4-3 to 98 percent was analyzed in this case.

The performances and costs of the Scenario 4 pathway cases are summarized in Exhibit 4-42 and Exhibit 4-43, respectively. The 1 absolute percent improvement in the inverter efficiency in Case 4-4 results in a 0.6 percentage point increase in plant HHV efficiency over Case 4-1. The decrease observed in the COE from Case 4-1 to Case 4-4 is mainly attributable to the assumptions of increased capacity factor and lower stack cost. The combined effects of stack cost and stack degradation rate on the COE and the cost of captured CO_2 are presented in Exhibit 4-44 and Exhibit 4-45. As in Scenario 3 cases, a stack degradation rate of 0.5 percent per 1000 h seems to be acceptable for the catalytic gasifier to be competitive with respect to IGCC systems. However, in reality, a stack degradation rate that is lower than 0.2 percent per 1000 h is desirable to compete with other technologies such as the NGCC systems.

Exhibit 4-42 Comparison of performance of Scenario 4 pathway cases

CASE	4-1	4-2	4-3	4-4
Dry Syngas CH_4 Content (%)	**31.6**			
SOFC Overpotential (mV)	**70**			
Inverter Efficiency (%)	**97**			**98**
POWER SUMMARY (Gross Power at Generator Terminals, kW_e)				
SOFC Power	575,200			575,100
Syngas Expander Power	8,900			8,800
Steam Turbine Power	66,900			66,100
TOTAL GROSS POWER (kW_e)	**651,000**			**650,000**
AUXILIARY LOAD SUMMARY (kW_e)				
Coal Handling	370			370
Coal size reduction	402			398
Catalyst-coal processing	1,562			1,546
Catalyst coal feeding	860			851
Sour Water Recycle Slurry Pump	114			113
Ash Handling	920			910
Air Separation Unit Auxiliaries	286			283
Air Separation Unit Main Air Compressor	13,220			13,080
Oxygen Compressor	2,270			2,250
Claus Plant TG Recycle Compressor	1,060			1,040
CO_2 Compressor	3,790			3,750
CO_2 Purification	22,641			22,407
Boiler Feedwater Pumps	1,161			1,147
Condensate Pump	87			85
Circulating Water Pump	1,120			1,110
Ground Water Pumps	296			293
Cooling Tower Fans	590			580
Scrubber Pumps	190			188
Selexol Auxiliary Power	1,757			1,739
Steam Turbine Auxiliaries	92			91
Claus Plant/TGTU Auxiliaries	28			28
Gas Turbine Auxiliaries	142			141
Cathode Air Compressor - Cathode Expander	42,830			42,380
Miscellaneous Balance of Plant[2]	2,775			2,771
Transformer Losses	2,230			2,220
TOTAL AUXILIARIES (kW_e)	**100,795**			**99,772**
NET POWER (kW_e)	**550,205**			**550,228**
NET PLANT EFFICIENCY, % (HHV)	**58.4**			**59.0**
NET PLANT HEAT RATE, kJ/kWh (Btu/kWh)	**6,166 (5,845)**			**6,102 (5,784)**
CO_2 Capture Rate (%)	**98.4**			**98.4**
CONDENSER COOLING DUTY 10^6 kJ/h (10^6 Btu/h)	**390 (370)**			**390 (370)**
As-Received Coal Feed, kg/h (lb/h)	125,033 (275,650)			123,740 (272,800)
Thermal Input[1], kWt	942,438			932,694
Raw Water Consumption, m^3/min (gpm)	2.9 (774)			2.9 (765)

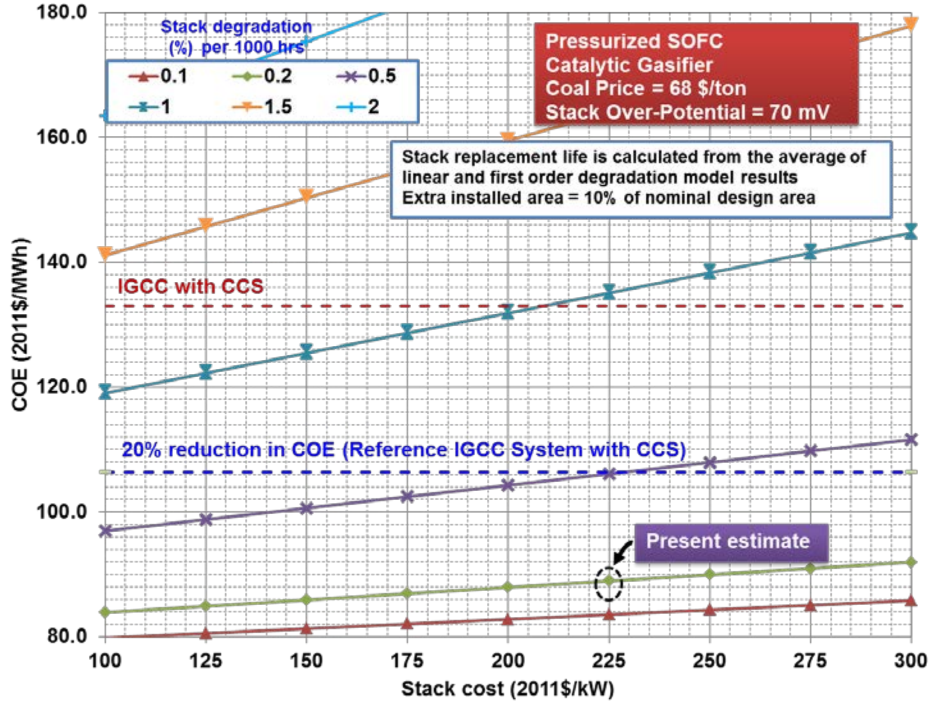
[1] HHV of as received Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb)

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

Exhibit 4-43 Comparison of Scenario 4 pathway cases costs

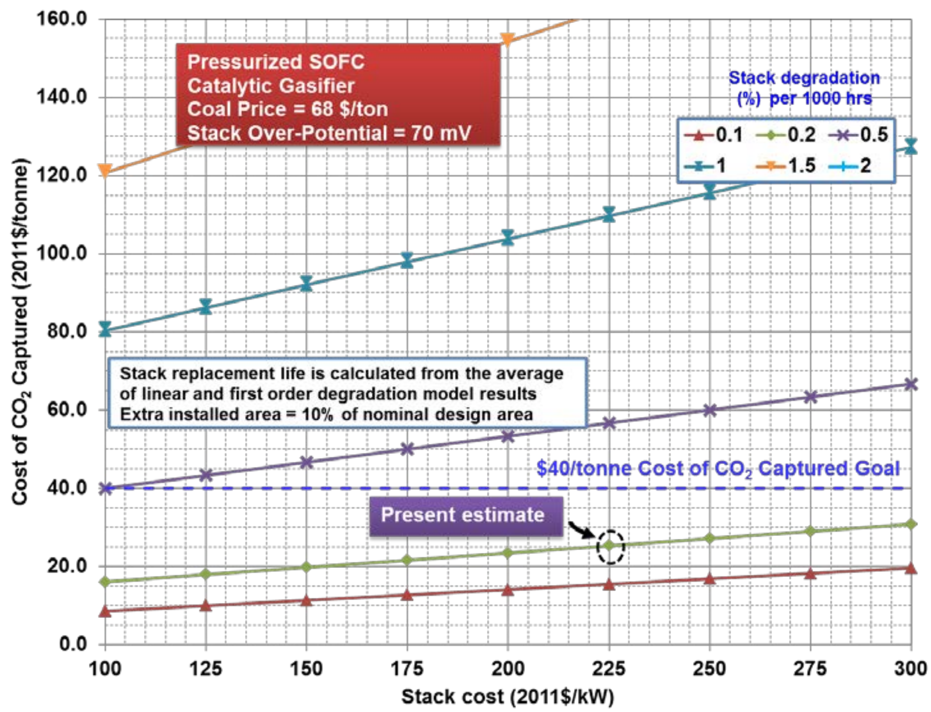
CASE	4-1	4-2	4-3	4-4
Dry Syngas CH_4 Content (%)	31.6			
SOFC Overpotential (mV)	70			
Capacity Factor (%)	85	90		
SOFC Stack Cost ($/kW)	225		200	
Inverter Efficiency (%)	97			98
PERFORMANCE				
Gross Power (MWe)	651.0	651.0	651.0	650.0
Auxiliary Loads (MWe)	100.8	100.8	100.8	99.8
Net Power (MWe)	550.2	550.2	550.2	550.2
Net Electric Efficiency, HHV (%)	58.4	58.4	58.4	59.0
CO_2 Capture rate (%)	98.4	98.4	98.4	98.4
CO_2 Emissions (lb/MWhgross)	17.8	17.8	17.8	22.0
CO_2 Emissions (lb/MWhnet)	21.0	21.0	21.0	26.0
Raw Water Consumption (gpm/MWnet)	1.4	1.4	1.4	1.4
COST				
Total Plant Cost (TPC) (1000$)	1,170,703	1,170,703	1,155,573	1,149,842
Total Overnight Cost (TOC) (1000$)	1,447,479	1,447,344	1,429,157	1,422,015
Total As-Spent Cost (TASC) (1000$)	1,650,126	1,649,972	1,629,238	1,621,098
Cost of Electricity ($/MWh)				
Variable COE	33.2	32.4	32.1	31.8
Fuel	17.1	17.1	17.1	16.9
Variable O&M	16.1	15.4	15.0	14.9
Fixed O&M	11.8	11.1	11.1	11.1
Capital Charges	43.9	41.5	41.0	40.7
Total First Year COE (excluding T&S)	88.9	85.1	84.1	83.6
CO_2 T&S	5.5	5.5	5.5	5.4
Total First Year COE (including T&S)	94.4	90.5	89.6	89.0
NETL Metrics				
% COE reduction $(COE_{IGFC} - COE_{IGCC\ with\ CCS}) / COE_{IGCC\ with\ CCS}$	33.1	36.0	36.8	37.1
Cost of Captured CO_2 ($/tonne CO_2) $(COE_{IGFC} - COE_{AUSC\ PC}) / CO_2$ Captured (tonnes/MWh)	27.9	20.1	18.2	17.3

Exhibit 4-44 Sensitivity of Case 4-1 COE to stack cost and degradation rate



Source: DOE/NETL

Exhibit 4-45 Sensitivity of Case 4-1 cost of CO_2 captured to stack cost and degradation rate



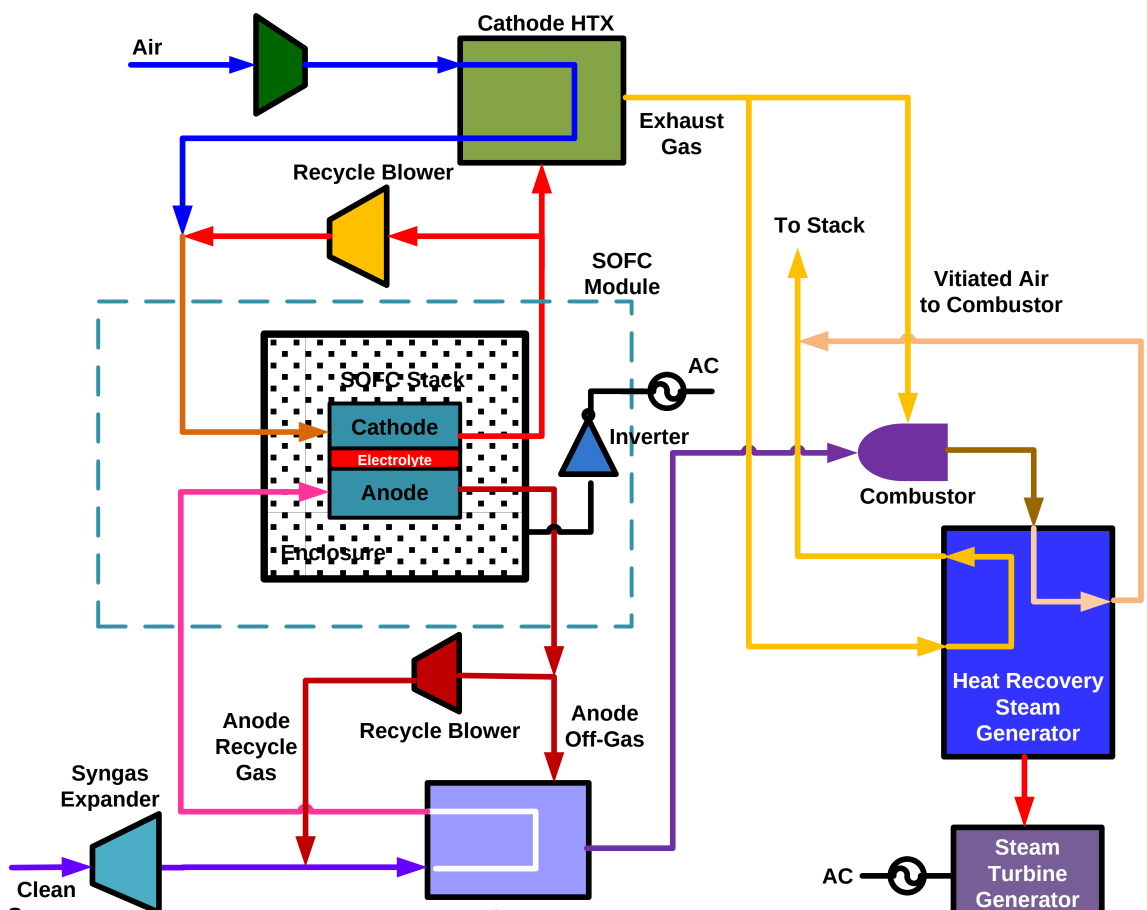
Source: DOE/NETL

5 IGFC Cases without CCS

Currently, the Environmental Protection Agency has proposed a GHG emissions limit of 1100 lb/MWh$_{GROSS}$ for coal-fueled plants. Salient IGFC cases were analyzed to evaluate their potential in meeting the GHG limits sans the performance and cost burden associated with carbon capture.

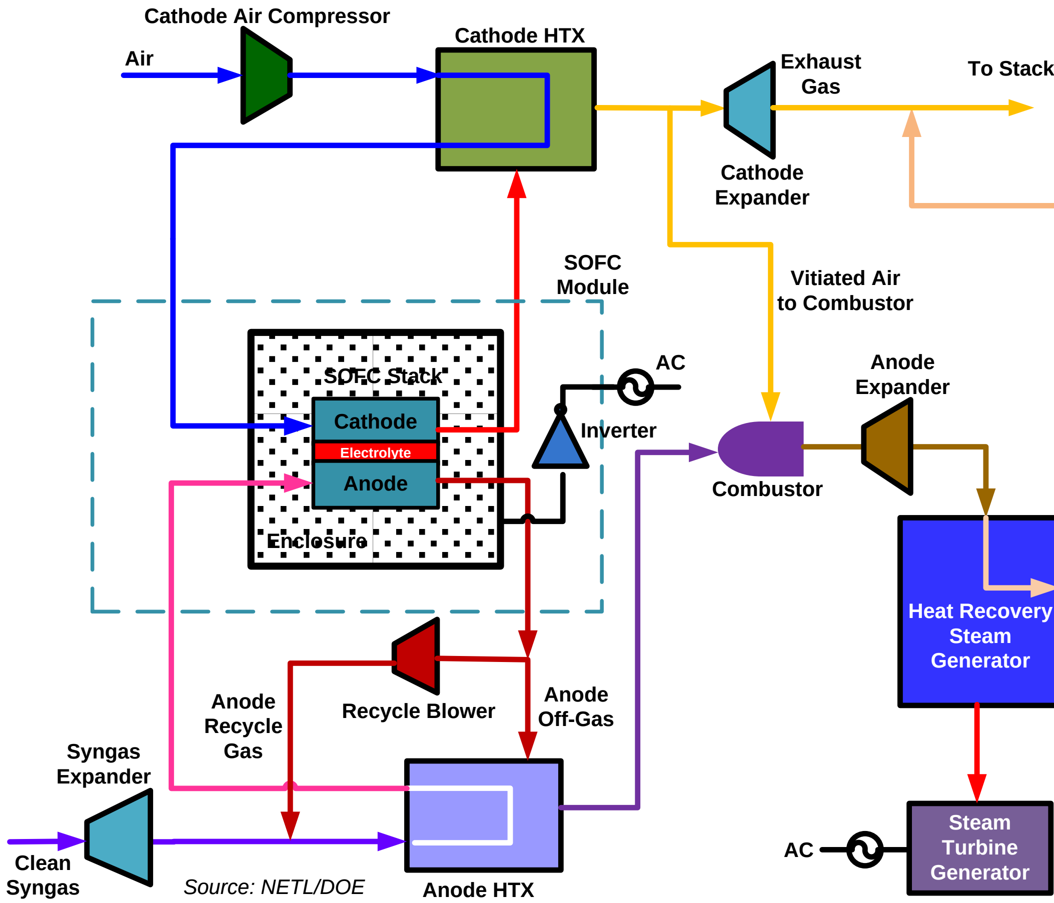
While the majority of the IGFC plant components remained unchanged, the CPU was eliminated and the SOFC power island was modified appropriately to optimize the system for heat recovery and work extraction. The ASU burden was reduced by routing the vitiated air from the cathode exhaust to provide the O_2 to the combustor. The modifications to the SOFC power island are shown in Exhibit 5-1 for the atmospheric SOFC case. For the pressurized SOFC case, the modified configuration, shown in Exhibit 5-2, includes an additional anode expander to expand the combustor off-gas to atmospheric pressure before heat recovery.

Exhibit 5-1 Power island configuration for atmospheric SOFC cases without CCS



Source: NETL/DOE

Exhibit 5-2 Power island configuration for pressurized SOFC cases without CCS



Source: NETL/DOE

The performance and cost of salient IGFC systems without carbon capture are summarized in Exhibit 5-3. An IGFC system with the CoP gasifier technology along with an atmospheric SOFC, both of which are projected to become commercially competitive in the GHG regulation time-frame, can be designed to meet the proposed 1100 lb/MWh$_{GROSS}$ GHG limit with 14 percent NG injection. Although the CoP gasifier case with the pressurized SOFC and the catalytic gasifier cases, with both atmospheric and pressurized SOFC, clearly exceed the proposed GHG limit, they reflect advanced futuristic systems in the NETL transformational time frame, and may be subjected to a more stringent GHG limit. In particular, even the emissions of the pressurized catalytic gasifier case are higher, albeit slightly, than the emissions of present NGCC systems.

Exhibit 5-3 Performance and cost of salient IGFC cases without CCS

Gasifier	CoP			Catalytic	
SOFC Operating Condition Operating Pressure (psia)	Atm. 15.6	Atm. 15.6	Press. 285	Atm. 15.6	Press. 285
Case	1-4	1-6	2-1	3-4	4-1
NG Injection (% of total input thermal energy)	-	14.2	-	-	-
PERFORMANCE					
Net Power (MWe)	551.0	550.1	550.2	550.0	550.1
Net Electric Efficiency, HHV (%)	**47.5**	**54.2**	**49.6**	**60.2**	**64.8**
COST					
Total Plant Cost (TPC) (1000$)	1,280,642	1,090,041	1,281,806	854,820	972,238
Total Overnight Cost (TOC) (1000$)	1,582,392	1,346,221	1,582,203	1,067,048	1,206,471
Total As-Spent Cost (TASC) (1000$)	1,803,927	1,534,692	1,803,712	1,216,434	1,375,377
Cost of Electricity ($/MWh)					
Variable COE	31.6	31.1	34.0	29.0	30.5
Fuel	21.1	20.9	20.2	16.6	15.4
Variable O&M	10.5	10.2	13.8	12.4	15.1
Fixed O&M	11.1	10.2	11.1	10.3	10.9
Capital Charges	47.9	40.9	48.0	32.4	36.6
Total First Year COE	**90.7**	**82.2**	**93.2**	**71.7**	**78.0**
CO_2 Emissions					
lb/MWhgross	**1252**	**1083**	**1060**	**1018**	**880**
lb/MWhnet	1432	1193	1373	1078	1002

6 Conclusions

The results of a pathway study for coal-based, IGFC power systems with CCS were presented in this report. They represent the potential future benefits of IGFC technology development by quantifying the performance and cost benefits for a series of projected gains made through the development of advanced technologies for improvements in plant operation and maintenance. In addition, the effectiveness of an IGFC system without CCS in meeting the proposed GHG emissions were discussed using results from select cases.

The overall plant performance and costs estimates of two parallel pathways of IGFC development were considered. The first pathway utilized conventional coal gasification technology, and featured the CoP E-Gas™ gasifier, while the second pathway utilizes an advanced, catalytic coal gasification technology projected to produce syngas having high methane content. The IGFC systems analyzed featured both atmospheric SOFC as well as pressurized SOFC configurations. The IGFC systems with CCS included a CO_2 purification unit that met EOR specifications for the product CO_2 stream.

The variable costs for the SOFC stack were estimated based on a model of stack operational scenarios, which effectively compensated for stack degradation. An SOFC stack with additional 10 percent area and an average stack replacement period of 7.3 years was used based on the model results for the Nth of a kind IGFC unit in the cost of electricity calculations. The costs are reported in 2011$ and are based on the NETL goal of $225/kW for the SOFC stack.

The 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.
- Significant reduction of SOFC performance degradation rate, in addition to enhancement of SOFC electrical performance, is required for the IGFC system to be economically competitive with other technologies.
- The IGFC consumes 50 percent less raw water than even the water-economical NGCC system with CCS.
- The cost of electricity with the IGFC system is projected to be significantly lower than IGCC and PC systems with CCS, while being competitive with NGCC systems with CCS (@ $6.13/MMBtu) and exceed the NETL goals. The IGFC systems using a catalytic coal gasifier and atmospheric-pressure SOFC result in the greatest benefit with a COE that is less than even the NGCC system with CCS @ $6.13/MMBtu). This IGFC system warrants the 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.
- A pressurized-SOFC provides no cost benefit over systems operating with an atmospheric-pressure SOFC. However, 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. In particular, the cost of the pressure vessel is an important parameter that needs further refinement. This IGFC configuration requires development of the pressurized-SOFC 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 comparable to that of an NGCC system with CCS @ $6.13/MMBtu). IGFC with natural gas injection can have a 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.
- The COE of all the IGFC plants considered herein are still higher than a NGFC system with capture (COE ~ $68/MMBtu @ $6.13/MMBtu).
- The natural gas injection case also represents an IGFC configuration that can meet the proposed EPA 1100 lb/MWh$_{GROSS}$ limit on GHG emissions for a coal power plant without any need for the CCS equipment. It is particularly attractive as it utilizes conventional gasification and SOFC technologies, which are likely to be developed within the regulation time-frame, unlike the pressurized SOFC and catalytic gasifier plants. However, the COE of this IGFC plant is still higher than an NGCC system without CCS (@ $6.13/MMBtu).

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.

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Wally Shelton
Walter.Shelton@netl.doe.gov

Dale Keairns
Dale.keairns@contr.netl.doe.gov

www.netl.doe.gov
Pittsburgh, PA • Morgantown, WV • Albany, OR • Sugar Land, TX • Anchorage, AK
(800) 553-7681