



NATIONAL ENERGY TECHNOLOGY LABORATORY



Integrated Gasification Fuel Cell Performance and Cost Assessment

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1.0 SUMMARY

This report presents an analysis of the levelized cost of electricity (LCOE) from two integrated coal gasification fuel cell (IGFC) power plants that use solid oxide fuel cell (SOFC) technology to convert syngas to electricity. The two plants are differentiated primarily by the Fuel Cell Island technology used, one representing an atmospheric-pressure fuel cell power island, and the other a pressurized fuel cell power island combined with gas-expander power recovery. This analysis is part of an overall effort to compare the economics of fuel cell-based systems in central station and distributed generation applications.

The integrated gasification fuel cell (IGFC) system includes an advanced, catalytic, high-methane gasifier (29 vol% in dry syngas) and syngas cleaning to achieve low sulfur levels in the fuel cell feed gas (0.1 ppmv). The fuel cell performance is based on what is projected to be achievable based on sub-scale testing (current density of 500 mA/cm², potential of 0.8 volts, lifetime of 5 years). The planar fuel cell design provides separation of carbon dioxide (CO₂) as a natural part of operation. Oxy-combustion of the fuel cell effluent enables nearly complete CO₂ capture (99+%). Unit operations were integrated in an ASPEN modeling framework to achieve an optimized system.

The fuel cell system cost is based on the Department of Energy (DOE) program goal adjusted to an installed cost in 2007 dollars of 700 \$/kW AC output from the stack. Capital cost estimates for other major unit operations are scaled from the costs reported in a range of studies.

The results of the analysis are summarized in Table 1 with comparison to an integrated gasification combined cycle (IGCC) with a conventional combustion turbine, taken from the 2007 NETL Baseline Studies. Note that the capital cost per kW of capacity is lower for the IGFC platform. The fuel cell system is more expensive than a combustion turbine but that expense is counterbalanced by the decrease in the unit cost of upstream equipment due to the higher IGFC system efficiency. The IGFC platform's capability to separate CO₂ results in its lower CO₂ emissions per kWh. Table 1 shows that the fuel cell platform offers nearly 100% CO₂ capture. This has a further positive effect on LCOE in scenarios with a cost for CO₂ emissions.

The assumptions and methodologies utilized in development of the results presented in Table 1 are the focus of the remainder of the report.

Table 1 – IGFC Comparison to IGCC with CO₂ Capture and Sequestration

	Conventional IGCC	Atm-Pressure IGFC Plant	Pressurized IGFC Plant
Efficiency (% , HHV)	32.5	49.4 ¹	56.2 ²
CO ₂ Emissions (lb CO ₂ /MWh net)	206	3	3
Water Usage (gal/MWh net)	494	193	172
Capital cost (2007\$/kW)	2,400	2,000	1,800
LCOE (cents/kWh)	10.2	8.8	7.9

¹ The efficiency without carbon dioxide compression is 52.6%.

² The efficiency without carbon dioxide compression is 60.1%

2.0 PLANT BASIS AND ASSUMPTIONS

The power plant design basis and key technology assumptions are listed in Table 2. The IGFC plants utilize both conventional and advanced technologies to achieve a 253 MWe output. The design-basis coal is Illinois #6. The plant is configured as a single-train plant having no spare major components. The plant site characteristics and the basis for water consumption estimates are the same as those used in the 2007 NETL Baseline Studies.

Table 2 – Plant Design Basis and Estimates

	Atm-Pressure IGFC	Pressurized IGFC
Ambient Conditions	ISO	ISO
Coal	Illinois #6	Illinois #6
Coal heating value (Btu/lb, as received HHV)	11,666	11,666
Coal heating value (Btu/lb, dry HHV)	13,126	13,126
Gasifier type	Advanced Catalytic steam-O ₂	Advanced Catalytic steam-O ₂
exit temperature (°F)	1300	1300
exit pressure (psia)	975	975
carbon loss with ash (% coal carbon)	5	5
gasifier cold gas efficiency (% HHV)	90.7	90.7
syngas methane content (dry vol%)	29.3	28.6
ASU technology	Cryogenic	Cryogenic
oxidant O ₂ content (vol%)	95.0	99.5
Gas Cleaning technology	Conventional Dry Gas Cleaning	Advanced Humid Gas Cleaning
Syngas Expander		
outlet pressure (psia)	26	275
efficiency (adiabatic, %)	90	90
Fuel Cell Stack		
inlet temperature (°F / °C)	1202 (650)	1202 (650)
outlet temperature (°F / °C)	1472 (800)	1472 (800)
outlet pressure (psia)	20	270
fuel single-step utilization (%)	70	70
fuel overall utilization (%)	82	85
anode gas recycle method	Syngas inductor	Syngas inductor
Nernst potential (estimation basis: outlet composition and temperature)	0.84	0.92
stack over-potential (mV)	40	50
current density (mA/cm ²)	500	500
Rectifier DC-to-AC efficiency (%)	97	97
Oxy-combustor outlet O ₂ (vol%)	1	1
Steam bottoming cycle conditions	subcritical	none

Both plants use an advanced, catalytic coal gasifier operated with oxygen and steam injection, with oxidant produced using conventional cryogenic air separation technology. The atmospheric-pressure SOFC IGFC plant uses conventional dry gas cleaning technologies for syngas cleaning, while the pressurized IGFC plant uses advanced humid gas cleaning technologies. The atmospheric-pressure SOFC IGFC plant is based on nearer-term, more mature, atmospheric-pressure SOFC cell technology, having the operating characteristics and performance assumptions listed in Table 2. In contrast, the pressurized IGFC plant utilizes more advanced, less mature, pressurized SOFC cell technology. Both plants use anode off-gas oxy-combustion to utilize the remaining fuel energy in the anode off-gas and to convert the gas into a CO₂-rich stream. This stream contains 1 vol% excess O₂ and is assumed to be acceptable for sequestration without further purification. It is dehydrated and compressed to 2215 psia for sequestration, also sequestering the coal contaminants that remain after syngas cleaning (small amounts of SO₂, HCN, etc.). The plant technologies are described in more detail below.

3.0 ATMOSPHERIC-PRESSURE SOFC IGFC PLANT DESCRIPTION

The atmospheric-pressure IGFC plant incorporating CO₂ capture is described here, with discussion divided into the major plant sections: the gasifier section, the air separation unit, the gas cleaning section, the fuel cell island, and the steam cycle. Figure 1 is a simplified flow diagram of the plant.

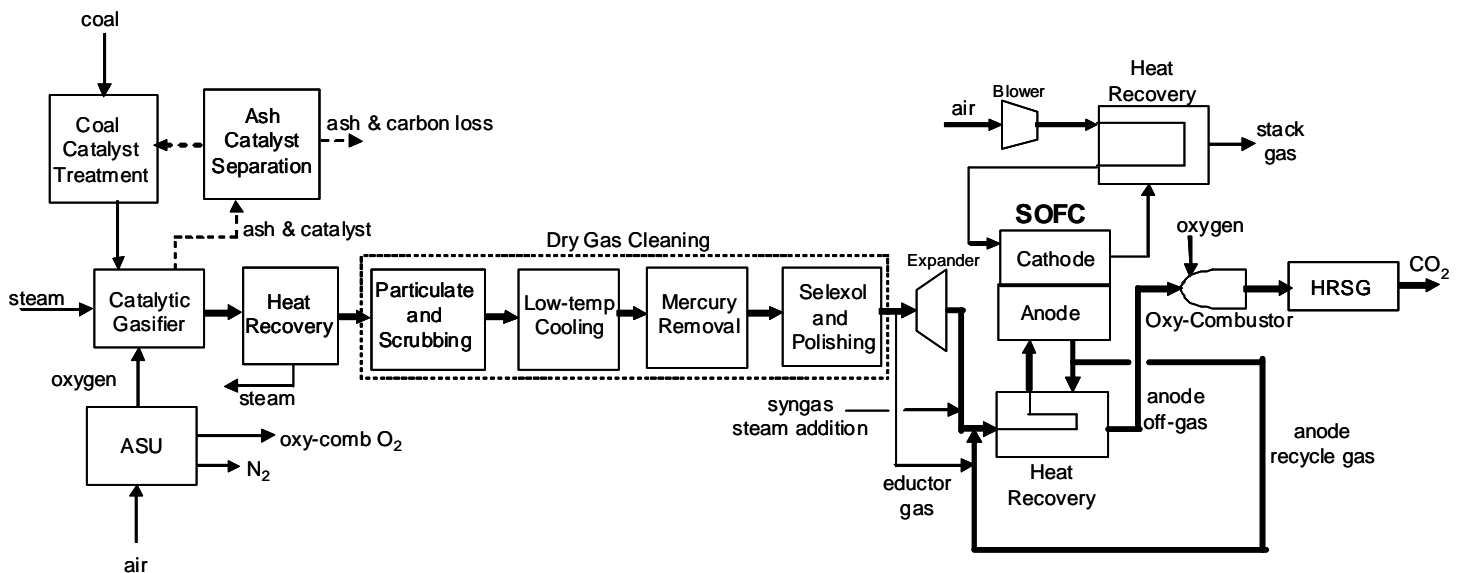
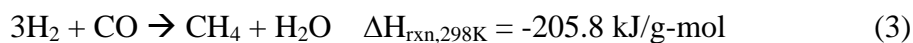
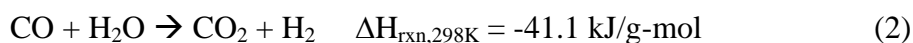
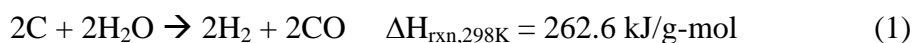


Figure 1 – Simplified Flow Diagram of the Atm-Pressure SOFC IGFC Plant

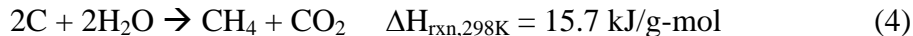
3.1 Gasifier Island

The gasifier modeled in this application is based on the concept of catalytic coal gasification to produce high methane content syngas at a relatively low operating temperature, which benefits both the gasifier efficiency and the SOFC performance. It is a hypothetical gasifier that has not yet been developed.

Coal gasification for methane as a primary reaction product is described by the following three reactions:



Reactions 1 through 3 can be combined to yield the following overall reaction:



The overall reaction (Reaction 4) requires a small amount of endothermic heat, therefore minimizing capital equipment for heat exchange and recovery. The extent of this reaction is limited by thermodynamics and the catalytic gasifier operates at conditions that could produce more methane than any existing conventional gasifier.

Catalyst is added to the coal as an aqueous solution, and the mixture is dried before being fed to the fluidized bed gasifier. The gasifier operates at 1000 psia and 1300 °F, and is fluidized by high-pressure steam. Oxygen (95% purity by volume) from a cryogenic air separation unit is fed to the gasifier to provide the endothermic reaction heat required for the gasification reactions. A mixture of ash, unburned carbon, and catalyst flows from the gasifier and enters a water scrubber, where catalyst separation occurs. Based on extensive development work by Exxon on a different type of catalytic gasifier, approximately two-thirds of the catalyst can be recovered; the remaining third is lost with the ash and must be replenished^{vi}. The reaction products, H₂, H₂O, CO, CO₂, CH₄, H₂S and other contaminants, are assumed to approach chemical equilibrium and are cooled in a waste heat recovery unit generating saturated, high-pressure steam. The steam recovered is used in the evaporation section of the steam-cycle heat recovery steam generator (HRSG).

3.2 Air Separation Unit

The conventional, cryogenic air separation unit is designed to produce 95% pure oxygen for use in the gasifier, Claus plant, and oxy-combustor. Ambient air is compressed to 190 psia, and is then concentrated in oxygen in the cryogenic air separation unit. The oxygen product is withdrawn from distillation columns as a liquid and is pressurized by a cryogenic pump. The pressurized liquid oxygen is then vaporized against high-pressure air feed before being warmed to ambient temperature. The gaseous oxygen is fed to a centrifugal compressor with intercooling between each stage of compression. The compressed oxygen is then fed to the gasifier, Claus plant, and oxy-combustor.

3.3 Gas Cleaning System

Particulate is removed from the raw syngas exiting the gasifier using a cyclone collector and a candle filter system. Halides (primarily HCl), and ammonia are removed from the syngas by cooling and condensate scrubbing in the conventional dry gas cleaning system, producing a sour water stream.

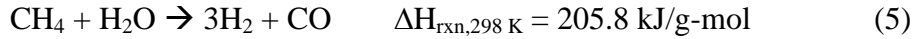
A sour water stripper removes NH₃, H₂S, and other impurities from the sour water stream. The sour gas released is sent to the sulfur recovery unit where the NH₃ is destroyed and the H₂S is converted to elemental sulfur. The remaining sour water contaminants (e.g., chlorides) are assumed to be treated at a wastewater treatment plant.

Sulfur is removed from the syngas in a two-step process. A single stage Selexol unit removes the bulk of the H₂S present in the cooled syngas, while a ZnO sorbent-based guard bed operating at about 700°F performs deep cleaning, bringing sulfur down to levels permitted by the SOFC

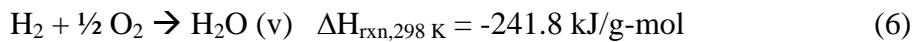
(<100 ppbv total sulfur). The cleaned syngas proceeds to the fuel cell island, while the acid gas stream is sent to the Claus plant for sulfur recovery.

3.4 Fuel Cell Island

This analysis assumes that methane will reform with steam within the SOFC to produce hydrogen (as opposed to direct anode oxidation) according to the endothermic methane reforming reaction:



The potential for oxidation of CH₄ is less than that for H₂, as shown by the higher open circuit voltage of hydrogen oxidation. This endothermic reaction can serve as a heat sink to the exothermic oxidation of hydrogen (Reaction 6), therefore lowering the stack thermal management requirement:



The rate of reforming of methane at the low operating temperature of the cell may not be sufficiently rapid, though, and the potential benefits need to be established experimentally. Sulfur free syngas from the Selexol unit serves as the fuel for the SOFC. Since the fuel cell requires the syngas to be near atmospheric pressure (with adequate head to overcome unit pressure losses), an expansion step is necessary to lower the incoming pressure. Preheated syngas is expanded to 20 psia to accommodate the atmospheric-pressure fuel cell. The introduction of low-pressure steam from the steam cycle occurs prior to the expansion step in order to achieve a hydrogen-to-carbon atomic ratio of 4:1 (to prevent anode coking). The SOFC inlet and outlet temperatures for this analysis were assumed to be 1,202 °F (650 °C) and 1,472 °F (800 °C), respectively. The anode fuel gas temperature is increased to 1,202 °F by mixing with recycled anode effluent gas and by recuperative heat exchange with a portion of the anode effluent.

Air is compressed to 24 psia by the cathode air blower. Oxygen is consumed as needed from this stream to complete the oxidation of hydrogen, Reaction 6. The remainder of the cathode air, consisting of nitrogen and unused oxygen, acts as a heat sink to aid in thermal management of the fuel cell stack, preventing the maximum temperature of 1,472 °F (800 °C) from being exceeded.

The fuel cell voltage was estimated by first calculating the Nernst potential based on the cell outlet operating temperature and the outlet anode gas and cathode gas compositions. The cell losses were then estimated from correlations for activation, ohmic, and concentration overpotentials. The fuel cell operating parameters used to develop this case are shown in Table 3.

Table 3 – Atm-Pressure Fuel Cell Operating Parameters

Parameter	Value
Nernst Potential, V	0.84
Operating Voltage, V	0.8
Anode Gas Recycle, %	50
Fuel Utilization (Single Pass / Recycle), %	70 / 82
Current Density, mA/cm ²	500

Upon exiting the fuel cell, the cathode effluent is used to preheat cathode inlet air, and the remaining heat is recovered in the economizer section of the HRSG.

The anode effluent is used for anode inlet preheating as described above. The resulting temperature of the anode effluent stream (495 °F) is not sufficiently high to generate any substantial amount of steam from which power can be derived. Also, there is still useful chemical energy remaining (8% H₂, 4% CO) which should be recovered, if possible.

Combustion with air is not an option since it is desired to keep the anode effluent rich in CO₂ and H₂O; this approach would result in dilution of the combustion products with nitrogen. However, combustion in the presence of purified oxygen from the air separation unit (oxy-combustion) will recover the chemical energy remaining in the syngas, and raise the stream temperature, while not diluting the resultant product stream with appreciable nitrogen.

Cooled anode effluent is combusted in the presence of 95% pure oxygen, recovering the unused fuel and raising the temperature to 1,742 °F. Based on discussions with researchers in the field of oxy-combustion technology, it is assumed that an excess of 1.0% O₂ (volume basis) in the product stream is required in order to recover all the fuel heating value initially present. This oxy-combustion product stream is used in the evaporator to generate supercritical steam, and in the process is cooled to 600 °F. This low quality heat is recovered in the economizer, resulting in further cooling to 150 °F. The remaining moisture is condensed out by heat exchange with cooling water, as well as by a conventional water absorber, further concentrating the stream in CO₂ (94% by volume). The purified CO₂ is finally compressed to 2,215 psia in a multi-stage compressor with intercooling to 100 °F. Although standard CO₂ pipeline specifications only allows for 100 ppmv O₂, the oxygen content of the purified CO₂ stream for this case is 3% (30,000 ppmv). Reduction of excess oxygen in oxy-combustion systems is an active research area, and this issue will be addressed further.

3.5 Steam Cycle

The subcritical steam conditions are 1,050 °F and 1,800 psia, with steam reheat to 1,050 °F. Water is pressurized to 2,135 psia by the boiler feedwater pump. The pressure loss through each feedwater heater was assumed to be approximately 3% of the incoming pressure. Therefore, the boiler feedwater pump must sufficiently pressurize water to overcome these losses, and still maintain 1,800 psia at the high-pressure (HP) steam turbine inlet.

Pressurized boiler feedwater is heated in the economizer to 375 °F, by recovering low quality heat from the cathode discharge (prior to exhaust out the stack) and anode effluent (prior to water knockout for CO₂ compression). The heat required for steam evaporation is from the gasifier's convective cooler. The evaporated steam is heated to 575 °F. The convective cooler also provides the energy required for steam reheat prior to the intermediate pressure (IP) steam turbine. The steam superheater energy is provided by the oxy-combustion products. Steam leaves the superheater at 1,050 °F and 1,800 psia, ready for expansion through the HP steam turbine.

The superheated steam flow is expanded to 715 psia in the HP steam turbine. Steam at 915 psia is extracted from the HP turbine to achieve the 4:1 hydrogen to carbon atomic ratio at the anode inlet. The HP steam discharge is reheated to 1,050 °F by heat recovery with the gasifier's convective cooler. The steam is expanded to 142 psia in the IP steam turbine, and finally expansion down to the condenser pressure of 1 psia occurs in the low-pressure (LP) steam

turbine. Low-pressure (65 psia) steam is extracted for use in the Selexol process (for deep cleaning of solvent), as well as for sour water stripper duty. One percent of the LP steam turbine discharge is blown down to prevent buildup of chlorides and dissolved solids. Any heat remaining in the LP steam turbine discharge is rejected at the condenser, producing water at 102 °F and 1 psia. Makeup water is added to the condenser as needed to maintain constant water flow throughout the steam cycle.

Cooling water is heated from 60 °F to 80 °F by absorbing the heat of condensation given up by LP steam at the condenser as well as other cooling duties in the plant. The heated water is cooled back to 60 °F in a mechanical-draft, counterflow wet cooling tower. Ambient air at 59°F and 60% relative humidity is pressurized to 0.5" H₂O to absorb the heat gained by the cooling water at the condenser. Evaporation of the heated cooling water simultaneously increases the air temperature to 75 °F and 90% relative humidity, and cools the water to 60 °F. Since a portion of the cooling water is lost to evaporation, makeup water at 50 °F is supplied to sustain a constant water flowrate at the condenser. In order to maintain acceptable levels of chlorides and total dissolved solids in the cooling loop, it is assumed that 0.5% of the total cooling water flow is blown down continuously. Makeup water is also added to account for the blowdown water loss.

4.0 PRESSURIZED-SOFC IGFC PLANT DESCRIPTION

The pressurized-SOFC IGFC plant incorporating CO₂ capture is described in the following. The coal gasifier section and the air separation section utilize the same type of technologies that were applied for the atm-pressure SOFC IGFC plant. There is no steam bottoming system in the pressurized IGFC plant, with turbine expanders providing power recovery from the SOFC off-gases. Advanced humid gas cleaning technology and a pressurized fuel cell power island are applied in this configuration. Figure 2 is a simplified flow diagram of the plant.

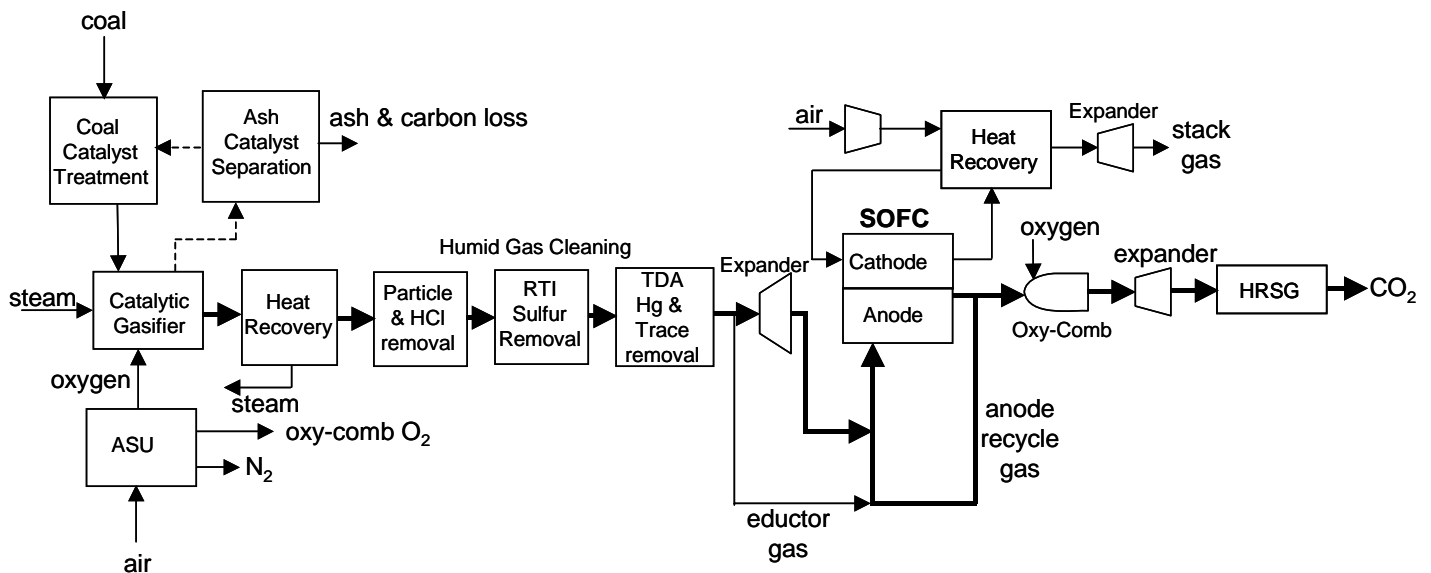


Figure 2 – Simplified Flow Diagram of the Pressurized-SOFC IGFC Plant

4.1 Gas Cleaning System

Following catalytic coal gasification and raw syngas heat recovery, the raw syngas is subjected to advanced humid gas cleaning in the series of cleaning steps described below. The raw syngas issued from the gasifier is cooled to 800°F using a convective heat exchanger generating saturated high-pressure steam.

Particle Removal

The raw syngas passes through a high-efficiency cyclone to separate out the bulk of the entrained slag, and this slag is recycled to the gasifier via catalyst-slurry preparation. The syngas then passes through a barrier filter (using ceramic or metal filter elements). The

outlet particulate content is expected to be less than 0.1 ppmw. The barrier filter technology has been demonstrated at three IGCC plants and has near-commercial status. Several volatile contaminant species in the raw syngas (e.g., alkali metals) condense out of the syngas onto the captured particulate within the barrier filter.

Halide Removal

Nahcolite or trona particles, sized at minus-325 mesh, are injected into the raw syngas before the barrier filter at a sodium-to-halide molar feed rate of 6-times the stoichiometric halide (primarily HCl) molar flow rate. It is estimated that 99.5% of the HCl is removed by the filter cake's nahcolite particulate content by reaction with the nahcolite particles, reducing the HCl content in the quenched syngas to about 1 ppmv. The separated nahcolite residue and coal ash stream is disposed of as a non-hazardous waste. Nahcolite is assumed to have a sodium bicarbonate content of 80 wt%, the remainder being inert material. The nahcolite is a cheap, natural mineral, having a cost of about \$50/ton, and sorbent regeneration is not warranted. This approach to halide removal has been demonstrated at a pilot scale by Siemens^{vii}.

Desulfurization

The syngas passes through an RTI transport desulfurizer coupled with an entrained sorbent regenerator. Total sulfur content is reduced to about 10 ppmv using an RTI, ZnO-based sorbent, effectively removing H₂S, COS and other sulfur species. The ZnO-based sorbent is assumed to have a ZnO-to-titanium molar ratio of 1.0. The desulfurizer reactions heat the off-gas to about 940°F and the outlet syngas, with entrained sorbent particles, passes through another barrier filter. Additional sulfur removal will occur within this second barrier filter, but no credit is taken for this further sulfur reduction. The captured sorbent particles are recycled to the sorbent regenerator. The air-fired regenerator operates with an exit temperature of about 1166°F, producing a regenerator gas having about 14.9 vol% SO₂. The regenerator off-gas passes through a small barrier filter that removes entrained sorbent particles and recycles them to the desulfurizer. The sorbent makeup rate is assumed to be about 0.005 lb per lb-mole of captured sulfur based on the assumption of long, cyclic sorbent life with low particle attrition and elutriation losses. Nitrogen is used as the aeration gas in the desulfurization unit. This technology has been demonstrated at pilot-scale. The transport desulfurizer system is followed by a conventional ZnO guard bed that polishes the syngas sulfur content down to less than 0.1 ppmv total sulfur. The commercial ZnO polishing-sorbent is periodically replaced.

Sulfur Recovery

The regenerator SO₂ off-gas is contacted with a small portion of clean syngas. The RTI Direct Sulfur Recovery Process (DSRP) is used to directly reduce about 99% of the SO₂ to elemental sulfur in a catalytic reactor having internal heat transfer cooling surface, approaching isothermal reactor behavior at 1100°F. The DSRP off-gas is cooled and the elemental sulfur is condensed from the stream at 290°F. A conventional, caustic tail gas cleaning process is used to capture about 98% of the remaining sulfur species to minimize sulfur emissions. The tail gas is recompressed and recycled to the clean syngas steam. This technology has been tested at field-scale and consumes about 1% of the clean syngas.

Mercury Removal

The syngas is cooled to 500°F, raising steam for use in the plant. The syngas enters a set of parallel, fixed bed adsorbers operated in a pressure-swing configuration. Regenerative mercury sorbent developed by TDA Research Inc. is used to remove more than 90% of the syngas mercury content as well as other trace contaminants of concern (e.g., As and P). The Hg removal-stages operate at 500°F, and the sorbent regenerator-stages operate at about 575°F in nitrogen to drive off and collect the mercury metal by condensation as a byproduct. The mercury sorbent is assumed to have a capacity of 1 lb Hg per 1000 lb of adsorbent. TDA has successfully tested this technology at pilot-scale gasifier slip-stream facilities.

4.2 Fuel Cell Island

Sulfur free syngas from the Selexol unit serves as the fuel for the SOFC. The fuel cell requires the syngas to be near the pressure of the gas expander that will recover power from the anode off-gas. Preheated syngas is expanded to 275 psia to accommodate the pressurized fuel cell. The SOFC inlet and outlet temperatures for this analysis were assumed to be 1,202 °F (650 °C) and 1,472 °F (800 °C), respectively. The expanded syngas acts as the anode fuel gas and is preheated to 1,202 °F by mixing with recycled anode effluent gas and by recuperative heat exchange with a portion of the anode effluent. The recycle of the anode off-gas provides all of the water vapor needed to protect against carbon deposition. Ambient air is compressed to 280 psia in the cathode air compressor. This compressed air provides the oxidant necessary for hydrogen oxidation, as well as cooling air for the fuel cell stack.

The fuel cell voltage was estimated by calculating the Nernst potential based on the outlet anode gas operating temperature, and outlet anode fuel and cathode gas compositions. The elevated pressure also contributes to the high Nernst potential of 0.92 V. The cell overpotential is assumed to be 50 mV for this advanced cell technology. The fuel cell operating parameters used to develop this case are shown in Table 4.

Table 4 – Pressurized Fuel Cell Operating Parameters

Parameter	Value
Nernst Potential, V	0.92
Operating Voltage, V	0.87
Anode Gas Recycle, %	59
Fuel Utilization (Single Pass / Recycle), %	70 / 85
Current Density, mA/cm ²	500

Upon exiting the fuel cell, the cathode effluent is used to preheat cathode inlet air, and the cathode gas is expanded through a gas expander assumed to be placed on the same shaft with the Cathode air compressor.

The anode effluent is used for anode inlet preheating as described above. It is combusted in the presence of purified oxygen from the air separation unit (oxy-combustion), releasing the chemical energy remaining in the syngas, to be expanded through a gas expander.

Cooled anode effluent is combusted in the presence of 99.5% pure oxygen, recovering the unused fuel and raising the temperature to 1,742 °F. An excess of 1.0 vol% O₂ in the product stream is used in order to recover all the fuel heating value. The oxy-combustion gas is expanded to 30 psia. The remaining moisture is condensed out by heat exchange with cooling water, as well as by a conventional water absorber, further concentrating the stream in CO₂ (87% by volume). The purified CO₂ is finally compressed to 2,215 psia in a multi-stage compressor with intercooling to 100 °F. Although standard CO₂ pipeline specifications only allows for 100 ppmv O₂, the oxygen content of the purified CO₂ stream for this case is 3% (30,000 ppmv).

5.0 PERFORMANCE

The IGFC plants described in the previous sections were modeled using AspenPlus, producing converged, steady-state material and energy balances. Major process stream flow rates, temperatures and pressures in each of the plants are listed in Table 5. Tables 6 and 7 list the compositions of the major process streams, including contaminant contents.

Table 5 – Major Process Stream Flows and Conditions

Stream	Atm-pressure SOFC IGFC Plant			Pressurized SOFC IGFC Plant		
	Flow lb/hr	Temperature °F	Pressure psia	Flow lb/hr	Temperature °F	Pressure psia
Coal	150,000	59	14.7	124,352	59	14.7
Gasifier oxidant	35,027	187	1050	34,950	272	1050
Gasifier steam	217,500	904	1100	179,652	580	1100
Raw syngas	383,202	1300	975	324,486	1300	975
Clean syngas	358,956	790	913	293,354	813	819
Anode feed gas	923,021	1202	22	1,057,101	1202	272
Anode off-gas	564,076	495	17	507,083	1427	267
Cathode air	2,024,759	59	14.7	1,143,244	59	14.7
Cathode off-gas	1,819,628	1472	22	963,554	1472	284
Oxy-combustor oxidant	53,451	90	145	38,371	287	280
CO ₂ gas	342,208	224	2215	322,067	100	2215

Table 6 – Atm-Pressure IGFC Plant Major Stream Compositions

	Raw syngas	Clean syngas	Anode feed gas	Anode Off-gas	CO ₂ gas
	vol%	vol%	vol%	vol%	vol%
H ₂	15.0	16.0	11.0	8.0	0.0
CH ₄	17.0	18.0	8.0	0.0	0.0
CO	5.0	5.0	5.0	4.0	0.0
CO ₂	21.0	22.0	26.0	29.0	95.0
H ₂ O	42.0	38.0	50.0	58.0	0.0
N ₂	1.0	1.0	1.0	0.0	2.0
Ar	0.0	0.0	0.0	0.0	0.0
O ₂	0.0	0.0	0.0	0.0	3.0
	ppmv	ppmv	ppmv	ppmv	ppmv
H ₂ S	6,800	0.05	0.05	0.04	0
COS	130	0	0	0	0
SO ₂	0	0	0	0	4
NH ₃	600	20	20	10	0
HCN	70	0	0	0	0
HCl	735	0	0	0	0
Hg (ppbv)	4	0.3	0.3	0.3	0

Table 7 – Pressurized IGFC Plant Major Stream Compositions

	Raw syngas	Clean syngas	Anode feed gas	Anode Off-gas	CO ₂ gas
	vol%	vol%	vol%	vol%	vol%
H ₂	14.8	13.9	9.0	6.5	0.0
CH ₄	17.5	17.8	6.0	0.0	0.0
CO	5.5	5.2	4.0	3.4	0.0
CO ₂	21.8	22.7	27.7	30.3	87.3
H ₂ O	38.9	35.2	48.9	55.9	0.0
N ₂	0.4	5.1	4.2	3.8	9.8
Ar	0.02	0.08	0.06	0.06	0.2
O ₂	0.0	0.0	0.0	0.0	2.7
	ppmv	ppmv	ppmv	ppmv	ppmv
H ₂ S	6,800	0.05	0.05	0.04	0
COS	130	0	0	0	0
SO ₂	0	0	0	0	4
NH ₃	600	600	210	20	0
HCN	70	70	20	0	0
HCl	735	3	3	2	6
Hg (ppbv)	4	0.3	0.3	0.3	0

Overall performance for the plant is shown in Table 8. For the atm-pressure IGFC plant, the net plant power is only 14 MW greater than the SOFC AC-power output, so the other generators in the plant (the syngas expander and the steam cycle) essentially counter the

plant auxiliary losses. The cathode blower and CO₂ compression train combined account for 66% of the entire parasitic power load (30% and 36%, respectively). Similarly, for the pressurized IGFC plant, the net plant power is only 32 MW greater than the SOFC AC-power output. Here, the cathode compressor and CO₂ compression train account for about 84% of the auxiliary power losses. While the pressurized IGFC plant has no steam cycle, it has significant power output resulting from the oxy-combustor expander. Note that the CO₂ compression energy is higher for the pressurized IGFC plant than for the Atm-pressure IGFC plant due to the higher inert gas content in the CO₂ stream of the Pressurized IGFC plant.

Table 8 – Coal IGFC Combined Cycle Performance Summary

	Atm-Pressure IGFC Plant	Pressurized IGFC Plant
Syngas Expander (kW)	25,425	8,375
SOFC	239,150	221,060
Oxy-combustor Expander	0	66,090
Cathode Gas Expander	0	51,963
Steam Cycle	34,895	0
Total	299,470	347,488
Auxiliary Power Summary		
Cathode Blower/Compressor (kWe)	13,760	60,823
Cooling Water Pump	75	50
ASU Compressor	9,005	8,453
ASU O ₂ Compressor	960	3,546
Boiler Feedwater Pump	1,185	200
Coal Handling	138	168
Coal Milling	699	200
Coal Feeding	226.5	254
Ash Handling	358.5	300
Condensate Pump	76.5	20
Cooling Tower Fan	260	250
Gas Cleaning Auxiliaries	1,048	1,691
Miscellaneous Balance of Plant	919	211
Transformer Loss	920	887
CO ₂ Compression	16,450	17,435
Total Auxiliaries	46,080	94,488
Net Total Power	253,390	253,000
Net Plant Efficiency (HHV)	49.4%	56.2%
Net Plant Heat Rate (Btu/kWh)	6,903	6,073

If these two IGFC plants were operated without CO₂ being captured and compressed for sequestration, the air separation unit capacity would be more than cut in half, the air separation unit power consumption and the oxygen compression power would be reduced significantly, and CO₂ compression power losses would be eliminated. The anode off-gas would be combusted using the cathode off-gas and the resulting product gas exhausted after energy recovery. This would push the atm-pressure IGFC plant efficiency up to 53.5% (HHV), and the pressurized IGFC plant efficiency to 61.8%.

An account of raw water usage is provided in Table 9. Water usage is defined as the difference between water consumption and water recovery. To reduce the total raw water usage, water is recovered and reused in the process where possible. Air cooling can also be used to replace water cooling for many duties in the plant.

Table 9 – Coal IGFC Combined Cycle Water Balance

	Atm-Pressure IGFC Plant	Pressurized IGFC Plant
Plant Water Consumption (lb/hr)		
Cooling Tower Losses	476,279	407,260
Steam Cycle Blow-down	1,532	0
Process Steam use	333,600	179,660
Total	811,411	586,920
Plant Water Recovery (lb/hr)		
CO ₂ -Stream Condensate	268,335	223,190
Syngas Condensate	135,062	0
Total	403,397	223,190
Raw Water Usage	408,013 lb/hr (193 gal/MWh)	363,730 lb/hr (172 gal/MWh)

6.0 IGFC PLANT COST

Total Plant Cost (TPC) and Levelized Cost-of-Electricity (LCOE) have been estimated for the 250 MW IGFC plants with carbon capture. Capital cost estimates for unit operations and operating cost estimates were derived from the 2007 NETL Baseline Studies¹ Integrated Gasification Combined Cycle (IGCC) cases where applicable. Cost of unit operations unique to the fuel cell platform were estimated using a range of sources.

6.1 Capital Cost

The approaches taken to estimate installed equipment costs for various process sections in the plant are summarized here. Table 10 provides detailed information on the cost estimation basis applied for each process section of the power plant and, where applicable, the capacity parameters used for scaling, the reference cost-year, and the reference reports consulted. Much of the costing is scaled from the 2007 NETL Baseline Studies. This scaling approach is approximate, but is expected to be sufficient for technology feasibility estimates.

Scaling from Prior Studies

For conventional process sections in the plant (i.e., coal handling, air separation unit, syngas heat recovery, gas cleaning, steam bottoming plant, flue gas heat recovery) prior, reported cost estimates for comparable conventional power plant process sections were scaled by extrapolation of the major process section capacity parameter for each section. For advanced plant sections (i.e., coal catalyst treatment, catalyst recovery, catalytic gasifier) equipment costs were estimated by scaling from costs reported in prior studies for similar advanced components.¹

For processing sections where multiple trains were employed, cost and capacity were reduced to a single train basis before scaling factors were applied. The general scaling formula used is as follows:

$$\text{Cost}_{\text{IGFC}} = \text{Cost}_{\text{Reference}} \times [\text{Capacity}_{\text{IGFC}} / \text{Capacity}_{\text{Reference}}]^S$$

¹ The catalytic coal gasifier cost was estimated by scaling available cost evaluation estimates for a conventional fluid bed gasifier, the KBR transport reactor. While these gasifier designs are significantly different, it is expected that this basis would provide a conservative cost estimate for the catalytic coal gasifier. The cost estimates for the higher cost elements associated with the oxygen-blown catalytic coal gasifier, the coal catalyst treatment system and the ash catalyst extraction system, were scaled from prior plant cost estimates made by Exxon for their catalytic coal gasification process applied for SNG production. While the specific catalyst to be used in an oxygen-blow catalytic gasifier has not been identified at this time, it was assumed for costing purposes that it would be likely to be a potassium-based material resulting in similar treatment equipment costs to those estimated for the Exxon catalytic gasifier process.

Scaling factor exponents, S , were estimated from tabulations of equipment cost-capacity scaling factors available in the open literature (e.g., Remer and Chai, "Design Cost Factors for Scaling-up Engineering Equipment," Chem. Eng. Progress, Aug 1990, p 77).

Other Cost Estimations

For some conventional process components (i.e., heat exchangers, compressors, blowers, expanders, CO₂ compression and dehydration) specific equipment cost correlations built into the ChemCad process simulator were applied.

For the SOFC module, the equipment cost was based on the DOE cost target set for that component of \$400/kW based on 2002-dollars. This represents the delivered equipment cost of the SOFC module, including the cost of its associated heat exchangers, blowers, controls and DC-to-AC rectifier. This 2002 equipment cost was escalated 23% to put it on a 2007-dollar basis based on an assessment of costs specific to manufacture of the SOFC module. The installed equipment cost was estimated to be 1.42 times the delivered equipment cost (consistent with the escalation of the gas turbine delivered costs, EPC costs and process and project contingencies in the NETL Baseline Studies, Case 1), resulting in an estimated installed equipment cost for the fuel cell system of \$700/kW AC power from the fuel cell.

Cost-Basis Year Adjustments

With the exception of the SOFC module, process section capital costs were escalated from their cost-basis year to the study year for this evaluation (2007) using the Chemical Engineering Equipment Cost Index.

Table 10 – Tabulation of Plant Section Cost Estimating Approach

Plant Section / Equipment	Cost Scaling Approach Applied
Gasifier, Coal, Ash, Catalyst Handling	
Coal & Sorbent Handling	Scaled on coal feed rate from SCS Draft-Report to NETL on Transport Gasifier IGCC Designs, Dec. 2005 (Jan 2003 \$) [Reference ii]
Coal Prep & Feed System	Scaled on coal feed rate from NETL Baseline Studies, 2007, Shell dry coal feed lock-hoppers, Case 6 [Reference i]
Catalyst Treatment	Scaled on coal feed rate from Exxon Catalytic Gasifier Process Simulation, Nov. 1982 [Reference iii]
Ash Handling	Scaled on ash/sorbent waste rate from Parsons FutureGen report to NETL on Transport Gasifier IGCC, Nov. 2005 (2004 \$) [Reference iv]
Catalyst Recovery	Scaled on coal feed rate from Exxon Catalytic Gasifier Process Simulation, Nov. 1982 [Reference iii]
Gasifier	Scaled on coal feed rate from Parsons FutureGen report to NETL on Transport Gasifier IGCC, Nov. 2005 (2004 \$) [Reference iv]
ASU	Scaled on O ₂ produced & gasifier pressure from Parsons FutureGen report to NETL on Transport Gasifier IGCC, Nov. 2005 (2004 \$) [Reference iv]
Gas clean up	
Dry Gas Cleaning	Scaled on syngas rate from NETL Baseline Studies, 2007, Texaco dry gas cleaning without CO ₂ removal, Case 1 [Reference i]
Humid Gas Cleaning	Scaled on syngas rate from prior IGCC plant studies conducted by Siemens [Reference v]
Sulfur Polishing	Engineering estimate: 2 parallel vessels each 5-ft ID with ZnO sulfur sorbent
Fuel cell system	
Syngas expander	Engineering estimate from ChemCad process simulator cost correlations
SOFC integrated module	Based on DOE SOFC Cost Target
Blowers, heat exchangers, expanders	Engineering estimate from ChemCad process simulator cost correlations
Oxy-Combustor	Engineering estimate based on conventional burner cost expectations
HRSG	Scaled on steam power from NETL Baseline Studies, 2007, Texaco dry gas cleaning without CO ₂ removal, Case 1 [Reference i]
Steam Turbine	Scaled on steam power from NETL Baseline Studies, 2007, Texaco dry gas cleaning without CO ₂ removal, Case 1 [Reference i]
CO₂ compression	Engineering estimate from ChemCad process simulator cost correlations
Other	
Cooling Water System	Scaled on cooling water circulation from NETL Baseline Studies, 2007, Texaco dry gas cleaning without CO ₂ removal, Case 1 [Reference i]
Feedwater & Misc BOP	Scaled on coal rate from NETL Baseline Studies, 2007, Texaco dry gas cleaning without CO ₂ removal, Case 1 [Reference i]
Accessory Electric Plant	
Instrumentation & Control	
Improvement to Site	
Building & Structures	

Total Plant Cost

Table 1111 provides the resulting capital cost by processing system for the 253 MW IGFC plant and the costs on a per net kW output from the overall system. The definition of total plant cost (TPC) is consistent with the NETL Baseline Studies which includes bare erected costs, contracting service fees and process and project contingencies. Owner's costs and costs "outside of the fence" are excluded. The costs are presented in 2007-dollars.

Table 11 – Total Plant Cost

	Atm-Pressure IGFC Plant		Pressurized IGFC Plant	
	Million \$	\$/kW net	Million \$	\$/kW net
Installed Capital Cost				
Gasifier, Coal, Ash, Catalyst Handling	138	545	123	487
Coal & Sorbent Handling	11	45	10	39
Coal Prep & Feed System	59	234	51	201
Catalyst Treatment	6	24	5	21
Ash Handling	14	54	16	62
Catalyst Recovery	10	39	9	35
Gasifier	37	147	32	128
ASU	38	150	32	127
Gas clean up	44	173	55	218
Dry Gas Cleaning / Humid Gas Cleaning	31	121	42	166
Sulfur Polishing	13	52	13	52
Fuel Cell System	167	661	155	612
Syngas Expander	2	7	2	7
SOFC Reactor	146	575	95	376
SOFC Air Blower & Heat Exch	20	79	58	229
Oxy-Combustor	5	18	5	18
HRSG	13	50	0	0
Steam Turbine	14	57	0	0
CO₂ capture	0	0	0	0
CO₂ compression	11	42	11	42
Other	73	287	59	234
Cooling Water System	8	31	10	38
Feedwater & Misc BOP	8	33	0	0
Accessory Electric Plant	31	120	27	106
Instrumentation & Control	10	40	9	35
Improvement to Site	8	32	7	28
Building & Structures	8	30	7	27
Overall TPC	503	1,983	440	1,738

6.2 Operating and Maintenance Expenses

The NETL Baseline Studies provided the primary basis for estimating the operating and maintenance (O&M) expenses. Additional costs specific to the IGFC-platform were also incorporated including SOFC stack replacement costs, polishing sorbent make-up price, and catalytic gasifier catalyst make-up costs.

Table 12 provides the information and sources used to develop the fixed and variable O&M costs. Table 13 provides the resulting fixed and variable O&M expenses for the 253MW IGFC plant both as total annual costs (assuming an 80% capacity factor) and on a per kW or per kWh basis.

Table 12 – O&M Expense Factors

Parameter	Value	Basis/Source
Fixed O&M		
Operating Jobs	12	Reference i, Case 2 requires 16 jobs for double-train 555 MW IGCC
Operating Labor Rate	\$375,800 per yr	Reference i, Case 2
Maintenance Labor	Scaled on TPC	Reference i, Case 2
Adm & Support Labor	25% of Operating & Maintenance Labor	Reference i, Case 2
Variable O&M		
Maintenance Costs	Scaled on TPC	Reference i, Case 2
Water Cost	\$1.03 per 1,000 gal	Reference i
Standard Chemicals/Waste Costs	Scaled on Coal Feed	Reference i, Case 2
Fuel Cell Stack Replacement		
SOFC Cost	\$100 per kW AC	DOE Program Goal
Stack Life	5 years	DOE Program Goal
Discount Rate	10%	Reference i, page 52, exhibit 2-15
Polishing Sorbent	9.85 cents per MMBtu HHV coal	Engineering Estimate
Catalytic Gasifier Catalyst	30 cents per MMBtu HHV coal	Reference vi
Fuel Cost	\$1.80 per MMBtu HHV	Reference i, page 30
CO ₂ Transport and Injection Cost	\$4 per mt	Reference i, page 13

Table 13 – Fixed and Variable O&M Expenses for 253MW IGFC Plant

	Atm-Pressure IGFC Plant		Pressurized IGFC plant	
	Annual Cost @ 80% Capacity Factor	Annual Unit Cost	Annual Cost @ 80% Capacity Factor	Annual Unit Cost
Fixed O&M Expenses	Million \$	\$/kW net	Million \$	\$/kW net
Total Fixed O&M	12.0	47	11.3	45
Variable O&M Expenses	Million \$	cents/kWh net	Million \$	cents/kWh net
Maintenance Material, Water, Chemicals	11.0	0.62	9.7	0.55
Stack Replacement	3.9	0.22	3.6	0.20
Sorbent	1.2	0.07	2.7	0.15
Gasifier Catalyst	3.7	0.21	3.2	0.18
Fuel	22.1	1.24	19.4	1.09
CO ₂ Storage	4.3	0.24	3.6	0.20
Total Variable O&M	46.2	2.60	42.2	2.38

6.3 Levelized Cost of Electricity

The LCOE is calculated consistent with the methodology and factors of the NETL Baseline Studies IGCC cases. The key inputs and resulting LCOE for the 250 MW IGFC plants are detailed in Table 14.

Table 14 – LCOE Key Inputs and Results

Key Inputs	Value	Source
Levelization Period	20 years	Reference i, page 51
Plant capacity factor	80%	Study Assumption
Capital expense factor	0.175 \$/yr/\$	Reference i, page 51
Coal cost levelization factor	1.2022	Reference i, page 51
General O&M levelization factor	1.1568	Reference i, page 51
Atm-Pressure IGFC LCOE	8.8 cents per kWh	
Pressurized IGFC LCOE	7.9 cents per kWh	

7.0 COMPARISON TO IGCC

The IGFC plant performance and cost is compared to that of a conventional IGCC plant with CO₂ removal in this section. This comparison illustrates the performance and cost advantages that IGFC technology potentially has over conventional coal-fired power generation technologies, especially when CO₂ removal is required.

Table 15 and Table 16 compare the performance, capital cost breakdown and LCOE of the three coal-fired power platforms, one based on a state-of-the-art IGCC plant and other two based on the solid oxide fuel cell plants described in this study. The capital cost of the system components is expressed in terms of \$ per net kW output from the system overall. As the system overall efficiency improves, the cost per net kWh of the balance-of-plant and other unit decreases proportionally.

Data on the combustion turbine system (IGCC) is taken from the 2007 NETL Baseline Studies. It represents today's state-of-the-art technology. The system presented includes a slurry-fed high-pressure/high-temperature gasifier, a General Electric 7FB turbine, and cold (dry) sulfur and pollutant removal from syngas. 90% CO₂ removal is accomplished with a two-stage syngas shift conversion and Selexol physical absorption prior to firing in the turbine. The system design was accomplished via ASPEN model simulations with an overall efficiency of 32.5% and net power output of 556 MW. The capital cost estimates were acquired by Parsons through original equipment manufacturer data.

The fuel cell systems offer a 14-22% lower LCOE compared to state-of-the-art IGCC (8.8 and 7.9 cents/kWh versus 10.2 for the IGCC). The efficiency is 17-24 percentage points higher and the capital cost is 17-27% less.

IGFC is especially favorable when comparing CO₂-capture platforms as the additional capital and power for the IGCC CO₂ separation systems is not required. This further increases the gap between the fuel cell and turbine systems because the fuel cell platform captures 99% of CO₂ versus 90% for the IGCC. The IGFC advantages are partially offset by the increased capital and O&M associated with the catalytic gasifier and SOFC.

Table 15 – Performance Comparison to IGCC

Parameter	IGCC	Atm-Pressure IGFC	Pressurized IGFC
Efficiency, % HHV	32.5	49.4	56.2
Capacity factor (hr/yr/8760hrs), %	80	80	80
% of net power from steam cycle	49	14	0
% of net power from fuel cell	84	94	87
% of net power from syngas expander	1	10	3
% of net power from turbine expanders	0	0	47
% of net power auxiliary load	(34)	(18)	(37)
Total Net Power, %	100	100	100
% CO ₂ capture	90	99+	99+
Total uncontrolled CO ₂ , lbCO ₂ /MWh	1765	1160	1020
Parasitic load due to CO ₂ capture, %-points	5.7	3.3	5.6

Table 16 – Capital Cost and LCOE Comparison to IGCC

Parameter	IGCC	Atm-Pressure IGFC	Pressurized IGFC
Capital cost, \$/kW net			
Gasifier, Coal, Ash, Catalyst Handling	744	545	487
Coal & Sorbent Handling	64	45	39
Coal Prep & Feed System	102	234	201
Catalyst Treatment	0	24	21
Ash Handling	80	54	62
Catalyst Recovery	0	39	35
Gasifier	498	147	128
ASU	342	150	127
Gas clean up	237	173	218
Dry/Humid Gas Cleaning	237	121	166
Sulfur Polishing	0	52	52
Combustion turbine/fuel cell	238	661	612
Combustion turbine	238	0	0
Syngas Expander	0	7	7
SOFC Reactor	0	575	376
SOFC Air Blower & Heat Exch	0	79	229
Oxy-Combustor	0	18	18
HRSB	99	50	0
Steam Turbine	116	57	0
CO ₂ capture	177	0	0
CO ₂ compression	68	42	42
Other	370	287	234
Cooling Water System	49	31	38
Feedwater & Misc BOP	69	33	0
Accessory Electric Plant	138	120	106
Instrumentation & Control	47	40	35
Improvement to Site	35	32	28
Building & Structures	32	30	27
Total Plant Cost, \$/kW net	2,391	1,983	1,738
Variable O&M, cents/kWh	0.81	1.11	1.09
General O&M	0.81	0.62	0.55
Stack Replacement Cost	0.00	0.22	0.20
Sorbent Cost	0.00	0.07	0.15
Gasifier Catalyst Cost	0.00	0.21	0.18
Fixed O&M, \$/kW/yr	43.7	47.3	44.7
Fuel cost, cents/kWh	1.89	1.24	1.09
CO ₂ cost, cents/kWh	0.34	0.24	0.20
LCOE (20-year), cents/kWh	10.2	8.8	7.9

Table 17 considers power plant emissions. Because almost all of the syngas environmental contaminants remaining after syngas cleaning are sequestered with the CO₂ stream, and almost complete CO₂ removal is achieved, the emissions from the IGFC plants are very small. The only emissions are likely to result from auxiliary process operations such as coal drying, sulfur recovery, and waste water processing, which are normally negligible. The IGFC plants water usage also compares very favorably to IGCC.

Table 17 – Plant Emissions and Water Usage

	IGCC	Atm-Pressure IGFC	Pressurized IGFC
Particulate (lb/MWh)	0.056	0.037	0.032
SO _x (lb/MWh)	0.0751	0 (sequestered)	0 (sequestered)
NO _x (lb/MWh)	0.366	0 (sequestered)	0 (sequestered)
Hg (lb/MWh)	4.48 x 10 ⁻⁶	0 (sequestered)	0 (sequestered)
CO ₂ (lb/MWh)	206	3	3
Water use (gal/MWh)	494	193	172

The SOFC system is considerably larger in footprint than the gas turbine power island, and the IGFC plant's air separation unit is roughly twice the footprint of the air separation unit for the IGCC plant. The overall power plant footprint of the IGFC plant is probably only slightly larger than the overall IGCC power plant footprint (includes the gasification section, the ASU, the gas cleaning section, the total power island with steam bottoming, and the associated coal receiving, storage, and preparation section).

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