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Performance and costs of advanced sustainable central power plants with CCS and H₂ co-production

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A B S T R A C T

With increasing concerns over global climate change caused by GHG emissions, carbon capture and storage (CCS) has become imperative for coal based power plants. Meanwhile, with the development and deployment of hybrid vehicles, electric vehicles, and alternative fuel vehicles, GHG reduction efforts in the power industry can also benefit the transportation sector. Power plants with H₂ co-production capability can contribute significantly in such development trends because H₂ powered fuel cell hybrid vehicles are very promising for future “zero emissions vehicles”. This work investigates the thermodynamic performance and cost advantage of employing advanced technologies currently under development for central power plants that (1) employ coal and biomass as feedstock; (2) co-produce power and high purity H₂; (3) capture most of the CO₂ evolved within the plants. Two system designs are developed: the first “base” case is an integrated gasification combined cycle (IGCC) system consisting of commercially ready technologies; the second “advanced” case is an integrated gasification fuel cell (IGFC) system. The feedstock employed consists of Utah bituminous coal along with two typical biomass resources, corn stover and cereal straw. The IGFC plant produces significantly higher amount of electricity for the same amounts of feedstock and H₂ export while the cost of producing the H₂ using a cost of electricity of $135/MW h is $1178/tonne for the IGFC case versus $2620/tonne for the IGCC case.

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1. Introduction

The world energy consumption is projected to continuously grow in the following decades; as a result, emissions of green house gas (GHG) CO₂ to the atmosphere are expected to increase by a significant amount. Coal fired power plants account for approximately 50% of the power generation in the United States and approximately 80% of the GHG emissions produced by the power generation sector [1]. With increasing concerns over global climate change caused by GHG emissions, CCS has become imperative for coal based power plants. Meanwhile, with the development and deployment of hybrid vehicles, electric vehicles, and alternative fuel vehicles, GHG reduction efforts in the power industry can also benefit the transportation sector, which accounts for one-fifth of global CO₂ emissions [2]. Power plants with H₂ co-production capability can contribute significantly in such development trends because H₂ powered fuel cell hybrid vehicles are very promising for future “zero emissions vehicles”.

Most of the H₂ co-production power plants investigated to date are based on coal gasification and gas turbine–steam turbine combined cycle for power generation (IGCC) [3–6]. High temperature fuel cells, such as solid oxide fuel cell (SOFC) and molten carbonate fuel cell (MCFC) are highly efficient energy conversion devices and the integration of coal gasification with high temperature fuel cell, the IGFC power plants are very promising for highly efficient utilization of coal for power production [7–9]. The high temperature fuel cell systems are also amenable to co-production of H₂ [10,11]; however system level investigations of such IGFC with H₂ co-production systems are relatively rare thus far.

Besides increasing the thermal efficiency of power plants, another dimension for GHG reduction is to use feedstock that has low carbon footprint; biomass, a nearly CO₂ neutral source of renewable fuel, is an important feedstock of this kind. Much research work has addressed H₂ and power production from biomass fuel [12]. However, due to its low energy density (and being a distributed resource), transportation of biomass collected from...
various locations over long distances to a central plant makes it economically prohibitive. The biomass facility has to be located in close proximity to the feedstock, which limits the size of the plant and the economies of scale of large plants cannot be taken advantage of to reduce the specific capital cost unless biomass is co-fed with coal to a “central station” type gasification facility (located close to the biomass sources); this type of plant may be a more practical solution.

The purpose of this work is to investigate the thermodynamic performance and cost advantage of employing advanced technologies currently under development for central power plants that (1) employ coal and biomass as feedstock; (2) co-produce power and high purity H2; (3) capture most of the CO2 evolved within the plants. Thus, two system designs are developed: the first “base” case is an IGCC system consisting of commercially ready technologies; the second “advanced” case is an IGFC system with large scale SOFCs which are yet to be developed for central power plant applications, currently projected to be demonstrated in the 2020 time frame.

2. Design basis and strategy

The feedstock employed in this research work consists of Utah bituminous coal and two typical biomass resources (corn stover [13] and cereal straw [14]). The characteristics of the two feedstocks are summarized in Table 1. The composition of the plant feedstock consists of 66 wt.% of Utah coal, 17 wt.% of corn stover, and 17 wt.% of cereal straw (all on a dry basis).

Site ambient conditions correspond to ISO conditions of 15 °C (59 °F) dry bulb temperature, 60% relative humidity and sea level barometric pressure. Mechanical draft cooling tower are utilized for plant heat rejection with a 3.9 °C (7 °F) approach to the wet bulb temperature; an 11.1 °C (20 °F) temperature rise is assumed for the cooling water.

The base case plant comprises of commercially offered subsystems for IGCC applications utilizes an entrained flow O2 blown gasifier, low temperature gas cleanup, pre-combustion CO2 separation using Selexol™ unit, and an F class gas turbine–steam turbine combined cycle for power generation. High purity H2 is produced using pressure swing adsorption (PSA) upstream of gas turbine combustor.

Based on previous research conducted on integrating gasifiers with SOFC stacks [8], a catalytic hydro-gasifier is chosen to be paired with SOFC since such gasifiers: (1) are capable of producing syngas with high CH4 content (CH4 can be internally reformed in SOFC and serves as a chemical heat sink, thus reducing cooling air requirement and the corresponding compression work); (2) have a high thermal efficiency (relatively low syngas exit temperature, thus most of the coal/biomass bound chemical energy is contained as syngas bound chemical energy); (3) have a reasonably high carbon conversion. To take full advantage of the cooling effect of CH4 contained in the syngas, CO2 separation is performed downstream of SOFC stacks while in the base IGCC case, the syngas is decarbonized upstream of the power block. By controlling the amount of fuel constituents of the syngas oxidized within the SOFC stacks, high purity H2 can be produced from the SOFC anode exhaust gas using PSA technology.

To compare the performance of the two cases on a consistent basis, the two systems are designed to consume the same amount of fuel and produce the same amount of H2, and the net power output compares the thermal efficiency of the two systems. The level of carbon capture of the two systems is maintained similar and the captured CO2 stream is compressed to the same pressure (151.70 bar, or 2200 psia) to be pipelined for sequestration. Aspen Plus® process engineering flow sheet simulation tool is utilized for modeling these systems. Major power block subsystem characteristics for both the IGCC and the IGFC cases are summarized in Table 2.

### Table 1
Summary of characteristics of feedstock used in this work.

|                      | Utah bituminous coal (emery) | Corn stover [13] | Cereal straw [14] |
|----------------------|-------------------------------|------------------|------------------|
|                      | As received | Dry              | As received | Dry              | As received | Dry              |
| **Proximate Analysis (wt.%)** |              |                  |              |                  |              |                  |
| Moisture             |  7.1       | –                | 30.00 (field air dried) | –          | 30.00 (field air dried) | –          |
| Ash                  | 12.0       | 12.92            | 4.71         | 6.73             | 3.71         | 5.30             |
| Volatile matter      | 34.6       | 37.24            | 46.61        | 66.61            | 47.13        | 67.33            |
| Fixed carbon         |  46.3      | 49.84            | 18.66        | 26.66            | 19.16        | 27.37            |
| **Higher heating value (kJ/kg)** | 26,162 | 28,161           | 12,551       | 17,930           | 12,250       | 17,500           |
| **Ultimate analysis (wt.%)** |              |                  |              |                  |              |                  |
| Carbon               | –          | 67.81            | –            | 45.44            | –            | 46.51            |
| Hydrogen             | –          | 4.96             | –            | 5.51             | –            | 6.27             |
| Nitrogen             | –          | 1.51             | –            | 0.69             | –            | 0.50             |
| Chlorine             | –          | 75–323 ppm       | –            | 0.10             | –            | 0.40             |
| Sulfur               | –          | 2.04             | –            | 0.04             | –            | 0.10             |
| Ash                  | –          | 12.92            | –            | 6.72             | –            | 5.28             |
| Oxygen               | –          | 10.76            | –            | 41.50            | –            | 40.94            |
tion, and produces N₂ at an intermediate pressure (IP) for injection into the gas turbine as a thermal diluent for NOₓ control. A portion of the IP N₂ is further compressed for use as transport gas to pneumatically convey the feedstock to the gasifier.

The gasifier operates at a nominal temperature and pressure of 1371 °C (2500 °F) and 42.4 bar (615 psia). The gasifier partially oxidizes the coal with the O₂ to generate hot raw syngas, slag and char. After dry particulate removal, the raw syngas provides heat for the steam system to generate high pressure (HP) and low pressure (LP) steam. The raw syngas is then water scrubbed to remove particulates, alkalis, chlorides and NH₃.

HP steam is injected into the syngas to increase H₂O content to a level sufficient for the downstream shift reactors. The syngas then enters the shift unit, where most of the CO present in the syngas is reacted with H₂O vapor to produce H₂ and CO₂. Heat generated by the exothermic shift reaction is recovered by generating IP and medium pressure (MP) steam. The shifted gas leaving the shift unit further provides heat to the vacuum condensate heater and is then cooled against cooling water. The cooled syngas is then superheated by about 11 °C (20 °F) to avoid pore condensation and then fed to a sulfided activated carbon bed for removal of Hg. The effluent gas is then cooled in a trim cooler against cooling water and fed into the Selexol™ acid gas removal unit. The high temperature condensate separated from the gas is recycled to the scrubber while the NH₃ laden colder condensate is fed to a sour water stripper. The sour gases stripped off from the water are routed to the acid gas combustor.

The Selexol™ process produces the clean syngas, acid gas which is supplied to the acid gas combustion unit, and CO₂ streams at four different pressures. The CO₂ streams are compressed to a pressure where the CO₂ liquefies near the cooling water temperatures, dehydrated and is then pumped to the pipeline pressure of 151.70 bar (2215 psia). The acid gas from the Selexol™ process, together with the sour gases from the sour water stripper, are combusted with 95 vol.% purity O₂ in a acid gas combustion unit. The heat generated in the acid gas combustion unit is recovered by producing IP and LP steam. The exhaust gas is mixed with the CO₂ stream prior to compression for sequestration.

A portion of the clean decarbonized syngas leaving the Selexol™ unit with ultra low sulfur content is treated in a PSA unit to produce high purity H₂ product which is compressed to 45.71 bar (663 psia) which is the same pressure of the H₂ stream produced in the IGFC case. The PSA also produces a tail gas stream which consists of the remaining fuel gas components (mostly CO, CO₂, H₂, and N₂).

The PSA tail gas is compressed and then combined with the remainder of the clean decarbonized syngas. This combined fuel gas stream is preheated to 288 °C (550 °F) using IP steam and then fed to the gas turbine combustor. IP N₂ from the ASU (preheated by IP steam) is also fed to the gas turbine combustor. The flow rate of IP N₂ is varied so that the input fuel and N₂ has a combined Lower Heating Value (LHV) of 4730 kJ/Nm³ (120 BTU/SCF). The introduction of N₂ reduces the formation of NOₓ within the combustor of

| Subsystem                | Value |
|--------------------------|-------|
| IGCC case – gas turbine  |       |
| Pressure ratio           | 16    |
| 1st rotor inlet temperature (°C) | 1320 |
| Generator efficiency (%) | 98.65 |
| IGCC case – steam turbine|       |
| HP isentropic efficiency (%) | 83.20 |
| LP isentropic efficiency (%) | 89.94 |
| IP isentropic efficiency (%) | 89.56 |
| Generator efficiency (%) | 98.0  |
| IGCC case – SOFC         |       |
| Nominal operating pressure (bar) | 10   |
| Operating temperature range (°C) | 650–850 |
| Fuel utilization (% per Pass) | 72    |
| Cell Voltage (V)         | 0.8   |
| Inverter efficiency (%)  | 97    |
the F class gas turbine by lowering the flame temperature. Note that because of the H₂ export, the relative amount of diluent N₂ available for gas turbine injection is high, thus there is no need to humidify the fuel gas as in some of the power-only IGCC plant configurations [16].

The exhaust gas exits the gas turbine at 564 °C (1047 °F) and enters the heat recovery steam generator (HRSG) to provide heat for the bottoming Rankine cycle consisting of a triple pressure reheat steam cycle. The flue gas is then discharged through the plant stack.

Bulk of the steam generated within the plant is used in a steam turbine for power generation while the remainder to satisfy process steam demands. The plant includes the necessary general facilities such as a cooling water system, instrument air, flare, etc.

### 4. Advanced design: IGFC plant with H₂ co-production and CCS

An overall block flow diagram for this IGFC case with CO₂ capture and H₂ co-production is presented in Fig. 2 and major stream data are presented in Table 4.

The gasification subsystem contains two gasifiers: a “top stage” catalytic hydro-gasifier and a “bottom stage” high temperature slagging entrained-bed O₂ blown gasifier. The majority of the fuel (coal and biomass) are gasified in the catalytic hydro-gasifier by a high temperature gas stream containing steam, H₂, and CO. The “as received” coal and biomass after size reduction are impregnated with potassium catalyst (in the form of KOH and K₂CO₃). Fluidized bed drying with vapor recompression is again employed here due to the high moisture content in the biomass. The catalyst requirement (on a K₂CO₃ basis) is assumed to be 15 wt.% of the total dry coal and biomass input. No credit was taken for the potassium present in the biomass. Some lime is used to regenerate the catalyst.

The unconverted carbon and accompanying “fine ash” from the catalytic hydro-gasifier, after catalyst recovery, is supplied to the high temperature slagging entrained-bed O₂ blown gasifier. The major function of this second gasifier is to produce a hot gas stream rich in H₂ and CO, which is then used in the catalytic hydro-gasifier as gasification agent. The carbon content left in the char from the top stage gasifier alone cannot produce sufficient H₂ and CO to meet the demand of the top stage gasifier and thus a portion of the dried coal, further reduced in size to meet the specifications of the O₂ blown entrained bed bottom stage gasifier, is provided to bottom stage gasifier as a supplement feed. Since the carbon content in the char and “fine ash” fed to the bottom stage gasifier is relatively low, only coal (which has higher carbon content than the biomass) is provided to the bottom stage O₂ blown gasifier. The O₂ blown gasifier helps increase the overall carbon conversion of the gasification process while converting the ash content into a vitrified non-leachable solid form. N₂ produced by the ASU unit, after further compression, is used as transport gas to pneumatically convey the feedstock to the gasifiers.

After dry particulate removal, the raw effluent from the catalytic hydro-gasifier at a temperature and pressure of 690 °C (1275 °F) and 70 bar (1015 psia) enters the heat recovery and gas cleanup system. The raw syngas is initially cooled against the HP steam which is utilized in the catalytic hydro-gasifier. After providing additional heat for the steam system, the raw syngas is then supplied to the syngas cleanup/low temperature gas cooling/heat recovery system, which includes water wash to remove particulates, alkalis, chlorides and NH₃, a carbonyl sulfide hydrolysis reactor and a sulfided activated carbon bed for capture of Hg, followed by a Selexol™ unit to remove the sulfur compounds.

Next, water vapor is introduced into the clean CH₄ rich syngas leaving the Selexol™ unit. The added moisture prevents deposition of carbon in the downstream reactors and the SOFC anode compartment. Instead of using HP steam for humidification, here the water vapor is introduced by directly contacting the syngas with liquid water flowing down through a counter-current column, which allows for recovery of low temperature heat generated within the plant. Clean process condensate collected from within the plant is used as the makeup water for this humidifier.

The clean syngas is then supplied to a reactor/expander topping cycle. In the shift reactor, the exothermic shift reaction takes place and the gases are heated up; the high temperature syngas then expands in the expander to recover some power.

Because an SOFC has more stringent requirements with respect to contaminants contained in the syngas than an F class gas turbine, a guard bed is included upstream of the expander as a final cleanup step to limit trace amounts of chlorides and sulfur compounds to <0.1 ppm V each. The guard bed consists of alternating layers of COS hydrolysis catalyst such as a Co–Mo, or a Ni–Mo catalyst and ZnO for capture of the H₂S and the chlorides.

The syngas coming out of the expander is preheated to 650 °C (1202 °F) and fed to the anode side of SOFC stacks; ambient air is compressed to 11.3 bar (164 psia), also preheated to 650 °C (1202 °F) and then sent to the cathode side of the SOFC stacks. The SOFC module design employed in this work is comprised of cascading four stages of identical SOFC stacks with air flowing in series and fuel flowing in parallel, and intra-stack introduction of fresh air to produce roughly identical operations for each stack in the module. The purpose of this design is to achieve high single pass air utilization in each of the SOFC stacks (thus maintaining effective cooling of the SOFC stacks) and low overall air utilization at the same time (thus reducing the parasitic air compression power); more details and discussions regarding this design can be found in [17].

| Component mole fractions | Syngas to power block | Flue gas to atmosphere | H₂ coproduct | CO₂ to sequestration |
|--------------------------|-----------------------|------------------------|--------------|---------------------|
| O₂                      | 0.1100                |                        | <200 ppm V  | 0.0878              |
| N₂ + Ar                 | 0.7633                |                        | <1 ppm V    | 0.0002              |
| H₂                      | 0.9998                |                        |              | 0.9738              |
| CO                      |                      |                        | <1 ppm V    | 0.0087              |
| CO₂                     |                      |                        |              | 0.0056              |
| H₂O                     |                      |                        |              | 0.1211              |
| CH₄                     |                      |                        |              | 0.0002              |
| H₂S                     |                      |                        |              | <1 ppm V            |
| SO₂                     |                      |                        |              | 0.0086              |
| Total flow (kg mol/h)   | 8855                  | 70845                  | 3194         | 7773                |
| Total flow (kg/h)       | 46846                 | 1944237                | 6438         | 339521              |
| Temperature (°C)        | 288                   | 151                    | 40           | 42                  |
| Pressure (bar)          | 30.81                 | 1.05                   | 45.71        | 151.70              |

Table 3: Major stream data for the IGCC case.
Seventy-two percent of the fuel in the syngas is oxidized in SOFC stacks. Anode exhaust leaves the SOFC stacks at 705°C (1302°F) and is cooled by providing heat to the SOFC stack syngas pre-heater, the shift reactor (in the reactor/expander topping cycle) pre-heater and the COS hydrolysis reactor pre-heater. The gas stream is then fed to a catalytic reactor for the water gas shift reaction, in which most of the CO content in the anode exhaust is converted into H₂ and CO₂. The effluent from the shift reactor, after further heat recovery and cooling, is fed to an activated methyl diethanol amine (aMDEA) unit for CO₂ separation. The CO₂ stripped from the aMDEA solvent is combined with the oxidized acid gas leaving the Selexol™ unit. This combined stream is then dehydrated and compressed (with inter-cooling) to 151.70 bar (2215 psia) for sequestration. A large fraction of the de-carbonized anode exhaust from the aMDEA unit is fed to a PSA unit, while the remaining gas is sent to the combustor downstream of the SOFC stacks. The PSA unit produces high purity H₂ product at pressure of 45.71 bar (663 psia) and a tail gas consist of the remaining fuel gas components. The PSA tail gas is compressed (with inter-cooling) and also fed to combustor. The oxidant for this combustor consists of the cathode exhaust from the SOFC stacks. The combustor exhaust is partially expanded in a turbine, fed to a recuperator to pre-heat the cathode inlet air supplied by the compressor, and then further expanded to near atmospheric pressure in the turbine to generate additional power before entering the HRSG. The exhaust gas from the HRSG is also used to provide some heat for the fuel drying operation.

5. System performance

Overall system performance of the two cases are compared in Table 5 while more detailed break-down of the power producers and consumers in the two plants are shown in Tables 6 and 7. Both plants consume the same amount of coal and biomass while exporting 154.6 tonne/D of H₂ (which is equivalent to
et al. [18], the IGCC plant can achieve an efficacy of 41.92% while (without taking credit for energy contained in the exported H2) of net electricity, which corresponds to an "electricity efficiency" of 23.41% of the input fuel bound energy on an HHV basis), and achieving similar levels of carbon capture (95% to ~98%), but the net electricity output of the IGFC case is significantly higher than that of the base IGCC case. The IGCC plant produces 200.64 MW of net electricity, which corresponds to an “electricity efficiency” (without taking credit for energy contained in the exported H2) of 18.51% while the IGFC plant produces as much as 342.80 MW of net electricity with a correspondingly higher electricity efficiency of 31.63% (both on HHV basis), a more than 13% points higher than the IGCC case primarily due to the more efficient power turbine produces only a small amount of power because most of the heat generated within the plant is employed for process heating and only a limited amount of the heat is left over for steam generation.

The power requirement of ASU in the IGCC case is significantly lower than the IGCC case since only a portion of the coal is fed to the O2 blown gasifier in the IGFC case. The CO2 compression work in the IGFC case on the other hand is much higher than that in the IGCC case, because in the IGCC case carbon is separated "post combustion" using aMDEA chemical absorption and as a result the CO2 released from the solvent is at lower pressure and requiring more energy for compression to the required final pressure for pipelining. In the IGCC case the carbon is separated from the gas stream at significantly higher pressure ("pre combustion" capture from shifted syngas) using Selexol™ physical absorption and the CO2 is released from the solvent by "pressure swing" at relatively higher pressures.

The IGFC plant has an efficacy as high as 55.04%. However, by assigning a thermal efficiency of say 60% (on LHV basis) for the conversion of the exported H2, an indication of the overall thermal efficiencies may be obtained. The resulting effective efficiencies are then 30.4% and 43.5% (on HHV basis) for the IGCC and the IGFC cases.

Table 5 shows that in the IGCC system, gas turbine and steam turbine are the two major power producers while the significant internal power consumers are ASU, acid gas removal (Selexol™) and CO2 compression. The gross power production and internal power consumption of the IGFC case are very different from the IGCC case. As can be seen in Table 7, the SOFC stack is the single largest power producer. It should be noted that in Table 7 the power produced by the cathode exhaust expander and the power required to compress the air for SOFC are reported separately even though these equipment may be mounted on a common shaft. Since the expander produces less power than that required by the compressor, a motor (rather than a generator as in the case of the gas turbine in the IGCC case) is mounted on this shaft. Steam turbine produces only a small amount of power because most of the heat generated within the plant is employed for process heating and only a limited amount of the heat is left over for steam generation.

The power requirement of ASU in the IGFC case is significantly lower than the IGCC case since only a portion of the coal is fed to the O2 blown gasifier in the IGFC case. The CO2 compression work in the IGFC case on the other hand is much higher than that in the IGCC case, because in the IGFC case carbon is separated "post combustion" using a MDEA chemical absorption and as a result the CO2 released from the solvent is at lower pressure and requiring more energy for compression to the required final pressure for pipelining. In the IGCC case the carbon is separated from the gas stream at significantly higher pressure ("pre combustion" capture from shifted syngas) using Selexol™ physical absorption and the CO2 is released from the solvent by “pressure swing” at relatively higher pressures.

Table 7

| Major gross power output (MW) | IGCC with H2 co-production and CCS | IGFC with H2 co-production and CCS |
|------------------------------|-----------------------------------|-----------------------------------|
| Coal feed rate (dry basis, tonne/D) | 2511 | 647 |
| Corn stover feed rate (dry basis, tonne/D) | 647 | 23.41 |
| Total energy input (HHV, GJ/h) | 3902 | |
| Total gross power (MW) | 330.56 | 607.51 |
| Total internal power consumption (MW) | 129.92 | 264.71 |
| Net Electric Power (MW) | 200.64 | 342.80 |
| H2 exported (tonne/D) | 154.57 | 31.63 |
| Carbon capture rate (%) | 95.0 | 97.5 |
| Net power generation efficiency (%) HHV) | 18.51 | 31.63 |
| Efficacy (%) HHV | 41.92 | 55.04 |

2 Efficacy = (energy contained in exported H2 on HHV basis + net electric power)/ (energy contained in feedstock on HHV basis).
6. Cost estimation and results

Total plant cost, operating and maintenance (O&M) costs are estimated in order to develop the relative economics of the two technologies as summarized in Table 8. Cost estimates are derived primarily from NETL studies on IGCC systems [16] and the IGFC systems [19]. The IGCC plant cost is lower at $1243 million versus $1440 million for the IGFC plant. However, due to the significantly higher efficiency of the IGFC case, the revenue stream generated by its larger export power more than compensates for its higher plant cost. The cost of producing the coproduct H2 is determined for the two cases using a 20 year levelized cost of electricity of $135/MWh, based on data developed by NETL [16] for IGCC and boiler plants fueled by a bituminous coal and equipped with 90% carbon capture. The resulting cost of producing the coproduct H2 is significantly lower for the IGFC case, $1178/tonne versus $2620/tonne for the IGCC case.

7. Summary and conclusion

The purpose of this work is to investigate the thermodynamic performance and cost advantage of employing advanced technologies currently under development for central power plants that (1) employ coal and biomass as feedstock; (2) co-produce power and high purity H2; (3) capture most of the CO2 evolved within the plants. Two system designs are developed: the first “base” case is an IGCC system consisting of commercially ready technologies; the second “advanced” case is an IGFC system with large scale SOFCs which are yet to be developed for central power plant applications, currently projected to be demonstrated in the 2020 time frame.

The feedstock employed in this research work consists of Utah bituminous coal and two typical biomass resources, corn stover and cereal straw. The composition of the plant feedstock consists of 66 wt.% of Utah coal, 17 wt.% of corn stover, and 17 wt.% of cereal straw (all on a dry basis). Site ambient conditions correspond to ISO conditions and mechanical draft cooling tower is utilized for plant heat rejection. Both the IGCC and the IGFC based plants consume the same amount of coal and biomass while exporting 154.6 tonnes/D of H2 (which is equivalent to 23.41% of the input fuel bound energy on an HHV basis), and achieving similar levels of carbon capture (95% to ~98%), but the net electricity output of the IGFC case is significantly higher than that of the base IGCC case. The IGCC plant produces 200.64 MW of net electricity, which corresponds to an electricity efficiency (without taking credit for energy contained in the exported H2) of 18.51% while the IGFC plant produces as much as 342.80 MW of net electricity with a correspondingly higher electricity efficiency of 31.63% (both on HHV basis), a more than 13% points higher than the IGCC case primarily due to the more efficient power block and synergy between the hydrogasification process and the SOFC. In terms of overall plant efficacy, the IGCC plant can achieve an efficacy of 41.92% while the IGFC plant has an efficacy as high as 55.04%.

The IGCC plant cost is lower at $1243 million versus $1440 million for the IGFC plant. However, due to the significantly higher efficiency of the IGFC case, the revenue stream generated by its larger export power more than compensates for its higher plant cost. The cost of producing the coproduct H2 is determined for the two cases using a 20 year levelized cost of electricity of $135/MWh, based on data developed by NETL [16] for IGCC and boiler plants fueled by a bituminous coal and equipped with 90% carbon capture. The resulting cost of producing the coproduct H2 is significantly lower for the IGFC case, $1178/tonne versus $2620/tonne for the IGCC case.

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