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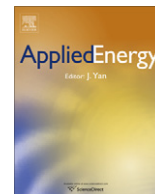
Performance and costs of advanced sustainable central power plants with CCS and H$_2$ co-production

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ABSTRACT

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 H$_2$ co-production capability can contribute significantly in such development trends because H$_2$ 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 feed stock; (2) co-produce power and high purity H$_2$; (3) capture most of the CO$_2$ 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 H$_2$ export while the cost of producing the H$_2$ 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) CO$_2$ 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 CO$_2$ emissions [2]. Power plants with H$_2$ co-production capability can contribute significantly in such development trends because H$_2$ powered fuel cell hybrid vehicles are very promising for future "zero emissions vehicles".

Most of the H$_2$ 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 H$_2$ [10,11]; however system level investigations of such IGFC with H$_2$ 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 CO$_2$ neutral source of renewable fuel, is an important feedstock of this kind. Much research work has addressed H$_2$ and power production from biomass fuel [12]. However, due to its low energy density (and being a distributed resource), transportation of biomass collected from

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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 feed stock; (2) co-produce power and high purity H_2; (3) capture most of the CO_2 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 comprising of commercially offered subsystems for IGCC applications utilizes an entrained flow O_2 blown gasifier, low temperature gas cleanup, pre-combustion CO_2 separation using Selexol™ unit, and an F class gas turbine–steam turbine combined cycle for power generation. High purity H_2 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 CH_4 content (CH_4 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 CH_4 contained in the syngas, CO_2 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 H_2 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 H_2, 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 CO_2 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.

3. Base case: IGCC plant with H_2 co-production and CCS

An overall block flow diagram for this IGCC case with CO_2 capture and H_2 co-production is presented in Fig. 1 and major stream data are presented in Table 3.

The gasification plant is based on a dry feed high pressure entrained bed gasifier. As received coal and biomass (corn stover and cereal straw) are fed to commutation equipment and then dried in fluid bed dryer. Biomass generally has higher moisture content than coal and requires more heat energy for drying. Fluidized bed drying with vapor recompression is employed. The technology utilizes the latent heat of the vapor produced in the dryer by pressurizing the vapor and supplying it to the heat exchange tubes located within the fluidized bed dryer. This type of drying technology has been successfully applied in drying of high moisture content brown coals [15]. The electricity demand to drive the vapor compressor is quite small as compared to the energy recovered from the vapor. The dried feedstock is then further reduced in size to meet the specifications of the entrained bed gasifier. The energy required for this fine grinding operation is greatly reduced when the biomass feedstocks are pre-dried.

The air separation unit (ASU) operating at elevated pressure supplies 95 vol.% purity O_2 to the gasifier and for acid gas combus-

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	46.63	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

Table 2
Major power block subsystem characteristics.

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
IP isentropic efficiency (%)	89.94
LP isentropic efficiency (%)	89.56
Generator efficiency (%)	98.0
IGFC 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

tion, and produces N_2 at an intermediate pressure (IP) for injection into the gas turbine as a thermal diluent for NO_x control. A portion of the IP N_2 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 and the biomass with the O_2 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 NH_3.

HP steam is injected into the syngas to increase H_2O 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 H_2O vapor to produce H_2 and CO_2. 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 next 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 NH_3 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 CO_2 streams at four different pressures. The CO_2 streams are compressed to a pressure where the CO_2 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 O_2 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 CO_2 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 H_2 product which is compressed to 45.71 bar (663 psia) which is the same pressure of the H_2 stream produced in the IGFC case. The PSA also produces a tail gas stream which consists of the remaining fuel gas components (mostly CO, CO_2, H_2, and N_2).

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 N_2 from the ASU (preheated by IP steam) is also fed to the gas turbine combustor. The flow rate of IP N_2 is varied so that the input fuel and N_2 has a combined Lower Heating Value (LHV) of 4730 kJ/Nm^3 (120 BTU/SCF). The introduction of N_2 reduces the formation of NO_x within the combustor of
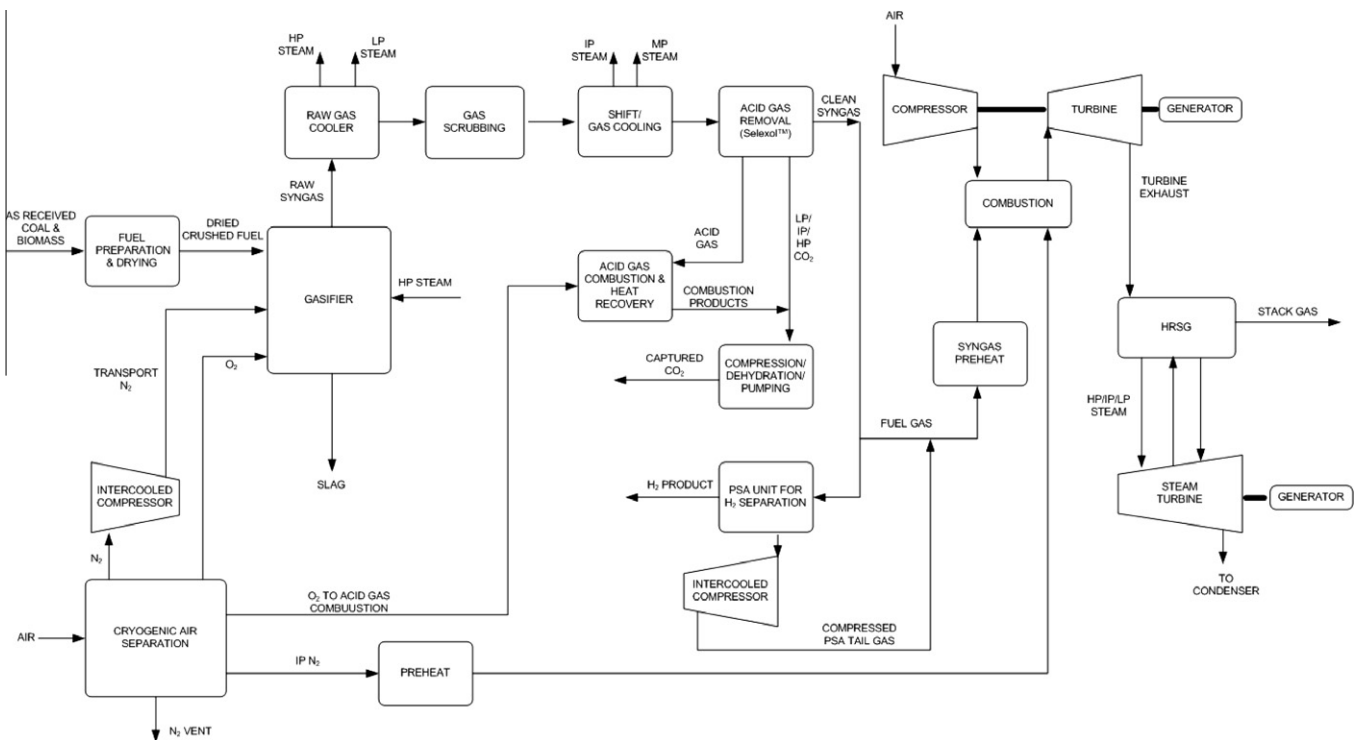

Fig. 1. Overall block flow diagram – IGCC with H_2 co-production and CO_2 sequestration.

Table 3
Major stream data for the IGCC case.

Component mole fractions	Syngas to power block	Flue gas to atmosphere	H_2 coproduct	CO_2 to sequestration
O_2		0.1100		
N_2 + Ar	0.0578	0.7633	<200 ppm V	0.0087
H_2	0.8985		0.9998	0.0087
CO	0.0120		<1 ppm V	0.0002
CO_2	0.0310	0.0056		0.9738
H_2O	0.0005	0.1211		
CH_4	0.0002			
H_2S	<1 ppm V			
SO_2				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

the F class gas turbine by lowering the flame temperature. Note that because of the H_2 export, the relative amount of diluent N_2 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 H_2 co-production and CCS

An overall block flow diagram for this IGFC case with CO_2 capture and H_2 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 O_2 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, H_2 and CO. The "as received" coal and biomass after size reduction are impregnated with potassium catalyst (in the form of KOH and K_2CO_3). Fluidized bed drying with vapor recompression is again employed here due to the high moisture content in the biomass. The catalyst requirement (on a K_2CO_3 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 O_2 blown gasifier. The major function of this second gasifier is to produce a hot gas stream rich in H_2 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 H_2 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 O_2 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 O_2 blown gasifier. The O_2 blown gasifier helps increase the overall carbon conversion

of the gasification process while converting the ash content into a vitrified non-leachable solid form. N_2 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 NH_3, 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 CH_4 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 H_2S 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].

Fig. 2. Overall block flow diagram – IGFC with H_2 co-production and CO_2 sequestration.

Table 4
Major stream data for the IGFC case.

Component mole fractions	Syngas to power block	Flue gas to atmosphere	H_2 coproduct	CO_2 to sequestration
O_2		0.1114		
N_2 + Ar	0.0311	0.8522	<200 ppm V	0.0063
H_2	0.2155		0.9998	0.0003
CO	0.0075		<1 ppm V	
CO_2	0.2343	0.0036		0.9849
H_2O	0.3040	0.0328		
CH_4	0.2076			
H_2S	<0.1 ppm V			
SO_2				0.0085
Total flow (kg mol/h)	14665	54350	3187	7879
Total flow (kg/h)	303062	1532442	6424	346875
Temperature (°C)	193	105	40	43
Pressure (bar)	11.17	1.05	45.71	151.70

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 H_2 and CO_2. The effluent from the shift reactor, after further heat recovery and cooling, is fed to an activated methyl diethanol amine (aMDEA) unit for CO_2 separation. The CO_2 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 H_2 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 H_2 (which is equivalent to

Table 5
Performance comparison of the two cases.

Case	IGCC with H_2 co-production and CCS	IGFC with H_2 co-production and CCS
Coal feed rate (dry basis, tonne/D)	2511	
Corn stover feed rate (dry basis, tonne/D)	647	
Cereal straw feed rate (dry basis, tonne/D)	647	
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
H_2 exported (tonne/D)	154.57	
H_2 exported (% of input fuel HHV)	23.41	
Carbon capture rate (%)	95.0	97.5
Net power generation efficiency (% HHV)	18.51	31.63
Efficacy (% HHV)	41.92	55.04

Table 6
Major gross power output and auxiliary power consumption of the IGCC case.

Major gross power output (MW)	
Gas turbine	230.00
Steam turbine	100.56
Total gross power generated	330.56
Major auxiliary power consumption (MW)	
Feedstock milling and handling	7.35
Feedstock solid pump	6.27
Vapor compressor in feedstock dryer	5.82
Slag handling and dewatering	0.84
ASU main air compressor	43.24
ASU O_2 compressor	5.78
ASU N_2 compressor	15.52
ASU auxiliaries	0.69
CO_2 compression	16.52
H_2 compression	0.81
PSA tail gas compression	3.23
Boiler feed water and water demineralization system pumps	3.57
Vacuum condensate pump	0.21
Process condensate and sour water stripping system	0.47
Scrubber pumps	0.06
Cooling water circulating pumps	2.53
Cooling tower fans	1.35
Selexol™ unit	10.88
Gas turbine auxiliaries	0.50
Steam turbine auxiliaries	0.04
Transformer losses	0.72
Miscellaneous balance of plant (BOP) and lighting	3.49
Total auxiliary power consumption	129.92

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 H_2) 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[2] as defined by McGurl et al. [18], the IGCC plant can achieve an efficacy of 41.92% while

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[2] Efficacy = (energy contained in exported H_2 on HHV basis + net electric power)/(energy contained in feedstock on HHV basis).

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 H_2, 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 6 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 CO_2 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 O_2 blown gasifier in the IGFC case. The CO_2 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 aMDEA chemical absorption and as a result the CO_2 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 CO_2 is released from the solvent by "pressure swing" at relatively higher pressures.

Table 7
Major gross power output and auxiliary power consumption of the IGFC case.

Major gross power output (MW)	
SOFC stacks	459.31
SOFC cathode exhaust expander	127.33
Steam turbine (as compressor driver)	0.80
Syngas expander (reactor/expander topping cycle)	19.10
Selexol™ hydraulic turbine	0.98
Total gross power generated	607.51
Major auxiliary power consumption (MW)	
Feedstock milling and handling	7.35
Feedstock solid pump	6.27
Vapor compressor in feedstock dryer	9.13
Slag handling and dewatering	3.45
ASU main air compressor	8.82
ASU O_2 compressor	4.94
ASU N_2 compressor	1.53
ASU auxiliaries	0.46
SOFC feeding air compressor	150.15
SOFC anode exhaust compressor	14.28
CO_2 compression	34.61
H_2 compression	4.36
PSA tail gas compression	2.29
Selexol™ unit	4.14
aMDEA unit	1.74
BFW feed pump and raw water demineralization system	1.51
Scrubber pumps	0.05
Condensate recycle pump	0.67
Syngas humidifier water pump	0.22
Cooling water circulating pumps	2.56
Cooling tower fans	1.37
Steam turbine auxiliaries	0.41
Transformer losses	1.32
Miscellaneous BOP & lighting	3.49
Total auxiliary power consumption	264.71

Table 8
Cost comparison of the IGCC case and IGFC case at capacity factor (CF) of 80%.

Case	IGCC with H_2 co-production and CCS	IGFC with H_2 co-production and CCS
Installed cost ($1000)		
ASU	164,084	66,522
Fuel (+catalyst) receiving, preparation and feeding	198,790	223,555
Gasifier and auxiliaries	349,758	241,261
Raw gas cooling and cleanup	13,537	175,061
Fuel gas expander	None	4091
CO_2 separation	161,721	152,780
CO_2 compression, dehydration and pumping	23,066	40,998
Fuel cell system	Not applicable	182,074
Gas turbine system	70,317	117,802[a]
Steam turbine system	32,890	1115
HRSG, ducting and stack	30,021	24,867
Oxy combustion system	1607	2263
Anode exhaust shifting	Not applicable	7469
PSA unit for H_2 production	9859	8923
Feedwater and miscellaneous BOP systems	53,066	29,455
Cooling water system	21,446	21,610
Accessory electric plant	52,058	79,709
Instrumentation and controls	26,766	26,766
Improvement to site	17,986	17,891
Building and structures	16,194	16,109
Total plant cost ($1000)	1,243,167	1,440,318
Initial operating cost ($)	4,835,833	2,894,967
Fixed O&M cost ($/year, at above CF)	25,775,385	29,794,020
Variable O&M cost ($/year, at above CF)	4,649,272	23,981,935
Annual fuel feed cost ($/year, at above CF)	57,563,479	57,563,479
Annual CO_2 emission cost ($/year, at above CF)	4,069,183	2,006,932
Annual H_2 export revenue ($/year, at above CF)	118,215,219	53,137,144
Cost of H_2 production ($/tonne)	2620	1178

[a] Consists of SOFC air blower and cathode exhaust combustor/expander system.

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 H_2 is determined for the two cases using a 20 year levelized cost of electricity of $135/MW h 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 H_2 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 feed stock; (2) co-produce power and high purity H_2; (3) capture most of the CO_2 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 tonne/D of H_2 (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 H_2) 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 H_2 is determined for the two cases using a 20 year levelized cost of electricity of $135/MW h based on data developed by NETL for IGCC and boiler plants fueled by a bituminous coal and equipped with 90% carbon capture. The resulting cost of producing the coproduct H_2 is significantly lower for the IGFC case, 1178/tonne versus $2620/tonne for the IGCC case.

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