1 Coproduction of transportation fuels 2 in advanced IGCCs via coal and 3 biomass mixtures

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7

8 **Abstract**

9 Converting abundant fossil resources of coal to alternative transportation fuels is a
10 promising option for countries heavily dependent on petroleum imports if plants are equipped
11 with carbon capture for sequestration and cofed with biomass (30% by weight of the total feed
12 on a dry basis), an essentially carbon neutral fuel, without penalizing the process economics
13 excessively. A potential exists to improve both thermal efficiency and economics of such plants
14 by taking advantage of the synergies of coproducing electricity using advanced technologies
15 under development. Three types of transportation fuels are considered. Fischer–Tropsch (F-T)
16 liquids consisting predominantly of waxes could be processed in existing refineries while
17 displacing petroleum and the refined products introduced into the market place at the present
18 time or in the near term without requiring changes to the existing infrastructure. Ethanol could
19 potentially serve in the not so distant future (or phased in by blending with conventional liquid
20 fuels). Hydrogen which could play a dominant role in the more distant future being especially
21 suitable to the fuel cell hybrid vehicle (FCHV). Two types of coal along with biomass cofeed are

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evaluated; bituminous coal at \$42.0/dry tonne, lignite at \$12.0/dry tonne, and switchgrass at \$99.0/dry tonne. The calculated cost for F-T liquids ranged from \$77.8/bbl to \$86.6/bbl (or \$0.0177 to 0.0197/MJ LHV) depending on the feedstock, which is comparable to the projected longer term market price of crude oil at ~\$80/bbl when supply and demand reach a new equilibrium [Lafakis. Moody's Analytics. <http://money.cnn.com/2015/01/09/investing/oil-price-bottom/index.html> (accessed on 1-12-15)][1] (or ~\$0.0172/MJ LHV). It should be noted, however, that F-T liquids contain no sulfur or nitrogen compounds and no inorganics. The calculated cost of fuel grade ethanol ranged from \$4.84 to 4.91/gal (or \$0.0566 to 0.0582/MJ LHV), while the price of gasoline in the U.S. amounted to \$0.0240 to 0.0279/MJ LHV when crude oil was at ~\$80/bbl. Ethanol coproduction may not appear to be as attractive as the other options at these scales, primarily due to the much lower plant efficiency. However, from a life cycle greenhouse gas emissions standpoint, ethanol produced with biomass cofeeding and CCS, have a lower carbon footprint than gasoline or diesel, especially when derived from petroleum. The calculated cost of hydrogen ranged from \$1.87 to 2.13/kg (or \$0.0156 to 0.0177/MJ LHV), which is significantly lower than the DoE announced goal of \$3.00/kg in 2005.

Keywords

IGCC, biomass, hydrogen, Fischer–Tropsch, ethanol

Nomenclature and acronyms

ASU	air separation unit
bbl	“blue” barrel (42 gallons)
BFW	boiler feed water
CCS	carbon capture and storage
COE	cost of electricity
DoE	Department of Energy
FCHV	fuel cell hybrid vehicle
F-T	Fischer-Tropsch

48	HHV	higher heating value
49	HRSG	heat recovery steam generator
50	IGCC	integrated gasification combined cycle
51	ISO	international organization for standardization
52	ITM	ion transport membrane
53	LHV	lower heating value
54	LP, MP, IP, HP	low, medium, intermediate, high pressure
55	tonne	metric ton (1000 kg)
56	ppmw	parts per million by weight
57	ppmv	parts per million by volume
58	PSA	pressure swing adsorption
59	RPS	Renewable Portfolio Standard
60	SMR	steam methane reforming
61	ST	short ton
62	TPC	total plant cost
63		

64 **1. Introduction**

65 Coal is a primary energy source with world consumption of this fossil fuel being 162
66 quadrillion kJ in 2012 accounting for 28.5% of total energy consumption on a world-wide basis.
67 Furthermore, it is projected to increase to 232 quadrillion kJ and is expected to maintain a
68 similar share of the total energy consumption in 2040. In the power generation sector, coal-fired
69 plants generated 8.4 trillion kWh of electricity in 2012 or 39.4% of total world electricity
70 generation while 13.9 trillion kWh of electricity or 35.6% of total world electricity generation is
71 projected to be generated by coal-based plants in 2040 [2]. Coal-fired power plants are the
72 major emitters of CO₂ emissions, and various other pollutants, such as particulate matter, sulfur
73 oxides, nitrogen oxides and mercury. For example, the U.S. emitted 5,290 million tonne of CO₂
74 in 2012 with electrical power generation accounting for 38.5% of the CO₂ emissions [3].
75 Integrated gasification combined cycle (IGCC) power plants are a clean approach for utilizing
76 coal [4]. IGCC plants designed for carbon capture shift the CO in the syngas generated by
77 gasification of the feedstock to CO₂ by reaction with steam while forming H₂, separating out the
78 CO₂ and combusting the decarbonized clean syngas in the gas turbine of the combined cycle.

79 Since the gas produced by gasification contains CO and H₂, which are the building blocks for a
80 host of coproducts, a potential exists to synergistically combine electricity generation along with
81 synthesis of a coproduct [4, 5, 6].

82 Biomass utilization in power generation has been gaining increasing attention due to its
83 near carbon neutral benefit while political level regulations such as the Renewable Portfolio
84 Standard (RPS) in the U.S. is accelerating this trend [7]. In particular, California is to achieve 33%
85 of renewables as its energy source by 2020 per the California Energy Commission [8]. State-of-
86 the-art gasifiers capable of cofeeding biomass along with coal are available [9, 10].

87 Converting abundant fossil resources of coal to alternative transportation fuels is a
88 promising option for countries heavily dependent on petroleum imports except that the carbon
89 footprint of such plants can be excessive. Equipping such plants with carbon capture for
90 sequestration and cofeeding biomass can overcome this environmental barrier if the captured
91 CO₂ can be sequestered in a safe and reliable manner without penalizing the process economics
92 excessively. A potential exists to improve both thermal efficiency and economics of such plants
93 by taking advantage of the synergies of coproducing electricity, the specific amount of CO₂
94 generated being inversely proportional to efficiency. Use of advanced technologies under
95 development would further improve the plant thermal efficiency and economics.

96 Current research and development activity for advancing IGCC technology include
97 membrane based air separation [11, 12] to separate O₂ at high temperatures, warm gas cleanup
98 processes to remove pollutants and CO₂ at relatively high temperatures [13, 14], and advanced
99 gas turbines to serve as high efficiency power generators [15]. In order to fully realize the
100 benefits of these advanced technologies, it is imperative to identify plant integration concepts

that not only improve the efficiency of one single unit of the plant, but also provide synergistic enhancement of the overall plant efficiency and process economics.

This paper investigates conceptual designs using these above feedstock blends and the advanced technologies for IGCC plants designed with CCS to assess the coproduction of three types of transportation fuels suitable for introduction into the market place at different time frames. F-T liquids could be processed in existing refineries while displacing petroleum and the refined products introduced into the market place at the present time or in the near term without requiring changes to the existing infrastructure. Fuel grade ethanol (97.7 mole % ethanol mixed with other alcohols resulting in an average carbon number of 2.01) could potentially serve in the not so distant future or could be phased in by blending with conventional liquid fuels. Hydrogen could play a dominant role in the more distant future being especially suitable to the FCHV. A number of system studies have been conducted to evaluate the merits of coproduction from coal [15, 16, 17, 18, 19, 20, 21, 22] but our paper focusses on the coproduction of three transportation fuels on a consistent basis to bridge the near term to the future while utilizing environmentally sustainable biomass as a cofeed and advanced technology under development.

2. Approach

About half the coal produced in the U.S. is bituminous with carbon content approximately ranging from 65% to 85% resulting in a high energy density. Lignite is another major coal resource despite its much lower energy density, and accounts for about 7% of total U.S. produced coal [23]. Due to the significant differences in heating value, ash, sulfur and moisture content between the two types of coals, plant performance and costs can differ significantly, and these two major coal types cover each end of the spectrum. Illinois No. 6 is

utilized as representative of the bituminous coal while the lignite is from N. Dakota. All data presented for these feedstocks except that for the Hg content of Illinois No. 6 coal and Cl content of N. Dakota lignite are taken from U.S. Department of Energy (DoE) Funding Opportunity Number: DE-FOA0000496 [24]. Data for the switchgrass is obtained from DoE/NETL Report 2012/1546 [25]. The moisture content for the switchgrass corresponds to covered field drying. Table 1 presents the coal and biomass ultimate analyses and heating value data. The proposed amount of biomass in the cofeed cases is 30% of the total feed on a dry weight basis based on successful operation of the Shell gasifier with coal and biomass mixtures [26].

Herbaceous energy crops such as switchgrass grown on purpose can provide constant and significant supply of the biomass resource. Switchgrass which is a fast growing crops has been successfully utilized in gasification facilities [27] and can be grown on marginal or degraded lands and avoid competition over agricultural lands, as promoted by the Conservation Reserve Program (CRP) [28] led by the U.S. Department of Agriculture (USDA) Natural Resources Conservation Service (NRCS). Cofeeding biomass along with coal appears to be a strategy for building large scale gasification plants which can take advantage of economies of scale. Furthermore, this strategy can circumvent any seasonal variability in the supply of biomass; the coal rate may be increased to compensate for the reduction in biomass supply as long as the feed handling system is properly designed.

Ambient conditions can have a significant effect on the performance and economics of power plants, especially gas turbine based, and it is essential that site specific conditions be specified. ISO ambient conditions are chosen to be representative for Illinois No. 6 bituminous coal case site, i.e., 15°C dry bulb temperature, 60% relative humidity, and sea-level barometric pressure. Since the lignite is from N. Dakota and lignite is typically not transported over

significant distances due to its much lower energy density, site ambient conditions more specific to N. Dakota are specified consisting of 4.4°C dry bulb temperature, 40% relative humidity, and elevation of 579 m above sea level, the corresponding lower barometric pressure ends up derating the gas turbine power output due to the lower air density.

Coal only IGCC plants without coproduction, developed to establish the cost of electricity for use in the economic analysis of the coproduction cases, capture 90% of the carbon present in the particulate free syngas. Biomass cofed cases “capture” 80% of the carbon (in the form of CO₂ for CCS and as carbon in the coproduct) in the particulate free syngas to take credit for the renewable nature of the biomass feedstock, consistent with basis defined in DoE Funding Opportunity Number: DE-FOA0000496 [24]. Even with this lower amount of capture for biomass cofed cases (which are cofed with 30% by weight of the total feed as biomass on a dry basis) a substantially lower CO₂ emission on a net MW basis is realized.

Coproduced F-T Liquids are stabilized but not refined and will have to be shipped to a refinery for further processing; the ethanol is made as fuel grade mixed alcohols; and the hydrogen is specified to be suitable for FCHV quality with molar purity of 99.99%.

The criteria used in setting the split between electricity and coproduct in each case is:

$$\frac{\text{Net MW} \times 3.6 \times 10^6 \text{ KJ/MWh}}{\text{Net MW} \times 3.6 \times 10^6 \text{ KJ/MWh} + \text{KJ/h HHV Coproduct}} = 50\%$$

A split of $\geq 50\%$ split was stipulated by the DoE who funded this project. For this study, the split was set at 50% for each case but it should be noted that this split could have a significant effect on the overall plant economics and a future study is recommended to optimize its value.

3. Plant configuration and integration concepts

The plants are configured with promising technologies under development with the main objective being to improve not only the performance of individual plant subsections but also the overall plant cost by identifying synergistic thermal integration of different plant subsections. Table 2 summarizes the technologies selected for major plant subsections for the gasification island and for the coproduction units. Technologies for the gasification island and their integration concepts are discussed in detail in a previous paper by the authors [29]. The plant steady state simulations are performed on Aspen Plus®, a process engineering flow sheet modeling system offered by AspenTech using thermodynamic models with appropriate approaches to equilibrium. The advanced technologies such as the ITM, syngas desulfurization and decarbonization, and ethanol synthesis using Rh based catalyst are under development. These systems were modeled using smaller scale test data and thus uncertainty exists on how the performance or yields would scale-up.

3.1 Electricity only

The resulting overall process scheme for the electricity only IGCC plants with CCS is depicted in Fig. 1 and only the advanced technology subsystems are described in the following.

Air Separation Unit (ASU). O₂ is separated from air using a high temperature ion transport membrane (ITM) which allows permeation of oxygen in ionic form. ITMs are projected to provide a 30% reduction in both power consumption and capital cost when compared to the conventional cryogenic unit. Feed air to this unit is extracted from the gas turbine compressor, further compressed to 25.7 bar in a booster compressor, and then heated up in an interchanger against the depleted air stream from the ITM followed by further heating to 850°C by directly firing decarbonized syngas into the air stream. It then enters the ITM unit which is modeled

using methodology developed by Air Products and Chemicals, Inc. [30]. Permeate oxygen stream is cooled in a heat recovery steam generator (HRSG) while generating high pressure (HP), intermediate pressure (IP) and low pressure (LP) steam. Finally it is cooled against cooling water and then compressed by a multistage intercooled compressor before it is provided to the gasifier, while non-permeate oxygen-depleted air is recycled back to the gas turbine. Some O_2 is also provided to a H_2SO_4 unit which produces this saleable byproduct from the sulfur compounds separated out from the syngas

Gasification and Syngas Cleanup. The dry feedstock is transported by CO_2 since there is no N_2 available (in a conventional IGCC utilizing a cryogenic ASU, high purity N_2 can be provided) and is fed to the gasifiers along with O_2 and steam via a top-mounted feed injector. Almost complete conversion (99.5%) of feedstock carbon occurs at a temperature of $1300^\circ C$ to $1400^\circ C$. Hot raw syngas and liquid slag are discharged from the gasifier reaction chamber and the raw gas is cooled to $677^\circ C$ in a radiant syngas cooler while generating HP steam. Coarser particulates are removed from raw syngas by barrier filter operating at this same temperature of $677^\circ C$ [31], followed by scrubbing with recycled water to remove water soluble components including alkalis and halides. The slag discharges into a water bath at the bottom of the radiant syngas cooler and is cooled to temperatures of $\sim 220^\circ C$ and the molten slag solidifies to an environmentally benign nonleachable vitrified granulate. The syngas now at $189^\circ C$ is then preheated in a feed/effluent interchanger and desulfurized by the regenerable ZnO process such as that developed by RTI[13] by the overall reactions: $H_2S + ZnO = ZnS + H_2O$ and $COS + ZnO = ZnS + CO_2$ $\sim 260^\circ C$, a temperature much higher than that of a conventional desulfurization process which requires cooling of the feed gas to near ambient temperatures causing both efficiency and plant cost penalties. The regenerator off-gas containing SO_2 obtained by the reaction: $ZnS + O_2 = ZnO + SO_2$, is fed to a H_2SO_4 unit to produce a saleable byproduct. The

desulfurized syngas after cooling in the feed/effluent interchanger enters fixed-bed sorption process such as that developed by TDA [32] for Hg removal. Some of the NH_3 and HCN are also captured by this sorbent.

Syngas Shifting and Decarbonization. The purpose of this unit is to convert most of the CO in the syngas to H_2 by means of the water gas shift reaction: $\text{CO} + \text{H}_2\text{O} = \text{H}_2 + \text{CO}_2$. This conversion step is crucial to the overall carbon capture of the IGCC plant. NH_3 in the feed passes through the shift reactor unchanged and without affecting the catalyst performance. On the other hand, HCN will be hydrogenated to CH_4 and N_2 . Sufficient steam is injected into the syngas and then fed to the shift unit. The heat evolved by the exothermic shift reaction is used to generate steam. The shifted syngas at a temperature of $\sim 200^\circ\text{C}$ is combined with recycle gas exiting the CO_2 purification unit and then fed to the TDA fixed-bed sorption unit [14, 33] for decarbonizing the syngas. Regeneration of the sorbent is accomplished utilizing steam. The mixture consisting of desorbed CO_2 , steam and residual syngas is cooled in a series of heat exchangers consisting of generating LP steam, vacuum condensate/makeup boiler feed water (BFW) heating and finally trim cooling against cooling water. The gas is then compressed, cooled while recovering bulk of the heat for vacuum condensate/makeup BFW heating, dehydrated and fed to a purification unit. The purification may be done cryogenically or simply fed to a catalytic oxidizer to burn off the combustibles present in this crude CO_2 stream depending on the required Ar content of the final CO_2 product. The purified CO_2 stream is then split into fractions as required for in-plant usage (such as gasifier feed transport), and pressurized to the required pressures. The decarbonized syngas leaving the adsorption unit at a temperature of $\sim 250^\circ\text{C}$ with its accompanying steam is supplied to the gas turbine along with the depleted air from the ASU. The accompany steam provides additional motive fluid in the turbine as well as a thermal diluent to reduce the flame temperature in the gas turbine combustor and thus NO_x formation.

Power Block. The combined-cycle power block consisting of the H class gas turbine exhausts into the HRSG which provides steam for the reheat steam turbine. The decarbonized syngas and the depleted air from the ASU are injected into the gas turbine combustor through separate nozzles. A reheat steam cycle consisting of HP steam at 166.5 bar pressure, and superheat and reheat temperatures of 538°C is utilized. Heat is rejected from the steam turbine exhaust via a mechanical draft cooling tower.

3.2 Coproduction

The coproducing IGCC configurations are similar to the previously described scheme for electricity only except in the feed handling systems (the coproduction plants being cofed with biomass), and due to the addition of the coproduction unit itself. Major differences in the configuration are described in the following:

Air Separation. Since a significant portion of the syngas is utilized for producing the coproduct, the relative size of the gasification island in relation to the power island is much larger than that in the Electricity Only case. This results in the amount of air that may be extracted from the gas turbine not being sufficient for the ASU, since the maximum amount that may be extracted is typically limited to 50% of the total gas turbine inlet air for reasons such as not starving the gas turbine combustor of its liner cooling medium. Thus, a compressor is included in the ASU to supply the additional air required by the ASU. In the F-T liquids and ethanol coproduction cases, the ASU provides O₂ to an autothermal reformer located in the synthesis unit, in addition to the gasifier and the H₂SO₄ unit.

F-T Liquids Synthesis. Once through synthesis of coproducts such as F-T liquids and ethanol while supplying the unconverted gas to the gas turbine has been shown to be advantageous but in the context of no CCS. These advantages include avoiding the buildup of

inerts in the feed to the synthesis reactor, and eliminating the large recycle compressor and associated equipment for the unconverted gas. Since our plant designs do require a significant degree of CCS, a screening analysis is first conducted to quantify the effect of purge rate from the synthesis loop on the overall system efficiency. For a F-T liquids coproduction plant, it is found that the electrical heat rate is increased by as much as 1.46% when the purge rate is increased to 10% of the total recycle. When the purge rate becomes higher than a threshold amount, decarbonization of the purge gas is also required to limit the overall plant carbon emissions to the required value. The reduction in efficiency is due to the requirement for shifting and reforming the purge gas before the carbon may be captured as CO₂ which creates major irreversibilities from a thermodynamic standpoint. These additional equipment end up also penalizing the plant cost on a net basis. It should also be noted that with the ITM technology utilized for air separation in our work, the inerts buildup due to a reduction in the purge rate is less significant than for a plant utilizing cryogenic air separation.

The resulting plant configuration is depicted in Fig. 2. A fraction of the clean syngas before it enters a 2nd shift reactor is combined with a fraction of the unshifted syngas in order to obtain the specified H₂/CO ratio of ~1 for the synthesis reactor feed gas, and passed through a bed of ZnO sandwiching a COS hydrolysis catalyst to remove the trace amounts of sulfur compounds present. The desulfurized syngas is then expanded through a power recovery turbine to near synloop pressure (nominally at 24 barA) and combined with the synloop recycle syngas. The combined stream is then cooled in a series of heat exchangers, fed to an amine wash unit to remove most of the CO₂, and preheated in a feed/effluent interchanger before being fed to a slurry reactor with Fe based catalyst particles suspending in an inert hydrocarbon liquid (a mineral oil). The mineral oil acts as a temperature moderator and a heat removal medium, transferring the heat of reaction from the catalyst surface via the liquid slurry to

boiling water in an internal tubular heat exchanger. IP steam is generated from the heat. The major overall reactions occurring are: $(2n+1) \text{H}_2 + n \text{CO} = \text{H}-(\text{CH}_2)_n-\text{H} + n \text{H}_2\text{O}$ and $\text{CO} + \text{H}_2\text{O} = \text{CO}_2 + \text{H}_2$. The reactor effluent at 260°C is cooled in a series of heat exchangers including the feed/effluent interchanger. The condensate collected is fed to the product stabilization unit which consists of a column to remove the dissolved light ends. The recycle gas which contains CH_4 and other undesirable hydrocarbons (undesirable in the context of maximizing the F-T liquids) is compressed and fed to an autothermal reformer to convert the hydrocarbons back to H_2 and CO . The F-T liquids contain significant amounts of high molecular weight waxes and require hydrocracking to produce lower molecular weight hydrocarbons followed by distillation for separating out the naphtha, diesel, and lighter molecular weight hydrocarbon fractions. It is assumed that these operations will be conducted at an existing refinery.

Ethanol Synthesis. The results obtained for the F-T liquids synthesis loop with respect to minimizing the purge rate are assumed to apply to ethanol coproduction where shifting and reforming of the unconverted gas is also required. The resulting plant configuration is depicted in Fig. 3. As in the previous case, a fraction of the clean syngas before it enters the 2nd shift reactor is combined with a fraction of the unshifted syngas but to obtain a higher H_2/CO ratio of 2.0 at the synthesis reactor inlet. Again, the combined gas is passed through a bed of ZnO sandwiching a COS hydrolysis catalyst to remove the trace amounts of sulfur compounds present. The desulfurized syngas is then expanded through a power recovery turbine to near synloop pressure which is lower in this case (14 barA at the synthesis reactor outlet) and combined with the synloop recycle syngas. The combined stream after cooling in a series of heat exchangers and removal of most of the CO_2 in an amine wash, is preheated in a feed/effluent interchanger before being fed to a fixed bed reactor containing Rh based catalyst. The exothermic reaction heat is transferred to boiling water in an internal tubular heat exchanger to

generate IP steam. The major overall reactions occurring are: $2n \text{ H}_2 + n \text{ CO} = \text{C}_n \text{ H}_{2n+1} \text{ OH} + (n-1) \text{ H}_2\text{O}$ (with n predominantly = 2), $3 \text{ H}_2 + \text{CO} = \text{CH}_4 + \text{H}_2\text{O}$ and $\text{CO} + \text{H}_2\text{O} = \text{CO}_2 + \text{H}_2$. The reactor effluent at 285°C is cooled in a series of heat exchangers including the feed/effluent interchanger. The recycle gas which contains CH_4 and CH_3OH is compressed and fed to an autothermal reformer to convert these compounds back to H_2 and CO in order to maximize the production of the ethanol. The condensate collected is fed to the purification unit which consists of a set of energy saving heat integrated distillation columns (HP and LP columns with the condenser of the HP column providing heat for the reboiler of the LP column) to produce the azeotropic mixture of $\text{C}_2\text{H}_5\text{OH}$ and H_2O which is then dehydrated using molecular sieves to meet the required specifications.

Hydrogen Coproduction. As seen in the plant configuration depicted in Fig. 4, a fraction of the H_2 rich decarbonized syngas (after shifting and CO_2 separation) before entering the power block is fed to a pressure swing adsorption (PSA) unit to produce the higher purity coproduct meeting the specifications as defined in the design basis. The tail gas from the PSA is compressed and combined with the remaining of the decarbonized syngas and combusted in the gas turbine. Coproduction of H_2 is thus easily accomplished in an IGCC designed for CCS.

4. Results and discussion

Tables 3 and 4 summarize the plant performances for the bituminous and the lignite cases. The plant feed rates for each of the cases are determined to fully load the H class gas turbine under the site specific ambient conditions. Thermal efficiencies for the electricity only cases with 90% carbon capture are 38.3% and 36.8% (HHV basis) with the bituminous and the lignite feedstocks respectively, the lignite case having a lower efficiency primarily due its higher moisture content. For the coproduction cases (where 50% of the energy exported is in the form

of electricity), the electrical efficiency, as expected, is highest for the hydrogen coproduction cases. The plant thermal efficiency reduces as a greater fraction of chemical energy contained in the raw syngas is degraded to heat. In the hydrogen coproduction case, this degradation occurs by the shift reaction but its exothermicity is quite small when compared to the reactions occurring in the F-T as well as the ethanol synthesis reactors.

The electrical efficiencies for F-T coproduction cases are higher than those for the ethanol coproduction cases but it should be noted that the ethanol coproduction cases produce the finished coproduct while the F-T coproduction cases produce a coproduct that requires further processing in a refinery. The cross comparison of the thermal performance between the various coproduct cases is further complicated by the fact that the carbon footprint is not the same when carbon leaving with the coproduct is accounted for. The lower efficiency of the ethanol cases is partly due to a low per pass conversion (< 10 mole % of the CO and ~15 mole % of the recycled methanol) to ethanol (and higher alcohols) requiring a large recycle rate. The thermal efficiency and consequently the overall plant economics can be enhanced if the methanol formed as a side-product is not recycled to the synthesis reactor but purified and sold as an additional coproduct.

The efficiency of both the F-T and the ethanol cases can be enhanced if the CH₄ formed during synthesis is sold as an additional coproduct, substitute natural gas after separation and purification. The impact on plant economics however is dependent on the prevailing natural gas price which is being impacted by shale gas prices. Tables 5 and 6 present the plant cost estimates and results of the economic analysis for the bituminous and the lignite cases. The 1st year levelized cost of electricity for each of the two Electricity Only (bituminous and lignite) cases is first developed and then the coproduct cost is derived such that the same 1st year

levelized cost of electricity as the corresponding Electricity Only case (i.e., with the same coal) results.

The first year cost of electricity calculated for the bituminous coal is \$103/MWh while that for the lignite is \$108/MWh. The calculated cost of the F-T liquids ranges from \$77.8/bbl to \$86.6/bbl (or \$0.0177 to 0.0197/MJ LHV) depending on the feedstock. The cost of the fuel grade ethanol ranges from \$4.84/gal to \$4.91/gal, while it ranged from \$2.20/gal to \$3.70/gal in a DoE funded study conducted by Louisiana State University [DOE Award: DE-FC26-06NT43024]. The Louisiana State University study consisted of a significantly larger plant than our study and benefited from economies of scale. When the plant size in our study is scaled up to similar size as in the Louisiana State University study, cost of alcohol is then reduced to a range of \$3.24/gal to \$4.28/gal, which is comparable. The cost of hydrogen ranges from \$1.87/kg to \$2.13/kg. In all cases, the coproduct cost is lower for the bituminous coal cofed plants than the lignite cofed plants.

5. Summary and conclusions

Three types of transportation fuels are considered, Fischer–Tropsch (F-T) liquids which could be refined in existing refineries while displacing petroleum and the refined products introduced into the market place at the present time or in the near term without requiring any changes to the existing infrastructure, ethanol could potentially serve in the not so distant future (or phased in by blending with conventional liquid fuels), and hydrogen which could be the fuel of the future being especially suitable to the FCHV. Two types of coal (bituminous and lignite) along with switchgrass as the biomass cofeed are included. Among the coproduction cases, the electrical efficiency as expected is highest for the hydrogen coproduction. The

electrical efficiency for F-T coproduction is higher than that for the ethanol coproduction but it should be noted that the ethanol coproduction plants produce the finished coproduct while the F-T coproduction plants produce a coproduct that requires further processing in a refinery. The cross comparison of the thermal performance between the various coproduct cases is further complicated by the fact that the carbon footprint is not the same when carbon leaving with the coproduct is accounted for. The F-T liquids cost is comparable to the current market price of crude oil. It should be noted, however, that F-T liquids contain no sulfur or nitrogen compounds and no inorganics. The cost of the ethanol is higher than the current price of gasoline in the U.S., and thus the ethanol coproduction may not appear to be as attractive as the other options at these scales, primarily due to the much lower plant efficiency. However, from a life cycle greenhouse gas emissions standpoint, ethanol produced with biomass cofeeding and CCS, have a lower carbon footprint than gasoline or diesel, especially when derived from petroleum. The calculated cost of hydrogen is significantly lower than the DoE announced goal of \$3.00/kg in 2005.

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FIGURE CAPTIONS

Fig.1. Block Flow Diagram – Electricity Only IGCC Plants with CCS (stream data in [43])

Fig.2. Block Flow Diagram – Electricity and Fischer Tropsch Liquids Coproduction IGCC Plants CCS (stream data in [43])

Fig.3. Block Flow Diagram – Electricity and Ethanol Coproduction IGCC Plants CCS (stream data in [43])

Fig.4. Block Flow Diagram – Electricity and H₂ Coproduction IGCC Plants CCS (stream data in [43])

Figure 1

Figure 2

Figure 3

Figure 4

Table 1

Table 1 Coal and Biomass Properties

	Illinois No. 6 Bituminous		N. Dakota Lignite		Switchgrass Biomass	
	Dry Basis, %	As Received, %	Dry Basis, %	As Received, %	Dry Basis, %	As Received, %
Carbon	71.72	63.75	61.88	39.55	42.60	36.21
Hydrogen	5.06	4.5	4.29	2.74	6.55	5.57
Nitrogen	1.41	1.25	0.98	0.63	1.31	1.11
Sulfur	2.82	2.51	0.98	0.63	0.01	0.01
Chlorine	0.33	0.29	18 ppmw [34]	12 ppmw [34]	0.04 [35]	0.03 [35]
Ash	10.91	9.7	15.43	9.86	7.41	6.30
Moisture	0.00	11.12	0.00	36.08	0.00	15.00
Oxygen	7.75	6.88	16.44	10.51	42.08	35.77
Total	100	100	100	100	100.00	100.00
HHV, kJ/kg	30,531	27,135	24,253	15,391	18,113	15,396
LHV, kJ/kg	29,568	26,172	23,334	14,803	16,659	14,161
Mercury, ppm		0.18 [36]		0.116		

Table 2**Table 2 Advanced technologies selected**

Plant subsection	Technology	Comments
Air Separation	ITM	Being developed by Air Products and Praxair [11, 12]
Biomass Pretreatment	Drying using vapor recompression and size reduction by milling	Co-feeding biomass with coal into a pressurized dry feed (Shell) gasifier at proposed levels demonstrated at Nuon IGCC (particle size < 1 mm) [26]. Experimental data for hammer mill power for switchgrass with 6 - 12% moisture available. Milling power not a significant impact on plant thermal performance.
Gasifier feeding	Solids pump	Using recycle CO ₂ as inerting and injection gas. Maximum discharge pressure limited to 8.3 MPa (1200 psi) based on PW gasifier demonstration program [38, 39].
Gasifier	Siemens type dry-feed entrained-bed operating at commercially offered operation pressure of ~41 bar (600 psi) and not at higher unproven pressure	Siemens gasifier with multiple feed nozzles is available [9]. The operating pressure of the Siemens gasifier for current projects is 4 MPa while their patents site a maximum pressure of 10 MPa.
Syngas Desulfurization	Humid gas cleanup using RTI's ZnO treating process	Regenerable process while producing H ₂ SO ₄ from the regenerator off-gas [13].
Syngas Mercury Removal	Humid gas cleanup using TDA's process	A "throw away" process, Hg content of feedstock being low [32].
Syngas Shifting	Sulfur tolerant catalyst	Adiabatic beds with intercooling in series depending on the degree of shifting required with bypass control for CO to H ₂ ratio [36].
Syngas CO ₂ Separation	Humid gas cleanup using absorption beds such as the TDA process	Regenerable process using steam to strip off the CO ₂ [14, 33].
F-T Synthesis	Slurry reactor containing Fe based catalyst	H ₂ /CO molar ratio of 1 in feed gas [40].
Ethanol Synthesis	Fixed bed reactor containing Rh based catalyst to synthesize primarily ethanol	H ₂ /CO molar ratio of 2 in feed gas [41].
Power Island	H class gas turbine based combined cycle	It is assumed that this type of gas turbine will be offered for syngas applications in the time frame of these advanced gasification plants. Firing temperature of the gas turbine

		reduced from 1428°C on natural gas to 1392°C for the syngas [42].
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Table 3 Plant thermal performance - bituminous coal cases

Cofeed		None	Switchgrass		
Coproduct	Units	None	F-T Liquids	Ethanol	Hydrogen
Coal feed rate	tonne/D (dry)	2,950	3,060	3,290	3,350
Biomass feed rate	tonne/D (dry)	-	1,310	1,410	1,440
Total HHV input	GJ/h	3,750	4,890	5,240	5,350
Gas turbine power output	MW	318	318	318	318
Steam turbine power output	MW	147	167	126	166
Total gross power output	MW	465	485	444	484
Total auxiliary consumption	MW	66	160	222	112
Net electric power output	MW	399	325	222	372
Net electric power efficiency, HHV	%	38.3	24.0	15.3	25.0
Coproduct produced	tonne/D	-	593	621	226
Coproduct HHV	GJ/h	-	1,160	792	1,340
Electricity/(Electricity + Coproduct HHV)	%	100	50	50	50
Carbon captured	%	90.0	80.0	80.0	80.0
Surface condenser duty	GJ/h	686	922	457	803
Raw water use	M ³ /m	13.2	17.6	18.7	16.0

Table 4

Table 4 Plant thermal performance - lignite cases

Cofeed		None	Switchgrass		
Coproduct	Units	None	F-T Liquids	Ethanol	Hydrogen
Coal feed rate	tonne/D (dry)	3,620	3,370	3,650	3,630
Biomass feed rate	tonne/D (dry)	-	1,440	1,560	1,560
Total HHV input	GJ/h	3,660	4,500	4,870	4,850
Gas turbine power output	MW	313	313	313	313
Steam turbine power output	MW	141	154	121	150
Total gross power output	MW	454	467	434	463
Total auxiliary consumption	MW	80	179	226	137
Net electric power output	MW	374	288	208	326
Net electric power efficiency, HHV	%	36.8	23.0	15.4	24.2
Coproduct produced	tonne/D	-	529	586	198
Coproduct HHV	GJ/h	-	1,030	748	1,170
Electricity/(Electricity + Coproduct HHV)	%	100	50	50	50
Carbon captured	%	90.0	80.0	80.0	80.0
Surface condenser duty	GJ/h	627	808	415	697
Raw water use	M ³ /m	12.9	16.0	17.0	14.6

Table 5 Plant cost estimates and economics - bituminous coal cases

Cofeed	None	Switchgrass		
Coproduct	None	F-T Liquids	Ethanol	Hydrogen
Feedstock Costs, \$/dry tonne				
Bituminous Coal [36]	42			
Switchgrass*		99		
Net Power, MW	399	325	222	372
Plant Cost, \$10 ⁶				
Fuel Preparation System	147	195	205	207
Air Separation Unit	42	52	61	53
Gasification System	320	386	406	410
Gas Cleanup and Conditioning Systems	267	201	208	316
Power Island	185	188	184	190
Co-production Unit	-	118	446	11
General Facilities	156	166	164	173
Total Plant Cost (TPC), \$10 ⁶	1,117	1,306	1,674	1,360
Total Fixed Operating Cost for Initial Year, \$10 ⁶ /yr	45	54	67	54
Total Variable Operating Costs, \$10 ⁶ /yr	71	121	138	135
1st Year Cost of Electricity (COE), \$/MWh**	103			
Coproduct Cost for Initial Year, \$/kg	-	\$77.8/bbl	\$4.91/gal	\$1.83/kg

* Switchgrass Cost (\$/dry tonne) = $(1.286 \times 10^{-11} \times X^3 - 3.028 \times 10^{-7} \times X^2 + 3.569 \times 10^{-3} \times X + 85.32) \times 1.1025$, where X = switchgrass production rate = feed rate, dry ST/day [25]. The cost for each case was found to be essentially constant due to the small variation in the feed rate and was held constant.

** Methodology consistent with [36].

Table 6 Plant cost estimates and economics - lignite cases

Cofeed	None	Switchgrass		
Coproduct	None	F-T Liquids	Ethanol	Hydrogen
Feedstock Costs, \$/dry tonne				
Lignite [36]	12			
Switchgrass [#]		99		
Net Power, MW	374	288	208	326
Plant Cost, \$10 ⁶				
Fuel Preparation System	213	246	261	259
Air Separation Unit	44	51	59	52
Gasification System	328	377	398	395
Gas Cleanup and Conditioning Systems	252	173	183	280
Power Island	180	181	178	181
Co-production Unit	-	109	384	10
General Facilities	154	160	159	166
Total Plant Cost (TPC), \$10 ⁶	1,171	1,297	1,622	1,343
Total Fixed Operating Cost for Initial Year, \$10 ⁶ /yr	47	53	65	53
Total Variable Operating Costs, \$10 ⁶ /yr	57	101	116	111
1st Year Cost of Electricity (COE), \$/MWh ^{##}	108			
Coproduct Cost for Initial Year, \$/kg	-	\$86.6/bbl	\$4.84/gal	\$2.08/kg

[#] Switchgrass Cost (\$/dry tonne) = $(1.286 \times 10^{-11} \times X^3 - 3.028 \times 10^{-7} \times X^2 + 3.569 \times 10^{-3} \times X + 85.32) \times 1.1025$, where X = switchgrass production rate = feed rate, dry ST/day [25]. The cost for each case was found to be essentially constant due to the small variation in the feed rate and was held constant.

^{##} Methodology consistent with [36].