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Performance and Economic Analysis of Biomass/Coal Co-Gasification IGCC Systems with Supercritical Steam Bottom Cycle, Part 2 – Pre-Combustion Carbon Capture

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ABSTRACT

One of the main advantages of IGCC technology is that it is the only form of power generation that is compatible with pre-combustion carbon capture. Part 2 of this paper will thus focus on analyzing pre-combustion CCS, utilizing both sour-shift and sweet-shift processes and comparing them to each other and to the results of post-combustion CCS from Part 1. Pre-combustion CCS plants are smaller than post-combustion ones, and usually require 25% less energy for CCS due to their compact size for processing fuel flow only under higher pressure (450 psi), versus processing the combusted gases at near-atmospheric pressure. For pre-combustion CCS, sour-shift appears to be superior both economically and thermally to sweet-shift in the current study. Sour-shift is always cheaper, (by a difference of about $600/kW and $0.02-$0.03/kW-hr), easier to implement, and also 2-3 percentage points more efficient.

Adding biomass to the system always reduces the emissions and can even make a plant carbon-negative with as little as 10% biomass by weight. In addition, the efficiency will improve (0.7 points) and power output will also improve (~1%-3% more) for up to 10% biomass ratio (BMR) for the right kind of biomass that has been properly pretreated. Beyond 10% BMR, however, the efficiency begins to drop due to the rising pretreatment costs, but the system itself still remains more efficient than from using coal alone (between 0.2-0.3 points on average). The economic difference is fairly marginal, but the trend is inversely proportional to the efficiency, with CoE decreasing by 0.5 cents/kW-hr from 0%-10% BMR and rising 2.5 cents/kW-hr from 10%-50% BMR. Finally, the CO_2 removal cost for sour-shift is around $20/ton, whereas sweet-shift's cost is around $30/ton, which is much cheaper than that of post-combustion CCS from Part 1: about $60-$70/ton.

NOMENCLATURE

ASU	Air Separation Unit
GT	Gas Turbine
ST	Steam Turbine
HRSG	Heat Recovery Steam Generator
IGCC	Integrated Gasification Combined Cycle
GHG	Greenhouse Gas(es)
AGR	Acid Gas Removal
HP	High Pressure (PSI)
IP	Intermediate Pressure (PSI)
DA	De-aerator
BMR	Biomass Ratio (biomass/feedstock) (wt%)
M.W.	Molecular Weight (lbs/lb-mol)
LHV	Lower Heating Value (Btu/lb)
HHV	Higher Heating Value (Btu/lb)
CoE	Cost of Electricity ($/kW-hr)
O&M	Overhead and Maintenance ($)

INTRODUCTION

Where Part 1 focused primarily on post-combustion CCS, Part 2 will focus on pre-combustion CCS. Pre-combustion CCS is unique to IGCC and allows for the removal of carbon dioxide during the gas cleanup phase of the cycle instead of having to wait until after the gas is already burned by the GT combustor. As such, pre-combustion CCS is not possible for ordinary PC plants or natural gas plants, since the CO_2 hasn't been produced before combustion occurs. Pre-combustion CCS is less costly than post-combustion CCS primarily because only the syngas (i.e. the GT fuel) needs to be processed, as opposed to the significantly larger volume of exhaust gases in the case of post-combustion CCS, which results in a much smaller unit. However, as previously stated, it isn't compatible with most power plants, and the process itself is continuously being improved and cost being reduced.

For pre-combustion CCS, the gas stream will be syngas. As such, since combustion has not occurred yet, there will be significant amounts of carbon monoxide (CO) in the gas mixture. Carbon monoxide cannot be captured at all: there are no solvents commercially available that can capture CO, and the molecules themselves are too small to be affected by any membrane or adsorption material. Therefore, this carbon monoxide has to be *converted* to carbon dioxide first, so that when the CCS process is performed, it can remove the maximum amount of carbon from the syngas stream. This is done by manipulating the water-shift reaction ($CO + H_2O \leftrightarrow$

$CO_2 + H_2$) by adding water vapor (steam) into the syngas (sometimes called the "CO-shift" process when done in this way), forcing the reaction to shift to the right: producing more CO_2, which can be captured or separated from H_2 in the syngas. This H_2 can also be burned as a fuel later on.

Unlike post-combustion CCS, pre-combustion CCS can be implemented in two different ways, based on where in the cycle the CO-shift reactor is located: before the *acid gas removal* (AGR) process or after it. When it occurs before AGR, it is called "*sour-shift*," and either "clean-shift" or "*sweet-shift*" when performed afterwards. This is an important consideration, because when sulfur is in the gas stream alongside the CO_2, and AGR and CCS absorption techniques are very similar processes, it means that the plant can perform AGR and CCS *at the same time* (through sour-shift CCS), using the same equipment. This means that sour-shift CCS is cheaper to implement and can sometimes be retro-fitted onto IGCC plants that have amine-based absorption AGR systems installed.

For sour-shift, this means that the CO-shift process can also be used at the same time as COS hydrolysis, using the same water supply, since the COS hydrolysis reaction ($COS + H_2O \leftrightarrow CO_2 + H_2S$) and the water-shift reaction are very similar. All in all, this makes sour-shift CCS a lot cheaper than sweet-shift CCS. The downside to this is the fact that sour shift requires significant amounts of water to shift, since many processes must occur all at once. Second, there is a lot of waste heat, since the water-shift reaction is *exothermic* in the direction of CO_2, the gas stream gets hot, and must be cooled immediately, especially before the AGR column. All in all, this translates to the fact that more cooling is needed to achieve the necessary temperature range to even perform acid gas removal than that of the same plant without any CO-shift or CCS [1].

As with Part 1, this part of the paper is meant as an addition to the research by Long and Wang [2]'s previous study on IGCC with supercritical steam cycles. Again, for this part, the same main parameters are examined as those of Part 1 (biomass, supercritical steam cycle, CCS), with the exception of using pre-combustion CCS instead of post-combustion CCS.

As well with Part 1, a supercritical steam cycle is also studied here, as it leads to increased efficiency for the Rankine cycle [3-5]. In addition, biomass is added to the feedstock in order to reduce emissions, and hopefully raise efficiency. Torrefaction is employed, as it is known to increase the feedstock heating value while making the physical properties more suitable for feeding [6-8].

PLANT DESIGN

For the pre-combustion plants, the design of the basic plant layout is nearly identical to that of the baseline from Part 1, and the only difference between the main cases outlined here is the location of the CCS plant and CO-shift reactor. The basic plant layout for Part 2 is outlined in Fig. 1, sour-shift CCS in the main schematic and sweet-shift CCS overlaid onto it with numbers corresponding to the subcritical steam cycle (no biomass in either case).

Since this is pre-combustion CCS, it allows for the use of a physical solvent, as opposed to post-combustion CCS, which demands the use of a chemical solvent. For the Selexol® process, the price for this solvent was assumed to be about \$2000/ton.

Figure 2 shows the layout of sour-shift's gas cleanup system, and Fig. 3 shows the layout of the complete sour-shift CCS system. The absorbers themselves operate in a cascade-like manner, with the lean solvent first absorbing CO_2 in one absorber, and then sliding down to absorb H_2S in a second absorber. Meanwhile, the syngas enters into the H_2S absorber and flows counter to the solvent, arriving at the CO_2 absorber to undergo carbon capture. This is necessary, because CO_2 and H_2S mix together when under the conditions for AGR, that is, the two compounds will dissolve at the same time. Sequestration implies that the CO_2 will be used for some other purpose, such as advanced oil recovery, which requires an extremely pure stream of CO_2 in order to work properly. Therefore, if H_2S is not removed beforehand, it will require even more work afterward to achieve the right level of CO_2 purity.

The sequestration system makes use of two flash tanks instead of the KO drum from post-combustion CCS. This is because, unlike in post-combustion CCS, the CO_2 absorber isn't directly connected to a stripper column: it will be much easier and less expensive to use flashing to pull the captured CO_2 out of solution, as there are no chemical bonds that need to be broken. In addition, there isn't very much water to separate from the mixture, so there is no need for a KO drum like there is at the end of the H_2S removal stage. The top flash tank strips about 70% of the capturable CO_2 from the solvent, while the lower tank handles the remaining 30% to achieve 90% CO_2 removal.

In addition, since this is physical absorption, there is no condensate to be removed before compression (and sequestration), and no additional cooling water needed, since there are no chemical reactions. Since sour shift occurs before AGR, sufficient steam already exists in the syngas stream and there is no additional steam needed to complete the shift: only the catalyst need be added. Hence, no additional steam need be taken from the steam cycle like for post-combustion CCS and sweet-shift CCS.

Figure 4, likewise shows the gas cleanup system for the sweet-shift cases, while Fig. 5 shows the layout for the sweet-shift CCS system. Notice from the overlay in Fig. 1, as well as in Fig. 4, that there is also additional steam (38.95 lbm/s) added to the CO-shift/CCS block. This is because, since the shifting occurs after the cooling stage, most of the steam that was present has already been condensed out of the syngas. Therefore, additional supplementary steam is necessary to complete the water-gas shift reaction. This is taken from one of the high-pressure (HP) nodes in the HRSG at 1100 PSI and 609°F. The HP stream is used in this case because the syngas is already at a very high pressure at this sweet shift point in the cycle. To avoid sacrificing potential GT power or adding unnecessary auxiliary losses by lowering the pressure of the fuel, "HP process" steam must be used for the CO-shift reaction, as shown in Fig. 6. Since the CO-shift is an exothermic process, a closed-loop cooling stream of water is supplied from the HRSG at 250°F (LTE) and returned to the IP stream (Figs. 4 and 6) as it is in the sour-shift plant.

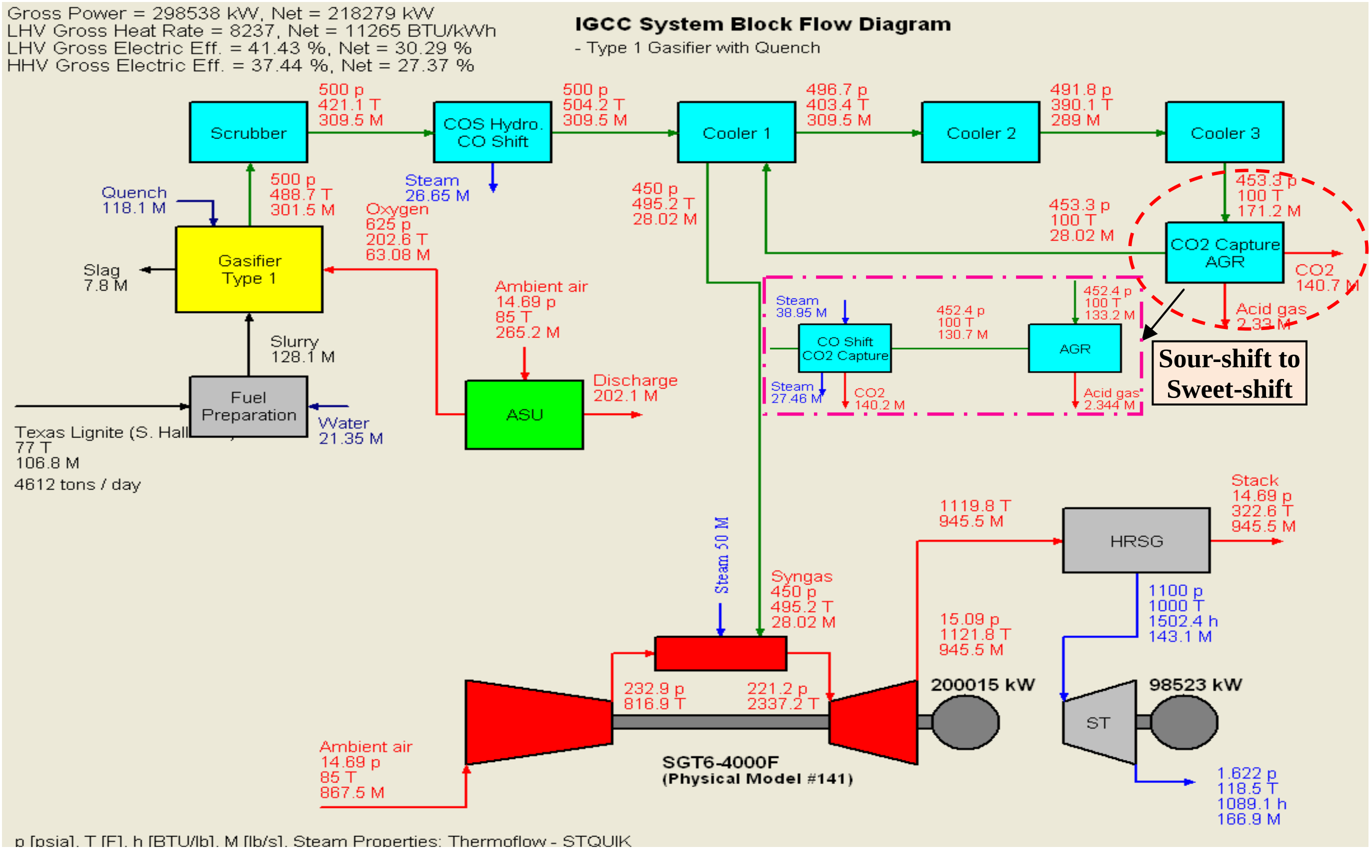

Figure 1 Overall plant layout with sour-shift with no biomass and subcritical steam. For sweet shift, the red dashed circle is replaced by the pink dashed box.

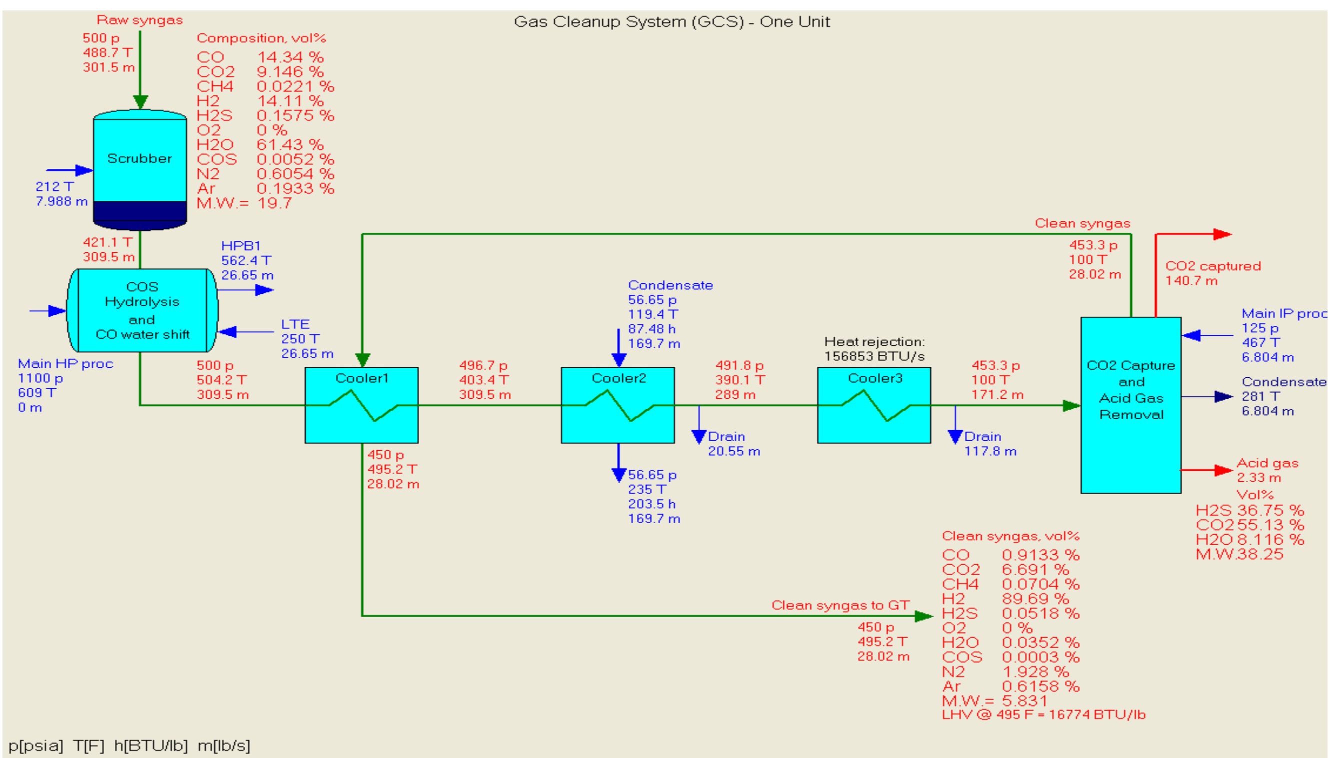

Figure 2 Pre-combustion CCS cleanup system with sour-shift

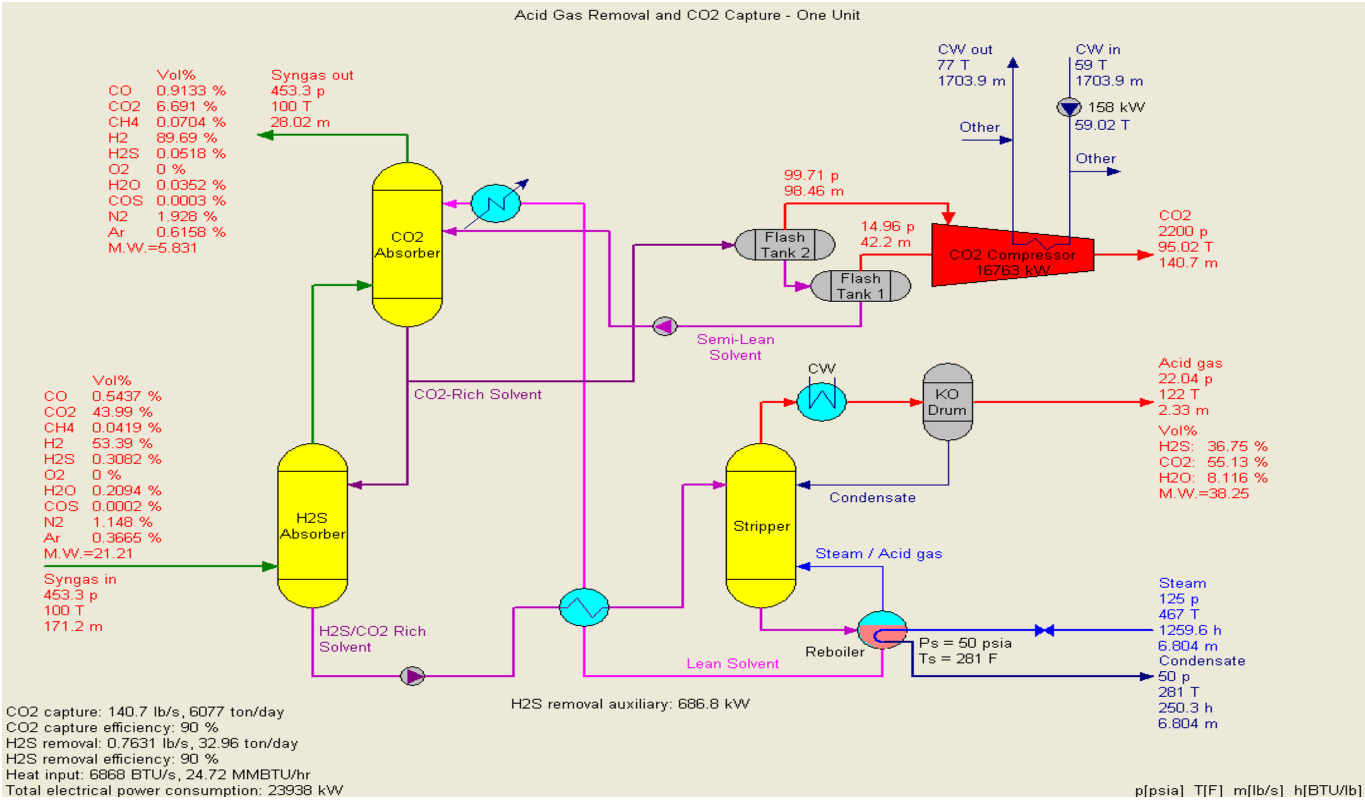

Figure 3 Sour-shift CCS system plus AGR plant

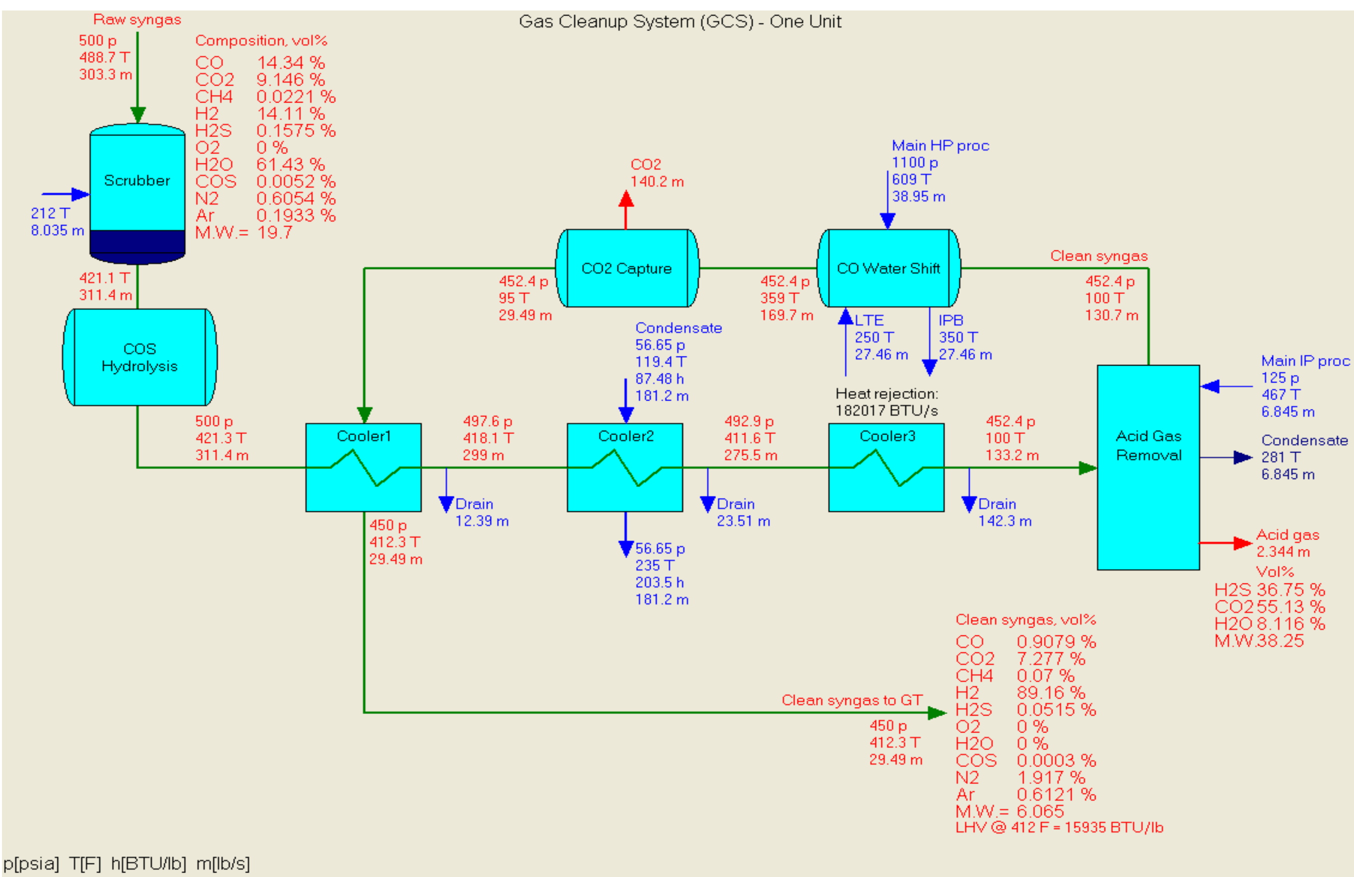

Figure 4 Pre-combustion CCS cleanup system with sweet-shift

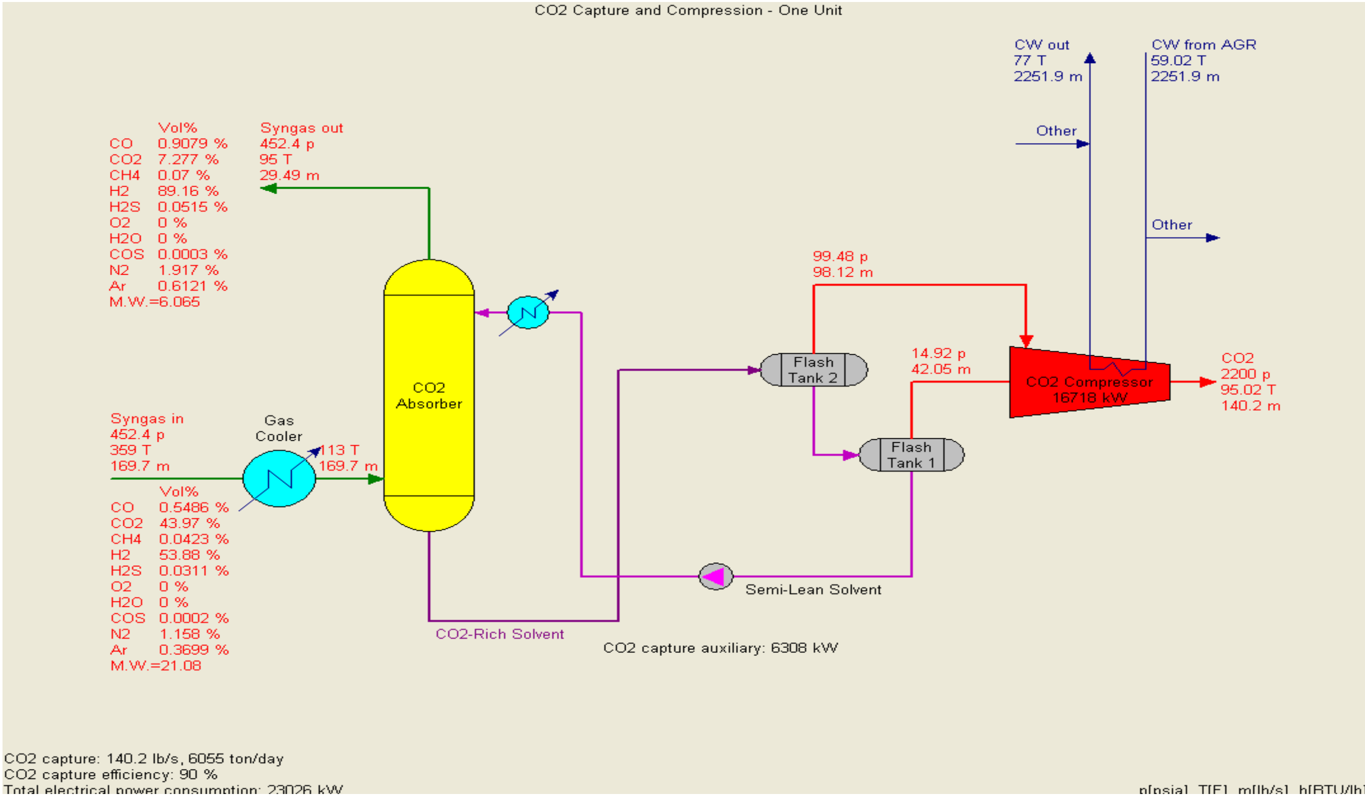

Figure 5 Sweet-shift CCS system

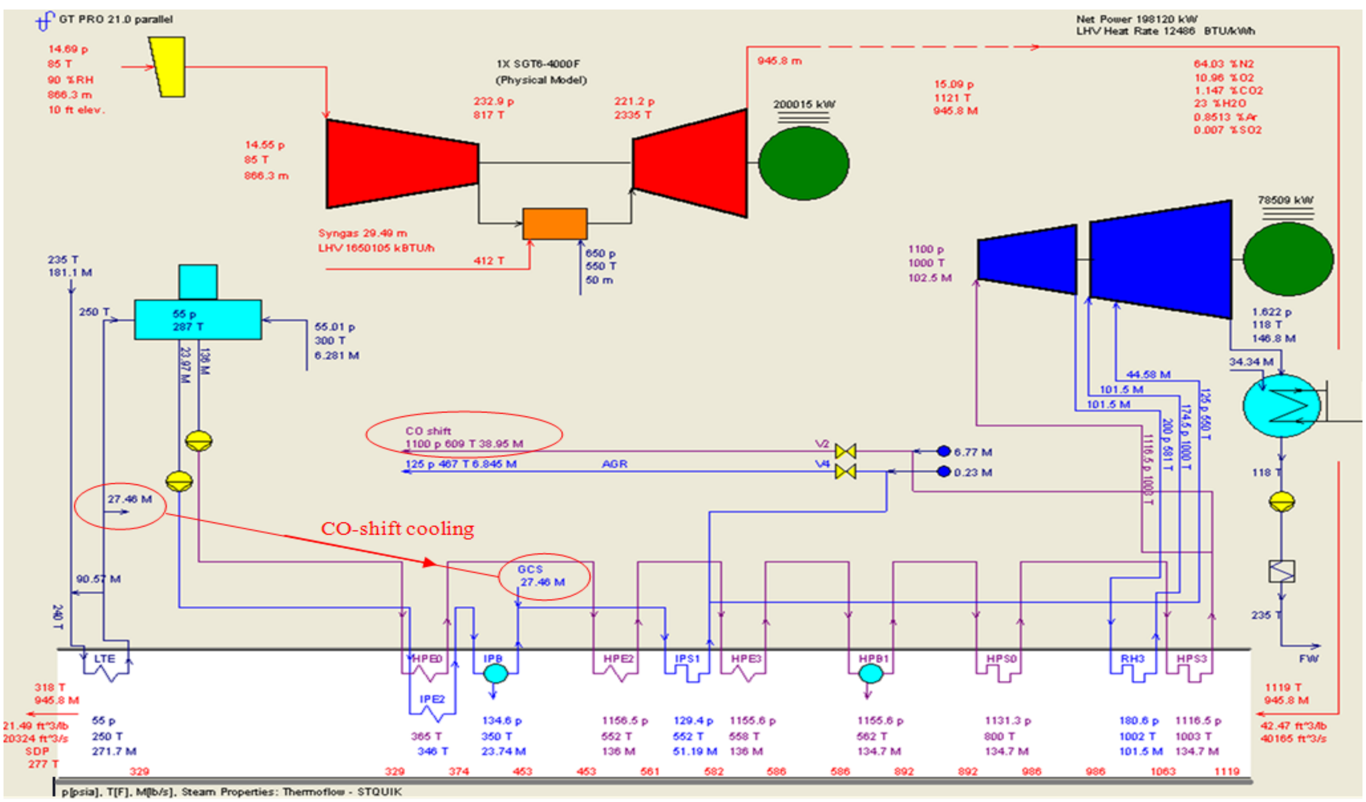

Figure 6 Steam cycle for sweet-shift CCS with CO-shift steam and cooling steam circled

The advantage of sweet shift is that the CO_2 capture plant is more simplistic than sour shift is, so there is less risk of losing some CO_2 during acid gas removal. The CO_2 capture plant for sweet-shift has the same setup and design criteria as those of the CO_2 capture section of the sour-shift plant, with two flash tanks (one stripping 70% of the capturable carbon dioxide and the next handling the remaining 30% to achieve 90% CO_2 capture.), and a CO_2 compressor bringing the pressure to 2200 PSI. The CO_2 compressor, like in the sour-shift case, is used for sequestration purposes. The CO_2 is assumed to be sequestered for a purpose, such as for advanced oil recovery or for permanent storage underground.

As in Part 1, the feedstocks used were Texas Lignite and sugarcane bagasse, whose ultimate analyses can be seen there. The prices assumed were, again, \$19/ton for lignite [9] and \$65/ton for bagasse [10]. The steam cycle's TIT and TIP are fixed for both cycles, with the subcritical cycle having inlet conditions of 1000°F/1100PSI, and the supercritical cycle's being 1200°F/2400PSI. The natural gas used in the duct burner was given a price of \$4.10/mmBtu [11]. All other costs are based on the EIA's 2010 report [12].

RESULTS AND DISCUSSION

The results are analyzed via a cross-comparison with post-combustion CCS and the baseline presented in Part 1. Since some trends are shared/maintained between these cases and those shown in Part 1, the common results will not be repeated here in Part 2. All assumptions and design conditions mentioned in Part 1 are carried over to Part 2. The output power and efficiency data is presented in Tables 1 and 2.

Table 1 Output power and Efficiency (LHV) – Sour-shift

Biomass/Coal Ratio	0%	10%	30%	50%
Subcritical Plants				
Aux. Losses (kW)	80,258	78,444	82,101	85,670
Net Power (kW)	**218,279**	**220,712**	**217,639**	**214,643**
Gross Elect. Eff.	41.43	42.05	42.41	42.76
Net Elect. Eff.	**30.29**	**31.03**	**30.79**	**30.56**
Supercritical Plants				
Aux. Losses (kW)	81,026	79,293	82,977	86,573
Net Power (kW)	**238,077**	**241,559**	**239,387**	**237,309**
Gross Elect. Eff.	42.03	42.72	43.12	43.52
Net Elect. Eff.	**31.36**	**32.16**	**32.02**	**31.89**

Table 2 Power and Efficiency – Sweet-shift

Biomass/Coal Ratio	0%	10%	30%	50%
Subcritical Plants				
Aux. Losses (kW)	80,404	78,586	82,310	85,934
Net Power (kW)	**198,120**	**200,290**	**196,906**	**193,598**
Gross Elect. Eff.	38.42	38.98	39.27	39.56
Net Elect. Eff.	**27.33**	**27.99**	**27.70**	**27.40**
Supercritical Plants				
Aux. Losses (kW)	82,313	80,494	84,221	87,847
Net Power (kW)	**225,406**	**227,536**	**224,130**	**220,819**
Gross Elect. Eff.	39.75	40.29	40.57	40.85
Net Elect. Eff.	**29.12**	**29.76**	**29.49**	**29.22**

Notice from Table 1 that the sour-shift cases are more efficient than the post-combustion cases. Where as many as 8 percentage points of efficiency were lost from the baseline for post-combustion CCS in Part 1, the sour-shift plants are only about 5 percentage points lower than the same baseline cases. An interesting thing to note as well is the fact that the total *steam turbine* power (not shown) actually *increases* for sour-shift when compared to the baseline in Part 1: about 9-10MW of extra power generated on average. This may be due to the fact that the CO-shift process makes use of a catalyst to convert extra water into hydrogen for burning. Since CO_2 is removed before it reaches the gas turbine, the loss of mass flow must be made up by pushing additional syngas through the gasifier (thus increasing the gasifier size so it can accept more feedstock.) This translates to extra *energy* to be given to the steam cycle, which, since the ST inlet temperature is constant, demands a higher steam mass flow to keep the same stack temperature. In addition, the sour CO-shift reaction itself requires no additional steam from the HRSG at all to go to completion: all necessary water is already present in the syngas. This water comes from the quench, the slurry water, and the already high moisture content of the coal used. However, this combination of circumstances makes it so that both the TIT constraint and the mass flow constraint on the GT cannot be met at the same time. As such, the TIT condition is held, while the total mass flow rate entering the turbine (air and fuel) is allowed to decrease, resulting in a higher TET.

However, the sour-shift system behaves slightly differently for the supercritical cycle than it does for the subcritical cycle. The most obvious change here is that the supercritical cycle *loses* steam power from this case. Where in the subcritical system, sour-shift *increases* the total steam power (not shown) by about 10MW compared to the baseline; in the supercritical case, it *decreases* the power by at least 3 MW. This change is most likely caused by the fact that the gasification system did not change with the steam cycle, so the quality of water given to the steam cycle remains the same, while the grade of steam taken from the HP stream is much higher for the supercritical cycle. This means that the additional water supplied here is not enough to make up for the direct loss of power from sacrificing such high grade steam, whereas, for the subcritical cycle, it was a much better trade. This is also why the efficiency doesn't increase as much between the subcritical and supercritical cycles for the sour-shift cases as it does for the baseline cases: only about 1.0-1.3 percentage points of improvement for sour-shift compared to the previous 1.6 percentage points without CCS.

Sweet-shift, however, appears to be the worst form of CCS in terms of efficiency. While the net power is still above that of post-combustion CCS, the net efficiency is at least one percentage point lower on average than the post-combustion cases. In fact, it consistently decreases the total power output by about 11MW in all instances compared to the baseline. This is due to the fact that, unlike sour-shift, sweet shift requires additional steam input from the steam cycle directly, resulting in a reduction of steam turbine output of approximately 11MW. Since it occurs after every other process in the gas cleanup system, the amount of water needed is largely independent of BMR. But, in the long run, this is still enough to cut the efficiency by about 8 percentage points: even more than in post-combustion CCS.

However, the supercritical cycle benefits sweet-shift CCS more than any other case set: about 1.8-1.9 percentage points of improvement from the subcritical cycle, making the efficiency higher than that of post-combustion CCS. Similar to sour-shift, the power output of the sweet shift case is also lost when compared with the baseline: although, still a great deal more than the loss from the sour-shift cases (20MW vs. 40MW). Interestingly enough, the auxiliary losses are not that much higher than those of the sour-shift cases, so the total net power is only about 10MW lower on average than that of sour-shift CCS.

Tables 3 and 4 show the emissions data for sour- and sweet-shift CCS, respectively. Notice in Table 3 that, like in post-combustion CCS, just 10% biomass by weight is enough to make the plant carbon-negative. On a per MW-year basis, only the pure coal cases have marginally lower effective CO_2 emissions than the pure-coal post-combustion cases; all biomass cases have lower negative CO_2 emissions than the post-combustion cases or, in other words, the post-combustion biomass cases produce about 10% less CO_2 emissions. This is due to the increased gasifier size, as, again, the mass flow to the GT cannot be maintained at the same power output without adding additional syngas mass flow. This can only be accomplished by a larger gasifier. Therefore, more CO_2 is being added due to simply having more carbon available from the beginning.

Table 3 Emissions (Tons/MW-year) – Sour-shift CCS

Biomass/ Coal Ratio	0%	10%	30%	50%
Subcritical Plants				
NO_x	0.842	0.831	0.843	0.855
SO_x	10.58	9.06	7.17	5.28
Gross CO_2	1,070.8	1,040.0	1,072.5	1,105.1
Eff. CO_2	**1,070.8**	**-32.0**	**-2,143.4**	**-4,256.3**
Supercritical Plants				
NO_x	0.763	0.752	0.759	0.766
SO_x	9.59	8.20	6.46	4.73
Gross CO_2	1,312.4	1,280.2	1,311.3	1,342.3
Eff. CO_2	**1,312.4**	**311.1**	**-1,584.3**	**-3,464.3**

Table 4 Emissions (Tons/MW-year) – Sweet-shift CCS

Biomass/ Coal Ratio	0%	10%	30%	50%
Subcritical Plants				
NO_x	0.937	0.925	0.940	0.955
SO_x	11.73	10.05	7.97	5.88
Gross CO_2	1,284.2	1,236.5	1,258.9	1,281.7
Eff. CO_2	**1,284.2**	**48.6**	**-2,316.7**	**-4,698.3**
Supercritical Plants				
NO_x	0.823	0.814	0.825	0.837
SO_x	10.31	8.84	7.00	5.16
Gross CO_2	1,508.3	1,463.7	1,486.8	1,510.2
Eff. CO_2	**1,508.3**	**418.0**	**-1,654.4**	**-3,732.6**

Even with all this taken into account, post-combustion CCS retains one advantage over sour-shift pre-combustion CCS: handling SO_x and NO_x. Only post-combustion's chemical absorption can process SO_x and NO_x, and only because post-combustion CCS occurs after those compounds are able to form. The gross CO_2 emissions of the sour-shift CCS cases are about 15% less than the sweet-shift cases in both subcritical and supercritical plants. When effective CO_2 emissions are examined, for the supercritical plant, the 10% BMR case isn't carbon-negative for either form of pre-combustion CCS. This is directly caused by the presence of the duct-burner, which adds 80,000-100,000 tons of CO_2 per year (or 336-420 tons/MW-year) to the emissions. This is also why the effective CO_2 emissions of other biomass cases in the supercritical cases are less negative than the corresponding subcritical cases. Only post-combustion CCS is capable of cleaning up the CO_2 emissions from the duct burner, which is why post-combustion's 10% biomass case is the only case for the supercritical cycle that is carbon-negative.

Table 5 Economics – Sour-shift CCS

Biomass/Coal Ratio	0%	10%	30%	50%
Subcritical Plants				
Capital cost (million $)	1,164.3	1,043.1	1,027.4	1,011.5
Capital Cost ($/kW)	**5,334**	**4,726**	**4,721**	**4,712**
CoE ($/kW-hr)	**0.1192**	**0.1146**	**0.1269**	**0.1392**
CCS cost ($/kW-hr)	0.0184	0.0167	0.0185	0.0202
CO_2 Removal Cost ($/ton)	**18.70**	**17.31**	**18.34**	**19.17**
Supercritical Plants				
Capital cost (million $)	1,206.2	1,086.8	1,072.3	1,060.0
Capital Cost ($/kW)	**5,066**	**4,499**	**4,479**	**4,467**
CoE ($/kW-hr)	**0.1159**	**0.1114**	**0.1222**	**0.1331**
CCS cost ($/kW-hr)	0.0187	0.0167	0.0181	0.0198
CO_2 Removal Cost ($/ton)	**21.37**	**19.51**	**20.16**	**21.09**

Table 6 Economics – Sweet-shift CCS

Biomass/Coal Ratio	0%	10%	30%	50%
Subcritical Plants				
Capital cost (million $)	1,181.7	1,059.9	1,044.2	1,028.2
Capital Cost ($/kW)	**5,964**	**5,292**	**5,303**	**5,311**
CoE ($/kW-hr)	**0.1316**	**0.1264**	**0.1405**	**0.1547**
CCS cost ($/kW-hr)	0.0308	0.0285	0.0321	0.0357
CO_2 Removal Cost ($/ton)	**32.18**	**29.85**	**31.15**	**32.19**
Supercritical Plants				
Capital cost (million $)	1,241.0	1,119.2	1,103.4	1,087.5
Capital Cost ($/kW)	**5,506**	**4,919**	**4,923**	**4,925**
CoE ($/kW-hr)	**0.1248**	**0.1203**	**0.1326**	**0.1449**
CCS cost ($/kW-hr)	0.0276	0.0256	0.0285	0.0316
CO_2 Removal Cost ($/ton)	**32.45**	**30.38**	**31.44**	**32.50**

Lastly, for the economic data, it can be seen from Tables 5 and 6 that the supercritical cycle has the same economic effect for pre-combustion CCS as it does for the baseline and post-combustion CCS: the supercritical cycle universally decreases the capital cost per unit energy ($400/kW) and CoE (0.6-1.0 cents/kW-hr) despite the increase in total cost. In addition, the same biomass trend can be observed here as well: 10% BMR is the most optimal ratio, as it boasts the lowest overall CoE (0.4-0.6 cents lower than pure coal and 1.2-1.5 cents lower than 30% BMR). While the sour-shift cases' capital cost/kW continues to decrease beyond 10% BMR, as opposed to those of the other forms of CCS, which increase again for 30% and 50% BMR, the actual difference between the amounts is small ($5-$20/kW) and not worth calling any special attention to. Overall, the CCS cost for pre-combustion CCS is about 11-12 cents/kW-hr, which is about the same as most of the cases (S1B, L1B, S3B, L3B, and S4B) in the U.S. Dept. of Energy's Vol. 3 report on fossil fuel plants [13], although the capital cost (called "total overnight cost" in the DoE report) is about $1,000/kW more expensive by comparison. This may be due to differing plant sizes (The DoE's plants are more than twice the size of these plants) and year-to-year inflation (This study is based on 2011 USD, while the DoE's is based on 2007 USD.)

Sour-shift CCS is the most economical form of CCS in this study, as it has universally lower capital and electrical costs than either other form of CCS: $500/kW cheaper in capital cost than sweet-shift, and $2500/kW cheaper than post-combustion, and ~1cent/kW-hr cheaper in CoE than sweet-shift and 5-6 cents/kW-hr cheaper than post-combustion. In other words, post-combustion CCS is the most expensive practice: about 3.7 times more costly than sour-shift CCS. This, again, mainly comes from the easy integration that sour-shift has with existing equipment: sour-shift can easily be retro-fitted onto existing devices, unlike sweet-shift, which requires that two entirely new sections be added to the cleanup system. Also, unlike post-combustion CCS, both forms of pre-combustion CCS use much smaller, less-expensive equipment. Both of these facts are major contributors to sour-shift's much cheaper price tag.

Finally, using the same formula as Part 1 shows that the CO_2 removal cost for sweet-shift is only around $30/ton, while sour-shift's is even lower, around $20/ton: once again demonstrating the economic superiority of pre-combustion CCS (recall that the CO_2 removal cost for post-combustion CCS was around $60-70/ton.) For a more global comparative view for all cases, see Figs. 7 through 10 and the appendix.

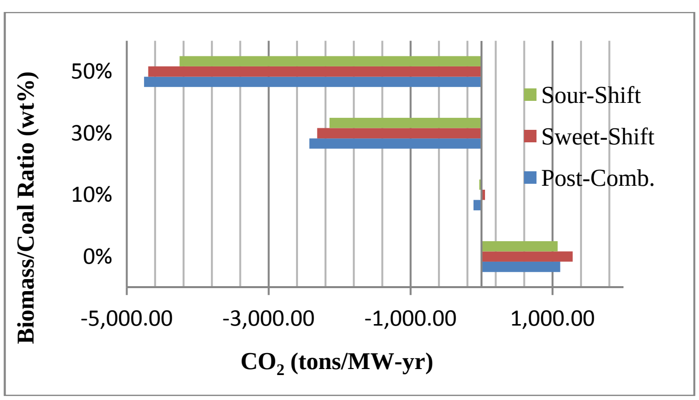

Figure 7 Effective CO_2 for subcritical plants with CCS

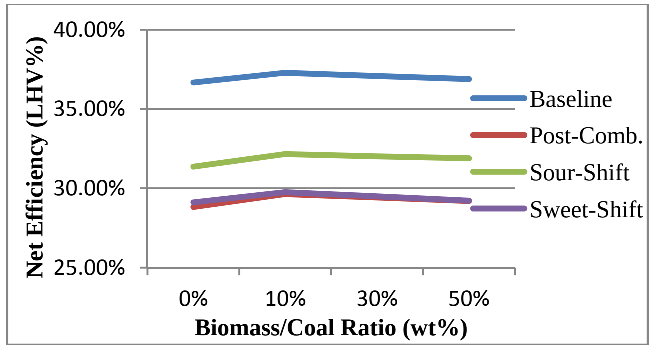

Figure 8 Efficiency for supercritical plants

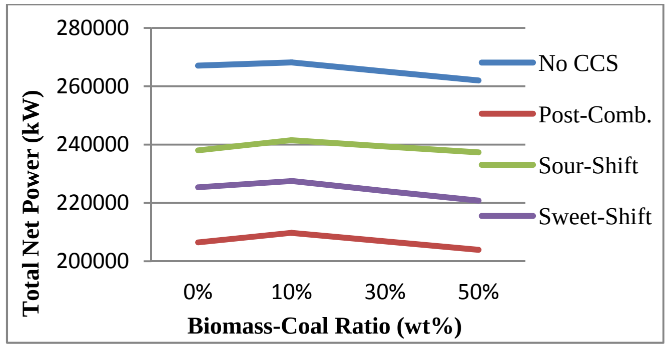

Figure 9 Total Net Power for supercritical plants

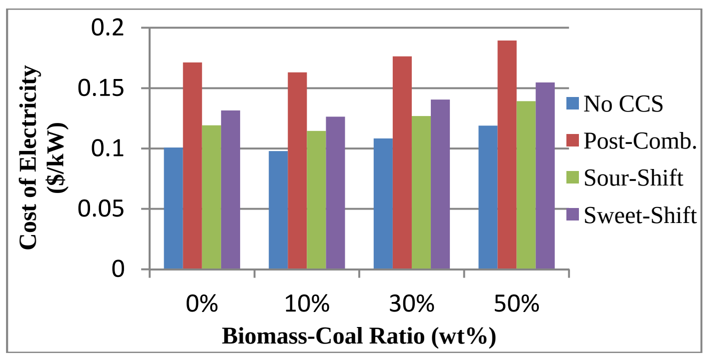

Figure 10 Cost of Electricity for subcritical plants

CONCLUSIONS

In summary, this study was performed using the same basic plant layout as Part 1: with a GE gasifier and Siemens SGT6-4000F gas turbine, and the plant was assumed to be constructed in southern Louisiana using Texas Lignite blended with sugarcane bagasse as the fuel. The Thermoflow® software suite's GTPro® program was used in this study. The results for Part 2 yield the following conclusions:

a. The optimal operating point in terms of fuel composition for all plants appears to be 10% BMR (about 0.7-0.8 percentage points better than pure coal), as the biomass chosen, after pretreatment, yields higher energy density syngas than the lignite. However, beyond 10% BMR, the energy needed for pretreatment becomes too high to counterbalance with the extra exergy from the new syngas blends. The 10% BMR cases are also cheaper (~$500/kW lower capital) and have about 0.5 cents/kW-hr lower CoE than the pure coal case.

b. As in Part 1, the supercritical cycle is universally superior operations-wise to the subcritical cycle. In other words, the previous trend mentioned in Part1 for post-combustion CCS is preserved in all the cases from Part 2 with pre-combustion CCS. The only difference is in the magnitude of the improvement: sweet-shift CCS benefits the most (1.8-1.9 percentage points of efficiency), due to the greater need for high-quality steam in more places Sour-shift, on the other hand, benefits the least (1.1-1.3 percentage points), due to the lower relative quality of the steam generated from the cleanup system. The economic costs, though, have generally the same differences as before (~$300-$400/kW lower capital and 0.3-0.4 cents/kW-hr lower CoE than the equivalent subcritical cycle.)

c. Sour-shift, pre-combustion CCS is the most optimal form of CCS in the current study: It has the highest efficiency (31-33%, compared to 27-29% sweet or 27.5-28% post-combustion with the lowest CCS impact on efficiency compared to the baseline,) and has the smallest increase in CCS cost compared to the baseline: $1000/kW ($0.018/kW-hr), compared to $1600/kW($0.029/kW-hr) sweet or $3700/kW ($0.068/kW-hr) post-combustion.

d. Sweet-shift, pre-combustion CCS is likewise cheaper to implement than post-combustion CCS (with $2100/kW difference in capital cost), and has lower CoE than post-combustion (by 4-5cents/kW), but, for the subcritical cycle cases, is thermally inferior (0.3-0.5 percentage points in efficiency). However, the supercritical cycle makes sweet-shift the better option due to the improvements from having access to higher quality steam, something that post-combustion CCS does not benefit from as greatly. The net efficiencies are comparable.

e. The CO_2 removal (or avoided) cost for <u>subcritical, sour-shift</u> is about $18/ton, <u>supercritical, sour-shift</u> is $21/ton, <u>sweet-shift</u> about $32/ton, and <u>post-combustion</u> is $70/ton. Blending 10% biomass can reduce the CO_2 cost by about $4-5/ton for post-combustion, and about $2/ton for pre-combustion CCS.

Finally, it needs to be noted that the better performance achieved by sour-shift over sweet-shift CCS in this study is subject to the specific design conditions imposed in this study. However, if a lower pressure, lower availability steam source is made available, it may be possible for sweet shift to have some advantage over sour-shift. (For example, using a third, lower pressure stream in the HRSG.)

ACKNOWLEDGEMENTS

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REFERENCES

[1] Long, H.A., 2011, "Analysis of Biomass/Coal Co-Gasification for Integrated Gasification Combined Cycle (IGCC) Systems with Carbon Capture." *University of New Orleans Theses and Dissertations.* Paper 1371.

[2] Long, H. A. and Wang, T., 2011. "A System Performance and Economic Analysis of IGCC with Supercritical Steam Bottom Cycle Supplied with Varying Blends of Coal and Biomass Feedstock," IMECE2011-65749, Proceedings of ASME 2011 International Mechanical Engineering Congress and Exposition, Denver, Colorado, U.S.A.

[3] Retzlaff, K. M., 1996. "Steam Turbines for Ultrasupercritical Power Plants." General Electric Company. Power Gen – Europe: Budapest, Hungary. GER-3945A.

[4] Voss, S.; Gould, G., 2001, "The Rankine Cycle: Workhorse of the Coal-fired Industry," Bruns and McDonnell: Public Company Papers.

[5] Hough, S., 2009, "Supercritical Rankine Cycle." University of Idaho. Technical Paper. ME-517 http://www.if.uidaho.edu/~gunner/ME443-543/HW/rankine.pdf

[6] Bergman, P. C. A.; Kiel, J. H. A., 2005. "Torrefaction for Biomass Upgrading." 14th European Biomass Conference and Exhibition. Paris, France. ECN-RX-05-180.

[7] Bergman, P.C.A., 2005. "Combined Torrefaction and Pelletization: the TOP Process." Renewable Energy in the Netherlands. ECN Biomass: ZG Petten, the Netherlands.

[8] Bridgeman, T.G., Jones, J., Shield, I., and Williams, P.T., 2007, "Torrefaction of Reed Canary Grass, Wheat Straw, and Willow to Enhance Solid Fuel Qualities and Combustion Properties." Energy and Resources Research Institute, University of Leeds: Leeds, UK.

[9] Energy Information Administration (EIA), 2009, "Average Sales Price of Coal by State and Coal Rank." U.S. Dept. of Energy. http://www.eia.gov/cneaf/coal/page/acr/table31.html

[10] Day, D., 2011, *Louisiana Sugar Cane Institute.* Louisiana State University – Agricultural Center. Private Correspondence.

[11] U.S. Department of Energy, 2011. "Natural Gas Weekly Update." <u>Independent Statistics and Analysis.</u> U.S. Energy Information Administration. http://www.eia.doe.gov/oog/info/ngw/ngupdate.asp

[12] Energy Information Administration (EIA), 2010, "Updated Capital Cost Estimates for Electricity Generation Plants." Office of Energy Analysis. U.S. Department of Energy: Washington, DC.

[13] U.S. Department of Energy, 2011, "Cost and Performance Baseline for Fossil Energy Plants." Vol. 3a. *Low Rank Coal to Electricity: IGCC Cases.* National Renewable Energy Laboratory, Pittsburgh, PA. NETL-2010/1399.

Appendix: Plant Summary for subcritical, sour-shift CCS plant with 10% biomass

GT PRO 21.0 parallel
1263 08-15-2011 15:40:52 file=C:\Documents and Settings\Hank\Desktop\Hank's Papers and Data\Grad Wo
rk\Thermoflow\Better Plant Design\SUBCRITICAL QUENCH - 10% BIOMASS - PRE-CCS (SOUR).GTP
Plant Configuration: GT, HRSG, and condensing reheat ST
One SGT6-4000F Engine (Physical Model #141), One Steam Turbine, GT PRO Type 10, Subtype 2
Steam Property Formulation: Thermoflow - STQUIK

SYSTEM SUMMARY

	Power Output kW		LHV Heat Rate BTU/kWh		Elect. Eff. LHV%	
	@ gen. term.	net	@ gen. term.	net	@ gen. term.	net
Gas Turbine(s)	200015		8205		41.59	
Steam Turbine(s)	99141					
Plant Total	299156	220712	8114	10998	42.05	31.03

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
25.88	20.74	27.94	8566

GT fuel HHV/LHV ratio =	1.179			
DB fuel HHV/LHV ratio =	1.179			
Total plant fuel HHV heat input / LHV heat input =	1.103			
Fuel HHV chemical energy input (77F/25C) =	2676998	kBTU/hr	743611	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2427428	kBTU/hr	674285	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2677209	kBTU/hr	743669	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2696010	kBTU/hr	748892	BTU/s

GAS TURBINE PERFORMANCE - SGT6-4000F (Physical Model #141)

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200015	41.59	8205	945	1122
Total	200015			945	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1935634	kBTU/hr	537676	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1641203	kBTU/hr	455890	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
77.56	99141	42.85	33.24	-249782

Number of steam turbine unit(s) =	1			
Fuel chemical HHV (77F/25C) to duct burners =	0	kBTU/hr	0	BTU/s
Fuel chemical LHV (77F/25C) to duct burners =	0	kBTU/hr	0	BTU/s
DB fuel chemical LHV + HRSG inlet sens. heat =	1017834	kBTU/hr	282732	BTU/s
Water/steam to gasification plant =	48597	kBTU/hr	13499	BTU/s
Water/steam from gasification plant =	194274	kBTU/hr	53965	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-49.62	%		

ESTIMATED PLANT AUXILIARIES (kW)

GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	831.3	kW
Condensate pump(s)*	199.5	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	3.234	kW
Cooling water pump(s)	1115	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	45	kW
Lights	75	kW
Aux. from PEACE running motor/load list	781.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	211.2	kW
Miscellaneous plant auxiliaries	299.2	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	36900	kW
Gasification plant, fuel preparation	10362	kW
Gasification plant, CO2 capture and AGR*	23186	kW
Gasification plant, other/misc	2577.6	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	76948	kW
Actual (user input) overall plant auxiliaries	76948	kW
Transformer losses	1495.8	kW
Total auxiliaries & transformer losses	**78444**	**kW**

* Heat balance related auxiliaries

IGCC PLANT HEAT BALANCE

Total Energy In	890505	BTU/s		
Power Block Energy In:				
Ambient air sensible	11290	BTU/s		
Ambient air latent	21492	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69384	BTU/s		
Process return & makeup	0	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	746833	BTU/s		
Gasifier slurry water	888.6	BTU/s		
Quench water	30491	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1376	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	9123	BTU/s		
Total Energy Out	890764	BTU/s		
Power Block Energy Out:				
Net power output	209204	BTU/s		
Stack gas sensible	75567	BTU/s		
Stack gas latent	159032	BTU/s		
GT cycle losses	5183	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	170136	BTU/s		
Process	0	BTU/s		
Steam cycle losses	4807	BTU/s		
Non-heat balance auxiliaries	15006	BTU/s		
Transformer losses	1417.8	BTU/s		
Gasifier Energy Out:				
Heat losses	67.43	BTU/s		
Slag	14813	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	4850	BTU/s		
CO2 removal	0	BTU/s		
Water condensed from syngas	14946	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
CO2 capture & AGR Qrej	29142	BTU/s		
CO2 capture & AGR heat loss	7160	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	143594	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	8345	BTU/s		
Heat rejection from compressor inter/after cooling	32978	BTU/s		
Compressors mechnical & electrical losses	1748.8	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-258.9	BTU/s	-0.0291	%

Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)