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Performance and Economic Analysis of Biomass/Coal Co-Gasification IGCC Systems with Supercritical Steam Bottom Cycle, Part 1 – Post-combustion Carbon Capture

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ABSTRACT

In recent years, Integrated Gasification Combined Cycle (IGCC) technology has been gaining steady popularity for use in clean coal power operations with Carbon Capture and Sequestration (CCS). This study focuses on investigating two approaches to improve efficiency and further reduce the greenhouse gas (GHG) emissions. First, replace the traditional subcritical Rankine steam cycle portion of the overall plant with a supercritical steam cycle. Second, add different amounts of biomass as feedstock to reduce emissions. Part 1 focuses on investigating post-combustion CCS and Part 2 on analyzing pre-combustion CCS, using both sour-shift and sweet-shift processes. Employing biomass as a feedstock has the advantage of being carbon neutral or even carbon negative if CCS is implemented. However, due to limited feedstock supply, such plants are usually small (2-50MW), which results in lower efficiency and higher capital and production costs. Considering these challenges, it is more economically attractive and less technically challenging to co-combust or co-gasify biomass wastes with coal.

Using the commercial software, Thermoflow®, this study analyzes the baseline plants around 235MW and 267 MW for the subcritical and supercritical designs, respectively. The results clearly show that utilizing a certain type of biomass with low-grade coals up to 50% (wt.) can, in most cases, not only improve the efficiency and reduce overall emissions, but may be economically advantageous, as well. The CO₂ emissions decrease by about 7,000 tons/MW-year, making both plants carbon-negative with only 10% biomass in the feedstock. In addition, implementing a *supercritical steam cycle* raises the efficiency (1.6 percentage points) and lowers the capital costs (\$300/kW), regardless of plant layout. Implementing post-combustion CCS consistently causes a drop in efficiency (at least 7-8 points) from the baseline, and increases the costs by \$3,000-\$4,000/kW and \$0.06-\$0.07/kW-hr. The SO_x emissions also decrease by about 190 tons/year (7.6 x 10⁻⁶ tons/MW-year), while the NO_x emissions are all but eliminated through post-combustion CCS. Finally, the CCS cost is around \$65-\$72 per ton of CO₂.

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%)
MEA	Monoethanolamine
M.W.	Molecular Weight (lbs/lb-mol)
LHV	Lower Heating Value (MJ/kg)
HHV	Higher Heating Value (MJ/kg)
CoE	Cost of Electricity (\$/kW-hr)
O&M	Overhead and Maintenance (\$)

OBJECTIVES

The primary objective of this study is to improve upon existing IGCC systems by (1) reducing the GHG emissions of such plants, (2) reducing their capital and electricity costs, and (3) increase the efficiency, if possible. A previous study by Long and Wang [1] was performed with these goals in mind using a baseline IGCC plant and changing the standard Rankine steam cycle to a supercritical steam cycle, where they noticed an improvement in efficiency of more than 1.5 percentage points. In addition, up to 10% biomass could be added for another 0.7 efficiency percentage point increment. This extra efficiency reduced the emissions of the plant by over 600 tons/MW-yr, and the addition of the supercritical cycle also reduced capital costs by about \$300/kW. However, in order to achieve the first goal of reducing emissions, the primary, most effective method remains to be utilizing carbon capture technology. The baseline plant has been adapted to make use of a post-combustion CCS system for Part One of this paper, while two different pre-combustion systems will be studied in Part Two. As in the previous study, biomass up to

50% (wt.) will be added to the baseline coal feedstock in order to further reduce the emissions, with the hope of reaching carbon-negative emissions eventually.

To achieve the second and third goals of reducing capital costs and raising the efficiency, a supercritical Rankine cycle is implemented to replace the traditional subcritical Rankine cycle as in the baseline. Although supercritical steam systems are more expensive than subcritical ones, the idea is that the extra provided power and efficiency will make up for the increased total cost, reducing the overall capital cost per unit power output.

MOTIVATIONS

Raising the inlet temperature and pressure of the steam turbine in a traditional Rankine cycle is the most direct way to increase the operating efficiency of said cycle. As early as the 1950's, scientists and engineers have been highly focused on this area of potential steam cycle improvement [2]. It was during this period where the maximum inlet pressure and temperature were raised from 2400PSI/1000°F to near 4500PSI/1150°F [2]. This was the onset of the first supercritical steam generation plant. The term "supercritical" comes from the idea that the steam running through the boiler or HRSG is *above* the "critical point" at the top of the vapor dome on a standard temperature-entropy diagram at around 3200PSI [3]. For reference, the typical efficiency of a standard subcritical Rankine (steam) cycle is around 30-38%, while a supercritical cycle under the same environmental conditions can achieve an efficiency of 42-45% [4]. So far, all of the research and industrial efforts going into supercritical cycle design are meant for standard pulverized coal (PC) plants. To the authors' knowledge, there is currently no literature available documenting a supercritical steam bottom system being used in any real-world IGCC system.

The use of biomass in IGCC is not a new idea. The first pure biomass IGCC plant was constructed in Värnamo, Sweden 1993. As a demonstration plant, it provided roughly 6 MW of net electricity to the grid by using a fuel equivalent energy input of approximately 18 MW [5]. Several other biomass plants in the range of 40-100MW have been constructed, such as the Hawaiian biomass gasification experimental plant developed by Siemens-Westinghouse [6] and the McNeil Station in Burlington, Vermont [7]. In addition, other, more traditional plants have been modified for use with biomass and gasification processes, such as the Chowchilla I in California and the Lahti Co-firing Project in Finland, which both used syngas derived from biomass to run a Rankine cycle [7]. All of these plants, however, have either failed or been removed from the commercial power sector due to not being economically competitive.

Due to the controversies surrounding issues of cultivating energy crops, in this study, only biomass wastes are considered. For convenience, the word "wastes" is dropped from the rest of the paper. The first and greatest challenge with utilizing biomass is associated with its availability, sustainability, and quantity. The supply of most biomass is seasonal and is limited by quantity. In addition, biomass cannot be economically transported over long distances due to its low mass density. A solution to some of these problems is by *co-*

feeding biomass alongside coal in a larger plant. This allows biomass to be used whenever it is available and on the same economy of scale that coal is. Doing this also reduces fossil fuel consumption, which is a benefit both for the environment and for energy providers, since most of biomass wastes are either free or bear very low costs. Next, since biomass is cleaner than coal is, co-feeding results in lower emissions than a pure coal plant, and is able to provide much more power than a pure biomass plant. Furthermore, because there is coal mixed in with the biomass, corrosion is less of an issue than it is with plants that use purely biomass.

However, there are still operational problems that biomass can cause to co-fed systems. For one, biomass has very low energy density. Coupled with its low mass density, this means that the required volumetric flow rates for providing the required energy to run the plant are higher than those of coal. [8] Limited biomass supplies and transport issues inhibit profitable operation of larger pure biomass plants, meaning that effectively utilizing pure biomass in any plant bigger than about 50-80MW is uneconomical at best [8]. Secondly, most types of biomass are very fibrous and tough, and tend to get stuck in various types of feeding machinery. Thirdly, biomass tends to contain many corrosive compounds that can damage other internal parts [9]. Lastly, biomass has an expiration date: it cannot be stored for any extended length of time due to its tendency to rot and decompose, being rendered almost useless as a fuel in the process.

To overcome this new set of challenges of biomass feeding and long-term storage, one available solution is employing *pretreatment*. Various chemical, thermal, and biological processes are available to transform raw biomass into a form that makes it more suitable for power generation. The type of pretreatment taken into consideration for this study is *torrefaction*. Torrefaction is a thermal process, wherein raw biomass is heated to about 200-300°C and essentially "cooked," removing a large portion of the moisture content, and altering the chemical structure of the biomass in such a way that it loses its tough, fibrous consistency, and "torrefied biomass," a reddish-brown, brittle, solid substance that has calorific properties that greatly approach those of low- to mid-grade coals [9]. During torrefaction, the biomass loses roughly 30% of its mass as torrefaction gases, and roughly 10% of its internal energy with them [10]. A simple algebraic calculation shows that this would result in about a 28% increase in the calorific value per unit mass for the feedstock [11].

In addition, torrefied biomass has a higher mass density than untreated biomass, is less corrosive, has higher grindability, and is much easier to store and transport [9-11]. Despite these benefits, using torrefaction at all requires that a separate, torrefaction plant be constructed on-site, which is a significant investment for most plants, especially the smaller ones. In fact, in one 1999 study done on a failed test plant by Siemens-Westinghouse in Maui, Hawaii, the researchers speculated that, while torrefaction itself is very effective at solving virtually all the feeding problems they had been having, investing in one might not be economically viable [6]. However, a 2005 study by P.C.A. Bergman of the Netherlands showed that torrefaction when combined with Pelletization

(another process that increases the mass density of the biomass) was not only viable in Europe, but perhaps *profitable* as well, albeit with a high dependency upon the price of the biomass feedstock and other factors [10].

While some biomass-coal co-feeding studies have been done in the realm of *co-combustion* of biomass with coal [12,13] they were mainly based on subcritical PC plant designs. In IGCC plants, the biomass and coal are *co-gasified* instead of co-combusted. For instance, the Polk IGCC plant performed several experiments in which a wood-based eucalyptus biomass feedstock was co-fed into an existing IGCC coal plant, in Tampa, Florida. The results showed that the existing Coal/Petcoke fed IGCC system was feasible to feed biomass, and the emissions of NO_x and SO_x were reduced about 10% [14]. The Buggenum IGCC plant in Netherlands also successfully co-gasified biomass (50% wt.) with coal using 3 major biomass sources: wood, sewer sludge, and manure, using about 300 tons of feedstock per year [15].

This study, like the previous study from Long and Wang (2011), focuses on investigating *co-gasification* of biomass and coal for application in IGCC systems with both subcritical and supercritical bottom Rankine cycle systems, but with the addition of carbon capture plants. Carbon capture is the next logical addition to such a plant given the current political situation: many countries around the world are or are considering implementing a “carbon tax” on industry, meaning there will be government-imposed fines for expelling too much carbon in the form of emissions into the atmosphere. A carbon capture system can drastically reduce such emissions, and potentially help many power companies to save money in the form of avoided tax penalties. In addition, the captured CO_2 may have other uses once captured, such as in advanced oil recovery [16]. When CCS is combined with biomass, it is possible, assuming biomass is carbon-neutral, for a plant to become *carbon-negative*. This means that, although emissions are produced, there is a net *decrease* in the amount of carbon put into the atmosphere, because plant-based biomass spends a great portion of its lifetime consuming CO_2 through photosynthesis.

In all, the focus of this study is to further investigate the effects of supercritical steam cycles and biomass on IGCC from the previous study, and how these affect systems with carbon capture: for a total of three main parameters.

PLANT DESIGN

The software used for this study was Thermoflow® program suite’s GTPro®. GTPro is a commercial software program that uses a top-down design approach for building gas turbine power plants and combined cycle plants. Figure 1 shows the general layout of the baseline case (No CCS), designed for around 240 MW of net power output. It consists of a single gasifier, based on the GE gasifier, which is slurry-fed and oxygen-blown with quench cooling. The gas cleanup system contains a section for particulate removal (a “scrubber”), a section for COS hydrolysis, a cooling segment, and Acid Gas Removal (AGR). The power block consists of a single GT, modeled after the Siemens SGT6-4000F turbine, with steam injection in the combustor to reduce NO_x

formation, and a single ST, with a fixed steam inlet temperature and pressure.

The plant is designed exclusively for power generation, so no chemicals or energy gases are exported anywhere in the middle of cleanup, and all waste products are assumed to be simply disposed of. The biomass is torrefied before entering the gasifier. Seen in Fig. 1, the fuel preparation block is where the fuel undergoes all related processes. Around 200kW-hr of energy per ton of biomass is assumed to be required to completely convert the biomass. All condensed water extracted from the raw syngas during cooling is used as makeup water for the steam system (not shown). The deaerator is assumed to be tray-type, and all process water is returned to it via a series of pipes. The deaerator also provides additional water to auxiliaries wherever more is needed and acts as the de-superheating source for all water streams that require cooler water/steam sources. Lastly, the ASU is assumed to be a cryogenic system with an operating pressure of 10 atm (147 PSI), and always delivers a stream of 95% pure oxygen at the required pressure to the gasifier.

Plant Location

The plant was assumed to be built around New Orleans, Louisiana, at an elevation of 10 feet above sea level. The climate condition was assumed to be an average of 85°F and 90% relative humidity in summer to provide a conservative plant output and thermal efficiency. ISO conditions (59°F and 60% R.H.) were not used as the baseline because those conditions are not common for Louisiana on the whole. It was deemed better to be more conservative with the model prediction by using conditions applicable to a Louisiana late summer/early fall. While both the temperature and humidity given above are highly unlikely to occur at the same time, they are meant to represent more of a weighted average: sometimes it will be 90+ degrees with 70% humidity, and at others, perhaps, 70-80 degrees with 80-95% humidity. As such, the conditions above were chosen to represent an “average” Louisiana summer day.

For coal, Louisiana is situated between two of the largest producers of lignite ore in the entire United States: Texas and Mississippi. In addition, Louisiana shares a very close relationship with both of these states in many areas from business to politics to tourism. For this reason, the coal chosen for the plant was South Hallsville Texas Lignite. In addition to being cheap, lignite is very easy to obtain and is abundant, especially in this region, and lignite from Texas is one of the best energy resources in the Southern United States.

As for biomass, Louisiana is one of the largest producers of sugarcane in the United States, and, out of those producers, Louisiana has the oldest and most historic part played in the sugar production industry [17,18]. About 16% of all sugar produced in the U.S. comes from Louisiana farms and factories [17], and around 16 million tons of raw sugarcane is harvested per year [18]. Since there is a lot of bagasse left as the waste product of the refinement process, sugarcane bagasse was selected as the feedstock. The fuel data, including ultimate analyses, for both lignite and bagasse can be seen in Table 1.

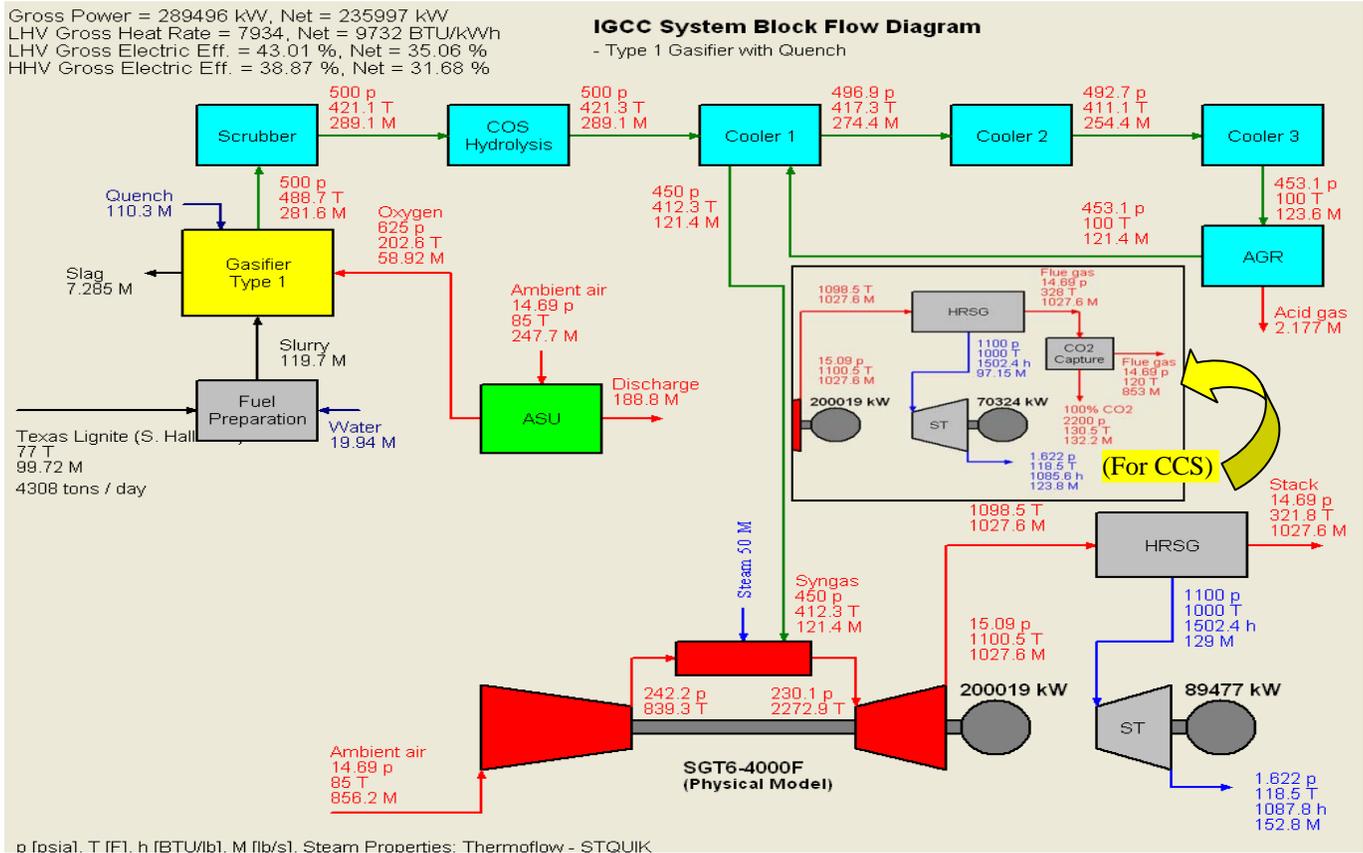


Figure 1 General plant layout without CCS (Subcritical cycle, 0% biomass) with insert showing post-combustion system attachment.

Table 1 Fuel Data

Component	S. Hallsville Texas Lignite (wt%)	Sugarcane Bagasse (wt%)
C	41.3	43.59
H ₂	3.053	5.26
N ₂	0.623	0.14
S	0.7476	0.04
O ₂	10.09	38.39
Cl ₂	0	0
H ₂ O	37.7	10.39
Ash	6.479	2.19
LHV (Btu/lb)	6398	6714
Price (\$/ton)	19.00	65.00

From: GTPro® internal fuel library, EIA[21], & D. Day[18]

Gasifier

The gasifier is modeled after the GE/Texaco gasifier. It is slurry-fed (35% water by weight) and oxygen-blown with an ASU pressure of 147PSI (10 bar), and has no coolers at all. The raw syngas is quenched with water at 300°F (149°C), up to a relative humidity of 50%. This selection of a quenched gasifier will allow for a more direct comparison when carbon capture is introduced to the design. Figure 5 shows the basic gasifier layout, with the numbers shown being representative

of the pure coal, subcritical case with no CCS. For both sets of cases (sub and supercritical), the GT used was a Siemens SGT6-4000F with steam injection, and the ST inlet temperature and pressure are fixed at 1100PSI/1000°F (76 bar/538°C) for the subcritical plant, and 2400PSI/1200°F (165.5 bar/650°C) for the supercritical plant. The supercritical plant’s turbine inlet pressure is lower than that of a typical PC plant (3500PSI) to avoid material failure and reduce maintenance for the first IGCC system having a supercritical bottom cycle. This is to serve as a stepping stone for moving to higher pressure and temperature supercritical steam conditions in the future. For all the cases, the turbine inlet temperature of GT is fixed and the total mass flow rate through the GT is also fixed.

Steam Turbine System

For the steam system itself, the HRSG connections and heat exchanger locations are consistent across all cases, both for post-combustion and the baseline, as are the ST layout, the condenser cooling system type, and the deaerator. This is all highlighted in Fig. 4. The only difference between the post-combustion CCS plant’s steam cycle and that of the baseline is the presence of a ~180lbs/sec mass source at the Intermediate Pressure (IP) boiler exit. Again, this is solely to enforce mass conservation within the HRSG, and to simulate an “external” steam source for the CCS steam, due to software limitations. The HRSG contains two pressure streams: HP and IP, which

both provide the steam necessary to provide power via the steam turbine. The HP stream is the main source of steam for the ST inlet, while the IP stream is used to drive plant auxiliaries and processes and also provides additional steam to the ST's reheat stages. All zones within the HRSG are fixed in all cases, with only temperatures and pressures varying from case to case. In addition, as stated previously, all HRSG connections are consistent for all cases (for instance, the main IP process stream at exchanger IPS1 always provides the water for Acid Gas Removal, the remaining IP stream always connects to the ST reheat section, etc.)

The steam turbine itself, as seen in Fig. 4, consists of 2 casings, for a total of 3 main stages: one large, high pressure stage and a second, lower pressure stage divided into two separate stages by an IP injection stream containing the leftover steam from the main IP process (acid gas removal for this study). It is a condensing turbine with a reheat section that links the high and intermediate pressure casings together. The temperatures and pressures vary according to which case is being studied. The second casing is split in two, due to the injection of supplementary steam from the IP process stream. The amount injected varies from case to case due to changes in steam demand on other components (CCS and AGR). The conditions were set so that the ST isentropic efficiency could be kept as high as possible in order to maintain the same TIT and TIP. Finally, the steam turbine condenser is connected to a natural draft cooling tower, which makes use of ambient air in the cooling process. All of the settings on the cooling tower are strictly enforced, and all parameters are fixed in each and every case.

The cleanup system consists of a particulate scrubber supplied with water at 215°F (102°C) for the procedure, a

section for Carbonyl-Sulfide (COS) Hydrolysis, a series of coolers and water “drains,” and Acid Gas Removal (AGR). The AGR unit is an amine-based system (single-stage), which is based on the Selexol® process and operates at 90% removal efficiency.

The NO_x production was based on the emissions specifications from the gas turbine manufacturer. As with the previous study, the NO_x specification used was based upon the SGT6-5000F, since data for the 6000F model was not available at the time. Both the 5000F and 6000F have similar combustors and the same NO_x control unit installed, so this comparison is a valid approximation. As such, the NO_x data was specified to be 9ppm at a 15% O₂ reference content [19].

Carbon Capture System

The second plant with CCS included is shown in Fig. 1 as an insert/addition to the baseline plant. The system makes use of an amine-based solvent separate the CO₂ from the rest of the GT exhaust. Because of the highly acid nature of the GT exhaust gases due to the presence of SO_x and NO_x, only chemical absorption is applicable in this case. The solvent chosen was Monoethanolamine (MEA), whose price tag was determined to be \$1600/ton [20]. On the whole, adding post-combustion CCS seems to take a drastic toll on the steam cycle's water supply. Among many possible options to provide the steam needed for the amine-based CCS process, the following two options are evaluated:

Option 1 is to tap the steam from the existing HRSG. In this way, the amount of steam delivered to the steam turbine will be reduced and the steam turbine output will suffer. This is a viable option, but will be inconvenient for comparison between non-CCS and CCS-based on similar power output.

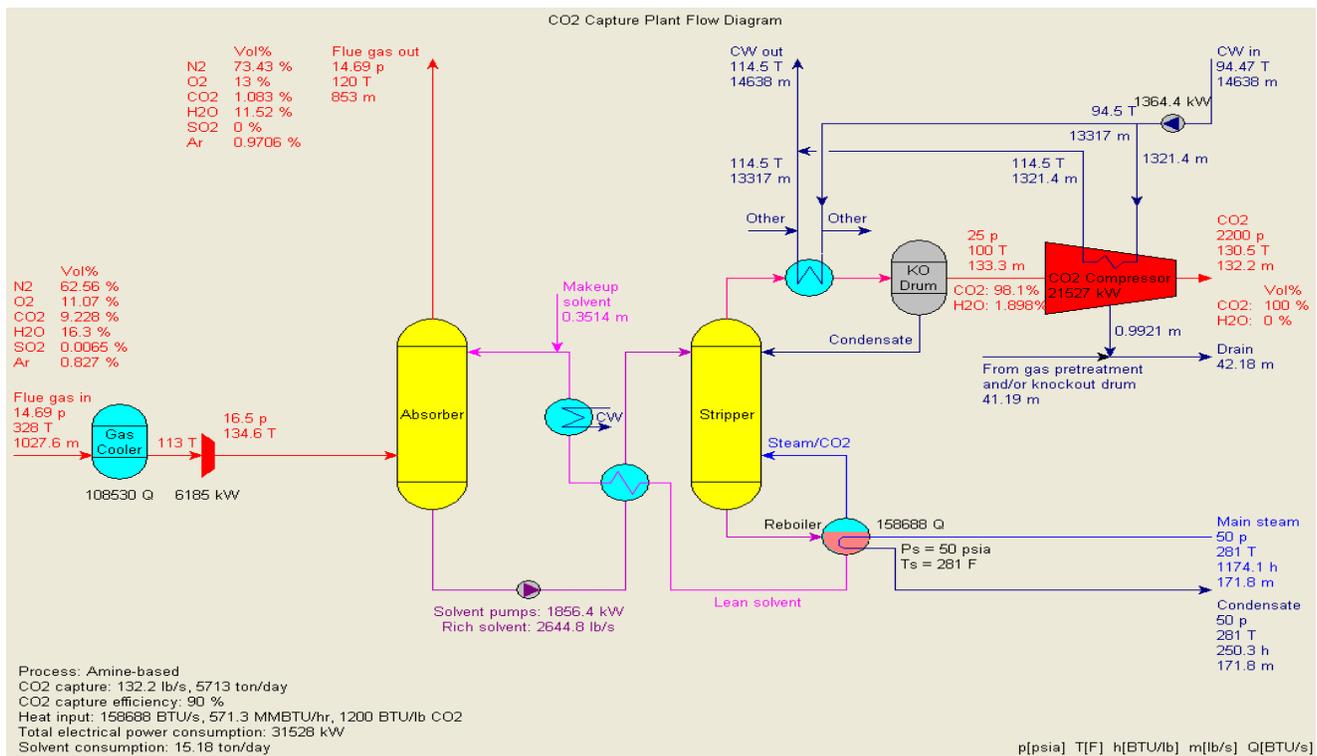


Figure 2 Post-combustion carbon capture system

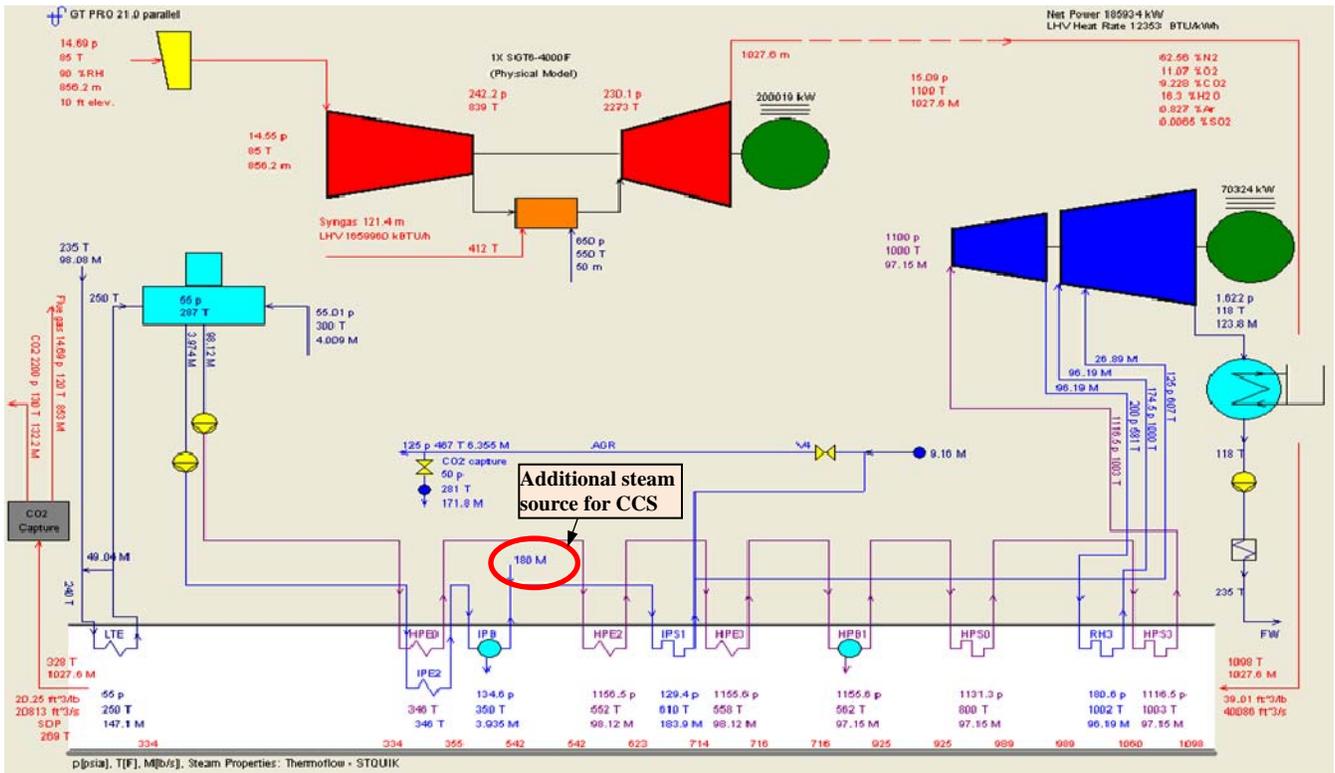


Figure 3 Post-combustion CCS steam cycle, with additional mass source imposed at IPB exit

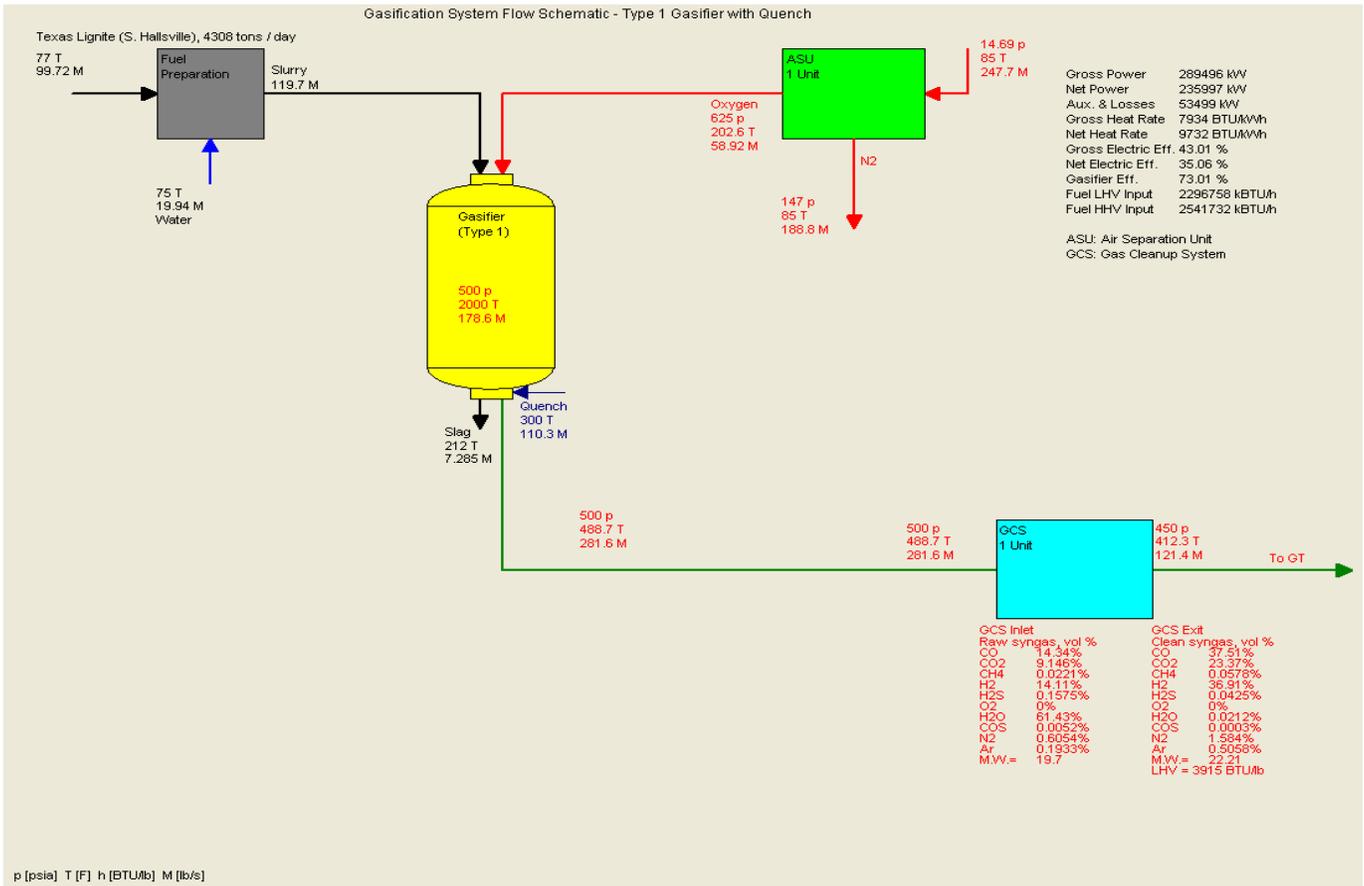


Figure 4 Gasifier design

Option 2 is to tap the steam from a bigger HRSG by increasing the total steam supply through bigger pumps. The steam that is used to drive the CCS process is thus taken from the IP stream at the same point as that of acid gas removal. This is the lowest pressure point in the system that this water can be taken without directly affecting the steam turbine.

The designed post-combustion system, seen in Fig. 3, has quite a demand for steam (171.8 lbs/sec). In an attempt to correct this problem, an additional mass source had to be created to force the steam turbine to operate more closely to that of the baseline case. This mass source was set in place to offset the cost of the CCS system and to make up for the head loss exhibited by the rest of the system, so the injection point in the second stage of the ST could remain close to the baseline case without reversing flow and becoming an extraction point. An enlarged rendition of the post-combustion plant's steam cycle can be seen in Fig. 4.

The carbon capture efficiency is targeted to be 90%. In addition, it should be noted that, due to the nature of chemical absorption, a small portion of the chemical solvent (< 0.01% wt) is lost during the capture process and must be replaced with fresh solvent. This adds up to a total of about 15 tons/day of solvent, or \$24,000 a day in maintenance. Finally, as this system also uses sequestration, a compressor is necessary to push the CO₂ into a proper storage unit. The compressor raises the pressure of the captured CO₂ to 2200 PSI, at a mass flow of about 130-140 lbs/s, with variation attributed to the different system designs and feedstock composition. In total, the CCS system uses around 32MW of electrical power, around 11% of the total expected plant capacity, to maintain operation.

The raw CO₂, however, is handled by simple conservation of mass and species within the program. However, when biomass is involved, the concept of carbon-neutrality must be observed. For this calculation, it was assumed that all biomass feedstock is completely carbon-neutral. This allows for the calculation of the so-called "effective" CO₂ output, which is obtained by taking the total CO₂ and subtracting the biomass's neutral CO₂ from it. The neutral CO₂ is determined by Equation 1:

$$\text{Neutral CO}_2(\text{ton/day}) = m_{\text{feed}} \left(\frac{\text{ton}}{\text{day}} \right) * \text{BMR} * \%C_{\text{bio}} * \frac{M.W._{\text{CO}_2}}{M.W._c} * 8000 \frac{\text{hrs}}{\text{year}} * \frac{1 \text{ day}}{24 \text{ hrs}} \quad (1)$$

This equation was derived under the assumptions that (1) all reactions involving carbon eventually result in CO₂, (2) start up and shutdown times are either neglected or assumed to be a part of the plant's recorded 8000 hour operating schedule (equivalent of 91.3% capacity factor), making no appreciable difference in the CO₂ emitted as compared to that of normal operating hours, and, finally, (3) the composition of biomass, particularly the carbon content, is constant and uniform, with no variation at any point in time.

ECONOMIC ANALYSIS

Beginning with fuel choice, lignite is cheap, and, according to the EIA's report [21], lignite from Texas costs approximately \$19.00/ton. While sugarcane itself is cheap at only \$30.00/ton [22], bagasse is only a portion of the sugarcane. All other by-products like cellulose and

hemicelluloses are left behind to be processed into sugar and/or ethanol. On average, about 200 lbs of dry bagasse will be produced from one ton of sugarcane. With this in mind, it becomes easy to make the mistake of assuming that bagasse will be cheaper than sugarcane on a per ton basis. The final price of the bagasse is around \$65 per ton of bagasse or \$13 per ton of the original weight of sugarcane [18].

The price of natural gas for the supercritical plant's duct burner was found to be \$4.10 per million Btu (\$/mmBtu) based on data from June, 2011, when this study was performed [23]. In addition, water consumed by the plant was assumed to be based on *utility*, and the price was set at \$2.00 per thousand gallons. Finally, overhead and maintenance (O&M) costs were taken from a report by the EIA, where they were determined to be \$60.00/kW (fixed) and \$0.006/kW-hr (variable) [24].

With the fuel prices known, the program used demands that the prices be input on a per unit energy basis, so the fuel costs of coal and biomass must be converted over, since all pricing is listed on a per unit weight basis, commercially. This simple conversion for coal is obtained through Equation 2:

$$\frac{\$}{\text{mmBtu}(\text{coal})} = \frac{\$}{\text{ton}(\text{coal})} * \frac{1 \text{ ton}}{2000 \text{ lbs}} * \frac{1}{\text{LHV}_{\text{coal}}} * \frac{10^6 \text{ Btu}}{1 \text{ mmBtu}} \quad (2)$$

which comes out to \$1.48/mmBtu for coal. Biomass, however, is received wet, so the heating value changes when it undergoes torrefaction and is dried. After the analysis is performed and some terms cancel out, the final conversion for biomass reduces to Equation 3:

$$\frac{\$}{\text{mmBtu}(\text{bio})} = \frac{\$}{\text{ton}(\text{wet})} * \frac{1 \text{ ton}}{2000 \text{ lbs}} * \frac{1 \text{ lb}(\text{wet})}{0.7 \text{ lbs}(\text{dry})} * \frac{1}{\text{LHV}_{\text{dry}}} * \frac{10^6 \text{ Btu}}{1 \text{ mmBtu}} \quad (3)$$

which comes to about \$6.92/mmBtu. For all blends of biomass and coal, the two prices obtained from equations 2 and 3 are linearly combined based on the biomass mass ratio (BMR) in the blended fuel and normalized by the blend's total LHV, which was calculated by the software. This final blend price is given by Equation 4:

$$\frac{\$}{\text{mmBtu}(\text{total})} = \frac{\text{BMR} * \frac{\$}{\text{mmBtu}(\text{bio})} * \text{LHV}_{\text{bio}} + (1 - \text{BMR}) * \frac{\$}{\text{mmBtu}(\text{coal})} * \text{LHV}_{\text{coal}}}{\text{LHV}_{\text{blend}}} \quad (4)$$

The cost of the blended biomass/coal feedstock is calculated under the assumption that it is linearly proportional to the biomass ratio (BMR). Finally, the energy consumption for coal grinding and drying was estimated to be 40kW-hrs/ton, while biomass, which must undergo torrefaction as well as grinding and drying, was estimated to be 200kW-hrs/ton. The average processing cost for each biomass case was also calculated to be linearly proportional to the BMR.

Finally, as mentioned previously, an overall plant life of 30 years is assumed, with a total operational capacity of 8,000 hours per year. In addition, 30% of the total initial investment is to be taken on *equity*, meaning that the plant owner must pay for these commodities out of his/her own pocket. Taxes on the plant were taken to be around 35%, with 10% flat-interest rates for all plant features. No inflation was considered for this study, so the analysis is based on 2011 USD. Lastly, the total package uses straight-line depreciation, but it was assumed that

only 75% of the total investment is available for depreciation for tax purposes.

RESULTS AND DISCUSSION

The results are analyzed based first on the effect of biomass within a single system, then upon the effect of the supercritical cycle, and, finally, with respect to the type of CCS used. Again, in Part one, only post-combustion CCS is considered. The results for the power output and plant efficiency can be seen in Tables 2 and 3.

Table 2 Power (kW) and Efficiency (LHV) – Baseline

Biomass/Coal Ratio	0%	10%	30%	50%
Subcritical Plants				
Aux. Losses (kW)	53,499	52,451	55,913	59,277
Net Power	235,997	237,356	234,296	231,291
Gross Efficiency	43.01	43.59	43.96	44.31
Net Efficiency	35.06	35.70	35.49	35.27
Supercritical Plants				
Aux. Losses (kW)	55,481	54,413	57,873	61,235
Net Power	267,111	268,207	265,090	262,043
Gross Efficiency	44.29	44.84	45.18	45.52
Efficiency	36.67	37.28	37.08	36.89

Table 3 Power (kW) and Efficiency (LHV) – Post-Combustion CCS

Biomass/Coal Ratio	0%	10%	30%	50%
Subcritical Plants				
Aux. Losses (kW)	84,409	82,668	86,098	89,434
Net Power	185,934	190,260	187,369	184,610
Gross Efficiency	40.16	41.06	41.42	41.79
Net Efficiency	27.62	28.62	28.38	28.15
Supercritical Plants				
Aux. Losses (kW)	86,731	84,935	88,368	91,702
Net Power	206,495	209,765	206,822	203,954
Gross Efficiency	40.93	41.65	41.99	42.33
Net Efficiency	28.82	29.64	29.42	29.20

As seen from Tables 2 and 3, the addition of biomass, even up to 50% by weight in the feedstock has a net positive effect on the efficiency. Beyond 10% BMR, however, the efficiency begins to decrease. This is due to the additional energy costs required to *process* the biomass, as it requires much more energy than raw coal to pre-treat. This added energy consumption is categorized as a part of the “auxiliary losses” in the tables, so the *gross* efficiency is unaffected by this, and, in fact, continues to increase since more gross power is generated. Despite the drop in efficiency, it still remains higher than the efficiency of the coal alone, due to the fact that this type of biomass has a higher heating value, as seen in Table 1.

The supercritical system has a clearly beneficial effect on the IGCC plant, with a consistent improvement in efficiency of over 1.6 percentage points when compared with Case A1. The efficiency for each amount of biomass also appears to follow the exact same trend as before, with an increase from 0-10%

BMR and decreases from 10-50%. The total net power also increases by about 25MW (9.8%) overall for each case. From this, it is clear that using a supercritical cycle provides an overall efficiency benefit for an IGCC plant as a whole.

The percentages of auxiliaries and various losses for the pure coal, supercritical baseline plant are shown in Fig. 5. It can be seen that the largest parasitic power is consumed by the ASU (68.21%) and the second largest one is from coal preparation (12.94%).

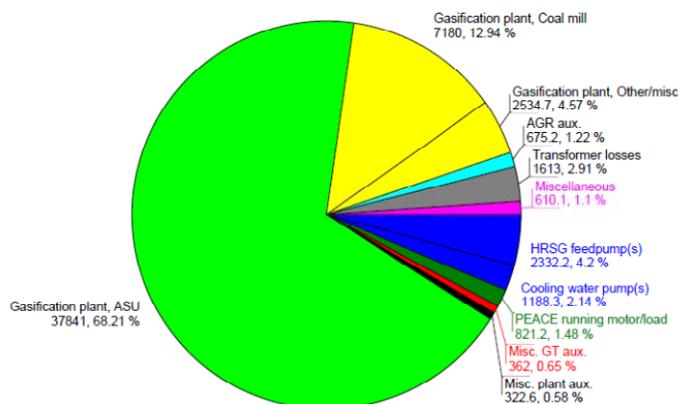


Figure 5 Auxiliaries & losses for supercritical pure-coal plant without CCS (See Appendix for details.)

Finally, Post-combustion CCS has a clearly negative impact on overall plant performance. This is not surprising, since CCS costs a significant amount of energy to perform, and offers nothing in the way of power or heat recovery to offset this, and post-combustion CCS seems to reduce the total net efficiency by nearly 8 percentage points in all subcritical cases. Even worse is the fact that the supercritical plant suffers *even greater losses in power* than the subcritical plant. While the subcritical plant lost about 17-19MW of steam power from the CCS plant, the supercritical plant loses over 30MW, nearly twice as much.

The main problem is that post-combustion CCS requires the use of chemical absorption, and the detriment to the total steam available to provide power is significantly increased, resulting in over 19MW of power directly lost due to reduced steam mass flow. In addition, the total auxiliary cost increases by about 60% due to the electrical energy consumed in order to maintain the pressure differences in the absorber and stripper columns and to compress the captured CO₂ at the end of the process. The GT power and other specifications not shown are unaffected by this, as all of the changes to the plant do not involve anything to do with the top cycle or the gasification block.

Tables 4 and 5 show the syngas compositions for both the subcritical and supercritical plants. The results correspond to the *raw* syngas leaving the gasifier, and, as such, relate to both the baseline case and all CCS cases (including the pre-combustion cases in Part 2.) This is mainly to demonstrate that the gasification block is virtually untouched by the effect of the supercritical cycle, aside from some scaling to maintain equilibrium. In addition, notice that the addition of further biomass continually increases the amount of CO, H₂, and CH₄, while simultaneously decreasing the amount of CO₂ and H₂O.

Table 4 Syngas Compositions (vol%) – Subcritical Plants

Biomass/Coal Ratio (wt%)	0%	10%	30%	50%
CO (vol%)	14.34	14.98	15.47	15.97
CO ₂	9.146	8.776	8.726	8.670
CH ₄	0.0221	0.0274	0.0299	0.0327
H ₂	14.11	14.76	14.91	15.06
H ₂ S	0.1575	0.1434	0.1142	0.0846
H ₂ O	61.43	60.56	60.03	59.51
COS	0.0052	0.0047	0.0038	0.0029
N ₂	0.6054	0.5726	0.5374	0.5016

Table 5 Syngas Compositions (vol%) – Supercritical plants

Biomass/Coal Ratio (wt%)	0%	10%	30%	50%
CO (vol%)	14.32	14.98	15.47	15.97
CO ₂	9.147	8.776	8.726	8.670
CH ₄	0.0221	0.0274	0.0299	0.0327
H ₂	14.11	14.76	14.91	15.06
H ₂ S	0.1575	0.1435	0.1142	0.0846
H ₂ O	61.42	60.55	60.03	59.51
COS	0.0052	0.0047	0.0038	0.0029
N ₂	0.6054	0.5727	0.5374	0.5016

Table 6 Emissions(Tons/MW-year) – Baseline without CCS

Biomass/Coal Ratio	0%	10%	30%	50%
Subcritical Plants				
NO _x	0.994	0.980	0.991	1.002
SO _x	9.14	6.84	6.22	4.57
Gross CO ₂	8,942	8,620	8,719	8,819
Eff. CO₂	8,942	7,688	5,928	4,173
Supercritical Plants				
NO _x	0.879	0.867	0.876	0.885
SO _x	8.08	6.97	4.37	4.03
Gross CO ₂	8,313	7,983	8,064	8,146
Eff. CO₂	8,313	7,159	5,598	4,045

Table 6 shows the emissions data for the baseline cases. Notice that the overall emissions for each type of pollutant universally decrease with the amount of biomass added. However, on a per unit output energy basis (ton/MW-year), the CO₂ emissions actually *increase* for increasing BMR in the supercritical IGCC cases, albeit by a very small amount. The only exception is from 0% to 10% BMR, where there is a sharp decrease just for adding biomass to the feedstock. While the CO₂ emissions beyond this point do increase, note the fact that the emissions for the biomass blends are still *always lower* than they are for pure coal. The *effective CO₂* on the other hand always decreases with increasing BMR. Again, the effective CO₂ is determined by calculating the neutral CO₂ from biomass and subtracting it from the gross CO₂.

The emissions for the post-combustion plants are shown in Table 7. In addition to NO_x emissions being virtually eliminated and SO_x emissions being cut by more than 98%, the

CO₂ emissions clearly drop by a significant amount. The reduction in SO_x and NO_x occurs because of (1) the fact that post-combustion CCS uses chemical absorption, which allows for the direct removal of SO_x and NO_x and (2) the fact that this form of capture is performed *after* SO_x and NO_x have already formed in addition to performing necessary cleaning beforehand (including Acid Gas Removal). Even the baseline cases make use of AGR, but only the post-combustion cases perform sulfur removal a second time, after combustion occurs. In addition, notice that after just 10% biomass is added, the plant has become carbon-negative. This is true even for the supercritical cycle, because post-combustion CCS is the only implementation of CCS that can clean the emissions released by the supercritical plants' duct burners.

Table 7 Emissions (Tons/MW-year)-Post-Combustion CCS

Biomass/Coal Ratio	0%	10%	30%	50%
Subcritical Plants				
NO _x	2.69x10 ⁻⁸	2.63x10 ⁻⁸	2.67x10 ⁻⁸	2.71x10 ⁻⁸
SO _x	0.058	0.049	0.039	0.029
Gross CO ₂	1,108.1	1,049.0	1,063.5	1,077.7
Eff. CO₂	1,108.1	-113.1	-2,426.2	-4,756.2
Supercritical Plants				
NO _x	2.42x10 ⁻⁸	2.38x10 ⁻⁸	2.42x10 ⁻⁸	2.45x10 ⁻⁸
SO _x	0.052	0.045	0.035	0.026
Gross CO ₂	1,034.4	987.0	999.5	1,012.0
Eff. CO₂	1,034.4	-67.0	-2,162.0	-4,356.8

Lastly, for the economic impact of these plants, see Tables 8 and 9. Note that, the additional \$10,000,000 for a torrefaction plant is ~1% of the total capital cost, meaning that it is insignificant compared to the total plant cost. Not to mention, co-gasifying biomass with coal actually *reduces* the total investment by a significant amount. The cost analysis program report (not shown) showed that the biggest saving is in the piping system and the gasifier itself. Not entirely clear at first, but taking into account that the GT inlet temperature and mass flow rate are fixed, this leads to the discovery that using biomass in the gasifier means that a *smaller gasifier can be used* and the plant will still get the same net GT power output. This is possible because of the reduced necessary syngas flow rate to the GT, since the syngas of the biomass blends have higher heating values than that of coal alone. This difference alone accounts for nearly 80% of the price reduction seen in the tables. This results in reductions of the capital cost for both subcritical and supercritical IGCC plants.

The cost of electricity (CoE) actually decreases from 0% to 10% BMR due to the reduced size of the cleanup and gasification islands for both sets of cases. However, it rises again beyond 30% due to the added extra cost of the biomass. CoE is calculated based on levelized capital cost, O&M costs, interest, and the costs of water and fuel.

Table 8 Economics - Baseline

Biomass/Coal Ratio	0%	10%	30%	50%
Subcritical Plants				
Capital cost (\$Mil)	1,029.8	926.74	911.62	897.44
Capital Cost (\$/kW)	4,363	3,904	3,891	3,880
CoE (\$/kW-hr)	0.1008	0.0979	0.1084	0.1190
Supercritical Plants				
Capital cost (\$Mil)	1,087.6	983.83	970.95	956.03
Capital Cost (\$/kW)	4,072	3,668	3,663	3,648
CoE (\$/kW-hr)	0.0972	0.0947	0.1041	0.1133

Table 9 Economics – Post-Combustion CCS

Biomass/Coal Ratio	0%	10%	30%	50%
Subcritical Plants				
Capital cost (\$Mil)	1,490.2	1,374.8	1,359.7	1,345.4
Capital Cost (\$/kW)	8,015	7,226	7,257	7,288
CoE (\$/kW-hr)	0.1713	0.1631	0.1763	0.1895
CCS cost (\$/kW-hr)	0.0705	0.0652	0.0679	0.0705
CO₂ Removal Cost (\$/ton)	71.99	66.86	65.02	66.12
Supercritical Plants				
Capital cost (\$Mil)	1,539.8	1,422.5	1,407.4	1,392.5
Capital Cost (\$/kW)	7,457	6,781	6,805	6,828
CoE (\$/kW-hr)	0.1626	0.1559	0.1678	0.1797
CCS Cost (\$/kW-hr)	0.0654	0.0612	0.0637	0.0664
CO₂ Removal cost (\$/ton)	71.88	67.75	65.67	63.22

Post-combustion CCS, however, has a clearly adverse effect on both CoE and capital cost. This is not surprising, considering the fact that post-combustion CCS requires the processing of thousands of tons of exhaust gases a day, all at atmospheric pressure. The unit itself must be hundreds of meters high, and uses up nearly \$24,000 worth of makeup solvent alone in a single day. At an **additional 6-7 cents/kW-hr** in CoE and nearly \$500/kW in capital costs, using post-combustion CCS may not be viable for this type of plant. The CCS cost follows the same overall pattern as the total CoE does: decreasing at 10% BMR, but increasing beyond this amount. Despite the later increase, the biomass blend cases still have a lower CCS cost than the pure coal cases.

The CCS cost in terms of COE can also be converted to the cost for removing CO₂ (per unit ton basis) using Eq. (5):

$$CO_2 \text{ Removal Cost } \left(\frac{\$}{\text{ton}} \right) = \frac{CoE_{CCS} - CoE_{base}}{\left[\frac{\text{tons}_{CO_2}}{MW-yr} \right]_{net} + \frac{1MW}{1000kW} \cdot \frac{1yr}{8000hrs}} \quad (5)$$

where the “net” tons of CO₂ refers to the difference in effective tons of CO₂ per unit power between the baseline case and the CCS case. The “CO₂ Removal Cost” is also called “CO₂ Avoided Cost” in the U.S. Department of Energy’s Volume 3 report on fossil power plants [25]. If a “carbon tax” is implemented, the CO₂ Removal Cost can also be referred to as the “Break-even Cost” for avoiding the carbon tax. Only when the carbon tax is higher than the CO₂ removal cost is implementing CCS justified.

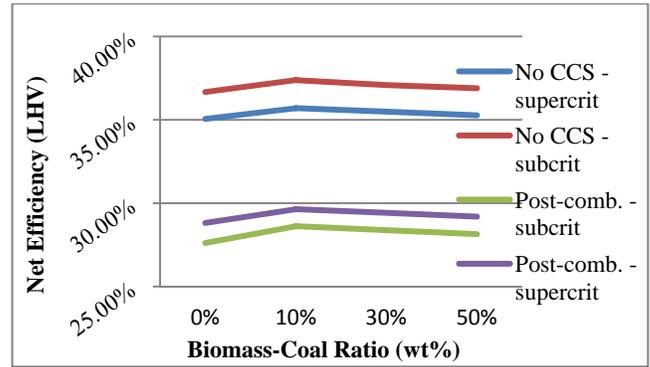


Figure 6 Net plant efficiency for all plants

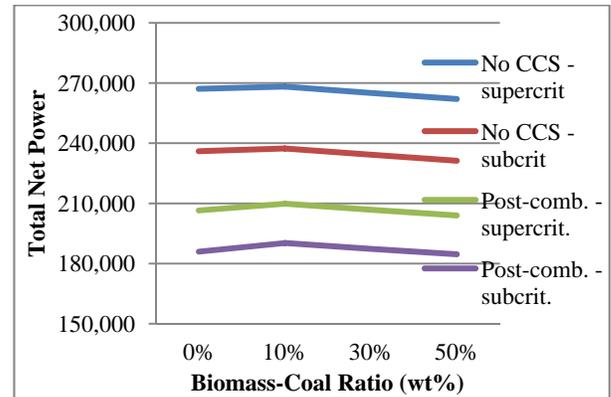


Figure 7 Total net power for all plants

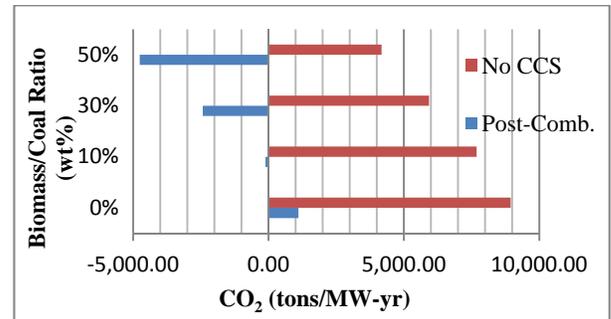


Figure 8 Effective CO₂ emissions for sub-critical plants

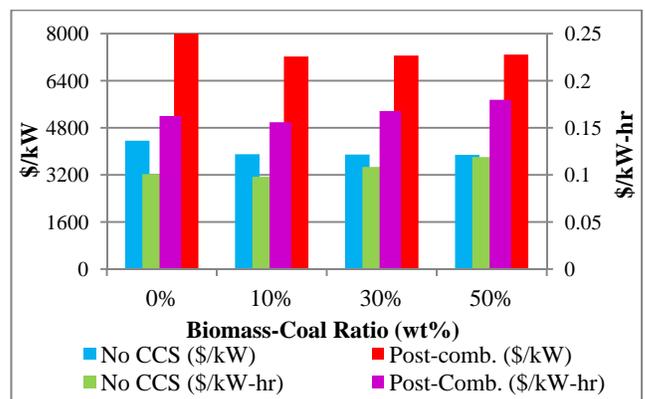


Figure 9 Economics for sub-critical plants

As seen in Tables 8 and 9, a \$70/ton carbon tax would only work for those plants that include biomass in the feedstock. For pure coal, a tax of about \$72/ton is necessary to justify CCS, whereas all of the biomass cases are below \$68/ton, with one case as low as \$63/ton. For more qualitative data on these plants, see Figs. 6 through 9 and the Appendix.

CONCLUSIONS

In summary, this study was performed using GTPro®, a program from the Thermoflow® software suite. It uses a GE/Texaco gasifier and Siemens-Westinghouse SGT6-4000F gas turbine, and the plant was assumed to be constructed in southern Louisiana using Texas Lignite and sugarcane bagasse as fuels. The results show that the net plant efficiency increases at 10% BMR for both sets of cases, but decreases thereafter. However, the efficiency of the blended cases remains higher than that of the baseline cases for *all blend ratios*.

The emissions (NO_x, SO_x, and effective CO₂) and the capital costs all decrease as the biomass ratio increases. However, the cost of electricity increases with BMR due to how expensive obtaining the biomass is. With these results in mind, the following conclusions can be drawn:

- The supercritical system is universally superior to the subcritical system regardless of which case is taken into consideration. It always provides 20-30MW (9-12%) more power at a reasonable extra cost (about \$300/kW less expensive) and always has the lower CoE (0.3 cents/kW-hr less). The net emissions also decrease by 400-500 tons/MW-yr (6-7% less) overall.
- Post-combustion carbon capture is both thermally and economically detrimental to the IGCC baseline: capital costs increase by \$3,700/kW (nearly double that of the baseline) and CoE increases by about \$0.07/kW-hr. However, the emissions decrease by nearly 7,000 tons/MW-yr through the use of CCS: enough for the plant itself to become carbon-negative using only 10% biomass.
- CO₂ removal cost is about \$72/ton for both subcritical and supercritical systems. Blending 10% biomass reduces removal cost about \$4-5/ton.
- Biomass blends are always better than pure coal, thermally and economically, but too much biomass (beyond 10% BMR) begins to inhibit power output and efficiency. Blending 10% biomass tends to increase the efficiency by about 0.7-1.0 percentage points and the output power by 2-5 MW. The capital costs per kW and CoE also decrease by \$400-\$800/kW and 0.3-0.8 cents/kW-hr, respectively. However, further biomass beyond 10 BMR decreases the efficiency and the power by up to 0.5 points and 6MW, respectively. The CoE increases by up to 2 cents/kW-hr, with or without CCS, but the capital cost can increase or decrease due to the effect of the CCS system.

ACKNOWLEDGEMENTS

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Appendix: Plant Data for subcritical, post-combustion CCS plant with 10% biomass

GT PRO 21.0 parallel
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 rk\Thermoflow\Better Plant Design\SUBCRITICAL QUENCH - 10% BIOMASS - POST-CCS.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)	200017		8307		41.08	
Steam Turbine(s)	72910					
Plant Total	272928	190260	8311	11923	41.06	28.62

PLANT EFFICIENCIES			
PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on chargeable energy, %	Canadian Class 43 Heat Rate, BTU/kWh
9.86	-8.90	20.39	76480

GT fuel HHV/LHV ratio =	1.083		
DB fuel HHV/LHV ratio =	1.083		
Total plant fuel HHV heat input / LHV heat input =	1.103		
Fuel HHV chemical energy input (77F/25C) =	2501654	kBTU/hr	694904 BTU/s
Fuel LHV chemical energy input (77F/25C) =	2268430	kBTU/hr	630120 BTU/s
Total energy input (chemical LHV + ext. addn.) =	3283522	kBTU/hr	912089 BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	3183488	kBTU/hr	884302 BTU/s

GAS TURBINE PERFORMANCE - SGT6-4000F (Physical Model #141)					
	Gross power output, kW	Gross LHV efficiency, %	Gross LHV Heat Rate BTU/kWh	Exh. flow lb/s	Exh. temp. F
per unit	200017	41.08	8307	1026	1102
Total	200017			1026	

Number of gas turbine unit(s) =	1		
Gas turbine load [%] =	100	%	
Fuel chemical HHV (77F/25C) per gas turbine =	1799919	kBTU/hr	499978 BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1661637	kBTU/hr	461566 BTU/s

STEAM CYCLE PERFORMANCE				
HRSG eff. %	Gross power output kW	Internal gross elect. eff., %	Overall elect. eff., %	Net process heat output kBTU/hr
61.10	72910	39.30	24.01	-851003

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 =	1036152	kBTU/hr	287820 BTU/s
Water/steam to gasification plant =	25133	kBTU/hr	6981 BTU/s
Water/steam from gasification plant =	47011	kBTU/hr	13059 BTU/s
Net process heat output as % of total output (net elec. + net heat) =	0	%	

ESTIMATED PLANT AUXILIARIES (kW)	
GT fuel compressor(s)*	0
GT supercharging fan(s)*	0
GT electric chiller(s)*	0
GT chiller/heater water pump(s)	0
HRSG feedpump(s)*	685.4
Condensate pump(s)*	118.6
HRSG forced circulation pump(s)	0
LTE recirculation pump(s)	1.458
Cooling water pump(s)	1039
Air cooled condenser fans	0
Cooling tower fans	0
HVAC	40
Lights	65
Aux. from PEACE running motor/load list	630.5
Miscellaneous gas turbine auxiliaries	362
Miscellaneous steam cycle auxiliaries	155.6
Miscellaneous plant auxiliaries	272.9
Constant plant auxiliary load	0
Gasification plant, ASU*	34484
Gasification plant, fuel preparation	9683
Gasification plant, AGR*	588.9
Gasification plant, other/misc	2408.8
Desalination plant auxiliaries	0
CO2 capture plant auxiliaries*	30769
Program estimated overall plant auxiliaries	81303
Actual (user input) overall plant auxiliaries	81303
Transformer losses	1364.6
Total auxiliaries & transformer losses	82668

* Heat balance related auxiliaries

IGCC PLANT HEAT BALANCE	
Total Energy In	1076128 BTU/s
Power Block Energy In:	
Ambient air sensible	11178
Ambient air latent	21278
External gas addition to combustor	0
Steam and water	281970
Process return & makeup	0
CO2 capture condensate return	41684
Gasifier Energy In:	
Gasifier fuel enthalpy	697915
Gasifier slurry water	830.4
Quench water	28494
Gas Cleanup System Energy In:	
Scrubber water	1285.9
Syngas moisturizer water	0
Syngas moisturizer heat addition	0
Air Separation Unit Energy In:	
Ambient air - sensible & latent	8525
Total Energy Out	1076401
Power Block Energy Out:	
Net power output	180340
Stack gas sensible	78963
Stack gas latent	114394
GT cycle losses	5198
GT ancillary heat rejected	0
GT process air bleed	0
Condenser	159671
Process	45580
Steam cycle losses	3984
Non-heat balance auxiliaries	13894
Transformer losses	1293.5
CO2 capture auxiliary	29164
Steam to CO2 capture	195549
Gasifier Energy Out:	
Heat losses	63.01
Slag	13843
Gas Cleanup System Energy Out:	
H2S removal	4532
CO2 removal	0
Cooling after CO shift	0
Water condensed from syngas	19418
Syngas export	0
H2 export	0
AGR Orej	5723
AGR heat loss	533
Other	0
Cooler heat rejection to external sink	164890
Air Separation Unit Energy Out:	
Discharge gas	7799
Heat rejection from compressor inter/after cooling	30818
Compressors mechanical & electrical losses	1634.3
ASU heat rejection to external sink	0
Energy In - Energy Out	-272.2
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)	-0.0283