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CASE STUDIES FOR VARIOUS IGCC PARAMETERS USING BLENDED COAL/BIOMASS WITH SUPERCRITICAL STEAM BOTTOM CYCLES

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ABSTRACT

Great efforts have been spent to reduce the greenhouse gas (GHG) emissions and improve the efficiency of the Integrated Gasification Combined Cycle (IGCC). This study focuses on investigating two approaches to achieve these goals. First, replace the traditional subcritical Rankine steam cycle of the overall plant with a supercritical steam cycle. Second, add different amounts of biomass as feedstock to reduce carbon footprint as well as the SO_x and NO_x emissions. The goal of this study is to examine the thermal and economic impact of different design implementations for an IGCC plant. The parametric dichotomies investigated were: radiant cooling vs. syngas quenching, dry-fed vs. slurry-fed gasification (particularly in relation to sour-shift and sweet-shift carbon capture systems), oxygen-blown vs. air-blown gasifiers, low-rank coals vs. high-rank coals, and options for using syngas or alternative fuels for the duct burner in the heat recovery steam generator (HRSG) to raise achieve the desired steam turbine inlet temperature.

Employing biomass as a feedstock has the advantage of being carbon neutral or even carbon negative if carbon is captured and sequestered (CCS), whether the goal is to generate chemicals or provide electrical power. However, due to a limited supply of feedstock, biomass plants are usually small, which results in higher capital and production costs. Considering these challenges, it is more economically attractive and less technically challenging to co-gasify biomass wastes with coal. Using the commercial software, Thermoflow®, the case studies were performed on a simulated 250 MW coal IGCC plant located near New Orleans and co-fed with biomass from 10% to 50% by weight.

The analysis is conducted using lower heating value (LHV) and 2011 USD as the standard. The results show that syngas coolers are more efficient than quench systems (by 5.5 percentage points), but are also more expensive (by \$500/kW and 0.6 cents/kW-hr). For the feeding system, dry-fed is more efficient than slurry-fed (by 2.2-2.5 points) and less expensive (by \$200/kW and 0.5 cents/kW-hr). Sour-shift CCS is both more efficient (by 3 percentage points) and cheaper (by \$600/kW or 1.5 cents/kW-hr) than sweet-shift CCS Natural gas is a better duct burner fuel than syngas (by 1.7 percentage points efficiency, \$400/kW capital, and 0.5 cents/kW-hr CoE). Higher-ranked coals are more efficient than lower-ranked coals (2.8 points without biomass, or 1.5 percentage points with biomass), and have lower capital cost (by \$600/kW without using biomass, or \$400/kW with biomass.) Without biomass, they produce a lower total CoE (by 0.1 cents/kW-hr), but are 0.21 cents/kW-hr *more* expensive with biomass. Finally, plants with biomass and low-rank coal feedstock are both more efficient and have lower costs than those with pure coal: just 10% biomass seems to increase the efficiency by 0.7 points and reduce costs by \$400/kW and 0.3 cents/kW-hr. However, for high-rank coals, this trend is different: efficiency decreases by 0.7 points and CoE *increases* by 0.1 cents/kW-hr, but capital costs still decrease by about \$160/kW.

NOMENCLATURE

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

1. INTRODUCTION

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. Previous studies by Long and Wang [1-3] focused on the plant as a whole, and incorporated a supercritical steam cycle into the power block to raise the output power and the efficiency while reducing capital costs. Employing biomass as a feedstock has the advantage of being carbon neutral or even carbon negative if carbon is captured and sequestered (CCS), whether the goal is to generate chemicals or provide electrical power. Therefore, biomass was also included in the feedstock, and various carbon capture schemes were implemented to reduce the emissions, as well [2, 3]. One additional objective of this study is to alter the *Biomass Ratio,* or BMR for short, with a primary interest in finding the effects of biomass on GHG emissions, as well as the overall changes in power, efficiency, and economics.

The goal of this study is to examine the thermal and economic impact of different design implementations for an IGCC plant. The parametric dichotomies investigated were: radiant cooling vs. syngas quenching, dry-fed vs. slurry-fed gasification (particularly in relation to sour-shift and sweet-shift carbon capture systems), oxygen-blown vs. air-blown gasifiers, low-rank coals vs. high-rank coals, and, options for using syngas or alternative fuels for the duct burner in the heat recovery steam generator (HRSG) to raise the steam turbine inlet temperature.

1.1 Supercritical Rankine Bottom Cycle

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. It was during this period where the maximum inlet pressure and temperature were raised from 2400PSI/1000^{o}F to near 4500PSI/1150^{o}F [4]. 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 [5]. 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%. [6] 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, as of this writing, there is still currently no literature available that documents a supercritical steam bottom system being used in any real-world IGCC system.

1.2 Biomass

As for using biomass, the first pure biomass IGCC plant was constructed in Värnamo, Sweden in 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 [7]. 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 [8] and the McNeil Station in Burlington, Vermont [9]. 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 [9]. All of these plants, however, have either failed due to some technical difficulties or been removed from the commercial power sector due to not being economically competitive. More detailed discussions about these plants are undertaken in previous papers [1-3, 10].

Due to the controversies surrounding issues of cultivating energy crops, in this study, only biomass wastes are considered. For convenience, the word "wastes" from this phrase 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 and short life-span. 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. Next, since biomass is cleaner than coal is, co-feeding results in lower emissions than a pure coal plant, and such plants are able to provide much more power than any 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 much higher than those of coal, to the point of being nearly impossible to achieve realistically. Also, 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 [10]. 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. 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.

1.3 Biomass Pretreatment

To overcome this 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^{o}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 [11]. During torrefaction, the biomass looses roughly 30% of its mass as torrefaction gases, and roughly 10% of its internal energy with them [12]. A simple algebraic calculation shows that this would result in about a 28% increase in the calorific value per unit mass for the feedstock [13].

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. 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'd been having, investing in one might not be economically viable [8]. 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 [12].

1.4 Coal-Biomass Co-gasification

While some biomass-coal co-feeding studies have been done in the realm of *co-combustion* of biomass with coal [14, 15] 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 for biomass, and the emissions of NO_x and SO_x were reduced about 10% [16]. 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 [17].

1.5 Carbon Capture and Storage (CCS)

Carbon capture is the next logical addition to such a plant given the concerns on GHG's effect on global warming: 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 [18]. When CCS is combined with biomass, it is possible, assuming biomass is carbon-neutral, for a plant to become *carbon-negative*, meaning that, although emissions are produced, there is a net *decrease* in the amount of carbon put into the atmosphere, given that plant-based biomass spends a great portion of its lifetime consuming CO_2 through photosynthesis.

This study, like the previous studies from the same authors (Long and Wang, [1-3]), focuses on investigating *co-gasification* of biomass and coal for application in IGCC systems with both subcritical and supercritical bottom Rankine cycle systems with and without carbon capture, but with several parametric changes that will, as mentioned previously, be analyzed/compared to the previous case studies. In all, the focus of this study is to investigate the effects of these special parameters, and how they affect the performance of the plants utilized in the previous studies. Note that the purpose of this paper is *not* to compare the supercritical and subcritical plants with each other. This study is part of a much larger, more exhaustive study, and that comparison has already been discussed at length in previous writings [1-3].

2. PLANT DESIGNS AND ECONOMIC ANALYSIS

2.1 BASELINE PLANT DESIGN

To keep things as concise as possible, a quick review of the baseline cases considered from the previous study is given here. The software used for this study is 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. All of the plants are based around a design of 200-240MW of NET power. 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, 2-stage ST, with a fixed steam inlet temperature at 1000^{o}F and pressure at 1100 PSI. In addition, a second baseline using a supercritical cycle with a fixed inlet temperature of 1200^{o}F and 2400 PSI also exists.

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 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, for all plants that require one, 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. The gasifier itself for the baseline cases is assumed to be modeled after the GE/Texaco gasifier, with an operating temperature and absolute pressure of 2000°F and 500 PSI, respectively. An outline of the baseline plant studied without any extras or improvements, such as carbon capture is shown in Fig. 1,

All plants are assumed to be built around New Orleans, Louisiana, at an elevation of 10 feet above sea level. The climate condition is 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.) are not used as the baseline because those conditions are not common for Louisiana on the whole. It is deemed better to be more conservative with the model prediction by using conditions applicable to a Louisiana late summer/early fall.

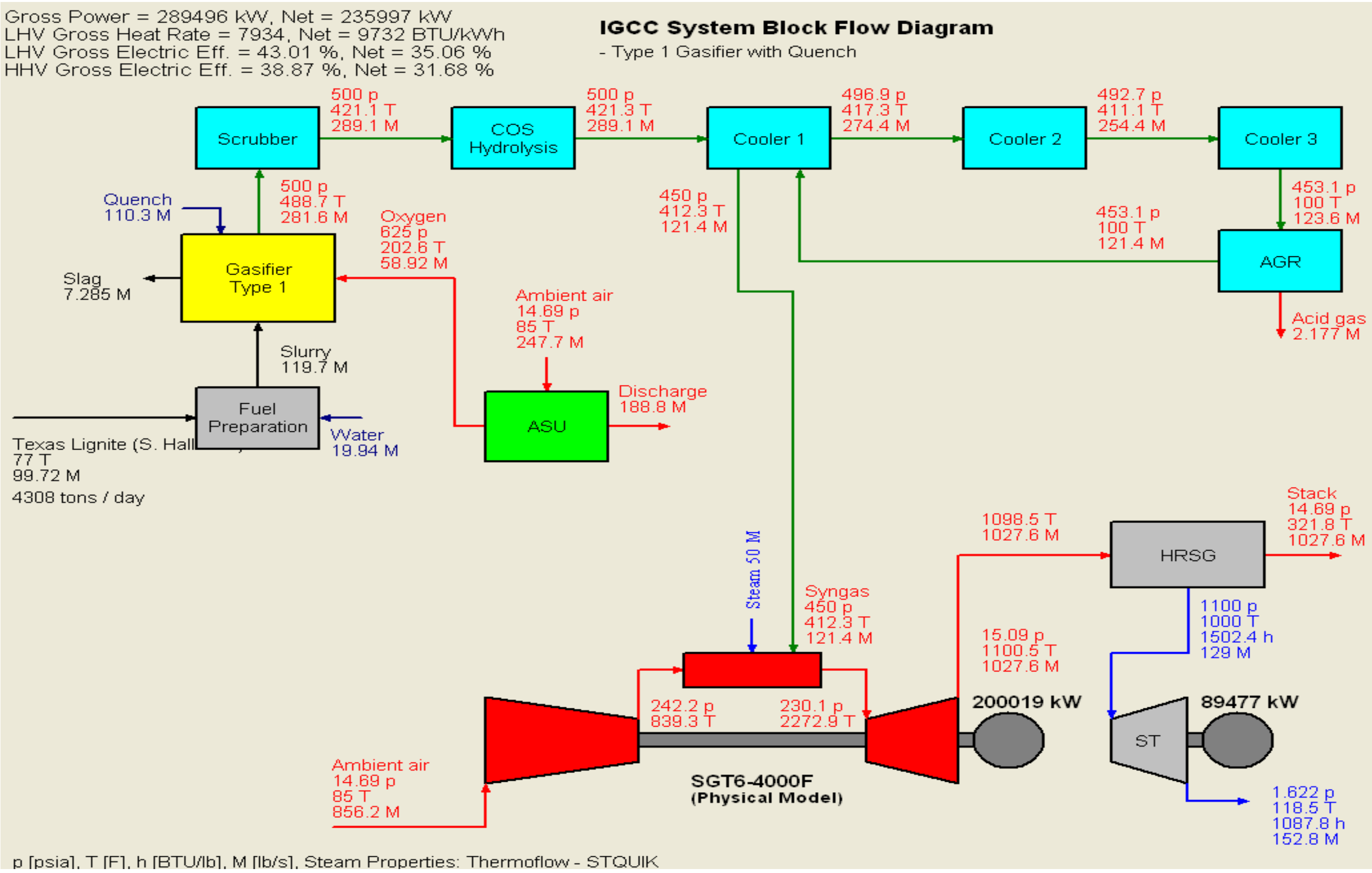

Figure 1 Baseline plant layout with quench cooling and no carbon capture

Before the syngas can enter the cleanup system and be cleaned, it must be cooled down. This is because the exit temperatures of most gasifiers is very high, and the known processes available commercially for cleaning the syngas can only be used within specific temperature ranges, since many of the processes, especially COS hydrolysis, are dependent upon specific chemical reactions. As such, it becomes necessary to reduce the temperature of the raw syngas before any cleaning can be done.

The baseline uses a quench design to cool the syngas. A quench system is simpler and cheaper to implement than another design (see Case 1) using radiant syngas coolers (RSCs) and convective syngas coolers (CSCs). Quench means that low-temperature water or steam, usually from an external source, is sprayed directly into the syngas stream upon exiting the gasifier. Efficiency-wise, quenched systems are generally inferior to radiant and convective cooled systems. For one, quenched systems heavily dilute the syngas produced, meaning lower heating values for the cleaned syngas if the water is not condensed and drained out at a later point. A quench can also be justified if the system is to make use of carbon capture, as the extra water can be used in the CO-shift reactor to avoid having to sacrifice too much extra water to complete the shift process.

The baseline case uses a quench of water at 300°F, provided up to the point that the syngas relative humidity reached 50%. The air separation unit (ASU) used is slurry-fed, and use 10% biomass alongside the otherwise Texas Lignite feedstock. No carbon capture is considered and a traditional subcritical Rankine cycle is used as bottom cycle. A comparison of the gasification block used by these plants can be seen in Fig. 2. (Note that the numbers shown in the quench gasifier are representative of pure coal or 0% BMR, while the one with coolers has the appropriate 10% BMR.)

Case 1: Radiant syngas coolers (RSCs) and Convective syngas coolers (CSCs)

Radiant syngas coolers (RSCs) and Convective syngas coolers (CSCs) are essentially large heat exchangers that typically use water or low-grade steam as the coolant source. This water or low-grade steam most often belongs to the closed-loop steam system of the bottom cycle as a way of "preserving" the energy that will be lost from the syngas stream before it can be cleaned. As such, designs that use RSCs and/or CSCs are at least 2-3% more efficient than quench designs. However, radiant coolers in particular are very expensive and bulky devices and, depending on the system involved, may not be worth the initial economic investment, despite the increase in efficiency. The result of Case 1 will provide detailed information for comparison with the baseline case.

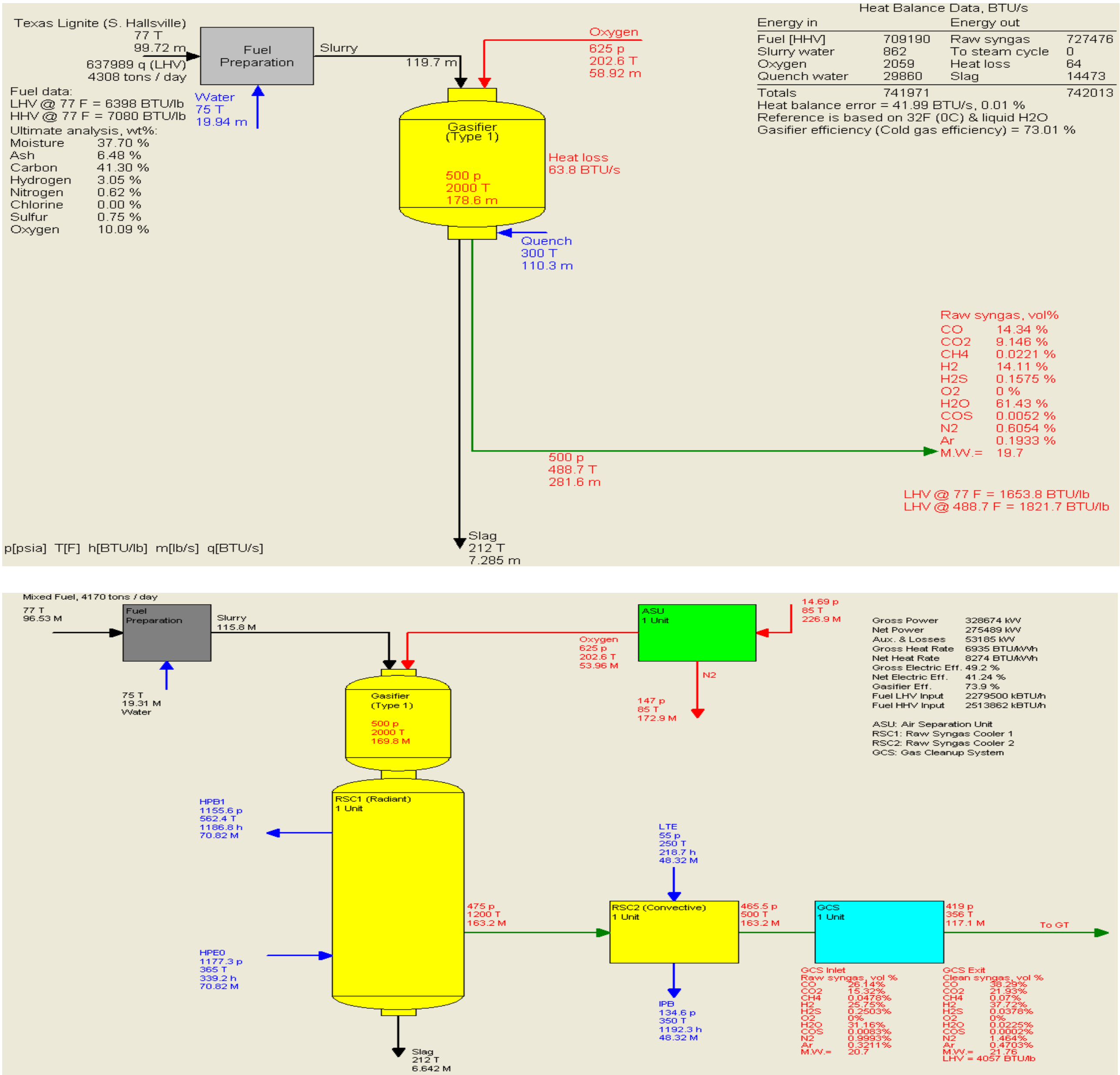

Figure 2 Quench system (top) vs. Radiant and convective coolers (bottom)

Cases 2 & 3: Dry-fed vs. Slurry-fed Gasification (30% BMR, with carbon capture)

Another parameter that is being considered is the idea of using a dry feedstock, where the baseline original study used a slurry feedstock. Dry-fed gasifiers are by and large cheaper than slurry-fed ones. Dry-fed systems use more feedstock for the same mass flow rates due to lack of water. Since biomass benefits greatly from CCS, as it has the advantage of being carbon-negative, 30% BMR with CCS is determined to be the best point at which to examine this parameter. Case 2 corresponds to a comparison with sour-shift CCS, while Case 3 corresponds to the same type of comparison, but with sweet-shift CCS. All other parameters are the same as the baseline.

Sour-shift means that the water-gas shift chemical reaction is performed *before* acid gas removal, while sweet-shift means that it is performed afterwards. With sour-shift, this means that the shift process can be integrated with COS hydrolysis, as both reactions are similar in nature, and that AGR and CCS can be performed at the same time, using the same system. For sweet-shift, the entire CCS process is completed at the end of the cleanup process. Because there is no integration with any other component, the process is easier to perform and control, however it can be more expensive, since more new equipment is needed. A more detailed discussion and analysis of sweet-shift versus sour-shift has been included in other studies [2, 3].

The gasifier chosen for the dry-fed scenario is the Shell gasifier, shown in Fig. 3, which uses an internal water-cooled membrane. Since there is no slurry, the feedstock is transported into the gasifier by means of a small amount of high-pressure nitrogen from the ASU, (notice in Fig. 3, it's only about 9 lbs/s, < 5% of the oxygen flow rate.) The gasifier also requires steam injection for additional gasification agent. The additional steam is taken from the HP stream, since it's the only place with high enough pressure to provide this steam. The cooling water for the membrane, meanwhile, is taken from the IP stream of the HRSG, since this stream doesn't have a pressure requirement. Finally, in order to keep the comparison as close as possible to the baseline case, the gasifier also uses water to quench the syngas, with the same stipulations as the baseline quench study (300°F water added until 50% R.H. is achieved.)

The CCS systems utilize an amine-based absorption system based on the Selexol® process: with an absorber column to remove the CO_2 and a reboiler to provide the energy for the absorption cycle. Sweet-shift uses a stripper column to pull the CO_2 out of solution with the solvent via temperature change, while sour-shift makes use of two flash tanks to do the same using pressure flashing. Both systems use CO_2 compression at the end for sequestration purposes. How the process of CO_2 after this point is left open-ended. A more detailed discussion and analysis of CCS has been included in other studies [2, 3].

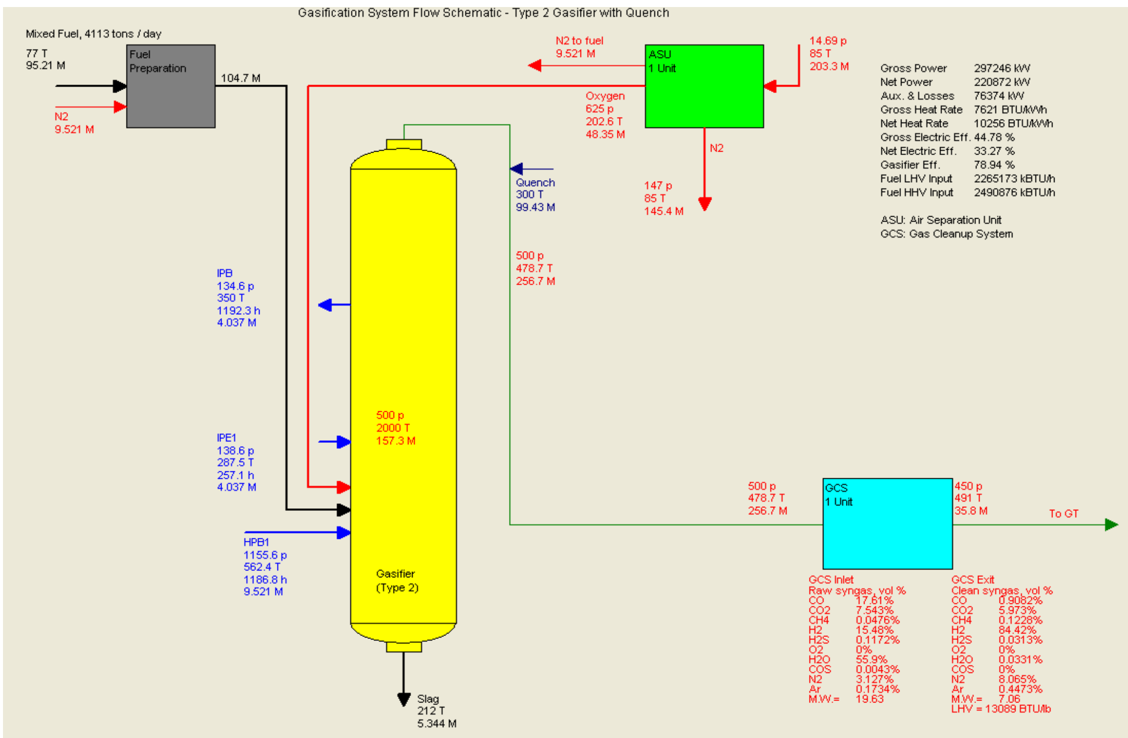

Figure 3 Dry-fed gasification system [P(psia), T(F), h (Btu/lb), m (lb/s), q (Btu/s)]

Case 4: Air-blown Gasification

Another parameter thought to have interesting implications is the idea of using an air-blown gasifier over an oxygen-blown gasifier. An air-blown gasifier saves a significant amount of energy and O&M costs from not using an ASU, but an oxygen-blown gasifier can produce syngas with a much more appreciable heating value, and it can be used with a smaller cleanup system. Which one ends up being more efficient or having a lower electricity cost is generally system-dependent [18]. The only large-scale commercial gasifier that has ever successfully used an air-blown layout is the 2-stage gasifier developed by Mitsubishi Heavy Industries (MHI), which is shown in Fig. 4. Like Shell's gasifier, MHI uses membrane-wall cooling in their design, which was connected to the IP stream.

Again, to keep with comparison, the gasifier was also subjected to a quench, with the same stipulations mentioned previously. A 35% slurry is used, and the air enters the gasifier with the same pressure as the other cases through the use of a boost compressor (not shown.) Although an MHI gasifier is usually dry-fed, a slurry was imposed upon the design in order to maintain a level of comparison between this case and the baseline case, since it is also slurry-fed. This approach causes much less of a discrepancy with reality, and is a lower source for error, than the alternative (making the GE gasifier dry-fed as well for comparison.) Finally, this gasifier is a two-stage gasifier, so part of the feedstock must be inserted at the second stage. Since this a readily adjustable criterion that is solely dependent on the application, a 50:50 split was arbitrarily chosen as the ratio for fuel sent to either stage.

It is only of concern to see how the efficiency, power, and heat transfer are affected by the use of such a system. Since the effect of an air-blown system is more deeply involved and entrenched within the actual chemistry than this study is meant to delve into, the pure coal case without CCS is sufficient to satisfy this level of interest.

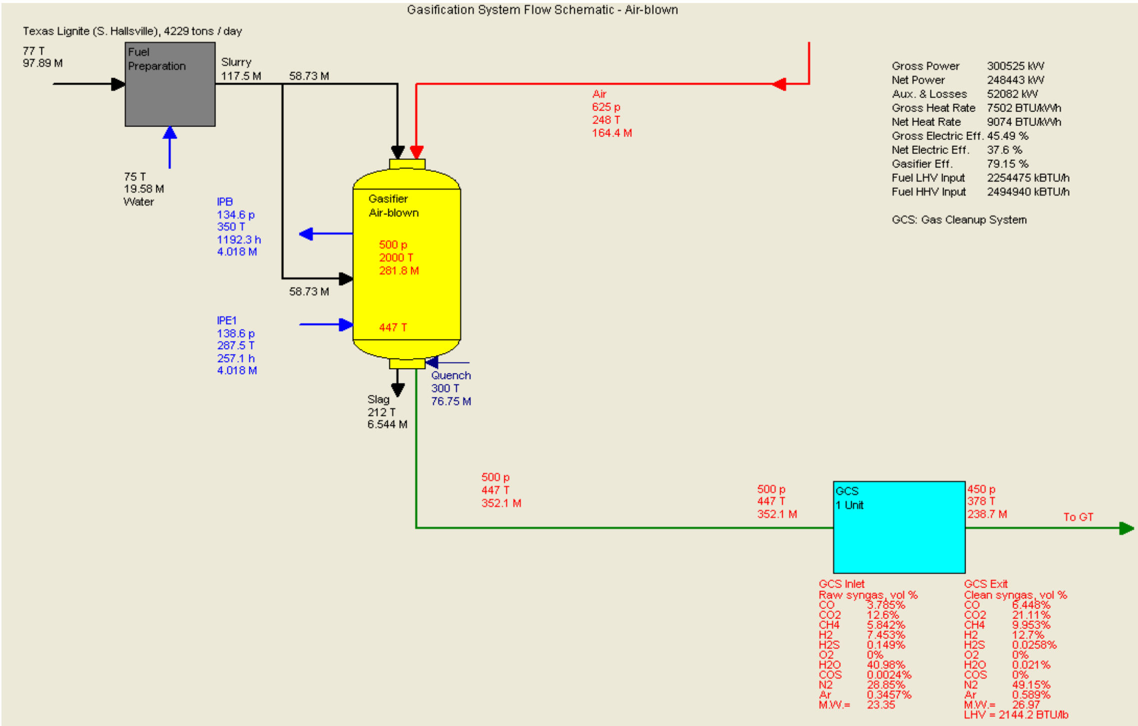

Figure 4 Air-blown gasification system

Case 5: Natural-gas duct burner vs. Syngas duct burner in HRSG

In the Baseline study, the supercritical steam cycle system makes use of natural gas to run the duct burner. A duct burner is required because the GT exhaust does not contain enough energy to heat the steam in the HRSG to the required ST inlet temperature (1200°F.) Using natural gas could cause a problem for some plants' operations, because natural gas prices are volatile and additional gas pipe lines or purchase contracts need to be secured. To make the plant more self-sufficient, the idea was taken to instead to take a small portion of syngas from the GT and burn that directly in the duct burner. The layout for this idea is shown in Fig. 5. Notice the split in the syngas stream: the off-shoot leads to the duct burner (DB.) Whether this is good for the efficiency or not will be seen in the results. Since it is not important for biomass or CCS whether this idea works or not, pure coal is used in the feedstock, and no CCS is implemented for this case.

Cases 6 & 7: Higher-ranked coal

In the baseline study, Texas Lignite is chosen as the fuel because it is readily available and cheap. But, what if a higher grade of coal could be used without the burden of additional transportation costs? Illinois is famous for bituminous coal, almost as much as Pennsylvania, and Illinois #6 is one of the most commonly favored bituminous coals because of its relatively low ash content, very low nitrogen content, and high heating value, despite its lower fixed carbon content (~39%.) Case 6 uses pure Illinois #6 (a bituminous coal), while Case 7 uses 10% biomass mixed in with this coal. The approximate analysis and other data on both coals as well as the biomass used (sugarcane bagasse) can be seen in Table 1.

Table 1 Fuel Approximate Analyses

Fuel	Texas Lignite (S. Hallsville)	Sugarcane Bagasse (dry)	Illinois #6
C (wt%)	41.3	43.59	55.35
H_2 (wt%)	3.053	5.26	4.00
N_2 (wt%)	0.623	0.14	1.08
S (wt%)	0.7476	0.04	4.00
O_2 (wt%)	10.09	38.39	7.47
Cl_2 (wt%)	0	0	0.1
H_2O (wt%)	37.7	10.39	12.0
Ash (wt%)	6.479	2.19	16.0
LHV (Btu/lb)	6398	6714	9599
Price ($/ton)	20.00	65.00	50.00

Taken from: GTPRo® fuel library, EIA, 2009[19], and Day, 2011[20]

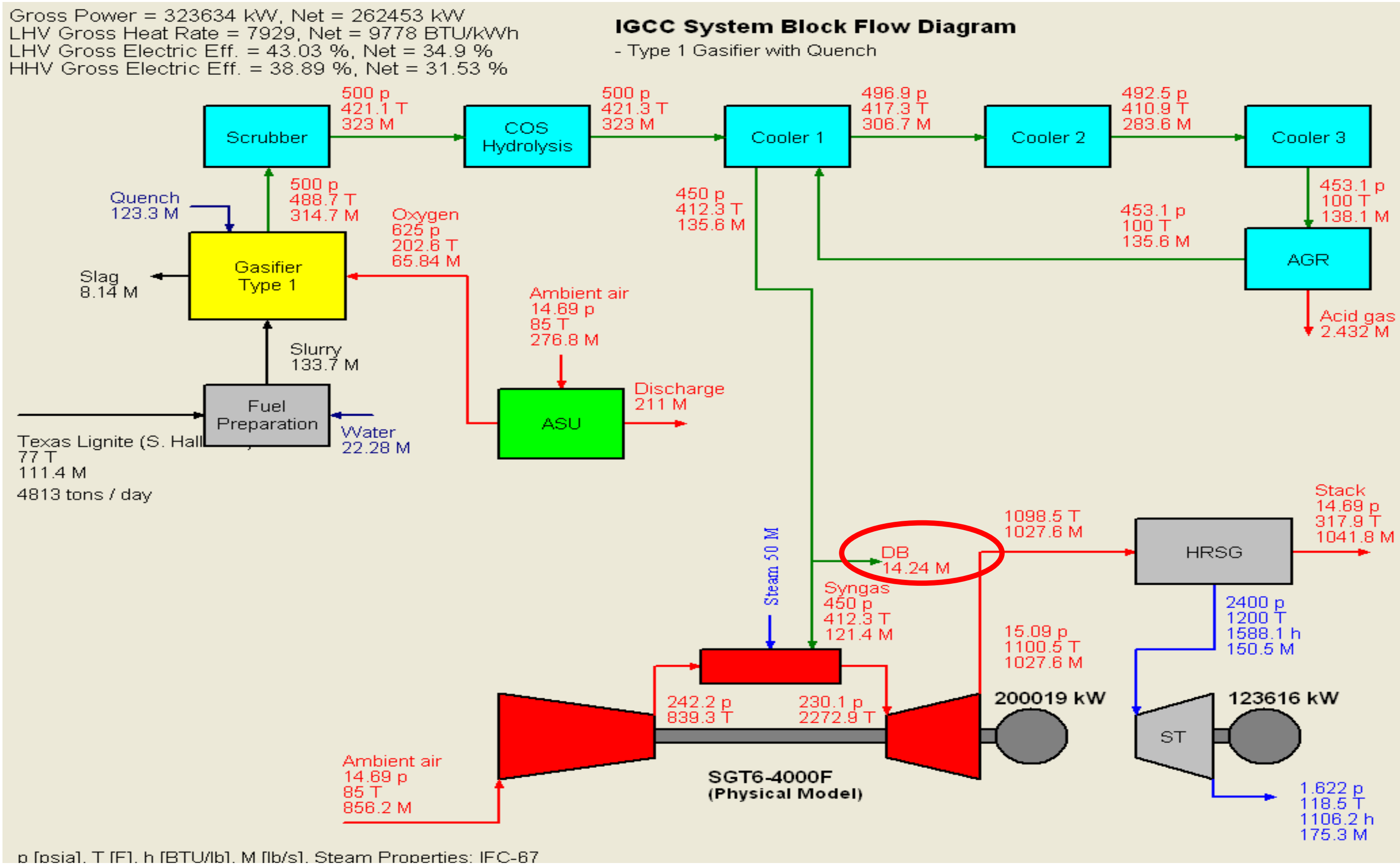

Figure 5 Supercritical plant layout with syngas bleed to duct burner (circled)

The main features of the baseline and deviation from the baseline of each case are summarized as follows:

Baseline: Oxygen-blown, GE gasifier, slurry-fed lignite, 0% BMR, Quench cooled, NO CCS

- Baseline A: subcritical steam cycle
- Baseline B: supercritical steam cycle with NG-fired DB

Case 1a: 10% BMR, **quench cooling**
Case 1b: 10% BMR, **radiant and convective syngas cooling**
Case 2a: Shell gasifier, 30% BMR, **sour-shift CCS, slurry-fed**
Case 2b: Shell gasifier, 30% BMR, **sour-shift CCS, dry fed**
Case 3a: Shell gasifier, 30% BMR,**sweet-shift CCS,slurry-fed**
Case 3b: Shell gasifier, 30% BMR, **sweet-shift CCS, dry fed**
Case 4: **Air-blown**, MHI gasifier
Case 5: **Syngas-fired duct burner**
Case 6: **Illinois bituminous #6 coal**
Case 7: **Illinois bituminous #6 coal**, 10% BMR

Note: Only Case 5 is based off of the supercritical baseline (B). All others are based off of the subcritical baseline (A).

2.2 Economic Analysis

Beginning with fuel choice, lignite is cheap, and, according to the EIA's report [19], lignite from Texas costs approximately $19.00/ton. For bagasse, 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, compared with $30 per ton of raw, unprocessed sugarcane [20].

The price of natural gas duct burner is assumed to be $4.10 per million Btu ($/mmBtu) based on data from June, 2011, when this study was performed [21]. In addition, water consumed by the plant is assumed to be based on *utility*, and the price is set at $2.00 per thousand gallons. Finally, overhead and maintenance (O&M) costs are taken from a report by the EIA, where they are determined to be $60.00/kW (fixed) and $0.006/kW-hr (variable) [22].

Finally, 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 are taken to be around 35%, with 10% flat-interest rates for all plant features. No inflation is considered for this study, so the analysis is based on 2011 USD. Lastly, the total package uses straight-line depreciation, but it is assumed that only 75% of the total investment is available for depreciation for tax purposes. Further details about the economic analysis can be found in the authors' other studies [1, 2].

3 RESULTS AND DISSCUSIONS

3.1 Results of Case 1 (10% BMR, radiant and convective syngas cooling)

For Case 1, using a radiant or convective cooler will always be more efficient than a direct syngas quench for sure, but, again, cost is an important factor in this study. Thus, the issue is whether or not the extra efficiency gained from using a radiant cooler can make up for increased costs.

Table 2 Power and Efficiency: Case 1

Case	Coolers	Quench
Gross GT Power (kW)	200,018	200,018
Gross ST Power (kW)	128,656	89,790
Auxiliary Losses (kW)	53,185	52,451
Total Net Power (kW)	**267,700**	**237,356**
Gross Efficiency (LHV)	49.20	43.59
Net Efficiency (LHV)	**41.24**	**35.70**

Table 2 shows the work and efficiency data for this special case. This shows that the total work output increases by nearly 30MW, all of it from increased steam power. The interesting thing of note is how this happens. The main advantage of using coolers over a quench is the fact that, during a quench, the quality (or grade) of heat from the gasifier is significantly downgraded to a low temperature. Coolers on the other hand, transfer the heat to the steam cycle, producing higher-grade (i.e. higher exergy) steam, which can be utilized more efficiently. Furthermore, in the quench system, the energy in the steam mixed with the syngas will not be as efficiently transferred to the steam cycle as the steam inside the pipes of the radiant or convective cooler will. Although doing this leads to bigger pumps and more auxiliary losses (also seen in the table), the larger mass flow rate through the steam turbine results in a bigger turbine and thus more net work output. In total, the net efficiency from all of this rises by nearly 6 percentage points.

Table 3 Economics: Case 1

Case	Coolers	Quench
Total capital cost (million $)	1,223.4	926.74
Capital cost ($/kW)	**4,441**	**3,904**
CoE ($/kW-hr)	**0.1044**	**0.0979**

The economic data, shown in Table 3, shows that employing coolers costs about $540/kW (12%) more capital and $0.007/kW-hr (7%) more in cost of electricity (CoE) than the quench process. In addition to the coolers themselves, this case requires the purchase of a larger steam turbine, larger pumps, and more piping to construct. This means that it will cost more, even on a per kilowatt basis, and will inevitably have a higher CoE as well. However, the larger total work output translates directly into a higher profit margin, and the economic feasibility of a project that uses coolers may yet be better still than the quench system under the right conditions; for example , when the feedstock cost pronouncedly increases, using syngas coolers would become more attractive. A deeper, more involved study is necessary to determine for certain which plant is a better investment for different applications.

Table 4 Emissions: Case 1

Case	Coolers	Quench
NO_x (tons/year)	233.6	232.5
SO_x (tons/year)	1,890.7	1,869.6
Gross CO_2 (tons/year)	2,055,898	2,045,916
Eff. CO_2 (tons/year)	**1,833,734**	**1,824,817**
Gross CO_2 (tons/MW-year)	7,462.7	8,619.6
Eff. CO_2 (tons/MW-year)	**6656.3**	**7,688.1**

Lastly, the emissions data is shown in Table 4. This shows that the system with the coolers actually produces more total emissions than the quench system. The NO_x and SO_x emissions are comparable, but the gross and effective CO_2 emissions per megawatt are 1100 tons/MW-year (13.5%) lower for the coolers due to the much larger power output (about 30 MW, or 13% more total net power). The increased gross emissions most likely results from the higher purity syngas with lower moisture content from the coolers. The greater mass flow rates of sulfur and nitrogen from the syngas would directly influence the creation of more SO_x and fuel NO_x. (Not shown: syngas from the coolers contains roughly 60-65% more H_2S and COS and 55% more N_2 than that of the quench system.) A similar relationship would be true for CO_2 in relation to itself and CO.

3.2 Results of Cases 2 & 3 (30% BMR, sour-shift and sweet-shift CCS, Dry-fed vs. Slurry-fed Gasification)

For the dry-fed system, the Shell gasifier is used as the base model, as mentioned previously. For this system, two cases are considered, both with pre-combustion CCS: one for sour-shift and one for sweet-shift. This is mainly to highlight the main differences in operation between dry-fed and slurry-fed systems in regards to CCS. Again, the Shell gasifier uses internal gasifier cooling via a wall membrane, but a quench is also added at the end for this case to reduce the exit syngas to the required temperature for gas cleaning like that of the slurry-fed case.

Table 5 Power and Efficiency: Cases 2 & 3

Case	Dry-fed, membrane wall + quench		Slurry-fed, Quenched	
CO-Shift type	Sour	Sweet	Sour	Sweet
Gross GT Power (kW)	200,017	200,017	200,014	200,014
Gross ST Power (kW)	97,229	76,577	99,725	79,202
Auxiliary Losses (kW)	76,374	76,951	82,101	82,310
Total Net Power (kW)	**220,872**	**199,643**	**217,639**	**196,906**
Gross Efficiency (LHV)	44.78	41.37	42.41	39.27
Net Efficiency (LHV)	**33.27**	**29.86**	**30.79**	**27.70**

Table 5 shows the work and efficiency data for all four plant designs. The result clearly shows that the dry-fed system results in a generally higher efficiency (about 2.5 percentage points) than the slurry-fed system for the same type of CCS. It is also observed that sour shift produces more net power output and net plant efficiency than sweet shift case for the same feeding method. A more detailed analysis and discussion are documented in [3].

From the economic data in Table 6, dry-fed is about \$150/kW cheaper in capital cost and 0.5 cents/kW-hr cheaper in CoE than slurry-fed cases. In addition, sour-shift wins again by being \$580/kW cheaper in capital cost and 1.5 cents/kW-hr cheaper in cost of electricity than sweet-shift in both feeding cases.

Table 6 Economics: Cases 2 & 3

Case	Dry-fed, membrane wall and quench		Slurry-fed, Quenched	
CO-Shift type	Sour	Sweet	Sour	Sweet
Total capital cost (million \$)	1,010.3	1,030.6	1,027.4	1,044.2
Capital Cost (\$/kW)	**4,574**	**5,162**	**4,721**	**5,303**
CoE (\$/kW-hr)	**0.1219**	**0.1354**	**0.1269**	**0.1405**

Finally, for the emissions data in Table 7, there appears to be no appreciable difference in the amount of CO_2, NO_x, or SO_x produced. The dry-fed system seems to produce less gross CO_2 on the whole, but the most carbon-negative plant is sweet-shift CCS with a slurry-fed system.

Table 7 Emissions: Cases 2 & 3

Case	Dry fed, membrane wall + quench		Slurry fed, Quenched	
CO Shift	Sour	Sweet	Sour	Sweet
NO_x (tons/year)	189.3	191	181.7	185.0
SO_x (tons/year)	1,465.5	1,475.9	1,545.5	1,569.7
Gross CO_2 (tons/year)	225,088	238,981	313,917	247,882
Eff. CO_2 (tons/year)	**-432,293**	**-423,035**	**-397,266**	**-456,169**
Gross CO_2 (tons/MW-year)	1,018.8	1,197.0	1,311.3	1,258.9
Eff. CO_2 (tons/MW-year)	**-1,956.6**	**-2,119.0**	**-1,584.3**	**-2,316.7**

3.3 Results of Case 4 (Air-blown with MHI Gasifier)

For the air-blown design, the chief concern is whether an air-blown system will increase the efficiency or not. The data for this special case is shown in Tables 8-10.

The work and efficiency data is shown in Table 8. Interestingly enough, the air-blown design is the one that produces the higher efficiency for this plant setup. In addition, more power is generated and there are predictably fewer auxiliary losses. All additional power generated comes from the steam cycle, like in the previous cases. To investigate the cause of this, the syngas composition is analyzed. This is shown in Table 9.

Table 8 Power and Efficiency: Case 4

Case	Air-blown	Oxygen-blown
Gross GT Power (kW)	200,015	200,019
Gross ST Power (kW)	100,511	89,477
Auxiliary Losses (kW)	52,082	53,499
Total Net Power (kW)	**248,443**	**235,997**
Gross Efficiency (LHV)	45.49	43.01
Net Efficiency (LHV)	**37.60**	**35.06**

Table 9 Syngas compositions: Case 4

Product (vol%)	Air-blown	Oxygen-blown
CO	3.785	14.34
CO_2	12.600	9.146
CH_4	5.842	0.0221
H_2	7.453	14.11
H_2S	0.149	0.1575
H_2O	40.98	61.43
COS	0.0024	0.0052
N_2	28.85	0.6054
LHV @ 77°F (Btu/lb)	1407.9	1653.8

The oxygen-blown system clearly has the higher LHV, and, as such, should produce more power. However, the air-blown system's syngas also contains less water overall than that of the oxygen-blown one, and more than 200 times as much CH_4 as the oxygen-blown design. This leads to a lower required syngas mass flow rate to the GT (to maintain the same TIT and work output), which means less need for cleaning, and fewer auxiliary losses.

Table 10 Economics: Case 4

Case	Air-blown	Oxygen-blown
Total capital cost (million \$)	1,009.5	1,029.75
Capital cost (\$/kW)	**4,063**	**4,363**
CoE (\$/kW-hr)	**0.0949**	**0.1008**

From an economic standpoint, as seen in Table 10, the air-blown system is much cheaper than the oxygen-blown system. And, since the efficiency of the cycle is also improved by the design, the CoE also decreases. However, this should not be taken as a universal trend: the air-blown vs. oxygen-blown argument is very complex and there are benefits of both systems, generally. For this set of design criteria, however, the air-blown system just so happens to be the better design.

Emissions were not considered for this case because of the new gasifier design. The air-blown system introduces additional nitrogen to the system, which results in *gasification* NO_x. This extra NO_x content cannot be taken into account, due to lack of data on MHI's gasifier in relation to NO_x production. The SO_x produced would simply be a result of the leftover sulfur from the cleanup system, and, since the gasifier was downsized in this case, can be assumed to decrease compared to the oxygen-blown system. The CO_2 emissions may be different, but they are not as important of a focus for this part of the study, as the main purpose behind switching to an air-blown design was to see how it affected the *efficiency* and overall thermal performance.

3.4 Results of Case 5 (Natural-gas duct burner vs. Syngas duct burner)

Another issue considered is the use of the cleaned syngas in the supercritical cycle's duct burner. The baseline cases used natural gas (approximated as 100% CH_4) for this device, but this introduces an additional cost into the system, and requires that the purchase contract of another input fuel be established. In order to avoid signing up an additional fuel purchase contract and decrease the amount of external heat input into the cycle and make it more self-sufficient, a case is examined to determine what the effects of simply using the cleaned syngas for the duct burner would be. The results of this case can be seen in Tables 11 and 12.

Table 11 Power and Efficiency: Case 5

Case	GT fuel	Natural Gas
Gross GT Power (kW)	200,019	200,019
Gross ST Power (kW)	123,616	122,573
Auxiliary Losses (kW)	61,181	55,481
Total Net Power (kW)	**262,453**	**267,111**
Gross Efficiency (LHV)	43.03	44.29
Net Efficiency (LHV)	**34.90**	**36.67**

Table 12 Economics: Case 5

Case	GT fuel	Natural Gas
Total capital cost (million $)	1,170.3	1,087.58
Capital cost ($/kW)	**4,459**	**4,072**
CoE ($/kW-hr)	**0.1024**	**0.0972**

From the power data in Table 11, it can be seen that using the GT fuel actually increases the total steam power, but the additional auxiliary losses offset this greatly, with the end result of reducing the total net power by about 5MW. The efficiency, of course, suffers as a result, decreasing by about 2 percentage points. These extra auxiliary losses arise due to the increased gasifier size, and required oxygen flow rate. This makes sense because the GT fuel (i.e. syngas) now has to be used for 2 purposes: to produce electricity in the GT itself and to produce more steam through the duct burner. This requires additional syngas flow rate in order to keep the GT mass flow rate and power output rating constant. The only way to achieve this is to increase the gasifier size, which results in greater heat losses and more auxiliary losses relating to gasification, especially the consumption of more ASU power and greater electrical power required to run the acid gas removal system.

Economically speaking, it would follow that the system using the syngas as the fuel for the HRSG's duct burner (Case 5) would be more expensive to build and more costly to operate than the natural gas cycles in the main case. This is confirmed in Table 12. The larger gasifier and more exhaustive cleanup system coupled with the total loss of net power practically guarantees that this would happen.

Emissions, again, are not considered for this case, either, because the objective of this special case is to observe the effects on performance of efficiency and output power.

3.5 Results of Cases 6 & 7 (Bituminous Coal vs. Lignite)

The final cases involve the use of a higher ranked coal: Illinois #6. Looking at Table 1, Illinois #6's price is more than double that of lignite, but its heating value is only about 1.5 times greater. It will be interesting to see how the use of this fuel will affect the efficiency and economics of the plant. The purpose of these two cases is to observe (1) how changing different ranks of coals affects the overall base plant statistics and (2) how Illinois #6 would behave when blended with 10% biomass by weight.

From the power and efficiency data in Table 13, Illinois #6 has a clear benefit regardless of whether it is used alone or blended with biomass. No matter what, the use of Illinois #6 always reduces auxiliary losses (due mostly to being of higher rank, and thus lowering the necessary gasifier flow rate and requiring less oxygen to fully gasify), raising total net power as a result, and raising the net efficiency by at least 2 points. <u>The interesting occurrence here is that using biomass with Illinois #6 *reduces* the total efficiency due to the additional auxiliary losses, whereas it raised the efficiency for the lignite cases.</u>

Table 13 Power and Efficiency: Cases 6 & 7

Case	Ill. #6	Texas Lignite	Ill. #6 + 10% bio	Tex. Lig. + 10% bio
Gross GT Power (kW)	200,017	200,019	200,018	200,018
Gross ST Power (kW)	86,115	89,477	86,291	89,790
Auxiliary Losses (kW)	41,677	53,499	43,689	52,451
Total Net Power (kW)	**244,455**	**235,997**	**242,621**	**237,356**
Gross Eff.(LHV)	44.33	43.01	43.90	43.59
Net Eff. (LHV)	**37.87**	**35.06**	**37.20**	**35.70**

Before there is cause for alarm, it should be known that the gasifier must have decreased in size for the bituminous coal cases. This is the only way to preserve the GT inlet temperature and mass flow conditions given Illinois #6's higher heating value and ability to generate syngases with higher heating values (see Table 14 for the syngas compositions.) The issue with many of these losses is that Illinois #6 contains more than 5 times as much sulfur as South Hallsville lignite. Naturally, this would require extensive cleaning later on. But, due to their similar ash contents, not much additional slag is produced (slag production actually decreases for pure Illinois #6 compared to pure lignite.

Table 14 Syngas compositions: Cases 6 & 7

Case	Ill. #6	Tex. Lig.	Ill. #6 + 10% bio	Tex. Lig. + 10% bio
CO (vol%)	22.05	14.34	20.77	14.98
CO_2 (vol%)	6.245	9.146	6.925	8.776
CH_4 (vol%)	0.1164	0.0221	0.0857	0.0274
H_2 (vol%)	17.7	14.11	16.92	14.76
H_2S (vol%)	0.7552	0.1575	0.6799	0.1434
H_2O (vol%)	52.19	61.43	53.7	60.56
COS (vol%)	0.0306	0.0052	0.0271	0.0047
N_2 (vol%)	0.7101	0.6054	0.6916	0.5726
LHV (Btu/lb)	2463.1	1653.8	2306	1739.8

Table 15 Economics: Cases 6 & 7

Case	Ill. #6	Texas Lignite	Ill. #6 + 10% bio	Tex. Lig. + 10% bio
Total capital cost (million $)	916.62	1,029.75	870.56	926.74
Capital cost ($/kW)	**3,750**	**4,363**	**3,588**	**3,904**
CoE ($/kW-hr)	**0.0996**	**0.1008**	**0.1005**	**0.0979**

From the economic data in Table 15, it can be seen that the Illinois #6 plants reduce the capital cost. However, the CoE doesn't quite follow this trend, as mixing biomass and Illinois #6 together actually causes the CoE to *increase* by about one-tenth of a cent per kW, or 1% overall. From the perspective of pure Illinois #6, adding biomass increases the cost of preparation, and requires a bigger increase in gasifier size than that of the pure lignite case when biomass is added (this is also a contributing factor to why the drop in total capital cost is lower for the bituminous cases), which increases CoE. From the perspective of lignite with biomass, changing out lignite to Illinois #6 does not reduce the auxiliary losses by as much as it did without the biomass. In the cases with biomass, while the total investment decreases by about $400/kW (~8%) due to the reduced prices of the gasifier and certain cleanup system components, the amount of operating expenses saved on things like water import and electrical usage cannot make up for the new, larger price tag of the coal, increasing the CoE by about 0.25 cents/kW (~2.6%).

Table 16 Emissions: Cases 6 & 7

Case	Ill. #6	Tex. Lig.	Ill. #6 + 10% bio	Tex. Lig. + 10% bio
NO_x (tons/year)	223.1	234.7	225.1	232.5
SO_x (tons/year)	7,396	2,157.5	7,031	1,869.6
Gross CO_2 (tons/year)	1,758,585	2,110,246	1,821,850	2,045,916
Eff. CO_2 (tons/year)	**1,758,585**	**2,110,246**	**1,667,081**	**1,824,817**
Gross CO_2 (tons/MW-year)	7,193.9	8,942.0	7,509.0	8,619.6
Eff. CO_2 (tons/MW-year)	**7,193.9**	**8,942.0**	**6,871.1**	**7,688.1**

Table 16 shows the emissions data for this set of plants. Since less coal needs to be burned in the Illinois #6 cases, the amount of NO_x and CO_2 is lowered. However, due to the higher sulfur content in Illinois #6, the SO_x emissions cannot do anything but increase. Aside from this, Illinois #6 is superior emissions-wise to lignite, producing less CO_2 overall (about 350,000 tons/year, or 27%) and less per MW as well (about 1,000 tons/MW-year, or 20%).

4. Conclusions

In summary, this study was performed using GTPro®, a program from the Thermoflow® software suite. The baseline plant was implemented with 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 conclusions are as follows:

- RSCs and CSCs are much more efficient than quench systems (6 points), but are also more expensive ($540/kW). The emissions are comparable.
- Dry-fed systems are universally more efficient than slurry-fed systems, given the same type of CCS (2-3 points). However, sour-shift CCS remains superior in both cases, being more efficient (3 points) and less costly (1.5 cents/kW-hr).
- In this study, the air-blown design is more efficient than the oxygen-blown one (2.6 points). It requires less capital cost (by $300/kW) and produces a cheaper CoE (by 0.6 cents/kW-hr) and fewer carbon emissions (500 tons/MW-yr).
- The duct burner is clearly better off with importing natural gas, at least in terms of 2011 USD. Using NG is actually even more attractive than what this study suggests because the price of natural gas has significantly decreased since this study was completed. The efficiency drops mainly due to the higher auxiliary losses from raising the gasifier mass flow rate (so that the GT fuel mass flow can be maintained along with the output power.)
- Illinois #6 is more efficient to use (2-3 points), as it has a higher heating value (3000 Btu/lb higher LHV) and produces better syngas (800 Btu/lb higher LHV) than Texas Lignite. However, it is also more expensive. When blended with biomass, the pure lignite case improves in efficiency, but the use of Illinois #6 reverses this trend: instead decreasing in efficiency by 0.67 points and raising the CoE by 0.1 cents/kW, although the Capital costs still decrease by about $160/kW.

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