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Analysis of Biomass/Coal Co-Gasification for Integrated Gasification Combined Cycle (IGCC) Systems with Carbon Capture

Henry A. Long III
University of New Orleans, hlong@uno.edu

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NOMENCLATURE

ASU	Air Separation Unit
AGR	Acid Gas Removal
CCS	Carbon Capture and Sequestration
DA	De-aerator
GHG	Greenhouse Gas(es)
IGCC	Integrated Gasification Combined Cycle
HP	High Pressure (referring to a steam condition, PSI)
IP	Intermediate Pressure (same as above, PSI)
BMR	Biomass Ratio (wt%)
GT	Gas Turbine
ST	Steam Turbine
HRSG	Heat Recovery Steam Generator
MW	Mega-Watt
M.W.	Molecular Weight (lbs/lb-mol)
HHV	Higher Heating Value (Btu/lb)
LHV	Lower Heating Value (Btu/lb)
O&M	Overhead and Maintenance Costs ($)

Subscripts

gross	efficiency without including parasitic power/total work done
net	efficiency with parasitic power included/gross work minus work input
th	thermal
e	electrical

ABSTRACT

In recent years, Integrated Gasification Combined Cycle Technology (IGCC) has become more common in clean coal power operations with carbon capture and sequestration (CCS). Great efforts have been spent on investigating ways to improve the efficiency, reduce costs, and further reduce greenhouse gas emissions. This study focuses on investigating two approaches to achieve these goals. First, replace the subcritical Rankine steam cycle with a supercritical steam cycle. Second, add different amounts of biomass as feedstock to reduce emissions. Finally, implement several types of CCS, including sweet- and sour-shift pre-combustion and post-combustion.

Using the software, Thermoflow®, this study shows that utilizing biomass with coal up to 50% (wt.) can improve the efficiency, and reduce emissions: even making the plant carbon-negative when CCS is used. CCS is best administered pre-combustion using sour-shift, and supercritical steam cycles are thermally and economically better than subcritical cycles. Both capital and electricity costs have been presented.

Keywords: Biomass, Gasification, Emissions, IGCC, CCS, Cost of Electricity, Carbon Capture

CHAPTER ONE

INTRODUCTION

1.1 Background

1.1.1 Biomass

Biomass is any material that is derived directly from living or previously living things. Examples include wood products, animal and plant wastes, and compost. The use of biomass as an alternative energy source is not a new idea: its use as a potential power source can be traced back to the first wood-fueled fire. Among all of the applications available to biomass, the most direct and obvious uses are in fireplaces, grill pits, and wood-burning stoves and ovens. Another application was in use as early as World War II, when military personnel would retrofit vehicles with wood-gasifying engines to mitigate fossil fuel dependency (Turare, 2002). In recent times, biomass has been used as a co-reactant, or even the main fuel source in both combustion-based power plants and in gasification plants as well. As shown in Fig. 1.1, the consumption of biomass in the United States is fairly large compared to those of other non-hydropower renewable resources, and is projected to experience the most significant growths in the coming years. The fact that biomass is one of the most often used types of renewable energy sources is attributed to its versatility and the range of the technology available for this energy source. In addition, biomass is typically free to use, and is universally available in most all regions, as opposed to fossil fuels, which can only be found in particular locations around the globe.

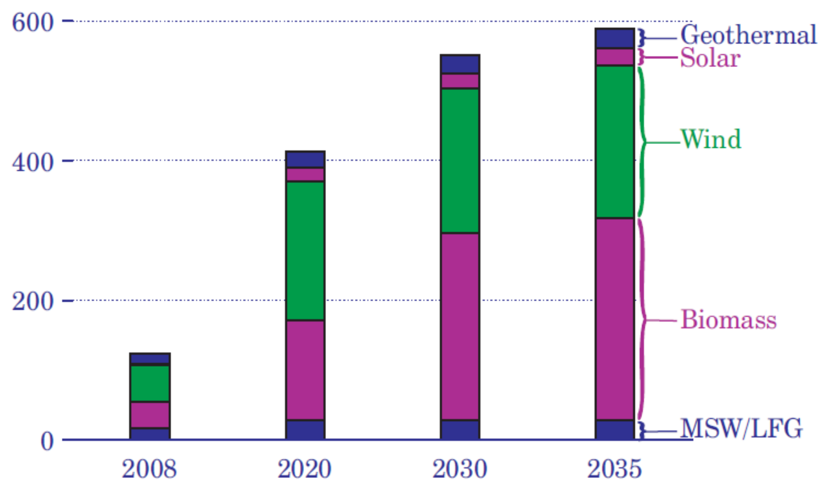

Figure 1.1 Non-hydropower renewable energy consumption and projected growth (EIA, 2010)

1.1.2 Biomass in Gasification

As a fuel, biomass contains no sulfur and very little ash, meaning it is relatively clean, especially compared to fossil fuels. Biomass also contains high concentrations of alkali metals, oxygen-rich compounds, and moisture. Compared to fossil fuels, biomass has very high volatile content, due to the presence of complex sugars, oxy-acids, proteins, and lipids within the organic sub-structures. However, most biomass has lower fixed carbon content than typical fossil fuels, like coal, so it is not as readily combustible. In particular, wood-based biomass is completely incombustible at room temperature (Reed et. al, 2005.) More importantly (plant-based) biomass is carbon-neutral. This means that, when biomass is used as a fuel, the CO_2 that it releases into the atmosphere is equal to the amount of CO_2 that has been removed from the atmosphere during its lifetime through natural processes like photosynthesis. This makes biomass environmentally friendly for use in producing energy. Table 1.1 shows typical biomass feedstocks with their average higher heating values (HHV).

Table 1.1 Typical biomass sources (Reed and Gaur, 2001)

Type	Higher Heating Value (Btu/lb)
Sewage Sludge	8217
Septage	8217
Fruit Pulp	3600
Wood Waste	8733
Mixed Solid Waste	4830

Gasification is a process that can transform a typical hydrocarbon material into synthesis gas (syngas for short) following *pyrolysis* (to be discussed later), which can then be used as a fuel to produce power, such as in a boiler or gas turbine combustor. Gasification is different from combustion: while combustion uses oxygen to “burn” a carbon-rich fuel to release heat energy via full oxidization, gasification only involves partial oxidation and actually absorbs heat from the surroundings. In other words, the goal of gasification isn’t to release the energy inside the feedstock, rather the purpose of gasification is to produce various alternative fuels or chemicals. Through this conversion, the idea is to allow the fuels to keep as much of their original energy content as possible, and the new fuels/chemicals can then be used in much “cleaner” ways than through combustion alone. Table 1.2 summarizes the key differences between combustion and gasification.

Table 1.2 Comparisons between Combustion and Gasification

Combustion	**Gasification**
✓ Requires the use of oxygen ✓ Exothermic process ✓ Produces high temperature gases for heating or power generation. ✓ Produces large amounts of pollutants.	✓ Performed with little oxygen ✓ Endothermic process ✓ Produces an alternative fuel (syngas) or chemicals. ✓ Products can be cleaned to remove sulfur and/or CO_2.

Gasification has a number of advantages over combustion, including:

1.) After combustion occurs, cleaning out the harmful pollutants, such as SO_x and NO_x, requires a large amount of energy and space because nitrogen is in the exhausted gases, whereas in gasification, the contaminants can be removed beforehand, which conserves a large portion of this energy. This is due to the lower amount of mass that needs to be cleaned. For reference, the mass of syngas is about 5-15% of that of the exhaust gases.

2.) By-products of gasification (such as COS, H_2S, HCN, CH_4, H_2SO_4, and slag) can be exported and sold for profit. Combustion by-products (such as H_2O, NO_x, and SO_x) cannot, as they have no market value.3.) Gasification can also be used to produce many chemicals such as methanol, ethanol, hydrogen, ammonia, urea, fertilizers, etc. See Fig. 1.2 for some of the many options gasification offers.

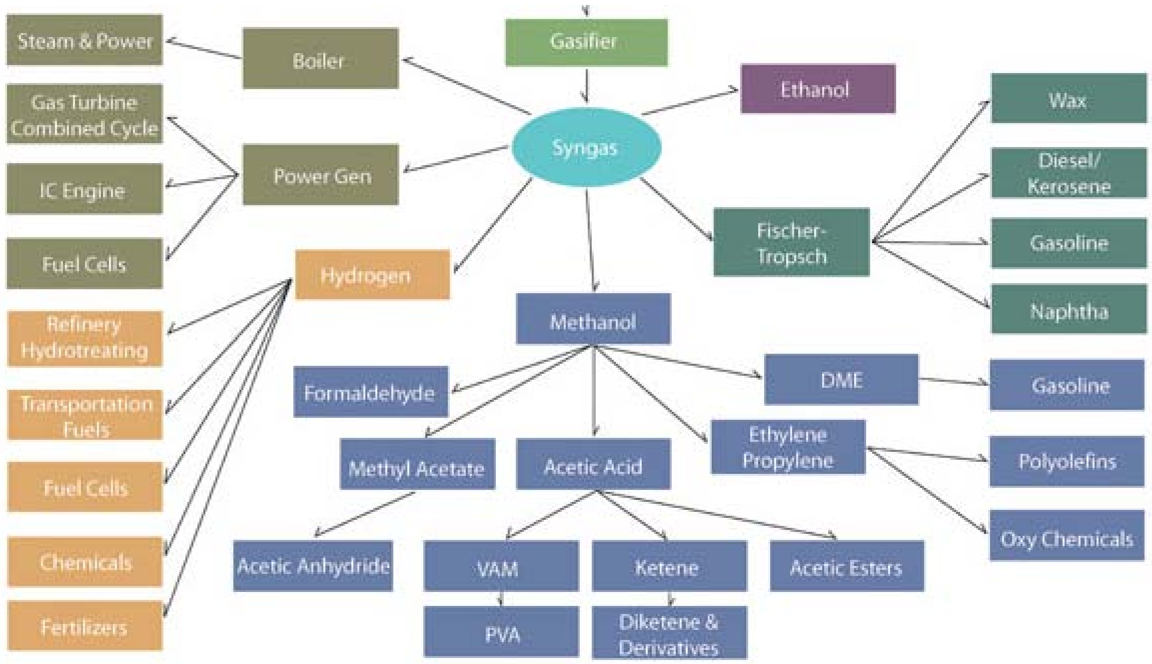

Figure 1.2 The many products that can be derived from coal gasification (EnviRes, 2010)

Because gasification is a cleaner process than the most conventional methods of power production, it makes sense to augment this cleanliness through the use of an even cleaner fuel, like biomass. However, biomass itself, due to its high volatile content, is more prone to producing *tar* than other fuels. In gasification systems, this can present a problem, as tar is very sticky and corrosive, and can easily damage heat exchangers and other internal gasifier components. This is usually solved in one of three ways: (a) prevent tar from forming by modifying the internal gasifier conditions or setup (such as using a fluidized bed), (b) thermally crack the tar into less dense products by raising the gasifier's operating temperature or burn the tar up as it is formed (like in entrained flow gasification), or (c) condense and remove the tar before it becomes a part of the syngas. Biomass also contains many other corrosive compounds, like alkali metal-oxides in the ash, and also has a tendency to produce *ammonia* (NH_3), which is both a highly corrosive compound and a deadly poison if released into the atmosphere, and must be cleaned out or burned before the syngas produced can be used.

On the other hand, an advantage of using biomass in gasification is the fact that biomass can be gasified at atmospheric pressure, unlike other fuels, such as coal, which require higher pressures to undergo efficient devolatilization. This means that the gasifier itself will be easier to operate and cheaper to construct if biomass is used. If any fossil fuels are present in the feedstock at all, however, this advantage is quickly lost, as the increased pressure becomes necessary. In addition, due to its low ash-content, biomass does not produce very much slag, which is usually a problem for other fuel types, especially coal.

1.1.3 IGCC – Integrated Gasification Combined Cycle, a Brief Overview

Using IGCC technology results in lower emissions and more energy efficiency than a standard pulverized coal (PC) plant (U.S. Dept. of Energy, 2009). In addition, because it uses *gasification*, IGCC allows for the implementation of pre-combustion carbon capture and storage (CCS), which is typically much cheaper to implement than the typical post-CCS system used in PC plants. It is capable of producing electrical power with a total output efficiency of near 50% (Jenkins, 2008). The basic outline of IGCC (Fig. 1.3) is as follows:

1.) Raw feedstock enters the gasifier and undergoes gasification.

2.) Syngas is extracted and particulates are removed.

3.) The syngas is cooled so it can be "cleaned." (Syngas can theoretically be cleaned at higher temperatures, but that technology is still under development.)

4.) The syngas is cleaned in a series of devices that remove particulates, COS, H_2S, SO_x, NO_x, and halides from the mixture.

5.) The gas is then burned in a combustor and run through a gas turbine.

6.) The turbine exhaust is then run through a heat recovery steam generator (HRSG), where the waste heat is used to generate steam.

7.) The steam is run through a steam turbine, where additional electrical power is generated from the recovered waste heat.

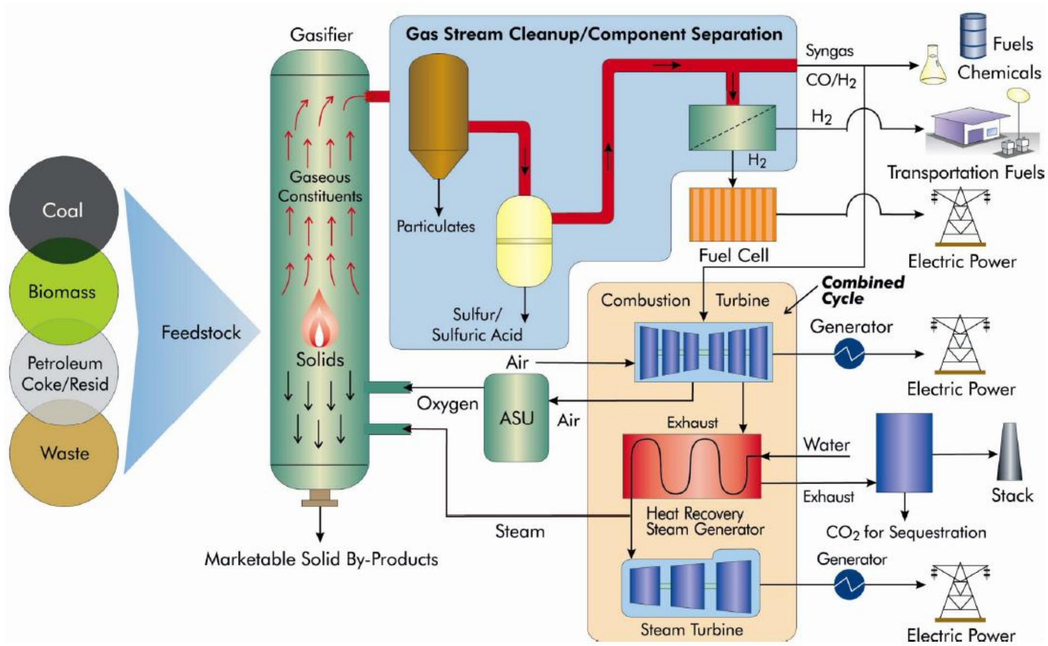

Figure 1.3 Typical IGCC plant (U.S. Department of Energy, 2011)

There are several successful commercially functioning IGCC plants in the world such as the Wabash River Station in West Terre Haute, Indiana, the Polk County Power Station in Tampa, Florida, the Buggenum plant in the Netherlands, and the ELCOGAS plant in Puertollano, Spain (Jenkins, 2008).

1.1.4 The Rankine Cycle – Sub-critical vs. Super-critical

The most common way of using the energy released during combustion is to convert water into superheated steam. This way the water can be run through a steam turbine to generate power, and return the water to its original state through a condenser and a pump, returning the working fluid to the boiler where heat from combustion is provided. This system is called the Rankine Cycle, after William Rankine, a Scottish physicist and engineer. Most of the electrical power in the world is generated using the Rankine Cycle, including solar, nuclear, and pulverized coal plants. A typical Rankine Cycle is shown in Fig. 1.4. In an IGCC system, the Rankine cycle is operated as the bottom cycle and the traditional boiler is replaced by an HRSG.

Of all of the potential improvements made to the Rankine Cycle in more modern times, raising the inlet temperature and pressure of the steam turbine in the traditional cycle is the most direct way to increase its operating efficiency and its total power output. Power generation specialists and engineers have been highly focused on this area of potential steam cycle improvement since the 1950's. 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 (Retzlaff, 1996). This was the onset of the first supercritical steam generation plants.

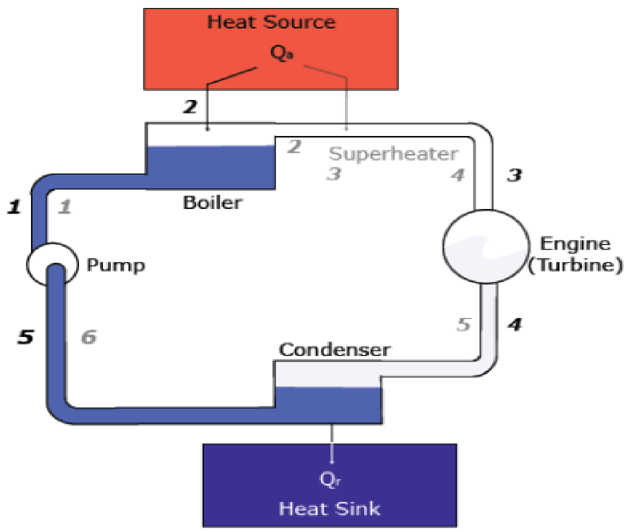

Figure 1.4 Typical Rankine Cycle (Hough, 2009)

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 (Voss and Gould, 2001). For reference, the typical efficiency of a standard sub-critical Rankine cycle is around 30-38%, while a supercritical cycle under the same environmental conditions can achieve efficiencies of 42-45%, and possibly near 47-50% for an ultra supercritical system (Hough, 2009).

So far, all of the research and industrial efforts going into supercritical cycle design are meant for standard PC plants, and no supercritical Rankine cycle has been operated in an IGCC system. However, due to rising materials costs and heavy system maintenance, overall unit availability is lacking for supercritical systems. Indeed, the vast majority of steam systems in the U.S. are sub-critical, mostly due to the rise in popularity of nuclear power systems in the early 1970s, and the continuously growing public disapproval of coal power in general (Voss and Gould, 2001).

1.2 Literature Review

1.2.1 Gasification

1.2.1.1 History of Gasification

The actual gasification process has been in use since at least 1792, when a Scottish engineer named William Murdoch used the gases from coal pyrolysis in lanterns and street lights. These gases became known as “town gas” (Also called “producer gas” - A more volatile-heavy mixture of gases produced at lower temperatures than typical syngas.) which replaced old candles and lanterns. This new technology made night shifts in factories possible, and helped pave the way for the Industrial Revolution. In Europe, this culminated in 1813, when the London and Westminster Gas Light and Coke Company used several tonnes of this gas to light up the Westminster Bridge on New Year’s Eve. In the U.S., town gas made its first appearance when the city of Baltimore, Maryland became the first city to use the gases to light the streetlamps (National Renewable Energy Laboratory, 2011). Eventually, many more plants were made to produce town gas, and, during World War II, vehicles were modified with wood-gasifying engines during a shortage of fossil fuels in order to reduce consumption. Over time, these gases started to fade from the public eye, and were eventually replaced with much cleaner and cheaper natural gas. Around 1960, syngas, as it was now called, became a topic of renewed interest overseas, and gasification first became an option for power generation in the mid 1980s.

1.2.1.2 Gasification Chemistry

The actual gasification process begins with *devolatilization* and *pyrolysis*, where a small part of the carbon-based feedstock is burned to provide heat that is needed to drive out moisture and volatiles, but in the absence or poor presence of oxygen. Figure 1.5 shows an overall flowchart of this process and the other major stages of gasification. After pyrolysis, more heat is needed to thermally crack the volatiles to break the long hydro-carbon chains into lighter gases as well as to gasify the remaining carbon left in the feedstock into syngas. The chemical makeup of syngas tends to consist predominantly of CO and H_2 with small amounts of CH_4 as fuel and CO_2, N_2, and water vapor as non-combustible gases. The syngas also contains other compounds like H_2S, COS, HCN, HCl, Hg, and other contaminants that will need to be removed before utilizing the syngas for power generation (Rezaiyan et. al, 2005).

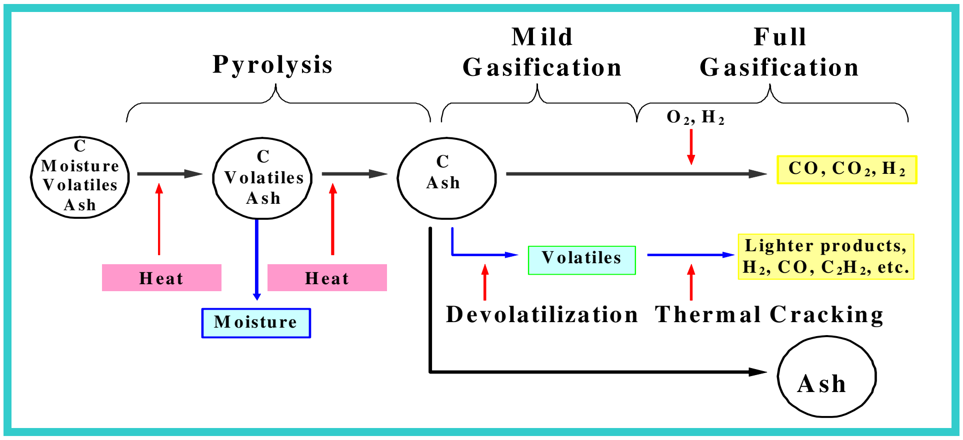

Figure 1.5 Basic pyrolysis and gasification flowchart

After the volatiles leave the feedstock, what is left in the gasifier is *char*, which is basically pure carbon. Several of the gases produced by the previous reactions react with the char, and produce more gas. These reactions are called Heterogeneous Phase Reactions, because they have a non-uniform reactant phase distribution: some reactants are solid and some are gaseous. All of this occurs within a single device, rightly called a *gasifier*. The most common fuel used in gasification is coal, but virtually any hydro-carbon-based substance can be gasified.

<u>Heterogeneous Reactions</u>

$\mathbf{2C + O_2 \rightarrow 2CO}$	(R1.1 – Incomplete Combustion)
$\mathbf{2CO + O_2 \rightarrow 2CO_2}$	(R1.2 – Complete Combustion)
$\mathbf{C + H_2O \rightarrow CO + H_2}$	(R1.3 – Steam *Gasification*)
$\mathbf{C + CO_2 \rightarrow 2CO}$	(R1.4 – CO_2 *Gasification*/the Boudouard Reaction)
$\mathbf{C + 2H_2 \rightarrow CH_4}$	(R1.5 – Direct Methanation)

Notice that reactions 1.1 and 1.2 represent combustion. This is because gasification is *endothermic,* and requires some heat energy to be provided in order for it to occur. In most gasifiers, this is accomplished using reactions 1.1 and 1.2, but it could just as easily be performed using some other side-reaction or process. It just so happens that using small, controlled amounts of combustion to initiate the process is the most available, controllable, and efficient method in most applications. Using air for these reactions is an acceptable substitute for pure oxygen, but it introduces extra nitrogen to the process, which usually results in more NO_x emissions, a lower heating value for the syngas, and requires larger pipes and a larger clean-up system. Pure oxygen is preferred in this sense, but using oxygen requires the use of an air-separation unit (ASU), like a distillation tower, which comes with additional energy costs, and a subsequent loss in net power output and efficiency. Which method to use is dependent upon the rest of the plant setup.

Reactions 1.3 and 1.4 are the actual *gasification* reactions, which aim to turn the raw carbon, or char, into syngas through exposure to the combustion gases or water vapor, and that the main fuel products are CO and H_2. CO and H_2 are the primary components of a good syngas mixture, as both are readily combustible fuels in their own right. For most applications, a good gasifier design will allow for maximum production of both of these gases. In most syngas applications, this will be accompanied by both water vapor and methane.

Reaction 1.5 is somewhat rarer in this family of processes, and usually CH_4 makes up a very minute portion of the final syngas mixture. Methane is usually not a major interest in most gasifiers, because its main use is as a substitute natural gas (SNG). It is possible for a gasifier to be designed specifically to produce SNGs, but producing syngas is typically cheaper and easier. In addition, despite its high heating value, most syngas-producing gasifiers will deliver very small amounts of CH_4: not nearly enough for its presence to make a difference. From here, among other reactions between existing volatiles, an equilibrium reaction (called the Water-gas Shift Reaction, R1.6) is established, along with two other methane producing reactions, all of which are called *Homogenous Phase Reactions,* named so because all reactants involved are gases. In other words, the phase distribution is homogeneous.

<u>Homogeneous Reactions</u>

$\mathbf{CO + H_2O \leftrightarrow CO_2 + H_2}$ (R1.6 – The Water-Gas Shift Reaction)

$\mathbf{CO + 3H_2 \rightarrow CH_4 + H_2O}$ (R1.7 – Homogeneous Methanation 1)

$\mathbf{2CO + 2H_2O \rightarrow CO_2 + CH_4}$ (R1.8 – Homogeneous Methanation 2)

As stated previously, methanation isn't of supreme importance in most gasification applications, so reactions 1.7 and 1.8 usually play a very minor role at this point. That being said, a few other reactions do occur, in which some carbon monoxide is converted further to carbon dioxide, and some hydrogen gets converted back into water. It is at this point that the concentrations of char, carbon monoxide, carbon dioxide, water, methane, and hydrogen (the main components in syngas) achieve a degree of equilibrium. At this point, reaction 1.6, the so-called "Water Shift" or "Water-Gas Shift" reaction is the most important chemical reaction for the entire process. This reaction is what defines the syngas's equilibrium state, and any number of external stimuli can drive this reaction in either direction.

On a strict energy-basis, CO is a more valuable fuel than H_2, and is much more usable as a fuel. CO_2 is not useful for generating power, and is generally considered to have no heating value at all. However, CO_2 is being considered as a diluent to reduce the flame temperature (and therefore reduced NO_x) and increase the power output via increased mass flow rate. Water on the other hand, while it does not react with anything during combustion, its heat of vaporization can be recovered and utilized during the process itself. This is the origin of the HHV vs. LHV dichotomy. As such, the *left* side of reaction 1.6 is generally more preferred for most IGCC plants. However, burning CO will create CO_2 during combustion anyway, and this of course increases the plant's carbon footprint. If CCS is implemented in the plant, *especially* if it is pre-combustion CCS, then the *right* side of the equation begins to carry more weight. This is because (a) the compound CO cannot be captured as is, so it must be converted to CO_2 using this reaction anyway, and (b) after capture, there will be excess amounts of hydrogen available for use in the GT, which will be greatly preferred over just water. In order to use CCS (discussed in greater detail later on,) industry often uses the term "CO-shift." This is referring directly to reaction 1.6, where CO is being "shifted" over to CO_2, where it may be captured and sequestered using whatever method is available.

After leaving the gasifier, the syngases are separated from the contaminants using *Gas Cleanup* Technologies. Processes like cyclone filters, misting technologies, and "scrubbers" are examples of such technologies. These Gas Cleanup systems "clean" the useable gases (namely CO and H_2) of their impurities, like COS, H_2S, HCN, and so on. The advantage of gasification is that many of these other impurities or contaminants can be removed before combustion, so they will not be released to the atmosphere through the exhaust. In addition, some of these contaminants, when separated, can be used in other applications or sold for profit as such. For example: H_2S is used to denature proteins and has use in other such chemical applications, COS is a primary ingredient in weed killers, elemental sulfur and H_2SO_4 are valuable byproducts which can be sold on the market, and, lastly, slag can be used as an ingredient in many types of concrete.

1.2.2 Coal vs. Biomass

With oil being the first, coal is the second most used energy resource in the world, and the most often used source of electrical power in the United States. As a fuel, coal enjoys one of the highest energy densities among fossil fuels, and is also one of the most readily available and easily obtainable. In addition, it is highly adaptable, and can be used in virtually any application where large amounts of energy are needed.

In the U.S., coal is a very important energy source, and accounts for over 40% of all energy generation in the country. Nearly 25% of the world's coal reserves are in the mainland United States, and global coal usage is projected to grow by at least 48% by 2030. Table 1.3 shows a few typical types of coal as well as their ultimate analyses. The various species of coal are widespread and varied, so the types listed in Table 1.3 should not be taken as representative of any particular type.

Biomass, on the other hand, is relatively unused as a fuel in the U.S., due to its lower energy density and unique challenges that must be overcome in order for it to be used as a fuel. However, with recent concerns about the environment and hazardous gas emissions from power plants and other energy applications, the use of bio-fuels is quickly becoming a hot topic in both the scientific and political arenas, and bio-fuels are seeing more and more usage in the mainstream, particularly ethanol and bio-diesel.

Table 1.3 Typical Ultimate Analyses for Various Coal Ranks (Rezaiyan et. al, 2005)

Rank	Higher Heating Value (Btu/lb)	Oxygen (wt%)	Hydrogen (wt%)	Carbon (wt%)	Nitrogen (wt%)	Sulfur (wt%)	Ash (wt%)
Anthracite	12700	5.0	2.9	80.0	0.9	0.7	10.5
Semi-Anthracite	13600	5.0	3.9	80.4	1.1	1.1	8.5
Low-volatile Bit.	14350	5.0	4.7	81.7	9.4	1.2	6.0
High-volatile Bit. A	13800	9.3	5.3	75.9	1.5	1.5	6.5
High-volatile Bit. B	12500	13.8	5.5	67.8	1.4	3.0	8.5
High-volatile Bit. C	11000	20.6	5.8	59.6	1.1	3.5	9.4
Sub-bituminous B	9000	29.5	6.2	53.5	1.1	1.0	9.8
Lignite A	6900	44.0	6.9	40.1	0.7	1.0	7.3

From Table 1.3, notice that there are several main categories of coal. The most commonly known type is black coal, more scientifically referred to as bituminous coal (Known as such because it contains the chemical compound, bitumen). Bituminous coals are the most common and complex variety of all coals. There are many different sub-types, all with various elemental contents. Just below bituminous and the loosely defined Sub-bituminous coals is lignite. Lignite, also called "brown coal," has less energy-producing capability than the other coal ranks, but it is, in turn, easier to gasify due to its high volatile content and reactivity. However, the presence and arrangement of these volatiles make lignite especially prone to *spontaneous combustion*, making its transport and handling dangerous. Lignite is also very moisture-rich, which makes it valuable in IGCC applications. The highest rank of coal is called Anthracite, sometimes called "black diamond." While anthracite possesses the largest energy-producing potential of all coals, it is expensive, and is reserved mainly for smaller-scale applications. Large power plants prefer to use bituminous coal or lignite (Rezaiyan et. al, 2005).

For their uses in gasification, coal and biomass are essentially chemical opposites. Where biomass has significant amounts of hydrogen, oxygen, and nitrous compounds, coal typically has higher carbon content as well as sulfur and ash. While biomass has a high concentration of volatiles, coals will typically have a higher concentration of fixed carbon, especially the higher ranked coals.

Unlike biomass, coal doesn't have nearly as great a risk of producing ammonia, and on the whole is less corrosive, making it much more suited for providing power. However, coal is still not as clean, and its high sulfur content leads to the creation of large amounts of sulfur-oxides (SO_x), which are directly responsible for the *acid rain* phenomenon. Coal also produces large amounts of ash, which, in most commercial gasification processes (where the average gasifier temperature is higher than the ash's fusion temperature,) it will melt down into *slag* inside the gasifier. Slag is essentially a collection of ore impurities that have coalesced into a single, nearly homogeneous mixture. Slag itself is not a problem for most gasification systems, but for certain types of gasifiers, slag production must be avoided since it can tend to clog up or damage certain parts of the gasifier (especially fluidized bed gasifiers, discussed later.)

Biomass on the other hand does not contain any sulfur, usually, and has much lower ash content, meaning there will be less slag produced when it is gasified. However, biomass has a different problem: it tends to produce large amounts of tar. As discussed earlier, there are solutions to this issue, and most gasifiers can more readily deal with tar production than excess slag. As discussed in the next few sections in more detail, most types of biomass have several feeding issues, unlike coal, which has a very high grindability index. Most biomass is very fibrous and tough, and cannot be broken down into a workable feedstock easily. Whereas coal can virtually be used as received, most of the time (with some minor amount of grinding, that is,) biomass, on the other hand, requires an extensive amount of pretreatment before it can effectively be used as a fuel (Richards, 2008).

1.2.2 Biomass

1.2.2.1 Issues with Biomass

While biomass is certainly a much cleaner source of energy than fossil fuels, and is certainly more abundant and cheaper to obtain than most other fuels, biomass has its own share of problems that make using it as a fuel somewhat challenging. In summary of the issues already mentioned: (1) biomass has low energy density, especially compared to fossil fuels, like coal. (2) Biomass has low mass density, meaning the same amount of biomass takes up much more space than the equivalent mass of other fuels. Combining these two issues means that the volumetric flow rates required to run certain larger plants are difficult to achieve using biomass in its raw form alone. For reference, a typical coal density is around 1600 kg/m^3. The density of raw oak wood is between 0.6 and 0.9 kg/m^3 depending on the specimen, meaning that for a coal plant requiring 1600 kg in an hour to operate, if the plant switched to oak wood biomass, those same 1600 kilograms would now take up between **1800 and 2700 cubic meters**! This fact alone makes the actual *feeding* part of the process very difficult. (3) Gasifying biomass releases large amounts of highly corrosive materials, most of which result from sources like the natural acids within most living cells, alkali metals that are found in the ashes, and natural lipids and enzymes, mostly in animal biomass sources. These substances can quickly and easily shorten the average lifespan of most plants due to the damage done to the system's internal parts, especially the gasifier, the piping system, and the GT combustor. Even worse is the fact that biomass's chemical makeup makes it relatively easy to produce *ammonia,* which is both damaging to the system at large and is also a deadly toxin. Finally, (4) biomass's physical structure is highly elastic and fibrous, unlike coal, which tends to be very brittle. This means that, for the purpose of getting the right particle size distribution (PSD), there aren't many physical processes that can be done to achieve this with biomass. Coal can easily be ground down into a fine powder, making feeding very simple. Biomass, however, typically contains lots of tiny fibers that simply cannot be "ground" down or torn easily. Even with extensive chopping and drying, this procedure is still very difficult and the biomass feedstock tends to get stuck in various types of machinery.

All in all, these issues combined make feeding biomass into any system a challenge. Besides feeding issues, there are more to consider that more directly affect plant viability, not just operation.

First and foremost, the greatest issue with biomass is *availability.* The supply of most biomass is seasonal and is limited by quantity. Secondly, 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 useless as a fuel in the process. This means that, for any given power plant, the biomass feedstock will only be available for a small window of time during that species of biomass's harvest season. (This is especially true for plant matter.) When coupled with biomass's low mass density, these issues has prohibited profitable operation of larger, pure biomass plants, meaning that effectively utilizing biomass alone in any plant bigger than about 50-80MW is uneconomical at best.

1.2.2.2 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 purpose of pretreatment is to improve the biomass feedstock in a way that will make it more suitable to be used in a wider array of power applications, especially so it may be used in larger plant designs. In particular, a good measure of pretreatment will be one that (a) increases the energy density of the fuel, so lower flow rates can be maintained and still allow for the same amount of net heat input (or produce more heat input for the same flow rate, depending on design specification, (b) reduce the acidity of the feedstock to extend the plant's life and improve performance, (c) raise the fuel's *mass* density, so that achieving higher mass flow rates possible for a given system, without requiring the higher, previously unachievable, *volumetric* flow rates, and, finally, (d) remove some of the harmful pollutants from the core structure, such as those that may produce *ammonia*.

One form of pretreatment for gasification purposes is *flash pyrolysis.* As has already been discussed, pyrolysis is the first phase of the overall gasification process. Flash pyrolysis is performed before the biomass enters the gasifier so that it is converted into both char and a substance that can easily be poured into the gasifier bed like typical coal-slurry or oil feedstock. This substance is usually called "bio-oil," and also has its uses outside of gasification, such as in the manufacturing of elemental Hydrogen (Hanssen, 2007). This is very beneficial for biomass, as liquids are usable feedstocks in just about every type of gasifier (except, perhaps, for down-draft gasifiers, depending on moisture content). In Fig. 1.6, the left picture is a sample of bio-oil.

Flash pyrolysis is referred to as such because it occurs at a very fast rate. Typically, the biomass is able to reach temperatures of about 1200°F (or 500°C) in less than 2 seconds. In addition to this, there is another, higher temperature form of flash pyrolysis that mainly results in a gaseous end-product, rather than a liquid one. This readily produces a highly reactive syngas that is about 80% carbon monoxide and hydrogen by weight. Many other forms of pyrolysis at various max temperatures, heating rates, and miscellaneous conditions are used, each with markedly different resulting compound compositions and heating values.

Another type of pretreatment, and the one that is taken into consideration for this study, is *torrefaction.* Torrefaction is a thermal process, wherein raw biomass is heated to about 500°F (200-300^{o}C) and essentially "cooked" for 5-10 minutes, 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 (Bergman and Kiel, 2005). During torrefaction, the biomass looses roughly 30% of its mass as torrefaction gases, and roughly 10% of its internal energy with them (Bergman, 2005). A simple algebraic calculation shows that this would result in about a 28% increase in the calorific value per unit mass for the feedstock (Bridgman et. al, 2007.) In Fig. 1.6, the right image is a sample of torrefied biomass, which was made from wood chips.

Figure 1.6 Bio-oil (left), torrefied biomass (right)

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 (Siemens-Westinghouse Corp., 1999). 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 (Bergman, 2005).

1.2.2.3 Co-Gasification of Biomass with Coal

Another solution to some of the problems associated with biomass is *co-feeding* biomass alongside coal or another fossil fuel in a larger plant for either combustion or gasification applications. *Co-gasification* is the main area of focus for this study, as not many plants in the U.S. make use of this technique, even though much research has been performed in the past 10 years in attempting to make it more mainstream. Using biomass alongside a bigger, more energy dense fuel will (a) allow for biomass to utilize the same economy of scale that coal and oil enjoy, (b) reduce the emissions of CO_2, SO_x, and NO_x into the atmosphere, (c) provide an easy and less costly way to use biomass in gasification by retrofitting an existing coal or oil power plant for use with biomass, (d) eliminate the seasonality factor involved with using biomass, meaning since the plant uses fossil fuels (available year-round), the plant can remain in operation even if there is no biomass fuel to use, as most biomass sources are seasonally dependent and, lastly, (e) because there is coal mixed in with the biomass, corrosion is less of an issue than it is with plants that use purely biomass.

1.2.3 Biomass in IGCC

1.2.3.1 BIGCC – Biomass Integrated Gasification Combined Cycles

The first pure biomass IGCC plant was constructed in Värnamo, Sweden, in 1993 (shown in Fig. 1.7). It was constructed as a demonstration plant, providing roughly 6 MW of net electricity to the grid. This was using a fuel equivalent energy input of approximately 18 MW: yielding a net plant efficiency of ~30%. Unfortunately, this plant was closed down in 2000, as the demonstration had ended, and it was not economically feasible to maintain operation any further. The plant is no longer providing commercial power, as such. However, the site was saved in 2003 by the Växjö Värnamo Biomass Gasification Centre (WBGC), and the current plant is being used as a research site for IGCC related issues, especially those related to biomass (Stahl, et. al, 2004).

Figure 1.7 Varnamo demonstration plant (Stahl, et. al, 2004.)

It was this plant that highlighted some of the key issues and concepts relating to biomass energy. For one, biomass can be gasified easily at atmospheric pressure and at much lower temperatures than fossil fuel feedstocks. Biomass contains virtually no sulfur, so most fuels will produce little to no SO_x emissions. Biomass does, however have large amounts of nitrogen and oxygen, which can lead to NO_x emissions equal to or greater than that of fossil fuel plants. In regards to IGCC, this can also lead to the production of ammonia, as discussed previously.

The most important aspect of BIGCC is that pure biomass is carbon-neutral. In modern times, where there is more carbon-dioxide in the air than any point in history in the past 400,000 years (O'Laughlin, 2010), it is important to find any convenient means to mitigate the amount of carbon released into the atmosphere or, if possible, remove CO_2 from the atmosphere directly using CCS technology, to either prevent or reduce the "carbon debt." Pure biomass plants, with CCS, can be *carbon-negative*, meaning that such plants actually subtract carbon dioxide from the atmosphere during normal operation. It can be demonstrated that with modern technology and proper investment, such plants can be both reasonably viable, and easily implemented, at least in the short-term (Rhodes & Keith, 2005), particularly plants that utilize biomass waste products (O'Laughlin, 2010).

As mentioned previously, the biggest challenge to overcome when using biomass is feeding. For reasons that vary with each type of biomass, the feedstock cannot usually be utilized in the gasification process without some form of pre-preparation. For instance, in 2002, the Tampa Electric Company performed several experiments in which a wood-based eucalyptus biomass feedstock was co-fed into an existing IGCC coal plant in Tampa, Florida, where it was found that biomass feedstock needed to be ground down and processed repeatedly before it could be fed into the system. Despite their efforts, the experimenters discovered three stray wood-chips that lodged themselves into one of the slurry feed-pumps. The process had to be stopped so that the three chips could be safely removed. Despite this, the researchers claim that, if the plant were to seriously adopt biomass as a feedstock, they are confident that a proper biomass feed-system could be constructed to prevent this from happening in the future (McDaniel, et. al, 2002).

Previously, around 1999, the Siemens-Westinghouse Corporation concluded a study of one of their test gasification plants in Hawaii, where sugarcane bagasse and charcoal were being used as feedstocks (shown in Fig. 1.8). The 1995 tests showed that simple drying, chopping, and conveyor-belt feeding were not sufficient solutions to feeding the bagasse into the system. Even recently dried, the bagasse became sticky and started to clog or plug up various components in future processes. After densification was attempted and employed, however, they were able to alleviate the problem somewhat, but the experimenters speculated that the densification system might not be justifiable for a commercial process (Siemens-Westinghouse, 1999).

Figure 1.8 Siemens-Westinghouse Test Plant in Makawao, Maui (1999)

Earlier, the Vermont Gasification Project began in 1994 on the McNeil Power Station in Burlington, Vermont, the goal of which was to produce a large-scale, integrated, Gasifier-Gas Turbine cycle facility using biomass (around $50MW_e$). The plant itself was constructed in 1984. One of the first problems encountered at Burlington was availability and transport: Since the wood used was not very dense, each of the trucks used for transport could only hold 25 tons of fuel. Given the size and scope of the plant, this meant that it took 3 full truckloads to keep the plant running for just one hour. In addition, acquiring wood requires deforesting, which, in Vermont, requires state licensing and approval by no less than four foresters from the State of Vermont before any cutting can be performed. Eventually, the City of Burlington added a recycling facility where citizens could send wastewood and compost instead of sending such material to the nearby landfill. This added significantly more fuel available and increased the plant's capacity. However, further problems developed when, due to political constraints, the owners had to reduce operations, resulting in large amounts of excess fuel (shown in Fig. 1.9). Since biomass cannot be stored for indefinite periods like fossil fuels, of course, the wood supply started to rot, which caused several complaints to be filed by nearby residents (Wiltsee, 2000).

Figure 1.9 McNeil Power Station, Burlington, Vermont

Overall, both the experience at Burlington and the experience in Hawaii have taught valuable lessons about biomass and its potential as a power resource. Ideally, any plant that uses biomass as a fuel should (a) be located suitably close to its fuel source, (b) be supplied in such a way that the feedstock can be used near-immediately upon harvest, (c) have adequate pre-treatment facilities available to ensure safe, continuous operation without damaging the system or forcing an unnecessary shut down for maintenance, (d) have access to enough fuel to maintain constant power output and mass flow within the system itself, and (e) possess the equipment and facilities necessary to replenish the fuel supply at the same rate that it is consumed by the main plant.

1.2.3.2 Co-fed IGCC systems

New gasification technologies are on the rise to further solidify biomass's place in this area, such as the MILENA gasifier from the Energy Research Centre of the Netherlands (Vreugdenhil, 2009), which has recently incorporated lignite fuel alongside its originally pure-biomass design for the production of methane gas.

Co-firing, or co-combustion as it is more commonly called, has similar advantages to that of co-gasification, and is actually a highly preferred method of using biomass in most of Europe, especially in Germany, since biomass co-combustion is always more efficient than burning both fuels by themselves with biomass in a much smaller plant. In addition, a single co-firing plant costs significantly less money per Kilowatt (300 Euros/kW_e vs. 2500-3000 Euros/kW_e) than two individual plants (VGB, 2008).

With regards to co-gasification, however, the bottom line is less than clear: no one can agree on whether or not the efficiency increases or decreases when biomass is incorporated. A few studies have claimed that biomass hindered plant efficiency (Matsuszewski, 2009), others have claimed that biomass offered drastic increases in efficiency using feedstock blends of up to 50% (wt.) biomass (Li et. al., 2008), and still others have noted no clear difference either way except in regards to emissions (McDaniel et. al, 2002). What is certain is that biomass does indeed reduce atmospheric emissions by a sizeable margin, and with the addition of CCS technologies, can even be *carbon-negative*.

Overall, there isn't much information available on the true nature of co-gasification mainly because a large-scale, commercial co-gasification plant has never been constructed anywhere in the world. As of this writing, co-gasification is still in the testing stages, and nearly all of the data available is from software simulations and small-scale experiments. But, that being said, there is hope that a commercial plant, if constructed, may yet be viable for providing commercial power.

1.3 Problem Statement

1.3.1 Objectives

The primary objectives of this study are to improve upon IGCC systems by (1) reducing the GHG emissions of such plants by blending biomass into the coal feedstock, improving the efficiencies, and implementing carbon capture and (2) reducing their capital and operating costs.

1.3.2 Specific Goals and Tasks

In order to meet the objectives, the following tasks were carried out:

1. Design a pure coal IGCC plant to establish a *basis* for comparison.
2. Design a second pure coal baseline plant with a *supercritical* Rankine bottom cycle.
3. Investigate the effects of *blending* various amounts of *biomass* to the coal feedstock.
4. Show how biomass affects similar systems equipped with *carbon capture processes.*
5. Evaluate the differences in performance of *different implementations of CCS* (sweet-shift vs. sour-shift and pre-combustion vs. post-combustion CCS)

6. Compare the performance and cost differences of the following parameters:
    - Radiant vs. quench syngas cooling
    - Slurry-fed vs. dry-fed systems
    - Lignite vs. bituminous coal
    - Oxygen-blown vs. air-blown

CHAPTER TWO

REVIEW OF COMPONENTS OF INTEGRATED GASIFICATION COMBINED CYCLE (IGCC)

2.1 Brief Component Summary

An IGCC system can essentially be broken down into four major parts (called “blocks” or “islands”): fuel preparation, gasification, gas cleanup, and power. First, there’s the fuel preparation island, which pre-processes the feedstock before it enters the gasification island. Since pre-processing has already been discussed at length in Chapter one, the information will not be duplicated here, but some supplementary information about torrefaction will be provided in the corresponding section of Chapter three.

After the preparation island, the gasification island transforms the fuel into syngas, which provides the energy necessary to produce power. The gasification island consists of the gasifier, an air separation unit (ASU) if necessary, and a cooling system. After this is the gas cleanup island, which removes the pollutants and other harmful compounds from the syngas before it is used for power generation. This block contains many devices designed specifically for removing certain substances, depending on the original feedstock, but nearly all cleanup systems contain a particulate scrubber or cyclone filter, a COS hydrolysis chamber, an acid gas removal (AGR) device, and several gas coolers to reduce the syngas temperature to that which is required by each component. Finally, the power block contains all of the equipment necessary to produce electrical power. The gas turbine, steam turbine, Heat Recovery Steam Generator, deaerator, and so forth are all contained within this block. The rest of this chapter is devoted to explaining the specifics of operation of all of these components and their relationships to each other.

2.2 The Gasifier

2.2.1 Types of Gasifiers

In an IGCC system, one of the most important components is the gasifier. Most gasifiers are named after the nature of the fuel’s flow pattern: fixed bed, moving bed, fluidized bed, entrained flow, and transport. They can be further described based on the feedstock feeding direction as: down-draft, up-draft, or cross-flow type, or the feedstock direction relative to the flow direction as co-current or counter-current. It is important to note that there are many different gasifiers out there, and each one is *unique*. However, most gasifiers follow a sort of generic pattern or layout. As such, it becomes important to know the information in the next few sections in order to better understand other literature on specific gasifiers. For reference, Table 2.1 contains a short-hand summary of the information within the next few sections.

Table 2.1 Various Gasifier Types

Type	Fixed bed, Down-draft	Fixed bed, Up-draft	Fluidized Bed	Entrained Flow	Transport
Gas Temperature (°F)	790-1200	790-1200	1700-1900	> 2000	1750-1825
Required grain size (mm)	< 50	< 50	< 6	< 0.15	< 0.4
Feedstock preparation	Drying	None	Grinding	Pulverization	Grinding
General Feedstock requirements	Low moisture content	Low tar content	High ash fusion temperature	None	None
Throughput ("flow rate")	Low	Low	Medium	High	Moderate-High
Ash Condition	Slag/Fly Ash	Slag/Fly Ash	Agglomeration	Slag	Fly Ash
Feed Conditions	Dry	Dry	Dry	Dry or slurry	Dry
Other requirements	None	None	Requires skilled operator	Must use Oxygen	Requires a transport agent

Compiled from: Rezaiyan and Cheremisinoff (2005), Turare (2002), Reed and Gaur (2001), and Loganbach et. al (2001).

2.2.1.1 The Down-Draft Gasifier

This gasifier gets its name from the fact that the air or oxygen combustion agent is injected into the top of the gasifier and flows towards the bottom. In other words, the "draft" goes downward. A typical down-draft gasifier schematic can be seen in Fig. 2.1. Since the typical feedstock is also fed from the top of the gasifier (resulting in both the input air and feedstock flowing in the same direction), this model is also called the “co-current” or “co-flow” gasifier. The internal temperature of a typical down-draft gasifier is between about 800 and 1200°F. Because both streams flow in the same direction, the highest temperatures in the whole process occur during the pyrolysis stage (Reed and Gaur, 2001).

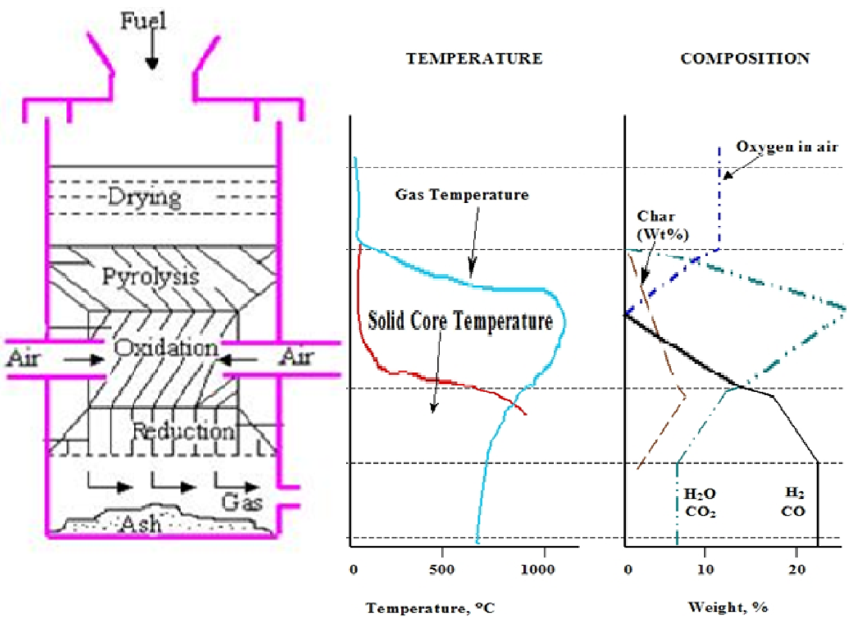

Figure 2.1 The down-draft gasifier (Turare, 2002)

The result is that there is very low tar production compared to other gasifier types, and, as such, there is less syngas cleanup necessary for this type of gasifier, due to the high temperature at the gasifier exit resulting from combustion and thermal cracking. Because of its ability to eliminate tar from the resulting syngas, this type of gasifier has been affixed to many existing combustion engines since early World War II. In fact, it was this very gasifier design that was used on several vehicles throughout Europe, using wood products as feedstock (Turare, 2002). Also due to its fairly low maximum temperature, most ash produced will be fly ash.

A typical problem that occurs in this type of gasifier is that the input feedstock cannot have very high moisture content, so it is not possible to send in a slurry-based feedstock, nor is non-dried biomass a useable fuel source. Another disadvantage is that a decent portion of the char produced during pyrolysis (about 6% or so) is left completely unconverted. And, lastly, unlike its cousin, the up-draft gasifier, the down-draft gasifier expels syngas at fairly high temperatures, which will result in much wasted heat if it is not recovered in some way.

2.2.1.2 The Up-Draft Gasifier

A close relative of the down-draft gasifier, this gasifier type is designed for the gasification agent (oxygen or air) to enter from the *bottom* of the gasifier (i.e. the "draft" blows *up* instead of down.) Since the feedstock is still fed from the top of the up-draft gasifier (as it is for this gasifier's cousin, the down-draft gasifier), this gasifier is also called the "counter-flow" or "counter current" gasifier. Simply changing the flow direction and origin of the input air has drastic effects on gasifier performance. For one, since the input air enters from the bottom of the gasifier, it acts as a cooling agent for the hotter syngases leaving from the same general location. Thus, there is much less wasted heat, which grants this gasifier design an efficiency boost over its cousin. In addition, after gasification and combustion, the leftover hot air, because it blows past the input feedstock entrance, dries the fuel. It is like having a built-in dryer section within the gasifier itself. Because of this "drying" effect, fuels with much higher moisture content can be utilized in up-draft gasifiers, especially raw biomass (Turare, 2002). Figure 2.2 is a diagram of a typical up-draft gasifier.

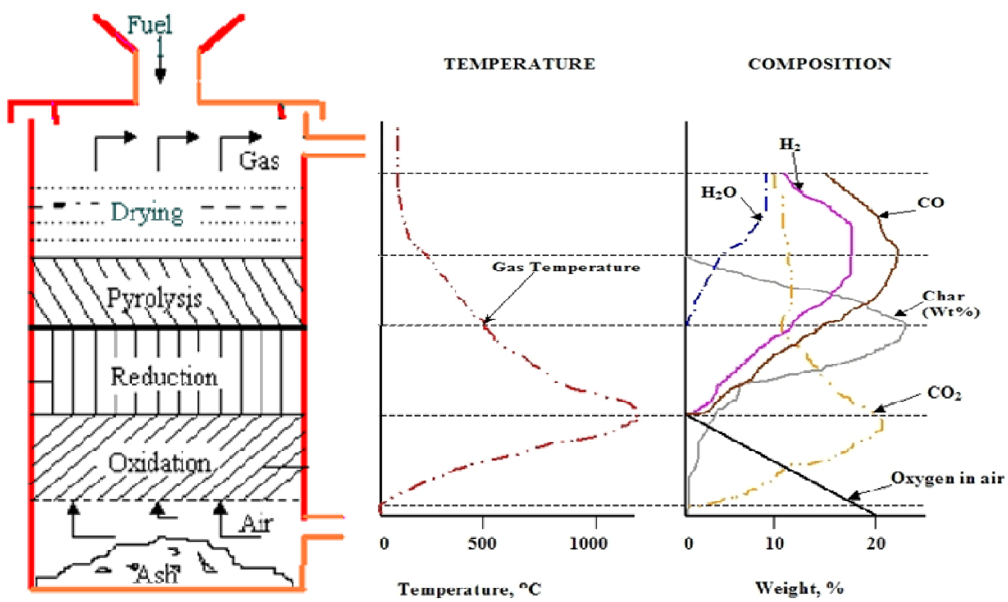

Figure 2.2 The up-draft gasifier (Turare, 2002)

The peak temperature inside an up-draft gasifier is much higher than in its down-draft cousin: so high, in fact, that there is applicable risk to the devices inside the gasifier. This means that these devices (such as the grate that holds the bulk of the unconverted char) must be either a.) made from stronger, less temperature-sensitive materials, or b.) protected by blowing in steam or some other gaseous coolant to maintain a more reasonable temperature in that area of the gasifier (Reed and Gaur, 2001).

Lastly, the biggest drawback that is readily observable in up-draft gasifiers is that they lack the down-draft gasifiers' abilities to eliminate tar. This is because, despite its high peak temperature, this gasifier has a very low outlet temperature. As such, the syngas very quickly cools down, forcing most of the tar to condense or de-gasify. As such, there is a great loss in efficiency to offset the reduced wasted heat, as most biomass feedstocks will tend to produce a lot of tar (As discussed later) that *must* be cleaned extensively before the resulting syngas mixture may be used in any sort of application, especially for traditional combustion engines and gas turbines.

2.2.1.3 The Fluidized Bed Gasifier

A very interesting and intuitive gasifier design, this type of gasifier uses fluidization to move the feedstock particles. Basically, the gasifier is filled with a bed of solid, dry feedstock particles (which may or may not actually be fuel particles. Sometimes, sand or gravel forms the bed and the fuel enters the bed with the gasifying agent), which is then met with a moving stream of fluid particles (usually the gasifying agent) that are allowed to seep through the pores and cracks in the solid medium. When the fluid flow rate reaches a certain "critical point," the solid particles become fully suspended in the fluid: they begin to levitate freely and essentially begin to behave as a fluid themselves; hence, they have been "fluidized." A diagram of a this type of gasifier can be seen in Fig. 2.3.

Fluidized bed gasifiers are not suitable for small-scale applications less than 10MW because of their high heat transfer rates. They are also difficult to operate, as the entire gasification process is very dependent upon a highly complicated equilibrium state, which must be maintained at all times. One particular strength of fluidized bed gasifiers is the fact that they do *not* produce slag, so they can use certain types of fuels that would ordinarily corrode the walls of slagging gasifiers. Instead, the stray ash is agglomerated into heavier particles that easily fall out of the fluidized mixture and are swept out of the bed when they reach the bottom. In addition, they can operate more readily at higher temperatures than any fixed-bed gasifier can, making them much more suitable for coal gasification, especially for high-ranking coals (Rezaiyan and Cheremisinoff, 2005).

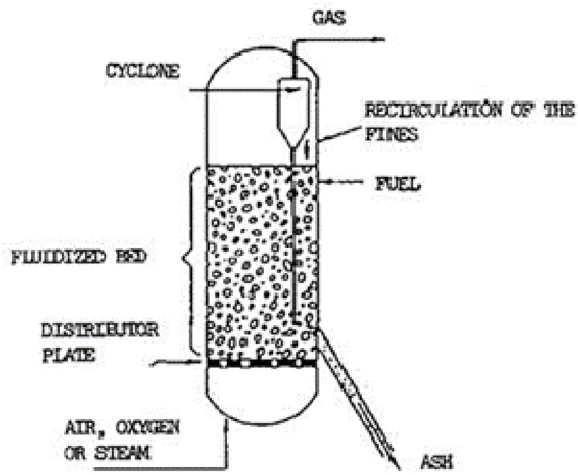

Figure 2.3 The fluidized bed gasifier (FAO, 1986)

However, fluidized bed gasifiers do not fair very well with feedstocks that have low ash fusion temperatures. Fluidized bed gasifiers must operate at generally higher temperatures than fixed bed gasifiers to be effective, so, naturally, using a fuel where the ash fusion temperature is too low will not allow for proper gasification. If the fuel is gasified anyway, the feedstock ash will melt, becoming slag, and begin to stick to the bed particles, resulting in rapid bed de-fluidization: a terribly undesirable effect. Second, despite its name, the fuel feedstock *must* be put in *dry*, as a slurry feedstock will only inhibit the gasifier's ability to produce a fluidized bed.

2.2.1.4 The Entrained Flow Gasifier

This particular gasifier gets its name from the fact that the feedstock particles and the gasification agent are a part of the same stream once inside the gasifier. In other words, the solid particles or liquid droplets of feedstock have been *entrained,* or "trapped" inside the gas stream. The *entrainment* that results is a matrix of solid or liquid particles within a gaseous medium. This allows for a much more even temperature distribution and a more steady reaction rate. Entrained flow gasifiers are very common in big power plants (> 200 MW) because they can achieve very high syngas mass flow rates and high yields, higher than any other gasifier type: a necessity for large plants. Figure 2.4 shows a diagram of a typical entrained flow gasifier.

Most entrained flow gasifiers produce slag. Part of the slag forms a protective coating along the sides of the gasifier, which protects the walls from more corrosive substances that may form during gasification. Entrained flow gasifiers are capable of undergoing gasification at high temperatures (> 2000°F), meaning that there is absolutely no risk of tar formation inside the gasifier walls. The greatest strength of the entrained flow gasifier, though, is that it can accept *any kind of feedstock*. Since the flow regime is basically just a gas with particles suspended in it, any liquid or powdered/pulverized solid is a viable fuel input for entrained flow gasifiers, regardless of its atomic makeup.

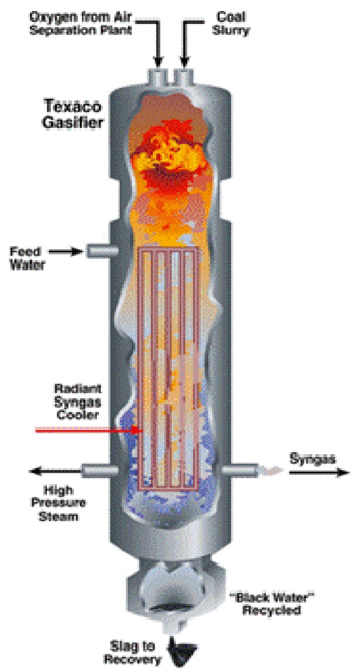

Figure 2.4 An entrained flow gasifier from GE/Texaco (Jenkins, 2008)

Although, for all of their strengths, entrained flow gasifiers have a few debilitating drawbacks. For one, the average feedstock particle size is very small: on the order of tenths to hundredths of millimeters in diameter. This is not a problem for liquid feedstocks, but solids like coal and biomass must be pretreated before they can be used in an entrained flow gasifier. This is usually not a problem for coal, because it can simply be ground down and pulverized mechanically. For biomass, however, as mentioned previously, this can be a major problem without proper pretreatment. Second, most entrained flow gasifiers typically require the use of oxygen, *not air*, as the gasifying agent. Very few entrained flow gasifiers use air, because (a) air volume makes the gasifier and associated piping bigger and (b) the conditions in the gasifier make the presence of nitrogen a problem for syngas production: the high temperatures and pressures can cause large amounts of unwanted NO_x production, rendering the resulting syngas mixture virtually unusable for power applications. Because of this strict oxygen requirement, most all entrained flow gasifiers require an ASU in order to operate. Finally, the syngas that leaves the gasifier will have an extremely high temperature compared to the other gasifier types, and there will be a large energy loss resulting from this during the cooling stage before the gas cleanup system (Rezaiyan and Cheremisinoff, 2005).

2.2.1.5 The Transport Gasifier

A recently produced model that has been under testing by Kellogg Brown and Root (KBR) since 1996, this gasifier type utilizes a similar structure to Circulating Fluidized Bed gasifiers (CFBs), except with higher velocities, riser densities, and circulation rates. Because the device can be run as both a combustor and a gasifier, it is sometimes called the “Transport Reactor” rather than transport gasifier. It is unique in that there is no true “bed” in the gasifier itself, as the feedstock, gasifying agent, and transport agent (sand) are constantly in motion throughout the system, much like an entrained flow gasifier, but with larger sand particles as the heat transfer agent. Ash and unconverted char particles are filtered out via a gravity-driven “disengager” (for larger particles) and a high-temperature cyclone filter (for smaller particles). Char particles separated in this fashion are sent back to the “mixing zone” (where the feedstock first enters the device) through a pipe called the J-leg, where they are re-gasified with a portion of the recycled syngas that came with them through the filters.

Currently, this gasifier can operate at temperatures up to 1825°F and gage pressures of up to 240 PSI (Loganbach et. al, 2001). The transport gasifier is still in the developmental stages. Plans were made to open a plant using this gasifier in Wilsonville, Alabama (Loganbach et. al, 2001) and another in Orlando, FL (Wallace et. al, 2006). The Wilsonville plant, as of 2005, has operated at nearly 8,000 hours, using both air-blown and oxygen-blown modes of operation, and has been largely successful in producing syngas. The Orlando plant has not been completed as of yet, but the plans are to make it a commercial, 285MW power plant with a gas turbine capable of running on either syngas or natural gas produced by the transport gasifier. It remains to be seen how this new gasifier will perform when used in the commercial sector. A schematic of the KBR transport gasifier can be seen in Fig. 2.5.

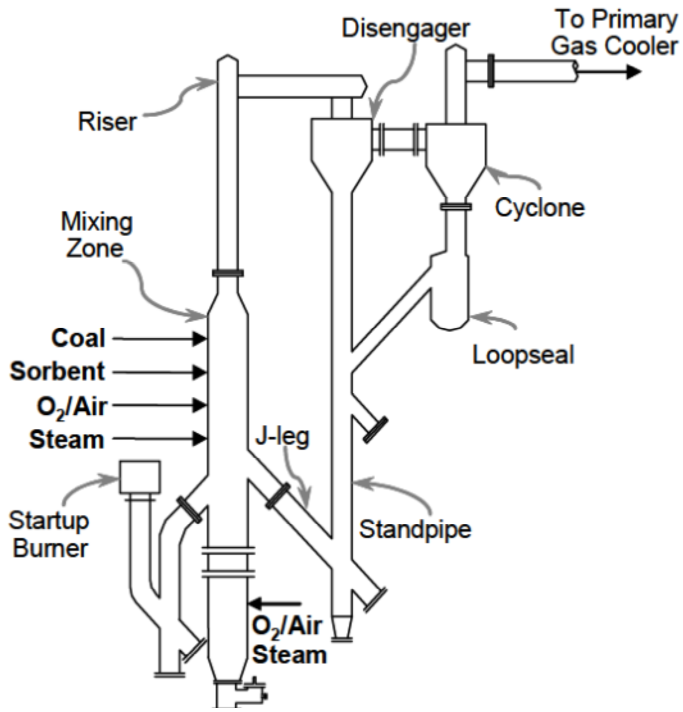

Figure 2.5 The transport gasifier from Kellogg Brown and Root (Wallace et. al, 2006)

2.2.1.6 The Indirect Gasifier

As mentioned previously, gasification is an *endothermic* process, meaning it requires some degree of energy input in order to take place. Most gasification applications use *direct combustion* to fill this energy need. Another option is to use indirect heating to provide thermal energy to feed the gasification process. A group of scientists and engineers from the Netherlands developed what they call *indirect* gasification, using a gasifier called the MILENA gasifier. In this process, the combustion and gasification reactions are performed in two separate chambers, with the heat of combustion being provided to the gasified feedstock through the walls of the chambers (van der Drift et. al, 2005.) Figure 2.6 is a schematic of the MILENA gasifier.

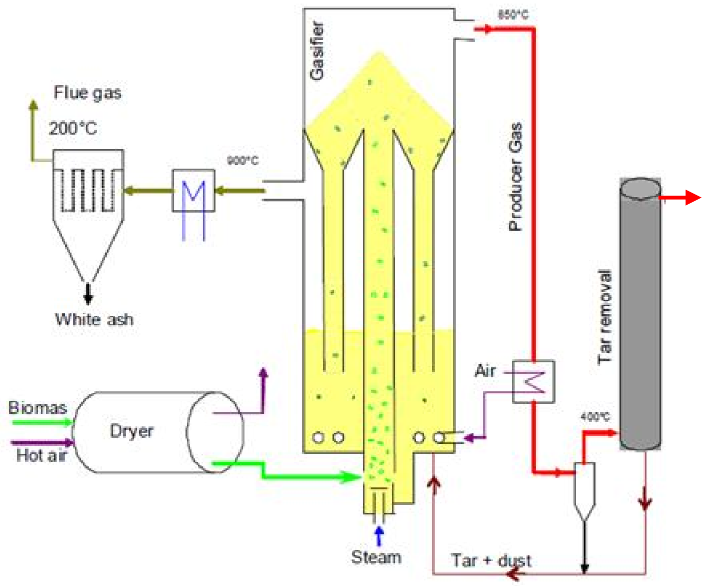

Figure 2.6 The MILENA gasifier (Vreugdenhil et. al, 2009)

The MILENA process's main use is in the production of a substitute natural gas (SNG) from methane. The primary gasification agent, as can be seen from Fig. 2.6, is steam. Notice that no air or oxygen ever enters the gasification zone, and the only direct exposure to oxygen the fuel gets is at the very bottom. In this manner, the resultant producer gas will contain high amounts of CO, H_2, and CH_4, with relatively little CO_2. (Note, however the externally combusted gases still contains CO_2.) Secondly, since no air is used in the gasification chamber, a virtually N_2-free producer gas stream can be exported without having to make use of an ASU, dramatically reducing power consumption.

The disadvantage to this type of gasifier is that it cannot achieve temperatures higher than 800-900°C (or 1470-1650°F) in the gasification chamber due to the temperature limit of the materials used in the heat exchanger and lack of adequate surface area for the heat to transfer, especially compared to entrained flow and fluidized bed gasifiers. Because of this, tar production is a major problem with this gasifier's operation. Several countermeasures have been performed in this area with varying degrees of success, but the most significant improvement came when the developers decided to change the design so that the initially pure biomass feedstock could be *co-gasified* alongside *lignite coal*. Lignite's lower volatile content and less complex molecular structure reduced the total amount of tar produced by the system. The results also showed that introducing lignite into the mix allowed more of the tar to be burned away due to lignite's higher heating value. In fact, overall tar production was reduced by more than *half* in one instance (Vreugdenhil et. al, 2009). That being said, as of 2009, this gasifier is still in the developmental/testing stages, and no commercial models have been produced yet.

2.2.2 Commercial Gasifier Models

When becoming familiar the basic types of gasifiers available, it becomes necessary to witness some real gasifier designs, particularly those that have enjoyed the greatest commercial success. This is primary focus of this section. In reading of these models, it becomes apparent that the vast majority of commercial gasifiers are classified as *entrained flow* gasifiers. This is because (a) most commercial applications have very large power demands, and, as such, these applications require significant syngas mass flow rates that only an entrained flow gasifier can provide, and (b) entrained flow gasifiers have the highest carbon conversion efficiency available (> 95%). Fluidized bed gasifiers are popular research topics due to the complex fluid mechanics involved in their operation, and a few do make their way into the commercial sector due to their non-slagging nature, moderately high throughputs, better ability to accept lower grade coals as feedstock, and less stringent need for fuel pretreatment (Spiegl, et. al, 2010). Among these is the Kellogg-Rust Westinghouse (KRW) model, which has not been widely used in part due to a failed IGCC demonstration project in Pinon Pine, Nevada in 2000.

Although a large amount of fixed-bed gasifiers have been successfully operated since the 1970s (Dennis, et. al, 2006,) especially with British-Gas/Lurgi (BGL) gasifiers in South Africa, in the end, more and more entrained flow and fluidized bed gasifiers are installed commercially.

2.2.2.1 The General Electric/Texaco Gasifier

The Texaco Gasifier has been in operation in the oil industry for nearly 45 years, using fuels such as oil, petcoke, natural gas, and coal. When GE acquired the technology in 2004, the company opened a test plant that has since operated in Aoio, Japan, currently running at 6 tons/day and operated by Ishikawajima-Harima Heavy Industries (IHI). In addition, there is a 15 tons/day test plant that opened in Montebello, California, that has since been relocated to China (NETL, 2011). The GE gasifier was the very first gasifier to be used for IGCC in the United States, during the Cool Water Project in 1984. Over 1.1 million tons of black coal were gasified over the course of its operation. The Tampa Electric Company in Polk County, Florida also uses this gasifier model in its Polk Power Station plant, which began operation in 1996 (Dennis, et. al, 2006). In total, over 100 commercial gasification projects around the world use this gasifier, making it one of the most successful gasifiers ever developed. Figure 2.4 from the previous few sections is a model representation of this commercial gasifier.

The process itself is slurry-fed, with about 65% water in typical operations, and is oxygen blown as well, demanding about 96% pure oxygen for most applications. The walls of the gasifier are protected by a special refractory coating to increase the longevity of the gasifier by protecting it from large amounts of slag, which gets collected by a lock-hopper at the bottom of the gasifier. Most models of this gasifier use a radiant syngas cooler to reduce the syngas temperature before cleanup, but some models do make use of a syngas *quench* process (This is discussed more in detail in later sections.)

As an entrained flow gasifier, the GE/Texaco model reaches very high temperatures of around 2,200 – 2,800°F, and pressures of over 300 PSI. In addition, the larger models in most commercial plants, like in Tampa, have throughputs of over 4,000 tons/day (Jenkins, 2008).

2.2.2.2 The Shell Gasifier

Shell Corporation first developed the Shell Gasification Process (SGP) around 1950 in order to produce syngas from gaseous and liquid feedstocks, particularly natural gas and oil. Then, in 1972, Shell extended this to solid fuels (i.e. coal) by developing the Shell Coal Gasification Process (SCGP), which makes use of a dry-fed, oxygen-blown, entrained flow gasifier (NETL, 2011). Figure 2.7 shows a basic layout of this gasifier. This project began as a collaboration between Shell Corp. and the Krupp Koppers Company.

Shell’s gasifier is used in many existing plants and other applications today. The Buggenum co-gasification plant in the Netherlands, for one, uses this gasifier model (Kanaar, 2006). In addition, Shell constructed several pilot plants as proof-of-concept shortly after development: a 6-ton/day plant in Amsterdam, another 15-ton/day plant in Harburg, Germany in 1978, and, eventually, a 250-ton/day power plant was constructed at Deer Park in Houston, Texas in 1987 (NETL, 2011).

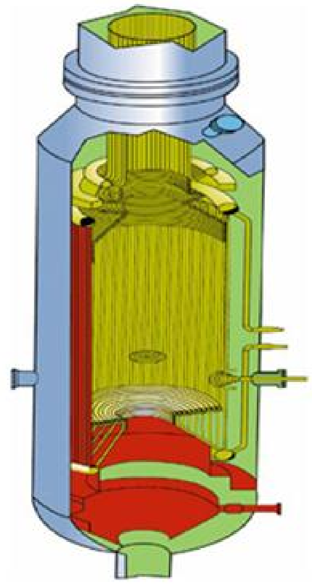

Figure 2.7 The Shell Gasifier (NETL, 2011)

Before the Houston plant was constructed, however, in 1981, Krupp Koppers left the SGCP Project, and both companies mutually decided to go their separate ways in the area of coal gasification. Krupp Koppers would eventually develop their own, competing dry-fed gasifier later on, called the PRENFLO gasifier. However, this gasifier would not enjoy nearly the amount of success Shell's gasifier would, and the only commercial plant in the world currently using the PRENFLO model is the Elcogas plant in Puertollano, Spain. The two companies joined forces again in 1999, but, currently, only the SGCP model is offered commercially.

The Shell gasifier, like GE's gasifier, is refractory-lined, to protect the walls from slag production. Shell's model, however, comes equipped with an inner membrane wall, with many steam-filled tubes to cool the syngas, as can be seen in Fig. 2.7. This wall allows the gasifier to operate for longer periods of time (20-year life cycle!) without maintenance, due to the fact that the slag will condense on the wall and form a further-protective layer, greatly reducing the erosion of the inner walls, as compared to the brick refractories used by most of Shell's competitors (NETL, 2011). The SGCP uses a *quench* design to help circumvent some of the extra capital costs from its membrane wall design (which is much more expensive than refractory brick), and dry-fed nature.

Shell's gasifier has very high throughput, even for an entrained flow gasifier, and can process coal feedstocks of nearly any grade without modification, as well as many other fuels, such as petcoke. This is mainly due to its built-in drying and milling section, which eliminates the effects of moisture and other compounds on gasifier performance. However, this gasifier is relatively expensive compared to other commercial models. Despite this drawback, Shell's gasifier still enjoys some degree of commercial success comparable to that of GE's gasifier, and Shell Corp. has sold at least 12 licenses for coal-to-chemical plants in mainland China (Dennis et. al, 2006).

2.2.2.3 The E-Gas/Conoco-Phillips Gasifier

In 1976, Dow Chemicals began working in collaboration with Global Energy and Destec Energy to produce a viable, coal-gasification reactor. Eventually, a 36-ton/day pilot plant followed by a 550-ton/day plant was constructed in Dow's main manufacturing complex in Plaquemines Parish, Louisiana using lignite as the main source of fuel. Then, in 1984, Dow created the plans for the Dow Syngas Project, in which, with support from the U.S. federal government's Synthetic Fuels Corporation (SFC), they constructed a commercial-scale IGCC plant in Plaquemine using their new "E-Gas" gasifier, which can be seen in Fig. 2.8. Louisiana Gasification Technologies, Incorporated (LGTI), a subsidiary of Dow Chemicals, operated the plant, and the plant itself began operations in 1987. Over 1600 tons of sub-bituminous coal taken from the Powder River Basin mines were used per day (Dennis, et. al, 2006). The exported syngas produced by this plant was taken to the main Plaquemine complex and burned in 2 Westinghouse 501D gas turbines, generating about 184 MW of total power. Unfortunately, in 1995, the SFC's support ended, and the plant was shut down. Earlier, however, in 1993, Destec Energy and PSI Energy entered a Joint Venture to replace an aging Pulverized Coal plant in West Terre Haute, Indiana with a brand new IGCC plant using Dow's E-Gas gasifier (NETL, 2011). After several business maneuvers and changes in administration, the Wabash River project opened in 1995, using Illinois #6 bituminous coal. This plant enjoyed immense success, and is still in operation today, now using petroleum coke as its sole source of energy (Dennis, et. al, 2006). When ConocoPhillips acquired the E-gas technology in 2003, the company began developing several new E-Gas projects as a part of the U.S. Department of Energy's Clean Coal Power Initiative (CCPI), including the Mesaba project in northern Minnesota and the Steelhead project in southern Illinois.

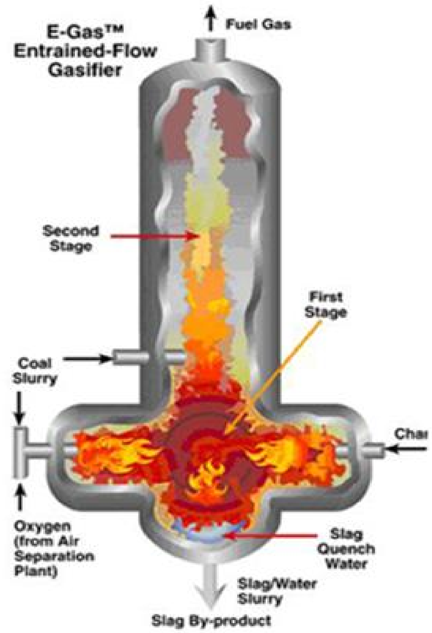

Figure 2.8 The Conoco-Phillips/E-Gas gasifier (NETL, 2011.)

This gasifier is slurry-fed and oxygen-blown, like the GE gasifier, but what makes the E-Gas gasifier unique is that the input oxygen and fuel are injected from the *bottom* of the gasifier, as opposed to GE's and Shell's designs, which both provide the fuel from the top. It has a unique, 2-stage gasification process, in which a portion of the slurry is injected into the first stage, and undergoes highly-exothermic oxidation reactions, producing slag, which exits at the bottom. Hot syngas then enters the second stage, just above the injection zone and the rest of the slurried feedstock is injected here, where it undergoes the more endothermic gasification and devolatization reactions. Char and heavy hydrocarbons from this process are recycled back into the first stage, where they are re-gasified, allowing this gasifier to achieve nearly complete carbon conversion.

The 2-stage design increases the efficiency of the gasification process, since it reduces the temperature of the exiting gases without any additional energy losses simply by isolating the gasification "zone." For reference, the first stage reaches temperatures of over 2600°F, while the raw syngas exits the gasifier from the second stage at around 1900°F, meaning that this gasifier has one of the lowest exit temperatures of all commercially available entrained flow designs.

2.2.2.4 The Mitsubishi Heavy Industries Gasifier

Mitsubishi Heavy Industries began working in collaboration with Combustion Engineering and several other utility companies, along with the Japanese federal government in the early 1980's. The goal of this joint venture, known collectively as the Clean Coal Power R&D Company, was to produce an entrained flow gasifier that was as efficient as possible. The project was completed completely internally within Japan, and the end result was an entrained flow demonstration gasifier that was used in a small, 2-ton per day process development plant, and, later, a 200-ton per day pilot plant in a small city called Nakoso, about 200 km north of Tokyo (NETL, 2011). A diagram of this gasifier is shown in Fig. 2.9.

MHI's gasifier is relatively new compared to the previously discussed models, and, as such, hasn't been used in any commercial plants, yet. However, in 2004, MHI began construction on a 250 MW_e (electric power) IGCC plant at the same site as their previous pilot plant, in Nakoso, Japan. This new IGCC plant would display an efficiency of over 42% LHV (MHI, 2011), which MHI attests to their gasifier's more efficient design, citing lower necessary auxiliary power and fewer heat losses than their European and American counterparts (Hashimoto, 2010).

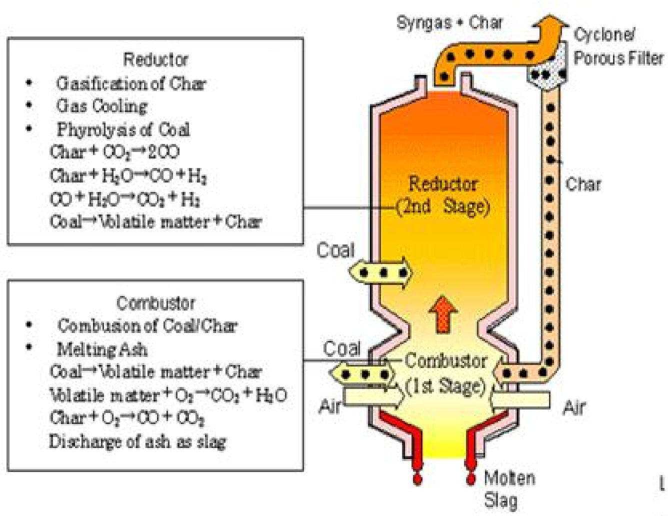

Figure 2.9 The Mitsubishi Heavy Industries gasifier (MHI, 2011)

The Nakoso plant is just the first step for this new gasifier, as there are plans for a new, full-scale, commercial IGCC plant with Carbon Capture and Sequestration (CCS) capabilities with MHI working heavily with the ZeroGen Pty Ltd. Company in Australia. The new plant will provide over 530 MW of electricity, and MHI's own gasifier will be used in its base design thanks to the success of the Nakoso plant. The new plant is scheduled to be fully commissioned by around 2015, and the expected efficiency is around 48-50% using Australian bituminous coal (Ishii, 2010 and Hashimoto, 2010). In addition, numerous chemical applications in Japan and elsewhere have also made use of MHI's unique design.

MHI's gasifier is dry-fed, and makes use of a water-cooled membrane wall, much like Shell's gasifier. In addition, it makes use of 2-stage gasification, like the E-Gas gasifier. What makes MHI's design unique is that their gasifier is *air-blown*: making it the only non-oxygen-blown entrained flow gasifier commercially available. Because of this fact, MHI's gasifier uses less auxiliary power than any other gasifier available due to the absence of an ASU, although some applications use "enriched" air, around 50% O_2 content, which does require an ASU, but only using partial-load operation. For chemical applications, however, MHI's gasifier does still require an oxygen-blown system to increase the rate of production (Ishii, 2010).

Like the E-gas gasifier, the first gasification stage is where exothermic oxidation reactions occur, raising the temperature enough to melt the ash content into slag, which is quenched and drained from the bottom. The gases released are then sent up through the second stage, where more fuel is injected into the stream, allowing the more endothermic reactions to occur. Unconverted char and heavier volatiles are then recycled back into the first stage to allow for more complete combustion (Dennis, et. al, 2006). Due to this design, MHI's gasifier boasts one of the highest carbon conversion efficiencies among any gasifier ever built, which MHI claims is above 99.9% (Hashimoto, 2010).

2.2.3 The Air Separation Unit

It has been mentioned that all gasifiers that are oxygen-blown require the use of an Air Separation Unit (ASU) in order to operate. There are several different types of ASU available commercially, but by far the most common type for large plants is the cryogenic ASU. A basic schematic of a typical cryogenic unit is shown in Fig. 2.10.

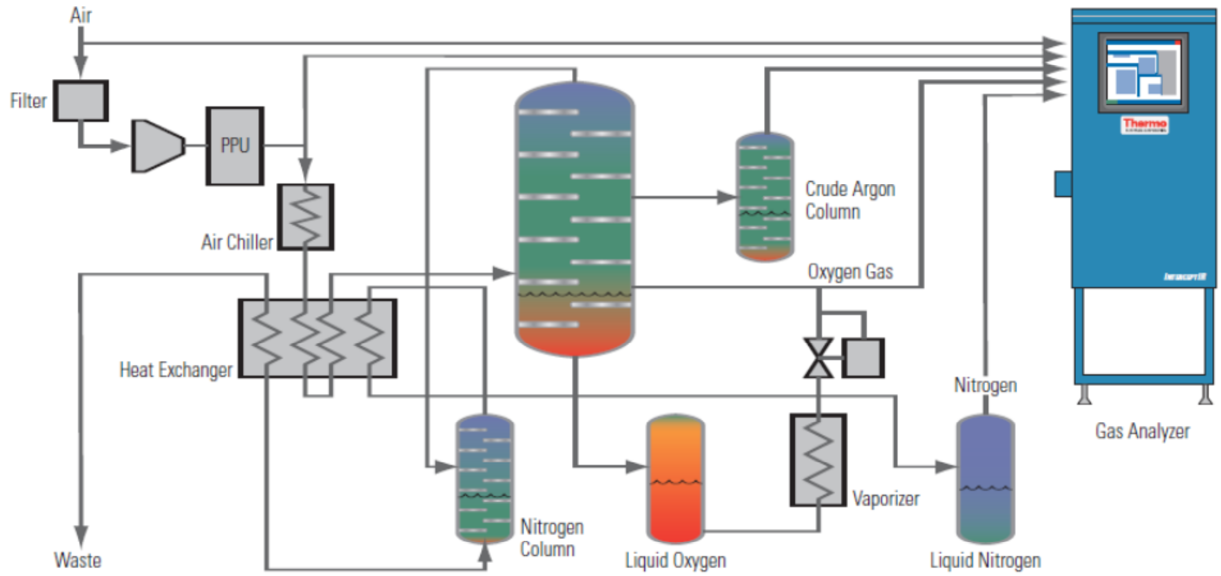

Figure 2.10 Cryogenic ASU schematic (Thermo Electron Corp., 2005.)

At the beginning of the air separation process, the atmospheric air is filtered of its impurities and then compressed, usually to around 6 bar of absolute pressure. Afterwards, the air will have heated up, so it must be cooled back down to the ambient temperature, usually through the use of a simple heat exchanger or chiller. During this phase, water (being a compressed liquid in this state) will condense out of the air mixture, and, thus, will be subsequently removed from the air stream.

Before this occurs, however, the air must pass through a Pre-Preparation Unit (PPU), where non-vaporized water, some CO_2, and nitrous oxides will be removed, so that they do not freeze inside the chiller or reboiler later on. Most commonly, this is accomplished via a molecular sieve or molecular adsorbers (see the section about adsorption in the section on Carbon Capture and Sequestration.), which can also removed stray hydrocarbons that can exist in manufacturing and power plant atmospheres, such as acetylene or benzene.

After having been pre-treated and cooled, the air now consists mostly of nitrogen, oxygen, and argon. At this point, it enters a large array of distillation towers and cryogenic equipment (collectively called the "cold box") which reduces the temperature of the air to cryogenic levels (-185°F or so,) taking advantage of the fact that these three primary components of air have different vaporization points, separating them one by one as the temperature drops. Since argon and oxygen have similar boiling points, most oxygen producing plants use two columns, one high pressure column and another, lower pressure column. Connecting these two will be a device called a “reboiler” (not shown in the figure), which vaporizes the liquids from the low pressure column while simultaneously condensing the gases from the high pressure column. This process purifies the oxygen from the higher pressure column (read: condenses it out) before they are delivered to whatever process for which the ASU is required. Most gasification power plants require streams of high purity oxygen, so the ASUs in these plants will have more than just the two stages mentioned for cleaning away the argon, but the overall process is the same (Thermo Electron Corp., 2005).

ASUs must be operated very carefully, especially when used in manufacturing or power production, as many hydrocarbons have higher boiling points than oxygen, meaning that they will be condensed out of the overall air mixture alongside the oxygen. This has lead to many instances of spontaneous combustion in several plants, which can destroy equipment and lead to serious injury to the plant operators. In addition, these stray “fugitive” hydrocarbons and other compounds can form thick blockages in the reboiler and distillation segments, greatly disrupting unit operation. On the whole, caution must be taken when using an ASU, and, before using any specific model of ASU, careful research should be performed on the contaminant contents of the local air supply before the unit is commissioned for use so that these problems can be safely avoided.

2.3 The Gas Cleanup System

After exiting the gasifier, the raw syngas must be cleaned of all impurities before it is usable in any further processes. Since these gases result from gasification, this process can be more easily achieved than in a combustion plant. The following sections contain information on all of the various components and processes used in a typical gas cleanup system, with special attention given to the various methods of carbon capture.

2.3.1. Cooling the syngas

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.

There are two methods currently in use for cooling off the syngas: a syngas *quench* and *radiant/convective coolers*. Radiant syngas coolers (RSCs) and Convective syngas coolers (CSCs) are essentially large heat exchangers that typically use water as the coolant source. This water most often comes from the steam system 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 tend to be very efficient: 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. For an example of radiant cooling, see Fig. 2.4's representation of the GE gasifier, which comes with an RSC built into it.

A quench, on the other hand, is much easier and cheaper to implement than RSCs and 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. This allows the temperature to be reduced without employing any other equipment aside from extra piping and control valves. However, 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. Second, recall the gasification reactions from Chapter One: in particular, reaction 1.6, shown again below:

$$\mathbf{CO + H_2O \leftrightarrow CO_2 + H_2}$$

Again, this is an equilibrium reaction called the 'Water-Shift' or 'Water-Gas Shift' (WGS) reaction. Quenching the syngas increases the concentration of H_2O on the left side of the reaction. In response, the chemical system will shift the reaction towards the right in order to maintain equilibrium. This means that a large amount of CO will transform into CO_2, which will waste its potential as a fuel later on in the GT combustor. Hydrogen is still useful, as it has a high enough heating value, but CO is the better fuel and there is no need to convert CO to CO_2 at the fuel conversion stage if it is not necessary. However, if the system uses *carbon capture*, pushing the WGS to the right side becomes a more attractive option, since shifting the equilibrium to the right side of the Water-Shift reaction to obtain more H_2 is preferable in this case for separating H_2 and CO_2 later, since the amount of "shifting" done later will be reduced, and some energy and additional money will be saved by using a quench to perform some of the water shifting. However, even with this benefit, the grade of energy from the temperature drop during cooling is still lost during quench, where it would be recovered partially by the steam cycle had radiant cooling been used.

In short, radiant/convective cooling is always more efficient than quench cooling, but quench cooling may be a better *economic* decision in the long run due to how expensive radiant coolers can be. When CCS is implemented, quench cooling becomes an obvious winner because it can also carry out part of the WGS reaction as an additional advantage. Later on, after particulates are removed and COS hydrolysis occurs, the syngas must be further cooled before acid gas removal occurs. This is almost always done with a series of simple heat exchangers. For clarification, when "cooler" is mentioned without further explanation, it is referring to one of *these* devices, and not the radiant/convective cooler attached to the gasifier.

2.3.2 Dealing with Ash

Ash is a problem in most thermal applications because it cannot be vaporized with the rest of the fuel, and it cannot be gasified or combusted. Ash particles either melt and become slag, settle to the bottom of the gasifier, or become entrained in the rest of the syngas and enter the gas cleanup system. This latter type is sometimes called "fly ash," mentioned briefly earlier. Fly ash *must* be removed from the syngas before GT combustion, as the stray solid particles can severely damage the turbine blades. Fortunately, ash is relatively easy to deal with using a particulate scrubber and/or a cyclone filter, which are usually included in most IGCC systems, even those with gasifiers that don't produce fly ash. This is because unconverted carbon particles still sometimes entrain out of the gasifier before they are able to be gasified, and these devices ensure that no solid matter will be allowed into the gas turbine.

In most commercial gasifiers, however, the temperatures are so high that the ash will *melt* before it has a chance to leave the gasifier. This molten ash, when solidified, is what is known as *slag*. Slag must be handled very carefully, as it tends to be extremely corrosive, and can damage internal gasifier parts if precautions are not taken, especially for gasifiers that use radiant coolers

(NETL, 2011). Most entrained flow gasifiers, as mentioned earlier, are not concerned with slag, as many take advantage of special refractory surfaces, which the slag can stick to freely, forming an even thicker protective coating against the high temperatures that occur inside the gasifier (this is especially true for the Shell Gasifier). For all these problems, slag is very easy to collect and extract: most gasifiers just have a simple exit port with a basic quench to solidify it on its way out. Also mentioned previously, slag is a useful and sought-after by-product of gasification, and has its uses in the production of concrete, ceramics applications, and as an ingredient in fertilizers for gardens and farms.

2.3.3 COS Hydrolysis

The COS hydrolysis reaction is an equilibrium reaction (much like the water shift reaction) shown below in reaction 2.1:

$$\mathbf{COS + H_2O \leftrightarrow CO_2 + H_2S} \qquad \text{(R2.1 – COS Hydrolysis)}$$

Adding more water to the syngas causes the reaction to shift to the right, just like in the water-shift process for CO_2 capture, creating more carbon dioxide and hydrogen sulfide. This stage is a fairly simple procedure, and, due to its simplicity, it is an ideal “place” to utilize the CO-shift process at the same time, since both the Hydrolysis and Water-Shift reactions depend upon the presence of water. When performed in this manner, it is called a “sour” shift (to be discussed later on in more detail.) The reason for this stage is because COS cannot be efficiently separated from the syngas stream, and usually isn’t worth the economic investment unless there is profit to be made by selling it. Performing COS Hydrolysis is generally a more economic practice, especially since all plants require some form of treatment for removing H_2S later on.

Most hydrolysis reactors can achieve conversion efficiencies of about 99%, and all of them use a catalyst to increase the reaction rate. While the process itself is technically exothermic, the reactants and products make up so little of the surrounding syngas, that the reactor conditions are very nearly isothermal (NETL, 2011). However, this is not the case when this reaction occurs alongside the CO-shift process.

2.3.4 Acid Gas Removal

The term "Acid Gas" refers to sulfurous compounds, particularly H_2S. Acid Gas Removal (AGR) is, of course, the generic term for any process that separates this compound from the rest of the syngas mixture in a gasification system. While power applications can typically allow for more sulfur in the syngas (10-30 ppm) than other gasification applications (NETL, 2011), sulfur is the main cause behind the acid rain phenomenon, and having large amounts of sulfuric emissions isn't looked highly upon politically or socially.

Generally, AGR makes use of a solvent to absorb the sulfur (H_2S) and separate it from the main syngas. The rich solvent is then later "stripped" of the sulfurous compounds in it and recycled back into the syngas to absorb more H_2S. In this way, AGR is very similar to most amine-based CCS sorption processes (to be discussed in more detail later). Figure 2.13 is a basic layout of this process. For acid gas removal, there are generally two main groups of processes: chemical absorption and physical absorption.

Chemical absorption is performed using compounds such as monoethanolamine (MEA), diethanolamine (DEA), and methyl-diethanolamine (MDEA), all of which are commercially available chemical solvents (Dennis et. al, 2006). The chemical solvents themselves are expensive: around \$1500-\$1600 per ton for MEA (ICIS pricing, 2010), but the chemical-based process in general is more efficient and less costly than the physical process (NETL, 2011). Physical absorption units themselves are about 2-4 times more expensive than chemical units, since they require cryogenic temperatures, as opposed to the chemical units, which operate slightly above ambient conditions.

Physical absorption is typically performed using dimethyl ethers of the substance polyethylene glycol. This is commercially known as the Selexol® process. Also available is the Rectisol® process, which uses cryogenic methanol as the solvent. Physical absorption solvents are more costly (≥\$2000), but they *last longer* than chemical ones, simply because removing the captured H_2S/CO_2 is more easily accomplished, making the units easier to clean out, which means that physical units on the whole will be cheaper to *maintain*. While physical units require more energy, resulting in greater heat/auxiliary losses, they use *less water* to run, meaning that they will take less energy from the steam cycle in IGCC plants than chemical units. This results in a comparable cost between the two processes (NETL, 2011). Physical solvents also tend to absorb more CO_2 than chemical solvents, making them more ideal for carbon capture plants. For this reason, physical solvents are most often used in chemical production applications, whereas for power generation, chemical solvents are more ideal.

2.3.5 Carbon Capture and Sequestration

Due to the imposition of the “carbon tax,” and other political and environmental concerns, it sometimes becomes necessary to institute a method of removing carbon-heavy pollutants from some power plants’ exhaust streams. The general term for the sets of processes that perform this function is “Carbon Capture and Storage” or sometimes “Carbon Capture and Sequestration” (CCS for short). There are many forms of CCS technology available in industry, and the purpose of these next few sections is to highlight the types of technology available as well as their implementation in both IGCC and PC plants.

2.3.5.1 CCS Overview

CCS technology on the whole is very broad, and there are many processes available, commercially or otherwise. Until recently, there has not been much focus on CCS, since removing CO_2 tends to significantly reduce plant thermal and electrical efficiencies, and historically, the primary focus has been to eliminate particulates and SO_x emissions. But, since 85% of all GHGs come from the energy-production industry and 95% of those are CO_2, more focus has been placed on incorporating and improving the implementation of CCS (NETL, 2011). As will be shown in the later sections, IGCC has an immense advantage over PC plants when CCS is included, both economically and thermally. The biggest advantage, as mentioned before, and will be discussed in detail later, is that IGCC allows for CCS to be implemented *before* full combustion, allowing the CO_2 to be removed *before* SO_x and NO_x can be formed, and while the gases are still under high pressure conditions, allowing them to be cleaned by a unit that can process the same mass of syngas, but at a much smaller (and cheaper) size than that in post-combustion CCS due to high pressure and without a large amount of combusted gases.

2.3.5.2 Types of CCS Technology

2.3.5.2.1 Sorption Processes

The first and most common classification of CCS technology involves the use of a material medium to “catch” the CO_2 gas, directly removing it from the gas stream. The two methods within this class are *absorption* and *adsorption.* Together, these two methods are called *Sorption Processes*. For this section, it is important to understand the difference between *AB*sorption and *AD*soprtion. The former process occurs when the contaminant substance passes through a medium and is trapped *within* the medium: think of a sponge that absorbs water. The latter occurs when the same substance passes *over* a medium and is trapped *upon* the surface. Figure 2.11 highlights the difference between these two similarly named concepts.

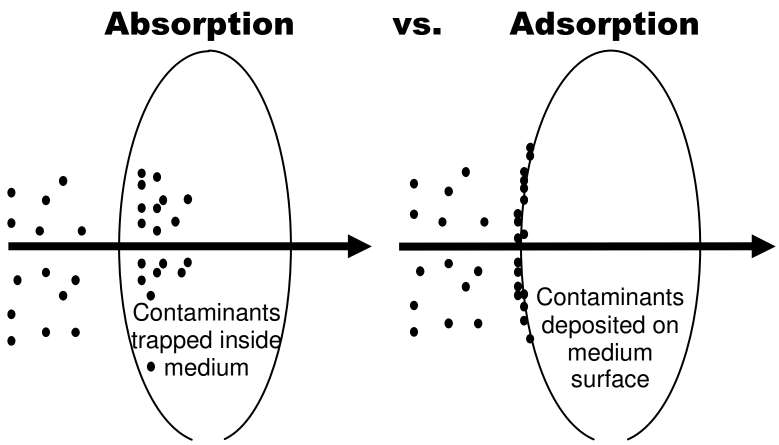

Figure 2.11 Differences between absorption and adsorption

2.3.5.2.1.1 Absorption

By and large, the most common form of CCS used is through absorption. Of all absorption methods, the most common is through the use of an amine-based solvent, such as monoethanolamine (MEA), in a manner similar to the Acid Gas Removal process. First, the flue gas is compressed and sent into a column called the "Absorber," where it is mixed with the solvent, which chemically bonds to the CO_2 in the gas stream, carries it out of the gas stream, and exits through the bottom of the column. The rest of the flue gas is carried out through the stack. The "rich solvent" is now sent to another column called the "Stripper," where the amine compound is de-bonded from the carbon content through the use of condensing steam. The CO_2 released from this process is then extracted and sent elsewhere for treatment and storage, while the amine-water mixture is sent down towards a re-boiler, which re-vaporizes the water and sends it back to the stripper and sends the clean solvent back to the absorber after passing it through a heat exchanger that will pre-heat the rich solvent exiting the absorber before it enters the stripper again (Tondeur, 2009). While the use of MEA is a chemical absorption process, the famous Selexol method is a physical absorption process (where the CO_2 is physically dissolved in the amine solution rather than being chemically bonded to it, as mentioned previously), and is typically more useful at higher stream pressures (> 300 PSI). Another concept currently under investigation is the use of ionic liquids for absorption, as CO_2 is highly soluble in ionic liquids, as are most sulfurous compounds. An ionic liquid is a salt that is a liquid at room temperature, and never evaporates (NETL, 2007). Such compounds are showing highly promising results, and may have future applications in both absorption processes and in supported liquid membrane processes (discussed more later on). Figure 2.12 shows an outline of a typical chemical, amine-based, CO_2 capture plant with compression and treatment.

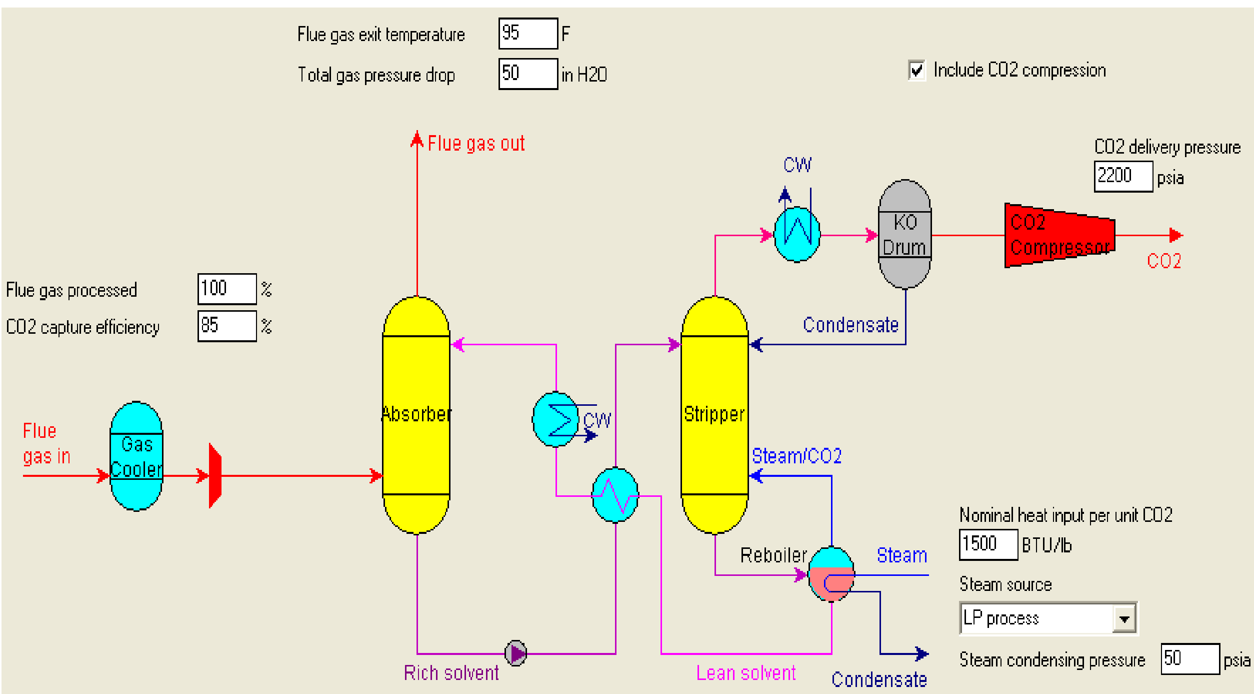

Figure 2.12 Amine-based carbon capture (From Thermoflow's GT Pro software)

For chemical solvents, the absorption reaction occurs at near room temperature. However, the *desorption* reaction (which is the absorption reaction occurring in reverse), requires higher temperatures (around 120-140°C/250-285°F), which can mean large heat losses (3GJ/metric ton of CO_2) for the plant. This extra heat usually comes from the steam cycle, meaning it costs a fair amount of water to use chemical methods. Also bear in mind that the energy provided must account for both the heat necessary to drive the desorption reaction itself (as it is endothermic) and the heat necessary to raise the system's conditions to the required temperature at around 130°C (Tondeur, 2009). Physical solvents however, do not need these increased temperatures to undergo desorption and require significantly less energy (only 1GJ/metric ton CO_2), as they rely solely upon CO_2's solubility, which is a stronger function of pressure than temperature. However, CO_2 will not dissolve in any liquid at ambient temperatures. Thus, the gas stream must be cooled down below ambient, sometimes below freezing by using electric chillers before entering the absorber. The energy cost, while lower than the cost for chemical absorption, is *electrical* energy, as opposed to just heat energy, so there is a direct loss of net power when physical absorption is used. As a result, as mentioned in the section on Acid Gas Removal, the two processes will have a comparable effect on plant efficiency. Which method ends up being better in the end is dependent upon system makeup and environmental conditions.

2.3.5.2.1.2 Adsorption

While amine-based solvents work through a chemical absorption process, it is also possible to use chemical *adsorption* to remove carbon dioxide. To reiterate: while absorption refers to the process by which a substance diffuses through a volume, adsorption is when the same substance accumulates over a surface and forms a film-like layer. Most adsorbent methods are implemented using a sort of fixed bed approach, with the incoming gases moving through a micro-porous solid insert, like a molecular sieve. Common adsorbents are activated carbon, silica gel, and aluminum oxide (Tondeur, 2009). In addition, metallic/organic hybrid crystal-like compounds are in development, and have been shown to have high capture capacities and CO_2 selectivity (NETL, 2007). Usually, in plants that use this type of CCS technology, there will be multiple filters in parallel with each other, and, when one of them reaches its adsorption limit, flow will be redirected to one of the others, and the dirty filter will either be swapped out for a new one or cleaned out and put back in place.

Another option is the so-called "regeneration method," where two filters will be placed in parallel with one another, and a series of valves regulates the flows between them. For each step, one set of filters will be the "adsorber" and the other will be the "regenerator." The adsorber column collects the CO_2 from the main gas stream, while the regenerator releases its captured

CO_2, thus *regenerating* the adsorbent within it, allowing it to have continuous operation. When the adsorber column gets "full," or reaches its capture limit, the valves switch the flow so that the old regenerator becomes an absorber and the old absorber becomes a regenerator, while a "purge gas" (usually hot steam) is forced through the new regenerator, forcing the captured CO_2 out. This process then repeats, forming a time-cycle. In order to maintain this cycle, the conditions inside each column are consistently in a state of flux. As such, adsorption methods like this are constantly transient in nature, so they are very difficult to model using equilibrium methods like those in Thermoflow and ASPEN Plus. Figure 2.13 shows a basic schematic of such a system.

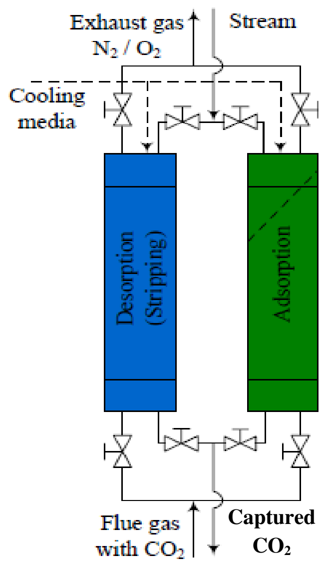

Figure 2.13 Adsorption-Regeneration system (NETL, 2007)

Like the absorption processes, different materials' CO_2 affinities are functions of temperature and pressure. For example, most adsorbents have *higher* carbon capture capacities at *greater pressures* and *lower temperatures.* Likewise, regeneration systems will utilize this concept to increase or decrease the amount of CO_2 that the adsorbent columns can hold, allowing the operators control which column is the adsorber and which one is the regenerator. Systems that use pressure as the primary driving potential are called *Pressure-Swing Adsorption* (PSA) systems, and likewise, those that use temperature are called *Temperature-Swing Adsorption* (TSA) systems. For the most part, PSAs are better than TSAs because manipulating pressure is easier, consumes less energy, and can occur at a faster rate, so there is less time-lag between switching columns. However, TSAs are, on the whole, less expensive to operate and can achieve higher CO_2 purity, meaning fewer other gases will accidentally be captured alongside the CO_2 in a TSA system. This last fact can very well make a difference for an IGCC plant using pre-combustion capture (discussed in detail later on), as that means that more hydrogen and methane will be available to be burned for power later on, meaning more power output and potentially higher efficiency, offsetting the initial energy cost. Finally, PSAs only work on gaseous compounds, but TSAs can be used for liquids as well as gases (ARI, 2011).

There also exists a third type of system called a *Vaccuum-Swing* or *Vaccuum-Pressure-Swing Adsorption* (VSA or VPSA) system, which involves reducing the pressure outside the adsorbent material instead of raising the pressure inside the column. In a way, it is essentially an extension of Pressure-Swing processes, but is much less costly, energy-wise. However, the reduced pressure inside the actual columns means that the adsorbent material will not have as high of an adsorbing capacity as it will for an ordinary PSA system. This means that the actual rate of capture and sequestration will be reduced overall.

2.3.5.2.2 Membrane Processes

It is also possible to remove carbon through a membrane that separates it from the source gas. The membrane is usually made of a thin layer of organic polymers, ceramics, or mineral materials, which selectively allows only certain substances to permeate through it. The contaminants are simply held back and redirected elsewhere after the free gas passes through the membrane. There have been membranes made of carbon nanotubes, ceramics, polymers, and many other substances. They are pressure-driven, much like pressure-swing adsorption processes, and in fact, can compete with adsorption techniques energy-wise for gas mixtures with CO_2 concentrations greater than 20% volume (Tondeur, 2009).

Membrane processes are hardly ever used in industry, mostly because they are not reliable: they have fairly low selectivity, meaning that many other substances aside from CO_2 will be prevented from passing through the membrane, and their mass transfer rates aren't anywhere near those achievable through chemical absorption. The advantages of using a membrane, though, are the facts that they are cheap, have low energy requirements, and are fairly small in size, with many possibilities to further enhance mass transfer. Because of their low selectivity, however, this means that the most effective use of membranes will be in the form of several-staged membrane systems, which can get to be very expensive fairly quickly. Otherwise, a single membrane will be unable to achieve the high purity CO_2 needed for sequestration, and another process will be necessary to refine it later on anyway (Tondeur, 2009).

There are two main types of membranes: *organic* membranes and *inorganic* membranes. Organic membranes are usually made from polymers, such as poly-ethylene-glycol (PEG,) and are used at lower temperatures. Organic polymers are highly *CO_2-selective*, meaning that the CO_2 will pass *through* the membrane *more readily* than other types of membranes. This means that the CO_2 obtained from membranes of this type will need to undergo a greater degree of compression before they can be sequestered compared to other processes. Inorganic membranes, on the other hand, are mostly made of ceramic materials, and, as such, are highly corrosion-resistant, and can withstand high temperatures and pressures. Inorganic ceramics are *H_2-selective*, so the CO_2 will usually be able to retain its higher pressure, saving on energy consumption due to compression. The downside to this is that ceramics *aren't N_2-selective*, so using these types of membranes in post-combustion CCS operations is not advisable. In pre-combustion, however, especially for oxygen-blown gasifier systems, there is relatively minimal N_2 in the syngas, so this is not as much of a concern for these systems (Tondeur, 2009).

While most membranes are solid, liquid membranes using organic enzymes, such as carbonic anhydrase have also been studied. In this manner, the actual "membrane" selectively blocks some of the CO_2, while the enzyme behind it *absorbs* the CO_2 and other contaminants that happen to pass through it. In this way, these membranes are more like a combination of a membrane and the chemical absorption methods discussed previously (NETL, 2007).

2.3.5.2.3 Cryogenic Processes

Last but not least, it has been discussed that cryogenic technology, such as that used in air separation and distillation units, may be applied to CCS. However, for the most part, this method isn't considered to be a viable method for carbon capture since it requires very low temperatures to work properly, and is projected to have prohibitively high costs. However, it may be possible to incorporate this technology with condensing units and liquefaction plants. It may also be possible to retrofit this technology with existing cryogenic ASU plants. On the whole, however, as mentioned before, integrating cryogenic technology with ordinary power plants isn't a viable option with current technology (Tondeur, 2009).

2.3.5.3 CCS Implementation

With the given technologies available for CCS understood, the next question to be answered is "how does one put it to use?" In other words, how is this technology incorporated into the core structure of an existing power plant? The purpose of the next few sections is to highlight these different methods of implementation and to discuss their various strengths and weaknesses.

2.3.5.3.1 Post-Combustion CCS

The most common carbon removal technique for gas turbine and steam turbine cycles is *post-combustion CCS*, wherein the CO_2 produced is removed from the flue gas and stored elsewhere (usually pumped underground). However, even though all CCS processes hurt the plant's thermal efficiency, post-combustion CCS is especially detrimental to that end, usually causing an efficiency drop of at least 7-8 percentage points in IGCC. For one, because the gases are cleaned *after* combustion occurs, the mixture is usually at atmospheric pressure, which means the density is very low, requiring larger cleaning units and more solvent/adsorbent in the case of sorption processes. This and the fact that extra N_2 from the air was added during the combustion process means that the total gas volume will be very large, making it more difficult to process. Secondly, because the flue gases include SO_x and NO_x, physical methods should not be used due to the shear amount of pretreatment the gases must go through to reach the required temperature conditions, and because of the fact that SO_x and NO_x cannot be cleaned out through physical means.

For PC plants, another option for CCS is *oxy-fuel combustion CCS,* wherein the combustion process is performed using pure oxygen from an air-separation unit. Then, the flue gas (largely CO_2 and H_2O) is re-circulated back into the combustion chamber to cool off the inside of the combustor to prevent overheating (Tondeur, 2009). After the water content is condensed out of the stream, the result is a continuous stream of almost pure CO_2, which can then be sequestered as a whole, which is why plants that use this method are often called "zero emission" plants, because they do not send the flue gas out through a stack. Rather, all potential emissions are stored and disposed of. However, sometimes some CO_2, and other contaminants can end up in the condensed water stream, and as such, this water must be treated properly in order for the plant to be a true “zero emission” plant. Another advantage of this alternative is that more inert gases usually found in flue gas may be left untreated, so the cleaning process tends to be more energy conservative than other forms of CCS.

2.3.5.3.2 Pre-Combustion CCS

Another method of incorporating CCS into commercial plants is *pre-combustion CCS*, a process unique to IGCC, which allows for the removal of carbon dioxide during the gas cleanup phase of the cycle instead of having to wait until after the gas is already burned by the GT combustor. Pre-combustion CCS is not possible for ordinary PC plants or natural gas plants, since the CO_2 hasn't been produced before combustion. Pre-combustion CCS is significantly less costly than post-combustion CCS, and does not impact the plant efficiency as much as post-combustion CCS does. However, as previously stated, it isn’t compatible with most power plants, and the process itself is not as well-understood as post-combustion CCS is.

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

$$CO + H_2O \leftrightarrow CO_2 + H_2$$

By adding significant amounts of water to the stream, the reaction shifts, as described previously, towards the right, thus converting most of the CO to CO_2. There are two points in the cycle that this can occur: before the *acid gas removal* (AGR) process or after it. When this occurs before AGR, it is called a “sour-shift” reaction, and either “clean-shift” or “sweet-shift” when performed afterwards.

2.3.5.3.2.1 Sour-Shift Pre-Combustion CCS

Sour-shift means that the "water-gas shift" reaction occurs *before* the Acid Gas Removal (AGR) stage of the gas cleanup system. This is an important consideration, because when sulfur is in the gas stream alongside the CO_2, and AGR and CCS absorption techniques are very similar processes, it means that the plant can perform AGR and CCS *at the same time*, using the same equipment. This means that sour-shift CCS is cheaper to implement and can sometimes be retro-fitted onto IGCC plants that have amine-based absorption AGR systems installed. Sour shift can also be used at the same time as COS hydrolysis, using the same water supply, since the COS hydrolysis reaction and water-shift are very similar. All in all, this makes sour-shift CCS a lot cheaper than sweet-shift CCS. The downside to this is the fact that sour shift requires significant amounts of water to shift, since many processes must occur all at once. Second, there is a lot of waste heat, since the water-shift reaction is *exothermic* in the direction of CO_2, the gas stream gets hot, and must be cooled immediately, especially before the AGR column. All in all, this translates to the fact that more cooling is needed to achieve the necessary temperature range to even perform acid gas removal than that of the same plant without any CO-shift or CCS. In this manner, sour shift performs very well in systems that use a *slurry*-fed reactor (Li and Wang, 2009). Since slurries add more water to the syngas stream by default, sour shift becomes more practical in this case, because a significant portion of the water needed for shifting is already present in the gas, and only needs additional catalyst to complete the reaction effectively.

2.3.5.3.2.2 Sweet-Shift Pre-Combustion CCS

Sweet-shift means that the CO-shift reaction occurs *after* Acid Gas Removal (AGR) in the cleanup process. Because of this, separate shift-reactors and CCS plants must be purchased and added to the cleanup system. However, because the process is done *after* AGR, the CCS process is more efficient and uses less water and energy (Li and Wang, 2009). In addition, the exothermic nature of the CO-shift process allows the resulting clean syngas to have a higher exit temperature from the cleanup system itself, since CCS is performed *last* in this portion of the cycle. Therefore, there is less net heat loss for sweet-shift CCS and less energy input needed for the syngas entering the gas turbine.

Sweet-shift CCS is more effective than sour-shift for systems with *dry-fed* reactors, due to the fact that it would require less additional input water from the steam cycle, and any water supplied to it from a slurry-fed gasifier would simply be drained out by the coolers within the cleanup system or consumed by the AGR process, having no effect on the Water-Shift reaction. In summary, sweet-shift is the better option for dry-fed systems, while sour-shift is better for slurry-fed systems, but sweet-shift is always more expensive than sour-shift. When taken in combination with the type of feeding, which one is better will generally depend on system layout, fuel type, and individual gasifier specifications.

2.4 The Power Block

2.4.1 The Gas Turbine System

Gas turbines are *internal combustion engines*, meaning that the energy needed to raise the working fluid's temperature is obtained from *within* the engine itself, most often through combustion of a fuel. In IGCC, this fuel is syngas, while for a standard Brayton cycle, it is usually natural gas, kerosene, diesel, or other gaseous/vaporizable hydrocarbon. There are many commercial gas turbine models available made by a large variety of companies. Gas turbines also have their uses not just in the power industry, but in the fields of jet propulsion and surface vehicle propulsion as well.

The GT block in an IGCC system is identical to that of a typical Brayton cycle: there is a compressor, which takes in and compresses atmospheric air to the desired pressure, a combustor that burns the fuel using the air obtained from the compressor as the combustion agent, and the actual turbine, which extracts work from the compressed air by expanding it. The one exception that the exhaust from the gas turbine isn't simply ejected into the atmosphere: it is the source of heat that drives the steam cycle. The GT cycle is often called the "top" cycle in all combined cycle applications, because it is the cycle that actually uses the high-grade energy in the syngas, and its exhausted gases provide the energy necessary to run the steam "bottom" cycle. Figure 2.14 is a flowchart of a standard open-air Brayton cycle.

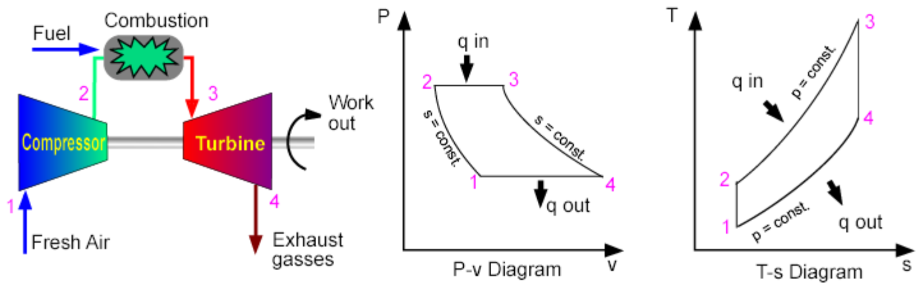

Figure 2.14 Standard, open-air Brayton Cycle (Wikipedia, 2011)

2.4.2 The Steam Cycle

As mentioned briefly in chapter one, the majority of all power generated in the world is acquired from *steam* engines. Steam power has been a part of human society for at least 300 years, and still plays a major part in our everyday lives. Steam turbines are *external combustion engines,* meaning that some fuel must be burned *separately* from the working fluid to provide heat to it. Because the steam cycle operates at typically lower pressures than gas turbine cycles, STs typically don't produce as much power as GTs. However, more often than not, the Rankine Cycle is a more efficient process than the Brayton Cycle. Combined Cycle technology was a way to combine the two together to achieve the highest possible efficiency. In fact, the natural gas combined cycle (NGCC) is the most efficient thermal cycle that is considered commercially viable today, easily reaching efficiencies above 55%.

For a combined cycle, IGCC or NGCC, the steam cycle is different than it is for a standard Rankine cycle. For instance, there is no boiler, meaning there is no fuel being burned: all the energy given to the steam in a combined cycle comes from the GT exhaust gases, through a device called the Heat Recovery Steam Generator (HRSG). However, all steam turbine cycles require the use of a cooling system of some sort in order to complete the cycle (or "close the loop.") This can be accomplished in one of two ways: with a water-based condenser ("condensing" steam cycle) or with an air-cooled heat exchanger ("non-condensing" steam cycle). Condensing turbine cycles are more expensive due to the much larger, more complex equipment needed, but also tend to be more efficient at cooling the steam down to the state required by the pumps that drive the cycle. However, air-cooled systems are much cheaper, and they also open up another option: using the heated air obtained from the water-cooling process to run another system, such as a heater. The heat released in this manner can thus be thought of as *useful heat output*. This means that air-cooled steam systems tend to have very high efficiencies for Combined Heat and Power (CHP) plants. However, for plants that are not intended to make use of this heat (such as systems constructed in warmer climates with large water supplies), condensing turbine systems nearly always have the higher efficiency.

2.4.2.1 The Heat Recovery Steam Generator

The main physical difference between the Rankine Cycle and the steam bottom cycle in IGCC is the existence of a device called a Heat Recovery Steam Generator (HRSG for short.) An HRSG is essentially a very large, complex heat exchanger. It is designed specifically to recover the waste heat from the exhaust of the GT to generate steam. In IGCC, this steam is used to run a steam turbine as a combined cycle, but it may also be used to provide *process steam* for industrial usages such as for heating (co-generation) or running auxiliaries, such as absorption cooling, drying, driving compressors, driving tools, and sanitization (for poly-generation). Figure 2.15 is a schematic of an HRSG with small velocity profile diagrams of the main hot gas flow.

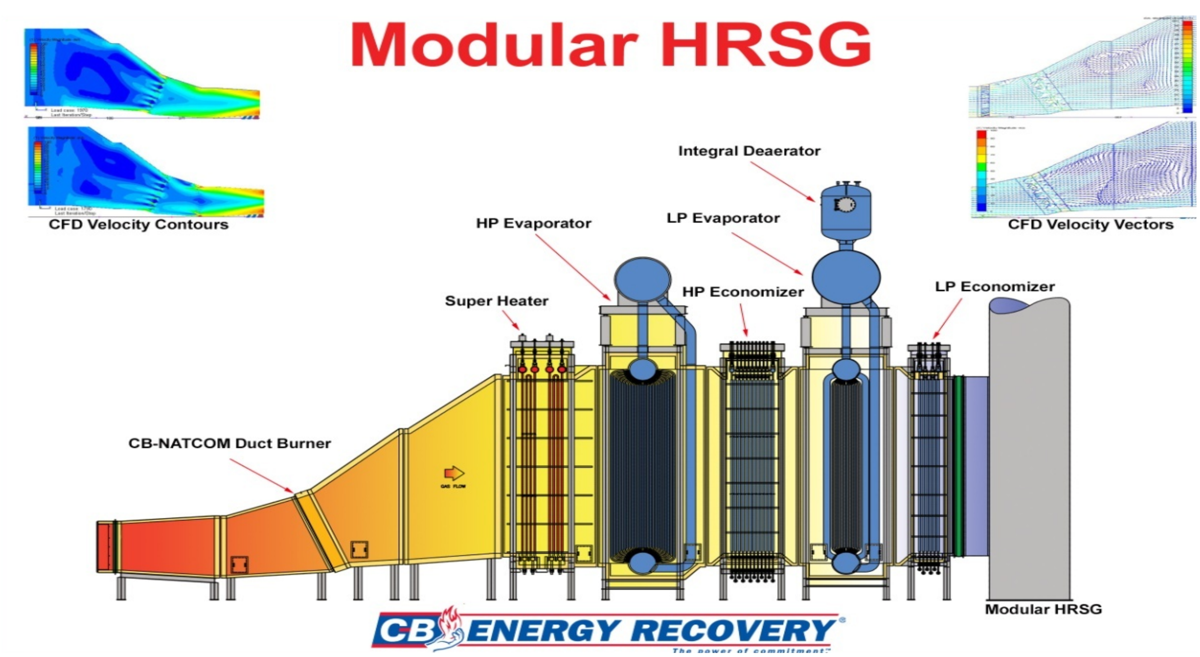

Figure 2.15 Heat Recovery Steam Generator, horizontal design (Wikipedia, 2011)

HRSGs consist of four main components: an evaporator, a superheater, a pre-heater, and an economizer. HRSGs are characterized by their basic layout: horizontal or vertical. A horizontal HRSG has the GT exhaust gas flowing horizontally over vertical steam pipes, while a vertical one has them flowing vertically over horizontal pipes. Second, they are also classified by the number of *pressure streams* that are used. The largest number of pressure streams used in commercial HRSGs is three: conveniently called the High Pressure (HP), Intermediate Pressure (IP), and Low Pressure (LP) streams. Each stream corresponds to a steam pressure *drum* or *barrel* that stores the water as a part of the evaporator, while the water turns into saturated steam (Nooter-Erikson, 2011). Preheated water will be pumped into each barrel after leaving the de-aerator, if there is one (explained in the next section.) After leaving each barrel, the individual steam streams will enter a *superheater* section, where the saturated steam will be provided with more energy from the hot gases running through the main section. After this section, an economizer section (usually one for each stream) will be used to further supply heat to the steam and reduce energy consumption. For most combined cycles, each pressure stream will be responsible for performing a different function: higher pressure streams will run STs or other power- or heat-producing devices, and lower pressure streams will be used to operate plant auxiliaries (such as AGR and CCS).

HRSGs can operate using either natural or forced convection. Forced convection designs use faster flow velocities, and of course, have higher heat transfer rates because of this. However, they are more costly and require more work input to operate the pumps than natural convection HRSGs.

Finally, there is a unique, specialized type of HRSG called a "Once-Through" Steam Generator, based on a design by German engineer, Mark Benson. What makes this design unique is the fact that there are no drums in the evaporator. All of the steam-carrying equipment is replaced by thin-walled pipes, which follow a continuous, un-segmented path through the device. This grants the Once-Through design a degree of flexibility compared to a traditional convection HRSG, allowing each section to be adjusted in size more freely based on the heat input from the GT, in addition to being easier to operate and cheaper.

2.4.2.2 The Duct Burner

As mentioned in chapter one, the easiest and most direct way to increase the efficiency of a standard Rankine Cycle is to raise the inlet conditions, particularly the pressure. For systems with fixed/known pressures, the inlet temperature becomes an area of primary importance. In a combined cycle, sometimes this is not achievable with the system specs given. For example, if the GT exhaust gases leave at 1200°F, it is probably not possible to push enough of the waste heat into the HP stream to get an ST inlet temperature of 1180°F, or at best, very difficult to do so. It is not feasible to raise the GT outlet temperature for this, because that would detract from the GT's power output, which would greatly affect the cycle efficiency.

In cases like this, a special device called a *duct burner* is inserted directly into the exhaust inlet port of the HRSG. A duct burner is simply a reactor that burns some kind of fuel and releases the heat from the fuel into the exhaust gas stream, raising its temperature and allowing it to transfer more heat into all of the steam pressure streams. Most duct burners will use a secondary fuel, such as natural gas, to perform this function, but it is also possible to burn some of the GT's fuel, since taking some of the mass flow from the GT will affect it less adversely than directly reducing its specific power output. Duct burner designs themselves are very similar to other reactors/combustion chambers, and tend to only differ by the fact that they must be outfitted to be attached to the inside of an HRSG economizer/superheater.

2.4.2.3 The Deaerator

Most water sources are not pure, and, in particular, raising the pressure of water streams high enough will result in many gases, particularly oxygen, becoming dissolved in the liquid, which can cause problems later on when the water is evaporated and run through a steam turbine. Under circumstances like this, it is necessary to remove these gases from the water so that it may be used more efficiently in the cycle. To perform this function, most plants will make use of a device appropriately called a deaerator. Deaerators come in 2 main varieties: *Tray-type* and *Spray-type*, both shown in Fig. 2.16.

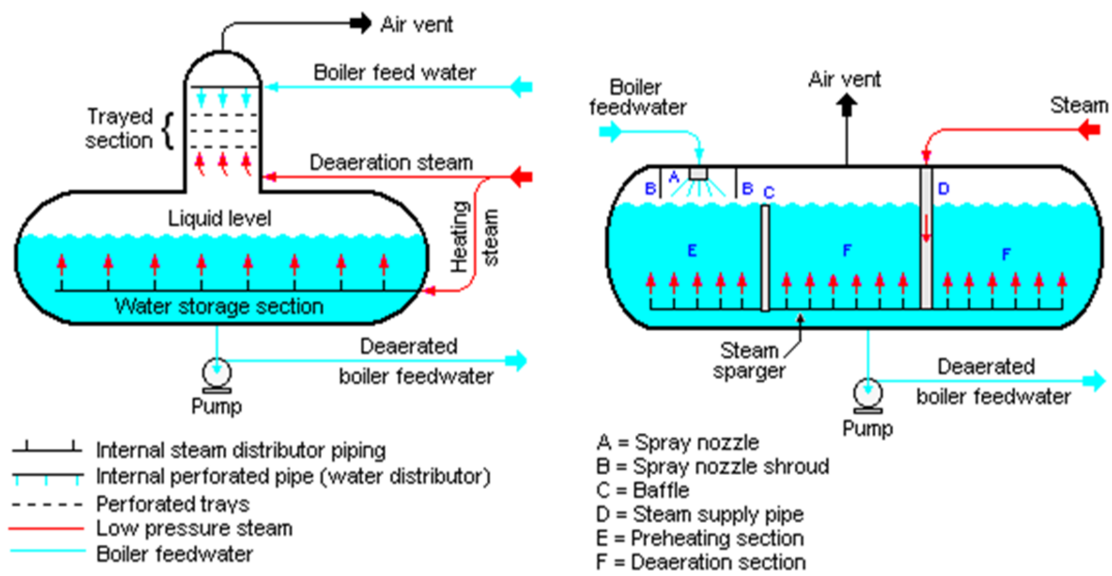

Figure 2.16 Deaerators, tray-type (left) and spray-type (right) (Wikipedia, 2011)

For tray-type deaerators, the boiler feed water is injected into the top through a perforated pipe and flows over a series of perforated "trays," hence the name. Lower pressure steam from outside the deaerator is then injected into the deaerator and sent upward through the deaeration section, mixing with the feed water as both streams pass over the trays. Since gases mix more readily with other gases than they do with liquids, the steam effectively "strips" the water of the dissolved gaseous components, and ferries them out through the top of the deaerator, while the deaerated water falls to the bottom of the lower tank and is pumped away after being heated by more low-pressure steam.

Spray-type deaerators, on the other hand, have the heating and deaeration sections integrated together, separated only by a thick baffle that rises just above the liquid level. In this type of deaerator, the feed water is "sprayed" into the top corner of the device and is heated by passing high-grade steam. This is to raise the water to its saturation temperature, to better facilitate deaeration. The higher temperature water then flows forward, into the deaeration section where more steam is injected, undergoing the deaeration (or "degasification") process. The hot steam atomizes the water by blasting it at very high flow rates, mechanically "shaking" the last of the dissolved gases from the liquid medium (Industrial Steam, 2011). The undissolved gases exit through an air vent at the top of the device, just like in the tray-type layout, and the water is then pumped away to whatever process it is needed for.

In general, both types of deaerators are capable of producing water at similar levels of purity and can be designed to operate at virtually any pressure, temperature, or flow-rate. However, tray-types are far more reliable than spray-types, due to the fact that the mechanism involved in the deaeration section is independent of flow rate. This means that tray-types have larger operating ranges and are less susceptible to spontaneous changes in design conditions. However, in unstable areas, such as aboard marine vessels, spray-types are easier to maintain and operate, because the perforated trays in a tray-type deaerator will not be able to be kept level on such vessels, severely inhibiting deaeration (Ketten, 1986).

CHAPTER THREE

CASE ORGANIZATION AND DESIGN METHODS

3.1 General Information

To reiterate, the primary objective of this study is to improve upon the basic IGCC system design by (1) reducing the emissions of such plants by introducing biomass into the coal feedstock, raising the efficiency, and implementing CCS and (2) reducing the capital and operating costs. This is achieved by performing a parametric study upon a baseline plant design, which will be described in detail throughout this chapter. The main parameters studied in the primary cases are: adding biomass to the plant's original coal feedstock (reduce emissions and hopefully raise efficiency), replace the existing Rankine steam cycle with a supercritical Rankine steam cycle (raise efficiency, lower costs), and add a carbon capture system (reduce emissions).

In addition, a series of special cases were examined, using other parameters, including dry-fed vs. slurry-fed systems, air-blown vs. oxygen-blown gasifiers, radiant cooling vs. quench cooling, sour shift vs. sweet shift, etc.

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. Each portion of the cycle is examined individually (such as "gasification block," "GT selection," "GT Inputs," "HRSG design," "Water Circuits," etc.), and inputs are entered sequentially in any order. When there is enough information to construct the plant, the user may run the simulation and the program will compute the result and compile all of the information in both textual and graphical forms.

Other programs available in the suite include: SteamPro®, which works just like GTPro, except with purely Rankine steam cycles, GTMaster® and SteamMaster®, which are used to evaluate plant designs from other programs using off-design conditions, and Thermoflex®, which is a fully-flexible thermal system design program similar to ASPEN® Plus. Thermoflex can be used to physically construct just about any system from its core components, provided that the needed components have been added to the database. This means that, unlike ASPEN, where each block is purely a tool for calculation and must be programmed separately to perform its intended function, Thermoflex's database contains pre-simulated blocks that represent real processes and devices. This makes for a much more user-friendly experience. Due to the large number of cases studied and the complexity of the changes made for each main case, all plants examined in this simulation were generated by GTPro. As such, all figures depicting plant layouts and components also originate from GTPro.

3.1.1 Overall plant setup

Figure 3.1 shows the general layout of the baseline case (Case A1a). It consists of a single 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 with steam injection in the combustor, 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. Though, not shown in Fig. 3.1, all condensed water extracted from the raw syngas during cooling is transported directly to the steam system via the deaerator (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.

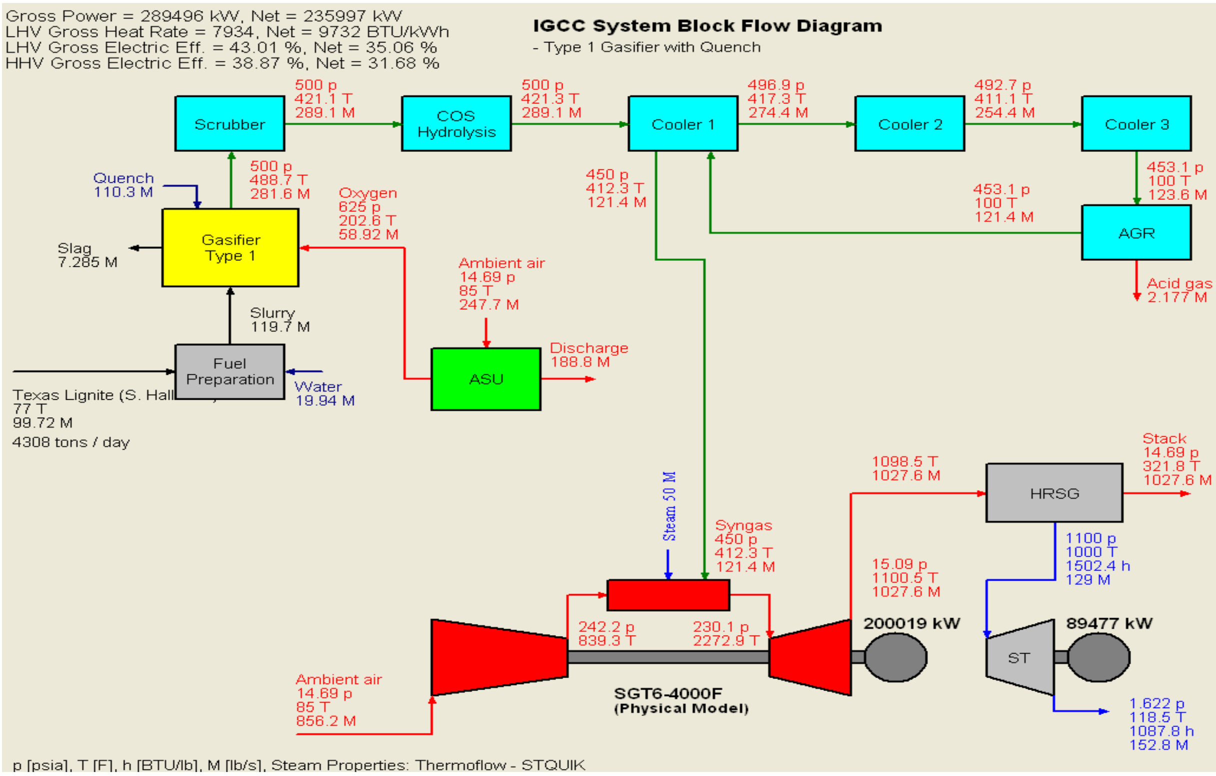

Figure 3.1 General plant layout (Case A1a)

3.1.2 Plant specifics

3.1.2.1 Plant location: New Orleans, Louisiana

Louisiana is one of the largest energy providers in the entire United States, coming in third behind Texas and Wyoming, ranking first in oil production and second in natural gas production (EIA, 2009 & LDNR, 2011). Energy is very important for the people of Louisiana; for a state that is, as of 2011, still recovering from the effects of natural disasters like Hurricane Katrina. As such, it was decided that, were a plant of this type to be built, Southern Louisiana, particularly in the vicinity of New Orleans would be an appropriate place to consider. In addition, the fact that the Pelican State is home to an incredibly large quantity of biomass makes this decision even more prudent. For one, 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. About 16% of all sugar produced in the U.S. comes from Louisiana farms and factories (Legendre et. al, 2000), and around 16 million tons of raw sugarcane is harvested per year (Day, 2011). In addition, Louisiana is one of the largest rice producers in the U.S. as well, with yields of between 500,000 and 600,000 tons of rice produced every year (Sanders, 2000). In addition, Northern Louisiana is home to a large amount of woodlands, which allows for the use of wood chips and bark as fuels. Finally, Louisiana also has fairly large claims in other food products such as soybeans, sweet potatoes, and corn.

The plant was assumed to be placed 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 realistic 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, 70-75 degrees with 90-100% 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 country: 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, the largest biomass crops produced in Louisiana are sugar cane and rice. Since rice is already a major contributor to the production of ethanol in Louisiana (Sanders, 2000), sugar cane was chosen to be the main source of biomass. However, to avoid using the crop itself and potentially impacting Louisiana's sugar production, only the bagasse, the waste product of the refinement process, was assumed to be used as the actual feedstock. The fuel data, including ultimate analyses, for both lignite and bagasse can be seen in Table 3.1.

Table 3.1 Fuel Data (Source(s): GTPRo® fuel library, EIA, 2009, and Day, 2011)

Fuel	**Texas Lignite (South Hallsville)**	**Sugarcane Bagasse (dry)**
C (wt%)	41.3	43.59
H_2 (wt%)	3.053	5.26
N_2 (wt%)	0.623	0.14
S (wt%)	0.7476	0.04
O_2 (wt%)	10.09	38.39
Cl_2 (wt%)	0	0
H_2O (wt%)	37.7	10.39
Ash (wt%)	6.479	2.19
LHV (Btu/lb)	6398	6714
Price ($/ton)	20.00	65.00

3.1.2.2 Feedstock preparation

Before the two fuels can be used in gasification, they must be prepared in a way that makes them able to be fed easily and cleanly into the gasifier itself. For coal, this is done by grinding and some drying. Since the gasifier is slurry-fed, the average particle size can be slightly larger than it would be in a dry-fed system, which can mean less work needed if not an easier operating procedure. In total, the amount of processing work needed was assumed to be 40kW-hrs per ton, based on the processing powers of commercial coal grinders. Biomass, however, as discussed extensively in Chapter 1, requires much more extensive pretreatment. Sugarcane bagasse is very tough, and becomes very sticky at high temperatures, making it not suitable for ordinary grinding (Siemens-Westinghouse Corp., 1999). It must therefore go through some other process before it can be added to the coal slurry. For this reason, it was decided to include a torrefaction plant in the pretreatment of the sugarcane bagasse. In total, the energy requirement for the grinding and torrefaction of biomass was assumed to be 200kW-hrs per ton. Finally, the total energy cost of pre-treating any blend was assumed to be a linear combination of the two numbers, based on the % biomass ratio (BMR), as shown in equation 1:

$$Total\ Fuel\ Work\ Input = 200 * BMR + 40 * (1 - BMR) \quad (1)$$

3.1.2.3 Gasifier design

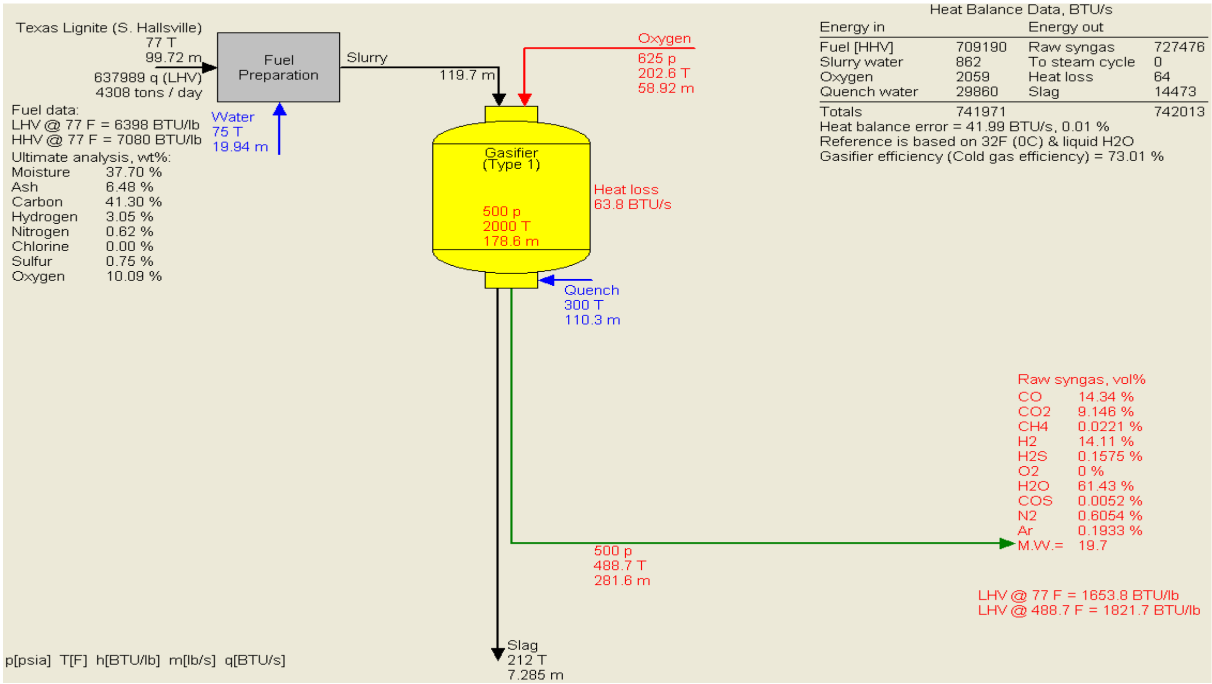

Figure 3.2 Gasifier design (Case A1a)

Figure 3.2 shows the basic design of the gasifier chosen for each of the plants in the main cases. It is modeled after GE's entrained-flow gasifier, being slurry-fed and oxygen-blown. The gasifier is fed from the top, with the slurried feedstock and oxygen streams traveling together in a co-current arrangement. The slurry is 35% water by weight, and is created at around 77°F (25°C), while the gasifier itself operates at a pressure of 500PSI and a temperature of 2000°F (1093°C). The oxygen is provided by an ASU, which, again, provides 95% pure oxygen with an operating pressure of 147 PSI.

Unlike the typical GE-model gasifier, the one chosen for this study is quench-cooled with water at 300°F (149°C) until a syngas relative humidity of 50% is reached. This leads to a final syngas output temperature of less than 500°F (260°C). All slag produced by the gasifier is assumed to be collected at the bottom, at a final temperature of 212°F (100°C).

3.1.2.4 Gas Cleanup system

The syngas cleanup system for this series of plants, shown in Fig. 3.3, consists of a particulate scrubber for removing ash, a COS-hydrolysis reactor, a series of heat exchangers for cooling, and finally an acid gas removal (AGR) plant.

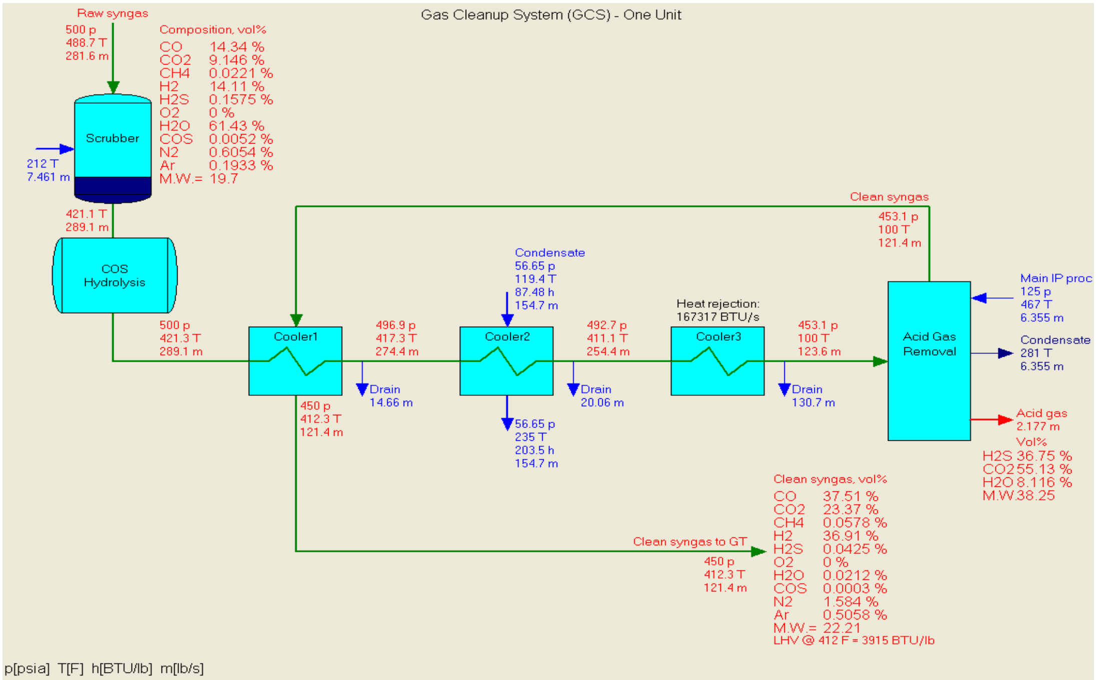

Figure 3.3 Gas cleanup system design (Case A1a)

The particulate scrubber takes the raw syngas and uses a spray jet of additional water that agglomerates all the stray ash particles, removing them as a "black ash" slurry. The cleaned syngas then undergoes COS-hydrolysis, discussed in Chapter 2, which shifts the COS to H_2S, so it can be removed by the AGR plant, a diagram of which can be seen in Fig. 3.4.

The AGR system uses the physical absorption method. Even though physical solvents are more expensive than chemical ones, the physical system will last longer and ultimately cost less due to the long-lasting nature of the solvent and lower maintenance on the system overall. The Absorber expels cooled syngas at 100°F (37.8°C), with a specified maximum output temperature of 120°F (48.9°C). The reboiler takes low-grade steam from the IP stream of the HRSG, and returns the waste water as condensate to the deaerator. It provides roughly 9000 Btu per pound of H_2S that travels through the stripper. Because amine-based solvents are also capable of absorbing CO_2, some CO_2 is in fact "captured" by the stripper, and the exit stream from the KO drum at the end contains roughly 1.5 moles of CO_2 for every 1 mole of H_2S. Finally, the cleaned syngas leaving the absorber returns to the first cooler in the cleanup system, where it will recover some of the heat lost from the earlier cooling, as shown in Figs. 3.1 and 3.2.

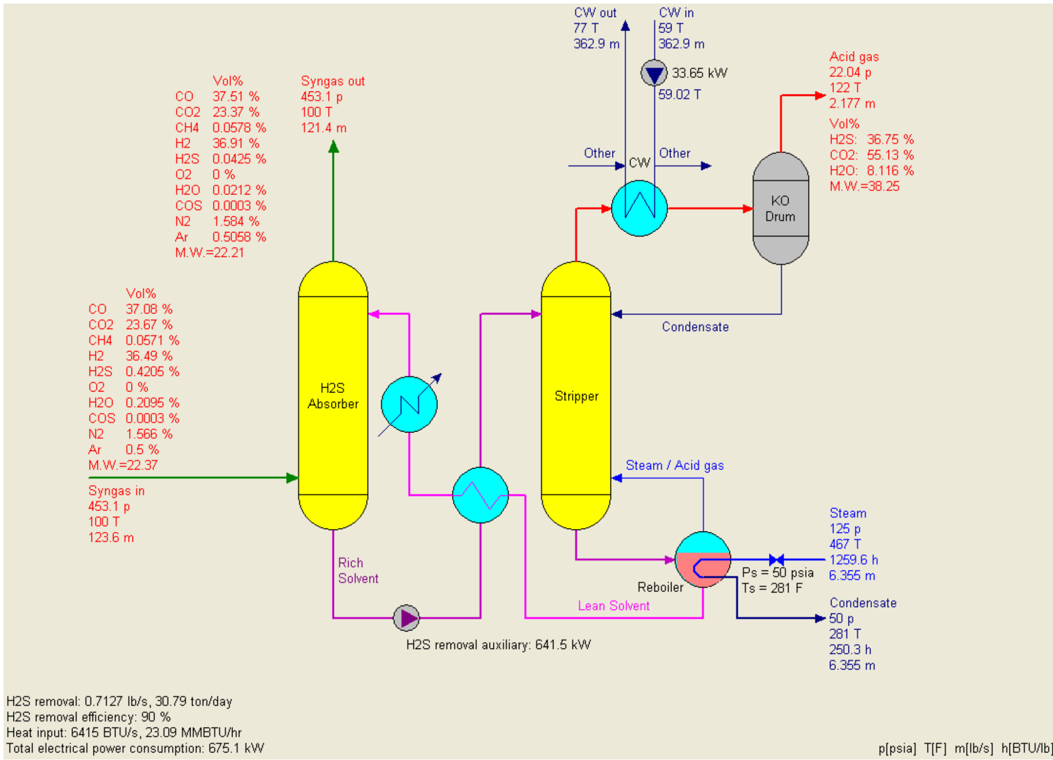

Figure 3.4 Acid gas removal system (Case A1a)

3.1.2.5 Gas Turbine specifications

The gas turbine used for all cases is the Siemens SGT6-4000F model, which makes use of a low NO_x, dry combustor and was recently used by Siemens to use syngas as the combustor fuel (Gadde, 2006). A total draft loss of 4 inches of H_2O was assumed in the compressor, with a 5% pressure loss in the combustor itself: a grand total of 11 inches of H_2O across the whole GT system. The turbine inlet pressure (TIP) is fixed at around 230 PSI, and the turbine inlet and exit temperatures (TIT and TET) are also roughly fixed at 2270°F (1,243°C) and 1100°F (593°C), respectively.

The total turbine gross output power is also fixed, roughly at 200,018 kW. The only mechanical modification to the GT itself is an increase in the turbine inlet nozzle cross-sectional area (assumed to be 6% larger than that of the base model.) Finally, the GT combustor also makes use of a steam injection port, taking in 50lbs/sec of external steam at 650 PSI and 550°F (287.8°C) to help reduce NO_x. A flowchart of the GT is shown in Fig. 3.5.

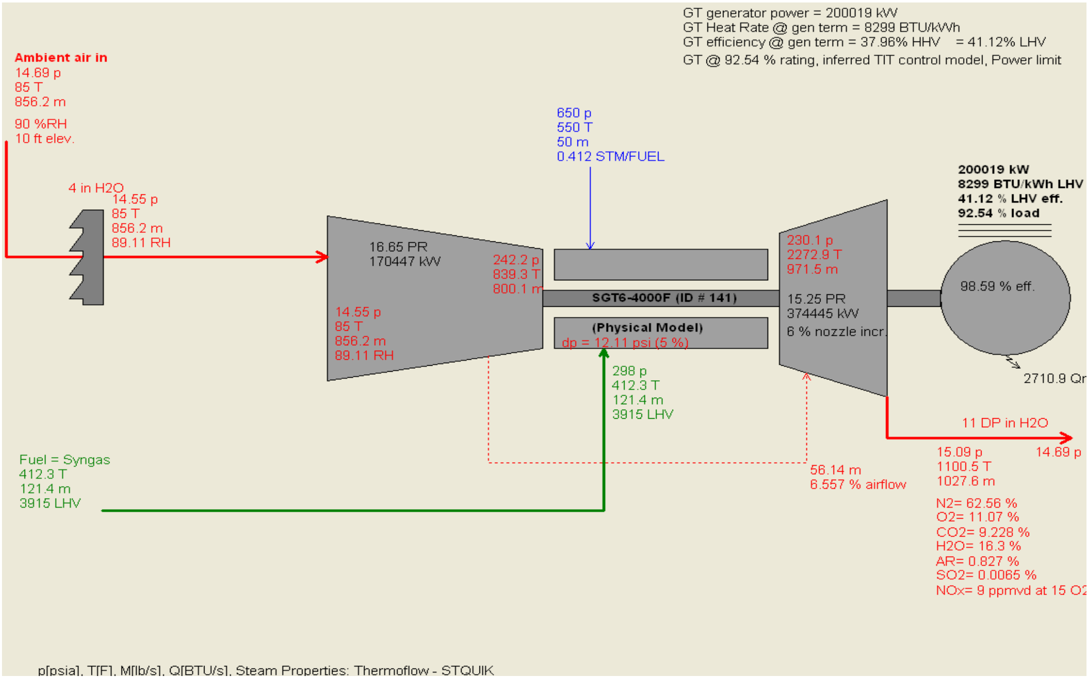

Figure 3.5 Gas turbine (Case A1a)

3.1.2.6 Steam Cycle general layout

A flowchart of the steam system (with the GT shown at the top) can be seen in Fig. 3.6. The HRSG connections and heat exchanger locations are consistent across all cases, as are the ST layout (highlighted in Fig. 3.7), the condenser cooling system type, and the deaerator.

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 deaerator is assumed to be a tray-type, and is the repository of all return water from all other processes, including: condensate from the gas cleanup system, return water from the ST condenser, and makeup water. The deaerator also provides the de-superheating water for processes that require lower-grade steam or water, and acts as the origin for both the IP and HP streams.

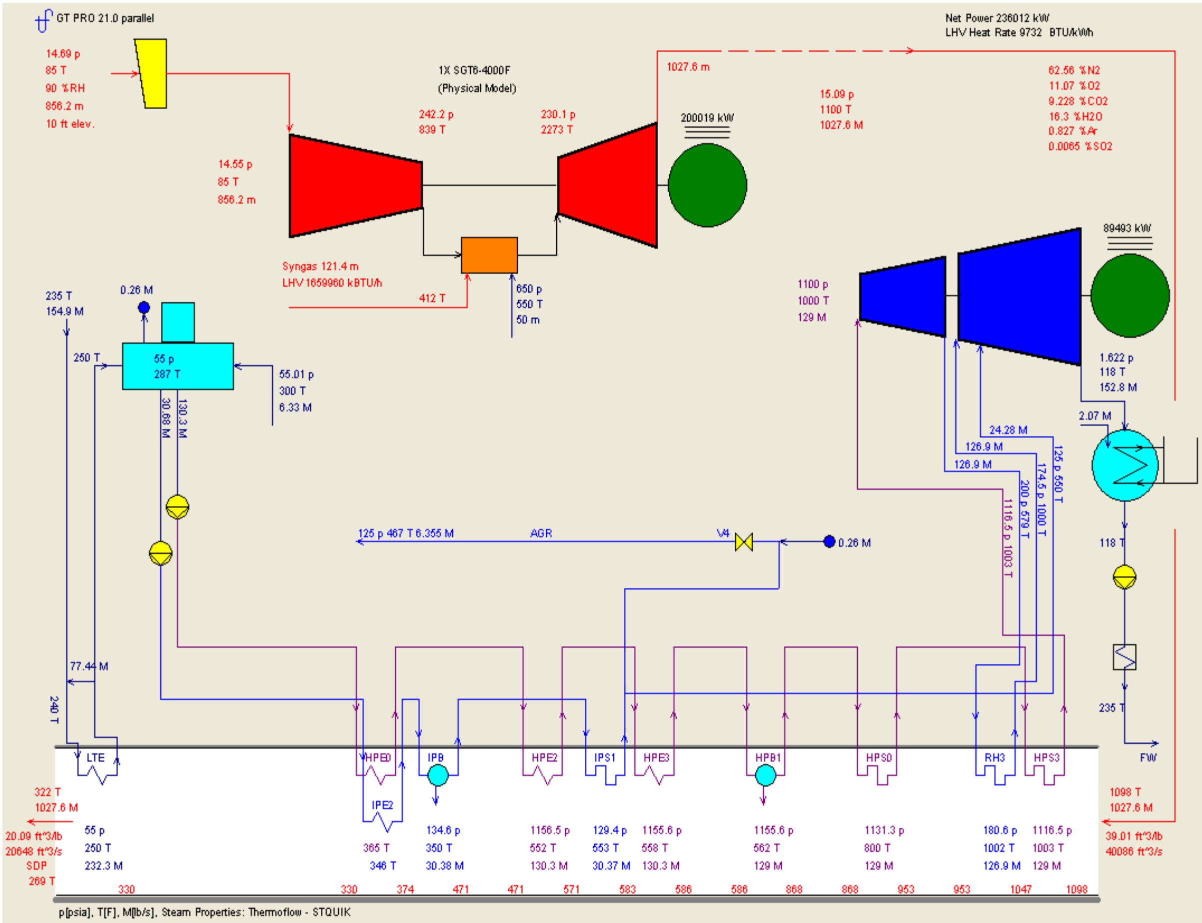

Figure 3.6 Power block flowchart with an emphasis on steam cycle (Case A1a)

The steam turbine itself, shown in greater detail in Fig. 3.7, consists of 2 casings, for a total of 3 main stages. The temperatures and pressures vary according to which case is being studied, but, in all cases, the ST uses a reheat system, sending a low pressure stream back to the HRSG to recover some of the heat energy from the GT exhaust before expanding again through the second casing. 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 (especially CCS and AGR). Finally, the controls were set so that the ST isentropic efficiency could be kept as high as possible in order to maintain the same TIT and TIP.

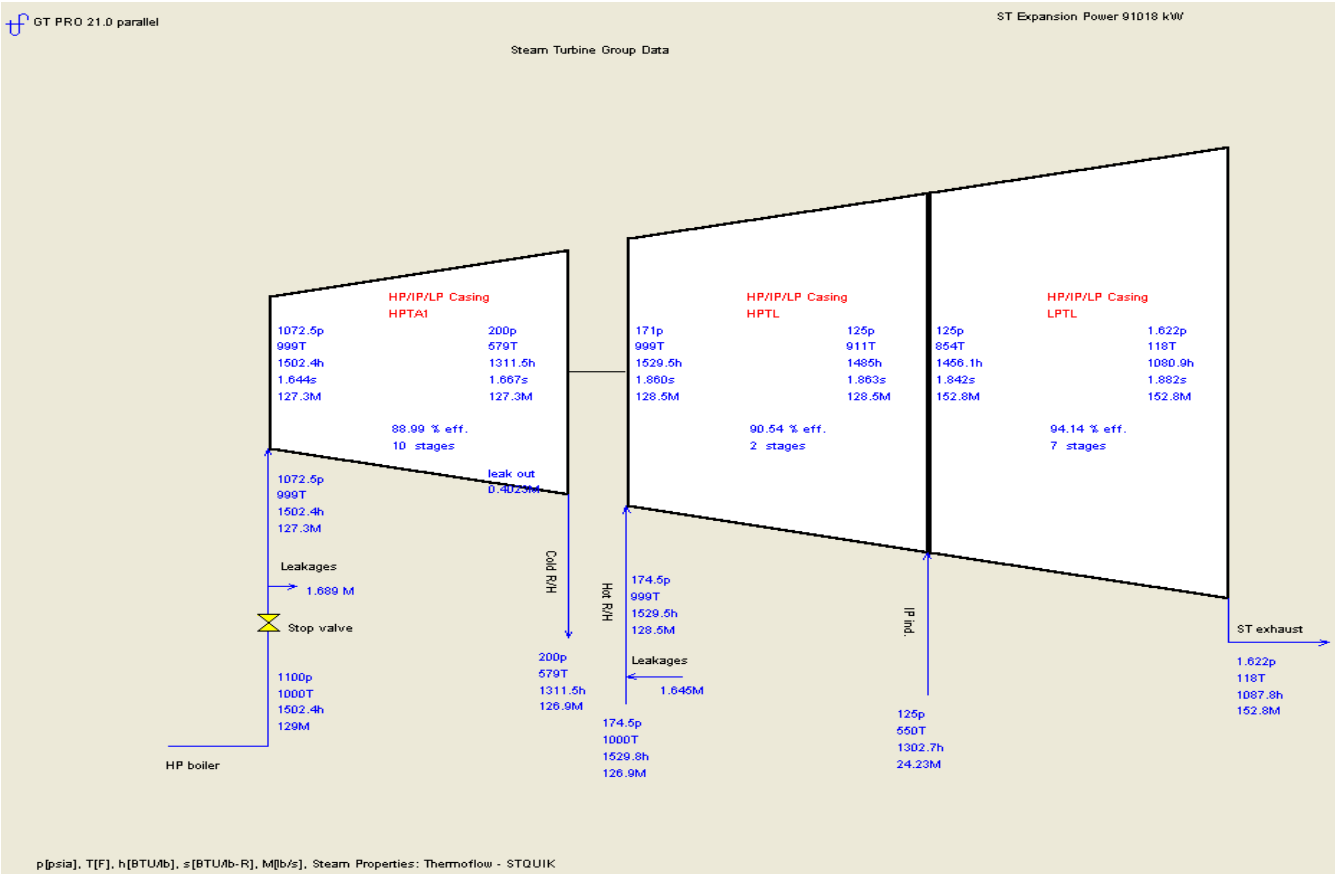

Figure 3.7 Steam turbine schematic (Case A1a)

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 controls on the cooling tower are strictly enforced, and all parameters are fixed in each and every case. A diagram of the cooling system, with all of these measurements can be seen in Fig. 3.8.

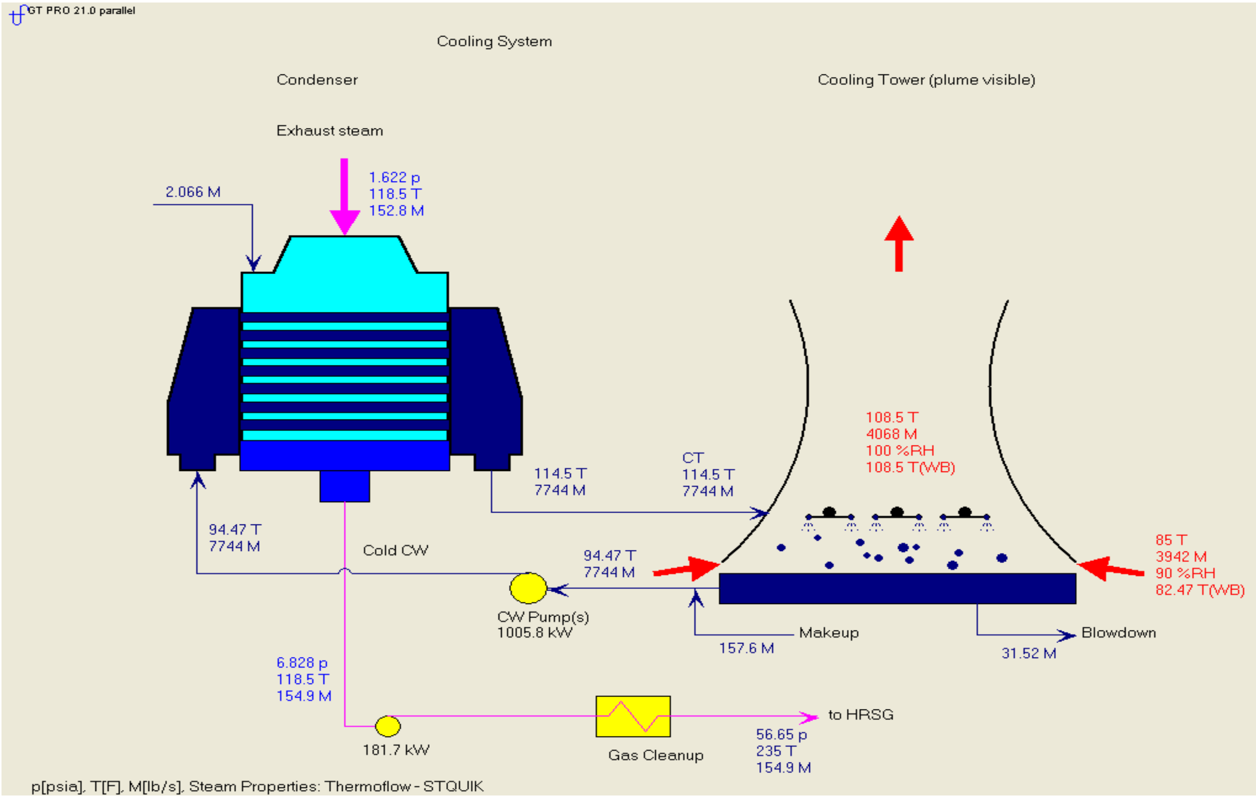

Figure 3.8 Cooling system (Natural draft cooling tower)

3.1.3 Economic considerations

3.1.3.1 Economic assumptions

For economic analysis, only costs were considered in analyzing the different cases. Revenues were not considered as a part of the analysis due to complexity and potential market volatility. As such, no potential profit margins for heat or fuel export were considered as a part of the analysis, but several cases do make note of the potential for such things. In addition, inflation and price scaling were also not considered for similar reasons. However, interest payments were considered, and a flat interest rate of 10% was imposed upon all costs and margins. For the plant capital cost, 30% of this was assumed to have been taken on *equity*, meaning 30% of the total cost was not covered by a loan, and was paid for out of the owner's personal funds. The tax rate was assumed to be 35%, and 75% of all costs were assumed to be eligible for depreciation (straight-line method.) Finally, the total plant life was taken to be 30 years in duration, and the plant is to operate at around 8,000 hours (91.3% capacity) per year.

3.1.3.2 Price settings and costs

Beginning with fuel choice, lignite is cheap, and, according to the EIA's report (2009), lignite from Texas costs approximately \$19.00/ton. While sugarcane itself is cheap at only \$30.00/ton (U.S. Dept. of Agriculture, 2011), bagasse is only a portion of the sugarcane. The key assumption of these plants' operations is that only the bagasse will be purchased from the producer: 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. However, note that even on a purely linear basis, the bagasse produced will of course be priced lower than the price of the entire plant, but the price *per ton* is under no such restrictions. For example, since 200 lbs is 1/10th of a short ton, a purely linear scale suggests that bagasse from 1 ton of cane would be \$3.00. This is NOT \$3.00/ton, it's \$3.00/200lbs. Converting back to tons yields the same price as that of the cane, or \$30.00/ton. However, though bagasse is technically a waste-product, it is not a non-valuable asset. Bagasse is typically burned for fuel to produce electricity via steam turbines or to drive the refining process of converting the rest of the plant into sugar. Because of this value and the additional treatment (like drying) the bagasse must undergo before it can be used as a fuel, the final price of the bagasse is around \$65/ton (Day, 2011).

The price of natural gas for the supercritical plant's duct burner (discussed later on) was found to be \$4.10 per million Btu (\$/mmBtu) based on data from June, 2011, when this study was performed (U.S. Dept. of Energy, 2011). 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 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) (EIA, 2010).

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:

$$\frac{\$}{mmBtu(coal)} = \frac{\$}{ton(coal)} * \frac{1\,ton}{2000lbs} * \frac{1}{LHV_{coal}} * \frac{10^6 Btu}{1\,mmBtu} \qquad (2)$$

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:

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

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:

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

One issue that arose during the course of this study was the independent derivation of another equation which could also calculate the total price for the blends as in equation 5:

$$\frac{\$}{mmBtu(total)} = \frac{BMR * \frac{\$}{ton\ (bio)} * \frac{1}{LHV_{bio}} + (1-BMR) * \frac{\$}{ton\ (coal)} * \frac{1}{LHV_{coal}}}{\frac{2000\ lbs}{1\ ton} * \frac{1\ mmBtu}{10^6 Btu}} \quad (5)$$

Notice that equation 5 reduces down directly to equation 2 when BMR = 0, and reduces down to an arbitrary biomass conversion equation when BMR = 1 (with the implicit assumption that the biomass is received dry and there is no change in heating value). Both of these equations were derived independently, and both seem to be valid methods of calculating the approximate cost of the blend on a per unit energy basis. However, these two equations result in different numbers for the final price, and will make a difference of between 0.1 and 0.9 cents/kW-hr in terms of cost of electricity when the final results are obtained. The main source of doubt is in the calculation of the "LHV_{blend}" quantity in equation 4, as it must be obtained numerically from GTPro: the total LHV of the blend is calculated by evaluating the energy available in all of the components within the blend using ultimate analysis, as opposed to the simple linear proportion used in determining equation 5. Further analysis is required to determine which one yields the more accurate result, but for this study, equation 4 was used, since it yields the larger, more conservative estimate. The estimates arrived at for both equations can be seen in Table 3.2.

Table 3.2 Total fuel cost for various blends of Biomass/Coal (prices listed as per million Btu)

Biomass Ratio (wt%)	Using Equation 4	Using Equation 5
0%	$1.480	$1.480
10%	$2.007	$1.820
30%	$3.110	$2.492
50%	$4.201	$3.160

Finally, for carbon capture, there has been some political discussion about a so-called "Carbon Tax" and issuing "Carbon Credits," in which a company may be fined for exporting too many carbon-based emissions, and in which financial rewards may be issued for those who reduce their emissions below a certain value. However, laws like these are not in effect at this time, and there is no indication that anything of the sort will be in the public spectrum anytime soon. Therefore, due to the uncertainty of carbon policy, the finance data presented in the results is based on the assumption that no Carbon Tax or Carbon Credits will be issued at all during any of the plants' 30-year life-spans.

3.1.4 Environmental concerns

3.1.4.1 NO_x and SO_x emissions

As mentioned in Chapter 1, NO_x and SO_x emissions are the leading causes of acid rain, and controls must be in effect to reduce the rate of release of both of these types of substances into the environment. For NO_x, this was done by choosing a GT with a low NO_x rating, and the calculation for NO_x emissions is based off of the GT performance rating for NO_x emissions. All NO_x produced during air-blown gasification is ignored: all N_2 at the beginning of the process is assumed to remain as N_2 until the syngas reaches the GT combustor.

There is no sophisticated NO_x solver in the program used, so data must be imported from the information provided by the GT supplier. Since the NO_x data for the Siemens SGT6-4000F was not available at time when this study was performed, the data from the Siemens SGT6-*5000*F was selected instead. This is a valid approach because both gas turbines make use of the same type of low-NO_x fuel combustor. The SGT6-5000F's average NO_x output was determined to be 9ppm at a reference O_2 content of 15% (Kovak, 2008).

Most SO_x, on the other hand, is handled ahead of time through Acid Gas Removal: removing the sulfurous compounds (especially H_2S) before they have a chance to even form SO_x. However, after this point in the system, there isn't much else left that can be done to prevent SO_x formation. For calculation purposes, all leftover sulfur and sulfurous compounds that reach the GT combustor are assumed to be completely combusted and released as SO_x.

3.1.4.2 Carbon-based emissions

There are many trace elements in emissions. For this study, the focus is placed on SO_x, NO_x, and CO_2. CO_2 is the most significant of these emissions types: a single 200MW power plant can produce upwards of several hundred-thousand pounds of CO_2 in just one hour.

The raw CO_2 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, <u>it was assumed that all biomass feedstock is completely carbon-neutral.</u> This allows for the calculation of the so-called "effective" or "net" CO_2 output, which is obtained by taking the total CO_2 and subtracting the biomass's neutral CO_2 from it. The neutral CO_2 was determined from equation 6:

$$Neutral\ CO_2(ton/day) = \dot{m}_{feed}\left(\frac{ton}{day}\right) * BMR * \%C_{bio} * \frac{M.W._{CO_2}}{M.W._{C}} * 8000\,\frac{hrs}{year} * \frac{1\,day}{24\,hrs} \quad (6)$$

where $\dot{m}_{feed}$ is the input mass flow rate of the blended feedstock in tons/day and M.W. stands for molecular weight. This equation was derived under the assumptions that (1) all carbon reactions result in CO_2, (2) start up and shutdown times are either neglected or assumed to be a part of the plant's recorded 8000 hour operating schedule, making no appreciable difference in the CO_2 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.

3.2 Specific Cases

3.2.1 Case Layout

The overall layout of the main cases is best highlighted by Fig. 3.9. Early on in the study it was determined to best categorize the different cases primarily by differences in hardware, i.e. subcritical vs. supercritical was placed as the primary dichotomy for the differences between cases. Below these two "groups" the main cases were classified by what type of carbon capture was used, with the no CCS case being declared Case 1, as it is the baseline. From there, post-combustion CCS is Case 2, sour-shift pre-combustion CCS is Case 3, and sweet/clean-shift pre-combustion CCS is Case 4. Below these main cases, 4 sub-cases, one for each level of biomass used in the feedstock, were considered. As per Fig. 3.9, henceforth, individual cases will be referenced by the formula - Case [capital letter-number-lowercase letter.] For example, Case B3c would be a plant with a supercritical steam cycle that makes use of pre-combustion CCS with sour CO shift, and with 30% biomass in the feedstock, while Case A1a, a baseline case, would be a standard IGCC plant with a subcritical steam cycle, no CCS, and pure coal feedstock.

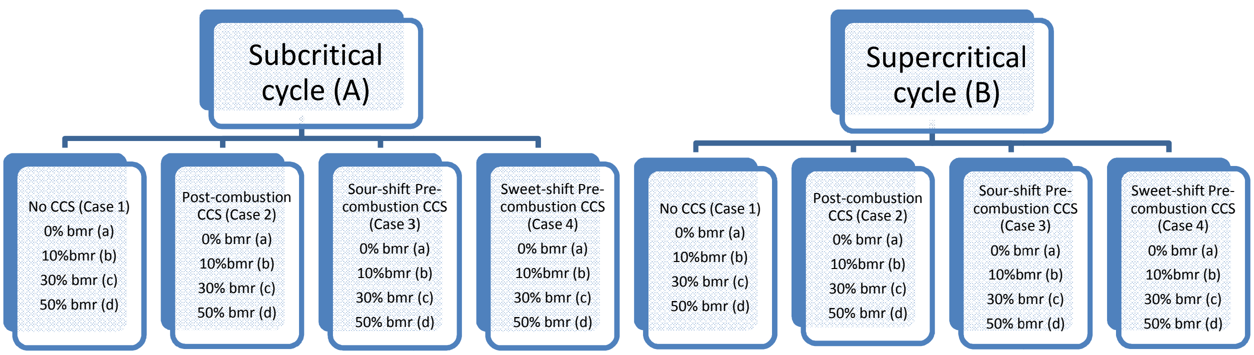

Figure 3.9 Case Layout

In addition to the main cases, a series of "special" cases studying different parameters were considered. Each of these cases is simply labeled with an "S" followed by a number that merely describes in which order the case was presented. For instance, "S1" is simply the first special case. There is no rhyme or reason to which number corresponds to which case, because each of these cases was made through an alteration of one of the main cases. In this instance, the only time a special case will be mentioned in the results is when the mention occurs with the special case alongside whatever this main case counterpart happens to be.

3.2.1.1 Groups A & B: subcritical vs. supercritical

Group A refers to the subcritical plant setup. The data and overall plant layout for this can be seen in Figs. 3.1 and 3.5 (for pure coal, without CCS). The subcritical steam plant has a TIT and TIP of 1000°F (537.8°C) and 1100 PSI, respectively. The reheat section occurs at an exit pressure of 200 PSI, and the inlet temperature of the second stage of the ST is set at 1000°F to match that of the first stage. In addition, all leftover water from the IP stream not used in running AGR and the other auxiliaries is injected directly into the second stage at a point where the pressure is 125 PSI. The temperature of this supplementary steam is around 550°F (287.8°C).

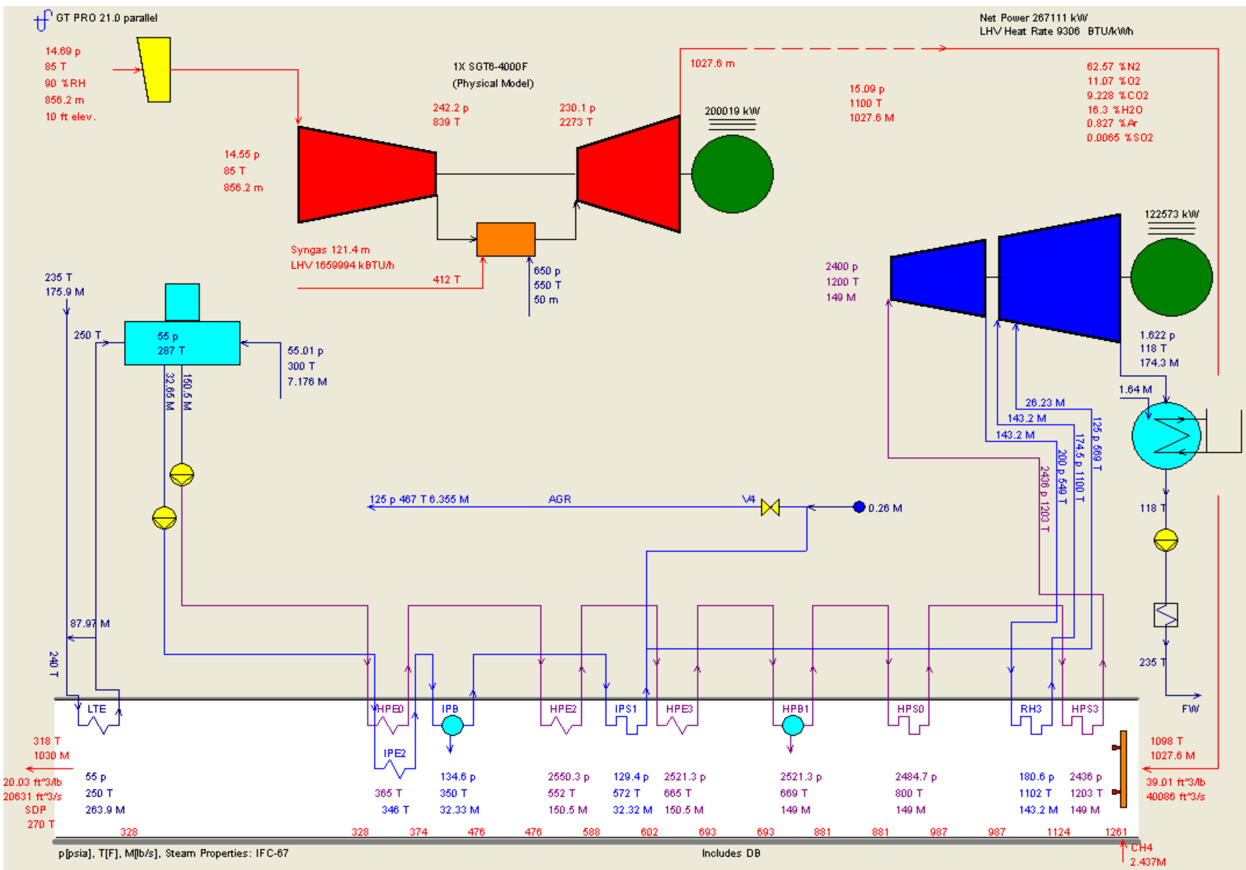

Figure 3.10 Supercritical Plant setup (Case B1a)

Group B is the supercritical plant setup. The different plant arrangement can be seen in Fig. 3.10 (Case B1a, specifically). The only difference in arrangement from the subcritical plant is the change in ST's TIT and TIP, which are now 1200°F (648.9°C) and 2400 PSI, respectively. In making these changes, however, some other parameters must change in order to meet this new demand. For one, all of the other temperatures and pressures in the HRSG must rise accordingly in order to keep the same pinch-points and pressure gains. For this to happen, the pumps providing the water must be upgraded to provide a higher pressure head, which equates to more work input needed. For temperature, however, it is noted that it is impossible to raise the temperature to the needed level, as the gas turbines exit temperature isn't high enough. In order for the steam to reach the indicated TIT, it becomes necessary to incorporate a device called a "duct burner" (as described in Chapter 2). The duct burner for these cases makes use of natural gas as the primary fuel (with heating value approximated as 100% CH_4, NO_x emissions set at 10ppm), and is set to provide enough heat in all cases for the GT exhaust gases to reach 1220°F (660°C, to account for piping heat losses on the way to the ST inlet). To accommodate the significantly increased steam pressure from 1100 psi to 2400 psi, the piping material and thickness had to be upgraded greatly. This piping change can't be seen in the typical plant layout figure, but the increased piping cost was included in economic analysis.

3.2.1.1.1 Case 1: No CCS

As mentioned previously, Case 1 is the baseline for the main cases. The cleanup system for Case 1 is exactly as it appears in Fig. 3.3, with no changes or alterations to the AGR system shown in Fig. 3.4. In addition, there are no added components or parts to the base system shown in Fig. 3.1 or 3.6 (3.10 for the supercritical case). The no CCS case is necessarily the most efficient of all the main cases, and as such should be the main source of comparison against all forms of CCS. Typically CCS costs a substantial amount for installation and operation and consumes an appreciable amount of energy to do, although one interesting exception did occur for Case 3 during the experiment (as will be shown in Chapter 4).

3.2.1.1.2 Case 2: Post-combustion CCS

Case 2 utilizes post-combustion CCS to reduce the emissions of the baseline case. The overall power-system layout is shown in Figs. 3.11 and 3.12 (subcritical type).

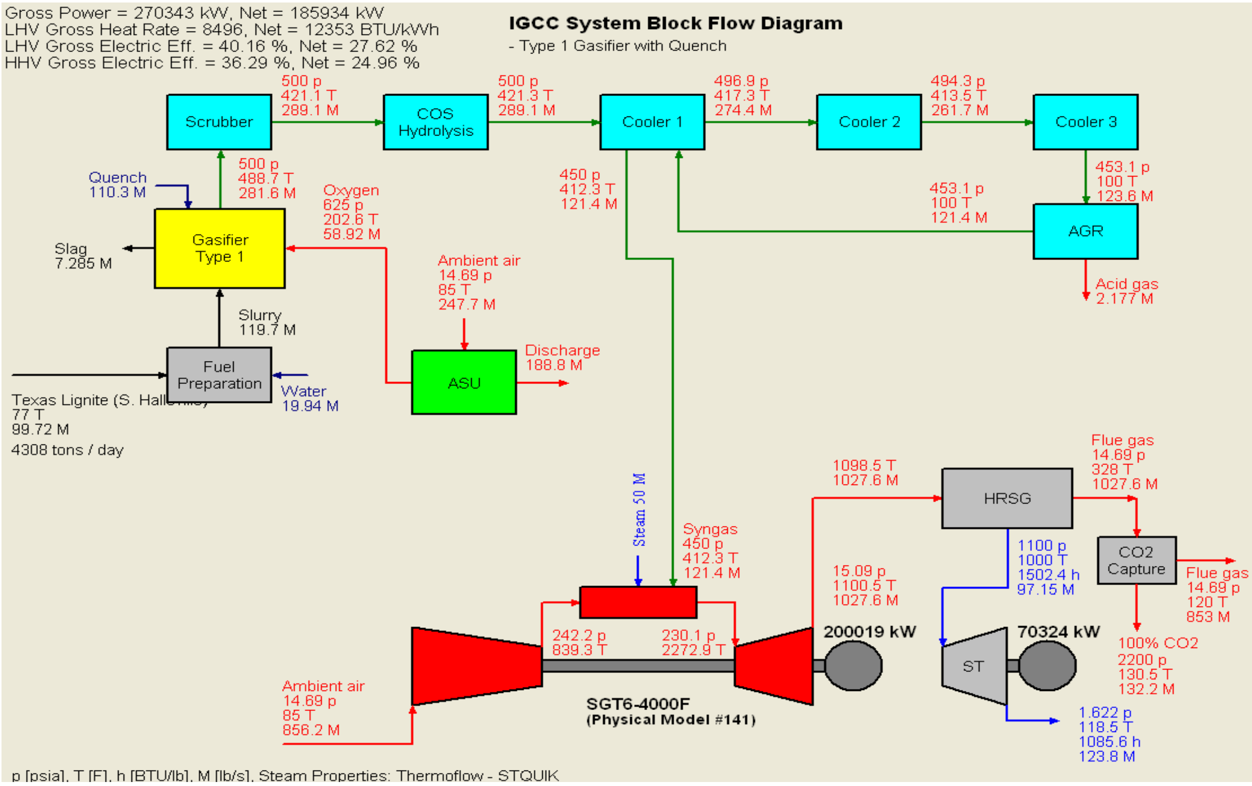

Figure 3.11 Overall plant layout: post-combustion CCS (Case A2a)

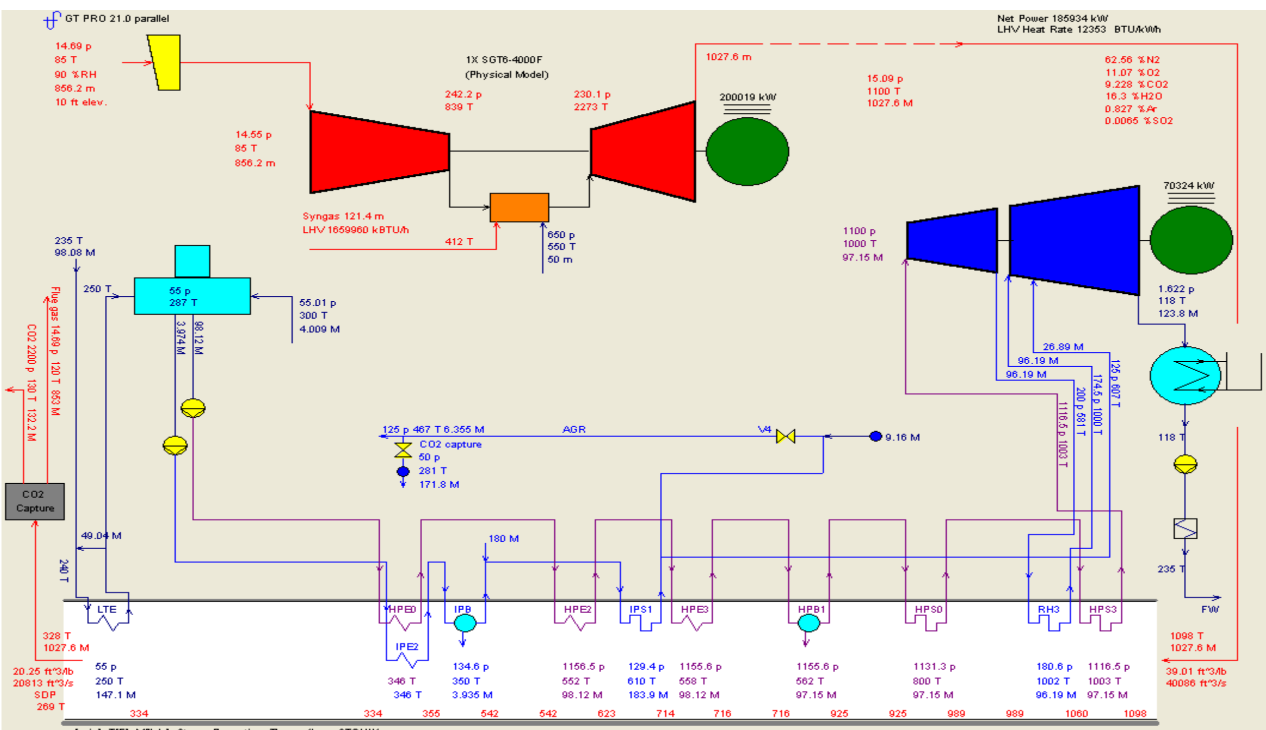

Figure 3.12 System design: Post-combustion CCS (Case A2a)

The CCS itself makes use of an amine-based absorption system, shown in Fig. 3.12. 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 (ICIS Pricing, 2010). On the whole, adding post-combustion CCS seems to take a drastic toll on the steam cycle's water supply. The original idea was to introduce a "carbon-ready" system that utilized steam from a low pressure source either outside the HRSG, or from a waste stream within the HRSG or at the ST exit, which would have allowed for the steam turbine to maintain the same power output, while the price for CO_2 capture would be paid in additional heat losses. This however, is not possible with current software limitations: an "external" source is not available for this type of system for whatever reason, and connecting the CCS system to the HRSG in any way reduces the amount of water delivered to the steam system. This does not make sense, since doing so increases the demand for steam from the whole system, and thus the steam flow rates should *increase* to compensate for this, which would be interpreted physically by changing the operating point of the pumps (or getting larger pumps) to draw more water from the source.

The steam that is used to drive the CCS process was 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. Unfortunately, the mass flow rate still suffers as a result, as mentioned previously (and elaborated on further on). In a real plant, it may not be necessary for even this, as the water used in the steam cycle has to be highly purified. The water required for amine-based CCS is purely a source of energy to help drive the stripping process. As such, it may be possible to reduce costs by using water without employing the *demineralization* process (where the water is cleaned and treated for use in the steam turbine). Using non-demineralized water would prevent the need to sacrifice ST power, and it would make better use of the water without having to push additional steam through the demineralization process when it may not in fact be necessary. Since this procedure can't be incorporated in GTPro, this idea cannot be examined as a part of the simulation.

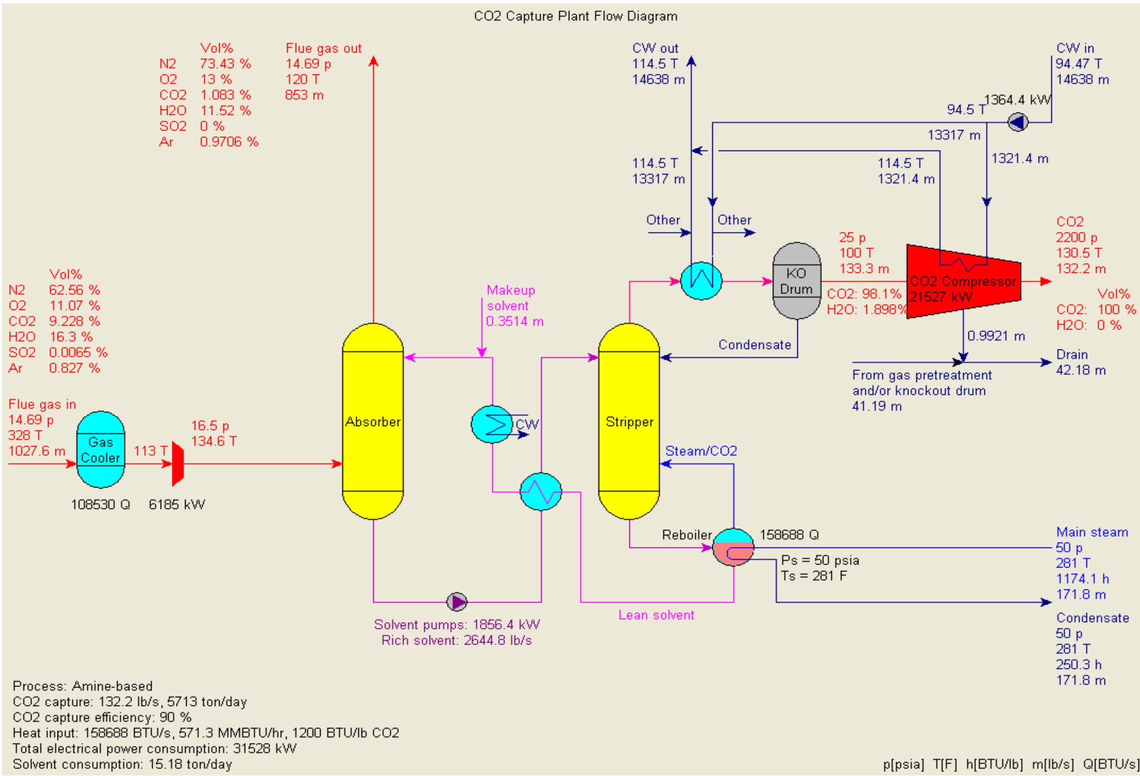

Figure 3.13 Post-combustion CCS (Case A2a)

As seen in Fig. 3.13, the demand for steam to operate the CO_2 capture system is quite high (171.8 lbs/s), and in the first stages of the simulation, this drastically cut into the ST's power output from the loss of mass flow. In an attempt to correct this problem, an artificial mass source had to be created (shown in Fig. 3.11) 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.

Also, Fig. 3.13 shows that the two columns in this CCS system are connected to a knock out (KO) drum. A KO drum is a liquid-vapor separator that relies mainly on gravity to take the liquid parts out of a liquid-vapor mixture, leaving the gases to exit at the top. In this case, it is needed to pull the extra water out of the otherwise pure CO_2 stream.

All CCS systems, including the pre-combustion ones, assume 90% capture efficiency. 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 CO_2 into a proper storage unit. The compressor raises the pressure of the captured CO_2 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.

3.2.1.1.3 Case 3: Pre-combustion CCS with sour shift

For Cases 3 and 4, the steam cycle and HRSG layout are virtually identical to those of Case 1. The only difference is in the external cooling needed during the CO-shift process, where a line from the low pressure water supply is delivered to the cleanup system and then subsequently delivered to the high pressure boiler (HPB) after receiving the heat from cooling the syngas. See Fig. 3.14 for the plant design.

Sour-shift is very economical, as it allows for the shift process to occur at the same time as COS hydrolysis, as shown in Fig. 3.15. This is possible because both COS hydrolysis and the water-gas shift reaction share not only a reactant, but can utilize the same catalyst in pushing the reaction forward. Other than the additional water supplied to the COS hydrolysis reactor, there are no major changes or additions to the cleanup system as a whole from the baseline cases without CCS. The acid gas removal plant, however, has an extra stage added to it, as shown in Fig. 3.16, to incorporate carbon capture. To account for this, additional water is taken from the same source as the AGR block in the baseline case (Case A1) and the total amount is supplied to the combined AGR/CCS plant.

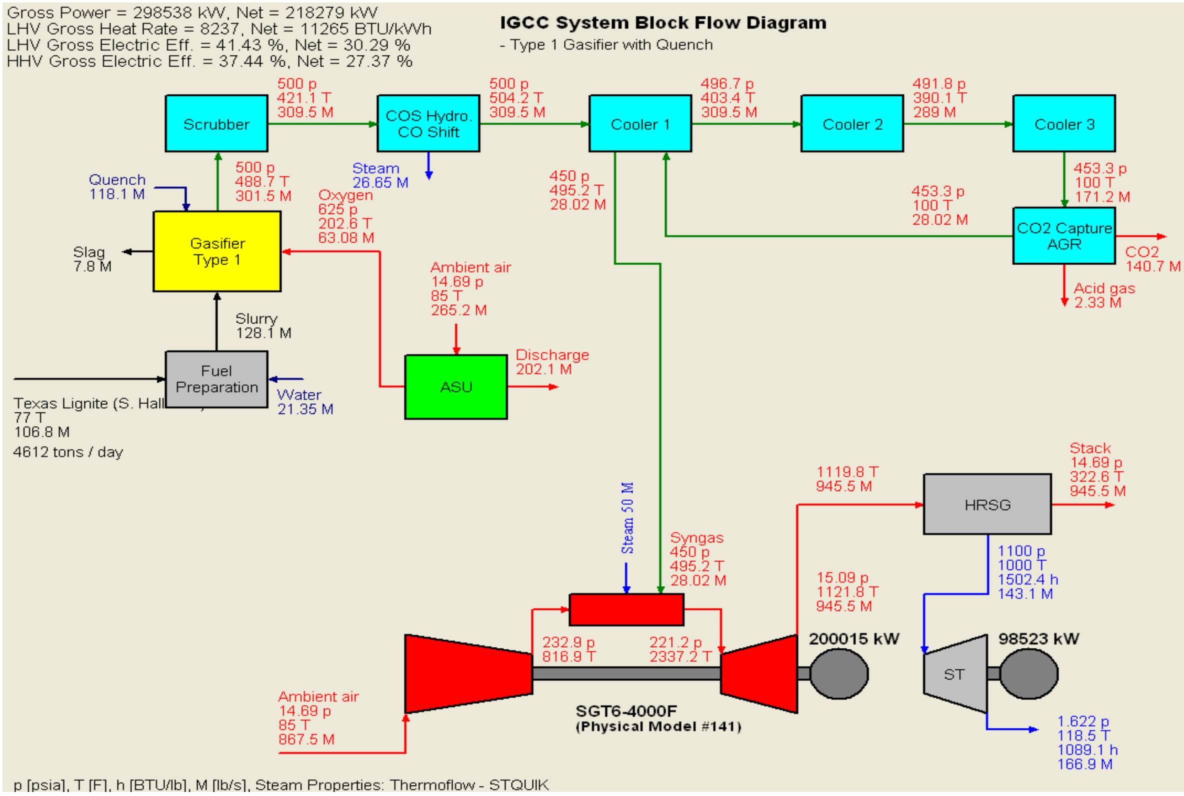

Figure 3.14 Overall plant layout: sour-shift pre-combustion CCS (Case A3a)

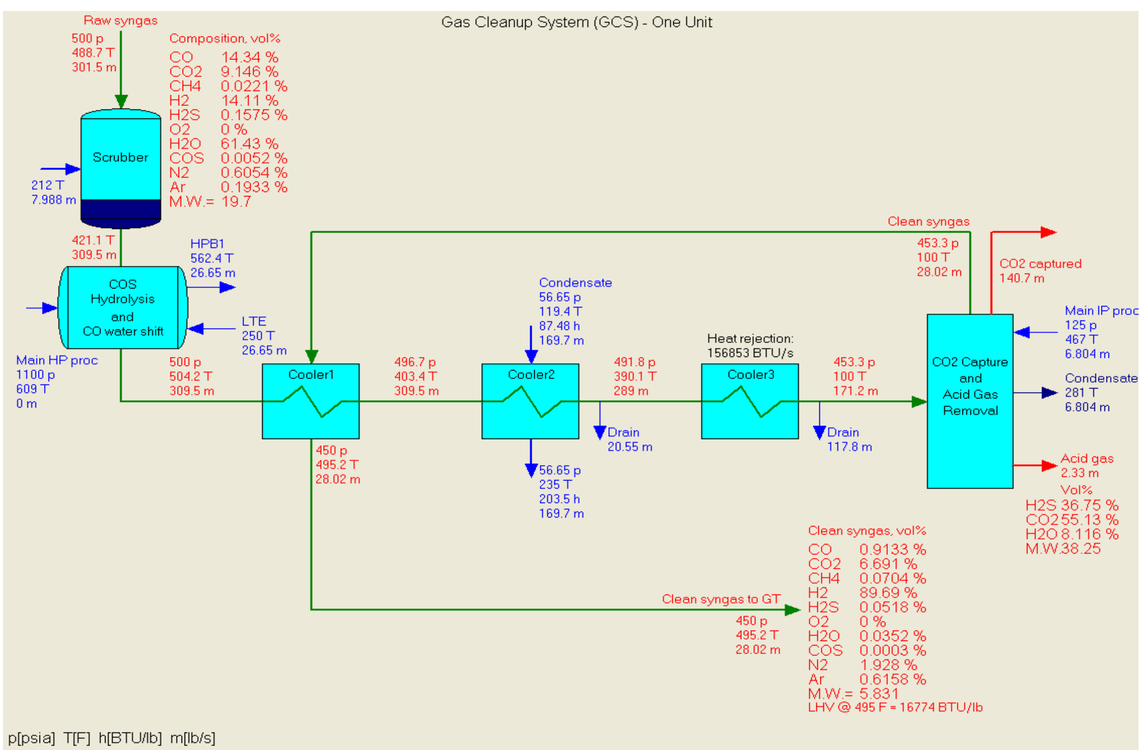

Figure 3.15 Pre-combustion CCS cleanup system with sour-shift (Case A3a)

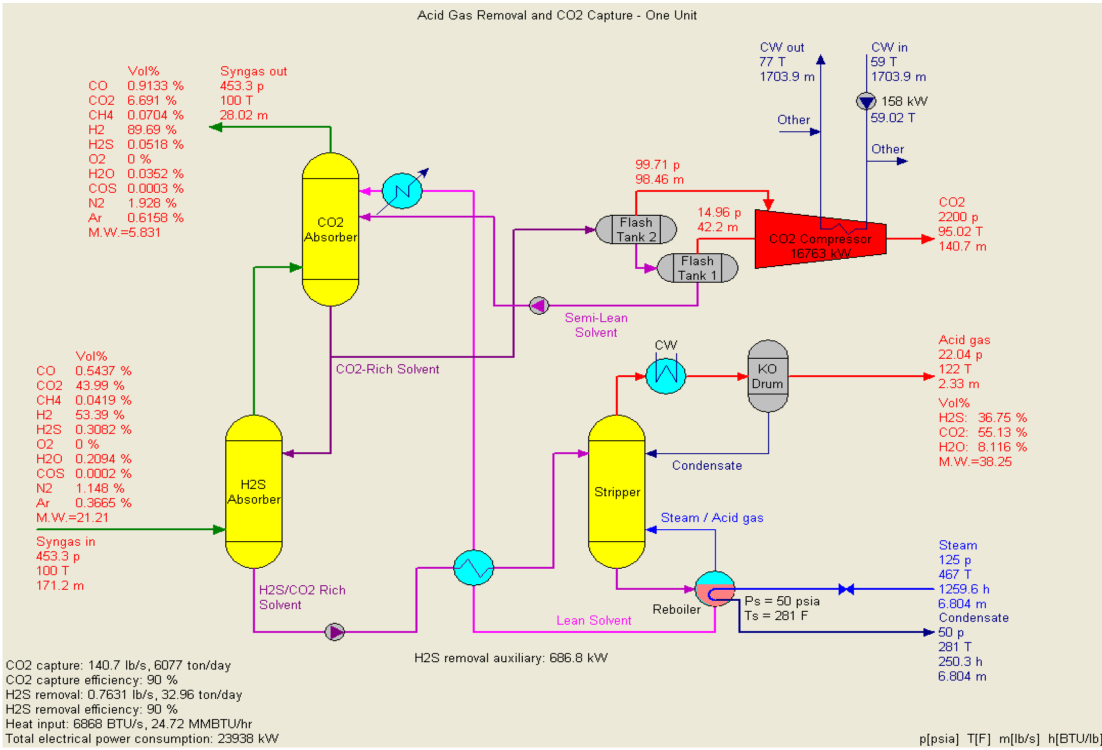

Figure 3.16 Sour shift case CCS + AGR plant (Case A3a)

Since this is pre-combustion CCS, it allows for the use of a physical solvent, meaning although the solvent itself is more expensive (assumed to be about $2000/ton), the process itself is less demanding on the system as a whole, as discussed in Chapter 2. The absorbers themselves operate in a cascade-like manner, with the lean solvent first absorbing CO_2 in one absorber, and then sliding down to absorb H_2S in a second absorber. Meanwhile, the syngas enters into the H_2S absorber and flows counter to the solvent, arriving at the CO_2 absorber to undergo carbon capture. This is necessary, because CO_2 and H_2S mix together when under the conditions for AGR, that is, the two compounds will dissolve at the same time. Sequestration implies that the CO_2 will be used for some other purpose, such as advanced oil recovery, which requires an extremely pure stream of CO_2 in order to work properly. Therefore, if H_2S is not removed beforehand, it will require even more work afterward to achieve the right level of CO_2 purity.

The sequestration system makes use of two flash tanks instead of the KO drum from post-combustion CCS. This is because, unlike in post-combustion CCS, the CO_2 absorber isn't directly connected to a stripper column: it will be much easier and less expensive to use flashing to pull the captured CO_2 out of solution. This means that there isn't very much water to separate from the mixture, so there is no need for a KO drum like there is at the end of the H_2S removal stage. The top flash tank strips about 70% of the CO_2 from the solvent, while the lower tank handles the remaining 30%. In addition, since this is physical absorption, there is no condensate to be removed before compression (and sequestration), and no additional cooling water needed, since there are no chemical reactions. Since sour shift occurs before AGR, there is no additional water needed to complete the shift: only the catalyst need be added (hence, this is why the "main HP process" water added to the shift reactor has zero mass flow).

3.2.1.1.4 Case 4: Pre-combustion CCS with sweet shift

For sweet shift pre-combustion CCS, additional changes must be made to the cleanup system, as outlined in Chapter 2, and shown in Figs. 3.17 and 3.18. As can be seen in Fig. 3.18, the flow rate for the additional water for the shift reactor is *not zero*. This is because the water within the syngas that can be used for sour shift has already been either condensed out and drained away or consumed in acid gas removal. In the end, there is not enough water left to complete the shifting process to the 99% required amount. To make matters worse, due to the very high pressure of the syngas stream, the only viable sources of steam available (external sources are assumed unavailable) are from an HP exchanger in the HRSG, or a high pressure turbine bleed. In either case, this will cause the ST performance to suffer, which will be reflected in the results. This will require an extra auxiliary stream in the steam cycle, which can be seen in Fig. 3.20.

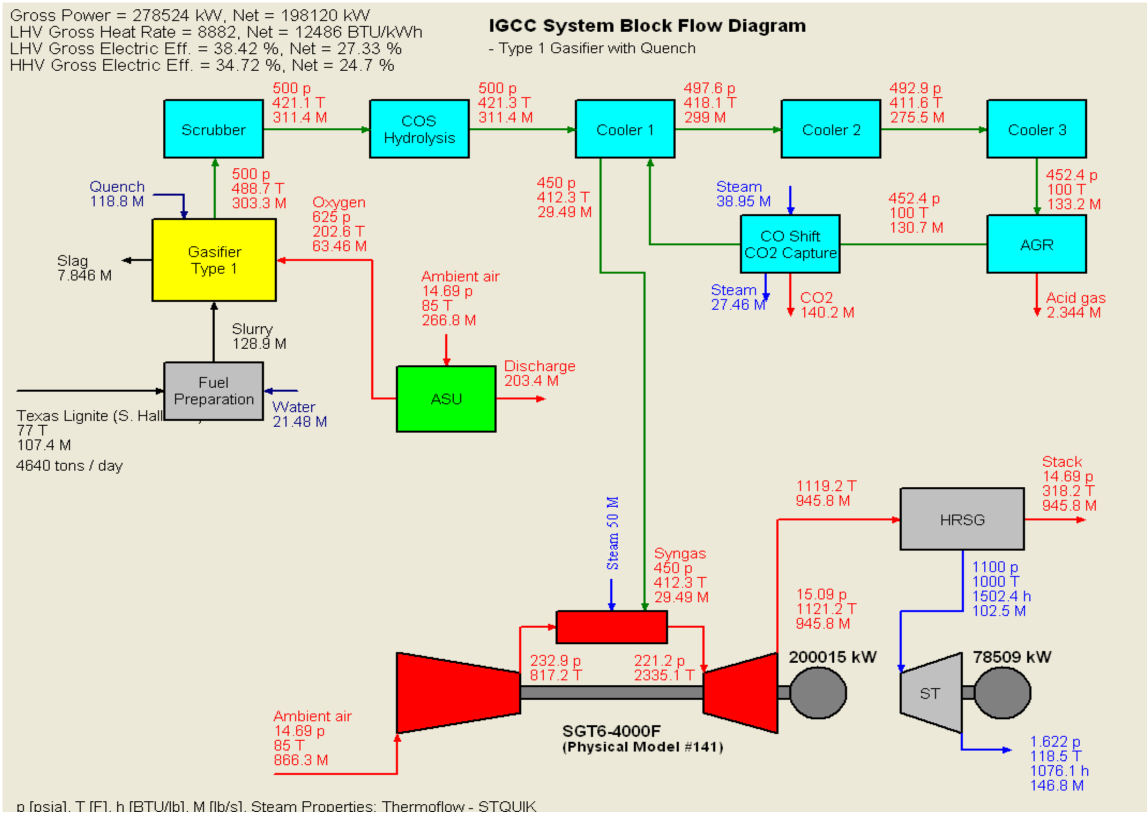

Figure 3.17 Overall plant layout: sweet-shift pre-combustion CCS (Case A4a)

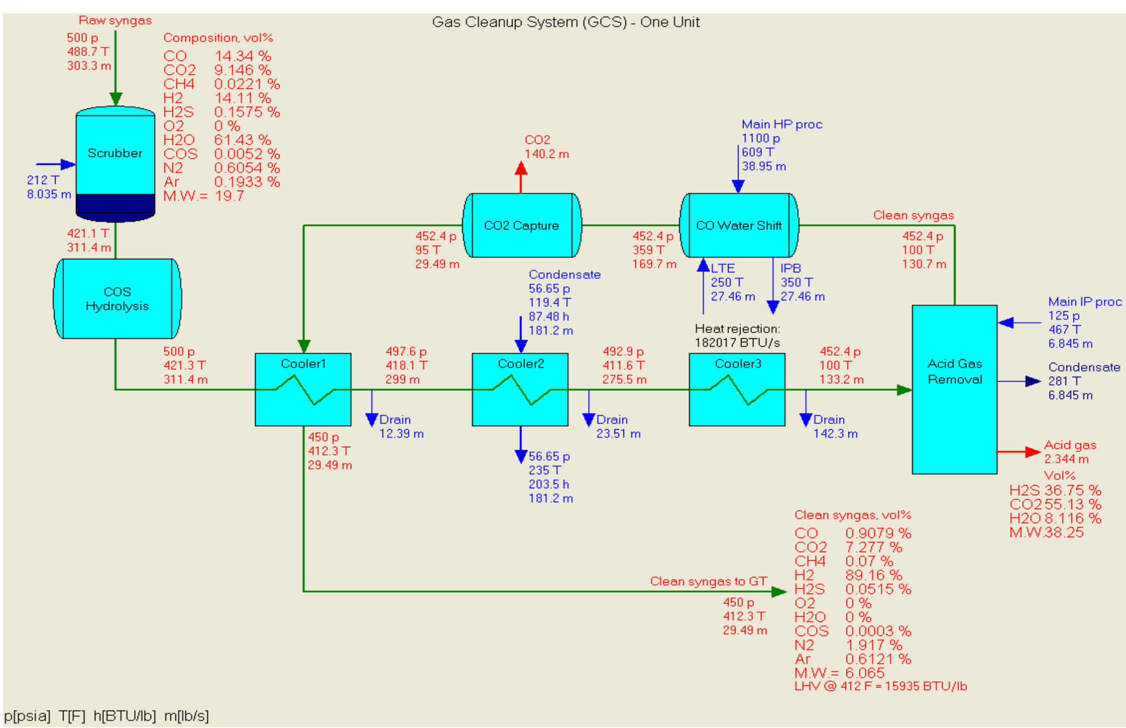

Figure 3.18 Pre-combustion CCS cleanup system with sweet shift (Case A4a)

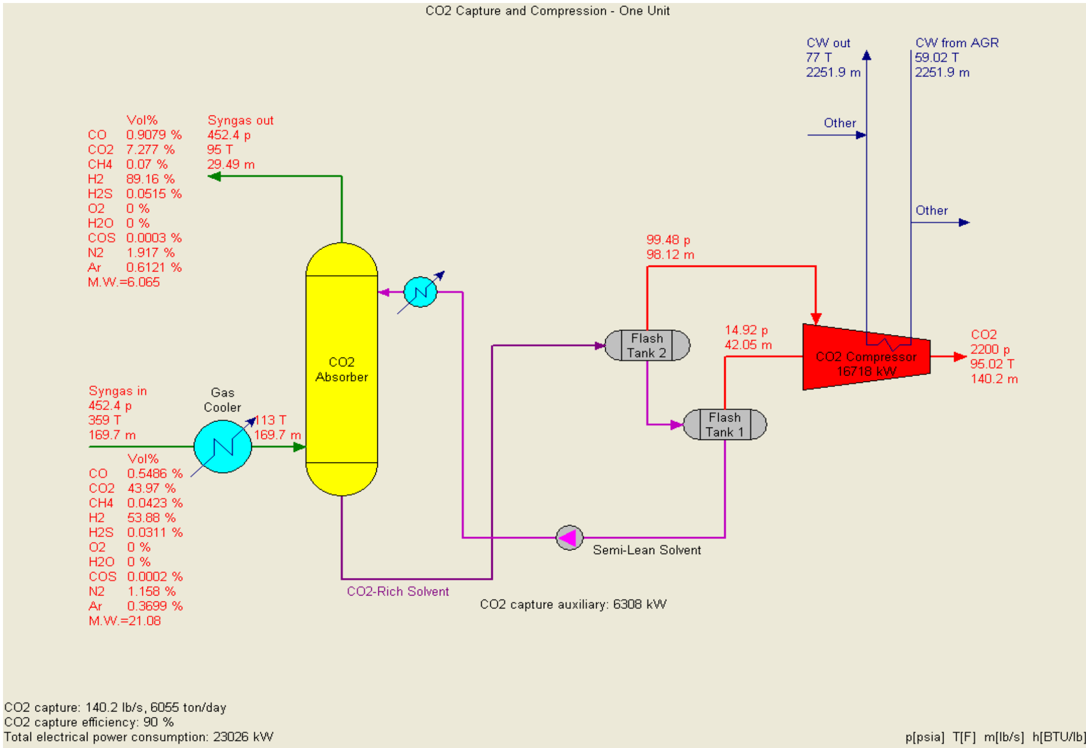

Figure 3.19 Sweet shift case CCS plant (Case A4a)

In order to save energy to help mitigate these losses, the additional supplementary cooling steam was returned to the IP stream rather than the HP stream (indicated in Figs. 3.17, 3.18, and 3.20). The logic was that the reduced necessary cooling would allow the syngas to reach the GT combustor at a higher temperature, allowing for more efficient use of the fuel, since the TIT must be held constant.

The advantage of sweet shift is that the CO_2 capture plant is more simplistic, so there is less risk of some CO_2 being lost during acid gas removal. The CO_2 capture plant in Case 4 has the same setup and controls as those of the CO_2 capture section in Case 3, with two flash tanks (one stripping 70% of the carbon dioxide and the next handling the remaining 30%), and a CO_2 compressor bringing the pressure to 2200 PSI. The CO_2 compressor, like in the sour-shift case, is used for sequestration purposes. The CO_2 is assumed to be sequestered for a purpose, but what that purpose is is purposefully left open-ended for this study, since the nature of this purpose may affect the economics in such a way that ordinary cost analysis cannot account for (For instance, advanced oil recovery would bring in additional profits, which cost analysis does not account for.) The layout of this is shown in Fig. 3.19. Notice that the total power consumption of sweet shift's CCS plant is only about 900kW less than sour shift's combined AGR/CCS plant.

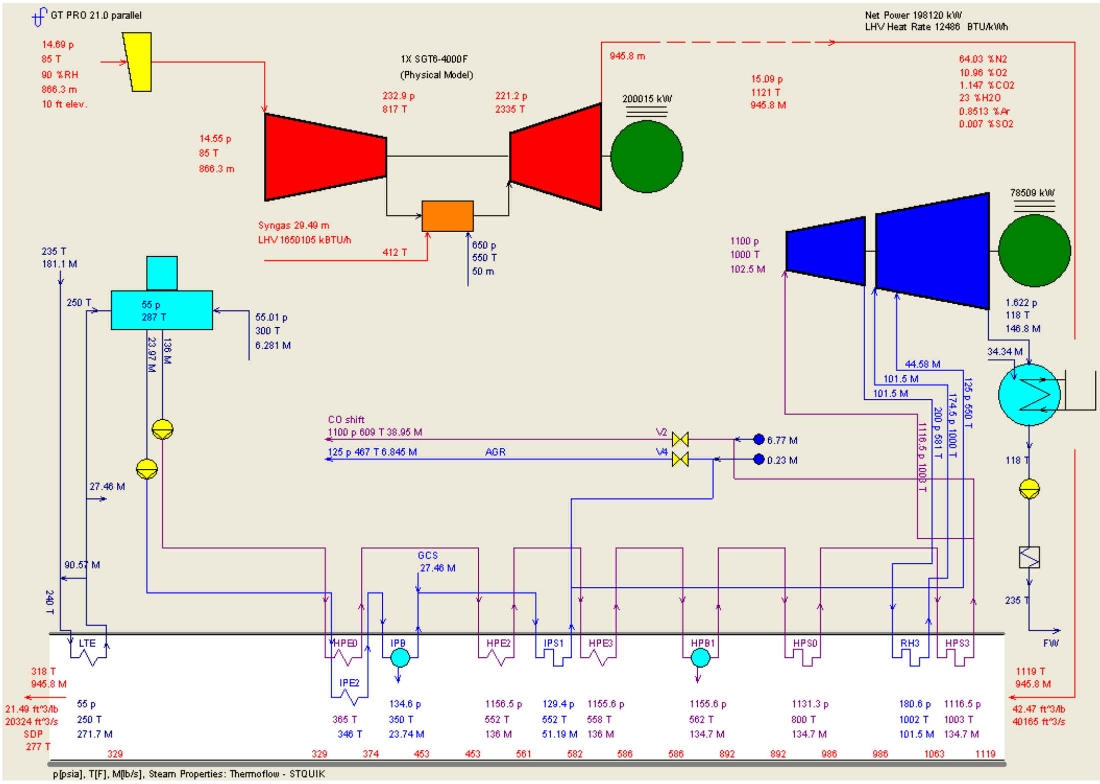

Figure 3.20 Power block for pre-combustion CCS with sweet shift (Case A4a)

3.2.1.2 Special Cases

As mentioned previously, in addition to the main cases, there are a series of "special" cases being studied. Each special case changes exactly one parameter and observes its effect on the overall cycle by comparing it to one of the main cases. They are not considered main cases because they do not have any sub-cases. Each case is run specifically to observe a specific situation and to evaluate the initial choices made during the design of the main cases, such as using a slurry feedstock or opting to use a quench-cooled system. The individual cases and their specifics are listed in the sections that follow.

3.2.1.2.1 Special Case 1: Radiant/Convective Coolers

The first and most obvious question to ask is what would have happened had RSCs or CSCs been chosen instead of the quench system used in the main cases. As discussed in Chapter

2, RSCs and CSCs are always more thermally efficient than any quench, but they are expensive, and cost is a major priority in this study. However, this raises a question: is the gain in efficiency good enough to circumvent these costs? This is the primary reason for running this as a special case.

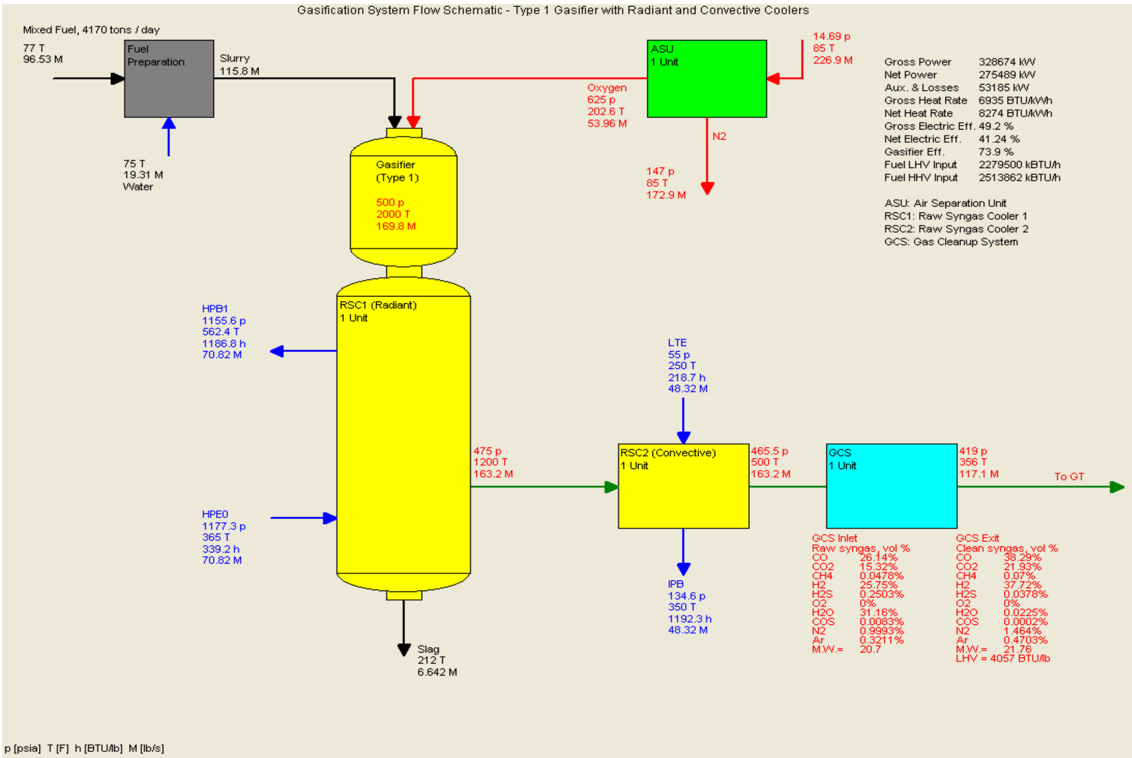

Figure 3.21 Gasifier with radiant and convective coolers

The gasifier, shown in Fig. 3.21, is still a GE gasifier with the same conditions on it as the subcritical baseline, but with the addition of 1 radiant and 1 convective syngas coolers. The radiant cooler is connected to the HP stream, while the convective cooler (with its much lower required temperature) is connected to the LP section right before the deaerator, but delivers the remaining steam to the IP stream near the auxiliary process. Thus, aside from some numbers, the steam cycle for this system is virtually unchanged aside from these 4 additional input/output connections. The radiant cooler has a fixed exit temperature of 1200°F (648.9°C), while the convective cooler's output temperature is fixed at 500°F (260°C): about the same as the final temperature of the raw syngas in the quenched system of the main cases. Both types of coolers were used in order to save money, as one very large radiant cooler is far more expensive than these two small coolers together. In this study, the difference is about $759,008,000!

The case as a whole is compared to main Case A1b (10% BMR, no CCS), because, as the results will show, sub-case b is almost uniformly the highest in efficiency for a specific case set. Since the idea of using a supercritical steam cycle is not yet in practice, it was determined that altering case A1b would be the best candidate to determine the highest efficiency possible while still within the realm of practicality.

3.2.1.2.2 Special Cases 2 & 3: Dry fed system

Another parameter, perhaps not raising as obvious of a question as Case S1, that is being considered is the idea of using a dry 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. A major cause for concern in this case is with carbon capture. Sour shift has a very big advantage in a slurry-fed system due to the extra water already present in the syngas after gasification occurs, so what happens if that extra water isn't there in the dry-fed systems?

Two cases were considered in this instance. The reason is to see first-hand the effect of dry-fed systems on both sour-shift CCS and sweet-shift CCS. The comparison cases run were Cases A3c and A4c. Biomass benefits greatly from CCS, as it has the advantage of being carbon-negative, so 30% biomass was determined to be the best point at which to examine this parameter.

The gasifier chosen was the Shell gasifier, shown in Fig. 3.22, which uses an internal water-cooled membrane, as mentioned in Chapter 2. Since there is no slurry, the feedstock is transported into the gasifier by means of excess high-pressure nitrogen from the ASU, but not enough to negate the value of using oxygen-blown approach (Notice in the figure, 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, was taken from the IP stream, since this stream doesn't have a pressure requirement. Finally, in order to keep the comparison as close as possible to the original cases, the gasifier also uses water to quench the syngas, with the same stipulations as the main cases (300°F added until 50% R.H.).

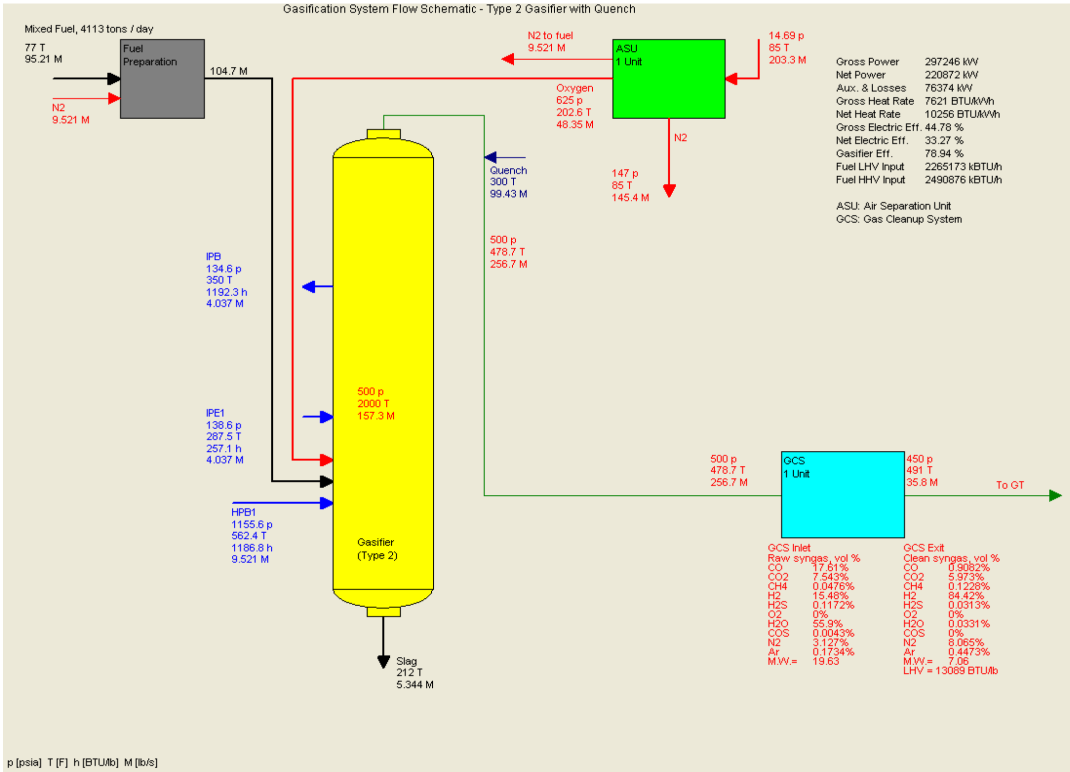

Figure 3.22 Dry-fed gasifier

3.2.1.2.3 Special Case 4: Air-blown gasifier

Another parameter thought to have interesting implications is the idea of using an air-blown gasifier over an oxygen-blown gasifier. As addressed in Chapter 2, 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.

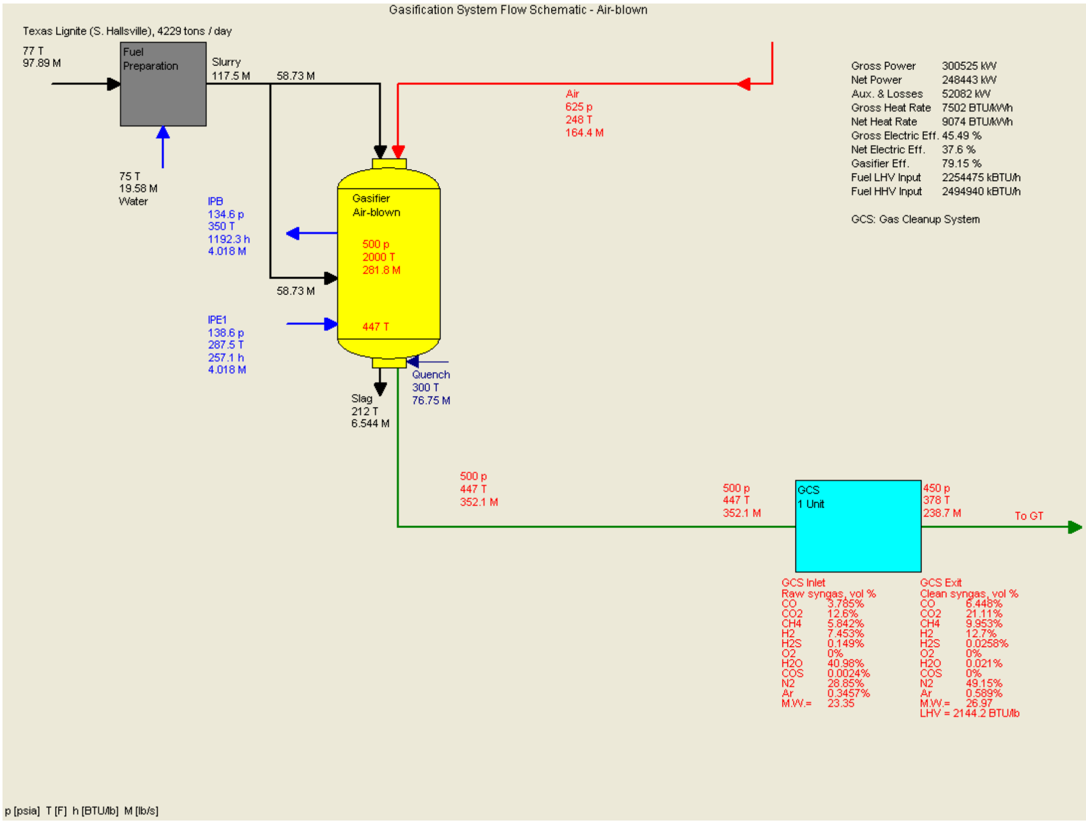

Figure 3.23 Air-blown gasifier

The only large-scale commercial gasifier that has ever successfully used an air-blown layout is Mitsubishi Heavy Industries's 2-stage gasifier, which is shown in Fig. 3.23. The main controls were set the same as the other gasifiers, with the same temperature and pressure. Like Shell's gasifier, MHI uses membrane-wall cooling in their design, which was connected to the IP stream, just like Cases S2 and S3. Again, to keep with comparison, the gasifier was also subjected to a quench, with the same stipulations mentioned previously (300°F water added until 50% R.H. is achieved in syngas.) A 35% slurry is used, and the air enters the gasifier with the same pressure as the main cases through the use of a boost compressor (not shown.) Although an MHI gasifier usually uses dry-fed mechanism, a slurry was imposed upon the design in order to maintain a level of comparison between this case and the main cases, since they are all slurry-fed designs. This approach causes much less of a discrepancy with reality than the alternative (taking the ASU out of the GE gasifier design and forcing it to use air. Finally, this gasifier is a two-stage gasifier, so part of the feedstock must be inserted at the second stage. Thus, a 50:50 split was arbitrarily chosen as the ratio for feedstock sent to either stage.

The comparison case here is Case A1a, 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. As such, it is only of concern to see how the efficiency, power, and heat transfer are affected by the use of such a system, so the pure coal case without CCS is sufficient to satisfy this level of interest.

3.2.1.2.4 Special Case 5: Duct Burner – use GT fuel

The supercritical system makes use of natural gas to run the duct burner. This 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.

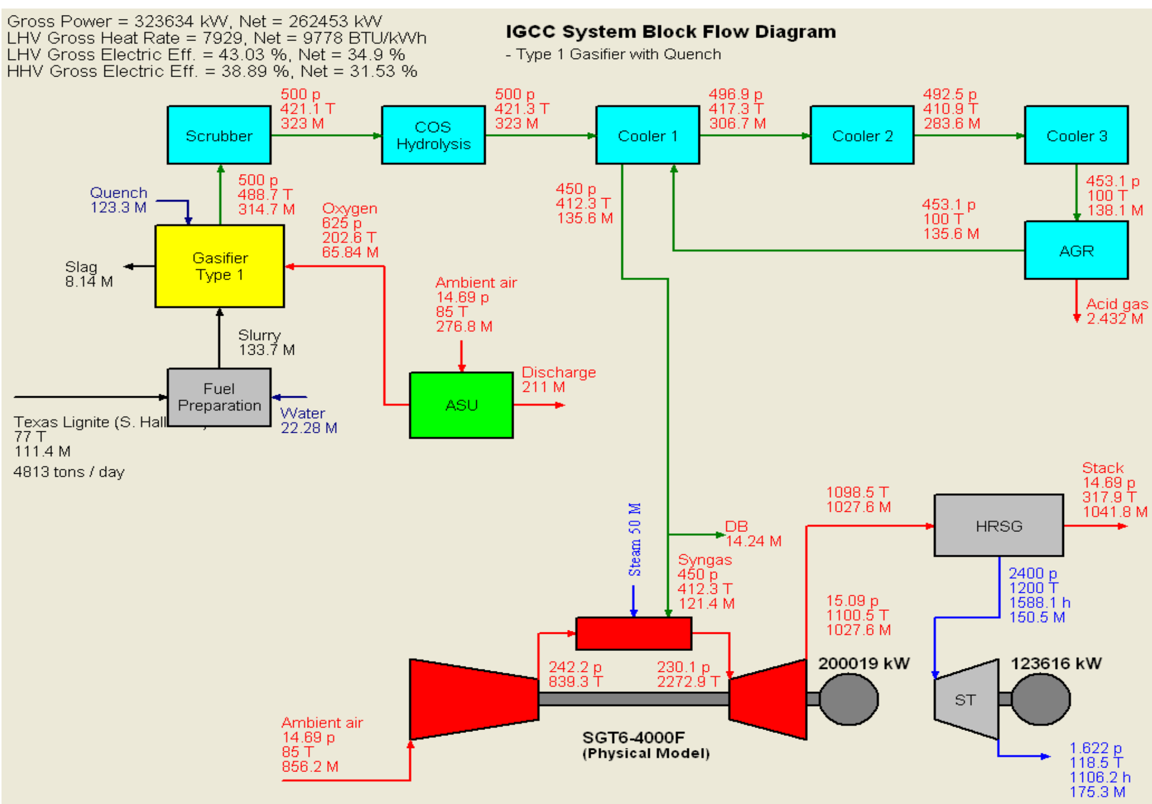

Figure 3.21 Supercritical cycle with syngas in duct burner

The layout for this idea is shown in Fig. 3.21. Notice where the split in the syngas stream is; where the offshoot leads to the duct burner (DB). Whether this is good for the efficiency or not will be seen in the results. The comparison case is Case B1a, since it is not important for biomass or CCS whether this idea works or not.

3.2.1.2.5 Special Cases 6 & 7: Illinois #6 – higher-rank coal

Texas Lignite was 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%). The ultimate analysis and other data on this coal can be seen in Table 3.3

Table 3.3 Various data for Illinois #6 high-volatile C bituminous coal

Fuel	**Illinois #6**
C (wt%)	55.35
H_2 (wt%)	4
N_2 (wt%)	1.08
S (wt%)	4
O_2 (wt%)	7.47
Cl_2 (wt%)	0.1
H_2O (wt%)	12
Ash (wt%)	16
LHV (Btu/lb)	9599
Price ($/ton)	50.00

Since Louisiana and Illinois are both along the Mississippi river, which is a frequently used avenue of transport for both states, it was considered that Illinois #6 could be fairly easily shipped to Louisiana by barge. The coal could then be considered to have the same availability as Texas or Mississippi Lignite. Despite this, Illinois #6 is more expensive, but it is still of higher rank. The idea here is to see (1) if the better fuel will be able to raise the efficiency enough to overcome its more than double price tag and (2) how the new fuel will be affected by adding biomass. Thus, there are two cases to compare to: Cases A1a and A1b.

CHAPTER FOUR
RESULTS AND DISCUSSION

4.1 Method of Analysis and Restatement of Objectives

To reiterate, the main objective of this study is to both reduce the emissions of IGCC systems and to reduce their cost. Reduction of emissions can be handled by using biomass in the feedstock, raising the thermal/electrical efficiency, and incorporating different forms of carbon capture technology. Biomass can reduce the emissions because it contains no sulfur, contains very little nitrogen, and is carbon-neutral. Raising the efficiency allows either (a) more power for the same fuel input, reducing the need to construct new plants or (b) the same power for *less* fuel input, directly reducing the amount of hazardous gases emitted by the plant. Finally, carbon capture technology provides a means to directly remove CO and CO_2 from the plant's exhaust gases, preventing them from ever entering the atmosphere and allowing CO_2 to either be used for other, more practical purposes, such as advanced oil recovery, or simply be sequestered underground.

The results are analyzed based first on the effect of the biomass in each group without CCS, then by the presence of CCS, and finally the difference between the subcritical and supercritical cycles across these two groups. In other words, all main cases within a specific group (subcritical or supercritical) are observed first. Within each main case, all sub-cases are compared and analyzed against each other, and then compared with their counterparts in other main cases as a whole. The analysis will start from the subcritical group, followed by the supercritical group. Cross-comparisons with the specific cases will be made between the two groups (for example, Case B2b will be compared to Case A2b).

As mentioned in Chapter 3, the special cases are treated as simple alterations to an individual case, and will be compared solely to that case, whichever it may be. The comparison cases are listed in each special case's individual section and also in the appropriate section back in Chapter 3. Each special case examines one additional parameter not addressed in the main cases, such as using a dry-fed system instead of a slurry-fed one, an air-blown rather than oxygen-blown gasifier, a higher ranked coal rather than the lignite, radiant and convective coolers instead of a quench system, etc.

4.2 The Main Cases

4.2.1 Subcritical Cycles (Group A)

As mentioned in Chapter 3, the subcritical plants all have a TIT and TIP of 1000°F (537.8°C) and 1100 PSI (74.83 bars), respectively. The steam turbine uses two casings with a reheat section between them, with the inlet temperature of the secondary casing determined to be 1000°F. The HP stream from the HRSG is used to feed the primary casing, while the IP stream is used to drive plant auxiliaries, and also provides additional steam to the secondary casing at a point where the internal steam pressure is 125 PSI. Overall, the subcritical plants deliver somewhere between 180-240MW_e of power to the grid, depending on the case.

4.2.1.1 No CCS (A1) --- Baseline

The baseline case does not make use of any CCS technology: it merely stands to provide a baseline efficiency, power output, cost, and set of emissions for the other plants to compare to. Like every main case, this setup has four sub-cases, one for each biomass ratio (BMR). Again, the biomass ratios studied are 0% (pure coal), 10%, 30%, and 50%, all by weight. All important data for this case can be seen in Tables 4.1-4.5. The basic layout for this plant was shown briefly in Chapter 3, in Fig. 3.1 and again with an emphasis on the steam cycle in Fig. 3.6.

Table 4.1 Work and efficiency for subcritical plants without CCS

Case Number	A1a	A1b	A1c	A1d
Biomass/Coal Ratio	0%	10%	30%	50%
Gross GT Power (kW)	200,019	200,018	200,017	200,017
Gross ST Power (kW)	89,477	89,790	90,191	90,551
Auxiliary Losses (kW)	53,499	52,451	55,913	59,277
Total Net Power (kW)	**235,997**	**237,356**	**234,296**	**231,291**
Gross Efficiency (LHV)	43.01	43.59	43.96	44.31
Net Efficiency (LHV)	**35.06**	**35.70**	**35.49**	**35.27**

From Table 4.1, it becomes apparent that adding biomass to the baseline plant slightly increases the efficiency with the highest efficiency occurring in Case A1b. The efficiency drops off beyond this point, but the cases run with biomass in the feedstock always seem to be more efficient than the case with coal alone (A1a). The efficiency increases most obviously due to the raised ST power, but also due to the reduced heat losses in the gas cleanup system, seen in Table

4.2. Beyond 10% BMR, 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 Table 4.1 and Fig. 4.1, so the *gross* efficiency is unaffected by this, and, in fact, continues to increase since more gross power is generated.

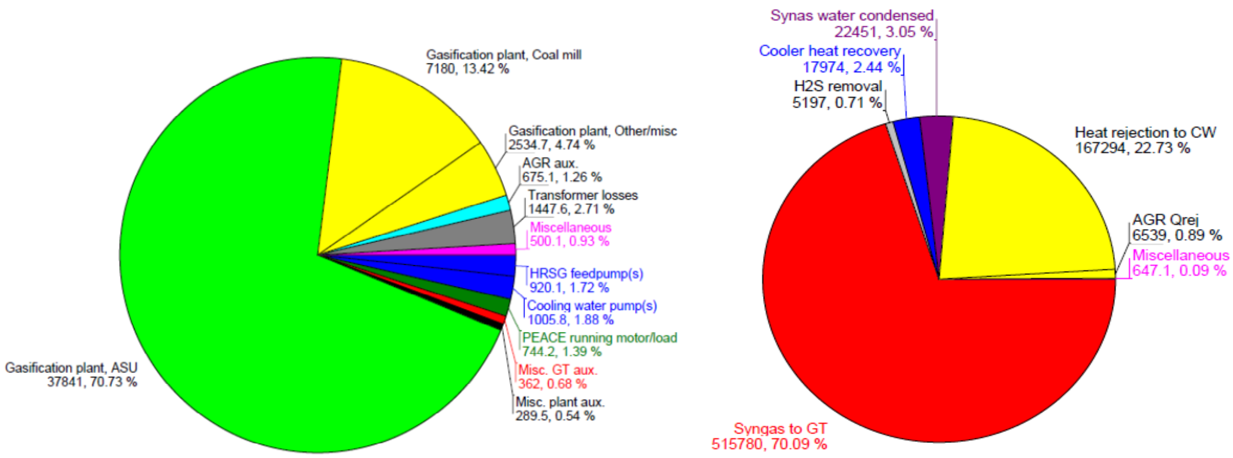

Figure 4.1 Auxiliaries & losses (left) and gas cleanup heat out (right) for Case A1a

Table 4.2 Selected parasitic energy & heat losses for subcritical plants without CCS

Case Number	A1a	A1b	A1c	A1d
Biomass/Coal Ratio	0%	10%	30%	50%
Acid Gas Removal (Btu/s)	5,197	4,532	3,534	2,562.9
Syngas Water Condensed (Btu/s)	22,445	21,584	21,168	20,765
AGR Heat Loss (Btu/s)	611.2	533.0	415.6	301.5
Slag Production (Btu/s)	14,473	13,843	13,451	13,071
Cooler Heat Rejection (Btu/s)	167,317	156,460	151,001	145,686

From Table 4.2, it is obvious that the sugarcane bagasse is much cleaner than lignite: there is less sulfur content in it, so its presence obviously reduces the energy needed to remove sulfurous compounds via AGR. (Note that the listed "Acid Gas Removal" category in the table represents the actual energy spent during the process. "AGR Heat Loss" is the energy removed due to H_2S's heating value.) There is also less ash, so, naturally, less slag is produced from the gasification. Less water is condensed out of the syngas from the cleanup system because, as shown in Table 3.1 in Chapter 3, after torrefaction, sugarcane bagasse has *less* water content than lignite does, so less water is available to condense out. However, due to the reduced water demand from the AGR plant, this results in more water available for the steam cycle, which translates directly into more ST power, as seen in Table 4.1. Lastly, because there is less water to condense, there is less heat needed to be rejected by the coolers in the gas cleanup system. The difference in energy largely contributes to the increased efficiency over pure coal, especially in Case A1b. A more qualitative representation of the data in Table 4.2 is shown in Fig. 4.1 for Case A1a, specifically.

From an economic perspective, biomass actually appears to be a *benefit* rather than a detriment, as shown in Table 4.3. The capital cost decreases as BMR increases, while the Cost of Electricity (CoE) decreases only with up to 10% BMR but increases with more added biomass.

Table 4.3 Economics for subcritical plants without CCS

Case Number	A1a	A1b	A1c	A1d
Biomass/Coal Ratio	0%	10%	30%	50%
Total capital cost (millions of $)	1,029.75	926.74	911.62	897.44
Capital cost ($/kW)	**4,363**	**3,904**	**3,891**	**3,880**
CoE ($/kW-hr)	**0.1008**	**0.0979**	**0.1084**	**0.119**

These reductions of both capital cost and CoE are not exactly clear at first, but, looking back at the work and efficiency data in Table 4.1, it shows that the GT output power is almost the same in all the cases. In addition, since the biomass has a higher LHV than the coal does, it follows that the syngas produced from this biomass should also have a higher heating value. This is shown in Table 4.4. With this being the case, the higher heating value gas will produce a higher inlet temperature when sent through the GT combustor. Since the TIT must be fixed, there are only 2 ways to counteract this: 1.) raise the oxygen (air) mass flow from the compressor, and 2.) *reduce* the syngas mass flow from the gasifier. Since the mass flow rate must also be held constant, the only choice available is to do both. Finally, since the syngas mass flow has to be reduced, the only way to accomplish this is to reduce the feedstock feed rate back at the gasifier. This, in turn, means that *a smaller, less expensive gasifier* can be used for this purpose, which is the main reason why more biomass can save on capital costs. The changes in CoE are associated with the changes in the overall efficiencies of the plants; however, in the 30% and 50% cases, both the more costly fuel price tag and the additional costs of pre-treating the new feedstock increase the CoE even though both have higher plant efficiency than the pure-coal case.

While \$4,300/kW may seem expensive at first, especially since the U.S. Dept. of Energy and the Energy Information Administration say that the average IGCC costs are around \$3,200/kW (EIA, 2010). New information and a changing economy suggest that real costs would be higher than this. In fact, a new IGCC plant in Edwardsport, Indiana recently reported a cost overrun: coming to over \$4,700/kW in capital costs (Coal Age, 2011). It is therefore believed that this set of plants is more in line with reality, particularly for 2011 dollars.

Table 4.4 Syngas compositions for subcritical plants without CCS

Case Number	A1a	A1b	A1c	A1d
Biomass/Coal Ratio	0%	10%	30%	50%
CO (vol%)	14.34	14.98	15.47	15.97
CO_2 (vol%)	9.146	8.776	8.726	8.670
CH_4 (vol%)	0.0221	0.0274	0.0299	0.0327
H_2 (vol%)	14.11	14.76	14.91	15.06
H_2S (vol%)	0.1575	0.1434	0.1142	0.0846
H_2O (vol%)	61.43	60.56	60.03	59.51
COS (vol%)	0.0052	0.0047	0.0038	0.0029
N_2 (vol%)	0.6054	0.5726	0.5374	0.5016
LHV (Btu/lb)	1653.8	1739.8	1775.2	1811.2

The emissions data for this case can be seen in Table 4.5. For this, it becomes obvious that biomass bears a significant impact on the emissions of this plant. For one, the biomass chosen, sugarcane bagasse, contains less nitrogen than lignite and less sulfur as well. Naturally, there will be less SO_x and NO_x produced as a result of this. The NO_x reduction is only slight, however, as fuel NO_x is only a portion of the total contribution to NO_x production. Gross CO_2 production, however, doesn't appear to be affected very much, either. For Gross CO_2 emissions, sugarcane actually has *higher* carbon content than the lignite coal, as shown in Chapter 3. The reduction in gross CO_2 is a result of the fact that there is less fuel needed to be used in the plant as a result of the higher plant efficiencies in the biomass blend cases. However, once the fact that biomass is carbon-neutral is taken into account, the *effective CO_2* can be found by taking off the CO_2 that originates from biomass, and it shows that adding biomass can make a significant impact on emissions, cutting the net effective CO_2 emissions by more than half in Case A1d.

Table 4.5 Emissions for subcritical plants without CCS

Case Number	A1a	A1b	A1c	A1d
Biomass/Coal Ratio	0%	10%	30%	50%
NO_x (tons/year)	234.7	232.5	232.1	231.8
SO_x (tons/year)	2,157.5	1,869.6	1,457.7	1,057.3
Gross CO_2 (tons/year)	2,110,246	2,045,916	2,042,789	2,039,757
Effective CO_2 (tons/year)	**2,110,246**	**1,824,817**	**1,388,924**	**965,167**
Gross CO_2 (tons/MW-year)	8,942.0	8,619.6	8,718.8	8,819.0
Effective CO_2 (tons/MW-year)	**8,942.0**	**7,688.1**	**5,928.1**	**4,173.0**

4.2.1.2 Post-combustion CCS (A2)

As discussed in Chapter 2, post-combustion CCS is performed after the GT has extracted power from the syngas. The CCS plant is placed at the exit of the stack of the HRSG, and, as mentioned in Chapter 3, for this case, chemical absorption is used with monoethanolamine (MEA) as the solvent of choice. The data for the post-combustion plants can be seen in Tables 4.6-4.9. The main differences between the post-combustion plant setup and the baseline case were highlighted in Chapter 3, in Figs. 3.11 and 3.12.

In Table 4.6, the total work and efficiencies are shown, just like Table 4.1 for the baseline plants. From this data, it can be seen that Post-combustion CCS has a clearly negative impact on overall plant performance. This is not surprising, since, as mentioned in Chapter 2, 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 cases. The main problem is that post-combustion CCS requires chemical absorption, as discussed in Chapter 3, 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 CO_2 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.

Table 4.6 Work and efficiency for subcritical plants with post-combustion CCS

Case Number	A2a	A2b	A2c	A2d
Biomass/Coal Ratio	0%	10%	30%	50%
Gross GT Power (kW)	200,019	200,017	200,017	200,017
Gross ST Power (kW)	70,324	72,910	73,450	74,027
Auxiliary Losses (kW)	84,409	82,668	86,098	89,434
Total Net Power (kW)	**185,934**	**190,260**	**187,369**	**184,610**
Gross Elect. Efficiency (LHV)	40.16	41.06	41.42	41.79
Net Elect. Efficiency (LHV)	**27.62**	**28.62**	**28.38**	**28.15**

For a comparison of the heat losses on carbon capture, see Table 4.7. From here, the table shows that post-combustion CCS has large heat losses, and the heat losses from CCS are easily the largest source of entropy production in the entire plant. This is another reason why the gross efficiency shown in Table 4.6 also decreased by an appreciable amount compared to that of the baseline. Meanwhile, all other heat data, such as water returned as condensate, acid gas removal heat losses, and slag production are nearly identical to what they were in the baseline cases. Again, this is because the gasification block itself is unaffected by the changes made to the bottom cycle to incorporate the CCS plant. An interesting change seen here, though, is that the heat rejection from the syngas coolers are between 8,000 and 9,000 Btu/s (about 5%) higher here than in the baseline cases. This is probably related to the efficiency loss from the CCS plant, which requires that more fuel be consumed. This implies larger mass flow rates in the syngas stream, which means more heat losses would occur from increased syngas cooling at the third syngas cooler.

Table 4.7 Parasitic energy & heat losses for subcritical plants with post-combustion CCS

Case Number	A2a	A2b	A2c	A2d
Biomass/Coal Ratio	0%	10%	30%	50%
Acid Gas Removal (Btu/s)	5,197	4,532	3,534	2,563
Syngas Water Condensed (Btu/s)	20,152	19,418	19,016	18,618
AGR heat loss (Btu/s)	611.2	533.0	415.6	301.5
Slag Production (Btu/s)	14,473	13,843	13,451	13,071
Cooler Heat Rejection (Btu/s)	176,166	164,890	159,421	154,127
Carbon Capture (Btu/s)	**231,563**	**224,713**	**224,383**	**224,064**

Table 4.8 Economics for subcritical plants with post-combustion CCS

Case Number	A2a	A2b	A2c	A2d
Biomass/Coal Ratio	0%	10%	30%	50%
Total capital cost (millions of $)	1,490.2	1,374.8	1,359.7	1,345.4
Capital Cost ($/kW)	**8,015**	**7,226**	**7,257**	**7,288**
CoE ($/kW-hr)	**0.1713**	**0.1631**	**0.1763**	**0.1895**

Of course, with the extra equipment required, and the efficiency losses, it is clear that the capital cost and CoE in the post-combustion cases rises significantly from that of the baseline. The additional mass flow needed from the gasifier translates to a *bigger, more expensive* gasifier. This coupled with the additional piping and equipment is what leads to the much larger cost (about 45% more than the baseline) seen in Table 4.8. In addition, due to the loss of power from the steam turbine, the cost of electricity rises by nearly 60% in most cases.

For all the costs and drawbacks of post-combustion CCS, the clear benefit from all of it is found in the emissions data, shown in Table 4.9. In addition to NO_x emissions being virtually eliminated and SO_x emissions being cut by more than 98%, the CO_2 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. Even Cases A1 and B1 make use of AGR, but only this set of cases and Case B2 perform sulfur removal a second time, after combustion occurs.

Table 4.9 Emissions for subcritical plants with post-combustion CCS

Case Number	A2a	A2b	A2c	A2d
Biomass/Coal Ratio	0%	10%	30%	50%
NO_x Produced (tons/year)	0.000005	0.000005	0.000005	0.000005
SO_x Produced (tons/year)	10.79	9.35	7.29	5.29
Gross CO_2 output (tons/year)	206,033	199,583	199,263	198,954
Effective CO_2 output (tons/year)	**206,033**	**-21,515**	**-454,601**	**-878,034**
Gross CO_2 (tons/MW-yr)	1,108.1	1,049.0	1,063.5	1,077.7
Effective CO_2 (tons/MW-yr)	**1,108.1**	**-113.1**	**-2,426.2**	**-4,756.2**

The reduction of CO_2 in each case is very significant, especially when neutral CO_2 is taken into account. Even as little as 10% biomass makes this plant carbon-negative. While additional biomass does not reduce the plant emissions by very much, as the two fuels used have similar carbon content, the effective CO_2 experiences drastic changes from case to case: Case A2d removes nearly four times as much CO_2 from the atmosphere as Case A2a can put out.

4.2.1.3 Sour-shift Pre-combustion CCS (A3)

Sour-shift, as mentioned in Chapter 2, means that the CO-shift reaction is performed *before* the acid gas removal stage of gas cleanup. This allows for AGR and CCS to occur simultaneously, saving on both equipment costs and energy consumption. Because this is a pre-combustion form of CCS, it allows for the use of physical absorption instead of chemical absorption. While the solvent needed is more expensive ($2000/ton), less solvent is lost between cycles, meaning the overall process is more cost-efficient than chemical absorption. The data for all sour-shift CCS plants can be seen in Tables 4.10-4.13.

Table 4.10 Work and efficiency for subcritical plants with sour-shift CCS

Case Number	A3a	A3b	A3c	A3d
Biomass/Coal Ratio	0%	10%	30%	50%
Gross GT Power (kW)	200,015	200,015	200,014	200,014
Gross ST Power (kW)	98,523	99,141	99,725	100,299
Auxiliary Losses (kW)	80,258	78,444	82,101	85,670
Total Net Power (kW)	**218,279**	**220,712**	**217,639**	**214,643**
Gross Elect. Efficiency (LHV)	41.43	42.05	42.41	42.76
Net Elect. Efficiency (LHV)	**30.29**	**31.03**	**30.79**	**30.56**

As seen in Table 4.10, plants that use sour-shift pre-combustion CCS are far more efficient than those with post-combustion CCS. Where as many as 8 percentage points of efficiency were lost from Case A1 to Case A2 (post-combustion), only 5 points were lost on average from Case A1 to Case A3 (sour-shift pre-combustion). An interesting thing to note as well is the fact that the total ST power actually *increases* for Case A3 when compared to the baseline: about 9-10MW of extra power generated on average. This may be due to the fact that the CO-shift process makes use of a catalyst to convert extra water into hydrogen for burning. Since CO_2 is removed before it reaches the gas turbine, the loss of mass flow must be made up by pushing additional syngas through the gasifier (thus increasing the gasifier size so it can accept more feedstock.) This translates to a decent supply of extra water available as condensate to be directed towards the steam cycle, while the reaction itself requires no additional water at all to go to completion: all necessary water is already present in the gas stream from the quench, the slurry feedstock, and the naturally high moisture content of the coal (Recall Fig. 3.13 in Chapter 3, where the additional water from the steam cycle in the CO-shift reaction has zero mass flow.) However, this combination of circumstances makes it so that both the TIT constraint and the mass flow constraint on the GT cannot be met at the same time. As such, the TIT condition is held, while the mass flow rate is allowed to decrease, unfortunately resulting in a higher TET.

Table 4.11 Selected parasitic energy & heat losses for subcritical plants with sour-shift CCS

Case Number	A3a	A3b	A3c	A3d
Biomass/Coal Ratio	0%	10%	30%	50%
Acid Gas Removal (Btu/s)	5,564	4,850	3,783	2,745
Syngas Water Condensed (Btu/s)	15,516	14,946	14,543	14,142
CO_2 and AGR losses (Btu/s)	**38,161**	**36,302**	**34,777**	**33,295**
Slag Production (Btu/s)	15,495	14,813	14,401	14,000
Cooler Heat Rejection (Btu/s)	156,853	143,594	137,036	130,659

As for heat losses, since CCS and AGR are performed within the same plant, their losses are combined into one, as per Table 4.11. It can be seen from this data that the combined heat losses of both processes is less than one-fifth of the heat losses of post-combustion's carbon capture system alone. While slag production and the energy provided by the steam cycle to the AGR/CCS plant increase (owing to the now larger gasifier size) from the baseline, the energy from condensed water and the heat rejection from the coolers both decrease. The reduced heat rejection from the third cooler is caused by the reduced syngas mass flow rate for the sour-shift case.

Table 4.12 Economics for subcritical plants with sour-shift CCS

Case Number	A3a	A3b	A3c	A3d
Biomass/Coal Ratio	0%	10%	30%	50%
Total capital cost (millions of $)	1,164.3	1,043.1	1,027.4	1,011.5
Capital Cost ($/kW)	**5,334**	**4,726**	**4,721**	**4,712**
CoE ($/kW-hr)	**0.1192**	**0.1146**	**0.1269**	**0.1392**

From an economic standpoint, sour-shift pre-combustion CCS is obviously superior to post-combustion CCS. All costs are universally lower, since no additional equipment is necessary for purchase: only additional catalyst in the COS-shift reactor (which now has to perform CO-shift as well).

Table 4.13 Emissions for subcritical plants with sour-shift CCS

Case Number	A3a	A3b	A3c	A3d
Biomass/Coal Ratio	0%	10%	30%	50%
NO_x Produced (tons/year)	183.8	183.5	183.5	183.5
SO_x Produced (tons/year)	2,309.9	2,000.7	1,560.5	1,132.3
Gross CO_2 output (tons/year)	233,739	229,544	233,410	237,198
Effective CO_2 output (tons/year)	**233,739**	**-7,057.7**	**-466,486**	**-913,578**
Gross CO_2 (tons/MW-yr)	1,070.8	1,040.0	1,072.5	1,105.1
Effective CO_2 (tons/MW-yr)	**1,070.8**	**-32.0**	**-2,143.4**	**-4,256.3**

Finally, the emissions data for Case A3 can be seen in Table 4.13. While this pre-combustion CCS system does not seem to be as effective at removing CO_2 directly as the post-combustion system in Case A2 is, based on total (gross) tons/year of CO2 emission (Table 4.9 vs. Table 4.13), the greater power output from this set of plants evens the two plants out in CO_2 per MW-yr. For effective CO_2, however, sour-shift benefits more from having biomass in the feedstock, and in raw tons/year, there are far fewer CO_2 emissions than in post-combustion for 30% and 50% BMR. On a per MW-year basis, only Case A3a (pure coal) has marginally lower effective CO_2 emissions than the pure-coal post-combustion case (Case A2); all biomass cases have lower negative CO_2 emissions than post-combustion cases. This is due to the increased gasifier size, as, again, the mass flow to the GT cannot be maintained at the same power output without adding additional syngas mass flow. This can only be accomplished by a larger gasifier. Therefore, more CO_2 is being added due to simply having more carbon available from the beginning. Even with all this taken into account, post-combustion CCS retains one advantage over sour-shift pre-combustion CCS: handling SO_x and NO_x. Only post-combustion's chemical absorption can process SO_x and NO_x, and only because it occurs after those compounds are able to form. However, sour-shift is cheaper and easier to implement in the case of new plants.

4.2.1.4 Sweet-shift Pre-combustion CCS (A4)

Finally, for the subcritical cases, the last main case is for sweet-shift pre-combustion CCS. Unlike sour-shift, as mentioned in Chapter 2, sweet-shift occurs *after* acid gas removal, so a new shift reactor and a separate carbon capture plant must be purchased, meaning sweet-shift is always more expensive than sour-shift. In this simulation, the sweet-shift plant makes use of the same solvent as the sour-shift plant, and all settings and controls on the CCS block are the same as those on the AGR/CCS block in the sour-shift plant. All data on this case can be seen in Tables 4.14-1.17.

The data in Table 4.14 shows that the sweet-shift process does *not* increase the total power of this plant like sour-shift does from the baseline case (Case A1a in Table 4.1). In fact, it actually consistently decreases the total power output by about 11MW in all instances. This is due to the fact that, unlike sour-shift, sweet shift requires additional steam input directly from the steam cycle directly, resulting in a reduction of steam turbine output of approximately 11MW. Since it occurs after every other process in the gas cleanup system, the amount of water needed is largely independent of BMR. But, in the long run, this is still enough to cut the efficiency by about 8 percentage points: even more than in post-combustion CCS.

For the heat loss data in Table 4.15, it shows that sweet-shift, in addition to taking energy from the steam cycle directly like in post-combustion CCS, it also experiences more heat losses than sour-shift does (Table 4.15 vs. Table 4.11), meaning it is thermally inferior to sour-shift CCS. Second, since the gasifier has to be resized for the same reasons as sour-shift, but the CO-shift reaction doesn't occur at the gasifier exit, more water is wasted from this process, and the coolers have to reject much more heat output in this case than in Case A3. This may not seem like much, but then it becomes apparent that CCS requires high-quality steam in order to work, so the only place to take the additional water is from an HP stream somewhere (recall Fig. 3.17), which, again, will contribute to a severe reduction in total ST power output.

Table 4.14 Work and efficiency for subcritical plants with sweet-shift CCS

Case Number	A4a	A4b	A4c	A4d
Biomass/Coal Ratio	0%	10%	30%	50%
Gross GT Power (kW)	200,015	200,015	200,014	200,015
Gross ST Power (kW)	78,509	78,861	79,202	79,518
Auxiliary Losses (kW)	80,404	78,586	82,310	85,934
Total Net Power (kW)	**198,120**	**200,290**	**196,906**	**193,598**
Gross Elect. Efficiency (LHV)	38.42	38.98	39.27	39.56
Net Elect. Efficiency (LHV)	**27.33**	**27.99**	**27.70**	**27.40**

Table 4.15 Selected parasitic energy & heat losses for subcritical plants with sweet-shift CCS

Case Number	A4a	A4b	A4c	A4d
Biomass/Coal Ratio	0%	10%	30%	50%
Acid Gas Removal (Btu/s)	5,597	4,878	3,805	2,761.3
Syngas Water Condensed (Btu/s)	23,699	22,888	22,502	22,126
CO_2 and AGR losses (Btu/s)	**55,245**	**53,100**	**51,599**	**50,139**
Slag Production (Btu/s)	15,588	14,898	14,485	14,083
Cooler Heat Rejection (Btu/s)	182,017	169,697	163,716	157,897

Also, the same problem with the gas turbine occurs in this case as happened in Case A3: both the TIT and the mass flow rate cannot be held constant. As such, again, the TIT was fixed while the mass flow was allowed to change. Predictably, the TET increased as a result, reducing GT efficiency.

Table 4.16 Economics for subcritical plants with sweet-shift CCS

Case Number	A4a	A4b	A4c	A4d
Biomass/Coal Ratio	0%	10%	30%	50%
Total capital cost (millions of $)	1,181.7	1,059.9	1,044.2	1,028.2
Capital Cost ($/kW)	**5,964**	**5,292**	**5,303**	**5,311**
CoE ($/kW-hr)	**0.1316**	**0.1264**	**0.1405**	**0.1547**

From Table 4.16, it is easy to see that the plant in Case A4 (sweet-shift) is universally more expensive overall than that of Case A3 (sour-shift in Table 4.12), but less than Post-combustion CCS of Case A2 in Table 4.8. (Sweet-shift has about $20 million more in capital and 1-2 cents higher CoE than sour shift, but about $300 million less in capital and 3-4 cents cheaper CoE than post-combustion.) This is not surprising because, as mentioned in both Chapters 2 and 3, the sour-shift process can be integrated directly into existing devices, while sweet-shift requires the purchase of completely new, separate units. Compared to post-combustion CCS, which directly interferes with the steam cycle by removing water from an IP stream, sweet-shift requires an HP stream due to its location in the cycle. However, post-combustion CCS also burdens the plant with additional auxiliary losses, which directly translate to more money paid in electric bills. The post-combustion system in Case A2 is also larger, as it processes the gases at atmospheric pressure, which makes for a more costly device. With this in mind, it becomes easier to see how a slightly more efficient design can still cost significantly more money (nearly $2000/kW capital and over $0.04/kW-hr CoE).

Table 4.17 Emissions for subcritical plants with sweet-shift CCS

Case Number	A4a	A4b	A4c	A4d
Biomass/Coal Ratio	0%	10%	30%	50%
NO_x Produced (tons/year)	185.6	185.2	185.0	184.9
SO_x Produced (tons/year)	2,323.8	2,012.2	1,569.7	1,139.1
Gross CO_2 output (tons/year)	254,416	247,667	247,882	248,120
Effective CO_2 output (tons/year)	**254,416**	**9,733**	**-456,169**	**-909,582**
Gross CO_2 (tons/MW-yr)	1,284.2	1,236.5	1,258.9	1,281.7
Effective CO_2 (tons/MW-yr)	**1,284.2**	**48.6**	**-2,316.7**	**-4,698.3**

From Tables 4.13 and 4.17, it can be seen that there is no appreciable difference between sweet- and sour-shift CCS in terms of emissions performance although sweet-shift is slightly worse. While the sweet-shift system in this case (Case A4a) is not carbon-negative like in Case A3, the difference in CO_2 emissions is so small that this can be attributed to round-off error or other numerical issues. The only other anomaly is that the gross CO_2 output in Case A4a is 20,000 tons/year greater than that in Case A3a. This is a result of a larger gasifier, the same issue that arose when comparing Case A2 with Case A3. This can be verified by the facts that there is slightly more SO_x and NO_x emitted and more slag is produced. Both of these facts directly point to a larger gasifier with a higher fuel mass flow rate due to the controls placed upon the system components.

4.2.2 Supercritical Cycles (Group B)

The supercritical steam system was created by modifying the subcritical system such that the TIT and TIP of the steam turbine were raised to 1200°F (648.9°C) and 2400PSI, respectively. The overall HRSG layout remains unchanged, aside from several pressure and temperature increases made at each junction to make way for the higher needed inlet temperature and pressure. The reheat section was also adjusted so that the inlet temperature of the second ST casing could be raised to 1100°F (594.4°C). Finally, in order to use the new TIT specified, the HRSG required the use of a duct burner, as seen in Fig. 3.10 back in Chapter 3. The duct burner was designed in order for the exit temperature of the exhaust gases to be fixed at 1220°F (660°C).

4.2.2.1 No CCS (Case B1)

The important data for Case B1 is shown in Tables 4.18-4.22. Like Case A1, this is mainly to serve as a basis for comparison to the other cases in this block, as well as to show the differences in performance and price between this baseline and the subcritical baseline, so as to obtain a more complete picture of how a supercritical steam cycle affects an IGCC plant.

As seen in Table 4.18, 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. From this, it is clear that using a supercritical cycle provides an overall efficiency benefit for an IGCC plant as a whole.

Table 4.18 Work and efficiency for supercritical plants without CCS

Case Number	B1a	B1b	B1c	B1d
Biomass/Coal Ratio	0%	10%	30%	50%
Gross GT Power (kW)	200,019	200,017	200,017	200,017
Gross ST Power (kW)	122,573	122,602	122,946	123,262
Auxiliary Losses (kW)	55,481	54,413	57,873	61,235
Total Net Power (kW)	**267,111**	**268,207**	**265,090**	**262,043**
Gross Elect. Efficiency (LHV)	44.29	44.84	45.18	45.52
Net Elect. Efficiency (LHV)	**36.67**	**37.28**	**37.08**	**36.89**

Table 4.19 Syngas compositions for supercritical plants without CCS

Case Number	B1a	B1b	B1c	B1d
Biomass/Coal Ratio	0%	10%	30%	50%
CO (vol%)	14.32	14.98	15.47	15.97
CO_2 (vol%)	9.147	8.776	8.726	8.670
CH_4 (vol%)	0.0221	0.0274	0.0299	0.0327
H_2 (vol%)	14.11	14.76	14.91	15.06
H_2S (vol%)	0.1575	0.1435	0.1142	0.0846
H_2O (vol%)	61.42	60.55	60.03	59.51
COS (vol%)	0.0052	0.0047	0.0038	0.0029
N_2 (vol%)	0.6054	0.5727	0.5374	0.5016
LHV (Btu/lb)	1653.9	1739.8	1775.3	1811.2

Looking at Table 4.19, it appears as though there is not much of a difference between Case A1 and Case B1 in terms of syngas composition. It is thus clear here that the production of syngas is independent of the steam cycle configuration. From Table 4.20, it appears that there is virtually no change in the heat loss data from the gas cleanup system, either. In other words, BMR and the steam cycle type are two independent parameters, and share no noticeable synergistic effects whatsoever. Like with Fig. 4.1 for the subcritical baseline, Fig. 4.2 shows a more qualitative representation of the heat balance data for Case B1a.

Table 4.20 Selected parasitic energy & heat losses for supercritical plants without CCS

Case Number	B1a	B1b	B1c	B1d
Biomass/Coal Ratio	0%	10%	30%	50%
Acid Gas Removal (Btu/s)	5,197	4,532	3,534	2,563
Syngas Water Condensed (Btu/s)	23,286	22,402	21,979	21,569
AGR heat loss (Btu/s)	611.2	533.1	415.6	301.5
Slag Production (Btu/s)	14,473	13,843	13,452	13,072
Cooler Heat Rejection (Btu/s)	164,026	153,195	147,746	142,441

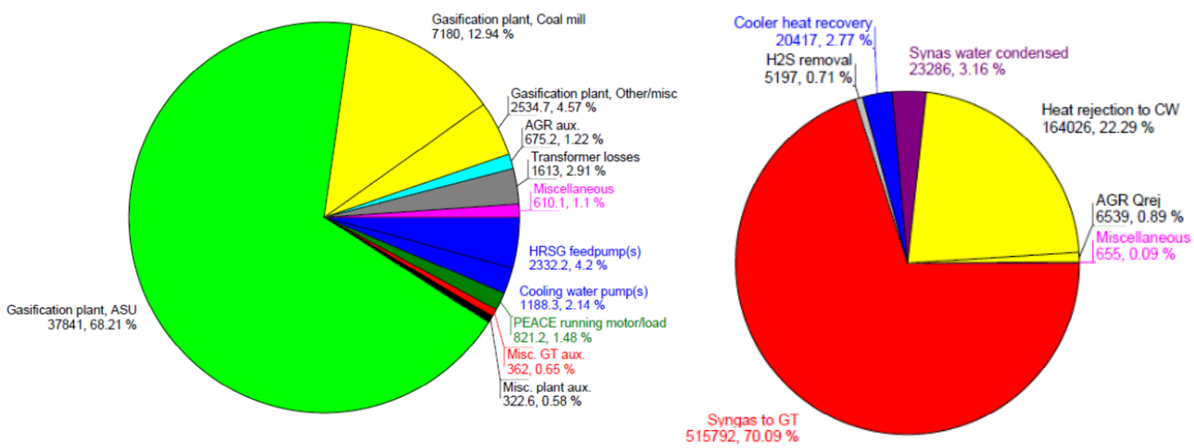

Figure 4.2 Auxiliaries & losses (left) and gas cleanup energy out (right) for Case B1a

Table 4.21 Economics for supercritical plants without CCS

Case Number	B1a	B1b	B1c	B1d
Biomass/Coal Ratio	0%	10%	30%	50%
Total capital cost (millions of $)	1,087.58	983.83	970.95	956.03
Capital cost ($/kW)	**4,072**	**3,668**	**3,663**	**3,648**
CoE ($/kW-hr)	**0.0972**	**0.0947**	**0.1041**	**0.1133**

From an economic standpoint (see Table 4.21), the supercritical cycle is approximately 10% superior to the subcritical cycle. While the overall capital cost is greater (as it should be, supercritical turbines and pipings are always more expensive than subcritical ones), the biggest expenditures are actually labor and labor interest: about $19 million more total. The next biggest cost is related to its installation and setting in concrete: about $12 million more. Then comes owner's soft costs (like fees and permits): $11 million more. And, finally, piping: $9-10 million more. The extra power gained from implementing the supercritical system more than circumvents this: <u>the capital cost per unit power is about 9% ($240-290/MW) less than in Case A1</u> (Table 4.21 vs. Table 4.3). Second, because of the additional power, <u>the CoE is also reduced by almost half of a cent per kW-hr (10%) in some cases</u>.

Table 4.22 Emissions for supercritical plants without CCS

Case Number	B1a	B1b	B1c	B1d
Biomass/Coal Ratio	0%	10%	30%	50%
NO_x (tons/year)	526.0	521.0	520.2	518.4
SO_x (tons/year)	2,157.6	1,868.7	1,457.7	1,057.3
Gross CO_2 (tons/year)	2,220,582	2,141,202	2,137,838	2,134,566
Eff. CO_2 (tons/year)	**2,220,582**	**1,920,103**	**1,483,973**	**1,059,975**
Gross CO_2 (tons/MW-year)	8,313.3	7,983.4	8,064.6	8,145.9
Eff. CO_2 (tons/MW-year)	**8,313.3**	**7,159.0**	**5,598.0**	**4,045.0**

Lastly, for the emissions data, see Table 4.21. This shows that the emissions of NO_x and SO_x for Case B1 are identical to those of Case A1 (Table 4.5), further emphasizing the fact that BMR and the steam cycle type are mutually independent parameters. However, the gross CO_2 output of Case B1 is about 100,000 tons/year greater than that of Case A1 and the NO_x emissions are about 200 tons/year greater as well. This is because of the presence of the duct burner, which burns natural gas and contributes to the total emissions of the plant as a whole. However, <u>due to the increased power output of the supercritical plant, the emissions per unit energy (MW-year) have clearly decreased</u>: 7% for pure-coal case, 9% for 10% BMR, and 5% for more than 10% BMR.

4.2.2.2 Post-combustion CCS (Case B2)

Post-combustion CCS for the supercritical cycle is exactly the same as that of the subcritical cycle. Since the overall layout of both plants is the same, all connections and modifications made from the subcritical system are carried over to the supercritical system. The only exception is a steam source is added to the HP stream to make up for the water loss consumed by the CCS plant: the amount of steam needed had to be increased to keep the same mass flow rates in the HP stream. All data for the supercritical post-combustion cases is shown in Tables 4.23-4.26.

Table 4.23 Work and efficiency for supercritical plants with post-combustion CCS

Case Number	B2a	B2b	B2c	B2d
Biomass/Coal Ratio	0%	10%	30%	50%
Gross GT Power (kW)	200,019	200,017	200,017	200,017
Gross ST Power (kW)	93,207	94,682	95,173	95,638
Auxiliary Losses (kW)	86,731	84,935	88,368	91,702
Total Net Power (kW)	**206,495**	**209,765**	**206,822**	**203,954**
Gross Elect. Efficiency (LHV)	40.93	41.65	41.99	42.33
Net Elect. Efficiency (LHV)	**28.82**	**29.64**	**29.42**	**29.20**

Table 4.23 shows the work and efficiency data for case B2. From this, it shows that post-combustion CCS also has a very detrimental effect on the performance of the supercritical cycle, as not even the high quality steam and additional power can make up for the loss of efficiency from this form of CCS. 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 (Table 4.24 vs. Table 4.6). The average efficiency losses work out to be around the same 8 percentage points, but the loss of power from this cycle very nearly makes it not worth using.

From Table 4.24, it appears as though the entire gas cleanup system downstream of the gasifier is, again, unaffected by the supercritical cycle. However, one object of note is the fact that the amount of energy necessary for carbon capture is universally about 8,000 Btu/s more costly for Case B2 than it is for Case A2 (Table 4.24 vs. Table 4.7). The reason for this will be explained later on by the emissions data.

Table 4.24 Parasitic energy & heat losses for supercritical plants with post-combustion CCS

Case Number	B2a	B2b	B2c	B2d
Biomass/Coal Ratio	0%	10%	30%	50%
Acid Gas Removal (Btu/s)	5,197	4,532	3,534	2,563
Syngas Water Condensed (Btu/s)	20,903	20,049	19,641	19,244
AGR heat loss (Btu/s)	611.2	533.1	415.6	301.5
Slag Production (Btu/s)	14,473	13,843	13,452	13,072
Cooler Heat Rejection (Btu/s)	173,281	162,448	156,981	151,683
Carbon Capture (Btu/s)	**239,587**	**232,631**	**232,300**	**231,965**

For economics, Case B2 has the same exact relationship to Case A2 as Case B1 did to Case A1, for the exact same reasons. However, the new system's CoE is nearly a full cent cheaper than before with CCS (B2 vs. A2), as compared to the half of a cent from earlier without CCS (B1 vs. A1.) Nevertheless, this system with CCS is still remarkably costly, and may be completely unviable overall, just like in Case A2. What this does show is that using a higher quality supercritical steam plant can help circumvent some of the costs associated with this type of carbon capture. The data for all of this is shown in Table 4.25.

Table 4.25 Economics for supercritical plants with post-combustion CCS

Case Number	B2a	B2b	B2c	B2d
Biomass/Coal Ratio	0%	10%	30%	50%
Total capital cost (millions of $)	1,539.8	1,422.5	1,407.4	1,392.5
Capital Cost ($/kW)	**7,457**	**6,781**	**6,805**	**6,828**
CoE ($/kW-hr)	**0.1626**	**0.1559**	**0.1678**	**0.1797**

Table 4.26 Emissions for supercritical plants with post-combustion CCS

Case Number	B2a	B2b	B2c	B2d
Biomass/Coal Ratio	0%	10%	30%	50%
NO_x Produced (tons/year)	0.000006	0.000006	0.000006	0.000006
SO_x Produced (tons/year)	10.79	9.35	7.29	5.29
Gross CO_2 output (tons/year)	213,589	207,036	206,715	206,392
Effective CO_2 output (tons/year)	**213,589**	**-14,062**	**-447,149.5**	**-868,198**
Gross CO_2 (tons/MW-yr)	1,034.4	987.0	999.5	1,012.0
Effective CO_2 (tons/MW-yr)	**1,034.4**	**-67.0**	**-2,162.0**	**-4,356.8**

Lastly, for the emissions, see Table 4.26. Case B2 and Case A2 are nearly identical, and the only clear difference between them is the presence of an extra 8,000 tons/year of CO_2. This comes from the duct burner, just like in Case B1 when compared to Case A1 without CCS. The amount of extra CO_2 from the duct burner is picked up at the end by the CCS system. This shows another clear advantage of post-combustion CCS, since it is actually capable of cleaning the emissions from the duct burner. This is also the reason for the additional carbon capture energy consumption seen in Table 4.24, as processing these extra emissions will, of course, cost additional energy. The effective CO_2 emissions are reduced by approximately 40 – 400 tons/MW-year between the subcritical post-combustion and supercritical post-combustion cases (Table 4.26 of Case B2 vs. Table 4.9 of Case A2) except the 10% BMR case, which produces about 40 ton/MW-year less than the corresponding subcritical case (Case B2b vs. Case A2b).

4.2.2.3 Sour-shift Pre-combustion CCS (B3)

When sour-shift pre-combustion CCS was introduced to the supercritical cycle, it was done in the same manner as was done previously for the subcritical cycle. Again, physical absorption was used, and the same carbon capture and AGR efficiencies were used. The data for this cycle is shown in Tables 4.27-4.30.

Table 4.27 Work and efficiency for supercritical plants with sour-shift CCS

Case Number	B3a	B3b	B3c	B3d
Biomass/Coal Ratio	0%	10%	30%	50%
Gross GT Power (kW)	200,014	200,014	200,014	200,014
Gross ST Power (kW)	119,089	120,838	122,350	123,868
Auxiliary Losses (kW)	81,026	79,293	82,977	86,573
Total Net Power (kW)	**238,077**	**241,559**	**239,387**	**237,309**
Gross Elect. Efficiency (LHV)	42.03	42.72	43.12	43.52
Net Elect. Efficiency (LHV)	**31.36**	**32.16**	**32.02**	**31.89**

From the work and efficiency data shown in Table 4.27, the sour-shift system behaves differently for the supercritical cycle than it did for the subcritical cycle. The most obvious change here is that the supercritical cycle *loses* steam power from this case. Where in the subcritical system, sour-shift *increases* the total steam power by about 10MW (Case A3 vs. Case 1 or Table 4.10 vs. Table 4.1), in this case, it *decreases* the power by at least 3 MW (Case B3 vs. B1 or Table 4.27 vs. Table 4.18). This change is most likely caused by the fact that the gasification system did not change with the steam cycle, so the quality of water given to the steam cycle in Case A3 remains the same, while the quality of steam taken from the HP stream is much higher for Case B3. This means that the additional water supplied here is not enough to make up for the direct loss of power from sacrificing such high grade steam, whereas in Case A3, it was a much better trade. This is also why the efficiency doesn't increase as much between Cases A3 and B3 as it did for the earlier supercritical cases when compared to their subcritical counterparts (Case A1 vs. B1, and Case A2 vs. B2.): only about 1.0-1.3 percentage points compared to the previous 1.6 percentage points.

Table 4.28 Selected parasitic energy & heat losses for supercritical plants with sour-shift CCS

Case Number	B3a	B3b	B3c	B3d
Biomass/Coal Ratio	0%	10%	30%	50%
Acid Gas Removal (Btu/s)	5,500	4,799	3,747	2,721
Syngas Water Condensed (Btu/s)	10,448	10,741	10,931	11,089
CO_2 and AGR heat loss (Btu/s)	**37,719**	**35,913**	**34,435**	**32,997**
Slag Production (Btu/s)	15,319	14,658	14,262	13,878
Cooler Heat Rejection (Btu/s)	175,250	159,150	150,544	142,215

From the heat loss data in Table 4.28, it appears that the supercritical plant actually has *fewer* heat losses than the subcritical plant (B3 vs. A3 or Table 4.28 vs. Table 4.11). One very peculiar point is that the energy loss from slag production actually decreases in this case, which leads one to suspect that the gasifier size is actually *smaller* than in Case A3. It is still clearly larger than the previous supercritical cases (Cases B1 and B2), but it may have actually needed to downsize compared to Case A3. Indeed, the fuel mass flow rate for Case A3a's gasifier is 4,612 tons/day, while the fuel mass flow rate for Case B3a's gasifier is 4,560 tons/day. While it is not clear at first where this comes from, looking back at Chapter 3, for Case A3a, the steam from the steam cycle used to perform the CO-shift reaction enters with a mass flow rate of about 26 lbs/s. In Case B3a, however, the steam is of higher grade, so only 5 lbs/s is needed. But, due to the higher steam temperature, the exit temperature of the CO-shift reactor is much higher. This leads to significantly higher temperatures throughout the gas cleanup system. And, while the coolers are able to reduce this temperature to the required 100°F, by the time CCS and AGR are to be performed, the recovery stage cooler (Cooler 1) regenerates the syngas stream to a much higher temperature (650°F in Case B3a compared to 490°F in Case A3a). This means that to get the same gas turbine inlet temperature, less fuel will need to be burned to reach it. Therefore, the syngas mass flow must be reduced, and the only way to do that while maintaining the same controls on the overall system is to decrease the feed rate of the input fuel (thus reducing the size of the gasifier). This would also explain all of the reduced energy losses in the rest of the gas cleanup system devices.

The economic data in Table 4.29 shows that the supercritical system affects the overall plant operation in the same manner as it has for both of the previous cases (Cases B1 and B2): a clear reduction in CoE and an increase in capital cost that is offset by the increased power output, resulting in a decrease in capital cost per unit power. Like with the subcritical cases, implementing sour-shift CCS is much less expensive than using post-combustion CCS: by about $2200/kW capital and about 5 cents/kW-hr CoE on average (Case B3, Table 4.29 vs. Case B2, Table 4.25).

Table 4.29 Economics for supercritical plants with sour-shift CCS

Case Number	B3a	B3b	B3c	B3d
Biomass/Coal Ratio	0%	10%	30%	50%
Total capital cost (millions of $)	1,206.2	1,086.8	1,072.3	1,060.0
Capital Cost ($/kW)	**5,066**	**4,499**	**4,479**	**4,467**
CoE ($/kW-hr)	**0.1159**	**0.1114**	**0.1222**	**0.1331**

Table 4.30 Emissions for supercritical plants with sour-shift CCS

Case Number	B3a	B3b	B3c	B3d
Biomass/Coal Ratio	0%	10%	30%	50%
NO_x Produced (tons/year)	409.4	409.9	409.9	410.4
SO_x Produced (tons/year)	2,283.6	1,979.7	1,545.5	1,122.5
Gross CO_2 output (tons/year)	312,449	309,251	313,917	318,549
Effective CO_2 output (tons/year)	**312,449**	**75,153**	**-397,266**	**-822,104**
Gross CO_2 (tons/MW-yr)	1,312.4	1,280.2	1,311.3	1,342.3
Effective CO_2 (tons/MW-yr)	**1,312.4**	**311.1**	**-1,584.3**	**-3,464.3**

From Table 4.30, the CO_2 emissions increased by about 80,000 tons/year. This is the same thing that happened between Cases A1 and B1: it is a result of the duct burner's CO_2 output. Unlike in post-combustion CCS, the emissions for the duct burner cannot be cleaned when using pre-combustion CCS, as the emissions are sent through the HRSG with the flue gases. But, since the TET is higher, the duct burner ends up burning less fuel to reach the desired temperature, which is why the DB's gross CO_2 emissions are about 20,000 tons/year lower in this case compared to Case A3. The disadvantage here is that the supercritical system does not become carbon negative at Case B3b: only Cases B3c and B3d are carbon-negative. This is a direct result of the additional CO_2 from the duct burner. In addition, the SO_x emissions decreased from subcritical Case A3, which is further evidence of the reduction in gasifier size mentioned earlier compared to that case.

4.2.2.4 Sweet-shift Pre-combustion CCS (B4)

Finally, Case B4 represents sweet-shift pre-combustion CCS used with the supercritical steam cycle. The CCS system and AGR plant used are the same as the ones in Case A4, and the methodology for comparison to Case B3 is the same as that used between Case A3 and A4. That is, the controls in place for the CCS system are the same as those for the AGR/CCS system in the sour-shift plant, and the CO-shift reactor has the same stipulations attached to it that the CO/COS-shift reactor did in the sour-shift case. All data on this specific case can be seen in Tables 4.31-4.34.

Table 4.31 Work and efficiency for supercritical plants with sweet-shift CCS

Case Number	B4a	B4b	B4c	B4d
Biomass/Coal Ratio	0%	10%	30%	50%
Gross GT Power (kW)	200,015	200,015	200,014	200,015
Gross ST Power (kW)	107,704	108,015	108,336	108,651
Auxiliary Losses (kW)	82,313	80,494	84,221	87,847
Total Net Power (kW)	**225,406**	**227,536**	**224,130**	**220,819**
Gross Elect. Efficiency (LHV)	39.75	40.29	40.57	40.85
Net Elect. Efficiency (LHV)	**29.12**	**29.76**	**29.49**	**29.22**

The work and efficiency data in Table 4.31 shows that the supercritical cycle benefits sweet-shift CCS more than any other case set: about 1.8-1.9 percentage points of improvement from Case A4. Similar to Case B3, the steam cycle also loses about 20MW of power when compared with Case B1, still a great deal more than the loss (3MW) from Case B3. Interestingly enough, the auxiliary losses are not that much higher than those of Case B3, so the total net power is only about 10MW lower on average than that of Case B3.

The data on the heat losses can be seen in Table 4.32. This shows that, just like in Case A4, the gasifier block has increased in size in order to increase the mass flow headed to the GT. This is most readily observed by the increased slag production, but the greater energy spent on acid gas removal and condensing water from the syngas stream are also good indicators of an increase in gasifier size. Also, like in Case A4, the CO_2 capture and AGR combined losses are about 20 Btu/s larger than those in the corresponding supercritical sour-shift case (Case B3).

Table 4.32 Selected parasitic energy & heat losses for supercritical plants with sweet-shift CCS

Case Number	B4a	B4b	B4c	B4d
Biomass/Coal Ratio	0%	10%	30%	50%
Acid Gas Removal (Btu/s)	5,597	4,878	3,805	2,761
Syngas Water Condensed (Btu/s)	24,176	23,354	22,961	22,579
CO_2 and AGR heat loss (Btu/s)	**55,246**	**53,102**	**51,601**	**50,141**
Slag Production (Btu/s)	15,588	14,899	14,485	14,083
Cooler Heat Rejection (Btu/s)	180,150	167,851	161,883	156,080

Table 4.33 Economics for supercritical plants with sweet-shift CCS

Case Number	B4a	B4b	B4c	B4d
Biomass/Coal Ratio	0%	10%	30%	50%
Total capital cost (millions of $)	1,241.0	1,119.2	1,103.4	1,087.5
Capital Cost ($/kW)	**5,506**	**4,919**	**4,923**	**4,925**
CoE ($/kW-hr)	**0.1248**	**0.1203**	**0.1326**	**0.1449**

For economics, the supercritical cycle has the exact same effect that it has had on every case so far: the total capital cost increases overall, but the extra power reduces the cost per kW, resulting in a net decrease in CoE, between 0.6 and 0.7 cents in this case, from the subcritical case (Case B4 vs. Case A4). All of this data can be seen in Table 4.33. And, in terms of overall cost, sweet-shift is still the second most expensive type of CCS to implement: about $500/kW more expensive than sour-shift (Case B4 vs. Case B3).

The emissions data for Case B4 is shown in Table 4.34. This shows that the supercritical plant has about the same effect on emissions as it has previously. All of the total emissions compared to Case A4 have increased by about 80,000 tons/year as a result of the duct burner. Again, the 20,000 tons/year lower CO_2 emissions come from the higher exit temperature of the gas turbine, as a result of the reduced mass flow through it. The SO_x and NO_x emissions, however, remain unchanged from Case A4. Lastly, like in all of the pre-combustion supercritical plants, the 10% BMR (B4b here) is not carbon-negative. And, again, there appears to be no appreciable difference in terms of performance between sweet- and sour-shift CCS. The SO_x emissions are greater in this case due to the larger gasifier size compared to Case B3, and the overall gross CO_2 emissions are greater for the same reasons. The larger gasifier size would also explain why the differences in effective CO_2 get smaller as more biomass is taken into account, eventually overtaking the sour-shift plant in terms of carbon-negativity (Case B4d vs. Case B3d).

Table 4.34 Emissions for supercritical plants with sweet-shift CCS

Case Number	B4a	B4b	B4c	B4d
Biomass/Coal Ratio	0%	10%	30%	50%
NO_x Produced (tons/year)	419.0	418.0	417.7	417.4
SO_x Produced (tons/year)	2,323.8	2,012.3	1,569.7	1,139.1
Gross CO_2 output (tons/year)	339,982	333,039	333,242	333,472
Effective CO_2 output (tons/year)	**339,982**	**95,105**	**-370,809**	**-824,230**
Gross CO_2 (tons/MW-yr)	1,508.3	1,463.7	1,486.8	1,510.2
Effective CO_2 (tons/MW-yr)	**1,508.3**	**418.0**	**-1,654.4**	**-3,732.6**

4.3 The Special Cases

What follows is a group of cases designed to track a single parameter, as discussed in Chapter 3. These so-called "special" cases are called as such because they cannot be broken down into sub-cases and integrated into the main case set: they only track one parameter and only in one specific instance. These cases, again, mainly serve as a source of comparison for some of the design parameters chosen at the beginning of this study.

4.3.1 Case S1: Radiant & Convective Syngas Coolers (vs. Case A1b)

The first special case deals with the use of syngas coolers to prepare the syngas for cleaning rather than using a quench based on Case A1b (subcritical, 10% BMR, and no CCS). 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. As mentioned in Chapter 3, both a radiant cooler and a convective cooler were used, and the output temperature of the syngas was set at 500°F (260°C). All data analyzed for this case can be seen in Tables 4.35-4.38.

Table 4.35 shows the work and efficiency data for this special case. This shows that the efficiency significantly increases about 5.7 percentage point (16%) and 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 of heat from the gasifier is significantly downgraded to a low temperature.

Coolers on the other hand, transfer the heat more effectively to the steam cycle, producing higher-grade (i.e., higher exergy) steam, which can be utilized more efficiently. Thus, with the right integration into the HRSG, it is possible to achieve HP steam temperatures of 1500-1600°F at this gasifier temperature. However, the TIT of the steam turbine is fixed based on the design specifications, mainly due to the metallurgical constraints of blade materials. However, in this study, the ST mass flow rate is not fixed, and it is still possible to transfer this energy away by simply raising the mass flow rates through the coolers. Doing this, however requires the mass flow rates throughout the steam system to increase along with them, which leads to bigger pumps and more auxiliary losses (also seen in the table). However, the larger mass flow rate through the steam turbine results in a bigger steam turbine and thus more work output. In the end, the net efficiency from all of this rises by nearly 6 percentage points.

Table 4.35 Work and efficiency for syngas coolers vs. syngas quench

Case	Coolers (S1)	Quench (A1b)
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**

The heat loss data in Table 4.36 shows that the coolers with a heat rejection of 25,094 Btu/s are much better at conserving energy than the quench, which have a heat rejection of 156,460 Btu/s. Because no water is added to the syngas stream, there is also much less heat lost from removing condensate.

Table 4.36 Selected parasitic energy & heat losses for syngas coolers vs. syngas quench

Case	Coolers (S1)	Quench (A1b)
Acid Gas Removal (Btu/s)	4,554	4,532
Syngas Water Condensed (Btu/s)	11,400	21,584
AGR heat loss (Btu/s)	535.7	533.0
Slag Production (Btu/s)	12,847	13,843
Cooler Heat Rejection (Btu/s)	25,094	156,460

While there is less heat loss from slag production, this does not mean that less slag is produced in this case. In fact, slag production *increases* for the coolers due to the increased gasifier size. Less heat is lost because the slag flows past the coolers during its trip to the bottom of the gasifier (which is how GE's gasifier is designed). In other words, some of the heat from the slag is given to the steam cycle in Case S1, as opposed to Case A1b, where the excess heat is completely discarded. However, the much purer, drier syngas is slightly more costly to clean up in regards to sulfur removal due to the higher concentrations of H_2S and COS.

Table 4.37 Economics for syngas coolers vs. syngas quench

Case	Coolers (S1)	Quench (A1b)
Total capital cost (millions of $)	1,223.4	926.74
Capital cost ($/kW)	**4,441**	**3,904**
CoE ($/kW-hr)	**0.1044**	**0.0979**

The economic data, shown in Table 4.37, shows that employing coolers is about $540/kW (12%) in capital cost and $0.007/kW-hr (7%) in CoE more expensive than the quench process. In addition to the coolers themselves, Case S1 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 still be better than the quench under the right conditions. <u>A deeper, more involved study is necessary to determine for certain which plant is a better investment for different conditions.</u>

Table 4.38 Emissions data for syngas coolers vs. syngas quench

Case	Coolers (S1)	Quench (A1b)
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 4.38. This shows that the system with the coolers actually produce more total emissions in gross tons than the quench system. The NO_x and SO_x emissions are comparable, but <u>the gross and effective CO_2 emissions per unit energy production are 1100 tons/MW-year (13.5%) lower for the coolers due to the much larger power output</u> (About 30 MW, or 13% more total net power).

4.3.2 Cases S2 & S3: Dry-fed Gasifier (vs. Cases A3c and A4c)

For the dry-fed system, the Shell gasifier was used as the base model, as mentioned in Chapter 3. For this system, two cases are considered, both with pre-combustion CCS: one for sour-shift (S2) and one for sweet-shift (S3). This was mainly to highlight the main differences in operation between dry-fed and slurry-fed systems in regards to CCS. Li and Wang (2009) showed that dry-fed systems were more suited to sweet-shift CCS, while insinuating that slurry-fed systems would be more suited to sour-shift CCS. All data for this part of the study is shown in Tables 4.39-4.42. Again, the Shell gasifier uses internal gasifier cooling via a wall membrane, but a quench was 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 4.39 Work and efficiency for dry-fed vs. slurry-fed systems

Case	Dry-fed, membrane wall + quench		Slurry-fed, Quenched	
CO Shift	Sour (S2)	Sweet (S3)	Sour (A3c)	Sweet(A4c)
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 4.39 shows the work and efficiency data for all four plant designs. The sour-shift system is always more efficient by 3-4 percentage points, and always results in greater work output for the reasons mentioned during the explanations of Case A3 and B3. 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, but only for the same type of CCS. The results contradict

those of Li and Wang (2009). This is probably due to different design considerations, such as allowing the steam mass flow to increase.

Table 4.40 Selected parasitic energy & heat losses for dry-fed vs. slurry-fed systems

Case	Dry-fed, membrane wall + quench		Slurry-fed, Quenched	
CO Shift	Sour (S2)	Sweet (S3)	Sour (A3c)	Sweet(A4c)
Acid Gas Removal (Btu/s)	3,553	3,578	3,783	3,805
Syngas Water Condensed (Btu/s)	12,691	20,764	14,543	22,502
CO_2 and AGR heat loss (Btu/s)	**33,307**	**50,119**	**34,777**	**51,599**
Slag Production (Btu/s)	7,735	7,790	14,401	14,485
Cooler Heat Rejection (Btu/s)	104,256	133,214	137,036	163,716

From the heat loss data in Table 4.40, the same relationship is observed between sweet- and sour-shift, regardless of whether the gasifier is dry-fed or slurry-fed. The sour-shift system is easier to operate, as it is fully integrated into the AGR plant, and the gasifier in these cases is smaller than that of the sweet-shift cases. Less water need be condensed because most of the water was already consumed by the CO-shift reaction. Finally, the coolers reject less heat because of the high recovery efficiency of cooler 1 (the low-temp gases leaving the AGR block absorb more heat from cooler 1, which has a higher hot gas inlet temperature. The end result is that more heat is removed at cooler 1, and does not have to be rejected at cooler 2 or 3. See the explanation in Cases A3 and B3, and compare Figs. 3.13 and 3.15 in Chapter 3).

Table 4.41 Economics for dry-fed vs. slurry-fed systems

Case	Dry-fed, membrane wall and quench		Slurry-fed, Quenched	
CO Shift	Sour (S2)	Sweet (S3)	Sour (A3c)	Sweet(A4c)
Total capital cost (millions of $)	1,010.3	1,030.6	1,027.4	1,044.2
Capital Cost ($/kW)	**4,574**	**5,162**	**4,721**	**5,303**
CoE ($/kW-hr)	**0.1219**	**0.1354**	**0.1269**	**0.1405**

From the economic data in Table 4.41, sour-shift wins again by being even cheaper in capital cost (as expected) and cost of electricity, almost by 1.5 cents/kW-hr in both sets of cases. The reasons for which are the same as explained in each main case's respective section in this chapter. There appears to be no qualitative difference between dry-fed and slurry-fed systems

<u>other than the cheaper cost of preparing the fuel in the case of the dry-fed system.</u> From this and the previous data, there appear to be no synergistic effects whatsoever between the feeding system type and the type of CCS used. Again, this is contradictory to the results found in Li's and Wang's report.

Table 4.42 Emissions for dry-fed vs. slurry-fed systems

Case	Dry fed, membrane wall + quench		Slurry fed, Quenched	
CO Shift	Sour (S2)	Sweet (S3)	Sour (A3c)	Sweet(A4c)
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**

Finally, for the emissions data in Table 4.42, there appears to be no appreciable difference in the amount of CO_2, NO_x, or SO_x produced. Sour-shift appears to produce less CO_2 on the whole in the dry-fed system, while the reverse is true for sweet-shift. However, when effective CO_2 is calculated, this is inconsequential, as all of the cases are undeniably carbon-negative. On a per MW-year basis, sweet-shift does appear to remove more CO_2 than sour-shift does on the whole.

4.3.3 Case S4: Air-blown Gasifier (vs. Case A1a)

For Case S4, the air-blown design, the MHI gasifier was chosen as the representative gasifier model, as it is the only entrained-flow gasifier to ever be commercially successful at using an air-blown design. Since this is a two-stage gasifier, the fuel must be split into 2 streams. For this, an exact 50:50 fuel ratio was chosen arbitrarily to get the result. A 35% slurry was used so as to keep an appreciable comparison with the main cases. The special case was only tested against the baseline case with no biomass and no CCS (Case A1a), because the effects of CCS were not deemed important for comparison. 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 4.43-4.46.

Table 4.43 Work and efficiency for air-blown vs. oxygen-blown designs

Case	Air-blown (S4)	Oxygen-blown (A1a)
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**

The work and efficiency data is shown in Table 4.43. 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 less auxiliary losses. All additional power generated comes from the steam cycle, like in previous cases (Cases A3, B1, and S1 compared to A1). To investigate the cause of this, the syngas composition is analyzed. This is shown in Table 4.44.

From Table 4.44, it is clear that the air-blown gasifier's syngas flow rate is higher than the oxygen-blown one. The GT output power must be maintained, as per specifications, but, to do this, the TIT and mass flow cannot remain the same because the syngas is diluted with N_2 and the combusted gases is further diluted by compressed air. Due to the change in syngas composition, various other GT modifications have to be made to keep the GT cycle efficiency and total pressure drop roughly equal. The best operating point was found by using a slightly higher mass flow rate *and also a lower TIT*, while the TET of course dropped as a result to the correct level needed to get the same output power. As a result of all of these changes, the gasifier must have more syngas output than in the oxygen-blown case. This usually means that it will be larger in size. In the end, this MHI gasifier is about $28 million more expensive than the GE gasifier used in Case A1a.

Table 4.44 Syngas compositions for air-blown vs. oxygen-blown designs

Biomass/Coal Ratio (vol%)	Air-blown (S4)	Oxygen-blown (A1a)
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
Raw Syngas Mass flow (lbs/s)	352.1	281.6

The oxygen-blown system clearly has the higher LHV, and, as such, should produce more power. However, Case S4's syngas also contains less water overall than that of Case A1a, and more than 200 times as much CH_4 as the oxygen-blown design. How this affects the cleanup system would be best gathered from the heat loss data shown in Table 4.45.

Table 4.45 Selected parasitic energy & heat losses for air-blown vs. oxygen-blown designs

Case	Air-blown (S4)	Oxygen-blown (A1a)
Acid Gas Removal (Btu/s)	5,103	5,197
Syngas Water Condensed (Btu/s)	19,454	22,445
AGR heat loss (Btu/s)	600.2	611.2
Slag Production (Btu/s)	4,921	14,473
Cooler Heat Rejection (Btu/s)	106,851	167,317

In this case, less energy and less cooling are needed for the AGR process. Less water is condensed because less water is present in the syngas to begin with, and there is less water needing to be added for the quench because the exit temperature of the gasifier is lower. Finally, the slag produced is so much lower for the air-blown case due to the 2-stage design: the temperature in the second stage isn't high enough to melt the ash in the second fuel injection stream, so it never turns into slag and is instead removed as fly ash in the particulate scrubber. Lastly, a portion of the slag is used to coat the walls of this gasifier, just like the Shell gasifier, so another fair amount of slag never even leaves the gasifier. While this is a transient process (i.e. the slag only coats the walls initially, and then the process stops and isn't allowed to start up again until the previous layer cracks or breaks off,) it can be approximated as a continuous process by taking the total amount of slag used for this function throughout the year of operation and distributing it evenly throughout this time period.

Table 4.46 Economics for air-blown vs. oxygen-blown designs

Case	Air-blown (S4)	Oxygen-blown (A1a)
Total capital cost (millions of $)	1,009.5	1,029.75
Capital cost ($/kW)	**4,063**	**4,363**
CoE ($/kW-hr)	**0.0949**	**0.1008**

From an economic standpoint, as seen in Table 4.46, 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 considerations, however, the air-blown system just so happens to be the better design.

Emissions are shown in Table 4.47. The air-blown system introduces additional nitrogen to the system, which results in more NO_x. The SO_x produced is simply a result of the leftover sulfur from the cleanup system, and, since the syngas mass flow rate is lower in this case, this SO_x decreases compared to Case A1a. The CO2 emissions for the air-blown case (S4) are about 5.6% lower than the oxygen-blown case (A1a).

Table 4.47 Emissions for air-blown vs. oxygen-blown designs

Case	Air-blown (S4)	Oxygen-blown (A1a)
NO_x (tons/year)	321.1	234.7
SO_x (tons/year)	2,105	2,157.5
Gross CO_2 (tons/year)	2,104,457	2,110,246
Eff. CO_2 (tons/year)	**2,104,457**	**2,110,246**
Gross CO_2 (tons/MW-year)	8,470.6	8,942.0
Eff. CO_2 (tons/MW-year)	**8,470.6**	**8,942.0**

4.3.4 Case S5: Supercritical Plant – Use GT Fuel in DB

Another issue considered is the use of the cleaned syngas in the supercritical cycle's duct burner. The main cases used natural gas (approximated as 100% CH_4) for this device, but this introduces an additional cost into the system, and requires set up the purchase contract of another input fuel. In order to avoid signing up 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 4.48-4.50.

Table 4.48 Work and efficiency for using GT fuel in duct burner

Case	GT fuel (S5)	Natural Gas (B1a)
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 4.49 Selected parasitic energy & heat losses for using GT fuel in duct burner

Case	Syngas (S5)	Natural Gas (B1a)
Acid Gas Removal (Btu/s)	5,806	5,197
Syngas Water Condensed (Btu/s)	25,262	23,284
AGR heat loss (Btu/s)	682.9	611.2
Slag Production (Btu/s)	16,171	14,473
Cooler Heat Rejection (Btu/s)	186,220	164,001

From the work data in Table 4.47, it can be seen that using the GT fuel actually increases the total steam power, but the additional auxiliary loss 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. Table 4.48 helps in explaining how this happens. For one, the gasifier size increased, as evidenced by the extra heat losses from AGR and slag produced. 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 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, including the consumption of more ASU power and greater electrical power required to run the acid gas removal system.

Table 4.50 Economics for using GT fuel in duct burner

Case	Syngas (S5)	Natural Gas (B1a)
Total capital cost (millions of $)	1,170.3	1,087.58
Capital cost ($/kW)	**4,459**	**4,072**
CoE ($/kW-hr)	**0.1024**	**0.0972**

Economically speaking, it would follow that the system using the syngas as the fuel for the HRSG's duct burner (Case S5) is more expensive to build and more costly to operate than the natural gas cycles in the main case. This is confirmed in Table 4.49. The larger gasifier and more exhaustive cleanup system coupled with the total loss of net power practically guarantees that this would happen.

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

4.3.5 Cases S6 & S7: Different Gasifier Fuel – Use Illinois #6 (vs. A1a and A1b)

The final special cases involve the use of a higher ranked coal, Illinois #6, in Cases A1a and A1b. As mentioned in Chapter 3, 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 of the plant, especially in regards to economics. The purpose of these two cases is to observe (1) how changing coals affects the overall base plant statistics (Case S6) and (2) how Illinois #6 would behave when blended with biomass (Case S7). All data for both of these cases can be seen in Tables 4.51-4.54.

From the work and efficiency data in Table 4.51, Illinois #6 has a clear benefit regardless of whether it is used alone or blended with biomass in comparison to lignite. The use of Illinois #6 always reduces auxiliary losses (due mostly to being of higher rank, and thus lowering the necessary gasifier flow rate,) raising total net power as a result, and raising the net efficiency by at least 2 points. 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 in the main cases (A1).

Table 4.51 Work and efficiency for Illinois #6 bituminous and Texas lignite coal plants

Case	Illinois #6 (S6)	Texas Lignite (A1a)	Ill. #6 + 10% biomass (S7)	Tex. Lig. + 10% biomass (A1b)
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 Efficiency (LHV)	44.33	43.01	43.90	43.59
Net Efficiency (LHV)	**37.87**	**35.06**	**37.20**	**35.70**

To determine how this happens, first of all, see the heat data in Table 4.52. Before there is cause for alarm at some of the numbers, it should be known that the gasifier must have decreased in size for Cases S6 and S7. This is 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 4.53 for the syngas compositions.)

The issue with many of these losses is that (1) Illinois #6 contains more than 5 times as much sulfur as South Hallsville lignite. Naturally, this would require extensive cleaning later on. But, (2) due to their similar ash contents, not much additional slag is produced (Slag actually decreases in Case S6 compared to Case A1a.) Finally, (3) the coolers reject less heat mostly from the downsizing of the gasifier and reduction in flow rate, but another contributing factor is the lower moisture content of the syngas. About 70lbs/s less moisture is removed from the syngas in Case S6 as it is from that of in Case A1a. This means there is less activity in the COS-reactor (Note that the COS to H_2S reaction is exothermic,) resulting in a lower exit desulfurized-syngas temperature, which in turn results in less heat needing to be rejected to cool the syngas down to the temperature it needs for removing acid gas.

Table 4.52 Parasitic energy & heat losses for Illinois #6 bituminous and Texas lignite coal plants.

Case	Illinois #6 (S6)	Texas Lignite (A1a)	Ill. #6 + 10% bio (S7)	Tex. Lig. + 10% bio (A1b)
Acid Gas Removal (Btu/s)	17,791	5,197	16,915	4,532
Syngas Water Condensed (Btu/s)	15,979	22,445	16,883	21,584
AGR heat loss (Btu/s)	2,092.2	611.2	1,989.2	533.0
Slag Production (Btu/s)	14,244	14,473	14,436	13,843
Cooler Heat Rejection (Btu/s)	91,403	167,317	101,400	156,460

Table 4.53 Syngas compositions for Illinois #6 bituminous and Texas lignite coal plants.

Case	Illinois #6 (S6)	Texas Lignite (A1a)	Ill. #6 + 10% bio (S7)	Tex. Lig. + 10% bio (A1b)
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 4.54 Economics for Illinois #6 bituminous and Texas lignite coal plants

Case	Illinois #6 (S6)	Texas Lignite (A1a)	Ill. #6 + 10% bio (S7)	Tex. Lig. + 10% bio (A1b)
Total capital cost (millions of $)	916.62	1,029.75	870.56	926.74
Capital cost ($/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 4.54, it can be seen that the Illinois #6 plants reduce the capital cost of plants using it by about $600/kW for pure coal case (Case S6 vs. Case A1a) and $400/kW (~8%) for 10% BMR (Case S6 vs. Case A1b). For the pure coal cases, the CoE is reduced by about 0.12 cents when Illinois #6 is used; however, 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 Case S6, adding biomass increases the cost of preparation, and requires a bigger gasifier than the case of using the lignite to make up for a loss of heating value in the syngas. Both of these contribute to extra auxiliary losses, which increase CoE directly. From the perspective of Case A1b, changing out lignite to Illinois #6 does not reduce the auxiliary losses by as much as Case S6 does from Case A1a. In this case, while the total investment decreases by about $400/kW (~8%) for 10% BMR (Case S6 vs. Case A1b) 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 4.55 Emissions for Illinois #6 bituminous and Texas lignite coal plants

Case	Illinois #6 (S6)	Texas Lignite (A1a)	Ill. #6 + 10% bio (S7)	Tex. Lig. + 10% bio (A1b)
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 4.55 shows the emissions data for this set of plants. As can be seen, less coal needs to be burned in Cases S6 and S7, so the amount of NO_x and carbon dioxide is lowered. However, due to the higher sulfur content in Illinois #6, the SO_x emissions cannot do anything but increase. Aside from this, Cases S6 and S7 are superior emissions-wise to the main cases, producing less CO_2 overall (about 300,000-400,000 tons/year, or 27%) and less per MW as well (about 1000 tons/MW-year, or 20%).

4.3.6 ISO Conditions

Finally, a short case was run using GTMaster to compare the effects of ISO conditions (59°F, 60% R.H.) on plant performance while keeping the physical hardware constant. The plant used was that of Case A1a (i.e. the subcritical baseline case: no CCS, 0% BMR.) The controls in place are the same as those of all previous cases, as described in detail in Chapter 3, with the one exception of GT output power. The GT output power is no longer fixed because the main source of efficiency gain lies in the additional availability to do work from the milder environmental conditions. In this regard, the GT power was allowed to fluctuate to account for this. In total, the gross efficiency only increases by about 0.15 percentage points, but the net efficiency increases by about 0.45 percentage points, about 3 times that amount. This is mostly due to increased gas turbine power, in the realm of about 15MW, plus an additional 3MW from the steam turbine, and about 1-2MW fewer auxiliary losses. Overall, the total net power improves by a grand total of 17MW (~6%).

CHAPTER FIVE
CONCLUSIONS

5.1 Overview and Primary Conclusions

In this study, several IGCC plants were simulated using Thermoflow's GTPro® software, and used to investigate the effects of biomass, carbon capture, and replacing the subcritical bottom cycle with a supercritical cycle. The objectives were to improve plant performance by increasing the efficiency and lowering the emissions while keeping costs to a minimum.

All plants were designed to be constructed in Southern Louisiana, with climate conditions of 85°F and 90% R.H. with the main cases using Texas Lignite and sugarcane bagasse as the fuel sources. The baseline power output is around 235MW total. For the major design criteria of all of the cases, the ST inlet temperature and pressure were fixed. The GT's inlet temperature, mass flow rate, and total output power were also fixed. However, for cases in which it is not possible to satisfy all three conditions at once, GT output power takes the first priority, followed by TIT, and finally mass flow rate.

The subcritical cycle was examined with different forms of CCS attached with one case without CCS to serve as the baseline. For each case, a set of sub-cases for the same plant were run using different amounts of biomass within the feedstock. After this, the supercritical cycle was used as a separate category to run the same cases again as a "group" with the same criteria breakdown. In addition, a set of miscellaneous "special" cases were observed to try to examine and/or validate some of the design decisions in the main cases. The special cases observed were: (1) using radiant and convective syngas coolers instead of a quench in cooling the syngas, (2) using a dry-fed gasifier instead of a slurry-fed gasifier and examining the effect of this on carbon capture with biomass, (3) using an air-blown system instead of an oxygen-blown one to see which system provided the higher efficiency/lower cost, (4) seeing if using the GT fuel for the duct burner in the supercritical system rather than using natural gas would raise or lower the efficiency/costs, and (5) upgrading to a higher-ranked coal and determining the effects of which on overall plant performance and emissions. Minor criteria that vary between cases is explained in detail in Chapters 3 and 4.

The results of these endeavors are summarized below. Bear in mind that the results presented in this study are subject to changes in design considerations and assumptions if different from those previously discussed.

1.) Supercritical vs. Subcritical: The supercritical system is universally superior to the subcritical system regardless of which case is taken into consideration. It always provides more power at a reasonable extra cost ($30 million extra, but 32MW extra power: about $300/kW less expensive) and always has the lower CoE (0.3 cents/kW-hr). The total emissions always increase (100,000 tons/year difference), but, due to increased power, this results in 400-500 tons/MW-year fewer emissions overall.

2.) Post-combustion CCS vs. Pre-combustion CCS: Post-combustion carbon capture is thermally and economically worse than pre-combustion carbon capture for IGCC. Even though it is easier to implement than pre-combustion CCS, the price tag for the system as a whole coupled with the inefficient use of solvent drive the total plant cost to unaffordably high levels (5 cents/kW-hr and $2500/kW more than sour-shift, 4 cents/kW-hr and $2000/kW more than sweet-shift). In addition, the average plant performance is impacted significantly for post-combustion (8 points below the baseline's net electrical efficiency, compared to sour-shift's mere 4 or 5 points).

3.) Sour-shift vs. Sweet-shift: For pre-combustion CCS, sour-shift is superior to sweet-shift regardless of the feedstock feeding system. It may be implemented yielding the same degree of capture at a fraction of the costs ($500/kW and $0.02/kW-hr cheaper), and, in this study, even results in better efficiency (only 4-5 points below the baseline, compared to sweet-shift's 5.0-8.5 points). In addition, sour-shift can actually *increase* total output power for subcritical (but not for supercritical) designs (10MW in this study). Even so, regardless of the system used, sour-shift will always be cheaper.

4.) Effect of Biomass: Adding biomass to the system always reduces the emissions and can even make a plant carbon-negative with as little as 10% biomass by weight. In addition, the efficiency will improve (0.7 points) and power output will also improve (~1%-3% more) for up to 10%

biomass ratio (BMR) for the right kind of biomass that has been properly pretreated. Beyond 10% BMR, however, the efficiency begins to drop due to the rising pretreatment costs, but the system itself still remains more efficient than from using coal alone (between 0.2-0.3 points on average). The economic difference is fairly marginal, but the trend is inversely proportional to the efficiency.

5.) Radiant and Convective Coolers vs. Quench: Radiant and convective syngas coolers are always more efficient than a quench (6 points). However, they are also more costly ($1000/kW).

6.) Dry-fed vs. Slurry-fed: Dry-fed systems are universally more efficient than slurry-fed systems (3 points), as long as the same kind of CCS is used. The trend between sour-shift and sweet-shift CCS in dry-fed system is the same as that of slurry-fed systems (see item 3): sour-shift remains cheaper and more efficient than sweet-shift even for dry-fed systems, but the dry-fed version of both cases is more efficient than its corresponding slurry-fed counterparts.

7.) Air-blown vs. Oxygen-blown: The air-blown gasifier in this case is superior to the oxygen-blown system used in the main cases. It results in lower costs ($300/kW and 0.6 cents/kW-hr), higher efficiency (2.6 points), and fewer carbon emissions (500 tons/MW-year). However, this is only valid for this plant setup. Other plants with different design criteria may not be affected in quite the same way.

8.) Syngas vs. Natural Gas in Duct Burner: Using natural gas in the duct burner for the supercritical plant is better than using the syngas because of the fact that additional mass flow must be provided to keep the gas turbine at the same level of performance. This additional mass flow will come from an enlarged, more expensive gasifier, which leads to greater auxiliary losses and more expenses. However, using syngas allows the plant to remain self-sufficient, while using NG exposes the plant to the volatility of the price of natural gas.

9.) <u>Bituminous Coal vs. Lignite:</u> Illinois #6 is always more efficient to use (2-3 points), as it has a higher heating value (3000Btu/lb higher LHV) and produces better syngas (800Btu/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 0.5 points and raising the CoE by 0.5 cents/kW (Capital costs still decrease by about $200/kW.)

5.2 Recommended Future Studies

Based on the results from this study, the following studies are recommended to be performed in order to form a more complete picture of the effects of the parameters examined:

a. Perform another simulation using an ultra-supercritical steam cycle instead. TIT and TIP should be at least 4500 PSI and 1500°F.

b. Perform at least one further case using the syngas in the duct burner: allow the total GT power to change based on the new mass flow rate and keep the gasification block the same.

c. Use the air-blown system in another plant or set of plants using different design considerations to further refine the difference between oxygen-blown and air-blown designs.

d. Find the "best operating point" of either this plant or another plant with other coals in the lignite-to-bituminous range using either bagasse or another type of biomass. The best operating or "optimum" point will be the BMR where the efficiency is the highest and/or the costs are the lowest.

e. Perform further analysis on sweet-shift CCS systems by considering other forms of CCS, such as adsorption-regeneration or membranes as well as other methods of CO-shift.

f. Develop a "Carbon-Ready" plant, and evaluate how a post-combustion CCS system will affect it if all of the base hardware is kept the same. In other words, a "carbon-ready" plant is simply one that does not use any CCS, but is "ready" to be retro-fitted with a post-combustion system at any time during its lifetime operation.

g. Consider air-integration as another parameter in a similar IGCC study. That is, connect the GT compressor to the ASU and use a portion of that air to aid in gasification in order to determine how this will affect the plant's performance.

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APPENDIX

Case A1a – Subcritical, No CCS, 0% biomass

GT PRO 21.0 parallel
1263 09-20-2011 18:06:57 file=C:\Documents and Settings\Hank\Desktop\Hank's Papers and Data\Grad Wo
rk\Thermoflow\Better Plant Design\SUBCRITICAL QUENCH - 0% BIOMASS.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)	200019		8299		41.12	
Steam Turbine(s)	89493					
Plant Total	289512	236012	7933	9732	43.01	35.06

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
29.58	24.10	31.37	8336

GT fuel HHV/LHV ratio =	1.083			
DB fuel HHV/LHV ratio =	1.083			
Total plant fuel HHV heat input / LHV heat input =	1.107			
Fuel HHV chemical energy input (77F/25C) =	2541735	kBTU/hr	706037	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2296760	kBTU/hr	637989	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2548661	kBTU/hr	707961	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2567622	kBTU/hr	713228	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200019	41.12	8299	1028	1100
Total	200019			1028	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1798008	kBTU/hr	499447	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1659960	kBTU/hr	461100	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
77.34	89493	38.12	29.48	-251901

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 =	1035892	kBTU/hr	287748	BTU/s
Water/steam to gasification plant =	28819	kBTU/hr	8005	BTU/s
Water/steam from gasification plant =	70434	kBTU/hr	19565	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-45.52	%		

Case A1b – Subcritical, No CCS, 10% biomass

GT PRO 21.0 parallel
1263 07-27-2011 16:29:58 file=C:\Documents and Settings\Hank\Desktop\Hank's Papers and Data\Grad Wo
rk\Thermoflow\Better Plant Design\SUBCRITICAL QUENCH - 10% BIOMASS.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)	200018		8307		41.08	
Steam Turbine(s)	89790					
Plant Total	289807	237356	7827	9557	43.59	35.70

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
30.15	24.60	31.90	8333

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) =	2501652	kBTU/hr	694903	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2268428	kBTU/hr	630119	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2520301	kBTU/hr	700084	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2539259	kBTU/hr	705350	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200018	41.08	8307	1026	1102
Total	200018			1026	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1799918	kBTU/hr	499977	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1661637	kBTU/hr	461566	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
77.38	89790	38.22	29.57	-251873

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 =	69632	kBTU/hr	19342	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-45.13	%		

Case A1a – Subcritical, No CCS, 0% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	920.1	kW
Condensate pump(s)*	181.7	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	2.705	kW
Cooling water pump(s)	1005.8	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	45	kW
Lights	80	kW
Aux. from PEACE running motor/load list	744.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	190.7	kW
Miscellaneous plant auxiliaries	289.5	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	37841	kW
Gasification plant, fuel preparation	7180	kW
Gasification plant, AGR*	675.1	kW
Gasification plant, other/misc	2534.7	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	52052	kW
Actual (user input) overall plant auxiliaries	52052	kW
Transformer losses	1447.6	kW
Total auxiliaries & transformer losses	**53500**	**kW**

Case A1b – Subcritical, No CCS, 10% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	921.6	kW
Condensate pump(s)*	181.5	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	2.702	kW
Cooling water pump(s)	1010	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	45	kW
Lights	80	kW
Aux. from PEACE running motor/load list	754.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	191.3	kW
Miscellaneous plant auxiliaries	289.8	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	34484	kW
Gasification plant, fuel preparation	9683	kW
Gasification plant, AGR*	588.9	kW
Gasification plant, other/misc	2408.8	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	51002	kW
Actual (user input) overall plant auxiliaries	51002	kW
Transformer losses	1449	kW
Total auxiliaries & transformer losses	**52451**	**kW**

Case A1a – Subcritical, No CCS, 0% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**637254**	**BTU/s**		
Ambient air sensible	11138	BTU/s		
Ambient air latent	21202	BTU/s		
GT syngas	515781	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69973	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	19565	BTU/s		
Energy Out	**637486**	**BTU/s**		
Net power output	223707	BTU/s		
Stack gas sensible	77545	BTU/s		
Stack gas latent	114251	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1425.2	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	154522	BTU/s		
Process steam	0.0002	BTU/s		
Process water	0	BTU/s		
Blowdown	920.2	BTU/s		
Heat radiated from steam cycle	2210.7	BTU/s		
ST/generator mech/elec/gear loss	1460.6	BTU/s		
Non-heat balance related auxiliaries	11786	BTU/s		
Transformer loss	1372.1	BTU/s		
ASU compressors	35868	BTU/s		
Water/steam to gasification plant	8005	BTU/s		
AGR auxiliary	639.9	BTU/s		
Energy In - Energy Out	**-232**	**BTU/s**	**-0.0364**	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -231.9 BTU/s				

Case A1b – Subcritical, No CCS, 10% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**637217**	**BTU/s**		
Ambient air sensible	11178	BTU/s		
Ambient air latent	21278	BTU/s		
GT syngas	515868	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69965	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	19342	BTU/s		
Energy Out	**637452**	**BTU/s**		
Net power output	224981	BTU/s		
Stack gas sensible	77446	BTU/s		
Stack gas latent	114395	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1425.4	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	155185	BTU/s		
Process steam	0	BTU/s		
Process water	0	BTU/s		
Blowdown	921	BTU/s		
Heat radiated from steam cycle	2210.9	BTU/s		
ST/generator mech/elec/gear loss	1463.4	BTU/s		
Non-heat balance related auxiliaries	14054	BTU/s		
Transformer loss	1373.5	BTU/s		
ASU compressors	32686	BTU/s		
Water/steam to gasification plant	6981	BTU/s		
AGR auxiliary	558.2	BTU/s		
Energy In - Energy Out	**-234.7**	**BTU/s**	**-0.0368**	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -234.7 BTU/s				

Case A1a – Subcritical, No CCS, 0% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	852528	BTU/s		
Power Block Energy In:				
Ambient air sensible	11138	BTU/s		
Ambient air latent	21202	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69973	BTU/s		
Process return & makeup	0	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	709190	BTU/s		
Gasifier slurry water	862	BTU/s		
Quench water	29860	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1352.6	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	9355	BTU/s		
Total Energy Out	852801	BTU/s		
Power Block Energy Out:				
Net power output	223707	BTU/s		
Stack gas sensible	77545	BTU/s		
Stack gas latent	114251	BTU/s		
GT cycle losses	5198	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	154522	BTU/s		
Process	0.0002	BTU/s		
Steam cycle losses	4591	BTU/s		
Non-heat balance auxiliaries	11786	BTU/s		
Transformer losses	1372.1	BTU/s		
Gasifier Energy Out:				
Heat losses	63.8	BTU/s		
Slag	14473	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	5197	BTU/s		
CO2 removal	0	BTU/s		
Cooling after CO shift	0	BTU/s		
Water condensed from syngas	22451	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
AGR Qrej	6539	BTU/s		
AGR heat loss	611.2	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	167294	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	8558	BTU/s		
Heat rejection from compressor inter/after cooling	33819	BTU/s		
Compressors mechnical & electrical losses	1793.4	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-273.1	BTU/s	-0.032	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case A1b – Subcritical, No CCS, 10% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	839057	BTU/s		
Power Block Energy In:				
Ambient air sensible	11178	BTU/s		
Ambient air latent	21278	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69965	BTU/s		
Process return & makeup	0	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	697914	BTU/s		
Gasifier slurry water	830.4	BTU/s		
Quench water	28494	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1285.9	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	8525	BTU/s		
Total Energy Out	839332	BTU/s		
Power Block Energy Out:				
Net power output	224981	BTU/s		
Stack gas sensible	77446	BTU/s		
Stack gas latent	114395	BTU/s		
GT cycle losses	5198	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	155185	BTU/s		
Process	0	BTU/s		
Steam cycle losses	4595	BTU/s		
Non-heat balance auxiliaries	14054	BTU/s		
Transformer losses	1373.5	BTU/s		
Gasifier Energy Out:				
Heat losses	63.01	BTU/s		
Slag	13843	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	4532	BTU/s		
CO2 removal	0	BTU/s		
Cooling after CO shift	0	BTU/s		
Water condensed from syngas	21584	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
AGR Qrej	5723	BTU/s		
AGR heat loss	533	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	156461	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	7799	BTU/s		
Heat rejection from compressor inter/after cooling	30818	BTU/s		
Compressors mechnical & electrical losses	1634.3	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-274.8	BTU/s	-0.0327	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case A1c – Subcritical, No CCS, 30% biomass

GT PRO 21.0 parallel							
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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		8312		41.05	
Steam Turbine(s)	90191					
Plant Total	290209	234296	7763	9616	43.96	35.49

PLANT EFFICIENCIES			
PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
29.90	24.31	31.68	8315

GT fuel HHV/LHV ratio =	1.082			
DB fuel HHV/LHV ratio =	1.082			
Total plant fuel HHV heat input / LHV heat input =	1.1			
Fuel HHV chemical energy input (77F/25C) =	2477354	kBTU/hr	688154	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2252876	kBTU/hr	625799	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2504770	kBTU/hr	695769	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2523730	kBTU/hr	701036	BTU/s

GAS TURBINE PERFORMANCE - SGT6-4000F (Physical Model #141)					
	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200017	41.05	8312	1026	1102
Total	200017			1026	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1799210	kBTU/hr	499780	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1662469	kBTU/hr	461797	BTU/s

STEAM CYCLE PERFORMANCE				
HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
77.41	90191	38.36	29.69	-251894

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 =	1036494	kBTU/hr	287915	BTU/s
Water/steam to gasification plant =	19595	kBTU/hr	5443	BTU/s
Water/steam from gasification plant =	68527	kBTU/hr	19035	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-46	%		

Case A1d – Subcritical, No CCS, 50% biomass

GT PRO 21.0 parallel							
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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		8316		41.03	
Steam Turbine(s)	90551					
Plant Total	290567	231290	7701	9675	44.31	35.27

PLANT EFFICIENCIES			
PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
29.64	24.01	31.46	8298

GT fuel HHV/LHV ratio =	1.081			
DB fuel HHV/LHV ratio =	1.081			
Total plant fuel HHV heat input / LHV heat input =	1.097			
Fuel HHV chemical energy input (77F/25C) =	2453744	kBTU/hr	681595	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2237765	kBTU/hr	621601	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2489670	kBTU/hr	691575	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2508630	kBTU/hr	696842	BTU/s

GAS TURBINE PERFORMANCE - SGT6-4000F (Physical Model #141)					
	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200017	41.03	8316	1026	1103
Total	200017			1026	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1798486	kBTU/hr	499579	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1663295	kBTU/hr	462026	BTU/s

STEAM CYCLE PERFORMANCE				
HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
77.41	90551	38.50	29.80	-251904

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 =	1036831	kBTU/hr	288009	BTU/s
Water/steam to gasification plant =	14213	kBTU/hr	3948	BTU/s
Water/steam from gasification plant =	67494	kBTU/hr	18748	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-46.88	%		

Case A1c – Subcritical, No CCS, 30% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	922.5	kW
Condensate pump(s)*	181.7	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	2.703	kW
Cooling water pump(s)	1016.8	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	45	kW
Lights	80	kW
Aux. from PEACE running motor/load list	754.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	192.2	kW
Miscellaneous plant auxiliaries	290.2	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	32798	kW
Gasification plant, fuel preparation	15000	kW
Gasification plant, AGR*	459.3	kW
Gasification plant, other/misc	2357.4	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	54462	kW
Actual (user input) overall plant auxiliaries	54462	kW
Transformer losses	1451	kW
Total auxiliaries & transformer losses	**55913**	**kW**
* Heat balance related auxiliaries		

Case A1d – Subcritical, No CCS, 50% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	923.1	kW
Condensate pump(s)*	181.7	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	2.705	kW
Cooling water pump(s)	1022.9	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	45	kW
Lights	80	kW
Aux. from PEACE running motor/load list	754.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	192.9	kW
Miscellaneous plant auxiliaries	290.6	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	31161	kW
Gasification plant, fuel preparation	20168	kW
Gasification plant, AGR*	333.4	kW
Gasification plant, other/misc	2307.4	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	57824	kW
Actual (user input) overall plant auxiliaries	57824	kW
Transformer losses	1452.8	kW
Total auxiliaries & transformer losses	**59277**	**kW**
* Heat balance related auxiliaries		

Case A1c – Subcritical, No CCS, 30% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**636601**	**BTU/s**		
Ambient air sensible	11192	BTU/s		
Ambient air latent	21305	BTU/s		
GT syngas	515509	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69971	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	19035	BTU/s		
Energy Out	**636896**	**BTU/s**		
Net power output	222080	BTU/s		
Stack gas sensible	77403	BTU/s		
Stack gas latent	113983	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1425.7	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	156236	BTU/s		
Process steam	0	BTU/s		
Process water	0	BTU/s		
Blowdown	921.7	BTU/s		
Heat radiated from steam cycle	2212	BTU/s		
ST/generator mech/elec/gear loss	1467.3	BTU/s		
Non-heat balance related auxiliaries	19052	BTU/s		
Transformer loss	1375.4	BTU/s		
ASU compressors	31088	BTU/s		
Water/steam to gasification plant	5443	BTU/s		
AGR auxiliary	435.4	BTU/s		
Energy In - Energy Out	**-295.4**	**BTU/s**	**-0.0464**	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -295.4 BTU/s				

Case A1d – Subcritical, No CCS, 50% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**635997**	**BTU/s**		
Ambient air sensible	11206	BTU/s		
Ambient air latent	21332	BTU/s		
GT syngas	515146	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69973	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	18748	BTU/s		
Energy Out	**636268**	**BTU/s**		
Net power output	219231	BTU/s		
Stack gas sensible	77401	BTU/s		
Stack gas latent	113569	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1425.9	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	157176	BTU/s		
Process steam	0	BTU/s		
Process water	0	BTU/s		
Blowdown	922.1	BTU/s		
Heat radiated from steam cycle	2212.4	BTU/s		
ST/generator mech/elec/gear loss	1470.8	BTU/s		
Non-heat balance related auxiliaries	23910	BTU/s		
Transformer loss	1377.1	BTU/s		
ASU compressors	29536	BTU/s		
Water/steam to gasification plant	3948	BTU/s		
AGR auxiliary	316	BTU/s		
Energy In - Energy Out	**-271.2**	**BTU/s**	**-0.0426**	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -271.2 BTU/s				

Case A1c – Subcritical, No CCS, 30% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	831167	BTU/s		
Power Block Energy In:				
Ambient air sensible	11192	BTU/s		
Ambient air latent	21305	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69971	BTU/s		
Process return & makeup	0	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	691071	BTU/s		
Gasifier slurry water	818.6	BTU/s		
Quench water	27858	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1255.1	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	8109	BTU/s		
Total Energy Out	831505	BTU/s		
Power Block Energy Out:				
Net power output	222080	BTU/s		
Stack gas sensible	77403	BTU/s		
Stack gas latent	113983	BTU/s		
GT cycle losses	5199	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	156236	BTU/s		
Process	0	BTU/s		
Steam cycle losses	4601	BTU/s		
Non-heat balance auxiliaries	19052	BTU/s		
Transformer losses	1375.4	BTU/s		
Gasifier Energy Out:				
Heat losses	62.58	BTU/s		
Slag	13451	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	3534	BTU/s		
CO2 removal	0	BTU/s		
Cooling after CO shift	0	BTU/s		
Water condensed from syngas	21169	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
AGR Qrej	4501	BTU/s		
AGR heat loss	415.6	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	151000	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	7417	BTU/s		
Heat rejection from compressor inter/after cooling	29312	BTU/s		
Compressors mechnical & electrical losses	1554.4	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-337.6	BTU/s	-0.0406	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case A1d – Subcritical, No CCS, 50% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	823497	BTU/s		
Power Block Energy In:				
Ambient air sensible	11206	BTU/s		
Ambient air latent	21332	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69973	BTU/s		
Process return & makeup	0	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	684421	BTU/s		
Gasifier slurry water	807.1	BTU/s		
Quench water	27239	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1223.4	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	7704	BTU/s		
Total Energy Out	823805	BTU/s		
Power Block Energy Out:				
Net power output	219231	BTU/s		
Stack gas sensible	77401	BTU/s		
Stack gas latent	113569	BTU/s		
GT cycle losses	5199	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	157176	BTU/s		
Process	0	BTU/s		
Steam cycle losses	4605	BTU/s		
Non-heat balance auxiliaries	23910	BTU/s		
Transformer losses	1377.1	BTU/s		
Gasifier Energy Out:				
Heat losses	62.16	BTU/s		
Slag	13071	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	2562.9	BTU/s		
CO2 removal	0	BTU/s		
Cooling after CO shift	0	BTU/s		
Water condensed from syngas	20765	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
AGR Qrej	3314	BTU/s		
AGR heat loss	301.5	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	145687	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	7047	BTU/s		
Heat rejection from compressor inter/after cooling	27849	BTU/s		
Compressors mechnical & electrical losses	1476.8	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-308.1	BTU/s	-0.0374	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case A2a – Subcritical, No CCS, 0% biomass

GT PRO 21.0 parallel

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rk\Thermoflow\Better Plant Design\SUBCRITICAL QUENCH - 0% 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)	200019		8299		41.12	
Steam Turbine(s)	70324					
Plant Total	270343	185934	8496	12353	40.16	27.62

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
9.32	-8.99	19.82	75313

GT fuel HHV/LHV ratio =	1.083			
DB fuel HHV/LHV ratio =	1.083			
Total plant fuel HHV heat input / LHV heat input =	1.107			
Fuel HHV chemical energy input (77F/25C) =	2541736	kBTU/hr	706038	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2296761	kBTU/hr	637989	BTU/s
Total energy input (chemical LHV + ext. addn.) =	3311416	kBTU/hr	919838	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	3201091	kBTU/hr	889192	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200019	41.12	8299	1028	1100
Total	200019			1028	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1798009	kBTU/hr	499447	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1659960	kBTU/hr	461100	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
60.08	70324	38.56	23.17	-841026

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 =	1035892	kBTU/hr	287748	BTU/s
Water/steam to gasification plant =	28819	kBTU/hr	8005	BTU/s
Water/steam from gasification plant =	46694	kBTU/hr	12970	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	0	%		

Case A2b – Subcritical, No CCS, 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	Canadian Class 43
%	%	chargeable energy, %	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	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	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	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	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	%		

Case A2a – Subcritical, No CCS, 0% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	681	kW
Condensate pump(s)*	115.6	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	1.406	kW
Cooling water pump(s)	992.8	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	40	kW
Lights	65	kW
Aux. from PEACE running motor/load list	620.5	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	150.1	kW
Miscellaneous plant auxiliaries	270.3	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	37841	kW
Gasification plant, fuel preparation	7180	kW
Gasification plant, AGR*	675.1	kW
Gasification plant, other/misc	2534.7	kW
Desalination plant auxiliaries	0	kW
CO2 capture plant auxiliaries*	31528	kW
Program estimated overall plant auxiliaries	83057	kW
Actual (user input) overall plant auxiliaries	83057	kW
Transformer losses	1351.7	kW
Total auxiliaries & transformer losses	**84409**	**kW**
* Heat balance related auxiliaries		

Case A2b – Subcritical, No CCS, 10% biomass

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

Case A2a – Subcritical, No CCS, 0% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**868863**	**BTU/s**		
Ambient air sensible	11138	BTU/s		
Ambient air latent	21202	BTU/s		
GT syngas	515781	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	281848	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	12970	BTU/s		
CO2 capture condensate return	42991	BTU/s		
Energy Out	**869117**	**BTU/s**		
Net power output	176240	BTU/s		
Stack gas sensible	79245	BTU/s		
Stack gas latent	114251	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1425.2	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	153087	BTU/s		
Process steam	48230	BTU/s		
Process water	0	BTU/s		
Blowdown	656	BTU/s		
Heat radiated from steam cycle	2011.8	BTU/s		
ST/generator mech/elec/gear loss	1260.8	BTU/s		
Non-heat balance related auxiliaries	11580	BTU/s		
Transformer loss	1281.2	BTU/s		
ASU compressors	35868	BTU/s		
Water/steam to gasification plant	8005	BTU/s		
AGR auxiliary	639.9	BTU/s		
CO2 capture auxiliary	29884	BTU/s		
Steam to CO2 capture	201679	BTU/s		
Energy In - Energy Out	**-253.4**	**BTU/s**	-0.0292	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -253.3 BTU/s				

Case A2b – Subcritical, No CCS, 10% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**868004**	**BTU/s**		
Ambient air sensible	11178	BTU/s		
Ambient air latent	21278	BTU/s		
GT syngas	515868	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	281970	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	13059	BTU/s		
CO2 capture condensate return	41684	BTU/s		
Energy Out	**868258**	**BTU/s**		
Net power output	180340	BTU/s		
Stack gas sensible	78963	BTU/s		
Stack gas latent	114394	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1425.4	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	159671	BTU/s		
Process steam	45580	BTU/s		
Process water	0	BTU/s		
Blowdown	665.3	BTU/s		
Heat radiated from steam cycle	2029.5	BTU/s		
ST/generator mech/elec/gear loss	1289.6	BTU/s		
Non-heat balance related auxiliaries	13894	BTU/s		
Transformer loss	1293.5	BTU/s		
ASU compressors	32686	BTU/s		
Water/steam to gasification plant	6981	BTU/s		
AGR auxiliary	558.2	BTU/s		
CO2 capture auxiliary	29164	BTU/s		
Steam to CO2 capture	195549	BTU/s		
Energy In - Energy Out	**-253.6**	**BTU/s**	-0.0292	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -253.6 BTU/s				

Case A2a – Subcritical, No CCS, 0% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	1090732	BTU/s		
Power Block Energy In:				
Ambient air sensible	11138	BTU/s		
Ambient air latent	21202	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	281848	BTU/s		
Process return & makeup	0	BTU/s		
CO2 capture condensate return	42991	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	709190	BTU/s		
Gasifier slurry water	862	BTU/s		
Quench water	29860	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1352.5	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	9355	BTU/s		
Total Energy Out	1091004	BTU/s		
Power Block Energy Out:				
Net power output	176240	BTU/s		
Stack gas sensible	79245	BTU/s		
Stack gas latent	114251	BTU/s		
GT cycle losses	5198	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	153087	BTU/s		
Process	48230	BTU/s		
Steam cycle losses	3929	BTU/s		
Non-heat balance auxiliaries	11580	BTU/s		
Transformer losses	1281.2	BTU/s		
CO2 capture auxiliary	29884	BTU/s		
Steam to CO2 capture	201679	BTU/s		
Gasifier Energy Out:				
Heat losses	63.8	BTU/s		
Slag	14473	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	5197	BTU/s		
CO2 removal	0	BTU/s		
Cooling after CO shift	0	BTU/s		
Water condensed from syngas	20152	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
AGR Qrej	6539	BTU/s		
AGR heat loss	611.2	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	176166	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	8558	BTU/s		
Heat rejection from compressor inter/after cooling	33819	BTU/s		
Compressors mechnical & electrical losses	1793.4	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-271.6	BTU/s	-0.0249	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case A2b – Subcritical, No CCS, 10% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	1076128	BTU/s		
Power Block Energy In:				
Ambient air sensible	11178	BTU/s		
Ambient air latent	21278	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	281970	BTU/s		
Process return & makeup	0	BTU/s		
CO2 capture condensate return	41684	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	697915	BTU/s		
Gasifier slurry water	830.4	BTU/s		
Quench water	28494	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1285.9	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	8525	BTU/s		
Total Energy Out	1076401	BTU/s		
Power Block Energy Out:				
Net power output	180340	BTU/s		
Stack gas sensible	78963	BTU/s		
Stack gas latent	114394	BTU/s		
GT cycle losses	5198	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	159671	BTU/s		
Process	45580	BTU/s		
Steam cycle losses	3984	BTU/s		
Non-heat balance auxiliaries	13894	BTU/s		
Transformer losses	1293.5	BTU/s		
CO2 capture auxiliary	29164	BTU/s		
Steam to CO2 capture	195549	BTU/s		
Gasifier Energy Out:				
Heat losses	63.01	BTU/s		
Slag	13843	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	4532	BTU/s		
CO2 removal	0	BTU/s		
Cooling after CO shift	0	BTU/s		
Water condensed from syngas	19418	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
AGR Qrej	5723	BTU/s		
AGR heat loss	533	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	164890	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	7799	BTU/s		
Heat rejection from compressor inter/after cooling	30818	BTU/s		
Compressors mechnical & electrical losses	1634.3	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-272.2	BTU/s	-0.0253	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case A2c – Subcritical, Post-combustion CCS, 30% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUBCRITICAL QUENCH - 30% 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		8312		41.05	
Steam Turbine(s)	73450					
Plant Total	273467	187369	8238	12024	41.42	28.38

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
9.49	-9.39	20.18	74731

GT fuel HHV/LHV ratio =	1.082			
DB fuel HHV/LHV ratio =	1.082			
Total plant fuel HHV heat input / LHV heat input =	1.1			
Fuel HHV chemical energy input (77F/25C) =	2477354	kBTU/hr	688154	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2252876	kBTU/hr	625799	BTU/s
Total energy input (chemical LHV + ext. addn.) =	3267960	kBTU/hr	907767	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	3167925	kBTU/hr	879979	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200017	41.05	8312	1026	1102
Total	200017			1026	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1799210	kBTU/hr	499780	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1662469	kBTU/hr	461797	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
61.11	73450	39.57	24.18	-850996

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 =	1036494	kBTU/hr	287915	BTU/s
Water/steam to gasification plant =	19595	kBTU/hr	5443	BTU/s
Water/steam from gasification plant =	45892	kBTU/hr	12748	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	0	%		

Case A2d – Subcritical, Post-combustion CCS, 50% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUBCRITICAL QUENCH - 50% 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		8316		41.03	
Steam Turbine(s)	74027					
Plant Total	274044	184610	8166	12122	41.79	28.15

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
9.14	-9.88	19.98	72931

GT fuel HHV/LHV ratio =	1.081			
DB fuel HHV/LHV ratio =	1.081			
Total plant fuel HHV heat input / LHV heat input =	1.097			
Fuel HHV chemical energy input (77F/25C) =	2453745	kBTU/hr	681596	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2237767	kBTU/hr	621602	BTU/s
Total energy input (chemical LHV + ext. addn.) =	3252826	kBTU/hr	903563	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	3152790	kBTU/hr	875775	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200017	41.03	8316	1026	1103
Total	200017			1026	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1798487	kBTU/hr	499580	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1663296	kBTU/hr	462027	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
61.12	74027	39.86	24.36	-850972

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 =	1036830	kBTU/hr	288008	BTU/s
Water/steam to gasification plant =	14213	kBTU/hr	3948	BTU/s
Water/steam from gasification plant =	44765	kBTU/hr	12435	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	0	%		

Case A2c – Subcritical, Post-combustion CCS, 30% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	687.6	kW
Condensate pump(s)*	118.5	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	1.457	kW
Cooling water pump(s)	1045.8	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	40	kW
Lights	65	kW
Aux. from PEACE running motor/load list	630.5	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	156.7	kW
Miscellaneous plant auxiliaries	273.5	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	32798	kW
Gasification plant, fuel preparation	15000	kW
Gasification plant, AGR*	459.3	kW
Gasification plant, other/misc	2357.4	kW
Desalination plant auxiliaries	0	kW
CO2 capture plant auxiliaries*	30735	kW
Program estimated overall plant auxiliaries	84730	kW
Actual (user input) overall plant auxiliaries	84730	kW
Transformer losses	1367.3	kW
Total auxiliaries & transformer losses	**86098**	**kW**
* Heat balance related auxiliaries		

Case A2d – Subcritical, Post-combustion CCS, 50% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	691.3	kW
Condensate pump(s)*	118.4	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	1.454	kW
Cooling water pump(s)	1051.9	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	40	kW
Lights	65	kW
Aux. from PEACE running motor/load list	630.5	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	157.9	kW
Miscellaneous plant auxiliaries	274	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	31161	kW
Gasification plant, fuel preparation	20168	kW
Gasification plant, AGR*	333.4	kW
Gasification plant, other/misc	2307.4	kW
Desalination plant auxiliaries	0	kW
CO2 capture plant auxiliaries*	30702	kW
Program estimated overall plant auxiliaries	88064	kW
Actual (user input) overall plant auxiliaries	88064	kW
Transformer losses	1370.2	kW
Total auxiliaries & transformer losses	**89434**	**kW**
* Heat balance related auxiliaries		

Case A2c – Subcritical, Post-combustion CCS, 30% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**867316**	**BTU/s**		
Ambient air sensible	11192	BTU/s		
Ambient air latent	21305	BTU/s		
GT syngas	515509	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	281968	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	12748	BTU/s		
CO2 capture condensate return	41621	BTU/s		
Energy Out	**867570**	**BTU/s**		
Net power output	177600	BTU/s		
Stack gas sensible	78963	BTU/s		
Stack gas latent	113983	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1425.7	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	160714	BTU/s		
Process steam	45580	BTU/s		
Process water	0	BTU/s		
Blowdown	666	BTU/s		
Heat radiated from steam cycle	2030.6	BTU/s		
ST/generator mech/elec/gear loss	1295.5	BTU/s		
Non-heat balance related auxiliaries	18893	BTU/s		
Transformer loss	1296	BTU/s		
ASU compressors	31088	BTU/s		
Water/steam to gasification plant	5443	BTU/s		
AGR auxiliary	435.4	BTU/s		
CO2 capture auxiliary	29132	BTU/s		
Steam to CO2 capture	195251	BTU/s		
Energy In - Energy Out	**-253.9**	**BTU/s**	-0.0293	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -253.9 BTU/s				

Case A2d – Subcritical, Post-combustion CCS, 50% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**866619**	**BTU/s**		
Ambient air sensible	11206	BTU/s		
Ambient air latent	21332	BTU/s		
GT syngas	515147	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	281961	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	12435	BTU/s		
CO2 capture condensate return	41559	BTU/s		
Energy Out	**866873**	**BTU/s**		
Net power output	174984	BTU/s		
Stack gas sensible	78972	BTU/s		
Stack gas latent	113569	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1425.9	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	161654	BTU/s		
Process steam	45580	BTU/s		
Process water	0	BTU/s		
Blowdown	667	BTU/s		
Heat radiated from steam cycle	2032.2	BTU/s		
ST/generator mech/elec/gear loss	1301.8	BTU/s		
Non-heat balance related auxiliaries	23751	BTU/s		
Transformer loss	1298.8	BTU/s		
ASU compressors	29536	BTU/s		
Water/steam to gasification plant	3948	BTU/s		
AGR auxiliary	316	BTU/s		
CO2 capture auxiliary	29101	BTU/s		
Steam to CO2 capture	194963	BTU/s		
Energy In - Energy Out	**-254.1**	**BTU/s**	-0.0293	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -254.1 BTU/s				

Case A2c – Subcritical, Post-combustion CCS, 30% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	1068170	BTU/s		
Power Block Energy In:				
Ambient air sensible	11192	BTU/s		
Ambient air latent	21305	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	281968	BTU/s		
Process return & makeup	0	BTU/s		
CO2 capture condensate return	41621	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	691071	BTU/s		
Gasifier slurry water	818.6	BTU/s		
Quench water	27858	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1255.1	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	8109	BTU/s		
Total Energy Out	1068447	BTU/s		
Power Block Energy Out:				
Net power output	177600	BTU/s		
Stack gas sensible	78963	BTU/s		
Stack gas latent	113983	BTU/s		
GT cycle losses	5199	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	160714	BTU/s		
Process	45580	BTU/s		
Steam cycle losses	3992	BTU/s		
Non-heat balance auxiliaries	18893	BTU/s		
Transformer losses	1296	BTU/s		
CO2 capture auxiliary	29132	BTU/s		
Steam to CO2 capture	195251	BTU/s		
Gasifier Energy Out:				
Heat losses	62.58	BTU/s		
Slag	13451	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	3534	BTU/s		
CO2 removal	0	BTU/s		
Cooling after CO shift	0	BTU/s		
Water condensed from syngas	19016	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
AGR Qrej	4501	BTU/s		
AGR heat loss	415.6	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	159421	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	7417	BTU/s		
Heat rejection from compressor inter/after cooling	29312	BTU/s		
Compressors mechnical & electrical losses	1554.4	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-276.4	BTU/s	-0.0259	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case A2d – Subcritical, Post-combustion CCS, 50% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	1060433	BTU/s		
Power Block Energy In:				
Ambient air sensible	11206	BTU/s		
Ambient air latent	21332	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	281961	BTU/s		
Process return & makeup	0	BTU/s		
CO2 capture condensate return	41559	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	684422	BTU/s		
Gasifier slurry water	807.1	BTU/s		
Quench water	27239	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1223.4	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	7704	BTU/s		
Total Energy Out	1060703	BTU/s		
Power Block Energy Out:				
Net power output	174984	BTU/s		
Stack gas sensible	78972	BTU/s		
Stack gas latent	113569	BTU/s		
GT cycle losses	5199	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	161654	BTU/s		
Process	45580	BTU/s		
Steam cycle losses	4001	BTU/s		
Non-heat balance auxiliaries	23751	BTU/s		
Transformer losses	1298.8	BTU/s		
CO2 capture auxiliary	29101	BTU/s		
Steam to CO2 capture	194963	BTU/s		
Gasifier Energy Out:				
Heat losses	62.16	BTU/s		
Slag	13071	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	2562.9	BTU/s		
CO2 removal	0	BTU/s		
Cooling after CO shift	0	BTU/s		
Water condensed from syngas	18618	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
AGR Qrej	3314	BTU/s		
AGR heat loss	301.5	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	154127	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	7047	BTU/s		
Heat rejection from compressor inter/after cooling	27849	BTU/s		
Compressors mechnical & electrical losses	1476.8	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-270.6	BTU/s	-0.0255	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case A3a – Subcritical, Sour-shift CCS, 0% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUBCRITICAL QUENCH - 0% BIOMASS - PRE-CCS (SOUR).gtp
Plant Configuration: GT, HRSG, and condensing reheat ST
One SGT6-4000F Engine (Physical Model #141), One Steam Turbine, GT PRO Type 10, Subtype 2
Steam Property Formulation: Thermoflow - STQUIK

SYSTEM SUMMARY

	Power Output kW		LHV Heat Rate BTU/kWh		Elect. Eff. LHV%	
	@ gen. term.	net	@ gen. term.	net	@ gen. term.	net
Gas Turbine(s)	200015		8201		41.61	
Steam Turbine(s)	98523					
Plant Total	298538	218279	8237	11265	41.43	30.29

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
25.21	20.13	27.31	8586

GT fuel HHV/LHV ratio =	1.179			
DB fuel HHV/LHV ratio =	1.179			
Total plant fuel HHV heat input / LHV heat input =	1.107			
Fuel HHV chemical energy input (77F/25C) =	2721234	kBTU/hr	755898	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2458959	kBTU/hr	683044	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2708789	kBTU/hr	752442	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2727594	kBTU/hr	757665	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200015	41.61	8201	945	1122
Total	200015			945	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1934546	kBTU/hr	537374	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1640257	kBTU/hr	455627	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
77.56	98523	42.60	33.04	-249830

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 =	1017614	kBTU/hr	282671	BTU/s
Water/steam to gasification plant =	51838	kBTU/hr	14399	BTU/s
Water/steam from gasification plant =	190894	kBTU/hr	53026	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-50.47	%		

Case A3b – Subcritical, Sour-shift CCS, 10% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUBCRITICAL QUENCH - 10% BIOMASS - PRE-CCS (SOUR).GTP
Plant Configuration: GT, HRSG, and condensing reheat ST
One SGT6-4000F Engine (Physical Model #141), One Steam Turbine, GT PRO Type 10, Subtype 2
Steam Property Formulation: Thermoflow - STQUIK

SYSTEM SUMMARY

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

PLANT EFFICIENCIES

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

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

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

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

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

STEAM CYCLE PERFORMANCE

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

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

Case A3a – Subcritical, Sour-shift CCS, 0% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	833.8	kW
Condensate pump(s)*	198.8	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	3.213	kW
Cooling water pump(s)	1106.6	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	45	kW
Lights	75	kW
Aux. from PEACE running motor/load list	781.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	209.9	kW
Miscellaneous plant auxiliaries	298.5	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	40513	kW
Gasification plant, fuel preparation	7687	kW
Gasification plant, CO2 capture and AGR*	23938	kW
Gasification plant, other/misc	2713.7	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	78766	kW
Actual (user input) overall plant auxiliaries	78766	kW
Transformer losses	1492.7	kW
Total auxiliaries & transformer losses	**80258**	**kW**
* Heat balance related auxiliaries		

Case A3b – Subcritical, Sour-shift CCS, 10% biomass

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

Case A3a – Subcritical, Sour-shift CCS, 0% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**709548**	**BTU/s**		
Ambient air sensible	11284	BTU/s		
Ambient air latent	21479	BTU/s		
GT syngas	554737	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69397	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	53026	BTU/s		
Energy Out	**709764**	**BTU/s**		
Net power output	206898	BTU/s		
Stack gas sensible	75556	BTU/s		
Stack gas latent	158979	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1409.9	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	168854	BTU/s		
Process steam	0	BTU/s		
Process water	0	BTU/s		
Blowdown	1016.6	BTU/s		
Heat radiated from steam cycle	2206.4	BTU/s		
ST/generator mech/elec/gear loss	1575.2	BTU/s		
Non-heat balance related auxiliaries	12590	BTU/s		
Transformer loss	1414.9	BTU/s		
ASU compressors	38401	BTU/s		
Water/steam to gasification plant	14399	BTU/s		
CO2 capture & AGR auxiliary	22690	BTU/s		
Energy In - Energy Out	**-215.1**	**BTU/s**	-0.0303	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -215.1 BTU/s				

Case A3b – Subcritical, Sour-shift CCS, 10% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**710591**	**BTU/s**		
Ambient air sensible	11290	BTU/s		
Ambient air latent	21492	BTU/s		
GT syngas	554832	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69384	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	53965	BTU/s		
Energy Out	**710806**	**BTU/s**		
Net power output	209204	BTU/s		
Stack gas sensible	75567	BTU/s		
Stack gas latent	159032	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1410.1	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	170136	BTU/s		
Process steam	0	BTU/s		
Process water	0	BTU/s		
Blowdown	1014.8	BTU/s		
Heat radiated from steam cycle	2208.3	BTU/s		
ST/generator mech/elec/gear loss	1584.1	BTU/s		
Non-heat balance related auxiliaries	15006	BTU/s		
Transformer loss	1417.8	BTU/s		
ASU compressors	34977	BTU/s		
Water/steam to gasification plant	13499	BTU/s		
CO2 capture & AGR auxiliary	21977	BTU/s		
Energy In - Energy Out	**-214.6**	**BTU/s**	-0.0302	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -214.5 BTU/s				

Case A3a – Subcritical, Sour-shift CCS, 0% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	905414	BTU/s		
Power Block Energy In:				
Ambient air sensible	11284	BTU/s		
Ambient air latent	21479	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69397	BTU/s		
Process return & makeup	0	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	759273	BTU/s		
Gasifier slurry water	922.9	BTU/s		
Quench water	31968	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1448.1	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	10016	BTU/s		
Total Energy Out	905684	BTU/s		
Power Block Energy Out:				
Net power output	206898	BTU/s		
Stack gas sensible	75556	BTU/s		
Stack gas latent	158979	BTU/s		
GT cycle losses	5183	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	168854	BTU/s		
Process	0	BTU/s		
Steam cycle losses	4798	BTU/s		
Non-heat balance auxiliaries	12590	BTU/s		
Transformer losses	1414.9	BTU/s		
Gasifier Energy Out:				
Heat losses	68.3	BTU/s		
Slag	15495	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	5564	BTU/s		
CO2 removal	0	BTU/s		
Water condensed from syngas	15516	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
CO2 capture & AGR Qrej	30701	BTU/s		
CO2 capture & AGR heat loss	7460	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	156853	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	9162	BTU/s		
Heat rejection from compressor inter/after cooling	36207	BTU/s		
Compressors mechnical & electrical losses	1920	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-269.9	BTU/s	-0.0298	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case A3b – Subcritical, Sour-shift CCS, 10% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	890505	BTU/s		
Power Block Energy In:				
Ambient air sensible	11290	BTU/s		
Ambient air latent	21492	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69384	BTU/s		
Process return & makeup	0	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	746833	BTU/s		
Gasifier slurry water	888.6	BTU/s		
Quench water	30491	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1376	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	9123	BTU/s		
Total Energy Out	890764	BTU/s		
Power Block Energy Out:				
Net power output	209204	BTU/s		
Stack gas sensible	75567	BTU/s		
Stack gas latent	159032	BTU/s		
GT cycle losses	5183	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	170136	BTU/s		
Process	0	BTU/s		
Steam cycle losses	4807	BTU/s		
Non-heat balance auxiliaries	15006	BTU/s		
Transformer losses	1417.8	BTU/s		
Gasifier Energy Out:				
Heat losses	67.43	BTU/s		
Slag	14813	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	4850	BTU/s		
CO2 removal	0	BTU/s		
Water condensed from syngas	14946	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
CO2 capture & AGR Qrej	29142	BTU/s		
CO2 capture & AGR heat loss	7160	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	143594	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	8345	BTU/s		
Heat rejection from compressor inter/after cooling	32978	BTU/s		
Compressors mechnical & electrical losses	1748.8	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-258.9	BTU/s	-0.0291	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case A3c – Subcritical, Sour-shift CCS, 30% biomass

GT PRO 21.0 parallel
1263 08-15-2011 15:51:19 file=C:\Documents and Settings\Hank\Desktop\Hank's Papers and Data\Grad Work\Thermoflow\Better Plant Design\SUBCRITICAL QUENCH - 30% BIOMASS - PRE-CCS (SOUR).gtp
Plant Configuration: GT, HRSG, and condensing reheat ST
One SGT6-4000F Engine (Physical Model #141), One Steam Turbine, GT PRO Type 10, Subtype 2
Steam Property Formulation: Thermoflow - STQUIK

SYSTEM SUMMARY

	Power Output kW		LHV Heat Rate BTU/kWh		Elect. Eff. LHV%	
	@ gen. term.	net	@ gen. term.	net	@ gen. term.	net
Gas Turbine(s)	200014		8206		41.58	
Steam Turbine(s)	99725					
Plant Total	299740	217639	8046	11082	42.41	30.79

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
25.61	20.44	27.71	8545

GT fuel HHV/LHV ratio =	1.179			
DB fuel HHV/LHV ratio =	1.179			
Total plant fuel HHV heat input / LHV heat input =	1.1			
Fuel HHV chemical energy input (77F/25C) =	2652132	kBTU/hr	736703	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2411817	kBTU/hr	669949	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2661571	kBTU/hr	739325	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2680370	kBTU/hr	744547	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200014	41.58	8206	945	1122
Total	200014			945	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1935807	kBTU/hr	537724	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1641370	kBTU/hr	455936	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
77.56	99725	43.10	33.43	-249754

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 =	1018002	kBTU/hr	282778	BTU/s
Water/steam to gasification plant =	43105	kBTU/hr	11974	BTU/s
Water/steam from gasification plant =	195560	kBTU/hr	54322	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-50.67	%		

Case A3d – Subcritical, Sour-shift CCS, 50% biomass

GT PRO 21.0 parallel
1263 08-15-2011 15:58:11 file=C:\Documents and Settings\Hank\Desktop\Hank's Papers and Data\Grad Work\Thermoflow\Better Plant Design\SUBCRITICAL QUENCH - 50% BIOMASS - PRE-CCS (SOUR).gtp
Plant Configuration: GT, HRSG, and condensing reheat ST
One SGT6-4000F Engine (Physical Model #141), One Steam Turbine, GT PRO Type 10, Subtype 2
Steam Property Formulation: Thermoflow - STQUIK

SYSTEM SUMMARY

	Power Output kW		LHV Heat Rate BTU/kWh		Elect. Eff. LHV%	
	@ gen. term.	net	@ gen. term.	net	@ gen. term.	net
Gas Turbine(s)	200014		8207		41.58	
Steam Turbine(s)	100299					
Plant Total	300313	214643	7981	11166	42.76	30.56

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
25.35	20.14	27.48	8524

GT fuel HHV/LHV ratio =	1.179			
DB fuel HHV/LHV ratio =	1.179			
Total plant fuel HHV heat input / LHV heat input =	1.097			
Fuel HHV chemical energy input (77F/25C) =	2627996	kBTU/hr	729999	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2396680	kBTU/hr	665745	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2646406	kBTU/hr	735113	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2665203	kBTU/hr	740334	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200014	41.58	8207	946	1122
Total	200014			946	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1935970	kBTU/hr	537769	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1641534	kBTU/hr	455982	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
77.55	100299	43.34	33.61	-249726

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 =	1018166	kBTU/hr	282824	BTU/s
Water/steam to gasification plant =	37772	kBTU/hr	10492	BTU/s
Water/steam from gasification plant =	196871	kBTU/hr	54686	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-51.74	%		

Case A3c – Subcritical, Sour-shift CCS, 30% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	829.8	kW
Condensate pump(s)*	200	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	3.247	kW
Cooling water pump(s)	1124.8	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	45	kW
Lights	75	kW
Aux. from PEACE running motor/load list	781.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	212.4	kW
Miscellaneous plant auxiliaries	299.7	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	35112	kW
Gasification plant, fuel preparation	16058	kW
Gasification plant, CO2 capture and AGR*	22975	kW
Gasification plant, other/misc	2523.7	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	80602	kW
Actual (user input) overall plant auxiliaries	80602	kW
Transformer losses	1498.7	kW
Total auxiliaries & transformer losses	**82101**	**kW**
* Heat balance related auxiliaries		

Case A3d – Subcritical, Sour-shift CCS, 50% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	828.3	kW
Condensate pump(s)*	200.4	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	3.26	kW
Cooling water pump(s)	1134.4	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	45	kW
Lights	75	kW
Aux. from PEACE running motor/load list	791.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	213.6	kW
Miscellaneous plant auxiliaries	300.3	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	33373	kW
Gasification plant, fuel preparation	21600	kW
Gasification plant, CO2 capture and AGR*	22770	kW
Gasification plant, other/misc	2471.2	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	84169	kW
Actual (user input) overall plant auxiliaries	84169	kW
Transformer losses	1501.6	kW
Total auxiliaries & transformer losses	**85670**	**kW**
* Heat balance related auxiliaries		

Case A3c – Subcritical, Sour-shift CCS, 30% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**710995**	**BTU/s**		
Ambient air sensible	11292	BTU/s		
Ambient air latent	21495	BTU/s		
GT syngas	554879	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69376	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	54322	BTU/s		
Energy Out	**711210**	**BTU/s**		
Net power output	206292	BTU/s		
Stack gas sensible	75585	BTU/s		
Stack gas latent	159036	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.8	BTU/s		
GT miscellaneous losses	1410.2	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	171481	BTU/s		
Process steam	0	BTU/s		
Process water	0	BTU/s		
Blowdown	1013.8	BTU/s		
Heat radiated from steam cycle	2209.4	BTU/s		
ST/generator mech/elec/gear loss	1592.5	BTU/s		
Non-heat balance related auxiliaries	20365	BTU/s		
Transformer loss	1420.6	BTU/s		
ASU compressors	33281	BTU/s		
Water/steam to gasification plant	11974	BTU/s		
CO2 capture & AGR auxiliary	21777	BTU/s		
Energy In - Energy Out	**-214.4**	**BTU/s**	-0.0302	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -214.4 BTU/s				

Case A3d – Subcritical, Sour-shift CCS, 50% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**711404**	**BTU/s**		
Ambient air sensible	11293	BTU/s		
Ambient air latent	21498	BTU/s		
GT syngas	554925	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69368	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	54686	BTU/s		
Energy Out	**711618**	**BTU/s**		
Net power output	203452	BTU/s		
Stack gas sensible	75602	BTU/s		
Stack gas latent	159039	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.8	BTU/s		
GT miscellaneous losses	1410.4	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	172798	BTU/s		
Process steam	0	BTU/s		
Process water	0	BTU/s		
Blowdown	1012.7	BTU/s		
Heat radiated from steam cycle	2210.5	BTU/s		
ST/generator mech/elec/gear loss	1600.8	BTU/s		
Non-heat balance related auxiliaries	25588	BTU/s		
Transformer loss	1423.3	BTU/s		
ASU compressors	31633	BTU/s		
Water/steam to gasification plant	10492	BTU/s		
CO2 capture & AGR auxiliary	21583	BTU/s		
Energy In - Energy Out	**-214.2**	**BTU/s**	-0.0301	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -214.2 BTU/s				

Case A3c – Subcritical, Sour-shift CCS, 30% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	882344	BTU/s		
Power Block Energy In:				
Ambient air sensible	11292	BTU/s		
Ambient air latent	21495	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69376	BTU/s		
Process return & makeup	0	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	739826	BTU/s		
Gasifier slurry water	876.3	BTU/s		
Quench water	29823	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1343.6	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	8681	BTU/s		
Total Energy Out	882603	BTU/s		
Power Block Energy Out:				
Net power output	206292	BTU/s		
Stack gas sensible	75585	BTU/s		
Stack gas latent	159036	BTU/s		
GT cycle losses	5183	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	171481	BTU/s		
Process	0	BTU/s		
Steam cycle losses	4816	BTU/s		
Non-heat balance auxiliaries	20365	BTU/s		
Transformer losses	1420.6	BTU/s		
Gasifier Energy Out:				
Heat losses	66.99	BTU/s		
Slag	14401	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	3783	BTU/s		
CO2 removal	0	BTU/s		
Water condensed from syngas	14543	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
CO2 capture & AGR Qrej	27764	BTU/s		
CO2 capture & AGR heat loss	7013	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	137036	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	7941	BTU/s		
Heat rejection from compressor inter/after cooling	31380	BTU/s		
Compressors mechnical & electrical losses	1664.1	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-258.6	BTU/s	-0.0293	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case A3d – Subcritical, Sour-shift CCS, 50% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	874418	BTU/s		
Power Block Energy In:				
Ambient air sensible	11293	BTU/s		
Ambient air latent	21498	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69368	BTU/s		
Process return & makeup	0	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	733026	BTU/s		
Gasifier slurry water	864.4	BTU/s		
Quench water	29173	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1310.4	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	8251	BTU/s		
Total Energy Out	874670	BTU/s		
Power Block Energy Out:				
Net power output	203452	BTU/s		
Stack gas sensible	75602	BTU/s		
Stack gas latent	159039	BTU/s		
GT cycle losses	5183	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	172798	BTU/s		
Process	0	BTU/s		
Steam cycle losses	4824	BTU/s		
Non-heat balance auxiliaries	25588	BTU/s		
Transformer losses	1423.3	BTU/s		
Gasifier Energy Out:				
Heat losses	66.57	BTU/s		
Slag	14000	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	2744.9	BTU/s		
CO2 removal	0	BTU/s		
Water condensed from syngas	14142	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
CO2 capture & AGR Qrej	26424	BTU/s		
CO2 capture & AGR heat loss	6871	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	130659	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	7547	BTU/s		
Heat rejection from compressor inter/after cooling	29826	BTU/s		
Compressors mechnical & electrical losses	1581.7	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-252.4	BTU/s	-0.0289	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case A4a – Subcritical, Sweet-shift CCS, 0% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUBCRITICAL QUENCH - 0% BIOMASS - PRE-CCS (SWEET).gtp
Plant Configuration: GT, HRSG, and condensing reheat ST
One SGT6-4000F Engine (Physical Model #141), One Steam Turbine, GT PRO Type 10, Subtype 2
Steam Property Formulation: Thermoflow - STQUIK

SYSTEM SUMMARY

	Power Output kW		LHV Heat Rate BTU/kWh		Elect. Eff. LHV%	
	@ gen. term.	net	@ gen. term.	net	@ gen. term.	net
Gas Turbine(s)	200015		8250		41.36	
Steam Turbine(s)	78509					
Plant Total	278524	198120	8882	12486	38.42	27.33

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
22.24	17.15	24.63	9504

GT fuel HHV/LHV ratio =	1.179			
DB fuel HHV/LHV ratio =	1.179			
Total plant fuel HHV heat input / LHV heat input =	1.107			
Fuel HHV chemical energy input (77F/25C) =	2737574	kBTU/hr	760437	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2473725	kBTU/hr	687146	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2725419	kBTU/hr	757061	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2744364	kBTU/hr	762323	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200015	41.36	8250	946	1121
Total	200015			946	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1946047	kBTU/hr	540569	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1650105	kBTU/hr	458362	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
77.96	78509	33.76	26.32	-251694

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 =	1017700	kBTU/hr	282694	BTU/s
Water/steam to gasification plant =	227381	kBTU/hr	63161	BTU/s
Water/steam from gasification plant =	199725	kBTU/hr	55479	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-59.31	%		

Case A4b – Subcritical, Sweet-shift CCS, 10% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUBCRITICAL QUENCH - 10% BIOMASS - PRE-CCS (SWEET).gtp
Plant Configuration: GT, HRSG, and condensing reheat ST
One SGT6-4000F Engine (Physical Model #141), One Steam Turbine, GT PRO Type 10, Subtype 2
Steam Property Formulation: Thermoflow - STQUIK

SYSTEM SUMMARY

	Power Output kW		LHV Heat Rate BTU/kWh		Elect. Eff. LHV%	
	@ gen. term.	net	@ gen. term.	net	@ gen. term.	net
Gas Turbine(s)	200015		8253		41.35	
Steam Turbine(s)	78861					
Plant Total	278876	200290	8755	12190	38.98	27.99

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
22.84	17.68	25.20	9491

GT fuel HHV/LHV ratio =	1.179			
DB fuel HHV/LHV ratio =	1.179			
Total plant fuel HHV heat input / LHV heat input =	1.103			
Fuel HHV chemical energy input (77F/25C) =	2692465	kBTU/hr	747907	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2441453	kBTU/hr	678181	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2693146	kBTU/hr	748096	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2712090	kBTU/hr	753358	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200015	41.35	8253	946	1122
Total	200015			946	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1946709	kBTU/hr	540753	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1650689	kBTU/hr	458525	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
77.96	78861	33.91	26.44	-251693

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 =	1017914	kBTU/hr	282754	BTU/s
Water/steam to gasification plant =	223539	kBTU/hr	62094	BTU/s
Water/steam from gasification plant =	200308	kBTU/hr	55641	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-58.3	%		

Case A4a – Subcritical, Sweet-shift CCS, 0% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	954.6	kW
Condensate pump(s)*	212.2	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	3.632	kW
Cooling water pump(s)	935.4	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	45	kW
Lights	75	kW
Aux. from PEACE running motor/load list	795.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	167.4	kW
Miscellaneous plant auxiliaries	278.5	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	40756	kW
Gasification plant, fuel preparation	7733	kW
Gasification plant, CO2 capture and AGR*	23963	kW
Gasification plant, other/misc	2730	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	79011	kW
Actual (user input) overall plant auxiliaries	79011	kW
Transformer losses	1392.6	kW
Total auxiliaries & transformer losses	**80404**	**kW**
* Heat balance related auxiliaries		

Case A4b – Subcritical, Sweet-shift CCS, 10% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	954.9	kW
Condensate pump(s)*	212.5	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	3.642	kW
Cooling water pump(s)	942.2	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	45	kW
Lights	75	kW
Aux. from PEACE running motor/load list	795.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	168.2	kW
Miscellaneous plant auxiliaries	278.9	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	37114	kW
Gasification plant, fuel preparation	10421	kW
Gasification plant, CO2 capture and AGR*	23227	kW
Gasification plant, other/misc	2592.5	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	77192	kW
Actual (user input) overall plant auxiliaries	77192	kW
Transformer losses	1394.4	kW
Total auxiliaries & transformer losses	**78586**	**kW**
* Heat balance related auxiliaries		

Case A4a – Subcritical, Sweet-shift CCS, 0% biomass

POWER BLOCK HEAT BALANCE			
Energy In	**714300**	**BTU/s**	
Ambient air sensible	11268	BTU/s	
Ambient air latent	21451	BTU/s	
GT syngas	555253	BTU/s	
External gas addition to combustor	0	BTU/s	
Steam and water	69915	BTU/s	
Makeup and process return	933.8	BTU/s	
N2 from ASU	0	BTU/s	
Water/steam from gasification plant	55479	BTU/s	
Energy Out	**714544**	**BTU/s**	
Net power output	187790	BTU/s	
Stack gas sensible	74438	BTU/s	
Stack gas latent	159420	BTU/s	
GT mechanical loss	1062	BTU/s	
GT gear box loss	0	BTU/s	
GT generator loss	2710.9	BTU/s	
GT miscellaneous losses	1410	BTU/s	
GT ancillary heat rejected	0	BTU/s	
GT process air bleed	0	BTU/s	
Air to ASU	0	BTU/s	
Air to gasifier	0	BTU/s	
Fuel compressor mech/elec loss	0	BTU/s	
Supercharging fan mech/elec loss	0	BTU/s	
Condenser	145002	BTU/s	
Process steam	0.0012	BTU/s	
Process water	0	BTU/s	
Blowdown	931.9	BTU/s	
Heat radiated from steam cycle	2161.1	BTU/s	
ST/generator mech/elec/gear loss	1349.8	BTU/s	
Non-heat balance related auxiliaries	12441	BTU/s	
Transformer loss	1320	BTU/s	
ASU compressors	38631	BTU/s	
Water/steam to gasification plant	63161	BTU/s	
CO2 capture & AGR auxiliary	22713	BTU/s	
Energy In - Energy Out	**-243.8**	**BTU/s**	**-0.0341** %
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)			
Gas Turbine and Steam Cycle: Energy In - Energy Out = -243.8 BTU/s			

Case A4b – Subcritical, Sweet-shift CCS, 10% biomass

POWER BLOCK HEAT BALANCE			
Energy In	**714558**	**BTU/s**	
Ambient air sensible	11277	BTU/s	
Ambient air latent	21466	BTU/s	
GT syngas	555326	BTU/s	
External gas addition to combustor	0	BTU/s	
Steam and water	69915	BTU/s	
Makeup and process return	933.3	BTU/s	
N2 from ASU	0	BTU/s	
Water/steam from gasification plant	55641	BTU/s	
Energy Out	**714778**	**BTU/s**	
Net power output	189847	BTU/s	
Stack gas sensible	74437	BTU/s	
Stack gas latent	159458	BTU/s	
GT mechanical loss	1062	BTU/s	
GT gear box loss	0	BTU/s	
GT generator loss	2710.9	BTU/s	
GT miscellaneous losses	1410.2	BTU/s	
GT ancillary heat rejected	0	BTU/s	
GT process air bleed	0	BTU/s	
Air to ASU	0	BTU/s	
Air to gasifier	0	BTU/s	
Fuel compressor mech/elec loss	0	BTU/s	
Supercharging fan mech/elec loss	0	BTU/s	
Condenser	145929	BTU/s	
Process steam	0	BTU/s	
Process water	0	BTU/s	
Blowdown	932	BTU/s	
Heat radiated from steam cycle	2161.7	BTU/s	
ST/generator mech/elec/gear loss	1353.5	BTU/s	
Non-heat balance related auxiliaries	14866	BTU/s	
Transformer loss	1321.7	BTU/s	
ASU compressors	35179	BTU/s	
Water/steam to gasification plant	62094	BTU/s	
CO2 capture & AGR auxiliary	22016	BTU/s	
Energy In - Energy Out	**-220.4**	**BTU/s**	**-0.0308** %
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)			
Gas Turbine and Steam Cycle: Energy In - Energy Out = -220.5 BTU/s			

Case A4a – Subcritical, Sweet-shift CCS, 0% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	912022	BTU/s		
Power Block Energy In:				
Ambient air sensible	11268	BTU/s		
Ambient air latent	21451	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69915	BTU/s		
Process return & makeup	933.8	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	763833	BTU/s		
Gasifier slurry water	928.4	BTU/s		
Quench water	32160	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1456.7	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	10076	BTU/s		
Total Energy Out	912307	BTU/s		
Power Block Energy Out:				
Net power output	187790	BTU/s		
Stack gas sensible	74438	BTU/s		
Stack gas latent	159420	BTU/s		
GT cycle losses	5183	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	145002	BTU/s		
Process	0.0012	BTU/s		
Steam cycle losses	4443	BTU/s		
Non-heat balance auxiliaries	12441	BTU/s		
Transformer losses	1320	BTU/s		
Gasifier Energy Out:				
Heat losses	68.71	BTU/s		
Slag	15588	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	5597	BTU/s		
CO2 removal	0	BTU/s		
Water condensed from syngas	23699	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
CO2 capture & AGR Qrej	47796	BTU/s		
CO2 capture & AGR heat loss	7449	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	182017	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	9217	BTU/s		
Heat rejection from compressor inter/after cooling	36424	BTU/s		
Compressors mechnical & electrical losses	1931.6	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-284.8	BTU/s	-0.0312	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case A4b – Subcritical, Sweet-shift CCS, 10% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	896859	BTU/s		
Power Block Energy In:				
Ambient air sensible	11277	BTU/s		
Ambient air latent	21466	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69915	BTU/s		
Process return & makeup	933.3	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	751147	BTU/s		
Gasifier slurry water	893.7	BTU/s		
Quench water	30668	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1384	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	9176	BTU/s		
Total Energy Out	897120	BTU/s		
Power Block Energy Out:				
Net power output	189847	BTU/s		
Stack gas sensible	74437	BTU/s		
Stack gas latent	159458	BTU/s		
GT cycle losses	5183	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	145929	BTU/s		
Process	0	BTU/s		
Steam cycle losses	4447	BTU/s		
Non-heat balance auxiliaries	14866	BTU/s		
Transformer losses	1321.7	BTU/s		
Gasifier Energy Out:				
Heat losses	67.82	BTU/s		
Slag	14898	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	4878	BTU/s		
CO2 removal	0	BTU/s		
Water condensed from syngas	22888	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
CO2 capture & AGR Qrej	45947	BTU/s		
CO2 capture & AGR heat loss	7153	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	169697	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	8393	BTU/s		
Heat rejection from compressor inter/after cooling	33169	BTU/s		
Compressors mechnical & electrical losses	1758.9	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-260.9	BTU/s	-0.0291	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case B1a – Supercritical, No CCS, 0% biomass

GT PRO 21.0 parallel
1263 07-27-2011 16:28:35 file=C:\Documents and Settings\Hank\Desktop\Hank's Papers and Data\Grad Wo
rk\Thermoflow\Better Plant Design\SUPERCRITICAL QUENCH - 0% BIOMASS.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: IFC-67

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)	200019		8299		41.12	
Steam Turbine(s)	122573					
Plant Total	322592	267111	7705	9306	44.29	36.67

PLANT EFFICIENCIES			
PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
31.53	26.39	33.02	8104

GT fuel HHV/LHV ratio =	1.083			
DB fuel HHV/LHV ratio =	1.11			
Total plant fuel HHV heat input / LHV heat input =	1.107			
Fuel HHV chemical energy input (77F/25C) =	2751301	kBTU/hr	764250	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2485624	kBTU/hr	690451	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2741115	kBTU/hr	761421	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2760346	kBTU/hr	766763	BTU/s

GAS TURBINE PERFORMANCE - SGT6-4000F (Physical Model #141)					
	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200019	41.12	8299	1028	1100
Total	200019			1028	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1798046	kBTU/hr	499457	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1659994	kBTU/hr	461109	BTU/s

STEAM CYCLE PERFORMANCE				
HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
80.94	122573	42.19	34.15	-255492

Number of steam turbine unit(s) =	1			
Fuel chemical HHV (77F/25C) to duct burners =	209505	kBTU/hr	58196	BTU/s
Fuel chemical LHV (77F/25C) to duct burners =	188809	kBTU/hr	52447	BTU/s
DB fuel chemical LHV + HRSG inlet sens. heat =	1224701	kBTU/hr	340195	BTU/s
Water/steam to gasification plant =	28818	kBTU/hr	8005	BTU/s
Water/steam from gasification plant =	79225	kBTU/hr	22007	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-38.95	%		

Case B1b – Supercritical, No CCS, 10% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUPERCRITICAL QUENCH - 10% BIOMASS.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: IFC-67

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		8308		41.07	
Steam Turbine(s)	122603					
Plant Total	322620	268207	7610	9154	44.84	37.28

PLANT EFFICIENCIES			
PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
32.07	26.87	33.53	8101

GT fuel HHV/LHV ratio =	1.083			
DB fuel HHV/LHV ratio =	1.11			
Total plant fuel HHV heat input / LHV heat input =	1.103			
Fuel HHV chemical energy input (77F/25C) =	2708926	kBTU/hr	752480	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2455228	kBTU/hr	682008	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2710672	kBTU/hr	752964	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2729899	kBTU/hr	758305	BTU/s

GAS TURBINE PERFORMANCE - SGT6-4000F (Physical Model #141)					
	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200017	41.07	8308	1026	1102
Total	200017			1026	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1799956	kBTU/hr	499988	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1661671	kBTU/hr	461575	BTU/s

STEAM CYCLE PERFORMANCE				
HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
80.93	122603	42.27	34.21	-255445

Number of steam turbine unit(s) =	1			
Fuel chemical HHV (77F/25C) to duct burners =	207219	kBTU/hr	57561	BTU/s
Fuel chemical LHV (77F/25C) to duct burners =	186749	kBTU/hr	51875	BTU/s
DB fuel chemical LHV + HRSG inlet sens. heat =	1222900	kBTU/hr	339694	BTU/s
Water/steam to gasification plant =	25132	kBTU/hr	6981	BTU/s
Water/steam from gasification plant =	78380	kBTU/hr	21772	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-38.72	%		

Case B1a – Supercritical, No CCS, 0% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	2332.2	kW
Condensate pump(s)*	206.1	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	3.007	kW
Cooling water pump(s)	1188.3	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	50	kW
Lights	90	kW
Aux. from PEACE running motor/load list	821.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	261	kW
Miscellaneous plant auxiliaries	322.6	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	37841	kW
Gasification plant, fuel preparation	7180	kW
Gasification plant, AGR*	675.2	kW
Gasification plant, other/misc	2534.7	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	53868	kW
Actual (user input) overall plant auxiliaries	53868	kW
Transformer losses	1613	kW
Total auxiliaries & transformer losses	**55481**	**kW**
* Heat balance related auxiliaries		

Case B1b – Supercritical, No CCS, 10% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	2328.4	kW
Condensate pump(s)*	205.8	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	2.999	kW
Cooling water pump(s)	1190.4	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	50	kW
Lights	90	kW
Aux. from PEACE running motor/load list	821.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	261	kW
Miscellaneous plant auxiliaries	322.6	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	34484	kW
Gasification plant, fuel preparation	9683	kW
Gasification plant, AGR*	589	kW
Gasification plant, other/misc	2408.8	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	52800	kW
Actual (user input) overall plant auxiliaries	52800	kW
Transformer losses	1613.1	kW
Total auxiliaries & transformer losses	**54413**	**kW**
* Heat balance related auxiliaries		

Case B1a – Supercritical, No CCS, 0% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**699051**	**BTU/s**		
Ambient air sensible	11138	BTU/s		
Ambient air latent	21199	BTU/s		
GT syngas & duct burner fuel	574200	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70970	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	22007	BTU/s		
Energy Out	**699261**	**BTU/s**		
Net power output	253184	BTU/s		
Stack gas sensible	76978	BTU/s		
Stack gas latent	120137	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1425.2	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	179599	BTU/s		
Process steam	0	BTU/s		
Process water	0	BTU/s		
Blowdown	1452.9	BTU/s		
Heat radiated from steam cycle	2612.8	BTU/s		
ST/generator mech/elec/gear loss	1912.5	BTU/s		
Non-heat balance related auxiliaries	12145	BTU/s		
Transformer loss	1528.9	BTU/s		
ASU compressors	35868	BTU/s		
Water/steam to gasification plant	8005	BTU/s		
AGR auxiliary	640	BTU/s		
Energy In - Energy Out	**-210.1**	**BTU/s**	-0.0301	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -210.1 BTU/s				

Case B1b – Supercritical, No CCS, 10% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**698373**	**BTU/s**		
Ambient air sensible	11178	BTU/s		
Ambient air latent	21276	BTU/s		
GT syngas & duct burner fuel	573650	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70957	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	21772	BTU/s		
Energy Out	**698584**	**BTU/s**		
Net power output	254223	BTU/s		
Stack gas sensible	76904	BTU/s		
Stack gas latent	120216	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1425.4	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	179915	BTU/s		
Process steam	0	BTU/s		
Process water	0	BTU/s		
Blowdown	1451	BTU/s		
Heat radiated from steam cycle	2608.8	BTU/s		
ST/generator mech/elec/gear loss	1912.9	BTU/s		
Non-heat balance related auxiliaries	14400	BTU/s		
Transformer loss	1529	BTU/s		
ASU compressors	32686	BTU/s		
Water/steam to gasification plant	6981	BTU/s		
AGR auxiliary	558.2	BTU/s		
Energy In - Energy Out	**-210.4**	**BTU/s**	-0.0301	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -210.4 BTU/s				

Case B1a – Supercritical, No CCS, 0% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	911899	BTU/s		
Power Block Energy In:				
Ambient air sensible	11138	BTU/s		
Ambient air latent	21199	BTU/s		
Duct burner fuel enthalpy	58409	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70970	BTU/s		
Process return & makeup	0	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	709207	BTU/s		
Gasifier slurry water	861.2	BTU/s		
Quench water	29844	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1378.4	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	9355	BTU/s		
Total Energy Out	912143	BTU/s		
Power Block Energy Out:				
Net power output	253184	BTU/s		
Stack gas sensible	76978	BTU/s		
Stack gas latent	120137	BTU/s		
GT cycle losses	5198	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	179599	BTU/s		
Process	0	BTU/s		
Steam cycle losses	5978	BTU/s		
Non-heat balance auxiliaries	12145	BTU/s		
Transformer losses	1528.9	BTU/s		
Gasifier Energy Out:				
Heat losses	63.8	BTU/s		
Slag	14473	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	5197	BTU/s		
CO2 removal	0	BTU/s		
Cooling after CO shift	0	BTU/s		
Water condensed from syngas	23286	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
AGR Qrej	6539	BTU/s		
AGR heat loss	611.2	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	164026	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	8557	BTU/s		
Heat rejection from compressor inter/after cooling	33819	BTU/s		
Compressors mechnical & electrical losses	1793.4	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-244	BTU/s	-0.0268	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case B1b – Supercritical, No CCS, 10% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	897772	BTU/s		
Power Block Energy In:				
Ambient air sensible	11178	BTU/s		
Ambient air latent	21276	BTU/s		
Duct burner fuel enthalpy	57772	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70957	BTU/s		
Process return & makeup	0	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	697930	BTU/s		
Gasifier slurry water	829.6	BTU/s		
Quench water	28479	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1285.5	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	8525	BTU/s		
Total Energy Out	898016	BTU/s		
Power Block Energy Out:				
Net power output	254223	BTU/s		
Stack gas sensible	76904	BTU/s		
Stack gas latent	120216	BTU/s		
GT cycle losses	5198	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	179915	BTU/s		
Process	0	BTU/s		
Steam cycle losses	5973	BTU/s		
Non-heat balance auxiliaries	14400	BTU/s		
Transformer losses	1529	BTU/s		
Gasifier Energy Out:				
Heat losses	63.01	BTU/s		
Slag	13843	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	4532	BTU/s		
CO2 removal	0	BTU/s		
Cooling after CO shift	0	BTU/s		
Water condensed from syngas	22402	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
AGR Qrej	5723	BTU/s		
AGR heat loss	533.1	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	153195	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	7798	BTU/s		
Heat rejection from compressor inter/after cooling	30819	BTU/s		
Compressors mechnical & electrical losses	1634.3	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-244	BTU/s	-0.0272	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case B2a – Supercritical, Post-combustion CCS, 0% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUPERCRITICAL QUENCH - 0% 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: IFC-67

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)	200019		8299		41.12	
Steam Turbine(s)	93207					
Plant Total	293226	206495	8338	11840	40.93	28.82

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
11.57	-5.67	21.02	42565

GT fuel HHV/LHV ratio =	1.083			
DB fuel HHV/LHV ratio =	1.11			
Total plant fuel HHV heat input / LHV heat input =	1.107			
Fuel HHV chemical energy input (77F/25C) =	2706077	kBTU/hr	751688	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2444868	kBTU/hr	679130	BTU/s
Total energy input (chemical LHV + ext. addn.) =	3462669	kBTU/hr	961853	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	3351598	kBTU/hr	930999	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200019	41.12	8299	1028	1100
Total	200019			1028	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1798043	kBTU/hr	499456	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1659993	kBTU/hr	461109	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
65.07	93207	41.29	26.86	-843259

Number of steam turbine unit(s) =	1			
Fuel chemical HHV (77F/25C) to duct burners =	164289	kBTU/hr	45636	BTU/s
Fuel chemical LHV (77F/25C) to duct burners =	148060	kBTU/hr	41128	BTU/s
DB fuel chemical LHV + HRSG inlet sens. heat =	1183951	kBTU/hr	328875	BTU/s
Water/steam to gasification plant =	28817	kBTU/hr	8005	BTU/s
Water/steam from gasification plant =	54419	kBTU/hr	15116	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	0	%		

Case B2b – Supercritical, Post-combustion CCS, 10% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUPERCRITICAL 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: IFC-67

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		8308		41.07	
Steam Turbine(s)	94682					
Plant Total	294700	209765	8193	11511	41.65	29.64

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
12.18	-5.28	21.55	41227

GT fuel HHV/LHV ratio =	1.083			
DB fuel HHV/LHV ratio =	1.11			
Total plant fuel HHV heat input / LHV heat input =	1.103			
Fuel HHV chemical energy input (77F/25C) =	2663786	kBTU/hr	739941	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2414546	kBTU/hr	670707	BTU/s
Total energy input (chemical LHV + ext. addn.) =	3432298	kBTU/hr	953416	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	3321223	kBTU/hr	922562	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200017	41.07	8308	1026	1102
Total	200017			1026	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1799956	kBTU/hr	499988	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1661672	kBTU/hr	461575	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
65.03	94682	42.02	27.33	-843209

Number of steam turbine unit(s) =	1			
Fuel chemical HHV (77F/25C) to duct burners =	162079	kBTU/hr	45022	BTU/s
Fuel chemical LHV (77F/25C) to duct burners =	146068	kBTU/hr	40575	BTU/s
DB fuel chemical LHV + HRSG inlet sens. heat =	1182220	kBTU/hr	328394	BTU/s
Water/steam to gasification plant =	25132	kBTU/hr	6981	BTU/s
Water/steam from gasification plant =	53567	kBTU/hr	14880	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	0	%		

Case B2a – Supercritical, Post-combustion CCS, 0% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	1727.1	kW
Condensate pump(s)*	137.2	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	1.785	kW
Cooling water pump(s)	1114.9	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	40	kW
Lights	75	kW
Aux. from PEACE running motor/load list	672.5	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	198.6	kW
Miscellaneous plant auxiliaries	293.2	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	37841	kW
Gasification plant, fuel preparation	7180	kW
Gasification plant, AGR*	675.2	kW
Gasification plant, other/misc	2534.7	kW
Desalination plant auxiliaries	0	kW
CO2 capture plant auxiliaries*	32411	kW
Program estimated overall plant auxiliaries	85265	kW
Actual (user input) overall plant auxiliaries	85265	kW
Transformer losses	1466.1	kW
Total auxiliaries & transformer losses	**86731**	**kW**
* Heat balance related auxiliaries		

Case B2b – Supercritical, Post-combustion CCS, 10% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	1724.3	kW
Condensate pump(s)*	136.9	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	1.778	kW
Cooling water pump(s)	1146.5	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	40	kW
Lights	75	kW
Aux. from PEACE running motor/load list	672.5	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	201.7	kW
Miscellaneous plant auxiliaries	294.7	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	34484	kW
Gasification plant, fuel preparation	9683	kW
Gasification plant, AGR*	589	kW
Gasification plant, other/misc	2408.8	kW
Desalination plant auxiliaries	0	kW
CO2 capture plant auxiliaries*	31641	kW
Program estimated overall plant auxiliaries	83461	kW
Actual (user input) overall plant auxiliaries	83461	kW
Transformer losses	1473.5	kW
Total auxiliaries & transformer losses	**84935**	**kW**
* Heat balance related auxiliaries		

Case B2a – Supercritical, Post-combustion CCS, 0% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**919123**	**BTU/s**		
Ambient air sensible	11138	BTU/s		
Ambient air latent	21199	BTU/s		
GT syngas & duct burner fuel	561594	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	282723	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	15116	BTU/s		
CO2 capture condensate return	44512	BTU/s		
Energy Out	**919365**	**BTU/s**		
Net power output	195728	BTU/s		
Stack gas sensible	78854	BTU/s		
Stack gas latent	118866	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1425.2	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	170170	BTU/s		
Process steam	48484	BTU/s		
Process water	0	BTU/s		
Blowdown	935.1	BTU/s		
Heat radiated from steam cycle	2319.5	BTU/s		
ST/generator mech/elec/gear loss	1496	BTU/s		
Non-heat balance related auxiliaries	11822	BTU/s		
Transformer loss	1389.7	BTU/s		
ASU compressors	35868	BTU/s		
Water/steam to gasification plant	8005	BTU/s		
AGR auxiliary	640	BTU/s		
CO2 capture auxiliary	30721	BTU/s		
Steam to CO2 capture	208866	BTU/s		
Energy In - Energy Out	**-242.5**	**BTU/s**	-0.0264	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -242.5 BTU/s				

Case B2b – Supercritical, Post-combustion CCS, 10% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**917175**	**BTU/s**		
Ambient air sensible	11178	BTU/s		
Ambient air latent	21276	BTU/s		
GT syngas & duct burner fuel	561065	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	282709	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	14880	BTU/s		
CO2 capture condensate return	43185	BTU/s		
Energy Out	**917418**	**BTU/s**		
Net power output	198828	BTU/s		
Stack gas sensible	78783	BTU/s		
Stack gas latent	118948	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1425.4	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	174046	BTU/s		
Process steam	48484	BTU/s		
Process water	0	BTU/s		
Blowdown	933.4	BTU/s		
Heat radiated from steam cycle	2315.4	BTU/s		
ST/generator mech/elec/gear loss	1519.7	BTU/s		
Non-heat balance related auxiliaries	14110	BTU/s		
Transformer loss	1396.7	BTU/s		
ASU compressors	32686	BTU/s		
Water/steam to gasification plant	6981	BTU/s		
AGR auxiliary	558.2	BTU/s		
CO2 capture auxiliary	29991	BTU/s		
Steam to CO2 capture	202640	BTU/s		
Energy In - Energy Out	**-242.7**	**BTU/s**	-0.0265	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -242.7 BTU/s				

Case B2a – Supercritical, Post-combustion CCS, 0% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	1138859	BTU/s		
Power Block Energy In:				
Ambient air sensible	11138	BTU/s		
Ambient air latent	21199	BTU/s		
Duct burner fuel enthalpy	45803	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	282723	BTU/s		
Process return & makeup	0	BTU/s		
CO2 capture condensate return	44512	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	709205	BTU/s		
Gasifier slurry water	861.2	BTU/s		
Quench water	29844	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1378.4	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	9355	BTU/s		
Total Energy Out	1139119	BTU/s		
Power Block Energy Out:				
Net power output	195728	BTU/s		
Stack gas sensible	78854	BTU/s		
Stack gas latent	118866	BTU/s		
GT cycle losses	5198	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	170170	BTU/s		
Process	48484	BTU/s		
Steam cycle losses	4751	BTU/s		
Non-heat balance auxiliaries	11822	BTU/s		
Transformer losses	1389.7	BTU/s		
CO2 capture auxiliary	30721	BTU/s		
Steam to CO2 capture	208866	BTU/s		
Gasifier Energy Out:				
Heat losses	63.8	BTU/s		
Slag	14473	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	5197	BTU/s		
CO2 removal	0	BTU/s		
Cooling after CO shift	0	BTU/s		
Water condensed from syngas	20903	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
AGR Qrej	6539	BTU/s		
AGR heat loss	611.2	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	173281	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	8557	BTU/s		
Heat rejection from compressor inter/after cooling	33819	BTU/s		
Compressors mechnical & electrical losses	1793.4	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-260.1	BTU/s	-0.0228	%

Case B2b – Supercritical, Post-combustion CCS, 10% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	1123491	BTU/s		
Power Block Energy In:				
Ambient air sensible	11178	BTU/s		
Ambient air latent	21276	BTU/s		
Duct burner fuel enthalpy	45187	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	282709	BTU/s		
Process return & makeup	0	BTU/s		
CO2 capture condensate return	43185	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	697930	BTU/s		
Gasifier slurry water	829.6	BTU/s		
Quench water	28479	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1310.5	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	8525	BTU/s		
Total Energy Out	1123750	BTU/s		
Power Block Energy Out:				
Net power output	198828	BTU/s		
Stack gas sensible	78783	BTU/s		
Stack gas latent	118948	BTU/s		
GT cycle losses	5198	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	174046	BTU/s		
Process	48484	BTU/s		
Steam cycle losses	4769	BTU/s		
Non-heat balance auxiliaries	14110	BTU/s		
Transformer losses	1396.7	BTU/s		
CO2 capture auxiliary	29991	BTU/s		
Steam to CO2 capture	202640	BTU/s		
Gasifier Energy Out:				
Heat losses	63.01	BTU/s		
Slag	13843	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	4532	BTU/s		
CO2 removal	0	BTU/s		
Cooling after CO shift	0	BTU/s		
Water condensed from syngas	20049	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
AGR Qrej	5723	BTU/s		
AGR heat loss	533.1	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	162448	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	7798	BTU/s		
Heat rejection from compressor inter/after cooling	30819	BTU/s		
Compressors mechnical & electrical losses	1634.3	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-259.1	BTU/s	-0.0231	%

Case B2c – Supercritical, Post-combustion CCS, 30% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUPERCRITICAL QUENCH - 30% 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: IFC-67

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		8312		41.05	
Steam Turbine(s)	95173					
Plant Total	295190	206822	8127	11599	41.99	29.42

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
11.84	-5.73	21.35	40790

GT fuel HHV/LHV ratio =	1.082			
DB fuel HHV/LHV ratio =	1.11			
Total plant fuel HHV heat input / LHV heat input =	1.1			
Fuel HHV chemical energy input (77F/25C) =	2639445	kBTU/hr	733179	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2398955	kBTU/hr	666376	BTU/s
Total energy input (chemical LHV + ext. addn.) =	3416702	kBTU/hr	949084	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	3305627	kBTU/hr	918230	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200017	41.05	8312	1026	1102
Total	200017			1026	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1799247	kBTU/hr	499791	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1662503	kBTU/hr	461806	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
65.04	95173	42.23	27.46	-843205

Number of steam turbine unit(s) =	1			
Fuel chemical HHV (77F/25C) to duct burners =	162037	kBTU/hr	45010	BTU/s
Fuel chemical LHV (77F/25C) to duct burners =	146030	kBTU/hr	40564	BTU/s
DB fuel chemical LHV + HRSG inlet sens. heat =	1182524	kBTU/hr	328479	BTU/s
Water/steam to gasification plant =	19594	kBTU/hr	5443	BTU/s
Water/steam from gasification plant =	52447	kBTU/hr	14569	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	0	%		

Case B2d – Supercritical, Post-combustion CCS, 50% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUPERCRITICAL QUENCH - 50% 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: IFC-67

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		8316		41.03	
Steam Turbine(s)	95638					
Plant Total	295655	203954	8062	11687	42.33	29.20

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
11.51	-6.18	21.15	40376

GT fuel HHV/LHV ratio =	1.081			
DB fuel HHV/LHV ratio =	1.11			
Total plant fuel HHV heat input / LHV heat input =	1.097			
Fuel HHV chemical energy input (77F/25C) =	2615526	kBTU/hr	726535	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2383567	kBTU/hr	662102	BTU/s
Total energy input (chemical LHV + ext. addn.) =	3401304	kBTU/hr	944807	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	3290228	kBTU/hr	913952	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200017	41.03	8316	1026	1103
Total	200017			1026	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1798522	kBTU/hr	499590	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1663328	kBTU/hr	462036	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
65.04	95638	42.43	27.60	-843195

Number of steam turbine unit(s) =	1			
Fuel chemical HHV (77F/25C) to duct burners =	161730	kBTU/hr	44925	BTU/s
Fuel chemical LHV (77F/25C) to duct burners =	145753	kBTU/hr	40487	BTU/s
DB fuel chemical LHV + HRSG inlet sens. heat =	1182583	kBTU/hr	328495	BTU/s
Water/steam to gasification plant =	14212	kBTU/hr	3948	BTU/s
Water/steam from gasification plant =	51354	kBTU/hr	14265	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	0	%		

Case B2c – Supercritical, Post-combustion CCS, 30% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	1726.6	kW
Condensate pump(s)*	136.8	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	1.778	kW
Cooling water pump(s)	1154.5	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	40	kW
Lights	75	kW
Aux. from PEACE running motor/load list	674.5	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	202.7	kW
Miscellaneous plant auxiliaries	295.2	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	32799	kW
Gasification plant, fuel preparation	15000	kW
Gasification plant, AGR*	459.4	kW
Gasification plant, other/misc	2357.3	kW
Desalination plant auxiliaries	0	kW
CO2 capture plant auxiliaries*	31607	kW
Program estimated overall plant auxiliaries	86892	kW
Actual (user input) overall plant auxiliaries	86892	kW
Transformer losses	1476	kW
Total auxiliaries & transformer losses	**88368**	**kW**
* Heat balance related auxiliaries		

Case B2d – Supercritical, Post-combustion CCS, 50% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	1728.7	kW
Condensate pump(s)*	136.8	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	1.777	kW
Cooling water pump(s)	1162.1	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	40	kW
Lights	75	kW
Aux. from PEACE running motor/load list	674.5	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	203.7	kW
Miscellaneous plant auxiliaries	295.7	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	31161	kW
Gasification plant, fuel preparation	20168	kW
Gasification plant, AGR*	333.5	kW
Gasification plant, other/misc	2307.3	kW
Desalination plant auxiliaries	0	kW
CO2 capture plant auxiliaries*	31573	kW
Program estimated overall plant auxiliaries	90223	kW
Actual (user input) overall plant auxiliaries	90223	kW
Transformer losses	1478.3	kW
Total auxiliaries & transformer losses	**91702**	**kW**
* Heat balance related auxiliaries		

Case B2c – Supercritical, Post-combustion CCS, 30% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**916475**	**BTU/s**		
Ambient air sensible	11192	BTU/s		
Ambient air latent	21303	BTU/s		
GT syngas & duct burner fuel	560695	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	282708	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	14569	BTU/s		
CO2 capture condensate return	43121	BTU/s		
Energy Out	**916729**	**BTU/s**		
Net power output	196039	BTU/s		
Stack gas sensible	78811	BTU/s		
Stack gas latent	118535	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1425.7	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	175107	BTU/s		
Process steam	48484	BTU/s		
Process water	0	BTU/s		
Blowdown	934	BTU/s		
Heat radiated from steam cycle	2316	BTU/s		
ST/generator mech/elec/gear loss	1526.8	BTU/s		
Non-heat balance related auxiliaries	19112	BTU/s		
Transformer loss	1399	BTU/s		
ASU compressors	31089	BTU/s		
Water/steam to gasification plant	5443	BTU/s		
AGR auxiliary	435.4	BTU/s		
CO2 capture auxiliary	29959	BTU/s		
Steam to CO2 capture	202341	BTU/s		
Energy In - Energy Out	**-253.4**	**BTU/s**	-0.0277	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -253.4 BTU/s				

Case B2d – Supercritical, Post-combustion CCS, 50% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**915703**	**BTU/s**		
Ambient air sensible	11206	BTU/s		
Ambient air latent	21329	BTU/s		
GT syngas & duct burner fuel	560246	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	282705	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	14265	BTU/s		
CO2 capture condensate return	43057	BTU/s		
Energy Out	**915956**	**BTU/s**		
Net power output	193320	BTU/s		
Stack gas sensible	78812	BTU/s		
Stack gas latent	118112	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1426	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	176108	BTU/s		
Process steam	48484	BTU/s		
Process water	0	BTU/s		
Blowdown	934.4	BTU/s		
Heat radiated from steam cycle	2316.3	BTU/s		
ST/generator mech/elec/gear loss	1533.5	BTU/s		
Non-heat balance related auxiliaries	23972	BTU/s		
Transformer loss	1401.2	BTU/s		
ASU compressors	29536	BTU/s		
Water/steam to gasification plant	3948	BTU/s		
AGR auxiliary	316.1	BTU/s		
CO2 capture auxiliary	29927	BTU/s		
Steam to CO2 capture	202038	BTU/s		
Energy In - Energy Out	**-253.6**	**BTU/s**	-0.0277	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -253.6 BTU/s				

Case B2c – Supercritical, Post-combustion CCS, 30% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	1115521	BTU/s		
Power Block Energy In:				
Ambient air sensible	11192	BTU/s		
Ambient air latent	21303	BTU/s		
Duct burner fuel enthalpy	45175	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	282708	BTU/s		
Process return & makeup	0	BTU/s		
CO2 capture condensate return	43121	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	691086	BTU/s		
Gasifier slurry water	817.8	BTU/s		
Quench water	27844	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1278.3	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	8108	BTU/s		
Total Energy Out	1115792	BTU/s		
Power Block Energy Out:				
Net power output	196039	BTU/s		
Stack gas sensible	78811	BTU/s		
Stack gas latent	118535	BTU/s		
GT cycle losses	5199	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	175107	BTU/s		
Process	48484	BTU/s		
Steam cycle losses	4777	BTU/s		
Non-heat balance auxiliaries	19112	BTU/s		
Transformer losses	1399	BTU/s		
CO2 capture auxiliary	29959	BTU/s		
Steam to CO2 capture	202341	BTU/s		
Gasifier Energy Out:				
Heat losses	62.58	BTU/s		
Slag	13452	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	3534	BTU/s		
CO2 removal	0	BTU/s		
Cooling after CO shift	0	BTU/s		
Water condensed from syngas	19641	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
AGR Qrej	4501	BTU/s		
AGR heat loss	415.6	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	156981	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	7417	BTU/s		
Heat rejection from compressor inter/after cooling	29313	BTU/s		
Compressors mechnical & electrical losses	1554.4	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-270.8	BTU/s	-0.0243	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case B2d – Supercritical, Post-combustion CCS, 50% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	1107699	BTU/s		
Power Block Energy In:				
Ambient air sensible	11206	BTU/s		
Ambient air latent	21329	BTU/s		
Duct burner fuel enthalpy	45089	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	282705	BTU/s		
Process return & makeup	0	BTU/s		
CO2 capture condensate return	43057	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	684436	BTU/s		
Gasifier slurry water	806.4	BTU/s		
Quench water	27225	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1246.9	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	7704	BTU/s		
Total Energy Out	1107969	BTU/s		
Power Block Energy Out:				
Net power output	193320	BTU/s		
Stack gas sensible	78812	BTU/s		
Stack gas latent	118112	BTU/s		
GT cycle losses	5199	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	176108	BTU/s		
Process	48484	BTU/s		
Steam cycle losses	4784	BTU/s		
Non-heat balance auxiliaries	23972	BTU/s		
Transformer losses	1401.2	BTU/s		
CO2 capture auxiliary	29927	BTU/s		
Steam to CO2 capture	202038	BTU/s		
Gasifier Energy Out:				
Heat losses	62.16	BTU/s		
Slag	13072	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	2563	BTU/s		
CO2 removal	0	BTU/s		
Cooling after CO shift	0	BTU/s		
Water condensed from syngas	19244	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
AGR Qrej	3314	BTU/s		
AGR heat loss	301.5	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	151683	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	7047	BTU/s		
Heat rejection from compressor inter/after cooling	27849	BTU/s		
Compressors mechnical & electrical losses	1476.8	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-269.6	BTU/s	-0.0243	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case B3a – Supercritical, Sour-shift CCS, 0% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUPERCRITICAL QUENCH - 0% BIOMASS - PRE-CCS (SOUR).gtp
Plant Configuration: GT, HRSG, and condensing reheat ST
One SGT6-4000F Engine (Physical Model #141), One Steam Turbine, GT PRO Type 10, Subtype 2
Steam Property Formulation: IFC-67

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)	200014		8107		42.09	
Steam Turbine(s)	119089					
Plant Total	319103	238077	8118	10881	42.03	31.36

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
26.46	21.56	28.37	8540

GT fuel HHV/LHV ratio =	1.179			
DB fuel HHV/LHV ratio =	1.11			
Total plant fuel HHV heat input / LHV heat input =	1.107			
Fuel HHV chemical energy input (77F/25C) =	2867317	kBTU/hr	796477	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2590537	kBTU/hr	719594	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2844439	kBTU/hr	790122	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2863550	kBTU/hr	795431	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200014	42.09	8107	946	1122
Total	200014			946	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1912537	kBTU/hr	531260	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1621594	kBTU/hr	450443	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
80.77	119089	42.74	34.52	-253902

Number of steam turbine unit(s) =	1			
Fuel chemical HHV (77F/25C) to duct burners =	177041	kBTU/hr	49178	BTU/s
Fuel chemical LHV (77F/25C) to duct burners =	159552	kBTU/hr	44320	BTU/s
DB fuel chemical LHV + HRSG inlet sens. heat =	1177142	kBTU/hr	326984	BTU/s
Water/steam to gasification plant =	34485	kBTU/hr	9579	BTU/s
Water/steam from gasification plant =	97673	kBTU/hr	27131	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-45.46	%		

Case B3b – Supercritical, Sour-shift CCS, 10% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUPERCRITICAL QUENCH - 10% BIOMASS - PRE-CCS (SOUR).gtp
Plant Configuration: GT, HRSG, and condensing reheat ST
One SGT6-4000F Engine (Physical Model #141), One Steam Turbine, GT PRO Type 10, Subtype 2
Steam Property Formulation: IFC-67

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)	200014		8120		42.03	
Steam Turbine(s)	120838					
Plant Total	320852	241559	7988	10610	42.72	32.16

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
27.21	22.26	29.07	8494

GT fuel HHV/LHV ratio =	1.179			
DB fuel HHV/LHV ratio =	1.11			
Total plant fuel HHV heat input / LHV heat input =	1.103			
Fuel HHV chemical energy input (77F/25C) =	2827639	kBTU/hr	785455	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2563032	kBTU/hr	711953	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2816685	kBTU/hr	782413	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2835777	kBTU/hr	787716	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200014	42.03	8120	946	1122
Total	200014			946	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1915380	kBTU/hr	532050	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1624030	kBTU/hr	451119	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
80.75	120838	43.32	34.98	-253653

Number of steam turbine unit(s) =	1			
Fuel chemical HHV (77F/25C) to duct burners =	178649	kBTU/hr	49625	BTU/s
Fuel chemical LHV (77F/25C) to duct burners =	161001	kBTU/hr	44723	BTU/s
DB fuel chemical LHV + HRSG inlet sens. heat =	1178812	kBTU/hr	327448	BTU/s
Water/steam to gasification plant =	33671	kBTU/hr	9353	BTU/s
Water/steam from gasification plant =	113331	kBTU/hr	31481	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-44.45	%		

Case B3a – Supercritical, Sour-shift CCS, 0% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	2187.1	kW
Condensate pump(s)*	201.4	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	2.987	kW
Cooling water pump(s)	1151.5	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	50	kW
Lights	90	kW
Aux. from PEACE running motor/load list	821.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	253.6	kW
Miscellaneous plant auxiliaries	319.1	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	40052	kW
Gasification plant, fuel preparation	7600	kW
Gasification plant, CO2 capture and AGR*	23657	kW
Gasification plant, other/misc	2682.8	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	79430	kW
Actual (user input) overall plant auxiliaries	79430	kW
Transformer losses	1595.5	kW
Total auxiliaries & transformer losses	**81026**	**kW**
* Heat balance related auxiliaries		

Case B3b – Supercritical, Sour-shift CCS, 10% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	2152.2	kW
Condensate pump(s)*	204.3	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	3.064	kW
Cooling water pump(s)	1175	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	50	kW
Lights	90	kW
Aux. from PEACE running motor/load list	821.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	257.3	kW
Miscellaneous plant auxiliaries	320.9	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	36514	kW
Gasification plant, fuel preparation	10253	kW
Gasification plant, CO2 capture and AGR*	22935	kW
Gasification plant, other/misc	2550.6	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	77689	kW
Actual (user input) overall plant auxiliaries	77689	kW
Transformer losses	1604.3	kW
Total auxiliaries & transformer losses	**79293**	**kW**
* Heat balance related auxiliaries		

Case B3a – Supercritical, Sour-shift CCS, 0% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**733143**	**BTU/s**		
Ambient air sensible	11295	BTU/s		
Ambient air latent	21498	BTU/s		
GT syngas & duct burner fuel	603130	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70528	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	27131	BTU/s		
Energy Out	**733421**	**BTU/s**		
Net power output	225664	BTU/s		
Stack gas sensible	74867	BTU/s		
Stack gas latent	163021	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.8	BTU/s		
GT miscellaneous losses	1409.9	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	174811	BTU/s		
Process steam	0.0005	BTU/s		
Process water	0	BTU/s		
Blowdown	1383.8	BTU/s		
Heat radiated from steam cycle	2509.6	BTU/s		
ST/generator mech/elec/gear loss	1864.8	BTU/s		
Non-heat balance related auxiliaries	12638	BTU/s		
Transformer loss	1512.3	BTU/s		
ASU compressors	37964	BTU/s		
Water/steam to gasification plant	9579	BTU/s		
CO2 capture & AGR auxiliary	22423	BTU/s		
Energy In - Energy Out	**-277.2**	**BTU/s**	-0.0378	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -277.3 BTU/s				

Case B3b – Supercritical, Sour-shift CCS, 10% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**738066**	**BTU/s**		
Ambient air sensible	11300	BTU/s		
Ambient air latent	21509	BTU/s		
GT syngas & duct burner fuel	603750	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70459	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	31481	BTU/s		
Energy Out	**738284**	**BTU/s**		
Net power output	228964	BTU/s		
Stack gas sensible	74983	BTU/s		
Stack gas latent	163193	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.8	BTU/s		
GT miscellaneous losses	1410.1	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	177908	BTU/s		
Process steam	0	BTU/s		
Process water	0	BTU/s		
Blowdown	1369.5	BTU/s		
Heat radiated from steam cycle	2514.8	BTU/s		
ST/generator mech/elec/gear loss	1888.8	BTU/s		
Non-heat balance related auxiliaries	15055	BTU/s		
Transformer loss	1520.6	BTU/s		
ASU compressors	34610	BTU/s		
Water/steam to gasification plant	9353	BTU/s		
CO2 capture & AGR auxiliary	21739	BTU/s		
Energy In - Energy Out	**-217.4**	**BTU/s**	-0.0295	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -217.4 BTU/s				

Case B3a – Supercritical, Sour-shift CCS, 0% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	946735	BTU/s		
Power Block Energy In:				
Ambient air sensible	11295	BTU/s		
Ambient air latent	21498	BTU/s		
Duct burner fuel enthalpy	49358	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70528	BTU/s		
Process return & makeup	0	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	750636	BTU/s		
Gasifier slurry water	911.6	BTU/s		
Quench water	31588	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1458.9	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	9901	BTU/s		
Total Energy Out	947057	BTU/s		
Power Block Energy Out:				
Net power output	225664	BTU/s		
Stack gas sensible	74867	BTU/s		
Stack gas latent	163021	BTU/s		
GT cycle losses	5183	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	174811	BTU/s		
Process	0.0005	BTU/s		
Steam cycle losses	5758	BTU/s		
Non-heat balance auxiliaries	12638	BTU/s		
Transformer losses	1512.3	BTU/s		
Gasifier Energy Out:				
Heat losses	67.53	BTU/s		
Slag	15319	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	5500	BTU/s		
CO2 removal	0	BTU/s		
Water condensed from syngas	10448	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
CO2 capture & AGR Qrej	30344	BTU/s		
CO2 capture & AGR heat loss	7375	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	175250	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	9057	BTU/s		
Heat rejection from compressor inter/after cooling	35795	BTU/s		
Compressors mechnical & electrical losses	1898.2	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-322.1	BTU/s	-0.034	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case B3b – Supercritical, Sour-shift CCS, 10% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	933109	BTU/s		
Power Block Energy In:				
Ambient air sensible	11300	BTU/s		
Ambient air latent	21509	BTU/s		
Duct burner fuel enthalpy	49807	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70459	BTU/s		
Process return & makeup	0	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	739019	BTU/s		
Gasifier slurry water	878.5	BTU/s		
Quench water	30156	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1387.7	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	9027	BTU/s		
Total Energy Out	933374	BTU/s		
Power Block Energy Out:				
Net power output	228964	BTU/s		
Stack gas sensible	74983	BTU/s		
Stack gas latent	163193	BTU/s		
GT cycle losses	5183	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	177908	BTU/s		
Process	0	BTU/s		
Steam cycle losses	5773	BTU/s		
Non-heat balance auxiliaries	15055	BTU/s		
Transformer losses	1520.6	BTU/s		
Gasifier Energy Out:				
Heat losses	66.72	BTU/s		
Slag	14658	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	4799	BTU/s		
CO2 removal	0	BTU/s		
Water condensed from syngas	10741	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
CO2 capture & AGR Qrej	28829	BTU/s		
CO2 capture & AGR heat loss	7084	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	159150	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	8257	BTU/s		
Heat rejection from compressor inter/after cooling	32633	BTU/s		
Compressors mechnical & electrical losses	1730.5	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-264.1	BTU/s	-0.0283	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case B3c – Supercritical, Sour-shift CCS, 30% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUPERCRITICAL QUENCH - 30% BIOMASS - PRE-CCS (SOUR).gtp
Plant Configuration: GT, HRSG, and condensing reheat ST
One SGT6-4000F Engine (Physical Model #141), One Steam Turbine, GT PRO Type 10, Subtype 2
Steam Property Formulation: IFC-67

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)	200014		8128		41.98	
Steam Turbine(s)	122350					
Plant Total	322364	239387	7913	10656	43.12	32.02

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
27.05	22.08	28.93	8454

GT fuel HHV/LHV ratio =	1.179			
DB fuel HHV/LHV ratio =	1.11			
Total plant fuel HHV heat input / LHV heat input =	1.1			
Fuel HHV chemical energy input (77F/25C) =	2806720	kBTU/hr	779644	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2550926	kBTU/hr	708590	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2804416	kBTU/hr	779004	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2823496	kBTU/hr	784304	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200014	41.98	8128	946	1122
Total	200014			946	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1917227	kBTU/hr	532563	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1625617	kBTU/hr	451560	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
80.76	122350	43.80	35.37	-253490

Number of steam turbine unit(s) =	1			
Fuel chemical HHV (77F/25C) to duct burners =	180039	kBTU/hr	50011	BTU/s
Fuel chemical LHV (77F/25C) to duct burners =	162254	kBTU/hr	45070	BTU/s
DB fuel chemical LHV + HRSG inlet sens. heat =	1180231	kBTU/hr	327842	BTU/s
Water/steam to gasification plant =	30044	kBTU/hr	8346	BTU/s
Water/steam from gasification plant =	123975	kBTU/hr	34438	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-45	%		

Case B3d – Supercritical, Sour-shift CCS, 50% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUPERCRITICAL QUENCH - 50% BIOMASS - PRE-CCS (SOUR).gtp
Plant Configuration: GT, HRSG, and condensing reheat ST
One SGT6-4000F Engine (Physical Model #141), One Steam Turbine, GT PRO Type 10, Subtype 2
Steam Property Formulation: IFC-67

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)	200014		8136		41.94	
Steam Turbine(s)	123868					
Plant Total	323882	237309	7840	10701	43.52	31.89

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
26.90	21.91	28.80	8415

GT fuel HHV/LHV ratio =	1.179			
DB fuel HHV/LHV ratio =	1.11			
Total plant fuel HHV heat input / LHV heat input =	1.097			
Fuel HHV chemical energy input (77F/25C) =	2786613	kBTU/hr	774059	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2539382	kBTU/hr	705384	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2792719	kBTU/hr	775755	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2811787	kBTU/hr	781052	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200014	41.94	8136	946	1122
Total	200014			946	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1919112	kBTU/hr	533087	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1627240	kBTU/hr	452011	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
80.79	123868	44.27	35.77	-253337

Number of steam turbine unit(s) =	1			
Fuel chemical HHV (77F/25C) to duct burners =	181496	kBTU/hr	50416	BTU/s
Fuel chemical LHV (77F/25C) to duct burners =	163567	kBTU/hr	45435	BTU/s
DB fuel chemical LHV + HRSG inlet sens. heat =	1181709	kBTU/hr	328252	BTU/s
Water/steam to gasification plant =	26534	kBTU/hr	7371	BTU/s
Water/steam from gasification plant =	134427	kBTU/hr	37341	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-45.53	%		

Case B3c – Supercritical, Sour-shift CCS, 30% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	2127.8	kW
Condensate pump(s)*	206.4	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	3.116	kW
Cooling water pump(s)	1196.7	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	50	kW
Lights	90	kW
Aux. from PEACE running motor/load list	821.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	260.5	kW
Miscellaneous plant auxiliaries	322.4	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	34775	kW
Gasification plant, fuel preparation	15904	kW
Gasification plant, CO2 capture and AGR*	22747	kW
Gasification plant, other/misc	2499.4	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	81365	kW
Actual (user input) overall plant auxiliaries	81365	kW
Transformer losses	1611.8	kW
Total auxiliaries & transformer losses	**82977**	**kW**
* Heat balance related auxiliaries		

Case B3d – Supercritical, Sour-shift CCS, 50% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	2105	kW
Condensate pump(s)*	208.6	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	3.173	kW
Cooling water pump(s)	1217.4	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	50	kW
Lights	90	kW
Aux. from PEACE running motor/load list	821.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	263.7	kW
Miscellaneous plant auxiliaries	323.9	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	33083	kW
Gasification plant, fuel preparation	21412	kW
Gasification plant, CO2 capture and AGR*	22565	kW
Gasification plant, other/misc	2449.6	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	84954	kW
Actual (user input) overall plant auxiliaries	84954	kW
Transformer losses	1619.4	kW
Total auxiliaries & transformer losses	**86573**	**kW**
* Heat balance related auxiliaries		

Case B3c – Supercritical, Sour-shift CCS, 30% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**741494**	**BTU/s**		
Ambient air sensible	11301	BTU/s		
Ambient air latent	21510	BTU/s		
GT syngas & duct burner fuel	604259	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70414	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	34438	BTU/s		
Energy Out	**741708**	**BTU/s**		
Net power output	226906	BTU/s		
Stack gas sensible	75033	BTU/s		
Stack gas latent	163307	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.8	BTU/s		
GT miscellaneous losses	1410.2	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	180707	BTU/s		
Process steam	0.0002	BTU/s		
Process water	0	BTU/s		
Blowdown	1359.7	BTU/s		
Heat radiated from steam cycle	2519.5	BTU/s		
ST/generator mech/elec/gear loss	1909.5	BTU/s		
Non-heat balance related auxiliaries	20388	BTU/s		
Transformer loss	1527.8	BTU/s		
ASU compressors	32962	BTU/s		
Water/steam to gasification plant	8346	BTU/s		
CO2 capture & AGR auxiliary	21561	BTU/s		
Energy In - Energy Out	**-214.8**	**BTU/s**	**-0.029**	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -214.7 BTU/s				

Case B3d – Supercritical, Sour-shift CCS, 50% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**744889**	**BTU/s**		
Ambient air sensible	11302	BTU/s		
Ambient air latent	21511	BTU/s		
GT syngas & duct burner fuel	604786	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70371	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	37341	BTU/s		
Energy Out	**745181**	**BTU/s**		
Net power output	224936	BTU/s		
Stack gas sensible	75074	BTU/s		
Stack gas latent	163424	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.8	BTU/s		
GT miscellaneous losses	1410.4	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	183521	BTU/s		
Process steam	0.0002	BTU/s		
Process water	0	BTU/s		
Blowdown	1350.5	BTU/s		
Heat radiated from steam cycle	2524.9	BTU/s		
ST/generator mech/elec/gear loss	1930.2	BTU/s		
Non-heat balance related auxiliaries	25586	BTU/s		
Transformer loss	1535	BTU/s		
ASU compressors	31358	BTU/s		
Water/steam to gasification plant	7371	BTU/s		
CO2 capture & AGR auxiliary	21388	BTU/s		
Energy In - Energy Out	**-291.8**	**BTU/s**	**-0.0392**	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -291.8 BTU/s				

Case B3c – Supercritical, Sour-shift CCS, 30% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	926058	BTU/s		
Power Block Energy In:				
Ambient air sensible	11301	BTU/s		
Ambient air latent	21510	BTU/s		
Duct burner fuel enthalpy	50194	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70414	BTU/s		
Process return & makeup	0	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	732727	BTU/s		
Gasifier slurry water	867.1	BTU/s		
Quench water	29521	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1355.4	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	8597	BTU/s		
Total Energy Out	926320	BTU/s		
Power Block Energy Out:				
Net power output	226906	BTU/s		
Stack gas sensible	75033	BTU/s		
Stack gas latent	163307	BTU/s		
GT cycle losses	5183	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	180707	BTU/s		
Process	0.0002	BTU/s		
Steam cycle losses	5789	BTU/s		
Non-heat balance auxiliaries	20388	BTU/s		
Transformer losses	1527.8	BTU/s		
Gasifier Energy Out:				
Heat losses	66.35	BTU/s		
Slag	14262	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	3747	BTU/s		
CO2 removal	0	BTU/s		
Water condensed from syngas	10931	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
CO2 capture & AGR Qrej	27490	BTU/s		
CO2 capture & AGR heat loss	6945	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	150544	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	7864	BTU/s		
Heat rejection from compressor inter/after cooling	31079	BTU/s		
Compressors mechnical & electrical losses	1648.1	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-261.3	BTU/s	-0.0282	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case B3d – Supercritical, Sour-shift CCS, 50% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	919268	BTU/s		
Power Block Energy In:				
Ambient air sensible	11302	BTU/s		
Ambient air latent	21511	BTU/s		
Duct burner fuel enthalpy	50600	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70371	BTU/s		
Process return & makeup	0	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	726644	BTU/s		
Gasifier slurry water	856.1	BTU/s		
Quench water	28904	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1323.7	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	8179	BTU/s		
Total Energy Out	919606	BTU/s		
Power Block Energy Out:				
Net power output	224936	BTU/s		
Stack gas sensible	75074	BTU/s		
Stack gas latent	163424	BTU/s		
GT cycle losses	5183	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	183521	BTU/s		
Process	0.0002	BTU/s		
Steam cycle losses	5806	BTU/s		
Non-heat balance auxiliaries	25586	BTU/s		
Transformer losses	1535	BTU/s		
Gasifier Energy Out:				
Heat losses	65.99	BTU/s		
Slag	13878	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	2721	BTU/s		
CO2 removal	0	BTU/s		
Water condensed from syngas	11089	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
CO2 capture & AGR Qrej	26187	BTU/s		
CO2 capture & AGR heat loss	6810	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	142215	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	7481	BTU/s		
Heat rejection from compressor inter/after cooling	29567	BTU/s		
Compressors mechnical & electrical losses	1567.9	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-337.5	BTU/s	-0.0367	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case B4a – Supercritical, Sweet-shift CCS, 0% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUPERCRITICAL QUENCH - 0% BIOMASS - PRE-CCS (SWEET).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: IFC-67

SYSTEM SUMMARY

	Power Output kW		LHV Heat Rate BTU/kWh		Elect. Eff. LHV%	
	@ gen. term.	net	@ gen. term.	net	@ gen. term.	net
Gas Turbine(s)	200015		8250		41.36	
Steam Turbine(s)	107704					
Plant Total	307719	225406	8584	11719	39.75	29.12

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
24.30	19.47	26.38	9149

GT fuel HHV/LHV ratio =	1.179			
DB fuel HHV/LHV ratio =	1.11			
Total plant fuel HHV heat input / LHV heat input =	1.107			
Fuel HHV chemical energy input (77F/25C) =	2923792	kBTU/hr	812164	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2641547	kBTU/hr	733763	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2896293	kBTU/hr	804526	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2915467	kBTU/hr	809852	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200015	41.36	8250	946	1121
Total	200015			946	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1946084	kBTU/hr	540579	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1650137	kBTU/hr	458371	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
81.11	107704	38.22	31.00	-254746

Number of steam turbine unit(s) =	1			
Fuel chemical HHV (77F/25C) to duct burners =	186161	kBTU/hr	51711	BTU/s
Fuel chemical LHV (77F/25C) to duct burners =	167771	kBTU/hr	46603	BTU/s
DB fuel chemical LHV + HRSG inlet sens. heat =	1185469	kBTU/hr	329297	BTU/s
Water/steam to gasification plant =	202960	kBTU/hr	56378	BTU/s
Water/steam from gasification plant =	180369	kBTU/hr	50102	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-49.52	%		

Case B4b – Supercritical, Sweet-shift CCS, 10% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUPERCRITICAL QUENCH - 10% BIOMASS - PRE-CCS (SWEET).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: IFC-67

SYSTEM SUMMARY

	Power Output kW		LHV Heat Rate BTU/kWh		Elect. Eff. LHV%	
	@ gen. term.	net	@ gen. term.	net	@ gen. term.	net
Gas Turbine(s)	200015		8253		41.35	
Steam Turbine(s)	108015					
Plant Total	308030	227536	8470	11466	40.29	29.76

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
24.88	20.00	26.93	9137

GT fuel HHV/LHV ratio =	1.179			
DB fuel HHV/LHV ratio =	1.11			
Total plant fuel HHV heat input / LHV heat input =	1.103			
Fuel HHV chemical energy input (77F/25C) =	2878261	kBTU/hr	799517	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2608894	kBTU/hr	724693	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2863630	kBTU/hr	795453	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2882804	kBTU/hr	800779	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200015	41.35	8253	946	1122
Total	200015			946	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1946747	kBTU/hr	540763	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1650722	kBTU/hr	458534	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
81.11	108015	38.34	31.10	-254736

Number of steam turbine unit(s) =	1			
Fuel chemical HHV (77F/25C) to duct burners =	185738	kBTU/hr	51594	BTU/s
Fuel chemical LHV (77F/25C) to duct burners =	167390	kBTU/hr	46497	BTU/s
DB fuel chemical LHV + HRSG inlet sens. heat =	1185302	kBTU/hr	329251	BTU/s
Water/steam to gasification plant =	199129	kBTU/hr	55314	BTU/s
Water/steam from gasification plant =	180913	kBTU/hr	50254	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-48.83	%		

Case B4a – Supercritical, Sweet-shift CCS, 0% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	2368	kW
Condensate pump(s)*	226.4	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	3.593	kW
Cooling water pump(s)	1088.8	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	50	kW
Lights	90	kW
Aux. from PEACE running motor/load list	864.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	229.4	kW
Miscellaneous plant auxiliaries	307.7	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	40757	kW
Gasification plant, fuel preparation	7733	kW
Gasification plant, CO2 capture and AGR*	23964	kW
Gasification plant, other/misc	2730	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	80774	kW
Actual (user input) overall plant auxiliaries	80774	kW
Transformer losses	1538.6	kW
Total auxiliaries & transformer losses	**82313**	**kW**
* Heat balance related auxiliaries		

Case B4b – Supercritical, Sweet-shift CCS, 10% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	2367.7	kW
Condensate pump(s)*	226.7	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	3.601	kW
Cooling water pump(s)	1095.5	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	50	kW
Lights	90	kW
Aux. from PEACE running motor/load list	864.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	230	kW
Miscellaneous plant auxiliaries	308	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	37114	kW
Gasification plant, fuel preparation	10422	kW
Gasification plant, CO2 capture and AGR*	23227	kW
Gasification plant, other/misc	2592.5	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	78954	kW
Actual (user input) overall plant auxiliaries	78954	kW
Transformer losses	1540.2	kW
Total auxiliaries & transformer losses	**80494**	**kW**
* Heat balance related auxiliaries		

Case B4a – Supercritical, Sweet-shift CCS, 0% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**761669**	**BTU/s**		
Ambient air sensible	11268	BTU/s		
Ambient air latent	21448	BTU/s		
GT syngas & duct burner fuel	607164	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70763	BTU/s		
Makeup and process return	922.5	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	50102	BTU/s		
Energy Out	**761947**	**BTU/s**		
Net power output	213653	BTU/s		
Stack gas sensible	74196	BTU/s		
Stack gas latent	164651	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1410	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	166636	BTU/s		
Process steam	0	BTU/s		
Process water	0	BTU/s		
Blowdown	1449.9	BTU/s		
Heat radiated from steam cycle	2531.5	BTU/s		
ST/generator mech/elec/gear loss	1706.2	BTU/s		
Non-heat balance related auxiliaries	12757	BTU/s		
Transformer loss	1458.4	BTU/s		
ASU compressors	38632	BTU/s		
Water/steam to gasification plant	56378	BTU/s		
CO2 capture & AGR auxiliary	22714	BTU/s		
Energy In - Energy Out	**-278.4**	**BTU/s**	-0.0365	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -278.4 BTU/s				

Case B4b – Supercritical, Sweet-shift CCS, 10% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**761795**	**BTU/s**		
Ambient air sensible	11277	BTU/s		
Ambient air latent	21464	BTU/s		
GT syngas & duct burner fuel	607120	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70760	BTU/s		
Makeup and process return	922.1	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	50254	BTU/s		
Energy Out	**762081**	**BTU/s**		
Net power output	215673	BTU/s		
Stack gas sensible	74192	BTU/s		
Stack gas latent	164677	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1410.2	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	167513	BTU/s		
Process steam	0.0011	BTU/s		
Process water	0	BTU/s		
Blowdown	1449.6	BTU/s		
Heat radiated from steam cycle	2531.4	BTU/s		
ST/generator mech/elec/gear loss	1710.6	BTU/s		
Non-heat balance related auxiliaries	15182	BTU/s		
Transformer loss	1459.8	BTU/s		
ASU compressors	35179	BTU/s		
Water/steam to gasification plant	55314	BTU/s		
CO2 capture & AGR auxiliary	22016	BTU/s		
Energy In - Energy Out	**-285.2**	**BTU/s**	-0.0374	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -285.1 BTU/s				

Case B4a – Supercritical, Sweet-shift CCS, 0% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	964783	BTU/s		
Power Block Energy In:				
Ambient air sensible	11268	BTU/s		
Ambient air latent	21448	BTU/s		
Duct burner fuel enthalpy	51901	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70763	BTU/s		
Process return & makeup	922.5	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	763848	BTU/s		
Gasifier slurry water	927.6	BTU/s		
Quench water	32144	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1484.6	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	10076	BTU/s		
Total Energy Out	965104	BTU/s		
Power Block Energy Out:				
Net power output	213653	BTU/s		
Stack gas sensible	74196	BTU/s		
Stack gas latent	164651	BTU/s		
GT cycle losses	5183	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	166636	BTU/s		
Process	0	BTU/s		
Steam cycle losses	5688	BTU/s		
Non-heat balance auxiliaries	12757	BTU/s		
Transformer losses	1458.4	BTU/s		
Gasifier Energy Out:				
Heat losses	68.72	BTU/s		
Slag	15588	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	5597	BTU/s		
CO2 removal	0	BTU/s		
Water condensed from syngas	24176	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
CO2 capture & AGR Qrej	47797	BTU/s		
CO2 capture & AGR heat loss	7449	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	180150	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	9217	BTU/s		
Heat rejection from compressor inter/after cooling	36425	BTU/s		
Compressors mechnical & electrical losses	1931.6	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-321	BTU/s	-0.0333	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case B4b – Supercritical, Sweet-shift CCS, 10% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	949499	BTU/s		
Power Block Energy In:				
Ambient air sensible	11277	BTU/s		
Ambient air latent	21464	BTU/s		
Duct burner fuel enthalpy	51783	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70760	BTU/s		
Process return & makeup	922.1	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	751163	BTU/s		
Gasifier slurry water	892.9	BTU/s		
Quench water	30652	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1410.5	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	9175	BTU/s		
Total Energy Out	949824	BTU/s		
Power Block Energy Out:				
Net power output	215673	BTU/s		
Stack gas sensible	74192	BTU/s		
Stack gas latent	164677	BTU/s		
GT cycle losses	5183	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	167513	BTU/s		
Process	0.0011	BTU/s		
Steam cycle losses	5692	BTU/s		
Non-heat balance auxiliaries	15182	BTU/s		
Transformer losses	1459.8	BTU/s		
Gasifier Energy Out:				
Heat losses	67.82	BTU/s		
Slag	14899	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	4878	BTU/s		
CO2 removal	0	BTU/s		
Water condensed from syngas	23354	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
CO2 capture & AGR Qrej	45948	BTU/s		
CO2 capture & AGR heat loss	7154	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	167851	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	8393	BTU/s		
Heat rejection from compressor inter/after cooling	33170	BTU/s		
Compressors mechnical & electrical losses	1759	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-325.7	BTU/s	-0.0343	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case B4c – Supercritical, Sweet-shift CCS, 30% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUPERCRITICAL QUENCH - 30% BIOMASS - PRE-CCS (SWEET).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: IFC-67

SYSTEM SUMMARY

	Power Output kW		LHV Heat Rate BTU/kWh		Elect. Eff. LHV%	
	@ gen. term.	net	@ gen. term.	net	@ gen. term.	net
Gas Turbine(s)	200015		8255		41.34	
Steam Turbine(s)	108336					
Plant Total	308351	224130	8411	11571	40.57	29.49

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
24.58	19.67	26.67	9126

GT fuel HHV/LHV ratio =	1.179			
DB fuel HHV/LHV ratio =	1.11			
Total plant fuel HHV heat input / LHV heat input =	1.1			
Fuel HHV chemical energy input (77F/25C) =	2853468	kBTU/hr	792630	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2593391	kBTU/hr	720387	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2848129	kBTU/hr	791147	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2867303	kBTU/hr	796473	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200015	41.34	8255	946	1122
Total	200015			946	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1947098	kBTU/hr	540861	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1651041	kBTU/hr	458623	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
81.12	108336	38.45	31.18	-254738

Number of steam turbine unit(s) =	1			
Fuel chemical HHV (77F/25C) to duct burners =	185711	kBTU/hr	51586	BTU/s
Fuel chemical LHV (77F/25C) to duct burners =	167366	kBTU/hr	46490	BTU/s
DB fuel chemical LHV + HRSG inlet sens. heat =	1185434	kBTU/hr	329287	BTU/s
Water/steam to gasification plant =	195233	kBTU/hr	54231	BTU/s
Water/steam from gasification plant =	181874	kBTU/hr	50521	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-49.94	%		

Case B4d – Supercritical, Sweet-shift CCS, 50% biomass

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\SUPERCRITICAL QUENCH - 50% BIOMASS - PRE-CCS (SWEET).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: IFC-67

SYSTEM SUMMARY

	Power Output kW		LHV Heat Rate BTU/kWh		Elect. Eff. LHV%	
	@ gen. term.	net	@ gen. term.	net	@ gen. term.	net
Gas Turbine(s)	200015		8256		41.33	
Steam Turbine(s)	108651					
Plant Total	308666	220819	8353	11676	40.85	29.22

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
24.28	19.34	26.42	9115

GT fuel HHV/LHV ratio =	1.179			
DB fuel HHV/LHV ratio =	1.11			
Total plant fuel HHV heat input / LHV heat input =	1.097			
Fuel HHV chemical energy input (77F/25C) =	2829416	kBTU/hr	785949	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2578372	kBTU/hr	716214	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2833112	kBTU/hr	786976	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2852286	kBTU/hr	792302	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200015	41.33	8256	946	1122
Total	200015			946	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1947441	kBTU/hr	540956	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1651357	kBTU/hr	458710	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
81.12	108651	38.55	31.27	-254740

Number of steam turbine unit(s) =	1			
Fuel chemical HHV (77F/25C) to duct burners =	185695	kBTU/hr	51582	BTU/s
Fuel chemical LHV (77F/25C) to duct burners =	167351	kBTU/hr	46486	BTU/s
DB fuel chemical LHV + HRSG inlet sens. heat =	1185572	kBTU/hr	329326	BTU/s
Water/steam to gasification plant =	191507	kBTU/hr	53196	BTU/s
Water/steam from gasification plant =	182870	kBTU/hr	50797	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-51.07	%		

Case B4c – Supercritical, Sweet-shift CCS, 30% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	2369.1	kW
Condensate pump(s)*	227.2	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	3.612	kW
Cooling water pump(s)	1103.7	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	50	kW
Lights	90	kW
Aux. from PEACE running motor/load list	867.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	230.7	kW
Miscellaneous plant auxiliaries	308.4	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	35319	kW
Gasification plant, fuel preparation	16153	kW
Gasification plant, CO2 capture and AGR*	23058	kW
Gasification plant, other/misc	2538.5	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	82680	kW
Actual (user input) overall plant auxiliaries	82680	kW
Transformer losses	1541.8	kW
Total auxiliaries & transformer losses	**84221**	**kW**
* Heat balance related auxiliaries		

Case B4d – Supercritical, Sweet-shift CCS, 50% biomass

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	2370.6	kW
Condensate pump(s)*	227.8	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	3.628	kW
Cooling water pump(s)	1110.6	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	50	kW
Lights	90	kW
Aux. from PEACE running motor/load list	867.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	231.4	kW
Miscellaneous plant auxiliaries	308.7	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	33573	kW
Gasification plant, fuel preparation	21729	kW
Gasification plant, CO2 capture and AGR*	22893	kW
Gasification plant, other/misc	2485.9	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	86303	kW
Actual (user input) overall plant auxiliaries	86303	kW
Transformer losses	1543.3	kW
Total auxiliaries & transformer losses	**87847**	**kW**
* Heat balance related auxiliaries		

Case B4c – Supercritical, Sweet-shift CCS, 30% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**762126**	**BTU/s**		
Ambient air sensible	11280	BTU/s		
Ambient air latent	21470	BTU/s		
GT syngas & duct burner fuel	607162	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70760	BTU/s		
Makeup and process return	933.7	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	50521	BTU/s		
Energy Out	**762404**	**BTU/s**		
Net power output	212444	BTU/s		
Stack gas sensible	74182	BTU/s		
Stack gas latent	164692	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1410.3	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	168606	BTU/s		
Process steam	0.0013	BTU/s		
Process water	0	BTU/s		
Blowdown	1450.1	BTU/s		
Heat radiated from steam cycle	2532	BTU/s		
ST/generator mech/elec/gear loss	1715.1	BTU/s		
Non-heat balance related auxiliaries	20575	BTU/s		
Transformer loss	1461.4	BTU/s		
ASU compressors	33477	BTU/s		
Water/steam to gasification plant	54231	BTU/s		
CO2 capture & AGR auxiliary	21856	BTU/s		
Energy In - Energy Out	**-277.9**	**BTU/s**	-0.0365	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -277.9 BTU/s				

Case B4d – Supercritical, Sweet-shift CCS, 50% biomass

POWER BLOCK HEAT BALANCE				
Energy In	**762467**	**BTU/s**		
Ambient air sensible	11283	BTU/s		
Ambient air latent	21476	BTU/s		
GT syngas & duct burner fuel	607205	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70761	BTU/s		
Makeup and process return	945.4	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	50797	BTU/s		
Energy Out	**762746**	**BTU/s**		
Net power output	209306	BTU/s		
Stack gas sensible	74175	BTU/s		
Stack gas latent	164705	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1410.4	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	169674	BTU/s		
Process steam	0	BTU/s		
Process water	0	BTU/s		
Blowdown	1450.5	BTU/s		
Heat radiated from steam cycle	2532.5	BTU/s		
ST/generator mech/elec/gear loss	1719.5	BTU/s		
Non-heat balance related auxiliaries	25818	BTU/s		
Transformer loss	1462.9	BTU/s		
ASU compressors	31823	BTU/s		
Water/steam to gasification plant	53196	BTU/s		
CO2 capture & AGR auxiliary	21699	BTU/s		
Energy In - Energy Out	**-278.2**	**BTU/s**	-0.0365	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -278.3 BTU/s				

Case B4c – Supercritical, Sweet-shift CCS, 30% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	941376	BTU/s		
Power Block Energy In:				
Ambient air sensible	11280	BTU/s		
Ambient air latent	21470	BTU/s		
Duct burner fuel enthalpy	51775	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70760	BTU/s		
Process return & makeup	933.7	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	744185	BTU/s		
Gasifier slurry water	880.7	BTU/s		
Quench water	29983	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1376.6	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	8731	BTU/s		
Total Energy Out	941701	BTU/s		
Power Block Energy Out:				
Net power output	212444	BTU/s		
Stack gas sensible	74182	BTU/s		
Stack gas latent	164692	BTU/s		
GT cycle losses	5183	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	168606	BTU/s		
Process	0.0013	BTU/s		
Steam cycle losses	5697	BTU/s		
Non-heat balance auxiliaries	20575	BTU/s		
Transformer losses	1461.4	BTU/s		
Gasifier Energy Out:				
Heat losses	67.39	BTU/s		
Slag	14485	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	3805	BTU/s		
CO2 removal	0	BTU/s		
Water condensed from syngas	22961	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
CO2 capture & AGR Qrej	44582	BTU/s		
CO2 capture & AGR heat loss	7019	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	161883	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	7987	BTU/s		
Heat rejection from compressor inter/after cooling	31565	BTU/s		
Compressors mechanical & electrical losses	1673.9	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-325.1	BTU/s	-0.0345	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Case B4d – Supercritical, Sweet-shift CCS, 50% biomass

IGCC PLANT HEAT BALANCE				
Total Energy In	933492	BTU/s		
Power Block Energy In:				
Ambient air sensible	11283	BTU/s		
Ambient air latent	21476	BTU/s		
Duct burner fuel enthalpy	51771	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70761	BTU/s		
Process return & makeup	945.4	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	737412	BTU/s		
Gasifier slurry water	868.8	BTU/s		
Quench water	29332	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1343.4	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	8300	BTU/s		
Total Energy Out	933813	BTU/s		
Power Block Energy Out:				
Net power output	209306	BTU/s		
Stack gas sensible	74175	BTU/s		
Stack gas latent	164705	BTU/s		
GT cycle losses	5183	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	169674	BTU/s		
Process	0	BTU/s		
Steam cycle losses	5703	BTU/s		
Non-heat balance auxiliaries	25818	BTU/s		
Transformer losses	1462.9	BTU/s		
Gasifier Energy Out:				
Heat losses	66.97	BTU/s		
Slag	14083	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	2761.4	BTU/s		
CO2 removal	0	BTU/s		
Water condensed from syngas	22579	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
CO2 capture & AGR Qrej	43253	BTU/s		
CO2 capture & AGR heat loss	6888	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	156080	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	7592	BTU/s		
Heat rejection from compressor inter/after cooling	30005	BTU/s		
Compressors mechanical & electrical losses	1591.1	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-320.8	BTU/s	-0.0344	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Special Case 1: Radiant/Convective Coolers (compare Case A1b)

GT PRO 21.0 parallel						
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rk\Thermoflow\Better Plant Design\Special Cases\Special Case 1 - Radiant Cooler - subcritical - 10%						
biomass - no 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)	200018		8348		40.88	
Steam Turbine(s)	128656					
Plant Total	328674	275489	6935	8274	49.20	41.24

PLANT EFFICIENCIES

PURPA efficiency %	CHP (Total) efficiency %	Power gen. eff. on chargeable energy, %	Canadian Class 43 Heat Rate, BTU/kWh
35.57	29.90	36.76	7152

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) =	2513863	kBTU/hr	698295	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2279500	kBTU/hr	633195	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2538062	kBTU/hr	705017	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2557524	kBTU/hr	710423	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	200018	40.88	8348	1026	1102
Total	200018			1026	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1808709	kBTU/hr	502419	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1669747	kBTU/hr	463819	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff. %	Gross power output kW	Internal gross elect. eff., %	Overall elect. eff., %	Net process heat output kBTU/hr
78.90	128656	53.70	42.37	-258562

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 =	1036223	kBTU/hr	287840	BTU/s
Water/steam to gasification plant =	149763	kBTU/hr	41601	BTU/s
Water/steam from gasification plant =	615831	kBTU/hr	171064	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-37.94	%		

Special Case 1: Radiant/Convective Coolers (compare Case A1b)

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	1181.4	kW
Condensate pump(s)*	282.1	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	8.796	kW
Cooling water pump(s)	1674.1	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	45	kW
Lights	90	kW
Aux. from PEACE running motor/load list	853.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	273.9	kW
Miscellaneous plant auxiliaries	328.7	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	34652	kW
Gasification plant, fuel preparation	9730	kW
Gasification plant, AGR*	591.8	kW
Gasification plant, other/misc	1468.4	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	51541	kW
Actual (user input) overall plant auxiliaries	51541	kW
Transformer losses	1643.4	kW
Total auxiliaries & transformer losses	**53185**	**kW**
* Heat balance related auxiliaries		

Special Case 1: Radiant/Convective Coolers (compare Case A1b)

POWER BLOCK HEAT BALANCE				
Energy In	**790851**	**BTU/s**		
Ambient air sensible	11172	BTU/s		
Ambient air latent	21268	BTU/s		
GT syngas	516089	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	71823	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	171064	BTU/s		
Energy Out	**791125**	**BTU/s**		
Net power output	261126	BTU/s		
Stack gas sensible	73087	BTU/s		
Stack gas latent	114577	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1425.5	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	241189	BTU/s		
Process steam	0	BTU/s		
Process water	0	BTU/s		
Blowdown	908.7	BTU/s		
Heat radiated from steam cycle	2418.7	BTU/s		
ST/generator mech/elec/gear loss	1994.9	BTU/s		
Non-heat balance related auxiliaries	14061	BTU/s		
Transformer loss	1557.7	BTU/s		
ASU compressors	32845	BTU/s		
Water/steam to gasification plant	41601	BTU/s		
AGR auxiliary	561	BTU/s		
Energy In - Energy Out	**-273.8**	**BTU/s**	**-0.0346**	**%**
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -273.7 BTU/s				

Special Case 1: Radiant/Convective Coolers (compare Case A1b)

IGCC PLANT HEAT BALANCE				
Total Energy In	815905	BTU/s		
Power Block Energy In:				
Ambient air sensible	11172	BTU/s		
Ambient air latent	21268	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	71823	BTU/s		
Process return & makeup	0	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	701321	BTU/s		
Gasifier slurry water	834.4	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1484.6	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	8567	BTU/s		
Total Energy Out	816202	BTU/s		
Power Block Energy Out:				
Net power output	261126	BTU/s		
Stack gas sensible	73087	BTU/s		
Stack gas latent	114577	BTU/s		
GT cycle losses	5198	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	241189	BTU/s		
Process	0	BTU/s		
Steam cycle losses	5322	BTU/s		
Non-heat balance auxiliaries	14061	BTU/s		
Transformer losses	1557.7	BTU/s		
Gasifier Energy Out:				
Heat losses	331	BTU/s		
Slag	12847	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	4554	BTU/s		
CO2 removal	0	BTU/s		
Cooling after CO shift	0	BTU/s		
Water condensed from syngas	11400	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
AGR Qrej	5763	BTU/s		
AGR heat loss	535.7	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	25094	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	7837	BTU/s		
Heat rejection from compressor inter/after cooling	30969	BTU/s		
Compressors mechnical & electrical losses	1642.3	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-297.7	BTU/s	-0.0365	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Special Case 2: Dry-fed, sour-shift CCS (compare Case A3c)

GT PRO 21.0 parallel

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rk\Thermoflow\Better Plant Design\Special Cases\SPECIAL CASE 2 - DRY-FED - SUBCRITICAL - 30% BIOMASS
- PRE-CCS (SOUR).GTP
Plant Configuration: GT, HRSG, and condensing reheat ST
One SGT6-4000F Engine (Physical Model #141), One Steam Turbine, GT PRO Type 10, Subtype 2
Steam Property Formulation: Thermoflow - STQUIK

SYSTEM SUMMARY

	Power Output kW		LHV Heat Rate BTU/kWh		Elect. Eff. LHV%	
	@ gen. term.	net	@ gen. term.	net	@ gen. term.	net
Gas Turbine(s)	200017		8164		41.80	
Steam Turbine(s)	97229					
Plant Total	297246	220872	7621	10256	44.78	33.27

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
27.74	22.21	29.73	8604

GT fuel HHV/LHV ratio =	1.179			
DB fuel HHV/LHV ratio =	1.179			
Total plant fuel HHV heat input / LHV heat input =	1.1			
Fuel HHV chemical energy input (77F/25C) =	2490878	kBTU/hr	691910	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2265174	kBTU/hr	629215	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2515829	kBTU/hr	698841	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2534695	kBTU/hr	704082	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200017	41.80	8164	944	1118
Total	200017			944	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1925584	kBTU/hr	534884	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1632974	kBTU/hr	453604	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
77.55	97229	42.25	32.77	-250655

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 =	1012486	kBTU/hr	281246	BTU/s
Water/steam to gasification plant =	87636	kBTU/hr	24343	BTU/s
Water/steam from gasification plant =	223086	kBTU/hr	61968	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-49.83	%		

Special Case 3: Dry-fed, sweet-shift CCS (compare Case A4c)

GT PRO 21.0 parallel

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rk\Thermoflow\Better Plant Design\Special Cases\Special Case 3 - Dry-fed - subcritical - 30% biomass
- pre-CCS (SWEET).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		8219		41.52	
Steam Turbine(s)	76577					
Plant Total	276594	199643	8247	11426	41.37	29.86

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
24.33	18.79	26.69	9568

GT fuel HHV/LHV ratio =	1.179			
DB fuel HHV/LHV ratio =	1.179			
Total plant fuel HHV heat input / LHV heat input =	1.1			
Fuel HHV chemical energy input (77F/25C) =	2508450	kBTU/hr	696792	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2281154	kBTU/hr	633654	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2533687	kBTU/hr	703802	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2552695	kBTU/hr	709082	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200017	41.52	8219	945	1118
Total	200017			945	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1938386	kBTU/hr	538440	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1643928	kBTU/hr	456647	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
78.07	76577	33.06	25.81	-252533

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 =	1012501	kBTU/hr	281250	BTU/s
Water/steam to gasification plant =	280116	kBTU/hr	77810	BTU/s
Water/steam from gasification plant =	237066	kBTU/hr	65852	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-58.91	%		

Special Case 2: Dry-fed, sour-shift CCS (compare Case A3c)

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	848.6	kW
Condensate pump(s)*	208	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	3.487	kW
Cooling water pump(s)	1106.9	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	45	kW
Lights	75	kW
Aux. from PEACE running motor/load list	802.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	207.1	kW
Miscellaneous plant auxiliaries	297.2	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	31560	kW
Gasification plant, fuel preparation	15082	kW
Gasification plant, CO2 capture and AGR*	21981	kW
Gasification plant, other/misc	2310.2	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	74888	kW
Actual (user input) overall plant auxiliaries	74888	kW
Transformer losses	1486.2	kW
Total auxiliaries & transformer losses	**76374**	**kW**
* Heat balance related auxiliaries		

Special Case 3: Dry-fed, sweet-shift CCS (compare Case A4c)

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	1000.7	kW
Condensate pump(s)*	220.6	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	3.813	kW
Cooling water pump(s)	914.1	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	45	kW
Lights	80	kW
Aux. from PEACE running motor/load list	805.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	163.3	kW
Miscellaneous plant auxiliaries	276.6	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	32085	kW
Gasification plant, fuel preparation	15188	kW
Gasification plant, CO2 capture and AGR*	22096	kW
Gasification plant, other/misc	2326.8	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	75568	kW
Actual (user input) overall plant auxiliaries	75568	kW
Transformer losses	1383	kW
Total auxiliaries & transformer losses	**76951**	**kW**
* Heat balance related auxiliaries		

Special Case 2: Dry-fed, sour-shift CCS (compare Case A3c)

POWER BLOCK HEAT BALANCE				
Energy In	**717098**	**BTU/s**		
Ambient air sensible	11169	BTU/s		
Ambient air latent	21262	BTU/s		
GT syngas	552936	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69626	BTU/s		
Makeup and process return	137	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	61968	BTU/s		
Energy Out	**717313**	**BTU/s**		
Net power output	209356	BTU/s		
Stack gas sensible	75258	BTU/s		
Stack gas latent	158283	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1405.7	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	168898	BTU/s		
Process steam	0	BTU/s		
Process water	0	BTU/s		
Blowdown	858.1	BTU/s		
Heat radiated from steam cycle	2191	BTU/s		
ST/generator mech/elec/gear loss	1556.6	BTU/s		
Non-heat balance related auxiliaries	19233	BTU/s		
Transformer loss	1408.7	BTU/s		
ASU compressors	29914	BTU/s		
Water/steam to gasification plant	24343	BTU/s		
CO2 capture & AGR auxiliary	20835	BTU/s		
Energy In - Energy Out	**-214.8**	**BTU/s**	**-0.03**	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -214.8 BTU/s				

Special Case 3: Dry-fed, sweet-shift CCS (compare Case A4c)

POWER BLOCK HEAT BALANCE				
Energy In	**723118**	**BTU/s**		
Ambient air sensible	11156	BTU/s		
Ambient air latent	21236	BTU/s		
GT syngas	553484	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70148	BTU/s		
Makeup and process return	1242.2	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	65852	BTU/s		
Energy Out	**723360**	**BTU/s**		
Net power output	189234	BTU/s		
Stack gas sensible	73808	BTU/s		
Stack gas latent	158784	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1405.7	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	142355	BTU/s		
Process steam	0.0018	BTU/s		
Process water	0	BTU/s		
Blowdown	930.7	BTU/s		
Heat radiated from steam cycle	2148.7	BTU/s		
ST/generator mech/elec/gear loss	1329.3	BTU/s		
Non-heat balance related auxiliaries	19114	BTU/s		
Transformer loss	1310.9	BTU/s		
ASU compressors	30413	BTU/s		
Water/steam to gasification plant	77810	BTU/s		
CO2 capture & AGR auxiliary	20944	BTU/s		
Energy In - Energy Out	**-241.6**	**BTU/s**	**-0.0334**	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -241.8 BTU/s				

Special Case 2: Dry-fed, sour-shift CCS (compare Case A3c)

IGCC PLANT HEAT BALANCE				
Total Energy In	832811	BTU/s		
Power Block Energy In:				
Ambient air sensible	11169	BTU/s		
Ambient air latent	21262	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69626	BTU/s		
Process return & makeup	137	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	694843	BTU/s		
Quench water	26908	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1188.6	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	7677	BTU/s		
Total Energy Out	833067	BTU/s		
Power Block Energy Out:				
Net power output	209356	BTU/s		
Stack gas sensible	75258	BTU/s		
Stack gas latent	158283	BTU/s		
GT cycle losses	5179	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	168898	BTU/s		
Process	0	BTU/s		
Steam cycle losses	4606	BTU/s		
Non-heat balance auxiliaries	19233	BTU/s		
Transformer losses	1408.7	BTU/s		
Gasifier Energy Out:				
Heat losses	62.92	BTU/s		
Slag	7735	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	3553	BTU/s		
CO2 removal	0	BTU/s		
Water condensed from syngas	12691	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
CO2 capture & AGR Qrej	26479	BTU/s		
CO2 capture & AGR heat loss	6649	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	104256	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	6903	BTU/s		
Heat rejection from compressor inter/after cooling	27751	BTU/s		
Compressors mechnical & electrical losses	1495.7	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-255.4	BTU/s	-0.0307	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Special Case 3: Dry-fed, sweet-shift CCS (compare Case A4c)

IGCC PLANT HEAT BALANCE				
Total Energy In	839560	BTU/s		
Power Block Energy In:				
Ambient air sensible	11156	BTU/s		
Ambient air latent	21236	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	70148	BTU/s		
Process return & makeup	1242.2	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	699745	BTU/s		
Quench water	27100	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1197.1	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	7735	BTU/s		
Total Energy Out	839838	BTU/s		
Power Block Energy Out:				
Net power output	189234	BTU/s		
Stack gas sensible	73808	BTU/s		
Stack gas latent	158784	BTU/s		
GT cycle losses	5179	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	142355	BTU/s		
Process	0.0018	BTU/s		
Steam cycle losses	4409	BTU/s		
Non-heat balance auxiliaries	19114	BTU/s		
Transformer losses	1310.9	BTU/s		
Gasifier Energy Out:				
Heat losses	63.37	BTU/s		
Slag	7790	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	3578	BTU/s		
CO2 removal	0	BTU/s		
Water condensed from syngas	20764	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
CO2 capture & AGR Qrej	43456	BTU/s		
CO2 capture & AGR heat loss	6663	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	133214	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	6955	BTU/s		
Heat rejection from compressor inter/after cooling	28384	BTU/s		
Compressors mechnical & electrical losses	1520.6	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-277.9	BTU/s	-0.0331	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Special Case 4: Air-blown gasifier (compare Case A1a)

GT PRO 21.0 parallel						
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rk\Thermoflow\Better Plant Design\Special Cases\Special Case 4 - Air-blown - subcritical - 0%biomass						
- no 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)	200014		7665		44.52	
Steam Turbine(s)	100511					
Plant Total	300525	248443	7502	9074	45.49	37.60

PLANT EFFICIENCIES

PURPA efficiency %	CHP (Total) efficiency %	Power gen. eff. on chargeable energy, %	Canadian Class 43 Heat Rate, BTU/kWh
31.93	26.26	33.52	8709

GT fuel HHV/LHV ratio =	1.111			
DB fuel HHV/LHV ratio =	1.111			
Total plant fuel HHV heat input / LHV heat input =	1.107			
Fuel HHV chemical energy input (77F/25C) =	2494940	kBTU/hr	693039	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2254475	kBTU/hr	626243	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2510168	kBTU/hr	697269	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2529414	kBTU/hr	702615	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	200014	44.52	7665	1155	902
Total	200014			1155	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1703839	kBTU/hr	473288	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1533142	kBTU/hr	425873	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff. %	Gross power output kW	Internal gross elect. eff., %	Overall elect. eff., %	Net process heat output kBTU/hr
77.15	100511	38.39	29.62	-255693

Number of steam turbine unit(s) =	1			
Fuel chemical HHV (77F/25C) to duct burners =	260883	kBTU/hr	72467	BTU/s
Fuel chemical LHV (77F/25C) to duct burners =	234747	kBTU/hr	65207	BTU/s
DB fuel chemical LHV + HRSG inlet sens. heat =	1158070	kBTU/hr	321686	BTU/s
Water/steam to gasification plant =	32016	kBTU/hr	8893	BTU/s
Water/steam from gasification plant =	96703	kBTU/hr	26862	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-43.19	%		

Special Case 4: Air-blown gasifier (compare Case A1a)

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	1014.5	kW
Condensate pump(s)*	207	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	3.033	kW
Cooling water pump(s)	1167.9	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	45	kW
Lights	90	kW
Aux. from PEACE running motor/load list	789.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	214.1	kW
Miscellaneous plant auxiliaries	300.5	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	35506	kW
Gasification plant, fuel preparation	7048	kW
Gasification plant, AGR*	663.5	kW
Gasification plant, other/misc	3168	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	50579	kW
Actual (user input) overall plant auxiliaries	50579	kW
Transformer losses	1502.6	kW
Total auxiliaries & transformer losses	**52082**	**kW**
* Heat balance related auxiliaries		

Special Case 4: Air-blown gasifier (compare Case A1a)

POWER BLOCK HEAT BALANCE				
Energy In	**702123**	**BTU/s**		
Ambient air sensible	11687	BTU/s		
Ambient air latent	22247	BTU/s		
GT syngas & duct burner fuel	570768	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	71026	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	26862	BTU/s		
Energy Out	**702344**	**BTU/s**		
Net power output	235490	BTU/s		
Stack gas sensible	90619	BTU/s		
Stack gas latent	132020	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1331.7	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	176780	BTU/s		
Process steam	0	BTU/s		
Process water	0	BTU/s		
Blowdown	1177.2	BTU/s		
Heat radiated from steam cycle	2449	BTU/s		
ST/generator mech/elec/gear loss	1603.8	BTU/s		
Non-heat balance related auxiliaries	12500	BTU/s		
Transformer loss	1424.3	BTU/s		
ASU compressors	33655	BTU/s		
Water/steam to gasification plant	8893	BTU/s		
AGR auxiliary	628.9	BTU/s		
Energy In - Energy Out	**-221.6**	**BTU/s**	**-0.0316**	**%**
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -221.6 BTU/s				

Special Case 4: Air-blown gasifier (compare Case A1a)

IGCC PLANT HEAT BALANCE				
Total Energy In	829744	BTU/s		
Power Block Energy In:				
Ambient air sensible	11687	BTU/s		
Ambient air latent	22247	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	71026	BTU/s		
Process return & makeup	0	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	696133	BTU/s		
Gasifier slurry water	846.1	BTU/s		
Quench water	20772	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1292.2	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Compressor Unit Energy In:				
Ambient air - sensible & latent	6208	BTU/s		
Total Energy Out	829969	BTU/s		
Power Block Energy Out:				
Net power output	235490	BTU/s		
Stack gas sensible	90619	BTU/s		
Stack gas latent	132020	BTU/s		
GT cycle losses	5105	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	176780	BTU/s		
Process	0	BTU/s		
Steam cycle losses	5230	BTU/s		
Non-heat balance auxiliaries	12500	BTU/s		
Transformer losses	1424.3	BTU/s		
Gasifier Energy Out:				
Heat losses	62.62	BTU/s		
Slag	4921	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	5103	BTU/s		
CO2 removal	0	BTU/s		
Cooling after CO shift	0	BTU/s		
Water condensed from syngas	19454	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
AGR Qrej	6539	BTU/s		
AGR heat loss	600.2	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	106851	BTU/s		
Air Compressor Unit Energy Out:				
Discharge gas	0	BTU/s		
Heat rejection from compressor inter/after cooling	25588	BTU/s		
Compressors mechnical & electrical losses	1682.7	BTU/s		
Energy In - Energy Out	-225.4	BTU/s	-0.0272	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Special Case 5: Use syngas in duct burner (compare Case B1a)

GT PRO 21.0 parallel
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rk\Thermoflow\Better Plant Design\Special Cases\Special Case 4 - use GT fuel in DB - supercritical -
0% biomass - no 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: IFC-67

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)	200019		8299		41.12	
Steam Turbine(s)	123616					
Plant Total	323634	262453	7929	9778	43.03	34.90

PLANT EFFICIENCIES

PURPA efficiency %	CHP (Total) efficiency %	Power gen. eff. on chargeable energy, %	Canadian Class 43 Heat Rate, BTU/kWh
29.91	24.93	31.52	8079

GT fuel HHV/LHV ratio =	1.083			
DB fuel HHV/LHV ratio =	1.083			
Total plant fuel HHV heat input / LHV heat input =	1.107			
Fuel HHV chemical energy input (77F/25C) =	2839938	kBTU/hr	788872	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2566222	kBTU/hr	712840	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2822005	kBTU/hr	783890	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2841258	kBTU/hr	789238	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	200019	41.12	8299	1028	1100
Total	200019			1028	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1798046	kBTU/hr	499457	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1659994	kBTU/hr	461110	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff. %	Gross power output kW	Internal gross elect. eff., %	Overall elect. eff., %	Net process heat output kBTU/hr
81.34	123616	42.14	34.28	-255783

Number of steam turbine unit(s) =	1			
Fuel chemical HHV (77F/25C) to duct burners =	210908	kBTU/hr	58586	BTU/s
Fuel chemical LHV (77F/25C) to duct burners =	194715	kBTU/hr	54088	BTU/s
DB fuel chemical LHV + HRSG inlet sens. heat =	1230607	kBTU/hr	341835	BTU/s
Water/steam to gasification plant =	32198	kBTU/hr	8944	BTU/s
Water/steam from gasification plant =	80591	kBTU/hr	22386	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-39.98	%		

Special Case 5: Use syngas in duct burner (compare Case B1a)

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	2355.4	kW
Condensate pump(s)*	208.1	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	3.057	kW
Cooling water pump(s)	1197.4	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	50	kW
Lights	90	kW
Aux. from PEACE running motor/load list	821.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	263.2	kW
Miscellaneous plant auxiliaries	323.6	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	42280	kW
Gasification plant, fuel preparation	8022	kW
Gasification plant, AGR*	754.4	kW
Gasification plant, other/misc	2832	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	59563	kW
Actual (user input) overall plant auxiliaries	59563	kW
Transformer losses	1618.2	kW
Total auxiliaries & transformer losses	**61181**	**kW**
* Heat balance related auxiliaries		

Special Case 5: Use syngas in duct burner (compare Case B1a)

POWER BLOCK HEAT BALANCE				
Energy In	**701598**	**BTU/s**		
Ambient air sensible	11138	BTU/s		
Ambient air latent	21199	BTU/s		
GT syngas & duct burner fuel	576293	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	71051	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	22386	BTU/s		
Energy Out	**701817**	**BTU/s**		
Net power output	248769	BTU/s		
Stack gas sensible	77638	BTU/s		
Stack gas latent	118856	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1425.2	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	180826	BTU/s		
Process steam	0.0003	BTU/s		
Process water	0	BTU/s		
Blowdown	1464.6	BTU/s		
Heat radiated from steam cycle	2632.5	BTU/s		
ST/generator mech/elec/gear loss	1926.7	BTU/s		
Non-heat balance related auxiliaries	13237	BTU/s		
Transformer loss	1533.8	BTU/s		
ASU compressors	40076	BTU/s		
Water/steam to gasification plant	8944	BTU/s		
AGR auxiliary	715.1	BTU/s		
Energy In - Energy Out	**-218.5**	**BTU/s**	**-0.0311**	**%**
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -218.5 BTU/s				

Special Case 5: Use syngas in duct burner (compare Case B1a)

IGCC PLANT HEAT BALANCE				
Total Energy In	941612	BTU/s		
Power Block Energy In:				
Ambient air sensible	11138	BTU/s		
Ambient air latent	21199	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	71051	BTU/s		
Process return & makeup	0	BTU/s		
Gasifier Energy In:				
Gasifier fuel enthalpy	792394	BTU/s		
Gasifier slurry water	962.3	BTU/s		
Quench water	33345	BTU/s		
Gas Cleanup System Energy In:				
Scrubber water	1540.2	BTU/s		
Syngas moisturizer water	0	BTU/s		
Syngas moisturizer heat addition	0	BTU/s		
Air Separation Unit Energy In:				
Ambient air - sensible & latent	10452	BTU/s		
Total Energy Out	941868	BTU/s		
Power Block Energy Out:				
Net power output	248769	BTU/s		
Stack gas sensible	77638	BTU/s		
Stack gas latent	118856	BTU/s		
GT cycle losses	5198	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Condenser	180826	BTU/s		
Process	0.0003	BTU/s		
Steam cycle losses	6024	BTU/s		
Non-heat balance auxiliaries	13237	BTU/s		
Transformer losses	1533.8	BTU/s		
Gasifier Energy Out:				
Heat losses	71.28	BTU/s		
Slag	16171	BTU/s		
Gas Cleanup System Energy Out:				
H2S removal	5806	BTU/s		
CO2 removal	0	BTU/s		
Cooling after CO shift	0	BTU/s		
Water condensed from syngas	25262	BTU/s		
Syngas export	0	BTU/s		
H2 export	0	BTU/s		
AGR Qrej	7306	BTU/s		
AGR heat loss	682.9	BTU/s		
Other	0	BTU/s		
Cooler heat rejection to external sink	186220	BTU/s		
Air Separation Unit Energy Out:				
Discharge gas	9561	BTU/s		
Heat rejection from compressor inter/after cooling	37786	BTU/s		
Compressors mechnical & electrical losses	2003.8	BTU/s		
ASU heat rejection to external sink	0	BTU/s		
Energy In - Energy Out	-255.3	BTU/s	-0.0271	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				

Special Case 6: Illinois #6 coal (compare Case A1a)

GT PRO 21.0 parallel

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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		8363		40.80	
Steam Turbine(s)	86115					
Plant Total	286131	244455	7697	9010	44.33	37.87

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
32.15	26.44	33.73	8465

GT fuel HHV/LHV ratio =	1.074			
DB fuel HHV/LHV ratio =	1.074			
Total plant fuel HHV heat input / LHV heat input =	1.052			
Fuel HHV chemical energy input (77F/25C) =	2317647	kBTU/hr	643791	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2202458	kBTU/hr	611794	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2454391	kBTU/hr	681775	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2473354	kBTU/hr	687043	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200017	40.80	8363	1021	1111
Total	200017			1021	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1797088	kBTU/hr	499191	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1672696	kBTU/hr	464638	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
77.60	86115	36.48	28.31	-251933

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 =	1038042	kBTU/hr	288345	BTU/s
Water/steam to gasification plant =	98660	kBTU/hr	27406	BTU/s
Water/steam from gasification plant =	84346	kBTU/hr	23429	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-43.27	%		

Special Case 7: Illinois #6 + 10% biomass (compare Case A1b)

GT PRO 21.0 parallel

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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)	200018		8354		40.85	
Steam Turbine(s)	86291					
Plant Total	286310	242621	7773	9173	43.90	37.20

PLANT EFFICIENCIES

PURPA efficiency	CHP (Total) efficiency	Power gen. eff. on	Canadian Class 43
%	%	chargeable energy, %	Heat Rate, BTU/kWh
31.54	25.88	33.16	8453

GT fuel HHV/LHV ratio =	1.075			
DB fuel HHV/LHV ratio =	1.075			
Total plant fuel HHV heat input / LHV heat input =	1.056			
Fuel HHV chemical energy input (77F/25C) =	2349581	kBTU/hr	652661	BTU/s
Fuel LHV chemical energy input (77F/25C) =	2225575	kBTU/hr	618215	BTU/s
Total energy input (chemical LHV + ext. addn.) =	2477498	kBTU/hr	688194	BTU/s
Energy chargeable to power (93.0% LHV alt. boiler) =	2496459	kBTU/hr	693461	BTU/s

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

	Gross power	Gross LHV	Gross LHV Heat Rate	Exh. flow	Exh. temp.
	output, kW	efficiency, %	BTU/kWh	lb/s	F
per unit	200018	40.85	8354	1022	1109
Total	200018			1022	

Number of gas turbine unit(s) =	1			
Gas turbine load [%] =	100	%		
Fuel chemical HHV (77F/25C) per gas turbine =	1796190	kBTU/hr	498942	BTU/s
Fuel chemical LHV (77F/25C) per gas turbine =	1670948	kBTU/hr	464152	BTU/s

STEAM CYCLE PERFORMANCE

HRSG eff.	Gross power output	Internal gross	Overall	Net process heat output
%	kW	elect. eff., %	elect. eff., %	kBTU/hr
77.56	86291	36.58	28.37	-251923

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 =	1037886	kBTU/hr	288302	BTU/s
Water/steam to gasification plant =	93802	kBTU/hr	26056	BTU/s
Water/steam from gasification plant =	83354	kBTU/hr	23154	BTU/s
Net process heat output as % of total output (net elec. + net heat) =	-43.74	%		

Special Case 6: Illinois #6 coal (compare Case A1a)

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	927.9	kW
Condensate pump(s)*	181.9	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	2.712	kW
Cooling water pump(s)	924.2	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	45	kW
Lights	75	kW
Aux. from PEACE running motor/load list	741.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	183.5	kW
Miscellaneous plant auxiliaries	286.1	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	27862	kW
Gasification plant, fuel preparation	4589	kW
Gasification plant, AGR*	2308.7	kW
Gasification plant, other/misc	1756.3	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	40246	kW
Actual (user input) overall plant auxiliaries	40246	kW
Transformer losses	1430.7	kW
Total auxiliaries & transformer losses	**41677**	**kW**
* Heat balance related auxiliaries		

Special Case 7: Illinois #6 + 10% biomass (compare Case A1b)

ESTIMATED PLANT AUXILIARIES (kW)		
GT fuel compressor(s)*	0	kW
GT supercharging fan(s)*	0	kW
GT electric chiller(s)*	0	kW
GT chiller/heater water pump(s)	0	kW
HRSG feedpump(s)*	926.6	kW
Condensate pump(s)*	181.8	kW
HRSG forced circulation pump(s)	0	kW
LTE recirculation pump(s)	2.711	kW
Cooling water pump(s)	929	kW
Air cooled condenser fans	0	kW
Cooling tower fans	0	kW
HVAC	45	kW
Lights	75	kW
Aux. from PEACE running motor/load list	741.2	kW
Miscellaneous gas turbine auxiliaries	362	kW
Miscellaneous steam cycle auxiliaries	183.9	kW
Miscellaneous plant auxiliaries	286.3	kW
Constant plant auxiliary load	0	kW
Gasification plant, ASU*	29615	kW
Gasification plant, fuel preparation	4843	kW
Gasification plant, AGR*	2195.1	kW
Gasification plant, other/misc	1871	kW
Desalination plant auxiliaries	0	kW
Program estimated overall plant auxiliaries	42257	kW
Actual (user input) overall plant auxiliaries	42257	kW
Transformer losses	1431.5	kW
Total auxiliaries & transformer losses	**43689**	**kW**
* Heat balance related auxiliaries		

Special Case 6: Illinois #6 coal (compare Case A1a)

POWER BLOCK HEAT BALANCE				
Energy In	**638261**	**BTU/s**		
Ambient air sensible	11421	BTU/s		
Ambient air latent	21741	BTU/s		
GT syngas	512135	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69981	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	23429	BTU/s		
Energy Out	**638531**	**BTU/s**		
Net power output	231709	BTU/s		
Stack gas sensible	76896	BTU/s		
Stack gas latent	110906	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1426.8	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	143394	BTU/s		
Process steam	0	BTU/s		
Process water	0	BTU/s		
Blowdown	925.5	BTU/s		
Heat radiated from steam cycle	2217	BTU/s		
ST/generator mech/elec/gear loss	1427.5	BTU/s		
Non-heat balance related auxiliaries	8498	BTU/s		
Transformer loss	1356.1	BTU/s		
ASU compressors	26410	BTU/s		
Water/steam to gasification plant	27406	BTU/s		
AGR auxiliary	2188.3	BTU/s		
Energy In - Energy Out	**-270.1**	**BTU/s**	**-0.0423**	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -270 BTU/s				

Special Case 7: Illinois #6 + 10% biomass (compare Case A1b)

POWER BLOCK HEAT BALANCE				
Energy In	**638123**	**BTU/s**		
Ambient air sensible	11381	BTU/s		
Ambient air latent	21664	BTU/s		
GT syngas	512390	BTU/s		
External gas addition to combustor	0	BTU/s		
Steam and water	69979	BTU/s		
Makeup and process return	0	BTU/s		
N2 from ASU	0	BTU/s		
Water/steam from gasification plant	23154	BTU/s		
Energy Out	**638393**	**BTU/s**		
Net power output	229971	BTU/s		
Stack gas sensible	77015	BTU/s		
Stack gas latent	111071	BTU/s		
GT mechanical loss	1062	BTU/s		
GT gear box loss	0	BTU/s		
GT generator loss	2710.9	BTU/s		
GT miscellaneous losses	1426.7	BTU/s		
GT ancillary heat rejected	0	BTU/s		
GT process air bleed	0	BTU/s		
Air to ASU	0	BTU/s		
Air to gasifier	0	BTU/s		
Fuel compressor mech/elec loss	0	BTU/s		
Supercharging fan mech/elec loss	0	BTU/s		
Condenser	144150	BTU/s		
Process steam	0.0005	BTU/s		
Process water	0	BTU/s		
Blowdown	924.8	BTU/s		
Heat radiated from steam cycle	2216.7	BTU/s		
ST/generator mech/elec/gear loss	1429.2	BTU/s		
Non-heat balance related auxiliaries	8852	BTU/s		
Transformer loss	1356.9	BTU/s		
ASU compressors	28070	BTU/s		
Water/steam to gasification plant	26056	BTU/s		
AGR auxiliary	2080.6	BTU/s		
Energy In - Energy Out	**-270.6**	**BTU/s**	**-0.0424**	%
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)				
Gas Turbine and Steam Cycle: Energy In - Energy Out = -270.5 BTU/s				

Special Case 6: Illinois #6 coal (compare Case A1a)

IGCC PLANT HEAT BALANCE		
Total Energy In	777496	BTU/s
Power Block Energy In:		
Ambient air sensible	11421	BTU/s
Ambient air latent	21741	BTU/s
External gas addition to combustor	0	BTU/s
Steam and water	69981	BTU/s
Process return & makeup	0	BTU/s
Gasifier Energy In:		
Gasifier fuel enthalpy	645373	BTU/s
Gasifier slurry water	974.8	BTU/s
Quench water	20669	BTU/s
Gas Cleanup System Energy In:		
Scrubber water	895.6	BTU/s
Syngas moisturizer water	0	BTU/s
Syngas moisturizer heat addition	0	BTU/s
Air Separation Unit Energy In:		
Ambient air - sensible & latent	6888	BTU/s
Total Energy Out	777792	BTU/s
Power Block Energy Out:		
Net power output	231709	BTU/s
Stack gas sensible	76896	BTU/s
Stack gas latent	110906	BTU/s
GT cycle losses	5200	BTU/s
GT ancillary heat rejected	0	BTU/s
GT process air bleed	0	BTU/s
Condenser	143394	BTU/s
Process	0	BTU/s
Steam cycle losses	4570	BTU/s
Non-heat balance auxiliaries	8498	BTU/s
Transformer losses	1356.1	BTU/s
Gasifier Energy Out:		
Heat losses	61.18	BTU/s
Slag	14244	BTU/s
Gas Cleanup System Energy Out:		
H2S removal	17791	BTU/s
CO2 removal	0	BTU/s
Cooling after CO shift	0	BTU/s
Water condensed from syngas	15979	BTU/s
Syngas export	0	BTU/s
H2 export	0	BTU/s
AGR Qrej	21890	BTU/s
AGR heat loss	2092.2	BTU/s
Other	0	BTU/s
Cooler heat rejection to external sink	91403	BTU/s
Air Separation Unit Energy Out:		
Discharge gas	6301	BTU/s
Heat rejection from compressor inter/after cooling	24901	BTU/s
Compressors mechnical & electrical losses	1320.5	BTU/s
ASU heat rejection to external sink	0	BTU/s
Energy In - Energy Out	-295.9	BTU/s -0.0381 %
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)		

Special Case 7: Illinois #6 + 10% biomass (compare Case A1b)

IGCC PLANT HEAT BALANCE		
Total Energy In	788201	BTU/s
Power Block Energy In:		
Ambient air sensible	11381	BTU/s
Ambient air latent	21664	BTU/s
External gas addition to combustor	0	BTU/s
Steam and water	69979	BTU/s
Process return & makeup	0	BTU/s
Gasifier Energy In:		
Gasifier fuel enthalpy	654358	BTU/s
Gasifier slurry water	1035.9	BTU/s
Quench water	21948	BTU/s
Gas Cleanup System Energy In:		
Scrubber water	958.2	BTU/s
Syngas moisturizer water	0	BTU/s
Syngas moisturizer heat addition	0	BTU/s
Air Separation Unit Energy In:		
Ambient air - sensible & latent	7322	BTU/s
Total Energy Out	788501	BTU/s
Power Block Energy Out:		
Net power output	229971	BTU/s
Stack gas sensible	77015	BTU/s
Stack gas latent	111071	BTU/s
GT cycle losses	5200	BTU/s
GT ancillary heat rejected	0	BTU/s
GT process air bleed	0	BTU/s
Condenser	144150	BTU/s
Process	0.0005	BTU/s
Steam cycle losses	4571	BTU/s
Non-heat balance auxiliaries	8852	BTU/s
Transformer losses	1356.9	BTU/s
Gasifier Energy Out:		
Heat losses	61.82	BTU/s
Slag	14436	BTU/s
Gas Cleanup System Energy Out:		
H2S removal	16915	BTU/s
CO2 removal	0	BTU/s
Cooling after CO shift	0	BTU/s
Water condensed from syngas	16883	BTU/s
Syngas export	0	BTU/s
H2 export	0	BTU/s
AGR Qrej	20825	BTU/s
AGR heat loss	1989.2	BTU/s
Other	0	BTU/s
Cooler heat rejection to external sink	101400	BTU/s
Air Separation Unit Energy Out:		
Discharge gas	6697	BTU/s
Heat rejection from compressor inter/after cooling	26467	BTU/s
Compressors mechnical & electrical losses	1403.5	BTU/s
ASU heat rejection to external sink	0	BTU/s
Energy In - Energy Out	-300.1	BTU/s -0.0381 %
Zero enthalpy: dry gases & liquid water @ 32 F (273.15 K)		

Cost Reports: Case B2a – Supercritical, Post-combustion CCS, 0% biomass

Project Cost Summary	Reference Cost	Estimated Cost	
Power Plant:			
I Specialized Equipment	105,161,000	110,419,000	USD
II Other Equipment	5,831,000	6,123,000	USD
III Civil	15,147,000	17,346,000	USD
IV Mechanical	23,177,000	26,566,000	USD
V Electrical Assembly & Wiring	3,052,000	3,565,000	USD
VI Buildings & Structures	5,045,000	5,801,000	USD
VII Engineering & Plant Startup	11,664,000	11,688,000	USD
Gasification Plant	544,643,000	604,554,000	USD
Desalination Plant	NA	NA	
CO2 Capture Plant	348,946,000	387,330,000	USD
Subtotal - Contractor's Internal Cost	1,062,666,000	1,173,393,000	USD
VIII Contractor's Soft & Miscellaneous Costs	98,722,000	109,738,000	USD
Contractor's Price	1,161,389,000	1,283,131,000	USD
IX Owner's Soft & Miscellaneous Costs	232,278,000	256,626,000	USD
Total - Owner's Cost (1 USD per US Dollar)	1,393,666,000	1,539,758,000	USD
Nameplate Net Plant Output	206	206	MW
Cost per kW - Contractor's	5624	6214	USD per kW
Cost per kW - Owner's	6749	7457	USD per kW

Cost Reports: Case B2a – Supercritical, Post-combustion CCS, 0% biomass

	Item Cost	Unit Cost	Quantity	Ref. Cost	Est. Cost
I Specialized Equipment (USD)				105,161,000	110,419,000
1. Gas Turbine Package		45,728,000	1	45,728,000	48,015,000
Combustion Turbine Genset	44,178,000				
Inlet Filter/Silencer System (w/ elements)	included				
Evaporative Cooling System					
Inlet Fogging System					
Exhaust Stack/Silencer System					
Electrical/Control/Instrumentation Package	included				
Gas Fuel Package	included				
Liquid Fuel Package	861,600				
Fuel Heating Package					
Steam Injection Package	689,200				
Water Injection Package					
Starting Package	included				
Lube Oil Package w/ main, auxiliary & emergency pump	included				
Compressor Water Wash System	included				
High Voltage Generator					
Transportation to Site	included				
2. Steam Turbine Package		25,234,000	1	25,234,000	26,496,000
Turbine	included				
Generator	included				
Exhaust System	included				
Electrical/Control/Instrumentation Package	included				
Lube Oil Package w/ main, auxiliary & emergency pump	included				
High Voltage Generator					
Transportation to Site	included				
3. Heat Recovery Boiler		20,312,000	1	20,312,000	21,328,000
Duct Burner & Burner Management System	497,200				
Gas Turbine Exhaust Transition	included				
Bypass Stack					
Main Stack	1,132,000				
Instrumentation	included				
SCR & Aqueous Ammonia System					
CO catalytic reactor for CO reduction					
Deaerator	included				
Steam Vents & Water Drains	included				
Non-Return Valves	included				
Blowdown Recovery System					
Forced Circulation Pumps					
Transportation to Site	included				
4. Water-cooled Condenser		1,713,000	1	1,713,000	1,798,000
Vacuum Pump	elsewhere				
Steam Jet Air Ejector					
Transportation to Site	included				

8. Continuous Emissions Monitoring System		**362,450**	**1**	**362,450**	**380,550**
Enclosures	included				
Electronics, Display Units, Printers & Sensors	included				
Transportation to Site	included				
9. Distributed Control System		**987,400**	**1**	**987,400**	**1,037,000**
Enclosures	included				
Electronics, Display Units, Printers & Sensors	included				
Transportation to Site	included				
10. Transmission Voltage Equipment		**5,439,000**	**1**	**5,439,000**	**5,711,000**
Transformers	4,721,000				
Circuit Breakers	459,050				
Miscellaneous Equipment	259,000				
Transportation to Site	included				
11. Generating Voltage Equipment		**5,385,000**	**1**	**5,385,000**	**5,654,000**
Generator Buswork	3,580,000				
Circuit Breakers	1,548,000				
Current Limiting Reactors					
Miscellaneous Equipment	256,400				
Transportation to Site	included				

Cost Reports: Case B2a – Supercritical, Post-combustion CCS, 0% biomass

	Unit Cost	Quantity	Ref. Cost	Est. Cost
II Other Equipment (USD)			5,831,000	6,123,000
1. Pumps			2,352,000	2,470,000
Integral Feedwater Pump				
HP Feedwater Pump	404,250	3	1,213,000	1,273,000
IP Feedwater Pump	48,660	3	146,000	153,250
LP Feedwater Pump				
Condensate Forwarding Pump	39,850	2	79,700	83,700
Condenser C.W. Pump	285,900	2	571,800	600,400
Condenser Vacuum Pump	44,680	2	89,350	93,800
Treated Water Pump	4,840	1	4,840	5,080
Demin Water Pump				
Raw Water Pump 1	4,610	1	4,610	4,850
Raw Water Pump 2	4,610	1	4,610	4,850
Raw Water Pump 3				
GT Water Injection Pump				
GT Evap Cooler Water Pump				
Auxiliary Boiler Feedwater Pump				
Fuel Oil Unloading Pump	12,000	1	12,000	12,600
Fuel Oil Forwarding Pump	11,340	2	22,670	23,800
Aux Cooling Water Pump (closed loop)	15,590	2	31,180	32,730
Diesel Fire Pump	68,650	2	137,300	144,200
Electric Fire Pump				
Jockey Fire Pump	4,120	1	4,120	4,320
GT Inlet Air Chiller/Heater Water Pump				
GT+Generator Lube Oil Coolant Pump				
GT Generator Lube Oil Coolant Pump				
GT Generator Cooling Pump				
GT Chiller Coolant Pump				
Fuel Compressor Coolant Pump				
ST+Generator Lube Oil Coolant Pump				
ST Generator Cooling Pump				
Aux Cooling Water Pump (open loop)	15,590	2	31,180	32,730
2. Tanks		8	1,247,000	1,309,000
Fuel Oil	446,750	2	893,500	938,200
Hydrous Ammonia				
Demineralized Water	54,650	1	54,650	57,350
Raw Water	54,650	1	54,650	57,350
Neutralized Water	41,150	1	41,150	43,200
Acid Storage	17,040	1	17,040	17,890
Caustic Storage	17,040	1	17,040	17,890
Waste Water				
Dedicated Fire Protection Water Storage	168,700	1	168,700	177,150
3. Cooling Tower				
4. Auxiliary Heat Exchangers			86,350	90,650
Auxiliary Cooling Water Heat Exchanger	86,350	1	86,350	90,650

Cost Reports: Case B2a – Supercritical, Post-combustion CCS, 0% biomass

	Material	Labor Hours	Labor Rate	Unit Cost	Quantity	Ref. Cost	Est. Cost
III Civil (USD)						15,147,000	17,346,000
1. Site Work	1,556,000	25,400	35			2,445,000	2,745,000
Site Clearing	included	included					
Demolition	included	included					
Culverts & Drainage	included	included					
Erosion Control	included	included					
Fencing, Controlled Access Gates	included	included					
Finish Grading	included	included					
Finish Landscaping	included	included					
Material (Dirt, Sand, Stone)	included	included					
Waste Material Removal	included	included					
Obstacles R&R	included	included					
Miscellaneous	included	included					
2. Excavation & Backfill	825,900	17,890	35	55.44	26,190 CY	1,452,000	1,650,000
Gas Turbine (1)	54,600	1,230	35	49.78	1,960 CY	97,650	111,150
Steam Turbine (1)	23,600	533	35	50.45	838 CY	42,260	48,110
Heat Recovery Boiler (1)	133,000	2,970	35	48.29	4,910 CY	237,100	269,800
Water Cooled Condenser (1)	37,660	838	35	47.31	1,420 CY	67,000	76,200
Cooling Tower							
Air Cooled Condenser							
Underground Piping	296,350	6,540	35	45.74	11,480 CY	525,200	597,200
Switchyard	6,340	141	35	142	79.5 CY	11,280	12,830
Other & Miscellaneous	274,250	5,640	35	85.65	5,510 CY	471,500	534,500
3. Concrete	5,263,000	161,700	35	1,250	8,740 CY	10,922,000	12,601,000
Gas Turbine (1)	703,800	19,880	35	1,360	1,030 CY	1,400,000	1,609,000
Turbine + Generator Package					1,010 CY		
Inlet Filter					22.17 CY		
Steam Turbine (1)	435,550	13,560	35	1,880	485 CY	910,100	1,051,000
Laydown pads:	41,090	1,220	35		41.71 CY		
Gas Turbine	24,010	709	35	1,790	27.21 CY	48,810	56,200
Steam Turbine	17,090	512	35	2,410	14.5 CY	34,990	40,320
Heat Recovery Boiler (1)	951,300	28,980	35	1,080	1,820 CY	1,966,000	2,267,000
Water Cooled Condenser (1)	224,550	7,420	35	1,230	394 CY	484,400	560,600
Cooling Tower							
Air Cooled Condenser							
Underground Piping:	55,650	1,940	35		85.12 CY		
Circulating Water	55,650	1,940	35	1,450	85.12 CY	123,500	143,250
Miscellaneous							
Makeup Water Treatment System							
Auxiliary Boiler (0)							
Electrical Power Equipment	571,900	18,000	35	1,390	864 CY	1,202,000	1,388,000
Inlet Chilling System (0)							
Fuel Gas Compressor (0)							
Pumps (10)	133,850	4,070	35	3,400	81.27 CY	276,200	318,500
Auxiliary Heat Exchangers							
Feedwater Heater(s) (0)							
Station/Instrument Air Compressors (2)	26,510	772	35	3,940	13.6 CY	53,550	61,600
Bridge Crane(s)							elsewhere
Recip Engine Genset(s) (0)							
Tanks:	1,469,000	50,100	35		3,070 CY		
Fuel Oil	1,350,000	46,110	35	1,030	2,890 CY	2,964,000	3,436,000
Hydrous Ammonia							
Demineralized Water	15,710	527	35	2,310	14.77 CY	34,150	39,550

	Material	Labor Hours	Labor Rate	Unit Cost	Quantity	Ref. Cost	Est. Cost
Raw Water	15,710	527	35	2,310	14.77 CY	34,150	39,550
Neutralized Water	12,060	411	35	3,020	8.763 CY	26,440	30,640
Acid Storage	5,850	211	35	5,690	2.326 CY	13,240	15,380
Caustic Storage	5,850	211	35	5,690	2.326 CY	13,240	15,380
Waste Water							
Dedicated Fire Protection Water Storage	63,200	2,100	35	990	138 CY	136,850	158,400
Switchyard	35,030	1,050	35	1,160	61.8 CY	71,800	82,750
Miscellaneous	614,700	14,700	35	1,420	795 CY	1,129,000	1,289,000
4. Roads, Parking, Walkways	**291,800**	**959**	**37.23**			**327,500**	**351,000**
Pavement, Curbing, Striping	217,100	603	35	4.26	55,950 ft^2	238,200	254,350
Lighting	74,700	356	41	5,580	16	89,300	96,650
5. User-defined						**0**	**0**

Cost Reports: Case B2a – Supercritical, Post-combustion CCS, 0% biomass

	Material	Labor Hours	Labor Rate	Unit Cost	Quantity	Ref. Cost	Est. Cost
IV Mechanical (USD)						**23,177,000**	**26,566,000**
1. On-Site Transportation & Rigging	**1,800,000**					**1,800,000**	**2,070,000**
2. Equipment Erection & Assembly	**1,036,000**	**136,750**	**40**			**6,505,000**	**7,924,000**
Gas Turbine Package	164,550	23,710	40	1,113,000	1	1,113,000	1,359,000
Steam Turbine Package	112,400	16,200	40	760,400	1	760,400	928,000
HRSG	309,450	44,590	40	2,093,000	1	2,093,000	2,555,000
Condenser	31,080	4,480	40	210,200	1	210,200	256,550
Cooling Tower							
Makeup Water Treatment System							
Auxiliary Boiler							
Electrical Power Equipment	116,900	16,850	40			790,700	965,000
Inlet Chilling System							
Fuel Gas Compressor							
Pumps	14,530	2,090	40			98,300	120,000
Tanks + Auxiliary Heat Exchangers	127,650	18,390	40			863,500	1,054,000
Feedwater Heater(s)							
Station/Instrument Air Compressors	1,640	237	40			11,110	13,550
Bridge Crane(s)	4,390	632	40			29,680	36,220
Recip Engine Genset(s)							
Miscellaneous	153,150	9,540	40			534,700	637,700
3. Piping	**9,514,000**	**112,700**	**40**	**528**	**26,560**	**14,022,000**	**15,624,000**
High Pressure Steam	2,873,000	3,950	40	8,020	378 ft	3,031,000	3,214,000
Cold Reheat Steam	150,650	3,440	40	680	424 ft	288,300	330,250
Hot Reheat Steam	636,200	9,260	40	2,470	407 ft	1,007,000	1,131,000
Intermediate Pressure Steam	435,950	11,210	40	634	1,390 ft	884,200	1,018,000
Low Pressure Steam							
Other Steam	444,700	10,820	40	542	1,620 ft	877,600	1,008,000
Circulating Water	1,121,000	8,410	40	729	2,000 ft	1,457,000	1,598,000
Auxiliary Cooling Water	307,000	8,760	40	249	2,640 ft	657,500	760,500
Feedwater	213,550	5,520	40	382	1,140 ft	434,300	500,200
Other Water							
GT Inlet Chilling/Heating System							
Raw Water	49,290	2,180	40	88.81	1,540 ft	136,700	161,000
Service Water	142,500	5,630	40	114	3,230 ft	367,600	431,000
Waste Water							elsewhere
Steam/Water Sampling							elsewhere
Sanitary Water							elsewhere
Vents							
Fuel Gas	732,500	10,600	40	717	1,610 ft	1,156,000	1,299,000
Fuel Oil	960,800	13,260	40	987	1,510 ft	1,491,000	1,672,000
Lube Oil	197,100	3,810	40	1,070	328 ft	349,550	397,500
Compressed Air							
GT Air Bleed							
Service Air	23,930	2,370	40	48.97	2,430 ft	118,800	143,700
Vacuum Air	29,930	841	40	388	164 ft	63,550	73,500
Trim							elsewhere
Chemical Feed							
Nitrogen							
Oxygen							
Carbon Dioxide							
Ammonia							

	Material	Labor Hours	Labor Rate	Unit Cost	Quantity	Ref. Cost	Est. Cost
Caustic							elsewhere
Acid							elsewhere
Boiler & Equipment Drain	31,890	214	40	99.91	405 ft	40,470	44,200
Boiler Blowdown	36,990	929	40	183	405 ft	74,150	85,300
Air Blowoff							
Steam Blowoff	220,350	1,980	40	1,180	254 ft	299,400	330,200
Chemical Cleaning							
Heat Tracing							
Fire Protection	242,150	1,660	40	109	2,830 ft	308,450	337,150
Miscellaneous	663,800	7,860	40	528	1,850 ft	978,300	1,090,000
4. Steel	**574,900**	**6,890**	**40**	**4,810**	**177 ton**	**850,600**	**948,300**
Racks, Supports, Ladders, Walkways, Platforms	574,900	6,890	40	4,810	177 ton	850,600	948,300
5. User-defined						**0**	**0**

	Material	Labor Hours	Labor Rate	Unit Cost	Quantity	Ref. Cost	Est. Cost
V Electrical (USD)						**3,052,000**	**3,565,000**
1. Controls	**101,650**	**11,040**	**41**			**554,500**	**672,800**
Gas Turbine Package	23,430	2,630	41	131,450	1	131,450	159,650
Steam Turbine Package	16,000	1,800	41	89,800	1	89,800	109,050
HRSG	44,060	4,950	41	247,200	1	247,200	300,200
Condenser	4,420	498	41	24,830	1	24,830	30,150
Cooling Tower							
Makeup Water Treatment System							
Auxiliary Boiler							
Electrical Power Equipment							
Inlet Chilling System							
Fuel Gas Compressor							
Pumps	7,980	898	41			44,780	54,400
Auxiliary Heat Exchangers							
Feedwater Heater(s)							
Station/Instrument Air Compressor	2,250	101	41			6,410	7,560
Bridge Crane(s)	3,510	158	41			10,000	11,790
Recip Engine Genset(s)							
2. Assembly & Wiring	**1,148,000**	**32,920**	**41**			**2,497,000**	**2,892,000**
Switchgear	5,470	470	41	24,740	1	24,740	29,830
Motor Control Centers	9,020	1,110	41	2,600	21	54,500	66,350
Feeders	308,300	11,970	41	11,260	71	799,200	937,300
Medium/Low Voltage Cable Bus	240,900	4,710	41	17,360	25	433,900	494,200
Cable Tray	125,850	2,640	41	234,250	1	234,250	267,650
General Plant Instrumentation	123,750	1,610	41	836	227	189,700	212,350
Generator to Step-up Transformer Bus	62,500	1,170	41	55,250	2	110,500	125,650
Transformers	95,750	5,380	41	63,300	5	316,450	376,400
Circuit Breakers	58,200	2,180	41	24,610	6	147,700	172,950
Miscellaneous	117,950	1,670	41	186,500	1	186,500	209,550
3. User-defined						**0**	**0**

Cost Reports: Case B2a – Supercritical, Post-combustion CCS, 0% biomass

	Area	Cost/Unit Area	Ref. Cost	Est. Cost
VI Buildings (USD)			5,045,000	5,801,000
1. Turbine Hall	22044 ft^2	164.43	3,625,000	4,168,000
2. Administration, Control Room, Machine Shop / Warehouse	13100 ft^2	106.73	1,398,000	1,608,000
3. Water Treatment System				
4. Guard House	200 ft^2	108.28	21,660	24,910
5. User-defined			0	0

	Material	Labor Hours	Labor Rate	Unit Cost	Quantity	Ref. Cost	Est. Cost
VII Engineering & Startup (USD)						11,664,000	11,688,000
1. Engineering						9,972,000	9,972,000
2. Start-Up	473,600	13,720	88.85	1,692,000		1,692,000	1,716,000
3. User-defined						0	0

	Ref. Cost	Est. Cost
VIII Soft & Miscellaneous Costs (USD)		
1. Contractor's Soft Costs	98,722,000	109,738,000
Contingency:		
Labor	5,121,000	6,401,000
Specialized Equipment	3,155,000	3,313,000
Other Equipment	233,250	244,900
Commodity	1,355,000	1,423,000
Profit:		
Labor	5,121,000	6,401,000
Specialized Equipment	7,361,000	7,729,000
Other Equipment	408,150	428,600
Commodity	1,581,000	1,660,000
Permits, Licenses, Fees, Miscellaneous	0	0
Bonds and Insurance	21,253,000	23,468,000
Spare Parts & Materials	0	0
Contractor's Fee	53,133,000	58,670,000
2. Owner's Soft Costs	232,278,000	256,626,000
Permits, Licenses, Fees, Miscellaneous	23,228,000	25,663,000
Land Cost	0	0
Utility Connection Cost	0	0
Legal & Financial Costs	23,228,000	25,663,000
Escalation and Interest During Construction	174,208,000	192,470,000
Spare Parts & Materials	0	0
Project Administration & Developer's Fee	11,614,000	12,831,000
3. Total of all user-defined costs displayed on each account	0	0

Cost Reports: Case B2a – Supercritical, Post-combustion CCS, 0% biomass

	Item Cost	Unit Cost	Quantity	Ref. Cost	Est. Cost
Gasification Plant (USD)				544,643,000	604,554,000
1. Gasification				222,164,000	246,602,000
Gasifier	122,594,000		1		
Radiant Syngas Cooler					
Convective Syngas Cooler					
Slag & Process Water Handling	included				
Feedstock Preparation	99,570,000		1		
Receiving & storage	included				
Grinding	included				
Slurry preparation	included				
Transportation to Site	included				
2. Air Separation Unit		100,605,000	1	100,605,000	111,671,000
3. Gas Cleanup System		120,433,000	1	120,433,000	133,680,000
Syngas Scrubbers	included				
COS Hydrolysis System	included				
Low Temperature Gas Cooling	included				
Acid Gas Removal	included				
Sulphur Recovery	included				
Clean Syngas Moisturization					
Clean Syngas Preheater					
Syngas energy recovery turbine					
Transportation to Site	included				
4. Gasification Plant Water Systems		27,438,000	1	27,438,000	30,456,000
5. Gasification Plant General Facilities		74,004,000	1	74,004,000	82,144,000
In-Plant Electrical Distribution	included				
Switchyard	included				
Buildings	included				
Miscellaneous Common Utilities	included				
6. User-defined				0	0

	Item Cost	Unit Cost	Quantity	Ref. Cost	Est. Cost
CO2 Capture Plant (USD)				348,946,000	387,330,000
1. CO2 Capture (Chemical Type)		348,946,000	1	348,946,000	387,330,000

VITA

Henry Allen Long, III was born in New Orleans, Louisiana on October 29th, 1986. He has lived in Louisiana all of his life everywhere from lower Saint Bernard Parish to West Monroe. He completed his primary education at Brother Martin High School in 2005, and received a Bachelor's Degree in Mechanical Engineering from the University of New Orleans (UNO) in 2009. He has since continued his education at UNO in pursuit of a Master of Science in Mechanical Engineering degree. During this time, he has performed most of his research at the Energy Conversion and Conservation Center under the supervision of Dr. Ting Wang, under whom he would publish three academic papers. He received his Master's Degree in December, 2011.