Title

Development and Application of Advanced Models for Steam Hydrogasification: Process Design and Economic Evaluation

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Publication Date

2012

UNIVERSITY OF CALIFORNIA
RIVERSIDE

Development and Application of Advanced Models for Steam Hydrogasification: Process Design and Economic Evaluation

A Dissertation submitted in partial satisfaction
of the requirements for the degree of

Doctor of Philosophy

in

Chemical and Environmental Engineering

by

Xiaoming Lu

December 2012

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ACKNOWLEDGEMENTS

This thesis would not have been possible without the support of many people. First and foremost, I would like to thank my adviser, Dr. Joseph M. Norbeck, for his intellectual supervision, continuous support and encouragement throughout my graduate studies. Working with him exposed me to many invaluable experiences that I will deeply cherish for the rest of my life. Special thanks are given to Dr. Chan S. Park, my co-advisor, for the guidance he provided. I benefited enormously from his comments and suggestions.

I would like to thank my reading committee members, Dr. Wayne Miller and Dr. Victor Rodgers and my advance candidacy members, Dr. David Cocker and Dr. Akua Asa-Awuku. I am very thankful for my seniors, Dr. Arun Raju and Dr. Wei He for helping me getting started. I also would like to thank my group mates Qian Luo, Xin Fan, Yang Li, Zhongzhe Liu, Yoothana Thanmongkhons, Amornrat Suemanotham, and Sangran Hu for their constructive suggestions and assistance. I also wish to thanks Junior Castillo for offering great help on providing experiments support. I would also like to express my thanks to all of my friends and colleagues in the Department of Chemical and Environmental Engineering for their encouragement and invaluable scientific discussions.

Finally, special thanks go to my parents and my girlfriend Yan Wang. I am deeply and forever indebted for their everlasting love, support and encouragement, for giving of themselves beyond the call of duty.

Riverside, USA, December 2012.

Xiaoming Lu

ABSTRACT OF THE DISSERTATION

Development and Application of Advanced Models for Steam Hydrogasification: Process Design and Economic Evaluation

by

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Doctor of Philosophy, Graduate Program in Chemical and Environmental Engineering
University of California, Riverside, December 2012
Dr. Joseph M. Norbeck, Chairperson

This thesis is aimed at the process development, design, modeling and optimization of synthetic fuels, power and Substitute Natural Gas (SNG) production from coal and biomass with economic analysis based on steam hydrogasification technology. The steam hydrogasification is a thermochemical process to convert carbonaceous materials into methane-rich gas in the Steam Hydrogasification Reactor (SHR) at high hydrogen and steam environment. The methane could be either converted into syngas in the Steam Methane Reformer (SMR) for synthetic fuels production or used as SNG after going through the Water Gas Shift (WGS) reactor.

Optimum operation conditions for the SHR are determined based on mass and heat balance analysis derived from the Aspen Plus simulation results as well as data collected from experiments. Facilities utilizing bituminous coal and biomass green waste for coproduction of synthetic fuels and electricity are designed in detail. Cases with design capacity of 4,000 TPD (coal, dry basis) and 2,000 TPD (green waste, dry basis) are investigated with process modeling and cost estimation. The plant performance and capital cost is used as major inputs in the power financial model for process economics evaluation. The analysis shows that the coal plant with 90% Carbon Capture and Storage (CCS) using a cobalt catalyst in Fischer Tropsch synthesis is expected to produce 8,548 barrels fuels per day with production cost of 2.07 \$/gal diesel equivalent at 12% Internal Rate of Return (IRR) and 54 \$/MWh electricity sale price. The biomass plant is expected to produce 2,430 barrels fuels per day with production cost ranging from 1.55 \$/gallon to 3.65 \$/gallon diesel equivalent. The green waste and biosolid-to-SNG plant is expected to produce 19,848 MMBTU SNG per day with production cost ranging from 2.53 \$/MMBTU to 15.23 \$/MMBTU.

The process simulation and economic analysis presented here demonstrate that the steam hydrogasification technology could potentially provide an effective pathway to convert coal and biomass to fuels with high conversion efficiency and less capital cost. The steam hydrogasification process appears to be suitable for commercialization in large scales with a coal feedstock and also in a distributed network of small scale facility utilizing localized renewable feedstocks.

Financial incentives such as tax incentives, waste tipping fee, and other mechanisms are significant parameters in addressing the economic and market challenges of biomass derived fuels. Prospective commercial economics benefits with increasing plant size and improvements from large-scale demonstration efforts on steam hydrogasification.

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1. Introduction

1.1 Background

Energy sustainability and global warming are always two intertwined issues of significant magnitude in the modern age. With oil prices rising from 12 \$/barrel to 120 \$/barrel during the period from 1945 to 2008 and atmospheric CO_2 levels increasing at a rate greater than 1.5 ppm each year [1, 2], it is time to develop cost effective and clean conversion processes for energy production.

Although there has been an increased emphasis on the transition to renewable alternatives over the last few years, the world energy consumption is dominated by fossil fuels as show in Figure 1 [3]. It can been see that global primary energy consumption rose 5.6% in 2010, the fastest annual increase since 1973 and easily taking 2010 past the previous peak year of 2008 and more than 87% of the energy consumption is derived from fossil fuels. A large portion of crude oil that represents 33.6% of global energy supply is used for transportation purpose.

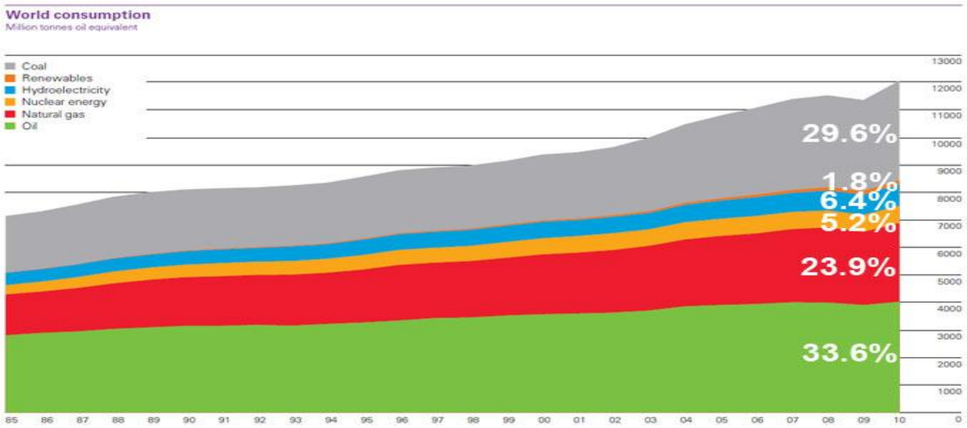
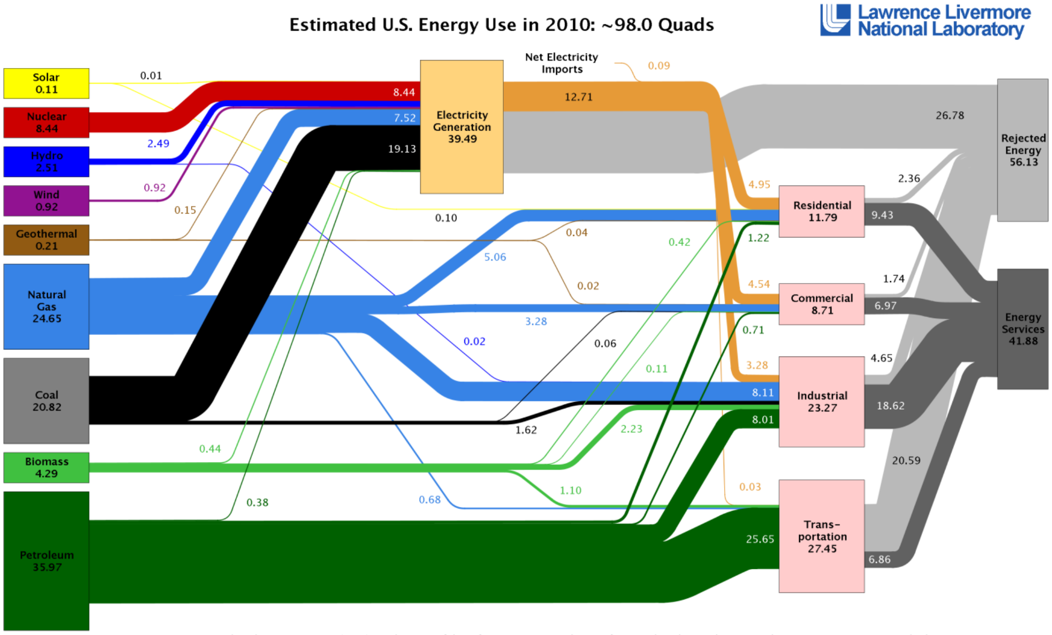


Figure 1.1 World energy consumption in 2010

Estimated U.S. energy use in 2010 was 98.0 quads with more than 83% of the energy consumption comes from fossil fuel as can been seen in Figure 1.2 [4]. Transportation accounts for the largest fraction and is about 39% of the total energy use. Furthermore, the transportation section (cars, ships, airplanes) relies heavily on crude oil and more than 93% of the transportation fuels are derived from it.


Estimated U.S. Energy Use in 2010: ~98.0 Quads

Source: LLNL 2011. Data is based on DOE/EIA-0384(2010), October 2011. If this information or a reproduction of it is used, credit must be given to the Lawrence Livermore National Laboratory and the Department of Energy, under whose auspices the work was performed. Distributed electricity represents only retail electricity sales and does not include self-generation. EIA reports flows for hydro, wind, solar and geothermal in BTU-equivalent values by assuming a typical fossil fuel plant "heat rate." (see EIA report for explanation of change to geothermal in 2010). The efficiency of electricity production is calculated as the total retail electricity delivered divided by the primary energy input into electricity generation. End use efficiency is estimated as 80% for the residential, commercial and industrial sectors, and as 25% for the transportation sector. Totals may not equal sum of components due to independent rounding. LLNL-MI-410527

Figure 1.2 U.S. energy consumption by sector in 2010

It can be expected that the fossil fuels (coal, petroleum and natural gas) will dominate energy consumption for decades even though the usage of renewable alternative fuels is increasing largely worldwide over the last few years. The amount of fossil fuel reserves, especially for crude oil, is a subject of ongoing debate [5, 6, 7]. However, even the most optimistic estimates on fossil fuel reserves and environmental challenges call for urgent actions towards a shift to alternative fuels [8]. This is especially true for alternative transportation fuels that are widely accepted as the solution to avoid an energy crisis and irrevocable impairment to the climate.

U.S. currently has almost one-third of the world’s automobiles (~230 million) and uses one quarter of the world’s crude oil [9]. The nation’s economy highly depends on liquid transportation fuels that are mainly derived from petroleum, to power cars, buses, trucks, locomotives, barges and airplanes. Use of petroleum has given rise to energy security concerns, contributions to climate change and other environmental challenges. In the absence of alternatives to petroleum products, the Energy Information Administration (EIA) projects that reliance on foreign producers for crude oil will increase 30% through 2030, and the transport sector’s GHG emissions will grow by nearly 40% [10]. Action is needed now to ensure that viable alternatives are developed to replace petroleum in conjunction with efficiency improvements to address these growing concerns.

Due to the magnitude of the demand and also the specific requirements of vehicle, transportation fuels have been posed a unique challenge and research efforts over the past have provided several options that have the potential to bring forth major changes. Alternatives such as Compressed Natural Gas (CNG), Liquefied Petroleum Gas (LPG), ethanol, biodiesel, hydrogen, Dimethyl Ether (DME) and Fischer-Tropsch (FT) liquids have been in use in several countries on different scales and have been reasonably effective in some specific cases [8, 11, 12]. The usage of alternative transportation fuels can provide the smooth transition and long term security of supply when crude oil shortfall occurs. Moreover, the alternative fuels are expected to produce cleaner exhaust gases that will improve the air quality and reduce the life cycle Green House Gas (GHG) emissions compared to petroleum based fuel if renewable feedstock is used or CO_2 sequestration technology is applied.

In U.S., a concerted effort is being made at a state and national level to increase the use of alternative fuels in transportation areas. The legislature in California passed AB1007 [13] that requires the California Air Resources Board (CARB) and the California Energy Commission (CEC) to develop a plan to increase alternative fuel use in California to reduce crude oil dependency and air pollution. The State Alternative Fuels plan, developed under Assembly Bill 1007, presents strategies and actions needed to increase the use of alternative non-petroleum fuels. This includes milestone goals for the years 2012, 2017, and 2022, with a longer term plan for a multi-fuel transportation fuel infrastructure in California by 2050. Other state initiatives include Assembly Bill AB 118 [14] to create an alternative fuel and vehicle technology program, AB32 [15], the Global Warming Solutions Act, and the Low Carbon Fuel Standard (LCFS) under CARB to reduce the carbon intensity of fuels.

The U.S. national goals for renewable transportation fuel are [16]:

- Reduce gasoline use by 20% in 10 years (2017).
- Replace 30% of current (2004) gasoline use with biofuels by 2030.
- Produce 7.5 billion gallons of renewable fuels by 2012.
- Produce 36 billion gallons of renewable fuels by 2022, with 21 billion gallons of this limit being produced by advanced technologies.

The focus of this thesis is mainly on synthetic hydrocarbon fuels, power and SNG production using renewable feedstock as well as conventional fossil fuel from a steam gasification process.

1.2 Gasification and its application

1.2.1 Gasification reactions

Gasification can be defined as the conversion of carbonaceous solid or heavy liquid feedstocks into product gases with a useable heating value [17]. This definition excludes combustion since the product flue gas has zero residual heating value. Commercial gasification processes date back to the late 18th century when coal was converted into town gas for lighting and cooking. Gasification has been used to produce chemicals and fuels from coal and biomass since 1920s [17]. Recently, interest in the research and commercial development of gasification has increased significantly due to the high crude oil price and also owing to considerable technology breakthroughs achieved over the past few decades. Based on the reactants used in the gasifier, gasification can be categorized as partial oxidation gasification, steam gasification (pyrolysis), hydrogasification and steam hydrogasification. The dominant technology is Partial Oxidation (POX) that produces synthesis gas (a mixture of H_2 and CO) in varying ratios of oxidant (oxygen, air and /or steam). The primary feedstock for gasification is coal, but in principle, any carbonaceous matter can be used including different types of biomass, Municipal Solid Wastes (MSW), sewage sludge, animal and food wastes, etc.

Usually, gasification occurs in three stages after initial feedstock drying process that includes feedstock devolatilization, volatile combustion and char gasification. The first step is the decomposition of the feedstock that results in the production of char from carbon along with gases from the volatile fraction of feedstock. This devolatilization takes place already at low temperatures (350–800 ℃) and in parallel with the heating up of the feedstock particles. The second step is the combustion of the volatile species generated in the previous stage with the surrounding oxidant. At last, the carbon in char form reacts with gasifying agent and gaseous products are produced as a result [17].

Devolatilization (pyrolysis, thermal decomposition):

Feedstock + Heat (350-800 ℃) → Coke (char) + Liquids (tar) + Gases

Combustion:

Carbon + Volatile + Oxidant → CO_2 + H_2O

Gasification:

Char + Gasifying agent (steam) + Heat (700-1400 ℃) → Gases (H_2, CO…) + Minerals (ash)

During the last two stages, the principle chemical reactions are those involving solid carbon, carbon monoxide, carbon dioxide, hydrogen, water (or steam) and methane.

Combustion reaction

$C + ½ O_2 = CO$ -111 MJ/kmol (1.1)

$CO + ½ O_2 = CO_2$ -283 MJ/kmol (1.2)

$H_2 + ½ O_2 = H_2O$ -242 MJ/kmol (1.3)

Boudouard reaction

$C + CO_2 \leftrightarrow 2CO$ +172 MJ/kmol (1.4)

Water gas reaction

$C + H_2O \leftrightarrow CO + H_2$ +131 MJ/kmol (1.5)

Methanation reaction

$C + 2H_2 \leftrightarrow CH_4$ -75 MJ/kmol (1.6)

Water gas shift reaction

$CO + H_2O \leftrightarrow CO_2 + H_2$ -41 MJ/kmol (1.7)

Steam methane reforming reaction

$CH_4 + H_2O \leftrightarrow CO + 3H_2$ +206 MJ/kmol (1.8)

Equations 1.1-1.3 show the combustion reactions traditionally employed for generating the required process heat by supplying oxygen or air into the gasifier. In the absence of the oxidants, these highly exothermic reactions are insignificant and hence the overall gasification process is endothermic and heat from external source is necessary to proceed under most conditions. Equation 1.5 shows the steam gasification reactions and Equation 1.6 is the hydrogasification reaction which essentially accounts for methane production. In general, the carbon conversion is also essentially complete under certain situation from the two homogeneous gas reactions given by Equation 1.7 and Equation 1.8.

In a word, Equation 1.1, 1.4, 1.5 and 1.6 describe the four ways in which carbonaceous materials are gasified. Equation 1.4 is very important for the production of high concentration of CO when gasifying feedstock with an oxygen/CO_2 mixture. Equation 1.5 plays a predominant role in the gasification process under high steam environment. Equation 1.6 is the basis of all hydrogenating gasification process. However, most gasification processes rely on a balance between Equation 1.1 (partial oxidation) and Equation 1.5 (water gas reaction).

Depending on the specific technology, gasification processes operate over a wide range of temperatures from 750 ℃ to 1800 ℃. Gasification temperatures are normally so high with oxidant (> 1200 ^{o}C) and almost there is no hydrocarbons other than methane can be present in significant quantities thermodynamically as well as in practice.

This may not be true for air-blown gasification, especially with biomass as feedstock. Gasification processes usually operate under high pressures, normally in the range of 10 to 100 bar. Advantages of high pressure processes include energy savings during the compression of syngas in the upstream and also an overall reduction in the equipment size [17]. There is a lot of literature available on the thermodynamics and kinetics study of gasification and research work is still being carried out [17-20]. In general, kinetic studies of gasification can be performed by simply studying the kinetics of carbon or char instead of the hydrocarbon feedstock [21, 22].

Currently, gasification processes using air or oxygen as the oxidant are attracting significant interest and large scale oxygen blown gasifiers are the primary choice for commercial syngas production from coal and other hydrocarbon feeds [17, 19].

1.2.2 Gasification reactors

A large number of coal gasification processes exist that are already commercialized or are ready to be commercialized, ranging from lower temperature and pressure, and larger grain size to larger body size and higher flow rate types.

Oxygen or air-blown processes are the primary focuses of current gasification development which are commonly known as POX technologies. Gasifiers can generally be grouped into three classes depending on their flow geometry which are fixed or moving bed gasifiers, entrained-flow gasifiers and fluidized-bed gasifiers. This section discusses some basic strengths and weaknesses and design layouts of typical commercial gasifiers.

Fixed or moving bed gasifiers

Fixed bed or moving bed gasifiers are the earliest gasification processes that still in commercial use. The preferred feedstocks are in general coke or anthracite and the outlet gas needs extensive cleaning to remove tars from the stream. Both fixed bed and moving bed operate at atmospheric pressure. During gasification process, large coal particles (6-50 mm) move slowly downward through a bed while reacting with gases moving in the opposite direction going upward through the bed as can be seen in Figure 1.3 [17].

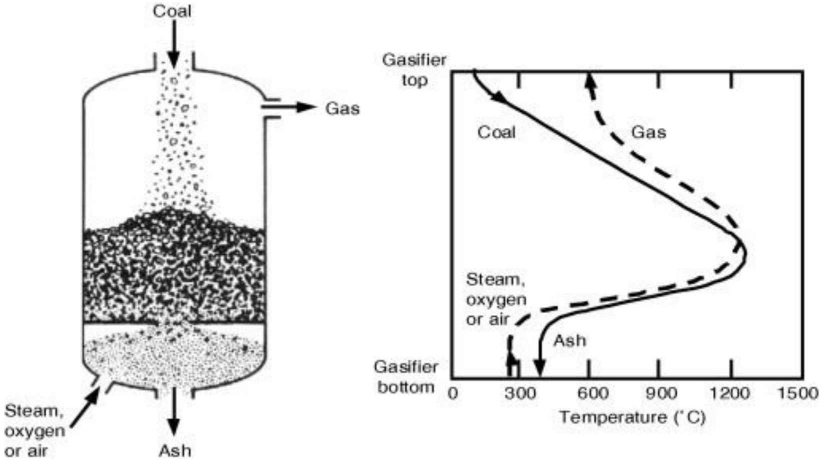


Figure 1.3 Schematic of moving bed gasifier with temperature profile

Moving bed gasifiers operate in two different modes which include dry ash and slagging types. The temperature is moderated to below the ash-slagging point by reaction of the char with excess steam in the dry-ash mode of operation. The ash below the combustion zone is then cooled by the entering steam and oxidant (oxygen or air) and leads a solid ash produced. In the slagging mode of operation, much less steam is used, and as the result, a much higher temperature is achieved in the combustion zone, melting the ash and producing it in the form of a solid slag.

The major disadvantage of moving bed gasifier is the limited ability in handling fine particles since a significant portion of ground coal is usually fine. Moving bed gasifiers normally consume large amounts of steam and the following product gas cleanup is more complex than other processes due to the presence of tars. Moving bed gasifiers provide some advantages such as high cold gas efficiencies due to efficient counter current heat transfer, high calorific value of product gas and low oxygen consumption. These gasifiers have the ability to produce syngas ratios that are applicable in FT synthesis. At present, there are three major commercial processes using moving bed gasifiers: The Sasol-Lurgi dry ash gasifier, the British Gas Lurgi slagging gasifier and a dry ash gasifier developed by Bharat Heavy Electricals Ltd (BHEL) in India. The Sasol-Lurgi dry ash gasifier is design by Lurgi and is used in the Sasol CTL plants in South Africa [18].

Fluidized bed gasifiers

Fluidized bed gasifiers, also referred to as fluid bed gasifiers, can be considered partially co-current gasifiers where the feed and the other reactants are completely mixed. These gasifiers use the incoming velocity of the air, steam or oxygen-steam mixture to fluidize the solid particles. The history and development of coal gasification using fluid-bed have been intimately linked since Winkler process was applied in the early 1920s.

The fluidized bed gasifiers normally consist of a vertical refractory lined cylinder with the gasification agents (e.g. steam or air) entering through the bottom and the feed, entering somewhere above, is effectively suspended in the gas flow. The region immediately above the feed zone is called the bed and is followed by the freeboard. The gasifiers are equipped with cyclones near the exit in order to capture the solid particles entrained in the gas stream. The difference between the mean gas velocity and the mean solids velocity in a fluid bed gasifier is known as slip velocity [17]. According to the velocities, fluid bed gasifiers can be classified into three categories. Stationary fluid bed gasifiers operate under a minimum slip velocity and the mean gas and solids velocities are relatively low. Circulating fluid beds operate under a maximum slip velocity and intermediate gas velocities and transport gasifiers operate under the highest gas velocities and all the solid particles are carried with the gas, resulting in pneumatic transport. In these cases, a clear distinction exists between the bed and the freeboard. An example of bubbling fluidized gasifier flow diagram is given in Figure 1.4.

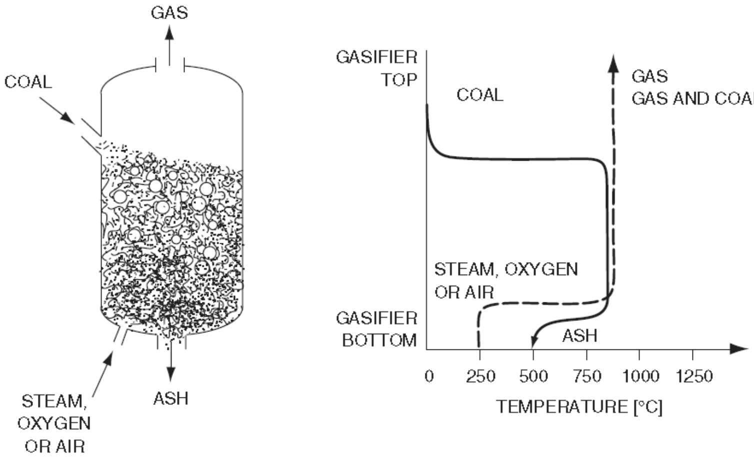


Figure 1.4 Schematic diagram of bubbling fluid bed gasifier with temperature profile

Most fluid bed gasifiers are operated below the softening point of ash and a certain feedstock that the ash fusion temperature is too low will not allow for gasification.

Higman and Van der Burgt [17] report that the temperature limits are in the range of 950-1100 ℃ for coal feed and typically 800-950 ℃ for biomass. Due to the relatively low temperatures, more reactive feedstocks with much more carbon in volatile matter such as lignite and biomass are preferred.

The particle size requirement is generally 6-10 mm and fluid beds can accept fines to a reasonable extent. Fluid bed gasifiers use moderate amounts of steam and oxygen and are not suitable for extremely small scale applications (< 10MW) because of their high heat transfer rates that results from the high amount of exposed feedstock surface area [23]. It is very complicated to operate the fluid bed gasifiers since the entire gasification process is very dependent upon a highly complicated equilibrium state, which must be maintained at all times. Major drawbacks of the fluidized bed gasifiers include the high tar content in the product gas, the incomplete carbon conversion, and poor response to load changes.

Well known fluid bed processes include the High Temperature Winkler (HTW) process, the Kellogg Brown and Root (KBR) transport gasifier, the Foster Wheeler (Ahlstrom) circulating fluid bed, the Kellogg Rust Westinghouse (KRW) agglomerating fluid bed process and the U-gas process developed by Institute of Gas Technology (IGT). Fluidized bed gasifiers are currently available on a semi-commercial basis from several manufacturers in Europe and U.S.

Entrained flow gasifier

The majority of the most successful coal gasification processes that have been developed after 1950s are entrained flow, slagging gasifiers operating at pressures of 20-70 bar and at high temperatures of at least 1300 ℃ with almost pure oxygen (99%) as oxidant. Such high temperatures make the use of catalysts unnecessary in most cases. Entrained flow gasifiers have become the preferred gasifier for hard coals, and have been selected for the majority of commercial scale Integrated Gasification Combined Cycle (IGCC) applications that will be discussed later in this chapter.

Entrained flow gasifiers can be considered as plug flow type reactors where the feed and the oxidant move concurrent to each other. Modern POX gasifiers are aimed at increasing the maximum amount of syngas in the product stream and reducing the amount of methane. Hence, they are operated at very high temperatures, in the order of 1400 ℃ or higher. The feed along with oxygen and a small amount of steam is fed from the top of the gasifier and moves downwards. This allows for a much more even temperature distribution and a more steady reaction rate.

The entrained flow gasifier is the most successful design for gasification and a majority of the commercial gasifiers in the world are entrained flow type. Entrained flow gasifiers are very common in very big power plants (> 200 MW) because they can achieve very high syngas mass flow rates, higher than any other gasifier type: a necessity for large plants. The ash is slagging in all entrained flow gasifiers and is removed in the exit quenching section as inert slags. The produced slagging inside the gasifier forms a protective coating along the sides of the gasifier, which protects the walls from more corrosive substances that may form during gasification. The greatest strength of the entrained flow gasifier is that it can accept a wide variety of feedstocks. Any liquid or powdered/pulverized solid is a viable fuel input for entrained flow gasifiers regardless of its atomic makeup. However, very high ash content in the feedstock may result in reduced thermal efficiencies with most heat loss due to slag formation. The product syngas from entrained flow gasifiers is totally free of tars and other heavy oils. The residence times are very short (normally 0.5-5 s) and there is no distinction between different reaction zones. The steam consumption is low and steam is generally added as a moderator.

Although, for all of their strengths, entrained flow gasifiers have a few debilitating drawbacks. First, the required average feedstock particle size is extremely small, typically 100 micron or even smaller. There is no problem for liquid feedstocks, but solids such as coal and biomass must receive a significant pretreatment before they can be used in an entrained flow gasifier. This is not an issue for coal since it can simply be ground down and pulverized mechanically.

Secondly, most entrained flow gasifiers typically require the use of oxygen as the gasifying agent instead of air. Very few entrained flow gasifiers are capable of using air, because the conditions in the gasifier make the presence of nitrogen a serious problem for syngas production. The high temperatures and pressures can cause large amounts of unwanted NO_x production that leads the resulting syngas mixture virtually unusable for power applications. Because of this strict oxygen requirement, almost all entrained flow gasifiers require an expensive Air Separation Unit (ASU) in order to operate. Finally, the syngas has an extremely high temperature after leaving the gasifier and there is high energy loss in the syngas cooling stage before it enters the gas cleanup system.

Entrained flow gasifier designs vary primarily in the feed type, feeding technique and also in the heat recovery technique. The feed can be either dry or in a slurry form with water. There are several entrained flow gasifiers available commercially. The Texaco gasifier technology, now owned by GE, is a slurry feed process (30 wt% water in the slurry) and is now considered an important option for IGCC projects due to less capital cost requirement shown in Figure 1.5.

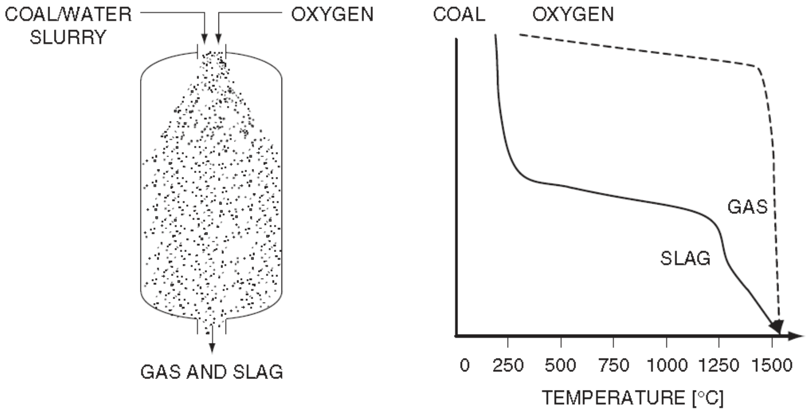


Figure 1.5 Schematic diagram of top-fired coal-water slurry feed slagging entrained flow gasifier with temperature profile (GE gasifier)

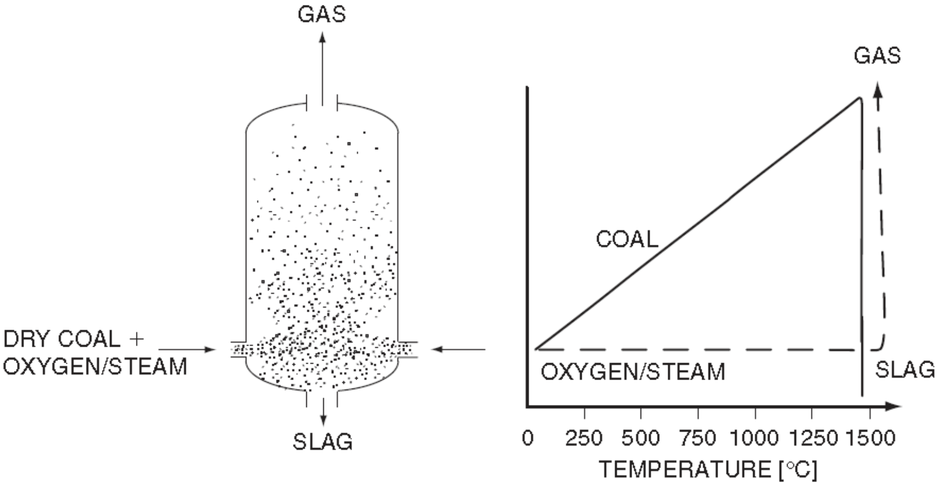
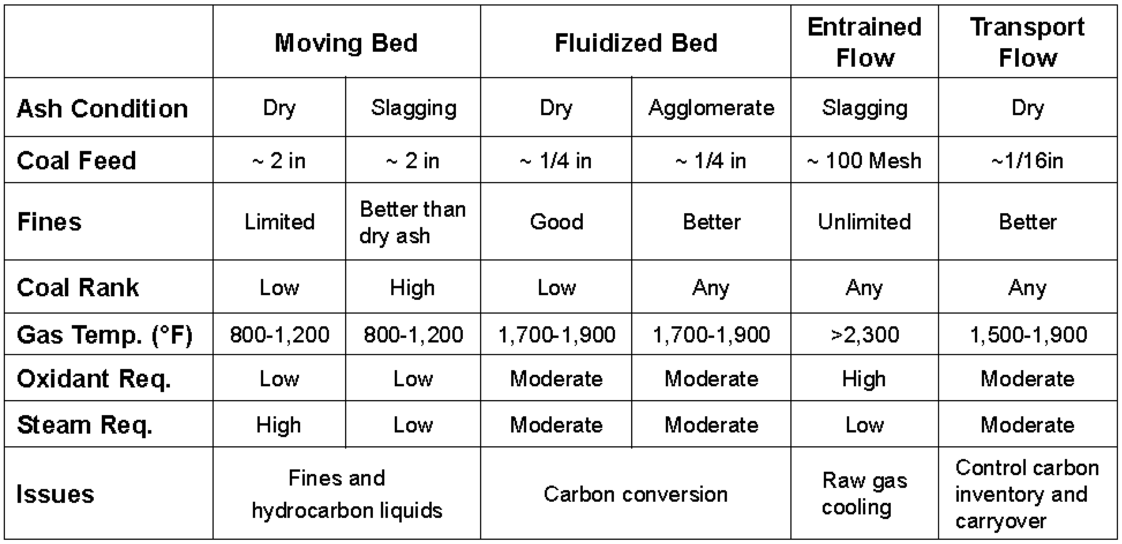


Figure 1.6 Side-fired dry coal feed slagging entrained flow gasifier (Shell Gasifier)

The Shell entrained flow gasifier is a dry feed pressurized slagging gasifier and is similar to the Prenflo process shown in Figure 1.6 [17]. Other well-known entrained flow gasifiers include the Noell gasifier, the E-Gas process and the Lurgi multi-purpose gasifier.

The features of three types of gasifier introduced in this section in commercialized coal gasification application are summarized in the Table 1.1 below [24].

Table 1.1 Pros and cons of three types of gasifier



	Moving Bed		Fluidized Bed		Entrained Flow	Transport Flow
Ash Condition	Dry	Slagging	Dry	Agglomerate	Slagging	Dry
Coal Feed	~ 2 in	~ 2 in	~ 1/4 in	~ 1/4 in	~ 100 Mesh	~1/16in
Fines	Limited	Better than dry ash	Good	Better	Unlimited	Better
Coal Rank	Low	High	Low	Any	Any	Any
Gas Temp. (°F)	800-1,200	800-1,200	1,700-1,900	1,700-1,900	>2,300	1,500-1,900
Oxidant Req.	Low	Low	Moderate	Moderate	High	Moderate
Steam Req.	High	Low	Moderate	Moderate	Low	Moderate
Issues	Fines and hydrocarbon liquids		Carbon conversion		Raw gas cooling	Control carbon inventory and carryover

1.2.3 Gasification application

This section presents a brief overview of some of the important gasification applications that are concerned with the non-catalytic conversion of carbonaceous solid feedstocks into synthesis gas.

Although catalytic conversion technologies for coal and biomass conversion have been proposed, these are still in the research stage and are not considered in this study. Until recently, gasification was used primarily as a means to generate electricity, fuel production and non-fuel chemicals synthesis. Integrated Gasification Fuel Cell (IGFC) system might be another option for gasification application although not considered in this thesis.

IGCC

IGCC is now a commercial technology and early plants in 1980s and 1990s have demonstrated the technical and environmental benefits of this technology. Interest in IGCC in the U.S. power industry started to appear in 2003, and this was strengthened by the acquisition of Texaco's gasification technology by GE Energy in 2004, which generated confidence in the possibility of an "overall wrap" for an integrated coal-in power-out plant supply arrangement and a number of commercial scale IGCC plants are in operation around the world currently [25].

The Department of Energy (DOE) clean coal demonstration project helped construct 2 prototype IGCC plants in U.S: Wabash River Power Station (250 MW, online 1995) in West Terre Haute, Indiana, Polk Power Station in Tampa, Florida (250 MW, online 1996). These plants have now been operating for more than 10 years, after an initial demonstration phase in a commercial setting. Considerable experience and a long list of "lessons learned" have been accumulated, which should provide a good basis for the next generation of plants [17].

IGCC plant with pre-combustion Carbon Capture and Storage (CCS) is shown in Figure 1.7 that converts carbonaceous materials into electricity [24]. In the plant, the carbon containing material is fed to the gasifier along with oxygen and steam to produce the raw syngas. The raw syngas is then cleaned of particulate matter, sulfur, chloride and other trace metal species. In order to increase the CO_2 recovery, the syngas produced passes through a shift reactor where CO is converted into H_2 and CO_2 and the primary H_2 stream after the acid gas removal is used as fuel in a gas turbine which produces electrical power.

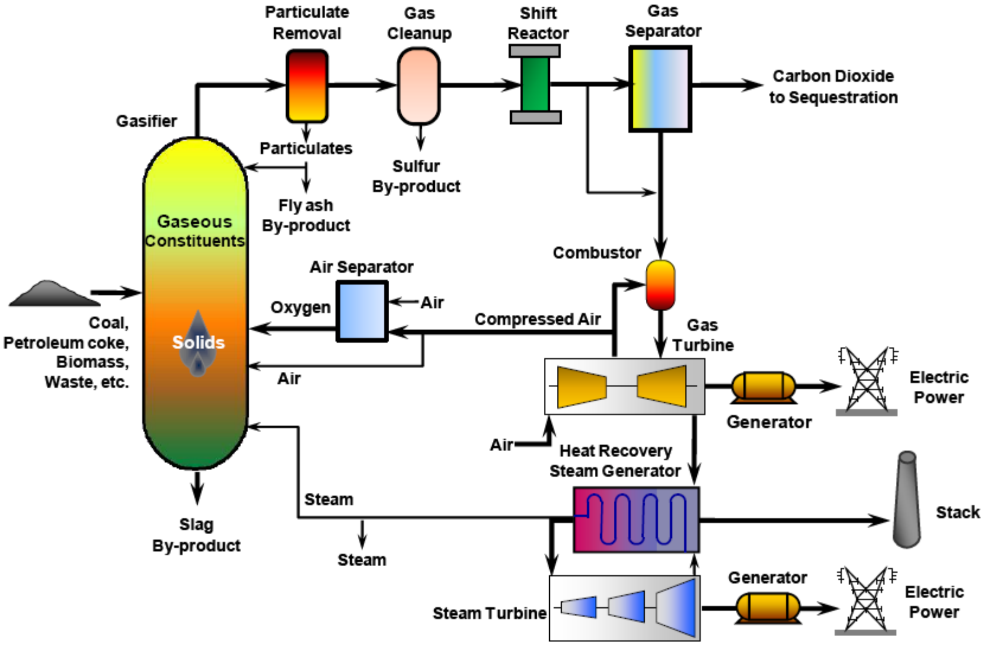


Figure 1.7 Schematic of IGCC plant with per-combustion CCS

In a normal combined cycle, flue gas from the gas turbine exhaust is used in a Heat Recovery Steam Generator (HRSG) to make additional steam for the steam turbine cycle. The steam is then used in multi-pressure-level steam turbines to produce additional electrical power. As a result, the IGCC plant improves its overall process efficiency by recovering the heat from the gasification and gas cooling process and adding the superheated steam to the steam turbine cycle.

The most important feature of power generation via IGCC is that it offers substantial advantages over pulverized coal combustion when CCS is required. It is obvious that removal of CO_2 from pressurized syngas is much easier and less expensive than removing CO_2 from a near atmospheric pressure flue gas produced by a pulverized coal plant. While in reality, a standard method for capturing CO_2 from pulverized coal plants is expected to reduce the plant's energy output by a quarter or more. This output loss for IGCC plants will be much smaller but reductions in output of greater than 15% are still expected [26, 27].

Under the same baseline with 90% carbon capture, Woods et al. estimated that the capital costs for IGCC with CCS were 2,400 \$/kW to 2,700 \$/kW versus 2,900 \$/kW for a subcritical pulverized coal combustion plant with CCS. IGCC without CCS has little to offer compared to traditional subcritical pulverized coal combustion process. On the contrary, the gasification is more expensive than traditional pulverized coal technology (typically 10-20%) and leads higher cost for electric power production [27, 28].

Fuels production and non-fuel chemicals synthesis

Instead of converting the energy content of the syngas to power, syngas produced from gasification can also be used for further processing to liquid fuels or non-fuel chemicals. The efficient commercial production of synthesis gas is gaining significant attention as the worldwide interest in synthetic fuels and chemicals is increasing. Coal is still the primary feedstock used in all the processes and the required syngas ratio varies depending on the desired product and the fuel production technology [18, 29, 30]. A list of some of the important industrial chemicals that can be produced from syngas and the required syngas feed ratio are given in Table 1.2.

Most applications of coal-derived syngas require a H_2/CO mixture with greater hydrogen content than that provided by the gasifier. Table 1.3 given the initial H_2/CO ratio in the raw product gas from most widely used commercial gasification technologies with oxygen and steam as gasification agents. The temperatures shown here are for the syngas prior to a quench section.

Table 1.2 Industrial chemicals production from synthesis gas

No	Product	H_2/CO ratio	No	Product	H_2/CO ratio
1	Fisher-Tropsch Fuel Co Catalyst	2.0-2.2	6	Acetic acid	1.0
2	Fisher-Tropsch Fuel Fe Catalyst	1.0	7	Ethyl acetate	1.0
3	Methanol	2.0	8	Vinyl acetate	1.5
4	Ethylene glycol	1.5	9	Ethanol	2.0
5	BTX	1.5	10	Ethylene	2.0

As shown in Table 1.3 [27, 31, 32], the high temperature entrained flow gasifiers generally produces H_2/CO ratios less than 1.0. On the other hand, the transport fluidized gasifier produces a high H_2/CO ratio due to a relatively low operating temperature and a high steam feed rate. The BGL gasifier has an even lower operating temperature, but the steam rate in is very low, producing a low H_2/CO ratio. In almost all these cases, the hydrogen concentration needs to be increased to meet the required ratio for fuels or chemicals synthesis (e.g., WGS).

Table 1.3 Syngas H_2/CO ratio for several most widely used gasifiers

Gasifier	GE	Conoco-Philips	Shell	Transport	BGL
Type	Entrained flow	Entrained flow	Entrained flow	Fluidized bed	Moving bed
Temp, ^oC	1316	1010	1427	907	537
H_2/CO	0.97	0.71	0.51	1.65	0.41

The conversion of syngas to fuels or chemicals can only occur in the presence of proper catalysts [33]. The catalytic reactions basically build up the small molecules in the syngas into larger compounds that are more easily stored and transported. The production of FT diesel fuel may require a H_2/CO syngas ratio from less than one to over two depending on the catalyst and specific technology used. Oxosynthesis and other processes such as DME synthesis generally require a H_2/CO ratio of 1.0. In most catalytic synthesis reactions, syngas cleanliness requirements are very high. The product syngas must be subjected to cleanup in order to remove contaminants such as sulfur components, heavy metals, chlorine before it can be used as a feed in the synthesis reactor. This step is especially important in the fuels and chemicals production since most catalysts are very quickly deactivated by contaminants such as sulfur. Most impurities and contaminants are removed to very low level (part per million, ppm) and even extreme low (parts per billion, ppb) concentrations. This also means that significant cost must be directed toward syngas cleaning. Syngas cleanup is in itself a major area of research and development and there are many studies available that provide a detailed review of this step [11, 19].

The composition of the syngas from gasification processes is generally not suited for direct use in the downstream fuel or chemical synthesis process. In most cases, H_2/CO syngas ratio adjustment techniques such as downstream shift reactors, membrane separators or pressure swing adsorption are employed to meet this ratio requirement [18, 34-36]. This also adds cost and complexity to the overall process in several cases.

Eastman methanol plant in Kingsport, started up in 1984 is the leader in the Coal-to-Chemicals field [17]. Currently, many coal or coke-based fertilizer, methanol and DME plants in Japan, U.S. and China exist with many of the major gasification technologies such as Lurgi, GE Energy and Shell. Sasol in South Africa is the pioneer in the coal derived liquid fuels process for synthetic gasoline/diesel production through FT synthesis and this technology is discussed in detail in the following section.

1.3 Synthetic fuel production based on gasification

As mentioned earlier, the focus of this thesis is mainly on the synthetic fuel production. The term synthetic fuels or synfuel is mainly used to describe liquid hydrocarbon fuels (< C24) derived from coal, biomass or natural gas. These fuels closely resemble gasoline and diesel derived from crude oil in terms of composition, energy density, properties and combustion efficiency. As a result, the synthetic fuels are often called synthetic gasoline or synthetic diesel. Coal can be liquefied directly or can be converted into synthetic fuels indirectly via FT process. Direct liquefaction technologies are undergone at very high temperatures and pressures based on the coal hydrogenation. The initial liquid hydrocarbons produced must go through a significant refinery process in order to generate qualified gasoline or diesel. Further information on direct liquefaction process can be found elsewhere [11, 12] and the focus of this thesis only on the indirect liquefaction method which is the synthetic fuels production from syngas by FT process. Gasification for the purpose of liquid fuel production from coal was originally developed in Germany during the 1930s and 40s and is now referred to as Coal-to-Liquids (CTL) process. CTL was later pioneered by Sasol (South African Coal, Oil and Gas Corporation) during the 1960s and 70s. Only two commercial CTL plants exist in the world as of 2010 and both are owned and operated by Sasol, a company that is considered to be the leader in the CTL industry by a wide margin. These plants, located in Secunda, South Africa, have a combined capacity of roughly 150,000 barrels/day [37] and supplies 41% of South Africa's transportation fuel requirements [33].

The FT process was developed around 1925 by Franz Fischer and Hans Tropsch at the Kaiser Wilhelm Institute for Coal Research in Germany. FT technology can be described as the catalytic conversion of syngas into a series of hydrocarbons ranging from methane to heavy waxes [38]. The basic FT reactions are:

$$nCO + 2nH_2 \rightarrow C_nH_{2n} + nH_2O \qquad 165\ \text{MJ/kmol} \quad (1.10)$$

Because of the highly exothermic reaction, the heat must be removed or the catalyst can be deactivated. Two main types of reactors have been designed: a fixed bed tubular reactor and a slurry phase reactor shown in Figure 1.8 [33].

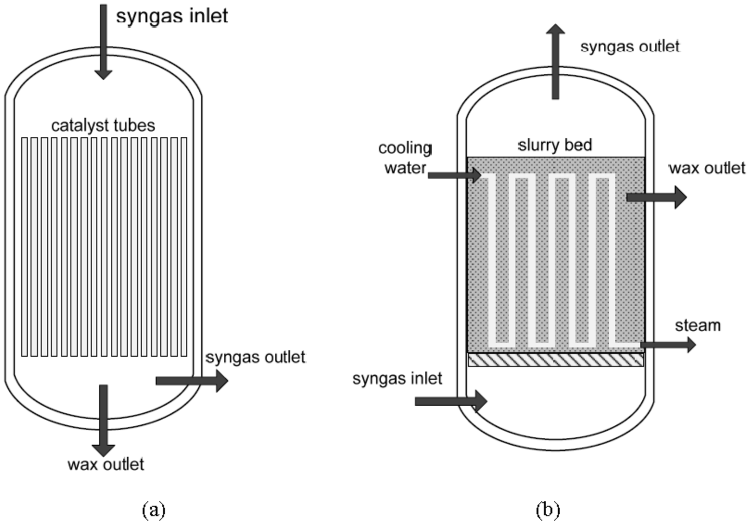


Figure 1.8 FT reactor types (a) multi-tubular fixed bed and (b) slurry bed

Heat removal is crucial to the process and has been the focus of reactor design development. The fixed bed reactor has many catalyst tubes with heat removal achieved by steam generation on the outside of the tubes [39]. The fixed bed reactor is simple to operate and is well suited for wax production due to simple liquid/wax removal. However, it is more expensive to build because of the many tubes and has a high pressure drop across the reactor [40]. The slurry phase reactor operates by suspending catalyst in a liquid and bubbling bed with syngas through from the bottom. Disadvantages of the slurry phase reactor are more complex operation and difficult wax removal. However, slurry phase reactor cost approximately 40% less to build than fixed bed reactors [40].

The product distribution is controlled by several factors, including the catalyst material, reaction temperature and pressure, and reactor type. Cobalt and iron are the two most important catalyst materials used for FT synthesis. Each catalyst has its own advantages and disadvantages. For cobalt catalysts, WGS activity is almost always negligible. Ruthenium is also active as a FT catalyst but it is not used due to availability and cost issues. The syngas ratio necessary for cobalt catalysts is from 2.0 to 2.2 whereas for iron catalysts, a ratio of 1.0 is sufficient. The reason for the lower hydrogen requirement of iron catalysts is due to the strong WGS activity during the reaction and syngas ratio of more than 1.8 is expected by the spontaneous shift reaction.

FT liquids synthesis are usually carried out at pressures in the range of 10-60 bars (150-900 psia) and temperatures from 200 to 350 °C [18]. Product distribution can be estimated using the Anderson-Schulz-Flory chain-growth-probability model, in which longer hydrocarbon chains form as the temperature decreases (see Appendix B).

For gasoline-range products, higher temperatures (300 °-350 °C) and iron catalysts are typically used. For diesel-range and wax products, lower temperatures (200 °-240 °C) and cobalt catalysts are typically used [39]. A significant amount of the FT products are liquids under ambient conditions, although gaseous and solid products are also generated. The liquid product consists of mostly paraffins that can be refined to produce clean transportation fuels such as diesel, kerosene/jet fuel or gasoline. The heavy waxes can also be hydrocracked to generate more fuel, although they are valuable products in separate markets such as the cosmetics industry. The by- product naphtha product can be also used in the manufacture of ethylene and propylene [18].

The conversion extent in the FT reactor is limited, depending on catalyst type and reactor size and technology. The reactor product stream thus contains unreacted carbon monoxide and hydrogen in addition to the FT products. The C5+ products are easily separated by a condensation step and sent to the recovery and upgrading section. Recovering the very small fraction of C4 is energy consuming and in general not economic [41].

To maximize the production of FT liquids, the off-gas containing unreacted H_2, CO and light hydrocarbon C1-C4 can be recycled to the entrance of the reactor. The recycle can contain a reformer to reconvert C1–C4 back into syngas, and a shift reactor. Furthermore, the FT off-gas may be recycled to the gasifier and subsequent tar cracker, which will work as reformer. This latter option is only sensible when the gasification pressure is the same as the FT pressure. With higher once through conversions in future, the recycling option could become obsolete.

Instead of maximizing fuel production, the system can also be optimized towards combined fuel and electricity production. In this case, the syngas passes only once through the FT reactor. The FT off-gas is not recycled, but completely purged to a combined cycle for electricity production. Investments are assumed to be lower (e.g. no reformer) and presence of inert is less problematic. Co-production of FT liquids and electricity may combine a high overall efficiency with lower investment costs.

FT based fuels has several advantages over crude oil based products, including the virtual absence of sulfur, very low aromatic content and high cetane numbers. FT diesel performs better than conventional diesel in terms of engine emissions [42] and is often blended with crude oil diesel in order to improve the grade of the crude oil diesel. Most important, FT based fuels are compatible with current automobile engines and clean burning with less GHG emission if the fuel is derived from non-fossil source. At last, since it can be produced, at least in principle, from any carbonaceous matter, they have the potential to contribute towards a sustainable energy future and also can provide an opportunity to convert waste matter into useful products.

Summary

Gasification processes provide a more efficient and cleaner method for the usage of syngas derived from carbonaceous material for power generation, fuels and chemicals synthesis. There are two major trends that prompt current interest in gasification. The first is the widely held belief that conventional petroleum supplies are declining, while demand for transportation fuels continues to rise. This has led to heightened interest in alternative energy supplies. The second major trend is concern about global warming. Gasification offers a relatively cost-effective means of using coal or other carbonaceous materials while minimizing GHG emissions if CCS process is utilized or carbon neutral renewable feedstocks (e.g. biomass and MSW) are used.

The focus of this thesis is to develop and design a new advanced gasification technology, called steam hydrogasification, utilizing fluidized bed type gasifier for synthetic fuels and/or power production that offers several potential advantages over conventional gasification technologies.

2. CE-CERT process description

2.1 Steam hydrogasification

As mentioned earlier, the most common gasification agents are oxygen, steam, air and hydrogen although CO_2 has also been used in some cases [43]. The gasification agents play an important role in determining the product composition, overall efficiency and the design basis of the gasifier. Generally, the gasification of solid feedstocks in the presence of hydrogen is called hydrogasification whereas gasification in the presence of steam is known as steam pyrolysis.

Hydrogasification was initially developed in the early 1900s and there was a revived interest in the process during the 1970s and 80s as a method to produce methane from coal. The basic reaction is the direct methanation of carbon. Significant amount of heat must be imported in bringing the reactants up to desired temperature and also to sustain the process although the reaction is mildly exothermic.

Since methane production is favored at high pressures (~100 atm) and the process is generally operated at temperatures ranging from 750 ℃ to 1000 ℃ [17]. A number of processes were developed including the HKV process (Hydrierende Kohlevergasung) by Rheinbraun in Germany, Hydrane gasifier by the U.S. Bureau of Mines (Department of Energy), and the Hygas gasifier by the Institute of Gas Technology (IGT) [11, 17]. The HKV fluidized bed pilot plant is successfully operated at temperatures of 850 to 930 ℃ at pressure of 60 to 95 bars with 24 ton per day feed rate. The Hydrane gasifier consisted of a two stage reactor with a free-fall section followed by a fluidized bed and was operated at 200 bar pressure and a temperature of 750 ℃ [11]. Both the HKV and the Hydrane processes produced hydrogen by gasification of the residual char from the hydrogasifier. The Hygas gasifier developed by IGT was also based on a fluidized bed but the hydrogen production was integrated with the gasifier [11, 44].

There was no further development of the hydrogasification due to the cheap and abundant natural gas supplies at that time [17]. The other major issue with hydrogasification processes was the source of hydrogen supply since hydrogen production can be expensive. Besides that, the low reaction activity of carbon with hydrogen was also a main reason that made this process unattractive. The reactivity of carbon with different species at 1073 K and 0.1 atmospheres are shown below [45].

$$\begin{array}{ccccccc} r_{O_2} & >> & r_{H_2O} & > & r_{CO_2} & > & r_{H_2} \\ 10^5 & & 3 & & 1 & & 3.1^{-3} \end{array}$$

Steam pyrolysis is one of the earliest known gasification processes to decompose the organic material into gas, liquid and solid phases. It is mostly restricted to small scale electricity generation plants and is also used in fast pyrolysis (400-600 ℃) to produce bio-oil from biomass at ambient pressure with a very short residence time (0.5-5 s). Due to the poor efficiency, steam pyrolysis is currently not considered to be a major gasification technology option but a pretreatment method.

After years of dedicated research, College of Engineering – Center for Environmental Research and Technology (CE-CERT) at the University of California, Riverside has found that the introduction of steam in hydrogasification process can increase the rate of methane formation significantly. Based on the study, the steam in the hydrogasification process overcomes the slow reactivity of hydrogen with carbon to some large extent. In a word, the combination of steam pyrolysis and hydrogasification, with a high methane yield with less residence time, is called steam hydrogasification [46].

2.2 CE-CERT process

Fuels (e.g. gasoline and diesel) and electricity is finally produced from the methane rich gas after a series of operation units. This energy production process based on steam hydrogasification is called CE-CERT process with block flow diagram given in Figure 2.1 below.

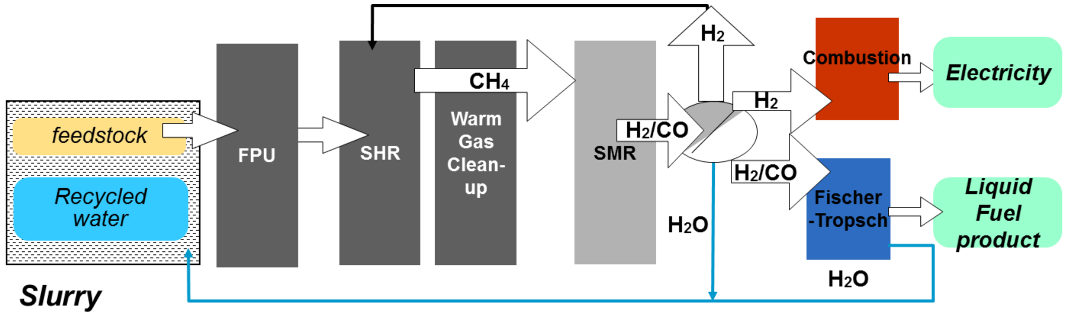


Figure 2.1 Block flow diagram of CE-CERT process

The basic chemical reactions taking place during the different stages of the CE-CERT process are given below.

SHR: $C + H_2O + 2H_2 \rightarrow CH_4 + H_2O$ + Others (CO, CO_2 and C_{2+}) 75 kJ/mol (2-1)

SMR: $CH_4 + H_2O \rightarrow 3H_2 + CO$ -206 kJ/mol (2-2)

FTR: $nCO + 2nH_2 \rightarrow C_nH_{2n} + nH_2O$ 165 kJ/mol (2-3)

The details of this process have been published elsewhere [47-50] and a short description of the important unit operations is presented here. Initially, the feedstock is made into slurry with water in the Feedstock Pretreatment Unit (FPU) and is fed into the SHR along with H_2 and steam (optional). The SHR produces CH_4, along with CO and CO_2 (Eq 2-1) heated by hot sand circulating from the regenerator. The CH_4 rich gas from the SHR is then subjected to a warm gas cleanup unit in order to remove the contaminants, primarily sulphur species. After that, most of CH_4 is converted into syngas, a mixture of H_2 and CO, in the SMR (Eq 2-2). Part of H_2 is separated and fed back to SHR in order to meet the system H_2 internal cycle requirements while liquid fuels are synthesized with an adjusted H_2/CO feed ratio in a Gas-to-Liquid (GTL) unit, usually a FT reactor (Eq 2-3). Finally, light hydrocarbons (C1-C4), unconverted syngas from FT reactor and surplus H_2 are burnt for heat and power generation.

Major attractive features of the CE-CERT process are summarized below.

1. CE-CERT process is suitable for small or medium scale distributed facilities which uses localized feedstock source. This feature is especially favorable in term of biomass and MSW gasification which have a restricted availability of feedstock within a reasonable transportation distance and usually do not lend themselves to large capacities. On the contrary, conventional oxygen-blown gasification plant is not considered to be economically viable in smaller operation scale due to the expensive capital cost in the ASU for oxygen generation.
2. The slurry feed method reduces the complexity related to the transportation of the feedstock into the gasifier by using a slurry pump. Wet feedstock with high moisture content can be used directly and this also reduces the feedstock drying costs.

3. Enhanced conversion of carbonaceous material to synthesis gas compared to other thermochemical processes. The non-usage of catalyst with moderate operation temperature and pressure of the SHR further reduces the capital and operation cost.
4. Closed-loop H_2 cycle, no external H_2 is needed and makes the process self sustainable.
5. The H_2/CO ratio of the product syngas can be controlled by varying the H_2O/feedstock and H_2/C ratio of the SHR feed. This avoids the need for more complex configurations currently employed in H_2/CO adjustment in other commercial processes.

Recently, the National Energy Technology Laboratory (NETL), a division of the DOE has just performed an in-depth independent techno-economic analysis of the steam hydrogasification process. The report concludes that this process has the potential to offer 3%-5% higher efficiency with less 14% capital costs compared to conventional state of the art gasification technologies [51].

After years of research, test results have shown that CE-CERT process is versatile and can handle a number of different feedstocks including conventional coal (lignite and bituminous) and biomass while some feedstocks with initial high moisture (>80%) can also be employed such as sewage sludge (biosolid) and microalgae. A detailed introduction of CE-CERT process with specific operation unit setup is given in the following sections.

2.3 Process description

Technologies for the commercial production of the synthetic hydrocarbon fuels using syngas (H_2 and CO) are currently considered to be mature. The CE-CERT process for fuels and electricity production consists of five major sections: (1) Feedstock pretreatment; (2) SHR with regenerator; (3) Gas cleanup; (4) SMR; (5) Gas conditioning; (6) Fuel synthesis and electricity generation.

Feedstock pretreatment

This unit operation aims to pump the feedstock slurry to the gasifier and specific treatment method is employed if necessary. The feedstock needs to go through a size reduction process before it is made into slurry form with water. The desired particle size and pretreatment procedure vary with initial moisture content and physical property of different kinds of feedstocks. For example, coal is easy to be ground and pumpable slurry feed can be simply made with H_2O/coal mass ratio of 1.0 due to its hydrophobic nature. The desired particle size in the coal slurry is about 70 microns. By contrast, biomass, with less carbon and energy density than coal, is highly hygroscopic and hydrophilic which gives rise to the difficulty to prepare suitable biomass slurries [52]. The hydroxyl groups in the polymeric structure of the biomass (e.g., cellulose, hemicellulose, and lignin) only allow for a small portion of solid in the biomass or biomass comingled mixtures. To prepare high solid biomass or biomass w/ biosolid comingled slurry, a Hydrothermal Pretreatment (HTP) process has been developed [53]. Slurry can be formed from the batch type feedstock HTP system with an initial solid loading of over 40 wt% (particle size < 100 microns). The example of untreated and HTP treated biomass (pine wood) and biosolid mixture samples are shown in Figure 2.2.

HTP treated biomass and biosolid slurry has a viscosity value of less than 1.5 Pa s at a shear rate of over $80 s^{-1}$, and such a slurry is pumpable [53]. The slurry feed of the feedstock instead of using a lock-hopper dry feed type makes the feeding system continuously, less energy-consuming and more economically. More important, it solves the technical problems posed in transportation of biomass into pressurized reactor in actual applications. Since water is utilized as a steam source in the CE-CERT process, a slurry formed homogeneous liquid and solid mixture is more preferred to be fed simultaneously into the SHR [54, 55].

Figure 2.2 Untreated and HTP treated biomass and biosolid mixture samples

SHR with regenerator

Gasifier is the core part of the CE-CERT process and many gasification technology options are explored in determining the type of the gasifier and finally the “low unit throughput” fluidized bed gasifier concept is chosen for application in the CE-CERT process due to the process requirements of long gas residence times that may be required to crack hydrocarbons (carbon molecules heavier than C4 or possibly as heavy as tar that could not easily be handled in recuperative heat exchangers or could physically condense and block the heat exchangers). These heavier molecules could also impact operation of warm gas cleanup or in the steam reformer.

The feed is transported into the SHR via a slurry pump along with recycled H_2. The water in the slurry provides the necessary steam source inside the gasifier and additionally steam, if necessary, is superheated and fed into the gasifier.

The issue of hydrogen supply has been resolved by recycling a portion of the hydrogen produced by the SMR. Although the methanation process inside SHR is mildly exothermic, large amount of heat is consumed in order to vaporize the water in the slurry and bring the feedstocks up to desired temperature, normally 700 ^{o}C-800 ^{o}C, and also to sustain the reaction. Therefore, the overall process is strongly endothermic and needs an external source of heat. This is the major disadvantage caused by the absence of an oxidizing agent such as oxygen or air. However, the issue of heat supply can be resolved through the use of a circulating fluidized bed reactor with solid recycle as heat carrier and this technology concept is one that is commercially utilized in ore roasting by Technip, Inc [56].

Sand is the most widely used material in the fluidized bed due to its stable property, cheap price and high melting point (>1600^{o}C). In this case, a circulating fluidized bed gasifier with a regenerator setup is designated with heat supply by combusting the leftover char schematically shown in Figure 2.3.

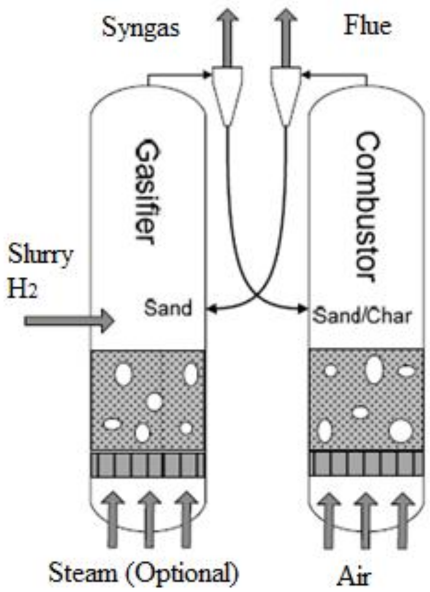


Figure 2.3 Schematic flow diagram of SHR with regenerator using sand as flow media

Other types of fuel source (e.g. natural gas, syngas or coal) might be used in the regenerator if the burning of unreacted char can't meet the gasification heat requirement. The carbon conversion in the SHR can be defined as functions of residence time and reaction condition (feed ratio, temperature and pressure). The gasification temperature and method to provide the process heat are of vital importance in determining the plant performance and process economy. Further analysis is confined to 750 ℃ operating temperature since the amount of feedstock carbon that has to be consumed in the gasifier sand regenerator is more for the 750℃ operating temperature than at the 850℃ temperature. However, the impact of hot sand solids recirculation to minimize hydrocarbons greater than C4 still needs to be confirmed based on pilot plant testing. The detailed design specification of SHR with regenerator has been performed based on process simulation as part of this thesis and the results are presented in Chapter 4.

Warm gas cleanup

After the particle removal, the raw gas from the SHR contains contaminates such as hydrogen sulfide, mercury, chloride and ammonia, etc. The downstream uses of syngas in CE-CERT process require that most of the contaminants present in the raw gas be removed to very low levels prior to use (in ppmv or ppb level). Many of these contaminants can lead to erosion and corrosion and act as poisons to the catalysts that are often used in the downstream gas processing and fuel synthesis steps. Conventional methods for removing sulfur and other contaminants from syngas typically rely on chemical or physical absorption processes (e.g. rectisol, sectisol and LO-CAT) operating at temperatures of 38 ℃ or less. Almost all of the steam present in the syngas condenses when the stream is cooled to this low temperature. After contaminants removal, the gas has to be reheated and sometimes additional steam is needed for downstream hydrogen production (SMR or WGS). These process swings adversely impact the plant's thermal efficiency and cost greatly. Economic analysis shows that gas cleanup processes under higher operating temperatures could significantly reduce this efficiency loss and improve the gasification plant performance [57]. In this thesis, Research Triangle Institute (RTI) warm gas cleanup technology is utilized all through the plant design and process simulation. The RTI process is in the leading position of warm gas cleanup field and has been tested extensively on a pilot plant scale slipstream at the Eastman Chemical gasifier site in Kingsport, TN. In over three thousand hours of operation at three operating pressures at Eastman Chemical, the desulfurization or sulfur removal achieved on this pilot unit was between 99.82 and 99.93% on inlet gases exceeding 8200 ppmv total H_2S

and COS content. RTI is developing modular fixed bed technologies for other contaminants including disposable sorbents for HCl, Hg, As and Se with operating temperature greater than 260°C [58]. Detail description of RTI gas cleanup process will be given in Chapter 4.

SMR

Steam reforming is a well-known process in gasification plant and also the most widely used commercial technology for syngas production. The primary application of steam reforming is hydrogen production. Methane or natural gas is used most time as the feedstock while hydrocarbons including naphtha are also applicable for this process. Since the reforming reaction is strongly endothermic, it is carried out in tubular reactors filled with catalyst with a fired furnace setup for heat supply. Most commercial steam reforming catalysts are nickel based loading on carrier such as alumina and zirconia, etc [18]. Cobalt and other noble metals are also active in the reforming reaction, however they are much more expensive compared to nickel.

The SMR catalysts are easily poisoned by contaminants such as sulfur species, halogens, arsenic and other heavy metals. This is especially truth for sulfur which is extremely sensitive in terms of the catalysts activity. The presence of sulfur will deactivate the catalyst permanently and it is recommended that the sulfur levels in steam reformer feed gases are kept below 0.1 ppmv. Besides that, carbon deposits on the catalyst surface can also cause deactivation and the steam/methane mole ratio of the reformer feed must be kept at a value of around 3 or higher in order to avoid this weakness.

Methane or other light hydrocarbons can also be reformed in the presence of CO_2 and this process is called dry reforming. There has been considerable interest in the dry reforming of methane since this allows two major greenhouse gases to be utilized simultaneously for the production of synthesis gas. A major disadvantage of dry reforming is also the carbon deposition which leads to catalyst deactivation. But this problem can be mitigated by the addition of steam to the process. The product stream of a steam reformer is generally rich in hydrogen and various techniques are employed to optimize the syngas ratio [35]. Methane conversion is favored by higher temperature, but metallurgy and heavy heat load limits the practical operating temperature. The methane conversion efficiency of more than 85% can be achieved from industry operation experiences based on nickel catalyst at around 850^{o}C. A detailed description of the SMR plant design including reactions involved and important parameters such as corresponding steam/methane ratio from SHR effluent is provided in Chapter 4.

Gas conditioning

The H_2 from the product gas stream is then separated using a Pressure Swing Adsorption (PSA) system or a polymeric membrane. A large stream of H_2 is recycled back to SHR to meet the system internal H_2 cycle requirement while the syngas ratio is adjusted to desired value for fuel synthesis application. Any excess H_2 is then burnt for power generation.

Fuels and electricity production

Like many other gasification plants, the CE-CERT process is also a heat intensive system that with massive heat imported and exported through each operation units. The process heat, although doesn’t contain in the fuels as major energy output, can be converted into useful co-product as electricity through HRSG and steam cycle in the steam turbine. To maximize this heat recovery, effective heat integration needs to be applied in the process where heat from process steams that require cooling is transferred to other streams that require heating. In industry, pinch analysis is a good way and widely used to optimize the heat integration within the plant by determining the system pinch point and the minimum delta T [59]. The export electricity as a co-product increases the overall process thermal conversion efficiency largely and lowers the fuels production cost which results in better economics, especially in large scale plant.

In this thesis, FT fuels with electricity co-production using coal or other renewable feedstock (e.g., biomass green waste and microalgae) based on steam hydrogasification will be further investigated. Detailed plant design and process simulation have been performed and the results are presented in Chapter 4 and 5. Other applications of steam hydrogasification such as power generation via IGCC and SNG production will also be discussed, but not as detailed as synthetic fuels production pathway.

2.3 Application of CE-CERT process

CE-CERT process generates a methane rich gas stream from carbon-containing feedstocks which is ideally suited for reforming to produce syngas directly due to the presence of a significant amount of unreacted steam. Also, the flexible control of H_2/CO ratio avoids the need for more complex configurations currently employed in several commercial processes as discussed earlier. The production of synthetic liquid fuels using syngas from steam hydrogasification directly coupled to a reforming process has been successfully demonstrated in our laboratory using several carbonaceous feedstocks [46, 48]. Moreover, a revised CE-CERT process is also under development aiming to produce SNG by using a WGS reactor instead of SMR and detail process description is given in Chapter 6.

The following configuration will be discussed and applied with process design and modeling based on steam hydrogasification using coal, biomass, microalgae and biosolid as feedstock.

1. FT fuels production based on iron catalyst (H_2/CO ratio=1.0) from bituminous coal.
2. FT fuels production based on cobalt catalyst (H_2/CO ratio=2.0) from bituminous coal.
3. FT fuels production based on cobalt catalyst (H_2/CO ratio=2.0) from biomass (green waste).
3. FT fuels production based on cobalt catalyst (H_2/CO ratio=2.0) from microalgae.
4. SNG production from biomass and biosolid comingle feedstock.
5. Power generation via IGCC from bituminous coal.

It should be noticed that last scenario power generation based on steam hydrogasification from coal via IGCC is not modeled in detail. Instead, plant performance from NETL report is used as baseline for the discussion and comparison.

The main objectives of this thesis are the development and optimization of conceptual plant design (commercial scale) for the scenarios listed above by means of advanced process simulation and economic analysis. The goal is to give objective assessments of steam hydrogasification process in commercialization applications by contrasting it with current completive main stream technologies within the field.

3. Process design and economic analysis methodology

3.1 Process modeling

Process modeling is a model-based representation of chemical, physical, biological, and other technical processes and unit operations in software. Basic prerequisites are a thorough knowledge of chemical and physical properties of pure components and mixtures, of reactions, and of mathematical models which, in combination, allow the calculation of a process with the aids of computers. During all systematic investigation, modeling is essential in order to understand and to analyze the various steps of experimentation, data analysis, process development, and engineering design. It is a tool that provides support for the planning, design, and evaluation of systems as well as the evaluation of strategies for system transformation and change.

Modeling has drawn the attention of scientists and engineers for many decades and now is still a subject of major importance for the knowledge of unitary processes as a fundamental key in process design and scale up. There is a wide range of possible reasons for undertaking a modeling and simulation study. Some of the most common are listed below [60]:

1. Education and training
2. Engineering design
3. Evaluation of decision or action alternatives
4. Evaluation strategies for transformation or change
5. Forecasting
6. Performance evaluation

7. Prototyping and concept evaluation

8. Risk/safety assessment

9. Sensitivity analysis

10. Support for acquisition/procurement decisions

In all, a model plays the role of an alternative for the system it represents and is to replace the system in experimental or practical studies. Especially when the underlying system does not exist or it may merely be an idea, concept or proposal, then developing a model is the only option as substitute.

Data is always easier to obtain from a model than from the system itself in most time and this is another important reason for supporting experimentation with a model. Furthermore, the fact that the platform for the modeling is a computer and this ensures reproducibility of results which is an essential requirement for establishing credibility of any investigations.

3.2 Process simulation software

Chemical process simulation software describes processes in flow diagrams where unit operations are positioned and connected by product or educt streams. The software has to solve the mass and energy balance to find a stable operating point. The goal of a process simulation is to find optimal conditions for an examined process with the help of the risk-free analysis. This is essentially an optimization problem which has to be solved in an iterative process. Some world widely used software in chemical engineering process simulation is listed below.

1. PRO/II by SimSci-Esscor Inc, USA

2. Aspen Plus and Hysys by AspenTech Inc, USA
3. gPROMS by PSE Inc, UK
4. ChemCAD by Chemstations Inc, USA
5. Design II by WinSim Inc, USA
6. VMGSim by Virtual Materials Group, CA

All these products have highly refined user interfaces and on-line component databases. They are widely used in real world applications from interpreting laboratory scale data to monitoring a full scale plant. Among these software, PRO/II, Aspen Plus and Hysys are the mostly widely used in chemical process simulation. PRO/II has mature simulation experience on petrochemical industry and builds a comprehensive database from industrial operation. Thus it has high simulation accuracy in oil refining field. Hysys is mostly used in oil refining and has a precise simulation module in oil gas. Dynamic simulation is its unique feature and the simulation is preferred in chemical engineering process that has a relative large and long pathway.

Aspen plus [61], a core element of AspenTech's aspenONE® Engineering applications, is a market-leading process modeling environment for conceptual design, optimization, and performance monitoring for the chemical, polymer, specialty chemical, metals and minerals, and coal power industries. It covers almost all the advantages of the chemical simulation software listed above and has some unique features, especially in electrolyte, solid non-conventional material and combustion unit. It includes the world's largest database of pure component and phase equilibrium data for conventional chemicals, electrolytes, solids, and polymers (more than 4 million experimental data points for over

24,000 pure components and 30,000 binary systems). Moreover, is integrated with AspenTech's industry-leading sizing and cost analysis software and heat exchanger design software to streamline the overall design and analysis workflow.

With reliable thermodynamic data, excellent performance in handling non-conventional solid matter, and rigorous equipment models, Aspen Plus is used to model CE-CERT process with different configurations all through this thesis.

Conceptual design of commercial scale facilities for varies energy product generation will be built with the aids of Aspen plus and overall plant capital and operation costs will be estimated based on the mass & energy acquired from the advanced simulation. It should be noted that the results from process simulation must be verified through experimental data before used for further application.

3.3 Process design methodology

The following steps were undertaken in this study:

- Conceptual design of facilities for synthetic fuels and SNG generation with relevant technologies under evaluation.
- Design detail process models using Aspen Plus process engineering software.
- Size and cost equipment using Aspen Icarus Process Evaluator software, literature references, and experimental data.
- Determine total plant cost and capital investments.
- Conduct discounted cash flow analysis on production cost, IRR, NPV and ROI evaluation.
- Perform sensitivity analysis on major process parameters.

The process simulation and economic evaluation methodology is schematically shown in Figure 3.1.

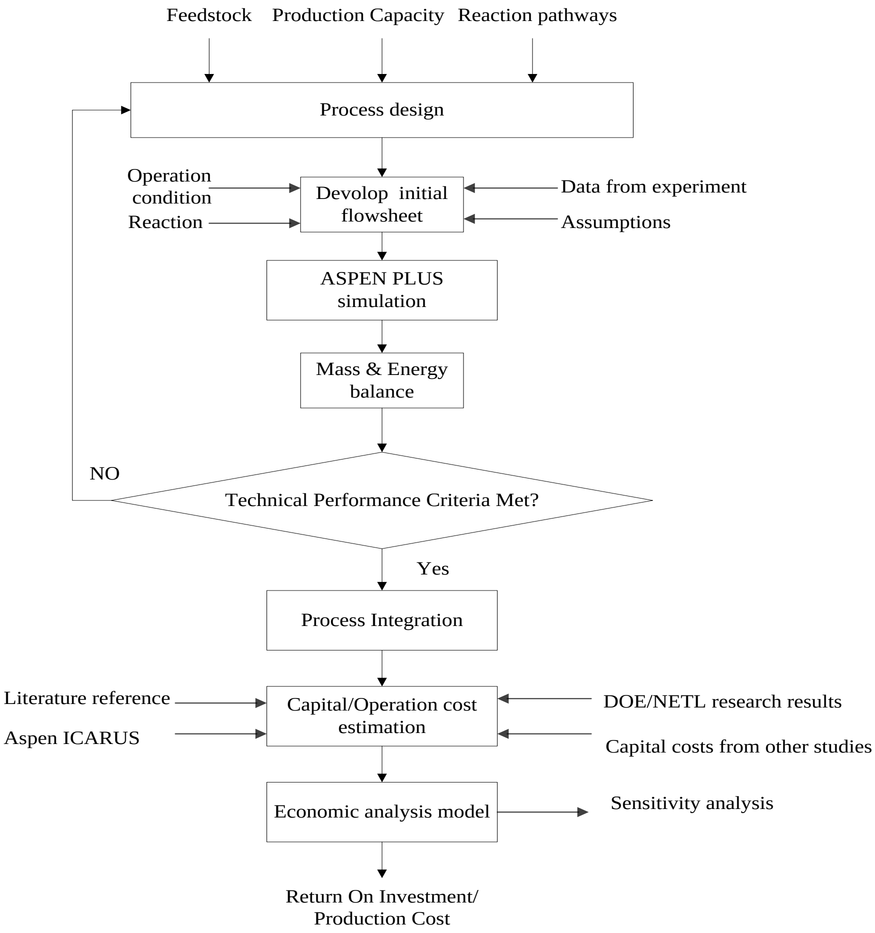


Figure 3.1 Schematic of the process design approach

Preliminary Criteria

The technology configuration options used in this thesis are reviewed and screened under three criteria. First, the technology used in the process design should be mature or commercially ready within next 5–8 years and preferably with high technology development. Since highly developed technology increases the possibility that a configuration will be performed at the large scale defined in the study.

For example, coal gasification has been commercially demonstrated at large scales for decades [62] while similar scale biomass or MSW gasification have not been proven commercially. In this case, the technology development of coal is assumed to apply for biomass in the plant design. Second, the size of facilities using biomass as feedstock should be feasible with typical agricultural productivity and within a realistic transportation and collection area. It is assumed that the reasonable and realistic collection radius should be within 100 miles and the total amount of biomass transported to the facility is around 2000 dry metric ton/day. This assumption is in accord with biorefinery plant scale study done by Tijmensen et al., Phillips et al., and Lau et al [63-65]. Third, the product produced should be compatible with current transportation fuel infrastructure, i.e., gasoline, diesel-range hydrocarbons, and SNG.

Major technologies selection

For the gasification area, non-slagging fluidized bed type gasifier with a regenerator setup is selected all through this thesis due to the relative long resident time and heat supply method. Warm gas cleanup technology developed by RTI, currently moving from commercial demonstration stage to commercial plant application, is utilized to remove hydrogen sulfide and other contaminants in the gas cleaning section in order to improve the overall process thermal efficiency. An amine-based chemical absorber/stripper configuration is chosen for carbon dioxide removal. This configuration is chosen due to data availability as compared to proprietary physical gas cleaning process such as Rectisol and Selexol process [17].

FT synthesis has been proven in operation at commercial scale for many years by Sasol. Because of more accessible data and long industrial experience, FT synthesis is the only liquid fuel synthesis option chosen with fuel synthesis and upgrading area. Other major technologies such as steam methane reforming, gas shifting and gas/steam turbine, with the first-hand data collected from industry, are considered as commercial mature and compatible with units operation all through the process design.

Major technologies not selected

The transport gasifier design, though a promising technology is not considered because of reactor complexity, unproven commercial-scale operation, and lack of public domain data. External fuel (e.g., natural gas or propane) options are not considered as heat supplier during the gasification process. Cold gas cleanup such as Rectisol and Selexol process with water quench and scrubber configuration is not considered due to large process heat loss. Tar removal process is not applied in the plant design for all the available feedstock introduced before. The primary reason is the lack of evidences and support from experiment work and also due to a low technological development in tar conversion and its inherent complexity. Coal and Biomass-to-Liquid (CBTL) is not considered because of time constraints and limited operational data. DME synthesis option is not considered because of limited commercial scale experience and incompatibility with present fuel infrastructure.

3.4 Project assumptions and nth plant economics

The process design is assumed to incorporate an “nth plant” level of implementation experience. The key theoretical assumption associated with nth plant economics is that several plants using the same technology have already been built and are operating. In other words, the assumption reflects a future in which a successful industry has been established with many operating plants and additional costs for risk financing, longer start-ups, equipment overdesign and other costs associated with first-of-a-kind plants are not included. At the very least, nth plant economics should help to provide justification and support for early technology adopters and pioneer plants.

Because equipment costs in the designated plant can been estimated explicitly from sources mentioned earlier, the nth plant assumptions apply primarily to the factored cost model used to determine the total capital investment from the purchased equipment cost and to the assumptions applied for plant financing. The nth plant assumption also applies to operating parameters, such as process uptime and start-up time. The main project assumptions for the nth plant economic analysis are listed in Table 3.1**.** A more extensive list can be found in Appendix A. These financial assumptions are consistent with assumptions used for other economic analyses done by NETL in DOE with some deviations to reflect the uniqueness of this process [66].

Examples have shown that the nth plant technoeconomic models can reasonably predict the production costs of biofuels and were able to predict the actual cost of production of these biofuels within an accuracy of ±30%. The variability in accuracy is largely a result of the variability in factored capital cost estimates used in these models.

Developing a technoeconomic model for a pre-commercial technology, such as the CTL and BTL based on steam hydrogasification described in this thesis, is based on less specific information and has to be rooted in current understanding of the state of the technology, feasible improvements at the time of the analysis, and good engineering practice. There will be more uncertainties in the cost predictions from such studies. Using the nth plant philosophy, it is assumed that the pre-commercial technoeconomic model reflects the production economics for the technology's mature future. Detailed plant technoeconomic analysis of configurations discussed will be given in the following chapters.

Table 3.1 Summary of nth plant assumption for technoeconomic analysis

Description of Assumption	Assumed Value
Discount rate	12%
Plant financing by equity/debt	30/70 for CTL plant, 45/55 for others
Plant life	30 years
Income tax rate	38%
Financing fee	3% of debt
Interest rate for debt financing	7.5% annually
Term for debt financing	15 years
Grace Period on Principal Repayment	1 year
Depreciation schedule	7-Year MACRS schedule [7]
Construction period (Spending schedule)	3 years (8% Y1, 60 Y2, 32% Y3)
Plant salvage value	No value
Plant annual operation hours	7884 hour

3.5 Methodology for economic analysis

Capital costs were estimated using a variety of resources. For sub-processes that utilize well-developed technologies and can be purchased as modular packages (e.g., gas cleanup unit and steam turbine), an overall package cost was used instead of the sum of costs for individual pieces of equipment.

Costs for common process equipment (e.g., tanks, drums, pumps, and simple heat exchangers) were estimated using Aspen ICARUS V7.2 costing software, which uses the Q1 2008 cost basis. Unit operations that are specific to thermochemical liquid fuels production (e.g., gasifier, SMR, and FT synthesis reactor) were priced through DOE studies, literature and other technical report. It is important to note that capital cost estimates for pre-commercial technologies like the circulating fluidized gasifier was derived from current estimation for first-of-its-kind fabrications, which should be higher than costs for mature nth plant technologies.

The original (base) purchased equipment costs reflect the base case for equipment size and cost year. Equipment sizes required for the process may vary from the original base case, requiring adjustment of the equipment costs. Instead of re-pricing equipment after minor changes in size, exponential scaling is applied to adjust the purchased equipment costs using Equation 3-1.

$$\text{Scale-Up Equipment Cost} = \text{Base Equipment Cost}\left(\frac{\text{Scale-Up Capacity}}{\text{Base Capacity}}\right)^{n} \tag{3-1}$$

The characteristic scaling exponent, *n*, is typically in the range of 0.6 to 0.7 for process equipment. The sizing parameters are based on a characteristic of the equipment related to production capacity, such as inlet flow for a process vessel or heat transfer duty for a heat exchanger. Equation 3-1 assumes that all other process parameters (pressure, temperature, etc.) remain constant relative to the base case.

For multiple unit operations that operate in parallel or in trains, a train cost factor is applied. The reason for the factor is because those units share some piping, electrical, and other installation costs. It is applied as shown in Equation 3-2, where m is the number of units in the train and n is the train factor with a value of 0.9.

$$\text{Cost}_{\text{train}} = \text{Cost}_{\text{unit}} * m^n \qquad (3\text{-}2)$$

Scaling exponents were determined from the following sources:

1. Vendors' estimates of scaling exponent or inference from vendor quotes when multiple quotes were available for equipment of various processing capacities.
2. Development of correlations by multiple estimates from Aspen ICARUS software.
3. Standard reference from published sources such as Harry [67], Peters, Timmerhaus and West [68], and Perry et al [69].

When cost data were not available in 2010 dollars, costs were adjusted with Chemical Engineering's (CE) Plant Cost Index (PCI) using Equation 3-3 [70].

$$\textit{Corrected Equipment Cost} = \textit{Base Equipment Cost}\left(\frac{\textit{2010 Cost Index Value}}{\textit{Base Year Cost Index Value}}\right) \qquad (3\text{-}3)$$

The CE indices used in this study are listed and plotted in Figure 3.2. The index data show a sharp increase after 2003 due to increases in global steel demand and a dip in 2009 due to the global recession. The August 2010 index was used for 2010 since that was roughly the time when most of the recent cost quotes were obtained.

Once the Total Purchased Equipment Costs (TPEC) were determined, scaled, and time-corrected, an equipment installation factor was applied to estimate the Total Installed Cost (TIC) for the equipment or process unit including associated piping, instrumentation and controls, electrical systems, buildings, yard improvements, and direct labor shown in Equation 3-4, where $f_{Installation}$ is the installation factor.

Year	Index	Year	Index
1990	357.6	2001	394.3
1991	361.3	2002	395.6
1992	358.2	2003	402.0
1993	359.2	2004	444.2
1994	368.1	2005	468.2
1995	381.1	2006	499.6
1996	381.7	2007	525.4
1997	386.5	2008	575.4
1998	389.5	2009	521.9
1999	390.6	2010 (August)	549.5
2000	394.1		

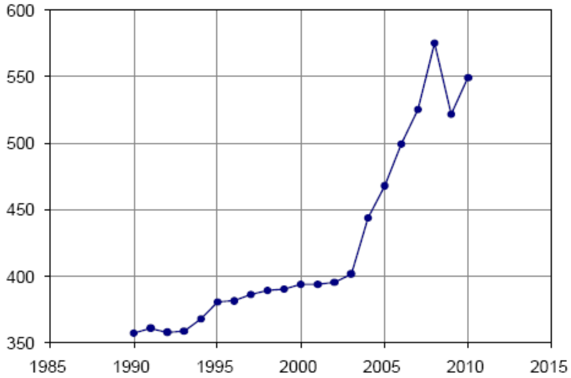


Figure 3.2 Chemical Engineering’s Plant Cost Index data

$$Total\ Installed\ Cost\ (TIC)\ =\ f_{Installation}\ *\ Total\ Purchased\ Equipment\ Cost\ (TPEC) \tag{3-4}$$

If possible, specific installation factors from vendor quotes or published data were applied to equipment or process units to estimate the TIC. When specific data was not available for equipment or package units, a default installation factor of 2.47 was applied to determine TIC. This value is based on the direct cost factor method presented by Peters, Timmerhaus, and West [68] for a solid-fluid processing plant.

Once the scaled TIC was determined, Indirect Cost (ID) was added to determine Engineering, Procurement and Construction (EPC) cost. Process contingency was then applied to the EPC cost which gave the Total Plant Cost (TPC). Finally, Total Required Capital (TRC) was determined by adding the Working Capital (WC), Financial Cost (FC) and TPC. TRC structure explanation and cost estimation methodology used in this thesis is given in Table 3.2 [67-69].

Table 3.2 Content and methodology for capital cost estimation

Parameter	Content	Method
Total Purchased Equipment Cost (TPEC)	Total purchased equipment cost	Aspen Icarus process evaluator, literature
Total Installed Cost (TIC)	Total installed equipment cost	Aspen Icarus process evaluator, literature
Indirect Cost (ID)	Construction and engineering & supervision cost, legal and contract fee	39% of TIC
Engineering Procurement and Construction (EPC) cost	Sum of TIC and ID	TIC+ID
Contingency	Compensate for uncertainty in cost estimates	20% of EPC
Total Plant Cost (TPC)	Sum of EPC and contingency	EPC + Contingency
Working Capital (WC)	Start-up cost, money tied up in receivables, payables and inventories	15% of EPC
Financial Cost (FC)	Financing fee and interest payment during construction	3% financing fee, 7.5% interest rate
Total Required Capital (TRC)	Sum of TPC, WC and FC	TPC+WC+FC

TRC along with plant operating expenses serves as the basis for the discount cash flow analysis. The discount cash flow analysis then yields the corresponding fuels production cost, the primary metric by which the steam hydrogasification based thermochemical process can be compared with other alternate fuel production designs. A detail explanation of capital structure and content is given in Appendix A.

It should be noted that for some technical mature units in current operating commercial coal gasification plants, such as coal handling and preparation, steam methane reforming and FT synthesis, unit capital cost acquired from DOE publications and other source with first hand data was used. In this thesis, a major study from DOE “The cost and performance baseline for fossil energy power plants, Volume 1: Bituminous coal and natural gas to electricity” [71] was highly relied on the cost estimation methodology for fossil energy plants. For the equipment cost quoted from these sources, costs including installation, indirect cost and process contingency were already combined and integrated in each operation unit. In this case, rather than using the capital cost estimation method mentioned in Table 3.2, the TPC was the sum of the each operation unit and TRC was determined by adding the rest of the capital expense.

4. Synthetic fuels and power production from coal

One current coal-based application of CE-CERT process is to produce diesel fuel and naphtha or generate power via IGCC. The large reserves of coal can improve the energy security greatly if production of syngas from coal, and successive conversion of the syngas to synthetic fuels via FT process. The CTL technology has received a lot of current interest [72-75] and is widely accepted as a possible pathway to avoid the potential energy crisis and any irrevocable impairment to the climate if the carbon is captured and securely stored in geological media or used for enhanced oil recovery [76].

In this chapter, effects of feed condition, gasification temperature and pressure on process performance are discussed while optimum gasification conditions are determined based on process simulation. Detailed CTL plant design, modeling and process economic analysis are employed to give comprehensive assessments of the plant performance as well as economics.

4.1 Process simulation methodology

Aspen Plus version 7.2 was used in the process simulation in this thesis. The gaseous and liquid components that are important for the process were included as distinct molecular species using Aspen Technology’s component properties database. The thermochemical conversion design requires processing of three different phases of matter (solid, liquid, and gas phases). Therefore, no single physical property package was sufficient to describe the entire plant from feed to products. Different property packages were used within the Aspen Plus simulation to more accurately represent chemical component behaviors in specific process areas. Peng Robinson cubic equation of state with the

Boston-Mathias alpha function (PRBM) was used in the gasification and much of the downstream unit simulation due to its high accuracy in estimating properties for gas-processing, refinery, and petrochemical applications. It was recommended for modeling nonpolar and mildly polar mixtures, including hydrocarbons and light gases like: CO_2, H_2S, and H_2. Reasonable results can be expected for all temperatures and pressures. The ELECNRTL package was used to model the electrolyte species in the amine-based CO_2 removed system.

The coal, biomass, MSW and ash components were modeled as non-conventional components. The SOLIDS property option was used for the coal crushing and screening section. The enthalpy model for coal and ash, the nonconventional components, was HCOALGEN and the density model is DCOALIGT. The HCOALGEN model included a number of empirical correlations for heat of combustion, heat of formation and heat capacity. All other values used were retrieved from the Aspen plus database.

Gasification section is the core part of the process and was simulated using decomposition and gasification units based on built-in Aspen reactor blocks. The decomposition block converted the non-conventional feedstock into its basic elements on the basis of yield information using the RYIELD block and the gasification block calculated the equilibrium product gas composition under the given conditions by means of Gibbs free energy minimization using the RGIBBS block.

Separation blocks were used to simulate contaminants removal (NH_3, Hg and Cl) in most cases. The H_2S removal step using regenerable ZnO was modeled by equilibrium module along with DSRP process. A stoichiometric block was used to simulate methane reforming by defining the chemical reactions occurred in the process.

In the FT synthesis, the -(CH_2)- is a basic structure for hydrocarbons with long chain. A main characteristic regarding the performance of the FT synthesis is the process liquid selectivity. The liquid selectivity is determined by chain growth probability. This is the chance that a hydrocarbon chain grows with another -(CH_2)- group instead of terminating. A high liquid selectivity (C_{5+} selectivity or S_{C5+}) is necessary to obtain a maximum amount of long hydrocarbon chains. The FT block in the process modeling used an external model called through a FORTRAN module. This external model was empirically developed by Hamelinck et al [77], to predict the selectivity of the FT process and is given in detail in Appendix B.

The whole simulation was controlled using FORTRAN routines (calculator blocks) and design specifications to reduce the number of independent specifications in order to adjust automatically those associated variables, i.e. the dependent variables were automatically adjusted when independent input variables were modified by the calculator block or a design specification. The main functional relationships (control structures) of the simulation were: the amount of H_2O input as a function of the feedstock mass input, the amount of H_2 input as a function of the carbon in the feedstock, and H_2/CO molar ratio in the syngas. Some major operation models used in the process simulation were shown in Table 4.1.

Mass and energy balance was determined based on the process simulation results. Due to the complexity of the simulation and large amount of blocks used, user model of

HIERARCHY was used and to separate the whole operation into seven major HIERARCHY areas:

- Feedstock pretreatment (Area 100)
- Gasification (Area 200)
- Gas Cleanup (Area 300)
- Steam Methane Reforming (Area 400)
- Syngas Conditioning (Area 500)
- Fuel Synthesis and Upgrading (Area 600)
- CO_2 Removal (Area 700)
- Power Island (Area 800)

Table 4.1 Representive unit operations used in the simulation

Unit operation	Aspen plus model	Specifications
Coal crushing	Crusher	Rigorous simulation of particle size distribution
Coal particles screening	Screen	Rigorous simulation of the separation efficiency of the screen
Coal gasification	RGIBBS	Specification of the possible products: H_2O, H_2 ,Cl_2, HCl, C, CO, CO_2, CH_4, COS, H_2S, CS_2
Regenerator	Rstoic	Rigorous simulation of char and gas combustion
Solid removing	Sep	Simplified simulation of gas/solid separation
Warm gas cleanup	Absorber, Rstoic	Rigorous simulation of the H_2S, trace metal and chloride removal
SMR w/furnace	Rstoic	Rigorous simulation of gas reforming and combustion
H_2 separation	Sep, Split	Simplified simulation of gas separation and split
FTR	RYield	Empirical simulation of FT liquids distribution
HRSG	Heat exchanger	Simplified simulation of steam cycle w/ heat recovery
CO_2 capture	Absorber	Rigorous simulation of the CO_2 removal
Gas and steam turbines	Turbine	Rigorous simulation of power generation

4.2 Feed condition determination

The feed H_2O/Coal ratio is calculated on a mass basis and the feed H_2/C ratio is calculated on a mole basis in this study. Coal is made into slurry with initial H_2O/coal mass ratio of 1:1 and additional water is then injected into gasifier as superheated steam in order to reach a high carbon conversion by increasing the H_2O/coal ratio [47]. One of the key aspects of the CE-CERT process is the internal H_2 supply by means of recycling the excess H_2 produced after SMR. It must be ensured that sufficient H_2 is available for feed under the desired operating conditions. The total amount of the recycled H_2 must be higher than the total H_2 input in the SHR in terms of H_2 system self sustainability. Meanwhile, the H_2/CO molar ratio in the syngas should be at least higher than 1.0 for the fuel synthesis purpose required by the FT catalyst configuration. Figure 4.1 shows the H_2/CO ratio in the syngas as functions of H_2/C molar ratio and H_2O/Coal mass ratio after H_2 separation to SHR. The gasification temperature is set at 750 °C while the unreacted char is fixed as 17.9% of the coal carbon and is assumed independent from feed condition although it might not be the case in reality.

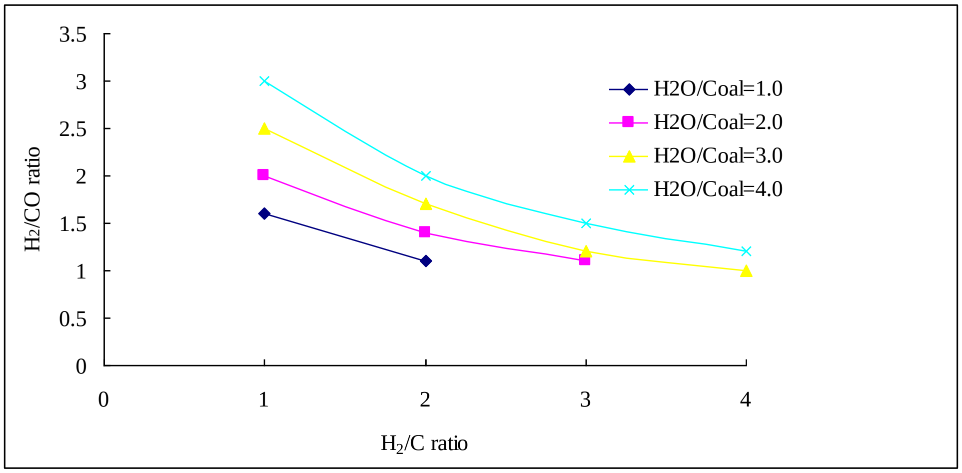


Figure 4.1 H_2/CO ratio as functions of H_2/C and H_2O/Coal input ratio

(Gasification temperature=750 °C, leftover char = 17.9%)

When H_2O/Coal mass ratio of 1.0 and 2.0 is used, the H_2/CO ratio in the syngas is either below minimum required value of 1.0 or cannot meet the system H_2 internal cycle when H_2/C molar ratio is beyond 2.0 and 3.0, respectively. In this case, such points are considered as non-eligible and not marked in the plot.

Regardless heat balance of the system, the trend of H_2/CO ratio in the syngas is pretty straight forward which decreases with H_2/C ratio and increases with H_2O/Coal ratio. This result is expected since the addition of H_2O in the gasifier promotes the H_2 generation. However, the optimum feed condition still cannot be decided from this plot from H_2/CO ratio analysis point of view alone.

Generally speaking, the H_2O/Coal mass ratio should not be so low since the reaction rate of carbon with other reactants within the gasifier, as shown early in chapter 2, is dominated by the totally amount of water inside the SHR. With same residence time, the conversion of coal into gas species is significant low if there is not sufficient water existing as gasification agent during the process. Although a higher gasification efficiency can be achieved by extending the reaction time even with less H_2O in the system, this is beyond the design of the fluidized bed type gasifier in reality and thus absolutely unpractical and uneconomic.

On the other hand, the overall H_2O/Coal mass ratio should not be so high from equipment design and plant thermal efficiency point of views. The steam takes up a lot of space inside the high temperature SHR and a larger gasifier is needed due to this volume increase which also adds the capital cost. Furthermore, the increased volume flow rate of SHR effluent, with almost steam in the gas stream, increases the scale of the downstream gas processing units. Not mention the additional operation and equipment cost. In addition, the large amount of water, either in liquid phrase or steam, is also a major challenge for the plant design in determining the heat balance. Since significant amount of superheated steam has to be generated via internal stream-to-stream heat exchangers in the heat exchanger network, the large amount of heat required by steam generation will take up most of the available process net heat and bring down the process thermal efficiency dramatically due to the heat loss in the condensation. In some cases, it may be even worse and the high H_2O/Coal input ratio makes the process not heat self-sufficient if no external heat source is found. Under this condition, the heavy heat burden caused by high water input ratio makes the process design more complicated and typically undesirable due to the process efficiency and plant economy.

In all, from reactor design, equipment cost and process thermal efficiency points of view, a medium H_2O/Coal mass ratio of 2.0 is selected that covers all the concerns listed above. The corresponding H_2/C molar ratio of 1.0 is determined to ensure an overall H_2/CO ratio above 2.0 is achieved after H_2 cycle back to SHR. This high H_2/CO ratio allows more excess H_2 for power generation if iron based catalyst is used while also meets the cobalt based catalyst requirement in FT synthesis.

It should be noted that the CH_4 formation and yield in the gasifier is simulated based on phase and chemical equilibrium by means of minimum Gibbs free energy. The H_2/CO ratio in the syngas after H_2 separation might be lower since the total amount of CH_4 reacted in the SMR may be varied since the gasifier is hardly to achieve the chemical equilibrium at the design temperature in reality. This is another reason for selecting a relative low H_2/C molar input ratio and high H_2/CO input ratio as feed condition in order to enable the system more tolerant and flexible with this potential difference between the simulation and real situation.

4.3 Temperature and pressure effect

As mentioned earlier, the SHR are modeled on the basis of Gibbs free energy minimization using built-in Aspen Plus reactor modules which calculates the entire potential products inside the reactor based on the phrase and chemical equilibrium. This method is especially suitable for estimating product composition in gasification process since there are hundreds of combinations and chemical reactions during gasification and it is hard to define each of the reaction with stoichiometric expressions.

CH_4 production is favored at high pressures and the process is generally operated at temperatures ranging from 750 ℃ to 1000 ℃ [17]. Based on preliminary experiment results, high gasification temperature leads high gasification efficiency while results in little leftover char available for heat generation. The fraction of leftover char is assumed as a function of gasification temperature only although this may not be the case in reality since carbon conversion efficiency also changes with H_2, H_2O or steam in the feed. The char production ratio at gasification temperatures ranges from 700-900 ^oC is given in the Table 4.2 below. The gasification pressure is set to be 27.6 bar (400 psi).

Table 4.2 Char production ratio as a function of gasification temperature

Temperature (^oC)	Char (%)
700	21.7
750	17.9
800	13.8
850	9.2
900	4.4

The purpose of this section is to estimate the gas composition using equilibrium model, especially the carbonaceous gases, and also to show the effect of the pressure and temperature on it. An optimum feed condition of H_2O/Coal mass ratio of 2.0 and H_2/C molar ratio of 1.0 is utilized all through the analysis. The effect of temperature on the carbonaceous gases is given in Figures 4.2.

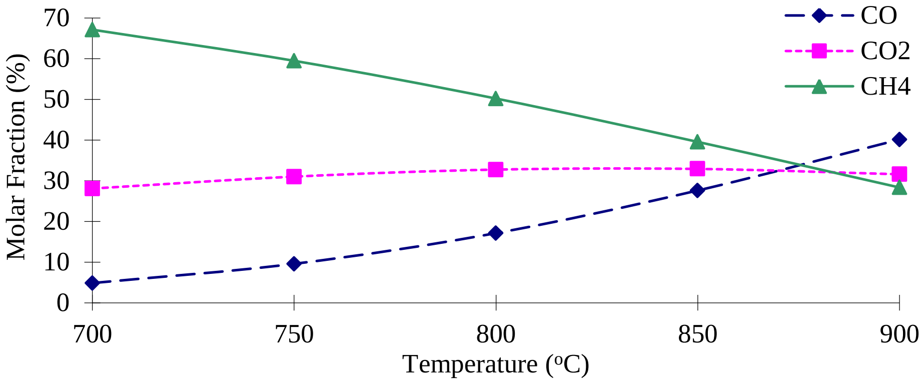


Figure 4.2 Effect of temperature on the equilibrium composition of carbonaceous SHR product gases with coal feed (H_2O/Coal = 2, H_2/C = 1, P = 400 psi)

All the carbon in the product gases was present in the form of CO, CO_2 or CH_4 and no hydrocarbon with carbon number higher than one is considered. The mole percentages presented here were calculated using these three species only. Other components present in the product gas such as unreacted steam and H_2 were not included in the product mole percentage calculations. Hence, the mole percentages presented have been normalized to 100 % based on the three carbon containing species, CO, CO_2 and CH_4.

It can be observed that the concentration of CH_4 decreases with temperature increase whereas the concentration of CO increases. This is expected since the hydrogenation of carbon which is also the primary CH_4 generation reaction is mildly exothermic. The CO producing reactions such as the Reversed Water Gas Shift (RWGS) reaction and carbon steam reaction are endothermic, which promote the CO formation with temperature increase. It is found out that CO_2 has a maximum concentration at the temperature of roughly 850 ℃ and this can be attributed to the WGS reaction which is reversed under high temperatures. Since these results are based on assumptions of equilibrium, it may not reflect the fact in reality especially for the reactions in CFB gasifier. However, equilibrium data can be used for experiments and equipment design to understand and evaluate process behavior.

The effect of pressure on the equilibrium composition of SHR product carbonaceous gases is given in Figure 4.3.

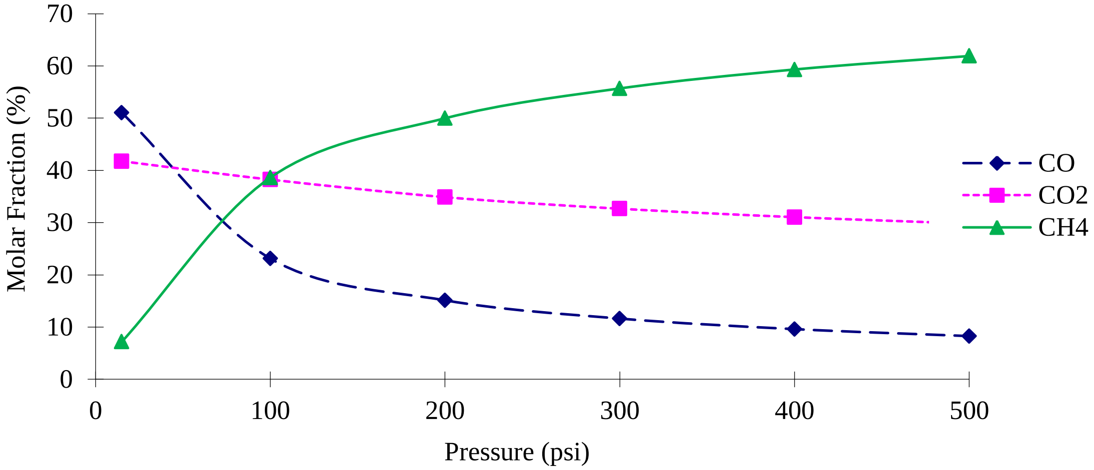


Figure 4.3 Effect of pressure on the equilibrium composition of carbonaceous SHR product gases with coal feed (H_2O/Coal = 2, H_2/C = 1, T = 750 ℃)

The temperature was set at 750 ℃ for the calculation and the H_2O and H_2 feed ratio is the same as used before. The molar concentrations of CO, CO_2 and CH_4 in this plot have also been normalized to 100, the same with the temperature effect analysis. The results show that pressure plays considerable role on determining the product gas composition, especially at lower pressures. From the perspective of the CE-CERT process, it is desirable to operate the SHR at high pressure (400 psi or higher) since the formation of CH_4 is favored at high pressure. This high operation pressure also reduces gas processing equipment costs as well as compression.

Based on the analysis results, it is recommended that the steam hydrogasification should take place at relative low temperature and high pressure in order to promote CH_4 formation. For the gasification pressure, 400 psi is selected which is consistent with most commercial applied gasification process and also the same operation pressure as required by the downstream FT synthesis. In terms of temperature, long residence time and low carbon conversion results if the temperature is too low and this is totally undesirable. While at high gasification temperature, sufficient amount of heat must be supplied to SHR. Theoretically, there has to be an existing temperature that satisfies both CH_4 yield and heat requirement. This optimum operation will be found out and discussed in the next section.

4.4 Optimum gasification temperature determination

As mentioned before, the SHR is the initial and most critical step in the whole process. The ability to supply the necessary process heat to the SHR is of vital importance in determining the plant performance and economy, especially at large scales.

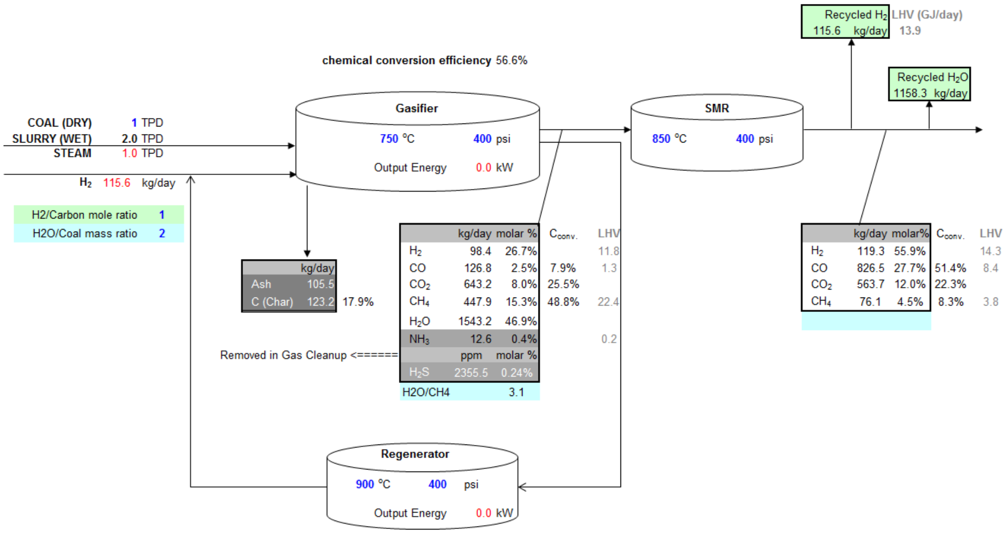
A CFB with a regenerator setup is proposed by combusting the leftover char from the SHR. The requirement for satisfactory operation is that the heat duty of both the gasifier and the regenerator equal to zero under the specified condition. The total available char for heat generation in the regenerator decreases with SHR temperature increase as shown in Table 4.2. It is recognized that the heat requirement of the gasification could not be met just with char combustion and would require additional fuel if the gasification is operated at high temperature. This section discussed the heat balance of the SHR with regenerator under different operation temperature.

A constraint was incorporated into this analysis requiring the process to remain "energy neutral." That is, the process was designed without allowing for additional usage of coal, natural gas, or electricity for heat generation. If the heat requirement of the process cannot be met through char combustion alone, the following options were considered:

1. Additional coal could be used to heat the process. This option increases the coal processing rate beyond design capacity.
2. External fossil fuel (natural gas) could be added directly to provide additional fuel.
3. Some syngas could be diverted from synthetic fuels production for heat production. This option makes the design energy self-sufficient but lowers the overall product yield.

It was decided that the first and second options would not be considered and the third option would be used to satisfy balance-of-process heat requirement.

To make a fair comparison, all input streams enter the SHR at the same design temperature at pressure of 400 psi. An optimum feed condition of H_2O/Coal mass ratio of 2.0 and H_2/C molar ratio of 1.0 is employed as determined early. Coal is made into slurry with solid loading of 50 wt% and the rest of water enters into the SHR as superheated steam. The regenerator operation temperature is determined by SHR temperature and is set as 150 ^{o}C higher than it. Due to the temperature difference of the regenerator, it is assumed that the hot flue gas exits the regenerator goes through a gas-to-gas heat exchanger and preheats the inlet air to design outlet temperature. The flue gas temperature at the exit of heat exchanger is set to be 500 °C for all cases discussed here. The coal feedstock input flow rate is normalized as 1 dry tonne per day for analysis purpose. The SHR product gas goes to the SMR block after contaminants removal and the CH_4 in rich gas is then converted into syngas. Part of H_2 is then split and recycled back to SHR as feed and all the steam is condensed before the determination of the syngas composition. Mass and energy balance of SHR and regenerator along with overall available syngas for energy production deducting the H_2 used in SHR is shown in Figure 4.4, 4.5 and 4.6 with corresponding gasification temperature of 750 °C, 800 °C and 850 °C, respectively.



	kg/day	
Ash	105.5	
C (Char)	123.2	17.9%

	kg/day	molar %	$C_{conv.}$	LHV
H_2	98.4	26.7%		11.8
CO	126.8	2.5%	7.9%	1.3
CO_2	643.2	8.0%	25.5%	
CH_4	447.9	15.3%	48.8%	22.4
H_2O	1543.2	46.9%		
NH_3	12.6	0.4%		0.2
	ppm	molar %		
H_2S	2355.5	0.24%		
H2O/CH4		3.1		

	kg/day	molar%	$C_{conv.}$	LHV
H_2	119.3	55.9%		14.3
CO	826.5	27.7%	51.4%	8.4
CO_2	563.7	12.0%	22.3%	
CH_4	76.1	4.5%	8.3%	3.8

Figure 4.4 Mass and energy balance of SHR & regenerator with syngas availability

(Gasification temperature = 750 °C, Regenerator temperature = 900 °C)

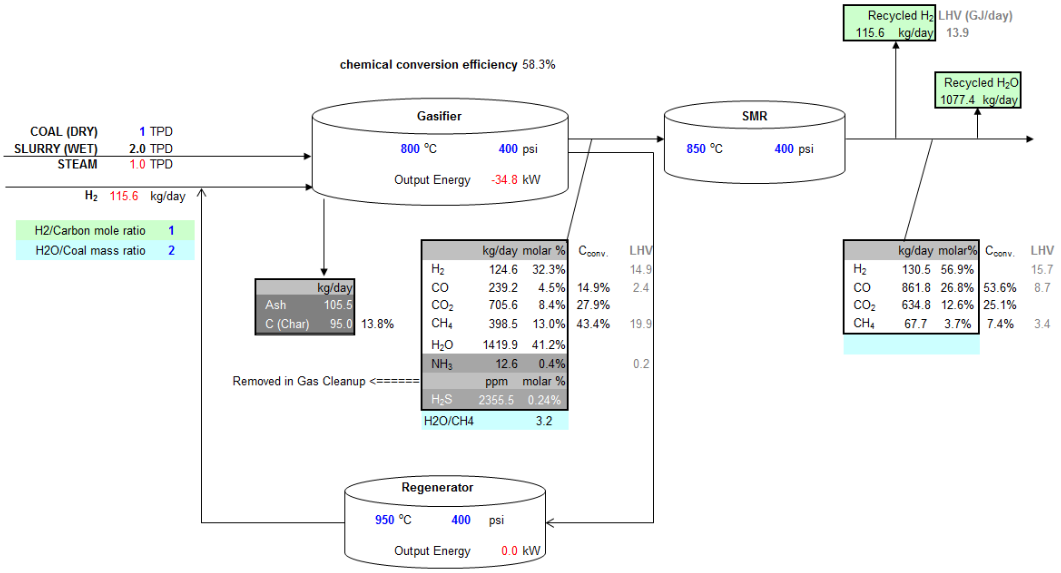


Figure 4.5 Mass and energy balance of SHR & regenerator with syngas availability

(Gasification temperature = 800 °C, Regenerator temperature = 950 °C)

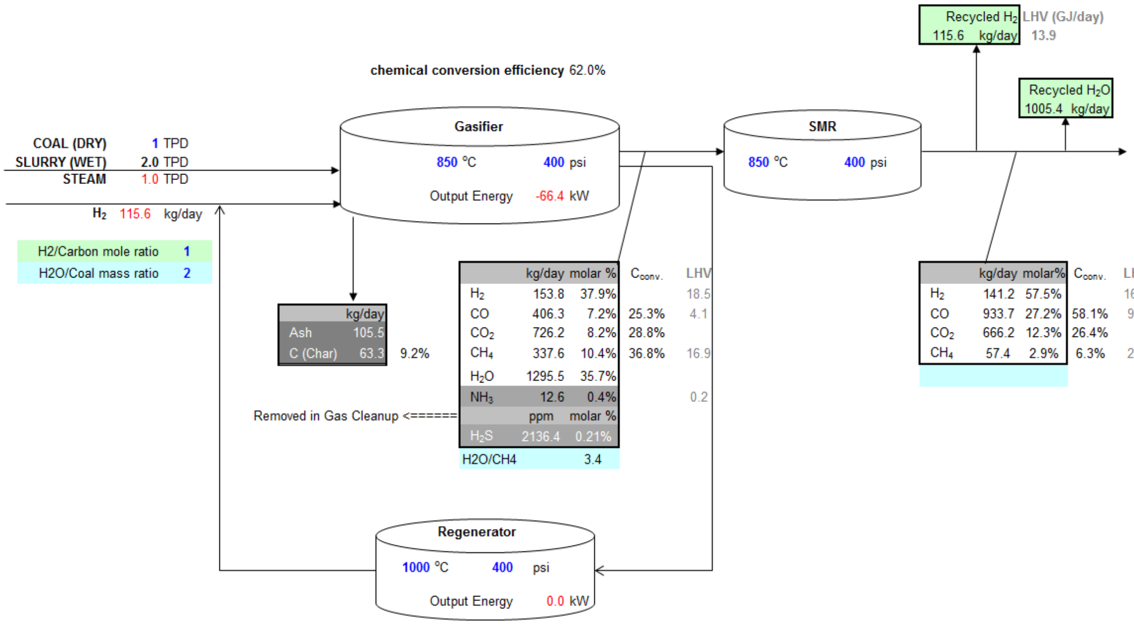


Figure 4.6 Mass and energy balance of SHR & regenerator with syngas availability

(Gasification temperature = 850 °C, Regenerator temperature = 1000 °C)

The composition of the gas composition at SHR effluent and syngas composition after H_2 separation are also given in the Figures. The results are initially generated by Aspen Plus simulation and then exported to an in-house developed user interface automatically linked by Microsoft excel macro VBA code.

The composition of carbonaceous gases, CO, CO_2 and CH_4, under different SHR operation temperature are consistent with the trend shown early in Figure 4.2. The H_2 yield in the SHR increased with operation temperature and this is expected since more coal carbon is involved in the chemical reactions with water at high temperature which also leads to less water in the SHR effluent.

It is noticed that the heat duty of SHR is not equal to zero when the operation temperature is at 800^oC and 850^oC, respectively. The negative value in the gasifier block means that specific amount of process heat has to be introduced into the SHR in order to achieve the zero net heat duty goal. Based on the simulation results, the combustion of leftover char in the regenerator can cover the SHR heat requirement when the gasification temperature is 750 ^oC while extra heat of 34.8 and 66.4 kw has to be added to the SHR by combusting other types of fuel when temperature is 800 ^oC and 850 ^oC, respectively.

The heating value of the bituminous coal used here is 26.2 GJ/ton (LHV, dry basis) and the results from Aspen Plus are directly imported into an excel worksheet and all conversion efficiencies are calculated using excel spreadsheet. The different efficiency values calculated using the simulation results are listed below.

Carbon Conversion Efficiency (CCE) of SHR = $1- (C_{Char} + C_{CO2})/C_{Coal}$

Where, CCE of SHR is calculated based on the fraction of coal carbon converted into hydrocarbons and CO. The operation condition, conversion efficiency and heat balance is summarized in Table 4.3. The high gasification temperature allows more carbon to be converted into gas product which results in a higher CCE in the SHR. On the other hand, the higher SHR CCE also leads to high energy content in the gas product after H_2 separation as also can be seen in the table. The energy content in the syngas is 306.3 kw, 321.2 kw and 338.4 kw for SHR operation temperature at 750^oC, 800^oC and 850^oC, respectively. Since the leftover char combustion alone cannot cover the SHR heat requirement when the gasifier is operated at high temperature, it is assumed that such energy deficit is then deducted from the syngas heating value and the energy content in the rest of syngas is finally considered as net energy which is the syngas diversion design option mentioned earlier. The net energy content of syngas after the energy compensation is 306.3, 286.4 and 272.0 at gasification temperature of 750^oC, 800^oC and 850^oC, respectively.

Table 4.3 Process conversion efficiency and heat balance of SHR w/ regenerator

Gasification temperature (^oC)	750	800	850
Leftover char (%)	17.9	13.8	9.2
CCE of SHR	56.6%	58.3%	62.0%
Gasifier net heat duty (kw)	0	-34.8	-66.4
Energy Content in Syngas (LHV, kw)	306.3	321.2	338.4
Net energy in syngas (LHV, kw)	306.3	286.4	272.0

Although the energy content of the syngas is 10.5% higher at gasification temperature of 850^oC than that of 750^oC, 11.2% less net energy results due to the large heat load in the gasifier. Based on the heat balance shown above, the SHR is more desired to operate under relative low temperature in order to have more available syngas for downstream usage.

Based on the mass and energy analysis in all cases, it is indicated that a lower SHR operation temperature is favored from the net syngas energy content point of view. While in reality, the syngas diversion option increased the downstream gas cleanup and SMR equipment cost greatly since it is a huge waste to burn the valuable clean syngas instead of producing fuels from it.

Gasification temperature under 750 ^{o}C is not considered in this study due to the long resident time to reach the chemical equilibrium as well as the tar and heavy hydrocarbon formation in such low temperature. After all, the gasification temperature is finally set at 750 ^{o}C and this temperature will be used all through the plant design in all the scenarios in this thesis. Again, there might be other types of fuel source (e.g. natural gas) to supply the heat needed in the gasification at high temperature but such options are beyond the scope of this thesis and will be not considered.

4.5 Simulation results of SHR with regenerator

The simulation was performed based on a demonstration pilot plant design capacity with feed rate of 20 dry Tonne Per Day (TPD) of bituminous coal. Block flow diagram of the SHR & regenerator setup under optimum gasification temperature and pressure of 750 ^{o}C and 400 psi is schematically shown in Figure 4.7 with process stream table given in Table 4.4 derived from simulation results.

The simulation was performed based on a pilot plant design capacity with feed rate of 20 dry ton/day of bituminous coal. The regenerator burns 17.9% of leftover char from gasifier in the presence of 20% excess air (based on the stoichiometric requirement) to ensure complete oxidation of char components.

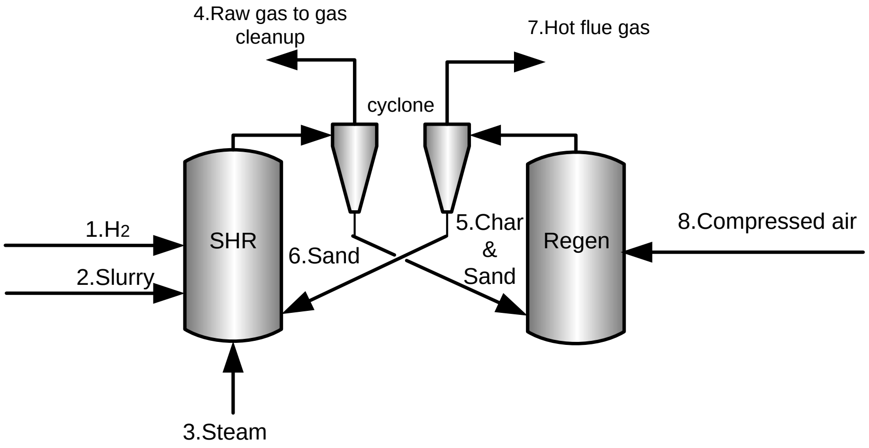


Figure 4.7 Schematic flow diagram of 20 TPD SHR w/ regenerator

Table 4.4 Stream tables of pilot plant simulation results

	Stream Number							
	1	2	3	4	5	6	7	8
Temp, ℃	324	220	738	750	750	900	900	742
Pressure, psi	400	400	400	400	400	400	400	400
Mass flow, ton/day	2.3	40.0	20.0	57.8	347.2	342.6	36.3	33.8
H_2	2.3			2.0				
CO				2.5				
CO_2				12.9			9.0	
CH_4				9.0				
H_2O (l)		20.0						
H_2O (g)			20.0	30.9				
NH_3				0.3				
H_2S				0.3				
O_2							1.3	7.9
N_2							26.0	26.0
Sand					342.6	342.6		
Char					2.5			
Ash					2.1			
Coal		20.0						

Within the gasifier, feedstock is thermally deconstructed to a mixture of syngas components (CO, H_2, CO_2, CH_4, etc.), tars, and solid char containing residual carbon and coke deposited on the sand. It is assumed that all tars will be decomposed in the expand section of the gasifier due to the long residence time and high steam concentration.

Cyclones at the exit of the gasifier separate the char and sand from the syngas. The solids flow to the regenerator where the char is burnt with air in a fluidized bed, resulting in sand temperatures greater than 900 ℃. The hot sand and residual ash is carried out of the regenerator by the effluent gases and separated using a pair of cyclones. The first cyclone captures sand while the second cyclone captures ash and sand fines. Hot sand flows back into the gasifier, completing the gasification loop. The hot flue gas from the regenerator is utilized for preheating the inlet cold air. Ash and sand fines are cooled, moistened to minimize dust, and removed as waste.

The operation of the gasifier and regenerator resembles that of a Fluid Catalytic Cracker (FCC), a common oil refinery unit operation. Like an FCC, the gasifier and char combustor will remain in heat balance by producing (gasifier) and burning (regenerator) enough char to satisfy the energy requirements of the operation. The amount of char formed in the gasifier is an inverse function of temperature. Unless supplemental fuel is provided to the char combustor for temperature control, the gasifier and combustor reach thermal equilibrium based on the amount of char formed in the gasifier. If the gasifier temperature is lower than the equilibrium temperature, then more char is formed and more heat is generated by char combustion. This results in more heat transfer from the combustor to the gasifier, thus bringing the gasifier temperature up toward equilibrium. The heat duty of the reactor and the regenerator are both zero under these conditions. It should be noted that no NO_x formation in the regenerator is considered due to the low combustion temperature (900^{o}C).

It is assumed that the full cycle time for the sand circulating between the SHR and regenerator is 30 second and the total amount of sand loaded in the pilot plant gasifier is actually 330 kg regardless the sand attrition. The Aspen Plus simulations of SHR and regenerator are based on equilibrium assumptions while it must be noted that experiments conducted in laboratory or pilot scale reactors may not be under equilibrium. Simulations based on rigorous kinetic and fluid dynamic models are necessary in order to predict the real world behavior of SHR more accurately. Developing such a model is out of the scope of this thesis.

4.6 CTL plant design

In this chapter, CTL facilities with design capacity of 400 and 4000 TPD (dry basis) are designed that consists of eight major sections: (1) Coal handling and preparation; (2) Gasification island; (3) Warm gas cleanup; (4) SMR; (5) Gas conditioning; (6) FTR and upgrading; (7) Power island and (8) CO_2 capture & compression.

Coal handling & preparation (Area 100)

The Utah bituminous coal feedstock for the plant is reclaimed from the storage area, elevated by conveyors to a crusher tower, crushed to minus 3 cm and then ground to a pulverized size distribution suitable for pumping as slurry. The size reduction is typically performed by initial impact crushing followed storage in a coal silo. A vibratory feeder from this silo loads a conveyer that moves the crushed coal to a rod mill surge tank and weighing feeder. Further size reduction occurs in a rod mill where water is added to make the initial slurry which enters the rod mill discharge slurry. A trommel screen is utilized to remove course material before circulating the final product in a loop around the slurry product tank. The desired particle size in the 50 % solids final slurry should be 70 microns.

Steam hydrogasification w/regenerator (Area 200)

Coal is injected via water slurry injectors into a bed of sand. The net temperature for gas leaving the fluidized bed gasifier is 150 ℃ lower than the sand temperature which uses steam and recycled hydrogen flow for fluidizations. Each gasifier is operated at 400 psia and the additional energy source for the gasifier is recirculated sand heated by carbon combustion with air in an external fluidized bed regenerator. The sand in the gasifier is mildly fluidized and the gas bubbles leaving the fluidized bed enter a large expanded section of the vessel where larger char particles can be disengaged from the gas. This also allows the gas to have a long residence time to promote cracking of tars and higher hydrocarbons. Fine char/ash produced in the moderate-temperature gasifier reactor flows with the gas to initial stage of a two stage internally supported cyclone, where disengaged solids falls into a standpipe back into the fluidized bed. Ash particles escaping the gasifier and regenerator cyclones are cooled and collected by hot gas filters, depressurized, and conveyed to ash bins for storage and eventual removal for onsite storage or disposal.

Gas leaving the sand regenerator goes through an internal cyclone that separates entrained particles from the regenerator flue gas prior to hot gas filtration, and reduction across an expansion turbine to reduce pressure to the inlet of the combustor for excess H_2

(Iron-based case). The particulate removal device on the gasifier is a sintered metal candle configuration, operating at the modest temperature of 510°C.

RTI warm gas removal system (Area 300)

Since the whole stream needs to be reheated to 850 ℃ in the SMR, it is not desirable to clean the gas by a conventional cold gas cleanup system (e.g., Selexol and Rectisol process) which would result in a heavy heat penalty in steam demand. Instead, the RTI warm gas cleanup is employed which consists of a number of parts. Initially the gas contacts a bed containing HCl adsorbents ($NaHCO_3$). The second step is a compact transport desulfurization reactor system that utilizes a regenerable ZnO sorbent for sulfur removal and conversion into SO_2. The particle free dirty syngas is routed to the Warm Gas Cleanup (WGCU) absorber where it is contacted with the circulating ZnO sorbent to remove H_2S and COS [78]. The following reactions are believed to take place in the absorber:

$$H_2S + ZnO = ZnS + H_2O \quad (4\text{-}1)$$

$$COS + ZnO = ZnS + CO_2 \quad (4\text{-}2)$$

The mixture leaves the top of the absorber into a cyclone where the solid sorbent containing ZnS is separated from the sulfur free syngas. Part of the solids is recycled to the absorber via a standpipe. A diverter valve located in the absorber standpipe takes a slip stream of these solids and feeds it to the regenerator. Within the sorbent regenerator, the ZnS containing sorbent comes into contact with a mixture of air. The oxygen in the air then reacts with the ZnS and forms SO_2 according to the following reactions:

$$ZnS + 1.5\ O_2 = ZnO + SO_2 \quad (4\text{-}3)$$

The above reaction is exothermic, raising the temperature of the resulting mixture (~815 ^oC). A cyclone is used to separate the solids and recycled it to the absorber. The regenerator offgas containing SO_2 is heat exchanged with the incoming air mixture before sending to next step. The SO_2 concentration in the regenerator offgas is adequate for feed to a RTI Direct Sulfur Recovery Process (DSRP) where equimolar H_2 contact produces elemental sulfur and water vapor. The reaction from SO_2 to SO_3 is an exothermic reversible reaction.

$$SO_2 + 2\ H_2 = 2\ H_2O + S \qquad (4\text{-}4)$$

H_2 is provided by a slip stream from the main H_2 stream after initial separation for SHR and fuel synthesis usage. The SO_2 is converted into elemental sulfur in the DSRP reactor. The reaction is exothermic and it raises the temperature of reaction outlet mixture. The product stream from the DSRP reactor is sent to a two stage sulfur condenser unit where the elemental sulfur is condensed and separated. Heat is recovered by making low pressure steam. The offgas passes through an absorber tower where most of existing SO_3 is removed with 98 percent H_2SO_4 before discharged to the atmosphere. A third step is the use of a regenerable sorbent for fixed-bed ammonia/HCN conversion. Finally, trace elements (Hg, As, and Se) can be removed at temperatures above 280°C.

Steam methane reformer and heat recovery (Area 400)

For the conversion of the gasifier product gas to the H_2/CO ratio suitable for FT, the first step is to convert most sulfur free syngas CH_4 content to H_2 and CO by reacting the CH_4 with steam from the syngas over a nickel based catalyst contained inside a system of high alloy steel tubes. Major reactions considered in the SMR are given below.

$$CH_4 + H_2O \leftrightarrow CO + 3H_2 \quad \Delta H=206\ kJ/mol \quad (4\text{-}5)$$

$$CH_4 + 2H_2O \leftrightarrow CO_2 + 4H_2 \quad \Delta H=165\ kJ/mol \quad (4\text{-}6)$$

$$CH_4 + CO_2 \leftrightarrow 2CO + 2H_2 \quad \Delta H=247\ kJ/mol \quad (4\text{-}7)$$

$$CO + H_2O \leftrightarrow CO_2 + H_2 \quad \Delta H=-40\ kJ/mol \quad (4\text{-}8)$$

The process takes place by first contacting the syngas with a ZnO polisher/sulfur guard bed. The SMR plant operation is similar to natural gas steam reforming which normally adds steam to the natural gas to reduce the SMR inlet to 22% CH_4 where the gasifier gas has 10%-15% CH_4 content when entering the SMR. The reforming furnace is strongly endothermic, with energy supplied by firing a furnace section of the reformer on the outside of the catalyst tubes with C1 to C4 fuel obtained in the CO_2 removal process that is part of the FT liquids synthesis unit.

Hydrogen separation and recycle (Area 500)

After leaving the reformer unit, the reformed gas is cooled in the syngas heater and then flows into a final series of heat exchangers that cool the reformed gas down to 38°C to condense the remaining H_2O before entry into the Pressure Swing Adsorption (PSA) process.

The PSA process is utilized for hydrogen separation to provide the necessary enriched hydrogen stream for recycle to the gasifier while providing the ability to adjust the reformed gas to design H_2/CO inlet ratio based on the catalyst used (1.0 for iron catalyst and 2.0 for cobalt catalyst). Recycled hydrogen (more that 98% purity) is boosted and used in the gasification system. The rest of the hydrogen is split in the streams for the FT unit, the supply for DSRP and any process excess used for the combustor/HRSG.

Synthesis and upgrading (Area 600)

Commercially deployed catalysts are used in FTR with desired H_2/CO ratio. To ensure that the hydrocarbon product distribution leans toward the production of diesel fuel, the value of chain growth factor alpha should be at least 0.85 and preferably greater than 0.9. The reactor operating temperature needed to achieve a chain growth value of 0.9 is approximately 220 ℃. To comply with reported requirements, the sorbent removes sulfur to approximately 200 ppb. In addition to sulfur, halides are also removed by the sorbent. Syngas contaminant level requirements for FT synthesis are shown in Table 4.5 [33].

Table 4.5 FT synthesis gas cleanliness requirements

Contaminant	**Tolerance Level**
Sulfur	0.2 ppm (200 ppb)
Ammonia	10 ppm
HCN	10 ppb
Halides	10 ppb

The gas bubbles up through the reactors that are filled with liquid hydrocarbons in which are suspended catalyst particles. Reaction heat is removed via heat exchangers immersed in the slurry bed to maintain the isothermal conditions that promote high synthesis gas conversions. Saturated steam generated in the FTR is then superheated and expand through a turbine for power generation. The liquid medium in FTR enables rapid heat transfer to the heat exchangers which allows high syngas conversion in a single pass through the reactor.

All exiting effluent is cooled to 35 ℃ and the liquid water and hydrocarbons are separated in a gas/liquid knock-out separator. Unconverted syngas is sent to the SMR furnace as fuel after CO_2 removal. Heavy product that is non-volatile under reaction conditions is removed from the reactor and separated from the catalyst. The raw FT products consisting of crude naphtha, crude middle distillate, and crude wax are sent directly to product upgrading.

The raw naphtha and middle distillate is sent to a hydrotreating unit to saturate the olefins that are produced in the FTR. The wax material is sent to a hydrocracking unit where the wax is converted into hydrocarbon gases (12.2 wt%), naphtha (26.1 wt%) and diesel fractions (61.7 wt%) [79].

CO_2 removal and compression (Area 700)

The FT tail gas containing light hydrocarbon gases, unconverted H_2 and CO and CO_2 is processed in an amine unit to remove the CO_2 that is inert and takes up space in the slurry FTR. The CO_2 removal process utilizes the Methyldiethanolamine (MDEA) solvent. The amine solution is costly, and measures are taken to conserve the solution during operations. As the presence of CO causes amine degradation in the form of heat stable salts, an amine reclaimer is included in the process. Also, additional water wash trays are included in the absorber tower to prevent excessive solvent loss due to vaporization. From the FT upgrading processor gas enters the absorber tower where it is contacted counter-currently with a lean aqueous amine solution. CO_2 in the feed is removed by the circulating lean amine gas. The rich amine from the stripper is pumped to a stripper column where the amine is regenerated by counter-current contact with CO_2 vapors generated in a steam-heated reboiler to remove the product CO_2 by fractionation. Regenerated lean amine is then cooled and sent back to the amine absorber tower.

To maximize the carbon recovery, a second CO_2 removal process, Fluor Economine FG Plus CO_2 recovery process, is utilized on the reformer heater stack flue gas to remove CO_2 resulting from reformer furnace, gasifiers sand regenerator and combustor [80]. The technology uses a formulation of Monoethanolamine (MEA) and a proprietary inhibitor to recover CO_2 from the flue gas. This process is specific designed to recover high-purity CO_2 from low-pressure streams that contain oxygen, such as flue gas from coal-fired power plants, gas turbine exhaust gas, and other waste gases.

The cooled flue gas from the reformer heater enters the bottom of the CO_2 absorber and flows up the tower countercurrent to a stream of lean MEA. The lean solvent enters the top of the absorber, absorbs the CO_2 from the flue gases and leaves the bottom of the absorber with the absorbed CO_2. A Solvent stripper is used to separate the CO_2 from the rich solvent feed exiting the bottom of the CO_2 Absorber. Most of the CO_2 in the stack gas is absorbed into the lean solvent, and the remaining gas leaves the top of the absorber section and eventually vented to the atmosphere. This, combined with the syngas CO_2 removal, results in an overall CO_2 capture of up to 90%. The uncondensed CO_2-rich gas is then delivered to the CO_2 product compressor.

In the compression section, the CO_2 is compressed to 15.2 MPa (2,215 psia) by a five-stage intercooled centrifugal compressor. The discharge pressures of the stages are balanced to give reasonable power distribution and discharge temperatures across the various stages as shown in Table 4.6.

Power consumption for this large compressor was estimated assuming an adiabatic efficiency of 75%. During compression process, the CO_2 stream is dehydrated to a dewpoint of -40 °C with triethylene glycol. The virtually moisture-free supercritical CO_2 stream is then ready for pipeline transport.

Table 4.6 CO_2 compressor multi-stage pressures

Stage	Outlet Pressure, MPa (psia)
1	0.32 (47)
2	1.15 (167)
3	3.80 (551)
4	12.1 (1,766)
5	15.2 (2,215)

Combustor, HRSG and power generation (Area 800)

The excess H_2 not used in the FT synthesis and upgrading is combusted in an atmospheric combustor with the flue gas streams from the steam reformer and the fluidized bed regenerator. (The flue gas from the regenerator is pressurized and expanded for power generation prior to the combustor). The exhaust from the combustor (> 950 ^oC) is used in a three-pressure level reheat steam cycle that was utilized for a CTL or IGCC plant. Major components include a HRSG, steam turbines (538 ^oC/1815 psia/ 538 ^oC / 400 psia / 65 psia), de-aerator, condenser, and integrations for heating coal slurry, regenerator air and clean syngas.

The combustor exhaust gas provides most of the heat input with large additional amounts of intermediate pressure and low level steam added that are generated in the FTR module. A small amount of steam is provided from heat exchangers in the RTI WGCU section. The exhaust gas leaves the HRSG at about 140^{o}C and either is directed to the stack or enters a CO_2 recovery unit according to the overall carbon capture ratio required in the plant design.

4.7 CTL plant performance based on steam hydrogasification

4.7.1 Plant design Cases selection

Conceptual designs of CTL plant are developed in this section to estimate the plant performance by means of process modeling. The cases discussed here are classified as conceptual studies which are comprised of defined yet unproven technology (steam hydrogasification) and non-commercialized technology (RTI WGCU) combined with currently commercial technology. Two different sizes of CTL plants with capabilities of 400 and 4000 TPD are selected to estimate the overall plant performance. The design is based on a generic location in Utah, using Utah bituminous coal as the design feedstock with chemical composition shown in Table 4.7.

Table 4.7 Composition of Utah bituminous coal

Ultimate Analysis			Proximate Analysis		
Constituent	Dry,%	As Received, %	Constituent	Dry, %	As Received, %
C	68.85	58.4	Moisture	--	15.18
H	4.74	4.02	Ash	10.55	8.95
N	1.04	0.88	Volatile Matter	40	33.93
S	1.18	1.39	Fixed Carbon	49.45	41.94
Ash	10.57	9.94			
O	11.39	13.43			

Higher Heating Value: 12,077 Btu/lb dry basis, 10,244 Btu/lb as received.

The CTL plant cases investigated are identified as the following. The carbon capture efficiency is defined as total amount of carbon fixed in the fuels (naphtha and diesel) and captured CO_2 as percentage of the overall carbon in the coal feedstock.

Case CERT-1

Synthetic fuels production from a demonstration plant size of 400 TPD (Utah bituminous coal, dry basis) with CO2 capture in the iron-based FT loop only (28.8% carbon capture).

Case CERT-2

Synthetic fuels production from a demonstration plant size of 4,000 TPD (Utah bituminous coal, dry basis) with CO_2 capture in the iron-based FT loop and sequestration off-site (65% carbon capture).

Case CERT-2B

Synthetic fuels production from a demonstration plant size of 4,000 TPD (Utah bituminous coal, dry basis) with CO_2 capture in the iron-based FT loop as well as in the power generation HRSG exhaust and sequestration off-site (90% carbon capture).

Case CERT-3

Synthetic fuels production from a demonstration plant size of 4,000 TPD (Utah bituminous coal, dry basis) with CO_2 capture in the cobalt-based FT effluent and sequestration off-site (58% carbon capture).

Case CERT-3B

Synthetic fuels production from a demonstration plant size of 4,000 TPD (Utah bituminous coal, dry basis) with CO_2 capture in the cobalt-based FT effluent as well as in the power generation HRSG exhaust and sequestration off-site (90% carbon capture).

To arrive at a cost estimate for the CTL plants with different configurations, the designs include commercial available process technology obtained from other NETL studies/verifiable sources [72]. Process technology includes commercially available technology except for the CE-CERT fluidized bed type gasifier and the RTI warm gas cleanup system with DSRP sulfur production.

Commercial technology on the synthetic fuel plant design based on the steam hydrogasification includes conventional synthesis gas cooling, commercial SMR, commercial PSA for H_2 separation, H_2 compression for recycle to the gasifier and finally FT technology with CO_2 capture in the effluent.

For all Cases, CO_2 is captured in the FT gas effluent and a single stage MDEA CO_2 removal process is included. The CO_2 from the HRSG exhaust in the power generation portion of the plant is further captured in the MEA units in Cases CERT-2B and CERT-3B which achieves an overall carbon capture and sequestration ratio greater than 90%. The criteria used in the plant design are listed in Table 4.8.

4.7.2 Case CERT-1 design and plant performance

This section is dedicated to the design and cost estimate for a 400 TPD dry basis demo-scale plant feed with Utah bituminous coal. The CERT-1 case is of interest primarily due to the small capacity and locating this size unit at a coal mining site will provide a quantity enough of liquids that can be supplied to a refiner or utilized in mining fleet operations. In turn, the operating experience gained will provide incentive for investment in a larger pre-commercial scale size plant.

Material and Energy Balance

The material and energy balance for the FT fuels production plant is based on the maximum diesel/naphtha production from 360 TPD of dry coal using an FT synthesis reactor for a ratio of 1.0 H_2/CO inlet specification (while allowing sufficient C1 to C4 production to fuel the SMR and any excess C1 to C4 to be burned along with excess H_2 in the combustor/HRSG). Ambient operating conditions are indicated in the plant design basis. One pressurized fluidized bed gasifier operating at 750 ℃ uses coal/water slurry feed pump, high steam partial pressure atmosphere and recycled H_2 to produce high CH_4 content and heating value synthesis gas.

Table 4.8 Design criteria for CTL plants based on steam hydrogasification

CTL Plant Parameter	CTL Plant Design Basis
Ambient Conditions	14.7 psia, 15 ℃, water access
Coal Feed	Utah bituminous coal slurried with water
Gasifier	CE-CERT gasifier with primary stage adjusted for 750 ℃ output heated by circulating sand in second stage regenerator at 900 ℃
Coal Feed Rate	400 TPD (CERT-1), 4000 TPD (CERT-2 and 2B, CERT-3 and 3B)
Hot Gas Temperature	~750 ℃
Gasifier Outlet Pressure	400 psia
Raw Gas Recuperator	750 ℃ down to 510 ℃, reheat clean syngas 299 ℃ up to 560 ℃
Candle Filter	Following recuperator, sulfur-tolerant
RTI Warm Gas Cleanup	Chloride guard, sulfur-tolerant
Mercury Removal	Carbon bed at 299 ℃
Desulfurization	Transport desulfurizer/regenerator part of RTI package
Sulfur Recovery	DSRP
Steam Methane Reformer	ZnO Sulfur polisher; single stage catalytic heated externally w/furnace fired by light hydrocarbons from FT unit
Hydrogen Purification	PSA
PSA Excess Gas	Fired in combustor for HRSG
Overall CO_2 Removal	65% Cases CERT-2, 90% Case CERT-2B with Fluor Econamine, 58% Cases CERT-3, 90% Case CERT-3B with Fluor Econamine,
Hydrogen Production	315 psia H_2 at recycle compressor inlet
Auxiliary Power Block	Steam turbine generator using steam from excess H_2 fired HRSG
Plant Design	Maximum diesel/naphtha production from 1:1 or 2:1 H_2:CO inlet specification
Plant Capacity Factor	90 percent

The syngas produced in the gasifier first stage (assumed to be free of hydrocarbons heavier than C4 and tar) at about 750 ℃ after passing through two stages of cyclones internal to the gasifier is cooled to 510°C by gas to gas heat exchange with clean syngas at 299°C from the downstream RTI warm gas cleanup system. The syngas passes through a candle filtration where all remaining ash & carbon containing solids are removed from the syngas. The fluidized bed gasifier bed material is sand that is continuously circulated from a fluid bed regeneration stage that burns char from the gasifier and any overhead ash/char material collected in the bottom outlet of the candle filter. This regenerator heats the sand to 900°C prior to reintroduction into the gasifier portion of the process. A dedicated air compressor provides air supply for the regenerator. The flue gas from the regenerator passes through a hot gas filter where fly ash is removed, and the particulate-free combustion is dropped in pressure across an expander to generate power. On the syngas flow side of the process, the RTI desulfurizer reduces total sulfur to less than 10 ppmv by means of an active, attrition resistant ZnO sorbent that circulates between the transport desulfurizer and its solids regenerator. The sorbent regenerator uses air to regenerate the sorbent and gives an SO_2 stream that can be processed by the DSRP into elemental sulfur.

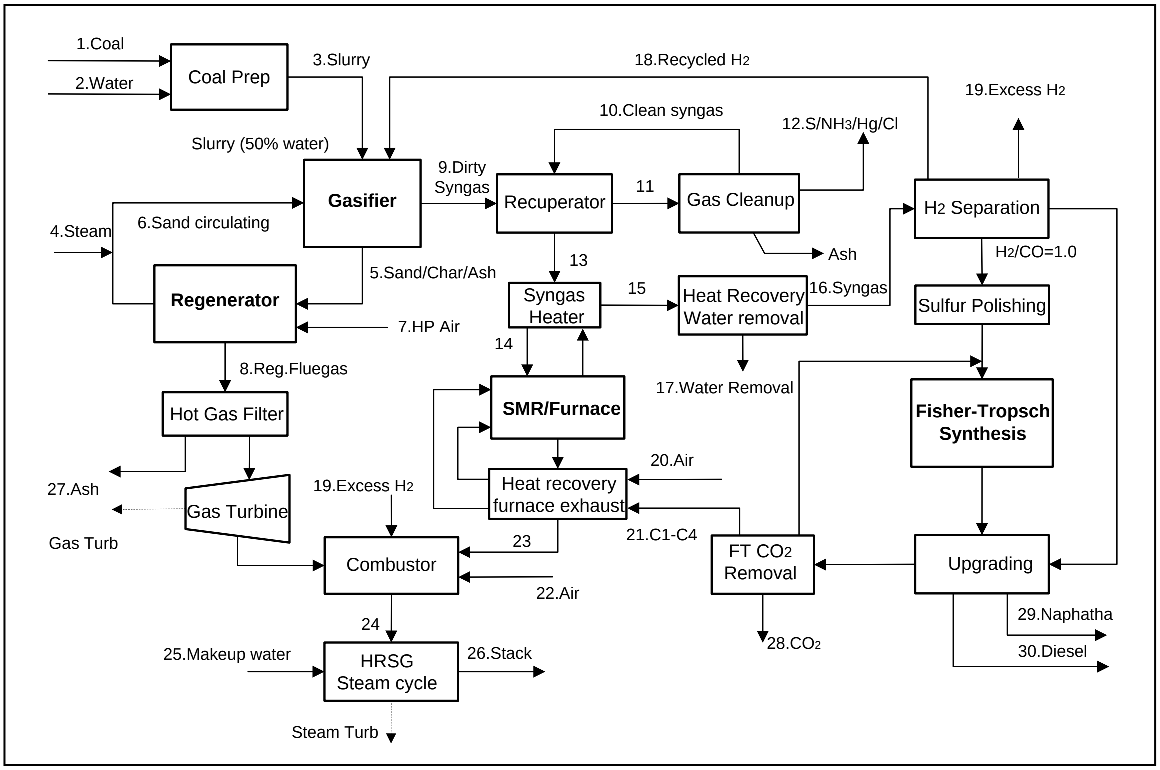
Prior to the SMR, the clean syngas is heated further above 800°C and polished of remaining sulfur species to less than 1 ppmv on the furnace side of the reformer. The gas goes through the catalytic steam reformer tubes where the 15.3% CH_4 content in the clean syngas is reduced to 2.0% CH_4 and H_2 content of the gas is increased to 51.4%. Fuel for the reformer furnace is the light hydrocarbons captured by the MDEA CO_2 removal unit featured on the FT synthesis gas loop.

The high H_2 syngas stream from the SMR enters a second gas cooler heat exchange that cools the gas from 600°C down to 38°C to condense water. High-pressure saturated steam is generated in the syngas cooler arrangement and is joined with the main steam supply. The clean high H_2 gas stream then passes through the PSA process for final separation of the H_2 needed for recycle to the gasifier, a small amount for FT liquid upgrade processing, and excess H_2 beyond that needed for the 1:1 H_2:CO inlet specification of FT synthesis. Regeneration gas from the PSA contains fuel value, and is fed to a combustor fired HRSG. Flue gases from the regenerator expander outlet and the SMR furnace provide the oxygen for the combustor-fired HRSG. The steam cycle is based on maximizing heat recovery from the gasifier cooler and HRSG, and utilizing steam generation opportunities in the FT unit. The steam turbine selected to match this cycle is a two-casing, reheat, double-flow (exhaust) machine, exhausting downward to the condenser. The HP and IP turbine sections are contained in one casing, with the LP section in a second casing. Overall process flow diagram of CERT-1 is given in Figure 4.8.

Overall performance for the entire plant of CERT-1 is summarized in Table 4.9, which includes auxiliary power requirements. For Case CERT-1, the net plant output power, after plant auxiliary power requirements are deducted, is nominally 10.62 MW_e. The overall plant thermal effective efficiency is 51.1% on an HHV basis.

4.7.3 Case CERT-2 and CERT-2B design and plant performance

Case CERT-2 is based on 10 times the coal feed and similar gasifier performance of Case CERT-1 except that heat loss in the gasifier simulation box is reduced and carbon conversion increased because of the larger size of this unit. Also, CERT-2 is configured to capture 65% amount of carbon (28.8% in fuels and 36.2% in CO_2) from the coal, compress the CO_2 and send it off-site for sequestration.



CERT-1

Stream	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
T (℃)	25	25	230	732	750	900	538	900	750	261	510	38	525	813	600
P (Bar)	1.0	1.0	32.1	30.3	30.0	30.0	31.0	30.0	30.0	28.1	29.9	1.0	28.0	27.9	27.5
m (kg/s)	4.2	4.2	8.4	4.2	85.2	84.2	6.8	7.8	12.1	11.9	12.1	0.1	11.9	11.9	11.9
Stream	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
T (℃)	204	38	324	204	16	36	16	382	993	15	133	261	35	25	25
P (Bar)	27.0	1.0	30.3	25.6	1.0	20.3	1.5	1.0	1.0	1.2	1.0	30.0	30.0	1.0	1.0
m (kg/s)	7.0	4.9	0.5	0.2	9.8	1.0	7.0	10.8	25.3	0.2	25.3	0.4	3.7	0.3	0.6

Figure 4.8 Overall process flow diagram for CERT-1 plant design with stream table

The total quantity of CO_2 captured for sequestration is 3,415 TPD from the MDEA system in the FT recycle loop. While CO_2 capture is increased from 65% to a level of 90% in CERT-2B. This case through the addition of a Fluor Economine scrubber to capture additional 25% amount of coal carbon in the flue gas which results in an overall 90% carbon capture efficiency. The total amount of CO_2 captured for sequestration is 5,518 TPD from the Fluor Econamine system and the MDEA system in the FT recycle loop.

Table 4.9 Case CERT-1 plant performance summary

Power generation (MW)	CERT-1
Gas turbine	2.39
Steam turbine	13.23
Total	15.62
FT liquid production (BBL/day)	
Naphtha	244
Diesel	435
Total	679
Auxiliary load (MW)	
Coal handling	0.05
Gasification	2.73
Warm gas cleanup	0.04
SMR	0.30
H_2 separation	0.29
FTR and upgrade	0.44
CO_2 capture and compression	0.37
Miscellaneous balance of plant	0.78
Total load	5.00
Plant performance	
Net plant power (MW)	10.62
Fuels efficiency (HHV base)	42.1%
Overall thermal efficiency (HHV base)	51.1%
Specific liquids production rate (BBL/dry metric ton)	1.87
Coal feed flow rate (ton/day)	364
Thermal input (MW)	118.2
CO_2 recovered (ton/day)	323

1 – (Diesel/Naphtha Heating Value + Electrical Power Equivalent)/ Fuel Heating Value, HHV.
2 – HHV of as-fed Utah bituminous coal is 10,244 Btu/lb.

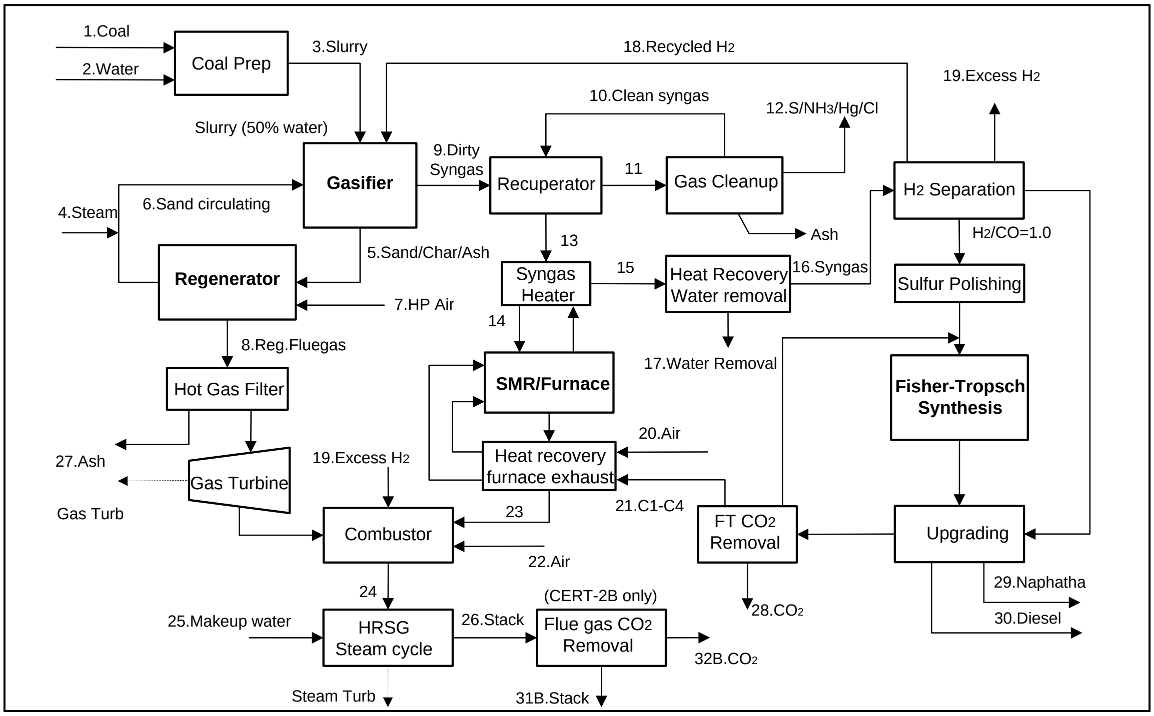
Material and Energy Balance

As with Case CERT-1, the material and energy balance for Case CERT-2 and CERT-2B is based on the maximum diesel/naphtha fuels production (while allowing sufficient C1 to C4 production to fuel the SMR) from 3,600 TPD of dry coal using an FT synthesis reactor for a ratio of 1.0 H_2/CO inlet specification. The SHR with regenerator and warm gas cleanup configurations are the same with CERT-1. The gas goes through the catalytic steam reformer tubes where the 15.4% CH_4 content in the clean syngas is reduced to 2.0% CH_4 and H_2 content of the gas is increased to 51.5%.

Fuel for the reformer furnace is the light hydrocarbons captured by the MDEA CO_2 removal unit featured on the FT synthesis gas recycle loop. CO_2 removed from the FT loop amounting to an overall 65% CO_2 capture is processed and is fed to a CO_2 compressor. In CERT-2B, Fluor Economine process is used to capture the CO_2 from the combustor flue gas with MEA as sorbent which increases the overall carbon capture ratio to 90%. In the compression section, the CO_2 is compressed to 2,215 psia by a five-stage intercooled centrifugal compressor. A pipeline transport, distribution and monitoring system is included. Detailed simulation block flow diagrams with stream tables of CERT-2 and 2B can be found in Appendix D. Overall process flow diagram with individual process stream tables is given in Figure 4.9.

Overall performance for the entire plant of CERT-2 and 2B is summarized in Table 4.10, which includes auxiliary power requirements.

For Case CERT-2, the net plant output power, after plant auxiliary power requirements are deducted, is nominally 106.45 MW_e. The overall plant thermal effective efficiency is 52.9% on an HHV basis. For Case CERT-2B, the net plant output power, after plant auxiliary power requirements are deducted, is nominally 73.73 MW_e. The overall plant thermal effective efficiency is 50.1% on an HHV basis.



CERT-2 & CERT-2B

Stream	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
T (℃)	15	15	230	735	900	900	538	900	750	290	480	38	601	808	650	38
P (Bar)	1.0	1.0	32.1	30.3	30.0	30.0	31.0	30.0	30.0	29.0	30.0	1.0	30.0	30.0	30.0	30.0
m (kg/s)	41.7	41.7	83.3	41.7	723.4	713.8	70.5	75.6	120.3	119.5	120.3	0.8	119.5	119.5	119.5	70.6
Stream	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31B	32B
T (℃)	38	324	204	16	36	16	368	1076	15	147	149	-40	25	25	33	-40
P (Bar)	1.0	30	25.6	1.0	30.0	1.5	1.0	1.0	1.0	1.0	30.0	151.7	1.0	1.0	1.0	151.7
m (kg/s)	48.9	4.8	2.3	102.0	7.0	65.8	109.1	252.8	0.0	252.8	4.4	39.5	2.9	6.8	225.8	24.3

Figure 4.9 Overall process flow diagram for CERT-2 and CERT-2B plant design with process stream table

4.7.4 Case CERT-3 and CERT-3B design and plant performance

Case CERT-3 and 3B has the same gasifier performance of Case CERT-2 and 2B with similar process configurations except for the fuel synthesis and power generation section. CERT-3 is configured to capture and compress the CO_2 in the mainstream only with overall 58.0% carbon capture efficiency (28.8% in fuels and 29.2% in CO_2) and send it off-site for sequestration. The total quantity of CO_2 captured for sequestration is 1,933 TPD from the MDEA system.

Table 4.10 Case CERT-2 and CERT-2B plant performance summary

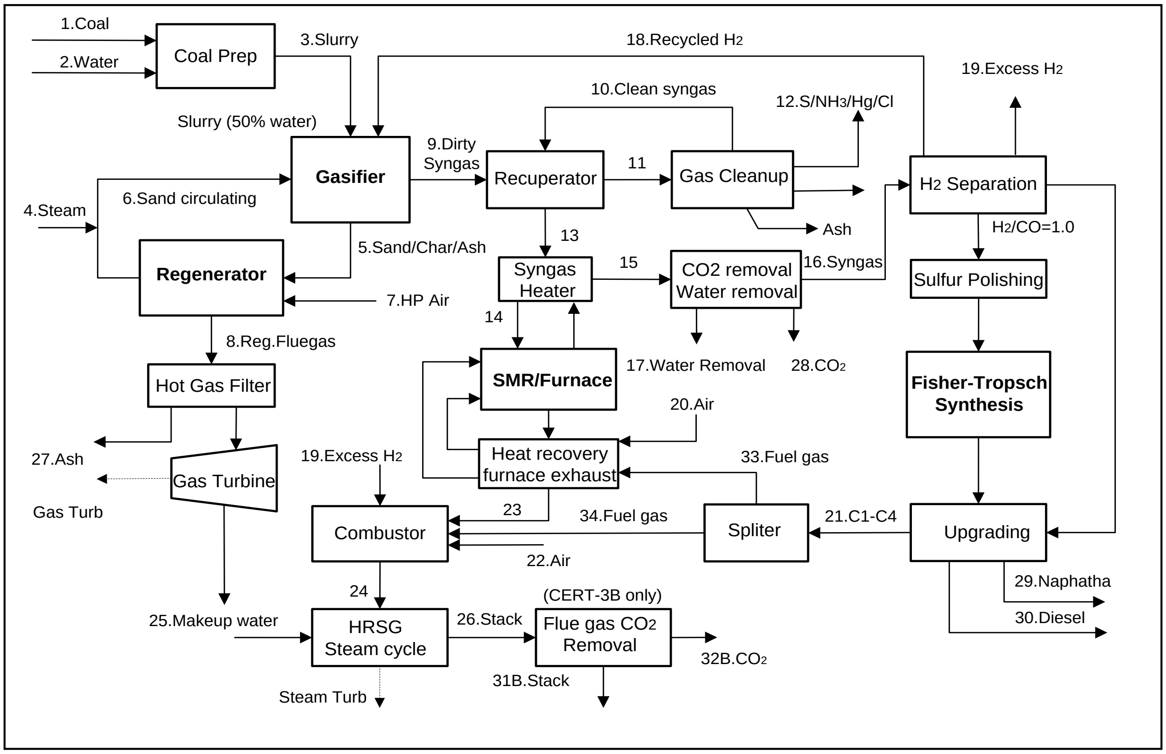
Power generation (MW)	CERT-2	CERT-2B
Gas turbine	24.11	24.11
Steam turbine	122.25	101.98
Total	146.35	126.08
FT liquid production (BBL/day)		
Naphtha	2,550	2,550
Diesel	4,535	4,535
Total	7,086	7,086
Auxiliary load (MW)		
Coal handling	0.25	0.25
Gasification	25.07	25.07
Warm gas cleanup	0.44	0.44
SMR	3.00	3.00
H_2 separation	2.90	2.90
FTR and upgrade	4.45	4.45
CO_2 capture and compression	3.80	16.26
Miscellaneous balance of plant	Included	Included
Total load	39.90	52.36
Plant performance		
Net plant power (MW)	106.45	73.73
Fuels efficiency (HHV base)	43.80%	43.80%
Overall thermal efficiency (HHV base)	52.90%	50.10%
Specific liquids production rate (BBL/dry metric ton)	1.97	1.97
Coal feed flow rate (ton/day)	3600	3600
Thermal input (MW)	1170.4	1170.4
CO_2 recovered (ton/day)	3415	5518

While CO_2 capture is increased from 58% to a level of 90% in CERT-3B. This case through the addition of a Fluor Economine scrubber to capture additional 32% of coal carbon in the flue gas which results in an overall 90% carbon capture efficiency. The total amount of CO_2 captured for sequestration is 4,950 TPD from the Fluor Econamine system and the MDEA system.

Material and Energy Balance

As with Case CERT-2 and 2B, the material and energy balance for Case CERT-3 and CERT-3B is based on the maximum diesel/naphtha fuels production (while allowing sufficient C1 to C4 production to fuel the SMR and combustor for power generation) from 3,600 tons per day of dry coal using an FT synthesis reactor for a ratio of 2.0 H_2/CO inlet specification.

The SHR with regenerator, warm gas cleanup and SMR configurations are the same with CERT-2 and 2B. Commercially deployed cobalt-based catalyst is used in the FTR with initial H_2/CO input ratio of 2.0 that has no water-gas-shift activity and long lifetime durability compared with iron-based catalyst. In this study, we have not considered the option of recycling unconverted syngas for additional liquid fuels production and, on the contrary, FT unit with syngas one-through (78% conversion efficiency) is employed to make the whole plant more heat self-sufficient. The unconverted syngas along with light hydrocarbon is used as fuel in the SMR furnace and combustor for power generation. CO_2 removed from the FT effluent in the MDEA section amounting to an overall 58% carbon capture is processed and is fed to a CO_2 compressor in case CERT-3. In CERT-3B, Fluor Economine process is used to capture the CO_2 from the combustor and regenerator flue gas with MEA as sorbent which increases the overall carbon capture ratio to 90%. Detailed simulation block flow diagrams with stream tables for CERT-3 and 3B can be found in Appendix D. Overall process flow diagram with process stream tables is given in Figure 4.10.



CERT-3 & CERT-3B

Stream	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
T (℃)	25	25	230	732	750	890	538	890	750	290	510	38	553	837	600	204	38
P (Bar)	1.0	1.0	32.1	30.3	30.0	30.0	31.0	30.0	30.0	28.1	29.9	1.0	28.0	27.9	27.5	27.0	1.0
m (kg/s)	42.0	42.0	84.0	42.0	849.6	840.0	62.8	72.3	121.3	119.4	121.3	1.4	119.4	119.4	119.4	71.1	48.3
Stream	18	19	20	21	22	23	24	25	26	27	28	29	30	31B	32B	33	34
T (℃)	324	204	16	36	16	276	1027	15	133	149	-40	25	25	33	-40	36	36
P (Bar)	30.3	25.6	1.0	20.3	1.5	1.0	1.0	1.0	1.0	30.0	151.7	1.0	1.0	1.0	151.7	20.3	20.3
m (kg/s)	4.9	0.1	95.6	14.3	44.2	104.6	154.2	1.2	154.2	4.4	22.4	3.5	8.4	119.3	34.9	9.0	5.3

Figure 4.10 Overall process flow diagram for CERT-3 and CERT-3B plant design with process stream table

Overall performance for the entire plant of CERT-3 and 3B is summarized in Table 4.11, which includes auxiliary power requirements. For Case CERT-3, the net plant output power, after plant auxiliary power requirements are deducted, is nominally 55.56 MW_e. The overall plant thermal effective efficiency is 57.59% on an HHV basis. For Case

CERT-3B, the net plant output power, after plant auxiliary power requirements are deducted, is nominally 13.44 MW_e. The overall plant thermal effective efficiency is 53.99% on a HHV basis.

Table 4.11 Case CERT-3 and CERT-3B plant performance summary

Power generation (MW)	CERT-3	CERT-3B
Gas turbine	24.11	24.11
Steam turbine	71.70	47.47
Total	95.80	71.57
FT liquid production (BBL/day)		
Naphtha	3078	3078
Diesel	5470	5470
Total	8548	8548
Auxiliary load (MW)		
Coal handling	0.25	0.25
Gasification	25.07	25.07
Warm gas cleanup	0.44	0.44
SMR	2.86	2.86
H_2 separation	3.74	3.74
FTR and upgrade	5.75	5.75
CO_2 capture and compression	2.15	20.04
Miscellaneous balance of plant	Included	Included
Total load	40.25	58.13
Plant performance		
Net plant power (MW)	55.56	13.44
Fuels efficiency (HHV base)	52.84%	52.84%
Overall thermal efficiency (HHV base)	57.59%	53.99%
Specific liquids production rate (BBL/dry metric ton)	2.37	2.37
Coal feed flow rate (ton/day)	3600	3600
Thermal input (MW)	1170.4	1170.4
CO_2 recovered (ton/day)	1933	4950

4.7.5 Summary

Plant performances in the five cases are summarized in Table 4.12. The CE-CERT process has a high percentage of input carbon that does not end up in liquids production but instead ends up as flue gas that is converted into steam and eventually power generation. The carbon content in this flue gas is a product of combustion of leftover char along with the C1 to C4 components separating from the FT outlet.

Table 4.12 Plant performance summary of CTL cases

	CERT-1	**CERT-2**	**CERT-2B**	**CERT-3**	**CERT-3B**	**POX-1**
Plant Output						
Gas Turbine Power (MWe)	2.39	24.11	24.11	24.11	24.11	n/a
Steam Turbine Power (MWe)	13.23	122.25	101.98	71.7	47.47	112.21
Total (MWe)	**15.62**	**146.35**	**126.08**	**95.8**	**71.57**	**112.21**
FT Liquid Production						
Diesel and Naphtha Product (bbl/day)	**679**	**7086**	**7086**	**8548**	**8548**	**7086**
Total Auxiliary Load (MWe)	**5.00**	**39.9**	**52.36**	**40.25**	**58.13**	**102.01**
Plant Performance						
Net Plant Power (MWe)	**10.62**	**106.45**	**73.73**	**55.56**	**13.44**	**10.20**
Net plant thermal efficiency (HHV)	51.1%	52.90%	50.10%	57.59%	53.99%	47.60 %
Specific liquids production rate (bbl/ton dry coal)	1.87	1.97	1.97	2.37	2.37	2.12
Coal feed flow rate, dry basis (ton/day)	364	3600	3600	3600	3600	3336
Thermal input (MW)	118.2	1170.4	1170.4	1170.4	1170.4	1093.3
CO_2 recovered (ton/day)	0	3415	5518	1933	4950	4615
Overall carbon capture efficiency	27.3%	65%	90%	58%	90%	88%

The unique design of SHR results in a large amount of coal carbon left as char for the internal heating requirement (19.3% in CERT-1, 17.9% in CERT-2, 2B, 3 and 3B) and more carbon exists as CO_2 at the high steam environment. The carbon loss reduces the overall available carbon fixed in the FT liquids (27.3% in CERT-1, 28.8% in CERT-2 and 2B, 34.7% in CERT-3 and 3B) and, however, more electricity is generated as a result due to large amount of fuel gas (e.g., H_2 and C1-C4) for combustion and process heat for HRSG.

It is found out that the 2.0 H_2O/coal mass input ratio used in the plant design is much higher than conventional gasification process with slurry feed. Although heavy heat load has to be introduced into SHR at the presence of large amount of water, enhance conversion of carbon with less resident time can be achieved. Moreover, the resulting H_2O/CH_4 molar input ratio is above 3.0 in the SMR which avoids the usage of an external steam source and thus compensates the heat load in SHR to some great extent.

In all five cases, the gasification island requires the highest power consumption for the regenerator air compression which accounts for more than half of the total plant auxiliary load. However, the high pressure hot flue gas from the regenerator produces additional power in the gas turbine expansion and thus largely compensates for this consumption.

CERT-2 and 2B are designed to have more power coproduction while case CERT-3 and 3B aim to produce more liquid fuels. As a result, CERT-3 and 3B have 20% more FT fuels yield than that of CERT-2 and 2B due to the cobalt catalyst (2.0 H_2/CO ratio) used in the FT synthesis while less net power results since less H_2 is left as fuel in the combustor.

The CO_2 capture and sequestration option in the plant design increases the power consumption a lot and the compression energy load is even more when the flue gas CO_2 removal is included. The power consumption to recover 1 ton CO_2 is 1.1 KW in the MDEA unit while 5.9 KW has to be used in the MEA flue gas CO_2 removal unit. Not mention the extra process heat needed for the CO_2 sorbent regeneration that enlarges the energy consumption further. As a result, the 90% overall carbon capture option leads less net export power and brings down the overall thermal efficiency to some large extent.

In order to make a comprehensive comparison of CTL technology via CE-CERT steam hydrogasification and conventional partial oxidation gasification, Case POX-1 is selected utilizing the same Utah bituminous coal, a dry feed entrained flow gasifier and iron-based FT liquids plant that processes 3,336 TPD of coal and includes 88% carbon capture to yield the same 7,086 BBL/day of liquid fuels as Case CERT-2 and 2B. It should be noted that the conventional CTL plant with a high throughput oxygen-blown dry feed entrained flow gasifier is not modeled here while plant performance is obtained from currently NETL gasification studies [51].

The comparison summary in Table 4.12 shows that although the specific liquids output of the entrained gasifier-based coal to liquids plant is nearly 8% higher than CERT-2B process at 90% carbon capture, the overall efficiency is 2.5% or approximately 5% lower. Thus steam hydrogasification process truly benefits from a lower gasifier operating temperature and no need for an ASU plant which in turn increases the amount of net power for export.

4.8 Economic analysis

The process economics are based on the assumption that this is the “nth” plant mentioned in chapter 3. The cost estimation has an expected accuracy of roughly ±30% which is consistent with the screening study level of information available for the various study power technologies [72]. All capital and O&M costs were presented as “overnight costs” expressed in 2010 dollar. The cost estimation methodology was highly relied on a major study for fossil energy plants [71] while inflation and electric price was estimated using data from the EIA [10]. The cost in the evaluation of the gasifier was based on an internal cost reported to NETL on the transport gasifier and the cost of warm gas cleanup system was based on cost information developed by Noblis, Inc.

The methodology for equipment cost and capital cost estimation as well as capital investment structure and other major economic parameters follows the instruction introduced in chapter 3. Uinta basin coal, although somewhat different in composition from the design Utah coal was used as a commercially available coal with average known market price of 42 \$/ metric ton (1.76 $\$/GJ_{HHV}$) [81]. Electricity sale price was set as 54 \$/MWh and escalation factor of 3% was employed in coal purchase price, electricity sale price and O&M expense to reflect inflation within plant lifetime. All financial values used in this paper were adjusted in 2010 August dollars while by-product NH_3, sulfur and captured CO_2 was not considered as credit in this study. Capital cost breakdown for the CTL plants is given in Table 4.13.

Table 4.13 Estimated plant capital cost (Overnight costs)

Plant economics	CERT-1 (\$MM)	CERT-2 (\$MM)	CERT-2B (\$MM)	CERT-3 (\$MM)	CERT-3B (\$MM)
Feedstock preparation	11.6	77.1	77.1	77.1	77.1
Gasification island	75.6	270.1	270.1	270.1	270.1
Gas cleanup	20.7	87.1	87.1	87.1	87.1
SMR	47.3	159.4	159.4	159.3	159.3
FTR	10.5	87.3	87.3	103.5	103.5
CO_2 capture& compression	19.3	97.5	157.0	69.3	148.7
Power island	21.7	118.6	108.0	84.7	66.5
BOP	32.9	82.4	87.4	78.6	84.3
Total plant cost	239.6	979.5	1033.3	929.6	996.6
Total required capital	**308.4**	**1366.1**	**1461.6**	**1301.5**	**1414.6**

In all five cases, gasification island accounts for the largest fraction of the cost, followed by SMR, power island and CO_2 capture & compression. Major difference between the

CERT-2, 2B, 3 and 3B is found in CO_2 capture & compression section which is expected since CERT-2B and 3B have more non-fuel carbon sequestrated in the flue gas. A Balance of Plant (BOP) cost is used in waste water treatment and cooling tower facility along with miscellaneous cost. The total required capital cost is 308.4, 1366.1, 1461.6, 1301.5, and 1414.6 $MM for CERT-1, CERT-2, CERT-2B, CERT-3 and CERT-3B, relatively. The CTL facility has always been a capital cost intensive process and this is especially true for relative small scale plant since the specific capital cost for CERT-1 is 454 $K/bbl while 192, 206, 152, and 165 $K/bbl for CERT-2, 2B, 3 and 3B respectively.

The intensive 90% CO_2 capture option increases the liquid fuels production by approximate 13 $/bbl in all cases which is accordance with CO_2 capture study in other gasification plants [82]. The sequestrated high pressure CO_2 can be potential used as enhance oil recover which could compensate the capture cost to some great extent although not discussed in this study.

Table 4.14 shows the levelized FT liquids Production Cos (PC), also called plant-gate selling price, at expected 12% Internal Rate of Return (IRR) calculated from the financial model. The production costs consist of several broad categories of cost elements which include operating labor, maintenance material and labor, coal purchase, consumables, byproduct power sale, loan expense, capital depreciation and tax. Credit is not taken for the sale of sulfur but is taken for the net electricity generated.

In addition, the cost is also shown in terms of Breakeven Crude Oil Price (BCOP) in $/bbl which is similar as Kreutz TG, et al [83]. The BCOP is the refiner acquisition price for crude oil and used to convert the coal derived FT fuels price to petroleum-derived fuels wholesale price. BCOP in this study is estimated as follows:

BCOP=42*(FTCOST-RM),

FTCOST is the average plant gate selling price of the FT liquids ($/gallon) as determined from the Discount Cash Flow Rate on Return (DCFROR) analysis. It is assumed average mix of naphtha and diesel produced (36/64) here has the same heating value as corresponding petroleum derived fuel (one gallon of FT liquids displace one gallon of the petroleum derived fuel). RM is the average refinery margins for petroleum naphtha and diesel. Since naphtha is the intermediate product of crude oil to gasoline in petroleum refinery, we assume the average RM is 18 ¢/gallon for naphtha (33 ¢/gallon for gasoline). For diesel, the RM is assumed to be 30 ¢/gallon [84].

Table 4.14 Levelized and breakdown FT liquids PC

Case	CERT-1	CERT-2	CERT-2B	CERT-3	CERT-3B
Breakdown FT liquids cost ($/gallon)					
O&M cost	2.43	0.49	0.51	0.44	0.45
Coal purchase	0.63	0.60	0.60	0.50	0.50
Loan expense	1.37	0.55	0.59	0.44	0.48
Income tax	0.25	0.10	0.11	0.08	0.09
Capital depreciation	0.89	0.33	0.35	0.26	0.28
Electricity sales	-0.48	-0.46	-0.32	-0.19	-0.04
Average return on investment	0.71	0.28	0.31	0.23	0.25
Total	**5.80**	**1.90**	**2.16**	**1.75**	**2.00**
PC ($/Diesel Gallon Equivalent)	5.99	1.97	2.23	1.81	2.07
BCOP ($/bbl)	232.7	68.9	79.8	62.6	73.1

The BCOP analysis provides an important insight regarding the risk mitigation that the CTL process might provide to the investors in the future. Major costs incurred are loan expense, feedstock purchase, O&M cost and capital depreciation. Electricity coproduct is sold as credit and shown as negative entries in the table.

Based on the discount cash flow analysis, the FT liquids PC for CERT-1, CERT-2, 2B, 3 and 3B is 5.80, 1.90, 2.16, 1.75 and 2.00 $/gallon (5.99, 1.97, 2.23, 1.81 and 2.07 $/DGE), respectively. It should be noticed that CERT-1 has almost three times PC compared with that in other four cases and the O&M (chemicals, utilities, labor and administrative support) cost in CERT-1 accounts for almost half of the PC due to the small design capacity. In this study, the corresponding BCOP for CERT-1, CERT-2, 2B, 3 and 3B is 232.7, 68.9, 79.8, 62.6 and 73.1 $/bbl, respectively. Based on the financial model outputs, the extremely high FT liquids PC in CERT-1 makes it extremely expensive to run the CTL facility in such a small capacity thus will not be further discussed in the following economic analysis.

Coal purchase price, electricity coproduct sale price and target IRR on equity are investigated in detail on the FT liquids PC analysis to illustrate the impact of changes on these most important variables. Figure 4.11 and 4.12 shows the corresponding BCOP as a function of the coal purchase price and electricity sale price at 12% IRR.

It is observed that for each $1 change in the purchase price per ton of coal at an electricity sale price of $54 MWh changes the corresponding BCOP by 0.59 $/bbl for case CERT-2 and 2B while 0.50 $/bbl for case CERT-3 and 3B, respectively. Similarly, for each $10 per MWh change in electricity sale price results in a BCOP change of 0.36, 0.25, 0.15 and 0.03 $/bbl for case CERT-2, 2B, 3 and 3B, respectively.

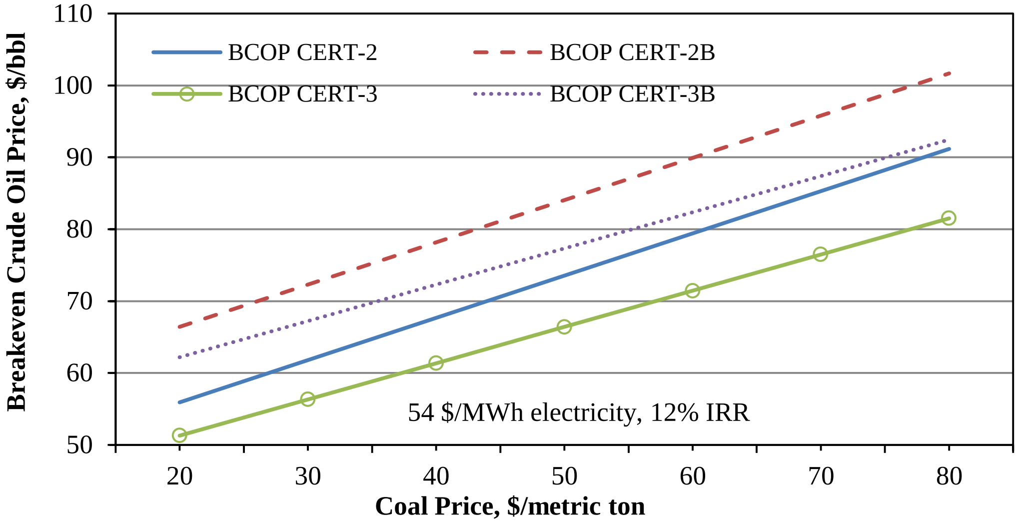


Figure 4.11 The effect of coal sale price on fuels production cost

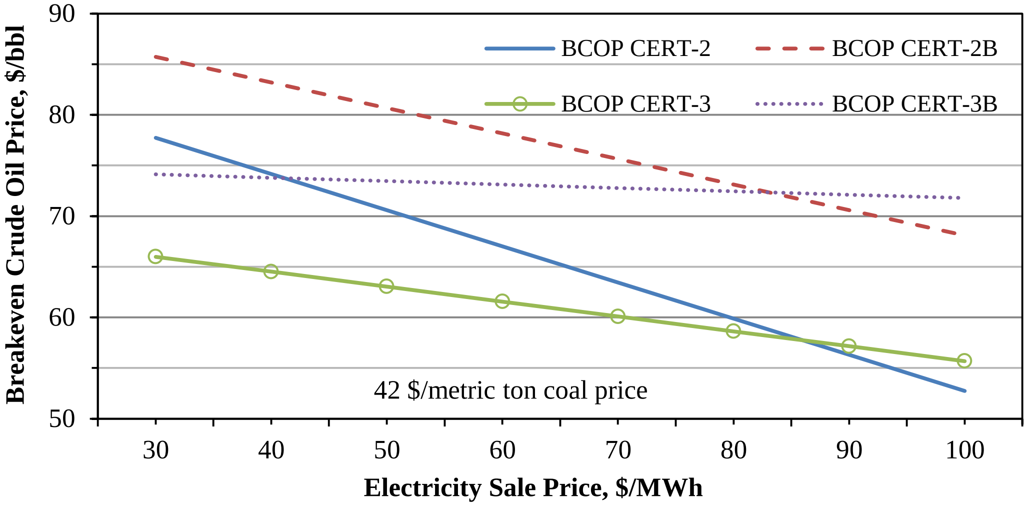


Figure 4.12 The effect of electricity sale price on fuels production cost

It can be indicated that export electricity, as the major revenue against the FT liquids PC, determines the plant economy greatly and FT fuels are typically less costly to produce when electricity is generated as a major coproduct [85]. This can be observed apparently from the plant performance in this study which has relative more export power due to the features of steam hydrogasification process described earlier.

Besides the coal purchase price and electricity sale price, FT fuels as a major revenue source, their sale price which is primarily dominated by the crude oil price is very important in determining the plant economy. The CTL plant has to compete with the crude oil market and Figure 4.13 shows the impact of the plant gate selling price on IRR values without any subsidy. The PC increases with IRR slightly parabolically and this trend is observed clearly on the BCOP. The BCOP is 43.3, 52.1, 42.0 and 51.1 $/bbl at IRR of 4% and raises up to 127.1, 141.8, 108.2 and 122.9 $/bbl when 28% IRR is desired for CERT-2, 2B, 3, and 3B, respectively.

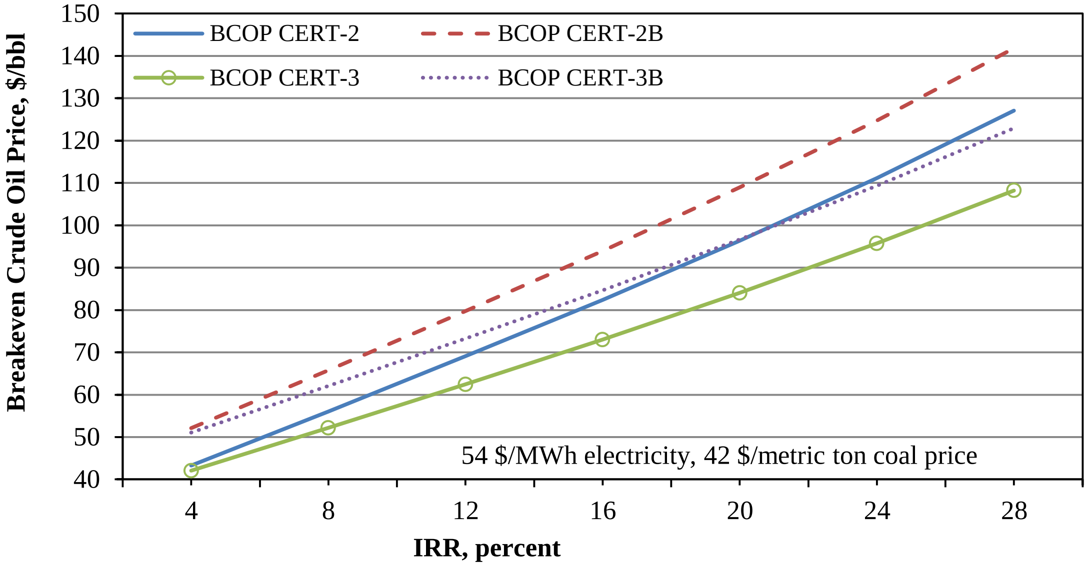


Figure 4.13 Minimum FT liquids sale price as a function of required IRR on equity

Additionally, the size of the plant plays a very important role in determining the process economy and is varied by feedstock input rate. The effect of plant size on PC is shown in Figure 4.14. Due to the strict policy in carbon emission control in the near and middle term, the plant scale up analysis only focuses on Case CERT-2B and 3B which have overall 90% carbon capture. No “scale effect” is considered in the plant scale up scenarios which means the thermal conversion efficiency is constant all through the plants with different size and the power and fuel output is linear with plant capacity. The capital cost at different plant capacity is estimated based on the "sixth-tenth" rule mentioned in chapter 3.

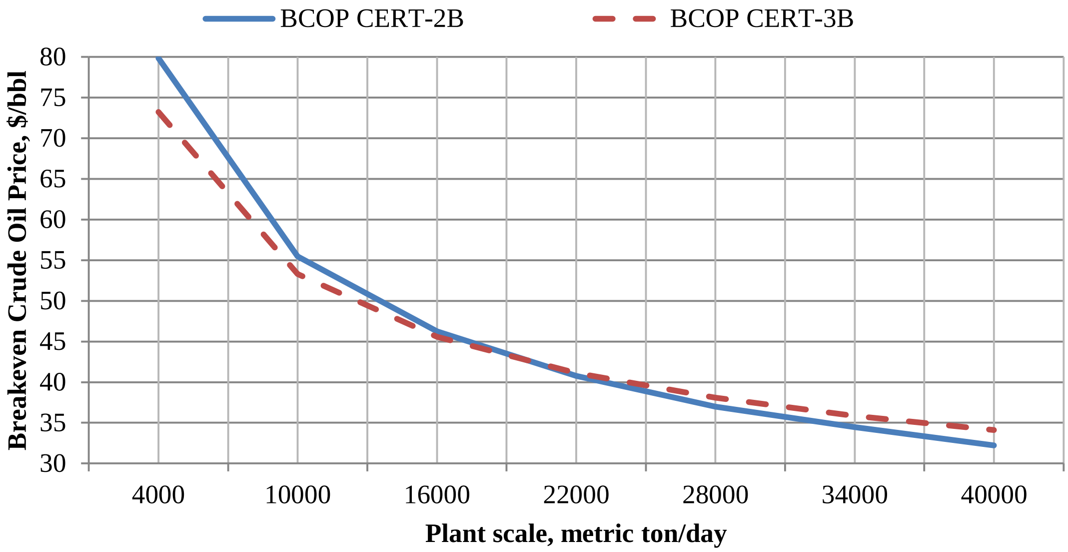


Figure 4.14 Minimum FT liquids sale price as a function of plant scale

The FT fuels PC drops dramatically when the plant size increases from 4,000 to 10,000 TPD. After that, the PC decreases with plant size gradually and when the plant size pass 20,000 TPD, the PC of CERT-3B exceeds that of CERT-2B due to the high capital cost in fuels synthesis and CO_2 capture as well as less electricity coproduction. Finally, the slope of the PC tends to level out when the plant size is over 35,000 TPD. The fuels production is below 35 $/bbl when the plant capacity is 40,000 TPD which is low enough to compete with the crude oil market and makes the CTL process more attractive.

Summary

Based on the plant performance and economic analysis results, CTL facilities utilizing steam hydrogasification technology benefit greatly from the relative low operation temperature and the non-usage of oxygen. Although there is more coal carbon loss during gasification, either in char or CO_2, the high CH_4 yield provides the system with sufficient amount of H_2 that can be used for FT synthesis and power generation. Furthermore, the high H_2/CO molar ratio also avoids the use of water gas shift unit which compensates for the upstream carbon loss to a large extent. All these features result in a potential higher overall plant thermal efficiency and lower PC for the CTL plant utilizing steam hydrogasification. However, key values and assumptions used in this study such as tar yield, char production and raw gas composition at desired gasification temperature still needs to be validated by further experiment via fluidized bed type Process Demonstration Unit (PDU) which has already been built and in extensive test right now.

4.9 Power generation via IGCC

As mentioned early, the other application for CE-CERT steam hydrogasification technology is to produce power via IGCC.

Case CERT-4

Power generation from a demonstration plant size of 4,000 TPD (Utah bituminous coal, dry basis) via IGCC w/o CO_2 capture.

Material and Energy Balance

The material and energy balance for Case CERT-4 is based on the maximum fuels gas production from 3,628 ton coal to satisfy site conditions for the 2015 advanced steam turbine configuration for power production. The coal handling, gasification island and gas clean configurations are the same as the operation units used in CTL plant. The flue gas from the regenerator passes through a hot gas filter where fly ash is removed, and the particulate-free fuel gas is directed to the gas turbine combustor to generate power. Prior to the steam methane reformer, the clean syngas is polished of remaining sulfur species to less than 1 ppmv level and cooled to pass a mercury and ammonia removal step.

The high H_2 syngas stream is cooled down in the heat exchanger where some moisture is condensed. The H_2 rich stream then combines with the bypass fuel gas for final separation of the H_2 needed for recycle to the gasifier. After the recycle H_2 is separated, the remaining fuel gas is cooled against boiler feeding water to generate LP steam that can be utilized in the LP section of the steam turbine set. Finally, the fuel gas is mixed with air from the gas turbine compressor in the gas turbine combustors. The results hot flue gas then goes through a serious heat exchanger in the HRSG for power generation in the steam turbine cycle. No CO_2 capture and sequestration option is considered in the configuration.

Overall performance for the Case CERT-4 IGCC plant is summarized in Table 4.15, which includes auxiliary power requirements. For Case CERT-4, the net plant output power, after plant auxiliary loads are deducted, is nominally 533.3 MW_e. The overall plant thermal effective efficiency is 45.2% on an HHV basis.

Additionally, the IGCC Case CERT-4 is compared to a conventional IGCC plant (Case POX-2) using an entrained flow gasifier with the same 2015 advanced gas turbine and RTI warm gas cleanup system. As can been seen in Table 4.15, major power generation difference happens in the steam turbine section and this result can be explained as steam hydrogasification process has a substantial amount of steam introduced into the gasifier when compared with the dry feed entrained flow gasifier which lowers the power output largely.

Table 4.15 Cases CERT-4 plant performance summary

	CERT-4	POX-2
Plant output		
Gas turbine power (MWe)	385.3	370.0
Steam turbine power (MWe)	175.7	245.1
Expander power (MWe)	--	27.0
Total auxiliary load (MWe)	27.7	81.4
Net plant power (MWe)	**533.3**	**560.8**
Coal fuel type	Utah	Utah
Net plant efficiency (HHV)	45.2%	45.5%
Coal feed flowrate, dry basis (ton/day)	3628	3789
Thermal input (MWth)	1179.8	1232.0

Although CERT-4 has less auxiliary loads compared with POX-2 associated with the entrained gasifier need for oxygen from the ASU, the net power output cannot be offset by the large amount of steam usage in SHR in CERT-4. As a result, unlike CTL process, the steam hydrogasification IGCC shows no advantage over the conventional partial oxidation gasification technology from plant thermal efficiency point of view.

4.10 Summary and conclusion

Optimum feed condition of H_2O/Coal mass ratio of 2.0 of range studies with H_2/C molar ratio of 1.0 is determined based on process mass and heat analysis. Detailed plant design, process simulation and cost analyses have been carried out for CTL plants using steam hydrogasification. The analyses provide the basis for discussion of the effect of CTL facility design capacity on the plant performance and economy. The unique design of SHR needs significant amount of heat to proceed and method for heat supply is of vital importance in determining the plant economy. Based on the simulation results, the SHR can be heat self-sufficient at 750 °C by char combustion in the regenerator. A major finding is that iron-based CTL plant designs in this study have more energy export as power due to large amount of excess H_2 available for power generation. The electricity coproduct reduces the fuels PC largely and is especially favored when the electricity value is high. With the cobalt-based CTL plant designs, more FT fuels are produced which results in less power coproduction and are especially favored when the crude oil price is high. For small scale CTL facility, the expensive plant capital investment, low fuel yield and high O&M expense result in an extremely high PC.

Synthetic diesel and naphtha, as the major product, their selling price impacts the IRR strongly while the impact may be even more sharply since the market price of the fuel is more volatile and less predictable than many other process variables. The studies show that for the FT fuels produced from the CTL facilities with capacity of 4,000 TPD utilizing steam hydrogasification, the cost of production is competitive with petroleum when crude oil price is as low as 95 $/bbl (20% IRR on equity) if coproduct electricity is sold at a sufficient high price and coal price keeps sufficient low. Further economic benefit comes with increasing plant size and the support from commercial scale demonstration efforts to make the process more competitive.

Major by-product sulfur and NH_3 in the fuel production process, although not considered as credit in the study, has great potential value to bring down the FT liquids PC further. Especially for the large amount of NH_3 (0.4 vol% in raw gas) that could be used for chemical synthesis or H_2 carrier. The CCS in the plant design offers a great approach for decarbonizing in both fuel and power and reduces the climate change risk. However, the additional capital cost and power consumption lowers the plant economy largely, especially for the plant design with 90% CCS. Policy on carbon sequestration credit from fossil energy source and revenue from CO_2 usage for enhanced oil recovery could probably drop the FT liquids PC further though is not discussed. Other financial incentives such as tax incentives, loan guarantees, and other mechanisms may also play a role in addressing the economic and market challenges of CTL, but these are beyond the scope of this study.

It can be indicated that export electricity, as the major revenue against the FT liquids PC, determines the plant economy greatly and FT fuels are typically less costly to produce when electricity is generated as a major coproduct [85]. This can be observed apparently from the plant performance in Case CERT-2 and 2B which have relative more export power.

The cases investigated here only compared the plant performance and economy with different plant design capacity and carbon sequestration ratio. It should be noticed that a lower FT liquids PC doesn’t guarantee a higher IRR in the plant design. Optimum operation condition and process design to maximize plant profits still needs to be further studied based on the kinetics data collected from future experiment work. Moreover, demonstration of commercial scale SHR with regenerator and warm gas clean up system is necessary to promote the CTL process based on steam hydrogasification in the future, but there are no fundamental research hurdles.

5. Synthetic fuels production from biomass

Biomass has long been a central point of development efforts that focused on using food crops for making renewable biofuels (primarily corn-based and starch-based ethanol but also biodiesel derived from soybeans and canola). However, concerns about food price impacts [86] and indirect land use impacts of growing biomass for energy on croplands [87,88] have led to growing recognition that emphasis should be shifted instead to exploiting for energy mainly lignocellulosic feedstocks that don't require use of food biomass for providing energy. These alternatives include various crop, forest residues and energy crops (e.g. short-rotation trees such as poplar and fast growing grasses such as switchgrass) that can be grown on degraded lands as well as municipal biomass waste (e.g. leaves, grass and prunings) that has large quantity of reserve in a certain area and can be collected within short radium.

Aside the conventional biofuel production using lipids extracted from biomass, there are several new pathways to provide renewable alternatives to the transportation fuels. These options include cellulosic ethanol produced biochemically and synthetic fuels derived thermochemically through gasification—so-called Biomass to Liquids (BTL) technologies. The BTL process is considered as "carbon neutral", since CO_2 released to the atmosphere is recycled via photosynthesis. Other processes such as fermentation of carbohydrates (anaerobic digestion) and liquefaction of biomass through fast pyrolysis can also produce power and fuels but not in the scope of this study.

In this study biomass green waste with considerable quantities in the Southern California (SC) is considered as feedstock in the BTL process. The green waste includes leaves, grass, prunings and stumps that can be categorized as MSW. Traditionally, green waste can be collected and used for landfill, recycle material production and power generation via combustion or anaerobic digestion. The geographic location and climate in SC allows the biomass feedstock continuously supplied throughout the year. The other important reason that green waste is selected as target feedstock is the slurry feed condition in CE-CERT process. Green waste can be used directly without drying which is very suitable for the conversion of high moisture content feedstock. This chapter focuses on process simulation and economic analysis of synthetic fuels production based on steam hydrogasification using municipal biomass green waste as feedstock.

5.1 Feedstock feasibility study

Much of the information required to construct an availability assessment is found from California Energy Commission (CEC) in Public Interest Energy Research (PIER) collaborative report and Statewide Waste Characterization Study [89.90].

Figure 5.1 Map of California with county lines defining Southern California

The plant is expected to locate in the center of SC and the green waste collection area taken into consideration in SC including Los Angeles County, San Bernardino County, Riverside County and Orange County schematically shown in Figure 5.1. Green waste in Ventura County, San Diego County and Imperial County is not considered due to the relative long transportation distance and less amount of feedstock generation compared with the selected area in this study. The total estimated green waste yield in SC is 348,770 Bone Dry Tonne (BDT) in 2007. Detailed green waste generation data is given in Table 5.1. Of the technical availability fraction in green waste, 50% is assumed to be available for fuels conversion.

Table 5.1 Gross green waste production and availability for conversion in 2007 in SC

SC region	Leaves, Grass	Prunings& Trimmings	Branches & Stumps	Available for Conversion
	BDT/yr	BDT/yr	BDT/yr	BDT/yr
Orange County	65,740	455,320	7,040	264,050
Los Angeles County	201,770	503,970	21,620	363,680
Riverside County	42,490	207,460	4,550	127,250
San Bernardino County	38,770	171,610	4,150	107,265
SC Total	348,770	1,338,360	37,360	862,245
CA Total	710,000	2,180,000	76,000	1,483,000
SC/CA	49.1%	61.4%	49.2%	58.1%

If the entire technical available portion of feedstock listed in the table is used for synthetic fuels production via CE-CERT technology, 1048 thousand barrels can be produced per month using the feedstock yield in 2007 which can replace more than 0.6% of the crude oil derived gasoline and diesel currently produced in CA [91]. Projections from the base 2007 year data are made for the years 2010, 2017, and 2020 [90]. The gross green waste production and availability for conversion in the year mentioned above is shown in Table 5.2. As the yields of green waste increased in SC in next 10 years, the synthetic fuels yield can be expected to rise up to 1232 thousand barrels per month in 2020 that can replace 0.7% of the crude oil derived gasoline and diesel currently produced in CA.

Table 5.2 Gross green waste production and availability for conversion in SC (BDT/yr)

SC region	Leaves, Grass	Prunings& Trimmings	Branches & Stumps	Available for Conversion
2010				
Orange County	69,640	482,330	7,460	279,715
Los Angeles County	208,680	521,230	22,360	376,135
Riverside County	50,380	245,990	5,400	150,885
San Bernardino County	43,280	191,560	4,630	119,735
SC Total	371,980	1,441,110	39,850	926,470
CA Total	760,000	2,330,000	80,000	1,585,000
SC/CA	48.9%	61.9%	49.8%	58.5%
2017				
Orange County	72,500	502,300	7,800	291,300
Los Angeles County	214,900	536,800	23,000	387,350
Riverside County	55,900	273,100	6,000	167,500
San Bernardino County	46,700	206,600	5,000	129,150
SC Total	390,000	1,518,800	41,800	975,300
CA Total	802,100	2,466,300	86,000	1,677,200
SC/CA	48.6%	61.6%	48.6%	58.2%
2020				
Orange County	74,700	517,400	8,000	300,050
Los Angeles County	219,000	547,000	23,500	394,750
Riverside County	60,800	296,600	6,500	181,950
San Bernardino County	49,300	218,400	5,300	136,500
SC Total	403,800	1,579,400	43,300	1,013,250
CA Total	833,000	2,563,900	88,900	1,742,900
SC/CA	48.5%	61.6%	48.7%	58.1%

Total amount of green waste that can be technically conversion to synthetic fuels in SC are summarized in Table 5.3.

Table 5.3 Technical available green waste yields in SC for synthetic fuel production

Year	BDT/yr	BDT/day
2007	862245	2362
2010	926470	2538
2017	975300	2672
2020	1013250	2776

5.2 Transportation cost estimation methodology

The development of the biomass transportation cost modeling consists of several parameters that contribute to the overall cost of transportation biomass feedstock to an energy conversion plant. The frameset of the equation consist of two areas, the distance variable cost (DVC) and the distance fixed cost (DFC) as shown in the following equation below [92]:

$$Total\ Transportation\ Cost = DFC + DVC \qquad (5\text{-}1)$$

For simplicity, it is assumed that the distance fixed cost is only associated with the loading and unloading costs of the biomass feedstock and will not be considered in this study. However, the distance variable cost takes into account for variations in biomass yield, collection radius, hauling load, and the fuel economy of the vehicle transporting the biomass.

The following transportation cost modeling was generated at CE-CERT with the aim of estimating the cost of transporting biomass feedstock to an energy conversion facility.

$$DVC\left(\$\ ton^{-1}\right) = \frac{4}{3} \times \frac{R}{l_t} \times \frac{f_p}{f_c} \times \frac{f_w}{(1-m)} \qquad (5.2)$$

Where the parameters are defined as the following:

- DVC: Distance variable cost [$/ton]

- R: Transportation radius [miles]
- lt: Truck load capacity [tons]
- fp: Fuel price [$/gallon]
- fc: Fuel economy [miles/ gallon]
- fw: Winding road factor [dimensionless]
- m: Moisture content of feedstock [percentage]

It is assumed that the location of the energy conversion plant is centrally located and the biomass feedstock that is being collected is surrounding within the transportation radium. The DVC is also a direct function of truck loading capacities, the fuel economy of the vehicle being driven, and the cost of fuel. Diesel is taken as fuel in the CE-CERT transportation model. The actual distance being traveled is corrected by applying the winding road factor which considers that the routes driven to transport the biomass are curved and not necessarily linear, therefore the winding road factor is set to be 1.35 [93-95]. The equation of the actual distance traveled is the following:

$$\text{Actual Distance Traveled [miles]} = \text{Distance} \times \text{Winding Road Factor} \tag{5.3}$$

Data Acquisition

Real world data was taken from several database sources to sufficiently estimate the cost of transporting biomass to the energy conversion facility. The cost of diesel fuel in California was taken from average diesel prices in the year 2011 as reported by the EIA and the average diesel price of $4.08 per gallon is used in the CE-CERT biomass transportation cost modeling.

The truck load capacity is estimated to be 10 tons which is the max load that is conventionally taken into the landfills by commercial haulers as reported by Riverside County Landfills [96]. Otherwise, waste haulers who exceed the allowed tonnage into the landfill will face penalties for any tonnage over the allowed weight. However, since the purpose is to remove or divert the feedstock from the landfill and collect the waste from the local residential area, it is assumed that each truck hauler will approximately haul 10 tons of feedstock from the residential areas and landfills. This is also a reasonable truck load capacity due to the diversity and scatter distribution of the green waste compared with energy crops such as corn stover.

The California Department of Motor Vehicles (DMV) reports that maximum load allowed on the roadways of California cannot exceed 40 tons so that transporting 10 tons to the energy conversion plant should not be an impediment since it is below the California weight limits [97]. As time progresses, advances in green waste management will allow truck to transport more feedstock to the energy conversion plant (e.g., 25 tons). The same could be said with the fuel economy of diesel truck where it is assumed that the fuel economy will get better over time with technological advances. The fuel economy of the diesel hauling truck is assumed to have a fuel economy of 5 miles per gallon. Parameters in the baseline case for green waste transportation cost estimation are given in Table 5.4.

The biomass delivery cost under the baseline case is 9.79 \$/ton dry basis and detailed analysis of the impact for biomass transportation radium on the plant capacity and economics will be given later.

Table 5.4 Baseline case for the parameters for CE-CERT

Parameter	Value
Radius, R [miles]	50.0
Truck Load Capacity, l_t [Ton]	10.0
Fuel Price, f_p [\$ gal^{-1}]	4.08
Fuel Economy, f_c [miles gal^{-1}]	5.00
Winding Road Factor, f_w [unitless]	1.35
Moisture Content, m, [percentage]	0.06

5.3 BTL process design and plant performance

Technologies for converting biomass into syngas and successive conversion to synthetic fuels have not yet been successfully demonstrated in industrialization. Most technologies such as biomass pretreatment, tar removal and cracking and gas cleanup are still under development. In this case, the technology developments of coal are assumed to apply for biomass in the plant design as well as in plant cost estimation.

The design is based on a geographic location in SC (includes Los Angeles County, Orange County, Riverside County and San Bernardino County), using green waste as the design feedstock with representive chemical composition shown in Table 5.5.

Table 5.5 Green waste chemical composition

	Compound	Green waste
Proximate	Volatile Matter	82.54
	Fixed Carbon	17.17
	Ash	0.29
Ultimate	C	49.25
	H	5.99
	O	44.36
	N	0.06
	S	0.05

One major prerequisite and also key assumption in the BTL process is the homogeneity of the green waste. Greene waste should have homogeneous and relative constant chemical composition although this might not be the case in reality due to diversity of the biomass and seasonal variation on the chemical composition.

BTL facility with design capacity of 2,000 TPD (dry basis) is selected and this is also a practical scale for the biomass based biorefinery with reasonable feedstock delivery radium. A 50 mile green waste collection radius is assumed and is considered to be realistic. Previous studies in biomass derived fuels plant design have used the similar plant size [63-65].

The plant is designed that consists of eight major sections: (1) Biomass preparation; (2) Gasification island; (3) Warm gas cleanup; (4) SMR; (5) Syngas conditioning; (6) FT synthesis and upgrading; (7) Power island.

All the operation units are the same with the CTL plant except for the feedstock preparation system which utilized a high temperature and high pressure environment to form the biomass slurry (Figure 3.2) in a batch type reactor. The application of the biomass pretreatment system is still in lab scale R&D therefore will not be discussed in detail. However, preliminary experiment results indicated that the biomass slurry is pumpable with a solid content as high as 40 wt% [53] which means a H_2O/biomass (dry basis) mass ratio of 1.5. The H_2/C molar ratio remains the same ratio of 1.0 based on simulation results. The gasification temperature remains at 750 ^{o}C at 400 psi. One key assumption in the BTL plant design is that the presence of tar and heavy hydrocarbon is not considered in the gasifier effluent although this might not be the case in reality for biomass gasification.

The BTL plant case investigated is identified as the following:

Case CERT-5

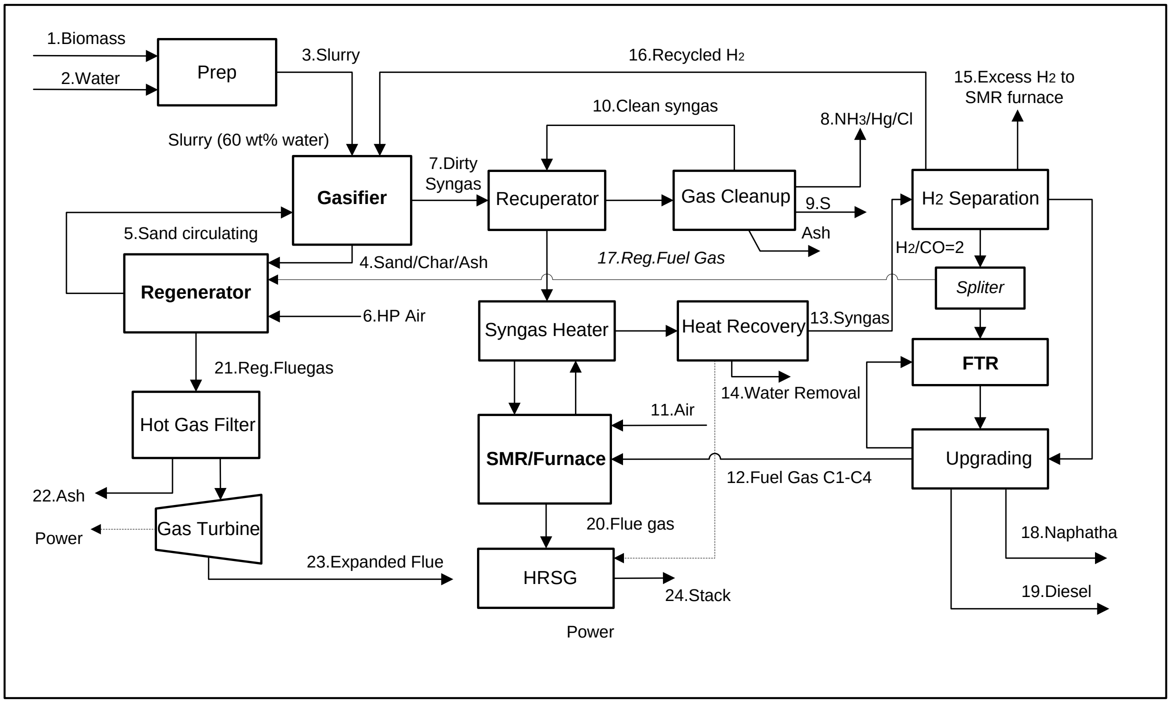
Synthetic fuels production from a 2,000 TPD biomass green waste (dry basis) demonstration plant using cobalt-based FT synthesis unit with fuels and power coproduction.

Material and Energy Balance

As with CTL cases, the material and energy balance for Case CERT-5 is based on the maximum diesel/naphtha fuels production (while allowing sufficient C1 to C4 production to fuel the SMR) from dry green waste of 1,800 TPD using a FT synthesis reactor for a ratio of 2.0 H_2/CO inlet specification.

The SHR with regenerator and warm gas cleanup configurations are the same with other cases introduced in chapter 4. The gas goes through the catalytic steam reformer tubes where the 11.8% CH_4 content in the clean syngas is reduced to 1.5% CH_4 and H_2 content of the gas is increased to 45.5%. No carbon capture option is considered in the BTL plant since the biomass feedstock is carbon neutral. Fuels for the reformer furnace are the light hydrocarbons from the FT synthesis effluent. Overall process flow diagram with individual process stream table for CERT-5 is given in Figure 5.2.

It should be noted that due to the volatile nature and low carbon content of biomass feedstock compared with coal, the leftover char in the SHR effluent only accounts for 11.7% of biomass carbon and cannot meet the heat requirement of SHR. In this case, part of the clean syngas after H_2 separation is diverted as fuel gas to compensate the heat consumption in the SHR. The syngas diversion option results in a lower fuels yield however, makes the process heat self-sufficient without considering external fuel source (e.g., natural gas and coal).



CERT-5

Stream	1	2	3	4	5	6	7	8	9	10	11	12
T (℃)	15	15	220	750	900	15	750	38	38	260	15	38
P (Bar)	1	1	32	28	28	28	28	28	28	28	28	1
m (ton/day)	1800	2700	4500	29823	29714	4359	4540	1	1	4538	2095	157
Stream	13	14	15	16	17	18	19	20	21	22	23	24
T (℃)	38	38	38	38	38	25	25	593	900	38	130	130
P (Bar)	28	1	28	28	28	1	1	1	1	1	1	1
m (ton/day)	2446	2091	4	149	618	103	207	2252	5086	5	5081	2252

Figure 5.2 Overall process flow diagram for CERT-5 plant design with process stream table

The low carbon content in the biomass also leads a low CH_4 yield and a high H_2O/CH_4 molar ratio of 4.3 can be achieved which avoids the need of additional steam in SHR and in turn reduces the plant heat burden to some large extent. The FT loop in the FT units results a high CO conversion efficiency up to 90% with the light hydrocarbons fuel the

SMR and no excess fuels gas is burnt for power generation in order to maximize the liquids yield.

Overall plant performance for CERT-5 is summarized in Table 5.6, which includes auxiliary power requirements. For Case CERT-5, the net plant output power, after plant auxiliary power requirements are deducted, is nominally 27.48 MW_e. The naphtha/diesel yield is 2430 bbl/day and the overall plant thermal effective efficiency is 51.63% on an HHV basis.

Table 5.6 Case CERT-5 plant performance summaries

Power generation (MW)	
Gas turbine	19.63
Steam turbine	37.30
Total	56.93
FT liquid production (BBL/day)	
Naphtha	862
Diesel	1,568
Total	2,430
Auxiliary load (MW)	
Biomass handling and preparation	1.90
HTP	1.20
Gasification	20.41
Warm gas cleanup	0.31
SMR	2.35
H2 separation	1.45
FTR and upgrade	1.83
Total load	29.45
Plant performance	
Net plant power (MW)	27.48
Fuels efficiency (HHV base)	44.65%
Overall thermal efficiency (HHV base)	51.63%
Specific liquids production rate (BBL/dry metric ton)	1.35
Coal feed flow rate (ton/day)	1800
Thermal input (MW)	393.75

The unique design of SHR results in 11.7% of biomass carbon left as char and about one quarter of syngas is diverted as fuel for the internal heating requirement shown in Figure 5.3. Moreover, more CO_2 is generated at the high steam environment as can be seen from the carbon balance in Figure 5.3. These carbon losses reduce the overall available carbon fixed in the FT liquids (27.6%) and, however, more electricity is generated as a result in the steam cycle.

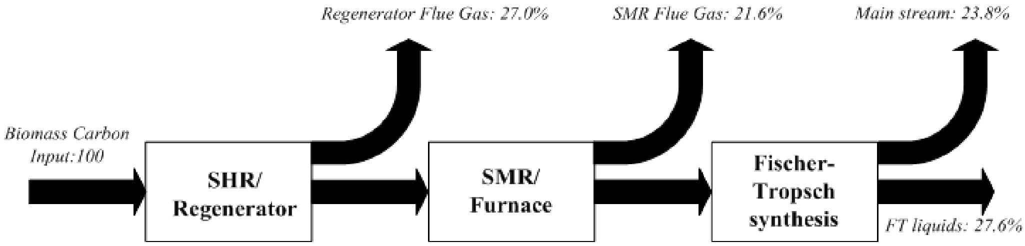


Figure 5.3 Carbon balance simulation results for CERT-5 plant design.

5.4 Economic analysis

The plant economic analysis methodology is the same as the CTL plant which uses an nth plant assumption and the capital investment structure follows the instruction introduced in chapter 3. The costs in the evaluation of SHR, warm gas cleanup, SMR and FT synthesis operation units are consistent with methods used in chapter 4. The cost of biomass hydrothermal pretreatment system is estimated from Aspen ICARUS. When a piece of equipment was scaled to a different size, a scale factor (0.6-0.9) is applied to adjust the initial cost to the design equipment cost. Generally, when green waste is sent to landfill, 30 to 60 dollar has to be paid for one ton biomass waste as tipping fee for the disposal cost [98]. In this design, trucks collect the waste within certain radius of the center plant and deliver green waste to the plant gate. Under some favorable conditions, the green waste tipping fee that would ordinarily be charged at the landfill gate would be subsidized to the BTL plant as part of the revenue since the plant helps with the diversion of waste disposal and management and converts the waste into clean energy meanwhile.

It is assumed here that the green waste feedstock cost is zero and this value is used as baseline all through the economic analysis. Electricity sale price was set as 54 \$/MWh and escalation factor of 3% was employed in biomass cost, electricity sale price and O&M expense to reflect inflation within plant lifetime. All financial values used in the economic analysis were adjusted in 2010 August dollars while by-product NH_3 and sulfur was not considered as credit in this study. Capital cost breakdown for the BTL plants is given in Table 5.7.

Table 5.7 Estimated plant capital cost (Overnight costs)

Plant economics	$MM	%
Feedstock preparation	10.6	3%
HTP	13.7	4%
Gasification island	74.8	24%
Gas cleanup	33.2	11%
SMR	56.9	18%
FTR	37.1	12%
Power island	51.7	17%
BOP	34.4	11%
Total installed cost	312.4	
Indirect cost	112.5	
EPC cost	424.9	
Contingency (20%)	85.0	
Total plant cost	509.9	
Working capital	76.5	
Financing cost	58.5	
Total Required Capital	**644.9**	

In the plant, gasification island accounts for the largest fraction of the cost, followed by SMR, power island and FT synthesis. A BOP cost accounts for 11% of TIC was used in waste water treatment and cooling tower facility along with miscellaneous cost [72]. The total required capital is 644.9 $MM and the specific capital investment for CERT-5 is 265 $K/bbl.

Table 5.8 shows the levelized FT liquids PC, also called plant-gate selling price, at 12% IRR calculated from the financial model along with corresponding BCOP introduced before in chapter 4.

Table 5.8 Levelized and breakdown FT liquids PC

Case	CERT-5
Breakdown FT liquids cost ($/gallon)	
O&M cost	0.96
Biomass purchase	0.00
Loan expense	0.60
Income tax	0.17
Capital depreciation	0.58
Electricity sales	-0.35
Average return on investment	0.54
Total	**2.51**
PC ($/Diesel Gallon Equivalent)	2.59
BCOP ($/bbl)	94.5

Based on the discount cash flow analysis, the FT liquids PC for CERT-5 is 2.51 $/gallon (2.59 $/DGE) with corresponding BCOP of 94.5 $/bbl. O&M cost accounts for more than 38% of overall PC due to the relative small capacity determined by the biomass feedstock availability. Other major costs incurred are loan expense and capital depreciation. Electricity coproduct is sold as credit and shown as negative entry in the table.

Biomass green waste cost is the most important factor in determining the whole plant economy due to its unpredictable nature (e.g., supply and policy). Based on this reason, a wide range of the biomass cost is discussed here to reflect the desirable and undesirable condition in the future shown in Figure 5.4. It should be noted that the feedstock cost in negative means that the plant is given the tipping fee for receiving the waste from individuals, facilities or government and the received tipping fee can be considered as an important revenue source through the plant lifetime.

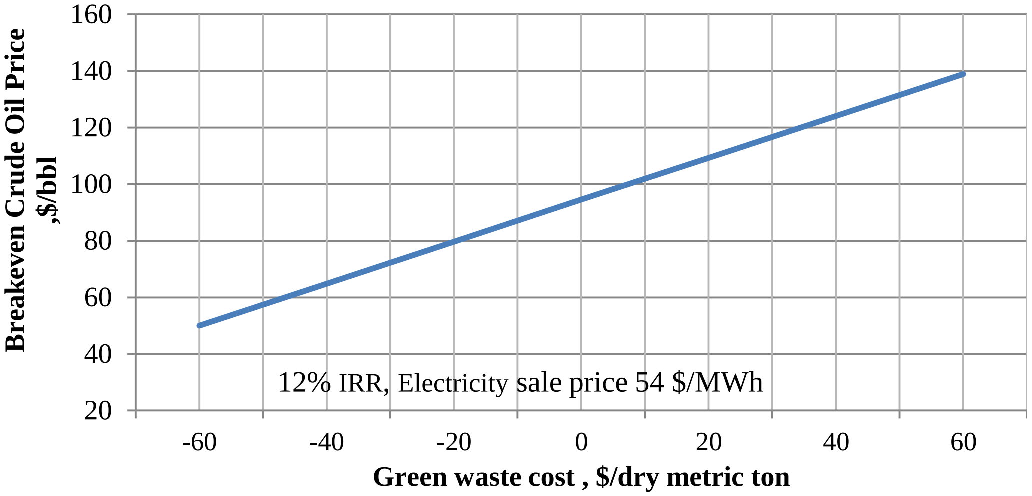


Figure 5.4 The effect of green waste cost on FT fuels PC

It can be observed that the fuels PC increases with green waste cost linearly and for a $10 change in green waste cost changes the corresponding BCOP by 7.4 $/bbl. If the biomass cost rises up to 60$/ton, the BCOP increases to 140 $/bbl and sufficient incentives have to be given to the BTL plant to make it competitive with the crude oil market.

Biomass waste transportation distance impact on the fuels PC is given in Figure 5.5. The green waste delivery radium is 50 miles for the BTL plant in the baseline. The delivery cost ($/dry ton biomass) is also converted into the FT liquids PC “$/gallon” by dividing the amount of gallon fuels derived from one dry ton of biomass. The green waste delivery cost is 0.17 $/gallon in the baseline 2,000 TPD BTL plant.

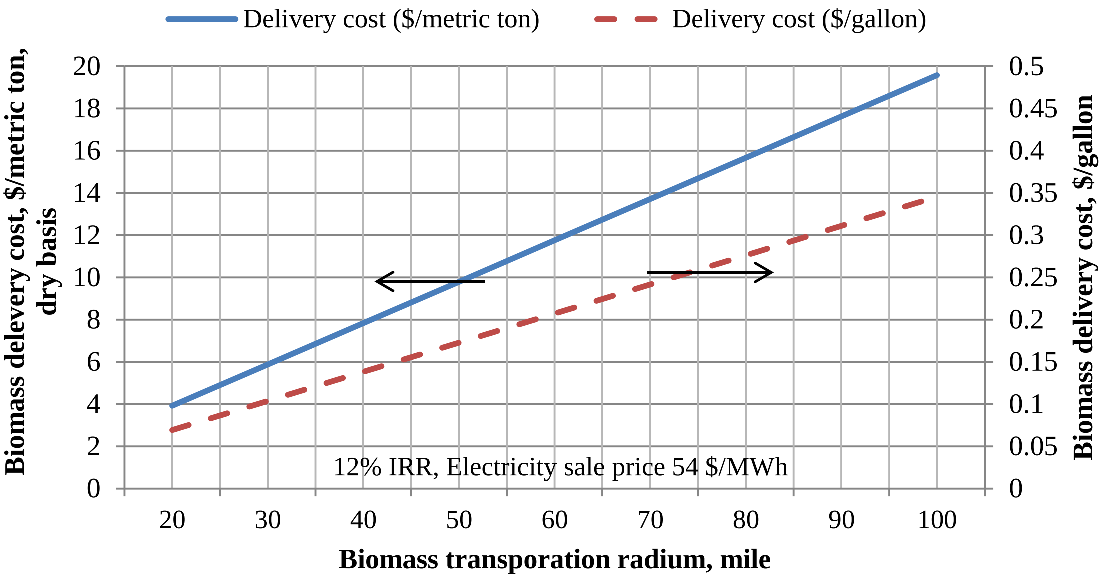


Figure 5.5 The effect of biomass waste transportation distance on fuels production cost

It is observed that the FT fuels PC increases with the biomass transportation radium linearly. For each 10 miles change in the biomass transportation radium changes the delivery cost 1.96 $/ton with corresponding fuels PC varied by 3.5 ¢/gallon.

Additionally, the size of the plant can be varied by feedstock availability and transportation distance. The overall fuels PC is the summation of the plant fuels PC with feedstock delivery cost shown in Figure 5.6. TRC of the BTL plants operating at different capacity is also given by using the plant scaling factor mentioned before and the result is schematically shown in Figure 5.7. No scale effect is considered in this plant scale up analysis which means all the operation units in the plant design have the some mechanical and thermochemical efficiency regardless the handling capacity. The biomass delivery radium for plant size of 1,000, 3,000, 4,000 and 5,000 TPD is 35, 65, 80 and 95 miles, respectively.

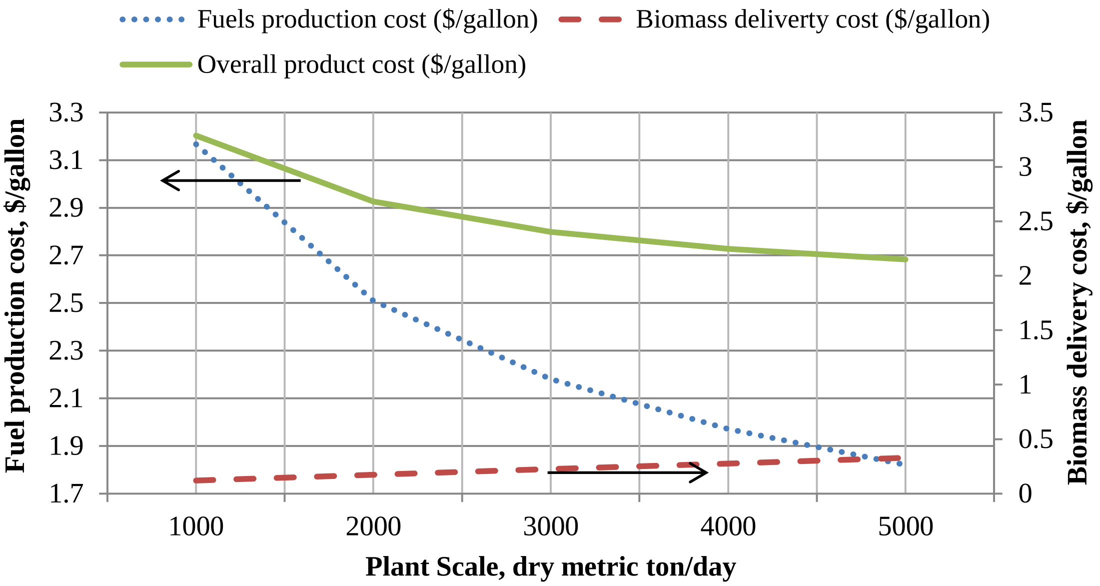


Figure 5.6 The effect of plant size on overall PC for CERT-5

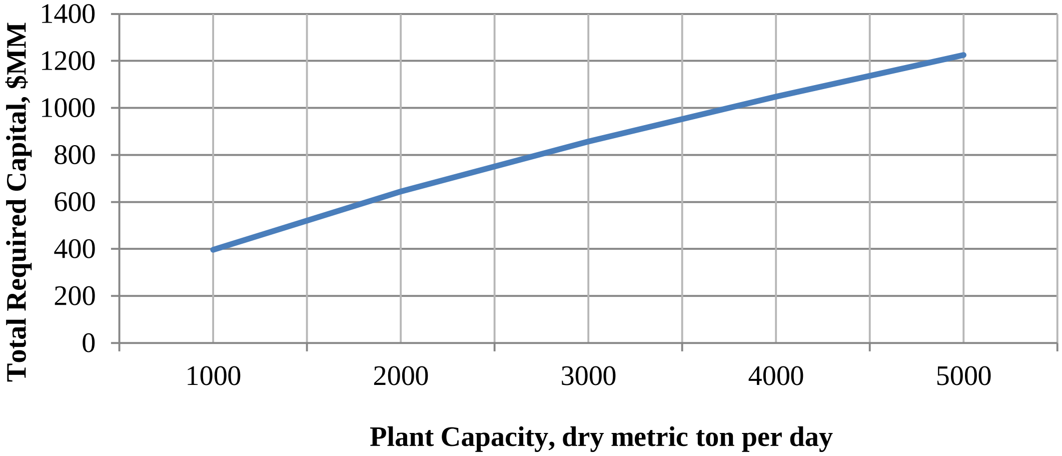


Figure 5.7 The effect of plant size on total required capital for CERT-5

When the plant size is below the 2,000 TPD baseline, the overall PC decreases sharply with the plant capacity increase. This trend is expected since the increase of plant size reduces the relative O&M cost and capital investment dramatically especially at small scale plant. As the plant size increases past the baseline, the slope of the overall PC decreases gradually and almost levels out when the plant scale is over 4000 TPD. Due to the large capital investment in the BTL plant (almost linear with plant size), it is suggested that the benefits of the lower overall PC may not be worth the significant increase of the capital cost.

Fuels PC sensitivity to several parameters is shown in Figure 5.8. The sensitivity bars are given to demonstrate the effect of different parameters on PC by varying the input value within ±25% of baseline.

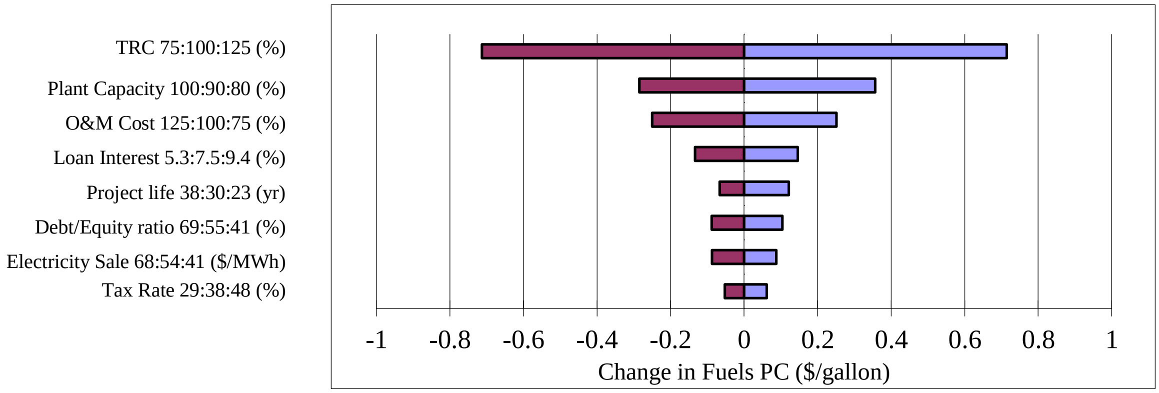


Figure 5.8 Fuels production cost sensitivity analysis with ±25% input change on baseline

As can be seen from the figure, the most influential factor is total required capital since it dominates the capital depreciation and total amount of loan expense. The PC changed by 0.71 $/gallon when TRC varied by ±25%. Plant capacity is the second most important factor that determines the total amount of liquid fuels production which is the major plant revenue. O&M cost is the third important factor due to the relatively small plant capacity compared with industrial CTL plant and PC varies 0.25 $/gallon while O&M cost changed by ±25%. Loan interest, debt/equity ratio and plant life time also have important effect on the PC greater than ±0.1 \$/gallon. Other economic parameters such as tax rate and electricity sale price has less impact on fuels PC within ± 0.1 \$/gallon.

5.5 BTL process using microalgae as feedstock

High crude oil prices, competing demands between foods and other biofuel sources, and the world food crisis have ignited the interest in making biofuels using land that is not suitable for agriculture. Microalgae are on the hotspot for next generation biofuel production due to their high lipid content and fast growth cycle than most of terrestrial plants. They can be grown with minimal impact on freshwater resources and produced using ocean and wastewater in wild water area, photo-bioreactor and inland race pond [99-101].

Normally, algae cost more per unit mass (as of 2010, food grade algae costs ~\$5000/tonne) due to high capital and operating costs [102]. It is claimed to yield between 10 and 100 times more fuel per unit area than other second generation terrestrial biofuel crops. The U.S. DOE estimates that if algae fuel replaced all the petroleum fuel in the U.S, it would require 15,000 square miles (39,000 km^2) which is only 0.42% of the U.S. map, or about half of the land area of Maine. This is less than $\frac{1}{7}$ the area of corn harvested in the U.S. in 2000 [103, 104]. Currently, researches are mainly focused on developing algae species with lipid content greater than 25% and mass yield higher than 80 dry ton per year per hectare. The production cost varies a lot (100-1000 \$/dry ton) based on cultivation methods, nutrient source, harvesting method and cultivation scale.

Microalgae can provide several different types of renewable biofuels including methane produced by anaerobic digestion and biodiesel derived from microalgae lipid. Among these applications, biodiesel production is the most widely used and a well-developed commercial mature technology that suitable for massive fuels production. The microalgae-to-biodiesel production process is the same as conventional biodiesel plant using vegetable oil and animal fat through transesterification reaction and the process schematic flow diagram is given in Figure 5.9.

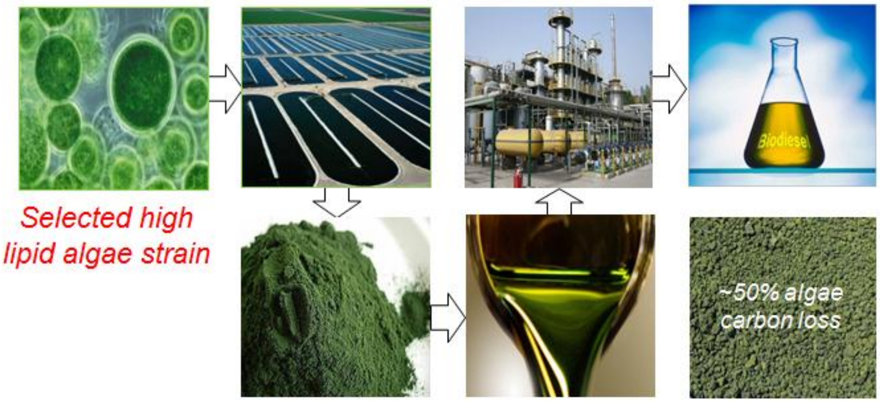


Figure 5.9 Schematic flow diagram for biodiesel production from microalgae

The microalgae-to-biodiesel process economy is strongly constrained by the upstream microalgae feedstock while algae strain with high lipid content is preferred since the rest part of microalgae is hydrocarbon and protein that have nothing to do with the biodiesel yield. The other shortcoming for biodiesel production from microalgae is that the feedstock has to be dried sufficiently in order to reduce the energy consumption in the lipid extraction and the drying process also requires additional power and heat consumption. The whole process can only convert less than 50% carbon in the microalgae to biodiesel and the rest carbon is left as microalgae cake (hydrocarbon and protein) that is normally used as animal feed thus cannot be used for thermal energy production.

A new thermochemical process for converting microalgae to synthetic fuels is developing based on steam hydrogasification. The goal of this study is to produce sustainable liquid transportation fuels with electricity coproduction from microalga and the process is schematically shown in Figure 5.10 (Case CERT-6).

Figure 5.10 Schematic flow diagram for synthetic fuels production from microalgae

The detail conversion process is the same as green waste-to-liquid process introduced earlier and algae stain with high mass yield and carbon content, no matter in the form of protein, lipid and hydrocarbon, can be used as feedstock theoretically. All the carbonaceous components are broken down into small molecules (e.g., CH_4, CO and CO_2) at high temperature and pressure gasification environment. The entire mass of the algae with high moisture content can be used as feedstock which enables the operation favorable and profitable at relatively small scale plant. Moreover, the resulting CO_2 in the flue gas can be used as carbon source and nutrient for microalgae cultivation which decreases the algae grow cost greatly.

Microalgae feedstock cost and plant capacity, the two most unpredictable and dominant factors in the BTL process are investigated to show the impact on the fuels PC. To simplify the study, no detailed process modeling is performed and instead, the plant performance and capital cost of the green waste-to-liquid plant (Case CERT-5) is used directly. This assumption is reasonable since the chemical composition and heating value of the green waste and microalgae are almost close to each other. Under these assumptions, the 1800 TPD microalgae-to-liquid plant has overall fuels yield of 2430 bbl/day with 27.48 MW electricity export to the grid and the TRC is 644.9 $MM.

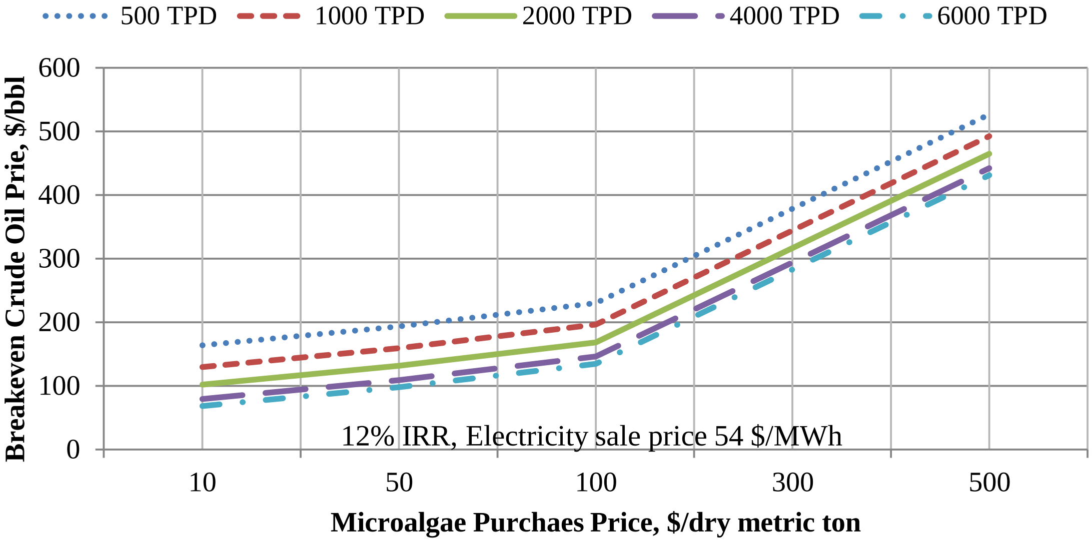


Figure 5.11 The effect of microalgae price and plant size on the fuels PC in CERT-6

The impacts of microalgae purchase price and plant scale on the FT fuels PC are given in Figure 5.11. It is observed that when the plant size is below the 2,000 TPD baseline, the fuel PC decreases sharply with the plant capacity increase at the same microalgae purchase price. This trend is the same with that in the green waste-to-liquid scenario due to less O&M cost and capital investment per gallon fuel produced in large scale plant. As the plant size increases past the baseline, the fuel PC decreases slowly within 10 $/bbl.

The fuel PC increases dramatically when the microalgae cost is beyond 100 $/ton as can be seen from the rising slope. This can be explained as the high feedstock cost requires higher return on investment to achieve the desired IRR thus increases the PC greatly.

5.6 Summary and conclusion

Biomass green waste-to-liquid process based on steam hydrogasification is mainly discussed in this chapter. Plant performance and economic analysis are given according to the process simulation results and financial model outputs.

Harvesting distance and transportation distance between feedstock sources and the BTL plant are one of the key factors to determine the profitability of the process. The biomass transportation model used here is based on the homogeneous distribution assumption however, determining the actual location of these feedstocks is essential in calculating the actual transportation distance in harvesting and collecting the feedstocks and transporting them to the plant to be built in SC. A Geographic Information System (GIS) can be used for this purpose to determine the optimal distance between biomass sources and the conversion plant in future study.

The 2000 TPD green waste-to-liquids plant is expected to produce fuels costing in the range of $1.5-$3.6 per gallon diesel equivalent with biomass waste cost varied from -$60 to $60 per dry metric ton using steam hydrogasification technology. The factors chiefly responsible for this PC are O&M cost, loan expense and return on investment in the 650 $MM BTL plant to process 1,800 metric tons of biomass green waste per day. When green waste transportation cost is considered, the product cost increases by 0.17 $/gallon at a feedstock delivery radius of 50 miles.

Parameters with the most sensitive effect on feels PC are TRC, plant capacity, and O&M cost, affecting the PC ±$0.20–$0.80 per gallon. The biomass waste cost is of vital importance in determining the plant economy and also the most unpredictable factor compared with other parameters. Research on the biomass waste distribution, disposal cost, and management policy will be focused in future study. It is desirable that a biomass waste tipping fee is given to the plant as part of the revenue so that it can compete with the crude oil market without any other incentives.

The CE-CERT process provides a new pathway to convert the high moisture content microalgae to transportation fuels through thermochemical method. The whole algae feedstock instead of the lipid is converted to fuels and power directly which improves the energy conversion efficiency greatly. Since CE-CERT process only focuses on the carbon content, algae species with mass yield of 200-300 dry ton per year per hectare are expected to be found in the future that reduces both the cultivation area usage as well as fuels PC. It has been widely accepted that if the cost of cultivation and harvest of microalgae could be reduced to as low as $50 per dry ton, it could become a significant source of fuels to replace fossil energy. Further economic benefit comes with improvement from commercial scale demonstration efforts on steam hydrogasification.

Plant effect analysis shown that BTL plant with large capacity (>2,000 TPD) decreases the fuels PC to some extent but may not receive much profit as expected due to the increased feedstock transportation cost and capital investment.

Because of time and resource constraints, the BTL process study presented includes a few shortcomings. The process configuration is not fully optimized for heat integration and tar handling (e.g. catalytic cracking and water scrubbing) is not included to simplify the process modeling. Further studies can benefit by optimizing heat exchange networks and modeling in detail areas.

This analysis tracks PC based on commercial technology for which sufficient public domain data existed in 2010. In order to evaluate the economics of fuels produced from steam hydrogasification based on future scenarios, particular attention will be needed on the most sensitive parameters-feedstock cost and capital cost. Attention should also focus on other commercial biomass-to-fuels gasification plants that come online. As these plants do come online, cost growth and plant performance factors will improve, thereby decreasing the plant PC.

6. Synthetic natural gas production from biomass and biosolid

6.1 Process description

Synthetic natural gas, also called substitute natural gas, is a sustainable gas from coal or biomass with natural gas specifications. Therefore, it can be transported through the existing gas infrastructure. There are many different issues associated with the deployment of SNG. Interest in developing SNG dated back to the 1970s when the energy crisis led researchers and policymakers to consider ways to convert coal into gaseous and liquid fuels. However, the later stabilization of the fuel market and increased availability of low-cost fuels led to the abandonment of most of coal-to-SNG projects [17].

Increasing demand for natural gas and high natural gas prices in the recent past has led many to pursue unconventional methods of natural gas production. According to DOE (2008), 90% of new U.S. power plants will be natural gas–fired plants. The ever increasing demand and high price of natural gas in recent past has led researchers to consider alternate methods of natural gas generation [104]. In California, natural gas provides almost one-third of the state's total energy requirements and will continue to be a major fuel in California's supply portfolio [105].

In this chapter, green waste biomass and biosolid commingle feedstock is used for SNG production based on steam hydrogasification technology. Biosolid, also refers to treated sludge, is generated from both aerobic and anaerobic waste water treatment process during sewage and wastewater treatment. The biosolid yield is a function of population within certain area (50-65 dry grams per person per day) and can be supplied continuously all through the year. Statistic has shown that the biosolid yield in SC is 396,200 BDT in 2007 which accounts for 50.4% of overall biosolid produced within CA [89, 90]. Generally, biosolid can be used for landfill or as feedstock in anaerobic digestion process. The landfill option usually brings heavy financial burden to the wastewater treatment facility since it has to pay certain amount of tipping fee for the waste disposal (20-50 $/ton). Current bacterial treatment of biosolid has very low energy conversion efficiency (< 35%) with limited treatment capacity and the biogas produced has very low CH_4 content and BTU value which can only be used for heat or power generation.

After years of dedicated research, it is found that the introduction of biosolid in biomass steam hydrogasification process can increase the rate of CH_4 formation greatly due to the metallic compound in the biosolid that behaves as catalyst. The high moisture content in biosolid (90 wt%) can also provide the necessary water requirement during steam hydrogasification. The usage of biosolid and green waste for SNG production can not only avoid the high waste disposal cost but also provides a potential method to dispose waste efficiently by converting it into green energy. The process flow diagram is schematically shown in Figure 6.1.

Initially, the green waste is made into slurry with biosolid in HTP process after proper preparation. The slurry feed is then pumped into the gasifier along with H_2 and a regenerator is setup to provide the necessary process heat of gasifier by combusting the leftover char or fuel gas if needed. The CH_4 rich gas from the gasifier is then subjected to a cleanup unit in order to remove contaminants, primarily sulphur species. The CO component is then converted into H_2 and CO_2 in the WGS reactor. Finally, a gas conditioning unit is used to separate and cycle excess H_2 back to gasifier as feed after H_2O condense and CO_2 removal. Process heat is recovered in the HRSG for power generation in steam cycle. The clean and high quality SNG product could be used in natural gas vehicle in transportation field therefore meets the low carbon fuel standard in California.

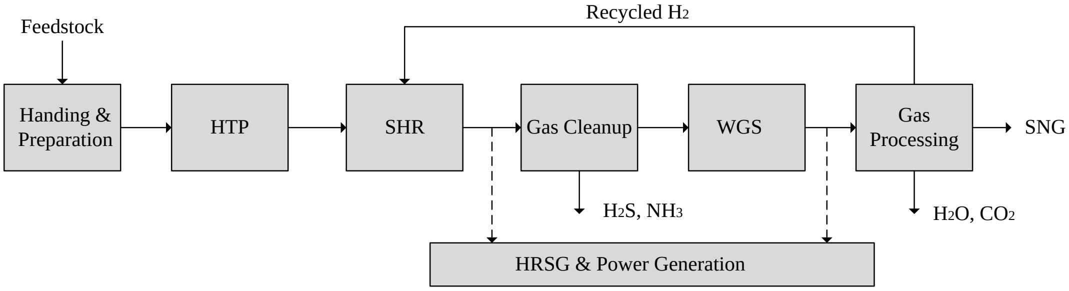


Figure 6.1 Process flow diagram of SNG production based on steam hydrogasification

Based on the feedstock feasibility, if the entire technical available portion of green waste and biosolid in SC is used for SNG production via steam hydrogasification technology, 15,357 Million Cubic Feet (MCF) can be generated using the feedstock yield in 2007 which accounts for about 4.9% of the natural gas produced in CA. As the yields of biosolid and green waste increased in SC in the next 10 years, the SNG yield can be expected to rise up to 17,997 MCF in 2020 that accounts for 5.8% of the natural gas produced in CA [105].

6.2 SNG process design and plant performance

The design is based on a geographic location in the waste water treatment plant in SC (Los Angeles County, Orange County, Riverside County and San Bernardino County), and biosolid with 90 wt% moisture content can be used as water source directly which avoids the biosolid dewatering cost as well as transportation cost. Chemical composition of biosolid used in this design is given in Table 6.1 collected from Riverside waste water treatment plant. The same assumptions used in the green waste-to-liquid process are applied in this section which means the feedstock has homogeneous and constant property and tar formation and removal are not considered in the design.

SNG plant with design capacity of 2,300 TPD (dry basis) is selected which utilizes 1,800 TPD green waste and 300 TPD biosolid as commingle feedstock. The plant is designed that consists of six major sections: (1) Feedstock handling and pretreatment; (2) Gasification island; (3) Warm gas cleanup; (4) WGS; (5) Gas conditioning; (6) Power island.

Table 6.1 Chemical composition of biosolid

	Compound	Biosolid
Proximate	Volatile Matter	60.25
	Fixed Carbon	10
	Ash	29.75
Ultimate	C	29.57
	H	5.39
	O	20.83
	N	5.79
	S	1.56

All the operation units such as HTP, SHR with regenerator setup and warm gas cleanup are the same with the BTL plant except for the gas shifting which utilizes a WGS reactor instead of SMR to convert CO into H_2 so that the H_2 can meet the internal cycle requirement while CH_4 can be left as product. 1 ton of fresh green waste (25 wt% water) mixes with 1.17 ton of biosolid (90 wt% water) to form the 40 wt% slurry in the HTP.

The H_2/C molar value of 1.0 is used which is consistent with that in BTL process. The gasification temperature remains at 750 ^{o}C at 400 psi based on simulation results.

The SNG plant case investigated is identified as the following:

Case CERT-7

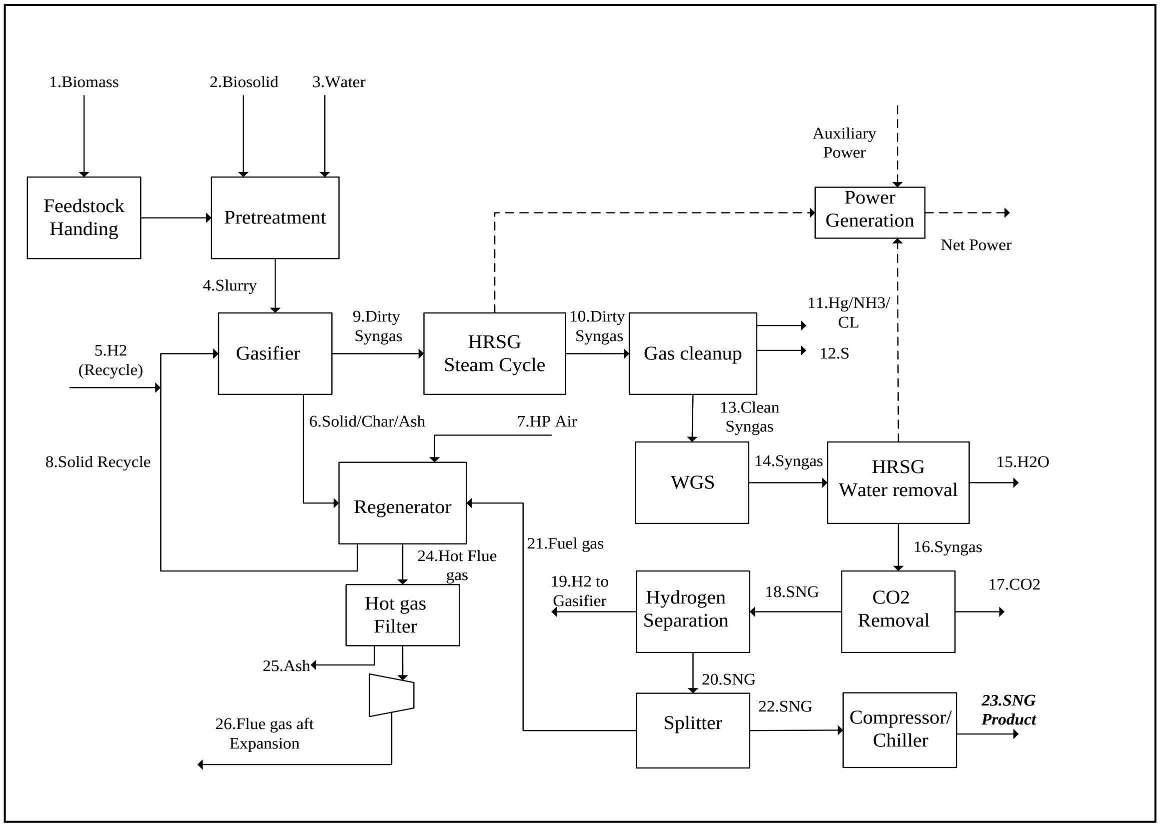
SNG production from a commercial size 2,300 TPD biomass green waste and biosolid (dry basis) plant with SNG and power coproduction.

Material and Energy Balance

The material and energy balance for Case CERT-7 is based on the maximum SNG production (while allowing sufficient H_2 for internal cycle) from the 2,100 TPD green waste and biosolid (dry basis) plant with WGS specification. The SHR with regenerator and warm gas cleanup configurations are the same with Case CERT-5. The gas goes through the catalytic WGS reactor where the 2.4% CO content in the clean syngas is reduced to 0.2% and H_2 content of the gas is increased to 27.6%.

Overall process flow diagram with individual process stream table for CERT-7 is given in Figure 6.2. Overall plant performance of CERT-7 is summarized in Table 6.2, which includes auxiliary power requirements.

Due to the low carbon content and heating value of biosolid feedstock, 32% of the clean syngas after H_2 separation is diverted as fuel gas to compensate the heat consumption in the SHR. The syngas diversion option lowers SNG yield however, make the process heat self-sufficient without considering external fuel source (e.g., natural gas and coal). The clean SNG product with CH_4 content higher than 98% can be obtained after CO_2 removal and water condense. Process heat recovered in the syngas cooling is used in HRSG for power generation.



CERT-7

Stream	1	2	3	4	5	6	7	8	9	10	11	12	13
T (℃)	15	15	15	180	38	750	15	900	750	350	15	15	350
P (Bar)	1.0	1.0	1.0	27.6	1.0	27.6	27.6	27.6	27.6	27.6	1.0	1.0	27.6
m (ton/day)	1800	300	3150	5250	168	28513	5468	28307	5212	5212	21	6	5185
Stream	14	15	16	17	18	19	20	21	22	23	24	25	26
T (℃)	350	38	38	38	38	38	38	38	38	-30	900	38	137
P (Bar)	27.6	1.0	27.6	1.0	27.6	27.6	27.6	27.6	27.6	27.6	27.6	1.0	1.0
m (ton/day)	5185	2818	2367	1579	788	168	620	198	422	422	5873	89	5783

Figure 6.2 Overall process flow diagram for CERT-7 plant design with stream table

For Case CERT-7, the net plant output power, after plant auxiliary power requirements are deducted, is nominally 32.90 MW_e. The SNG yield is 23.88 MSCF/day (19,848 MMBTU/day) and the overall plant thermal efficiency is 61.22% on an HHV basis.

Table 6.2 Case CERT-7 plant performance summaries

Power generation (MW)		
	Gas turbine	22.90
	Steam turbine	39.50
	Total	62.40
SNG production (MMBTU/day)	SNG	19,848
Auxiliary load (MW)		
	Biomass handling and preparation	1.90
	HTP	1.40
	Gasification	23.81
	Warm gas cleanup	0.36
	WGS	0.20
	Gas conditioning	1.82
	Total load	29.49
Plant performance		
	Net plant power (MW)	32.90
	SNG efficiency (HHV base)	53.90%
	SNG production (MSCF/day)	23.88
	Overall thermal efficiency (HHV base)	61.22%
	Specific SNG production rate (KSCF/dry metric ton)	11.37
	Feed flow rate (ton/day)	2100
	Thermal input (MW)	449.7

6.3 Economic analysis

The economic analysis methodology in the SNG plant is the same with that in the BTL plant which uses an nth plant assumption and the capital investment structure follows the instruction introduced in chapter 3. The cost of WGS reactor was estimated from Aspen ICARUS. When a piece of equipment was scaled to a different size, a scale factor (0.6-0.9) is applied to adjust the initial cost to the design equipment cost.

It is assumed here that the green waste and biosolid mixture cost is zero and this value is used as baseline all through the economic analysis. Electricity sale price was set as 54 $/MWh and escalation factor of 3% was employed in biomass cost, electricity sale price and O&M expense to reflect inflation within plant lifetime. All financial values used in the economic analysis were adjusted in 2010 August dollars while by-product NH_3 and sulfur were not considered as credit. Capital cost breakdown for the SNG plant is given in Table 6.3.

Table 6.3 Estimated SNG plant capital cost (Overnight costs)

Plant economics	$MM	%
Feedstock preparation	10.6	4%
HTP	15.3	6%
Gasification island	83.3	33%
Gas cleanup	37.0	15%
WGA	9.4	4%
Gas conditioning	13.7	5%
Power island	53.8	21%
BOP	27.6	11%
Total installed cost	250.7	
Indirect cost	90.2	
EPC cost	340.9	
Contingency (20%)	68.2	
Total plant cost	409.1	
Working capital	61.4	
Financing cost	47.0	
Total Required Capital	**517.4**	

Like the BTL process, gasification island accounts for the largest fraction of the cost, followed by power island and gas cleanup. The total required capital is 517.4 $MM and the specific capital investment for CERT-7 is 26.1 $K/MMBTU.

Table 6.4 shows the levelized SNG production cost at 12% IRR calculated from the financial model using baseline parameters.

Based on the discount cash flow analysis, the SNG production cost for CERT-7 is 8.87 $/MMBTU (7.37 $/KSCF). O&M cost accounts for more than 37% of PC due to the relative small plant capacity determined by the feedstock availability. Other major costs incurred are loan expense and capital depreciation. Electricity coproduct is sold as credit and shown as negative entry in the table.

Table 6.4 Levelized and breakdown SNG production cost

CERT-7	SNG
Breakdown SNG PC ($/MMBtu)	
O&M cost	3.25
Feedstock purchase	0.00
Loan expense	2.47
Income tax	0.57
Capital depreciation	2.40
Electricity sales	-2.15
Average return on investment	2.33
Total	**8.87**
SNG PC ($/KSCF)	7.37

Biomass green waste and biosolid cost is the most important factor in determining the whole SNG plant economy. Feedstock cost ranges from -60 $/ton to 60 $/ton (dry basis) is discussed to reflect the desirable and undesirable condition in the future shown in Figure 6.3. The value in negative means that the SNG plant is given certain amount of tipping fee for receiving the waste from individuals, facilities or government and the received money can be considered as an important revenue source through the plant lifetime.

It can be observed that the SNG PC increases with green waste and biosolid cost linearly and for a $10 change in the commingle feedstock changes the SNG production cost by 0.88 $/KSCF. If a 60 $/ton feedstock tipping fee is given to the SNG plant, SNG sale price as low as 2.1 $/KSCF can be achieved to receive the desired 12% IRR which is very attractive.

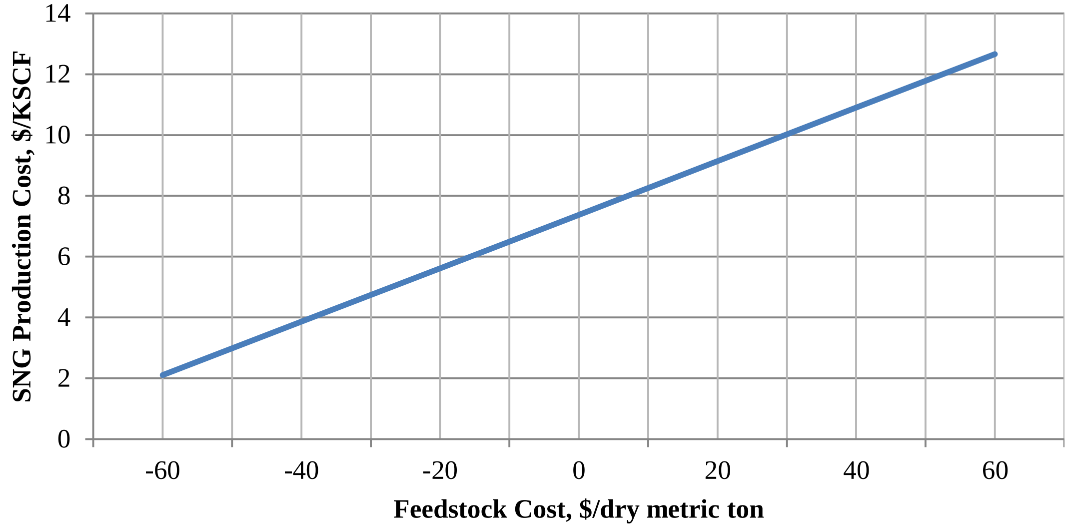


Figure 6.3 The effect of feedstock cost on SNG PC

The impact of feedstock tipping fee on SNG sale price, also known as plant gate price, in achieving different target IRR is given in Figure 6.4. The SNC sale price increases with IRR slightly parabolically and for each 1% change in IRR requires the SNG sale price varied by 0.53 $/KSCF averagely.

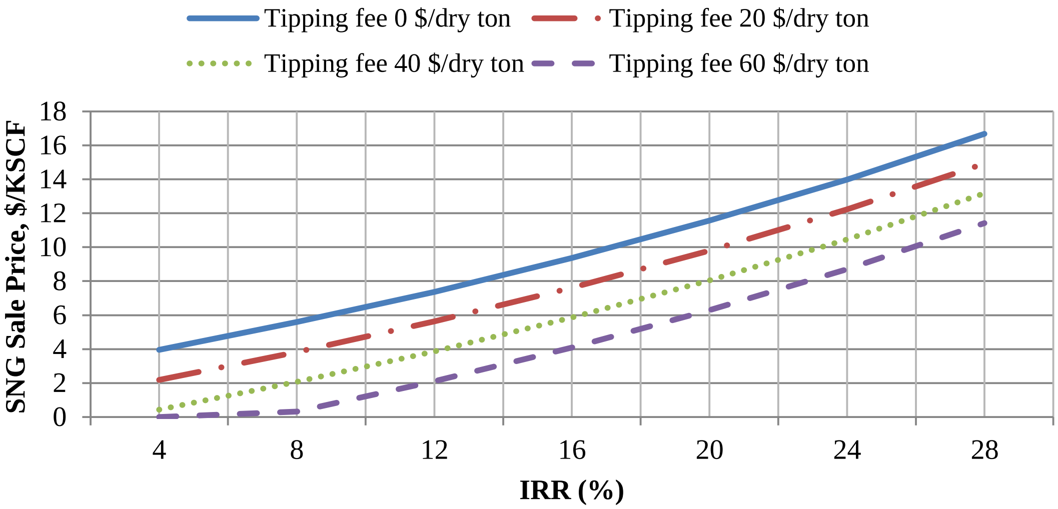


Figure 6.4 The effect of feedstock tipping fee on SNG sale price at different IRR

6.4 Summary and conclusion

A new thermochemical process for SNG production using green waste and biosolid based on steam hydrogasification is discussed in the chapter with process modeling and economic analysis. This one step SNG production process improves the energy conversion efficiency large without the need of catalyst or methanation step.

The 2300 TPD green waste and biosolid-to-SNG plant is expected to produce SNG costing in the range of \$2.53-\$15.23 per MMBTU (\$2.01-\$12.66 per KSCF) with feedstock cost varied from -\$60 to \$60 per dry metric ton. The factors chiefly responsible for this SNG PC are O&M cost, loan expense, capital depreciation and return on investment in the 520 \$MM SNG plant to process 1,800 metric tons of biomass green waste and 300 metric tons of biosolid per day.

Like BTL process, economic analysis has shown that the green waste and biosolid to SNG process relies on the feedstock tipping fee heavily. This is expected since the natural gas has relative low volume energy density and thus much cheaper compared with synthetic liquid fuels. The other factor that impact the plant economy largely is the natural gas price since the SNG has to compete with the natural gas market to gain profit. The natural gas price in California in different consumption sections from year 2006 to 2011 is given in Table 6.5 [106]. The SNG process could be feasible and profitable if the product is sold for residential, commercial and industrial usage at relative high price. Since the plant is geographic located near the waste water treatment facility as well as commercial and residential area, it is desirable to supply the SNG to the consumers within the area which avoids the product distribution cost compared with natural gas supply. The process can be more favorable if the waste tipping fee higher than 40 $/ton is given to the plant to promote the process IRR higher than 20%.

Table 6.5 Natural gas price in California in different consumption area

California natural gas price	2006	2007	2008	2009	2010	2011
Wellhead Price	6.47	6.62	8.38	3.96	4.87	n/a
Citygate Price	6.76	6.82	8.11	4.17	4.86	4.47
Residential Price	11.79	11.57	12.75	9.43	9.92	9.92
Commercial Price	10.43	10.2	11.75	7.75	8.3	8.27
Industrial Price	9.3	9.07	10.8	6.56	7.02	7.04
Vehicle Fuel Price	7.92	7.72	11.32	7.61	5.55	n/a
Electric Power Price	6.71	6.72	8.23	4.44	4.99	4.7

Like BTL plant design, the process configuration is not fully optimized for heat integration and tar handling (e.g., catalytic cracking and water scrubbing) is not included to simplify the process modeling. Further studies can benefit by optimizing heat exchange networks and modeling in detail areas.

7. Conclusion and future work

This section summarizes the conclusions derived from the process design, simulation and economic analysis performed as part of this thesis. Suggestions for further research are also provided.

Process model using Aspen Plus simulation tool has been developed for different configurations. The simulation results have been used to determine process operation condition and process performance as well as to understand process behavior. Some of the conclusions derived from the simulation work are listed below.

1. Based on the Aspen Plus simulation results for SHR modeling, the optimum gasification temperature has been determined at 750 ^oC with H_2O/feedstock mass ratio of 2.0 and H_2/C mole ratio of 1.0 using coal as feedstock.
2. Based on the Aspen Plus simulation results for BTL process modeling, it can be concluded that a H_2/C ratio of 1.0 and H_2O/feedstock ratio of 1.5 provides optimal feed conditions. 25% of clean syngas is diverted as fuel in the regenerator to supply the process heat in the SHR.
3. Based on the Aspen Plus simulation result for SNG process modeling, 32% of clean syngas is diverted as fuel in the regenerator to supply the process heat in the SHR.

Conceptual design of commercial scale plants have been developed for synthetic fuels and SNG production and the simulation results have been used as major input for process economic analysis in the financial model. The equipment cost in the designated synthetic fuel plant and SNG plant is evaluated based on Aspen Plus ICARUS software, vendor quote, technical report and other literature. Total required capital is determined by adding the total installed cost, indirect cost, project contingency, working capital and financing cost. Fuels or SNG PC at 12% IRR is then estimated based on discount cash rate on return analysis.

The electricity coproduct in the iron-based CTL plant reduces the fuels production cost largely and is especially favored when the electricity value is high. While for the cobalt-based CTL plant design, more FT fuels are produced with less power coproduction which is preferred when the crude oil price is high. For small scale CTL facility, the expensive plant capital investment, low fuel yield and high O&M expense result in an extremely high PC.

The BTL and SNG plant rely on the feedstock cost heavily in determining the process economy due to the small plant capacity with high operation cost and capital investment. Financial incentives such as tax incentives, loan guarantees, waste tipping fee, and other mechanisms play very important role in addressing the economic and market challenges of biomass derived fuels. Biomass plant with large capacity (>2000 TPD) decreases the liquid fuels PC to some extent but may not receive much profit as expected due to the high feedstock transportation cost and capital investment.

The process simulation and economic analysis results presented here demonstrate that the steam hydrogasification technology could potentially provide an effective pathway to convert coal and biomass to fuels with high conversion efficiency and less capital cost. The steam hydrogasification process appears to be suitable for commercialization in large scales with a coal feedstock and also in a distributed network of small scale facility utilizing localized renewable feedstocks.

Future work

The research on the CE-CERT process will be focused on feedstock hydrothermal pretreat, steam hydrogasification reactor and warm gas cleanup in the near term. Process feasibility must be validated in demonstration or small pilot plant in order to gain the knowledge and collect kinetic data to modify the process model and move towards to a larger pilot or a pre-commercial unit. This is especially important for feedstocks such as biomass, biosolid and other MSW. The important areas of focus for further research and development are listed below.

1. Develop kinetic model for SHR simulation based on experiments results in determining the carbon conversion efficiency, gas composition and tar yield. Dynamic modeling of the circulating fluidized bed type SHR & regenerator with mass transfer coefficient and temperature profile are necessary to further understand and predict the process behavior.
2. The co-products in the plant design such as NH_3, element sulfur and especially for the captured high pressure CO_2 stream are not considered as credit in the thesis. Further process benefits come with sale of these products for chemical synthesis or enhanced oil recovery.
3. Harvesting distance and transportation distance between feedstock sources and the biomass plant are one of the key factors to determine the profitability of the process. Determining the actual location of these feedstocks is essential in

calculating the actual transportation distance in harvesting and collecting the feedstocks and transporting it to the plant. A Geographic Information System (GIS) can be used for this purpose to determine the optimal distance between biomass sources and the conversion plant in future study.

4. It should be noticed that a lower fuels or SNG production cost doesn't guarantee a higher IRR in the plant design. Optimum operation condition and process design to maximize plant profits still needs to be further studied based on the data collected from future work.
5. The biomass waste and biosolid cost is of vital importance in determining the plant economy and also the most unpredictable factor compare with other parameters. Research on the biomass waste distribution, disposal cost, and management policy will be focused in future study. It is desirable that a biomass waste tipping fee is given to the plant as part of the revenue so that it can compete with the crude oil market without any other subsidies.
6. Because of time and resource constraints, the conceptual plant design with process optimization and modeling presented includes a few shortcomings. The process optimization is only performed for individual operation unit. The optimal plant operation conditions are determined by combining each optimized units together. A globe process optimization is necessary to conduct by considering all the process parameters within the plant (e.g., temperature, pressure and feed flow rate) using numerical optimization method.

7. The process configuration is not fully optimized for heat integration and tar handling (e.g., catalytic cracking and water scrubbing) is not included to simplify the process modeling. Further studies can benefit by optimizing heat exchange networks and modeling in detail areas.

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Appendix A. Technoeconomic financial model assumptions

Major financial model entries

- Capital structure
    - Percentage debt: 70% for coal plant, 55% for biomass plant
    - Percentage equity: 30% for coal plant, 45% for biomass plant
- Project debt terms
    - Interest rate: 7.5%
    - Financing fee: 3%
    - Repayment term (in Years): 15
    - Grace period on principal repayment: 1
    - First year of principal repayment: 2017
- Depreciation model
    - Salvage value for general plant and steam/power plant: 0
    - Type of depreciation: Double-Declining-Balance (DDB) depreciation method
    - General plant depreciation period (in years): 7
    - Steam/Power system depreciation period (in years): 20
- Working capital
    - Days receivable (in days): 30
    - Days payable (in days): 30
    - Initial working capital (% of EPC): 20%

Economic assumptions

- Cash flow analysis period
    - Plant economic life (in years):30
    - (Expected IRR) Discount rate: 12%
- Escalation factors
    - Electricity: 3.0%
    - FT liquids: 3.0%
    - SNG: 3.0%
    - Coal: 3.0%
    - Biomass & Biosolid: 3.0%
    - Variable O&M: 3.0%
    - Fixed O&M: 3.0%
- Tax assumptions
    - Income tax rate: 38%
- Construction schedule
    - Construction start date: 1/1/2013
    - Construction period (in months): 36
    - Plant start-up date: 1/1/2016
- Percentage of cost for construction periods
    - Construction year 1: 10%
    - Construction year 1: 60%
    - Construction year 1: 30%
- Plant Ramp-up assumptions (% of full capacity)

    - Year 1: 51%
    - Year 2: 78%
    - Year 3: 90%
- Operation hours
    - Annual operation hours: 7884

Capital Cost

- Cost year for analysis: 2010 August dollars. When cost data were not available in 2010 dollars, costs were adjusted with Chemical Engineering's Plant Cost Index.
- The plant is designed based on the state of the technology at the nth plant level of experience.
- Most equipment installation factors are applied using Aspen ICARUS software and chemical process design handbooks for solid-fluid plants [68].
- Materials of construction are carbon steel, stainless steel, alloys, and refractory where necessary.
- Sensitivity parameters involving changes in equipment size or capacity use scaling exponents available in literature.

Capital costs and operating expenses

Engineering, Procurement, and Construction (EPC) Costs

The EPC cost category includes all relevant direct costs, indirect costs, and design services. It can be defined as the equipment purchase and installation plus all detailed design, construction, and project management costs.

- Direct cost include: process equipment, on-site facilities and infrastructure that support the plant and the direct labor required for their installation and/or construction at the site.
- Indirect cost elements cover all field costs (materials, subcontracts, manual and non-manual labor), which cannot be specifically assigned to items in the direct cost category. The indirect field costs include temporary facilities, construction equipment, labor, field office costs, and consumable supplies.

Process Contingency

Process contingency is designed to compensate for uncertainty in cost estimates caused by performance uncertainties associated with the development status of one or more plant sections. Usually, this is not applied to the whole plant, but only to the technologically developing units such as gasification (15%) or warm gas cleanup (20%). The process contingency allowances range from 0 to 40% of the plant section, with the value depending upon the technology status.

Project Contingency

The project contingency category covers all unforeseen costs that may impact the construction cost of a project. Contingency funds are expected to be spent. In the model, project contingency costs are calculated as a percentage of total EPC costs. Although contingency factors vary by project, a 20% contingency factor is set as an initial default value due to the process complexity.

Total Plant Cost (TPC)

TPC is the sum of the EPC cost, process contingency and project contingency.

Start-up Costs

Start-up costs include labor, materials, and consumable items directly linked to the start-up of a plant. This includes all start-up capital cost items (including chemicals and catalysts). For the purposes of analysis conducted using this model, the start-up cost of a project is calculated as a percentage of total EPC costs. As an initial default value, start-up costs were set equal to 2% of total EPC costs.

Working Capital (WC)

Initial working capital needs were assumed to be equal to 15% of TPC. Days payable and accounts receivable were both assumed to be 30 days. Working capital is calculated in each year as the sum of accounts receivable, inventories, operating cash, less accounts payable. It reflects the amount of capital that is tied up in receivables, money invested in inventories, and payables, plus cash on hand. Working capital is the fund that is set up in the year prior to operations to initially fund the account.

Interest During Construction

Interest charges accumulated during the construction period.

Financing Fees

Additional fees associated with the debt portion of financing.

Total Required Capital (TRC)

TRC is the sum of the TPC, start-up costs, working capital, interest during construction and financing fee.

Variable O&M

Variable costs are dependent on the output level at a given plant. Variable O&M costs include all consumable items, spare parts, and labor that fluctuate with the actual plant output. Variable costs are calculated as a percentage of total EPC costs. As an initial default value, variable O&M costs were assumed to equal 3.0 % of total EPC costs. Variable O&M costs can also be directly input, or calculated using key cost components.

Fixed O&M

Fixed costs include labor and other costs that are independent of the plant output level. Fixed cost items must be paid whether or not the plant produces any output. Fixed costs are calculated as a percentage of total EPC costs. As an initial default value, fixed O&M was assumed to equal 3.5 % of total EPC costs. Fixed O&M costs can also be directly input, or calculated using key cost components.

Economic and financial

Tax Options

Standard tax rates can be used to calculate annual income taxes for a project under a regular tax schedule (i.e., during periods where there are no tax credits or benefits).

Grace Period

Lenders often grant projects a grace period on the initial repayment of principal. Therefore, the model contains an option for using a grace period (1 year as default value) on the repayment of principal for each loan.

Depreciation

Construction costs can be depreciated using a straight-line method (variable number of years) or a 150% declining balance method over a period of 15 or 20 years. Financing charges are can be separately depreciated using either method.

Escalation of Operating Costs and Revenues

The model escalates power costs, fuel costs, and operating expenses by an annual escalation rate. An escalation factor of 3% is employed in product sale price and O&M cost to reflect the inflation within plant lifetime.

Discount Rate

All discount rates are assumed to be in nominal form.

Construction Schedule

Construction schedule in the model is defined by the plant start-up year and length of construction period. Using these parameters, the model calculates a start date. Plant production is assumed to start on January 1 of the start-up year.

Plant Ramp-up Option

To account for the plant start-up period, the model includes an option to allow the plant to gradually reach full capacity. Specifically, the model calculates an average annual capacity percentage for up to the first two full years of operation.

Appendix B. Scenario modeling details

Stream/block nomenclature

All streams and blocks within the model follow a specific alphanumeric notation with the purpose of clarity and consistency across scenarios and across platforms. Each area within the model, such as Area 200 Gasification, has a two letter abbreviation GS. These abbreviations are used for naming streams as well as blocks. In addition to serving purposes mentioned above, the notation is descriptive (e.g., the notation REAC describes a block as a reactor). Another example is SGAS, which describes a stream that contains syngas. Aspen Plus software limits block and stream names to be eight characters. Figure B.1 below shows the pattern of notation for a syngas stream in the gasification area.

Area		Number		Description			
G	**S**	**0**	**1**	**S**	**G**	**A**	**S**

Figure B.1 Stream nomenclature used in model

Similarly, the notation for the first reactor block in the gasification area is shown in Figure B.2.

Area		Description				Number	
G	**S**	**R**	**E**	**A**	**C**	**0**	**1**

Figure B.2 Block nomenclature used in model

The Q or W sets the stream apart as a heat or work stream. The block description is limited to three characters and the number is limited to one character listed in Figure B.3

below. The following definitions and assumptions in Table B.1 are included to provide the user with a more detailed explanation of the model's inputs and parameters.

Q or W	Area		Description				Number
Q	**G**	**S**	**R**	**E**	**A**	**C**	**1**

Figure B.3 Heat and work stream nomenclature used in model

Aspen plus calculator block descriptions

CINERT

This block let user define the ratio of unconverted carbon as inert material in the gasification simulation. Definition of the stream and calculation expression is given as follows:

- CIN= mole flow of the carbon (solid) in the gasifier entrance
- RINERT=mole flow of the carbon (solid) in the gasifier exit
- R=Raito of inert carbon
- R=RINERT/ CIN

DECOMP

This block converts each element composition in the feedstock ultimate analysis into wet basis and then calculates the decomposed outputs based on the mass and element balance. The calculation expression is given as follows.

- WATER= Moisture content (weight percentage) in feedstock the proximate analysis
- Fact is the factor to convert the ultimate analysis to a wet basis.

  FACT= (100-WATER)/100

Table B.1 Detailed description of stream and block nomenclature

Area	Description	Name	Block	Name	Stream	Name
Plant	All Areas	PL	Reactor	REAC	Coal	COAL
A100	Pretreatment	PR	Mixer	MIX	Steam	STM
A200	Gasification	GS	Heat Mixer	QMIX	Flue Gas	FLUE
A300	Gas Cleanup	GC	Work Mixer	WMIX	Syngas	SGAS
A400	Steam Reforming	SM	Splitter	SPL	Ash	ASH
A500	Syngas Conditioning	SC	Separator	SEP	Carbon Dioxide	CO2
A600	Fuel Synthesis and Upgrading	FS	Cyclone	CYC	Air	AIR
A700	Carbon Dioxide Removal	CR	Column	COL	Hydrogen	HYD
A800	Power Generation	PG	Grinder	GRIN	FT Products	FTP
			Heater	HEAT	FT Liquids	FTL
			Heat Exchanger	HX	FT Gas	FTG
			Pump	PMP	Light Hydrocarbon	LHYC
			Compressor	COMP	Sulfur	SUL
			Turbine	TURB	Sulfur Dioxide	SULD
			Crusher	CRSH	Water	WAT
			Screen	SCRN	Naphtha	NAPH
					Diesel	DISL
					Char	CHAR
					Sand	SAND
					Zinc oxide	ZNO
					Zinc sulfide	ZNS
					Lean MDEA Soln	MDEAL
					Rich MDEA Soln	MDEAR
					Lean MEA Soln	MEAL
					Rich MEA Soln	MEAR
					Slurry	SLRY
					Solid/Gas mixture	SGMX
					Solid mixture	SMIX
					Other Contaminants	OC
					Ammonia	NH3

- Mass fraction of H_2O in the block output= WATER/100

- Mass fraction of ASH in the block output= ASH mass fraction in the ultimate analysis * FACT
- Mass fraction of C in the block output = Mass fraction of C in the ultimate analysis *FACT
- Mass fraction of H_2 in the block output = Mass fraction of H in the ultimate analysis *FACT
- Mass fraction of N_2 in the block output = Mass fraction of N in the ultimate analysis *FACT
- Mass fraction of Cl_2 in the block output = Mass fraction of Cl in the ultimate analysis *FACT
- Mass fraction of S in the block output = Mass fraction of S in the ultimate analysis *FACT
- Mass fraction of O_2 in the block output = Mass fraction of O in the ultimate analysis *FACT

FTR

This block calculates the hydrocarbon product (C1-C30) yield distribution based on the Anderson–Schulz–Flory (ASF) distribution. Parameters definition and calculation expression is given as follows:

- X_{H2}= Molar fraction of H_2 in the FT synthesis block input stream
- X_{CO}= Molar fraction of CO in the FT synthesis block input stream
- T_{FTR}= Reaction temperature of the FTR (in K)
- P_{FTR}= Reaction pressure of FTR (in bar)

- S_{C5+} = Mass fraction of C5+ in the hydrocarbon product

  $S_{C5+} = 1.7 – 0.0024*T_{FTR} – 0.088* X_{H2}/ X_{CO} + 0.18*(X_{H2} + X_{CO}) + 0.0078* P_{FTR}$

- α_{Cn} is the molar fraction of Cn in the hydrocarbon product

  $\alpha_{Cn} = \alpha^{n-1} (1-\alpha)$

- $\alpha = 0.75 – 0.373 *[–\log (S_{C5+})]^{0.5} + 0.25*S_{C5+}$

Aspen plus model design specifications

H2OIN

The required H_2O mass input in the slurry is set as a function of overall feedstock mass flow rate. A H_2O/feedstock mass ratio of R is selected based on the plant design configuration and the necessary mass flow rate of H_2O is calculated by the expression given below:

Mass flow rate of H_2O = R* Mass flow rate of feedstock.

H2IN

The required H_2 molar input is set as a function of overall carbon molar flow rate in the feedstock. A H_2/carbon molar ratio of R is selected based on the plant design configuration and the necessary molar flow rate of H_2 is calculated by the expression given below:

Molar flow rate of H_2 = R* Molar flow rate of carbon in the feedstock.

REGENAIR

The required air flow rate in the regenerator is supplied as 20% excess. The necessary mass flow rate of air is calculated by the expression given below:

Mass flow rate of air = 13.7* Mass flow of char in the regenerator input.

SAND

The required sand circulating between the SHR and regenerator (temperature difference 150 ^oC) is calculated based on the regenerator heat duty. The net heat duty of the regenerator is set as zero and the necessary sand, with physical property of silicon dioxide, is calculated according to the system heat balance.

SHRSTEAM (Optional)

The required steam mass input in the gasifier is similar as H2OIN and the necessary mass flow rate of steam is calculated by the expression given below:

Mass flow rate of steam = R* Mass flow rate of feedstock.

ZNOIN

The required ZnO input with 40% excess supply in the H_2S removal block is calculated by the expression given below:

Molar flow rate of ZnO = 1.4* Molar flow rate of H_2S in the block input

DSRPAIR

The required air input in the ZnO regeneration stage during DSRP process is supplied as 20% excess. The necessary mass flow rate of air is calculated by the expression given below:

Mass flow rate of air = 1.7* Mass flow rate of ZnS in the regenerator input stream.

DSRPH2IN

The required H_2 input in the SO_2 reduction reaction in DSRP process is calculated by the expression given below:

Molar flow rate of H_2 = 2.2* Molar flow rate of $SO_{2.}$

SMRAIRIN

The required air flow rate in the SMR furnace is supplied as 20% excess and the fuel contains H_2 , CO and light hydrocarbon (C1-C4) is from the FT upgrading section. A system of stoichiometric combustion reactions is set up to sum all the oxygen required to fully combust the fuel gas as can been seen in the Table B.2 below.

The necessary molar flow rate of air is calculated by the expression given below:

Molar flow of air (M_{air}) = 5.7 * (0.5* M_{air} + 0.5*M_{CO} + 2*M_{CH4} + 3*M_{C2H4} + 3.5*M_{C2H6} +4.5*M_{C3H6} + 5*M_{C3H8}+6*M_{C4H8}+6.5*M_{C4H10})

Table B.2 Combustion reactions to determine required oxygen

Component	Reaction
H_2	$H_2 + 0.5O_2 \rightarrow H_2O$
CO	$CO + 0.5O_2 \rightarrow CO_2$
CH_4	$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$
C_2H_4	$C_2H_4 + 3O_2 \rightarrow 2CO_2 + 2H_2O$
C_2H_6	$C_2H_6 +3.5O_2 \rightarrow 2CO_2 + 3H_2O$
C_3H_6	$C_3H_6 + 4.5O_2 \rightarrow 3CO_2 + 3H_2O$
C_3H_8	$C_3H_8 + 5O_2 \rightarrow 3CO_2 + 4H_2O$
C_4H_8	$C_4H_8 + 6O_2 \rightarrow 4CO_2 +4 H_2O$
C_4H_{10}	$C_4H_{10} + 6.5O_2 \rightarrow 4CO_2 + 5H_2O$

H2RECYCL

The required H_2 flow rate in the syngas is adjusted using this design specification and a H_2/CO molar ratio of R is selected based on the catalyst utilized in the FTR. The necessary molar flow rate of H_2 in the FT synthesis is calculated by the expression given below:

Molar flow rate of H_2 = R * Molar flow rate of CO in the FT synthesis input stream

MDEA

The required molar flow of MDEA for CO_2 removal arriving from FT reactor effluent at high pressure (400 psia) is calculated here. The MDEA is able to capture 0.5 moles CO_2 per mole MDEA. Additionally, the MDEA is diluted with concentration of 50 wt%. The necessary molar flow rate of MDEA is calculated by the expression given below:

Molar flow rate of MDEA = Molar flow rate of CO_2 in the gas stream/0.5

The necessary molar flow rate of H_2O in the MDEA solution is calculated as below:

Molar flow rate of H_2O in the solution = 6.6 *Molar flow rate of MDEA

MEA

The required molar flow of MEA for CO_2 removal in the flue gas (SHR regenerator, SMR furnace and combustor) at ambient pressure is calculated here. The MEA is able to capture 0.35 moles CO_2 per mole MEA. Additionally, the MEA is diluted with concentration of 20 wt%. The necessary molar flow rate of MEA is calculated by the expression given below:

Molar flow rate of MEA = Molar flow rate of CO_2 in the gas stream/0.35.

The necessary molar flow rate of H_2O in the MEA solution is calculated as below:

Molar flow rate of H_2O in the solution = 13.6 *Molar flow rate of MEA.

COMBAIR

Excess H_2 is sent to the combustor for power generation in the downstream steam cycle. Part of O_2 in supplied the combustor comes from the SHR regenerator flue gas and SMR furnace flue gas. The rest necessary air in the combustor with overall 20% excess supply is calculated below:

Molar flow rate of air (M_{air}) = ($0.6 * M_{H2\text{-}Excess} – M_{o2\text{-}regen\ flue} – M_{o2\text{-}Furnace\ flue}$) /0.21

If light hydrocarbon from FT reactor effluent is used as fuel in the power generation, the necessary air with overall 20% excess supply in the combustor is calculated as following:

Molar flow rate of air (M_{air}) = ($0.6 * M_{H2\text{-}Excess} + 0.6 * M_{H2\text{-}Fuel} + 0.6 * M_{CO\text{-}Fuel} + 2.4 * M_{CH4\text{-}Fuel} + 3.6 * M_{C2H4\text{-}Fuel} + 4.2 * M_{C2H6\text{-}Fuel} + 5.4 * M_{C3H6\text{-}Fuel} + 6.0 * M_{C3H8\text{-}Fuel} + 7.2 * M_{C4H8\text{-}Fuel} + 7.8 * M_{C4H10\text{-}Fuel} – M_{o2\text{-}regen\ flue} – M_{o2\text{-}Furnace\ flue}$) /0.21

H2SPLIT

This design spec calculates the required H_2 that to be reserved for SHR input, DSRP and hydrocracking section. This unit operation happens after the syngas ratio adjustment step and the rest of H_2 after the split is then sent to the combustor for power generation. A typical yield from hydrocracking is shown in Table B.3.

Table B.3 Hydroprocessing product blend

Component	Mass Fraction
Fuel gas (methane)	0.034
LPG (propane)	0.088
Gasoline (n-octane)	0.261
Diesel (n-hexadecane)	0.617

The carbon mass flow is the same as that of the final blend stream flow. Using the blend fractions in Table B.3, the amount of H_2 in the hydrocracking process is calculated in the final blend and the difference in H_2 is determined. The difference is multiplied by 1.1 to obtain the delivered H_2 mass flow rate to hydrocracking area.

Process assumptions

For all cases, most of the process was modeled with the aid of Aspen Plus software. The process was divided by logical process areas, which are named below.

A 100 – Feedstock handling and preparation

- Coal is delivered to plant from coal mine nearby with moisture content of 15%.
- Biomass is transported to the plant with initial moisture content of 25%.
- Feedstock is grinded to 70 micron meter or less.
- The biomass is made into slurry in the hydrothermal pretreatment system (not modeled).
- No carbon loss is considered in the pretreatment process.
- Heat for feedstock pre-heating in this area is provided from the plant heat integration (e.g., hot flue gas).
- Energy consumption for feedstock grinding is calculated separately using literature correlations.

A 200 – Gasification

- Overall unconverted char ratio at different gasification temperature is obtained from lab experiments and used as major input in the gasification simulation.

- The fluidized bed gasifier is simply modeled using thermodynamic equilibrium model. The regenerator is simulated with combustion model using leftover char or diverted syngas as fuel with 20% excess air supply.
- Sand, as the heat media, between the gasifier and regenerator is modeled with physical property of silica dioxide. No mass loss is considered during the circulating.
- Energy consumption in the air compression is calculated separately using empirical expression from literature.
- The pressurized hot flue gas goes through a gas turbine for power generation before further usage. The mechanical efficiency of pressure changers such as compressors and expanders is assumed to be 1.0 while the isentropic efficiency is 0.8~0.9.

A 300 – Gas Cleanup

- Particulate handling and ash removal process is simply modeled using separation blocks
- Chloride, trace metal element and ammonia removal process is not modeled but using separation blocks to achieve this purpose.
- Regenerable ZnO circulating between the reactor and regenerator is used in the H_2S removal process. A stream of air is utilized to regenerate the ZnO and convert the sulfur specie in the form of SO_2.

- Small part of H_2 diverted from the main H_2 stream after syngas adjustment is used in the DSRP process to convert the SO_2 into element sulfur with more than 95% efficiency.
- The leftover SO_2 in the DSRP exit gas stream is trapped by passing through sulfuric acid solution with high concentration (not modeled). The almost SO_2 free gas is then vented to the atmosphere.
- More than 99% of the sulfur is removed in this area and the contaminant free clean gas stream enters the next section at temperature higher than 280 ^{o}C.

A 400 – Steam Methane Reforming

- ZnO and activated carbon guard bed polishing at the entrance assumed (not modeled in detail).
- Steam methane reforming occurs at equilibrium and is modeled as such.
- An integrated furnace setup, modeled with combustion block, is built to provide the necessary heat for the strong endothermic reaction. Light hydrocarbons diverted from the FT synthesis section are used as fuel combusted with 20% excess air supply and the operation temperature of the furnace is assumed 8 ^{o}C higher than the reaction temperature.
- The hot flue gas from the furnace then goes to the final combustor for power generation or heat exchangers for heat recovery depending on the design configurations.

A 500 – Syngas Conditioning

- PSA is employed to separate H_2 at an efficiency of 90% and 98% purity. The absorbers are filled with activated carbon (2/3) and (1/3) molecular sieve.
- H_2 separation membrane is used to adjust the H_2 fraction in the syngas based on the downstream FT synthesis configuration (not modeled).
- The H_2 after the syngas adjustment is then spitted in the streams used in the SHR feed, DSRP and hydroprocessing section. The result of H_2 is then sent to the combustor for power generation.

A 600 – FT Synthesis and Upgrading

- The yield of hydrocarbon product in the FT synthesis follows the Anderson–Schulz–Flory distribution.
- Water-gas-shift occurs at equilibrium when Iron-based catalyst is used in FT synthesis and is modeled as such.
- CO conversion efficiency of up to 90% is achieved in the looped FTR with Iron-based catalyst.
- CO conversion efficiency of 78% is achieved in the one-through FTR with Cobalt-based catalyst.
- All light hydrocarbons (C1-C4) along with unreacted CO and H_2 from FTR goes to the SMR furnace as fuel when Iron-based catalyst is used while a portion of the fuel gas is split in the combustor for power generation when Cobalt-based catalyst is used.

A 700 – CO_2 removal

- Captured CO_2 stream is compressed to 2217 psia in multi-stage, intercooled compressor, dehydrated and sent to pipeline transportation (not modeled in detail).
- The removal efficiency is up to 95% in the MDEA CO_2 capture section.
- The removal efficiency ranges from 50%-78% in the MEA CO_2 capture section based on the requirement of overall plant carbon capture efficiency.

Appendix C. Case performance summary

FT liquids Production Process Engineering Analysis
Design Report: CERT-1

360 Dry Metric Tonnes Coal per Day w/ 29% Carbon Capture

Steam Hydrogasification, Warm Gas Cleanup, Steam Methane Reformer, Iron FT Synthesis, Steam-Power Cycle

All Values in 2010 $

Minimum FT Liquids Selling Price ($/gallon)	**$5.80**	$43.52 $46.08	($/GJ, HHV) ($/GJ, LHV)
FT Liquids Production at operating capacity (MM gallon/year)	9.4	679	(BBL/day)
FT Liquids Yield (gallon/Dry Metric Ton Coal)	78.5	0.22	(ton/ton coal)
Delivered Feedstock Cost $/Dry Metric Ton	$49.5	2.07	($/GJ, HHV)
Internal Rate of Return (After-Tax)	12%		
Equity Percent of Total Investment	30%		

Capital Costs	
Feed Handling & Preparation	$7,400,000
Gasification	$48,400,000
Warm Gas Cleanup	$13,200,000
Steam Methane Reforming	$30,300,000
F-T Synthesis and Upgrading	$6,700,000
Power Island	$13,900,000
Carbon Dioxide Capture	$12,400,000
Cooling Water and Other Utilities	$21,100,000
Total Installed Equipment Cost	$153,400,000
Indirect Costs	$86,300,000
Percentage of TPC (%)	36.0%
Total Plant Cost (TPC)	$239,700,000
Total Required Capital (TRC)	$308,400,000
Loan Rate	7.5%
Term (years)	15
Grace Period on Principal Repayment (year)	1
Income Tax Rate	38.0%
Salvage Value (% of TPC)	0.0%
Plant Ramp-up (% of full capacity)	
Ramp-Up-Year 1	51.0%
Ramp-Up-Year 2	78.0%
Ramp-Up-Year 3	100.0%
Gasifier Efficiency - HHV	84.10%
Gasifier Efficiency - LHV	82.27%
Feedstock-to-Fuel Efficiency - HHV	42.1%
Feedstock-to-Fuel Efficiency - LHV	41.3%

Operating Costs ($/gallon FT Liquids)	
Feedstock	0.63
Loan expense	1.37
Variable Costs	1.02
Fixed Costs	1.41
Electricity	-0.48
Capital Depreciation	0.89
Average Income Tax	0.25
Average Return on Investment	0.71

Operating Costs ($/yr)	
Feedstock	$5,900,000
Loan expense	$12,800,000
Variable Costs	$9,600,000
Fixed Costs	$13,200,000
Electricity	-$4,500,000
Capital Depreciation	$8,300,000
Average Income Tax	$2,300,000
Average Return on Investment	$6,700,000

Electricity Produced Onsite (MW)		15.62
Gas Turbine (MW)	2.39	
Steam Turbine (MW)	13.23	
Total Plant Electricity Usage (MW)		5.00
Electricity Export to Grid (MW)		10.62
Plant Electricity Use (KWh/gallon FT Liquid)		4.21
Plant Electricity Generation (KWh/gallon FT Liquid)		13.15
Coal-to-Power Efficiency - HHV		9.0%
Coal-to-Power Efficiency - LHV		9.3%
Overall Plant Efficiency - HHV		51.1%
Overall Plant Efficiency - LHV		50.6%

Figure C.1 Economic analysis summary for Case CERT-1

FT liquids Production Process Engineering Analysis
Design Report: CERT-2

3600 Dry Metric Tonnes Coal per Day w/ 65% Carbon Capture

Steam Hydrogasification, Warm Gas Cleanup, Steam Methane Reformer, Iron FT Synthesis, CO_2 capture, Steam-Power Cycle

All Values in 2010 $

Minimum FT Liquids Selling Price ($/gallon)	**$1.90**	$14.26	($/GJ, HHV)
		$15.10	($/GJ, LHV)
FT Liquids Production at operating capacity (MM gallon/year)	97.8	7,086	(BBL/day)
FT Liquids Yield (gallon/Dry Metric Ton Coal)	82.7	0.23	(ton/ton coal)
Delivered Feedstock Cost $/Dry Metric Ton	$49.5	2.07	($/GJ, HHV)
Internal Rate of Return (After-Tax)	12%		
Equity Percent of Total Investment	30%		

Capital Costs	
Feed Handling & Preparation	$49,300,000
Gasification	$172,900,000
Warm Gas Cleanup	$55,700,000
Steam Methane Reforming	$102,000,000
F-T Synthesis and Upgrading	$55,900,000
Power island	$75,900,000
Carbon Dioxide Capture	$62,400,000
Cooling Water and Other Utilities	$52,700,000
Total Installed Equipment Cost	$626,800,000
Indirect Costs	$352,600,000
Percentage of TPC (%)	36.0%
Total Plant Cost (TPC)	$979,400,000
Total Required Capital (TRC)	$1,366,100,000
Loan Rate	7.5%
Term (years)	15
Grace Period on Principal Repayment (year)	1
Income Tax Rate	38.0%
Salvage Value (% of TPC)	0.0%
Plant Ramp-up (% of full capacity)	
Ramp-Up-Year 1	51.0%
Ramp-Up-Year 2	78.0%
Ramp-Up-Year 3	100.0%
Gasifier Efficiency - HHV	84.60%
Gasifier Efficiency - LHV	82.74%
Feedstock-to-Fuel Efficiency - HHV	43.8%
Feedstock-to-Fuel Efficiency - LHV	42.9%

Operating Costs ($/gallon FT Liquids)	
Feedstock	0.60
Loan expense	0.55
Variable Costs	0.24
Fixed Costs	0.25
Electricity	-0.46
Capital Depreciation	0.33
Average Income Tax	0.10
Average Return on Investment	0.28

Operating Costs ($/yr)	
Feedstock	$70,800,000
Loan expense	$64,900,000
Variable Costs	$28,300,000
Fixed Costs	$29,500,000
Electricity	-$54,300,000
Capital Depreciation	$38,900,000
Average Income Tax	$11,800,000
Average Return on Investment	$33,000,000

Electricity Produced Onsite (MW)		146.36
Gas Turbine (MW)	24.11	
Steam Turbine (MW)	122.25	
Total Plant Electricity Usage (MW)		39.9
Electricity Export to Grid (MW)		106.46
Plant Electricity Use (KWh/gallon FT Liquid)		3.22
Plant Electricity Generation (KWh/gallon FT Liquid)		11.80
Coal-to-Power Efficiency - HHV		9.1%
Coal-to-Power Efficiency - LHV		9.4%
Overall Plant Efficiency - HHV		52.9%
Overall Plant Efficiency - LHV		52.4%

Figure C.2 Economic analysis summary for Case CERT-2

FT liquids Production Process Engineering Analysis

Design Report: CERT-2B

3600 Dry Metric Tonnes Coal per Day w/ 90% Carbon Capture

Steam Hydrogasification, Warm Gas Cleanup, Steam Methane Reformer, Iron FT Synthesis, CO_2 Capture Steam-Power Cycle

All Values in 2010 $

Minimum FT Liquids Selling Price ($/gallon)	**$2.16**	$16.21	($/GJ, HHV)
		$17.16	($/GJ, LHV)
FT Liquids Production at operating capacity (MM gallon/year)	97.8	7,086	(BBL/day)
FT Liquids Yield (gallon/Dry Metric Ton Coal)	82.7	0.23	(ton/ton coal)
Delivered Feedstock Cost $/Dry Metric Ton	$49.5	2.07	($/GJ, HHV)
Internal Rate of Return (After-Tax)	12%		
Equity Percent of Total Investment	30%		

Capital Costs		Operating Costs ($/gallon FT Liquids)		
Feed Handling & Preperation	$49,300,000	Feedstock		0.60
Gasification	$172,900,000	Loan expense		0.59
Warm Gas Cleanup	$55,700,000	Variable Costs		0.26
Steam Methane Reforming	$102,000,000	Fixed Costs		0.25
F-T Synthesis and Upgrading	$55,900,000	Electricity		-0.32
Power island	$69,100,000	Capital Depreciation		0.35
Carbon Dioxide Capture	$100,500,000	Average Income Tax		0.11
Cooling Water and Other Utilities	$55,900,000	Average Return on Investment		0.31
Total Installed Equipment Cost	$661,300,000			
		Operating Costs ($/yr)		
Indirect Costs	$372,000,000	Feedstock		$58,700,000
Percentage of TPC (%)	36.0%	Loan expense		$57,700,000
		Variable Costs		$25,400,000
Total Plant Cost (TPC)	$1,033,300,000	Fixed Costs		$24,400,000
		Electricity		-$31,300,000
Total Required Capital (TRC)	$1,461,600,000	Capital Depreciation		$34,200,000
		Average Income Tax		$10,800,000
Loan Rate	7.5%	Average Return on Investment		$30,300,000
Term (years)	15			
Grace Period on Principal Repayment (year)	1	Electricity Produced Onsite (MW)		126.09
Income Tax Rate	38.0%	Gas Turbine (MW)	24.11	
Salvage Value (% of TPC)	0.0%	Steam Turbine (MW)	101.98	
		Total Plant Electricity Usage (MW)		52.36
Plant Ramp-up (% of full capacity)		Electricity Export to Grid (MW)		73.73
Ramp-Up-Year 1	51.0%			
Ramp-Up-Year 2	78.0%	Plant Electricity Use (KWh/gallon FT Liquid)		4.22
Ramp-Up-Year 3	100.0%	Plant Electricity Generation (KWh/gallon FT Liquid)		10.17
Gasifier Efficiency - HHV	84.60%	Coal-to-Power Efficiency - HHV		6.3%
Gasifier Efficiency - LHV	82.74%	Coal-to-Power Efficiency - LHV		6.5%
Feedstock-to-Fuel Efficiency - HHV	43.8%	Overall Plant Efficiency - HHV		50.1%
Feedstock-to-Fuel Efficiency - LHV	42.9%	Overall Plant Efficiency - LHV		49.5%

Figure C.3 Economic analysis summary for Case CERT-2B

FT liquids Production Process Engineering Analysis

Design Report: CERT-3

3600 Dry Metric Tonnes Coal per Day w/ 58% Carbon Capture

Steam Hydrogasification, Warm Gas Cleanup, Steam Methane Reformer, Cobalt FT Synthesis,CO_2 Capture Steam-Power Cycle

All Values in 2010 $

Minimum FT Liquids Selling Price ($/gallon)	**$1.75**	$13.13	($/GJ, HHV)
		$13.90	($/GJ, LHV)
FT Liquids Production at operating capacity (MM gallon/year)	117.9	8,548	(BBL/day)
FT Liquids Yield (gallon/Dry Metric Ton Coal)	99.5	0.29	(ton/ton coal)
Delivered Feedstock Cost $/Dry Metric Ton	$49.5	2.07	($/GJ, HHV)
Internal Rate of Return (After-Tax)	12%		
Equity Percent of Total Investment	30%		

Capital Costs	
Feed Handling & Preparation	$49,300,000
Gasification	$172,900,000
Warm Gas Cleanup	$55,700,000
Steam Methane Reforming	$102,000,000
F-T Synthesis and Upgrading	$66,200,000
Power island	$54,200,000
Carbon Dioxide Capture	$44,400,000
Cooling Water and Other Utilities	$50,300,000
Total Installed Equipment Cost	$595,000,000
Indirect Costs	$334,700,000
Percentage of TPC (%)	36.0%
Total Plant Cost (TPC)	$929,700,000
Total Required Capital (TRC)	$1,301,500,000
Loan Rate	7.5%
Term (years)	15
Grace Period on Principal Repayment (year)	1
Income Tax Rate	38.0%
Salvage Value (% of TPC)	0.0%
Plant Ramp-up (% of full capacity)	
Ramp-Up-Year 1	51.0%
Ramp-Up-Year 2	78.0%
Ramp-Up-Year 3	100.0%
Gasifier Efficiency - HHV	84.60%
Gasifier Efficiency - LHV	82.74%
Feedstock-to-Fuel Efficiency - HHV	52.8%
Feedstock-to-Fuel Efficiency - LHV	51.7%

Operating Costs ($/gallon FT Liquids)	
Feedstock	0.50
Loan expense	0.44
Variable Costs	0.22
Fixed Costs	0.22
Electricity	-0.19
Capital Depreciation	0.26
Average Income Tax	0.08
Average Return on Investment	0.23

Operating Costs ($/yr)		
Feedstock		$59,000,000
Loan expense		$51,900,000
Variable Costs		$25,900,000
Fixed Costs		$25,900,000
Electricity		-$22,400,000
Capital Depreciation		$30,700,000
Average Income Tax		$9,400,000
Average Return on Investment		$27,100,000
Electricity Produced Onsite (MW)		95.80
Gas Turbine (MW)	24.11	
Steam Turbine (MW)	71.70	
Total Plant Electricity Usage (MW)		40.25
Electricity Export to Grid (MW)		55.56
Plant Electricity Use (KWh/gallon FT Liquid)		2.69
Plant Electricity Generation (KWh/gallon FT Liquid)		6.40
Coal-to-Power Efficiency - HHV		4.7%
Coal-to-Power Efficiency - LHV		4.9%
Overall Plant Efficiency - HHV		57.5%
Overall Plant Efficiency - LHV		56.7%

Figure C.4 Economic analysis summary for Case CERT-3

FT liquids Production Process Engineering Analysis

Design Report: CERT-3B

3600 Dry Metric Tonnes Coal per Day w/ 90% Carbon Capture

Steam Hydrogasification, Warm Gas Cleanup, Steam Methane Reformer, Cobalt FT Synthesis,CO_2 Capture Steam-Power Cycle

All Values in 2010 $

Minimum FT Liquids Selling Price ($/gallon)	**$2.00**	$15.01	($/GJ, HHV)
		$15.89	($/GJ, LHV)
FT Liquids Production at operating capacity (MM gallon/year)	117.9	8,548	(BBL/day)
FT Liquids Yield (gallon/Dry Metric Ton Coal)	99.5	0.29	(ton/ton coal)
Delivered Feedstock Cost $/Dry Metric Ton	$47	2.07	($/GJ, HHV)
Internal Rate of Return (After-Tax)	12%		
Equity Percent of Total Investment	30%		

Capital Costs		Operating Costs ($/gallon FT Liquids)		
Feed Handling & Preperation	$49,300,000	Feedstock		0.50
Gasification	$172,900,000	Loan expense		0.48
Warm Gas Cleanup	$55,700,000	Variable Costs		0.23
Steam Methane Reforming	$102,000,000	Fixed Costs		0.22
F-T Synthesis and Upgrading	$66,200,000	Electricity		-0.04
Power island	$42,600,000	Capital Depreciation		0.28
Carbon Dioxide Capture	$95,200,000	Average Income Tax		0.09
Cooling Water and Other Utilities	$54,000,000	Average Return on Investment		0.25
Total Installed Equipment Cost	$637,900,000			
		Operating Costs ($/yr)		
Indirect Costs	$358,800,000	Feedstock		$59,000,000
Percentage of TPC (%)	36.0%	Loan expense		$56,600,000
		Variable Costs		$27,100,000
Total Plant Cost (TPC)	$996,700,000	Fixed Costs		$25,900,000
		Electricity		-$4,700,000
Total Required Capital (TRC)	$1,414,600,000	Capital Depreciation		$33,000,000
		Average Income Tax		$10,600,000
Loan Rate	7.5%	Average Return on Investment		$29,500,000
Term (years)	15			
Grace Period on Principal Repayment (year)	1	Electricity Produced Onsite (MW)		71.58
Income Tax Rate	38.0%	Gas Turbine (MW)	24.11	
Salvage Value (% of TPC)	0.0%	Steam Turbine (MW)	47.47	
		Total Plant Electricity Usage (MW)		58.13
Plant Ramp-up (% of full capacity)		Electricity Export to Grid (MW)		13.44
Ramp-Up-Year 1	51.0%			
Ramp-Up-Year 2	78.0%	Plant Electricity Use (KWh/gallon FT Liquid)		3.89
Ramp-Up-Year 3	100.0%	Plant Electricity Generation (KWh/gallon FT Liquid)		4.79
Gasifier Efficiency - HHV	84.60%	Coal-to-Power Efficiency - HHV		1.1%
Gasifier Efficiency - LHV	82.74%	Coal-to-Power Efficiency - LHV		1.2%
Feedstock-to-Fuel Efficiency - HHV	52.8%	Overall Plant Efficiency - HHV		54.0%
Feedstock-to-Fuel Efficiency - LHV	51.7%	Overall Plant Efficiency - LHV		52.9%

Figure C.5 Economic analysis summary for Case CERT-3B

Appendix D. Process flow diagrams

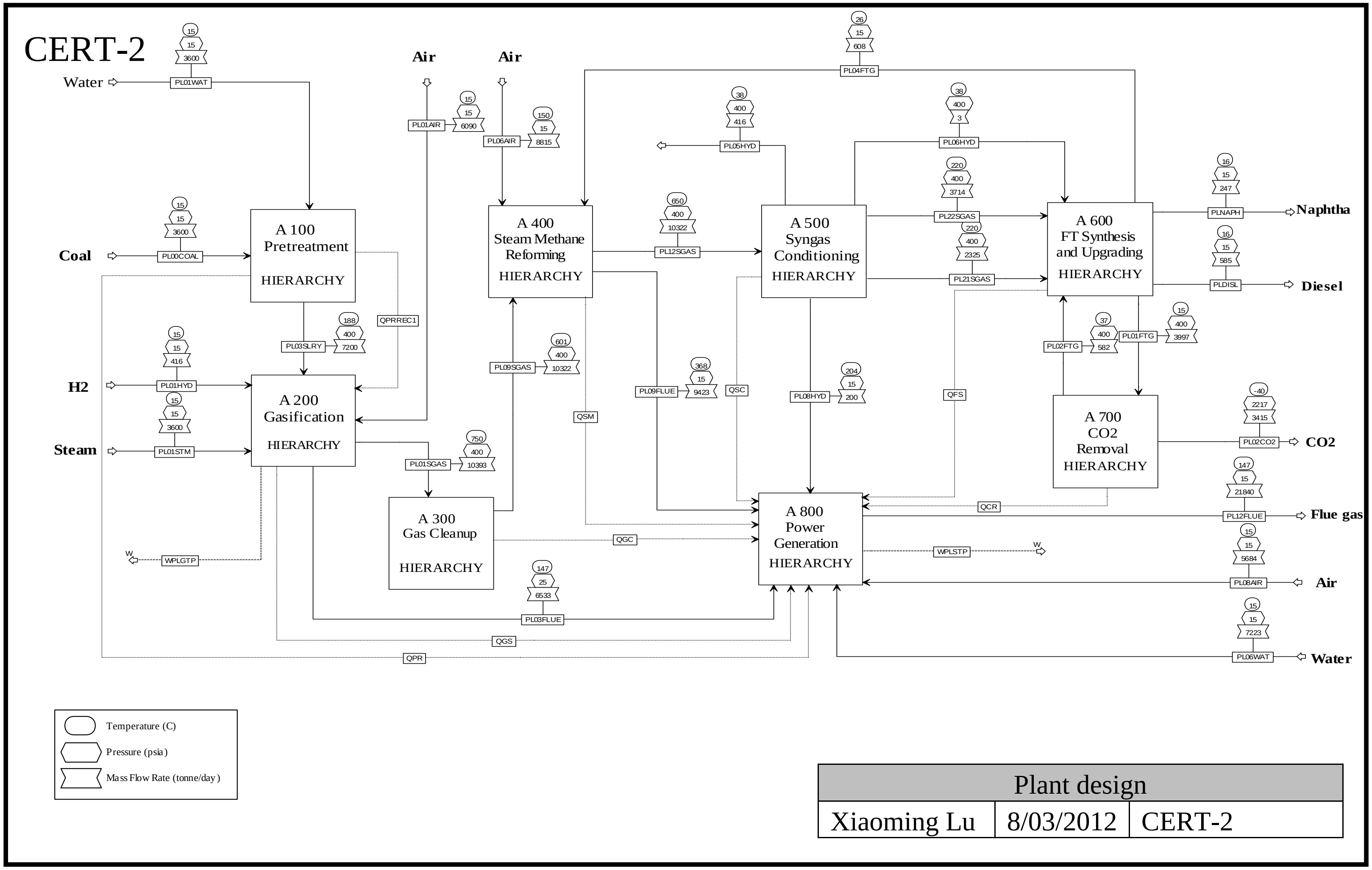

Figure D.1. Overall plant area process flow diagram for Case CERT-2

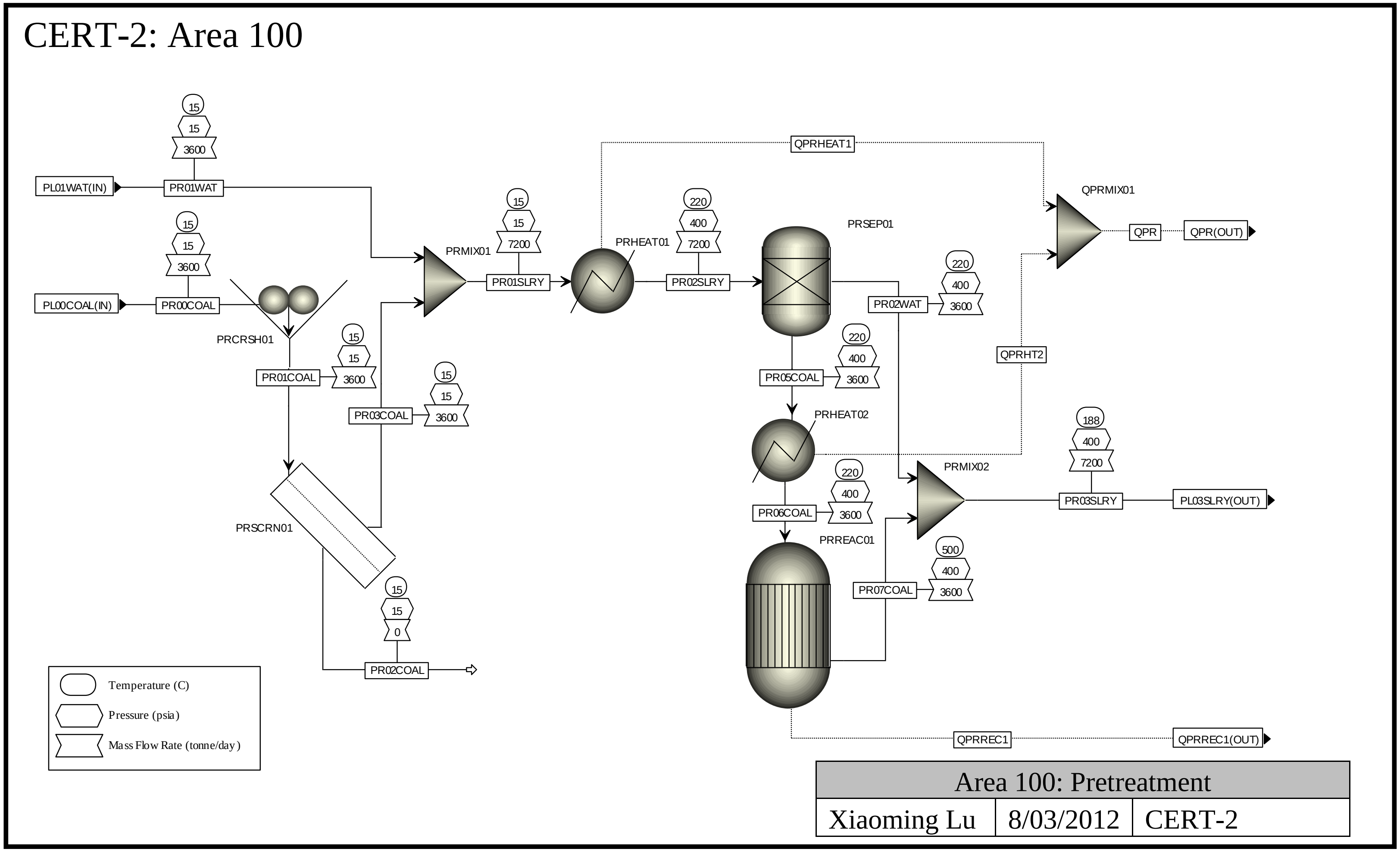

Figure D.2. Coal handling and preparation area process flow diagram for Case CERT-2

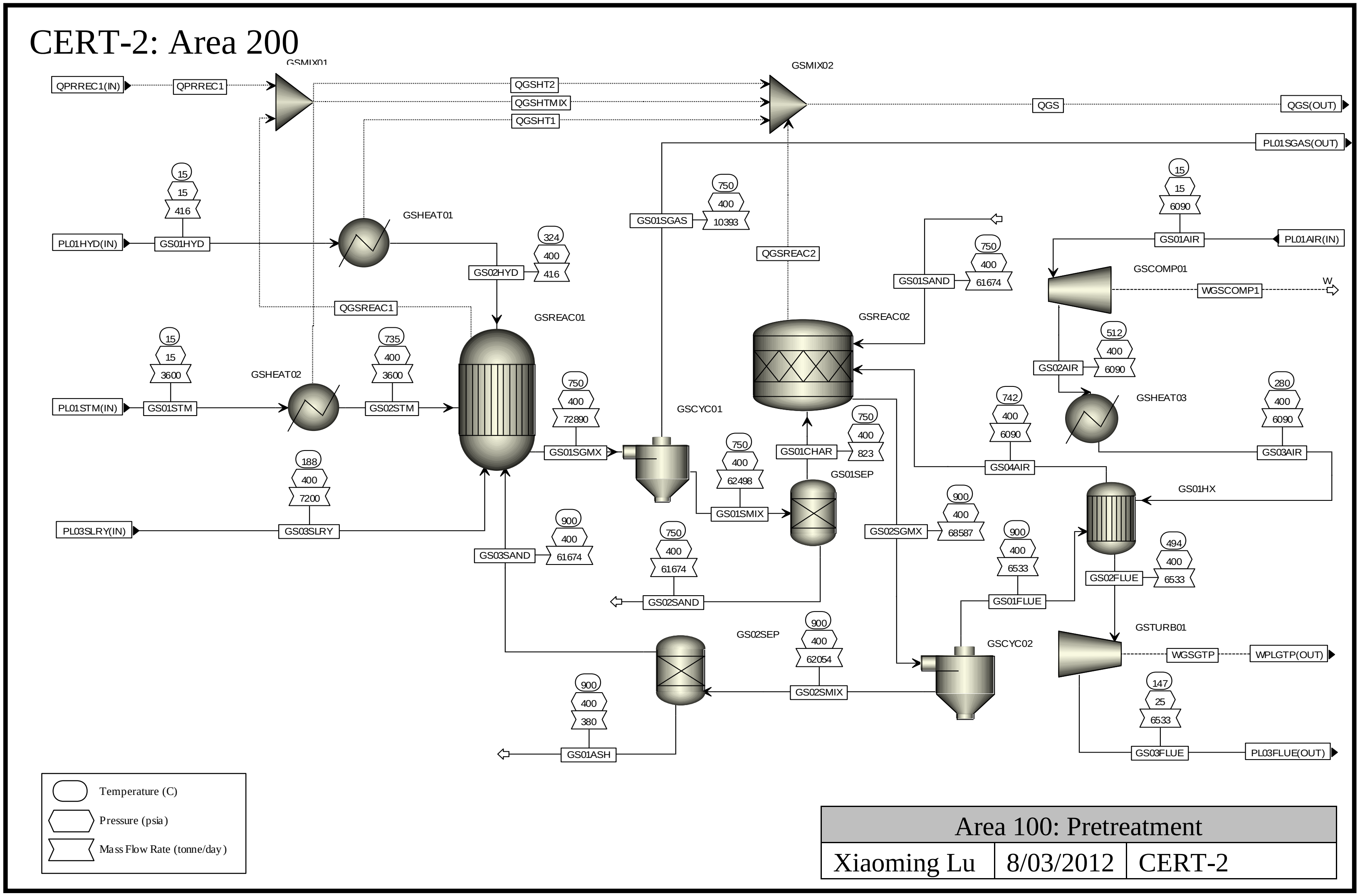

Figure D.3. Gasification area process flow diagram for Case CERT-2

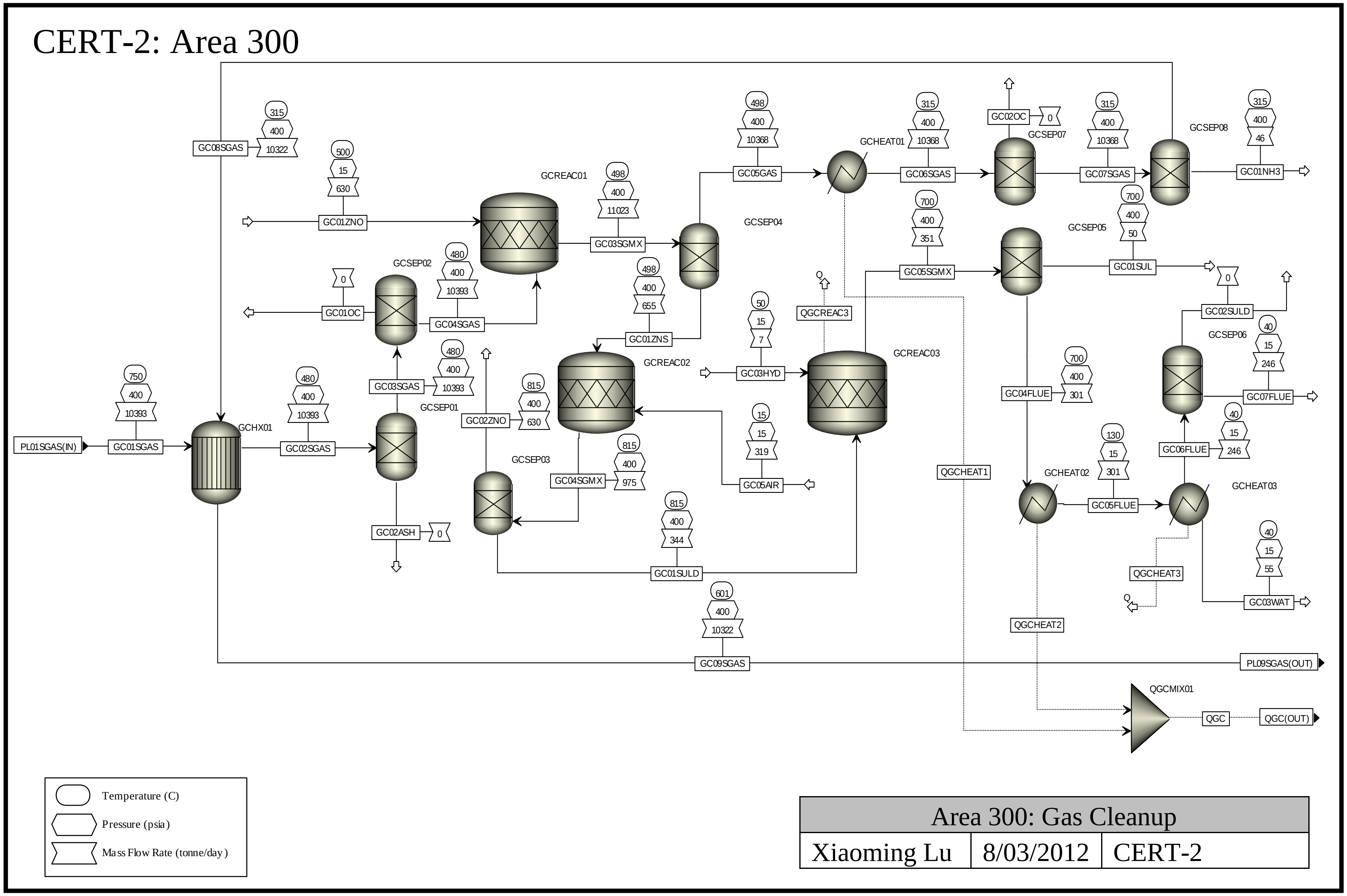

Figure D.4 Gas cleanup area process flow diagram for Case CERT-2

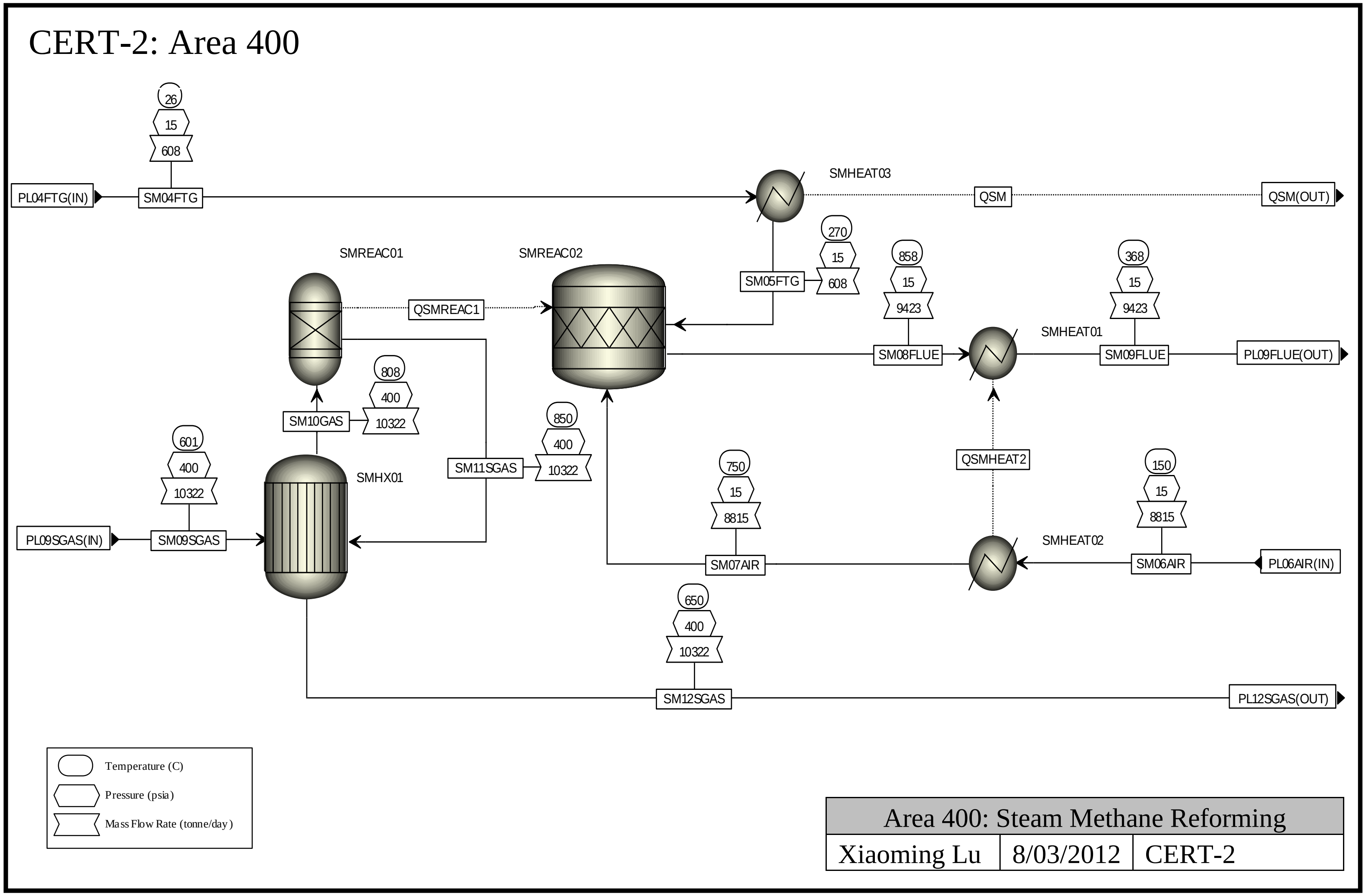

Figure D.5 Steam methane reforming area process flow diagram for Case CERT-2

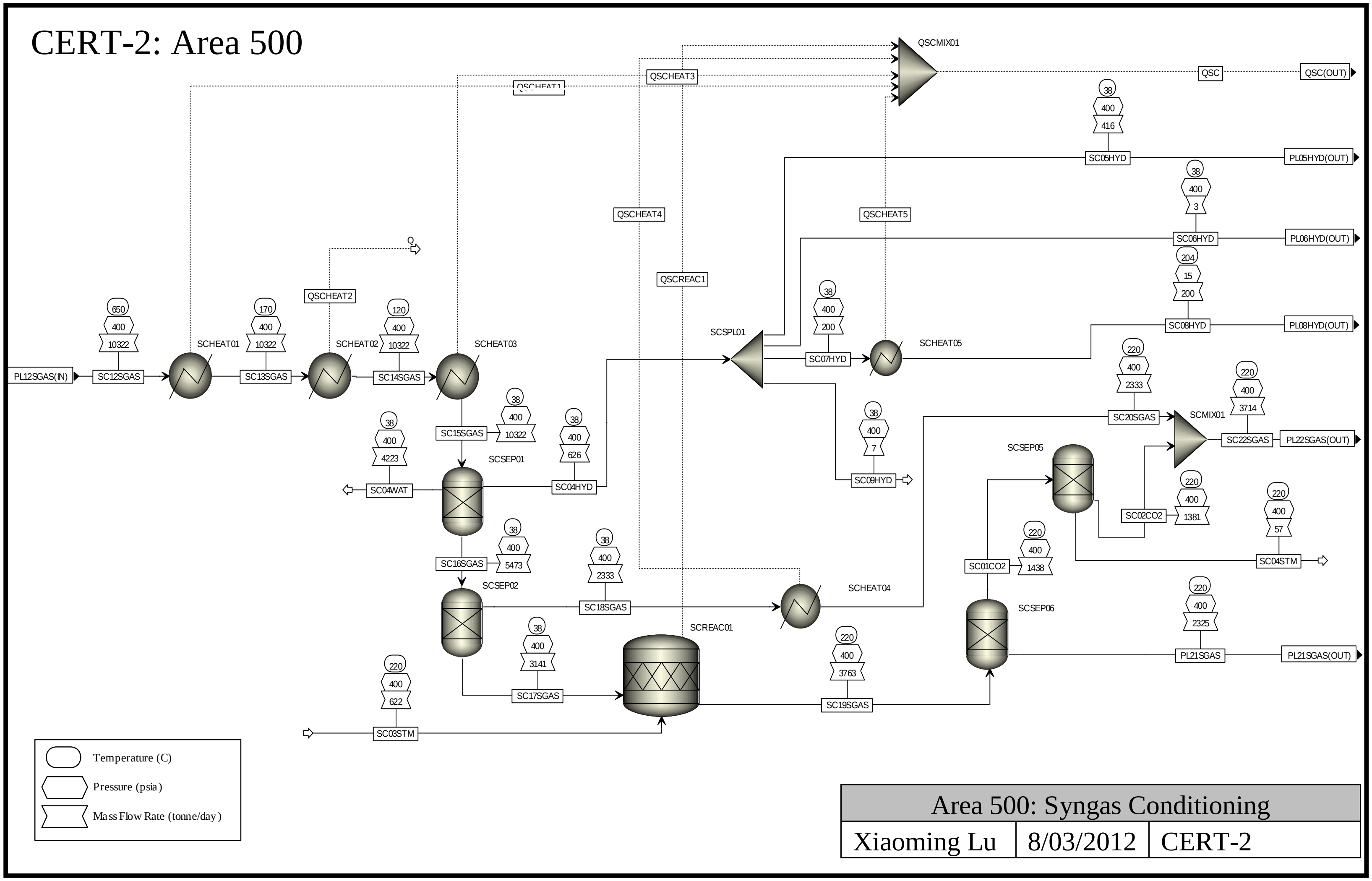

Figure D.6 Syngas conditioning area process flow diagram for Case CERT-2

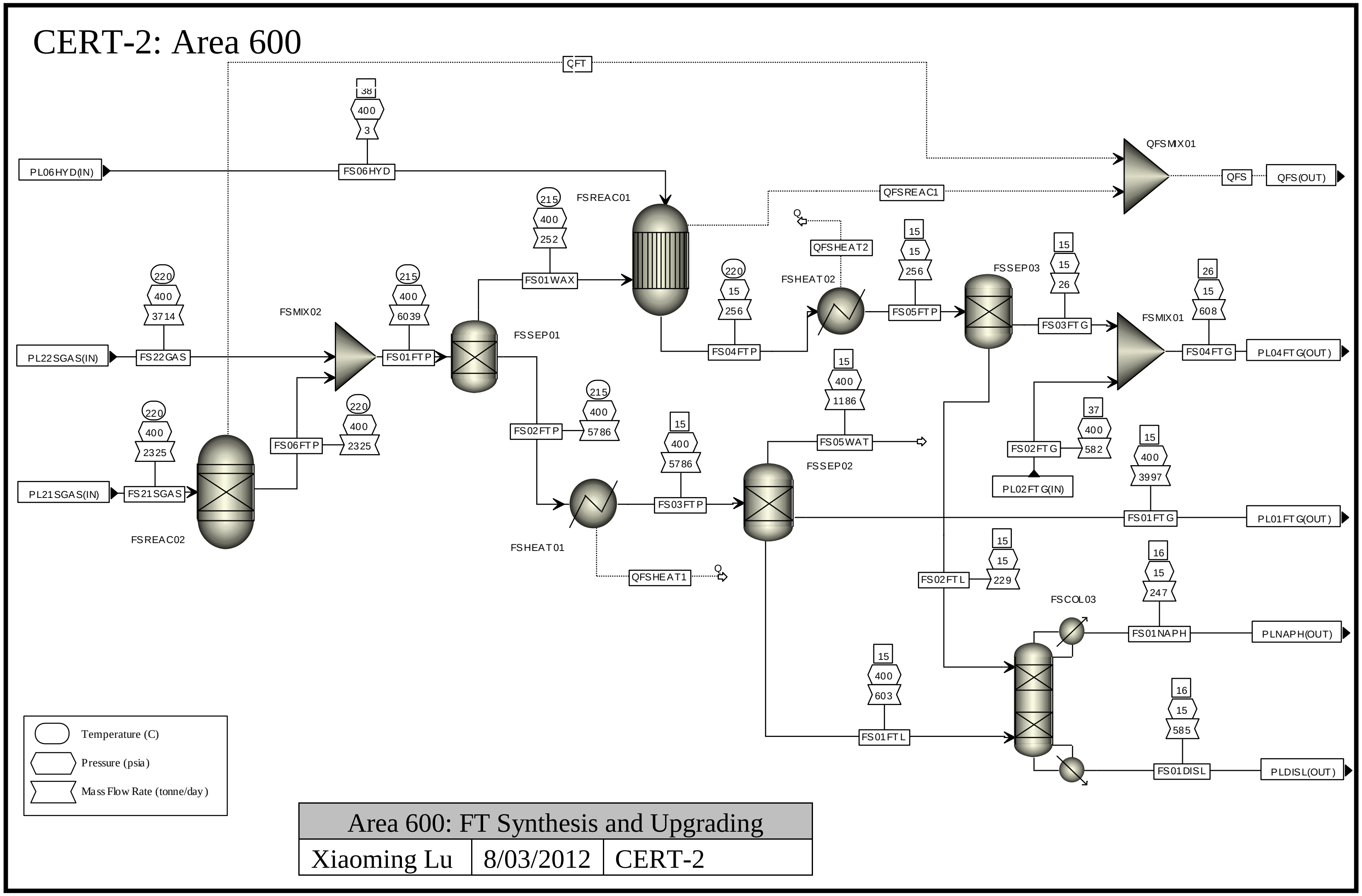

Figure D.7 Fuel synthesis and upgrading area process flow diagram for Case CERT-2

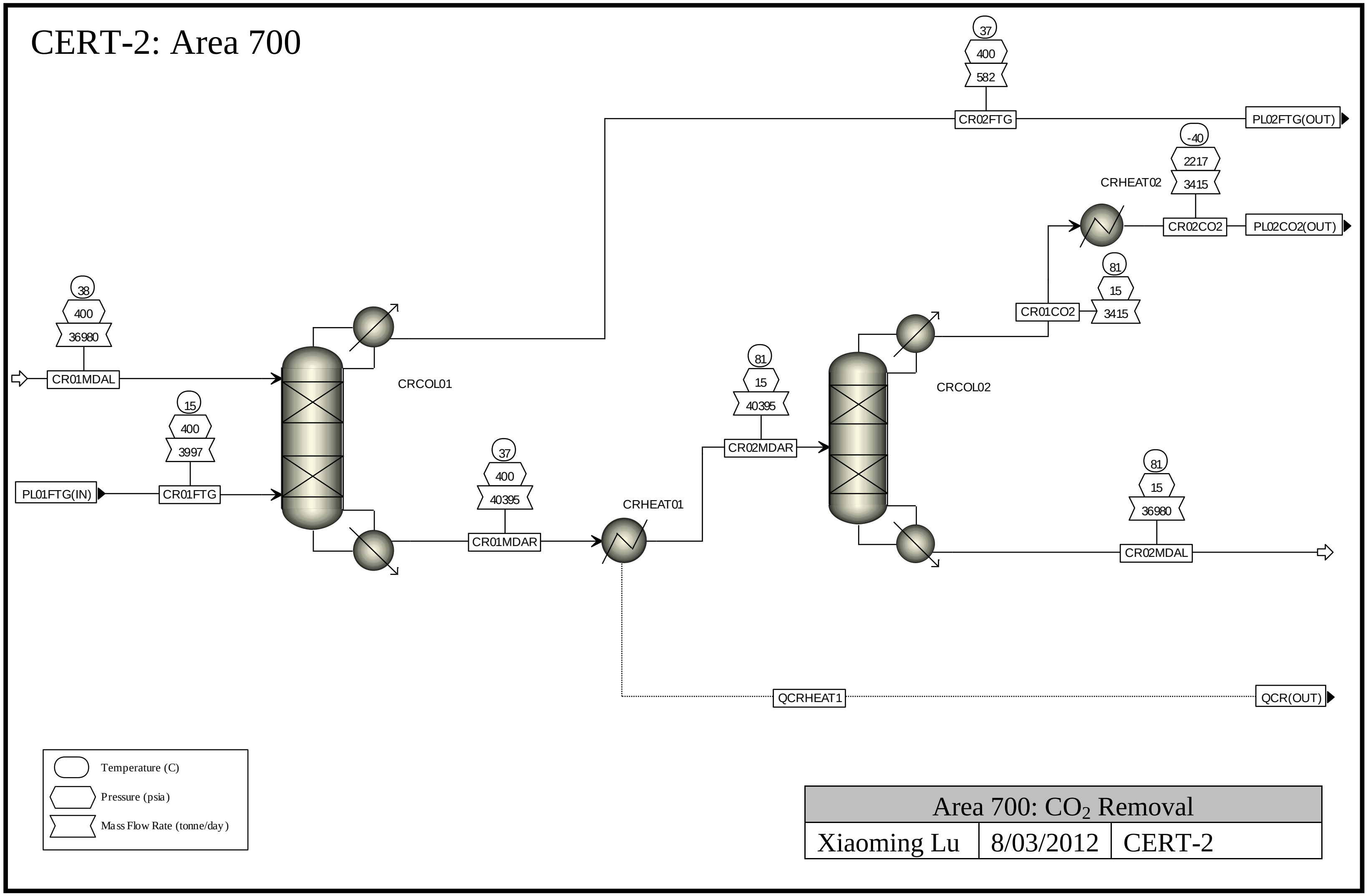

Figure D.8 CO_2 removal area process flow diagram for Case CERT-2

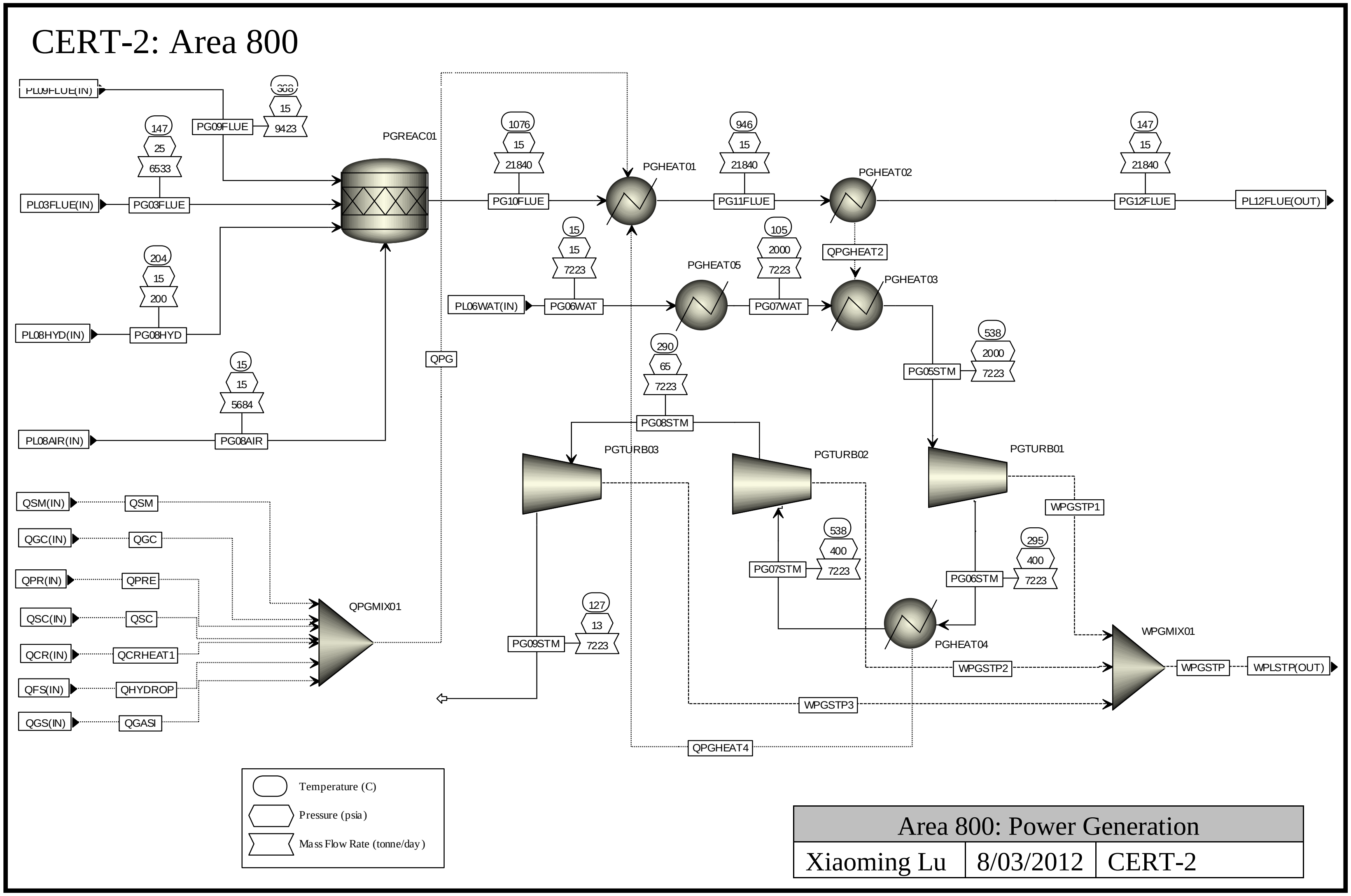

Figure D.9 Power Generation area process flow diagram for Case CERT-2

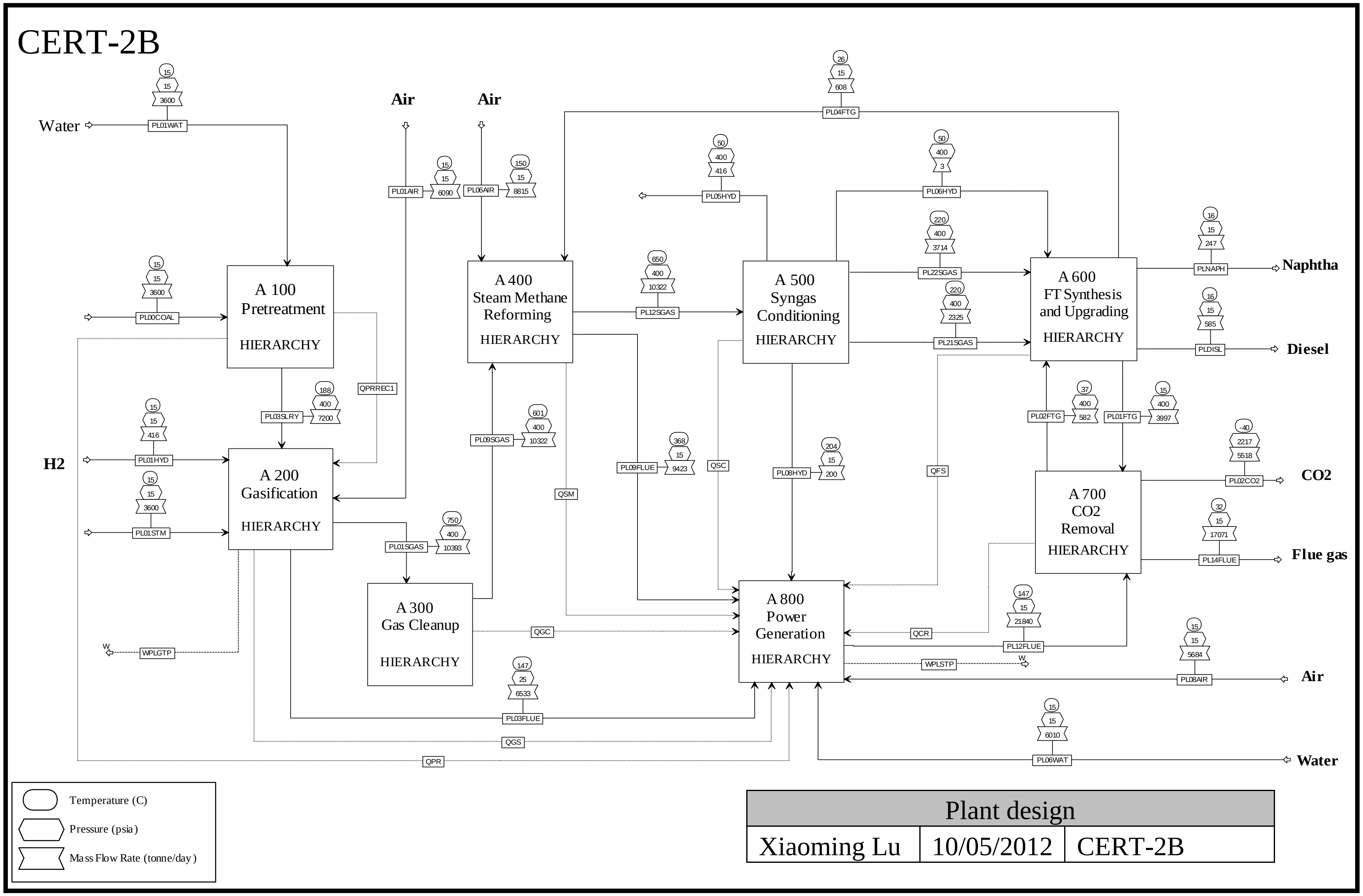

Figure D.10 Overall plant area process flow diagram for Case CERT-2B

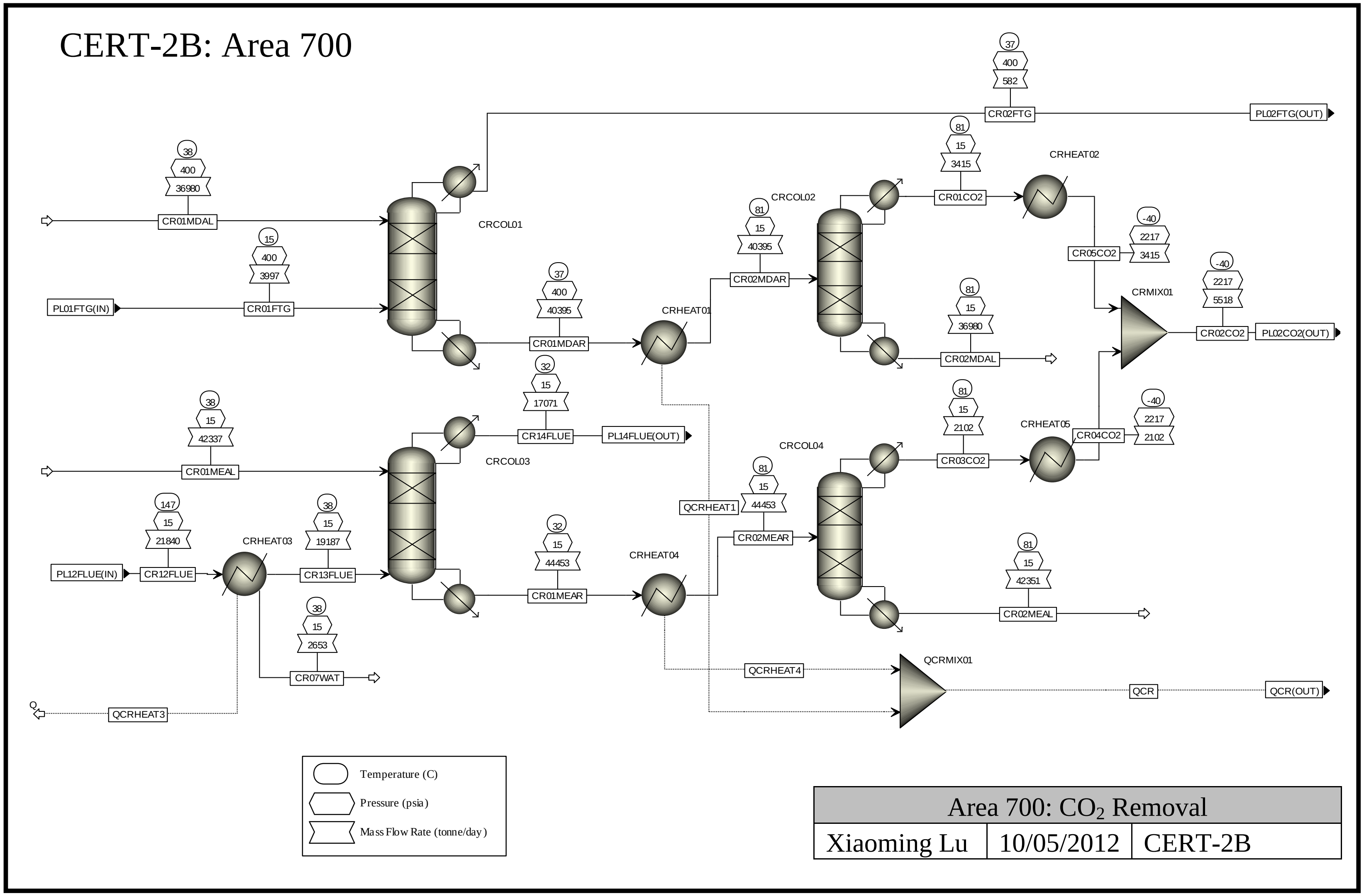

Figure D.11 CO_2 removal area process flow diagram for Case CERT-2B

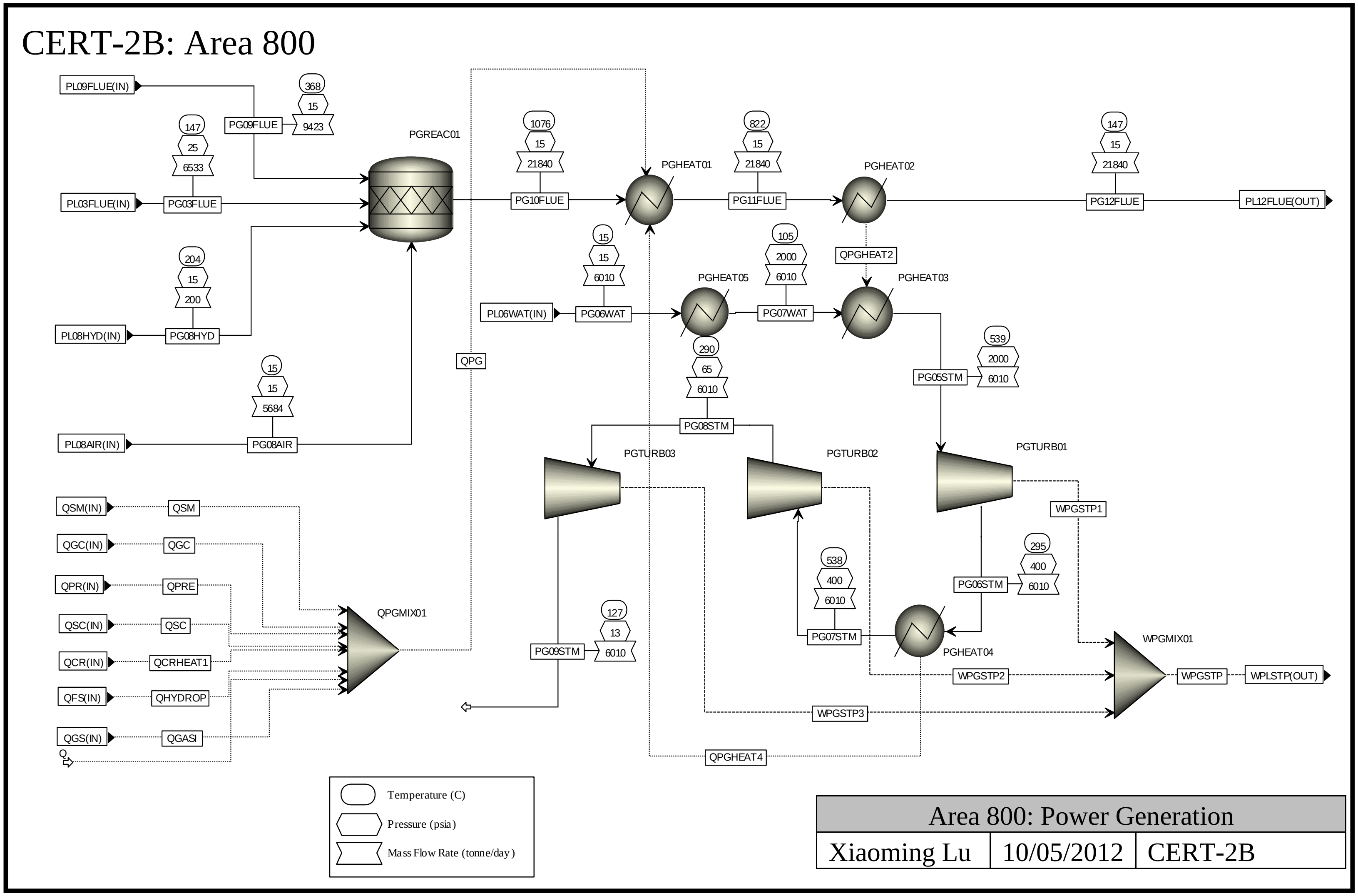

Figure D.12 Power Generation area process flow diagram for Case CERT-2B

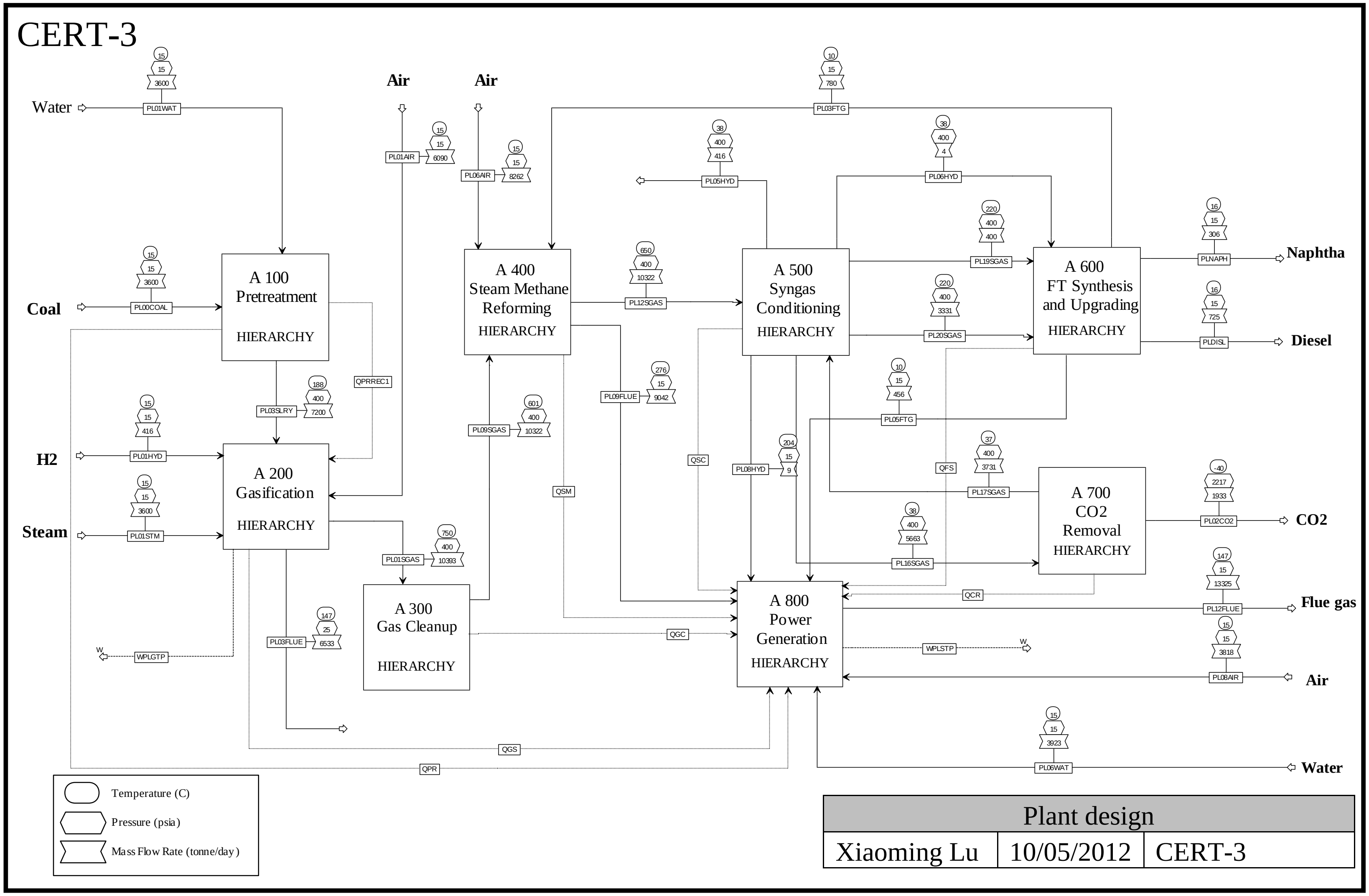

Figure D.13 Overall plant area process flow diagram for Case CERT-3

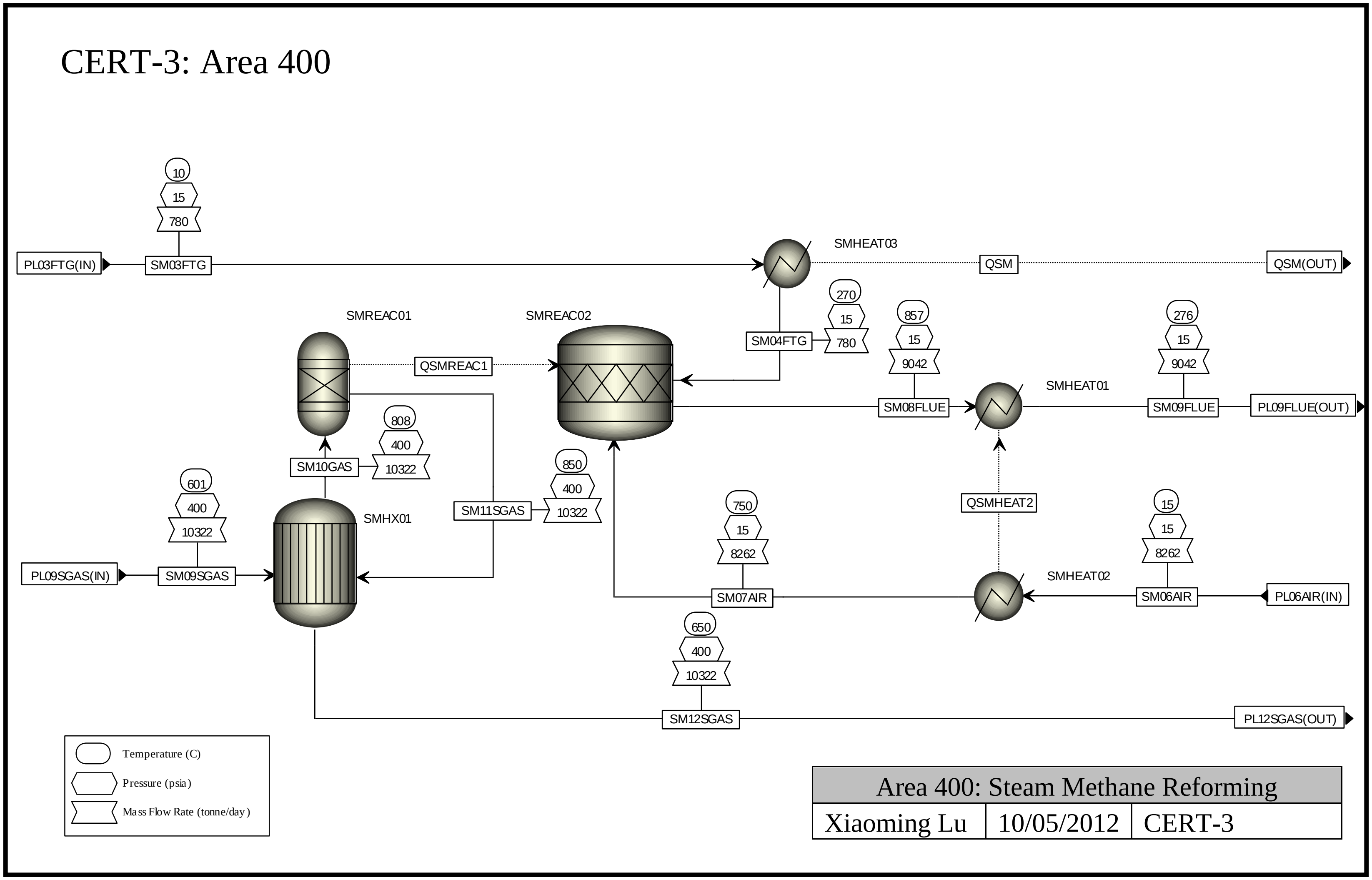

Figure D.14 Steam methane reforming area process flow diagram for Case CERT-3

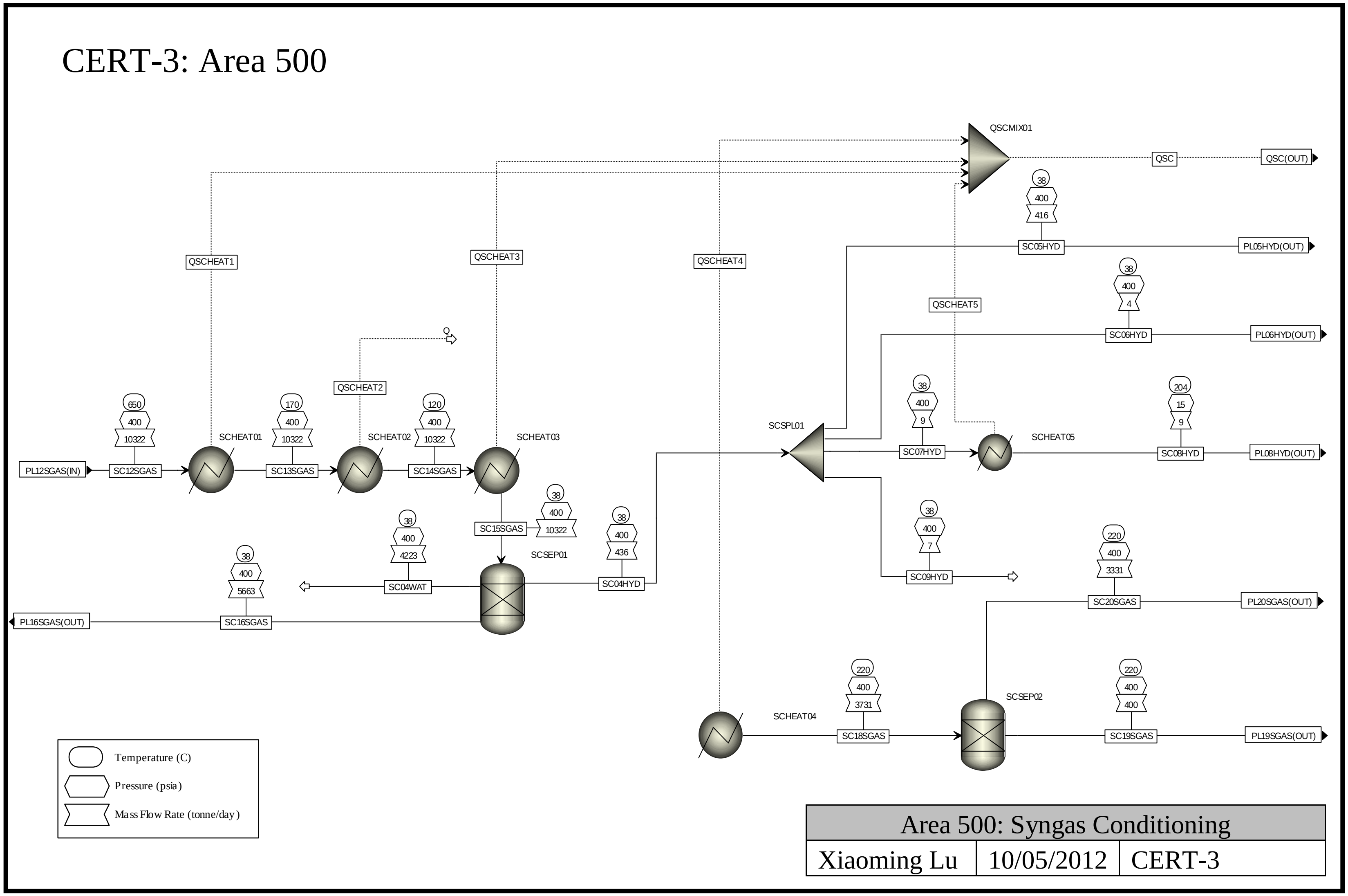

Figure D.15. Syngas conditioning area process flow diagram for Case CERT-3

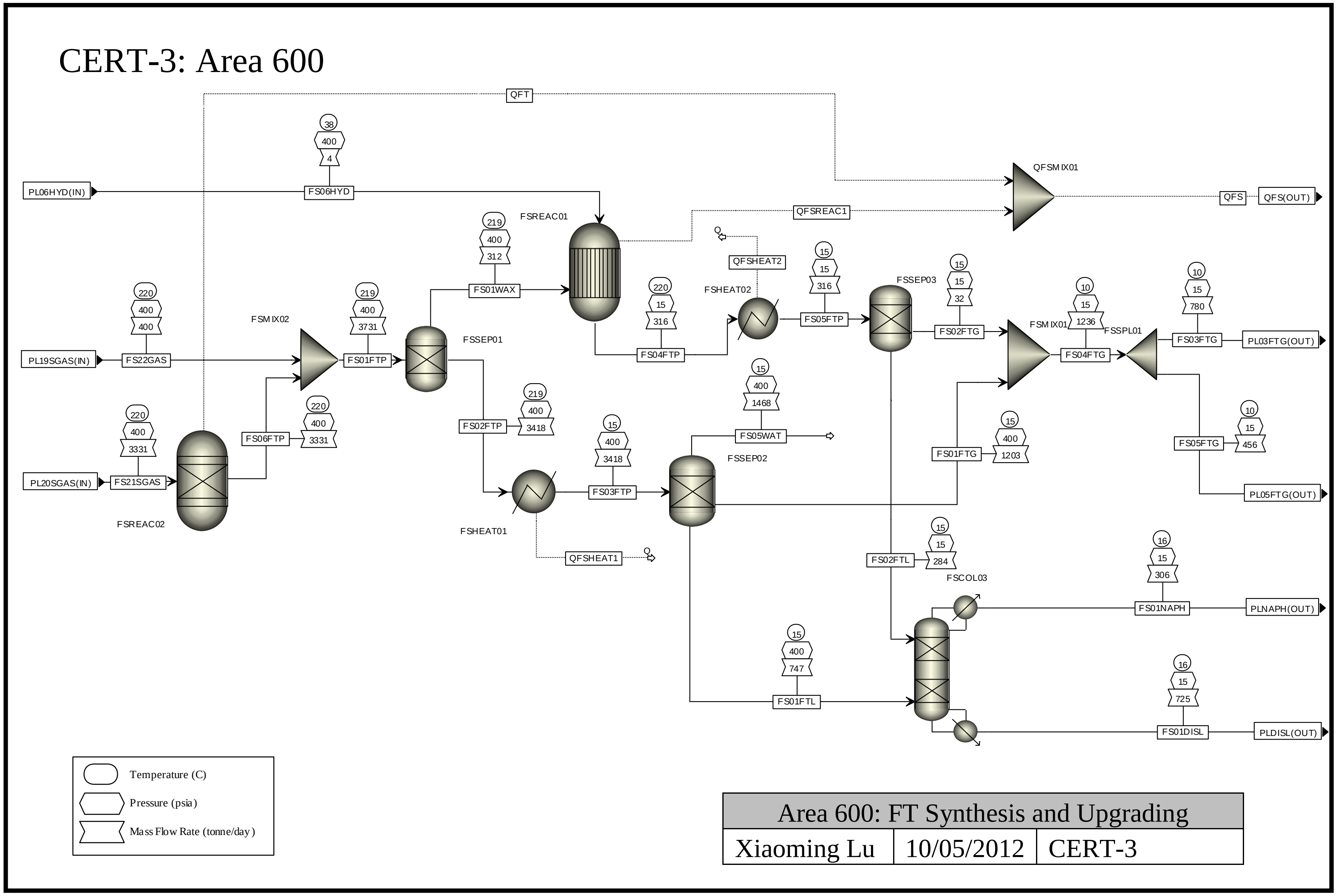

Figure D.16 Fuel synthesis and upgrading area process flow diagram for Case CERT-3

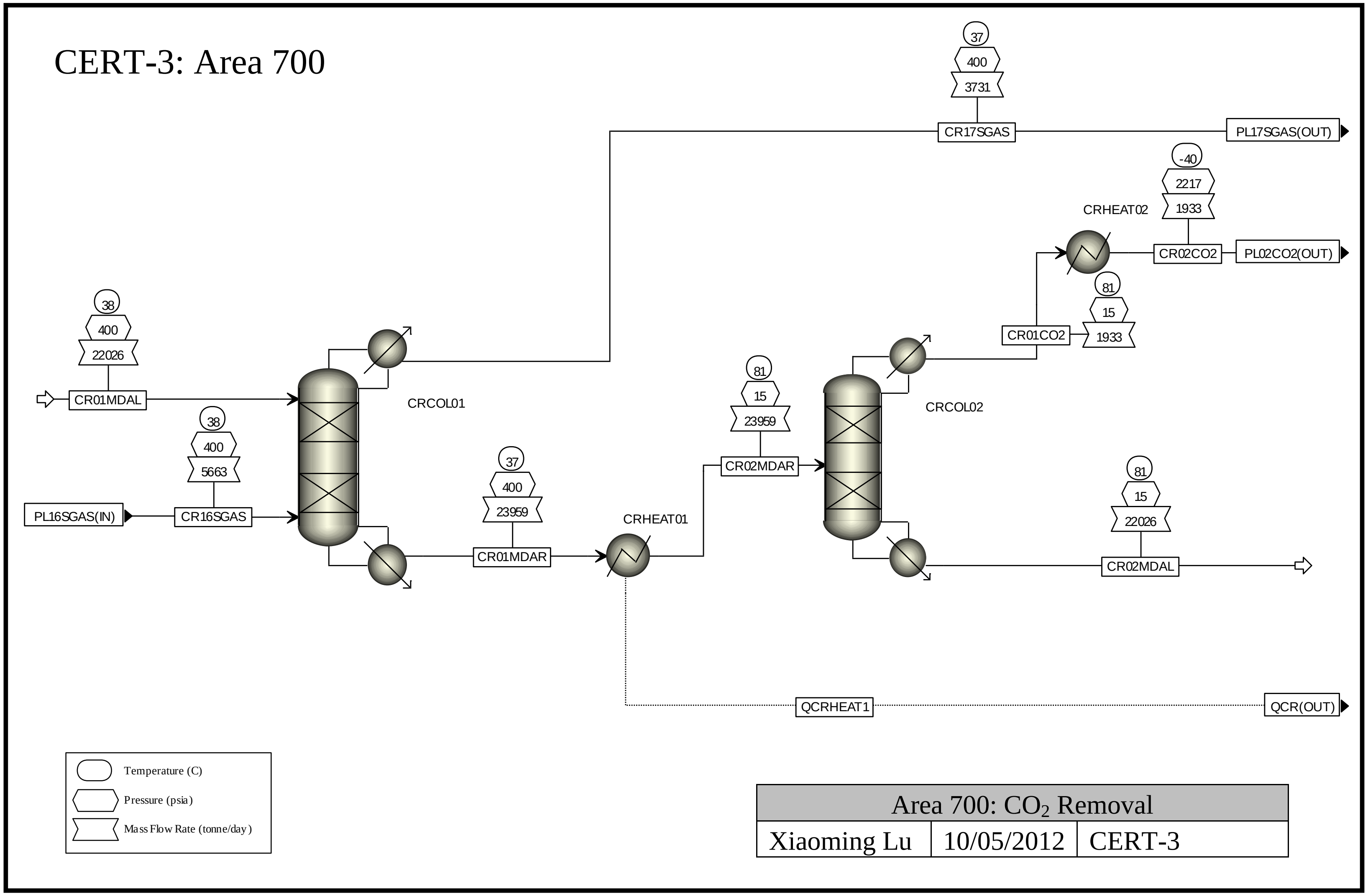

Figure D.17 CO_2 removal area process flow diagram for Case CERT-3

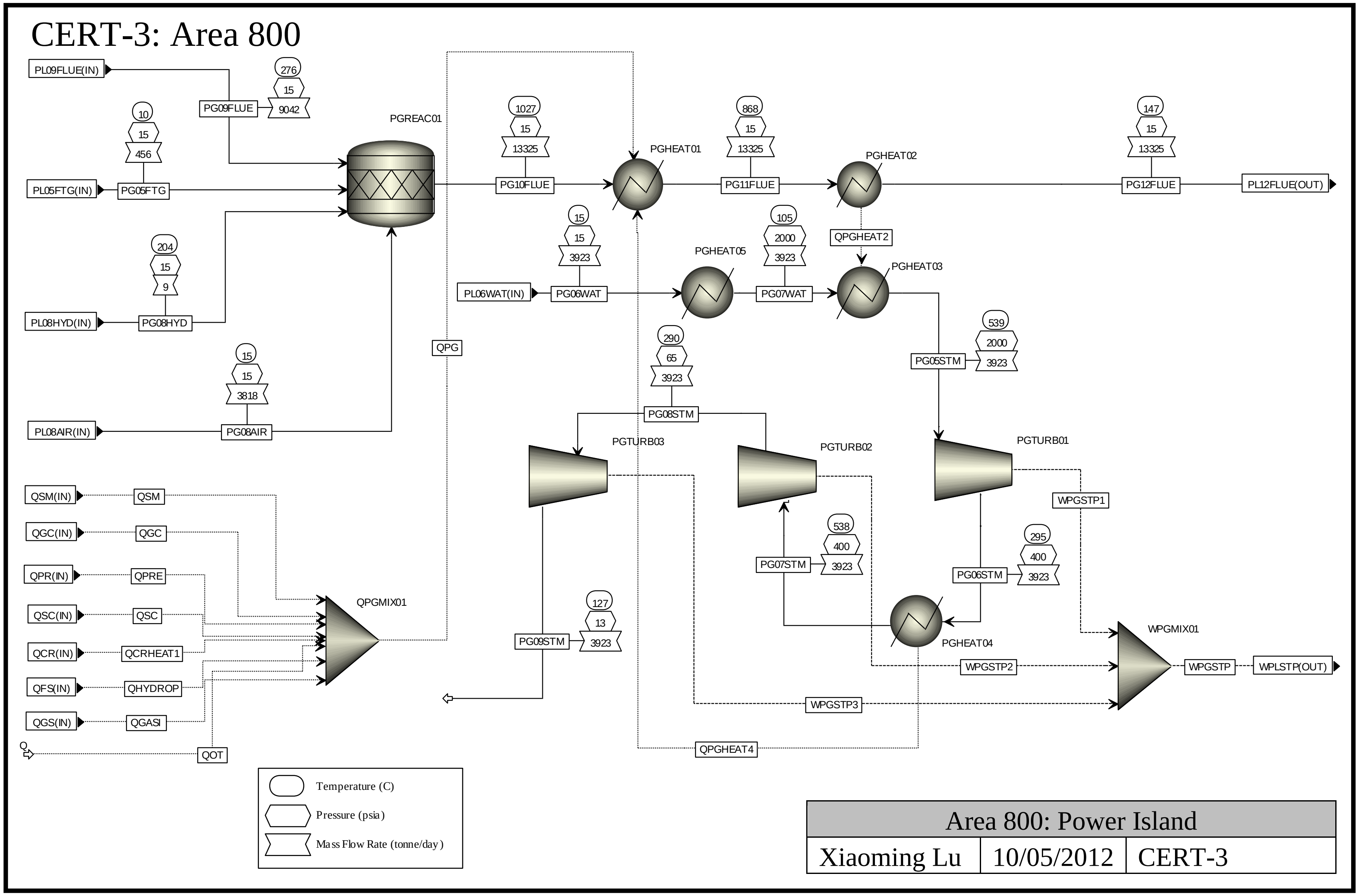

Figure D.18 Power generation area process flow diagram for Case CERT-3

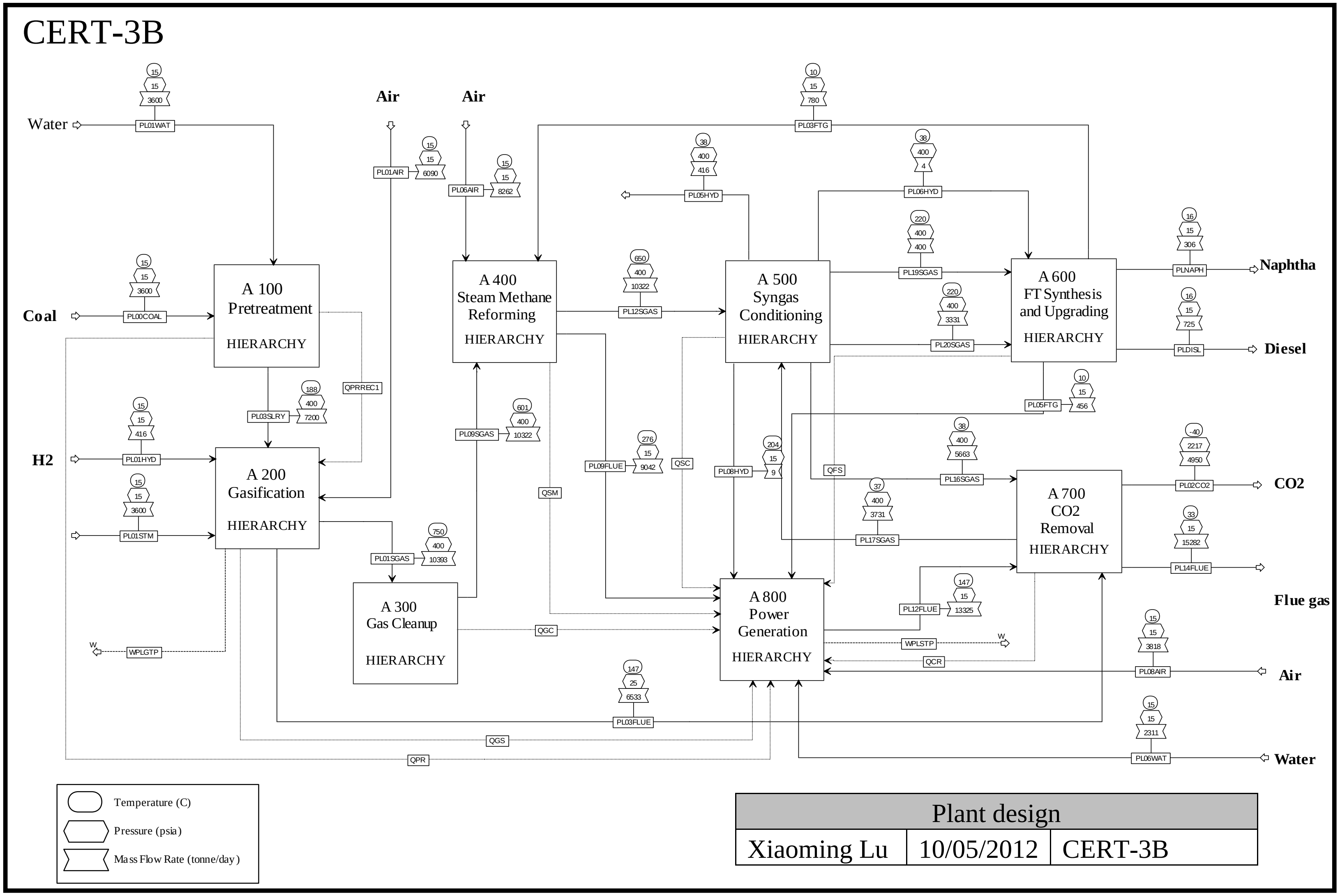

Figure D.19 Overall plant area process flow diagram for Case CERT-3B

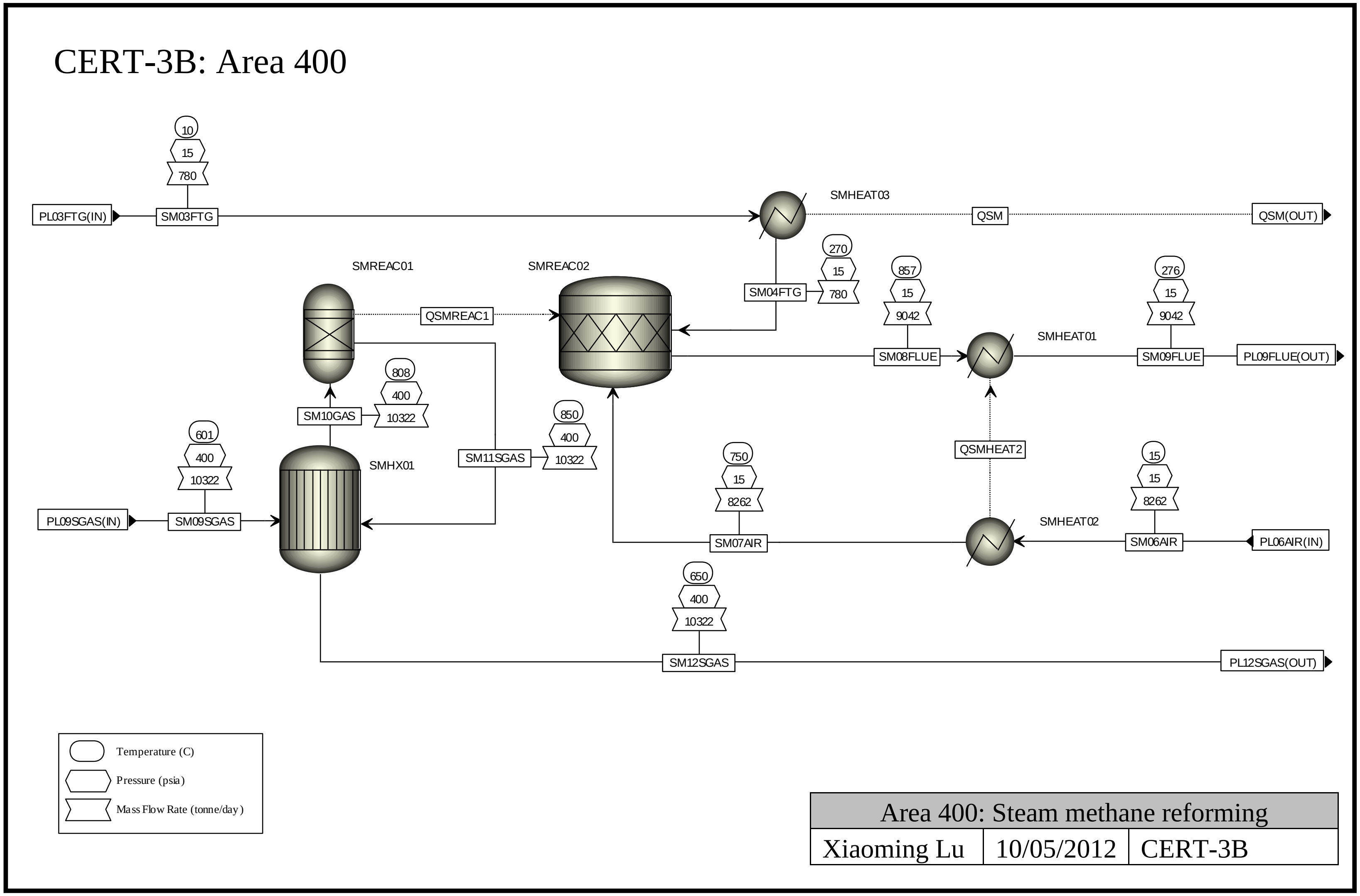

Figure D.20 Steam methane reforming area process flow diagram for Case CERT-3B

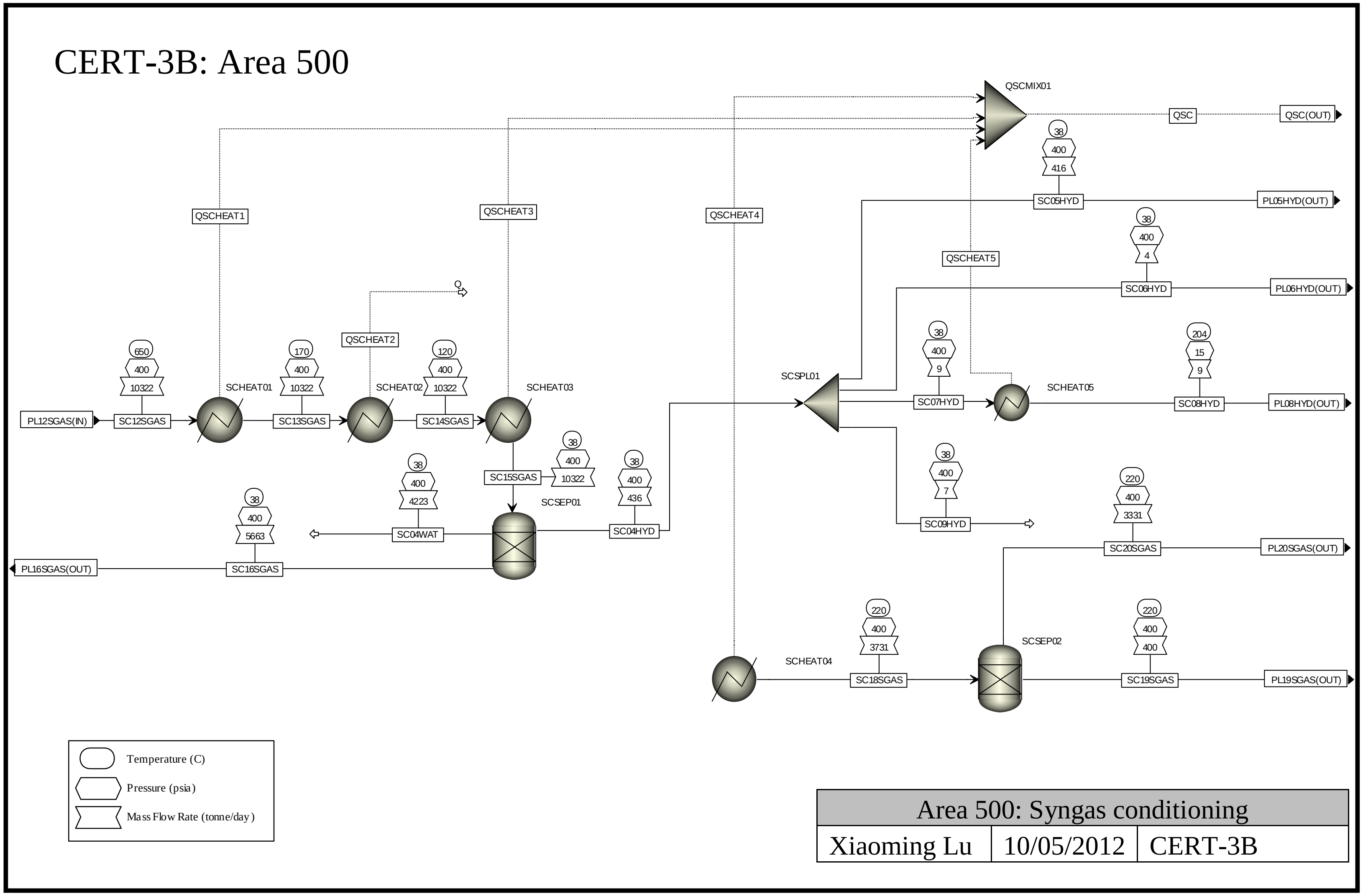

Figure D.21 Syngas conditioning area process flow diagram for Case CERT-3B

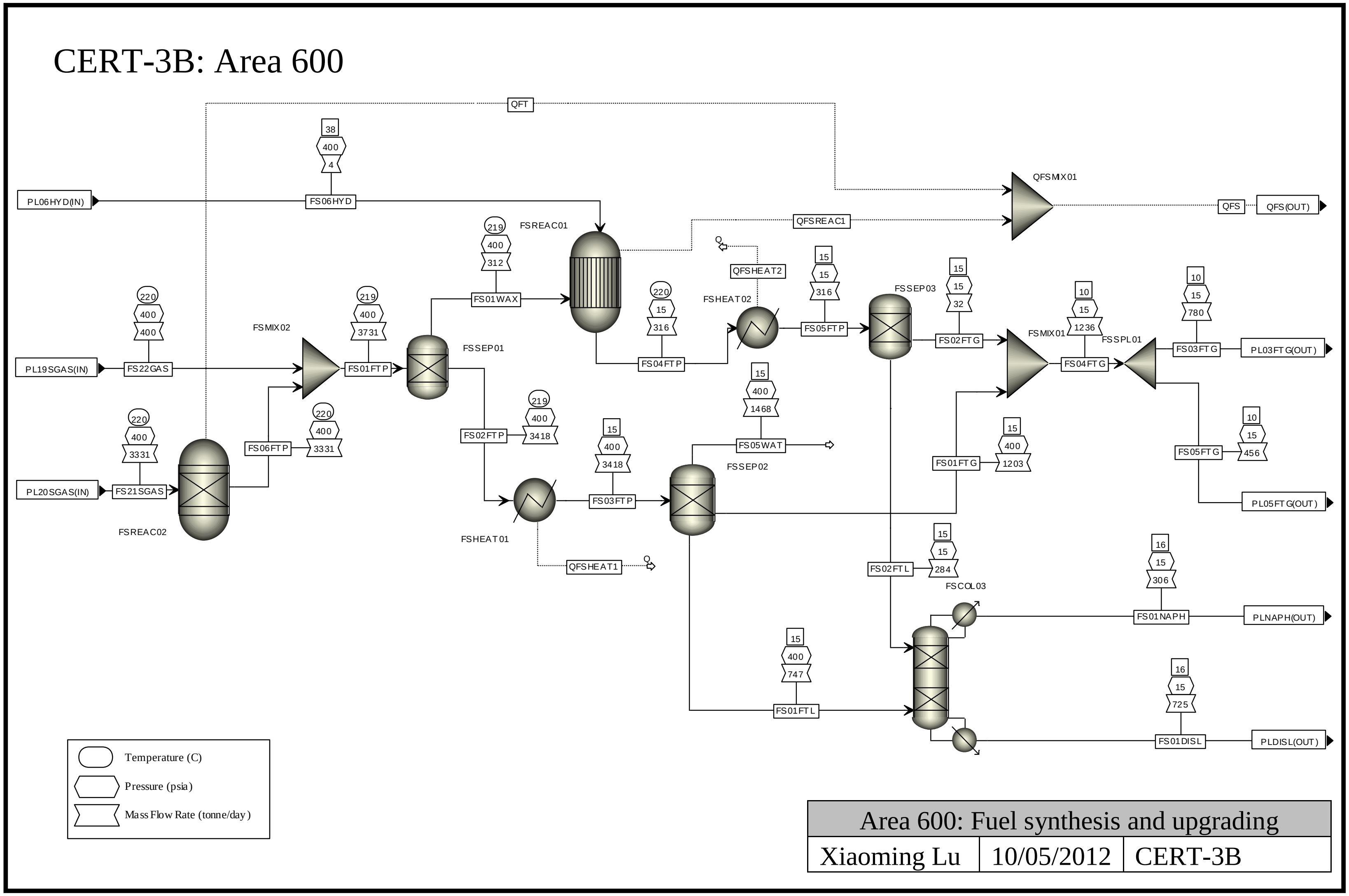

Figure D.22 Fuel synthesis and upgrading area process flow diagram for Case CERT-3B

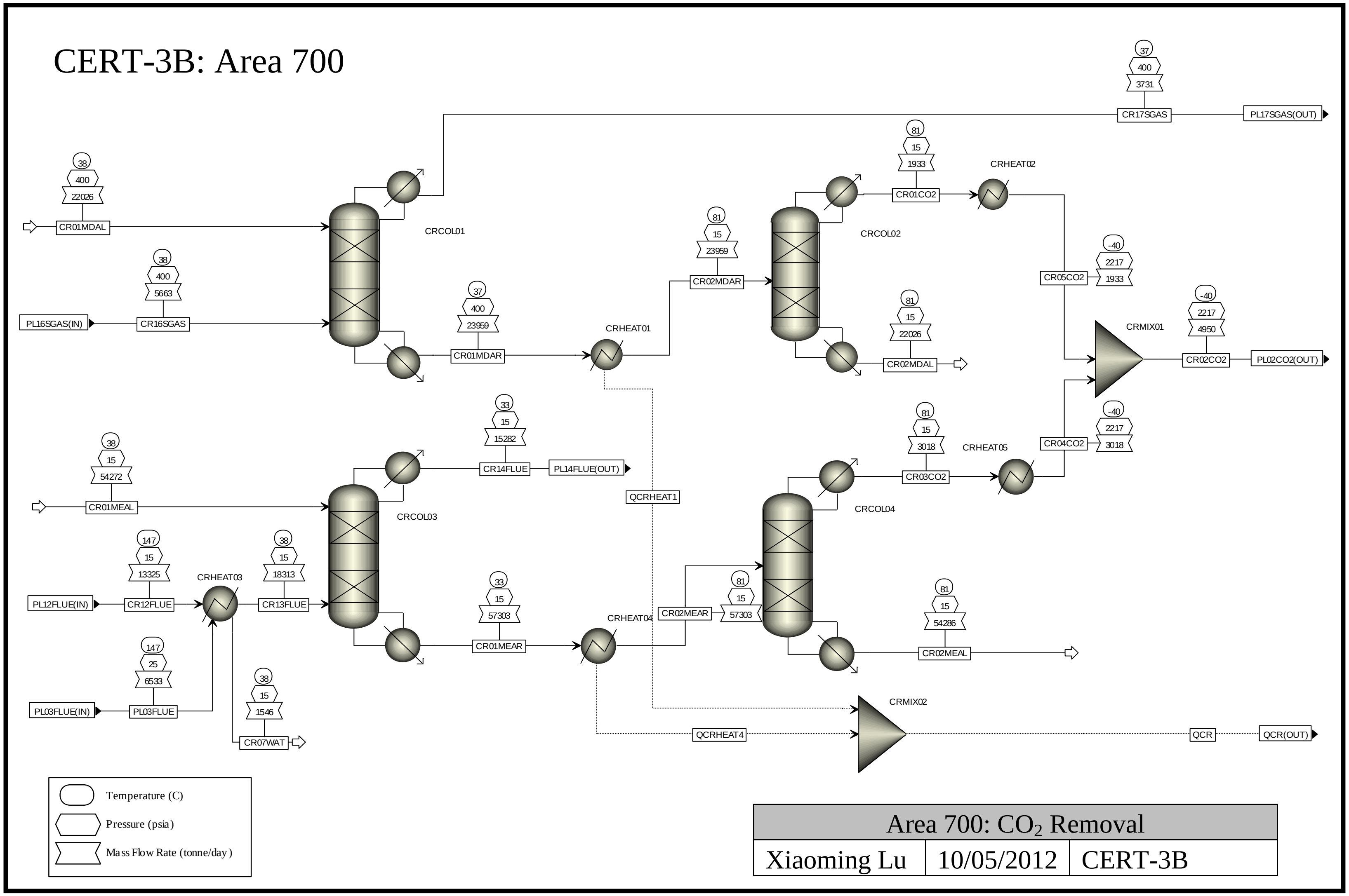

Figure D.23 CO_2 removal area process flow diagram for Case CERT-3B

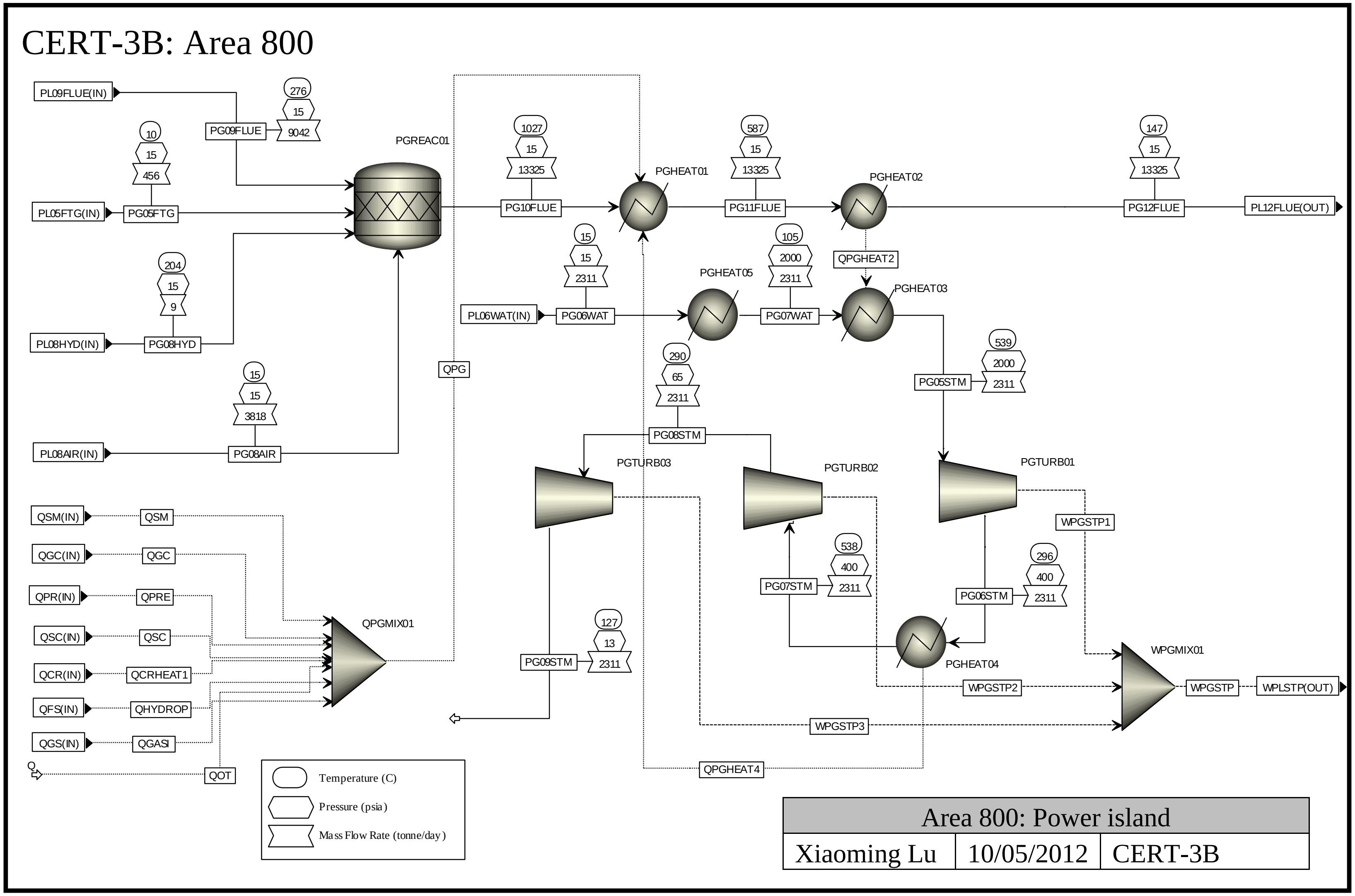

Figure D.24 Power Generation area process flow diagram for Case CERT-3B

Appendix E: Stream Data

CERT-2

Table E.1. Overall plant stream data for Case CERT-2

	PL00 COAL	PL0 1AIR	PL01 FTG	PL01 HYD	PL01 SGAS	PL01 STM	PL01 WAT	PL02 CO2	PL02 FTG	PL03 FLUE	PL03 SLRY	PL04 FTG	PL05 HYD	PL0 6AIR	PL06 HYD	PL06 WAT	PL0 8AIR	PL08 HYD	PL09 FLUE	PL09 SGAS	PL12 FLUE	PL12 SGAS	PL21 SGAS	PL22 SGAS	PLD ISL	PLN APH
T (°C)		15	15	15	750	15	15	-40	37	147	188	28	50	150	50	15	15	204	368	601	147	650	220	220	16	16
P (psia)	15	15	400	15	400	15	15	2217	400	25	400	15	400	15	400	15	15	15	15	400	15	400	400	400	15	15
Mass Flow ton/day	0	6090	3997	416	10393	3600	3600	3415	582	6533	7200	608	416	8815	3	7139	5684	200	9423	10322	21840	10322	2325	3714	585	247
C	0	0	0	0	0	0	0	0	0	0	2479	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	0	0	3	416	354	0	0	0	3	0	171	3	416	0	3	0	0	200	0	354	0	837	274	0	0	0
CO	0	0	205	0	457	0	0	0	205	0	0	205	0	0	0	0	0	0	0	457	0	2930	2051	0	0	0
CO2	0	0	3415	0	2316	0	0	3415	0	1626	0	0	0	0	0	0	0	0	1437	2316	3063	2034	0	3415	0	0
H2O	0	0	0	0	5555	3600	3600	0	0	0	3600	0	0	0	0	7139	0	0	882	5583	2667	4223	0	0	0	0
CH4	0	0	341	0	1612	0	0	0	341	0	0	350	0	0	0	0	0	0	0	1612	0	298	0	298	0	0
C2-C4	0	0	33	0	0	0	0	0	33	0	0	51	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5-C11	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	247
C12-C24	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	585	0
C24+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	1418	0	0	0	0	0	0	0	236	484	0	0	2053	0	0	1324	0	342	0	317	0	0	0	0	0
N2	0	4671	0	0	0	0	0	0	0	4671	37	0	0	6762	0	0	4360	0	6762	0	15793	0	0	0	0	0
H3N	0	0	0	0	46	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	50	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	53	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MDEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	380	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	3600	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

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Table E.2. Coal handling and preparation area stream data for Case Case-2

	PR00COAL	PR01COAL	PR01SLRY	PR01WAT	PR02COAL	PR02SLRY	PR02WAT	PR03COAL	PR03SLRY	PR05COAL	PR06COAL	PR07COAL
T (^{o}C)	15	15	15	15	15	220	220	15	188	220	220	500
P (psia)	15	15	15	15	15	400	400	15	400	400	400	400
Mass Flow tonne/day	3600	3600	7200	3600	0	7200	7200	3600	7200	3600	3600	3600
C	0	0	0	0	0	0	0	0	2479	0	0	2479
H2	0	0	0	0	0	0	0	0	171	0	0	171
CO	0	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0	0	0	0
H2O	0	0	3600	3600	0	3600	3600	0	3600	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0
C2-C4	0	0	0	0	0	0	0	0	0	0	0	0
C5-C11	0	0	0	0	0	0	0	0	0	0	0	0
C12-C24	0	0	0	0	0	0	0	0	0	0	0	0
C24+	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	484	0	0	484
N2	0	0	0	0	0	0	0	0	37	0	0	37
H3N	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	50	0	0	50
H2S	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0	0	0	0	0
MDEA	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	380	0	0	380
Coal	3600	3600	3600	0	0	3600	0	3600	0	3600	3600	0

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Table E.3. Gasification area stream data for Case Case-2

Streams	GS01 AIR	GS01 ASH	GS01C HAR	GS01F LUE	GS01 HYD	GS01S AND	GS01S GAS	GS01S GMX	GS01S MIX	GS01 STM	GS02 AIR	GS02F LUE	GS02 HYD	GS02S AND	GS02S GMX	GS02S MIX	GS02 STM	GS03 AIR	GS03F LUE	GS03S AND	GS03 SLRY	GS04 AIR
T (^{o}C)	15	900	750	900	15		750	750	750	15	512	494	324	750	900	900	735	280	147	900	188	743
P (psia)	15	400	400	400	15	400	400	400	400	15	400	400	400	400	400	400	400	400	25	400	400	400
Mass Flow tonne/day	6090	380	823	6533	416	61674	10393	72890	62498	3600	6090	6533	416	61674	68587	62054	3600	6090	6533	61674	7200	6090
C	0	0	444	0	0	0	0	444	444	0	0	0	0	0	0	0	0	0	0	0	2479	0
H2	0	0	0	0	416	0	354	354	0	0	0	0	416	0	0	0	0	0	0	0	171	0
CO	0	0	0	0	0	0	457	457	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	1626	0	0	2316	2316	0	0	0	1626	0	0	1626	0	0	0	1626	0	0	0
H2O	0	0	0	0	0	0	5555	5555	0	3600	0	0	0	0	0	0	3600	0	0	0	3600	0
CH4	0	0	0	0	0	0	1612	1612	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C2-C4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5-C11	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C12-C24	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C24+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	1418	0	0	236	0	0	0	0	0	0	1418	236	0	0	236	0	0	1418	236	0	484	1418
N2	4671	0	0	4671	0	0	0	0	0	0	4671	4671	0	0	4671	0	0	4671	4671	0	37	4671
NH3	0	0	0	0	0	0	46	46	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	50	0
H2S	0	0	0	0	0	0	53	53	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	61674	0	61674	61674	0	0	0	0	61674	61674	61674	0	0	0	61674	0	0
MDEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	380	380	0	0	0	0	380	380	0	0	0	0	0	380	380	0	0	0	0	380	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

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Table E.4 Warm gas cleanup area stream data for Case Case-2

Streams	GC01NH3	GC01 SGAS	GC01SUL	GC01 SULD	GC01ZNO	GC01ZNS	GC02 SGAS	GC02ZNO	GC03HYD	GC03 SGAS	GC03 SGMX	GC03 WAT	GC04 FLUE	GC04 SGAS	GC04 SGMX	GC05AIR	GC05 FLUE	GC05GAS	GC05 SGMX	GC06 FLUE	GC06 SGAS	GC07 FLUE	GC07 SGAS	GC08 SGAS	GC09 SGAS
T (°C)	315	750	140	815	500	498	480	815	50	480	498	40	700	480	815	15	130	498	700	40	315	40	315	315	601
P (psia)	400	400	400	400	15	400	400	400	15	400	400	15	400	400	400	15	15	400	400	15	400	15	400	400	400
Mass Flow tonne/day	46	10393	50	344	630	655	10393	630	7	10393	11023	55	301	10393	975	319	301	10368	351	246	10368	246	10368	10322	10322
C	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	0	354	0	0	0	0	354	0	7	354	354	0	1	354	0	0	1	354	1	1	354	1	354	354	354
CO	0	457	0	0	0	0	457	0	0	457	457	0	0	457	0	0	0	457	0	0	457	0	457	457	457
CO2	0	2316	0	0	0	0	2316	0	0	2316	2316	0	0	2316	0	0	0	2316	0	0	2316	0	2316	2316	2316
H2O	0	5555	0	0	0	0	5555	0	0	5555	5583	55	56	5555	0	0	56	5583	56	0	5583	0	5583	5583	5583
CH4	0	1612	0	0	0	0	1612	0	0	1612	1612	0	0	1612	0	0	0	1612	0	0	1612	0	1612	1612	1612
C2-C4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5-C11	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C12-C24	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C24+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	74	0	0	0	0	0	0	0	0	0
N2	0	0	0	245	0	0	0	0	0	0	0	0	245	0	245	245	245	0	245	245	0	245	0	0	0
NH3	46	46	0	0	0	0	46	0	0	46	46	0	0	46	0	0	0	46	0	0	46	0	46	0	0
SULFUR	0	0	50	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	50	0	0	0	0	0	0
H2S	0	53	0	0	0	0	53	0	0	53	0	0	0	53	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	99	0	0	0	0	0	0	0	0	0	0	99	0	0	0	0	0	0	0	0	0	0
ZNO	0	0	0	0	630	504	0	630	0	0	504	0	0	0	630	0	0	0	0	0	0	0	0	0	0
ZNS	0	0	0	0	0	151	0	0	0	0	151	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

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Table E.5 Steam methane reforming area stream data for Case Case-2

Streams	SM04FTG	SM05FTG	SM06AIR	SM07AIR	SM08FLUE	SM09FLUE	SM09SGAS	SM10GAS	SM11SGAS	SM12SGAS
T (^{o}C)	38	270	150	750	858	368	601	808	850	650
P (psia)	15	15	15	15	15	15	400	400	400	400
Mass Flow tonne/day	608	608	8815	8815	9423	9423	10322	10322	10322	10322
C	0	0	0	0	0	0	0	0	0	0
H2	3	3	0	0	0	0	354	354	837	837
CO	205	205	0	0	0	0	457	457	2930	2930
CO2	0	0	0	0	1437	1437	2316	2316	2034	2034
H2O	0	0	0	0	882	882	5583	5583	4223	4223
CH4	350	350	0	0	0	0	1612	1612	298	298
C2-C4	51	51	0	0	0	0	0	0	0	0
C5-C11	0	0	0	0	0	0	0	0	0	0
C12-C24	0	0	0	0	0	0	0	0	0	0
C24+	0	0	0	0	0	0	0	0	0	0
O2	0	0	2053	2053	342	342	0	0	0	0
N2	0	0	6762	6762	6762	6762	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0	0	0
MDEA	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0

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Table E.6 Gas conditioning area stream data for Case Case-2

Streams	PL21 SGAS	SC01 CO2	SC02 CO2	SC03 STM	SC04 HYD	SC04 STM	SC04 WAT	SC05 HYD	SC06 HYD	SC07 HYD	SC08 HYD	SC09 HYD	SC12 SGAS	SC13 SGAS	SC14 SGAS	SC15 SGAS	SC16 SGAS	SC17 SGAS	SC18 SGAS	SC19 SGAS	SC20 SGAS	SC22 SGAS
T (^{o}C)	220	220	220	220	38	220	38	38	38	38	204	38	650	170	120	38	38	38	38	220	220	220
P (psia)	400	400	400	400	400	400	400	400	400	400	15	400	400	400	400	400	400	400	400	400	400	400
Mass Flow tonne/day	2325	1438	1381	622	626	57	4223	416	3	200	200	7	10322	10322	10322	10322	5473	3141	2333	3763	2333	3714
C																						
H2	274	0	0	0	626	0	0	416	3	200	200	7	837	837	837	837	211	211	0	274	0	0
CO	2051	0	0	0	0	0	0	0	0	0	0	0	2930	2930	2930	2930	2930	2930	0	2051	0	0
CO2	0	1381	1381	0	0	0	0	0	0	0	0	0	2034	2034	2034	2034	2034	0	2034	1381	2034	3415
H2O	0	57	0	622	0	57	4223	0	0	0	0	0	4223	4223	4223	4223	0	0	0	57	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	298	298	298	298	298	0	298	0	298	298
C2-C4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5-C11	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C12-C24	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C24+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MDEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

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Table E.7 Fuel synthesis and upgrading area stream data for Case Case-2

Streams	FS01DISL	FS01FTG	FS01FTL	FS01FTP	FS01NAPH	FS01WAX	FS02FTG	FS02FTL	FS02FTP	FS03FTG	FS03FTP	FS04FTG	FS04FTP	FS05FTP	FS05WAT	FS06FTP	FS06HYD	FS21SGAS	FS22GAS
T (°C)	15	15	215	16	215	37	15	215	15	15	26	220	15	15	220	38	220	220	220
P (psia)	400	400	400	15	400	400	15	400	15	400	15	15	15	400	400	400	400	400	400
Mass Flow tonne/day	585	3997	603	6039	247	252	582	229	5786	27	5786	608	256	256	1186	2325	3	2325	3714
C	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	0	3	0	3	0	0	3	0	3	0	3	3	0	0	0	3	3	274	0
CO	0	205	0	205	0	0	205	0	205	0	205	205	0	0	0	205	0	2051	0
CO2	0	3415	0	3415	0	0	0	0	3415	0	3415	0	0	0	0	0	0	0	3415
H2O	0	0	0	1186	0	0	0	0	1186	0	1186	0	0	0	1186	1186	0	0	0
CH4	0	341	0	341	0	0	341	0	341	9	341	350	9	9	0	42	0	0	298
C2-C4	0	33	0	33	0	0	33	0	33	17	33	51	17	17	0	33	0	0	0
C5-C11	0	0	206	206	247	0	0	41	206	0	206	0	41	41	0	206	0	0	0
C12-C24	585	0	397	397	0	0	0	188	397	0	397	0	188	188	0	397	0	0	0
C24+	0	0	0	252	0	252	0	0	0	0	0	0	0	0	0	252	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MDEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

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Table E.8 CO$_2$ removal area stream data for Case Case-2

Streams	CR01CO2	CR01FTG	CR01MDAL	CR01MDAR	CR02CO2	CR02FTG	CR02MDAL	CR02MDAR
T (°C)	81	25	38	38	-40	38	81	81
P (psia)	15	400	400	400	2217	400	15	15
Mass Flow tonne/day	3415	3997	36980	40395	3415	582	36980	40395
C	0	0	0	0	0	0	0	0
H2	0	3	0	0	0	3	0	0
CO	0	205	0	0	0	205	0	0
CO2	3415	3415	0	3415	3415	0	0	3415
H2O	0	0	18484	18484	0	0	18484	18484
CH4	0	341	0	0	0	341	0	0
C2-C4	0	33	0	0	0	33	0	0
C5-C11	0	0	0	0	0	0	0	0
C12-C24	0	0	0	0	0	0	0	0
C24+	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0
MDEA	0	0	18496	18496	0	0	18496	18496
ASH	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0

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Table E.9 Power generation area stream data for Case Case-2

Streams	PG03FLUE	PG05STM	PG06STM	PG06WAT	PG07STM	PG07WAT	PG08AIR	PG08HYD	PG08STM	PG09FLUE	PG09STM	PG10FLUE	PG11FLUE	PG12FLUE
T (^{o}C)	147	538	295	15	538	105	15	204	290	368	127	1076	946	147
P (psia)	25	2000	400	15	400	2000	15	15	65	15	13	15	15	15
Mass Flow tonne/day	6533	7223	7223	7223	7223	7223	5684	200	7223	9423	7223	21840	21840	21840
C														
H2	0	0	0	0	0	0	0	200	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	1626	0	0	0	0	0	0	0	0	1437	0	3063	3063	3063
H2O	0	7223	7223	7223	7223	7223	0	0	7223	882	7223	2667	2667	2667
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C2-C4	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5-C11	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C12-C24	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C24+	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	236	0	0	0	0	0	1324	0	0	342	0	317	317	317
N2	4671	0	0	0	0	0	4360	0	0	6762	0	15793	15793	15793
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MDEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0

Case-2B

Table E.10 Overall plant stream data for Case CERT-2B

	PL00COAL	PL01 AIR	PL01 FTG	PL01 HYD	PL01 SGAS	PL01 STM	PL01 WAT	PL02 CO2	PL02 FTG	PL03 FLUE	PL03 SLRY	PL04 FTG	PL05 HYD	PL06AIR	PL06 HYD	PL06 WAT	PL08AIR	PL08 HYD	PL09 FLUE	PL09 SGAS	PL14 FLUE	PL12 SGAS	PL21 SGAS	PL22 SGAS	PLDISL	PLN APH
T (°C)		15	15	15	750	15	15	-40	37	147	188	28	50	150	50	15	15	204	368	601	147	650	220	220	16	16
P (psia)	15	15	400	15	400	15	15	2217	400	25	400	15	400	15	400	15	15	15	15	400	15	400	400	400	15	15
Mass Flow ton/day	0	6090	3997	416	10393	3600	3600	5518	582	6533	7200	608	416	8815	3	6010	5684	200	9423	10322	17071	10322	2325	3714	585	247
C	0	0	0	0	0	0	0	0	0	0	2479	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	0	0	3	416	354	0	0	0	3	0	171	3	416	0	3	0	0	200	0	354	0	837	274	0	0	0
CO	0	0	205	0	457	0	0	0	205	0	0	205	0	0	0	0	0	0	0	457	0	2930	2051	0	0	0
CO2	0	0	3415	0	2316	0	0	5518	0	1626	0	0	0	0	0	0	0	0	1437	2316	961	2034	0	3415	0	0
H2O	0	0	0	0	5555	3600	3600	0	0	0	3600	0	0	0	0	6010	0	0	882	5583	0	4223	0	0	0	0
CH4	0	0	341	0	1612	0	0	0	341	0	0	350	0	0	0	0	0	0	0	1612	0	298	0	298	0	0
C2-C4	0	0	33	0	0	0	0	0	33	0	0	51	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5-C11	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	247
C12-C24	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	585	0
C24+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	1418	0	0	0	0	0	0	0	236	484	0	0	2053	0	0	1324	0	342	0	317	0	0	0	0	0
N2	0	4671	0	0	0	0	0	0	0	4671	37	0	0	6762	0	0	4360	0	6762	0	15793	0	0	0	0	0
H3N	0	0	0	0	46	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	50	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	53	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MDEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	380	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	3600	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

CERT-2B

Table E.11 CO_2 removal stream data for Case Case-2B

	CR01 CO2	CR01 FTG	CR01M DAL	CR01M DAR	CR01 MEAL	CR01M EAR	CR02 CO2	CR02 FTG	CR02M DAL	CR02M DAR	CR02 MEAL	CR02M EAR	CR03 CO2	CR04 CO2	CR05 CO2	CR07 WAT	CR12F LUE	CR13F LUE	CR14F LUE
T (°C)	81	15	38	37	38	32	-40	37	81	81	81	81	81	-40	-40	38	147	38	32
P (psia)	15	400	400	400	15	15	2217	400	15	15	15	15	15	2217	2217	15	15	15	15
Mass Flow tonne/day	3415	3997	36980	40395	42337	44453	5518	549	36980	40395	42351	44453	2102	2102	3415	2653	21840	19187	17071
C	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	0	3	0	0	0	0	0	3	0	0	0	0	0	0	0	0	0	0	0
CO	0	205	0	0	0	0	0	205	0	0	0	0	0	0	0	0	0	0	0
CO2	3415	3415	0	3415	0	2102	5518	0	0	3415	0	2102	2102	2102	3415	0	3063	3063	961
H2O	0	0	18484	18484	33834	33848	0	0	18484	18484	33848	33848	0	0	0	2653	2667	14	0
CH4	0	341	0	0	0	0	0	341	0	0	0	0	0	0	0	0	0	0	0
C2-C4	0	33	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5-C11	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C12-C24	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C24+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	317	317	317
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	15793	15793	15793
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MDEA	0	0	18496	18496	0	0	0	0	18496	18496	0	0	0	0	0	0	0	0	0
MEA	0	0	0	0	8502	8502	0	0	0	0	8502	8502	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

CERT-2B

Table E.12 Power generation area stream data for Case Case-2B

	PG03FLUE	PG05STM	PG06STM	PG06WAT	PG07STM	PG07WAT	PG08AIR	PG08HYD	PG08STM	PG09FLUE	PG09STM	PG10FLUE	PG11FLUE	PG12FLUE
T (°C)	147	539	295	15	538	105	15	204	290	368	127	1076	822	147
P (psia)	25	2000	400	15	400	2000	15	15	65	15	13	15	15	15
Mass Flow tonne/day	6533	6010	6010	6010	6010	6010	5684	200	6010	9423	6010	21840	21840	21840
C	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	200	0	0	0	0	0	0
CO	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	1626	0	0	0	0	0	0	0	0	1437	0	3063	3063	3063
H2O	0	6010	6010	6010	6010	6010	0	0	6010	882	6010	2667	2667	2667
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C2-C4	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5-C11	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C12-C24	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C24+	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	236	0	0	0	0	0	1324	0	0	342	0	317	317	317
N2	4671	0	0	0	0	0	4360	0	0	6762	0	15793	15793	15793
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MDEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0

CERT-3

Table E.13 Overall plant stream data for Case CERT-3

	PL00COAL	PL01AIR	PL01HYD	PL01SGAS	PL01STM	PL01WAT	PL02CO2	PL03FLUE	PL03FTG	PL03SLRY	PL05FTG	PL05HYD	PL06AIR	PL06HYD	PL06WAT	PL08AIR	PL08HYD	PL09FLUE	PL09SGAS	PL12FLUE	PL12SGAS	PL16SGAS	PL17SGAS	PL19SGAS	PL20SGAS	PLDISL	PLNAPH
T (°C)	15	15	15	750	15	15	-40	147	10	188	10	38	15	38	15	15	204	276	601	147	650	38	38	220	220	16	16
P (psia)	15	15	15	400	15	15	2217	25	15	400	15	400	15	400	15	15	15	15	400	15	400	400	400	400	400	15	15
Mass Flow tonne/day	3600	6090	416	10393	3600	3600	1933	6533	780	7200	456	416	8262	4	3923	3818	9	9042	10322	13325	10322	5663	3731	400	3331	725	306
C	0	0	0	0	0	0	0	0	0	2479	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	0	0	416	354	0	0	0	0	40	171	24	416	0	4	0	0	9	0	354	0	837	401	401	0	401	0	0
CO	0	0	0	457	0	0	0	0	407	0	238	0	0	0	0	0	0	0	457	0	2930	2930	2930	0	2930	0	0
CO2	0	0	0	2316	0	0	1933	1626	64	0	38	0	0	0	0	0	0	1452	2316	2301	2034	2034	102	102	0	0	0
H2O	0	0	0	5555	3600	3600	0	0	0	3600	0	0	0	0	3923	0	0	932	5583	1559	4223	0	0	0	0	0	0
CH4	0	0	0	1612	0	0	0	0	228	0	133	0	0	0	0	0	0	0	1612	0	298	298	298	298	0	0	0
C2-C4	0	0	0	0	0	0	0	0	40	0	23	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5-C11	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	306
C12-C24	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	725	
C24+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	1418	0	0	0	0	0	236	0	484	0	0	1924	0	0	889	0	321	0	199	0	0	0	0	0	0	0
N2	0	4671	0	0	0	0	0	4671	0	37	0	0	6338	0	0	2929	0	6338	0	9266	0	0	0	0	0	0	0
NH3	0	0	0	46	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	50	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	53	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MDEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	380	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	3600	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

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Table E.14 Steam methane reforming area stream data for Case CERT-3

	SM03FTG	SM04FTG	SM06AIR	SM07AIR	SM08FLUE	SM09FLUE	SM09SGAS	SM10GAS	SM11SGAS	SM12SGAS
T (^{o}C)	10	270	15	750	857	276	601	808	850	650
P (psia)	15	15	15	15	15	15	400	400	400	400
Mass Flow tonne/day	780	780	8262	8262	9042	9042	10322	10322	10322	10322
C	0	0	0	0	0	0	0	0	0	0
H2	40	40	0	0	0	0	354	354	837	837
CO	407	407	0	0	0	0	457	457	2930	2930
CO2	64	64	0	0	1452	1452	2316	2316	2034	2034
H2O	0	0	0	0	932	932	5583	5583	4223	4223
CH4	228	228	0	0	0	0	1612	1612	298	298
C2-C4	40	40	0	0	0	0	0	0	0	0
C5-C11	0	0	0	0	0	0	0	0	0	0
C12-C24	0	0	0	0	0	0	0	0	0	0
C24+	0	0	0	0	0	0	0	0	0	0
O2	0	0	1924	1924	321	321	0	0	0	0
N2	0	0	6338	6338	6338	6338	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0	0	0
MDEA	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0

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Table E.15 Gas conditioning area stream data for Case CERT-3

	SC04HYD	SC04WAT	SC05HYD	SC06HYD	SC07HYD	SC08HYD	SC09HYD	SC12SGAS	SC13SGAS	SC14SGAS	SC15SGAS	SC16SGAS	SC17SGAS	SC18SGAS	SC19SGAS	SC20SGAS
T (°C)	38	38	38	38	38	204	38	650	170	120	38	38	38	220	220	220
P (psia)	400	400	400	400	400	15	400	400	400	400	400	400	400	400	400	400
Mass Flow tonne/day	436	4223	416	4	9	9	7	10322	10322	10322	10322	5663	3731	3731	400	3331
C	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	436	0	416	4	9	9	7	837	837	837	837	401	401	401	0	401
CO	0	0	0	0	0	0	0	2930	2930	2930	2930	2930	2930	2930	0	2930
CO2	0	0	0	0	0	0	0	2034	2034	2034	2034	2034	102	102	102	0
H2O	0	4223	0	0	0	0	0	4223	4223	4223	4223	0	0	0	0	0
CH4	0	0	0	0	0	0	0	298	298	298	298	298	298	298	298	0
C2-C4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5-C11	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C12-C24	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C24+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MDEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

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Table E.16 Fuel synthesis and upgrading area stream data for Case CERT-3

	FS01D ISL	FS01F TG	FS01 FTL	FS01 FTP	FS01N APH	FS01 WAX	FS02F TG	FS02 FTL	FS02 FTP	FS03F TG	FS03 FTP	FS04F TG	FS04 FTP	FS05F TG	FS05 FTP	FS05 WAT	FS06 FTP	FS06H YD	FS21S GAS	FS22 GAS
T (^{o}C)	16	15	15	219	16	219	15	15	219	10	15	10	220	10	15	15	220	38	220	220
P (psia)	15	400	400	400	15	400	15	15	400	15	400	15	15	15	15	400	400	400	400	400
Mass Flow tonne/day	725	1203	747	3731	306	312	32	284	3418	780	3418	1236	316	456	316	1468	3331	4	3331	400
C	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	0	64	0	64	0	0	0	0	64	40	64	64	0	24	0	0	64	4	401	0
CO	0	645	0	645	0	0	0	0	645	407	645	645	0	238	0	0	645	0	2930	0
CO2	0	102	0	102	0	0	0	0	102	64	102	102	0	38	0	0	0	0	0	102
H2O	0	0	0	1468	0	0	0	0	1468	0	1468	0	0	0	0	1468	1468	0	0	0
CH4	0	351	0	351	0	0	11	0	351	228	351	362	11	133	11	0	53	0	0	298
C2-C4	0	41	0	41	0	0	21	0	41	40	41	63	21	23	21	0	41	0	0	0
C5-C11	0	0	255	255	306	0	0	51	255	0	255	0	51	0	51	0	255	0	0	0
C12-C24	725	0	491	491	0	0	0	233	491	0	491	0	233	0	233	0	491	0	0	0
C24+	0	0	0	312	0	312	0	0	0	0	0	0	0	0	0	0	312	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MDEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

CERT-3

Table E.17 CO_2 removal area stream data for Case CERT-3

	CR01CO2	CR01MDAL	CR01MDAR	CR02CO2	CR02MDAL	CR02MDAR	CR16SGAS	CR17SGAS
T (°C)	81	38	38	-40	81	81	38	38
P (psia)	15	400	400	2217	15	15	400	400
Mass Flow tonne/day	1933	22026	23959	1933	22026	23959	5663	3731
C	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	401	401
CO	0	0	0	0	0	0	2930	2930
CO2	1933	0	1933	1933	0	1933	2034	102
H2O	0	11010	11010	0	11010	11010	0	0
CH4	0	0	0	0	0	0	298	298
C2-C4	0	0	0	0	0	0	0	0
C5-C11	0	0	0	0	0	0	0	0
C12-C24	0	0	0	0	0	0	0	0
C24+	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0
MDEA	0	11017	11017	0	11017	11017	0	0
ASH	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0

CERT-3

Table E.18 Power generation area stream data for Case CERT-3

	PG05FTG	PG05STM	PG06STM	PG06WAT	PG07STM	PG07WAT	PG08AIR	PG08HYD	PG08STM	PG09FLUE	PG09STM	PG10FLUE	PG11FLUE	PG12FLUE
T (^{o}C)	10	539	295	15	538	105	15	204	290	276	127	1027	868	147
P (psia)	15	2000	400	15	400	2000	15	15	65	15	13	15	15	15
Mass Flow tonne/day	456	3923	3923	3923	3923	3923	3818	9	3923	9042	3923	13325	13325	13325
C	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	24	0	0	0	0	0	0	9	0	0	0	0	0	0
CO	238	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	38	0	0	0	0	0	0	0	0	1452	0	2301	2301	2301
H2O	0	3923	3923	3923	3923	3923	0	0	3923	932	3923	1559	1559	1559
CH4	133	0	0	0	0	0	0	0	0	0	0	0	0	0
C2-C4	23	0	0	0	0	0	0	0	0	0	0	0	0	0
C5-C11	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C12-C24	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C24+	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	889	0	0	321	0	199	199	199
N2	0	0	0	0	0	0	2929	0	0	6338	0	9266	9266	9266
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MDEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0

CERT-3B

Table E.19 Overall plant stream data for Case CERT-3B

	PL00COAL	PL01AIR	PL01HYD	PL01SGAS	PL01STM	PL01WAT	PL02CO2	PL03FLUE	PL03FTG	PL03SLRY	PL05FTG	PL05HYD	PL06AIR	PL06HYD	PL06WAT	PL08AIR	PL08HYD	PL09FLUE	PL09SGAS	PL12FLUE	PL12SGAS	PL16SGAS	PL17SGAS	PL19SGAS	PL20SGAS	PLDISL	PLNAPH
T (°C)	15	15	15	750	15	15	-40	147	10	188	10	38	15	38	15	15	204	276	601	147	650	38	38	220	220	16	16
P (psia)	15	15	15	400	15	15	2217	25	15	400	15	400	15	400	15	15	15	15	400	15	400	400	400	400	400	15	15
Mass Flow tonne/day	3600	6090	416	10393	3600	3600	4950	6533	780	7200	456	416	8262	4	2311	3818	9	9042	10322	13325	10322	5663	3731	400	3331	725	306
C	0	0	0	0	0	0	0	0	0	2479	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	0	0	416	354	0	0	0	0	40	171	24	416	0	4	0	0	9	0	354	0	837	401	401	0	401	0	0
CO	0	0	0	457	0	0	0	0	407	0	238	0	0	0	0	0	0	0	457	0	2930	2930	2930	0	2930	0	0
CO2	0	0	0	2316	0	0	4950	1626	64	0	38	0	0	0	0	0	0	1452	2316	2301	2034	2034	102	102	0	0	0
H2O	0	0	0	5555	3600	3600	0	0	0	3600	0	0	0	0	2311	0	0	932	5583	1559	4223	0	0	0	0	0	0
CH4	0	0	0	1612	0	0	0	0	228	0	133	0	0	0	0	0	0	0	1612	0	298	298	298	298	0	0	0
C2-C4	0	0	0	0	0	0	0	0	40	0	23	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5-C11	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	306
C12-C24	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	725	0
C24+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	1418	0	0	0	0	0	236	0	484	0	0	1924	0	0	889	0	321	0	199	0	0	0	0	0	0	0
N2	0	4671	0	0	0	0	0	4671	0	37	0	0	6338	0	0	2929	0	6338	0	9266	0	0	0	0	0	0	0
NH3	0	0	0	46	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	50	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	53	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	380	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	3600	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

CERT-3B

Table E.20 CO_2 removal area stream data for Case CERT-3B

	CR01 CO2	CR01 MDAL	CR01 MDAR	CR01 MEAL	CR01 MEAR	CR02 CO2	CR02 MDAL	CR02 MDAR	CR02 MEAL	CR02 MEAR	CR03 CO2	CR04 CO2	CR05 CO2	CR07 WAT	CR12 FLUE	CR13 FLUE	CR14 FLUE	CR16 SGAS	CR17 SGAS	CR03 FLUE
T (^{o}C)	81	38	38	38	33	-40	81	81	81	81	81	-40	-40	38	147	38	33	38	38	147
P (psia)	15	400	400	15	15	2217	15	15	15	15	15	2217	2217	15	15	15	15	400	400	25
Mass Flow tonne/day	1933	22026	23959	54272	57303	4950	22026	23959	54286	57303	3018	3018	1933	1546	13325	18313	15282	5663	3731	6533
C	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	401	401	0
CO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2930	2930	0
CO2	1933	0	1933	0	3018	4950	0	1933	0	3018	3018	3018	1933	0	2301	3926	909	2034	102	1626
H2O	0	11010	11010	43373	43386	0	11010	11010	43386	43386	0	0	0	1546	1559	13	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	298	298	0
C2-C4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C5-C11	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C12-C24	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C24+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	199	435	435	0	0	236
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	9266	13938	13938	0	0	4671
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MDEA	0	11017	11017	0	0	0	11017	11017	0	0	0	0	0	0	0	0	0	0	0	0
MEA	0	0	0	10899	10899	0	0	0	10899	10899	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

CERT-3B

Table E.21 Power generation area stream data for Case CERT-3B

	PG05FTG	PG05STM	PG06STM	PG06WAT	PG07STM	PG07WAT	PG08AIR	PG08HYD	PG08STM	PG09FLUE	PG09STM	PG10FLUE	PG11FLUE	PG12FLUE
T (^oC)	10	539	296	15	538	105	15	204	290	276	127	1027	587	147
P (psia)	15	2000	400	15	400	2000	15	15	65	15	13	15	15	15
Mass Flow tonne/day	456	2311	2311	2311	2311	2311	3818	9	2311	9042	2311	13325	13325	13325
C	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	24	0	0	0	0	0	0	9	0	0	0	0	0	0
CO	238	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	38	0	0	0	0	0	0	0	0	1452	0	2301	2301	2301
H2O	0	2311	2311	2311	2311	2311	0	0	2311	932	2311	1559	1559	1559
CH4	133	0	0	0	0	0	0	0	0	0	0	0	0	0
C2-C4	23	0	0	0	0	0	0	0	0	0	0	0	0	0
C5-C11	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C12-C24	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C24+	0	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	889	0	0	321	0	199	199	199
N2	0	0	0	0	0	0	2929	0	0	6338	0	9266	9266	9266
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SULFUR	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SAND	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MDEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MEA	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Coal	0	0	0	0	0	0	0	0	0	0	0	0	0	0

Appendix F. Discounted cash flow

Note: Negative cash flow is shown in the bracket

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Table F.1 Discounted cash flow sheet for construction period and years 1-12 of Case CERT-1

Project lifetime include construction	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Year	2013	2014	2015	2016	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027
Project lifetime	**CY 1**	**CY 2**	**CY 3**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**
Loan	$22.6	$135.6	$67.8												
Equity Expense	($9.7)	($58.1)	($29.1)												
Loan Expense				($25.6)	($25.6)	($25.6)	($25.6)	($25.6)	($25.6)	($25.6)	($25.6)	($25.6)	($25.6)	($25.6)	($25.6)
Variable O&M Cost				($4.0)	($6.3)	($8.3)	($8.5)	($8.8)	($9.0)	($9.3)	($9.6)	($9.9)	($10.2)	($10.5)	($10.8)
Fixed O&M Cost				($17.1)	($17.6)	($18.2)	($18.7)	($19.3)	($19.9)	($20.5)	($21.1)	($21.7)	($22.4)	($23.0)	($23.7)
Coal Cost				($3.3)	($5.2)	($6.9)	($7.1)	($7.3)	($7.5)	($7.7)	($8.0)	($8.2)	($8.4)	($8.7)	($9.0)
TOTAL EXPENSES	($9.7)	($58.1)	($29.1)	($50.0)	($54.7)	($58.9)	($59.9)	($60.9)	($62.0)	($63.1)	($64.2)	($65.4)	($66.6)	($67.8)	($69.0)
Diesel Revenues				$19.4	$30.6	$40.4	$41.6	$42.8	$44.1	$45.4	$46.8	$48.2	$49.6	$51.1	$52.7
Naphtha Revenues				$10.9	$17.1	$22.6	$23.3	$24.0	$24.7	$25.5	$26.2	$27.0	$27.8	$28.7	$29.5
Power Revenues				$2.5	$4.0	$5.2	$5.4	$5.6	$5.7	$5.9	$6.1	$6.3	$6.4	$6.6	$6.8
Sale of FT Liquid				$30.3	$47.7	$63.0	$64.9	$66.8	$68.8	$70.9	$73.0	$75.2	$77.5	$79.8	$82.2
TOTAL REVENUES BEF TAX				$32.8	$51.7	$68.2	$70.3	$72.4	$74.6	$76.8	$79.1	$81.5	$83.9	$86.4	$89.0
Annual Depreciation															
General Plant															
DDB				$65.2	$46.6	$33.3	$23.8	$17.0	$12.1	$8.7					
SL				$32.6	$27.2	$23.3	$20.8	$19.8	$19.8	$19.8					
Remaining Value				$163.0	$116.4	$83.2	$59.4	$42.4	$30.3	$21.6					
Actual				$65.2	$46.6	$33.3	$23.8	$19.8	$19.8	$19.8					
Steam Plant															
DDB				$1.7	$1.6	$1.5	$1.3	$1.2	$1.2	$1.1	$1.0	$0.9	$0.8	$0.8	$0.7
SL				$1.1	$1.1	$1.2	$1.2	$1.3	$1.3	$1.3	$1.3	$1.3	$1.3	$1.3	$1.3
Remaining Value				$21.0	$19.4	$18.0	$16.6	$15.4	$14.2	$13.2	$12.2	$11.3	$10.4	$9.6	$8.9
Actual				$1.7	$1.6	$1.5	$1.3	$1.3	$1.3	$1.3	$1.3	$1.3	$1.3	$1.3	$1.3
Net Revenue				($84.1)	($51.2)	($25.4)	($14.7)	($9.7)	($8.5)	($7.4)	$13.6	$14.8	$16.1	$17.3	$18.7
Losses Forward					($84.1)	($135.3)	($160.7)	($175.4)	($185.1)	($193.6)	($201.0)	($187.4)	($172.6)	($156.6)	($139.2)
Taxable Income				($84.1)	($135.3)	($160.7)	($175.4)	($185.1)	($193.6)	($201.0)	($187.4)	($172.6)	($156.6)	($139.2)	($120.6)
Income Tax				$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
Annual cash income				($17.2)	($3.0)	$9.3	$10.4	$11.5	$12.6	$13.7	$14.9	$16.1	$17.4	$18.7	$20.0
Annual Net Cash Flow	($9.7)	($58.1)	($29.1)	($17.2)	($3.0)	$9.3	$10.4	$11.5	$12.6	$13.7	$14.9	$16.1	$17.4	$18.7	$20.0
Discount factor	$0.9	$0.8	$0.7	$0.6	$0.6	$0.5	$0.5	$0.4	$0.4	$0.3	$0.3	$0.3	$0.2	$0.2	$0.2
Annual Present Value	($8.6)	($46.3)	($20.7)	($10.9)	($1.7)	$4.7	$4.7	$4.6	$4.5	$4.4	$4.3	$4.1	$4.0	$3.8	$3.7
Cumulative Cash Flow	**($8.6)**	**($55.0)**	**($75.7)**	**($86.6)**	**($88.3)**	**($83.6)**	**($78.9)**	**($74.3)**	**($69.7)**	**($65.3)**	**($61.0)**	**($56.9)**	**($52.9)**	**($49.1)**	**($45.5)**

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Table F.2 Discounted cash flow sheet for years 13-30 of Case CERT-1

16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33
2028	2029	2030	2031	2032	2033	2034	2035	2036	2037	2038	2039	2040	2041	2042	2043	2044	2045
13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
($25.6)	($25.6)	($25.6)															
($11.1)	($11.4)	($11.8)	($12.1)	($12.5)	($12.9)	($13.2)	($13.6)	($14.1)	($14.5)	($14.9)	($15.4)	($15.8)	($16.3)	($16.8)	($17.3)	($17.8)	($18.3)
($24.4)	($25.2)	($25.9)	($26.7)	($27.5)	($28.3)	($29.2)	($30.0)	($30.9)	($31.9)	($32.8)	($33.8)	($34.8)	($35.9)	($37.0)	($38.1)	($39.2)	($40.4)
($9.2)	($9.5)	($9.8)	($10.1)	($10.4)	($10.7)	($11.0)	($11.3)	($11.7)	($12.0)	($12.4)	($12.8)	($13.1)	($13.5)	($13.9)	($14.4)	($14.8)	($15.2)
($70.3)	($71.7)	($73.1)	($48.9)	($50.4)	($51.9)	($53.4)	($55.0)	($56.7)	($58.4)	($60.1)	($61.9)	($63.8)	($65.7)	($67.7)	($69.7)	($71.8)	($74.0)
$54.2	$55.9	$57.5	$59.3	$61.0	$62.9	$64.8	$66.7	$68.7	$70.8	$72.9	$75.1	$77.3	$79.6	$82.0	$84.5	$87.0	$89.6
$30.4	$31.3	$32.3	$33.2	$34.2	$35.3	$36.3	$37.4	$38.5	$39.7	$40.9	$42.1	$43.4	$44.7	$46.0	$47.4	$48.8	$50.3
$7.0	$7.3	$7.5	$7.7	$7.9	$8.2	$8.4	$8.7	$8.9	$9.2	$9.5	$9.8	$10.0	$10.3	$10.7	$11.0	$11.3	$11.6
$84.7	$87.2	$89.8	$92.5	$95.3	$98.1	$101.1	$104.1	$107.2	$110.5	$113.8	$117.2	$120.7	$124.3	$128.0	$131.9	$135.8	$139.9
$91.7	$94.4	$97.3	$100.2	$103.2	$106.3	$109.5	$112.8	$116.2	$119.6	$123.2	$126.9	$130.7	$134.7	$138.7	$142.9	$147.1	$151.6
$0.7	$0.6	$0.6	$0.5	$0.5	$0.5	$0.4	$0.4										
$1.3	$1.3	$1.3	$1.3	$1.3	$1.3	$1.3	$1.3										
$8.2	$7.6	$7.0	$6.5	$6.0	$5.6	$5.2	$4.8										
$1.3	$1.3	$1.3	$1.3	$1.3	$1.3	$1.3	$1.3										
$20.0	$21.4	$22.9	$50.0	$51.5	$53.1	$54.8	$56.4	$59.5	$61.3	$63.1	$65.0	$66.9	$69.0	$71.0	$73.2	$75.3	$77.6
($120.6)	($100.5)	($79.1)	($56.2)	($6.2)	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
($100.5)	($79.1)	($56.2)	($6.2)	$45.3	$53.1	$54.8	$56.4	$59.5	$61.3	$63.1	$65.0	$66.9	$69.0	$71.0	$73.2	$75.3	$77.6
$0.0	$0.0	$0.0	$0.0	$17.2	$20.2	$20.8	$21.4	$22.6	$23.3	$24.0	$24.7	$25.4	$26.2	$27.0	$27.8	$28.6	$29.5
$21.3	$22.8	$24.2	$51.3	$35.6	$34.2	$35.3	$36.3	$36.9	$38.0	$39.1	$40.3	$41.5	$42.8	$44.0	$45.4	$46.7	$48.1
$21.3	$22.8	$24.2	$51.3	$35.6	$34.2	$35.3	$36.3	$36.9	$38.0	$39.1	$40.3	$41.5	$42.8	$44.0	$45.4	$46.7	$48.1
$0.2	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
$3.5	$3.3	$3.1	$6.0	$3.7	$3.2	$2.9	$2.7	$2.4	$2.2	$2.1	$1.9	$1.7	$1.6	$1.5	$1.4	$1.2	$1.1
($42.0)	**($38.7)**	**($35.5)**	**($29.6)**	**($25.9)**	**($22.7)**	**($19.8)**	**($17.1)**	**($14.7)**	**($12.4)**	**($10.4)**	**($8.5)**	**($6.8)**	**($5.2)**	**($3.7)**	**($2.3)**	**($1.1)**	**$0.0**

CERT-2

Table F.3 Discounted cash flow sheet for construction period and years 1-12 of Case CERT-2

Project lifetime include construction	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Year	2012	2013	2014	2015	2016	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026
Project lifetime	**CY 1**	**CY 2**	**CY 3**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**
Loan	$95.63	$573.76	$286.88												
Equity Expense	($41.0)	($245.9)	($122.9)												
Loan Expense				($108.3)	($108.3)	($108.3)	($108.3)	($108.3)	($108.3)	($108.3)	($108.3)	($108.3)	($108.3)	($108.3)	($108.3)
Variable O&M Cost	$0.0	$0.0	$0.0	($15.4)	($24.2)	($32.0)	($32.9)	($33.9)	($34.9)	($36.0)	($37.1)	($38.2)	($39.3)	($40.5)	($41.7)
Fixed O&M Cost	$0.0	$0.0	$0.0	($22.7)	($23.4)	($24.1)	($24.8)	($25.5)	($26.3)	($27.1)	($27.9)	($28.7)	($29.6)	($30.5)	($31.4)
Coal Cost	$0.0	$0.0	$0.0	($32.6)	($51.4)	($67.9)	($69.9)	($72.0)	($74.2)	($76.4)	($78.7)	($81.0)	($83.5)	($86.0)	($88.5)
TOTAL EXPENSES	($41.0)	($245.9)	($122.9)	($179.0)	($207.3)	($232.2)	($235.9)	($239.8)	($243.7)	($247.8)	($252.0)	($256.3)	($260.7)	($265.3)	($270.0)
Diesel Revenues	$0.0	$0.0	$0.0	$66.4	$104.6	$138.1	$142.3	$146.5	$150.9	$155.5	$160.1	$164.9	$169.9	$175.0	$180.2
Naphtha Revenues	$0.0	$0.0	$0.0	$37.3	$58.8	$77.7	$80.0	$82.4	$84.9	$87.4	$90.0	$92.7	$95.5	$98.4	$101.3
Power Revenues	$0.0	$0.0	$0.0	$25.3	$39.8	$52.5	$54.1	$55.7	$57.4	$59.1	$60.9	$62.7	$64.6	$66.6	$68.5
Sale of FT Liquid	$0.0	$0.0	$0.0	$103.7	$163.4	$215.8	$222.3	$228.9	$235.8	$242.9	$250.2	$257.7	$265.4	$273.4	$281.6
TOTAL REVENUES BEF TAX	$0.0	$0.0	$0.0	$129.0	$203.2	$268.3	$276.4	$284.7	$293.2	$302.0	$311.1	$320.4	$330.0	$339.9	$350.1
Annual Depreciation															
General Plant															
DDB				$245.6	$175.4	$125.3	$89.5	$63.9	$45.7	$32.6					
SL				$122.8	$102.3	$87.7	$78.3	$74.6	$74.6	$74.6					
Remaining Value				$613.9	$438.5	$313.2	$223.7	$159.8	$114.2	$81.5					
Actual				$245.6	$175.4	$125.3	$89.5	$74.6	$74.6	$74.6					
Steam Plant															
DDB				$9.0	$8.3	$7.7	$7.1	$6.6	$6.1	$5.6	$5.2	$4.8	$4.5	$4.1	$3.8
SL				$6.0	$5.8	$6.2	$6.5	$6.9	$6.9	$6.9	$6.9	$6.9	$6.9	$6.9	$6.9
Remaining Value				$111.0	$102.7	$95.0	$87.9	$81.3	$75.2	$69.5	$64.3	$59.5	$55.0	$50.9	$47.1
Actual				$9.0	$8.3	$7.7	$7.1	$6.9	$6.9	$6.9	$6.9	$6.9	$6.9	$6.9	$6.9
Net Revenue				($304.6)	($187.8)	($96.9)	($56.2)	($36.6)	($32.0)	($27.3)	$52.2	$57.2	$62.4	$67.7	$73.2
Losses Forward					($304.6)	($492.4)	($589.3)	($645.5)	($682.1)	($714.1)	($741.4)	($689.2)	($632.0)	($569.6)	($501.9)
Taxable Income				($304.6)	($492.4)	($589.3)	($645.5)	($682.1)	($714.1)	($741.4)	($689.2)	($632.0)	($569.6)	($501.9)	($428.7)
Income Tax				$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
Annual cash income				($50.0)	($4.1)	$36.1	$40.4	$44.9	$49.5	$54.2	$59.1	$64.1	$69.3	$74.6	$80.1
Annual Net Cash Flow	($41.0)	($245.9)	($122.9)	($50.0)	($4.1)	$36.1	$40.4	$44.9	$49.5	$54.2	$59.1	$64.1	$69.3	$74.6	$80.1
Discount factor	$0.9	$0.8	$0.7	$0.6	$0.6	$0.5	$0.5	$0.4	$0.4	$0.3	$0.3	$0.3	$0.2	$0.2	$0.2
Annual Present Value	($36.6)	($196.0)	($87.5)	($31.8)	($2.3)	$18.3	$18.3	$18.1	$17.9	$17.5	$17.0	$16.5	$15.9	$15.3	$14.6
Cumulative Cash Flow	**($36.6)**	**($232.6)**	**($320.1)**	**($351.9)**	**($354.2)**	**($335.9)**	**($317.6)**	**($299.5)**	**($281.7)**	**($264.2)**	**($247.2)**	**($230.7)**	**($214.9)**	**($199.6)**	**($184.9)**

Table F.4 Discounted cash flow sheet for years 13-30 of Case CERT-2

16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33
2027	2028	2029	2030	2031	2032	2033	2034	2035	2036	2037	2038	2039	2040	2041	2042	2043	2044
13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
($108.3)	($108.3)	($108.3)															
($43.0)	($44.3)	($45.6)	($47.0)	($48.4)	($49.8)	($51.3)	($52.8)	($54.4)	($56.1)	($57.7)	($59.5)	($61.3)	($63.1)	($65.0)	($66.9)	($69.0)	($71.0)
($32.3)	($33.3)	($34.3)	($35.3)	($36.4)	($37.5)	($38.6)	($39.8)	($41.0)	($42.2)	($43.4)	($44.7)	($46.1)	($47.5)	($48.9)	($50.4)	($51.9)	($53.4)
($91.2)	($93.9)	($96.8)	($99.7)	($102.6)	($105.7)	($108.9)	($112.2)	($115.5)	($119.0)	($122.6)	($126.2)	($130.0)	($133.9)	($138.0)	($142.1)	($146.4)	($150.7)
($274.8)	($279.8)	($285.0)	($181.9)	($187.4)	($193.0)	($198.8)	($204.8)	($210.9)	($217.2)	($223.8)	($230.5)	($237.4)	($244.5)	($251.8)	($259.4)	($267.2)	($275.2)
$185.6	$191.2	$197.0	$202.9	$208.9	$215.2	$221.7	$228.3	$235.2	$242.2	$249.5	$257.0	$264.7	$272.6	$280.8	$289.2	$297.9	$306.8
$104.4	$107.5	$110.7	$114.0	$117.5	$121.0	$124.6	$128.4	$132.2	$136.2	$140.3	$144.5	$148.8	$153.3	$157.9	$162.6	$167.5	$172.5
$70.6	$72.7	$74.9	$77.2	$79.5	$81.9	$84.3	$86.8	$89.4	$92.1	$94.9	$97.7	$100.7	$103.7	$106.8	$110.0	$113.3	$116.7
$290.0	$298.7	$307.7	$316.9	$326.4	$336.2	$346.3	$356.7	$367.4	$378.4	$389.7	$401.4	$413.5	$425.9	$438.7	$451.8	$465.4	$479.3
$360.6	$371.4	$382.6	$394.1	$405.9	$418.1	$430.6	$443.5	$456.8	$470.5	$484.6	$499.2	$514.2	$529.6	$545.5	$561.8	$578.7	$596.0
$3.5	$3.3	$3.0	$2.8	$2.6	$2.4	$2.2	$2.0										
$6.9	$6.9	$6.9	$6.9	$6.9	$6.9	$6.9	$6.9										
$43.6	$40.3	$37.3	$34.5	$31.9	$29.5	$27.3	$25.2										
$6.9	$6.9	$6.9	$6.9	$6.9	$6.9	$6.9	$6.9										
$78.8	$84.7	$90.7	$205.2	$211.5	$218.1	$224.8	$231.8	$245.9	$253.3	$260.9	$268.7	$276.8	$285.1	$293.6	$302.4	$311.5	$320.8
($428.7)	($349.9)	($265.2)	($174.5)	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
($349.9)	($265.2)	($174.5)	$30.6	$211.5	$218.1	$224.8	$231.8	$245.9	$253.3	$260.9	$268.7	$276.8	$285.1	$293.6	$302.4	$311.5	$320.8
$0.0	$0.0	$0.0	$11.6	$80.4	$82.9	$85.4	$88.1	$93.4	$96.2	$99.1	$102.1	$105.2	$108.3	$111.6	$114.9	$118.4	$121.9
$85.8	$91.6	$97.6	$200.5	$138.1	$142.2	$146.3	$150.7	$152.5	$157.0	$161.7	$166.6	$171.6	$176.7	$182.0	$187.5	$193.1	$198.9
$85.8	$91.6	$97.6	$200.5	$138.1	$142.2	$146.3	$150.7	$152.5	$157.0	$161.7	$166.6	$171.6	$176.7	$182.0	$187.5	$193.1	$198.9
$0.2	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
$14.0	$13.3	$12.7	$23.3	$14.3	$13.2	$12.1	$11.1	$10.0	$9.2	$8.5	$7.8	$7.2	$6.6	$6.1	$5.6	$5.1	$4.7
($170.9)	**($157.6)**	**($144.9)**	**($121.6)**	**($107.3)**	**($94.2)**	**($82.1)**	**($71.0)**	**($60.9)**	**($51.7)**	**($43.2)**	**($35.4)**	**($28.2)**	**($21.6)**	**($15.5)**	**($9.9)**	**($4.8)**	**($0.0)**

CERT-2B

Table F.5 Discounted cash flow sheet for construction period and years 1-12 of Case CERT-2B

Project lifetime include construction	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Year	2012	2013	2014	2015	2016	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026
Project lifetime	**CY 1**	**CY 2**	**CY 3**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**
Loan	$102.3	$613.9	$306.9												
Equity Expense	($43.8)	($263.1)	($131.5)												
Loan Expense				($115.9)	($115.9)	($115.9)	($115.9)	($115.9)	($115.9)	($115.9)	($115.9)	($115.9)	($115.9)	($115.9)	($115.9)
Variable O&M Cost				($15.9)	($25.1)	($33.1)	($34.1)	($35.1)	($36.2)	($37.3)	($38.4)	($39.6)	($40.7)	($42.0)	($43.2)
Fixed O&M Cost				($23.8)	($24.5)	($25.2)	($26.0)	($26.7)	($27.6)	($28.4)	($29.2)	($30.1)	($31.0)	($31.9)	($32.9)
Coal Cost				($32.6)	($51.4)	($67.9)	($69.9)	($72.0)	($74.2)	($76.4)	($78.7)	($81.0)	($83.5)	($86.0)	($88.5)
TOTAL EXPENSES	($43.8)	($263.1)	($131.5)	($188.2)	($216.9)	($242.1)	($245.9)	($249.8)	($253.8)	($258.0)	($262.2)	($266.6)	($271.1)	($275.8)	($280.6)
Diesel Revenues				$75.2	$118.5	$156.5	$161.2	$166.0	$171.0	$176.1	$181.4	$186.9	$192.5	$198.2	$204.2
Naphtha Revenues				$42.3	$66.6	$88.0	$90.6	$93.3	$96.1	$99.0	$102.0	$105.0	$108.2	$111.4	$114.8
Power Revenues				$17.5	$27.6	$36.4	$37.5	$38.6	$39.8	$41.0	$42.2	$43.5	$44.8	$46.1	$47.5
Sale of FT Liquid				$117.5	$185.1	$244.5	$251.8	$259.4	$267.1	$275.2	$283.4	$291.9	$300.7	$309.7	$319.0
TOTAL REVENUES BEF TAX				$135.0	$212.7	$280.9	$289.3	$298.0	$306.9	$316.1	$325.6	$335.4	$345.4	$355.8	$366.5
Annual Depreciation															
General Plant															
DDB				$263.8	$188.4	$134.6	$96.1	$68.7	$49.0	$35.0					
SL				$131.9	$109.9	$94.2	$84.1	$80.1	$80.1	$80.1					
Remaining Value				$659.5	$471.1	$336.5	$240.3	$171.7	$122.6	$87.6					
Actual				$263.8	$188.4	$134.6	$96.1	$80.1	$80.1	$80.1					
Steam Plant															
DDB				$8.3	$7.6	$7.1	$6.5	$6.0	$5.6	$5.2	$4.8	$4.4	$4.1	$3.8	$3.5
SL				$5.5	$5.4	$5.7	$6.0	$6.4	$6.4	$6.4	$6.4	$6.4	$6.4	$6.4	$6.4
Remaining Value				$101.8	$94.1	$87.1	$80.5	$74.5	$68.9	$63.7	$59.0	$54.5	$50.4	$46.7	$43.2
Actual				$8.3	$7.6	$7.1	$6.5	$6.4	$6.4	$6.4	$6.4	$6.4	$6.4	$6.4	$6.4
Net Revenue				($325.3)	($200.2)	($102.9)	($59.3)	($38.3)	($33.4)	($28.3)	$57.0	$62.4	$67.9	$73.6	$79.5
Losses Forward					($325.3)	($525.5)	($628.4)	($687.7)	($726.0)	($759.4)	($787.7)	($730.7)	($668.3)	($600.4)	($526.7)
Taxable Income				($325.3)	($525.5)	($628.4)	($687.7)	($726.0)	($759.4)	($787.7)	($730.7)	($668.3)	($600.4)	($526.7)	($447.2)
Income Tax				$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
Annual cash income				($53.2)	($4.2)	$38.7	$43.4	$48.2	$53.1	$58.2	$63.4	$68.8	$74.3	$80.0	$85.9
Annual Net Cash Flow	($43.8)	($263.1)	($131.5)	($53.2)	($4.2)	$38.7	$43.4	$48.2	$53.1	$58.2	$63.4	$68.8	$74.3	$80.0	$85.9
Discount factor	$0.9	$0.8	$0.7	$0.6	$0.6	$0.5	$0.5	$0.4	$0.4	$0.3	$0.3	$0.3	$0.2	$0.2	$0.2
Annual Present Value	($39.2)	($209.7)	($93.6)	($33.8)	($2.4)	$19.6	$19.6	$19.5	$19.1	$18.7	$18.2	$17.6	$17.0	$16.4	$15.7
Cumulative Cash Flow	**($39.2)**	**($248.9)**	**($342.5)**	**($376.3)**	**($378.7)**	**($359.1)**	**($339.4)**	**($320.0)**	**($300.9)**	**($282.1)**	**($263.9)**	**($246.3)**	**($229.2)**	**($212.9)**	**($197.2)**

CERT-2B

Table F.6 Discounted cash flow sheet for years 13-30 of Case CERT-2B

16	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**	**27**	**28**	**29**	**30**	**31**	**32**	**33**
2027	2028	2029	2030	2031	2032	2033	2034	2035	2036	2037	2038	2039	2040	2041	2042	2043	2044
13	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**	**27**	**28**	**29**	**30**
($115.9)	($115.9)	($115.9)															
($44.5)	($45.9)	($47.2)	($48.7)	($50.1)	($51.6)	($53.2)	($54.8)	($56.4)	($58.1)	($59.8)	($61.6)	($63.5)	($65.4)	($67.4)	($69.4)	($71.5)	($73.6)
($33.9)	($34.9)	($35.9)	($37.0)	($38.1)	($39.3)	($40.5)	($41.7)	($42.9)	($44.2)	($45.5)	($46.9)	($48.3)	($49.8)	($51.3)	($52.8)	($54.4)	($56.0)
($91.2)	($93.9)	($96.8)	($99.7)	($102.6)	($105.7)	($108.9)	($112.2)	($115.5)	($119.0)	($122.6)	($126.2)	($130.0)	($133.9)	($138.0)	($142.1)	($146.4)	($150.7)
($285.5)	($290.6)	($295.8)	($185.3)	($190.9)	($196.6)	($202.5)	($208.6)	($214.9)	($221.3)	($227.9)	($234.8)	($241.8)	($249.1)	($256.6)	($264.3)	($272.2)	($280.3)
$210.3	$216.6	$223.1	$229.8	$236.7	$243.8	$251.1	$258.7	$266.4	$274.4	$282.6	$291.1	$299.9	$308.9	$318.1	$327.7	$337.5	$347.6
$118.2	$121.8	$125.4	$129.2	$133.1	$137.1	$141.2	$145.4	$149.8	$154.3	$158.9	$163.7	$168.6	$173.6	$178.8	$184.2	$189.7	$195.4
$48.9	$50.4	$51.9	$53.4	$55.0	$56.7	$58.4	$60.1	$61.9	$63.8	$65.7	$67.7	$69.7	$71.8	$74.0	$76.2	$78.5	$80.8
$328.5	$338.4	$348.6	$359.0	$369.8	$380.9	$392.3	$404.1	$416.2	$428.7	$441.5	$454.8	$468.4	$482.5	$497.0	$511.9	$527.2	$543.0
$377.4	$388.8	$400.4	$412.4	$424.8	$437.6	$450.7	$464.2	$478.1	$492.5	$507.3	$522.5	$538.2	$554.3	$570.9	$588.1	$605.7	$623.9

$3.2	$3.0	$2.8	$2.6	$2.4	$2.2	$2.0	$1.9										
$6.4	$6.4	$6.4	$6.4	$6.4	$6.4	$6.4	$6.4										
$39.9	$36.9	$34.2	$31.6	$29.2	$27.0	$25.0	$23.1										
$6.4	$6.4	$6.4	$6.4	$6.4	$6.4	$6.4	$6.4										
$85.6	$91.8	$98.2	$220.7	$227.6	$234.6	$241.8	$249.3	$263.3	$271.2	$279.3	$287.7	$296.3	$305.2	$314.4	$323.8	$333.5	$343.5
($447.2)	($361.7)	($269.9)	($171.6)	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
($361.7)	($269.9)	($171.6)	$49.1	$227.6	$234.6	$241.8	$249.3	$263.3	$271.2	$279.3	$287.7	$296.3	$305.2	$314.4	$323.8	$333.5	$343.5
$0.0	$0.0	$0.0	$18.7	$86.5	$89.1	$91.9	$94.7	$100.0	$103.0	$106.1	$109.3	$112.6	$116.0	$119.5	$123.0	$126.7	$130.5
$91.9	$98.2	$104.6	$208.4	$147.4	$151.8	$156.3	$160.9	$163.2	$168.1	$173.2	$178.4	$183.7	$189.2	$194.9	$200.8	$206.8	$213.0
$91.9	$98.2	$104.6	$208.4	$147.4	$151.8	$156.3	$160.9	$163.2	$168.1	$173.2	$178.4	$183.7	$189.2	$194.9	$200.8	$206.8	$213.0
$0.2	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
$15.0	$14.3	$13.6	$24.2	$15.3	$14.1	$12.9	$11.9	$10.8	$9.9	$9.1	$8.4	$7.7	$7.1	$6.5	$6.0	$5.5	$5.1
($182.2)	**($167.9)**	**($154.3)**	**($130.1)**	**($114.8)**	**($100.8)**	**($87.8)**	**($76.0)**	**($65.2)**	**($55.3)**	**($46.2)**	**($37.9)**	**($30.2)**	**($23.1)**	**($16.6)**	**($10.6)**	**($5.1)**	**($0.0)**

CERT-3

Table F.7 Discounted cash flow sheet for construction period and years 1-12 of Case CERT-3

Project lifetime include construction	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Year	2013	2014	2015	2016	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027
Project lifetime	CY 1	CY 2	CY 3	1	2	3	4	5	6	7	8	9	10	11	12
Loan	$91.1	$546.6	$273.3												
Equity Expense	($39.0)	($234.3)	($117.1)												
Loan Expense				($103.2)	($103.2)	($103.2)	($103.2)	($103.2)	($103.2)	($103.2)	($103.2)	($103.2)	($103.2)	($103.2)	($103.2)
Variable O&M Cost				($16.5)	($26.0)	($34.3)	($35.3)	($36.4)	($37.5)	($38.6)	($39.8)	($40.9)	($42.2)	($43.4)	($44.7)
Fixed O&M Cost				($23.8)	($24.5)	($25.2)	($26.0)	($26.7)	($27.6)	($28.4)	($29.2)	($30.1)	($31.0)	($31.9)	($32.9)
Coal Cost				($32.6)	($51.4)	($67.9)	($69.9)	($72.0)	($74.2)	($76.4)	($78.7)	($81.0)	($83.5)	($86.0)	($88.5)
TOTAL EXPENSES	($39.0)	($234.3)	($117.1)	($176.1)	($205.0)	($230.6)	($234.4)	($238.3)	($242.4)	($246.6)	($250.9)	($255.3)	($259.9)	($264.6)	($269.4)
Diesel Revenues				$73.4	$115.7	$152.8	$157.3	$162.1	$166.9	$171.9	$177.1	$182.4	$187.9	$193.5	$199.3
Naphtha Revenues				$41.3	$65.1	$86.0	$88.5	$91.2	$93.9	$96.7	$99.6	$102.6	$105.7	$108.9	$112.2
Power Revenues				$12.7	$20.0	$26.4	$27.2	$28.0	$28.9	$29.8	$30.6	$31.6	$32.5	$33.5	$34.5
Sale of FT Liquid				$114.8	$180.8	$238.7	$245.9	$253.3	$260.9	$268.7	$276.7	$285.0	$293.6	$302.4	$311.5
TOTAL REVENUES BEF TAX				$127.5	$200.8	$265.1	$273.1	$281.3	$289.7	$298.4	$307.4	$316.6	$326.1	$335.9	$346.0
Annual Depreciation															
General Plant															
DDB				$241.0	$172.2	$123.0	$87.8	$62.7	$44.8	$32.0					
SL				$120.5	$100.4	$86.1	$76.9	$73.2	$73.2	$73.2					
Remaining Value				$602.6	$430.4	$307.4	$219.6	$156.9	$112.0	$80.0					
Actual				$241.0	$172.2	$123.0	$87.8	$73.2	$73.2	$73.2					
Steam Plant															
DDB				$6.5	$6.0	$5.5	$5.1	$4.7	$4.4	$4.0	$3.7	$3.5	$3.2	$3.0	$2.7
SL				$4.3	$4.2	$4.4	$4.7	$5.0	$5.0	$5.0	$5.0	$5.0	$5.0	$5.0	$5.0
Remaining Value				$79.6	$73.6	$68.1	$63.0	$58.2	$53.9	$49.8	$46.1	$42.6	$39.4	$36.5	$33.7
Actual				$6.5	$6.0	$5.5	$5.1	$5.0	$5.0	$5.0	$5.0	$5.0	$5.0	$5.0	$5.0
Net Revenue				($296.1)	($182.4)	($93.9)	($54.2)	($35.2)	($30.8)	($26.3)	$51.5	$56.3	$61.3	$66.4	$71.6
Losses Forward					($296.1)	($478.5)	($572.4)	($626.6)	($661.9)	($692.7)	($719.0)	($667.4)	($611.1)	($549.8)	($483.5)
Taxable Income				($296.1)	($478.5)	($572.4)	($626.6)	($661.9)	($692.7)	($719.0)	($667.4)	($611.1)	($549.8)	($483.5)	($411.9)
Income Tax				$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
Annual cash income				($48.6)	($4.3)	$34.6	$38.7	$43.0	$47.3	$51.9	$56.5	$61.3	$66.2	$71.3	$76.6
Annual Net Cash Flow	($39.0)	($234.3)	($117.1)	($48.6)	($4.3)	$34.6	$38.7	$43.0	$47.3	$51.9	$56.5	$61.3	$66.2	$71.3	$76.6
Discount factor	$0.9	$0.8	$0.7	$0.6	$0.6	$0.5	$0.5	$0.4	$0.4	$0.3	$0.3	$0.3	$0.2	$0.2	$0.2
Annual Present Value	($34.9)	($186.8)	($83.4)	($30.9)	($2.4)	$17.5	$17.5	$17.4	$17.1	$16.7	$16.2	$15.7	$15.2	$14.6	$14.0
Cumulative Cash Flow	**($34.9)**	**($221.6)**	**($305.0)**	**($335.9)**	**($338.3)**	**($320.8)**	**($303.3)**	**($285.9)**	**($268.9)**	**($252.2)**	**($235.9)**	**($220.2)**	**($205.0)**	**($190.4)**	**($176.4)**

CERT-3

Table F.8 Discounted cash flow sheet for years 13-30 of Case CERT-3

16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33
2028	2029	2030	2031	2032	2033	2034	2035	2036	2037	2038	2039	2040	2041	2042	2043	2044	2045
13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
($103.2)	($103.2)	($103.2)															
($46.1)	($47.5)	($48.9)	($50.4)	($51.9)	($53.4)	($55.0)	($56.7)	($58.4)	($60.1)	($61.9)	($63.8)	($65.7)	($67.7)	($69.7)	($71.8)	($74.0)	($76.2)
($33.9)	($34.9)	($35.9)	($37.0)	($38.1)	($39.3)	($40.5)	($41.7)	($42.9)	($44.2)	($45.5)	($46.9)	($48.3)	($49.8)	($51.3)	($52.8)	($54.4)	($56.0)
($91.2)	($93.9)	($96.8)	($99.7)	($102.6)	($105.7)	($108.9)	($112.2)	($115.5)	($119.0)	($122.6)	($126.2)	($130.0)	($133.9)	($138.0)	($142.1)	($146.4)	($150.7)
($274.4)	($279.5)	($284.8)	($187.0)	($192.7)	($198.4)	($204.4)	($210.5)	($216.8)	($223.3)	($230.0)	($236.9)	($244.1)	($251.4)	($258.9)	($266.7)	($274.7)	($282.9)
$205.3	$211.5	$217.8	$224.3	$231.1	$238.0	$245.1	$252.5	$260.1	$267.9	$275.9	$284.2	$292.7	$301.5	$310.5	$319.8	$329.4	$339.3
$115.5	$119.0	$122.6	$126.2	$130.0	$133.9	$137.9	$142.1	$146.3	$150.7	$155.2	$159.9	$164.7	$169.6	$174.7	$180.0	$185.4	$190.9
$35.5	$36.6	$37.7	$38.8	$40.0	$41.2	$42.4	$43.7	$45.0	$46.4	$47.7	$49.2	$50.7	$52.2	$53.7	$55.3	$57.0	$58.7
$320.8	$330.4	$340.4	$350.6	$361.1	$371.9	$383.1	$394.6	$406.4	$418.6	$431.1	$444.1	$457.4	$471.1	$485.3	$499.8	$514.8	$530.3
$356.3	$367.0	$378.0	$389.4	$401.1	$413.1	$425.5	$438.3	$451.4	$464.9	$478.9	$493.3	$508.1	$523.3	$539.0	$555.2	$571.8	$589.0
$2.5	$2.3	$2.2	$2.0	$1.9	$1.7	$1.6	$1.5										
$5.0	$5.0	$5.0	$5.0	$5.0	$5.0	$5.0	$5.0										
$31.2	$28.9	$26.7	$24.7	$22.9	$21.1	$19.6	$18.1										
$5.0	$5.0	$5.0	$5.0	$5.0	$5.0	$5.0	$5.0										
$77.0	$82.5	$88.3	$197.4	$203.4	$209.7	$216.1	$222.8	$234.6	$241.6	$248.8	$256.3	$264.0	$271.9	$280.1	$288.5	$297.1	$306.1
($411.9)	($334.9)	($252.4)	($164.1)	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
($334.9)	($252.4)	($164.1)	$33.3	$203.4	$209.7	$216.1	$222.8	$234.6	$241.6	$248.8	$256.3	$264.0	$271.9	$280.1	$288.5	$297.1	$306.1
$0.0	$0.0	$0.0	$12.6	$77.3	$79.7	$82.1	$84.6	$89.1	$91.8	$94.6	$97.4	$100.3	$103.3	$106.4	$109.6	$112.9	$116.3
$82.0	$87.5	$93.2	$189.7	$131.1	$135.0	$139.0	$143.1	$145.4	$149.8	$154.3	$158.9	$163.7	$168.6	$173.7	$178.9	$184.2	$189.8
$82.0	$87.5	$93.2	$189.7	$131.1	$135.0	$139.0	$143.1	$145.4	$149.8	$154.3	$158.9	$163.7	$168.6	$173.7	$178.9	$184.2	$189.8
$0.2	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
$13.4	$12.7	$12.1	$22.0	$13.6	$12.5	$11.5	$10.6	$9.6	$8.8	$8.1	$7.5	$6.9	$6.3	$5.8	$5.3	$4.9	$4.5
($163.0)	**($150.3)**	**($138.2)**	**($116.2)**	**($102.6)**	**($90.1)**	**($78.6)**	**($68.0)**	**($58.4)**	**($49.6)**	**($41.5)**	**($34.1)**	**($27.2)**	**($20.9)**	**($15.1)**	**($9.8)**	**($4.9)**	**($0.4)**

CERT-3B

Table F.9 Discounted cash flow sheet for construction period and years 1-12 of Case CERT-3B

Project lifetime include construction	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**
Year	2013	2014	2015	2016	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027
Project lifetime	**CY 1**	**CY 2**	**CY 3**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**
Loan	$99.0	$594.1	$297.1												
Equity Expense	($42.4)	($254.6)	($127.3)												
Loan Expense				($112.2)	($112.2)	($112.2)	($112.2)	($112.2)	($112.2)	($112.2)	($112.2)	($112.2)	($112.2)	($112.2)	($112.2)
Variable O&M Cost				($17.6)	($27.7)	($36.6)	($37.7)	($38.8)	($40.0)	($41.2)	($42.4)	($43.7)	($45.0)	($46.4)	($47.8)
Fixed O&M Cost				($23.8)	($24.5)	($25.2)	($26.0)	($26.7)	($27.6)	($28.4)	($29.2)	($30.1)	($31.0)	($31.9)	($32.9)
Coal Cost				($32.6)	($51.4)	($67.9)	($69.9)	($72.0)	($74.2)	($76.4)	($78.7)	($81.0)	($83.5)	($86.0)	($88.5)
TOTAL EXPENSES	($42.4)	($254.6)	($127.3)	($186.2)	($215.8)	($241.9)	($245.8)	($249.8)	($253.9)	($258.1)	($262.5)	($267.0)	($271.7)	($276.5)	($281.4)
Diesel Revenues				$84.2	$132.7	$175.2	$180.5	$185.9	$191.5	$197.2	$203.2	$209.2	$215.5	$222.0	$228.7
Naphtha Revenues				$47.4	$74.7	$98.6	$101.6	$104.6	$107.8	$111.0	$114.3	$117.7	$121.3	$124.9	$128.7
Power Revenues				$2.7	$4.3	$5.6	$5.8	$6.0	$6.2	$6.4	$6.5	$6.7	$6.9	$7.2	$7.4
Sale of FT Liquid				$131.6	$207.4	$273.9	$282.1	$290.5	$299.2	$308.2	$317.5	$327.0	$336.8	$346.9	$357.3
TOTAL REVENUES BEF TAX				$134.4	$211.7	$279.5	$287.9	$296.5	$305.4	$314.6	$324.0	$333.7	$343.7	$354.1	$364.7
Annual Depreciation															
General Plant															
DDB				$265.0	$189.3	$135.2	$96.6	$69.0	$49.3	$35.2					
SL				$132.5	$110.4	$94.7	$84.5	$80.5	$80.5	$80.5					
Remaining Value				$662.6	$473.3	$338.0	$241.5	$172.5	$123.2	$88.0					
Actual				$265.0	$189.3	$135.2	$96.6	$80.5	$80.5	$80.5					
Steam Plant															
DDB				$5.2	$4.8	$4.4	$4.1	$3.8	$3.5	$3.2	$3.0	$2.8	$2.6	$2.4	$2.2
SL				$3.5	$3.4	$3.5	$3.8	$4.0	$4.0	$4.0	$4.0	$4.0	$4.0	$4.0	$4.0
Remaining Value				$63.8	$59.0	$54.6	$50.5	$46.7	$43.2	$40.0	$37.0	$34.2	$31.6	$29.3	$27.1
Actual				$5.2	$4.8	$4.4	$4.1	$4.0	$4.0	$4.0	$4.0	$4.0	$4.0	$4.0	$4.0
Net Revenue				($322.0)	($198.2)	($102.0)	($58.6)	($37.7)	($33.0)	($28.0)	$57.5	$62.7	$68.1	$73.6	$79.3
Losses Forward					($322.0)	($520.2)	($622.2)	($680.8)	($718.5)	($751.5)	($779.5)	($722.0)	($659.3)	($591.2)	($517.6)
Taxable Income				($322.0)	($520.2)	($622.2)	($680.8)	($718.5)	($751.5)	($779.5)	($722.0)	($659.3)	($591.2)	($517.6)	($438.3)
Income Tax				$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
Annual cash income				($51.8)	($4.1)	$37.6	$42.1	$46.8	$51.5	$56.4	$61.5	$66.7	$72.1	$77.6	$83.3
Annual Net Cash Flow	($42.4)	($254.6)	($127.3)	($51.8)	($4.1)	$37.6	$42.1	$46.8	$51.5	$56.4	$61.5	$66.7	$72.1	$77.6	$83.3
Discount factor	$0.9	$0.8	$0.7	$0.6	$0.6	$0.5	$0.5	$0.4	$0.4	$0.3	$0.3	$0.3	$0.2	$0.2	$0.2
Annual Present Value	($37.9)	($203.0)	($90.6)	($32.9)	($2.3)	$19.1	$19.1	$18.9	$18.6	$18.2	$17.7	$17.1	$16.5	$15.9	$15.2
Cumulative Cash Flow	**($37.9)**	**($240.9)**	**($331.5)**	**($364.4)**	**($366.8)**	**($347.7)**	**($328.6)**	**($309.8)**	**($291.2)**	**($273.0)**	**($255.3)**	**($238.2)**	**($221.7)**	**($205.8)**	**($190.6)**

CERT-3B

Table F.10 Discounted cash flow sheet for years 13-30 of Case CERT-3B

16	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**	**27**	**28**	**29**	**30**	**31**	**32**	**33**
2028	2029	2030	2031	2032	2033	2034	2035	2036	2037	2038	2039	2040	2041	2042	2043	2044	2045
13	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**	**27**	**28**	**29**	**30**
($112.2)	($112.2)	($112.2)															
($49.2)	($50.7)	($52.2)	($53.8)	($55.4)	($57.0)	($58.7)	($60.5)	($62.3)	($64.2)	($66.1)	($68.1)	($70.1)	($72.3)	($74.4)	($76.7)	($79.0)	($81.3)
($33.9)	($34.9)	($35.9)	($37.0)	($38.1)	($39.3)	($40.5)	($41.7)	($42.9)	($44.2)	($45.5)	($46.9)	($48.3)	($49.8)	($51.3)	($52.8)	($54.4)	($56.0)
($91.2)	($93.9)	($96.8)	($99.7)	($102.6)	($105.7)	($108.9)	($112.2)	($115.5)	($119.0)	($122.6)	($126.2)	($130.0)	($133.9)	($138.0)	($142.1)	($146.4)	($150.7)
($286.5)	($291.7)	($297.1)	($190.4)	($196.2)	($202.0)	($208.1)	($214.4)	($220.8)	($227.4)	($234.2)	($241.3)	($248.5)	($255.9)	($263.6)	($271.5)	($279.7)	($288.1)
$235.5	$242.6	$249.9	$257.3	$265.1	$273.0	$281.2	$289.6	$298.3	$307.3	$316.5	$326.0	$335.8	$345.9	$356.2	$366.9	$377.9	$389.3
$132.5	$136.5	$140.6	$144.8	$149.2	$153.6	$158.2	$163.0	$167.9	$172.9	$178.1	$183.4	$188.9	$194.6	$200.5	$206.5	$212.7	$219.0
$7.6	$7.8	$8.1	$8.3	$8.5	$8.8	$9.1	$9.3	$9.6	$9.9	$10.2	$10.5	$10.8	$11.1	$11.5	$11.8	$12.2	$12.5
$368.0	$379.1	$390.4	$402.2	$414.2	$426.7	$439.5	$452.6	$466.2	$480.2	$494.6	$509.4	$524.7	$540.5	$556.7	$573.4	$590.6	$608.3
$375.6	$386.9	$398.5	$410.5	$422.8	$435.4	$448.5	$462.0	$475.8	$490.1	$504.8	$519.9	$535.5	$551.6	$568.2	$585.2	$602.8	$620.8
$2.0	$1.9	$1.7	$1.6	$1.5	$1.4	$1.3	$1.2										
$4.0	$4.0	$4.0	$4.0	$4.0	$4.0	$4.0	$4.0										
$25.0	$23.2	$21.4	$19.8	$18.3	$17.0	$15.7	$14.5										
$4.0	$4.0	$4.0	$4.0	$4.0	$4.0	$4.0	$4.0										
$85.2	$91.2	$97.4	$216.0	$222.6	$229.4	$236.4	$243.6	$255.0	$262.7	$270.6	$278.7	$287.1	$295.7	$304.5	$313.7	$323.1	$332.8
($438.3)	($353.1)	($261.9)	($164.5)	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
($353.1)	($261.9)	($164.5)	$51.5	$222.6	$229.4	$236.4	$243.6	$255.0	$262.7	$270.6	$278.7	$287.1	$295.7	$304.5	$313.7	$323.1	$332.8
$0.0	$0.0	$0.0	$19.6	$84.6	$87.2	$89.8	$92.6	$96.9	$99.8	$102.8	$105.9	$109.1	$112.4	$115.7	$119.2	$122.8	$126.5
$89.2	$95.2	$101.4	$200.4	$142.0	$146.2	$150.6	$155.0	$158.1	$162.9	$167.8	$172.8	$178.0	$183.3	$188.8	$194.5	$200.3	$206.3
$89.2	$95.2	$101.4	$200.4	$142.0	$146.2	$150.6	$155.0	$158.1	$162.9	$167.8	$172.8	$178.0	$183.3	$188.8	$194.5	$200.3	$206.3
$0.2	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
$14.5	$13.9	$13.2	$23.3	$14.7	$13.5	$12.4	$11.4	$10.4	$9.6	$8.8	$8.1	$7.5	$6.9	$6.3	$5.8	$5.3	$4.9
($176.0)	**($162.2)**	**($149.0)**	**($125.7)**	**($111.0)**	**($97.5)**	**($85.0)**	**($73.6)**	**($63.2)**	**($53.6)**	**($44.8)**	**($36.7)**	**($29.2)**	**($22.4)**	**($16.1)**	**($10.3)**	**($4.9)**	**($0.0)**

CERT-5

Table F.11 Discounted cash flow sheet for construction period and years 1-12 of Case CERT-5

Project lifetime include construction	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Year	2013	2014	2015	2016	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027
Project lifetime	CY 1	CY 2	CY 3	1	2	3	4	5	6	7	8	9	10	11	12
Loan	$35.5	$212.8	$106.4												
Equity Expense	($29.0)	($174.1)	($87.1)												
Loan Expense				($40.2)	($40.2)	($40.2)	($40.2)	($40.2)	($40.2)	($40.2)	($40.2)	($40.2)	($40.2)	($40.2)	($40.2)
Variable O&M Cost				($8.2)	($12.9)	($17.0)	($17.5)	($18.0)	($18.6)	($19.1)	($19.7)	($20.3)	($20.9)	($21.5)	($22.2)
Fixed O&M Cost				($19.2)	($19.8)	($20.4)	($21.0)	($21.6)	($22.3)	($23.0)	($23.6)	($24.4)	($25.1)	($25.8)	($26.6)
Biomass Cost				$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
TOTAL EXPENSES	($29.0)	($174.1)	($87.1)	($67.6)	($72.9)	($77.6)	($78.7)	($79.8)	($81.0)	($82.3)	($83.5)	($84.8)	($86.2)	($87.5)	($89.0)
Diesel Revenues				$30.2	$47.6	$62.8	$64.7	$66.7	$68.7	$70.7	$72.9	$75.0	$77.3	$79.6	$82.0
Naphtha Revenues				$16.6	$26.2	$34.6	$35.6	$36.7	$37.8	$38.9	$40.1	$41.3	$42.5	$43.8	$45.1
Power Revenues				$6.5	$10.3	$13.6	$14.0	$14.4	$14.8	$15.3	$15.7	$16.2	$16.7	$17.2	$17.7
Sale of FT Liquid				$46.8	$73.8	$97.4	$100.3	$103.3	$106.4	$109.6	$112.9	$116.3	$119.8	$123.4	$127.1
TOTAL REVENUES BEF TAX				$53.3	$84.0	$111.0	$114.3	$117.7	$121.3	$124.9	$128.6	$132.5	$136.5	$140.6	$144.8
Annual Depreciation															
General Plant															
DDB				$139.1	$99.3	$70.9	$50.7	$36.2	$25.9	$18.5					
SL				$69.5	$57.9	$49.7	$44.3	$42.2	$42.2	$42.2					
Remaining Value				$347.7	$248.3	$177.4	$126.7	$90.5	$64.6	$46.2					
Actual				$139.1	$99.3	$70.9	$50.7	$42.2	$42.2	$42.2					
Steam Plant															
DDB				$7.5	$6.9	$6.4	$5.9	$5.5	$5.1	$4.7	$4.3	$4.0	$3.7	$3.4	$3.2
SL				$5.0	$4.9	$5.1	$5.4	$5.8	$5.8	$5.8	$5.8	$5.8	$5.8	$5.8	$5.8
Remaining Value				$92.2	$85.3	$78.9	$73.0	$67.5	$62.4	$57.8	$53.4	$49.4	$45.7	$42.3	$39.1
Actual				$7.5	$6.9	$6.4	$5.9	$5.8	$5.8	$5.8	$5.8	$5.8	$5.8	$5.8	$5.8
Net Revenue				($160.8)	($95.1)	($43.9)	($21.0)	($10.1)	($7.8)	($5.4)	$39.4	$41.9	$44.6	$47.3	$50.1
Losses Forward					($160.8)	($255.8)	($299.8)	($320.8)	($330.9)	($338.7)	($344.0)	($304.7)	($262.7)	($218.2)	($170.9)
Taxable Income				($160.8)	($255.8)	($299.8)	($320.8)	($330.9)	($338.7)	($344.0)	($304.7)	($262.7)	($218.2)	($170.9)	($120.9)
Income Tax				$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
Annual cash income				($14.2)	$11.2	$33.4	$35.6	$37.9	$40.2	$42.6	$45.1	$47.7	$50.3	$53.0	$55.8
Annual Net Cash Flow	($29.0)	($174.1)	($87.1)	($14.2)	$11.2	$33.4	$35.6	$37.9	$40.2	$42.6	$45.1	$47.7	$50.3	$53.0	$55.8
Discount factor	$0.9	$0.8	$0.7	$0.6	$0.6	$0.5	$0.5	$0.4	$0.4	$0.3	$0.3	$0.3	$0.2	$0.2	$0.2
Annual Present Value	($25.9)	($138.8)	($62.0)	($9.0)	$6.3	$16.9	$16.1	$15.3	$14.5	$13.7	$13.0	$12.2	$11.5	$10.9	$10.2
Cumulative Cash Flow	**($25.9)**	**($164.7)**	**($226.7)**	**($235.7)**	**($229.4)**	**($212.5)**	**($196.4)**	**($181.1)**	**($166.6)**	**($152.8)**	**($139.9)**	**($127.6)**	**($116.1)**	**($105.2)**	**($95.0)**

CERT-5

Table F.12 Discounted cash flow sheet for years 13-30 of Case CERT-5

16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33
2028	2029	2030	2031	2032	2033	2034	2035	2036	2037	2038	2039	2040	2041	2042	2043	2044	2045
13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
($40.2)	($40.2)	($40.2)															
($22.8)	($23.5)	($24.2)	($25.0)	($25.7)	($26.5)	($27.3)	($28.1)	($28.9)	($29.8)	($30.7)	($31.6)	($32.6)	($33.5)	($34.5)	($35.6)	($36.7)	($37.8)
($27.4)	($28.2)	($29.1)	($29.9)	($30.8)	($31.8)	($32.7)	($33.7)	($34.7)	($35.8)	($36.8)	($37.9)	($39.1)	($40.2)	($41.5)	($42.7)	($44.0)	($45.3)
$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
($90.4)	($91.9)	($93.5)	($54.9)	($56.6)	($58.3)	($60.0)	($61.8)	($63.7)	($65.6)	($67.5)	($69.6)	($71.6)	($73.8)	($76.0)	($78.3)	($80.6)	($83.1)
$84.5	$87.0	$89.6	$92.3	$95.1	$97.9	$100.9	$103.9	$107.0	$110.2	$113.5	$116.9	$120.4	$124.0	$127.8	$131.6	$135.5	$139.6
$46.4	$47.8	$49.3	$50.7	$52.3	$53.8	$55.4	$57.1	$58.8	$60.6	$62.4	$64.3	$66.2	$68.2	$70.2	$72.3	$74.5	$76.7
$18.2	$18.8	$19.4	$19.9	$20.5	$21.1	$21.8	$22.4	$23.1	$23.8	$24.5	$25.2	$26.0	$26.8	$27.6	$28.4	$29.3	$30.1
$130.9	$134.8	$138.9	$143.0	$147.3	$151.7	$156.3	$161.0	$165.8	$170.8	$175.9	$181.2	$186.6	$192.2	$198.0	$203.9	$210.1	$216.4
$149.1	$153.6	$158.2	$163.0	$167.9	$172.9	$178.1	$183.4	$188.9	$194.6	$200.4	$206.4	$212.6	$219.0	$225.6	$232.4	$239.3	$246.5
$2.9	$2.7	$2.5	$2.3	$2.1	$2.0	$1.8	$1.7										
$5.8	$5.8	$5.8	$5.8	$5.8	$5.8	$5.8	$5.8										
$36.2	$33.5	$31.0	$28.6	$26.5	$24.5	$22.7	$21.0										
$5.8	$5.8	$5.8	$5.8	$5.8	$5.8	$5.8	$5.8										
$52.9	$55.9	$59.0	$102.3	$105.5	$108.9	$112.3	$115.9	$125.3	$129.0	$132.9	$136.9	$141.0	$145.2	$149.6	$154.1	$158.7	$163.4
($120.9)	($67.9)	($12.0)	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
($67.9)	($12.0)	$47.0	$102.3	$105.5	$108.9	$112.3	$115.9	$125.3	$129.0	$132.9	$136.9	$141.0	$145.2	$149.6	$154.1	$158.7	$163.4
$0.0	$0.0	$17.8	$38.9	$40.1	$41.4	$42.7	$44.0	$47.6	$49.0	$50.5	$52.0	$53.6	$55.2	$56.8	$58.5	$60.3	$62.1
$58.7	$61.7	$46.9	$69.2	$71.2	$73.3	$75.4	$77.6	$77.7	$80.0	$82.4	$84.9	$87.4	$90.0	$92.7	$95.5	$98.4	$101.3
$58.7	$61.7	$46.9	$69.2	$71.2	$73.3	$75.4	$77.6	$77.7	$80.0	$82.4	$84.9	$87.4	$90.0	$92.7	$95.5	$98.4	$101.3
$0.2	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
$9.6	$9.0	$6.1	$8.0	$7.4	$6.8	$6.2	$5.7	$5.1	$4.7	$4.3	$4.0	$3.7	$3.4	$3.1	$2.8	$2.6	$2.4
($85.5)	**($76.5)**	**($70.4)**	**($62.4)**	**($55.0)**	**($48.2)**	**($42.0)**	**($36.2)**	**($31.1)**	**($26.4)**	**($22.1)**	**($18.1)**	**($14.5)**	**($11.1)**	**($8.0)**	**($5.1)**	**($2.5)**	**($0.1)**

CERT-7

Table F.13 Discounted cash flow sheet for construction period and years 1-12 of Case CERT-7

Project lifetime include construction	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Year	2013	2014	2015	2016	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027
Project lifetime	CY 1	CY 2	CY 3	1	2	3	4	5	6	7	8	9	10	11	12
Loan	$28.5	$170.8	$85.4												
Equity Expense	($23.3)	($139.7)	($69.9)												
Loan Expense				($32.2)	($32.2)	($32.2)	($32.2)	($32.2)	($32.2)	($32.2)	($32.2)	($32.2)	($32.2)	($32.2)	($32.2)
Variable O&M Cost				($5.2)	($8.3)	($10.9)	($11.2)	($11.6)	($11.9)	($12.3)	($12.6)	($13.0)	($13.4)	($13.8)	($14.2)
Fixed O&M Cost				($12.9)	($13.2)	($13.6)	($14.0)	($14.5)	($14.9)	($15.3)	($15.8)	($16.3)	($16.8)	($17.3)	($17.8)
Biomass Cost				$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
TOTAL EXPENSES	($23.3)	($139.7)	($69.9)	($50.3)	($53.7)	($56.8)	($57.5)	($58.3)	($59.1)	($59.9)	($60.7)	($61.5)	($62.4)	($63.3)	($64.3)
SNG Revenues				$32.3	$50.8	$67.1	$69.1	$71.2	$73.3	$75.5	$77.8	$80.1	$82.5	$85.0	$87.6
Power Revenues				$7.8	$12.3	$16.2	$16.7	$17.2	$17.7	$18.3	$18.8	$19.4	$20.0	$20.6	$21.2
TOTAL REVENUES BEF TAX				$40.1	$63.1	$83.4	$85.9	$88.4	$91.1	$93.8	$96.6	$99.5	$102.5	$105.6	$108.8
Annual Depreciation															
General Plant															
DDB				$104.9	$74.9	$53.5	$38.2	$27.3	$19.5	$13.9					
SL				$52.4	$43.7	$37.4	$33.4	$31.8	$31.8	$31.8					
Remaining Value				$262.1	$187.2	$133.7	$95.5	$68.2	$48.7	$34.8					
Actual				$104.9	$74.9	$53.5	$38.2	$31.8	$31.8	$31.8					
Steam Plant															
DDB				$7.8	$7.2	$6.6	$6.1	$5.7	$5.3	$4.9	$4.5	$4.2	$3.8	$3.6	$3.3
SL				$5.2	$5.0	$5.3	$5.6	$6.0	$6.0	$6.0	$6.0	$6.0	$6.0	$6.0	$6.0
Remaining Value				$95.7	$88.6	$81.9	$75.8	$70.1	$64.8	$60.0	$55.5	$51.3	$47.5	$43.9	$40.6
Actual				$7.8	$7.2	$6.6	$6.1	$6.0	$6.0	$6.0	$6.0	$6.0	$6.0	$6.0	$6.0
Net Revenue				($122.9)	($72.7)	($33.6)	($16.0)	($7.7)	($5.8)	($3.9)	$30.0	$32.0	$34.1	$36.3	$38.5
Losses Forward					($122.9)	($195.6)	($229.1)	($245.2)	($252.8)	($258.6)	($262.5)	($232.6)	($200.6)	($166.5)	($130.2)
Taxable Income				($122.9)	($195.6)	($229.1)	($245.2)	($252.8)	($258.6)	($262.5)	($232.6)	($200.6)	($166.5)	($130.2)	($91.7)
Income Tax				$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
Annual cash income				($10.3)	$9.4	$26.6	$28.3	$30.2	$32.0	$34.0	$35.9	$38.0	$40.1	$42.3	$44.5
Annual Net Cash Flow	($23.3)	($139.7)	($69.9)	($10.3)	$9.4	$26.6	$28.3	$30.2	$32.0	$34.0	$35.9	$38.0	$40.1	$42.3	$44.5
Discount factor	$0.9	$0.8	$0.7	$0.6	$0.6	$0.5	$0.5	$0.4	$0.4	$0.3	$0.3	$0.3	$0.2	$0.2	$0.2
Annual Present Value	($20.8)	($111.4)	($49.7)	($6.5)	$5.3	$13.5	$12.8	$12.2	$11.5	$10.9	$10.3	$9.7	$9.2	$8.6	$8.1
Cumulative Cash Flow	**($20.8)**	**($132.2)**	**($181.9)**	**($188.4)**	**($183.1)**	**($169.6)**	**($156.8)**	**($144.6)**	**($133.1)**	**($122.2)**	**($111.8)**	**($102.1)**	**($92.9)**	**($84.2)**	**($76.1)**

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Table F.14 Discounted cash flow sheet for years 13-30 of Case CERT-7

16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33
2028	2029	2030	2031	2032	2033	2034	2035	2036	2037	2038	2039	2040	2041	2042	2043	2044	2045
13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
($32.2)	($32.2)	($32.2)															
($14.7)	($15.1)	($15.6)	($16.0)	($16.5)	($17.0)	($17.5)	($18.0)	($18.6)	($19.1)	($19.7)	($20.3)	($20.9)	($21.5)	($22.2)	($22.8)	($23.5)	($24.2)
($18.3)	($18.9)	($19.4)	($20.0)	($20.6)	($21.2)	($21.9)	($22.5)	($23.2)	($23.9)	($24.6)	($25.4)	($26.1)	($26.9)	($27.7)	($28.6)	($29.4)	($30.3)
$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
($65.2)	($66.2)	($67.2)	($36.0)	($37.1)	($38.2)	($39.4)	($40.6)	($41.8)	($43.0)	($44.3)	($45.7)	($47.0)	($48.4)	($49.9)	($51.4)	($52.9)	($54.5)
$90.2	$92.9	$95.7	$98.6	$101.5	$104.6	$107.7	$110.9	$114.3	$117.7	$121.2	$124.9	$128.6	$132.5	$136.4	$140.5	$144.8	$149.1
$21.8	$22.5	$23.2	$23.8	$24.6	$25.3	$26.1	$26.8	$27.6	$28.5	$29.3	$30.2	$31.1	$32.0	$33.0	$34.0	$35.0	$36.1
$112.0	$115.4	$118.8	$122.4	$126.1	$129.9	$133.8	$137.8	$141.9	$146.2	$150.6	$155.1	$159.7	$164.5	$169.4	$174.5	$179.8	$185.2
$3.0	$2.8	$2.6	$2.4	$2.2	$2.1	$1.9	$1.8										
$6.0	$6.0	$6.0	$6.0	$6.0	$6.0	$6.0	$6.0										
$37.6	$34.8	$32.1	$29.7	$27.5	$25.4	$23.5	$21.8										
$6.0	$6.0	$6.0	$6.0	$6.0	$6.0	$6.0	$6.0										
$40.8	$43.2	$45.6	$80.4	$83.0	$85.6	$88.4	$91.2	$100.1	$103.1	$106.2	$109.4	$112.7	$116.1	$119.6	$123.1	$126.8	$130.6
($91.7)	($50.9)	($7.7)	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
($50.9)	($7.7)	$38.0	$80.4	$83.0	$85.6	$88.4	$91.2	$100.1	$103.1	$106.2	$109.4	$112.7	$116.1	$119.6	$123.1	$126.8	$130.6
$0.0	$0.0	$14.4	$30.5	$31.5	$32.5	$33.6	$34.7	$38.0	$39.2	$40.4	$41.6	$42.8	$44.1	$45.4	$46.8	$48.2	$49.6
$46.8	$49.2	$37.2	$55.8	$57.4	$59.1	$60.8	$62.5	$62.1	$63.9	$65.9	$67.8	$69.9	$72.0	$74.1	$76.3	$78.6	$81.0
$46.8	$49.2	$37.2	$55.8	$57.4	$59.1	$60.8	$62.5	$62.1	$63.9	$65.9	$67.8	$69.9	$72.0	$74.1	$76.3	$78.6	$81.0
$0.2	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.1	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0	$0.0
$7.6	$7.2	$4.8	$6.5	$6.0	$5.5	$5.0	$4.6	$4.1	$3.8	$3.5	$3.2	$2.9	$2.7	$2.5	$2.3	$2.1	$1.9
($68.5)	**($61.3)**	**($56.5)**	**($50.0)**	**($44.0)**	**($38.6)**	**($33.6)**	**($28.9)**	**($24.8)**	**($21.1)**	**($17.6)**	**($14.4)**	**($11.5)**	**($8.8)**	**($6.4)**	**($4.1)**	**($2.0)**	**($0.1)**