# Techno-Economic Analysis of Biofuels Production Based on Gasification

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# Foreword

The purpose of this techno-economic analysis is to compare a set of biofuel conversion technologies selected for their promise and near-term technical viability. Every effort is made to make this comparison on an equivalent basis using common assumptions. The process design and parameter value choices underlying this analysis are based on public domain literature only. For these reasons, these results are not indicative of potential performance, but are meant to represent the most likely performance given the current state of public knowledge.

# List of Acronyms

| | |
|---|---|
| AGR | acid gas removal |
| ASU | air separation unit |
| BTL | biomass to liquids |
| CFB | circulating fluidized bed |
| DCFROR | discounted cash flow rate of return |
| DME | dimethyl-ether |
| FCI | fixed capital investment |
| FT | Fischer-Tropsch |
| GGE | gallon of gasoline equivalent |
| HRSG | heat recovery steam generator |
| HT | high temperature |
| IC | indirect costs |
| IGCC | integrated gasification combined cycle |
| IRR | internal rate of return |
| ISU | Iowa State University |
| LHV | lower heating value |
| LT | low temperature |
| MEA | monoethanolamine |
| MJ | megajoule |
| MM | million |
| MTG | methanol to gasoline |
| MW | megawatt |
| $Nm^3$ | normal cubic meter |
| NREL | National Renewable Energy Laboratory |
| PSA | pressure swing adsorption |
| PV | product value |
| Sasol | South African Coal, Oil, and Gas Corporation |
| SPR | slurry phase reactor |
| SMR | steam methane reforming |
| SWGS | sour water-gas-shift |
| TCI | total capital investment |
| TDIC | total direct and indirect cost |
| TIC | total installed cost |
| tpd | tons per day |
| TPEC | total purchased equipment cost |
| WGS | water-gas-shift |

# Executive Summary

This study compares capital and production costs of two biomass-to-liquid production plants based on gasification. The goal is to produce liquid transportation fuels via Fischer-Tropsch synthesis with electricity as a co-product. The biorefineries are fed by 2,000 metric tons per day of corn stover. The first biorefinery scenario is an oxygen-fed, low-temperature (870°C), non-slagging, fluidized bed gasifier. The second scenario is an oxygen-fed, high-temperature (1,300°C), slagging, entrained flow gasifier. Both are followed by catalytic Fischer-Tropsch synthesis and hydroprocessing to naphtha-range (gasoline blend stock) and distillate-range (diesel blend stock) liquid fractions. (Hydroprocessing is a set of refinery processes that removes impurities and breaks down large molecules to fractions suitable for use in commercial formulations.)

Process modeling software (Aspen Plus) is utilized to organize the mass and energy streams and cost estimation software is used to generate equipment costs. Economic analysis is performed to estimate the capital investment and operating costs. A 20-year discounted cash flow rate of return analysis is developed to estimate a fuel product value (PV) at a net present value of zero with 10% internal rate of return. All costs are adjusted to the year 2007. The technology is limited to commercial technology available for implementation in the next 5–8 years, and as a result, the process design is restricted to available rather than projected data.

Results show that the total capital investment required for $n^{th}$ plant scenarios is \$610 million and \$500 million for high-temperature and low-temperature scenarios, respectively. PV for the high-temperature and low-temperature scenarios is estimated to be \$4.30 and \$4.80 per gallon of gasoline equivalent (GGE), respectively, based on a feedstock cost of \$75 per dry short ton. The main reason for a difference in PV between the scenarios is because of a higher carbon efficiency and subsequent higher fuel yield for the high-temperature scenario. Sensitivity analysis is also performed on process and economic parameters. This analysis shows that total capital investment and feedstock cost are among the most influential parameters affecting the PV, while least influential parameters include per-pass Fischer-Tropsch-reaction-conversion extent, inlet feedstock moisture, and catalyst cost.

In order to estimate the cost of a pioneer plant (first of its kind), an analysis is performed that inflates total capital investment and deflates the plant output for the first several years of operation. Base case results of this analysis estimate a pioneer plant investment to be \$1.4 billion and \$1.1 billion for high-temperature and low-temperature scenarios, respectively. Resulting PVs are estimated to be \$7.60/GGE and \$8.10/GGE for high-temperature and low-temperature pioneer plants, respectively.

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# Introduction

This study investigates the economic feasibility of the thermochemical pathway of gasification of biomass to renewable transportation fuels. The objective is to compare capital investment costs and production costs for $n^{th}$ plant biorefinery scenarios based on gasification. The selected scenarios are high-temperature (slagging) gasification and low-temperature (dry-ash) gasification, each followed by Fischer-Tropsch synthesis and hydroprocessing. They are designed to produce liquid hydrocarbon fuels from 2,000 dry metric tons (2,205 dry short tons) per day of agricultural residue, namely corn stover. Corn stover is chosen as a feedstock in order to facilitate comparisons with biochemical and pyrolysis biofuels scenarios [1, 2].

The two scenarios were chosen from many options according to the following criteria:

1. The technology under consideration should be commercially ready in the next 5–8 years.
2. The size of the biorefinery should be feasible with current agricultural productivity and within a realistic feedstock collection area.
3. In addition, the desired end product should be compatible with the present fuel infrastructure, i.e., gasoline and/or diesel.

The high-temperature gasification scenario is based on a steam/oxygen-fed entrained flow, slagging gasifier similar to that described in Frey and Akunuri [3]. The low-temperature gasification scenario is based on a pressurized, steam/oxygen-fed fluidized bed gasifier developed by the Gas Technology Institute and reported by Bain [4]. The main areas of operation are feedstock preprocessing, gasification, syngas cleaning, syngas conditioning/upgrading, fuel synthesis, power generation, and air separation (for oxygen production), as shown in Figure 1. Process modeling software is utilized to organize the mass and energy streams, and cost estimation software is used to generate equipment costs. Economic analysis is performed to estimate the capital investment and operating costs. A 20-year discounted cash flow rate of return (DCFROR) analysis is developed to estimate a fuel product value (PV) at a net present value of zero with 10% internal rate of return. All costs are adjusted to the year 2007.

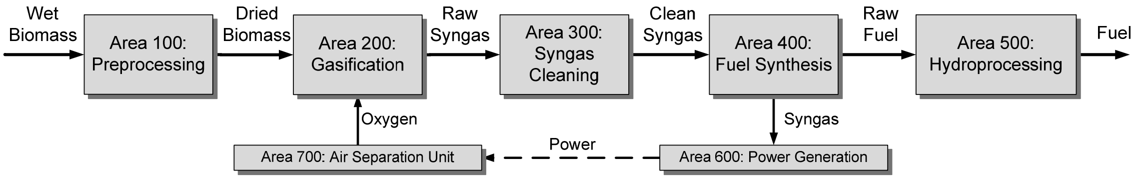


**Figure 1. Overall process flow diagram for both scenarios**

# Background

## Biorenewable Resources

The world population has long utilized materials that were in close proximity. The nearest resource available to the human population is the organic matter in the environment around it. This organic matter is present for a limited amount of time due to its decomposable nature. Brown [5] defines this material, or biorenewable resources, as organic material of recent biological origin. It is a renewable resource if the rate of consumption is equal to the regeneration or growth and therefore must be used only if preserving biodiversity [6]. As a result, these resources have been important contributors to the world economy, providing foodstuffs, transportation, energy, and construction materials, as well as serving many other functions.

Biorenewable resources for generating energy can be classified as woody biomass, energy crops, residues, and municipal waste [6]. The first two are primary resources while the remaining are secondary resources, meaning that their primary use has already occurred. Woody biomass includes logging products and energy crops include short-rotation trees such as poplar and fast-growing grasses such as switchgrass. Residues can come from logging processing or agricultural processing (e.g., corn stover). According to Perlack et al. [7], the energy crop and agricultural residue potential in the United States is 1.4 billion annual tons. According to the U.S. Department of Energy's "Roadmap for Agriculture Biomass Feedstock Supply in the U.S.," there is potential for 2 billion annual tons, including municipal waste and biosolids such as manure.

Many end products can be produced from these resources. Aside from the conventional use of biomass for human food consumption, livestock feed, and building materials, there are many new pathways to provide renewable alternatives to our transportation, infrastructure, and energy. Combustion of biomass offers a way to provide heat and power and displaces coal and fuel oil. Fermentation of carbohydrates and liquefaction of biomass through fast pyrolysis yield liquid products with the potential to displace petrochemicals. Gasification of biomass allows for chemical and liquid fuel synthesis, which is the focus of this study.

Developing an economy that involves biorenewable resources, especially biofuels, has many benefits. According to Greene et al. [8], biofuel production has the potential to provide a new source of revenue for farmers by generating $5 billion per year. Additionally, toxic and greenhouse gas emissions can be reduced by the use of biofuels. In the same study, Greene et al. report that 22% of the United States's total greenhouse gas emissions could be reduced if biofuels were developed to replace half of the petroleum consumption. Arguably, the most important benefit of biofuel production is the potential for closing the carbon cycle.

## Gasification

Gasification is a high-temperature and catalytic pathway for producing biofuels. It is defined as the partial oxidation of solid, carbonaceous material with air, steam, or oxygen into a flammable gas mixture called producer gas or synthesis gas [5]. The synthesis gas contains mostly carbon monoxide and hydrogen with various amounts of carbon dioxide, water vapor, and methane. Typical volumetric energy content of synthesis gas is 4–18 $MJ/Nm^3$ [9]. Comparatively, natural gas (composed of mostly methane) energy content is 36 $MJ/Nm^3$ [9]. Much of the energy content of the biomass is retained in the gas mixture by partial oxidation rather than full oxidation of the biomass, which would result in the release of mostly thermal energy. Historically, gasification of coal and wood produced "town gas," which was subsequently burned in street lamps [10]. Additionally, during the World Wars, vehicles were adapted to operate with gasification reactors [10]. During this same time period, Germany developed the catalytic synthesis of transportation fuels from synthesis gas [11]. The same concept is still in use today by the South African Coal, Oil, and Gas Corporation (Sasol) to produce motor fuels and liquid by-products using coal [11].

### ***Reactions***

Four stages occur during gasification of carbonaceous material: drying, devolatilization, combustion, and reduction [9]. First, the moisture within the material is heated and removed through a drying process. Second, continued heating devolatilizes the material where volatile matter exits the particle and comes into contact with the oxygen. Third, combustion occurs, where carbon dioxide and carbon monoxide are formed from carbon and oxygen. The combustion stage is very exothermic and provides enough heat for the last stage, the reduction reactions, to occur. The last stage includes water gas reaction, Boudouard reaction, water-gas-shift reaction, and methanation reaction (Table 1). As all these stages progress, solid fixed carbon remains present. The amount of fixed carbon varies depending on the equivalence ratio.

**Table 1. Reactions Occurring within the Reduction Stage of Gasification**

| Name | Reaction |
|---|---|
| Water Gas | $C + H_2O \rightarrow CO + H_2$ |
| Boudouard | $C + CO_2 \rightarrow 2CO$ |
| Water-Gas-Shift | $CO + H_2O \rightarrow CO_2 + H_2$ |
| Methanation | $CO + 3H_2 \rightarrow CH_4 + H_2O$ |

When the equivalence ratio (defined as the actual air/fuel ratio all divided by the stoichiometric air/fuel ratio) increases, solid fixed carbon (i.e., char) decreases until enough oxidizer is available for complete conversion (Figure 2). This point of complete conversion occurs at an equivalence ratio of approximately 0.25. At nearly the same point, the maximum synthesis gas energy content (without accounting for sensible energy) is reached.

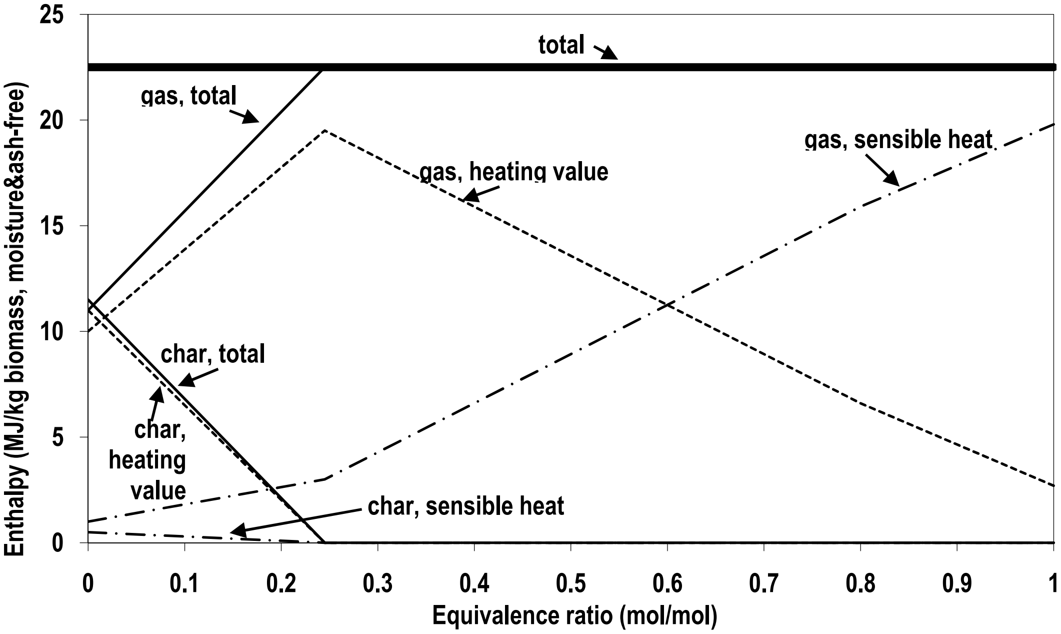


**Figure 2. Typical energy content of the products of gasification of wood using air varied by equivalence ratio [12]**

## *Gasifier Types*

Three types of reactors are used for gasification: fixed bed, fluidized bed, and entrained flow [13]. Fixed beds are fed with biomass from the top of the reactor and form a bed, which gasifies as air moves through the bed (Figure 3). As the material releases volatile components, the char and ash exit through a grate at the bottom. Typical operating temperature range is 750°–900°C. The two main types of fixed-bed gasifiers are updraft and downdraft. The advantage of fixed-bed technology is its simplicity, but the technology is limited in scale-up and has low heat mixing due to high channeling potential within the reactor [14].

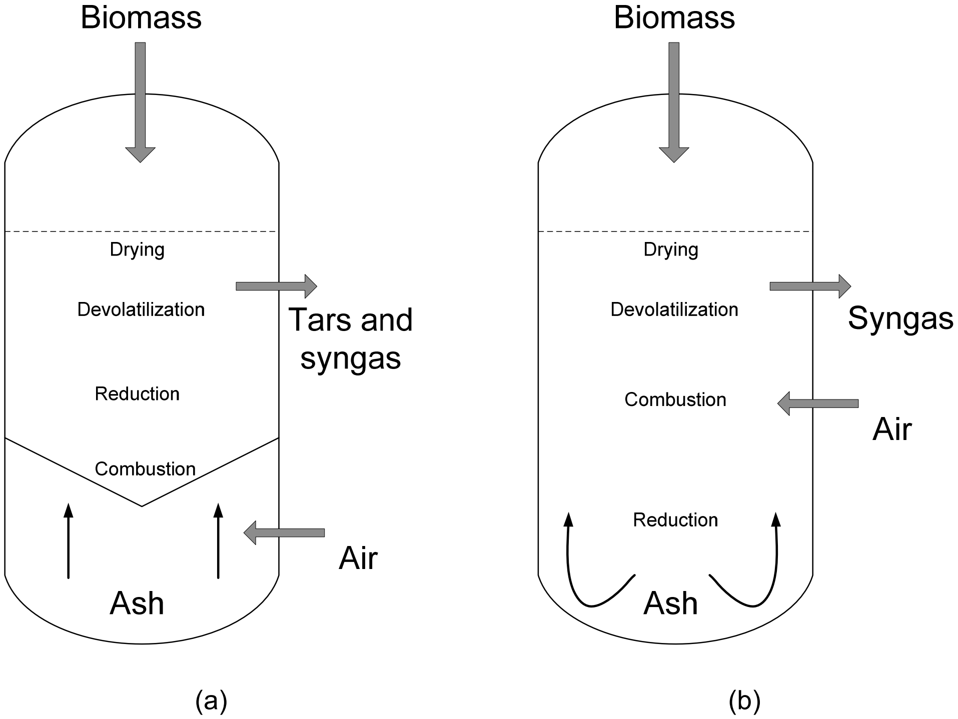
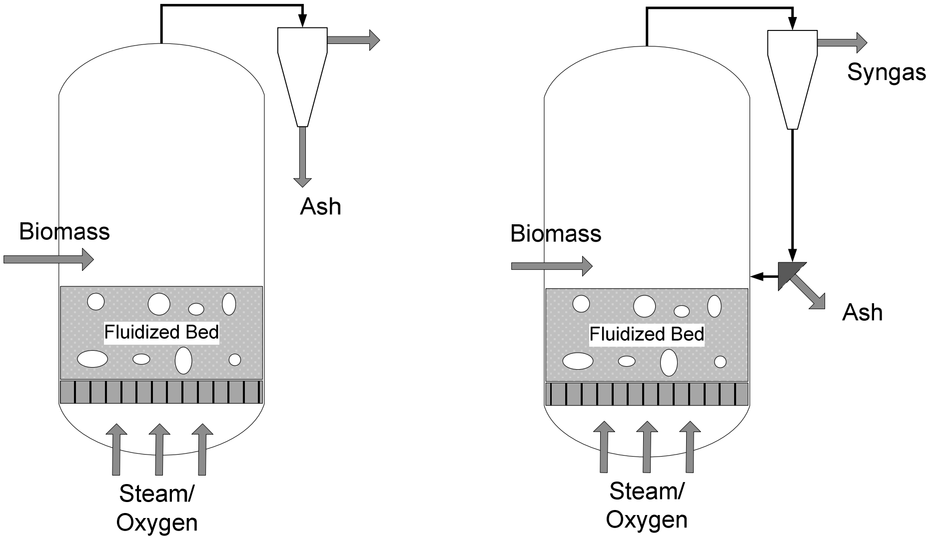


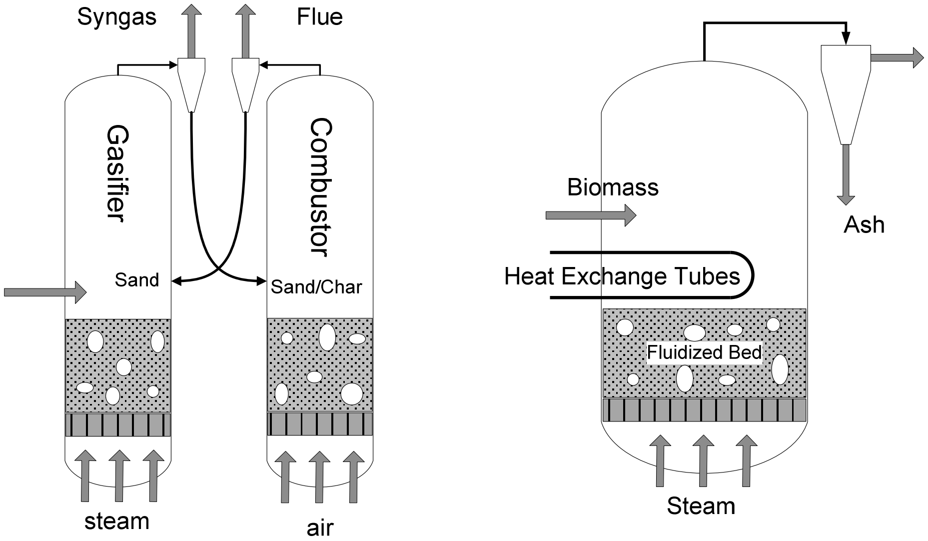
**Figure 3. Design of fixed-bed (a) updraft and (b) downdraft gasifiers showing reaction zones [13]**

When the volumetric gas flow is increased through the grate, the fixed bed becomes a fluidized bed. Fluidized-bed gasifiers are so named because of the inert bed material that is fluidized by oxidizing gas creating turbulence through the bed material (Figure 4). Biomass enters just above the top of the bed and mixes with hot, inert material, creating very high heat and mass transfer. Operating temperature range is the same as for the fixed bed. Advantages of the fluidized bed include flexible feeds, uniform temperature distribution across the bed, and large volumetric flow capability [15]. The main types of fluidized-bed gasifiers are circulating fluidized bed (CFB) and bubbling fluidized bed. Bubbling-bed gasifiers are directly heated from the combustion reactions occurring in the bed. They produce gas and the ash and char fall out the bottom or the side. The CFB recycles the char through a cyclone while the product leaves out the top of the cyclone. There are also indirectly heated fluidized beds that use a hot material such as sand to provide the heat needed for gasification, as shown in Figure 4. Fluidized beds have high carbon conversion efficiencies and can scale up easily [14].



(a) Bubbling fluidized-bed gasifier

(b) Circulating fluidized-bed gasifier



(c) Indirectly heated gasifier via combustor

(d) Indirectly heated gasifier via heat exchange tubes

**Figure 4. Fluidized bed gasifier designs of (a) and (b) directly heated type and (c) and (d) indirectly heated type [16]**

Another type of gasifier is the entrained-flow gasifier (Figure 5). Normally operated at elevated pressures (up to 50 bar), it requires very fine fuel particles gasified at high temperatures to ensure complete gasification during the short residence times in the reactor. The Energy Research Centre of the Netherlands has investigated this type of gasification and reported promise with biomass, as long as the biomass is pretreated to certain requirements [17]. To keep the residence time at approximately the time for a particle to fall the length of the reaction zone, fuel particles smaller than 1 mm and high temperatures (1100°–1500°C) are necessary for successful operation.

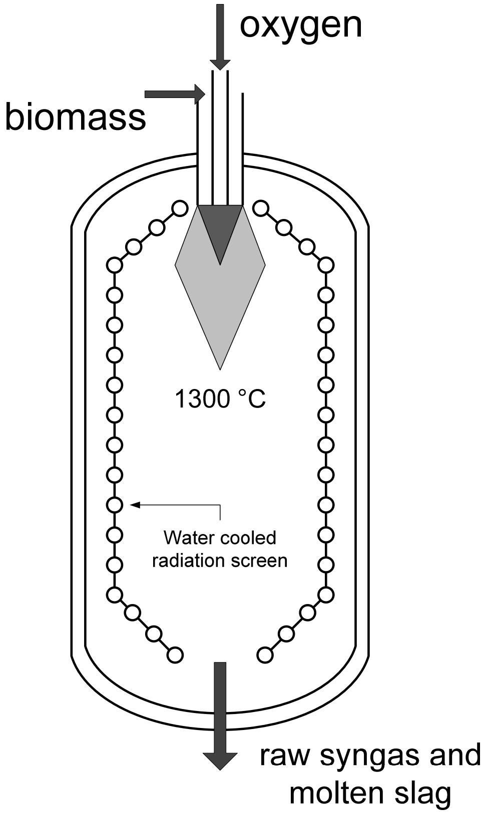


**Figure 5. Entrained-flow gasifier [18]**

Entrained-flow gasification mixes the fuel with a steam/oxygen stream to form a turbulent flow within the gasifier. Ash-forming components melt in the gasifier and form a liquid slag on the inside wall of the gasifier, effectively protecting the wall itself. The liquid flows down and is collected at the bottom. To form the slag, limestone can be added as a fluxing material. For herbaceous biomass, such as switchgrass or corn stover, which is high in alkali content, there may be sufficient inherent fluxing material present [18]. Advantages of entrained-flow gasification are that tar and methane content are negligible and high carbon conversion occurs due to more complete gasification of the char. Syngas cleanup is simplified because slag is removed at the bottom of the gasifier, negating the need for cyclones and tar removal [19]. The disadvantages are that very high temperatures need to be maintained and the design and operation is more complex. An entrained-flow gasifier co-firing up to 25% biomass with coal has been developed by Shell in Buggenum, Netherlands. Another gasifier developed by Future Energy in Freiburg, Germany, uses waste oil and sludges. Both are operating at commercial scale [17].

## Biomass Preprocessing

A degree of biomass processing is required before gasification can occur. Most gasifiers require smaller size feedstock than is typically collected during harvest. Therefore, a significant degree of size reduction is necessary. A typical setup for size reduction is using a two-step process in which a chipper accomplishes the primary reduction and a hammer mill is used for the secondary reduction [20]. In addition, the maximum moisture content for biomass gasification is 20%–30%

(wet basis), and for normal operation it is less than 15% (wet basis) [9]. Therefore, a drying process is required to prepare the feedstock for gasification.

The main benefit of drying biomass is to avoid using energy within the gasifier to heat and dry the feedstock [21]. Drier biomass also makes for more stable temperature control within the gasifier. Rotary dryers typically operate utilizing hot flue gas from a downstream process as the drying medium. They have high capacity but require long residence times. In addition, rotary dryers have a high fire hazard when using flue gas [21]. To avoid using flue gas, rotary dryers can use superheated steam, essentially an inert gas, when a combined-cycle heat and power system is used downstream. That system has significant steam available for use because of the steam produced in the steam cycle. An advantage of using steam for drying is better heat transfer and therefore shorter residence time.

Pretreatment options for entrained-flow gasification include torrefaction followed by grinding to 0.1-mm particles, grinding to 1-mm particles, pyrolysis to produce bio-oil/char slurry (bioslurry), and initial fluidized-bed gasification of larger particles coupled to an entrained-flow gasifier. Torrefaction, essentially an oxygen-free roasting process, causes the biomass particles to be brittle, which makes for easy grinding but releases up to 15% of the energy in the biomass via volatile compounds [17]. The fluidized/entrained coupled option is attractive because of an overall energy efficiency of 80%–85%, but it is expensive because two gasifiers are used in series.

The bioslurry option is illustrated in Figure 6. Basically, a flash-pyrolysis process yields bio-oil and char and is followed by a slagging, entrained-flow gasifier. Because this process utilizes an entrained-flow gasifier, the feed must be pressurized. Fortunately, the pyrolysis slurry, already in an emulsified liquid state, can be pressurized easily. Technology for slurry feeding is state of the art as a result of experience with coal slurries [17]. The bioslurry still contains 90% of the energy contained in the original biomass [22]. Another advantage is that no inert gas is needed for solids pressurization, avoiding dilution of the feed, which would dilute the syngas. In the search for cost-effective methods for production of syngas, this option has potential, but it isn't as developed as other technologies such as fluidized-bed gasification. The biggest challenge is constructing and operating a large-scale pyrolysis process, because large-scale systems have not been demonstrated [17].

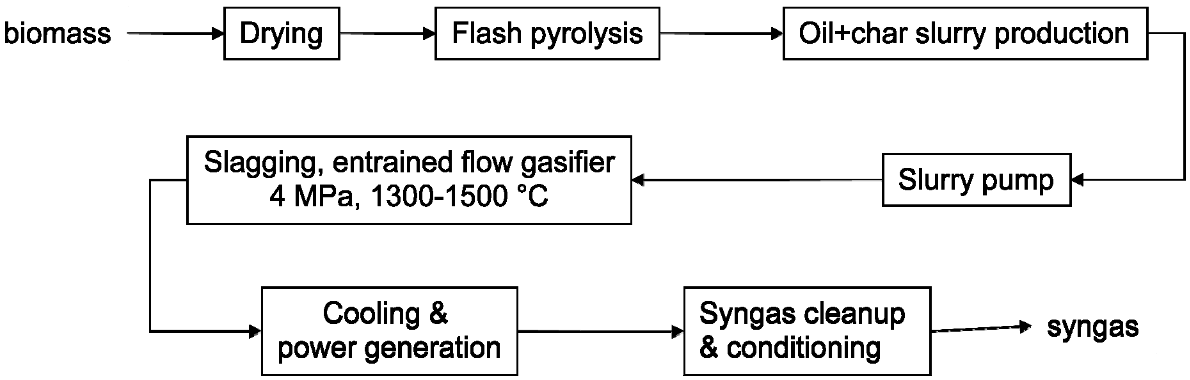


**Figure 6. Schematic of a biomass pretreatment via fast pyrolysis followed by an entrained-flow gasifier [17]**

## Syngas Cleaning

Because the raw syngas leaving the gasifier contains particulates, tars, alkali compounds, sulfur compounds, nitrogen compounds, and other contaminants, those components must be removed or reduced significantly. Particulates and tars have the potential for clogging downstream processes. Sulfur and nitrogen have the potential to poison downstream processes, especially catalysts used in fuel synthesis applications. Moreover, another motivation for cleaning syngas is meeting environmental emissions limits.

The syngas must be cooled before conventional gas cleanup is utilized. This can be accomplished in two ways: direct quench by injection of water and indirect quench via a heat exchanger. Direct quench is less expensive but dilutes the syngas. Direct quenching can also be used to clean up the gas by removing alkali species, particulates, and tars [23].

Particulates are defined as inorganic mineral material, ash, and unconverted biomass or char [24]. In addition, bed material from the gasifier is included in the particulates. Switchgrass feedstock typically has 10% inorganic material in the form of minerals. Many gasifiers operate with a 98%–99% carbon conversion efficiency where 1%–2% of the solid carbon is in the form of char [24].

Particulates are primarily removed by physical systems such as cyclones, in which the heavy particles fall down the center while the gases rise up and out of the cyclone. The initial step for particulate removal is usually a cyclone. Importantly, particulates should be removed before the gas is cooled down for cold gas cleaning. If removed after gas cooling, tars can condense onto the particulates and potentially plug equipment. Alternatively, barrier filters, which operate above tar condensation temperatures, use metal or ceramic screens or filters to remove particulates while allowing the gas to remain hot. Barrier filters, however, have presented problems in sintering and breaking [24].

Even more critical to downstream syngas applications is tar removal. Tars are defined as higher-weight organics and oxygenated aromatics heavier than benzene (78 g/mol). They are produced from volatized material after polymerization [24]. A review by Milne et al. [25] of tars produced during gasification covers different removal methods. One method uses physical removal via wet gas scrubbing of tars in a scrubbing tower for the “heavy tars” followed by a venturi scrubber for lighter tars. This setup is similar to the direct quench cooling mentioned previously, as cooling occurs as well. Tar concentration is reported to be lower than 10 ppm by volume at the exit of this setup [23]. The disadvantage of this setup is that wastewater treatment is required to dispose of the tar and can be expensive.

The other method for tar removal is catalytic or thermal conversion to non-condensable gas. This is also known as hot gas cleaning, as it occurs at temperatures at or above gasification temperatures. Catalytic conversion can occur at temperatures as low as 800°C, and thermal conversion can occur at temperatures up to 1200°C. The energy required for thermal tar cracking may not be cost-competitive because of the temperature rise required from the gasification temperature to crack the high refractory tars [24].

Alkali compounds such as calcium oxide and potassium oxide are present in biomass. When gasified, both either vaporize or concentrate in the ash. Condensation of these compounds begins at 650°C, and they can deposit on cool surfaces causing equipment clogging, equipment corrosion, and catalyst deactivation [26]. According to Stevens [26], research on alkali adsorption filters using bauxite has been promising but not demonstrated on a large scale. Stevens concludes that the best current method for alkali removal is using proven syngas cooling followed by wet scrubbing, where the addition of water cools the syngas and physically removes small particles and liquid droplets.

Wet scrubbing also removes ammonia that forms during gasification from the nitrogen in the biomass. Without proper removal, ammonia can deactivate catalysts as well. Complete ammonia removal can be accomplished through wet scrubbing [27]. For gasifiers coupled to a catalytic or thermal tar reformer, most of the ammonia can be reformed to hydrogen and nitrogen [27]. Sulfur in the biomass mostly forms hydrogen sulfide ($H_2S$) with small amounts of carbonyl sulfide (COS). Hydrogen sulfide is removed by three main ways: chemical solvents, physical solvents, and catalytic sorbents. For chemical removal, amine-based solvents are typically utilized to chemically bond with $H_2S$. Physical removal takes advantage of the high solubility of $H_2S$ using an organic solvent. Typical setups of both chemical and physical removal involve an absorber unit followed by a solvent regenerator unit, known as a stripper. Operation usually occurs at temperatures lower than 100°C and medium to high pressures (150–500 psi) [27]. Sulfur leaving these two systems is around 1–4 ppm and can require further removal, especially for fuel synthesis. In that case, a syngas polishing step using a fixed-bed zinc oxide-activated carbon catalyst removes $H_2S$ and COS to achieve the parts per billion levels necessary for fuel synthesis. Halides, present in trace amounts in the biomass, can also be removed with the zinc oxide catalyst [27]. The water streams with the scrubbed-out impurities are sent to wastewater treatment.

## End-Use Product

After syngas is cleaned of particulates, impurities, and contaminants, there is sufficient energy content for producing a higher valued product. There are three main large-scale biomass gasification pathways that have been researched and suggested for producing higher valued products: power generation, liquid-fuel synthesis, and chemical synthesis. According to Wender [28], the three largest commercial uses for syngas are ammonia production from hydrogen, methanol synthesis, and hydrocarbon synthesis via the Fischer-Tropsch process.

### *Power Generation*

Power is most effectively generated using gasification by combusting the syngas in a gas turbine to provide mechanical work for a generator. Steam can then be generated by recovering heat from the hot syngas, and the steam in turn is used for mechanical work via a steam turbine. This gasifier plus gas turbine and steam turbine setup is known as integrated gasification combined cycle (IGCC) power generation. The level of particulates, alkali metals, and tar can decrease the performance of the gas turbine. Consonni and Larson [29] found that particulates can cause turbine blade erosion; 99% of 10-micron or smaller particles should be removed. In addition, they also report that alkali metals corrode the turbine blades and tars condense on the turbine blades, both hindering operation and escalating turbine failure. Fortunately, nearly all alkali metals and tars can be removed using proven wet scrubbing techniques.

Using the IGCC approach to generate power, Bridgwater et al. [30] and Craig and Mann [23] expect biomass to produce power with efficiencies in the range of 35%–40% with large scale systems (greater than 100 MW net output) at the high end of the range. Moreover, Craig and Mann suggest that future advanced turbine systems could reach 50% biomass-to-power efficiency.

### *Synthetic Fuels and Chemicals*

Instead of converting the energy content of the syngas to power, the energy content can be condensed into a liquid energy carrier, or fuel. The conversion of syngas to fuels can only occur in the presence of proper catalysts [31]. The catalytic reactions basically build up the small molecules in the syngas (i.e., carbon monoxide and hydrogen) into larger compounds that are more easily stored and transported. A summary of many catalytic pathways to fuels and chemicals is shown in Figure 7. In most catalytic synthesis reactions, syngas cleanliness requirements are very high. Most impurities and contaminants are removed to low parts-per-million and even parts-per-billion concentrations. This means that significant cost must be directed toward syngas cleaning.

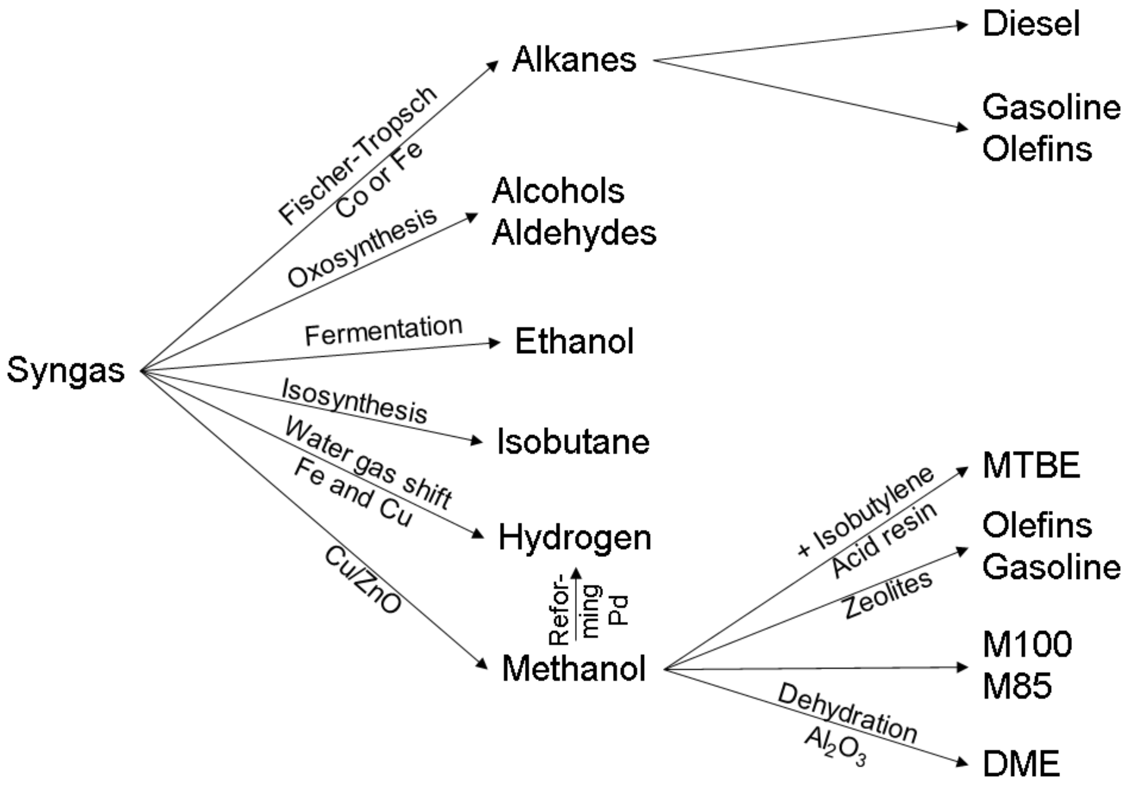


**Figure 7. Main syngas conversion pathways [32]**

### *Methanol to Gasoline*

Methanol is one of the top chemicals produced in the world [32]. Most commercially produced methanol is synthesized via steam methane reforming and autothermal reforming. The synthesis of methanol from syngas is highly exothermic (equation 1). The reaction occurs over a

$Cu/ZnO/Al_2O_3$ catalyst at temperatures of 220°–275°C and pressures of 50–100 bar with a catalyst lifetime of 2–5 years [31]. Wender [28] reports syngas-to-methanol conversion efficiency can reach 99% with recycle, but per-pass efficiency is only about 25%.

$$CO + 2H_2 \rightarrow CH_3OH \qquad \text{(eqn. 1)}$$

Although methanol can be used directly as a liquid fuel, it can also be converted into the conventional transportation fuel range. This process is known as the methanol-to-gasoline (MTG) process and was developed by the Mobil Oil Corporation [31]. In that process, methanol is heated to 300°C and dehydrated over alumina catalyst at 27 atm, yielding methanol, dimethyl ether (DME), and water. The exiting mixture reacts with a zeolite catalyst at 350°C and 20 atm to produce 56% water and 44% hydrocarbons by weight. Of the hydrocarbon product, 85% is in the gasoline range, with 40% of the gasoline-range product being aromatic. However, limitations on the aromatic content of gasoline have been proposed in legislation [31]. The thermal efficiency of the MTG process is 70% [11]. The overall MTG process usually includes multiple MTG reactors in parallel to perform periodic catalyst regeneration by burning off coke deposits [11]. Mobil operated a commercial plant producing 14,500 barrels per day in New Zealand during the 1980s [32]. Alternatively, the reaction process could be stopped directly after the methanol synthesis to focus on producing DME, because DME can be used as a diesel fuel because it has a high cetane number. DME is formed from the dehydration reaction of methanol over an acid catalyst γ-alumina. Per-pass efficiency can be as high as 50%. Overall, syngas-to-DME production efficiency is higher than that for syngas to methanol [31]. However, DME is in gaseous form at atmospheric conditions and needs to be pressurized for use in diesel engines [33]. Therefore, engine modification is required; this is the main disadvantage for DME use as transportation fuel.

### Fischer-Tropsch

Fischer-Tropsch (FT) catalytic synthesis is a highly exothermic reaction producing a wide variety of alkanes (equation 2).

$$CO + 2.1H_2 \rightarrow -(CH_2)- + H_2O \qquad \text{(eqn. 2)}$$

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 [34]. Operating pressures are in the range of 10–40 bar. Product distribution can be estimated using the Anderson-Schulz-Flory chain-growth-probability model, in which longer hydrocarbon chains form as the temperature decreases. At high temperatures, selectivity favors methane and light gases. This is a disadvantage if liquid fuel production is the focus. At low temperatures, selectivity favors long-carbon-chain wax products requiring further hydrocracking to the diesel range in a separate unit, which adds more construction cost but is necessary for liquid fuel production.

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 (Figure 8). Heat removal is crucial to the process and has been the focus of reactor design development [31]. The fixed-bed reactor has many catalyst tubes with heat removal achieved by steam generation on the outside of the tubes [35]. 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 [36]. The slurry-phase reactor (SPR) operates by suspending catalyst in a liquid and bubbling the syngas through from the bottom. Disadvantages of the SPR are more complex operation and difficult wax removal. However, SPRs cost approximately 40% less to build than fixed-bed reactors [36].

FT diesel is very low in sulfur, low in aromatic content, and has a high cetane number, making it very attractive as a conventional fuel alternative. Emissions across the board decrease when using FT diesel. A South Africa-based company, Sasol, has been producing transportation fuel since 1955 using the FT process and supplies 41% of South Africa’s transportation fuel requirements [31].

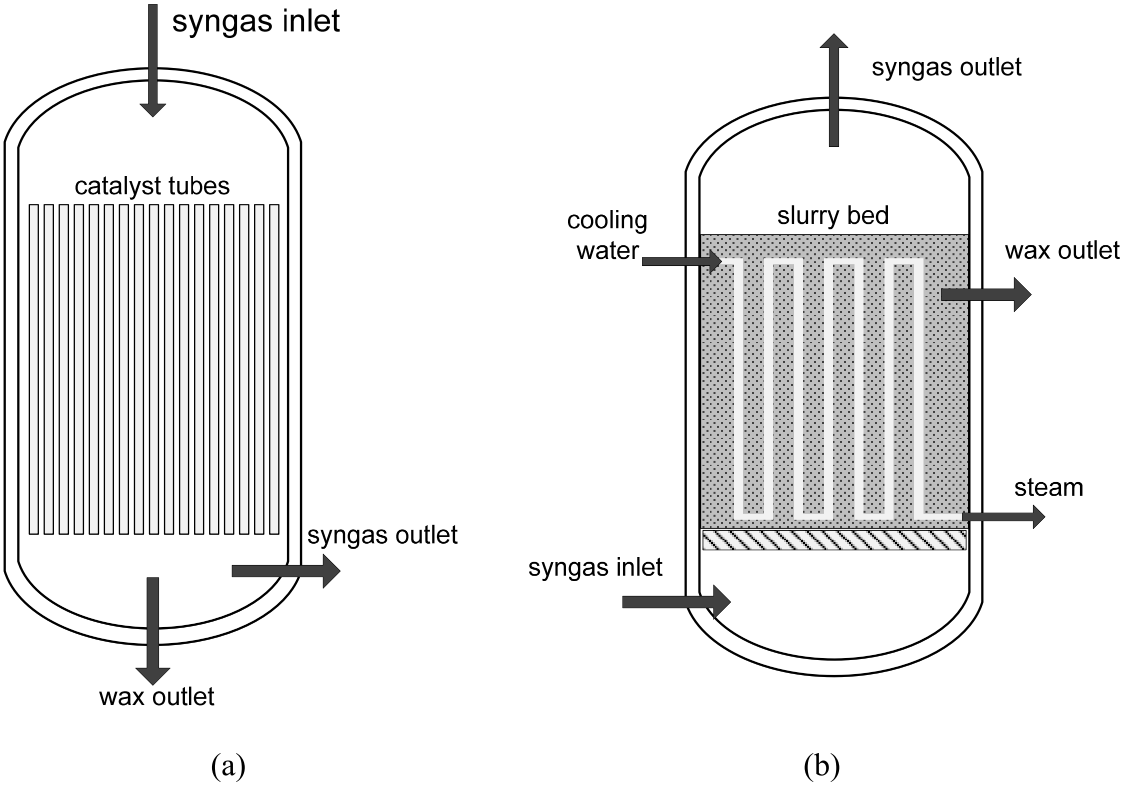


**Figure 8. Fischer-Tropsch reactor types (a) multi-tubular fixed bed and (b) slurry bed [31]**

## Techno-Economic Analysis

For biofuels technologies to be utilized in commercial applications, the economic feasibility must be determined. A feasibility analysis is also called a techno-economic analysis, in which the technical aspects of a project are coupled to the economic aspects. First, the basic theoretical configuration is developed and a mass and energy balance is performed. Second, cost estimation allows the investment and production cost of a biorefinery to be determined. With rising interest in biorenewable resources, many techno-economic studies have been performed on power generation and biofuel scenarios. These studies assist in understanding how the physical process relates to the cost of producing renewable alternatives. Accuracy of results from these studies is usually ±30% of the actual cost [5].

Previous studies of gasification-based biomass-to-liquid production plants estimate the cost of transportation fuels to range from \$12/GJ to \$16/GJ (\$1.60–\$2.00 per gallon of gasoline equivalent) [16, 37-40]. The same studies estimate total capital investment in the range of \$191 million for 2,000 dry metric ton per day (tpd) input [39] to \$541 million for 4,500 dry metric tpd input [38].

A 1,650 dry metric tpd biomass-to-methanol plant based on gasification, with a production cost of \$15/GJ (\$0.90/gal of methanol), is reported by Williams et al. [16] in 1991\$ for \$45/dry metric ton of biomass. Williams et al. also shows the production cost of methanol-derived natural gas to be \$10/GJ (\$0.60/gal of methanol). However, that study concludes that if a carbon tax system was developed for lifecycle carbon emissions, then renewable methanol could become competitive with natural-gas-derived methanol at a tax of approximately \$90 per metric ton of carbon. A more recent study by Larson et al. of switchgrass-to-hydrocarbons production in 2009 reports a production cost of \$15.3/GJ (\$1.90/gal of gasoline) in 2003\$ for a 4,540 dry metric tpd (5,000 dry short tpd) plant based on gasification [38].

Table 2 compares four biofuel production studies based on gasification. A range of cost year, plant size, and feedstock cost show the diversity of characteristics and assumptions that techno-economic studies use. In addition, resulting capital investment costs of the studies have a large range. For example, the capital investment costs of the Phillips et al. [39] and Tijmensen et al. [40] studies are \$191 million and \$387 million, respectively, at similar plant sizes. Reasons for such a significant difference are choice of technologies and level of technology development. The Phillips et al. study is a target study, meaning that it assumes future technology improvement and therefore lower costs. Direct comparison is difficult because of the varying assumptions used by each study.

**Table 2. Previous Techno-Economic Studies of Biomass-Gasification Biofuel Production Plants**

| | **Williams et al. [16]** | **Phillips et al. [39]** | **Tijmensen et al. [40]** | **Larson et al. [38]** |
|---|---|---|---|---|
| Cost Year | 1991 | 2005 | 2000 | 2003 |
| Plant Size (dry metric ton per day) | 1,650 | 2,000 | 1,741 | 4,540 |
| Feedstock | Generic biomass | Poplar | Poplar | Switchgrass |
| Fuel Output | Methanol | Ethanol | FT liquids | Diesel, gasoline |
| Feedstock Cost ($/dry short ton) | 41 | 35 | 33 | 46 |
| Capital Investment ($MM) | N/A | 191 | 387 | 541 |
| Product Value ($/GJ) | 15 | 12 | 16 | 15 |
| Product Value ($/GGE) | 1.90 | 1.60 | 2.00 | 1.85 |

# Methodology

The following steps were undertaken to perform the analysis in this study:

- Collect performance information on relevant technologies for systems under evaluation
- Perform down-selection process with developed criteria to identify most appropriate scenarios
- Design process models using Aspen Plus process engineering software
- Size and cost equipment using Aspen Icarus Process Evaluator software, literature references, and experimental data
- Determine capital investments and perform discounted cash flow analysis
- Perform sensitivity analysis on process and economic parameters
- Perform pioneer plant cost growth and performance analysis.

## Down-Selection Process

A number of process configurations for the gasification-based, biomass-to-liquids (BTL) route were initially considered. These configurations are listed in Table 3 and discussed in the following sections.

**Table 3. Process Configurations Considered in Down Selection Process**

| | |
|---|---|
| **Gasifier Block** | Entrained flow, slagging gasifier |
| | Fluid bed, dry ash gasifier |
| | Transport gasifier, dry ash (e.g., Kellog, Brown, and Root) |
| | Indirect gasifier, dry ash (e.g., Battelle-Columbus Labs) |
| **Syngas Cleaning** | Water scrubbing |
| | Catalytic tar conversion/reduction |
| | Thermal tar conversion/reduction |
| | Amine-based acid gas removal |
| | Physical sorbent-based acid gas removal (e.g., Selexol, Rectisol) |
| **Fuel Synthesis** | Fischer-Tropsch |
| | Mixed alcohols |
| | Methanol to gasoline (MTG) |
| | Dimethyl ether |
| | Syngas fermentation |

### *Preliminary Criteria*
The initial technology configuration options are reviewed and screened in accordance with three criteria.

First, the technology under consideration should be commercially ready in the next 5–8 years and preferably with high technology development. High technology development increases the likelihood that a configuration will perform at the scale defined in this study. For example, coal gasification has been demonstrated commercially at large scales [11]. While similar scale biomass gasifiers have not been proven commercially, the technology development of coal is assumed to apply for biomass in 5–8 years.

Second, the size of biorefinery should be feasible with typical agricultural productivity and within a realistic collection area. For example, if one-third of total land use surrounding the biorefinery is for stover collection and each acre provides conservatively 1 short dry ton per year, then the required collection radius is 35 miles and the amount of biomass transported to the biorefinery is approximately 2,300 short tons (2,090 metric tons) per day. The collection area with a 35-mile radius is assumed to be realistic. In addition, previous studies by Tijmensen et al., Phillips et al., and Lau et al. have used similar plant sizes [39-41].

Third, the desired product should be compatible with the present transportation fuel infrastructure, i.e., gasoline and diesel-range hydrocarbons.

### *Scenario Selection*
For the gasification area, two gasifiers are selected for modeling. First, an entrained-flow, slagging gasifier is chosen because of its commercial application with coal (GE, Siemens, Shell, and ConocoPhillips) and its potential for use with biomass. Moreover, process modeling of this gasifier is simple because it can be closely approximated at thermodynamic equilibrium [3]. Second, a fluidized-bed, dry-ash gasifier is chosen because of experience at the Gas Technology Institute and because of data availability. Also, a report by Bain [4] at the National Renewable Energy Laboratory contains collected and analyzed data for fluidized-bed gasification. In addition, Iowa State University is currently operating an atmospheric-pressure fluidized-bed gasifier as either air or oxygen/steam fed.

The syngas cleaning area is chosen to include configurations that have less technological complexity than previous studies. Phillips et al. [39] and Larson et al. [38] both employ an external catalytic tar-reforming process for dry-ash gasification. Because of low technological development in tar conversion and its inherent complexity, a direct-contact syngas quenching and scrubbing system is instead chosen for this study. In the case of the slagging gasifier, high temperatures inhibit tar formation, yet quenching and particulate and ammonia removal are still required. An amine-based chemical absorber/stripper configuration is chosen for removal of hydrogen sulfide and carbon dioxide. This configuration is chosen due to data availability, as compared to proprietary physical gas cleaning process such as Rectisol and Selexol.

Two fuel synthesis configurations under consideration produce liquid hydrocarbons: Fischer-Tropsch (FT) synthesis and MTG. FT synthesis has been proven in operation at commercial scale for many years by Sasol [11]. Because of more accessible data and long industrial experience, FT synthesis is the only fuel synthesis option chosen. Because of this selection, a post-synthesis fuel upgrading area is also necessary, as FT products need to be separated and hydroprocessed.

### Scenarios Not Selected

The indirect, dry-ash gasifier and the mixed alcohol synthesis configurations are not considered due to previous work by Phillips et al. [39]. 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. Tar conversion via external thermal or catalytic cracking is not considered due to lack of public domain data and commercial scale experience. Acid gas removal using proprietary technology (e.g., Rectisol or Selexol) is not considered because of a lack of public operational data. MTG, including methanol synthesis, is not considered because of time constraints and limited operational data. DME and syngas fermentation are not considered because of limited commercial scale experience and incompatibility with present fuel infrastructure.

### Project Assumptions

The main project assumptions for process and economic analysis are listed in Table 4**.** A more extensive list can be found in Appendix A. The process design is assumed to incorporate an "$n^{th}$ plant" level of implementation experience. This is a significant assumption for a process concept that has yet to undergo detailed engineering development and see its first commercial application. The design and operation of an $n^{th}$ plant is likely to diverge from the design and operation considered in this study. However, the $n^{th}$ plant evaluation is chosen to provide an analysis similar to studies completed by other groups, including NREL and Pacific Northwest National Laboratory.

**Table 4. Main Assumptions Used in $n^{th}$ Plant Scenarios**

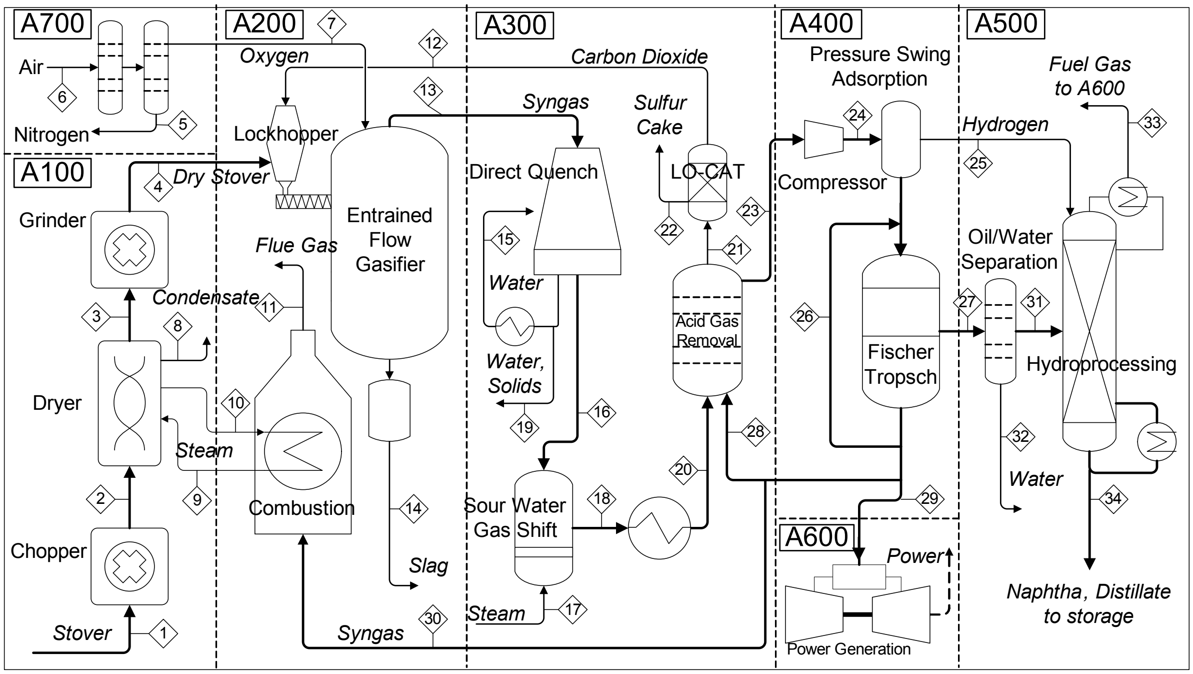
| Main Assumptions |
|---|
| The plant is modeled as $n^{th}$ plant |
| Plant capacity is 2,000 dry metric ton/day |
| Feedstock is corn stover at 25% moisture |
| Feedstock ash content at 6% |
| Feedstock is purchased at plant gate for $75/dry short ton |
| All financial values are adjusted to 2007 cost year |
| Plant is 100% equity financed |
| Fuel PV is evaluated at 10% internal rate of return |
| Plant initiates operation in 5–8 year time frame |
| Plant life is 20 years |
| Plant availability is 310 days per year (85%) |

## Process Description

### *High-Temperature Scenario Overview*

The high-temperature (HT) or slagging scenario is a 2,000 dry metric ton (2,205 dry short ton) per day corn stover-fed gasification biorefinery that produces naphtha-range and distillate-range liquid fractions to be used as blend stock for gasoline and diesel, respectively, as well as electricity for export. It is based on pressurized, oxygen-blown, entrained-flow gasification. The HT scenario is an $n^{th}$ plant design, meaning significant design, engineering, and operating experience has been achieved.

Figure 9 shows seven main areas of operation for the HT scenario. Feedstock preprocessing (Area 100) is where the stover is chopped, dried, and ground to 1-mm size and 10% moisture. Gasification (Area 200) contains the stover pressurization for solids feeding, gasification, and slag removal. Synthesis gas cleaning (Area 300) contains cold gas cleaning technologies where the syngas is quenched and scrubbed of particulates, ammonia, hydrogen sulfide, and carbon dioxide. Area 300 also contains the water-gas-shift reaction, which occurs before the hydrogen sulfide and carbon dioxide removal. This adjusts the ratio of hydrogen to carbon monoxide for optimal fuel synthesis. Fuel synthesis (Area 400) contains syngas boost pressurization, contaminant polishing via zinc oxide guard beds, Fischer-Tropsch reactor, and hydrocarbon gas/liquid separation. Hydroprocessing (Area 500) produces the final fuel blend and is treated as a black box utilizing published data. Power generation (Area 600) contains gas and steam turbines along with a heat recovery steam generator. Area 700 contains the air separation unit (ASU) where oxygen is separated from air and pressurized for use in the gasifier. For cost analysis uses only, a balance-of-plant area includes the cooling tower area, cooling water system, waste solids and liquids handling area, and feed water system. Detailed process flow diagrams can be found in Appendix D and detailed stream data can be found in Appendix E.



| Stream | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| T (°C) | 25 | 25 | 90 | 90 | 16 | 32 | 149 | 120 | 200 | 120 | 220 | 249 | 1300 | 50 | 203 | 203 | 190 |
| P (bar) | 1.01 | 1.01 | 1.01 | 1.01 | 1.1 | 1.01 | 28 | 1.98 | 1.98 | 1.98 | 1.03 | 28 | 26.62 | 26.62 | 26.62 | 25.93 | 10 |
| m (Mg/day) | 2667 | 2667 | 2222 | 2222 | 2189 | 2903 | 744 | 444 | 4000 | 4000 | 271 | 180 | 3823 | 114 | 4000 | 3867 | 550 |
| Stream | 18 | 19 | 20 | 21 | 22 | 23 | 24 | 25 | 26 | 27 | 28 | 29 | 30 | 31 | 32 | 33 | 34 |
| T (°C) | 240 | 40 | 40 | 50 | 50 | 62 | 76 | 30 | 200 | 42 | 45 | 45 | 45 | 35 | 35 | 35 | 37 |
| P (bar) | 24.8 | 24.82 | 24.82 | 3.45 | 1.93 | 22.75 | 26 | 1.01 | 24.96 | 23.6 | 23.58 | 23.58 | 23.58 | 22.2 | 22.2 | 22.2 | 1.03 |
| m (Mg/day) | 4417 | 1502 | 2965 | 1755 | 3.1 | 3570 | 3570 | 4.38 | 1404 | 1068 | 2308 | 156 | 33 | 427 | 641 | 52 | 378 |

**Figure 9. Overall process flow diagram for HT scenario (parallelograms enclosing numbers in the diagram designate individual process streams, which are detailed in the accompanying table)**

Recycle streams are utilized to provide better syngas-to-FT-products conversion. Unconverted syngas in the fuel synthesis area is recycled to the syngas cleaning area to remove carbon dioxide and allow for further conversion in the Fischer-Tropsch reactor. A small portion of unconverted syngas is sent to a steam boiler to raise steam required for drying the biomass. The balance of unconverted syngas is combusted in a gas turbine and waste heat is recovered in a steam generator for steam turbine power. Power generated is used throughout the plant and excess is sold.

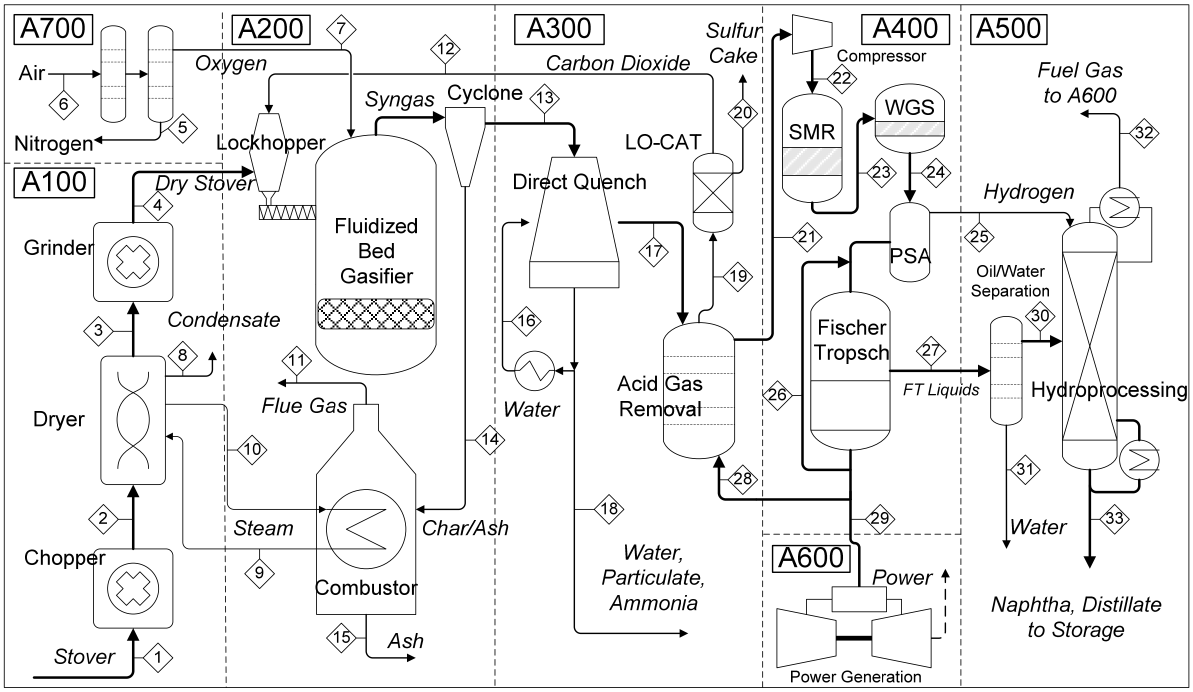
Some of the largest consumers of power are the ASU and hydroprocessing area at 11.6 MW and 2.2 MW, respectively. Another consumer of power is the hammer mill for grinding the dried biomass in Area 100; it requires 3.0 MW. The amine/water solution recirculation pump in Area 300 requires approximately 0.9 MW. Syngas compressors throughout the plant require a significant amount of power as well. Gross plant power production is 48.6 MW and net electricity for export is 13.8 MW.

## *Low-Temperature Scenario Overview*

The low-temperature (LT) or dry-ash scenario is a 2,000 dry metric ton (2,205 dry short ton) per day corn stover-fed gasification biorefinery that produces naphtha-range and distillate-range liquid fractions to be used as blend stock for gasoline and diesel, respectively, as well as electricity for export. It is based on a pressurized, oxygen/steam-blown fluidized-bed gasifier developed by Gas Technology Institute. The LT scenario is an $n^{th}$ plant design, meaning significant design, engineering, and operating experience has been achieved.

Figure 10 shows the seven main areas of operation for the LT scenario. Feedstock preprocessing (Area 100) is where the stover is chopped, dried, and ground to 6-mm size and 10% moisture. Gasification (Area 200) contains the stover pressurization for solids feeding, gasification, and char and ash removal. Synthesis gas cleaning (Area 300) contains cold gas cleaning technologies where the syngas is quenched and scrubbed of particulates, ammonia, hydrogen sulfide, and carbon dioxide. Fuel synthesis (Area 400) contains syngas boost pressurization, contaminant polishing via zinc oxide beds, Fischer-Tropsch reactor, and hydrocarbon gas/liquid separation. Also included within Area 400 is the steam methane reformer (SMR) to reduce methane content and water-gas-shift (WGS) to adjust the ratio of hydrogen and carbon monoxide.

Hydroprocessing (Area 500) produces the final fuel blend and is treated as a black box utilizing published data. Power generation (Area 600) contains gas and steam turbines along with a heat recovery steam generator. Area 700 contains the ASU that separates oxygen from air and pressurizes it for use in the gasifier. Detailed process flow diagrams can be found in Appendix D and detailed stream data can be found in Appendix E.



| Stream | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| T (°C) | 25 | 25 | 90 | 90 | 16 | 32 | 149 | 120 | 200 | 120 | 200 | 243 | 870 | 870 | 1500 | 40 | 40 |
| P (bar) | 1.01 | 1.01 | 1.01 | 1.01 | 1.10 | 1.01 | 28.00 | 1.98 | 1.98 | 1.98 | 1.00 | 28.00 | 27.55 | 27.55 | 1.00 | 26.17 | 26.17 |
| m (Mg/day) | 2667 | 2667 | 2222 | 2222 | 1744 | 2313 | 569 | 444 | 4000 | 4000 | 1471 | 180 | 2930 | 215 | 119 | 6000 | 2198 |
| Stream | 18 | 19 | 20 | 21 | 22 | 23 | 24 | 25 | 26 | 27 | 28 | 29 | 30 | 31 | 32 | 33 | |
| T (°C) | 40 | 50 | 50 | 62 | 78 | 870 | 304 | 60 | 200 | 32 | 32 | 32 | 35 | 35 | 50 | 50 | |
| P (bar) | 26.17 | 3.45 | 2.10 | 22.89 | 26.50 | 25.81 | 25.12 | 1.01 | 25.00 | 22.89 | 22.89 | 22.89 | 22.89 | 22.89 | 22.89 | 22.89 | |
| m (Mg/day) | 1380 | 1593 | 2.3 | 2709 | 2709 | 3707 | 3708 | 3.4 | 560 | 1463 | 2073 | 168 | 331 | 1132 | 41 | 293 | |

**Figure 10. Overall process flow diagram for LT scenario (parallelograms enclosing numbers in the diagram designate individual process streams, which are detailed in the accompanying table)**

Recycle streams are utilized to provide better FT products conversion. Unconverted syngas in the fuel synthesis area is recycled to the syngas cleaning area to remove carbon dioxide and allow for further conversion in the Fischer-Tropsch reactor. The balance of unconverted syngas is combusted in a gas turbine and waste heat is recovered in a steam generator for steam turbine power. Power generated is used throughout the plant and excess is sold. Unconverted carbon within the gasifier in the form of char is collected and combusted in a furnace to produce heat, thereby generating steam for the drying of the biomass.

Some of the largest consumers of power are the ASU and hydroprocessing area at 9.1 MW and 1.7 MW, respectively. Another consumer of power is the hammer mill for grinding the dried biomass in Area 100; it requires 1.1 MW. The amine/water solution recirculation pump in Area 300 requires approximately 0.7 MW. Syngas compressors throughout the plant require a significant amount of power as well. Gross plant power production is 40.7 MW and net electricity for export is 16.3 MW.

### *Area 100 Preprocessing*

The preprocessing area contains all the unit operations required for preparing the biomass for feeding into the gasifier. Biomass enters the plant gate at 25 wt% moisture on wet basis in bales. The corn stover composition is shown below in Table 5. Ash content is assumed to be 6% by weight. The composition of char formed in the gasifier is also shown in Table 5. Forklifts transport the bales to conveyors where the stover is separated from any metal in a magnetic separator. The first modeled operational area is a primary biomass chopper to complete the initial size reduction step and prepare stover for drying.

**Table 5. Stover and Char Elemental Composition (wt %)**

| Element | Stover | Char |
|---|---|---|
| Ash | 6.00 | 0 |
| Carbon | 47.28 | 68.05 |
| Hydrogen | 5.06 | 3.16 |
| Nitrogen | 0.80 | 0.29 |
| Chlorine | 0 | 0 |
| Sulfur | 0.22 | 0.15 |
| Oxygen | 40.63 | 28.34 |

The next area of operation is the direct-contact steam drying, which is modeled as a rotary steam dryer with exiting biomass moisture of 10% on wet basis. For steam dryers, Amos [21] suggests a 9:1 ratio of steam to evaporated moisture. Therefore, 4,000 metric tpd steam is utilized in a loop and heated to 200°C from the hot combustion flue gases exiting the syngas- or char-fired combustor in Area 200. Steam mixes with 25°C biomass and enters the dryer. At the exit, steam at 120°C returns to the combustor for reheating, and dried biomass exits at 90°C and is conveyed to the grinding area.

The grinding area has the same configuration as the chopping area except that the grinder requires significantly more power because of the larger size reduction. The grinder reduces the size of the biomass to 1-mm and 6-mm particles for the HT and LT scenarios, respectively. The power requirements of the grinder for the HT and LT scenarios are 3,000 kW and 1,100 kW, respectively. Energy requirements for grinding are determined using the correlations for specific energy (kWh per short ton), which is adapted from Mani et al. [42].

### *Area 200 Gasification*

The gasification area of the plant produces synthesis gas using pressurized gasifiers. Also in this area, slag, char, and ash are removed. This area also includes lockhoppers for biomass pressurization and a fired combustor that provides heat to raise steam for drying the stover.

Dried and ground stover enters the area and is immediately conveyed to a lockhopper system for pressurized feeding. Carbon dioxide from the syngas cleaning area is used as pressurization gas. According to Lau et al. [41], a lockhopper system is the best setup for pressurized feeding of solids, despite higher operating costs due to high inert gas usage. A proven track record with biomass is the main reason for their recommendation. The power requirement of a lockhopper system using biomass is 0.082 kW/metric tpd, resulting in a 180-kW system. Higman and van der Burgt [43] report inert gas usage as 0.09 kg/kg for 25-bar applications. This results in a 180 metric ton per day carbon dioxide addition into the hopper. It is assumed that only 5% of the inert gas leaks into the gasifier while the rest is vented by the lockhopper.

Pressurized biomass is then conveyed into the gasifier. Oxygen at 95% purity is produced from the air separation unit. A fixed 0.35 mass ratio of oxygen to biomass is used for the entrained-flow gasifier as reported by Henrich [18]. Steam addition to the gasifier is set at 0.48 mass ratio of steam to biomass in accordance with Probstein and Hicks [11] and explained further in Appendix C. This gasifier operates at a temperature of 1,300°C, meaning that equilibrium can be modeled according to Frey and Akunuri [3]. The reactions shown in equations 3–9 are modeled using equilibrium constants.

$$C + 2H_2 \leftrightarrow CH_4 \qquad \text{(eqn. 3)}$$

$$2C + 1.5O_2 \leftrightarrow CO + CO_2 \qquad \text{(eqn. 4)}$$

$$CO + H_2O \leftrightarrow CO_2 + H_2 \qquad \text{(eqn. 5)}$$

$$2CO + O_2 \leftrightarrow 2CO_2 \qquad \text{(eqn. 6)}$$

$$S + H_2 \leftrightarrow H_2S \qquad \text{(eqn. 7)}$$

$$0.5N_2 + 1.5H_2 \leftrightarrow NH_3 \qquad \text{(eqn. 8)}$$

$$CO + H_2S \leftrightarrow COS + H_2 \qquad \text{(eqn. 9)}$$

The LT scenario gasifier uses a 0.26 mass ratio of oxygen to biomass at a gasification temperature of 870°C. This ratio is developed from the data found in an IGT gasifier study by Bain [4]. In that study, Bain develops mass balances for an IGT gasifier operating with woody biomass. Steam addition to the gasifier is calculated using a 40/60 steam to oxygen mass ratio, consistent with experiments performed at Iowa State University using corn stover feedstock and a steam/oxygen-blown fluidized-bed gasifier. Low-temperature gasification cannot be modeled at equilibrium with or without approach temperatures for reactions. Instead, an elemental mass balance calculation and adjustment is performed to ensure all inlet and outlet streams are accounted for across the gasifier. For details on the LT gasifier mass balance calculation, see Appendix C.

Yield from each gasifier is different. As Table 6 shows, hydrocarbons and tars are not produced in the high-temperature gasifier because of near equilibrium conditions. Also, more hydrogen formation occurs in the high-temperature gasifier as a result of the water-gas-shift reaction (equation 5) and because thermodynamically, nearly no methane, ethane, and ethylene are produced. The low-temperature gasifier, on the other hand, produces a significant amount of methane, ethane, and ethylene in the syngas, requiring downstream reforming. In the HT scenario, slag forms from the ash when the ash melts and flows on the inside walls. It is then collected at the bottom and removed for storage and subsequent waste removal. In accordance with Frey and Akunuri [3], it is assumed that 95% of the ash in the stover becomes slag while the rest becomes fly ash.

**Table 6. Syngas Composition (Mole Basis) Leaving Gasifier for Gasification Scenarios Evaluated**

| Component | High Temperature (mole fraction) | Low Temperature (mole fraction) |
|---|---|---|
| Carbon Monoxide | 0.264 | 0.240 |
| Hydrogen | 0.310 | 0.200 |
| Carbon Dioxide | 0.137 | 0.274 |
| Water | 0.280 | 0.194 |
| Nitrogen | 0.002 | 0 |
| Methane | $6\times10^{-6}$ | 0.055 |
| Ethane | 0 | 0.0061 |
| Ethylene | 0 | 0.013 |
| Ammonia | $3.1\times10^{-5}$ | 0.0094 |
| Hydrogen Sulfide | $6.72\times10^{-4}$ | 0.00112 |
| Carbonyl Sulfide | $2.6\times10^{-5}$ | 0 |
| Tar (Anthracene) | 0 | $5.00\times10^{-4}$ |
| Oxygen | 0 | 0 |
| Argon | 0.006 | 0.006 |

Directly after the low-temperature gasifier, initial syngas cleaning by cyclones captures char and ash. The cyclones are split into two trains because of high volumetric gas flow. Each train contains a medium efficiency cyclone, followed by high efficiency cyclones to capture particulates. Overall particulate removal efficiency for the cyclone area is 99%. Nearly particulate-free syngas travels to the more rigorous syngas cleaning in Area 300. Captured char in the LT scenario is collected and combusted in a fluidized-bed combustor providing energy for heating low-pressure steam, which is used for drying the stover. Syngas produced in the HT scenario contains fly ash, which is subsequently removed in a direct-water-quench unit. The combustion area in the HT scenario receives unconverted syngas from the fuel synthesis area, as char is not produced. For both scenarios, the combustor is assumed to operate adiabatically, resulting in an exit flue gas temperature of approximately 1,800°C. Hot flue gas heats 120°C steam to 200°C and loops to the stover drying area.

### *Area 300 Syngas Cleaning*

After the initial particulate removal accomplished by the cyclones, the syngas still contains some particulates and all of the ammonia, hydrogen sulfide, and other contaminants. Area 300 removes these species using a cold-gas-cleaning approach, which is presently proven in many commercial configurations. Hydrogen sulfide and carbon dioxide, collectively known as acid gas, are absorbed via amine scrubbing. The LO-CAT hydrogen sulfide oxidation process is used to separate carbon dioxide from hydrogen sulfide with subsequent recovery of solid sulfur. In addition, the HT scenario contains a sour water-gas-shift (SWGS) process (sour because of the presence of sulfur), whereas the LT scenario situates the water-gas-shift (WGS) process directly upstream from the Fischer-Tropsch reactor.

Because of a less than optimal hydrogen to carbon monoxide ratio from the gasifier, a WGS reaction is necessary at some point in the process to adjust to the optimum Fischer-Tropsch ratio of 2.1. Therefore, a significant WGS activity is required, so a sizable amount of carbon dioxide is produced. To keep that carbon dioxide from building up in downstream processes, the SWGS reactor is located before the acid gas removal area. This SWGS unit operation is the most significant difference between the HT and LT scenarios in this area.

In the HT scenario, the syngas arriving from the gasifier is cooled by direct-contact water quench to the operating temperature of the SWGS unit. In addition to cooling, the direct water quench removes all of the fly ash, sludge, and black water to prevent downstream plugging. At this point, a portion of the syngas is diverted to the SWGS unit, which is modeled at equilibrium conditions and has an exit gas temperature of 300°C. A ratio of 3:1 water to carbon monoxide is reached by addition of steam to the SWGS reactor. After the syngas is combined, the gas is further cooled to prepare for the acid gas removal. In the LT scenario, the direct quench unit condenses the syngas, removing approximately 90% of ammonia and 99% of solids. Tar is condensed in this unit and can be recycled back into the gasifier using a slurry pump, but this configuration is not modeled. A water treatment facility for the direct-quench effluent is not modeled but is accounted for in a balance-of-plant cost.

The next step for cleanup is the removal of acid gas (carbon dioxide and hydrogen sulfide) through the use of an amine-based solvent in a chemical gas absorption system. At this point in the cleaning process, hydrogen sulfide and carbon dioxide content is approximately 900 ppm and 30% on a molar basis, respectively. Sulfur must be removed to at least 0.2 ppm for Fischer-Tropsch synthesis [31]. According to the Gas Processors Suppliers Association *Engineering Data Book* [44], amine-based systems are only capable of removing sulfur down to 4 ppm. Therefore, a zinc oxide guard bed is required to remove the difference. In this study, 20% concentrated monoethanolamine (MEA), capable of absorbing 0.4 mole acid gas per mole amine, is used as the absorbent. The process setup is based on a report by Nexant Inc. [27]. Hydrogen sulfide leaves the top of the absorber at 4 ppm and $CO_2$ leaves at 2%, which is 99% and 90% removal, respectively. The clean syngas is now ready for polishing to final cleanliness requirements. A stripper is utilized to desorb the acid gas and regenerate the amine solution. Before the acid gas and amine solution enters the stripper, a heat exchanger raises the temperature to 90°C.

Acid gas is brought to the LO-CAT sulfur recovery system to isolate hydrogen sulfide and convert it to solid sulfur. The LO-CAT system sold and owned by Gas Technology Products uses oxygen and a liquid solution of ferric iron to oxidize hydrogen sulfide to elemental solid sulfur [45]. This system is suitable for a range of 150 lb to 20 tons per day sulfur recovery and 100 ppm to 10% $H_2S$ concentration in sour gas, as reported by Nexant Inc. [27]. The sulfur production in this model is approximately 3 metric tpd with $H_2S$ concentration of approximately 150 ppm, which is within the reported ranges. First, the $H_2S$ is absorbed/oxidized, forming solid sulfur and water, while the ferric iron converts to ferrous iron. The second vessel oxidizes the ferrous iron back to ferric iron and the sulfur cake is removed while the iron solution is recycled back into the absorber [46]. The carbon dioxide gas stream from the absorber is split, and a portion is compressed and used in biomass pressurization while the rest is vented to the atmosphere.

### *Area 400 Fuel Synthesis*

Conversion from syngas to liquid fuel occurs in Area 400, the fuel synthesis area. The major operations in this area are zinc oxide/activated carbon gas polishing, steam methane reforming (SMR) (only in the LT scenario), water-gas-shift (only in the LT scenario), Fischer-Tropsch (FT) synthesis, hydrogen separation via pressure swing adsorption (PSA), FT products separation, and unconverted syngas distribution. Another major difference between the LT and HT scenarios is in this area. Area 400 in the LT scenario contains the water-gas-shift reaction and steam methane reformer, as recycle streams contain a high enough content of methane and ethylene to significantly accumulate and cause dilution.

A compressor is the first operation in Area 400, boosting the pressure to 25 bar for FT synthesis. Then the syngas is heated to 200°C and passes through zinc oxide/activated carbon fixed-bed sorbent. This polishing guard bed acts as a barrier to any upstream non-normal contaminant concentrations and provides sulfur removal down to synthesis requirements. To limit downstream catalyst poisoning, the syngas stream must be cleaned of these components. Removal to 50 ppb sulfur is possible with zinc oxide sorbent [27]. To comply with reported requirements, the sorbent removes sulfur to approximately 200 ppb. In addition to sulfur, halides are removed by the sorbent. Syngas contaminant level requirements for Fischer-Tropsch synthesis are shown in Table 7.

**Table 7. Fischer-Tropsch Gas Cleanliness Requirements [31]**

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

Methane, nitrogen, and carbon dioxide act as inert gases in the FT synthesis. At this point in the LT scenario, an SMR step is utilized. Syngas is heated to 870°C through a fired heater and passed through a reformer nickel-based catalyst to reduce methane, ethylene, and ethane content. It is assumed that the SMR can be modeled to operate at equilibrium. Steam is added to bring the steam to methane ratio to approximately 6.0, which at 870°C and 26 bar results in about 1.5% equilibrium methane content in the exit stream [47]. For the HT scenario, the SMR step is not necessary. The WGS reaction is now employed for the LT scenario to increase the $H_2$:CO ratio. A portion of the gas is diverted through the fixed catalyst bed while the rest bypasses the reactor, similar to the SWGS unit in the HT scenario.

The exiting $H_2$:CO ratio after WGS is slightly above 2.1 in order for the excess hydrogen to be separated and used in the hydroprocessing area. A PSA process is employed to isolate a stream of hydrogen. Because only a small amount of hydrogen needs to be separated from the syngas stream for downstream use, a small percentage of the syngas is directed to the PSA unit. Hydrogen removal efficiency within the PSA unit is assumed to be 85%, and the unit produces pure hydrogen [41]. After the PSA, the syngas rejoins the main gas line and enters the FT reactor.

The Fischer-Tropsch synthesis reactor operates at 200°C and 25 bar using a cobalt catalyst according to equation 10. Per-pass carbon monoxide conversion in the reactor is set at 40%. The product distribution follows the Anderson-Schulz-Flory alpha distribution where chain growth factor, $\alpha$, depends on partial pressures of $H_2$ and CO and the temperature of the reactor, as reported by Song et al. [48] for cobalt catalyst and shown in equation 11, where $y$ is the molar fraction of carbon monoxide or hydrogen and $Temp$ is the reactor operating temperature in kelvin. The reactor is based on a fixed-bed type reactor and that choice is reflected by the low per-pass CO conversion.

$$CO + 2.1H_2 \rightarrow\ -(CH_2)-\ + H_2O \qquad \text{(eqn. 10)}$$

$$\alpha = \left[0.2332 \cdot \frac{y_{CO}}{y_{CO} + y_{H2}} + 0.6330\right] \cdot [1 - 0.0039(Temp - 533)] \qquad \text{(eqn. 11)}$$

To ensure that the hydrocarbon product distribution leans toward the production of diesel fuel, the value of alpha should be at least 0.85 and preferably greater than 0.9, as shown in Figure 11. The reactor operating temperature needed to achieve a chain growth value of 0.9 is approximately 200°C. This produces 30 wt% wax in the FT products, requiring hydrocracking before addition to the final fuel blend. All exiting effluent is cooled to 35°C and the liquid water and hydrocarbons are separated in a gas/liquid knock-out separator. Unconverted syngas is split into four streams: direct recycle to the FT reactor, recycle to the acid gas removal area, purge to the combustor in Area 200, and to the gas turbine in the power generation area. The LT scenario does not include sending a syngas stream to the combustor in Area 200 because char is used. Overall CO conversion is 66% due to recycling syngas. The recycle ratio is approximately 1.95 for both scenarios.

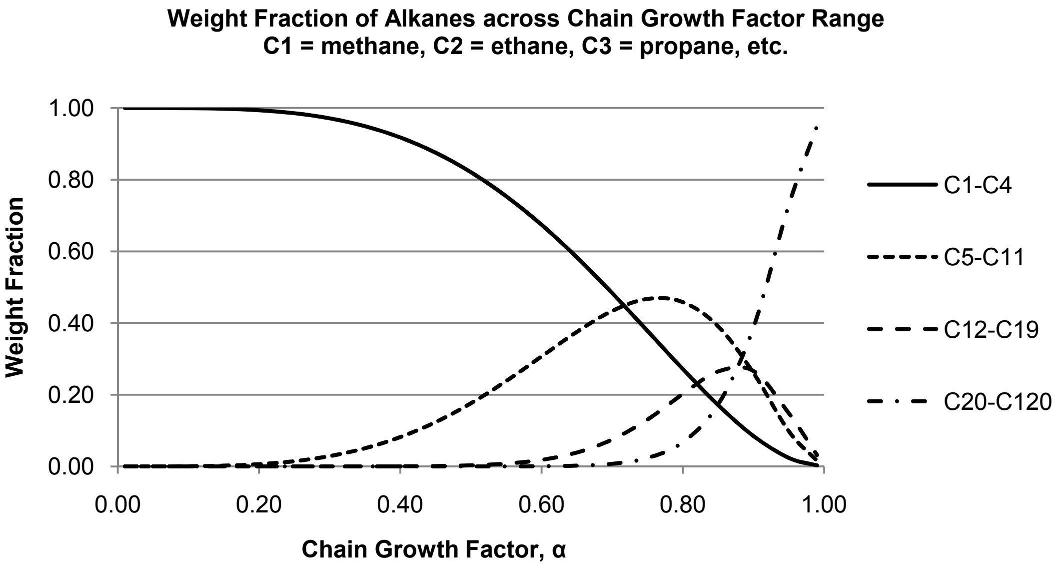


**Figure 11. Fischer-Tropsch product distribution as a function of chain growth factor ($\alpha$) using equation 11 [48]**

### *Area 500 Hydroprocessing*

FT products from the fuel synthesis area contain significant amounts of high-molecular-weight wax. Hydrogen is required to crack these high-molecular-weight parrafins to low-molecular-weight hydrocarbons. A product distribution is specified in Table 8, as detailed in Shah et al. [49]. It is assumed that the hydroprocessing area contains a hydrocracker for converting the wax fraction and a distillation section for separating naphtha, diesel, and lighter-molecular-weight hydrocarbons. Also, hydrogen is assumed to be recycled within this area as needed. Methane and propane are separated and used to fuel the gas turbine in the power generation area. The hydroprocessing area is modeled as a "black box."

**Table 8. Hydroprocessing Product Distribution [49]**

| Component | Mass Fraction |
|---|---|
| Methane | 0.0346 |
| LPG (propane) | 0.0877 |
| Gasoline (octane) | 0.2610 |
| Diesel (hexadecane) | 0.6167 |

### *Area 600 Power Generation*

A gas turbine and steam turbine provide the means to produce the power that is required throughout the plant and the excess power that is generated for export. Unconverted syngas from Fisher-Tropsch synthesis and fuel gas from hydroprocessing are combusted in a gas turbine, producing hot flue gas and shaft work. The flue gas exchanges heat with water in a heat-recovery steam generator to produce steam for the steam turbines, which subsequently produce more shaft work. Electric generators attached to both the gas turbine and the steam turbine produce electricity from the shaft work.

### *Area 700 Air Separation*

Because 95% purity oxygen is used for both scenarios, a cryogenic ASU is employed rather than purchasing oxygen. A two-column cryogenic oxygen/nitrogen separation system is employed with subsequent oxygen compression and nitrogen vent. Air pre-cooling is accomplished by exchanging heat with exiting nitrogen. This area requires a significant amount of power, as explained in the results section, which is provided by the power generation area.

## Methodology for Economic Analysis

Capital investment and PV of each scenario are determined by finding all equipment costs and operating costs for the construction and operation of a plant for 20 years. Total capital investment is based on the total equipment cost plus additional installation costs and indirect costs (such as engineering, construction, and contingency costs). Annual operating costs are determined and a discounted cash flow rate of return analysis is developed. PV per unit volume of fuel is determined at a net present value of zero and a 10% internal rate of return. The major economic assumptions used in this analysis are listed in Table 9. A detailed list of assumptions can be found in Appendix A.

**Table 9. Main Economic Assumptions for n$^{th}$ Plant Scenarios**

| Parameter | Assumption |
|---|---|
| Financing | 100% equity |
| Internal rate of return (after taxes) | 10% |
| General plant depreciation period | 7 years (all areas except Area 600) |
| Steam plant depreciation period | 20 years (Area 600 only) |
| Construction period | 2.5 years with total capital investment spent at 8%, 60%, and 32% per year during years before operation |
| Startup time | 0.5 years, where during that time revenues, variable operating costs, and fixed operating costs are 50%, 75%, and 100% of normal, respectively. |
| Income tax rate | 39% |
| Contingency | 20% of fixed capital investment |
| Electricity cost | 5.4 cents/kWh |
| Working capital | 15% of fixed capital investment |
| Land purchase | 6% of total purchased equipment cost |
| Plant availability | 310 days per year (85%) |

Unit operations from the scenarios are sized and costs estimated using Aspen Icarus Process Evaluator software based on the Aspen Plus simulation data. Unique equipment costs for such equipment as the gasifier and Fischer-Tropsch synthesis reactor are estimated externally using literature references. Additionally, some equipment, such as the biomass dryer and lockhoppers, require literature references to determine the sizing, whereby their costs are subsequently estimated using Aspen Icarus. The hydroprocessing plant area is modeled as a “black box” and therefore its costs are estimated as an overall scaled area cost from literature.

The costs of each equipment item and process area are scaled based on a scaling stream and scaling size factor ($n$) according to equation 12, where the size factor is between 0.6 and 1.0, depending on the equipment type.

$$Cost_{new} = Cost_0 * \left[\frac{Stream\ Size_{new}}{Stream\ Size_0}\right]^n \qquad \text{(eqn. 12)}$$

All purchased equipment costs determined via Aspen Icarus contain an installation factor that accounts for piping, electrical, and other costs required for installation. However, this installation factor tends to be significantly lower than metrics suggested by Peters et al. [50]. Therefore, rather than using the software-derived installation factors, an overall installation factor is applied to most equipment. A 3.02 overall installation factor is used, as suggested by Peters et al. for solid-liquid plants. Basically, the purchased equipment cost of a piece of equipment is multiplied by the installation cost to determine its installed cost. For the gasification unit, a 2.35 installation factor is used according to a National Energy Technology Laboratory study by Reed et al. [51]. It is assumed that all gas compressors receive a 1.2 installation factor, which is consistent with Aspen Icarus. The Chemical Engineering Plant Cost Index is used to bring the cost to 2007 U.S. dollars whenever a source for an estimated cost is from a previous year [52]. For multiple unit operations that operate in parallel or in trains, a train cost factor is applied. The reason for the factor, as reported by Larson et al. [38], is because those units share some piping, electrical, and other installation costs. It is applied as shown in equation 13, where $n$ is the number of units in the train and $m$ is the train factor with a value of 0.9.

$$Cost_{train} = Cost_{unit} * n^m \qquad \text{(eqn. 13)}$$

Table 10 explains the methodology used to estimate capital investment. After total purchased equipment cost (TPEC) and total installed cost (TIC) are determined, indirect costs are applied. Indirect costs (IC) include engineering and supervision, construction expenses, and legal and contractors’ fees at 32%, 34%, and 23% of TPEC, respectively [50]. Project contingency is added as 20% of total direct and indirect cost (TDIC). TDIC is set as the sum of IC and total installed costs (TIC). Adding project contingency to TDIC gives the fixed capital investment. Total capital investment (TCI) is determined by adding working capital to fixed capital investment and thereby represents the overall investment required for each scenario.

**Table 10. Methodology for Capital Cost Estimation for $n^{th}$ Plant Scenarios**

| Parameter | Method |
|---|---|
| Total Purchased Equipment Cost (TPEC) | Aspen Icarus Process Evaluator, references |
| Total Installed Cost (TIC) | TPEC Installation Factor |
| Indirect Cost (IC) | 89% of TPEC[a] |
| Total Direct and Indirect Costs (TDIC) | TIC + IC |
| Contingency | 20% of TDIC |
| Fixed Capital Investment (FCI) | TDIC + Contingency |
| Working Capital (WC) | 15% of FCI |
| Total Capital Investment | FCI + WC |

[a] Indirect costs are broken down into engineering and supervision, construction expenses, and legal and contractors' fees at 32%, 34%, and 23%, respectively, for a total of 89% of TPEC.

Raw material costs are inflated to 2007 U.S. dollars using the Industrial Inorganic Chemical Index also used by Phillips et al. [39]. Annual variable operating costs are determined from material stream flows. Variable operating costs and their respective cost methods are shown in Table 11. Natural gas for use in the gas turbine to produce power during startup and backup periods is assumed to be employed 5% of the annual operating time. Solids disposal costs are for the handling and removal of ash in the LT scenario and slag in the HT scenario. Wastewater disposal cost is applied to the sludge and black water produced during direct syngas quench. Catalyst costs are not calculated on an annual basis because the catalysts for all reactors are assumed to be replaced every 3 years. Instead they are accounted for in the discounted cash flow analysis.

**Table 11. Variable Operating Cost Parameters Adjusted to $2007**

| Variable Operating Costs | Cost information |
|---|---|
| Feedstock | \$75/dry short ton |
| LO-CAT Chemicals | \$176/metric ton of sulfur produced as reported in Phillips et al. [39] |
| Amine Makeup | \$1.09/lb as reported in Phillips et al. and set as 0.01% of the circulating rate [39] |
| Process Steam | \$8.20/ton (Peters et al.) [50] |
| Cooling Water | \$0.31/ton (Peters et al.) |
| Hydroprocessing | \$4.00/barrel produced as reported by Robinson and Dolbear [53] |
| Natural Gas (for backup) | \$6.40/thousand standard cubic feet as the average wellhead price for 2007 [54] |
| Ash/Char Disposal | \$23.52/ton [39] |
| Wastewater Disposal | \$3.30/hundred cubic feet [39] |
| Electricity | \$0.054/kWh[a] |
| Sulfur | \$40.00/ton [39] |
| Fischer-Tropsch Catalyst (cobalt) | \$15/lb and 64 lb/ft$^3$ density; applied on first operation year and then every three years[a] |
| Water-Gas-Shift Catalyst (copper-zinc) | \$8/lb and 900 kg/m$^3$; applied on first operation year and then every three years. Sour shift and normal WGS are assumed to operate with same catalyst[a] |
| Steam Methane Reforming Catalyst (nickel-aluminum) | \$15/lb and 70 lb/ft$^3$; applied on first operation year and then every three years[a] |
| Pressure Swing Adsorption | \$2/lb[a] |

[a] Assumed.

Fixed operating costs include employee salaries, overhead and maintenance, and insurance and taxes. Salaries are calculated similarly to Phillips et al. [39], where employees include a plant manager, shift supervisors, lab technicians, maintenance technicians, shift operators, yard workers, and office clerks. The labor index developed by the Bureau of Labor Statistics [55] is used to adjust the labor cost to $2007. Overhead is calculated as 60% of total salaries; maintenance cost and taxes/insurance cost are both 2% of total installed equipment cost in accordance with Aden et al. [56].

For the DCFROR analysis, the capital investment is spent over a 2.5 year construction period, with 8% in the first half-year, followed by 60% and 32% for the second and third years. Working capital is applied in the year before operation and recovered at the end of the plant life. A standard modified accelerated cost recovery system (MACRS) is used, with the steam plant depreciating over 20 years and the rest of the plant over a 7-year period consistent with IRS allowances. The project life is 20 years. Plant availability of 310 days per year (85%) is assumed and affects raw materials purchase as well as fuel production. The PV per gallon of gasoline equivalent is calculated for a set net present value of zero including a 10% internal rate of return.

### *Methodology for Major Equipment Costs*

The software used for determining equipment costs is not capable of estimating every unit in this study. Some units such as the gasifiers and Fischer-Tropsch reactors are unique pieces of equipment that are underestimated if estimated as a simple vertical pressure vessel. Therefore, literature sources are used to help estimate sizes and costs of many units. The following section details a few of the more important units.

The biomass dryer costs are estimated by determining the drying contact area. According to Couper [57], typical rotary dryers have a diameter of 6 feet and solids holdup of 8%. Assuming a bulk density of 100 kg/m$^3$ for ground stover and 1,000 kg/m$^3$ for moisture in the stover, the resulting total surface area required for drying is 1,880 m$^2$. The surface area provides enough information for estimating the costs because rotary dryer costs are estimated based on surface area in Aspen Icarus. Details on dryer sizing can be found in the Detailed Calculations section of Appendix C.

The lockhopper system sizes are estimated by referring to a U.S. Department of Energy report completed by Combustion Engineering, Inc. [58], in which residence times and operating pressures are given. The biomass receiving bin, lockhopper, and feed bin costs are then estimated with Aspen Icarus. Details on lockhopper sizing can be found in the Detailed Calculations section of Appendix C.

The high-temperature gasifier cost is estimated from Reed et al. [51]. The total bare erected cost (installed cost) of a train of eight high-temperature E-Gas gasifiers (2,500 metric tpd coal) including syngas cooling costs is \$638 million (\$2006). It is assumed that the syngas cooling accounts for 20% of that cost, and therefore the estimated installed cost in millions of \$2006 for a 2,000 metric tpd high-temperature gasifier follows the formula in equation 14, resulting in \$57 million installed cost.

$$Cost_{HTgasifier} = \frac{(638 \cdot 80\%)}{8} \cdot \left[\frac{2000MT}{2500MT}\right]^{0.7} \qquad \text{(eqn. 14)}$$

Installed cost for a fluidized-bed gasifier is described in Larson et al. [38] and is calculated as shown in equation 15, where $Cost_{0\_gasifier}$ is \$6.41 million (\$2003), $Stream\ Size_0$ is 41.7 metric ton per hour, and $n$ is 0.7. The gasifier is evaluated at 300 short tpd because that appears to be the highest proven capacity for a GTI gasifier. Therefore, seven fluidized-bed gasifiers are used in parallel. It is assumed that the gasifier train follows the train cost formula (equation 13), resulting in \$19 million installed cost.

$$Cost_{LTgasifier} = Cost_{0\_gasifier} * \left[\frac{Stream\ Size}{Stream\ Size_0}\right]^{n} \qquad \text{(eqn. 15)}$$

In a similar manner, the FT reactor is estimated as described in Larson et al. [38], where base installed cost is $10.5 million ($2003), base sizing value is 2.52 million scf/h of synthesis gas flow, and sizing exponent is 0.72. An installation factor of 3.6 is assumed for the FT reactor as found in Peters et al. [50] for liquid production plants. This allows the purchased cost of the unit to be back-calculated.

The acid gas removal (AGR) area cost is evaluated using information from Phillips et al. [39], following equation 12 where the base stream size is 4,000 short tpd and the base cost is $5.45 million. The stream size is the mass flow of the synthesis gas entering the AGR as the sum of fresh syngas from gas scrubbing and unconverted syngas from the fuel synthesis area.

Capital investment for the hydroprocessing area is found in Robinson et al. [53]. That study reports a volumetric unit cost of $4,000 per barrel per standard day. Assuming the typical hydroprocessing refinery produces 25,000 barrels per day, the base cost, $C_0$, is $100 million. Assuming a scaling exponent of 0.65, the cost of Area 500 is found using equation 12. The cost details for both of the gasifiers, the AGR area, the FT reactor, and the hydroprocessing area can be found in the Detailed Calculations section of Appendix C.

### *Methodology for Sensitivity Analysis*

Sensitivity parameters are chosen to reflect the change in PV. The parameters are either economic or process parameters. The sensitivity bounds are chosen based on what is expected to be observed in the construction and operation of a biomass-to-liquids production plant. The chosen favorable, baseline, and unfavorable sensitivity variables are shown in Table 12.

**Table 12. Sensitivity Parameters for n^th^ Plant Scenarios**

| Parameter | Favorable | Baseline | Unfavorable |
|---|---|---|---|
| Availability (hours/year) | 8,000 | 7,446 | 7,000 |
| Balance of Plant (% of TPEC)[a] | 8 | 12 | 16 |
| Catalyst Cost (%)[b] | 50 | 100 | 200 |
| Catalyst Lifetime (years) | 5 | 3 | 1 |
| CO Conversion in FT Reactor (%) | 30 | 40 | 50 |
| Compressor Install Factor | 1.0 | 1.2 | 3.0 |
| Contingency (% of TDIC)[c] | 10 | 20 | 30 |
| Feedstock Cost ($/dry short ton) | 50 | 75 | 100 |
| Feedstock Moisture (% wet) | 20 | 25 | 30 |
| Price of Electricity (¢/kWh) | 7.0 | 5.4 | 3.0 |
| Total Capital Investment (% of baseline) | 70 | 100 | 130 |

[a] TPEC=total purchased equipment cost.

[b] All catalyst costs are varied over this range.

[c] TDIC=total direct and indirect cost.

### *Methodology for Pioneer Plant Analysis*

Economic analysis is based on an nth plant design, and before a project is undertaken the pioneer (1st) plant cost is important to estimate. We use a method begun by the RAND Corporation that estimates pioneer plant costs and plant performance. Using this methodology, two main areas of the nth plant economic analysis are adjusted: capital investment and plant performance. Through a series of parameters, a cost inflation factor is generated to inflate the capital investment. In addition, a plant performance factor is calculated that reduces the fuel sales, feedstock purchase, and variable operating costs for the first several years that the plant is in operation. Each year, the plant performance factor is increased until full performance is attained. For the purpose of determining a range of pioneer plant costs, baseline, optimistic, and pessimistic values are chosen. The details of the RAND methodology can be found in Merrow et al. [59]. The following section explains the reasoning behind the parameters chosen for the scenarios.

Cost growth and plant performance factors are calculated as shown in equations 16 and 17 in accordance with Merrow et al. [59].

$$\begin{aligned} Cost\ Growth = 1.1219 - 0.00297 * PCTNEW - 0.02125 * IMPURITIES \\ - 0.01137 * COMPLEXITY + 0.00111 * INCLUSIVENESS \\ - 0.06351 * PROJECT\ DEFINITION \end{aligned} \qquad \text{(eqn. 16)}$$

$$\begin{aligned} Plant\ Perf. = 85.77 - 9.69 * NEWSTEPS + 0.33 * BALEQS - 4.12 \\ * WASTE - 17.91 * SOLIDS \end{aligned} \qquad \text{(eqn. 17)}$$

The variables used in determining pioneer plant performance (equation 17) are defined as follows:

> NEWSTEPS (0+): The feedstock handling area is chosen as a new step because of the large scale, which has not been demonstrated with biomass. The gasifier and solids feeding are also included as a new step because a pressurized biomass feeding system has not been demonstrated at a commercial scale except for in limited campaigns.
>
> BALEQS (0 to 100): The mass and energy balances cannot be validated with current plant data, so a value of zero is chosen.
>
> WASTE (0 to 5): Waste streams for gasification include scrubber sludge, black water, gasifier slag, fly ash, and sulfur. The scrubber sludge and black water require chemical treatment and the sulfur requires special handling. A mid-range value of 2.5 is chosen.
>
> SOLIDS (0 or 1): Solids are present; therefore a value of 1 is used.

Variables used in determining pioneer plant cost growth (equation 16) are defined as follows:

> PCTNEW (0 to 100%): The percentage cost of the gasifier, solids pressurizing, and solids feeding out of the total purchased equipment cost.

IMPURITIES (0 to 5): There are two major recycle streams in the gasification process, and there is the possibility of inert component buildup. There is also a potential for equipment corrosion due to sulfur components, hydrogen chloride, and hydrogen, so a value of 4 is assigned.

COMPLEXITY (0+): There are nine continuously linked steps in the gasification process. These include feedstock handling, solids feeding, gasification, amine scrubbing, sour water-gas-shift, pressure swing adsorption, Fischer-Tropsch synthesis, hydroprocessing, and air separation.

INCLUSIVENESS (0 to 100): Land costs and startup costs are considered in the TCI; however, they have not been rigorously investigated. A value of 33% is used.

PROJECT DEFINITION (2 to 8): The gasification platform is considered to be in the study design stage so a value of 7 is assigned.

The factors are applied to the capital investment and plant performance as shown in equations 18 and 19. Expenses and revenues affected by the plant performance factor are fuel sales, feedstock purchase, co-product credits, and variable operating costs.

$$TCI_{pioneer} = \frac{TCI_{nth}}{Cost\ Growth} \tag{eqn. 18}$$

The $Cost\ Growth$ factor causes the TCI of the pioneer plant to increase from the n$^{th}$ plant cost.

$$Cost_{pioneer}(t) = \ Cost_{nth}(t) * \frac{(Plant\ Perf. + 20 * (t-1))}{100} \tag{eqn. 19}$$

$Cost_{nth}(t)$ is the n$^{th}$ plant expense or revenue at year $t$. The plant performance factor is applied at year 1 and increases by 20% each year until 100% performance is reached. The chosen parameters and calculated factors for baseline, optimistic, and pessimistic scenarios are shown in Table 13.

**Table 13. Pioneer Plant Analysis Parameters and Factors**

| **Parameter** | **Baseline** | **Optimistic** | **Pessimistic** | **Range** |
|---|---|---|---|---|
| NEWSTEPS | 2 | 1 | 3 | 0+ |
| BALEQS | 0 | 0 | 0 | 0–100 |
| WASTE | 4 | 3 | 5 | 0–5 |
| SOLIDS | 1 | 1 | 1 | 0 or 1 |
| **Plant Performance** | **38.18** | **49.93** | **22.31** | **0–100** |
| | | | | |
| PCTNEW | 19 (9)[a] | 10 (5)[a] | 25 (20)[a] | 0–100 |
| IMPURITIES | 4 | 3 | 5 | 0–5 |
| COMPLEX | 9 | 6 | 12 | 0+ |
| INCLUSIV. | 33 | 50 | 0 | 0–100 |
| PROJ. DEF. | 7 | 6 | 8 | 2–8 |
| **Cost Growth (HT)** | **0.47** | **0.63** | **0.30** | **0–1** |
| **Cost Growth (LT)** | **0.50** | **0.65** | **0.31** | **0–1** |

[a] Value in parentheses is value chosen for LT scenario.

# Results and Discussion

## Process Results

With a higher carbon conversion efficiency (see mass and energy balance discussions below), the HT scenario produces 61.0 GGE of fuels per dry metric ton of feedstock compared to 47.2 GGE/metric ton for the LT scenario. Resulting annual fuel outputs are 41.7 million GGE/yr for the HT scenario and 32.3 million GGE/yr for the LT scenario.

Along with lower fuel yield, the LT scenario consumes less power (Table 14). The LT scenario and HT scenario total power usage are 15 and 22 MW, respectively. Major contributions to this result are a lower grinder power due to less strict biomass size requirement, lower pressurized oxygen consumption in the gasifier, and generally lower downstream mass flow rates throughout the plant for the LT scenario. Lower syngas yield also means that there is less unconverted syngas and fuel gas from the hydroprocessing area available for the gas turbine. Therefore, the LT scenario generates 31 MW compared to 36 MW generated by the HT scenario. Due to unoptimized flow rates of the recycle streams, the LT scenario actually generates a net 16 MW of power, which is more than the 14 MW produced in the HT scenario. Reducing the net power generation is achievable by increasing the recycle ratio and thereby increasing conversion, but a consequence is higher flow rates and therefore larger and more expensive equipment. The focus of this study is to produce liquid fuels. However, procedures to optimize recycle ratios, equipment sizes, and fuel production rates are not within the scope of this study and are not undertaken.

**Table 14. Power Generation and Usage**

| **Power (MW)** | **HT Scenario** | **LT Scenario** |
|---|---:|---:|
| **USAGE** | | |
| Chopper | 0.50 | 0.50 |
| Grinder | 2.96 | 1.10 |
| Lockhopper System | 0.18 | 0.18 |
| Lean Amine Solution Pump | 0.86 | 0.69 |
| Syngas Booster Compressor | 1.25 | 0.96 |
| PSA Compressor | 0.15 | 0.11 |
| Recycle Compressor | 0.39 | 0.29 |
| Hydroprocessing Area | 2.24 | 1.73 |
| Oxygen Compressor (ASU) | 3.61 | 2.80 |
| Air Compressor (ASU) | 7.94 | 6.31 |
| Sour Gas Shift Steam Compressor | 1.59 | 0 |
| $CO_2$ Compressor | 0.39 | 0.39 |
| **Total Usage** | **22.06** | **15.06** |
| **GENERATION** | | |
| Gas Turbine | 26.25 | 21.02 |
| Steam Turbine | 9.63 | 10.40 |
| **Total Generated** | **35.88** | **31.42** |
| **Net Export** | **13.82** | **16.36** |

An energy balance of the scenarios shows that the biomass-to-fuels efficiency for the LT and HT scenarios is 39% and 50% on a lower-heating-value (LHV) basis, respectively (Table 15). When the net electricity is added, the efficiencies are 43% and 53% on an LHV basis, respectively. The LT scenario is expected to be lower because mass and energy loss occurs in the production and removal of char and tar. Char and tar energy loss sums to 7.5% of the energy in the biomass. In this scenario, char is combusted in a fluidized-bed combustor to provide heat for biomass drying. Biomass drying in the HT scenario is accomplished by a syngas purge. However, the most significant energy loss in the LT scenario, about 25%, occurs across the gasifier. One reason for higher energy loss in the LT scenario is because conversion efficiency increases with increasing operating temperature. A second reason is the loss of energy during the cooling of the syngas after the gasifier. More effective capture of the energy in the hot syngas would increase the overall energy efficiency.

High exothermicity of the FT reaction causes a significant portion of the chemical energy in the syngas to leave as thermal energy in both scenarios. A higher loss across the FT reactor is observed in the HT scenario due to higher flow rates. Energy closure, as shown in Table 15, is approximately 90% for both scenarios. It is assumed that the last 10% is due mostly to heat loss from the cooling of the syngas by direct quench rather than capturing the heat and raising steam.

**Table 15. Overall Energy Balance on LHV Basis**

| | HT Scenario | LT Scenario |
|---|---|---|
| **IN** | | |
| Biomass | 100.0 | 100.0 |
| **OUT** | | |
| Fuel | 49.7 | 38.5 |
| Net Electricity | 3.5 | 4.2 |
| Power Gen Losses | 4.2 | 3.1 |
| FT Reactor Losses | 16.2 | 12.5 |
| Gasifier Losses | 12.1 | 24.9 |
| Char | - | 6.3 |
| Tar | - | 1.2 |
| Syngas Purge | 1.8 | - |
| Total[a] | 87.5 | 90.7 |

[a] The balance of energy is assumed to come from various heating and cooling losses.

A carbon balance analysis shows that 26% and 34% of the carbon in the biomass is passed on to the fuels for the LT and HT scenarios, respectively (Table 16). Approximately 99% of the carbon is accounted for. Major carbon losses include carbon dioxide flue gases, LO-CAT venting, and lockhopper venting. Char leaving the LT scenario is accounted for in the A200 flue gas because the char is combusted for process heat. Also, because the LT scenario produces low-molecular-weight hydrocarbons in the gasification process, a small fraction is dissolved in the liquid effluent of the wet scrubber. Carbon dioxide also dissolves in the wet scrubber effluent stream. Another carbon loss comes from the hydrocarbons that dissolve in the acid gas removal area.

**Table 16. Overall Carbon Balance**

| | HT Scenario | | LT Scenario | |
|---|---|---|---|---|
| | kmol/h | % | kmol/h | % |
| **IN** | | | | |
| Biomass | 3,281 | 100 | 3,281 | 100 |
| **OUT** | | | | |
| Fuel | 1,111 | 34 | 862 | 26 |
| A300 $CO_2$ Vent | 1,458 | 45 | 1,294 | 39 |
| A600 Flue Gas | 334 | 10 | 302 | 9 |
| A200 Flue Gas | 39 | 1 | 227 | 7 |
| Lockhopper Vent | 159 | 5 | 162 | 5 |
| Wet Scrubber Effluent | 154 | 5 | 318 | 10 |
| Tar | 0 | - | 35 | 1 |
| Dissolved Hydrocarbons | 0 | - | 46 | 1 |
| Total | 3,257 | 99 | 3,245 | 99 |

Throughout both scenarios, steam and cooling water are required as utilities. Because a pinch analysis (a method to optimize heat exchange) is not undertaken for this study, integration of the heat streams is not optimized. Therefore, it is assumed that the resulting heating and cooling requirements within the model represent steam and cooling water utilities, whereby they are recycled at a ratio of 9:1. In other words, fresh steam and cooling water utility input to the scenarios are calculated at 10% of the required circulating rate.

## Cost Estimating Results

### *Capital and Operating Costs for n<sup>th</sup> Plant*

The breakdown of costs by area and resulting total capital investment is shown in Table 17. Total capital investment amounts for the HT and LT scenarios are $606 million and $498 million, respectively. Major areas of investment are the gasification area in the HT scenario and the fuel synthesis area in the LT scenario. Moreover, these two areas contain significant differences in capital investment between the scenarios. The installed cost of the entrained-flow gasifier is significantly higher than the cost of the fluidized-bed gasifiers, even when seven of the latter are used in parallel. Area 400 costs for the LT scenario are higher than for the HT scenario due to the steam methane reformer and additional heat-exchange equipment required for the high operational temperature. A significant portion of the capital cost is due to gas compression, such as the air compressor in the air separation unit and the syngas booster compressor. Due to high purchase costs, compressors make up approximately 18% of the TPEC for each scenario. Detailed accounting of equipment found in each process area can be found in Appendix B.

**Table 17. Capital Investment Breakdown for n$^{th}$ Plant Scenarios**

| Area | HT Scenario | | LT Scenario | |
|---|---|---|---|---|
| | Installed Cost | | Installed Cost | |
| | $MM | % | $MM | % |
| A100: Preprocessing | 23 | 7 | 23 | 9 |
| A200: Gasification | 68 | 22 | 28 | 11 |
| A300: Syngas Cleaning | 34 | 11 | 29 | 12 |
| A400: Fuel Synthesis | 49 | 16 | 59 | 23 |
| A500: Hydroprocessing | 33 | 11 | 30 | 12 |
| A600: Power Generation | 46 | 15 | 39 | 15 |
| A700: Air Separation Unit | 24 | 8 | 20 | 8 |
| Balance of Plant | 33 | 11 | 27 | 11 |
| **Total Installed Cost** | **309** | | **254** | |
| **Indirect Cost** | **130** | | **107** | |
| **Total Direct and Indirect Cost** | **439** | | **361** | |
| **Contingency** | **88** | | **72** | |
| **Fixed Capital Investment** | **527** | | **433** | |
| **Working Capital** | **79** | | **65** | |
| **Total Capital Investment** | **606** | | **498** | |

Annualized costs for operation of the plant are shown in Table 18. The percentage displayed also represents percentage of PV. The largest annual incurred costs for both scenarios are the average return on investment and feedstock purchase. Utilities such as steam and cooling water are higher for the LT scenario due to heating and cooling of the syngas before and after the SMR and steam input to the SMR. Waste disposal costs are equal because equal amounts of ash or slag are by-products of both plants. Annual hydroprocessing area costs and income taxes are higher for the HT scenario because of its higher fuel production rate. Fixed costs and capital depreciation are also higher due to higher TCI.

Catalyst costs are not determined on an annual basis because catalysts are assumed to be replaced only every 3 years. Table 19 contains catalyst replacement costs. The catalyst cost for the ZnO guard bed and PSA unit are equal across the scenarios because the volumes of the units are assumed to be the same. The FT catalyst for the HT scenario is significantly more expensive because of a higher gas flow rate and hence, higher catalyst use. Using a DCFROR analysis, the PVs at a net present value of zero for the LT and HT scenarios are $4.83/GGE and $4.27/GGE, respectively. Further detail of the yearly cash flow of the life of the plant can be found in Appendix B.

**Table 18. Annual Operating Cost Breakdown for $n^{th}$ Plant Scenarios**

| | HT Scenario | | LT Scenario | |
|---|---|---|---|---|
| | $MM | % | $MM | % |
| Average Return on Investment | 58.2 | 32.7 | 48.3 | 31.0 |
| Feedstock | 51.3 | 28.9 | 51.3 | 32.9 |
| Capital Depreciation | 26.3 | 14.8 | 21.7 | 13.9 |
| Average Income Tax | 21.9 | 12.3 | 18.0 | 11.6 |
| Fixed Costs | 14.3 | 8.1 | 12.4 | 8.0 |
| Hydroprocessing | 4.4 | 2.5 | 3.0 | 2.0 |
| Steam | 2.7 | 1.5 | 3.5 | 2.2 |
| Cooling Water | 2.3 | 1.3 | 3.5 | 1.6 |
| Waste Disposal | 1.5 | 0.3 | 1.5 | 0.3 |
| Other Raw Material Costs | 1.4 | 0.8 | 1.3 | 0.8 |
| Co-product Credits | (5.6) | -3.1 | (6.6) | -4.2 |

**Table 19. Catalyst Replacement Costs for Both Scenarios (3-Year Replacement Period)**

| Catalyst | HT scenario | LT scenario |
|---|---|---|
| Water-Gas-Shift (copper-zinc) | $115,000 | $105,000 |
| Steam Reforming (nickel-aluminum) | N/A | $103,000 |
| ZnO Guard Bed | $424,000 | $424,000 |
| PSA Packing | $497,000 | $497,000 |
| Fischer-Tropsch (cobalt) | $7,687,000 | $6,128,000 |

### *Sensitivity Results for $n^{th}$ Plant*

The results of sensitivity analysis are summarized in Figure 12 and Figure 13 for the HT and LT scenarios, respectively. Total capital investment and feedstock purchase cost have the highest effect on the PV at approximately ±$0.80 and ±$0.40 per GGE, respectively, for both scenarios. Due to the high percentage of equipment cost for compressors, the compressor installation factor has a very high effect on PV as well. When the compressor installation factor is increased to 3.0, which is the usual installation factor for most of the equipment, the PV increases by $0.71 and $0.78 per GGE for the LT and HT scenarios, respectively. Parameters with a lesser, but still significant, effect are the contingency factor (as percentage of total direct and indirect costs) and plant availability, both with an effect of approximately ±$0.20 per GGE. Parameters with the least effect are generally characteristic of the process rather than of the economics. For example, catalyst life, feedstock moisture, and carbon monoxide conversion in the FT reactor affect the PV less than ±$0.15 per GGE.

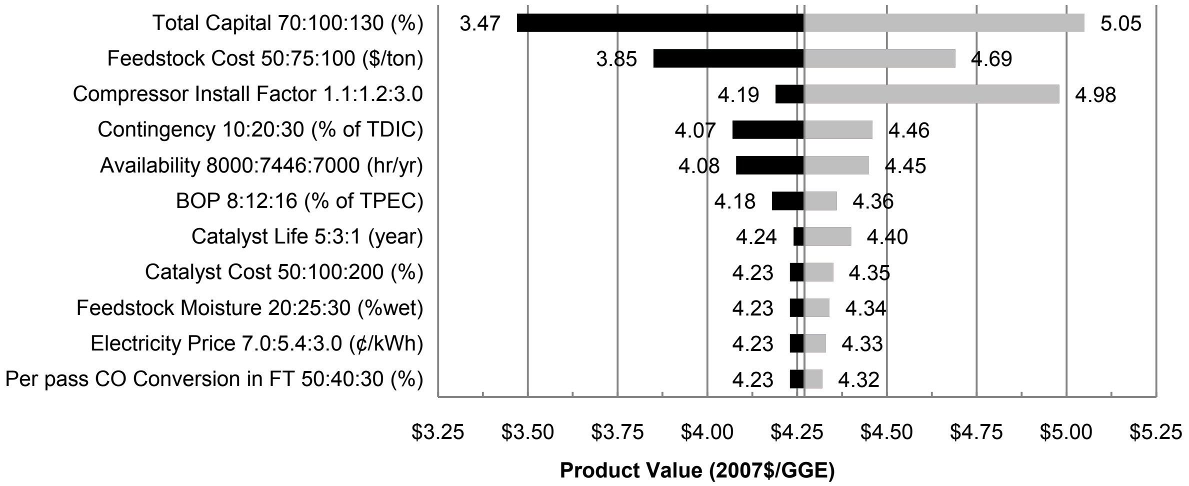


**Figure 12. Sensitivity results for HT $n^{th}$ plant scenario**

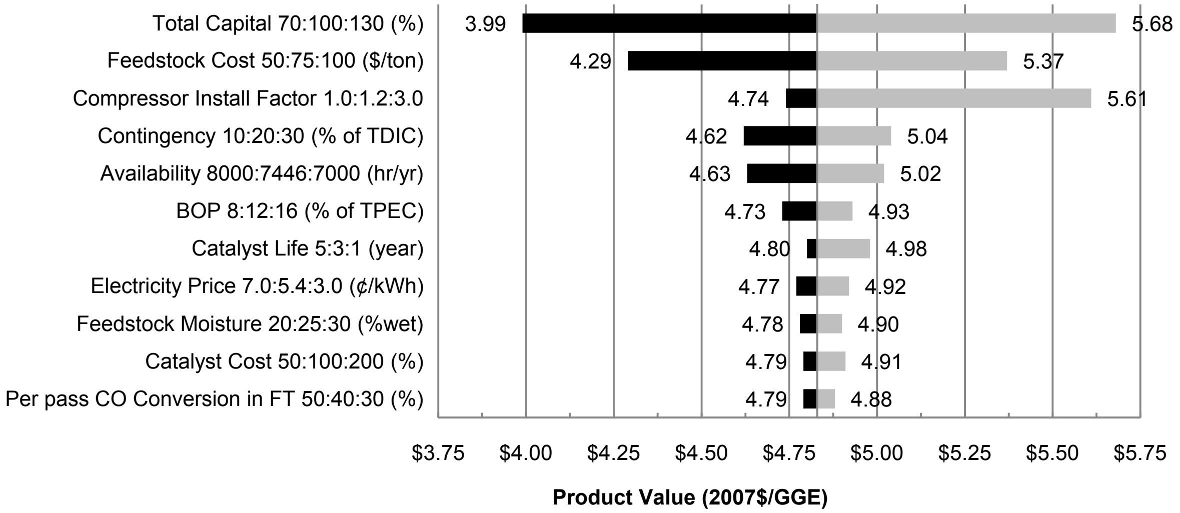


**Figure 13. Sensitivity results for LT $n^{th}$ plant scenario**

Additionally, the size of the plants can be varied by feedstock input rate. The effects of plant size on PV and TCI are shown in Figure 14 and Figure 15, respectively. When the plant size is reduced to 500 metric ton per day, the two scenarios approach equal PV. Also, as the plant size is reduced from the baseline, the difference in capital investment decreases. As the plant size increases past the baseline, the slope of the PV levels out, suggesting that the benefits of lower PV may not be worth the significant increase in capital cost (Figure 14).

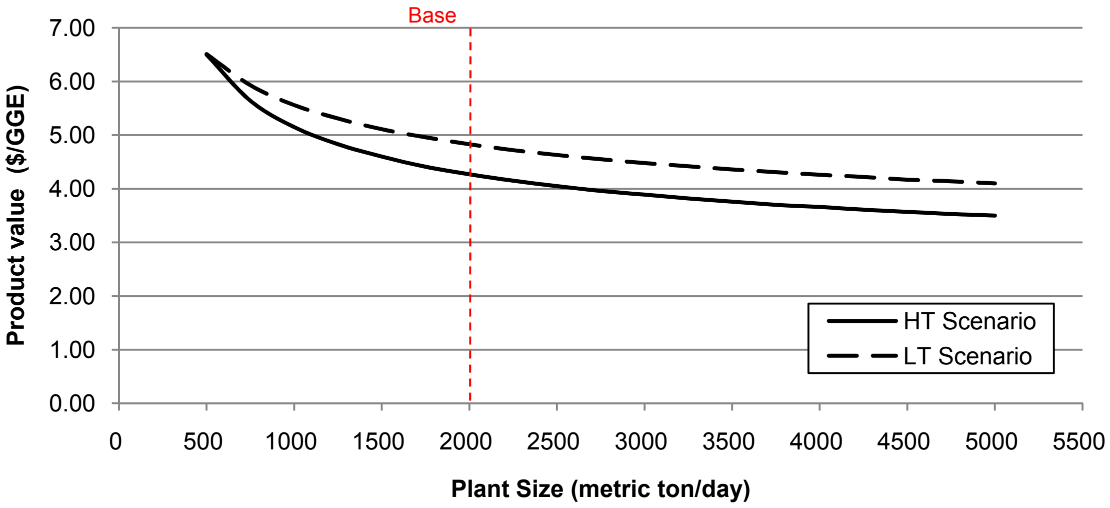


**Figure 14. The effect of plant size on product value (per gallon of gasoline equivalent) for nth plant scenarios**

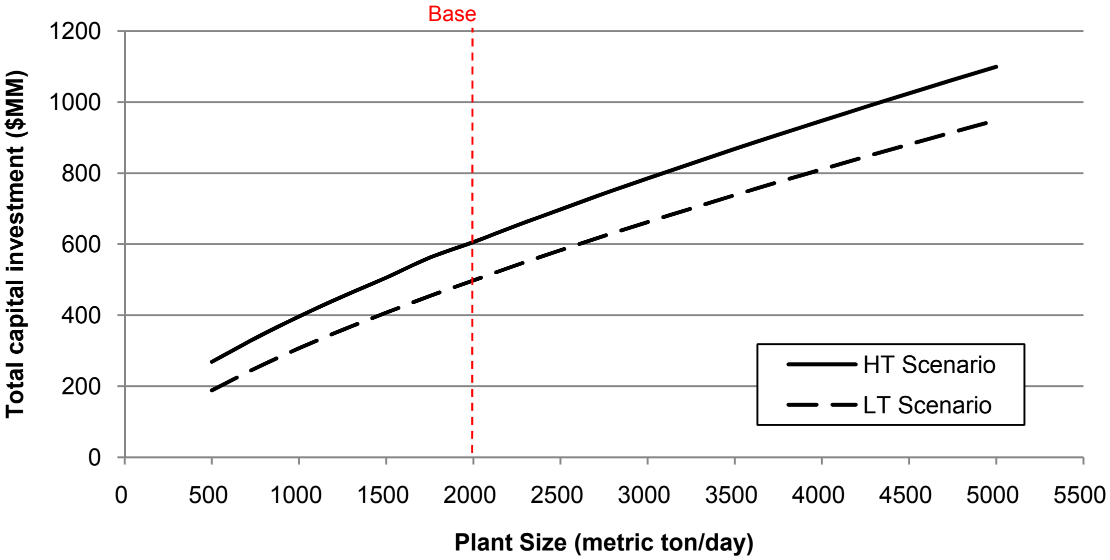


**Figure 15. The effect of plant size on total capital investment for nth plant scenarios**

### *Pioneer Plant Analysis Results*

The total capital investment for a base-case pioneer plant is expected to double from the nth plant scenario, as detailed in Table 20. PV for a base case pioneer plant is estimated to increase to $7.60 and $8.10 per GGE for the LT and HT scenarios, respectively. Table 20 also shows estimates for the optimistic and pessimistic cases. An important observation is that the pioneer-plant PV for the LT scenario is actually lower than for the HT scenario. The reason behind this inverted result is because of the higher capital cost inflation (cost growth factor) in the HT scenario due to higher gasification area capital costs.

**Table 20. Pioneer Plant Analysis Results**

| Analysis | HT Scenario | | LT Scenario | |
|---|---|---|---|---|
| | TCI ($MM) | PV ($/GGE) | TCI ($MM) | PV ($/GGE) |
| n$^{th}$ Plant | 606 | 4.27 | 498 | 4.83 |
| 1$^{st}$ Plant Base | 1,400 | 8.10 | 1,080 | 7.60 |
| 1$^{st}$ Plant Optimistic | 1,030 | 6.30 | 830 | 6.30 |
| 1$^{st}$ Plant Pessimistic | 2,200 | 12.60 | 1,740 | 11.50 |

## Comparison with Previous Techno-Economic Studies

Two previous BTL studies that specifically use biomass feedstock, low-temperature gasification, and Fischer-Tropsch synthesis technology are Tijmensen et al. [40] and Larson et al. [38]. In order to compare with these studies, major economic and process parameters from the present n$^{th}$ plant LT scenario are adjusted to reflect similar values of the previous studies. First, the plant size of the present study is adjusted to increase equipment costs and raw materials purchases. As a result, the annual biomass input and TCI are affected. Second, availability in hours per year, rate of return, cost year, and feedstock cost are adjusted. The combined effect of all adjusted parameters causes the present study's product value to reflect the comparison study.

A comparison to the IGT-R scenario (which employs a low-temperature IGT gasifier and a steam methane reformer) in Tijmensen et al. shows that fuel product value is higher in the present study, as summarized in Table 21. Of all the scenarios developed by Tijmensen et al., the IGT-R scenario is most similar to the present study because of the reformer. The IGT-R scenario has a TCI of $387 million, a feedstock cost of $33 per short ton, and a product value of $1.90/GGE. An important characteristic of the Tijmensen et al. study is that it does not include a hydroprocessing area. Therefore, it might be expected that the TCI would be higher for the present study, which includes hydroprocessing. However, that is not the case, as the TCI of the present study using Tijmensen et al. parameters is $339 million, which is lower than the reported $387 million of the earlier study. Another important observation is that the annual fuel production for the present study, with adjusted parameters, is 30.2 million gallons per year, compared to 39.8 million gallons per year of FT products reported by Tijmensen et al. One reason for lower annual fuel production in the present study is because of a loss during hydroprocessing. Therefore, due to lower annual fuel production and hence lower fuel revenue, the present study has a higher product value compared to Tijmensen et al.

**Table 21. Comparison of $n^{th}$ Plant LT Scenario to Tijmensen et al. Study IGT-R Scenario [40]**

| **Parameter** | **Tijmensen et al. Study (IGT-R Scenario)** | **Present Study $n^{th}$ Plant LT Scenario** | **Present Study w/ Tijmensen et al. Parameters** |
|---|---:|---:|---:|
| Plant Size (dry tons/day) | 1,920 | 2,205 | 1,920 |
| Annual Biomass Input (tons) | 640,000 | 684,100 | 640,000 |
| Total Capital Investment ($MM) | 387 | 498 | 339 |
| Availability (hour/year) | 8,000 | 7,446 | 8,000 |
| Rate of Return (%) | 10 | 10 | 10 |
| Cost Year | 2000 | 2007 | 2000 |
| Feedstock Cost ($/short ton) | 33.00 | 75.00 | 33.00 |
| Efficiency (%, LHV, incl. elec.) | 50.1 | 42.7 | 42.7 |
| Fuel Yield (MMGGE/yr) | 39.8 | 32.3 | 30.2 |
| Product Value ($/GJ) | 16.50 | 39.80 | **25.17** |
| Product Value ($/GGE) | 2.00 | 4.83 | **3.05** |

A comparison to the FT-OT-VENT scenario (which is low-temperature gasification with carbon dioxide vent and once through FT synthesis) reported by Larson et al. is summarized in Table 22. In a similar fashion to the previous comparison, the parameters are adjusted to approximate the comparison study. Some important observations are made from this comparison. First, the TCI of the present study with adjusted parameters is significantly higher. Second, the net electricity is significantly lower for the present study. Third, the PV is significantly higher for the present study. Essentially, the Larson et al. study generates more revenue from selling electricity and recovers the capital investment in less time. In addition, annual operating costs for the Larson et al. study are lower than for the present study. Therefore, because of lower revenue from electricity and higher annual operating costs, the present study has a higher fuel product value when compared on a similar basis to Larson et al.

**Table 22. Comparison of nth Plant LT Scenario to Larson et al. Study FT-OT-VENT Scenario [38]**

| Parameter | Larson et al. Study (FT-OT-VENT Scenario) | Present Study nth Plant LT Scenario | Present Study w/ Larson et al. Parameters |
|---|---|---|---|
| Plant Size (dry tons/day) | 5,000 | 2,205 | 5,001 |
| Annual Biomass Input (tons) | 1,458,000 | 684,000 | 1,459,000 |
| Total Capital Investment ($MM) | 541 | 498 | 678 |
| Availability (hour/year) | 7,000 | 7,446 | 7,000 |
| Debt/Equity (% Equity) | 60 | 100 | 60 |
| Rate of Return (%) | 12 | 10 | 12 |
| Cost Year | 2003 | 2007 | 2003 |
| Electricity Price (cents/kWh) | 4.0 | 5.4 | 4.0 |
| Net Electricity (MW) | 207 | 16.3 | 37.1 |
| Feedstock Cost ($/short ton) | 46.00 | 75.00 | 46.00 |
| Plant Yield (MMGGE/yr) | 63.3 | 32.3 | 68.9 |
| Product Value ($/GJ) | 15.25 | 39.80 | **26.80** |
| Product Value ($/GGE) | 1.85 | 4.83 | **3.25** |

## Summary of nth Plant Scenarios

The HT scenario requires more power and capital investment, yields more fuel per ton of feedstock, and subsequently produces more fuel per year compared to the LT scenario. The total capital investment requirements for the LT and HT scenarios are $498 million and $606 million, respectively. Despite higher capital investment for the HT scenario, the product value (PV) is lower. PVs for the LT and HT scenarios are $4.83 and $4.27 per gallon of gasoline equivalent, respectively. The main reason for a lower PV is because of increased fuel revenue. The main nth plant scenario results are shown in Table 23. A detailed summary of costs can be found in Appendix B.

**Table 23. Main Scenario nth Plant Results**

| Scenario | TCI ($MM) | TPEC ($MM) | Fuel Yield (GGE/metric ton) | Annual Fuel Output (MMGGE/yr) | Net Electricity Export (MW) | PV ($/GGE) |
|---|---|---|---|---|---|---|
| High Temperature | 605.9 | 145.7 | 61.0 | 41.7 | 13.8 | 4.27 |
| Low Temperature | 498.3 | 120.4 | 47.2 | 32.3 | 16.4 | 4.83 |

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## Conclusions

This analysis compares capital and operating cost for two biomass-to-liquids scenarios: high-temperature (HT) gasification and low-temperature (LT) gasification. The selection of these scenarios allows for direct comparison between two modes of gasification: slagging and non-slagging. The slagging, entrained-flow gasifier employed for the HT scenario results in higher plant costs (about 20%) than the LT scenario, which employs a fluidized bed gasifier. On the other hand, the higher carbon conversions for the HT gasifier result in a lower PV compared to the LT scenario. A biomass-to-liquids plant is expected to produce fuels costing in the range of \$4–\$5 per gallon gasoline equivalent using present gasification and Fischer-Tropsch synthesis technology. The factors chiefly responsible for this relatively high PV are feedstock costs and investment return on the capital to build a \$500 million to \$650 million plant to process 2,000 metric tons of biomass per day. A pioneer plant analysis estimates that the total capital investment for a pioneer plant would double and PV would increase by approximately 60% from the $n^{th}$ plant scale. This uncertainty suggests that economics are still a major challenge for biomass-to-liquids production plants.

Parameters with the most sensitive effect on PV are total capital cost, feedstock purchase cost, and compressor installation factor, affecting the PV ±\$0.40–\$0.80 per gallon. Less expensive biomass feedstock that is lower in ash content than that used in the present study will have higher fuel yield and have the potential to significantly decrease PV. Gas compression is a major portion of capital investment, and sensitivity analysis shows that installation costs of compressors have a high effect on PV. Factors with little effect on the PV are mostly related to the process, such as carbon monoxide conversion in the FT reactor, feedstock inlet moisture, and catalyst lifetime.

Because of time and resource constraints, the techno-economic study presented includes a few shortcomings. The process configuration is not fully optimized for heat integration. While some recycle streams are included, a complete heat exchange network for heat recovery is not conceptualized. In addition, some areas such as FT product separation and hydroprocessing are not modeled rigorously and can be improved with detailed mass and energy flows. Further studies can benefit by optimizing heat exchange networks and modeling in detail areas such as FT product separation and hydroprocessing. More detailed modeling will likely result in changes to economic components such as equipment cost, installation factors, and contingency. However, because increasing model detail may result in tradeoffs between the different economic components, the net effect of increased model detail on PV is difficult to predict.

This analysis tracks PV based on commercial technology for which sufficient public domain data existed in 2007. In order to evaluate the economics of biofuels produced from gasification based on future scenarios, particular attention will be needed on the most sensitive parameters—feedstock cost and capital cost. Attention should also focus on 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 pioneer plant PV.

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# Appendix A. Techno-Economic Model Assumptions

## Financial Assumptions

- Capital investment
    - Equity: 100%
    - Working capital (% of FCI): 15%
- Depreciation model
    - Zero Salvage Value for both general plant and steam/ power plant
    - Type of depreciation: Double-Declining-Balance Depreciation Method (DDB) as per IRS Modified Accelerated Cost Recovery System (MACRS) guidelines
        - Depreciation period (years):
            - General plant: 7
            - Steam/power system: 20
- Construction & startup:
    - Construction period (years): 2.5
        - % spent in year "-3": 8%
        - % spent in year "-2": 60%
        - % spent in year "-1": 32%
    - Startup time (years): 0.5
        - Revenues (% of normal): 50%
        - Variable costs (% of normal): 75%
        - Fixed cost (% of normal): 100%
- Other
    - Internal rate of return: 10%
    - Income tax rate: 39%
    - Operating hours per year: 8,406

## Capital Costs

- Cost year for analysis: 2007; cost escalation is applied using the Chemical Engineering Plant Cost Index
- The plant is designed based on the state of the technology, at the n$^{th}$ plant level of experience
- Most equipment installation factors are applied using Peters et al. for solid-fluid plants (i.e., 3.02 installation factor)

- 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

### Operating Costs

- Working capital is assumed to be 15% of the total capital investment
- Annual maintenance materials are 2% of the total installed equipment cost
- Boiler feedwater and wastewater treatment costs are derived from prior NREL work
- Fresh cooling water and steam costs are calculated at 10% of the required circulation rate, meaning a 9:1 ratio of water recycling
- Employee salary estimation is same as that chosen by Phillips et al.
- Employee salaries are indexed to the year of 2007 following the data of the Bureau of Labor Statistics

### Feedstock, Products, and By-Products

- Feedstock is corn stover (comprising stalks, leaves, cobs, and husks)
    - Moisture content in the feedstock is 25%
- Feed rate is 2,000 dry metric ton per day
    - The feedstock delivery logistics are not considered
    - The feedstock is delivered to the feed handling area of the plant
- Feed cost is assumed to be $75/dry short ton at the gate
- Gasoline and diesel products are sold for over-the-fence prices
- Gasoline energy content is 115,000 BTU/gallon
- Fly ash and slag incur a solids waste disposal cost
- Solid sulfur and electricity are sold as by-products

### Process Assumptions

For both scenarios, 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.

Area 100 - Preprocessing

- Biomass is dried down to 10%
    - Steam raised from hot flue gas is used to dry the feedstock
    - Steam to moisture removal ratio is set at 9:1 in accordance with Amos
    - Heat is provided by combusting char and unreacted syngas
- Grinder reduces biomass to 6-mm or less

    - The energy required for grinding is calculated separately using literature correlations by Mani et al.

Area 200 - Gasification

- Scenario 1: Entrained-flow gasifier is modeled using thermodynamic equilibrium
- Scenario 2: Fluidized-bed gasifier is modeled using a mass balance calculation
- 95% purity oxygen produced from air separation unit provides oxidizer
- Carbon dioxide is used as solids pressurization gas
- All char produced in LT scenario is combusted for process heat

Area 300 - Syngas Cleaning

- Particulates, tar, and partial ammonia removal via wet scrubbing
    - Scrubbing water is recycled at 90% rate
    - Particulate handling (not modeled)
        - High-temperature gasifier: particulate decant slurry is sent back into slagging gasifier
        - Low-temperature gasifier: particulate decant slurry is piled and landfilled; excess water is sent to aerobic water treatment (not modeled)
    - Makeup water compensates for water lost via particulate slurry
        - Process water condensate is used as makeup water
- Sour water-gas-shift occurs at equilibrium and is modeled as such
- Carbon dioxide, hydrogen sulfide, and excess ammonia removal via amine scrubbing acid gas removal (AGR) at pressure
    - 99% of sulfur is removed and 90% of carbon dioxide
    - Monoethanolamine (MEA) is the scrubbing solvent
    - Carbon dioxide is vented following LO-CAT removal of $H_2S$
- Hydrogen sulfide is converted to solid sulfur via LO-CAT oxidation (99% conversion)
- Ammonia can be disposed of by decomposition (not modeled) in
    - Gasifier burner (slagging gasifier)
    - Char and syngas combustor (fluidized-bed gasifier)
- Zinc oxide and activated carbon guard bed polishing assumed (not modeled in detail)

Area 400 - Fuel Synthesis

- Water-gas-shift occurs at equilibrium and is modeled as such
- Pressure swing adsorption (PSA) is employed to remove excess $H_2$ at an efficiency of 85% and 99% purity
  - The PSA system employs two trains with six reactors each to account for all stages of pressurization, depressurization, purging, etc.
  - PSA adsorbers are filled 2/3 with activated carbon and 1/3 with molecular sieve
- Syngas is catalytically converted to fuels by one-step Fischer-Tropsch synthesis followed by wax hydrocracking and fuel separation
  - FT synthesis employs cobalt catalyst
  - 40% syngas conversion to fuels
  - Part of the unconverted syngas is recycled
    - A fraction of the recycle is sent to the AGR to prevent $CO_2$ buildup
    - The overall recycle ratio is about 1.9
- A syngas purge is used as fuel in the combustor side of the biomass dryer (only in HT scenario)
- Excess syngas is sent to a gas turbine for power production

Areas 500, 600, 700

- Hydroprocessing and product distillation costs are estimated as a “black box” based on literature capital cost and operating cost information from Robinson et al.
  - Literature yield data is used for estimating the relative yields of gasoline and diesel

## Miscellaneous

- Combustion occurs with 20% excess oxygen

# Appendix B. Detailed Costs

## Cost Summaries

### HT Biomass-to-Liquids Scenario Summary

2,000 Dry Metric Tonnes Biomass per Day
High Temperature Entrained Flow Gasifier, Sulfur Removal, Fischer-Tropsch Synthesis, Hydroprocessing, Combined Cycle Power
All Currency in 2007$ and Volume in Gallons Gasoline Equivalent (GGE)

| | |
|---|---|
| Product Value ($/gal) | **$4.26** |
| Total Production at Operating Capacity (MM gal / year) | 41.7 |
| Product Yield (gal / Dry US Ton Feedstock) | 61.0 |
| Delivered Feedstock Cost $/Dry US Ton | $75 |
| Internal Rate of Return (After-Tax) | 10% |
| Equity Percent of Total Investment | 100% |

| Capital Costs | | |
|---|---|---|
| Area 100: Pretreatment | $22,700,000 | 7% |
| Area 200: Gasification | $67,800,000 | 22% |
| Area 300: Syngas Cleaning | $33,500,000 | 11% |
| Area 400: Fuel Synthesis | $49,400,000 | 16% |
| Area 500: Hydrocracking/Hydrotreating | $33,000,000 | 11% |
| Area 600: Power Generation | $45,600,000 | 15% |
| Area 700: Air Separation | $24,300,000 | 8% |
| Balance of Plant | $33,100,000 | 11% |
| Total Installed Equipment Cost | $309,400,000 | |
| Indirect Costs | 129,700,000 | |
| (% of TPI) | 21.4% | |
| Project Contingency | 79,000,000 | |
| Total Project Investment (TPI) | $605,900,000 | |
| Installed Equipment Cost per Annual Gallon | $7.42 | |
| Total Project Investment per Annual Gallon | $14.52 | |
| Loan Rate | N/A | |
| Term (years) | N/A | |
| Capital Charge Factor | 0.176 | |
| Gasifier Efficiency - HHV % | 82.1 | |
| Gasifier Efficiency - LHV % | 87.9 | |
| Overall Plant Efficiency (incl. electricity) - HHV % | 52.7 | |
| Overall Plant Efficiency - LHV % | 53.0 | |
| Availability (%) | 85.0% | |
| Plant Hours per year | 7446 | |

| Operating Costs (cents/gal product) | | |
|---|---|---|
| Feedstock | 123.0 | 28.9% |
| Steam | 6.4 | 1.5% |
| Cooling Water | 5.5 | 1.3% |
| Other Raw Materials | 3.4 | 0.8% |
| Waste Disposal | 1.3 | 0.3% |
| Hydroprocessing | 10.6 | 2.5% |
| Fixed Costs | 34.4 | 8.1% |
| Co-product credits | -13.3 | -3.1% |
| Capital Depreciation | 63.0 | 14.8% |
| Average Income Tax | 52.4 | 12.3% |
| Average Return on Investment | 139.5 | 32.7% |

| Operating Costs ($/yr) | |
|---|---|
| Feedstock | $51,300,000 |
| Steam | $2,700,000 |
| Cooling Water | $2,300,000 |
| Other Raw Matl. Costs | $1,400,000 |
| Waste Disposal | $1,500,000 |
| Hydroprocessing | $4,400,000 |
| Fixed Costs | $14,300,000 |
| Co-product credits | -$5,600,000 |
| Capital Depreciation | $26,300,000 |
| Average Income Tax | $21,900,000 |
| Average Return on Investment | $58,200,000 |

| | |
|---|---|
| Total Plant Electricity Usage (KW) | 22,065 |
| Electricity Produced Onsite (KW) | 35,880 |
| Electricity Purchased from Grid (KW) | 0 |
| Electricity Sold to Grid (KW) | 13,815 |
| Plant Electricity Use (KWh/gal product) | 6.1 |

**Figure B-1. Economic analysis summary for HT scenario**

# LT Biomass-to-Liquids Process Engineering Analysis

2,000 Dry Metric Tonnes Biomass per Day
Low Temperature Fluidized Gasifier, Sulfur Removal, Fischer-Tropsch Synthesis, Hydroprocessing, Combined Cycle Power
All Currency in 2007$ and Volume in Gallons Gasoline Equivalent (GGE)

| | |
|---|---|
| Product Value ($/gal) | **$4.83** |
| Total Production at Operating Capacity (MM gal / year) | 32.3 |
| Product Yield (gal / Dry US Ton Feedstock) | 47.2 |
| Delivered Feedstock Cost $/Dry US Ton | $75 |
| Internal Rate of Return (After-Tax) | 10% |
| Equity Percent of Total Investment | 100% |

| Capital Costs | | |
|---|---|---|
| Area 100: Pretreatment | $22,700,000 | 9% |
| Area 200: Gasification | $28,200,000 | 11% |
| Area 300: Syngas Cleaning | $29,300,000 | 12% |
| Area 400: Fuel Synthesis | $58,700,000 | 23% |
| Area 500: Hydrocracking/Hydrotreating | $29,500,000 | 12% |
| Area 600: Power Generation | $38,900,000 | 15% |
| Area 700: Air Separation | $19,500,000 | 8% |
| Balance of Plant | $27,200,000 | 11% |
| Total Installed Equipment Cost | $253,900,000 | |
| Indirect Costs | 107,200,000 | |
| (% of TPI) | 21.5% | |
| Project Contingency | 65,000,000 | |
| Total Project Investment (TPI) | $498,300,000 | |
| Installed Equipment Cost per Annual Gallon | $7.86 | |
| Total Project Investment per Annual Gallon | $15.43 | |
| Loan Rate | N/A | |
| Term (years) | N/A | |
| Capital Charge Factor | 0.177 | |
| Gasifier Efficiency - HHV % | 64.3 | |
| Gasifier Efficiency - LHV % | 68.8 | |
| Overall Plant Efficiency - HHV % | 43.0 | |
| Overall Plant Efficiency - LHV % | 43.3 | |
| Availability (%) | 85.0% | |
| Plant Hours per year | 7446 | |

| Operating Costs (cents/gal product) | | |
|---|---|---|
| Feedstock | 158.9 | 32.9% |
| Steam | 10.9 | 2.2% |
| Cooling Water | 7.8 | 1.6% |
| Other Raw Materials | 4.1 | 0.8% |
| Waste Disposal | 1.5 | 0.3% |
| Hydroprocessing | 9.4 | 2.0% |
| Fixed Costs | 38.4 | 8.0% |
| Co-product credits | -20.4 | -4.2% |
| Capital Depreciation | 67.2 | 13.9% |
| Average Income Tax | 55.9 | 11.6% |
| Average Return on Investment | 149.5 | 31.0% |

| Operating Costs ($/yr) | |
|---|---|
| Feedstock | $51,300,000 |
| Steam | $3,500,000 |
| Cooling Water | $3,500,000 |
| Other Raw Matl. Costs | $1,300,000 |
| Waste Disposal | $1,500,000 |
| Hydroprocessing | $3,000,000 |
| Fixed Costs | $12,400,000 |
| Co-product credits | -$6,600,000 |
| Capital Depreciation | $21,700,000 |
| Average Income Tax | $18,000,000 |
| Average Return on Investment | $48,300,000 |

| | |
|---|---|
| Total Plant Electricity Usage (KW) | 15,044 |
| Electricity Produced Onsite (KW) | 31,420 |
| Electricity Purchased from Grid (KW) | 0 |
| Electricity Sold to Grid (KW) | 16,376 |
| Plant Electricity Use (KWh/gal product) | 5.4 |

**Figure B-2. Economic analysis summary for LT scenario**

# Detailed Equipment Lists

## *High-Temperature Scenario Equipment Lists*

### Table B-1. Detailed Equipment List for Areas 100 and 200 of HT Scenario

| Equipment Number | Number Required | Number Spares | Equipment Name | Original Equip Cost (per unit) | Base Year | Total Original Equip Cost (Req'd & Spare) in Base Year | Scaled Uninstalled Cost in 2007$ | Installed Cost Base Year | Installed Cost in 2007$ | Cost Source |
|---|---|---|---|---|---|---|---|---|---|---|
| A100.CONV1 | 2 | | Bale Transport Conveyor | $400,000 | 2000 | $800,000 | $1,066,531 | $1,296,000 | $1,727,781 | Aden et al. 2002 |
| A100.CONV2 | 2 | | Bale Unwrapping Conveyor | $150,000 | 2000 | $300,000 | $399,949 | $357,000 | $475,940 | Aden et al. 2002 |
| A100.CONV3 | 1 | | Belt Press Discharge Conveyor | $50,000 | 2000 | $50,000 | $66,658 | $94,500 | $125,984 | Aden et al. 2002 |
| A100.SCALE | 2 | | Truck Scales | $34,000 | 2000 | $68,000 | $90,655 | $167,960 | $223,918 | Aden et al. 2002 |
| A100.FORK1 | 4 | 1 | Truck Unloading Forklift | $18,000 | 2000 | $90,000 | $119,985 | $90,000 | $119,985 | Aden et al. 2002 |
| A100.FORK2 | 4 | | Bale Moving Forklift | $18,000 | 2000 | $72,000 | $95,988 | $72,000 | $95,988 | Aden et al. 2002 |
| A100.SLAB | 1 | | Concrete Feedstock-Storage Slab | $450,655 | 2000 | $450,655 | $600,797 | $991,441 | $1,321,754 | Aden et al. 2002 |
| A100.MAGSEP | 1 | | Magnetic Separator | $13,863 | 1998 | $13,863 | $18,700 | $18,022 | $24,310 | Aden et al. 2002 |
| A100.A100CHOP.CHGRIN01 | 4 | | Chopper | $105,100 | 2007 | $420,400 | $420,400 | $1,105,258 | $1,105,258 | Process Evaluator software<br>Aspen Icarus |
| A100.A100CHOP.CHMIX01 | 1 | | Chopper Conveyor | $61,400 | 2007 | $61,400 | $61,400 | $185,428 | $185,428 | Aspen Icarus |
| A100.A100CHOP.CHSEP01 | 1 | | Chopper Screen with Recycle Conveyor | $20,800 | 2007 | $20,800 | $20,800 | $62,816 | $62,816 | Aspen Icarus |
| A100.A100DRY.DRDRY01 | 10 | | Dryer | $633,700 | 2007 | $6,337,000 | $6,337,000 | $15,201,647 | $15,201,647 | Aspen Icarus |
| A100.A100GRIN.GRGRIN01 | 4 | | Grinder | $167,100 | 2007 | $668,400 | $668,400 | $1,757,266 | $1,757,266 | Aspen Icarus |
| A100.A100GRIN.GRMIX01 | 1 | | Grinder Conveyor | $61,400 | 2007 | $61,400 | $61,400 | $185,428 | $185,428 | Aspen Icarus |
| A100.A100GRIN.GRSEP01 | 1 | | Grinder Screen with Recycle Conveyor | $20,800 | 2007 | $20,800 | $20,800 | $62,816 | $62,816 | Aspen Icarus |
| **A100** | | | | | **Subtotal** | **$9,434,718** | **$10,049,464** | **$21,647,582** | **$22,676,317** | |
| A200.A200COMB.CBREAC01 | 1 | | Combustor - Steam Boiler | $1,450,500 | 2007 | $1,450,500 | $1,450,500 | $4,380,510 | $4,380,510 | Aspen Icarus |
| A200.A200SLAG.SLREAC01 | 1 | | Entrained Flow, Slagging Gasifier | $23,234,043 | 2006 | $23,234,043 | $24,433,879 | $54,600,000 | $57,419,616 | Reed et al. 2007 |
| A200.A200SLAG.SLSEP01 | 1 | | Slag collector/separator | $35,100 | 2007 | $35,100 | $35,100 | $106,002 | $106,002 | Aspen Icarus |
| A200.A200SLAG.SLSEP03 | 3 | | Direct Quench Syngas Cooler | $396,200 | 2007 | $1,188,600 | $1,188,600 | $3,589,572 | $3,589,572 | Aspen Icarus |
| A200.GSHOP01 | 1 | | Biomass Receiving Hopper | $151,400 | 2007 | $297,900 | $297,900 | $899,658 | $899,658 | Aspen Icarus |
| A200.GSTANK01 | 1 | | Lockhopper | $229,100 | 2007 | $229,100 | $229,100 | $691,882 | $691,882 | Aspen Icarus |
| A200.GSTANK02 | 1 | | Biomass Feeding Bin | $228,900 | 2007 | $228,900 | $228,900 | $691,278 | $691,278 | Aspen Icarus |
| **A200** | | | | | **Subtotal** | **$26,664,143** | **$27,863,979** | **$64,958,902** | **$67,778,518** | |

# Table B-2. Detailed Equipment List for Areas 300, 400, and 500 of HT Scenario

| Equipment Number | Number Required | Number Spares | Equipment Name | Original Equip Cost (per unit) | Base Year | Total Original Equip Cost (Req'd & Spare) in Base Year | Scaled Uninstalled Cost in 2007$ | Installed Cost Base Year | Installed Cost in 2007$ | Cost Source |
|---|---|---|---|---|---|---|---|---|---|---|
| A300.A300AGR.AGRarea | 1 | | High Pressure Amine System | $6,949,800 | 2005 | $6,949,800 | $7,798,857 | $20,988,396 | $23,552,549 | Phillips et al. 2007 |
| A300.A300SGS.SGCOMP01 | 2 | | Sour Water Gas Shift Steam Compressor | $1,381,900 | 2007 | $2,763,800 | $2,763,800 | $3,316,560 | $3,316,560 | Aspen Icarus |
| A300.A300SGS.SGREAC01 | 1 | | Sour Water Gas Shift Reactor | $66,600 | 2007 | $66,600 | $66,600 | $201,132 | $201,132 | Aspen Icarus |
| A300.A300SUL.SUCOL01 | 1 | | LO-CAT Absorber | $23,800 | 2007 | $23,800 | $23,800 | $71,876 | $71,876 | Aspen Icarus |
| A300.A300SUL.SUREAC01 | 1 | | LO-CAT Oxidizer Vessel | $1,000,000 | 2007 | $1,000,000 | $1,000,000 | $3,020,000 | $3,020,000 | Phillips et al. 2007 |
| A300.A300SUL.SUSEP01 | 1 | | Sulfur Separator | $15,900 | 2007 | $15,900 | $15,900 | $48,018 | $48,018 | Aspen Icarus |
| A300.CLCOMP01 | 2 | | Carbon Dioxide Compressor | $1,181,200 | 2007 | $2,362,400 | $2,362,400 | $2,834,880 | $2,834,880 | Aspen Icarus |
| A300.CLDRUM01 | 1 | | Liquid Collection Tank | $29,600 | 2007 | $29,600 | $29,600 | $89,392 | $89,392 | Aspen Icarus |
| A300.CLHEAT03 | 1 | | Direct Quench Syngas Cooler | $91,500 | 2007 | $91,500 | $91,500 | $276,330 | $276,330 | Aspen Icarus |
| A300.CLMIX01 | 1 | | Venturi Scrubber | $27,100 | 2007 | $27,100 | $27,100 | $81,842 | $81,842 | Aspen Icarus |
| **A300** | | | | | **Subtotal** | **$13,330,500** | **$14,179,557** | **$30,928,426** | **$33,492,579** | |
| A400.FSCOMP01 | 2 | | Booster Syngas Compressor | $1,007,100 | 2007 | $2,014,200 | $2,014,200 | $2,417,040 | $2,417,040 | Asen Icarus |
| A400.FSCOMP02 | 1 | | Recycle Syngas Booster Compressor | $748,400 | 2007 | $748,400 | $748,400 | $898,080 | $898,080 | Asen Icarus |
| A400.FSCOMP03 | 1 | | PSA Booster Compressor | $1,461,700 | 2007 | $1,461,700 | $1,461,700 | $1,754,040 | $1,754,040 | Asen Icarus |
| A400.FSHEAT01 | 1 | | Syngas Heater | $73,400 | 2007 | $73,400 | $73,400 | $221,668 | $221,668 | Asen Icarus |
| A400.FSHEAT03 | 1 | | Syngas Cooler | $137,400 | 2007 | $137,400 | $137,400 | $414,948 | $414,948 | Asen Icarus |
| A400.FSHEAT04 | 1 | | Recycle Syngas Pre-heater | $21,500 | 2007 | $21,500 | $21,500 | $64,930 | $64,930 | Asen Icarus |
| A400.FSREAC01 | 1 | | Fischer-Tropsch Reactor | $8,888,889 | 2003 | $8,888,889 | $11,617,468 | $32,000,000 | $41,822,886 | Larson et al. 2005 |
| A400.FSSEP01 | 2 | | ZnO Sulfur Removal Beds | $61,000 | 2007 | $122,000 | $122,000 | $368,440 | $368,440 | Asen Icarus |
| A400.FSSEP02 | 12 | | Pressure Swing Absorption Unit | $33,300 | 2007 | $399,600 | $399,600 | $1,206,792 | $1,206,792 | Asen Icarus |
| A400.FSSEP03 | 1 | | FT knock-out Column | $39,600 | 2007 | $39,600 | $39,600 | $119,592 | $119,592 | Asen Icarus |
| A400.FSSEP04 | 1 | | Water Separator | $47,900 | 2007 | $47,900 | $47,900 | $144,658 | $144,658 | Asen Icarus |
| **A400** | | | | | **Subtotal** | **$13,954,589** | **$16,683,168** | **$39,610,188** | **$49,433,074** | |
| A500.HYREAC01 | 1 | | Hydroprocessing Unit | $9,377,483 | 2007 | $9,377,483 | $9,377,483 | $28,320,000 | $28,320,000 | Robinson & Dolbear 2007 |
| A500.HYTANK01 | 1 | | Diesel 30-day Storage Tank | $1,167,600 | 2007 | $1,167,600 | $1,167,600 | $3,526,152 | $3,526,152 | Aspen Icarus |
| A500.HYTANK02 | 1 | | Gasoline 30-day Storage Tank | $371,900 | 2007 | $371,900 | $371,900 | $1,123,138 | $1,123,138 | Aspen Icarus |
| **A500** | | | | | **Subtotal** | **$10,916,983** | **$10,916,983** | **$32,969,290** | **$32,969,290** | |

# Table B-3. Detailed Equipment List for Areas 600 and 700 of HT Scenario

| Equipment Number | Number Required | Number Spares | Equipment Name | Original Equip Cost (per unit) | Base Year | Total Original Equip Cost (Req'd & Spare) in Base Year | Scaled Uninstalled Cost in 2007$ | Installed Cost Base Year | Installed Cost in 2007$ | Cost Source |
|---|---|---|---|---|---|---|---|---|---|---|
| A600.COMBB | 1 | | Combustion Turbine - Electric Generator | $22,404,000 | 2007 | $22,404,000 | $22,404,000 | $26,884,800 | $26,884,800 | Aspen Icarus |
| A600.CWPUMP | 1 | 1 | Cooling Water Pump | $5,900 | 2007 | $11,800 | $11,800 | $35,636 | $35,636 | Aspen Icarus |
| A600.ECON1_HRSG | 1 | | Heat Recovery Steam Generator | $202,200 | 2007 | $202,200 | $202,200 | $610,644 | $610,644 | Aspen Icarus |
| A600.HPPUMP | 1 | 1 | High Pressure Steam Pump | $266,700 | 2007 | $533,400 | $533,400 | $1,610,868 | $1,610,868 | Aspen Icarus |
| A600.HPSEP | 1 | | High Pressure Steam/Water Separation | $107,400 | 2007 | $107,400 | $107,400 | $324,348 | $324,348 | Aspen Icarus |
| A600.LPEXP_ELECGEN | 1 | | Combined Steam Turbine - Electric Gen. | $4,709,600 | 2007 | $4,709,600 | $4,709,600 | $5,651,520 | $5,651,520 | Aspen Icarus |
| A600.LPSEP | 1 | | Low Pressure Water/Steam Separation | $108,800 | 2007 | $108,800 | $108,800 | $328,576 | $328,576 | Aspen Icarus |
| A600.O2COMP | 1 | | Air Compressor | $8,431,900 | 2007 | $8,431,900 | $8,431,900 | $10,118,280 | $10,118,280 | Aspen Icarus |
| **A600** | | | | | **Subtotal** | **$36,509,100** | **$36,509,100** | **$45,564,672** | **$45,564,672** | |
| A700.COMP1 | 2 | | Air Compressor | $3,346,500 | 2007 | $6,693,000 | $6,693,000 | $8,031,600 | $8,031,600 | Aspen Icarus |
| A700.COOLER | 1 | | Air Cooler | $27,200 | 2007 | $27,200 | $27,200 | $82,144 | $82,144 | Aspen Icarus |
| A700.GOXCLR-1 | 1 | | Oxygen Compressor Cooler | $23,300 | 2007 | $23,300 | $23,300 | $70,366 | $70,366 | Aspen Icarus |
| A700.GOXCLR-2 | 1 | | Oxygen Compressor Cooler | $23,000 | 2007 | $23,000 | $23,000 | $69,460 | $69,460 | Aspen Icarus |
| A700.GOXCMP-1 | 2 | | Oxygen Compressor | $1,489,600 | 2007 | $2,979,200 | $2,979,200 | $3,575,040 | $3,575,040 | Aspen Icarus |
| A700.HIGH-P.cond | 1 | | High Pressure Column Condenser | $20,300 | 2007 | $20,300 | $20,300 | $61,306 | $61,306 | Aspen Icarus |
| A700.HIGH-P.cond acc | 1 | | High Pressure Column Condenser Accumulator | $40,500 | 2007 | $40,500 | $40,500 | $122,310 | $122,310 | Aspen Icarus |
| A700.HIGH-P.reflux pump | 1 | 1 | High Pressure Column Reflux Pump | $14,300 | 2007 | $28,600 | $28,600 | $86,372 | $86,372 | Aspen Icarus |
| A700.HIGH-P.tower | 1 | | High Pressure Column Tower | $314,300 | 2007 | $314,300 | $314,300 | $949,186 | $949,186 | Aspen Icarus |
| A700.INTRC1 | 1 | | Air Compressor Intercooler | $338,300 | 2007 | $338,300 | $338,300 | $1,021,666 | $1,021,666 | Aspen Icarus |
| A700.INTRC2 | 1 | | Air Compressor Intercooler | $304,500 | 2007 | $304,500 | $304,500 | $919,590 | $919,590 | Aspen Icarus |
| A700.INTRC3 | 1 | | Air Compressor Intercooler | $222,500 | 2007 | $222,500 | $222,500 | $671,950 | $671,950 | Aspen Icarus |
| A700.LOW-P.reb | 1 | | Low Pressure Column Reboiler | $19,600 | 2007 | $19,600 | $19,600 | $59,192 | $59,192 | Aspen Icarus |
| A700.LOW-P.tower | 1 | | Low Pressure Column Tower | $2,581,600 | 2007 | $2,581,600 | $2,581,600 | $7,796,432 | $7,796,432 | Aspen Icarus |
| A700.TSA | 1 | | Water Knock-out Drum | $35,900 | 2007 | $35,900 | $35,900 | $108,418 | $108,418 | Aspen Icarus |
| A700.TURB-1 | 2 | | Gas Expander | $86,100 | 2007 | $172,200 | $172,200 | $520,044 | $520,044 | Aspen Icarus |
| A700.WK01 | 1 | | Water Knock-out Drum | $57,700 | 2007 | $57,700 | $57,700 | $174,254 | $174,254 | Aspen Icarus |
| **A700** | | | | | **Subtotal** | **$13,881,700** | **$13,881,700** | **$24,319,330** | **$24,319,330** | |
| | | | | Total | | $124,691,733 | $130,083,951 | $259,998,390 | $276,233,779 | |
| | | | | **Total (with BOP)** | | **$139,654,741** | **$145,694,026** | **$291,198,196** | **$309,381,833** | |

## *Low-Temperature Scenario Equipment Lists*

### Table B-4. Detailed Equipment List for Areas 100 and 200 of LT Scenario

| Equipment Number | Number Required | Number Spares | Equipment Name | Original Equip Cost (per unit) in Base Year | Base Year | Total Original Equip Cost (Req'd & Spare) in Base Year | Scaled Uninstalled Cost in 2007$ | Installed Cost Base Year | Installed Cost in 2007$ | Cost Source |
|---|---|---|---|---|---|---|---|---|---|---|
| A100.CONV1 | 2 | | Bale Transport Conveyor | $400,000 | 2000 | $800,000 | $1,066,531 | $1,296,000 | $1,727,781 | Aden et al. 2002 |
| A100.CONV2 | 2 | | Bale Unwrapping Conveyor | $150,000 | 2000 | $300,000 | $399,949 | $357,000 | $475,940 | Aden et al. 2002 |
| A100.CONV3 | 1 | | Belt Press Discharge Conveyor | $50,000 | 2000 | $50,000 | $66,658 | $94,500 | $125,984 | Aden et al. 2002 |
| A100.SCALE | 2 | | Truck Scales | $34,000 | 2000 | $68,000 | $90,655 | $167,960 | $223,918 | Aden et al. 2002 |
| A100.FORK1 | 4 | 1 | Truck Unloading Forklift | $18,000 | 2000 | $90,000 | $119,985 | $90,000 | $119,985 | Aden et al. 2002 |
| A100.FORK2 | 4 | | Bale Moving Forklift | $18,000 | 2000 | $72,000 | $95,988 | $72,000 | $95,988 | Aden et al. 2002 |
| A100.SLAB | 1 | | Concrete Feedstock-Storage Slab | $450,655 | 2000 | $450,655 | $600,797 | $991,441 | $1,321,754 | Aden et al. 2002 |
| A100.MAGSEP | 1 | | Magnetic Separator | $13,863 | 1998 | $13,863 | $18,700 | $18,022 | $24,310 | Aden et al. 2002 |
| A100.A100CHOP.CHGRIN01 | 4 | | Chopper | $105,100 | 2007 | $420,400 | $420,400 | $1,105,258 | $1,105,258 | Aspen Icarus |
| A100.A100CHOP.CHMIX01 | 1 | | Chopper Conveyor | $61,400 | 2007 | $61,400 | $61,400 | $185,428 | $185,428 | Aspen Icarus |
| A100.A100CHOP.CHSEP01 | 1 | | Chopper Screen with Recycle Conveyor | $20,800 | 2007 | $20,800 | $20,800 | $62,816 | $62,816 | Aspen Icarus |
| A100.A100DRY.DRDRY01 | 10 | | Dryer | $633,700 | 2007 | $6,337,000 | $6,337,000 | $15,201,647 | $15,201,647 | Aspen Icarus |
| A100.A100GRIN.GRGRIN01 | 4 | | Grinder | $167,100 | 2007 | $668,400 | $668,400 | $1,757,266 | $1,757,266 | Aspen Icarus |
| A100.A100GRIN.GRMIX01 | 1 | | Grinder Conveyor | $61,400 | 2007 | $61,400 | $61,400 | $185,428 | $185,428 | Aspen Icarus |
| A100.A100GRIN.GRSEP01 | 1 | | Grinder Screen with Recycle Conveyor | $20,800 | 2007 | $20,800 | $20,800 | $62,816 | $62,816 | Aspen Icarus |
| **A100** | | | | | **Subtotal** | **$9,434,718** | **$10,049,464** | **$21,647,582** | **$22,676,317** | |
| A200.A200COMB.CBCYC01 | 3 | | Combustor Cyclone (medium efficiency) | $35,400 | 2007 | $106,200 | $106,200 | $320,724 | $320,724 | Aspen Icarus |
| A200.A200COMB.CBCYC02 | 3 | | Combustor Cyclone (high efficiency) | $6,700 | 2007 | $20,100 | $20,100 | $60,702 | $60,702 | Aspen Icarus |
| A200.A200COMB.CBMIX01 | 1 | | Ash Storage Vessel | $142,800 | 2007 | $142,800 | $142,800 | $431,256 | $431,256 | Aspen Icarus |
| A200.A200COMB.CBREAC01 | 1 | | Combustor - Steam Boiler | $1,450,500 | 2007 | $1,450,500 | $1,450,500 | $4,380,510 | $4,380,510 | Aspen Icarus |
| A200.A200CYC.CYCYC01 | 2 | | 1st train, medium efficiency cyclone | $20,300 | 2007 | $40,600 | $40,600 | $122,612 | $122,612 | Aspen Icarus |
| A200.A200CYC.CYCYC02 | 4 | | 1st train, high efficiency cyclone | $24,900 | 2007 | $99,600 | $99,600 | $300,792 | $300,792 | Aspen Icarus |
| A200.A200CYC.CYMIX02 | 1 | | Char Collector and conveyor | $84,400 | 2007 | $84,400 | $84,400 | $254,888 | $254,888 | Aspen Icarus |
| A200.GSREAC01 | 7 | | Fluidized Bed Gasifier (Pressurized) | $1,096,170 | 2003 | $7,673,191 | $10,028,594 | $14,843,424 | $19,399,838 | Larson et al. 2005 |
| A200.GSTANK01 | 7 | | Biomass Receiving Hopper | $71,700 | 2007 | $501,900 | $501,900 | $1,247,712 | $1,247,712 | Aspen Icarus |
| A200.GSTANK02 | 7 | | Lockhopper | $47,700 | 2007 | $333,900 | $333,900 | $830,068 | $830,068 | Aspen Icarus |
| A200.GSTANK03 | 7 | | Biomass Feeding Bin | $47,700 | 2007 | $333,900 | $333,900 | $830,068 | $830,068 | Aspen Icarus |
| **A200** | | | | | **Subtotal** | **$10,787,091** | **$13,142,494** | **$23,622,756** | **$28,179,170** | |

# Table B-5. Detailed Equipment List for Areas 300, 400, and 500 of LT Scenario

| Equipment Number | Number Required | Number Spares | Equipment Name | Original Equip Cost (per unit) in Base Year | Base Year | Total Original Equip Cost (Req'd & Spare) in Base Year | Scaled Uninstalled Cost in 2007$ | Installed Cost Base Year | Installed Cost in 2007$ | Cost Source |
|---|---|---|---|---|---|---|---|---|---|---|
| A300.A300AGR.AGRarea | 1 | | High Pressure Amine System | $6,050,000 | 2005 | $6,050,000 | $6,789,129 | $18,271,000 | $20,503,168 | Phillips et al. 2007 |
| A300.A300SUL.SUCOL01 | 1 | | LO-CAT Absorber | $16,200 | 2007 | $16,200 | $16,200 | $48,924 | $48,924 | Aspen Icarus |
| A300.A300SUL.SUREAC01 | 1 | | LO-CAT Oxidizer Vessel | $1,000,000 | 2007 | $1,000,000 | $1,000,000 | $3,020,000 | $3,020,000 | Phillips et al. 2007 |
| A300.A300SUL.SUSEP01 | 1 | | Sulfur Separator | $16,200 | 2007 | $16,200 | $16,200 | $48,924 | $48,924 | Aspen Icarus |
| A300.CLCMP01 | 2 | | Carbon Dioxide Compressor | $1,176,900 | 2007 | $2,353,800 | $2,353,800 | $2,824,560 | $2,824,560 | Aspen Icarus |
| A300.CLHEAT01 | 2 | | Direct Quench Recycle Cooling | $188,800 | 2007 | $377,600 | $377,600 | $1,140,352 | $1,140,352 | Aspen Icarus |
| A300.CLHEAT02 | 1 | | Venturi Recycle Cooling | $91,500 | 2007 | $91,500 | $91,500 | $276,330 | $276,330 | Aspen Icarus |
| A300.CLMIX01 | 1 | | Venturi Scrubber | $26,800 | 2007 | $26,800 | $26,800 | $80,936 | $80,936 | Aspen Icarus |
| A300.CLSEP03 | 2 | | Direct Quench Syngas Cooler | $188,800 | 2007 | $377,600 | $377,600 | $1,140,352 | $1,140,352 | Aspen Icarus |
| A300.CLSEP04 | 1 | | Venturi Liquid Collection Tank | $74,500 | 2007 | $74,500 | $74,500 | $224,990 | $224,990 | Aspen Icarus |
| **A300** | | | | | **Subtotal** | **$10,384,200** | **$11,123,329** | **$27,076,368** | **$29,308,536** | |
| A400.A400COND.CDHEAT01 | 1 | | Syngas Heater | $60,500 | 2007 | $60,500 | $60,500 | $182,710 | $182,710 | Aspen Icarus |
| A400.A400COND.CDHEAT02 | 1 | | Syngas Pre-heater Furnace | $1,949,500 | 2007 | $1,949,500 | $1,949,500 | $5,887,490 | $5,887,490 | Aspen Icarus |
| A400.A400COND.CDHEAT03 | 1 | | Reformed Syngas Waste Heat Boiler | $396,600 | 2007 | $396,600 | $396,600 | $1,197,732 | $1,197,732 | Aspen Icarus |
| A400.A400COND.CDHEAT04 | 1 | | Syngas Cooler #2 | $41,200 | 2007 | $41,200 | $41,200 | $124,424 | $124,424 | Aspen Icarus |
| A400.A400COND.CDREAC01 | 1 | | Steam Methane Reformer | $1,650,800 | 2007 | $1,650,800 | $1,650,800 | $4,985,416 | $4,985,416 | Aspen Icarus |
| A400.A400COND.CDREAC02 | 1 | | Water Gas Shift Reactor | $136,600 | 2007 | $136,600 | $136,600 | $412,532 | $412,532 | Aspen Icarus |
| A400.A400COND.CDSEP01 | 2 | | ZnO Sulfur Removal Beds | $46,400 | 2007 | $92,800 | $92,800 | $280,256 | $280,256 | Aspen Icarus |
| A400.FSCOMP01 | 2 | | Booster Syngas Compressor | $921,600 | 2007 | $1,843,200 | $1,843,200 | $2,211,840 | $2,211,840 | Aspen Icarus |
| A400.FSCOMP02 | 1 | | Recycle Syngas Booster Compressor | $725,400 | 2007 | $725,400 | $725,400 | $870,480 | $870,480 | Aspen Icarus |
| A400.FSCOMP03 | 1 | | PSA Booster Compressor | $1,482,100 | 2007 | $1,482,100 | $1,482,100 | $1,778,520 | $1,778,520 | Aspen Icarus |
| A400.FSDRUM01 | 1 | | PSA Knock-out | $1,482,100 | 2007 | $1,482,100 | $1,482,100 | $4,475,942 | $4,475,942 | Aspen Icarus |
| A400.FSHEAT03 | 1 | | Syngas Cooler | $165,200 | 2007 | $165,200 | $165,200 | $498,904 | $498,904 | Aspen Icarus |
| A400.FSHEAT04 | 1 | | Recycle Syngas Pre-heater | $24,300 | 2007 | $24,300 | $24,300 | $73,386 | $73,386 | Aspen Icarus |
| A400.FSREAC01 | 1 | | Fischer-Tropsch Reactor | $7,303,889 | 2003 | $7,303,889 | $9,545,928 | $26,294,000 | $34,365,342 | Larson et al. 2005 |
| A400.FSSEP02 | 12 | | Pressure Swing Absorption Unit | $30,500 | 2007 | $366,000 | $366,000 | $1,105,320 | $1,105,320 | Aspen Icarus |
| A400.FSSEP03 | 1 | | FT knock-out Column | $72,100 | 2007 | $72,100 | $72,100 | $217,742 | $217,742 | Aspen Icarus |
| A400.FSSEP04 | 1 | | Water Separator | $39,200 | 2007 | $39,200 | $39,200 | $118,384 | $118,384 | Aspen Icarus |
| **A400** | | | | | **Subtotal** | **$17,792,289** | **$20,034,328** | **$50,596,694** | **$58,668,036** | |
| A500.HYREAC01 | 1 | | Hydrocracking/Hydrotreating Unit | $7,927,152 | 2007 | $7,927,152 | $7,927,152 | $23,940,000 | $23,940,000 | Robinson & Dolbear 2007 |
| A500.HYTANK01 | 1 | | Gasoline 30-day Storage Tank | $646,300 | 2007 | $646,300 | $646,300 | $1,951,826 | $1,951,826 | Aspen Icarus |
| A500.HYTANK02 | 1 | | Diesel 30-day Storage Tank | $1,200,700 | 2007 | $1,200,700 | $1,200,700 | $3,626,114 | $3,626,114 | Aspen Icarus |
| **A500** | | | | | **Subtotal** | **$9,774,152** | **$9,774,152** | **$29,517,940** | **$29,517,940** | |

# Table B-6. Detailed Equipment List for Areas 600 and 700 of LT Scenario

| Equipment Number | Number Required | Number Spares | Equipment Name | Original Equip Cost (per unit) in Base Year | Base Year | Total Original Equip Cost (Req'd & Spare) in Base Year | Scaled Uninstalled Cost in 2007$ | Installed Cost Base Year | Installed Cost in 2007$ | Cost Source |
|---|---|---|---|---|---|---|---|---|---|---|
| A600.COMBB | 1 | | Combustion Turbine - Electric Generator | $18,607,700 | 2007 | $18,607,700 | $18,607,700 | $22,329,240 | $22,329,240 | Aspen Icarus |
| A600.CWPUMP | 1 | 1 | Cooling Water Pump | $5,900 | 2007 | $11,800 | $11,800 | $35,636 | $35,636 | Aspen Icarus |
| A600.ECON1_HRSG | 1 | | Heat Recovery Steam Generator | $202,200 | 2007 | $202,200 | $202,200 | $610,644 | $610,644 | Aspen Icarus |
| A600.HPPUMP | 1 | 1 | High Pressure Steam Pump | $266,700 | 2007 | $533,400 | $533,400 | $1,610,868 | $1,610,868 | Aspen Icarus |
| A600.HPSEP | 1 | | High Pressure Steam/Water Separation | $107,400 | 2007 | $107,400 | $107,400 | $324,348 | $324,348 | Aspen Icarus |
| A600.LPEXP_ELECGEN | 1 | | Combined Steam Turbine - Electric Gen. | $5,056,300 | 2007 | $5,056,300 | $5,056,300 | $6,067,560 | $6,067,560 | Aspen Icarus |
| A600.LPSEP | 1 | | Low Pressure Water/Steam Separation | $108,800 | 2007 | $108,800 | $108,800 | $328,576 | $328,576 | Aspen Icarus |
| A600.O2COMP | 1 | | Air Compressor | $6,331,200 | 2007 | $6,331,200 | $6,331,200 | $7,597,440 | $7,597,440 | Aspen Icarus |
| **A600** | | | | | **Subtotal** | **$30,958,800** | **$30,958,800** | **$38,904,312** | **$38,904,312** | |
| A700.COMP1 | 2 | | Air Compressor | $3,119,600 | 2007 | $6,239,200 | $6,239,200 | $7,487,040 | $7,487,040 | Aspen Icarus |
| A700.COOLER | 1 | | Air Cooler | $24,300 | 2007 | $24,300 | $24,300 | $73,386 | $73,386 | Aspen Icarus |
| A700.GOXCLR-1 | 1 | | Oxygen Compressor Cooler | $23,300 | 2007 | $23,300 | $23,300 | $70,366 | $70,366 | Aspen Icarus |
| A700.GOXCLR-2 | 1 | | Oxygen Compressor Cooler | $23,000 | 2007 | $23,000 | $23,000 | $69,460 | $69,460 | Aspen Icarus |
| A700.GOXCMP-1 | 2 | | Oxygen Compressor | $1,514,700 | 2007 | $3,029,400 | $3,029,400 | $3,635,280 | $3,635,280 | Aspen Icarus |
| A700.HIGH-P.cond | 1 | | High Pressure Column Condenser | $20,300 | 2007 | $20,300 | $20,300 | $61,306 | $61,306 | Aspen Icarus |
| A700.HIGH-P.cond acc | 1 | | High Pressure Column Condenser Accumulator | $36,300 | 2007 | $36,300 | $36,300 | $109,626 | $109,626 | Aspen Icarus |
| A700.HIGH-P.reflux pump | 1 | 1 | High Pressure Column Reflux Pump | $14,300 | 2007 | $28,600 | $28,600 | $34,320 | $34,320 | Aspen Icarus |
| A700.HIGH-P.tower | 1 | | High Pressure Column Tower | $279,900 | 2007 | $279,900 | $279,900 | $335,880 | $335,880 | Aspen Icarus |
| A700.INTRC1 | 1 | | Air Compressor Intercooler | $338,300 | 2007 | $338,300 | $338,300 | $405,960 | $405,960 | Aspen Icarus |
| A700.INTRC2 | 1 | | Air Compressor Intercooler | $304,500 | 2007 | $304,500 | $304,500 | $919,590 | $919,590 | Aspen Icarus |
| A700.INTRC3 | 1 | | Air Compressor Intercooler | $222,500 | 2007 | $222,500 | $222,500 | $671,950 | $671,950 | Aspen Icarus |
| A700.LOW-P.reb | 2 | | Low Pressure Column Reboiler | $19,600 | 2007 | $39,200 | $39,200 | $118,384 | $118,384 | Aspen Icarus |
| A700.LOW-P.tower | 1 | | Low Pressure Column Tower | $1,538,900 | 2007 | $1,538,900 | $1,538,900 | $4,647,478 | $4,647,478 | Aspen Icarus |
| A700.TSA | 1 | | Water Knock-out Drum | $30,100 | 2007 | $30,100 | $30,100 | $90,902 | $90,902 | Aspen Icarus |
| A700.TURB-1 | 2 | | Gas Expander | $89,200 | 2007 | $178,400 | $178,400 | $538,768 | $538,768 | Aspen Icarus |
| A700.WK01 | 1 | | Water Knock-out Drum | $64,800 | 2007 | $64,800 | $64,800 | $195,696 | $195,696 | Aspen Icarus |
| **A700** | | | | | **Subtotal** | **$12,421,000** | **$12,421,000** | **$19,465,392** | **$19,465,392** | |
| | | | | Total | | $101,552,251 | $107,503,567 | $210,831,043 | $226,719,704 | |
| | | | | **Total (with BOP)** | | **$113,738,521** | **$120,403,995** | **$236,130,768** | **$253,926,068** | |

# Discounted Cash Flow

## *High-Temperature Scenario*

### Table B-7. Discounted Cash Flow Sheet for Construction Period and Years 1-8 of HT Scenario

**DCFROR Worksheet**

| Year | | -2 | -1 | 0 | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Fixed Capital Investment | | $50,890,395 | $316,115,651 | $168,595,014 | | | | | | | | |
| Working Capital | | | | $79,028,913 | | | | | | | | |
| Loan Payment | | | | | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| Loan Interest Payment | | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| Loan Principal | | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| GGE (Gallon of Gasoline Equiv.) Sales | | | | | $133,364,635 | $177,819,513 | $177,819,513 | $177,819,513 | $177,819,513 | $177,819,513 | $177,819,513 | $177,819,513 |
| Diesel Sales | | | | | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| By-Product Credit | | | | | $4,173,208 | $5,564,277 | $5,564,277 | $5,564,277 | $5,564,277 | $5,564,277 | $5,564,277 | $5,564,277 |
| Plant Performance | | | | | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 |
| Total Annual Sales | | | | | $137,537,843 | $183,383,791 | $183,383,791 | $183,383,791 | $183,383,791 | $183,383,791 | $183,383,791 | $183,383,791 |
| Annual Manufacturing Cost | | | | | | | | | | | | |
| Raw Materials | | | | | $44,894,145 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 |
| SWGS catalysts | | | | | $114,621 | $0 | $0 | $114,621 | $0 | $0 | $114,621 | $0 |
| Steam reforming catalysts | | | | | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| ZnO | | | | | $424,410 | $0 | $0 | $424,410 | $0 | $0 | $424,410 | $0 |
| Pressure Swing Adsorption Packing | | | | | $497,135 | $0 | $0 | $497,135 | $0 | $0 | $497,135 | $0 |
| FT catalysts | | | | | $7,686,720 | $0 | $0 | $7,686,720 | $0 | $0 | $7,686,720 | $0 |
| Other Variable Costs | | | | | $11,727,856 | $13,403,264 | $13,403,264 | $13,403,264 | $13,403,264 | $13,403,264 | $13,403,264 | $13,403,264 |
| Fixed Operating Costs | | | | | $14,345,785 | $14,345,785 | $14,345,785 | $14,345,785 | $14,345,785 | $14,345,785 | $14,345,785 | $14,345,785 |
| Total Product Cost | | | | | $79,690,672 | $79,056,643 | $79,056,643 | $87,779,529 | $79,056,643 | $79,056,643 | $87,779,529 | $79,056,643 |
| Annual Depreciation | | | | | | | | | | | | |
| General Plant | | | | | | | | | | | | |
| DDB | | | | | $128,361,546 | $91,686,818 | $65,490,585 | $46,778,989 | $33,413,564 | $23,866,831 | $17,047,737 | |
| SL | | | | | $64,180,773 | $53,483,977 | $45,843,409 | $40,931,615 | $38,982,491 | $38,982,491 | $38,982,491 | |
| Remaining Value | | | | | $320,903,864 | $229,217,046 | $163,726,461 | $116,947,472 | $83,533,909 | $59,667,078 | $42,619,341 | |
| Actual | | | | | $128,361,546 | $91,686,818 | $65,490,585 | $46,778,989 | $38,982,491 | $38,982,491 | $38,982,491 | |
| Steam Plant | | | | | | | | | | | | |
| DDB | | | | | $5,819,551 | $5,383,084 | $4,979,353 | $4,605,902 | $4,260,459 | $3,940,925 | $3,645,355 | $3,371,954 |
| SL | | | | | $3,879,700 | $3,777,603 | $3,688,410 | $3,612,472 | $3,550,382 | $3,503,044 | $3,471,767 | $3,458,414 |
| Remaining Value | | | | | $71,774,458 | $66,391,374 | $61,412,021 | $56,806,119 | $52,545,660 | $48,604,736 | $44,959,380 | $41,587,427 |
| Actual | | | | | $5,819,551 | $5,383,084 | $4,979,353 | $4,605,902 | $4,260,459 | $3,940,925 | $3,645,355 | $3,458,414 |
| Net Revenue | | | | | ($76,333,925) | $7,257,245 | $33,857,210 | $44,219,371 | $61,084,198 | $61,403,732 | $52,976,416 | $100,868,733 |
| Losses Forward | | | | | | ($76,333,925) | ($69,076,681) | ($35,219,471) | $0 | $0 | $0 | $0 |
| Taxable Income | | | | | ($76,333,925) | ($69,076,681) | ($35,219,471) | $8,999,900 | $61,084,198 | $61,403,732 | $52,976,416 | $100,868,733 |
| Income Tax | | | | | $0 | $0 | $0 | $3,509,961 | $23,822,837 | $23,947,455 | $20,660,802 | $39,338,806 |
| Annual Cash Income | | | | | $57,847,171 | $104,327,147 | $104,327,147 | $92,094,301 | $80,504,310 | $80,379,692 | $74,943,459 | $64,988,341 |
| Discount Factor | | 1.21 | 1.1 | 1 | 0.909090909 | 0.826446281 | 0.751314801 | 0.683013455 | 0.620921323 | 0.56447393 | 0.513158118 | 0.46650738 |
| Annual Present Value | $645,181,377 | | | | $52,588,337 | $86,220,783 | $78,382,530 | $62,901,646 | $49,986,843 | $45,372,241 | $38,457,845 | $30,317,541 |
| Total Capital Investment + Interest | | $61,577,378 | $347,727,216 | $247,623,927 | | | | | | | | |
| Net Present Worth | | | | $0 | | | | | | | | |

# Table B-8. Discounted Cash Flow Sheet for Years 9-20 of HT Scenario

**DCFROR Worksheet**

| Year | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Fixed Capital Investment | | | | | | | | | | | | |
| Working Capital | | | | | | | | | | | | ($79,028,913) |
| Loan Payment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| Loan Interest Payment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| Loan Principal | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| GGE (Gallon of Gasoline Equiv.) Sales | $177,819,513 | $177,819,513 | $177,819,513 | $177,819,513 | $177,819,513 | $177,819,513 | $177,819,513 | $177,819,513 | $177,819,513 | $177,819,513 | $177,819,513 | $177,819,513 |
| Diesel Sales | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| By-Product Credit | $5,564,277 | $5,564,277 | $5,564,277 | $5,564,277 | $5,564,277 | $5,564,277 | $5,564,277 | $5,564,277 | $5,564,277 | $5,564,277 | $5,564,277 | $5,564,277 |
| Plant Performance | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 |
| Total Annual Sales | $183,383,791 | $183,383,791 | $183,383,791 | $183,383,791 | $183,383,791 | $183,383,791 | $183,383,791 | $183,383,791 | $183,383,791 | $183,383,791 | $183,383,791 | $183,383,791 |
| Annual Manufacturing Cost | | | | | | | | | | | | |
| Raw Materials | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 |
| SWGS catalysts | $0 | $114,621 | $0 | $0 | $114,621 | $0 | $0 | $114,621 | $0 | $0 | $114,621 | $0 |
| Steam reforming catalysts | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| ZnO | $0 | $424,410 | $0 | $0 | $424,410 | $0 | $0 | $424,410 | $0 | $0 | $424,410 | $0 |
| Pressure Swing Adsorption Packing | $0 | $497,135 | $0 | $0 | $497,135 | $0 | $0 | $497,135 | $0 | $0 | $497,135 | $0 |
| FT catalysts | $0 | $7,686,720 | $0 | $0 | $7,686,720 | $0 | $0 | $7,686,720 | $0 | $0 | $7,686,720 | $0 |
| Other Variable Costs | $13,403,264 | $13,403,264 | $13,403,264 | $13,403,264 | $13,403,264 | $13,403,264 | $13,403,264 | $13,403,264 | $13,403,264 | $13,403,264 | $13,403,264 | $13,403,264 |
| Fixed Operating Costs | $14,345,785 | $14,345,785 | $14,345,785 | $14,345,785 | $14,345,785 | $14,345,785 | $14,345,785 | $14,345,785 | $14,345,785 | $14,345,785 | $14,345,785 | $14,345,785 |
| Total Product Cost | $79,056,643 | $87,779,529 | $79,056,643 | $79,056,643 | $87,779,529 | $79,056,643 | $79,056,643 | $87,779,529 | $79,056,643 | $79,056,643 | $87,779,529 | $79,056,643 |
| Annual Depreciation | | | | | | | | | | | | |
| General Plant | | | | | | | | | | | | |
| DDB | | | | | | | | | | | | |
| SL | | | | | | | | | | | | |
| Remaining Value | | | | | | | | | | | | |
| Actual | | | | | | | | | | | | |
| Steam Plant | | | | | | | | | | | | |
| DDB | $3,119,057 | $2,885,128 | $2,668,743 | $2,468,587 | $2,283,443 | $2,112,185 | $1,953,771 | $1,807,238 | $1,671,696 | $1,546,318 | $1,430,344 | $1,323,069 |
| SL | $3,458,414 | $3,458,414 | $3,458,414 | $3,458,414 | $3,458,414 | $3,458,414 | $3,458,414 | $3,458,414 | $3,458,414 | $3,458,414 | $3,458,414 | $3,458,414 |
| Remaining Value | $38,468,370 | $35,583,242 | $32,914,499 | $30,445,912 | $28,162,468 | $26,050,283 | $24,096,512 | $22,289,273 | $20,617,578 | $19,071,260 | $17,640,915 | $16,317,847 |
| Actual | $3,458,414 | $3,458,414 | $3,458,414 | $3,458,414 | $3,458,414 | $3,458,414 | $3,458,414 | $3,458,414 | $3,458,414 | $3,458,414 | $3,458,414 | $3,458,414 |
| Net Revenue | $100,868,733 | $92,145,848 | $100,868,733 | $100,868,733 | $92,145,848 | $100,868,733 | $100,868,733 | $92,145,848 | $100,868,733 | $100,868,733 | $92,145,848 | $100,868,733 |
| Losses Forward | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| Taxable Income | $100,868,733 | $92,145,848 | $100,868,733 | $100,868,733 | $92,145,848 | $100,868,733 | $100,868,733 | $92,145,848 | $100,868,733 | $100,868,733 | $92,145,848 | $100,868,733 |
| Income Tax | $39,338,806 | $35,936,881 | $39,338,806 | $39,338,806 | $35,936,881 | $39,338,806 | $39,338,806 | $35,936,881 | $39,338,806 | $39,338,806 | $35,936,881 | $39,338,806 |
| Annual Cash Income | $64,988,341 | $59,667,381 | $64,988,341 | $64,988,341 | $59,667,381 | $64,988,341 | $64,988,341 | $59,667,381 | $64,988,341 | $64,988,341 | $59,667,381 | $64,988,341 |
| Discount Factor | 0.424097618 | 0.385543289 | 0.350493899 | 0.318630818 | 0.28966438 | 0.263331254 | 0.239392049 | 0.217629136 | 0.197844669 | 0.17985879 | 0.163507991 | 0.148643628 |
| Annual Present Value | $27,561,401 | $23,004,358 | $22,778,017 | $20,707,288 | $17,283,515 | $17,113,461 | $15,557,692 | $12,985,361 | $12,857,597 | $11,688,724 | $9,756,094 | $9,660,103 |
| Total Capital Investment + Interest | | | | | | | | | | | | ($11,747,144.32) |
| Net Present Worth | | | | | | | | | | | | |

## *Low-Temperature Scenario*

### Table B-9. Discounted Cash Flow Sheet for Construction Period and Years 1-8 of LT Scenario

**DCFROR Worksheet**

| Year | | -2 | -1 | 0 | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Fixed Capital Investment | | $41,888,460 | $259,981,649 | $138,656,880 | | | | | | | | |
| Working Capital | | | | $64,995,412 | | | | | | | | |
| Loan Payment | | | | | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| Loan Interest Payment | | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| Loan Principal | | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| GGE (Gallon of Gasoline Equiv.) Sales | | | | | $117,025,289 | $156,033,719 | $156,033,719 | $156,033,719 | $156,033,719 | $156,033,719 | $156,033,719 | $156,033,719 |
| Diesel Sales | | | | | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| By-Product Credit | | | | | $4,945,498 | $6,593,998 | $6,593,998 | $6,593,998 | $6,593,998 | $6,593,998 | $6,593,998 | $6,593,998 |
| Plant Performance | | | | | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 |
| Total Annual Sales | | | | | $121,970,788 | $162,627,717 | $162,627,717 | $162,627,717 | $162,627,717 | $162,627,717 | $162,627,717 | $162,627,717 |
| Annual Manufacturing Cost | | | | | | | | | | | | |
| Raw Materials | | | | | $44,894,145 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 |
| WGS catalysts | | | | | $104,732 | $0 | $0 | $104,732 | $0 | $0 | $104,732 | $0 |
| Steam reforming catalysts | | | | | $103,412 | $0 | $0 | $103,412 | $0 | $0 | $103,412 | $0 |
| ZnO | | | | | $424,410 | $0 | $0 | $424,410 | $0 | $0 | $424,410 | $0 |
| Pressure Swing Adsorption Packing | | | | | $497,135 | $0 | $0 | $497,135 | $0 | $0 | $497,135 | $0 |
| FT catalysts | | | | | $6,127,680 | $0 | $0 | $6,127,680 | $0 | $0 | $6,127,680 | $0 |
| Other Variable Costs | | | | | $11,238,097 | $12,843,539 | $12,843,539 | $12,843,539 | $12,843,539 | $12,843,539 | $12,843,539 | $12,843,539 |
| Fixed Operating Costs | | | | | $12,404,834 | $12,404,834 | $12,404,834 | $12,404,834 | $12,404,834 | $12,404,834 | $12,404,834 | $12,404,834 |
| Total Product Cost | | | | | $75,794,444 | $76,555,967 | $76,555,967 | $83,813,336 | $76,555,967 | $76,555,967 | $83,813,336 | $76,555,967 |
| Annual Depreciation | | | | | | | | | | | | |
| General Plant | | | | | | | | | | | | |
| DDB | | | | | $104,833,121 | $74,880,801 | $53,486,286 | $38,204,490 | $27,288,922 | $19,492,087 | $13,922,919 | |
| SL | | | | | $52,416,561 | $43,680,467 | $37,440,400 | $33,428,929 | $31,837,075 | $31,837,075 | $31,837,075 | |
| Remaining Value | | | | | $262,082,803 | $187,202,002 | $133,715,716 | $95,511,226 | $68,222,304 | $48,730,217 | $34,807,298 | |
| Actual | | | | | $104,833,121 | $74,880,801 | $53,486,286 | $38,204,490 | $31,837,075 | $31,837,075 | $31,837,075 | |
| Steam Plant | | | | | | | | | | | | |
| DDB | | | | | $4,979,012 | $4,605,586 | $4,260,167 | $3,940,654 | $3,645,105 | $3,371,722 | $3,118,843 | $2,884,930 |
| SL | | | | | $3,319,341 | $3,231,990 | $3,155,679 | $3,090,709 | $3,037,588 | $2,997,087 | $2,970,327 | $2,958,903 |
| Remaining Value | | | | | $61,407,813 | $56,802,227 | $52,542,060 | $48,601,405 | $44,956,300 | $41,584,577 | $38,465,734 | $35,580,804 |
| Actual | | | | | $4,979,012 | $4,605,586 | $4,260,167 | $3,940,654 | $3,645,105 | $3,371,722 | $3,118,843 | $2,958,903 |
| Net Revenue | | | | | ($63,635,790) | $6,585,363 | $28,325,297 | $36,669,236 | $50,589,570 | $50,862,953 | $43,858,462 | $83,112,848 |
| Losses Forward | | | | | | ($63,635,790) | ($57,050,426) | ($28,725,129) | $0 | $0 | $0 | $0 |
| Taxable Income | | | | | ($63,635,790) | ($57,050,426) | ($28,725,129) | $7,944,107 | $50,589,570 | $50,862,953 | $43,858,462 | $83,112,848 |
| Income Tax | | | | | $0 | $0 | $0 | $3,098,202 | $19,729,932 | $19,836,551 | $17,104,800 | $32,414,011 |
| Annual Cash Income | | | | | $46,176,343 | $86,071,750 | $86,071,750 | $75,716,179 | $66,341,818 | $66,235,199 | $61,709,581 | $53,657,740 |
| Discount Factor | | 1.21 | 1.1 | 1 | 0.909090909 | 0.826446281 | 0.751314801 | 0.683013455 | 0.620921323 | 0.56447393 | 0.513158118 | 0.46650738 |
| Annual Present Value | $530,655,988 | | | | $41,978,494 | $71,133,678 | $64,666,980 | $51,715,169 | $41,193,049 | $37,388,043 | $31,666,772 | $25,031,732 |
| Total Capital Investment + Interest | | $50,685,036 | $285,979,814 | $203,652,292 | | | | | | | | |
| Net Present Worth | | | | $0 | | | | | | | | |

# Table B-10. Discounted Cash Flow Sheet for Years 9-20 of LT Scenario

**DCFROR Worksheet**

| Year | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Fixed Capital Investment | | | | | | | | | | | | |
| Working Capital | | | | | | | | | | | | ($64,995,412) |
| Loan Payment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| Loan Interest Payment | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| Loan Principal | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| GGE (Gallon of Gasoline Equiv.) Sales | $156,033,719 | $156,033,719 | $156,033,719 | $156,033,719 | $156,033,719 | $156,033,719 | $156,033,719 | $156,033,719 | $156,033,719 | $156,033,719 | $156,033,719 | $156,033,719 |
| Diesel Sales | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| By-Product Credit | $6,593,998 | $6,593,998 | $6,593,998 | $6,593,998 | $6,593,998 | $6,593,998 | $6,593,998 | $6,593,998 | $6,593,998 | $6,593,998 | $6,593,998 | $6,593,998 |
| Plant Performance | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 |
| Total Annual Sales | $162,627,717 | $162,627,717 | $162,627,717 | $162,627,717 | $162,627,717 | $162,627,717 | $162,627,717 | $162,627,717 | $162,627,717 | $162,627,717 | $162,627,717 | $162,627,717 |
| Annual Manufacturing Cost | | | | | | | | | | | | |
| Raw Materials | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 | $51,307,594 |
| WGS catalysts | $0 | $104,732 | $0 | $0 | $104,732 | $0 | $0 | $104,732 | $0 | $0 | $104,732 | $0 |
| Steam reforming catalysts | $0 | $103,412 | $0 | $0 | $103,412 | $0 | $0 | $103,412 | $0 | $0 | $103,412 | $0 |
| ZnO | $0 | $424,410 | $0 | $0 | $424,410 | $0 | $0 | $424,410 | $0 | $0 | $424,410 | $0 |
| Pressure Swing Adsorption Packing | $0 | $497,135 | $0 | $0 | $497,135 | $0 | $0 | $497,135 | $0 | $0 | $497,135 | $0 |
| FT catalysts | $0 | $6,127,680 | $0 | $0 | $6,127,680 | $0 | $0 | $6,127,680 | $0 | $0 | $6,127,680 | $0 |
| Other Variable Costs | $12,843,539 | $12,843,539 | $12,843,539 | $12,843,539 | $12,843,539 | $12,843,539 | $12,843,539 | $12,843,539 | $12,843,539 | $12,843,539 | $12,843,539 | $12,843,539 |
| Fixed Operating Costs | $12,404,834 | $12,404,834 | $12,404,834 | $12,404,834 | $12,404,834 | $12,404,834 | $12,404,834 | $12,404,834 | $12,404,834 | $12,404,834 | $12,404,834 | $12,404,834 |
| Total Product Cost | $76,555,967 | $83,813,336 | $76,555,967 | $76,555,967 | $83,813,336 | $76,555,967 | $76,555,967 | $83,813,336 | $76,555,967 | $76,555,967 | $83,813,336 | $76,555,967 |
| Annual Depreciation | | | | | | | | | | | | |
| General Plant | | | | | | | | | | | | |
| DDB | | | | | | | | | | | | |
| SL | | | | | | | | | | | | |
| Remaining Value | | | | | | | | | | | | |
| Actual | | | | | | | | | | | | |
| Steam Plant | | | | | | | | | | | | |
| DDB | $2,668,560 | $2,468,418 | $2,283,287 | $2,112,040 | $1,953,637 | $1,807,115 | $1,671,581 | $1,546,212 | $1,430,246 | $1,322,978 | $1,223,755 | $1,131,973 |
| SL | $2,958,903 | $2,958,903 | $2,958,903 | $2,958,903 | $2,958,903 | $2,958,903 | $2,958,903 | $2,958,903 | $2,958,903 | $2,958,903 | $2,958,903 | $2,958,903 |
| Remaining Value | $32,912,244 | $30,443,825 | $28,160,538 | $26,048,498 | $24,094,861 | $22,287,746 | $20,616,165 | $19,069,953 | $17,639,706 | $16,316,728 | $15,092,974 | $13,961,001 |
| Actual | $2,958,903 | $2,958,903 | $2,958,903 | $2,958,903 | $2,958,903 | $2,958,903 | $2,958,903 | $2,958,903 | $2,958,903 | $2,958,903 | $2,958,903 | $2,958,903 |
| Net Revenue | $83,112,848 | $75,855,478 | $83,112,848 | $83,112,848 | $75,855,478 | $83,112,848 | $83,112,848 | $75,855,478 | $83,112,848 | $83,112,848 | $75,855,478 | $83,112,848 |
| Losses Forward | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 | $0 |
| Taxable Income | $83,112,848 | $75,855,478 | $83,112,848 | $83,112,848 | $75,855,478 | $83,112,848 | $83,112,848 | $75,855,478 | $83,112,848 | $83,112,848 | $75,855,478 | $83,112,848 |
| Income Tax | $32,414,011 | $29,583,637 | $32,414,011 | $32,414,011 | $29,583,637 | $32,414,011 | $32,414,011 | $29,583,637 | $32,414,011 | $32,414,011 | $29,583,637 | $32,414,011 |
| Annual Cash Income | $53,657,740 | $49,230,744 | $53,657,740 | $53,657,740 | $49,230,744 | $53,657,740 | $53,657,740 | $49,230,744 | $53,657,740 | $53,657,740 | $49,230,744 | $53,657,740 |
| Discount Factor | 0.424097618 | 0.385543289 | 0.350493899 | 0.318630818 | 0.28966438 | 0.263331254 | 0.239392049 | 0.217629136 | 0.197844669 | 0.17985879 | 0.163507991 | 0.148643628 |
| Annual Present Value | $22,756,120 | $18,980,583 | $18,806,710 | $17,097,009 | $14,260,393 | $14,129,760 | $12,845,236 | $10,714,044 | $10,615,898 | $9,650,816 | $8,049,620 | $7,975,881 |
| Total Capital Investment + Interest | | | | | | | | | | | | ($9,661,153.89) |
| Net Present Worth | | | | | | | | | | | | |

# Appendix C. Scenario Modeling Details

## Property Method

The model setup includes a particle size distribution in order to better estimate the solids simulation in the grinding and cyclone operations. It operates globally with the Redlich-Kwong-Soave with Boston-Mathias modification (RKS-BM) property method, which is recommended for medium-temperature refining and gas processing operations including combustion and gasification. During acid-gas absorption and stripping, another property method, ELECNRTL, is used for more accurate simulation. For solids handling, such as in the pretreatment area and cyclones, the SOLIDS property method is used.

## 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 (e.g., Area 200 gasification) has a two letter abbreviation (e.g., gasification is 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 C-1 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 C-1. Stream nomenclature used in model**

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

| Area | | Description | | | | Number | |
|---|---|---|---|---|---|---|---|
| G | S | R | E | A | C | 0 | 1 |

**Figure C-2. Block nomenclature used in model**

Table C-1 contains the abbreviations for areas, unit operation block descriptions, and stream descriptions.

**Table C-1. Detailed Description of Stream and Block Nomenclature**

| Area | Description | Name | Block | Name | Stream | Name |
|---|---|---|---|---|---|---|
| **Plant** | All Areas | **PL** | Reactor | **REAC** | Biomass | **BMAS** |
| **A100** | Pretreatment | **PR** | Mixer | **MIX** | Steam | **STM** |
| **A100CHOP** | Chopping | **CH** | Heat Mixer | **QMX** | Flue Gas | **FLUE** |
| **A100DRY** | Drying | **DR** | Work Mixer | **WMX** | Syngas | **SGAS** |
| **A100GRIN** | Grinding | **GR** | Splitter | **SPL** | Ash | **ASH** |
| **A200** | Gasification | **GS,SL** | Separator | **SEP** | Carbon Dioxide | **$CO_2$** |
| **A200CYC** | Cyclones | **CY** | Cyclones | **CYC** | Air | **AIR** |
| **A200COMB** | Combustion | **CB** | Flash Drum | **DRUM** | Hydrogen | **HYD** |
| **A300** | Syngas Cleaning | **CL** | Column | **COL** | FT Products | **FT** |
| **A300AGR** | Acid Gas Removal | **AG** | Distillation | **DIST** | Water | **WAT** |
| **A300SUL** | Sulfur Recovery | **SU** | Grinder | **GRIN** | Oxygen | **OX** |
| **A400** | Fuel Synthesis | **FS** | Dryer | **DRY** | Sulfur | **SUL** |
| **A400COND** | Syngas Conditioning | **CD** | Heater | **HEAT** | Fuel | **FUEL** |
| **A400MTG** | Methanol to Gasoline | **MG** | Heat Exchanger | **HX** | Tar | **TAR** |
| **A500** | Hydrocracking | **HY** | Tank/Hopper | **TANK** | Char | **CHAR** |
| **A500** | Fuel Separation | **SE** | Pump | **PMP** | Acid Gas | **AG** |
| **A600** | Power Generation | **PG** | Compressor | **COMP** | Lean MEA Soln. | **MEAL** |
| **A700** | Air Separation Unit | | Turbine | **TURB** | Rich MEA Soln. | **MEAR** |
| | | | | | Light Gases | **LGAS** |
| | | | | | Nitrogen | **NTGN** |

A special notation is used for heat and work streams. In the case that the first reactor in the gasification area includes a heat stream leaving the unit, it follows the nomenclature shown in Figure C-3.

| Q or W | | Area | | Block Description | | | Number |
|---|---|---|---|---|---|---|---|
| Q | - | G | S | R | E | A | 1 |

**Figure C-3. Heat and work stream 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.

## Aspen Plus Calculator Block Descriptions

### High-Temperature Scenario

#### *AIRCOMB*

This block calculates the nitrogen that accompanies the oxygen in the air inlet for the combustion of unconverted syngas. Molar nitrogen flow (in kmol/h) is calculated as follows:

$$\dot{M}_{N2} = \left(\frac{0.79}{0.21}\right) \cdot \dot{M}_{O2} \qquad \text{(eqn. 20)}$$

where $\dot{M}_{O2}$ is molar flow of oxygen in kmol/h.

#### *AMINE*

This block calculates the molar flow of monoethanolamine (MEA) needed for the required acid gas removal ($CO_2$ and $H_2S$) arriving from syngas quench and FT unconverted syngas recycle stream. The MEA is able to capture 0.35 moles acid gas per mole MEA. Additionally, the MEA is diluted as explained in DILUTH2O.

Molar MEA flow (in kmol/h) is calculated by

$$\dot{M}_{MEA} = (\dot{M}_{CO2,syn} + \dot{M}_{CO2,rec} + \dot{M}_{H2S,syn})/0.35 \qquad \text{(eqn. 21)}$$

where $\dot{M}_{CO2,syn}$ is molar flow of $CO_2$ from the syngas after the syngas quench, $\dot{M}_{CO2,rec}$ is the molar flow of $CO_2$ from the unconverted syngas recycle after the FT synthesis, and $\dot{M}_{H2S,syn}$ is the molar flow rate of $H_2S$ from the syngas quench.

Because the MEA solution in the amine absorption unit is to be 20 wt% concentrated with water, the flow of water must be calculated. Mole flow of water is calculated as

$$\dot{M}_{H2O} = \frac{\dot{M}_{MEA} * MW_{MEA}/0.20}{MW_{H2O}} \qquad \text{(eqn. 22)}$$

#### *BIOELEM*

Because the high-temperature gasifier is modeled at equilibrium, the simulation software requires that all components in the input are located in the conventional stream. Therefore, this block splits the biomass into the following compounds based on its ultimate analysis: carbon, hydrogen, oxygen, sulfur, nitrogen, and ash. Water in the biomass is not affected because it is already a conventional component. Biomass in the exit stream is set to zero.

***FTDISTR***

This calculator block calculates an alpha chain growth parameter using the equation by Song et al. (2004) for cobalt catalyst. Inlet and outlet streams are defined and calculated. FT products include paraffins from C1 through C20. FT waxes are paraffins at C30.

FT reaction is as follows:

$$\mathrm{CO} + 2.1 * \mathrm{H_2} \; -- > \; -(C\mathrm{H_2})- \; + \mathrm{H_2O} \qquad \text{(eqn. 23)}$$

Section 100 sets the CO conversion.

Section 200 calculates the reaction extent (in lbmol) based on an alpha value of 0.9.

-------------Section 100------------------------

Percent conversion of CO is calculated as follows, and then the molar amount of converted CO (COCONV) is calculated knowing the molar amount of CO entering (COIN).

PERCEN = 40

CONV=PERCEN/100.0

COCONV=COIN*CONV

------------Section 200-------------------------

R1, R2, R3, etc. represent the molar reaction extent (lbmol/h) that is utilized in the FT reactor for each reaction (i.e., CO + 3*H2 → CH4 + H2O, 2*CO + 5*H2 → C2H6 + 2*H2O, etc.). The coefficients of each reaction extent are calculated by solving a set of 21 equations shown below in Table C-2 and as described in the Detailed Calculations section of this Appendix.

**Table C-2. Reaction Extent Equations for Each Alkane Hydrocarbon**

| Alkane Component | Equation |
|---|---|
| C1 | $R1 = COCONV * 0.01$ |
| C2 | $R2 = COCONV * 0.018/2$ |
| C3 | $R3 = COCONV * 0.0243/3$ |
| C4 | $R4 = COCONV * 0.02916/4$ |
| C5 | $R5 = COCONV * 0.0328/5$ |
| C6 | $R6 = COCONV * 0.03543/6$ |
| C7 | $R7 = COCONV * 0.0372/7$ |
| C8 | $R8 = COCONV * 0.03826/8$ |
| C9 | $R9 = COCONV * 0.03874/9$ |
| C10 | $R10 = COCONV * 0.03874/10$ |
| C11 | $R11 = COCONV * 0.03835/11$ |
| C12 | $R12 = COCONV * 0.03766/12$ |
| C13 | $R13 = COCONV * 0.03672/13$ |
| C14 | $R14 = COCONV * 0.03559/14$ |
| C15 | $R15 = COCONV * 0.03432/15$ |
| C16 | $R16 = COCONV * 0.03294/16$ |
| C17 | $R17 = COCONV * 0.0315/17$ |
| C18 | $R18 = COCONV * 0.03002/18$ |
| C19 | $R19 = COCONV * 0.02852/19$ |
| C20 | $R20 = COCONV * 0.02702/20$ |
| C30 | $R21 = COCONV * 0.36473/30$ |

### *GRIND*

This block calculates the power requirement (in kW) for grinding the biomass from the chop size of 15 mm to the final size of 1 mm. This power requirement data is found in Mani et al. and is for 12% exiting moisture. The correlation was changed from a polynomial (quadratic) regression, which Mani et al. used, to a power regression because the power regression more accurately matched the data. $S_{cut}$ is the final grind size in the units of millimeters.

$$P_{grind} = \left(28.76 \cdot {S_{cut}}^{-0.81}\right) \cdot \dot{m}_{biomass} \qquad \text{(eqn. 24)}$$

### *HRSG*

This calc block totals the heat transfer areas of all the heat exchangers in A600 Power Generation for use in Aspen Icarus software costing of a heat recovery steam generator, which is estimated as a waste heat boiler.

### *HUMIDITY*

This block sets humidity of the air entering the air separation unit.

### *HV-101, HV-203, HV-445*

This block calculates the lower and higher heating values of the following streams: biomass, syngas, and FT products.

### *LOCKHOP*

This block calculates the $CO_2$ required for pressurizing the lockhopper. Higman et al. reports 0.09 kg of pressurization gas is required per kg of biomass.

$$XCO2 = 0.09 * BIOMAS \qquad \text{(eqn. 25)}$$

### *MEATEMP*

This block sets the temperature of the incoming monoethanolamine solution entering the absorber column in the AGR area.

### *MOISTURE*

This block sets the moisture content of the entering biomass to the preprocessing area and sets the biomass moisture content exiting the biomass dryer. Also, the steam loop flow rate for drying the biomass is set at 9 times the amount of moisture removed during the drying process.

Moisture content (% wet basis) of entering biomass feed, $\mathrm{XMOIS1} = 25$. Inlet mass flow of moisture, $\mathrm{WATERI}$, is computed.

$$WATERI = FEED * XMOIS1/100/(1 - XMOIS1/100) \qquad \text{(eqn. 26)}$$

Moisture content (% wet basis) of biomass exiting the dryer, $\mathrm{XMOIS2} = 10$. Mass flow of moisture, $\mathrm{WATERO}$, is computed.

$$\mathrm{WATERO} = \mathrm{FEED} * \mathrm{XMOIS2}/100/(1 - \mathrm{XMOIS2}/100) \qquad \text{(eqn. 27)}$$

Specify steam required to remove moisture, $\mathrm{STEAMI}$.

$$\mathrm{STEAMI} = 9 * (\mathrm{WATERI} - \mathrm{WATERO}) \qquad \text{(eqn. 28)}$$

### *O2COMB*

Oxygen is required to combust the char and syngas that provides the energy necessary for drying the biomass. A system of stoichiometric combustion reactions is set up to sum all the oxygen required to fully combust the unconverted syngas purge from the FT synthesis outlet. The reactions are shown in Table C-3.

**Table C-3. Combustion Reactions to Determine Required Oxygen**

| Component | Reaction |
|---|---|
| CO | $CO + 0.5 \cdot O2 \rightarrow CO2$ |
| H2 | $H2 + 0.5 \cdot O2 \rightarrow H2O$ |
| CH4 | $CH4 + 2O2 \rightarrow 2H2O + CO2$ |
| C2H6 | $C2H6 + 3.5 \cdot O2 \rightarrow 3H2O + 2CO2$ |
| C2H4 | $C2H4 + 3O2 \rightarrow 2H2O + 2CO2$ |
| C2H2 | $C2H2 + 2.5O2 \rightarrow H2O + CO2$ |
| C3H8 | $C3H8 + 5O2 \rightarrow 4H2O + 3CO2$ |
| C4H10 | $C4H10 + 6.5O2 \rightarrow 5H2O + 4CO2$ |
| C5H12 | $C5H12 + 8O2 \rightarrow 6H2O + 5CO2$ |
| C6H14 | $C6H14 + 9.5O2 \rightarrow 7H2O + 6CO2$ |
| C7H16 | $C7H16 + 11O2 \rightarrow 8H2O + 7CO2$ |
| C8H18 | $C8H18 + 12.5O2 \rightarrow 9H2O + 8CO2$ |
| C9H20 | $C9H20 + 14O2 \rightarrow 10H2O + 9CO2$ |
| Tar | $C14H10(tar) + 16.5O2 \rightarrow 5H2O + 14CO2$ |
| H2S | $H2S + 1.5O2 \rightarrow H2O + SO2$ |
| NH3 | $NH3 + 1.75O2 \rightarrow 1.5H2O + NO2$ |

The molar flow rate of oxygen entering the combustor is summed and multiplied by a factor of 1.25 in order to combust with 25% excess air as shown in equation 29 below.

$$\begin{aligned}\dot{M}_{O2,in} = 1.25 \cdot (&\dot{M}_{CHAR,in} + 0.5\dot{M}_{CO,in} + 0.5\dot{M}_{H2,in} + 2\dot{M}_{CH4,in} + 3.5\dot{M}_{C2H6,in} + \\ &3\dot{M}_{C2H4,in} + 2.5\dot{M}_{C2H2,in} + 5\dot{M}_{C3H8,in} + 6.5\dot{M}_{C4H10,in} + 8\dot{M}_{C5H12,in} + \\ &9.5\dot{M}_{C6H14,in} + 11\dot{M}_{C7H16,in} + 12.5\dot{M}_{C8H18,in} + 14\dot{M}_{C9H20,in} + \\ &16.5\dot{M}_{TAR,in} + 1.5\dot{M}_{H2S,in} + 1.75\dot{M}_{NH3,in})\end{aligned} \quad \text{(eqn. 29)}$$

### *O2TURB*

This block calculates the molar flow rate of air (oxygen and nitrogen) required to combust syngas obtained from FT synthesis and the fuel gas obtained from Area 500 in the gas turbine of Area 600. A value of 25% excess air is assumed. The calculations are similar to the methodology in *O2COMB*.

### *OXYSET*

This block sets the entering oxygen at 0.35 lb oxygen per lb dry biomass into the gasifier.

$$\dot{m}_{O2,gas} = 0.35/100 \cdot \dot{m}_{biomass} \quad \text{(eqn. 30)}$$

### *SWGSSTM*

This block sets the steam flow into the sour water-gas-shift reactor to be at a ratio of 3:1 water to carbon monoxide. This ratio ensures enough water-gas-shift activity occurs within the reactor.

$$\dot{m}_{STM,addition} = 3.0 \cdot \dot{m}_{CO} - \dot{m}_{H2O} \qquad \text{(eqn. 31)}$$

### ***Low-Temperature Scenario***

#### *AMINE*

This block calculates the molar flow of monoethanolamine (MEA) needed for the required acid gas removal ($CO_2$ and $H_2S$) arriving from syngas quench and FT unconverted syngas recycle stream. The MEA is able to capture 0.35 moles acid gas per mole MEA. Additionally, the MEA is diluted as explained in *DILUTH2O.*

Molar MEA flow (in kmol/h) is calculated by

$$\dot{M}_{MEA} = (\dot{M}_{CO2,syn} + \dot{M}_{CO2,rec} + \dot{M}_{H2S,syn})/0.35 \qquad \text{(eqn. 32)}$$

where $\dot{M}_{CO2,syn}$ is molar flow of $CO_2$ from the syngas after the syngas quench, $\dot{M}_{CO2,rec}$ is the molar flow of $CO_2$ from the unconverted syngas recycle after the FT synthesis, and $\dot{M}_{H2S,syn}$ is the molar flow rate of $H_2S$ from the syngas quench.

Because the MEA solution in the amine absorption unit is to be 20 wt% concentrated with water, the flow of water must be calculated.

Mole flow of water is calculated as

$$\dot{M}_{H2O} = \frac{\dot{M}_{MEA} * MW_{MEA}/0.20}{MW_{H2O}} \qquad \text{(eqn. 33)}$$

#### *BIOELEM*

Same as for the HT scenario.

#### *DILUTH2O*

This block sets the MEA solution to be 20% concentrated with water.

#### *FTDISTR*

Same as for the high-temperature scenario.

#### *GASYIELD*

The following model describes how the fluidized-bed gasifier keeps an elemental mass balance. Experiments performed at Iowa State University provide the initial gasifier product distribution, and the model adjusts the yields of those experiments in order to balance carbon, hydrogen, sulfur, nitrogen, oxygen, and ash.

The approach taken to balance each element across the gasifier is by "floating" a component of each element. The "floating" component for carbon is the char. All sulfur and nitrogen not found in the char is assumed to form hydrogen sulfide and ammonia, respectively. Therefore, sulfur and nitrogen are balanced. Next, elemental hydrogen is adjusted in the model by either converting diatomic hydrogen to steam or decomposing steam to diatomic hydrogen. Oxygen balance is more complex. Because gasification operates at fuel-rich conditions, diatomic oxygen should not be present in the syngas leaving the gasifier. Therefore, diatomic oxygen cannot be the "floating" component. Instead, oxygen is balanced by adjusting the carbon monoxide or carbon dioxide in the exiting syngas. Because there is one oxygen difference between those two components, the oxygen can be adjusted to help close the balance.

Carbon balance follows the flow chart shown in Figure C-4. If there is less gaseous carbon out than total carbon in, then the difference is made up of char carbon, CCARB. Char is assumed to be composed of 68% carbon with the rest as H, O, N, and S. Ash is considered apart from the char and is considered inert in the model. Because the char is now fixed, the only pathway for sulfur and nitrogen to take is to form hydrogen sulfide and ammonia. Therefore, the sulfur and nitrogen are balanced.

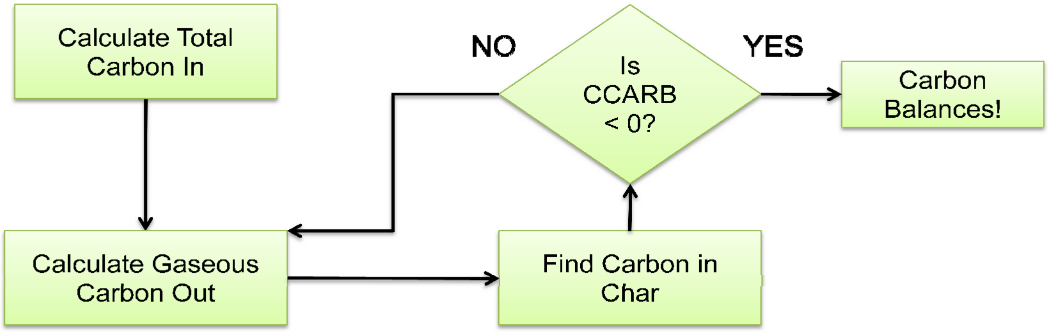


**Figure C-4. Decision diagram for carbon balance**

Next, as shown in Figure C-5, hydrogen is balanced. Knowing hydrogen in the char and in gaseous products, the hydrogen required (HREQD) is calculated as the sum of those two components. If the hydrogen required is less than hydrogen available (HAVAIL), made up of hydrogen in steam, biomass moisture, and in the biomass itself (THYD), then there is enough hydrogen available to balance. To balance hydrogen, the product yield swings toward either steam or diatomic hydrogen.

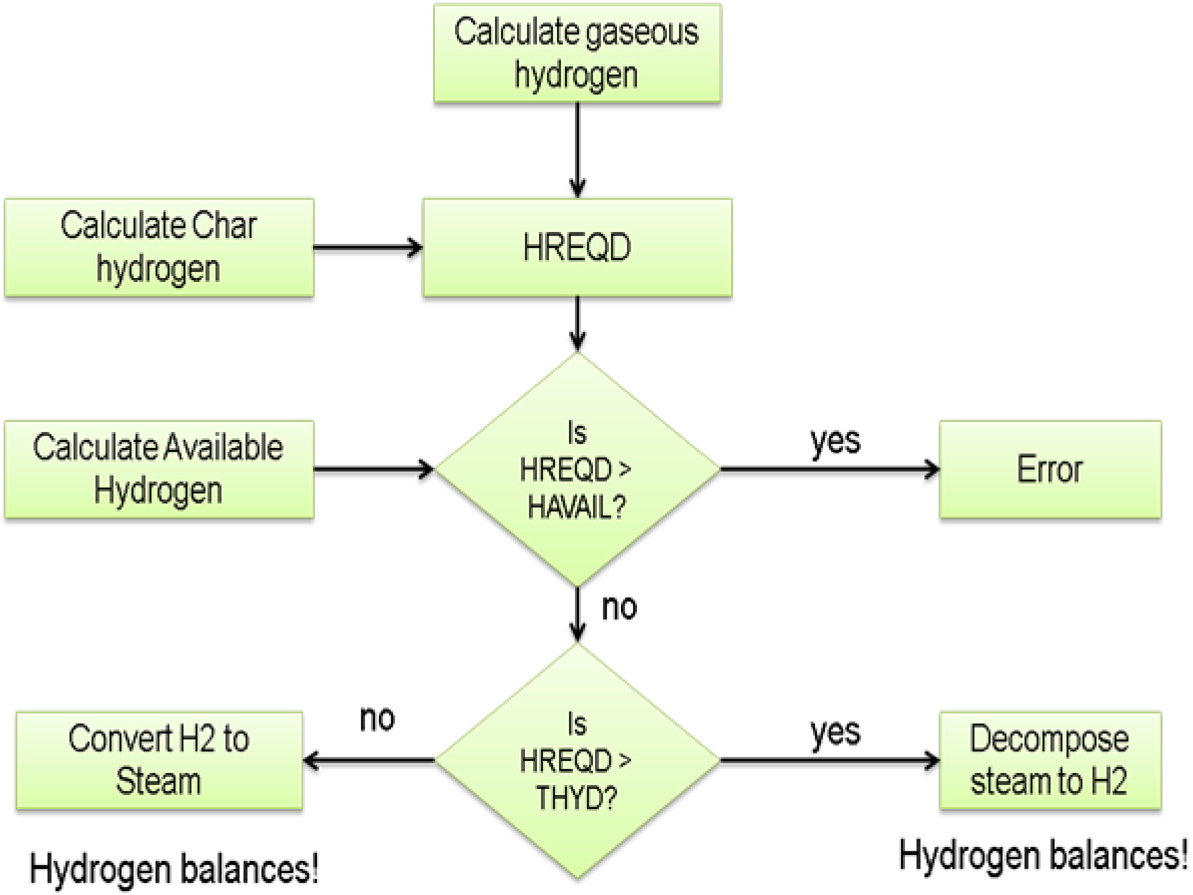


**Figure C-5. Decision diagram for hydrogen balance**

The only element left to balance is oxygen, which is accomplished by forcing creation of carbon monoxide or creation of carbon dioxide as shown in Figure C-6. The required oxygen (OREQD), made up of oxygen in char and oxygen in syngas, is checked against the available oxygen found in the entering oxygen, steam, and biomass. If there is more oxygen available than required, then the option is to move the excess oxygen to $CO_2$ by decreasing CO. If there is still oxygen present when CO is decreased to zero, then the yields need to be adjusted because excess oxygen is still present. If there is an oxygen deficit (OREQD > OAVAIL), then CO is increased and $CO_2$ is decreased. After that, if there is still an oxygen deficit, then insufficient oxygen is present and yields need to be adjusted. When all these steps are completed and no errors are generated, there is an elemental mass balance across the gasifier.

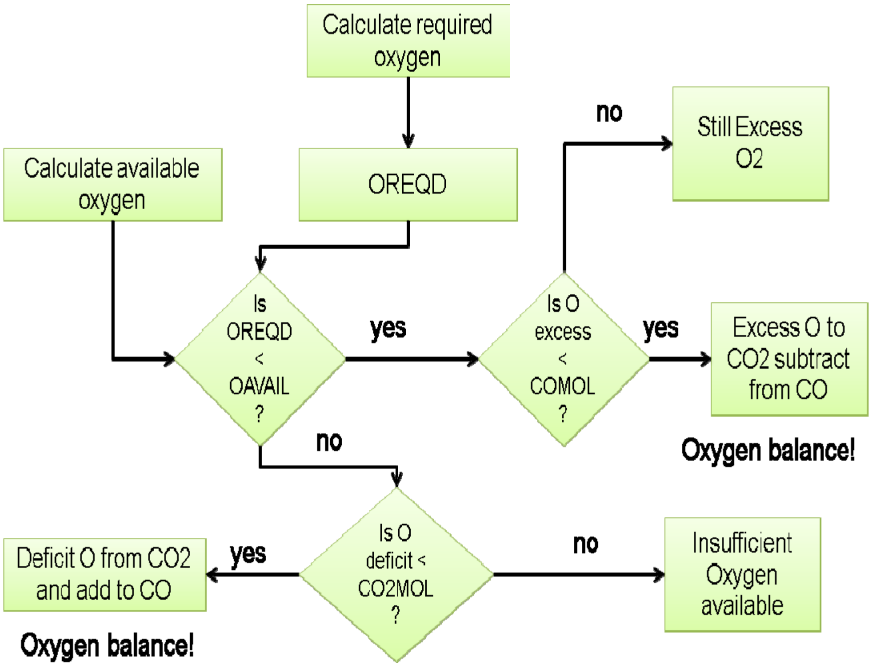


**Figure C-6. Decision diagram for oxygen balance**

### *GRIND*

This block calculates the power requirement (in kW) for grinding the biomass from the chop size of 12 mm to the final size of 6 mm. This power requirement data is found in Mani et al. and is for 12% exiting moisture. The correlation has been changed from a polynomial regression (which Mani et al. used) to a power regression because the power regression fit the data better. $S_{cut}$ is in millimeters.

$$P_{grind} = \left(28.76 \cdot S_{cut}{}^{-0.81}\right) \cdot \dot{m}_{biomass} \qquad \text{(eqn. 34)}$$

### *HUMIDITY*

This block sets humidity of the air entering the air separation unit.

### *HV-101, HV-203, HV-445*

This block calculates the lower and higher heating values of the following streams: biomass, syngas, and FT products.

### *MOISTURE*

This block is the same as found in the HT scenario.

### *O2COMB*

This block is the same as found in the HT scenario.

### *O2TURB*

This block is the same as found in the HT scenario.

### *OXYSET*

This block sets the entering oxidizing agents, oxygen and steam, into the gasifier. A linear correlation with temperature, $T_{gas}$ (in Fahrenheit), adapted from Bain for oxygen, is used because as oxygen increases in the gasifier, the temperature increases. Mass flow of oxygen, $\dot{m}_{O2,gas}$ , is in percentage of dry feedstock.

$$\dot{m}_{O2,gas} = (-11.567 + 0.02375 \cdot T_{gas})/100 \cdot \dot{m}_{biomass} \tag{eqn. 35}$$

The steam feed rate is set at 0.66 lb steam per lb oxygen.

$$\dot{m}_{steam,gas} = 0.66 \cdot \dot{m}_{O2,gas} \tag{eqn. 36}$$

Because 95% purity oxygen is produced in the air separation unit, argon mass flow is set at 5% of molar oxygen flow.

$$\dot{m}_{argon} = 0.05 \cdot \left(\frac{\dot{m}_{O2,gas}}{MW_{O2}}\right) * MW_{Ar} \tag{eqn. 37}$$

## Aspen Plus Model Design Specifications

### *High-Temperature Scenario*

#### *DS-1*

The exiting temperature of air in the heat exchanger used to pre-cool the air entering the cryogenic distillation column is varied until a net duty of zero is observed.

#### *FSSPL02*

This design specification varies the fraction of unconverted syngas that is piped to Area 200 for the combustion of syngas. The syngas, in turn, provides the heat required to dry the biomass.

#### *H2SPLIT*

This design spec calculates the required hydrogen that needs to be reserved by the PSA unit for use in Area 500: Hydrocracking. A typical yield from hydrocracking is shown in Table C-4. Because the FT products are hydrogen deficient relative to the final blend, makeup hydrogen is required. The syngas purge amount going to the pressure swing adsorption (PSA) unit is varied so that the calculated delivered hydrogen matches the required hydrogen to Area 500. Without showing the detailed calculations, the basic steps are first calculating the carbon and hydrogen content in the FT product stream. The carbon mass flow is the same as that of the final blend stream flow. Using the blend fractions in Table C-4, the amount of hydrogen is calculated in the final blend and the difference in hydrogen is determined. The difference is multiplied by 1.1 to obtain the delivered hydrogen mass flow rate to hydrocracking area.

**Table C-4. 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 |

#### *O2-101, O2-203, O2-445*

These design specifications vary the amount of oxygen inlet to the heating value blocks (HV-101, HV-203, HV-445) so as to be stoichiometric in the combustion of the duplicate stream.

#### *O2-SULF*

This design specification varies the amount of oxygen into the LO-CAT oxidizer unit to fully oxidize the $H_2S$ into solid sulfur.

#### *SGSTEMP*

The temperature of operation in the sour water-gas-shift reactor is varied until the exiting equilibrium molar ratio of $H_2/CO$ is just above the optimal FT ratio (2.1). A small amount of hydrogen is captured in the PSA unit bringing that ratio down to the optimum for FT synthesis.

### Low-Temperature Scenario

#### DS-1

This design specification is the same as for the HT scenario.

#### H2SPLIT

This design specification is the same as for the HT scenario.

#### O2-101, O2-203, O2-445

These design specifications are the same as in the HT scenario.

#### O2-SULF

This design specification is the same as for the HT scenario.

#### STMRECOV

Heat can be recovered from the combustion of syngas and char. This specification varies the steam flow rate (stream 280) to bring the combustion flue gas (stream 252) down to 200°C via heat exchanging.

#### WGSTEMP

The temperature of operation in the water-gas-shift reactor is varied until the exiting equilibrium molar ratio of $H_2/CO$ is just above the optimal FT ratio (2.1). A small amount of hydrogen is captured in the PSA unit, bringing that ratio down to the optimum for FT synthesis.

## Detailed Calculations

### *Aspen Plus Model Calculations and Notes*

### *Outline*

- Plant Input
- Plant Output
- Carbon Efficiency to Fuels
- Energy Content
- FT Reaction Conversion Solver
- Equipment Sizing
  - Dryer
  - Lockhoppers
  - Slag/Char Collection
  - PSA Unit
  - Fuel Storage
- LT Gasifier Cost
- FT Reactor Cost
- Acid Gas Removal Area Cost
- A500 Hydroprocessing Area Cost
- Reactors and Catalysts
- Natural Gas Utility Usage

### *Defining Units*

$\mathrm{MJ} := 10^{6}\,\mathrm{J}$

$\mathrm{kPa} := 10^{3}\cdot\mathrm{Pa}$

$\rho_{water} := 1000\frac{\mathrm{kg}}{\mathrm{m}^{3}}$

$\mathrm{kmol} := 1000\mathrm{mol}$

$\mathrm{lbmol} := \frac{\mathrm{kmol}}{2.2}$

$\mathrm{bbl} := 42\mathrm{gal}$

$\rho_{gas} := 737.22\frac{\mathrm{kg}}{\mathrm{m}^{3}}$

$\rho_{diesel} := 840\frac{\mathrm{kg}}{\mathrm{m}^{3}}$

$\mathrm{MMgal} := 10^{6}\mathrm{gal}$

$\mathrm{kJ} := 1000\,\mathrm{J}$

$\mathrm{bpsd} := \frac{42\mathrm{gal}}{\mathrm{day}}$

$\mathrm{PJ} := 10^{15}\mathrm{J}$

$\mathrm{MMcf} := 10^{6}\mathrm{ft}^{3}$

$\mathrm{Cp} := 100\mathrm{poise}$

$\mathrm{MW}_{H2O} := 18.02\frac{\mathrm{gm}}{\mathrm{mol}}$

$\mathrm{MMBTU} := 10^{6}\mathrm{BTU}$

$\mathrm{HHV}_{stover} := 17.65\frac{\mathrm{MJ}}{\mathrm{kg}}$

$\mathrm{HHV}_{stover} = 7.588\times 10^{3}\,\frac{\mathrm{BTU}}{\mathrm{lb}}$

$100\frac{\mathrm{kg}}{\mathrm{m}^{3}} = 6.243\frac{\mathrm{lb}}{\mathrm{ft}^{3}}$

$\mathrm{therm} := 100000\,\mathrm{BTU}$

$\mathrm{dekatherm} := 10\mathrm{therm}$

$P_{ref} \equiv 1\mathrm{atm}$

$T_{ref} \equiv 298\,\mathrm{K}$

$\mathrm{GJ} := 10^{9}\,\mathrm{J}$

## Plant Input

## Biomass

$m_{dot\_biomass} := 2000 \frac{tonne}{day}$ Availability $:= 310\,day$ Load $:= 7446\ hr$

### Elemental Composition

Carbon $Frac_{C\_biomass} \equiv 0.4728$ $MW_C := 12.01 \frac{gm}{mol}$

Oxygen $Frac_{O\_biomass} \equiv 0.4063$ $MW_O := 16. \frac{gm}{mol}$

Hydrogen $Frac_{H\_biomass} \equiv 0.0506$ $MW_H := 1.01 \frac{gm}{mol}$

Sulfur $Frac_{S\_biomass} \equiv 0.0022$ $MW_S := 32.07 \frac{gm}{mol}$

Nitrogen $Frac_{N\_biomass} \equiv 0.008$ $MW_N := 14.01 \frac{gm}{mol}$

Ash $Frac_{A\_biomass} \equiv 0.0600$

### Elemental Mass Flow

$m_{dot\_C\_in} := m_{dot\_biomass} \cdot Frac_{C\_biomass}$ $m_{dot\_C\_in} = 945.6 \frac{tonne}{day}$

$m_{dot\_O\_in} := m_{dot\_biomass} \cdot Frac_{O\_biomass}$ $m_{dot\_O\_in} = 812.6 \frac{tonne}{day}$

$m_{dot\_H\_in} := m_{dot\_biomass} \cdot Frac_{H\_biomass}$ $m_{dot\_H\_in} = 101.2 \frac{tonne}{day}$

$m_{dot\_S\_in} := m_{dot\_biomass} \cdot Frac_{S\_biomass}$ $m_{dot\_S\_in} = 4.4 \frac{tonne}{day}$

$m_{dot\_N\_in} := m_{dot\_biomass} \cdot Frac_{N\_biomass}$ $m_{dot\_N\_in} = 16 \frac{tonne}{day}$

$m_{dot\_A\_in} := m_{dot\_biomass} \cdot Frac_{A\_biomass}$ $m_{dot\_A\_in} = 120 \frac{tonne}{day}$

## Elemental Mole Flow

$n_{dot\_C\_in} := \frac{m_{dot\_C\_in}}{MW_C}$ $\quad n_{dot\_C\_in} = 911.278\ \frac{mol}{s}$

$n_{dot\_O\_in} := \frac{m_{dot\_O\_in}}{MW_O}$ $\quad n_{dot\_O\_in} = 587.818\ \frac{mol}{s}$

$n_{dot\_H\_in} := \frac{m_{dot\_H\_in}}{MW_H}$ $\quad n_{dot\_H\_in} = 1160\ \frac{mol}{s}$

$n_{dot\_S\_in} := \frac{m_{dot\_S\_in}}{MW_S}$ $\quad n_{dot\_S\_in} = 1.588\ \frac{mol}{s}$

$n_{dot\_N\_in} := \frac{m_{dot\_N\_in}}{MW_N}$ $\quad n_{dot\_N\_in} = 13.218\ \frac{mol}{s}$

## ***Biomass Moisture***

$moist_{in} := 0.25$ $\quad moist_{dried} := 0.10$

$m_{dot\_moist\_in} := \frac{moist_{in} \cdot m_{dot\_biomass}}{1 - moist_{in}}$ $\quad m_{dot\_moist\_in} = 666.667\ \frac{tonne}{day}$

$m_{dot\_moist\_dried} := \frac{moist_{dried} \cdot m_{dot\_biomass}}{1 - moist_{dried}}$ $\quad m_{dot\_moist\_dried} = 222.222\frac{tonne}{day}$

$\rho_{bulk\_stover} := 100\frac{kg}{m^3}$ Source: *Kaliyan and Morey, 2005* for 0.66-0.8 mm sized particles

## ***HT Gasifier Steam/Oxygen Addition***

Source: *Probstein and Hicks, 2006*

Stoichiometric/thermo-neutral requirement for synthesis gas according to following equation:

$1.34C + 0.34\ O_2 + H_2O \text{ --> } 0.34CO_2 + CO + H_2$

Oxygen to Carbon: 0.25
Steam to Carbon: 0.75

$m_{dot\_O2\_in} := 0.35 \cdot m_{dot\_biomass}$ $\qquad m_{dot\_O2\_in} = 700 \frac{tonne}{day}$

$n_{dot\_O2\_in} := \frac{m_{dot\_O2\_in}}{2 \cdot MW_O}$ $\qquad n_{dot\_O2\_in} = 253.183 \frac{mol}{s}$

$Ratio_{O2\_to\_C} := \frac{n_{dot\_O2\_in}}{n_{dot\_C\_in}}$ $\qquad Ratio_{O2\_to\_C} = 0.278$

Steam addition ratio is then three times that of oxygen minus the moisture in the biomass

$n_{dot\_H2O\_in} := 3 \cdot Ratio_{O2\_to\_C} \cdot n_{dot\_C\_in} - \frac{m_{dot\_moist\_dried}}{MW_{H2O}}$ $\qquad n_{dot\_H2O\_in} = 616.817 \frac{mol}{s}$

$m_{dot\_H2O\_in} := n_{dot\_H2O\_in} \cdot MW_{H2O}$ $\qquad m_{dot\_H2O\_in} = 960 \frac{tonne}{day}$

## Plant Output

### HT Fuel Production

$m_{dot\_gasHT} := 112.78 \frac{tonne}{day}$

$m_{dot\_dieselHT} := 266.5 \frac{tonne}{day}$

$v_{dot\_gasHT} := \frac{m_{dot\_gasHT}}{\rho_{gas}}$

$v_{dot\_dieselHT} := \frac{m_{dot\_dieselHT}}{\rho_{diesel}}$

$v_{dot\_gasHT} = 40413 \frac{gal}{day}$

$v_{dot\_dieselHT} = 83812 \frac{gal}{day}$

$v_{dot\_gasHT} = 962 \frac{bbl}{day}$

$v_{dot\_dieselHT} = 1996 \frac{bbl}{day}$

$v_{dot\_gasoline\_per\_year} := v_{dot\_gasHT} \cdot Load$

$v_{dot\_gasoline\_per\_year} = 12.538\ MMgal$

$v_{dot\_diesel\_per\_year} := v_{dot\_dieselHT} \cdot Load$

$v_{dot\_diesel\_per\_year} = 26.003\ MMgal$

### LT Fuel Production

$m_{dot\_gasLT} := 87.12 \frac{tonne}{day}$

$m_{dot\_dieselLT} := 205.86 \frac{tonne}{day}$

$v_{dot\_gasLT} := \frac{m_{dot\_gasLT}}{\rho_{gas}}$

$v_{dot\_dieselLT} := \frac{m_{dot\_dieselLT}}{\rho_{diesel}}$

$v_{dot\_gasLT} = 31218 \frac{gal}{day}$

$v_{dot\_dieselLT} = 64741 \frac{gal}{day}$

$v_{dot\_gasLT} = 743 \frac{bbl}{day}$

$v_{dot\_dieselLT} = 1541 \frac{bbl}{day}$

$v_{dot\_gasoline\_per\_yearLT} := v_{dot\_gasLT} \cdot Load$

$v_{dot\_gasoline\_per\_yearLT} = 9.685\ MMgal$

$v_{dot\_diesel\_per\_yearLT} := v_{dot\_dieselLT} \cdot Load$

$v_{dot\_diesel\_per\_yearLT} = 20.086\ MMgal$

## Carbon Efficiency to Fuels

### HT Scenario

### Gasoline Carbon

$$Frac_{C\_gasoline} := \frac{8 \cdot 12.01}{8 \cdot 12.01 + 18 \cdot 1.01} \qquad Frac_{C\_gasoline} = 0.841$$

$$m_{dot\_gasHT} = 112.78 \frac{tonne}{day}$$

$$m_{dot\_C\_gasHT} := Frac_{C\_gasoline} \cdot m_{dot\_gasHT} \qquad m_{dot\_C\_gasHT} = 94.835 \frac{tonne}{day}$$

### Diesel Carbon

$$m_{dot\_dieselHT} = 266.5 \frac{tonne}{day}$$

$$Frac_{C\_diesel} := \frac{16 \cdot 12.01}{16 \cdot 12.01 + 34 \cdot 1.01} \qquad Frac_{C\_diesel} = 0.848$$

$$m_{dot\_C\_dieselHT} := Frac_{C\_diesel} \cdot m_{dot\_dieselHT} \qquad m_{dot\_C\_dieselHT} = 226.096 \frac{tonne}{day}$$

$$m_{dot\_C\_outHT} := m_{dot\_C\_gasHT} + m_{dot\_C\_dieselHT}$$

$$C\_effHT = 0.339$$

$$m_{dot\_C\_outHT} = 320.931 \frac{tonne}{day} \qquad C\_effHT := \frac{m_{dot\_C\_outHT}}{m_{dot\_C\_in}}$$

### *LT Scenario*

## Gasoline Carbon

$m_{dot\_gasLT} = 87.12 \frac{tonne}{day}$

$m_{dot\_C\_gasLT} := Frac_{C\_gasoline} \cdot m_{dot\_gasLT}$ $m_{dot\_C\_gasLT} = 73.258 \frac{tonne}{day}$

## Diesel Carbon

$m_{dot\_dieselLT} = 205.86 \frac{tonne}{day}$

$m_{dot\_C\_dieselLT} := Frac_{C\_diesel} \cdot m_{dot\_dieselLT}$ $m_{dot\_C\_dieselLT} = 174.649 \frac{tonne}{day}$

$m_{dot\_C\_outLT} := m_{dot\_C\_gasLT} + m_{dot\_C\_dieselLT}$

$m_{dot\_C\_outLT} = 247.908 \frac{tonne}{day}$ $C\_effLT := \frac{m_{dot\_C\_outLT}}{m_{dot\_C\_in}}$ $C\_effLT = 0.262$

## Energy Content

This section acquires the energy content (on a LHV basis) from the Aspen Plus data and converts to megawatts for use in developing an energy balance

### Biomass

$E_{biomass} := 1400313 \frac{MJ}{hr}$ $E_{biomass} = 388.976\ MW$

### Fuel

$E_{fuelHT} := 695598 \frac{MJ}{hr}$ $E_{fuelHT} = 193.222\ MW$

$E_{fuelLT} := 539292 \frac{MJ}{hr}$ $E_{fuelLT} = 149.803\ MW$

### Char/Tar

$E_{char\_LT} := 87792 \frac{MJ}{hr}$ $E_{char\_LT} = 24.387\ MW$

$E_{tar\_LT} := 16980 \frac{MJ}{hr}$ $E_{tar\_LT} = 4.717\ MW$

### Raw Syngas

$E_{rawsyngas\_HT} := 1230712 \frac{MJ}{hr}$ $E_{rawsyngas\_HT} = 341.864\ MW$

$E_{rawsyngas\_LT} := 964054 \frac{MJ}{hr}$ $E_{rawsyngas\_LT} = 267.793\ MW$

### Energy Loss across the Gasifier

Energy lost across the gasifier is calculated as difference in energy in the biomass and energy in the raw syngas and char (only in LT scenario)

$E_{gasifierloss\_HT} := E_{biomass} - E_{rawsyngas\_HT}$ $E_{gasifierloss\_HT} = 47.111\ MW$

$E_{gasifierloss\_LT} := E_{biomass} - E_{rawsyngas\_LT} - E_{char\_LT}$ $E_{gasifierloss\_LT} = 96.796\ MW$

## Unconverted Syngas used in A600 Power Generation

$E_{syngasA600\_HT} := 129332\,\frac{MJ}{hr}$ $E_{syngasA600\_HT} = 35.926\ MW$

$E_{syngasA600\_LT} := 109708\,\frac{MJ}{hr}$ $E_{syngasA600\_LT} = 30.474\ MW$

## Fuel Gas from A500 used in A600 Power Generation

$E_{fuelgas\_HT} := 104114\,\frac{MJ}{hr}$ $E_{fuelgas\_HT} = 28.921\ MW$

$E_{fuelgas\_LT} := 80718\,\frac{MJ}{hr}$ $E_{fuelgas\_LT} = 22.422\ MW$

## Fischer-Tropsch Product

$E_{FTliquids\_HT} := 782894\,\frac{MJ}{hr}$ $E_{FTliquids\_HT} = 217.471\ MW$

$E_{FTliquids\_LT} := 606801\,\frac{MJ}{hr}$ $E_{FTliquids\_LT} = 168.556\ MW$

## Electricity Generated

$E_{elecgenOUT\_HT} := 48.55\,MW$ $E_{elecgenOUT\_LT} := 40.73\,MW$

## Net Electricity (exported)

$E_{elecnet\_HT} := 13.8\,MW$ $E_{elecnet\_LT} := 16.3\ MW$

## Power Generation Loss

The loss is the difference between electric generation out and the gas energy in

$E_{A600losses\_HT} := E_{syngasA600\_HT} + E_{fuelgas\_HT} - E_{elecgenOUT\_HT}$

$E_{A600losses\_HT} = 16.296\ MW$

$E_{A600losses\_LT} := E_{syngasA600\_LT} + E_{fuelgas\_LT} - E_{elecgenOUT\_LT}$

$E_{A600losses\_LT} = 12.166\ MW$

## Loss across FT Reactor

$E_{FTreactorlosses\_HT} := 226737\,\frac{MJ}{hr}$ $E_{FTreactorlosses\_HT} = 62.983\ MW$

$E_{FTreactorlosses\_LT} := 175128\,\frac{MJ}{hr}$ $E_{FTreactorlosses\_LT} = 48.647\ MW$

## Unconverted Syngas used for Biomass Drying

Only in HT scenario

$E_{biomass\_drying\_HT} := 24663\,\frac{MJ}{hr}$ $E_{biomass\_drying\_HT} = 6.851\ MW$

## *Fischer-Tropsch Reaction Conversion Solver*

This section solves for the reaction fractional conversion for each reaction in the Fischer-Tropsch reactor. A set of equations is developed and solved. The resulting ε values (ε1 - ε30) are used directly in the Aspen Plus conversion reactor block. The reactions in the reactor block are defined as molar extent.

Depending on the alpha chain growth probability, the reactor forms different product composition.

Step 1: choose the expected alpha chain growth value

$\alpha_{FT} := 0.9$

Step 2: using the $\alpha_{FT}$ chain growth, the mole fraction of each hydrocarbon in the FT product is calculated.

$M1 := \alpha_{FT}^{1-1}\cdot\left(1-\alpha_{FT}\right)$ $M1 = 0.1$

$M2 := \alpha_{FT}^{2-1}\cdot\left(1-\alpha_{FT}\right)$ $M2 = 0.09$

$M3 := \alpha_{FT}^{3-1}\cdot\left(1-\alpha_{FT}\right)$ $M3 = 0.081$

$M4 := \alpha_{FT}^{4-1}\cdot\left(1-\alpha_{FT}\right)$ $M4 = 0.073$

$M5 := \alpha_{FT}^{5-1}\cdot\left(1-\alpha_{FT}\right)$ $M5 = 0.066$

$M6 := \alpha_{FT}^{6-1}\cdot\left(1-\alpha_{FT}\right)$ $M6 = 0.059$

$M7 := \alpha_{FT}^{7-1}\cdot\left(1-\alpha_{FT}\right)$ $M7 = 0.053$

$M8 := \alpha_{FT}^{8-1}\cdot\left(1-\alpha_{FT}\right)$ $M8 = 0.048$

$M9 := \alpha_{FT}^{9-1}\cdot\left(1-\alpha_{FT}\right)$ $M9 = 0.043$

$M10 := \alpha_{FT}^{10-1}\cdot\left(1-\alpha_{FT}\right)$ $M10 = 0.039$

$M11 := \alpha_{FT}^{11-1}\cdot\left(1-\alpha_{FT}\right)$ $M11 = 0.035$

$M12 := \alpha_{FT}^{12-1}\cdot\left(1-\alpha_{FT}\right)$ $M12 = 0.031$

$M13 := \alpha_{FT}^{13-1}\cdot\left(1-\alpha_{FT}\right)$ $M13 = 0.028$

$M14 := \alpha_{FT}^{14-1}\cdot\left(1-\alpha_{FT}\right)$ $M14 = 0.025$

$M15 := \alpha_{FT}^{15-1}\cdot\left(1-\alpha_{FT}\right)$ $M15 = 0.023$

$M16 := \alpha_{FT}^{16-1}\cdot\left(1-\alpha_{FT}\right)$ $M16 = 0.021$

$M17 := \alpha_{FT}^{17-1}\cdot\left(1-\alpha_{FT}\right)$ $M17 = 0.019$

$M18 := \alpha_{FT}^{18-1}\cdot\left(1-\alpha_{FT}\right)$ $M18 = 0.017$

$M19 := \alpha_{FT}^{19-1}\cdot\left(1-\alpha_{FT}\right)$ $M19 = 0.015$

$M20 := \alpha_{FT}^{20-1}\cdot\left(1-\alpha_{FT}\right)$ $M20 = 0.014$

All hydrocarbons greater than C20 make up the balance and are modeled using C30.

$$M30 := 1 - \begin{pmatrix} M1 + M2 + M3 + M4 + M5 + M6 + M7 + M8 + M9 + M10 \ldots \\ + M11 + M12 + M13 + M14 + M15 + M16 + M17 + M18 + M19 + M20 \end{pmatrix}$$

$M30 = 0.122$

Step 3: Set up a series of equations to solve along with guess values (required for Mathcad)

For a nominal 1000 moles of CO input, the expected CO output is 600 moles because 40% is converted.

Known $CO_{out} := 600$ $CO_{in} := 1000$ <---------------- 40% conversion of CO

Guess $\varepsilon 1 := 20$ $\varepsilon 2 := 20$ $\varepsilon 3 := 20$ $\varepsilon 4 := 20$ $\varepsilon 5 := 20$

$\varepsilon 6 := 20$ $\varepsilon 7 := 20$ $\varepsilon 8 := 20$ $\varepsilon 9 := 20$ $\varepsilon 10 := 20$

$\varepsilon 11 := 20$ $\varepsilon 12 := 20$ $\varepsilon 13 := 20$ $\varepsilon 14 := 20$ $\varepsilon 15 := 20$

$\varepsilon 16 := 20$ $\varepsilon 17 := 20$ $\varepsilon 18 := 20$ $\varepsilon 19 := 20$ $\varepsilon 20 := 20$

$\varepsilon 30 := 20$ $D := 0.1$ <----- This value to be varied until COconv is equal to desired.

A nominal 400 moles of CO are converted in the FT reactor. The sum of the exiting amount of moles in the FT product distribution will not be 400, because moles are not conserved. Mass is conserved, however. Therefore, the variable "D" represents a factor that adjusts all the conversions (ε1, ε2, etc.).

The resulting value of D is 0.1 meaning that 40 moles of FT products exit the reactor.

Given

$$D = \left( \begin{array}{l} \varepsilon 1 + \frac{1}{2}\varepsilon 2 + \frac{1}{3}\varepsilon 3 + \frac{1}{4}\varepsilon 4 + \frac{1}{5}\varepsilon 5 + \frac{1}{6}\varepsilon 6 + \frac{1}{7}\varepsilon 7 + \frac{1}{8}\varepsilon 8 + \frac{1}{9}\varepsilon 9 + \frac{1}{10}\varepsilon 10 + \frac{1}{11}\varepsilon 11 + \frac{1}{12}\varepsilon 12 \ldots \\ + \frac{1}{13}\varepsilon 13 + \frac{1}{14}\varepsilon 14 + \frac{1}{15}\varepsilon 15 + \frac{1}{16}\varepsilon 16 + \frac{1}{17}\varepsilon 17 + \frac{1}{18}\varepsilon 18 + \frac{1}{19}\varepsilon 19 + \frac{1}{20}\varepsilon 20 + \frac{1}{30}\varepsilon 30 \end{array} \right)$$

$$M1 = \frac{\varepsilon 1}{D} \quad M2 = \frac{\frac{1}{2}\varepsilon 2}{D} \quad M3 = \frac{\frac{1}{3}\varepsilon 3}{D} \quad M4 = \frac{\frac{1}{4}\varepsilon 4}{D} \quad M5 = \frac{\frac{1}{5}\varepsilon 5}{D} \quad M6 = \frac{\frac{1}{6}\varepsilon 6}{D}$$

$$M7 = \frac{\frac{1}{7}\varepsilon 7}{D} \quad M8 = \frac{\frac{1}{8}\varepsilon 8}{D} \quad M9 = \frac{\frac{1}{9}\varepsilon 9}{D} \quad M10 = \frac{\frac{1}{10}\varepsilon 10}{D} \quad M11 = \frac{\frac{1}{11}\varepsilon 11}{D} \quad M12 = \frac{\frac{1}{12}\varepsilon 12}{D}$$

$$M13 = \frac{\frac{1}{13}\varepsilon 13}{D} \quad M14 = \frac{\frac{1}{14}\varepsilon 14}{D} \quad M15 = \frac{\frac{1}{15}\varepsilon 15}{D} \quad M16 = \frac{\frac{1}{16}\varepsilon 16}{D} \quad M17 = \frac{\frac{1}{17}\varepsilon 17}{D} \quad M18 = \frac{\frac{1}{18}\varepsilon 18}{D}$$

$$M19 = \frac{\frac{1}{19}\varepsilon 19}{D} \quad M20 = \frac{\frac{1}{20}\varepsilon 20}{D} \quad M30 = \frac{\frac{1}{30}\varepsilon 30}{D}$$

$$Solve := Find(\varepsilon 1, \varepsilon 2, \varepsilon 3, \varepsilon 4, \varepsilon 5, \varepsilon 6, \varepsilon 7, \varepsilon 8, \varepsilon 9, \varepsilon 10, \varepsilon 11, \varepsilon 12, \varepsilon 13, \varepsilon 14, \varepsilon 15, \varepsilon 16, \varepsilon 17, \varepsilon 18, \varepsilon 19, \varepsilon 20, \varepsilon 30)$$

Solve =

| | 0 |
|---|---|
| 0 | 0.01 |
| 1 | 0.018 |

$e1 := Solve_{0}$ $e1 = 0.01$

$e2 := Solve_{1}$ $e2 = 0.018$

$e3 := Solve_{2}$ $e3 = 0.0243$

$e4 := Solve_{3}$ $e4 = 0.02916$

$e5 := Solve_{4}$ $e5 = 0.0328$

$e6 := Solve_{5}$ $e6 = 0.03543$

$e7 := Solve_{6}$ $e7 = 0.0372$

$e8 := Solve_{7}$ $e8 = 0.03826$

$e9 := Solve_{8}$ $e9 = 0.03874$

$e10 := Solve_{9}$ $e10 = 0.03874$

$e11 := Solve_{10}$ $e11 = 0.03835$

$e12 := Solve_{11}$ $e12 = 0.03766$

$e13 := Solve_{12}$ $e13 = 0.03672$

$e14 := Solve_{13}$ $e14 = 0.03559$

$e15 := Solve_{14}$ $e15 = 0.03432$

$e16 := Solve_{15}$ $e16 = 0.03294$

$e17 := Solve_{16}$ $e17 = 0.0315$

$e18 := Solve_{17}$ $e18 = 0.03002$

$e19 := Solve_{18}$ $e19 = 0.02852$

$e20 := Solve_{19}$ $e20 = 0.02702$

$e30 := Solve_{20}$ $e30 = 0.36473$

$D = 0.1$

Step 4: The guess value of D is varied until the sum of all reaction conversions (ε1, ε2, etc.) sum to 1.0 as seen below. This means that all 400 moles of CO are converted as expected.

$$CO_{conv} := e1 + e2 + e3 + e4 + e5 + e6 + e7 + e8 + e9 + e10 + e11 + e12 + e13 + e14 + e15 + e16 \ldots + e17 + e18 + e19 + e20 + e30$$

$$CO_{conv} = 1$$

Step 5: Each value for ε is imported into Aspen Plus

## *Equipment Sizing*

### ***Rotary Dryer*** Source: *Process Engineering Economics* by James Couper, 2003

Typical rpm of rotary dryers $rpm_{dryer} := 4$

Typical product of $rpm \cdot diameter(feet)$ equals 15-25. Assume value of 25 for larger end

$$D_{dryer} := \frac{25ft}{rpm_{dryer}} \qquad D_{dryer} = 6.25\ ft$$

Typical residence times are 5-90 minutes and holdup of solids is 7%-8%. Assume 5 minutes and 8%.

$$t_{res} := 5min \qquad holdup := 0.08$$

Typical exit gas temperature is 10°-20°C above the entering solids.

Feed rate into plant is 2000 ton/day with bulk density of stover equal to 100kg/m^3. Water density is accounted for as well.

$$m_{dot\_feed} := 2000\frac{tonne}{day} \qquad m_{dot\_moist\_in} = 666.667\frac{tonne}{day}$$

$$\rho_{bulk\_stover} = 100\frac{kg}{m^3} \qquad \rho_{water} = 1000\frac{kg}{m^3}$$

Volume of solids in dryer $V_{solids} := \left(\frac{m_{dot\_feed}}{\rho_{bulk\_stover}} + \frac{m_{dot\_moist\_in}}{\rho_{water}}\right) \cdot t_{res}$ $V_{solids} = 71.759m^3$

Volume of solids and steam $V_{dryer\_total} := \frac{V_{solids}}{holdup}$ $V_{dryer\_total} = 896.991m^3$

Length of theoretical dryer $length_{dryer} := \frac{V_{dryer\_total}}{D_{dryer}^2 \cdot \frac{\pi}{4}}$ $length_{dryer} = 314.708\ m$

Surface area of theoretical dryer $A_{surf\_dryer} := length_{dryer} \cdot \pi \cdot D_{dryer}$ $A_{surf\_dryer} = 1883.4m^2$

Max surface area as reported by Aspen Icarus is 185 $m^2$; therefore approximately 10 dryers are required.

Feed throughput in each dryer (used for Icarus input) $\frac{m_{dot\_feed} + m_{dot\_moist\_in}}{10} = 24495.8\frac{lb}{hr}$

## *Lockhopper System*

Source: *CE IGCC Repowering Project Bins and Lockhoppers,* Combustion Eng. 1993

Note: this report's feedstock is coal

Assumptions from report
-A receiving bin is situated before the lockhopper with a 40 minute residence time
-Design pressure is for 50 psia.
-Cycle time for lockhopper system is designed for 10 minutes resulting in approximately 50,000 cycles per year
-Storage volume for lockhopper and feed bin is assumed to be 10 minutes
-Approximate lockhopper and feed bin vessel thickness is 1.5 inches and design pressure is for 450 psia
-Volume is theoretical + 33%

### Residence Time

biomass receiving bin $t_{res\_rbin} := 40\,min$ $\varepsilon_{void} := 25\%$

biomass lockhopper $t_{res\_lock} := 10\,min$

biomass feed bin $t_{res\_fbin} := 10\,min$

$$m_{dot\_feed\_lock} := m_{dot\_feed} + m_{dot\_moist\_dried}$$

$$m_{dot\_feed\_lock} = 2222\,\frac{tonne}{day}$$

Density of feed $$\rho_{stover\_10\%moist} := \frac{\rho_{bulk\_stover}\cdot 2000 + \rho_{water}\cdot 222}{2222}$$

$$\rho_{stover\_10\%moist} = 189.919\,\frac{kg}{m^3}$$

### HT Scenario Lockhopper System (1 train)

Volume of biomass receiving bin $$V_{r\_bin} := \frac{t_{res\_rbin}\cdot m_{dot\_feed\_lock}}{\rho_{stover\_10\%moist}}\cdot\frac{1}{1-\varepsilon_{void}}$$

$V_{r\_bin} = 433\,m^3$

$V_{lock} = 108\,m^3$

Volume of biomass lockhopper $$V_{lock} := \frac{t_{res\_lock}\cdot m_{dot\_feed\_lock}}{\rho_{stover\_10\%moist}}\cdot\frac{1}{1-\varepsilon_{void}}$$

$V_{f\_bin} = 108\,m^3$

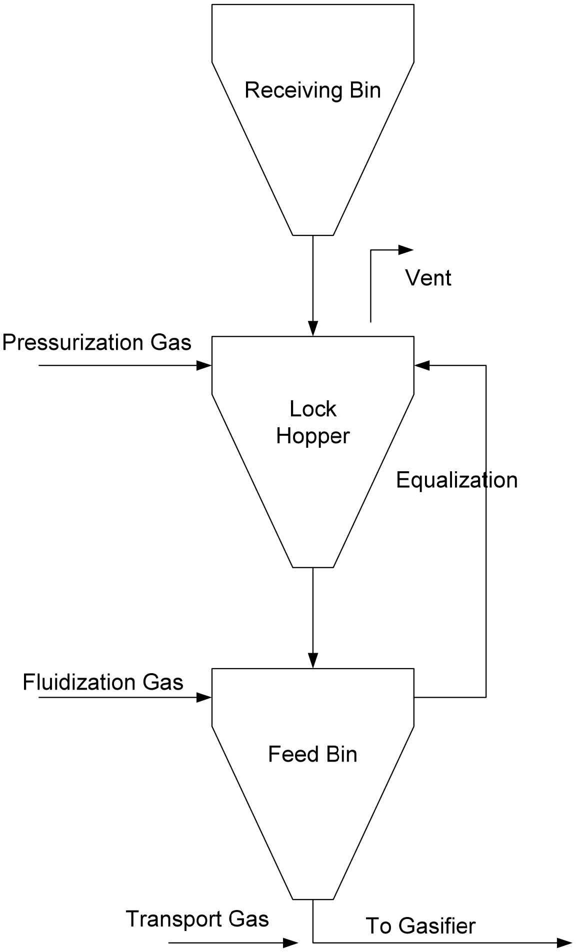
Volume of biomass feed bin $$V_{f\_bin} := \frac{t_{res\_fbin}\cdot m_{dot\_feed\_lock}}{\rho_{stover\_10\%moist}}\cdot\frac{1}{1-\varepsilon_{void}}$$

## LT Scenario Lockhopper System (7 trains)

Volume of biomass receiving bin $V_{r\_binLT} := \frac{V_{r\_bin}}{7}$ $V_{r\_binLT} = 61.909\,m^3$

Volume of biomass lockhopper $V_{lockLT} := \frac{V_{lock}}{7}$ $V_{lockLT} = 15.477\,m^3$

Volume of biomass feed bin $V_{f\_binLT} := \frac{V_{f\_bin}}{7}$ $V_{f\_binLT} = 15.477\,m^3$



Source: Combustion Engineering 1993

## *Lockhopper Power Consumption*

Source: *Techno-Economic Analysis of Hydrogen Production by Gasification of Biomass* by Lau et al. [2002]

Specific power of lockhopper, kW/tonne/day $SP_{lock} := 0.082\frac{kW}{\frac{tonne}{day}}$

Biomass inlet to gasifier $m_{dot\_gasifier} := m_{dot\_biomass} + m_{dot\_moist\_dried}$

$Power_{lock} := SP_{lock} \cdot m_{dot\_gasifier}$ $Power_{lock} = 182.222\,kW$

## *Fly Ash Collection Storage Tank (assume 7 days storage)*

$\rho_{ash} := 700\frac{kg}{m^3}$ (assumed) $m_{dot\_ash} := 5.88\frac{tonne}{day}$

$V_{tank} := \frac{m_{dot\_ash}}{\rho_{ash}} \cdot 7day$ $V_{tank} = 58.8\,m^3$ $V_{tank} = 2.077\times 10^3\,ft^3$

## *Slag Separation Drum (5 minute residence time, 20% volume)*

$\rho_{slag} := 2700\frac{kg}{m^3}$ $m_{dot\_slag} := 114\frac{tonne}{day}$ $\varepsilon_{void\_slag} := 0.8$

$V_{drum} := \frac{m_{dot\_slag}}{\rho_{slag}} \cdot 5min \cdot \frac{1}{1 - \varepsilon_{void\_slag}}$ $V_{drum} = 0.733\,m^3$

## *Slag Collection Storage Tank (7 days storage)*

$V_{slag\_storage} = 295.6\,m^3$

$V_{slag\_storage} := \frac{m_{dot\_slag}}{\rho_{slag}} \cdot 7day$

## *Char Collection Storage Bin (1-day residence time, 80% volume)*

$\rho_{char} := 2700\frac{kg}{m^3}$ $m_{dot\_char} := 214\frac{tonne}{day}$ $\varepsilon_{void\_char} := 0.2$ $V_{chardrum} = 99.074\,m^3$

assume 20% voidage

$V_{chardrum} := \frac{m_{dot\_char}}{\rho_{char}} \cdot 1day \cdot \frac{1}{1 - \varepsilon_{void\_char}}$

Note: the resulting volumes are used to assist in costing using Aspen Icarus

## ***Pressure Swing Adsorption Unit Sizing***

References in parentheses are given at the end of this section.

$\mathrm{Pi} := 3.1415 \quad \mathrm{nm} := 10^{-9} \cdot \mathrm{m}$

The adsorption unit is 1/3 molsieve and 2/3 activated carbon

Molsieve 13X

(a) $\mathrm{BulkDens} := 43 \cdot \frac{\mathrm{lb}}{\mathrm{ft}^3}$

(b) $\mathrm{SA} := 1320 \cdot \frac{\mathrm{m}^2}{\mathrm{gm}}$

(b) $\mathrm{PoreVol} := 0.51 \cdot \frac{\mathrm{cm}^3}{\mathrm{gm}}$

Determine dry volumetric flow rate of the syngas stream at atmospheric pressure and 25 deg C

$$\mathrm{VolFlowRate} := (167 - 1) \cdot \frac{\mathrm{kmol}}{\mathrm{hr}} \cdot 22.414 \cdot \frac{\mathrm{m}^3}{\mathrm{kmol}} \cdot \frac{14.696 \cdot \mathrm{psi}}{400 \cdot \mathrm{psi}} \cdot \frac{(273.15 + 25) \cdot \mathrm{K}}{273.15 \cdot \mathrm{K}}$$

$$\mathrm{VolFlowRate} = 149.211 \frac{\mathrm{m}^3}{\mathrm{hr}}$$

Mole fraction of components that are adsorbed

$\mathrm{CO} := 23 \quad \mathrm{CO2} := 1 \quad \mathrm{CH4} := 1$

Actual flow rate of components adsorbed

$$\mathrm{FlowRateAds} := \mathrm{VolFlowRate} \cdot \frac{\mathrm{CO} + \mathrm{CO2} + \mathrm{CH4}}{100}$$

$$\mathrm{FlowRateAds} = 37.303 \frac{\mathrm{m}^3}{\mathrm{hr}}$$

Adsorbent capacity

$$\mathrm{AdsCap} := 0.34 \cdot \frac{\mathrm{ft}^3}{\mathrm{lb}} \cdot \frac{14.696 \cdot \mathrm{psi}}{400 \cdot \mathrm{psi}} \cdot \frac{(273.15 + 25) \cdot \mathrm{K}}{273.15 \cdot \mathrm{K}}$$

(d) SCF/lb corrected for P and T to actual $\mathrm{cm}^3$/gm; PSA occurs at ambient temperature

$$\mathrm{AdsCap} = 0.851 \frac{\mathrm{cm}^3}{\mathrm{gm}}$$

Mass of molsieve required

$\mathrm{CycleTime} := 5 \cdot \mathrm{min}$

$$\mathrm{MolSieveMass} := \frac{\mathrm{FlowRateAds} \cdot \mathrm{CycleTime}}{\mathrm{AdsCap}} \qquad \mathrm{MolSieveMass} = 3.652 \times 10^3 \ \mathrm{kg}$$

Determine volume and length of molsieve bed and activated carbon bed

$\mathrm{BedVolume} := \dfrac{\mathrm{MolSieveMass}}{\mathrm{BulkDens}}$ $\mathrm{BedVolume} = 5.302\ \mathrm{m}^3$

$\mathrm{Diam} := 4 \cdot \mathrm{ft}$ $\mathrm{Diam} = 1.219\,\mathrm{m}$ (assumed)

$\mathrm{Length} := \dfrac{\mathrm{BedVolume}}{\mathrm{Pi} \cdot \mathrm{Diam}^2} \times 4$ $\mathrm{Length} = 3.725\,\mathrm{ft}$ (Just molsieve bed)

$\mathrm{RxtrLength} := 3 \cdot \mathrm{Length}$ (bed is 1/3 molsieve, 2/3 activated carbon)

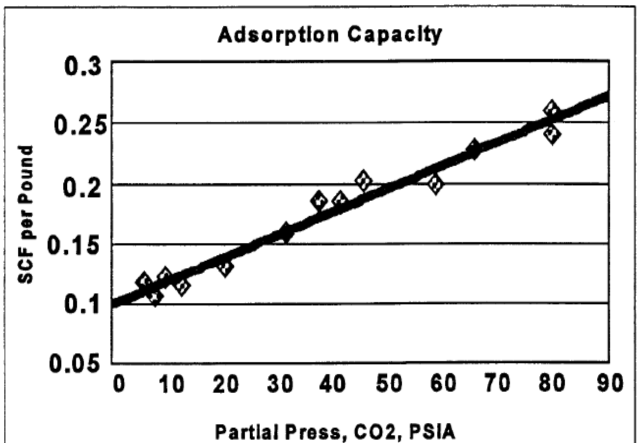
$\mathrm{RxtrLength} = 11.175\,\mathrm{ft}$ $\mathrm{RxtrLength} = 3.406\,\mathrm{m}$

$\mathrm{RxtrVolume} := \mathrm{RxtrLength} \cdot \mathrm{Diam}^2 \cdot 0.25 \cdot \pi$ $\mathrm{RxtrVolume} = 3.977\,\mathrm{m}^3$

(a) http://www.sigmaaldrich.com/Brands/Aldrich/Tech_Bulletins/AL_143/Molecular_Sieves.html

(b) US Pat 6117810

(d) WO/1998/058726 BULK SEPARATION OF CARBON DIOXIDE FROM METHANE USING NATURAL CLINOPTILOLITE --extrapolate to partial pressure of $CO_2+CH_4+N_2+CO=32.6\%*400$ psi



### *HT Scenario Fuel Storage*

## Gasoline Storage Tank (30 days storage)

$$m_{dot\_gasHT} = 112.78\frac{tonne}{day}$$

$$v_{dot\_gasHT} = 4.041\times 10^{4}\frac{gal}{day}$$

$$V_{gas\_tankHT} := v_{dot\_gasHT}\cdot 30day$$

$$V_{gas\_tankHT} = 4589m^{3}$$

## Diesel Storage Tank (30 days storage)

$$m_{dot\_dieselHT} = 266.5\frac{tonne}{day}$$

$$v_{dot\_dieselHT} = 8.381\times 10^{4}\frac{gal}{day}$$

$$V_{diesel\_tankHT} := v_{dot\_dieselHT}\cdot 30day$$

$$V_{diesel\_tankHT} = 9518m^{3}$$

Note: the resulting volumes are used to assist in costing using Aspen Icarus software

### *LT Scenario Fuel Storage*

Gasoline Storage Tank (30 days storage)

$$m_{dot\_gasLT} = 87.12\frac{\text{tonne}}{\text{day}}$$

$$v_{dot\_gasLT} = 3.122\times 10^{4}\ \frac{\text{gal}}{\text{day}}$$

$$V_{gas\_tankLT} := v_{dot\_gasLT}\cdot 30\text{day}$$

$$V_{gas\_tankLT} = 3545\ \text{m}^{3}$$

Diesel Storage Tank (30 days storage)

$$m_{dot\_dieselLT} = 205.86\frac{\text{tonne}}{\text{day}}$$

$$v_{dot\_dieselLT} = 6.474\times 10^{4}\ \frac{\text{gal}}{\text{day}}$$

$$V_{diesel\_tankLT} := v_{dot\_dieselLT}\cdot 30\text{day}$$

$$V_{diesel\_tankLT} = 7352\ \text{m}^{3}$$

Note: the resulting volumes are used to assist in costing using Aspen Icarus

## *LT Gasifier Cost*

Source: *Larson et al. 2005* in 2003$

$C_{0\_gasifier} := 6.41 \cdot 10^{6}$ $MM $S_{0\_gasifier} := 41.7 \frac{tonne}{hr}$ $S_{max} := 120 \frac{tonne}{hr}$ $f := 0.7$

Biomass throughput of 300 tpd

$S_{gasifierLT} := 300 \frac{ton}{day}$ $S_{gasifierLT} = 11.34 \frac{tonne}{hr}$

The cost ($MM) of one train at 300 ton per day

$$C_{gasifierLT} := C_{0\_gasifier} \cdot \left( \frac{S_{gasifierLT}}{\frac{tonne}{hr}} \cdot \frac{1}{\frac{S_{0\_gasifier}}{\frac{tonne}{hr}}} \right)^{f}$$

$C_{gasifierLT} = 2.576 \times 10^{6}$ $MM

Because 2,205 ton/day we need 7 gasifiers but we can apply the multiple train scaling exponent

$m_{train} := 0.9$

$C_{gasifierLTtrain} := C_{gasifierLT} \cdot 7^{m_{train}}$ $C_{gasifierLTtrain} = 1.484 \times 10^{7}$ $MM

## *FT Reactor Costing*

Source: *Larson et al. 2005* in 2003$

$C_{FT\_base} := 10.5$ $MM $\qquad f_{FT2} := 0.72 \qquad S_{FT\_base} := 2.52\frac{MMcf}{hr}$

## *HT Scenario*

$M_{dot\_FTHT} := 13829\frac{kmol}{hr} \qquad V_{standard\_FTHT} := M_{dot\_FTHT}\cdot 22.4\frac{L}{mol}$

$$V_{standard\_FTHT} = 10.939\frac{MMcf}{hr}$$

$$C_{FTHT\_reac} := C_{FT\_base}\cdot\left(\frac{V_{standard\_FTHT}}{S_{FT\_base}}\right)^{f_{FT2}}$$

$C_{FTHT\_reac} = 30.217$ Installed cost $MM (assume 3.6 install factor consistent with Peters et al.)

## *LT Scenario*

$M_{dot\_FTLT} := 11400\frac{kmol}{hr} \qquad V_{standard\_FTLT} := M_{dot\_FTLT}\cdot 22.4\frac{L}{mol}$

$$V_{standard\_FTLT} = 9.018\frac{MMcf}{hr}$$

$$C_{FTLT\_reac} := C_{FT\_base}\cdot\left(\frac{V_{standard\_FTLT}}{S_{FT\_base}}\right)^{f_{FT2}}$$

$C_{FTLT\_reac} = 26.294$ Installed cost $MM (assume 3.6 install factor consistent with Peters et al.)

## Acid-Gas Removal Area Cost

Source: Phillips et al. 2007 in 2005$

Calculated by adding the input syngas streams to the absorber column

$S_{AGR\_base} := 332910 \frac{lb}{hr}$ $\quad f_{AGR} := 0.65$ $\quad C_{AGR\_base} := 5446503$

## HT Scenario

$S_{AGR\_HT} := (2965 + 2308) \cdot \frac{tonne}{day}$ $\quad S_{AGR\_HT} = 484374 \frac{lb}{hr}$

$$C_{AGR\_HT} := C_{AGR\_base} \cdot \left( \frac{S_{AGR\_HT}}{\frac{lb}{hr}} \cdot \frac{1}{\frac{S_{AGR\_base}}{\frac{lb}{hr}}} \right)^{f_{AGR}}$$

$C_{AGR\_HT} = 6949808$

## LT Scenario

$S_{AGR\_LT} := (2070 + 2190) \frac{tonne}{day}$ $\quad S_{AGR\_LT} = 391321 \frac{lb}{hr}$

$$C_{AGR\_LT} := C_{AGR\_base} \cdot \left( \frac{S_{AGR\_LT}}{\frac{lb}{hr}} \cdot \frac{1}{\frac{S_{AGR\_base}}{\frac{lb}{hr}}} \right)^{f_{AGR}}$$

$C_{AGR\_LT} = 6049946$

## *A500 Hydroprocessing Area Cost*

Source: Robinson et al. 2007 in 2007$

Note: "bpsd" is barrels per standard day

$AreaCost_0 := \frac{4000}{bpsd}$    $S_{0\_HY} := 25000\ bpsd$    $f_{HY} := 0.65$ (assumed)

$C_{0\_HY} := AreaCost_0 \cdot S_{0\_HY}$    $C_{0\_HY} = 100000000$

## *HT Scenario*

$m_{dot\_FTL\_HT} := 428 \frac{tonne}{day}$    $\rho_{FTL} := 750 \frac{kg}{m^3}$    (from Aspen Plus model)

$v_{dot\_FTL\_HT} := \frac{m_{dot\_FTL\_HT}}{\rho_{FTL}}$    $v_{dot\_FTL\_HT} = 3.589 \times 10^3\ bpsd$

$$C_{HY\_HT} := C_{0\_HY} \cdot \left( \frac{v_{dot\_FTL\_HT}}{bpsd} \cdot \frac{1}{\frac{S_{0\_HY}}{bpsd}} \right)^{f_{HY}}$$

$C_{HY\_HT} = 2.832 \times 10^7$

### Power required for A500

$Power_{per\_bpsd} := \frac{15kW \cdot hr}{bpsd \cdot day}$

$Power_{areaHT} := Power_{per\_bpsd} \cdot v_{dot\_FTL\_HT}$    $Power_{areaHT} = 2.243\ MW$

## *LT Scenario*

$m_{dot\_FTL\_LT} := 330.42 \frac{tonne}{day}$    (from Aspen Plus model)

$v_{dot\_FTL\_LT} := \frac{m_{dot\_FTL\_LT}}{\rho_{FTL}}$    $v_{dot\_FTL\_LT} = 2.771 \times 10^3\ bpsd$

$$C_{HY\_LT} := C_{0\_HY} \cdot \left( \frac{v_{dot\_FTL\_LT}}{bpsd} \cdot \frac{1}{\frac{S_{0\_HY}}{bpsd}} \right)^{f_{HY}}$$

$C_{HY\_LT} = 2.394 \times 10^7$

### Power required for A500

$Power_{areaLT} := Power_{per\_bpsd} \cdot v_{dot\_FTL\_LT}$    $Power_{areaLT} = 1.732\ MW$

## Reactors and Catalysts

### Fischer-Tropsch Reactor and Cobalt Catalyst

#### FT reactor volume

Using gas hourly space velocity and actual volumetric flow rate, the volume of the reactor is determined

$$\mathrm{GHSV} = \frac{\nu_0}{V}$$

$\mathrm{GHSV_{FT}} := 100 \cdot \mathrm{hr}^{-1}$ (assumed)

$\nu_{\mathrm{rate\_actHT}} := 6.298 \frac{\mathrm{m}^3}{\mathrm{s}}$ (from Aspen Plus model)

$V_{\mathrm{FTHT}} := \frac{\nu_{\mathrm{rate\_actHT}}}{\mathrm{GHSV_{FT}}}$ $V_{\mathrm{FTHT}} = 226.728\mathrm{m}^3$ $V_{\mathrm{FTHT}} = 8.007 \times 10^3\ \mathrm{ft}^3$

$\nu_{\mathrm{rate\_actLT}} := 5.021 \frac{\mathrm{m}^3}{\mathrm{s}}$

$V_{\mathrm{FTLT}} := \frac{\nu_{\mathrm{rate\_actLT}}}{\mathrm{GHSV_{FT}}}$ $V_{\mathrm{FTLT}} = 180.756\mathrm{m}^3$ $V_{\mathrm{FTLT}} = 6.383 \times 10^3\ \mathrm{ft}^3$

#### FT catalyst cost

$\mathrm{Co_{cost}} := \frac{15}{\mathrm{lb}}$ (assumed) $\rho_{\mathrm{Co}} := 64 \frac{\mathrm{lb}}{\mathrm{ft}^3}$ (assumed)

$\mathrm{Co_{vol\_cost}} := \mathrm{Co_{cost}} \cdot \rho_{\mathrm{Co}}$ $\mathrm{Co_{vol\_cost}} = 960 \frac{1}{\mathrm{ft}^3}$

Replacement cost of cobalt catalyst

$\mathrm{Co_{total\_costHT}} := \mathrm{Co_{vol\_cost}} \cdot V_{\mathrm{FTHT}}$ $\mathrm{Co_{total\_costHT}} = 7.687 \times 10^6$

$\mathrm{Co_{total\_costLT}} := \mathrm{Co_{vol\_cost}} \cdot V_{\mathrm{FTLT}}$ $\mathrm{Co_{total\_costLT}} = 6.128 \times 10^6$

## *Water Gas Shift Reactor and Catalyst*

### Sour WGS reactor volume (HT scenario)

Using gas hourly space velocity and actual volumetric flow rate, the volume of the reactor is determined

$$\mathrm{GHSV} = \frac{\nu_0}{V}$$

$\mathrm{GHSV}_{WGS} := 1000 \cdot \mathrm{hr}^{-1}$ (assumed)

$\nu_{rate\_actSWGS} := 2.008 \frac{m^3}{s}$ (from Aspen Plus)

$V_{SWGS} := \frac{\nu_{rate\_actSWGS}}{\mathrm{GHSV}_{WGS}}$ $V_{SWGS} = 7.229 m^3$ $V_{SWGS} = 255.283 ft^3$

### WGS reactor volume (LT scenario)

$\nu_{rate\_actWGS} := 1.834 \frac{m^3}{s}$ (from Aspen Plus)

$V_{WGS} := \frac{\nu_{rate\_actWGS}}{\mathrm{GHSV}_{WGS}}$ $V_{WGS} = 6.602 m^3$ $V_{WGS} = 233.162 ft^3$

### WGS and SWGS Catalyst Cost

$\mathrm{CatCost}_{WGS} := \frac{8}{lb}$ $\rho_{cat\_WGS} := 56 \frac{lb}{ft^3}$ $\rho_{cat\_WGS} = 897.034 \frac{kg}{m^3}$

$\mathrm{CatCost}_{vol\_WGS} := \mathrm{CatCost}_{WGS} \cdot \rho_{cat\_WGS}$ $\mathrm{CatCost}_{vol\_WGS} = 448 \frac{1}{ft^3}$

Replacement cost of WGS catalyst

$\mathrm{TotalCatCost}_{SWGS} := \mathrm{CatCost}_{vol\_WGS} \cdot V_{SWGS}$ $\mathrm{TotalCatCost}_{SWGS} = 114367$

$\mathrm{TotalCatCost}_{WGS} := \mathrm{CatCost}_{vol\_WGS} \cdot V_{WGS}$ $\mathrm{TotalCatCost}_{WGS} = 104456$

## *Steam Methane Reformer Reactor and Catalyst (LT Scenario)*

$GHSV_{SMR} := 2600 hr^{-1}$ (assumed)

$\nu_{rate\_actSMR} := 7.082 \frac{m^3}{s}$ (from Aspen Plus model)

$V_{SMR} := \frac{\nu_{rate\_actSMR}}{GHSV_{SMR}}$ $V_{SMR} = 9.806 m^3$ $V_{SMR} = 346.29 ft^3$

## SMR Catalyst Cost

$CatCost_{SMR} := \frac{4.67}{lb}$ $\rho_{cat\_SMR} := 64 \frac{lb}{ft^3}$ $\rho_{cat\_SMR} = 1.025 \times 10^3 \frac{kg}{m^3}$

$CatCost_{vol\_SMR} := CatCost_{SMR} \cdot \rho_{cat\_SMR}$ $CatCost_{vol\_SMR} = 298.88 \frac{1}{ft^3}$

### Replacement cost of SMR catalyst

$TotalCatCost_{SMR} := CatCost_{vol\_SMR} \cdot V_{SMR}$ $TotalCatCost_{SMR} = 103499$

## Natural Gas Utility Consumption

Annual natural gas requirement at 5% of yearly operating hours

Natural gas properties

Price (Source: Energy Information Administration)

$$HHV_{ng} := 54\frac{MJ}{kg} \qquad MW_{ng} := 16.04\frac{gm}{mol} \qquad \rho_{ng} := 22.4\frac{L}{mol} \qquad Cost_{ng} := \frac{6.4}{1000\,ft^3}$$

$$Cost_{ng} \cdot \frac{\rho_{ng}}{MW_{ng}} = 286.335\frac{1}{ton}$$

## HT Scenario

$P_{required\_plantHT} := 32.813MW$ (from Aspen Plus model, includes power required for gas turbine air compressor)

$Eff_{ng\_to\_power} := 0.35$ (assumed)

$$m_{dot\_ngHT} := \frac{\frac{P_{required\_plantHT}}{Eff_{ng\_to\_power}}}{HHV_{ng}} \qquad m_{dot\_ngHT} = 1.378\times 10^{4}\frac{lb}{hr}$$

Annual natural gas requirement

$$M_{ngHT} := m_{dot\_ngHT} \cdot Availability \cdot 0.05 \qquad M_{ngHT} = 2563\ ton$$

Average flow rate of natural gas

$$m_{dot\_ng\_5\%HT} := \frac{M_{ngHT}}{8760hr} \qquad m_{dot\_ng\_5\%HT} = 585.14\frac{lb}{hr}$$

## LT Scenario

$P_{required\_plantLT} := 24.3MW$ (from Aspen Plus model, includes power required for gas turbine air compressor)

$$m_{dot\_ngLT} := \frac{\frac{P_{required\_plantLT}}{Eff_{ng\_to\_power}}}{HHV_{ng}} \qquad m_{dot\_ngLT} = 1.02\times 10^{4}\,\frac{lb}{hr}$$

Annual natural gas requirement

$$M_{ngLT} := m_{dot\_ngLT}\cdot Availability\cdot 0.05 \qquad \boxed{M_{ngLT} = 1898\ ton}$$

Average flow rate of natural gas

$$m_{dot\_ng\_5\%LT} := \frac{M_{ngLT}}{8760\,hr} \qquad m_{dot\_ng\_5\%LT} = 433.331\,\frac{lb}{hr}$$

# Appendix D. Process Flow Diagrams

## High-Temperature Scenario

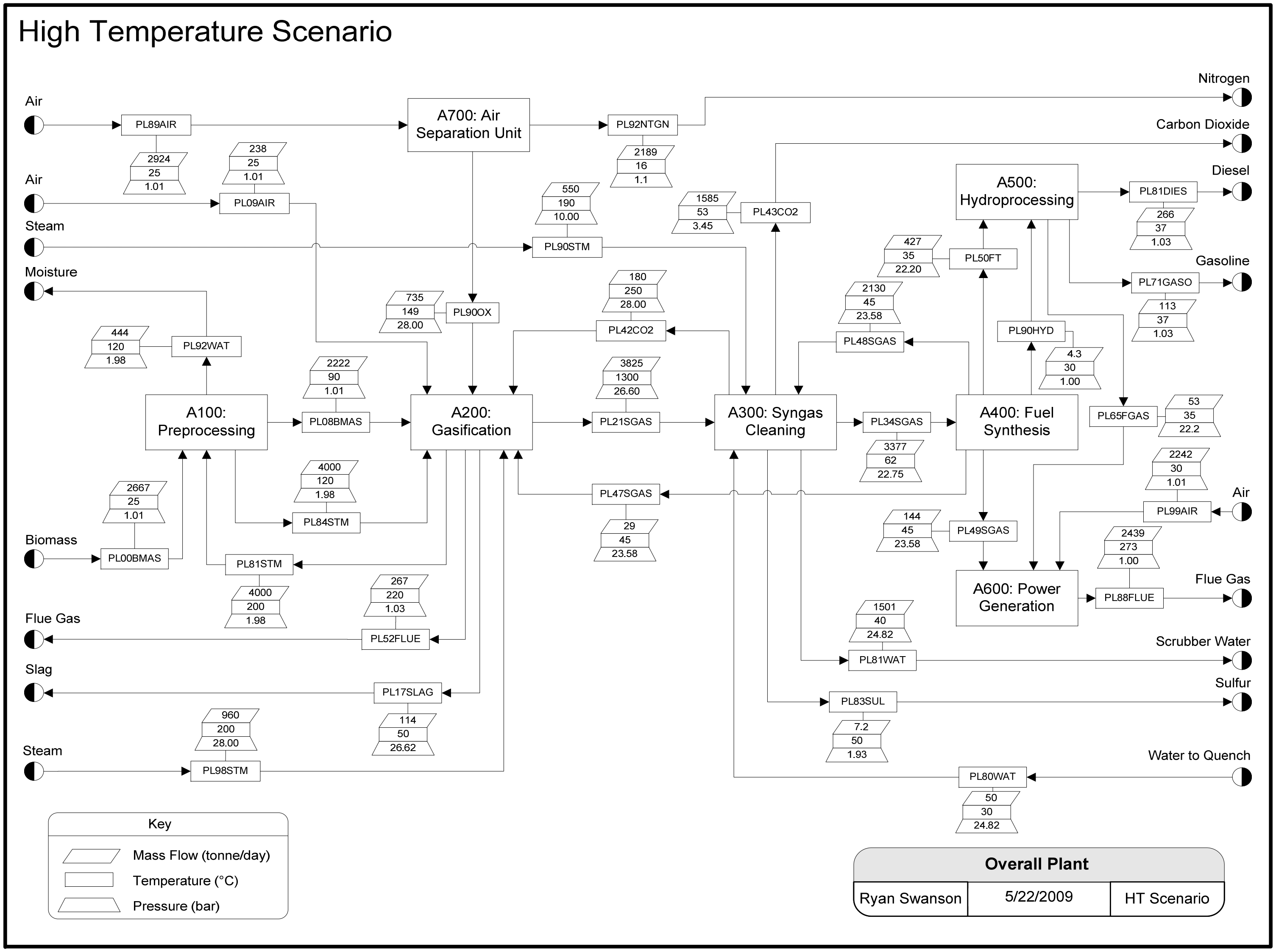


**Figure D-1. Overall plant area process flow diagram for HT scenario**

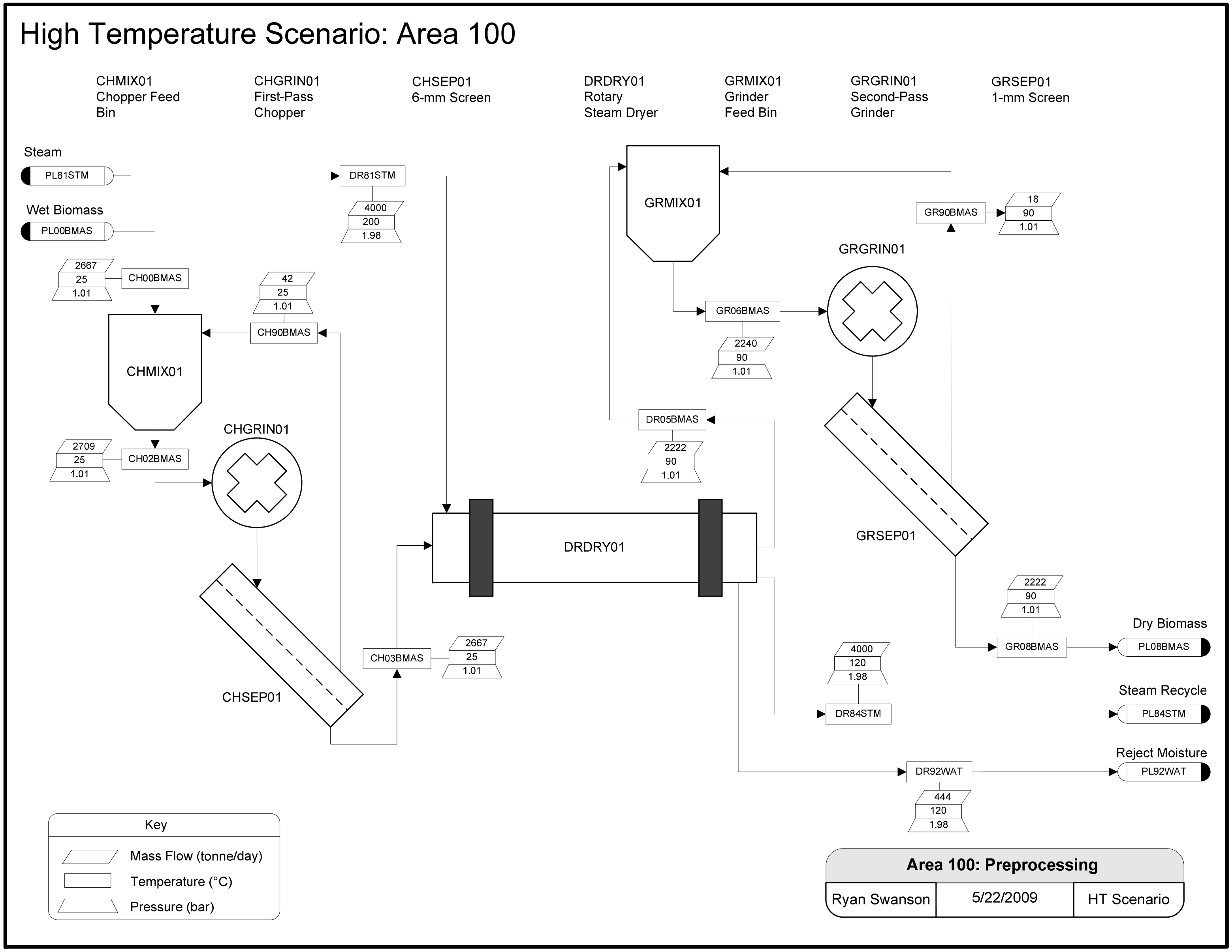


**Figure D-2. Preprocessing area process flow diagram for HT scenario**

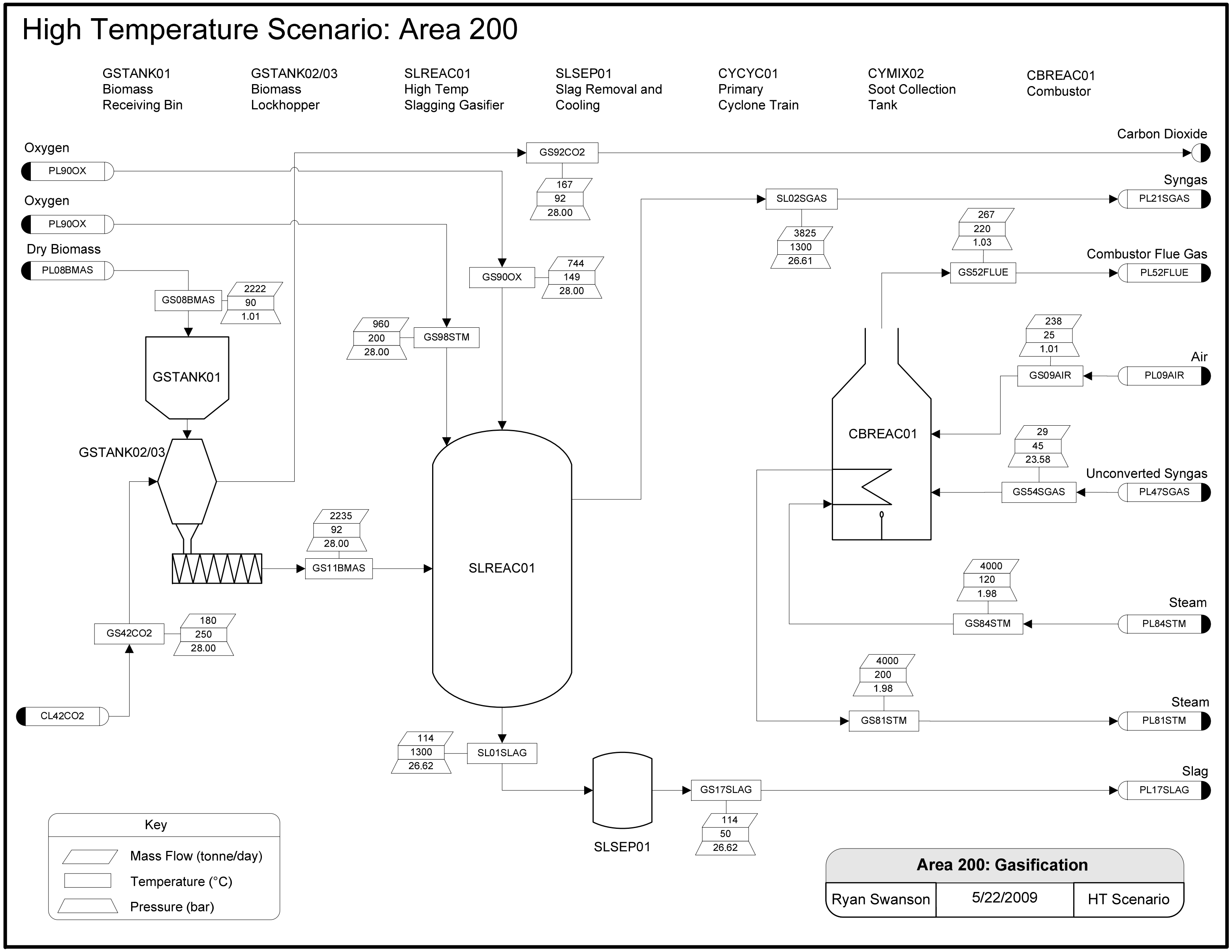


**Figure D-3. Gasification area process flow diagram for HT scenario**

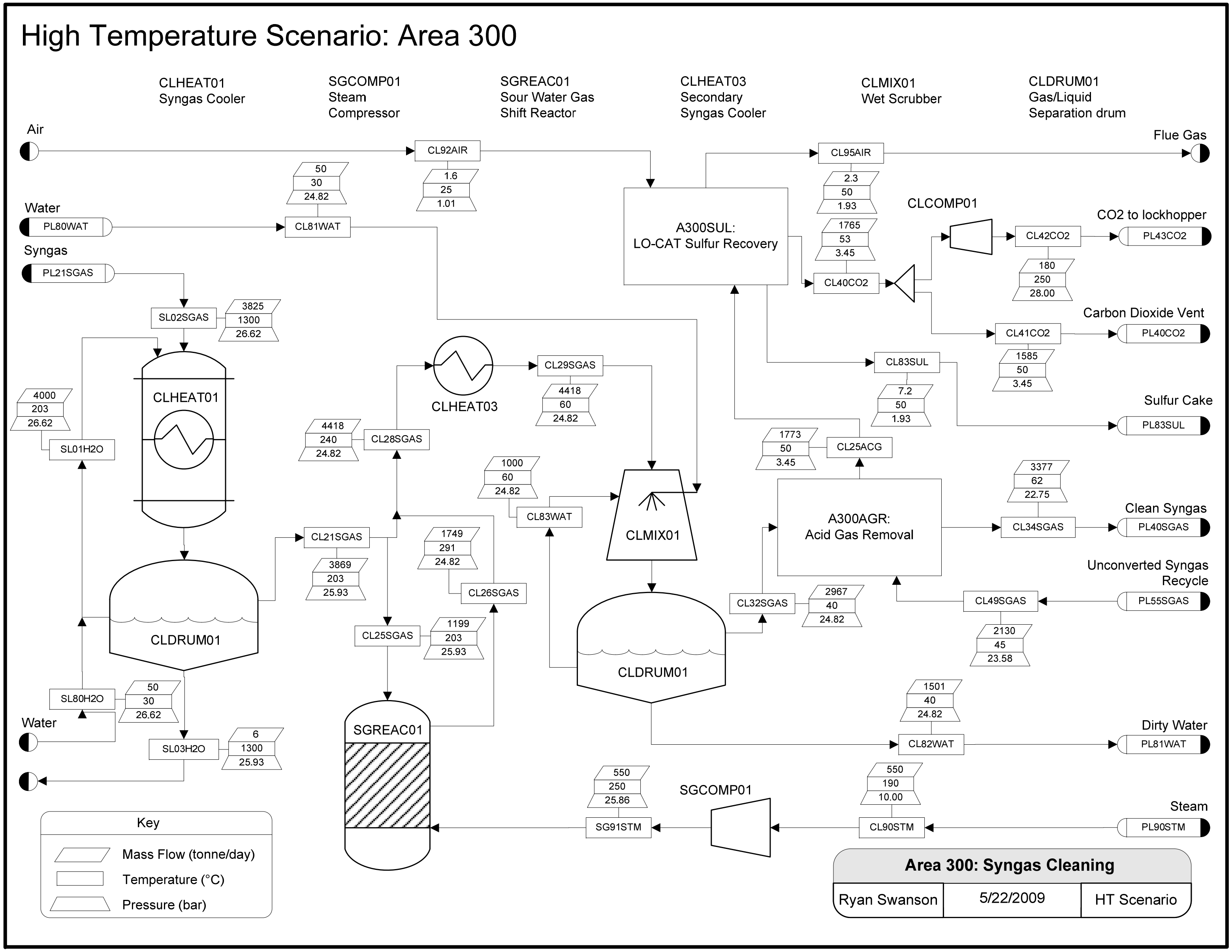


**Figure D-4. Syngas cleaning area process flow diagram for HT scenario**

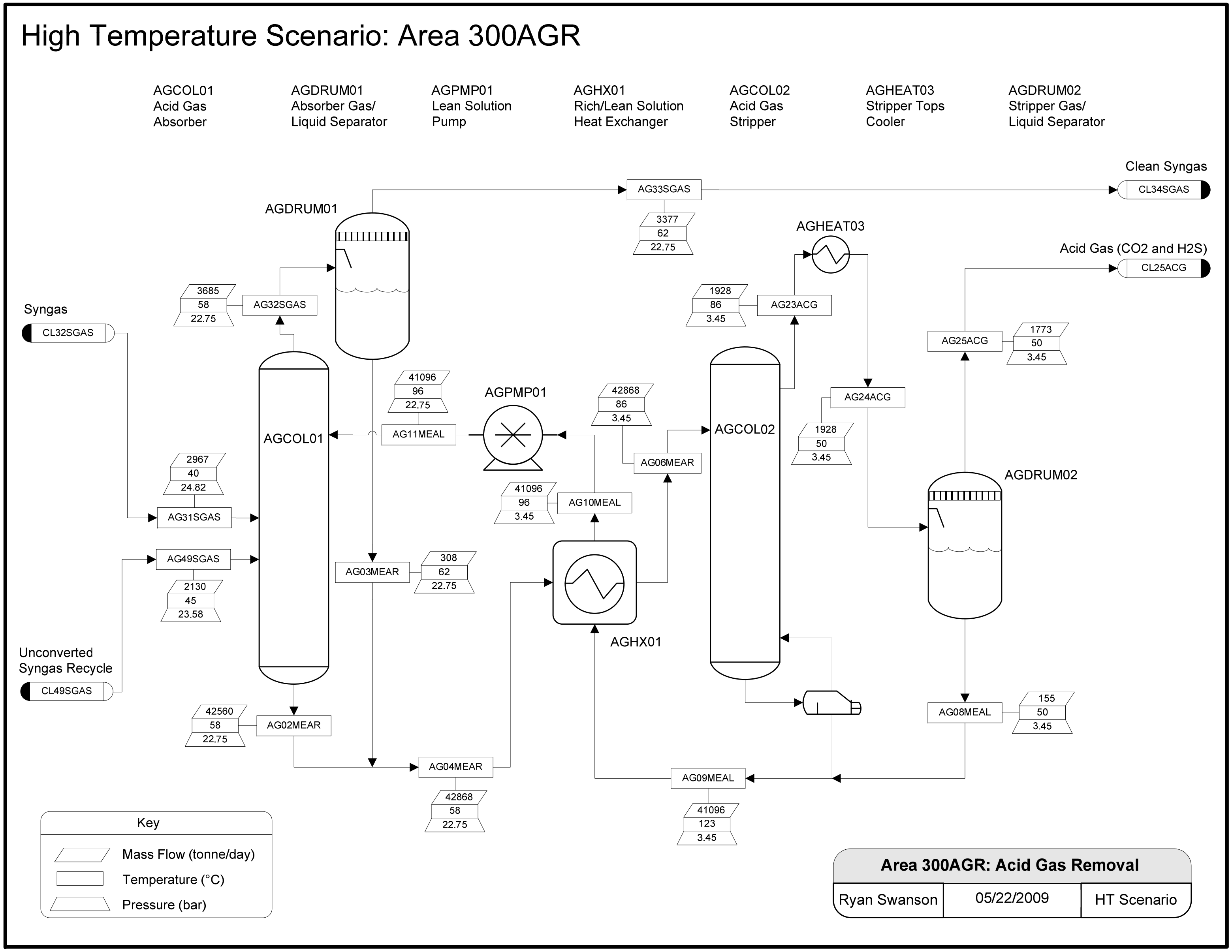


Figure D-5. Acid gas removal area process flow diagram for HT scenario

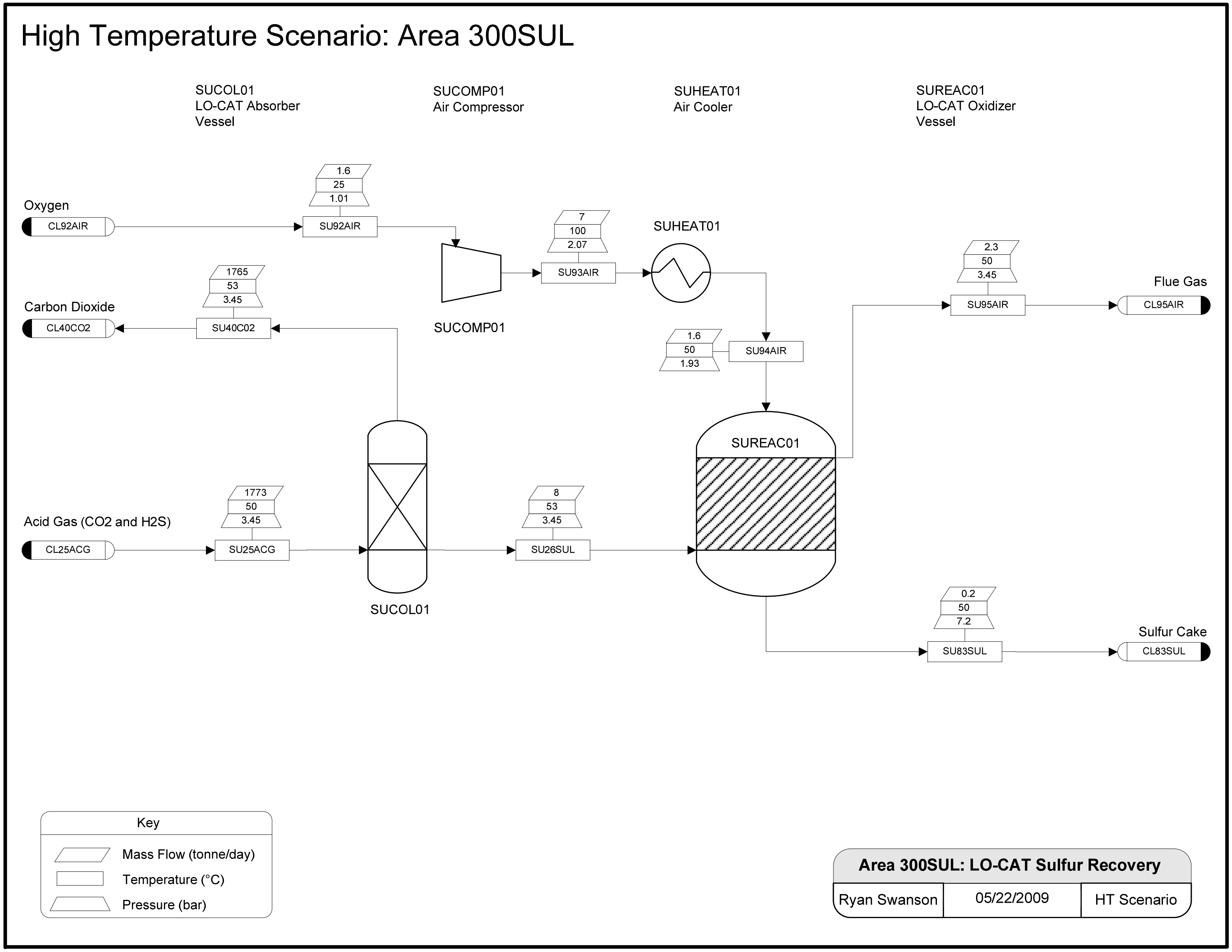


**Figure D-6. Sulfur recovery area process flow diagram for HT scenario**

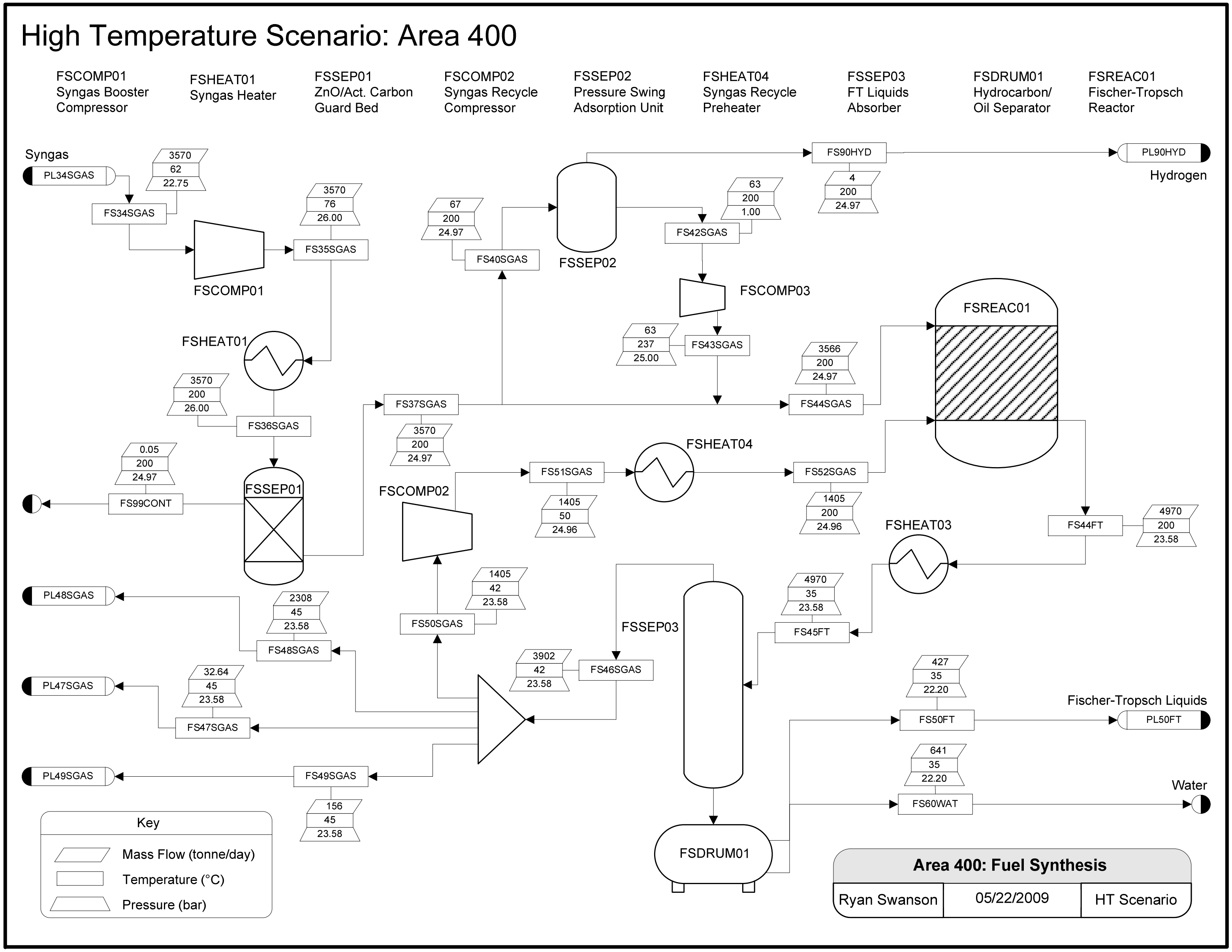


Figure D-7. Fuel synthesis area process flow diagram for HT scenario

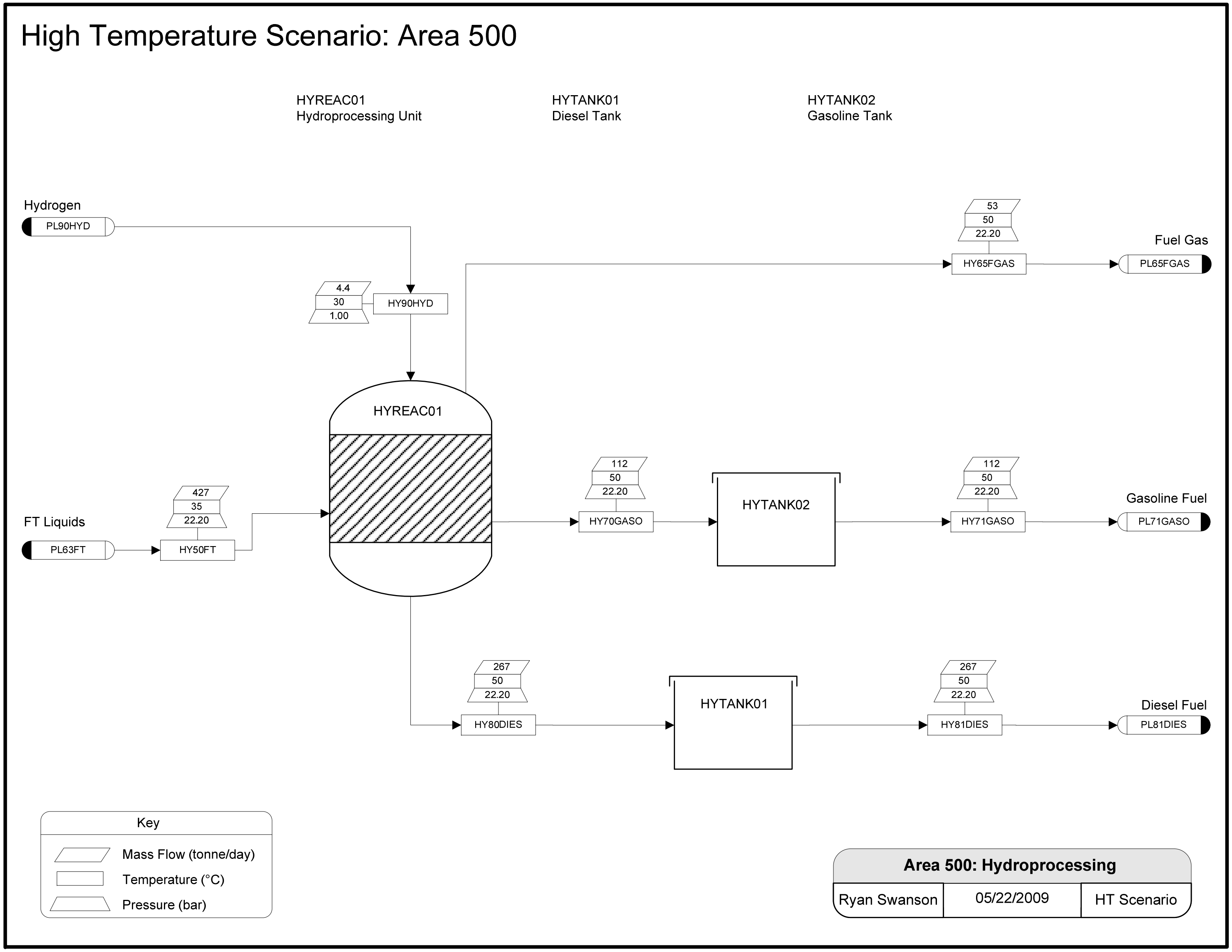


**Figure D-8. Hydroprocessing area process flow diagram for HT scenario**

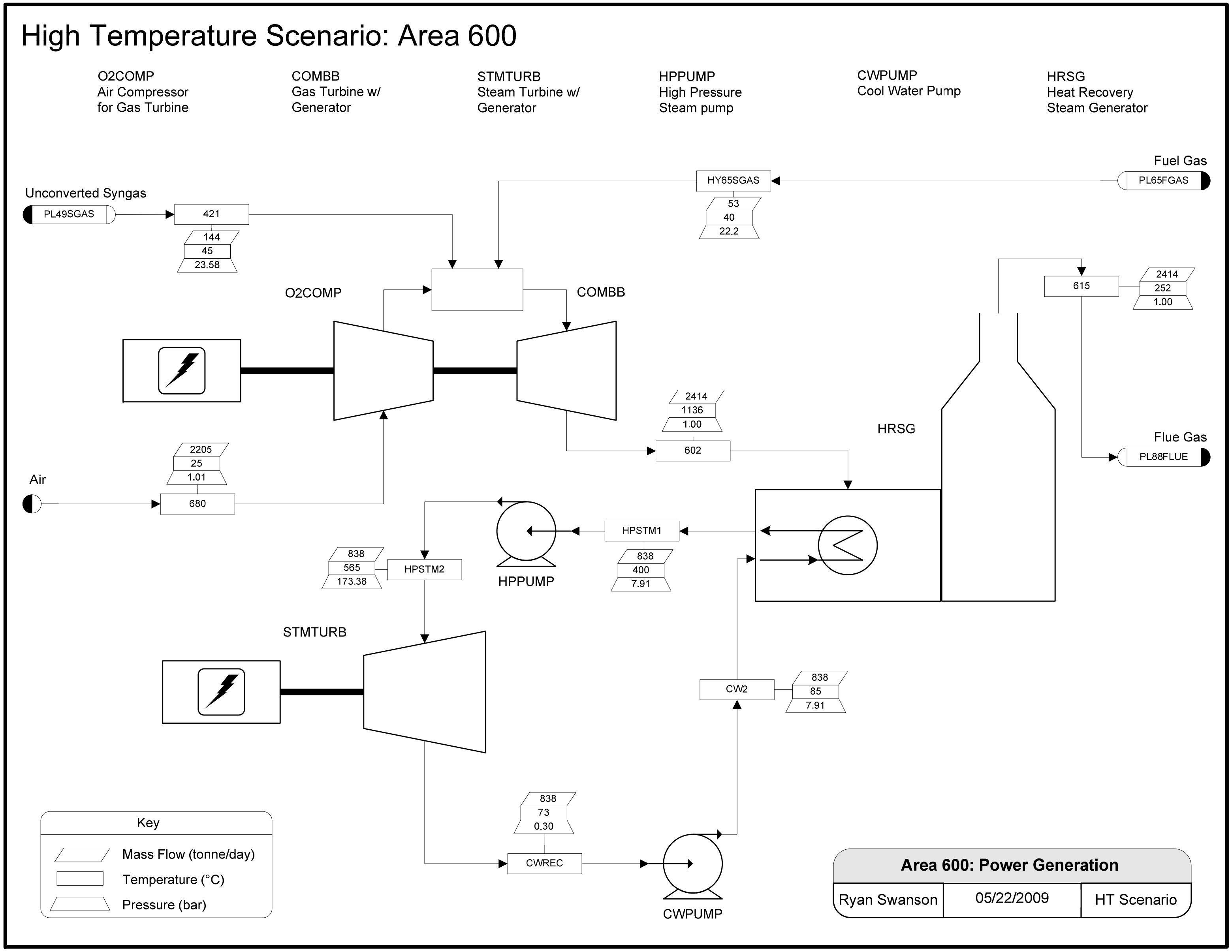


Figure D-9. Power generation area process flow diagram for HT scenario

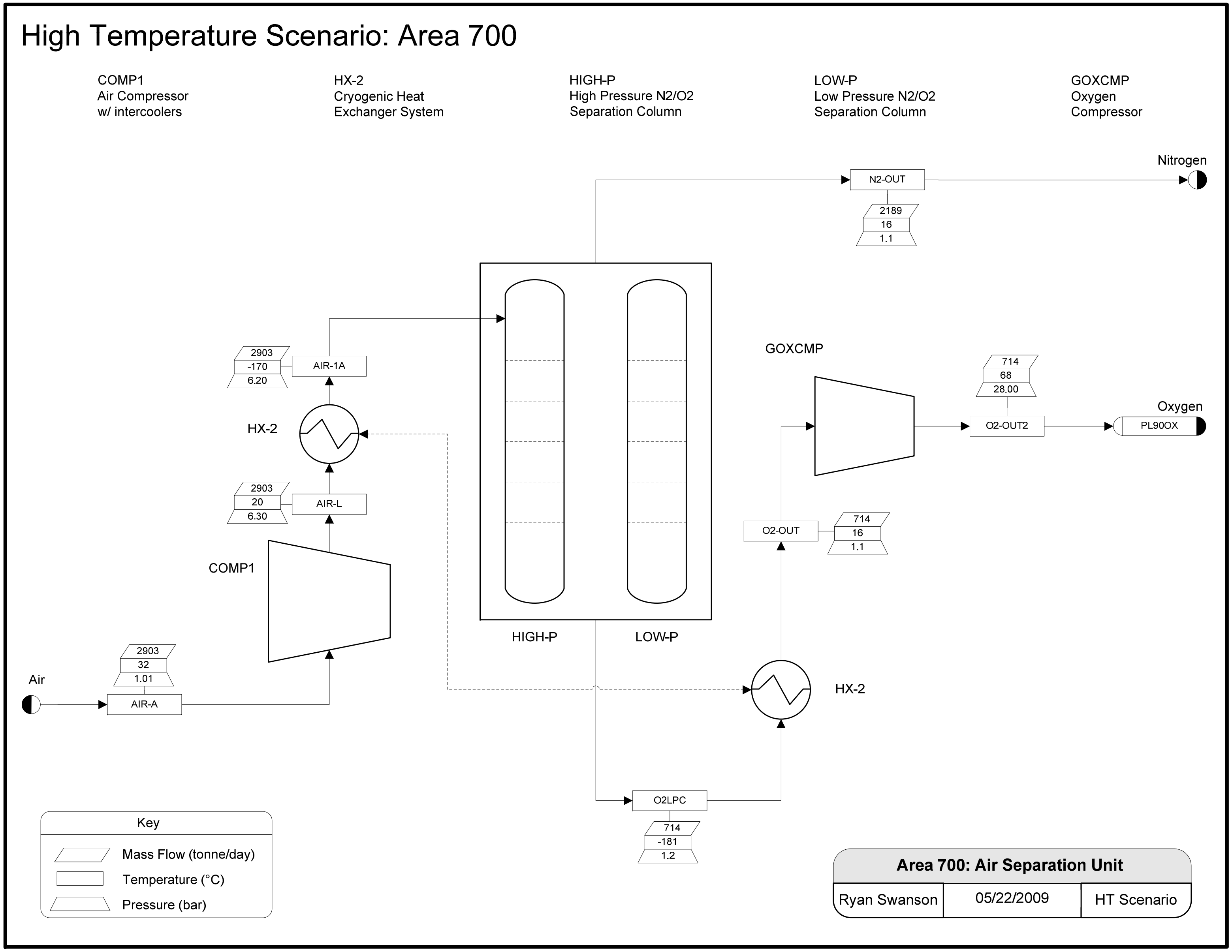


**Figure D-10. Air separation unit process flow diagram for HT scenario**

## Low-Temperature Scenario

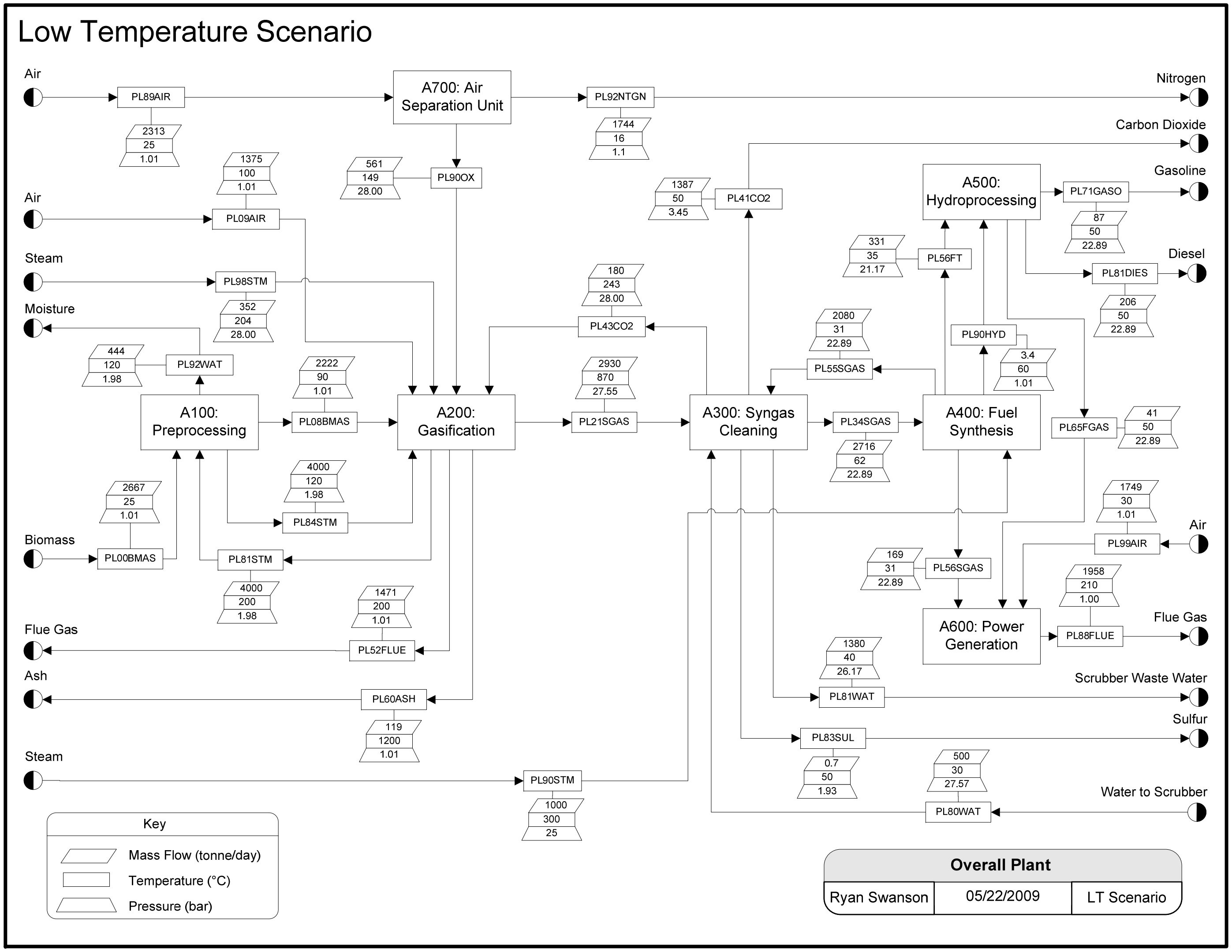


Figure D-11. Overall plant area process flow diagram for LT scenario

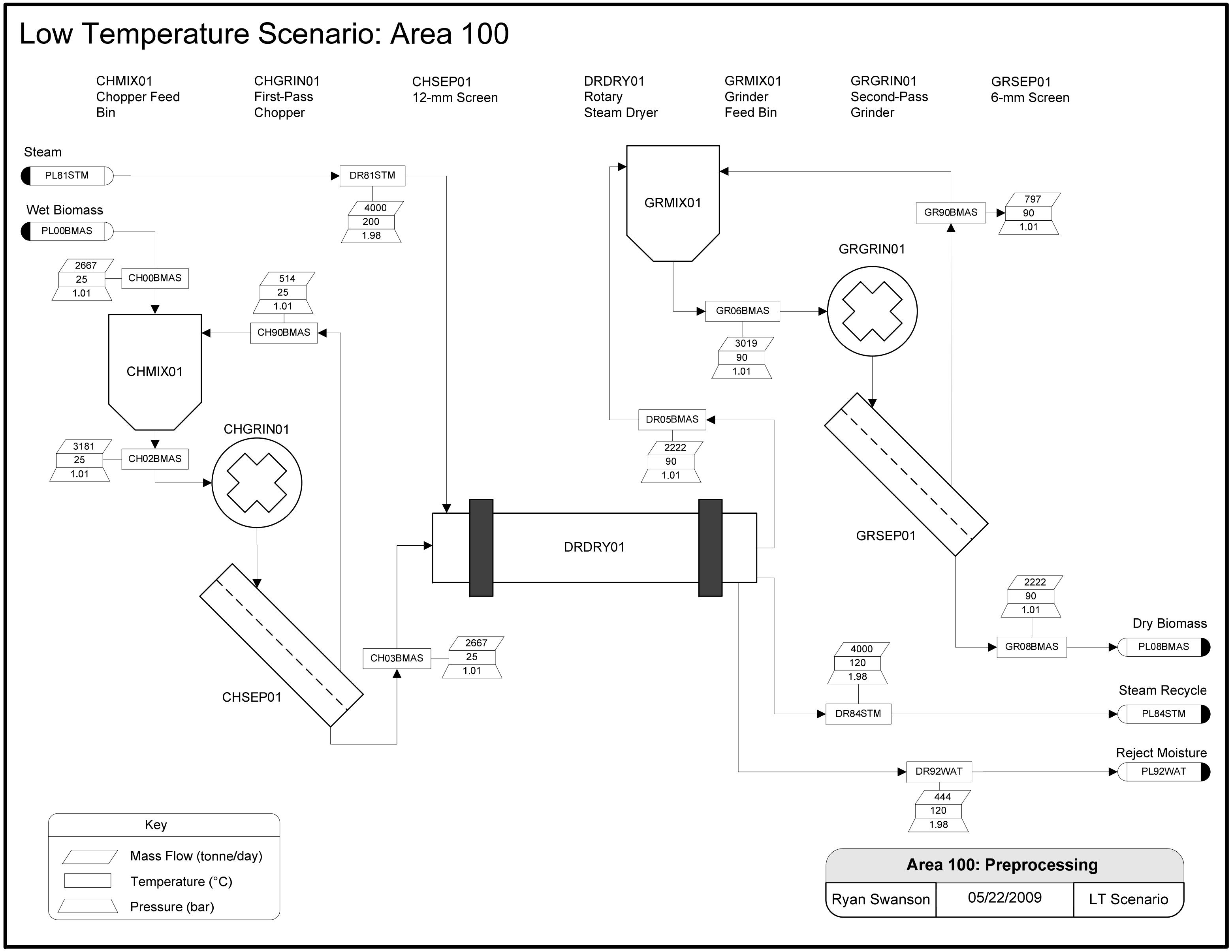


Figure D-12. Preprocessing area process flow diagram for LT scenario

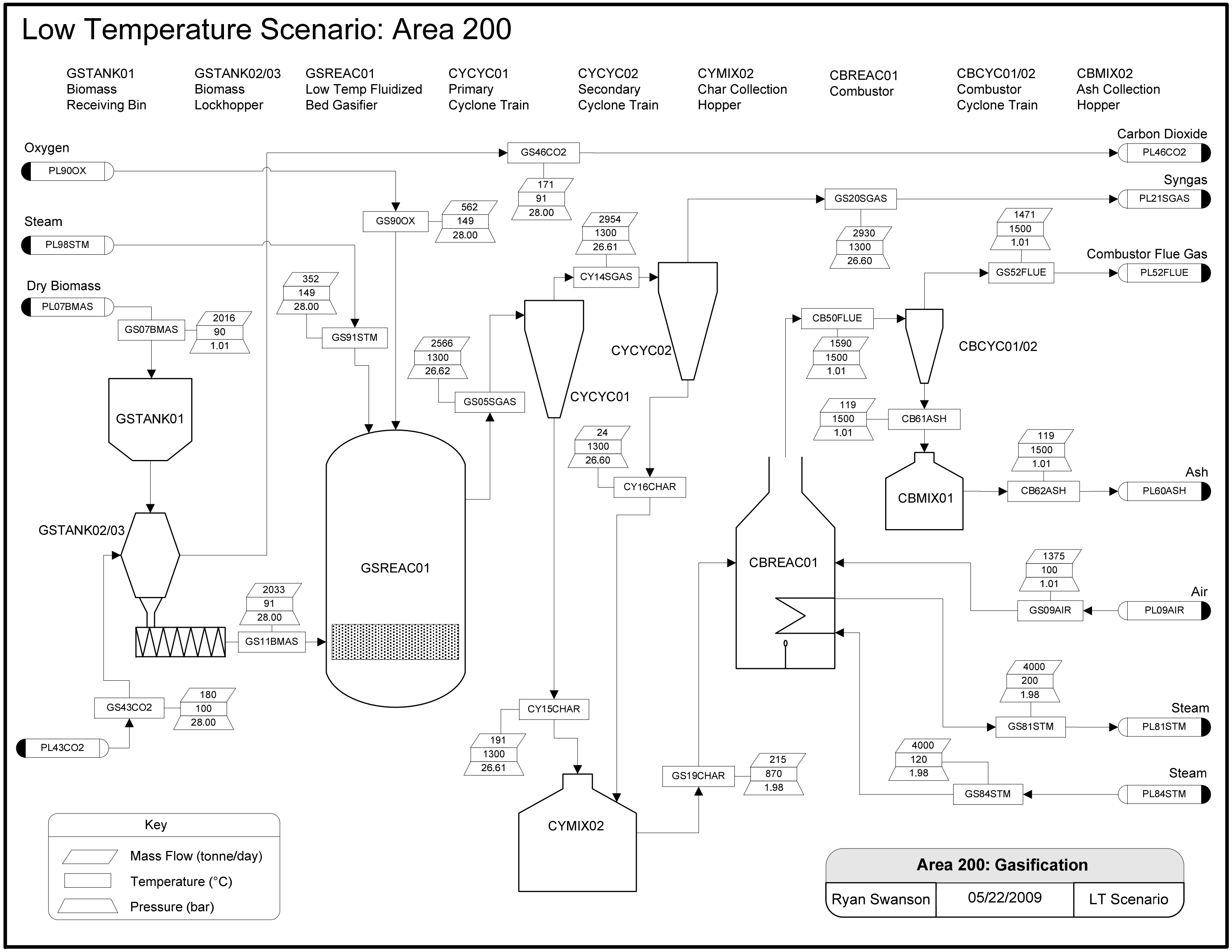


Figure D-13. Gasification area process flow diagram for LT scenario

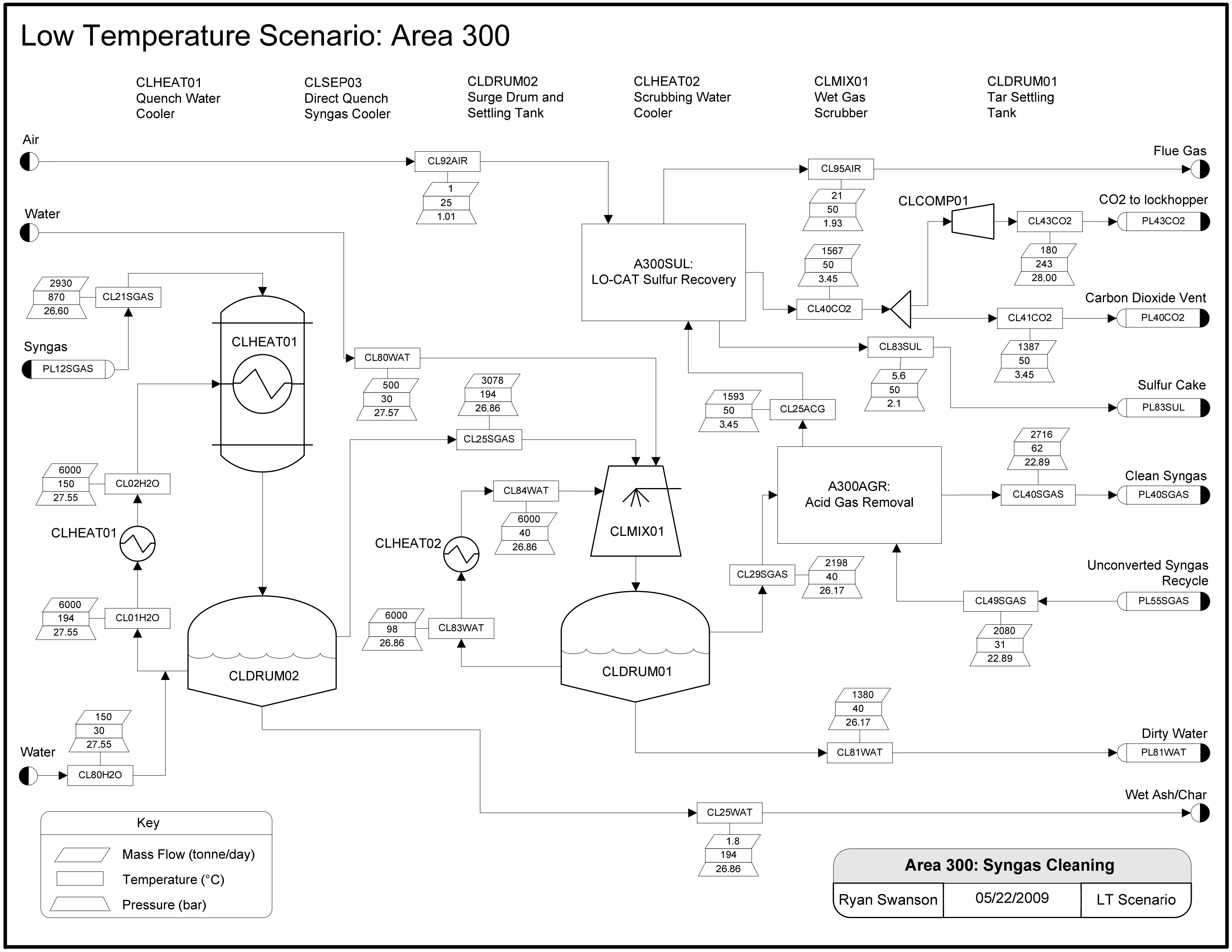


Figure D-14. Syngas cleaning area process flow diagram for LT scenario

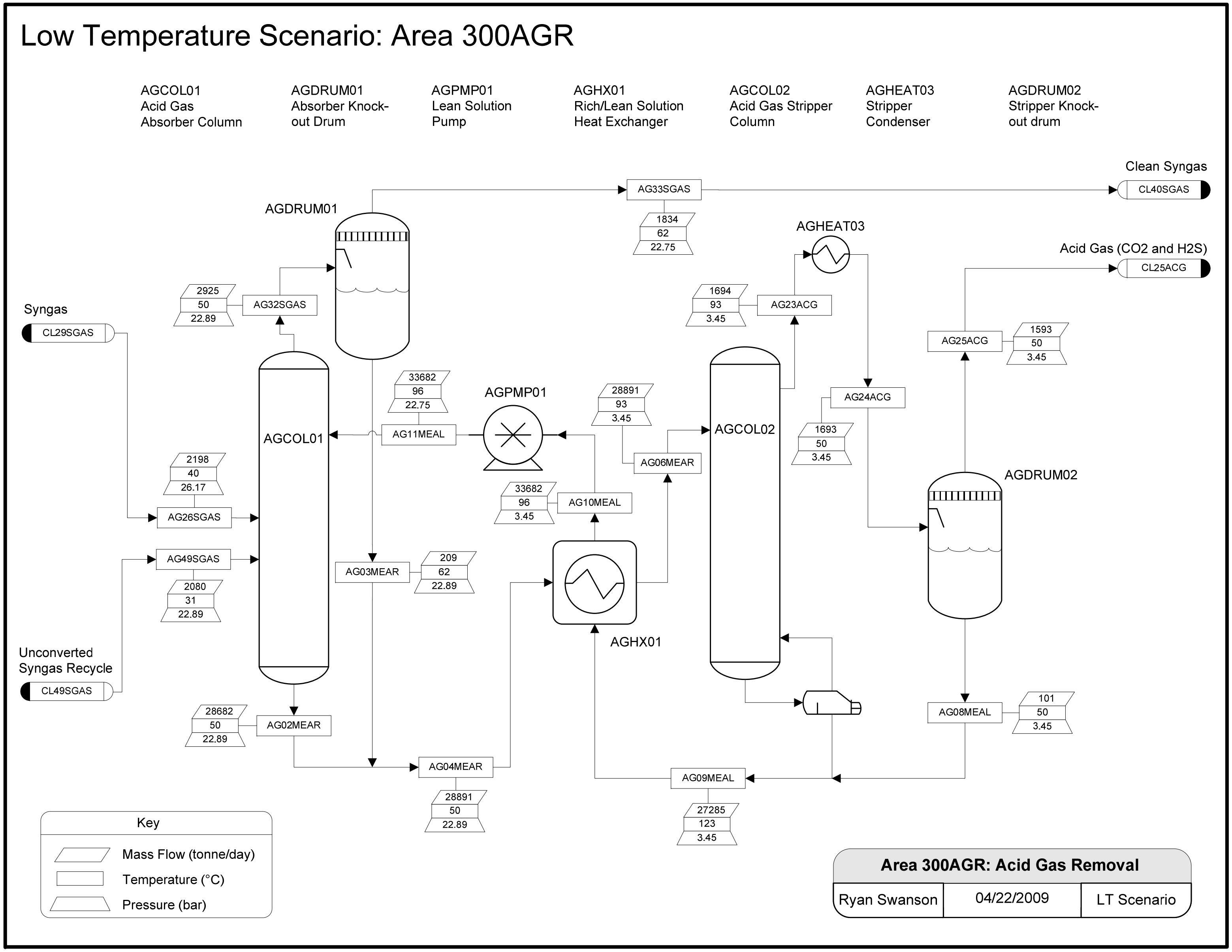


**Figure D-15. Acid gas removal area process flow diagram for LT scenario**

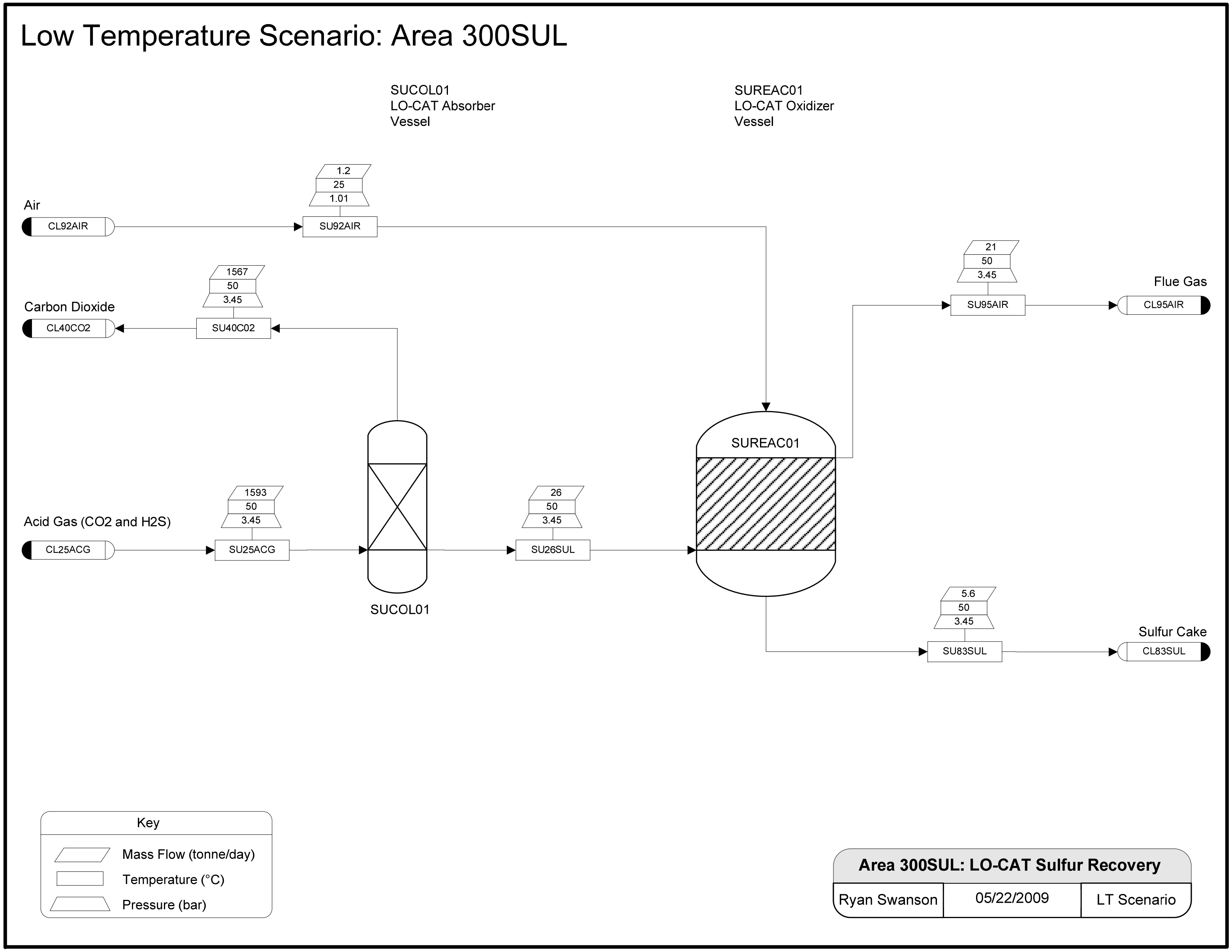


**Figure D-16. Sulfur recovery process flow diagram for LT scenario**

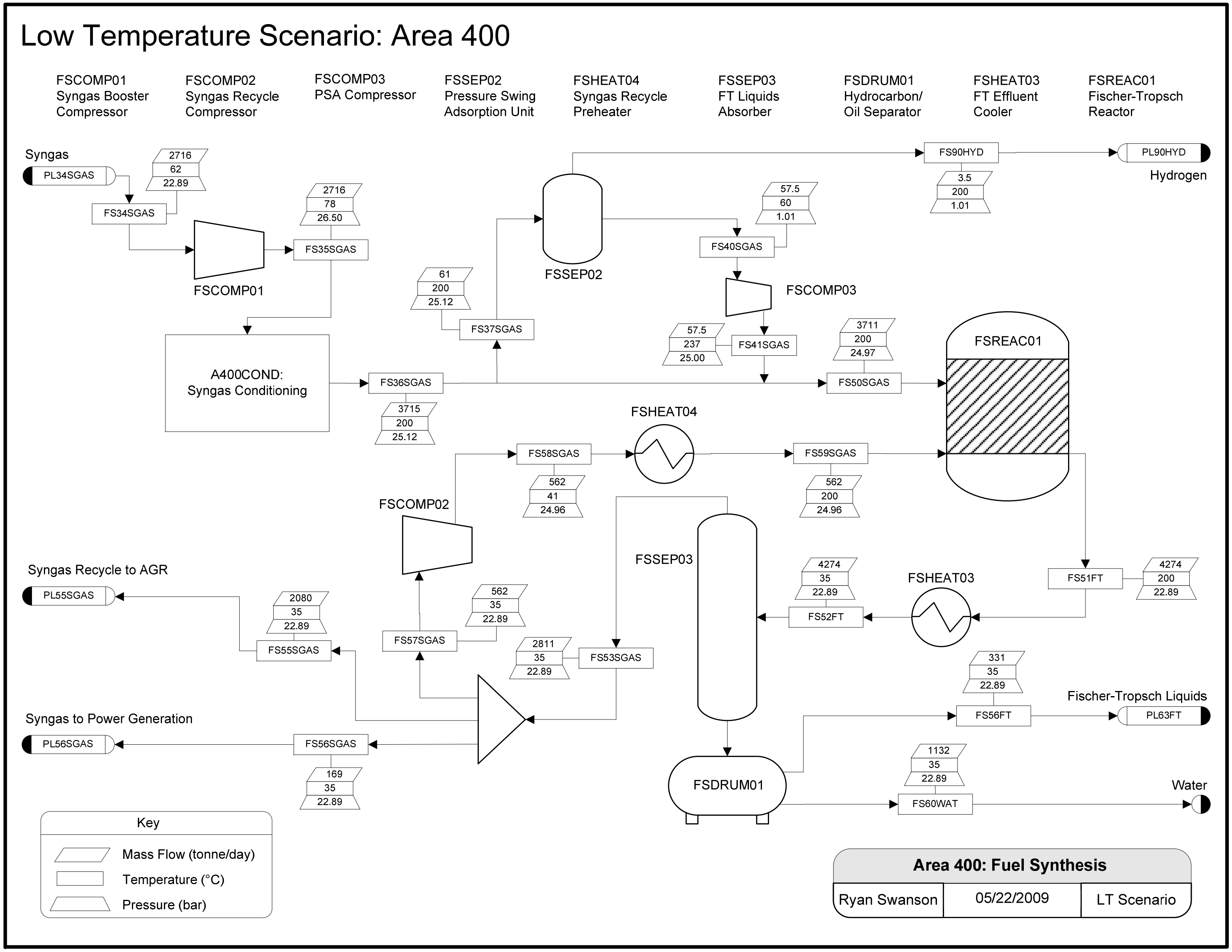


Figure D-17. Fuel synthesis area process flow diagram for LT scenario

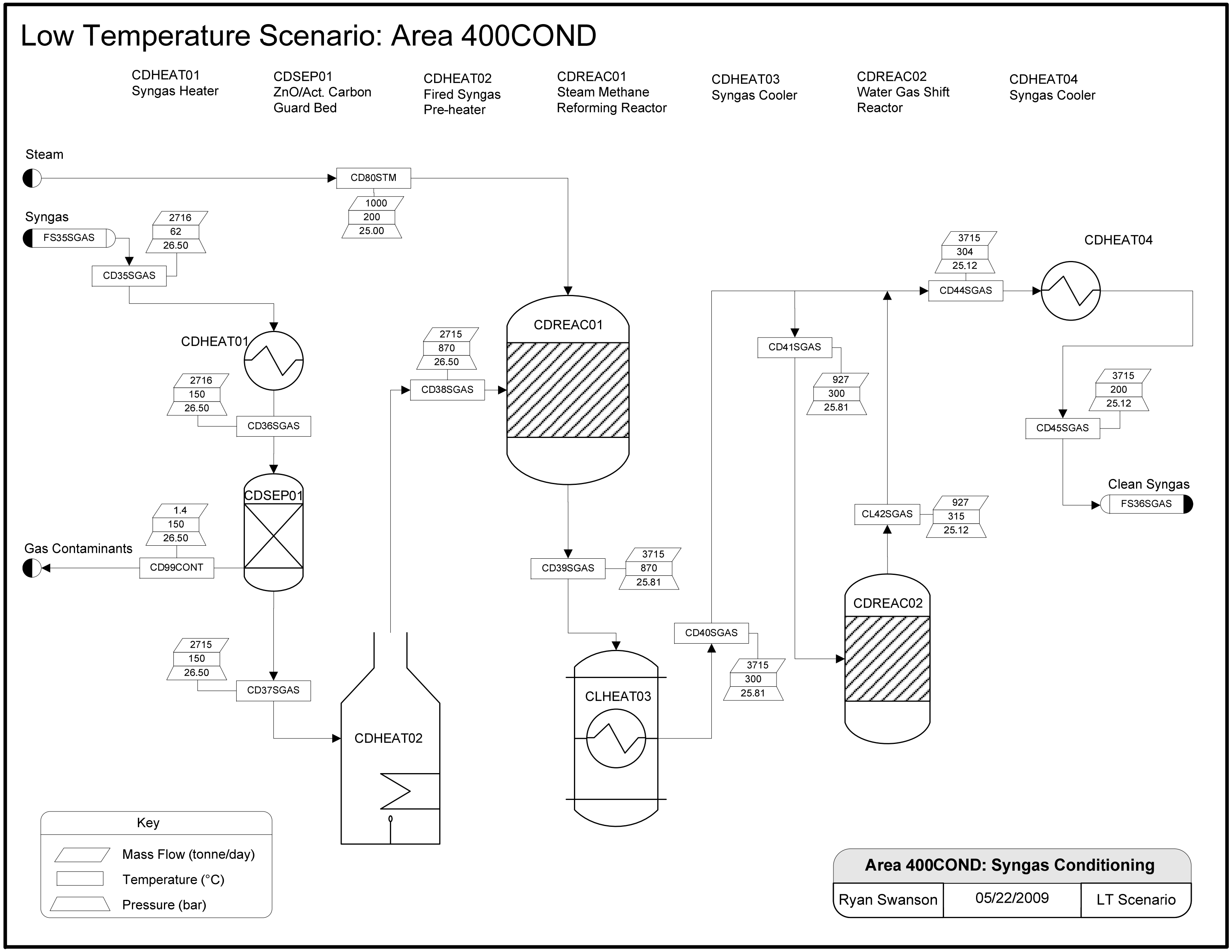


**Figure D-18. Syngas conditioning area process flow diagram for LT scenario**

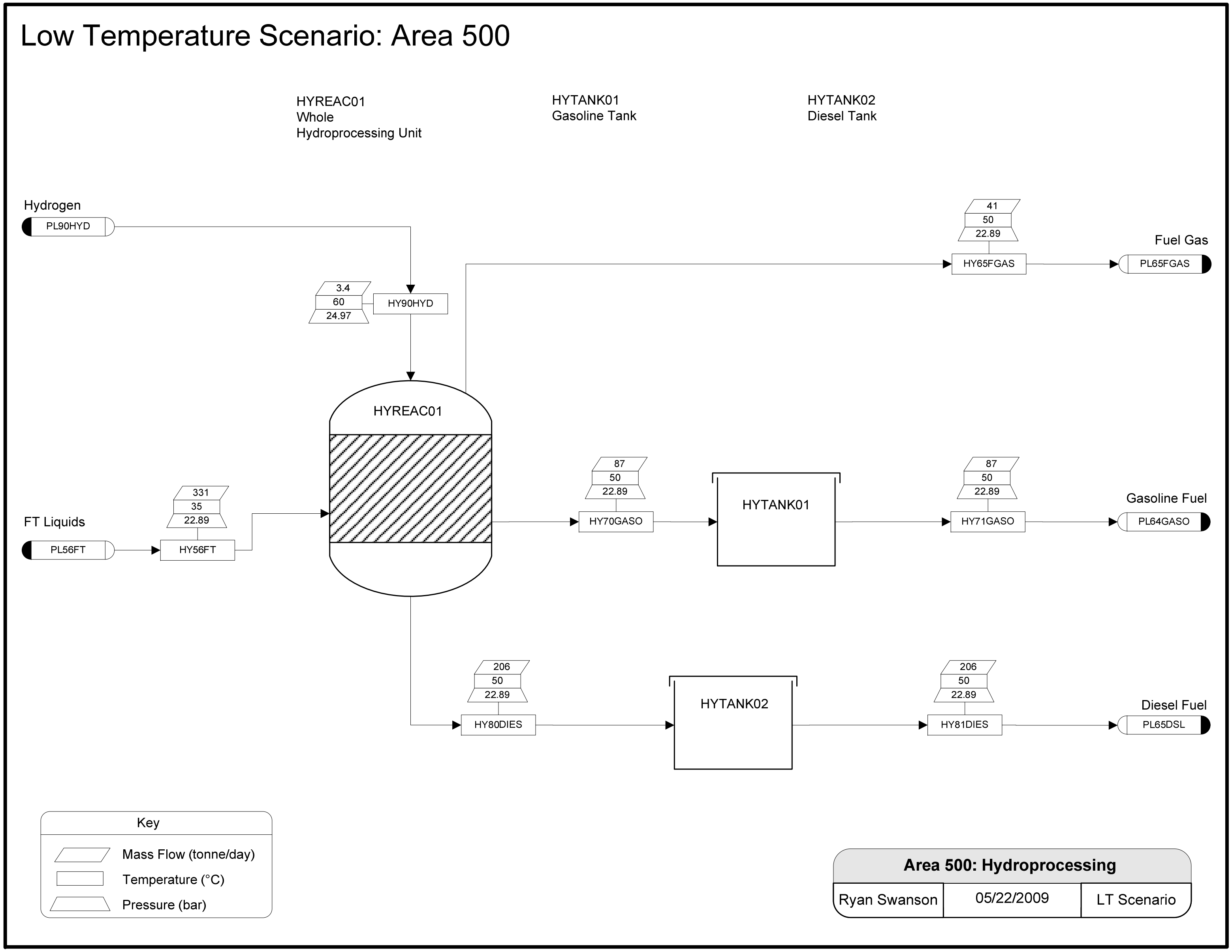


**Figure D-19. Hydroprocessing area process diagram for LT scenario**

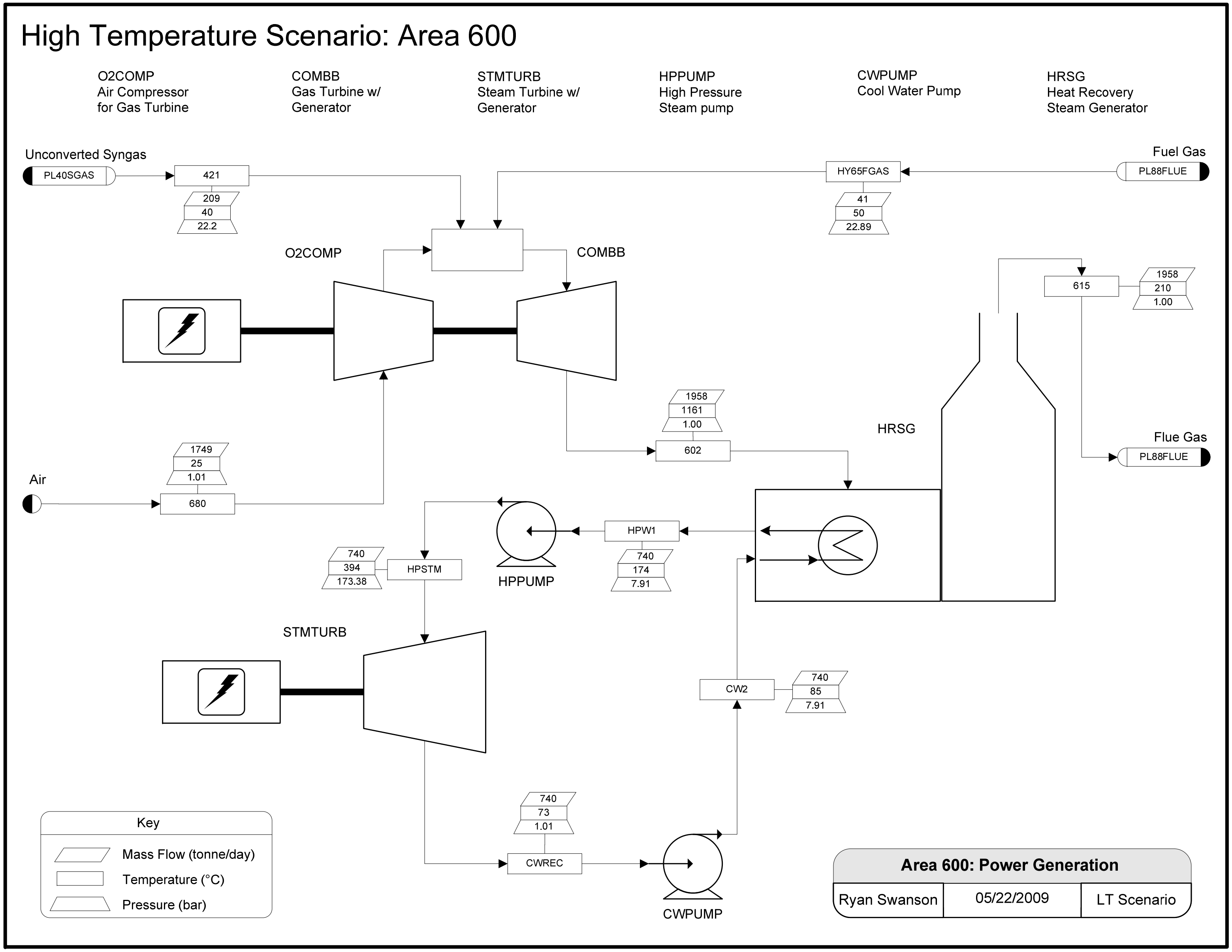


Figure D-20. Power generation area process flow diagram for LT scenario

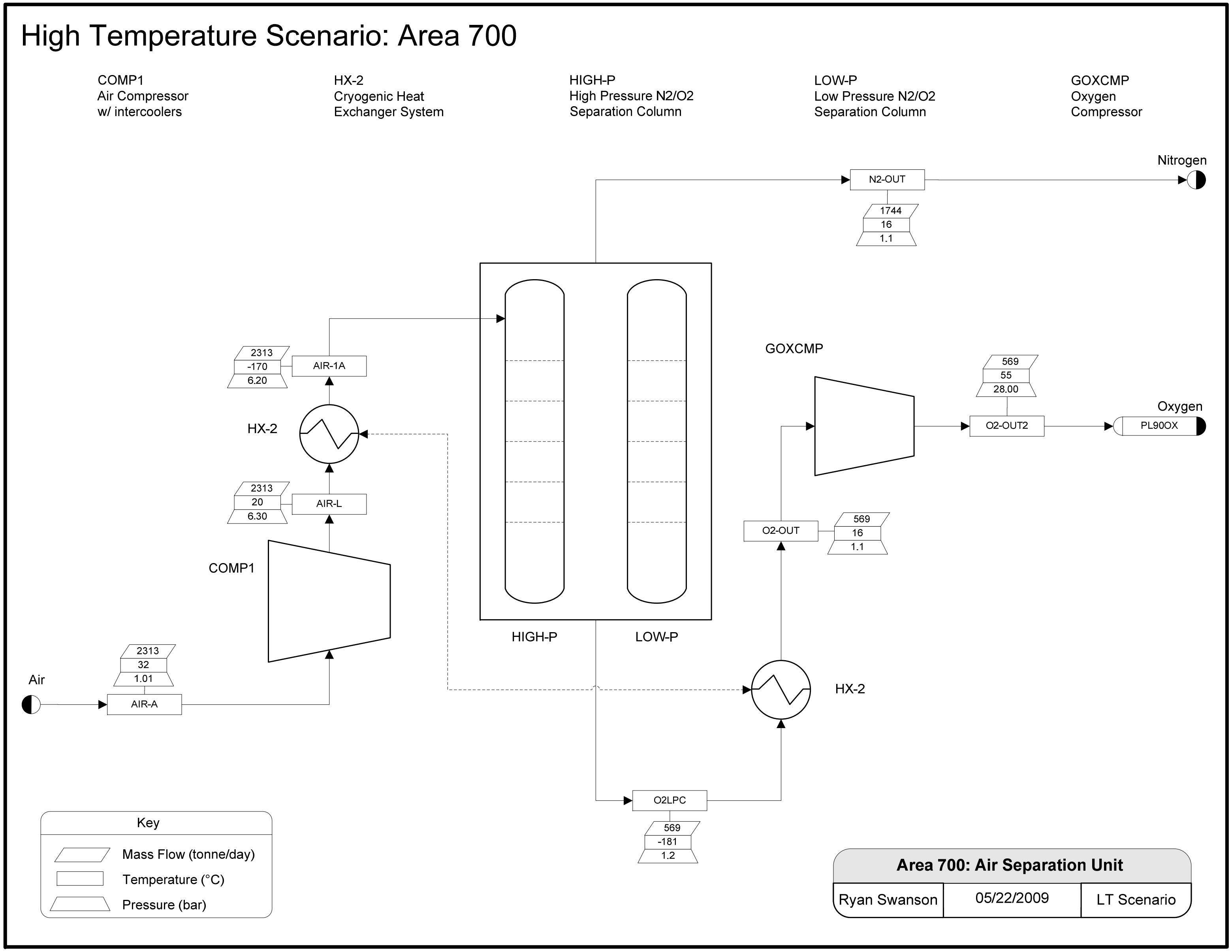


Figure D-21. Air separation unit process flow diagram for LT scenario

## Appendix E. Stream Data

# High-Temperature Scenario

## Table E-1. Overall Plant Stream Data for HT Scenario

| HT Overall Plant | PL00BMAS | PL08BMAS | PL09CAIR | PL17SLAG | PL21SGAS | PL34SGAS | PL42CO2 | PL43CO2 | PL47SGAS | PL48SGAS | PL49SGAS | PL50FT | PL52FLUE | PL65FGAS | PL71GASO | PL81DIES | PL81STM | PL81WAT | PL83SUL | PL84STM | PL88FLUE | PL89AIR | PL90HYD | PL90OX | PL90STM | PL92NTGN | PL92WAT | PL98STM | PL99AIR |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Temperature (C) | 25 | 90 | 25 | 50 | 1300 | 62 | 250 | 53 | 45 | 45 | 45 | 35 | 220 | 35 | 36 | 37 | 200 | 40 | 50 | 120 | 273 | 32 | 30 | 149 | 190 | -179 | 120 | 200 | 30 |
| Pressure (bar) | 1.01 | 1.01 | 1.01 | 26.62 | 26.62 | 22.75 | 28.00 | 3.45 | 23.58 | 23.58 | 23.58 | 22.20 | 1.03 | 22.20 | 1.03 | 1.03 | 1.98 | 24.82 | 1.93 | 1.98 | 1.00 | 1.01 | 1.00 | 28.00 | 10.00 | 1.20 | 1.98 | 28.00 | 1.01 |
| Vapor Fraction | 0.00 | 0.00 | 1.00 | 0.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 0.00 | 1.00 | 1.00 | 0.00 | 0.00 | 1.00 | 0.00 | 0.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 0.00 | 1.00 |
| Volume Flow** (m$^3$/sec) | 0.43 | 0.14 | 2.33 | 0 | 11.23 | 3.45 | 0.07 | 3.32 | 0.02 | 1.64 | 0.11 | 0.01 | 4.33 | 0.02 | 0 | 0.01 | 50.53 | 0.02 | 0 | 42.34 | 45.37 | 29.07 | 0.63 | 0.34 | 1.30 | 5.76 | 4.70 | 0.02 | 22.37 |
| Mole Flow** (kmol/hr) | 1542 | 513.97 | 343.36 | 0 | 8192 | 10089 | 175.05 | 1541 | 71.82 | 5260 | 355.46 | 89.49 | 393.42 | 74.55 | 41.08 | 48.97 | 9251 | 3276 | 13.46 | 9251 | 3595 | 4177 | 90.27 | 957.07 | 1271 | 3255 | 1028 | 2220 | 3237 |
| Mass Flow (tonnes/day) | 2667 | 2222 | 237.82 | 114.00 | 3825 | 3377 | 180.00 | 1585 | 29.08 | 2130 | 143.95 | 427.14 | 266.91 | 52.77 | 112.62 | 266.11 | 4000 | 1501 | 7.20 | 4000 | 2439 | 2903 | 4.37 | 743.70 | 549.62 | 2189 | 444.44 | 960.00 | 2242 |
| H2O | 666.67 | 222.22 | 0 | 0 | 988.43 | 45.53 | 2.50 | 21.97 | 0 | 0 | 0 | 0 | 27.89 | 0 | 0 | 0 | 0 | 1348 | 4.04 | 0 | 233.43 | 0 | 0 | 0 | 549.62 | 0 | 444.44 | 960.00 | 0 |
| CO | 0 | 0 | 0 | 0 | 1457 | 1818 | 0.47 | 4.15 | 11.24 | 823.54 | 55.65 | 0 | 0 | 0 | 0 | 0 | 0 | 22.10 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| H2 | 0 | 0 | 0 | 0 | 122.88 | 288.34 | 0 | 0 | 1.83 | 134.08 | 9.06 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 4.37 | 0 | 0 | 0 | 0 | 0 | 0 |
| CO2 | 0 | 0 | 0 | 0 | 1184 | 190.01 | 175.21 | 1543 | 2.40 | 175.74 | 11.88 | 0 | 40.00 | 0 | 0 | 0 | 0 | 128.26 | 0 | 0 | 352.24 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| O2 | 0 | 0 | 55.97 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 10.28 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 104.25 | 672.19 | 0 | 700.00 | 0 | 0 | 0 | 0 | 527.27 |
| N2 | 0 | 0 | 181.85 | 0 | 17.68 | 0 | 1.78 | 15.63 | 0 | 0 | 0 | 0 | 181.85 | 0 | 0 | 0 | 0 | 0.26 | 0 | 0 | 1715 | 2194 | 0 | 0 | 0 | 2189 | 0 | 0 | 1715 |
| CH4 | 0 | 0 | 0 | 0 | 0.02 | 63.41 | 0.01 | 0.07 | 0.87 | 63.47 | 4.29 | 0 | 0 | 14.94 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H6 | 0 | 0 | 0 | 0 | 0 | 106.90 | 0.01 | 0.10 | 1.46 | 107.01 | 7.23 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H4 | 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 |
| C2H2 | 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 |
| C3 | 0 | 0 | 0 | 0 | 0 | 141.45 | 0.01 | 0.10 | 1.93 | 141.56 | 9.57 | 0 | 0 | 37.83 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C4 | 0 | 0 | 0 | 0 | 0 | 167.38 | 0.02 | 0.15 | 2.29 | 167.55 | 11.32 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| H2S | 0 | 0 | 0 | 0 | 4.50 | 0.03 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1.02 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| NH3 | 0 | 0 | 0 | 0 | 0.11 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.10 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| TAR | 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 |
| SULFUR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 3.16 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CARBON | 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 |
| CHAR | 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 |
| STEAM | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 4000 | 0 | 0 | 4000 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| SO2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.02 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.08 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| NO2 | 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 |
| 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 | 0 | 0 |
| AIR | 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 |
| WAXES | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 170.22 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C5 | 0 | 0 | 0 | 0 | 0 | 2.29 | 0 | 0 | 0.03 | 2.29 | 0.15 | 15.49 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C6 | 0 | 0 | 0 | 0 | 0 | 2.46 | 0 | 0 | 0.03 | 2.46 | 0.17 | 16.65 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C7 | 0 | 0 | 0 | 0 | 0 | 1.15 | 0 | 0 | 0.02 | 1.15 | 0.08 | 17.54 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C8 | 0 | 0 | 0 | 0 | 0 | 1.18 | 0 | 0 | 0.02 | 1.18 | 0.08 | 17.99 | 0 | 0 | 112.62 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C9 | 0 | 0 | 0 | 0 | 0 | 1.11 | 0 | 0 | 0.02 | 1.19 | 0.08 | 18.10 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C10 | 0 | 0 | 0 | 0 | 0 | 1.16 | 0 | 0 | 0.02 | 1.19 | 0.08 | 18.12 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C11 | 0 | 0 | 0 | 0 | 0 | 0.83 | 0 | 0 | 0.02 | 1.16 | 0.08 | 17.62 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C12 | 0 | 0 | 0 | 0 | 0 | 0.13 | 0 | 0 | 0.01 | 0.54 | 0.04 | 17.25 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C13 | 0 | 0 | 0 | 0 | 0 | 0.02 | 0 | 0 | 0.01 | 0.52 | 0.04 | 16.70 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C14 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.01 | 0.50 | 0.03 | 16.16 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C15 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.01 | 0.49 | 0.03 | 15.57 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C16 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.01 | 0.47 | 0.03 | 14.93 | 0 | 0 | 0 | 266.11 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C17 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 14.75 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C18 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 14.05 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C19 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 13.35 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C20 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 12.64 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| COS | 0 | 0 | 0 | 0 | 0.30 | 1.20 | 0 | 0 | 0.02 | 1.12 | 0.08 | 0 | 0 | 0 | 0 | 0 | 0 | 0.13 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| AR | 0 | 0 | 0 | 0 | 43.70 | 544.37 | 0 | 0 | 6.87 | 502.93 | 33.99 | 0 | 6.87 | 0 | 0 | 0 | 0 | 1.03 | 0 | 0 | 33.99 | 37.28 | 0 | 43.70 | 0 | 0.12 | 0 | 0 | 0 |
| BIOMASS | 2000 | 2000 | 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 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| SOOT | 0 | 0 | 0 | 0 | 6.00 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| SLAG | 0 | 0 | 0 | 114.00 | 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 |

**Volumetric and Mole flow values do not include biomass, ash, soot, or slag

# Table E-2. Preprocessing Area Stream Data for HT Scenario

| HT A100 | CH00BMAS | CH02BMAS | CH03BMAS | CH90BMAS | DR05BMAS | DR81STM | DR84STM | DR92WAT | GR06BMAS | GR08BMAS | GR90BMAS |
|---|---|---|---|---|---|---|---|---|---|---|---|
| Temperature (C) | 25 | 25 | 25 | 25 | 90 | 200 | 120 | 120 | 90 | 90 | 90 |
| Pressure (bar) | 1.01 | 1.01 | 1.01 | 1.01 | 1.01 | 1.98 | 1.98 | 1.98 | 1.01 | 1.01 | 1.01 |
| Vapor Fraction | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 1.00 | 1.00 | 1.00 | 0.00 | 0.00 | 0.00 |
| Volume Flow** ($m^3$/sec) | 0.43 | 0.43 | 0.43 | 0 | 0.14 | 50.53 | 42.34 | 4.70 | 0.14 | 0.14 | 0 |
| Mole Flow** (kmol/hr) | 1542 | 1542 | 1542 | 0 | 513.97 | 9251 | 9251 | 1028 | 513.97 | 513.97 | 0 |
| Mass Flow (tonnes/day) | 2667 | 2709 | 2667 | 42.20 | 2222 | 4000 | 4000 | 444.44 | 2240 | 2222 | 17.60 |
| H2O | 666.67 | 666.67 | 666.67 | 0 | 222.22 | 0 | 0 | 444.44 | 222.22 | 222.22 | 0 |
| CO | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| H2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CO2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| O2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| N2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CH4 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H6 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H4 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C3 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C4 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| H2S | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| NH3 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| TAR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| SULFUR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CARBON | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CHAR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| STEAM | 0 | 0 | 0 | 0 | 0 | 4000 | 4000 | 0 | 0 | 0 | 0 |
| SO2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| NO2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| MEA | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| AIR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| WAXES | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C5 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C6 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C7 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C8 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C9 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C10 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C11 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C12 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C13 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C14 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C15 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C16 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C17 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C18 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C19 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C20 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| COS | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| AR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| BIOMASS | 2000 | 2042 | 2000 | 42.20 | 2000 | 0 | 0 | 0 | 2018 | 2000 | 17.60 |
| ASH | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| SOOT | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| SLAG | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |

**Volumetric and Mole flow values do not include biomass, ash, soot, or slag

## Table E-3. Gasification Area Stream Data for HT Scenario

| HT A200 | GS02SLAG | GS08BMAS | GS09CAIR | GS11BMAS | GS42CO2 | GS52FLUE | GS54SGAS | GS81STM | GS84STM | GS90OX | GS92CO2 | GS98STM | S06SGAS | SL01SLAG | SL02SGAS |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Temperature (C) | 50 | 90 | 25 | 90 | 250 | 220 | 45 | 200 | 120 | 149 | 92 | 200 | 203 | 1300 | 1300 |
| Pressure (bar) | 26.62 | 1.01 | 1.01 | 1.01 | 28.00 | 1.03 | 23.58 | 1.98 | 1.98 | 28.00 | 28.00 | 28.00 | 25.93 | 26.62 | 26.62 |
| Vapor Fraction | 0.00 | 0.00 | 1.00 | 0.53 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 0.00 | 1.00 | 0.00 | 1.00 |
| Volume Flow** ($m^3$/sec) | 0 | 0.14 | 2.33 | 30.59 | 0.07 | 4.33 | 0.02 | 50.53 | 42.34 | 0.34 | 0.04 | 0.02 | 3.50 | 0 | 11.23 |
| Mole Flow** (kmol/hr) | 0 | 513.97 | 343.36 | 6974 | 175.05 | 393.42 | 71.82 | 9251 | 9251 | 957.07 | 158.25 | 2220 | 8308 | 0 | 8192 |
| Mass Flow (tonnes/day) | 114.00 | 2222 | 237.82 | 2222 | 180.00 | 266.91 | 29.08 | 4000 | 4000 | 743.70 | 166.90 | 960.00 | 3869 | 114.00 | 3825 |
| H2O | 0 | 222.22 | 0 | 222.22 | 2.50 | 27.89 | 0 | 0 | 0 | 0 | 0 | 960.00 | 1038 | 0 | 988.43 |
| CO | 0 | 0 | 0 | 0 | 0.47 | 0 | 11.24 | 0 | 0 | 0 | 0.45 | 0 | 1457 | 0 | 1457 |
| H2 | 0 | 0 | 0 | 101.20 | 0 | 0 | 1.83 | 0 | 0 | 0 | 0 | 0 | 122.88 | 0 | 122.88 |
| CO2 | 0 | 0 | 0 | 0 | 175.21 | 40.00 | 2.40 | 0 | 0 | 0 | 166.45 | 0 | 1184 | 0 | 1184 |
| O2 | 0 | 0 | 55.97 | 812.60 | 0 | 10.28 | 0 | 0 | 0 | 700.00 | 0 | 0 | 0 | 0 | 0 |
| N2 | 0 | 0 | 181.85 | 16.00 | 1.78 | 181.85 | 0 | 0 | 0 | 0 | 0 | 0 | 17.68 | 0 | 17.68 |
| CH4 | 0 | 0 | 0 | 0 | 0.01 | 0 | 0.87 | 0 | 0 | 0 | 0 | 0 | 0.02 | 0 | 0.02 |
| C2H6 | 0 | 0 | 0 | 0 | 0.01 | 0 | 1.46 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H4 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C3 | 0 | 0 | 0 | 0 | 0.01 | 0 | 1.93 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C4 | 0 | 0 | 0 | 0 | 0.02 | 0 | 2.29 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| H2S | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 4.50 | 0 | 4.50 |
| NH3 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.11 | 0 | 0.11 |
| TAR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| SULFUR | 0 | 0 | 0 | 4.40 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CARBON | 0 | 0 | 0 | 945.60 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CHAR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| STEAM | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 4000 | 4000 | 0 | 0 | 0 | 0 | 0 | 0 |
| SO2 | 0 | 0 | 0 | 0 | 0 | 0.02 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| NO2 | 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 |
| AIR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| WAXES | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C5 | 0 | 0 | 0 | 0 | 0 | 0 | 0.03 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C6 | 0 | 0 | 0 | 0 | 0 | 0 | 0.03 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C7 | 0 | 0 | 0 | 0 | 0 | 0 | 0.02 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C8 | 0 | 0 | 0 | 0 | 0 | 0 | 0.02 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C9 | 0 | 0 | 0 | 0 | 0 | 0 | 0.02 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C10 | 0 | 0 | 0 | 0 | 0 | 0 | 0.02 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C11 | 0 | 0 | 0 | 0 | 0 | 0 | 0.02 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C12 | 0 | 0 | 0 | 0 | 0 | 0 | 0.01 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C13 | 0 | 0 | 0 | 0 | 0 | 0 | 0.01 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C14 | 0 | 0 | 0 | 0 | 0 | 0 | 0.01 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C15 | 0 | 0 | 0 | 0 | 0 | 0 | 0.01 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C16 | 0 | 0 | 0 | 0 | 0 | 0 | 0.01 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C17 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C18 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C19 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C20 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| COS | 0 | 0 | 0 | 0 | 0 | 0 | 0.02 | 0 | 0 | 0 | 0 | 0 | 0.30 | 0 | 0.30 |
| AR | 0 | 0 | 0 | 0 | 0 | 6.87 | 6.87 | 0 | 0 | 43.70 | 0 | 0 | 43.70 | 0 | 43.70 |
| BIOMASS | 0 | 2000 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| ASH | 0 | 0 | 0 | 120.00 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| SOOT | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 6.00 |
| SLAG | 114.00 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 114.00 | 0 |

**Volumetric and Mole flow values do not include biomass, ash, soot, or slag

# Table E-4. Syngas Cleaning Area Stream Data for HT Scenario

| HT A300 | CL25ACG | CL25SGAS | CL26SGAS | CL28SGAS | CL29SGAS | CL32SGAS | CL40CO2 | CL41CO2 | CL42CO2 | CL49SGAS | CL81WAT | CL82WAT | CL83SUL | CL83WAT | CL90STM | CL92AIR | CL95AIR | GS84STM | SG91STM | SL01H2O | SL02SGAS | SL03H2O | SL80H2O |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Temperature (C) | 50 | 203 | 291 | 240 | 60 | 40 | 53 | 53 | 250 | 45 | 30 | 40 | 50 | 60 | 190 | 25 | 50 | 62 | 250 | 203 | 1300 | 203 | 30 |
| Pressure (bar) | 3.45 | 25.93 | 24.82 | 24.82 | 24.82 | 24.82 | 3.45 | 3.45 | 28.00 | 23.58 | 24.82 | 24.82 | 1.93 | 24.82 | 10.00 | 1.01 | 1.93 | 22.75 | 25.86 | 26.62 | 26.62 | 25.93 | 26.62 |
| Vapor Fraction | 1.00 | 1.00 | 1.00 | 1.00 | 0.69 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 0.00 | 0.00 | 0.00 | 0.00 | 1.00 | 1.00 | 0.93 | 1.00 | 1.00 | 0.00 | 1.00 | 0.00 | 0.00 |
| Volume Flow** ($m^3$/sec) | 3.69 | 1.08 | 2.01 | 4.54 | 2.07 | 1.86 | 3.70 | 0.38 | 0.07 | 1.64 | 0 | 0.02 | 0 | 0.02 | 1.30 | 0.01 | 0.01 | 3.45 | 0.54 | 0.07 | 11.23 | 0 | 0 |
| Mole Flow** (kmol/hr) | 1728 | 2576 | 3847 | 9579 | 9579 | 6419 | 1716 | 175.05 | 175.05 | 5260 | 115.64 | 3276 | 13.46 | 2313 | 1271 | 2.07 | 2.58 | 10089 | 1271 | 9251 | 8192 | 0 | 115.64 |
| Mass Flow (tonnes/day) | 1773 | 1199 | 1749 | 4418 | 4418 | 2967 | 1765 | 180.00 | 180.00 | 2130 | 50.00 | 1501 | 7.20 | 1000 | 549.62 | 1.59 | 2.32 | 3377 | 549.62 | 4000 | 3825 | 6.00 | 50.00 |
| H2O | 27.18 | 321.91 | 591.14 | 1308 | 1308 | 9.19 | 24.46 | 2.50 | 2.50 | 0 | 50.00 | 1348 | 4.04 | 1000 | 549.62 | 0 | 0.45 | 45.53 | 549.62 | 4000 | 988.43 | 0 | 50.00 |
| CO | 4.62 | 451.69 | 15.74 | 1021 | 1021 | 999.01 | 4.62 | 0.47 | 0.47 | 823.54 | 0 | 22.10 | 0 | 0 | 0 | 0 | 0 | 1818 | 0 | 0 | 1457 | 0 | 0 |
| H2 | 0 | 38.09 | 69.47 | 154.26 | 154.26 | 154.26 | 0 | 0 | 0 | 134.08 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 288.34 | 0 | 0 | 122.88 | 0 | 0 |
| CO2 | 1718 | 367.08 | 1052 | 1869 | 1869 | 1741 | 1718 | 175.21 | 175.21 | 175.74 | 0 | 128.26 | 0 | 0 | 0 | 0 | 0 | 190.01 | 0 | 0 | 1184 | 0 | 0 |
| O2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1.59 | 0.02 | 0 | 0 | 0 | 0 | 0 | 0 |
| N2 | 17.41 | 5.48 | 5.48 | 17.68 | 17.68 | 17.43 | 17.41 | 1.78 | 1.78 | 0 | 0 | 0.26 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 17.68 | 0 | 0 |
| CH4 | 0.08 | 0.01 | 0.01 | 0.02 | 0.02 | 0.02 | 0.08 | 0.01 | 0.01 | 63.47 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 63.41 | 0 | 0 | 0.02 | 0 | 0 |
| C2H6 | 0.11 | 0 | 0 | 0 | 0 | 0 | 0.11 | 0.01 | 0.01 | 107.01 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 106.90 | 0 | 0 | 0 | 0 | 0 |
| C2H4 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C3 | 0.11 | 0 | 0 | 0 | 0 | 0 | 0.11 | 0.01 | 0.01 | 141.56 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 141.45 | 0 | 0 | 0 | 0 | 0 |
| C4 | 0.17 | 0 | 0 | 0 | 0 | 0 | 0.17 | 0.02 | 0.02 | 167.55 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 167.38 | 0 | 0 | 0 | 0 | 0 |
| H2S | 3.39 | 1.40 | 1.40 | 4.50 | 4.50 | 3.48 | 0 | 0 | 0 | 0 | 0 | 1.02 | 0 | 0 | 0 | 0 | 0.03 | 0.03 | 0 | 0 | 4.50 | 0 | 0 |
| NH3 | 0 | 0.03 | 0.03 | 0.11 | 0.11 | 0.01 | 0 | 0 | 0 | 0 | 0 | 0.10 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.11 | 0 | 0 |
| TAR | 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 | 0 | 0 | 0 | 3.16 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CARBON | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CHAR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| STEAM | 0 | 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 | 0 |
| NO2 | 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 |
| AIR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| WAXES | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C5 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 2.29 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 2.29 | 0 | 0 | 0 | 0 | 0 |
| C6 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 2.46 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 2.46 | 0 | 0 | 0 | 0 | 0 |
| C7 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1.15 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1.15 | 0 | 0 | 0 | 0 | 0 |
| C8 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1.18 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1.18 | 0 | 0 | 0 | 0 | 0 |
| C9 | 0.08 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1.19 | 0 | 0 | 0 | 0 | 0 | 0 | 0.08 | 1.11 | 0 | 0 | 0 | 0 | 0 |
| C10 | 0.03 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1.19 | 0 | 0 | 0 | 0 | 0 | 0 | 0.03 | 1.16 | 0 | 0 | 0 | 0 | 0 |
| C11 | 0.23 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1.16 | 0 | 0 | 0 | 0 | 0 | 0 | 0.23 | 0.83 | 0 | 0 | 0 | 0 | 0 |
| C12 | 0.13 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.54 | 0 | 0 | 0 | 0 | 0 | 0 | 0.13 | 0.13 | 0 | 0 | 0 | 0 | 0 |
| C13 | 0.04 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.52 | 0 | 0 | 0 | 0 | 0 | 0 | 0.04 | 0.02 | 0 | 0 | 0 | 0 | 0 |
| C14 | 0.01 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.50 | 0 | 0 | 0 | 0 | 0 | 0 | 0.01 | 0 | 0 | 0 | 0 | 0 | 0 |
| C15 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.49 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C16 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.47 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C17 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C18 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C19 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C20 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| COS | 0.08 | 0.09 | 0.09 | 0.30 | 0.30 | 0.17 | 0 | 0 | 0 | 1.12 | 0 | 0.13 | 0 | 0 | 0 | 0 | 0.08 | 1.20 | 0 | 0 | 0.30 | 0 | 0 |
| AR | 1.22 | 13.55 | 13.55 | 43.70 | 43.70 | 42.66 | 0 | 0 | 0 | 502.93 | 0 | 1.03 | 0 | 0 | 0 | 0 | 1.22 | 544.37 | 0 | 0 | 43.70 | 0 | 0 |
| BIOMASS | 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 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| SOOT | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 6.00 | 6.00 | 0 |
| SLAG | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |

**Volumetric and Mole flow values do not include biomass, ash, soot, or slag

# Table E-5. Acid Gas Removal and Sulfur Recovery Areas Stream Data for HT Scenario

| HT A300AGR | AG02MEAR | AG03MEAR | AG04MEAR | AG06MEAR | AG08MEAL | AG09MEAL | AG10MEAL | AG11MEAL | AG23ACG | AG24ACG | AG25ACG | AG31SGAS | AG32SGAS | AG33SGAS | AG49SGAS | A300SUL | SU25ACG | SU26SUL | SU40CO2 | SU83SUL | SU92AIR | SU93AIR | SU94AIR | SU95AIR |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Temperature (C) | 58 | 62 | 58 | 86 | 50 | 123 | 96 | 96 | 86 | 50 | 50 | 40 | 58 | 62 | 45 | | 50 | 53 | 53 | 50 | 25 | 100 | 50 | 50 |
| Pressure (bar) | 22.75 | 22.75 | 22.75 | 3.45 | 3.45 | 3.45 | 3.45 | 20.68 | 3.45 | 3.45 | 3.45 | 24.82 | 22.75 | 22.75 | 23.58 | | 3.45 | 3.45 | 3.45 | 1.93 | 1.01 | 2.07 | 1.93 | 1.93 |
| Vapor Fraction | 0.00 | 0.00 | 0.00 | 0.01 | 0.00 | 0.00 | 0.00 | 0.00 | 0.98 | 0.83 | 1.00 | 1.00 | 0.93 | 1.00 | 1.00 | | 1.00 | 0.49 | 1.00 | 0.00 | 1.00 | 1.00 | 1.00 | 0.93 |
| Volume Flow** ($m^3$/sec) | 0.480 | 0 | 0.480 | 1.590 | 0 | 0.480 | 0.470 | 0.470 | 4.820 | 3.690 | 3.690 | 1.860 | 3.410 | 3.450 | 1.640 | | 3.690 | 0.010 | 3.700 | 0 | 0.010 | 0.010 | 0.010 | 0.010 |
| Mole Flow** (kmol/hr) | 84877 | 699.000 | 85576 | 85576 | 342.110 | 83848 | 83848 | 83848 | 2070 | 2070 | 1728 | 6419 | 10788 | 10089 | 5260 | | 1728 | 11.910 | 1716 | 13.460 | 2.070 | 2.070 | 2.070 | 2.580 |
| Mass Flow (tonnes/day) | 42560 | 308.340 | 42868 | 42868 | 155.170 | 41096 | 41096 | 41096 | 1928 | 1928 | 1773 | 2967 | 3685 | 3377 | 2130 | | 1773 | 7.930 | 1765 | 7.200 | 1.590 | 1.590 | 1.590 | 2.320 |
| H2O | 33956 | 297.460 | 34254 | 34254 | 144.090 | 34226 | 34226 | 34226 | 171.270 | 171.270 | 27.180 | 9.190 | 342.990 | 45.530 | 0 | | 27.180 | 2.720 | 24.460 | 4.040 | 0 | 0 | 0 | 0.450 |
| CO | 0 | 4.640 | 4.640 | 4.640 | 0.010 | 0.010 | 0.010 | 0.010 | 4.630 | 4.630 | 4.620 | 999.010 | 1823 | 1818 | 823.540 | | 4.620 | 0 | 4.620 | 0 | 0 | 0 | 0 | 0 |
| H2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 154.260 | 288.340 | 288.340 | 134.080 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CO2 | 1725 | 1.650 | 1727 | 1727 | 8.690 | 8.690 | 8.690 | 8.690 | 1727 | 1727 | 1718 | 1741 | 191.660 | 190.010 | 175.740 | | 1718 | 0 | 1718 | 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 | 1.590 | 1.590 | 1.590 | 0.020 |
| N2 | 17.430 | 0 | 17.430 | 17.430 | 0.020 | 0.020 | 0.020 | 0.020 | 17.430 | 17.430 | 17.410 | 17.430 | 0 | 0 | 0 | | 17.410 | 0 | 17.410 | 0 | 0 | 0 | 0 | 0 |
| CH4 | 0 | 0.080 | 0.080 | 0.080 | 0 | 0 | 0 | 0 | 0.080 | 0.080 | 0.080 | 0.020 | 63.490 | 63.410 | 63.470 | | 0.080 | 0 | 0.080 | 0 | 0 | 0 | 0 | 0 |
| C2H6 | 0 | 0.110 | 0.110 | 0.110 | 0 | 0 | 0 | 0 | 0.110 | 0.110 | 0.110 | 0 | 107.010 | 106.900 | 107.010 | | 0.110 | 0 | 0.110 | 0 | 0 | 0 | 0 | 0 |
| C2H4 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C3 | 0 | 0.110 | 0.110 | 0.110 | 0 | 0 | 0 | 0 | 0.110 | 0.110 | 0.110 | 0 | 141.560 | 141.450 | 141.560 | | 0.110 | 0 | 0.110 | 0 | 0 | 0 | 0 | 0 |
| C4 | 0 | 0.170 | 0.170 | 0.170 | 0 | 0 | 0 | 0 | 0.170 | 0.170 | 0.170 | 0 | 167.550 | 167.380 | 167.550 | | 0.170 | 0 | 0.170 | 0 | 0 | 0 | 0 | 0 |
| H2S | 3.450 | 0 | 3.450 | 3.450 | 0.060 | 0.060 | 0.060 | 0.060 | 3.450 | 3.450 | 3.390 | 3.480 | 0.030 | 0.030 | 0 | | 3.390 | 3.390 | 0 | 0 | 0 | 0 | 0 | 0.030 |
| NH3 | 0.010 | 0 | 0.010 | 0.010 | 0.010 | 0.010 | 0.010 | 0.010 | 0.010 | 0.010 | 0 | 0.010 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| TAR | 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 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 3.160 | 0 | 0 | 0 | 0 |
| CARBON | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CHAR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| STEAM | 0 | 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 | 0 |
| NO2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| MEA | 6858 | 0 | 6858 | 6858 | 0 | 6858 | 6858 | 6858 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| AIR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| WAXES | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C5 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 2.290 | 2.290 | 2.290 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C6 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 2.460 | 2.460 | 2.460 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C7 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1.150 | 1.150 | 1.150 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C8 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1.180 | 1.180 | 1.180 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C9 | 0 | 0.080 | 0.080 | 0.080 | 0.010 | 0.010 | 0.010 | 0.010 | 0.080 | 0.080 | 0.080 | 0 | 1.190 | 1.110 | 1.190 | | 0.080 | 0.080 | 0 | 0 | 0 | 0 | 0 | 0.080 |
| C10 | 0 | 0.030 | 0.030 | 0.030 | 0 | 0 | 0 | 0 | 0.030 | 0.030 | 0.030 | 0 | 1.190 | 1.160 | 1.190 | | 0.030 | 0.030 | 0 | 0 | 0 | 0 | 0 | 0.030 |
| C11 | 0 | 0.330 | 0.330 | 0.330 | 0.100 | 0.100 | 0.100 | 0.100 | 0.320 | 0.320 | 0.230 | 0 | 1.160 | 0.830 | 1.160 | | 0.230 | 0.230 | 0 | 0 | 0 | 0 | 0 | 0.230 |
| C12 | 0 | 0.410 | 0.410 | 0.410 | 0.280 | 0.280 | 0.280 | 0.280 | 0.410 | 0.410 | 0.130 | 0 | 0.540 | 0.130 | 0.540 | | 0.130 | 0.130 | 0 | 0 | 0 | 0 | 0 | 0.130 |
| C13 | 0 | 0.500 | 0.500 | 0.500 | 0.460 | 0.460 | 0.460 | 0.460 | 0.500 | 0.500 | 0.040 | 0 | 0.520 | 0.020 | 0.520 | | 0.040 | 0.040 | 0 | 0 | 0 | 0 | 0 | 0.040 |
| C14 | 0 | 0.500 | 0.500 | 0.500 | 0.490 | 0.490 | 0.490 | 0.490 | 0.500 | 0.500 | 0.010 | 0 | 0.500 | 0 | 0.500 | | 0.010 | 0.010 | 0 | 0 | 0 | 0 | 0 | 0.010 |
| C15 | 0 | 0.490 | 0.490 | 0.490 | 0.490 | 0.490 | 0.490 | 0.490 | 0.490 | 0.490 | 0 | 0 | 0.490 | 0 | 0.490 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C16 | 0 | 0.470 | 0.470 | 0.470 | 0.470 | 0.470 | 0.470 | 0.470 | 0.470 | 0.470 | 0 | 0 | 0.470 | 0 | 0.470 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C17 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C18 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C19 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C20 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| COS | 0 | 0.080 | 0.080 | 0.080 | 0 | 0 | 0 | 0 | 0.080 | 0.080 | 0.080 | 0.170 | 1.290 | 1.200 | 1.120 | | 0.080 | 0.080 | 0 | 0 | 0 | 0 | 0 | 0.080 |
| AR | 0 | 1.220 | 1.220 | 1.220 | 0 | 0 | 0 | 0 | 1.220 | 1.220 | 1.220 | 42.660 | 545.590 | 544.370 | 502.930 | | 1.220 | 1.220 | 0 | 0 | 0 | 0 | 0 | 1.220 |
| BIOMASS | 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 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| SOOT | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| SLAG | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |

**Volumetric and Mole flow values do not include biomass, ash, soot, or slag

## Table E-6. Fuel Synthesis Area Stream Data for HT Scenario

| HT A400 | FS34SGAS | FS35SGAS | FS36SGAS | FS37SGAS | FS40SGAS | FS42SGAS | FS43SGAS | FS44FT | FS44SGAS | FS45FT | FS45SGAS | FS46FT | FS46SGAS | FS47SGAS | FS48SGAS | FS49SGAS | FS50FT | FS50SGAS | FS51SGAS | FS52SGAS | FS60WAT | FS89HYD | FS90HYD | FS99CONT |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Temperature (C) | 62 | 76 | 200 | 200 | 30 | 30 | 417 | 200 | 202 | 35 | 202 | 43 | 0 | 45 | 45 | 45 | 35 | 45 | 50 | 200 | 35 | 30 | 30 | 200 |
| Pressure (bar) | 22.75 | 26.00 | 26.00 | 24.97 | 1.00 | 1.00 | 24.97 | 23.58 | 24.97 | 23.58 | 24.96 | 23.58 | 0.00 | 23.58 | 23.58 | 23.58 | 22.20 | 23.58 | 24.96 | 24.96 | 22.20 | 1.00 | 1.00 | 24.97 |
| Vapor Fraction | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 0.86 | 1.00 | 0.00 | 1.00 | 1.00 | 1.00 | 1.00 | 0.00 | 1.00 | 1.00 | 1.00 | 0.00 | 1.00 | 1.00 | 1.00 |
| Volume Flow** ($m^3$/sec) | 3.45 | 3.16 | 4.29 | 4.47 | 1.26 | 0.63 | 0.06 | 4.84 | 4.45 | 2.71 | 5.86 | 0.02 | 2.74 | 0.02 | 1.64 | 0.11 | 0.01 | 1.00 | 0.96 | 1.42 | 0.01 | 0.63 | 0.63 | 0 |
| Mole Flow** (kmol/hr) | 10089 | 10089 | 10089 | 10089 | 179.78 | 89.51 | 89.51 | 10438 | 9998 | 10438 | 13198 | 1575 | 8887 | 71.82 | 5260 | 355.46 | 89.49 | 3199 | 3199 | 3199 | 1483 | 90.27 | 90.27 | 0.04 |
| Mass Flow (tonnes/day) | 3377 | 3377 | 3377 | 3377 | 60.18 | 55.81 | 55.81 | 4668 | 3372 | 4668 | 4668 | 1069 | 3599 | 29.08 | 2130 | 143.95 | 427.14 | 1296 | 1296 | 1296 | 641.34 | 4.37 | 4.37 | 0.03 |
| H2O | 45.53 | 45.53 | 45.53 | 45.53 | 0.81 | 0.81 | 0.81 | 642.07 | 45.53 | 642.07 | 45.53 | 642.07 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 641.34 | 0 | 0 | 0 |
| CO | 1818 | 1818 | 1818 | 1818 | 32.40 | 32.40 | 32.40 | 1391 | 1818 | 1391 | 2319 | 0 | 1391 | 11.24 | 823.54 | 55.65 | 0 | 500.87 | 500.87 | 500.87 | 0 | 0 | 0 | 0 |
| H2 | 288.34 | 288.34 | 288.34 | 288.34 | 5.14 | 0.77 | 0.77 | 225.34 | 283.97 | 225.34 | 365.52 | 0 | 226.52 | 1.83 | 134.08 | 9.06 | 0 | 81.55 | 81.55 | 81.55 | 0 | 4.37 | 4.37 | 0 |
| CO2 | 190.01 | 190.01 | 190.01 | 190.01 | 3.39 | 3.39 | 3.39 | 296.89 | 190.01 | 296.89 | 296.89 | 0 | 296.90 | 2.40 | 175.74 | 11.88 | 0 | 106.88 | 106.88 | 106.88 | 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 | 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 | 0 | 0 |
| CH4 | 63.41 | 63.41 | 63.41 | 63.41 | 1.13 | 1.13 | 1.13 | 107.33 | 63.41 | 107.33 | 102.02 | 0 | 107.23 | 0.87 | 63.47 | 4.29 | 0 | 38.60 | 38.60 | 38.60 | 0 | 0 | 0 | 0 |
| C2H6 | 106.90 | 106.90 | 106.90 | 106.90 | 1.91 | 1.91 | 1.91 | 180.95 | 106.90 | 180.95 | 171.99 | 0 | 180.79 | 1.46 | 107.01 | 7.23 | 0 | 65.08 | 65.08 | 65.08 | 0 | 0 | 0 | 0 |
| C2H4 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C3 | 141.45 | 141.45 | 141.45 | 141.45 | 2.52 | 2.52 | 2.52 | 239.37 | 141.45 | 239.37 | 227.54 | 0 | 239.16 | 1.93 | 141.56 | 9.57 | 0 | 86.10 | 86.10 | 86.10 | 0 | 0 | 0 | 0 |
| C4 | 167.38 | 167.38 | 167.38 | 167.38 | 2.98 | 2.98 | 2.98 | 283.31 | 167.38 | 283.31 | 269.28 | 0 | 283.06 | 2.29 | 167.55 | 11.32 | 0 | 101.90 | 101.90 | 101.90 | 0 | 0 | 0 | 0 |
| H2S | 0.03 | 0.03 | 0.03 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.00 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.03 |
| NH3 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.00 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| TAR | 0 | 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 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CARBON | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CHAR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| STEAM | 0 | 0 | 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 | 0 | 0 |
| NO2 | 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 |
| AIR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| WAXES | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 170.22 | 0 | 170.22 | 0 | 170.22 | 0 | 0 | 0 | 0 | 170.22 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C5 | 2.29 | 2.29 | 2.29 | 2.29 | 0.04 | 0.04 | 0.04 | 19.36 | 2.29 | 19.36 | 3.69 | 15.49 | 3.87 | 0.03 | 2.29 | 0.15 | 15.49 | 1.39 | 1.39 | 1.39 | 0 | 0 | 0 | 0 |
| C6 | 2.46 | 2.46 | 2.46 | 2.46 | 0.04 | 0.04 | 0.04 | 20.81 | 2.46 | 20.81 | 3.96 | 16.65 | 4.16 | 0.03 | 2.46 | 0.17 | 16.65 | 1.50 | 1.50 | 1.50 | 0 | 0 | 0 | 0 |
| C7 | 1.15 | 1.15 | 1.15 | 1.15 | 0.02 | 0.02 | 0.02 | 19.49 | 1.15 | 19.49 | 1.85 | 17.54 | 1.95 | 0.02 | 1.15 | 0.08 | 17.54 | 0.70 | 0.70 | 0.70 | 0 | 0 | 0 | 0 |
| C8 | 1.18 | 1.18 | 1.18 | 1.18 | 0.02 | 0.02 | 0.02 | 19.99 | 1.18 | 19.99 | 1.90 | 17.99 | 2.00 | 0.02 | 1.18 | 0.08 | 17.99 | 0.72 | 0.72 | 0.72 | 0 | 0 | 0 | 0 |
| C9 | 1.11 | 1.11 | 1.11 | 1.11 | 0.02 | 0.02 | 0.02 | 20.11 | 1.11 | 20.11 | 1.83 | 18.10 | 2.01 | 0.02 | 1.19 | 0.08 | 18.10 | 0.72 | 0.72 | 0.72 | 0 | 0 | 0 | 0 |
| C10 | 1.16 | 1.16 | 1.16 | 1.16 | 0.02 | 0.02 | 0.02 | 20.13 | 1.16 | 20.13 | 1.88 | 18.12 | 2.01 | 0.02 | 1.19 | 0.08 | 18.12 | 0.72 | 0.72 | 0.72 | 0 | 0 | 0 | 0 |
| C11 | 0.83 | 0.83 | 0.83 | 0.83 | 0.01 | 0.01 | 0.01 | 19.58 | 0.83 | 19.58 | 1.54 | 17.62 | 1.96 | 0.02 | 1.16 | 0.08 | 17.62 | 0.71 | 0.71 | 0.71 | 0 | 0 | 0 | 0 |
| C12 | 0.13 | 0.13 | 0.13 | 0.13 | 0 | 0 | 0 | 18.16 | 0.13 | 18.16 | 0.45 | 17.25 | 0.91 | 0.01 | 0.54 | 0.04 | 17.25 | 0.33 | 0.33 | 0.33 | 0 | 0 | 0 | 0 |
| C13 | 0.02 | 0.02 | 0.02 | 0.02 | 0 | 0 | 0 | 17.58 | 0.02 | 17.58 | 0.33 | 16.70 | 0.88 | 0.01 | 0.52 | 0.04 | 16.70 | 0.32 | 0.32 | 0.32 | 0 | 0 | 0 | 0 |
| C14 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 17.01 | 0 | 17.01 | 0.31 | 16.16 | 0.85 | 0.01 | 0.50 | 0.03 | 16.16 | 0.31 | 0.31 | 0.31 | 0 | 0 | 0 | 0 |
| C15 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 16.39 | 0 | 16.39 | 0.30 | 15.57 | 0.82 | 0.01 | 0.49 | 0.03 | 15.57 | 0.29 | 0.29 | 0.29 | 0 | 0 | 0 | 0 |
| C16 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 15.72 | 0 | 15.72 | 0.28 | 14.93 | 0.79 | 0.01 | 0.47 | 0.03 | 14.93 | 0.28 | 0.28 | 0.28 | 0 | 0 | 0 | 0 |
| C17 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 14.75 | 0 | 14.75 | 0 | 14.75 | 0 | 0 | 0 | 0 | 14.75 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C18 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 14.05 | 0 | 14.05 | 0 | 14.05 | 0 | 0 | 0 | 0 | 14.05 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C19 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 13.35 | 0 | 13.35 | 0 | 13.35 | 0 | 0 | 0 | 0 | 13.35 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C20 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 12.64 | 0 | 12.64 | 0 | 12.64 | 0 | 0 | 0 | 0 | 12.64 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| COS | 1.20 | 1.20 | 1.20 | 1.20 | 0.02 | 0.02 | 0.02 | 1.88 | 1.20 | 1.88 | 1.88 | 0 | 1.88 | 0.02 | 1.12 | 0.08 | 0 | 0.68 | 0.68 | 0.68 | 0 | 0 | 0 | 0 |
| AR | 544.37 | 544.37 | 544.37 | 544.37 | 9.70 | 9.70 | 9.70 | 850.25 | 544.37 | 850.25 | 850.25 | 0 | 849.66 | 6.87 | 502.93 | 33.99 | 0 | 305.88 | 305.88 | 305.88 | 0 | 0 | 0 | 0 |
| BIOMASS | 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 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| SOOT | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| SLAG | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |

**Volumetric and Mole flow values do not include biomass, ash, soot, or slag

## Table E-7. Hydroprocessing, Power Generation, and Air Separation Areas Stream Data for HT Scenario

| HT A500 | HY50FT | HY65FGAS | HY70GASO | HY71GASO | HY80DIES | HY81DIES | HY90HYD | HT A600 | 421 | 602 | 615 | 680 | CW2 | CWREC | HPSTM1 | HPSTM2 | HT A700 | AIR-1A | AIR-A | AIR-L | N2-OUT | O2LPC | O2-OUT | O2-OUT2 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Temperature (C) | 35 | 35 | 35 | 36 | 35 | 37 | 30 | | 45 | 1144 | 273 | 30 | 85 | 73 | 170 | 565 | | -170 | 32 | 20 | 16 | -177 | 16 | 68 |
| Pressure (bar) | 22.20 | 22.20 | 22.20 | 1.03 | 22.20 | 1.03 | 1.00 | | 23.58 | 1.00 | 1.00 | 1.01 | 7.91 | 0.30 | 7.91 | 173.38 | | 6.20 | 1.01 | 6.30 | 1.10 | 1.88 | 1.10 | 29.97 |
| Vapor Fraction | 0.00 | 1.00 | 0.00 | 0.00 | 0.00 | 0.00 | 1.00 | | 1.00 | 1.00 | 1.00 | 1.00 | 0.00 | 0.97 | 0.00 | 0.00 | | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 0.98 | 1.00 |
| Volume Flow** ($m^3$/sec) | 0.01 | 0.02 | 0 | 0 | 0.01 | 0.01 | 0.63 | | 0.11 | 117.68 | 45.37 | 22.37 | 0.01 | 49.13 | 0.01 | 0.01 | | 1.33 | 29.07 | 4.49 | 19.73 | 1.04 | 5.58 | 0.24 |
| Mole Flow** (kmol/hr) | 89.49 | 74.55 | 41.08 | 41.08 | 48.97 | 48.97 | 90.27 | | 355.46 | 3595 | 3595 | 3237 | 1939 | 1939 | 1939 | 1939 | | 3968 | 4177 | 4177 | 3255 | 921.35 | 921.35 | 921.35 |
| Mass Flow (tonnes/day) | 427.14 | 52.77 | 112.62 | 112.62 | 266.11 | 266.11 | 4.37 | | 143.95 | 2439 | 2439 | 2242 | 838.24 | 838.23 | 838.23 | 838.23 | | 2758 | 2903 | 2903 | 2189 | 714.26 | 714.26 | 714.26 |
| H2O | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 233.43 | 233.43 | 0 | 838.24 | 838.23 | 838.23 | 838.23 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CO | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 55.65 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| H2 | 0 | 0 | 0 | 0 | 0 | 0 | 4.37 | | 9.06 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CO2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 11.88 | 352.24 | 352.24 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| O2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 104.25 | 104.25 | 527.27 | 0 | 0 | 0 | 0 | | 638.58 | 672.19 | 672.19 | 0 | 672.19 | 672.19 | 672.19 |
| N2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 1715 | 1715 | 1715 | 0 | 0 | 0 | 0 | | 2084 | 2194 | 2194 | 2189 | 4.91 | 4.91 | 4.91 |
| CH4 | 0 | 14.94 | 0 | 0 | 0 | 0 | 0 | | 4.29 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H6 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 7.23 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H4 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C3 | 0 | 37.83 | 0 | 0 | 0 | 0 | 0 | | 9.57 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C4 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 11.32 | 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 |
| NH3 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| TAR | 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 |
| CARBON | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CHAR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| STEAM | 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 | 0.08 | 0.08 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| NO2 | 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 |
| AIR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| WAXES | 170.22 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C5 | 15.49 | 0 | 0 | 0 | 0 | 0 | 0 | | 0.15 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C6 | 16.65 | 0 | 0 | 0 | 0 | 0 | 0 | | 0.17 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C7 | 17.54 | 0 | 0 | 0 | 0 | 0 | 0 | | 0.08 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C8 | 17.99 | 0 | 112.62 | 112.62 | 0 | 0 | 0 | | 0.08 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C9 | 18.10 | 0 | 0 | 0 | 0 | 0 | 0 | | 0.08 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C10 | 18.12 | 0 | 0 | 0 | 0 | 0 | 0 | | 0.08 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C11 | 17.62 | 0 | 0 | 0 | 0 | 0 | 0 | | 0.08 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C12 | 17.25 | 0 | 0 | 0 | 0 | 0 | 0 | | 0.04 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C13 | 16.70 | 0 | 0 | 0 | 0 | 0 | 0 | | 0.04 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C14 | 16.16 | 0 | 0 | 0 | 0 | 0 | 0 | | 0.03 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C15 | 15.57 | 0 | 0 | 0 | 0 | 0 | 0 | | 0.03 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C16 | 14.93 | 0 | 0 | 0 | 266.11 | 266.11 | 0 | | 0.03 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C17 | 14.75 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C18 | 14.05 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C19 | 13.35 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C20 | 12.64 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| COS | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0.08 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| AR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 33.99 | 33.99 | 33.99 | 0 | 0 | 0 | 0 | 0 | | 35.42 | 37.28 | 37.28 | 0.12 | 37.16 | 37.16 | 37.16 |
| BIOMASS | 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 |
| SOOT | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| SLAG | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 | 0 |

**Volumetric and Mole flow values do not include biomass, ash, soot, or slag

## Low-Temperature Scenario

### Table E-8. Overall Plant Stream Data for LT Scenario

| HT Overall Plant | PL00BMAS | PL08BMAS | PL09CAIR | PL21SGAS | PL34SGAS | PL41CO2 | PL43CO2 | PL52FLUE | PL55SGAS | PL56FT | PL56SGAS | PL60ASH | PL65FGAS | PL71GASO | PL81DIES | PL81STM | PL81WAT | PL83SUL | PL84STM | PL88FLUE | PL90HYD | PL90OX | PL90STM | PL92WAT | PL98STM | PL89AIR | PL92NTGN | PL99AIR |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Temperature (C) | 25 | 90 | 100 | 870 | 62 | 50 | 243 | 200 | 32 | 35 | 32 | 0 | 50 | 50 | 50 | 200 | 40 | 50 | 120 | 344 | 60 | 149 | 300 | 120 | 204 | 32 | 16 | 30 |
| Pressure (bar) | 1.01 | 1.01 | 1.01 | 27.55 | 22.89 | 3.45 | 28.00 | 1.00 | 22.89 | 22.89 | 22.89 | 1.00 | 22.89 | 22.89 | 22.89 | 1.98 | 26.86 | 2.07 | 1.98 | 1.00 | 1.01 | 22.00 | 25.00 | 1.98 | 22.00 | 1.01 | 1.10 | 1.01 |
| Vapor Fraction | 0.00 | 0.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 0.00 | 1.00 | 0.00 | 1.00 | 0.00 | 0.00 | 1.00 | 0.00 | 0.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 0.00 | 1.00 | 1.00 | 1.00 |
| Volume Flow** ($m^3$/sec) | 0.43 | 0.14 | 16.89 | 4.76 | 2.40 | 2.84 | 0.07 | 22.48 | 1.50 | 0.01 | 0.12 | 0 | 0.02 | 0 | 0 | 50.53 | 0.02 | 0 | 42.34 | 40.84 | 0.53 | 0.32 | 1.15 | 4.70 | 0.01 | 23.17 | 15.73 | 17.42 |
| Mole Flow** (kmol/hr) | 1542 | 513.97 | 1985 | 4939 | 7066 | 1334 | 170.42 | 2053 | 4869 | 69.40 | 394.75 | 0 | 57.67 | 31.78 | 37.88 | 9251 | 2804 | 5.56 | 9251 | 2863 | 69.96 | 722.81 | 2313 | 1028 | 814.33 | 3328 | 2594 | 2522 |
| Mass Flow (tonnes/day) | 2667 | 2222 | 1375 | 2930 | 2706 | 1380 | 180.00 | 1471 | 2071 | 330.42 | 167.90 | 118.88 | 40.83 | 87.12 | 205.86 | 4000 | 1388 | 3.10 | 4000 | 1955 | 3.38 | 561.66 | 1000 | 444.44 | 352.09 | 2313 | 1744 | 1746 |
| H2O | 666.67 | 222.22 | 0 | 413.42 | 30.18 | 20.50 | 0 | 27.16 | 0 | 0.09 | 0 | 0 | 0 | 0 | 0 | 0 | 1060 | 2.40 | 0 | 189.23 | 0 | 0 | 1000 | 444.44 | 352.09 | 0 | 0 | 0 |
| CO | 0 | 0 | 0 | 797.86 | 1575 | 0 | 0 | 0 | 795.10 | 0 | 64.47 | 0 | 0 | 0 | 0 | 0 | 15.27 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| H2 | 0 | 0 | 0 | 47.75 | 168.38 | 0 | 0 | 0 | 120.71 | 0 | 9.79 | 0 | 0 | 0 | 0 | 0 | 0.07 | 0 | 0 | 0 | 3.38 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CO2 | 0 | 0 | 0 | 1427 | 167.80 | 1359 | 180.00 | 239.52 | 559.11 | 0 | 45.33 | 0 | 0 | 0 | 0 | 0 | 263.09 | 0 | 0 | 318.35 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| O2 | 0 | 0 | 320.19 | 0 | 0 | 0 | 0 | 148.99 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 80.99 | 0 | 528.66 | 0 | 0 | 0 | 535.65 | 0 | 410.78 |
| N2 | 0 | 0 | 1055 | 0 | 0 | 0 | 0 | 1055 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1336 | 0 | 0 | 0 | 0 | 0 | 1748 | 1744 | 1336 |
| AR | 0 | 0 | 0 | 33.00 | 411.50 | 0 | 0 | 0 | 380.61 | 0 | 30.86 | 0 | 0 | 0 | 0 | 0 | 0.69 | 0 | 0 | 30.86 | 0 | 33.00 | 0 | 0 | 0 | 29.71 | 0.10 | 0 |
| CH4 | 0 | 0 | 0 | 103.81 | 149.06 | 0 | 0 | 0 | 52.01 | 0 | 4.22 | 0 | 11.56 | 0 | 0 | 0 | 4.58 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H6 | 0 | 0 | 0 | 21.82 | 28.49 | 0 | 0 | 0 | 13.55 | 0 | 1.10 | 0 | 0 | 0 | 0 | 0 | 5.75 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H4 | 0 | 0 | 0 | 44.96 | 36.05 | 0 | 0 | 0 | 0.36 | 0 | 0.03 | 0 | 0 | 0 | 0 | 0 | 8.11 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C6H6 | 0 | 0 | 0 | 4.10 | 39.50 | 0 | 0 | 0 | 36.54 | 0 | 2.96 | 0 | 0 | 0 | 0 | 0 | 0.98 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C3 | 0 | 0 | 0 | 0 | 54.18 | 0 | 0 | 0 | 58.62 | 0 | 4.75 | 0 | 29.27 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C4 | 0 | 0 | 0 | 0 | 34.31 | 0 | 0 | 0 | 41.73 | 0 | 3.38 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| H2S | 0 | 0 | 0 | 4.51 | 0.13 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1.87 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| NH3 | 0 | 0 | 0 | 19.09 | 1.52 | 0 | 0 | 0 | 0.28 | 0 | 0.02 | 0 | 0 | 0 | 0 | 0 | 17.36 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| TAR | 0 | 0 | 0 | 10.57 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 10.57 | 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 | 0 | 0 | 0 | 0 | 0 | 0 |
| CARBON | 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 |
| STEAM | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 4000 | 0 | 0 | 4000 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| SO2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.30 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| NO2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.06 | 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 | 0 |
| AIR | 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 |
| WAXES | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 131.46 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C5 | 0 | 0 | 0 | 0 | 2.21 | 0 | 0 | 0 | 2.21 | 11.92 | 0.18 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C6 | 0 | 0 | 0 | 0 | 2.37 | 0 | 0 | 0 | 2.37 | 12.82 | 0.19 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C7 | 0 | 0 | 0 | 0 | 1.11 | 0 | 0 | 0 | 1.11 | 13.53 | 0.09 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C8 | 0 | 0 | 0 | 0 | 1.14 | 0 | 0 | 0 | 1.14 | 13.88 | 0.09 | 0 | 0 | 87.12 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C9 | 0 | 0 | 0 | 0 | 1.08 | 0 | 0 | 0 | 1.15 | 13.95 | 0.09 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C10 | 0 | 0 | 0 | 0 | 1.13 | 0 | 0 | 0 | 1.15 | 13.98 | 0.09 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C11 | 0 | 0 | 0 | 0 | 0.50 | 0 | 0 | 0 | 0.54 | 13.86 | 0.04 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C12 | 0 | 0 | 0 | 0 | 0.44 | 0 | 0 | 0 | 0.53 | 13.54 | 0.04 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C13 | 0 | 0 | 0 | 0 | 0.34 | 0 | 0 | 0 | 0.51 | 13.10 | 0.04 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C14 | 0 | 0 | 0 | 0 | 0.20 | 0 | 0 | 0 | 0.49 | 12.57 | 0.04 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C15 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.46 | 11.93 | 0.04 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C16 | 0 | 0 | 0 | 0 | 0.04 | 0 | 0 | 0 | 0.45 | 11.48 | 0.04 | 0 | 0 | 0 | 205.86 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C17 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 11.39 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C18 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 10.85 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C19 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 10.31 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C20 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 9.76 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| BIOMASS | 2000 | 2000 | 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 | 1.01 | 0 | 0 | 0 | 0.01 | 0 | 0 | 0 | 118.88 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CHAR | 0 | 0 | 0 | 0.81 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |

**Volumetric and Mole flow values do not include biomass, ash, or char

## Table E-9. Preprocessing Area Stream Data for LT Scenario

| HT A100 | CH00BMAS | CH02BMAS | CH03BMAS | CH90BMAS | DR05BMAS | .DR81STM | .DR84STM | .DR92WAT | GR06BMAS | GR08BMAS | GR90BMAS |
|---|---|---|---|---|---|---|---|---|---|---|---|
| Temperature (C) | 25 | 25 | 25 | 0 | 90 | 200 | 120 | 120 | 90 | 90 | 0 |
| Pressure (bar) | 1.01 | 1.01 | 1.01 | 0.00 | 1.01 | 1.98 | 1.98 | 1.98 | 1.01 | 1.01 | 0.00 |
| Vapor Fraction | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 1.00 | 1.00 | 1.00 | 0.00 | 0.00 | 0.00 |
| Volume Flow** ($m^3$/sec) | 0.43 | 0.43 | 0.43 | 0 | 0.14 | 50.53 | 42.34 | 4.70 | 0.14 | 0.14 | 0 |
| Mole Flow** (kmol/hr) | 1542 | 1542 | 1542 | 0 | 513.97 | 9251 | 9251 | 1028 | 513.97 | 513.97 | 0 |
| Mass Flow (tonnes/day) | 2667 | 3181 | 2667 | 514.02 | 2222 | 4000 | 4000 | 444.44 | 3019 | 2222 | 796.81 |
| H2O | 666.67 | 666.67 | 666.67 | 0 | 222.22 | 0 | 0 | 444.44 | 222.22 | 222.22 | 0 |
| CO | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| H2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CO2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| O2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| N2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| AR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CH4 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H6 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H4 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C6H6 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C3 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C4 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| H2S | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| NH3 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| TAR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| SULFUR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CARBON | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| STEAM | 0 | 0 | 0 | 0 | 0 | 4000 | 4000 | 0 | 0 | 0 | 0 |
| SO2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| NO2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| MEA | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| AIR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| WAXES | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C5 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C6 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C7 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C8 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C9 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C10 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C11 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C12 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C13 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C14 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C15 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C16 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C17 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C18 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C19 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C20 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| BIOMASS | 2000 | 2514 | 2000 | 514.02 | 2000 | 0 | 0 | 0 | 2797 | 2000 | 796.81 |
| ASH | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CHAR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |

**Volumetric and Mole flow values do not include biomass, ash, or char

## Table E-10. Gasification Area Stream Data for LT Scenario

| HT A200 | CB52FLUE | CB61ASH | CB62ASH | CY14SGAS | CY15CHAR | CY16CHAR | GS05SGAS | GS07SGAS | GS09CAIR | GS11BMAS | GS19CHAR | GS20SGAS | GS43CO2 | GS46CO2 | GS52FLUE | GS81STM | GS84STM | GS90OX | GS91STM |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Temperature (C) | 1200 | 0 | 0 | 870 | 0 | 0 | 871 | 870 | 100 | 96 | 0 | 870 | 243 | 100 | 200 | 200 | 120 | 149 | 204 |
| Pressure (bar) | 1.00 | 1.00 | 1.00 | 27.57 | 27.57 | 27.55 | 28.00 | 27.58 | 1.01 | 1.01 | 27.55 | 27.55 | 28.00 | 27.58 | 1.00 | 1.98 | 1.98 | 22.00 | 22.00 |
| Vapor Fraction | 1.00 | 0.00 | 0.00 | 1.00 | 0.00 | 0.00 | 1.00 | 1.00 | 1.00 | 0.00 | 0.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 0.00 |
| Volume Flow** ($m^3$/sec) | 69.99 | 0 | 0 | 2.38 | 0 | 0 | 4.69 | 4.76 | 16.89 | 0 | 0 | 4.76 | 0.07 | 0.05 | 22.48 | 50.53 | 42.34 | 0.32 | 0.01 |
| Mole Flow** (kmol/hr) | 2053 | 0 | 0 | 2470 | 0 | 0 | 4931 | 4939 | 1985 | 513.97 | 0 | 4939 | 170.42 | 161.90 | 2053 | 9251 | 9251 | 722.81 | 814.33 |
| Mass Flow (tonnes/day) | 1471 | 17.82 | 118.88 | 1477 | 95.38 | 12.09 | 3136 | 3145 | 1375 | 2222 | 214.95 | 2930 | 180.00 | 171.00 | 1471 | 4000 | 4000 | 561.66 | 352.09 |
| H2O | 27.16 | 0 | 0 | 206.71 | 0 | 0 | 413.42 | 413.42 | 0 | 222.22 | 0 | 413.42 | 0 | 0 | 27.16 | 0 | 0 | 0 | 352.09 |
| CO | 0 | 0 | 0 | 398.93 | 0 | 0 | 797.86 | 797.86 | 0 | 0 | 0 | 797.86 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| H2 | 0 | 0 | 0 | 23.87 | 0 | 0 | 47.75 | 47.75 | 0 | 0 | 0 | 47.75 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CO2 | 239.52 | 0 | 0 | 713.66 | 0 | 0 | 1418 | 1427 | 0 | 0 | 0 | 1427 | 180.00 | 171.00 | 239.52 | 0 | 0 | 0 | 0 |
| O2 | 148.99 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 320.19 | 0 | 0 | 0 | 0 | 0 | 148.99 | 0 | 0 | 528.66 | 0 |
| N2 | 1055 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1055 | 0 | 0 | 0 | 0 | 0 | 1055 | 0 | 0 | 0 | 0 |
| AR | 0 | 0 | 0 | 16.50 | 0 | 0 | 33.00 | 33.00 | 0 | 0 | 0 | 33.00 | 0 | 0 | 0 | 0 | 0 | 33.00 | 0 |
| CH4 | 0 | 0 | 0 | 51.90 | 0 | 0 | 103.81 | 103.81 | 0 | 0 | 0 | 103.81 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H6 | 0 | 0 | 0 | 10.91 | 0 | 0 | 21.82 | 21.82 | 0 | 0 | 0 | 21.82 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H4 | 0 | 0 | 0 | 22.48 | 0 | 0 | 44.96 | 44.96 | 0 | 0 | 0 | 44.96 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C6H6 | 0 | 0 | 0 | 2.05 | 0 | 0 | 4.10 | 4.10 | 0 | 0 | 0 | 4.10 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C3 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C4 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| H2S | 0 | 0 | 0 | 2.25 | 0 | 0 | 4.51 | 4.51 | 0 | 0 | 0 | 4.51 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| NH3 | 0 | 0 | 0 | 9.54 | 0 | 0 | 19.09 | 19.09 | 0 | 0 | 0 | 19.09 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| TAR | 0 | 0 | 0 | 5.29 | 0 | 0 | 10.57 | 10.57 | 0 | 0 | 0 | 10.57 | 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 |
| CARBON | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| STEAM | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 4000 | 4000 | 0 | 0 |
| SO2 | 0.30 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.30 | 0 | 0 | 0 | 0 |
| NO2 | 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 |
| AIR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| WAXES | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C5 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C6 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C7 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C8 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C9 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C10 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C11 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C12 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C13 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C14 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C15 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C16 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C17 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C18 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C19 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C20 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| BIOMASS | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 2000 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| ASH | 0.01 | 17.82 | 118.88 | 7.19 | 52.75 | 6.69 | 119.90 | 119.90 | 0 | 0 | 118.89 | 1.01 | 0 | 0 | 0.01 | 0 | 0 | 0 | 0 |
| CHAR | 0 | 0 | 0 | 5.81 | 42.63 | 5.40 | 96.87 | 96.87 | 0 | 0 | 96.06 | 0.81 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |

**Volumetric and Mole flow values do not include biomass, ash, or char

## Table E-11. Syngas Cleaning Area Stream Data for LT Scenario

| HT A300 | CL01H2O | CL02H2O | CL21SGAS | CL25ACG | CL25SGAS | CL25WAT | CL29SGAS | CL40CO2 | CL40SGAS | CL41CO2 | CL43CO2 | CL49SGAS | CL80H2O | CL80WAT | CL81WAT | CL83SUL | CL83WAT | CL84WAT | CL92AIR | CL95AIR |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Temperature (C) | 194 | 150 | 870 | 50 | 194 | 0 | 40 | 50 | 62 | 50 | 243 | 32 | 30 | 30 | 40 | 50 | 98 | 40 | 50 | 50 |
| Pressure (bar) | 27.55 | 27.55 | 27.55 | 3.45 | 27.55 | 27.55 | 26.86 | 3.45 | 22.89 | 3.45 | 28.00 | 22.89 | 27.57 | 27.57 | 26.86 | 2.07 | 27.55 | 27.55 | 2.07 | 2.07 |
| Vapor Fraction | 0.00 | 0.00 | 1.00 | 1.00 | 1.00 | 0.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 1.00 | 1.00 |
| Volume Flow** ($m^3$/sec) | 0.08 | 0.08 | 4.76 | 3.28 | 2.03 | 0 | 0.98 | 3.21 | 2.40 | 2.84 | 0.07 | 1.50 | 0 | 0.01 | 0.02 | 0 | 0.07 | 0.07 | 0 | 0.10 |
| Mole Flow** (kmol/hr) | 13877 | 13877 | 4939 | 1538 | 5286 | 0 | 3638 | 1505 | 7066 | 1334 | 170.42 | 4869 | 346.93 | 1156 | 2804 | 5.56 | 13877 | 13877 | 0.45 | 27.74 |
| Mass Flow (tonnes/day) | 6000 | 6000 | 2930 | 1586 | 3078 | 1.82 | 2190 | 1560 | 2706 | 1380 | 180.00 | 2071 | 150.00 | 500.00 | 1388 | 3.10 | 6000 | 6000 | 0.35 | 22.89 |
| H2O | 6000 | 6000 | 413.42 | 22.78 | 563.42 | 0 | 3.78 | 20.50 | 30.18 | 20.50 | 0 | 0 | 150.00 | 500.00 | 1060 | 2.40 | 6000 | 6000 | 0 | 0.27 |
| CO | 0 | 0 | 797.86 | 3.07 | 797.86 | 0 | 782.59 | 0 | 1575 | 0 | 0 | 795.10 | 0 | 0 | 15.27 | 0 | 0 | 0 | 0 | 3.07 |
| H2 | 0 | 0 | 47.75 | 0 | 47.75 | 0 | 47.68 | 0 | 168.38 | 0 | 0 | 120.71 | 0 | 0 | 0.07 | 0 | 0 | 0 | 0 | 0 |
| CO2 | 0 | 0 | 1427 | 1539 | 1427 | 0 | 1164 | 1539 | 167.80 | 1359 | 180.00 | 559.11 | 0 | 0 | 263.09 | 0 | 0 | 0 | 0 | 0 |
| O2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.35 | 0 |
| N2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| AR | 0 | 0 | 33.00 | 1.42 | 33.00 | 0 | 32.31 | 0 | 411.50 | 0 | 0 | 380.61 | 0 | 0 | 0.69 | 0 | 0 | 0 | 0 | 1.42 |
| CH4 | 0 | 0 | 103.81 | 2.18 | 103.81 | 0 | 99.23 | 0 | 149.06 | 0 | 0 | 52.01 | 0 | 0 | 4.58 | 0 | 0 | 0 | 0 | 2.18 |
| C2H6 | 0 | 0 | 21.82 | 1.11 | 21.82 | 0 | 16.07 | 0 | 28.49 | 0 | 0 | 13.55 | 0 | 0 | 5.75 | 0 | 0 | 0 | 0 | 1.11 |
| C2H4 | 0 | 0 | 44.96 | 1.14 | 44.96 | 0 | 36.85 | 0 | 36.05 | 0 | 0 | 0.36 | 0 | 0 | 8.11 | 0 | 0 | 0 | 0 | 1.14 |
| C6H6 | 0 | 0 | 4.10 | 0.15 | 4.10 | 0 | 3.11 | 0 | 39.50 | 0 | 0 | 36.54 | 0 | 0 | 0.98 | 0 | 0 | 0 | 0 | 0.15 |
| C3 | 0 | 0 | 0 | 4.29 | 0 | 0 | 0 | 0 | 54.18 | 0 | 0 | 58.62 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 4.29 |
| C4 | 0 | 0 | 0 | 6.73 | 0 | 0 | 0 | 0 | 34.31 | 0 | 0 | 41.73 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 6.73 |
| H2S | 0 | 0 | 4.51 | 2.48 | 4.51 | 0 | 2.64 | 0 | 0.13 | 0 | 0 | 0 | 0 | 0 | 1.87 | 0 | 0 | 0 | 0 | 1.73 |
| NH3 | 0 | 0 | 19.09 | 0.42 | 19.09 | 0 | 1.73 | 0 | 1.52 | 0 | 0 | 0.28 | 0 | 0 | 17.36 | 0 | 0 | 0 | 0 | 0.42 |
| TAR | 0 | 0 | 10.57 | 0 | 10.57 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 10.57 | 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 |
| CARBON | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| STEAM | 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 |
| NO2 | 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 |
| AIR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| WAXES | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C5 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 2.21 | 0 | 0 | 2.21 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C6 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 2.37 | 0 | 0 | 2.37 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C7 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1.11 | 0 | 0 | 1.11 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C8 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1.14 | 0 | 0 | 1.14 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C9 | 0 | 0 | 0 | 0.07 | 0 | 0 | 0 | 0 | 1.08 | 0 | 0 | 1.15 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.07 |
| C10 | 0 | 0 | 0 | 0.02 | 0 | 0 | 0 | 0 | 1.13 | 0 | 0 | 1.15 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.02 |
| C11 | 0 | 0 | 0 | 0.03 | 0 | 0 | 0 | 0 | 0.50 | 0 | 0 | 0.54 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.03 |
| C12 | 0 | 0 | 0 | 0.07 | 0 | 0 | 0 | 0 | 0.44 | 0 | 0 | 0.53 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.07 |
| C13 | 0 | 0 | 0 | 0.10 | 0 | 0 | 0 | 0 | 0.34 | 0 | 0 | 0.51 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.10 |
| C14 | 0 | 0 | 0 | 0.09 | 0 | 0 | 0 | 0 | 0.20 | 0 | 0 | 0.49 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.09 |
| C15 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.46 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C16 | 0 | 0 | 0 | 0.02 | 0 | 0 | 0 | 0 | 0.04 | 0 | 0 | 0.45 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.02 |
| C17 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C18 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C19 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C20 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| BIOMASS | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| ASH | 0 | 0 | 1.01 | 0 | 0 | 1.01 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CHAR | 0 | 0 | 0.81 | 0 | 0 | 0.81 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |

**Volumetric and Mole flow values do not include biomass, ash, or char

# Table E-12. Acid Gas Removal and Sulfur Recovery Areas Stream Data for LT Scenario

| HT A300AGR | AG02MEAR | AG03MEAR | AG04MEAR | AG06MEAR | AG08MEAL | AG09MEAL | AG10MEAL | AG11MEAL | AG23ACG | AG24ACG | AG25ACG | AG26SGAS | AG32SGAS | AG33SGAS | AG49SGAS | A300SUL | SU25ACG | SU26SUL | SU40CO2 | SU83SUL | SU92AIR | SU95AIR |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Temperature (C) | 50 | 62 | 50 | 93 | 50 | 123 | 80 | 80 | 93 | 50 | 50 | 40 | 50 | 62 | 32 | | 50 | 50 | 50 | 50 | 50 | 50 |
| Pressure (bar) | 22.89 | 22.89 | 22.89 | 3.45 | 3.45 | 3.45 | 3.45 | 26.00 | 3.45 | 3.45 | 3.45 | 26.86 | 22.89 | 22.89 | 22.89 | | 3.45 | 3.45 | 3.45 | 2.07 | 2.07 | 2.07 |
| Vapor Fraction | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 1.00 | 0.88 | 1.00 | 1.00 | 0.93 | 1.00 | 1.00 | | 1.00 | 0.82 | 1.00 | 0.00 | 1.00 | 1.00 |
| Volume Flow** ($m^3$/sec) | 0.31 | 0 | 0.31 | 0.33 | 0 | 0.31 | 0.30 | 0.30 | 4.24 | 3.28 | 3.28 | 0.98 | 2.30 | 2.40 | 1.50 | | 3.28 | 0.06 | 3.21 | 0 | 0 | 0.10 |
| Mole Flow** (kmol/hr) | 54067 | 449.42 | 54516 | 54516 | 207.83 | 52978 | 52978 | 52978 | 1746 | 1746 | 1538 | 3638 | 7515 | 7066 | 4869 | | 1538 | 33.30 | 1505 | 5.56 | 0.45 | 27.74 |
| Mass Flow (tonnes/day) | 28640 | 208.71 | 28848 | 28848 | 101.05 | 27263 | 27263 | 27263 | 1687 | 1687 | 1586 | 2190 | 2915 | 2706 | 2071 | | 1586 | 25.65 | 1560 | 3.10 | 0.35 | 22.89 |
| H2O | 20923 | 181.16 | 21104 | 21104 | 82.74 | 21081 | 21081 | 21081 | 105.52 | 105.52 | 22.78 | 3.78 | 211.34 | 30.18 | 0 | | 22.78 | 2.28 | 20.50 | 2.40 | 0 | 0.27 |
| CO | 0 | 3.07 | 3.07 | 3.07 | 0 | 0 | 0 | 0 | 3.07 | 3.07 | 3.07 | 782.59 | 1578 | 1575 | 795.10 | | 3.07 | 3.07 | 0 | 0 | 0 | 3.07 |
| H2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 47.68 | 168.38 | 168.38 | 120.71 | | 0 | 0 | 0 | 0 | 0 | 0 |
| CO2 | 1551 | 4.53 | 1556 | 1556 | 16.15 | 16.15 | 16.15 | 16.15 | 1556 | 1556 | 1539 | 1164 | 172.33 | 167.80 | 559.11 | | 1539 | 0 | 1539 | 0 | 0 | 0 |
| O2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0.35 | 0 |
| N2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 |
| AR | 0 | 1.42 | 1.42 | 1.42 | 0 | 0 | 0 | 0 | 1.42 | 1.42 | 1.42 | 32.31 | 412.92 | 411.50 | 380.61 | | 1.42 | 1.42 | 0 | 0 | 0 | 1.42 |
| CH4 | 0 | 2.19 | 2.19 | 2.19 | 0.01 | 0.01 | 0.01 | 0.01 | 2.19 | 2.19 | 2.18 | 99.23 | 151.24 | 149.06 | 52.01 | | 2.18 | 2.18 | 0 | 0 | 0 | 2.18 |
| C2H6 | 0 | 1.13 | 1.13 | 1.13 | 0.02 | 0.02 | 0.02 | 0.02 | 1.13 | 1.13 | 1.11 | 16.07 | 29.62 | 28.49 | 13.55 | | 1.11 | 1.11 | 0 | 0 | 0 | 1.11 |
| C2H4 | 0 | 1.16 | 1.16 | 1.16 | 0.01 | 0.01 | 0.01 | 0.01 | 1.16 | 1.16 | 1.14 | 36.85 | 37.21 | 36.05 | 0.36 | | 1.14 | 1.14 | 0 | 0 | 0 | 1.14 |
| C6H6 | 0 | 0.15 | 0.15 | 0.15 | 0 | 0 | 0 | 0 | 0.15 | 0.15 | 0.15 | 3.11 | 39.65 | 39.50 | 36.54 | | 0.15 | 0.15 | 0 | 0 | 0 | 0.15 |
| C3 | 0 | 4.44 | 4.44 | 4.44 | 0.15 | 0.15 | 0.15 | 0.15 | 4.44 | 4.44 | 4.29 | 0 | 58.62 | 54.18 | 58.62 | | 4.29 | 4.29 | 0 | 0 | 0 | 4.29 |
| C4 | 0 | 7.42 | 7.42 | 7.42 | 0.69 | 0.69 | 0.69 | 0.69 | 7.41 | 7.41 | 6.73 | 0 | 41.73 | 34.31 | 41.73 | | 6.73 | 6.73 | 0 | 0 | 0 | 6.73 |
| H2S | 2.51 | 0.01 | 2.52 | 2.52 | 0.04 | 0.04 | 0.04 | 0.04 | 2.52 | 2.52 | 2.48 | 2.64 | 0.13 | 0.13 | 0 | | 2.48 | 2.48 | 0 | 0 | 0 | 1.73 |
| NH3 | 0 | 0.49 | 0.49 | 0.49 | 0.07 | 0.07 | 0.07 | 0.07 | 0.49 | 0.49 | 0.42 | 1.73 | 2.01 | 1.52 | 0.28 | | 0.42 | 0.42 | 0 | 0 | 0 | 0.42 |
| TAR | 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 |
| CARBON | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 |
| STEAM | 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 |
| NO2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 |
| MEA | 6163 | 0 | 6163 | 6163 | 0 | 6163 | 6163 | 6163 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 |
| AIR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 |
| WAXES | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 |
| C5 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 2.21 | 2.21 | 2.21 | | 0 | 0 | 0 | 0 | 0 | 0 |
| C6 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 2.37 | 2.37 | 2.37 | | 0 | 0 | 0 | 0 | 0 | 0 |
| C7 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1.11 | 1.11 | 1.11 | | 0 | 0 | 0 | 0 | 0 | 0 |
| C8 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1.14 | 1.14 | 1.14 | | 0 | 0 | 0 | 0 | 0 | 0 |
| C9 | 0 | 0.07 | 0.07 | 0.07 | 0 | 0 | 0 | 0 | 0.07 | 0.07 | 0.07 | 0 | 1.15 | 1.08 | 1.15 | | 0.07 | 0.07 | 0 | 0 | 0 | 0.07 |
| C10 | 0 | 0.02 | 0.02 | 0.02 | 0 | 0 | 0 | 0 | 0.02 | 0.02 | 0.02 | 0 | 1.15 | 1.13 | 1.15 | | 0.02 | 0.02 | 0 | 0 | 0 | 0.02 |
| C11 | 0 | 0.04 | 0.04 | 0.04 | 0 | 0 | 0 | 0 | 0.04 | 0.04 | 0.03 | 0 | 0.54 | 0.50 | 0.54 | | 0.03 | 0.03 | 0 | 0 | 0 | 0.03 |
| C12 | 0 | 0.08 | 0.08 | 0.08 | 0.02 | 0.02 | 0.02 | 0.02 | 0.08 | 0.08 | 0.07 | 0 | 0.53 | 0.44 | 0.53 | | 0.07 | 0.07 | 0 | 0 | 0 | 0.07 |
| C13 | 0 | 0.17 | 0.17 | 0.17 | 0.08 | 0.08 | 0.08 | 0.08 | 0.17 | 0.17 | 0.10 | 0 | 0.51 | 0.34 | 0.51 | | 0.10 | 0.10 | 0 | 0 | 0 | 0.10 |
| C14 | 0 | 0.29 | 0.29 | 0.29 | 0.20 | 0.20 | 0.20 | 0.20 | 0.29 | 0.29 | 0.09 | 0 | 0.49 | 0.20 | 0.49 | | 0.09 | 0.09 | 0 | 0 | 0 | 0.09 |
| C15 | 0 | 0.46 | 0.46 | 0.46 | 0.46 | 0.46 | 0.46 | 0.46 | 0.46 | 0.46 | 0 | 0 | 0.46 | 0 | 0.46 | | 0 | 0 | 0 | 0 | 0 | 0 |
| C16 | 0 | 0.41 | 0.41 | 0.41 | 0.38 | 0.38 | 0.38 | 0.38 | 0.41 | 0.41 | 0.02 | 0 | 0.45 | 0.04 | 0.45 | | 0.02 | 0.02 | 0 | 0 | 0 | 0.02 |
| C17 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 |
| C18 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 |
| C19 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 |
| C20 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 |
| BIOMASS | 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 |
| CHAR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | | 0 | 0 | 0 | 0 | 0 | 0 |

**Volumetric and Mole flow values do not include biomass, ash, or char

## Table E-13. Fuel Synthesis Area Stream Data for LT Scenario

| HT A400 | FS34SGAS | FS35SGAS | FS36SGAS | FS37SGAS | FS40SGAS | FS41GAS | FS50SGAS | FS51FT | FS52FT | FS53SGAS | FS54FT | FS55SGAS | FS56FT | FS56SGAS | FS57SGAS | FS58SGAS | FS59SGAS | FS60WAT | FS90HYD |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Temperature (C) | 62 | 78 | 200 | 200 | 60 | 235 | 200 | 200 | 35 | 32 | 35 | 32 | 35 | 32 | 32 | 41 | 200 | 35 | 60 |
| Pressure (bar) | 22.89 | 26.50 | 25.12 | 25.12 | 1.01 | 25.00 | 24.96 | 22.89 | 22.89 | 22.89 | 22.89 | 22.89 | 22.89 | 22.89 | 22.89 | 25.30 | 24.96 | 22.89 | 1.01 |
| Vapor Fraction | 1.00 | 1.00 | 1.00 | 1.00 | 0.88 | 1.00 | 1.00 | 1.00 | 0.72 | 1.00 | 0.00 | 1.00 | 0.00 | 1.00 | 1.00 | 1.00 | 1.00 | 0.00 | 1.00 |
| Volume Flow** ($m^3$/sec) | 2.40 | 2.18 | 4.44 | 0.07 | 0.65 | 0.05 | 5.02 | 4.39 | 2.08 | 2.02 | 0.01 | 1.50 | 0.01 | 0.12 | 0.41 | 0.38 | 0.58 | 0.01 | 0.53 |
| Mole Flow** (kmol/hr) | 7066 | 7066 | 10153 | 166.97 | 97.01 | 97.01 | 11399 | 9268 | 9268 | 6579 | 70.21 | 4869 | 69.40 | 394.75 | 1316 | 1316 | 1316 | 2619 | 69.96 |
| Mass Flow (tonnes/day) | 2706 | 2706 | 3705 | 60.93 | 57.55 | 57.55 | 4261 | 4261 | 4261 | 2798 | 330.77 | 2071 | 330.42 | 167.90 | 559.66 | 559.66 | 559.66 | 1132 | 3.38 |
| H2O | 30.18 | 30.18 | 671.91 | 11.05 | 11.05 | 11.05 | 671.91 | 1133 | 1133 | 0 | 0.43 | 0 | 0.09 | 0 | 0 | 0 | 0 | 1132 | 0 |
| CO | 1575 | 1575 | 1576 | 25.92 | 25.92 | 25.92 | 1791 | 1074 | 1074 | 1074 | 0 | 795.10 | 0 | 64.47 | 214.89 | 214.89 | 214.89 | 0 | 0 |
| H2 | 168.38 | 168.38 | 242.13 | 3.98 | 0.60 | 0.60 | 271.37 | 163.12 | 163.12 | 163.12 | 0 | 120.71 | 0 | 9.79 | 32.62 | 32.62 | 32.62 | 0 | 3.38 |
| CO2 | 167.80 | 167.80 | 604.44 | 9.94 | 9.94 | 9.94 | 755.55 | 755.55 | 755.55 | 755.55 | 0 | 559.11 | 0 | 45.33 | 151.11 | 151.11 | 151.11 | 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 |
| AR | 411.50 | 411.50 | 411.50 | 6.77 | 6.77 | 6.77 | 514.36 | 514.36 | 514.36 | 514.34 | 0 | 380.61 | 0 | 30.86 | 102.87 | 102.87 | 102.87 | 0 | 0 |
| CH4 | 149.06 | 149.06 | 52.13 | 0.86 | 0.86 | 0.86 | 66.19 | 70.29 | 70.29 | 70.29 | 0 | 52.01 | 0 | 4.22 | 14.06 | 14.06 | 14.06 | 0 | 0 |
| C2H6 | 28.49 | 28.49 | 7.72 | 0.13 | 0.13 | 0.13 | 11.38 | 18.30 | 18.30 | 18.30 | 0 | 13.55 | 0 | 1.10 | 3.66 | 3.66 | 3.66 | 0 | 0 |
| C2H4 | 36.05 | 36.05 | 0.39 | 0.01 | 0.01 | 0.01 | 0.49 | 0.49 | 0.49 | 0.49 | 0 | 0.36 | 0 | 0.03 | 0.10 | 0.10 | 0.10 | 0 | 0 |
| C6H6 | 39.50 | 39.50 | 39.50 | 0.65 | 0.65 | 0.65 | 49.38 | 49.38 | 49.38 | 49.38 | 0 | 36.54 | 0 | 2.96 | 9.88 | 9.88 | 9.88 | 0 | 0 |
| C3 | 54.18 | 54.18 | 54.18 | 0.89 | 0.89 | 0.89 | 70.02 | 79.16 | 79.16 | 79.22 | 0 | 58.62 | 0 | 4.75 | 15.84 | 15.84 | 15.84 | 0 | 0 |
| C4 | 34.31 | 34.31 | 34.31 | 0.56 | 0.56 | 0.56 | 45.59 | 56.43 | 56.43 | 56.40 | 0 | 41.73 | 0 | 3.38 | 11.28 | 11.28 | 11.28 | 0 | 0 |
| H2S | 0.13 | 0.13 | 0.01 | 0 | 0 | 0 | 0.01 | 0.01 | 0.01 | 0.01 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| NH3 | 1.52 | 1.52 | 0.30 | 0 | 0 | 0 | 0.38 | 0.38 | 0.38 | 0.38 | 0 | 0.28 | 0 | 0.02 | 0.08 | 0.08 | 0.08 | 0 | 0 |
| TAR | 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 |
| CARBON | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| STEAM | 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 |
| NO2 | 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 |
| AIR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| WAXES | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 131.46 | 131.46 | 0 | 131.46 | 0 | 131.46 | 0 | 0 | 0 | 0 | 0 | 0 |
| C5 | 2.21 | 2.21 | 2.21 | 0.04 | 0.04 | 0.04 | 2.80 | 14.91 | 14.91 | 2.98 | 11.92 | 2.21 | 11.92 | 0.18 | 0.60 | 0.60 | 0.60 | 0 | 0 |
| C6 | 2.37 | 2.37 | 2.37 | 0.04 | 0.04 | 0.04 | 3.01 | 16.02 | 16.02 | 3.20 | 12.82 | 2.37 | 12.82 | 0.19 | 0.64 | 0.64 | 0.64 | 0 | 0 |
| C7 | 1.11 | 1.11 | 1.11 | 0.02 | 0.02 | 0.02 | 1.41 | 15.03 | 15.03 | 1.50 | 13.53 | 1.11 | 13.53 | 0.09 | 0.30 | 0.30 | 0.30 | 0 | 0 |
| C8 | 1.14 | 1.14 | 1.14 | 0.02 | 0.02 | 0.02 | 1.45 | 15.42 | 15.42 | 1.54 | 13.88 | 1.14 | 13.88 | 0.09 | 0.31 | 0.31 | 0.31 | 0 | 0 |
| C9 | 1.08 | 1.08 | 1.08 | 0.02 | 0.02 | 0.02 | 1.39 | 15.50 | 15.50 | 1.55 | 13.95 | 1.15 | 13.95 | 0.09 | 0.31 | 0.31 | 0.31 | 0 | 0 |
| C10 | 1.13 | 1.13 | 1.13 | 0.02 | 0.02 | 0.02 | 1.44 | 15.54 | 15.54 | 1.55 | 13.98 | 1.15 | 13.98 | 0.09 | 0.31 | 0.31 | 0.31 | 0 | 0 |
| C11 | 0.50 | 0.50 | 0.50 | 0.01 | 0.01 | 0.01 | 0.65 | 14.59 | 14.59 | 0.73 | 13.86 | 0.54 | 13.86 | 0.04 | 0.15 | 0.15 | 0.15 | 0 | 0 |
| C12 | 0.44 | 0.44 | 0.44 | 0.01 | 0.01 | 0.01 | 0.59 | 14.26 | 14.26 | 0.71 | 13.54 | 0.53 | 13.54 | 0.04 | 0.14 | 0.14 | 0.14 | 0 | 0 |
| C13 | 0.34 | 0.34 | 0.34 | 0.01 | 0.01 | 0.01 | 0.47 | 13.79 | 13.79 | 0.69 | 13.10 | 0.51 | 13.10 | 0.04 | 0.14 | 0.14 | 0.14 | 0 | 0 |
| C14 | 0.20 | 0.20 | 0.20 | 0 | 0 | 0 | 0.34 | 13.23 | 13.23 | 0.66 | 12.57 | 0.49 | 12.57 | 0.04 | 0.13 | 0.13 | 0.13 | 0 | 0 |
| C15 | 0 | 0 | 0 | 0 | 0 | 0 | 0.13 | 12.55 | 12.55 | 0.63 | 11.93 | 0.46 | 11.93 | 0.04 | 0.13 | 0.13 | 0.13 | 0 | 0 |
| C16 | 0.04 | 0.04 | 0.04 | 0 | 0 | 0 | 0.16 | 12.08 | 12.08 | 0.60 | 11.48 | 0.45 | 11.48 | 0.04 | 0.12 | 0.12 | 0.12 | 0 | 0 |
| C17 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 11.39 | 11.39 | 0 | 11.39 | 0 | 11.39 | 0 | 0 | 0 | 0 | 0 | 0 |
| C18 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 10.85 | 10.85 | 0 | 10.85 | 0 | 10.85 | 0 | 0 | 0 | 0 | 0 | 0 |
| C19 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 10.31 | 10.31 | 0 | 10.31 | 0 | 10.31 | 0 | 0 | 0 | 0 | 0 | 0 |
| C20 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 9.76 | 9.76 | 0 | 9.76 | 0 | 9.76 | 0 | 0 | 0 | 0 | 0 | 0 |
| BIOMASS | 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 |
| CHAR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |

**Volumetric and Mole flow values do not include biomass, ash, or char

## Table E-14. Syngas Conditioning Area Stream Data for LT Scenario

| HT A400COND | CD35SGAS | CD36SGAS | CD37SGAS | CD38SGAS | CD39SGAS | CD40SGAS | CD41SGAS | CD42SGAS | CD44SGAS | CD45SGAS | CD80STM | CD81STM | CD99CONT |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Temperature (C) | 78 | 150 | 150 | 870 | 870 | 300 | 300 | 474 | 363 | 200 | 300 | 870 | 150 |
| Pressure (bar) | 26.50 | 26.50 | 26.50 | 26.50 | 25.81 | 25.81 | 25.81 | 25.12 | 25.12 | 25.12 | 25.00 | 27.00 | 26.50 |
| Vapor Fraction | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 |
| Volume Flow** ($m^3$/sec) | 2.18 | 2.64 | 2.64 | 7.08 | 10.44 | 5.24 | 1.83 | 2.46 | 5.98 | 4.44 | 1.15 | 2.26 | 0 |
| Mole Flow** (kmol/hr) | 7066 | 7066 | 7063 | 7063 | 10153 | 10153 | 3554 | 3554 | 10153 | 10153 | 2313 | 2313 | 3.12 |
| Mass Flow (tonnes/day) | 2706 | 2706 | 2705 | 2705 | 3705 | 3705 | 1297 | 1297 | 3705 | 3705 | 1000 | 1000 | 1.34 |
| H2O | 30.18 | 30.18 | 30.18 | 30.18 | 850.65 | 850.65 | 297.73 | 118.99 | 671.91 | 671.91 | 1000 | 1000 | 0 |
| CO | 1575 | 1575 | 1575 | 1575 | 1854 | 1854 | 648.81 | 370.91 | 1576 | 1576 | 0 | 0 | 0 |
| H2 | 168.38 | 168.38 | 168.38 | 168.38 | 222.13 | 222.13 | 77.75 | 97.75 | 242.13 | 242.13 | 0 | 0 | 0 |
| CO2 | 167.80 | 167.80 | 167.80 | 167.80 | 167.80 | 167.80 | 58.73 | 495.37 | 604.44 | 604.44 | 0 | 0 | 0 |
| O2 | 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 |
| AR | 411.50 | 411.50 | 411.50 | 411.50 | 411.50 | 411.50 | 144.02 | 144.02 | 411.50 | 411.50 | 0 | 0 | 0 |
| CH4 | 149.06 | 149.06 | 149.06 | 149.06 | 52.13 | 52.13 | 18.25 | 18.25 | 52.13 | 52.13 | 0 | 0 | 0 |
| C2H6 | 28.49 | 28.49 | 28.49 | 28.49 | 7.72 | 7.72 | 2.70 | 2.70 | 7.72 | 7.72 | 0 | 0 | 0 |
| C2H4 | 36.05 | 36.05 | 36.05 | 36.05 | 0.39 | 0.39 | 0.14 | 0.14 | 0.39 | 0.39 | 0 | 0 | 0 |
| C6H6 | 39.50 | 39.50 | 39.50 | 39.50 | 39.50 | 39.50 | 13.83 | 13.83 | 39.50 | 39.50 | 0 | 0 | 0 |
| C3 | 54.18 | 54.18 | 54.18 | 54.18 | 54.18 | 54.18 | 18.96 | 18.96 | 54.18 | 54.18 | 0 | 0 | 0 |
| C4 | 34.31 | 34.31 | 34.31 | 34.31 | 34.31 | 34.31 | 12.01 | 12.01 | 34.31 | 34.31 | 0 | 0 | 0 |
| H2S | 0.13 | 0.13 | 0.01 | 0.01 | 0.01 | 0.01 | 0 | 0 | 0.01 | 0.01 | 0 | 0 | 0.12 |
| NH3 | 1.52 | 1.52 | 0.30 | 0.30 | 0.30 | 0.30 | 0.11 | 0.11 | 0.30 | 0.30 | 0 | 0 | 1.21 |
| TAR | 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 |
| CARBON | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| STEAM | 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 |
| NO2 | 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 |
| AIR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| WAXES | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C5 | 2.21 | 2.21 | 2.21 | 2.21 | 2.21 | 2.21 | 0.77 | 0.77 | 2.21 | 2.21 | 0 | 0 | 0 |
| C6 | 2.37 | 2.37 | 2.37 | 2.37 | 2.37 | 2.37 | 0.83 | 0.83 | 2.37 | 2.37 | 0 | 0 | 0 |
| C7 | 1.11 | 1.11 | 1.11 | 1.11 | 1.11 | 1.11 | 0.39 | 0.39 | 1.11 | 1.11 | 0 | 0 | 0 |
| C8 | 1.14 | 1.14 | 1.14 | 1.14 | 1.14 | 1.14 | 0.40 | 0.40 | 1.14 | 1.14 | 0 | 0 | 0 |
| C9 | 1.08 | 1.08 | 1.08 | 1.08 | 1.08 | 1.08 | 0.38 | 0.38 | 1.08 | 1.08 | 0 | 0 | 0 |
| C10 | 1.13 | 1.13 | 1.13 | 1.13 | 1.13 | 1.13 | 0.40 | 0.40 | 1.13 | 1.13 | 0 | 0 | 0 |
| C11 | 0.50 | 0.50 | 0.50 | 0.50 | 0.50 | 0.50 | 0.18 | 0.18 | 0.50 | 0.50 | 0 | 0 | 0 |
| C12 | 0.44 | 0.44 | 0.44 | 0.44 | 0.44 | 0.44 | 0.16 | 0.16 | 0.44 | 0.44 | 0 | 0 | 0 |
| C13 | 0.34 | 0.34 | 0.34 | 0.34 | 0.34 | 0.34 | 0.12 | 0.12 | 0.34 | 0.34 | 0 | 0 | 0 |
| C14 | 0.20 | 0.20 | 0.20 | 0.20 | 0.20 | 0.20 | 0.07 | 0.07 | 0.20 | 0.20 | 0 | 0 | 0 |
| C15 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C16 | 0.04 | 0.04 | 0.04 | 0.04 | 0.04 | 0.04 | 0.01 | 0.01 | 0.04 | 0.04 | 0 | 0 | 0 |
| C17 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C18 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C19 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C20 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| BIOMASS | 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 |
| CHAR | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |

**Volumetric and Mole flow values do not include biomass, ash, or char

## Table E-15. Hydroprocessing, Power Generation, and Air Separation Areas Stream Data for LT Scenario

| HT A500 | HY56FT | HY65FGAS | HY70GASO | HY71GASO | HY80DIES | HY81DIES | HY90HYD | A600 | 421 | 602 | 615 | 680 | CW2 | CWREC | HPW1 | HPSTM | A700 | AIR-1A | AIR-A | AIR-L | N2-OUT | O2LPC | O2-OUT | O2-OUT2 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Temperature (C) | 35 | 50 | 50 | 50 | 50 | 50 | 60 |  | 32 | 1161 | 344 | 30 | 85 | 23 | 170 | 393 |  | -170 | 32 | 20 | 16 | 35 | 16 | 55 |
| Pressure (bar) | 22.89 | 22.89 | 22.89 | 22.89 | 22.89 | 22.89 | 1.01 |  | 22.89 | 1.00 | 1.00 | 1.01 | 7.91 | 0.02 | 7.91 | 173.38 |  | 6.20 | 1.01 | 6.30 | 1.10 | 23.08 | 1.10 | 27.00 |
| Vapor Fraction | 0.00 | 1.00 | 0.00 | 0.00 | 0.00 | 0.00 | 1.00 |  | 1.00 | 1.00 | 1.00 | 1.00 | 0.00 | 0.89 | 0.00 | 1.00 |  | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 | 1.00 |
| Volume Flow** ($m^3$/sec) | 0.01 | 0.02 | 0 | 0 | 0 | 0 | 0.53 |  | 0.12 | 94.88 | 40.84 | 17.42 | 0.01 | 442.16 | 0 | 0.03 |  | 1.06 | 23.17 | 3.58 | 15.73 | 0.22 | 4.45 | 0.21 |
| Mole Flow** (kmol/hr) | 69.40 | 57.67 | 31.78 | 31.78 | 37.88 | 37.88 | 69.96 |  | 394.75 | 2863 | 2863 | 2522 | 1709 | 1709 | 1709 | 1709 |  | 3162 | 3328 | 3328 | 2594 | 734.20 | 734.20 | 734.20 |
| Mass Flow (tonnes/day) | 330.42 | 40.83 | 87.12 | 87.12 | 205.86 | 205.86 | 3.38 |  | 167.90 | 1955 | 1955 | 1746 | 740.10 | 740.10 | 740.10 | 740.10 |  | 2198 | 2313 | 2313 | 1744 | 569.18 | 569.18 | 569.18 |
| H2O | 0.09 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 189.23 | 189.23 | 0 | 740.10 | 740.10 | 740.10 | 740.10 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CO | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 64.47 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| H2 | 0 | 0 | 0 | 0 | 0 | 0 | 3.38 |  | 9.79 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CO2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 45.33 | 318.35 | 318.35 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| O2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 80.99 | 80.99 | 410.78 | 0 | 0 | 0 | 0 |  | 508.87 | 535.65 | 535.65 | 0 | 535.65 | 535.65 | 535.65 |
| N2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 1336 | 1336 | 1336 | 0 | 0 | 0 | 0 |  | 1661 | 1748 | 1748 | 1744 | 3.91 | 3.91 | 3.91 |
| AR | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 30.86 | 30.86 | 30.86 | 0 | 0 | 0 | 0 | 0 |  | 28.22 | 29.71 | 29.71 | 0.10 | 29.61 | 29.61 | 29.61 |
| CH4 | 0 | 11.56 | 0 | 0 | 0 | 0 | 0 |  | 4.22 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H6 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 1.10 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C2H4 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0.03 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C6H6 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 2.96 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C3 | 0 | 29.27 | 0 | 0 | 0 | 0 | 0 |  | 4.75 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C4 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 3.38 | 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 |
| NH3 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0.02 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| TAR | 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 |
| CARBON | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| STEAM | 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 |
| NO2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0.06 | 0.06 | 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 |
| AIR | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| WAXES | 131.46 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C5 | 11.92 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0.18 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C6 | 12.82 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0.19 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C7 | 13.53 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0.09 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C8 | 13.88 | 0 | 87.12 | 87.12 | 0 | 0 | 0 |  | 0.09 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C9 | 13.95 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0.09 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C10 | 13.98 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0.09 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C11 | 13.86 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0.04 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C12 | 13.54 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0.04 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C13 | 13.10 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0.04 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C14 | 12.57 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0.04 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C15 | 11.93 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0.04 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C16 | 11.48 | 0 | 0 | 0 | 205.86 | 205.86 | 0 |  | 0.04 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C17 | 11.39 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C18 | 10.85 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C19 | 10.31 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| C20 | 9.76 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| BIOMASS | 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 |
| CHAR | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |  | 0 | 0 | 0 | 0 | 0 | 0 | 0 |

**Volumetric and Mole flow values do not include biomass, ash, or char

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**14. ABSTRACT** *(Maximum 200 Words)*

This study compares capital and production costs of two biomass-to-liquid production plants based on gasification. The first biorefinery scenario is an oxygen-fed, low-temperature (870°C), non-slagging, fluidized bed gasifier. The second scenario is an oxygen-fed, high-temperature (1,300°C), slagging, entrained flow gasifier. Both are followed by catalytic Fischer-Tropsch synthesis and hydroprocessing to naphtha-range (gasoline blend stock) and distillate-range (diesel blend stock) liquid fractions. Process modeling software (Aspen Plus) is utilized to organize the mass and energy streams and cost estimation software is used to generate equipment costs. Economic analysis is performed to estimate the capital investment and operating costs. Results show that the total capital investment required for nth plant scenarios is $610 million and $500 million for high-temperature and low-temperature scenarios, respectively. Product value (PV) for the high-temperature and low-temperature scenarios is estimated to be $4.30 and $4.80 per gallon of gasoline equivalent (GGE), respectively, based on a feedstock cost of $75 per dry short ton. Sensitivity analysis is also performed on process and economic parameters. This analysis shows that total capital investment and feedstock cost are among the most influential parameters affecting the PV.

**15. SUBJECT TERMS**
biomass; biofuels; gasification; Fischer-Tropsch; techno-economic analysis; biomass-to-liquid; thermochemical

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