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Biomass to Hydrogen Production Detailed Design and Economics Utilizing the Battelle Columbus Laboratory Indirectly-Heated Gasifier

Technical Report
NREL/TP-510-37408
May 2005

P. Spath, A. Aden, T. Eggeman,
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Executive Summary

The U.S. Department of Energy (DOE) Biomass Program promotes the development of technologies for converting biomass into valuable fuels, chemicals, and power that foster the growth of biorefineries with the goal of reducing foreign oil imports. With this in mind, in 2003, the National Renewable Energy Laboratory (NREL) conducted an extensive literature search and examined the technical and economic feasibility of numerous fuels and chemicals from biomass-derived syngas (Spath and Dayton, 2003). Hydrogen was one product that emerged as highly favorable in this technical and economic feasibility study. Therefore, hydrogen was chosen as a model product to conduct further analysis and examine the process integration effects and economics of a final product from biomass gasification.

This analysis developed detailed process flow diagrams and an Aspen Plus® model, evaluated energy flows including a pinch analysis, obtained process equipment and operating costs, and performed an economic evaluation of two process designs based on the syngas clean up and conditioning work being performed at NREL. One design, the current design, attempts to define today's state of the technology. The other design, the goal design, is a target design that attempts to show the effect of meeting specific research goals. Each process design broadly consists of feed handling, drying, gasification, gas clean up and conditioning, shift conversion, and purification with some unit operation differences. The main difference between the current design and goal design is in the tar reformer. The tar reformer in the current design is a bubbling fluidized bed reactor with 1% per day catalyst replacement. In the goal design, there is a tar reformer/catalyst regenerator system and because the conversion of methane is higher for this case, the steam methane reformer can be eliminated from the process design.

Several parts of the system operate at a high temperature, therefore, heat integration and recovery are important. Each process design recovers process heat in a steam cycle with an extraction steam turbine/generator to produce some power and supply steam for gasification and steam methane reforming or shift conversion.

Both designs utilize the Battelle Columbus Laboratory (BCL) low pressure indirectly-heated gasifier. The base case plant size is 2,000 dry tonne/day and the feedstock cost is $30/dry ton. The current design plant produces 57 MM kg/yr or 66 MM scf/day of hydrogen at 100% capacity. The goal design plant produces 61 MM kg/yr or 71 MM scf/day of hydrogen at 100% capacity.

The results of this analysis show a minimum hydrogen selling price of $1.38/kg ($11.48/GJ, lower heating value [LHV]) for the current design base case analysis and a price of $1.24/kg ($10.34/GJ, LHV) for the goal design. The hydrogen price decreases for the goal design mainly because of an increase in the hydrogen yield. The decrease in the total project investment also has some affect. This result shows that the research at NREL in catalytic tar destruction and heteroatom removal is moving in a direction that has the potential to decrease the cost of producing clean syngas and any subsequent fuel products via biomass gasification.

Several sensitivity cases were run to examine the effects of different parameters on the analysis. The feedstock cost contributes the most to the product hydrogen price (about 30%), and thus this variable will always have a large impact on the economics. Overall, the sensitivity analysis shows that any parameter that significantly affects the heat balance of the system will greatly affect the minimum hydrogen selling price.

As a benchmark for thermochemical conversion, the DOE Biomass Program is setting program targets based on intermediate syngas prices to track progress toward reducing the technical barriers associated with biomass gasification. Therefore, this analysis included calculations in determining both an intermediate and a stand-alone clean, reformed syngas price. The intermediate syngas price for the current and goal designs are $6.88/GJ ($7.25/MMBtu) and $4.98/GJ ($5.25/MMBtu), respectively. This is the price for clean, reformed syngas as an intermediate in the integrated biomass-to-hydrogen design. Stand-alone syngas plants are not being built today, but for a stand-alone plant based on the current design, the syngas price would be $8.22/GJ ($8.67/MMBtu), and $6.73/GJ ($7.10/MMBtu) for a plant based on the goal design. The lower intermediate syngas price shows the importance of integration within the fuels synthesis process plant.

More detailed capital costs in the feed handling, gasification, and clean up areas would improve the accuracy of the analysis. Additionally, more work needs to be done to compare indirect gasification with direct gasification to determine the most suitable and economically viable gasification system for different fuels products. Future work will entail examining other biomass feedstocks and other products along with the integration of thermochemical and biochemical conversion processes into biorefinery concepts.

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Acronyms

Aspen Plus – Advanced Simulator for Process Engineering
BCL – Battelle Columbus Laboratory
BDW – bone dry wood
BFW – boiler feed water
DCFROR – discounted cash flow rate of return
DDB – double declining balance
DOE – Department of Energy
EIA – Energy Information Administration
EOS – equation of state
GHSV – gas hourly space velocity
GJ – gigajoule
gpm – gallons per minute
GTI – Gas Technology Institute
HHV – higher heating value
HTS – high temperature shift
IP – intermediate pressure
IRR – internal rate of return
K.O. – knock out
kW - kilowatt
kWh – kilowatt-hour
LHV – lower heating value
LP – low pressure
LTS – low temperature shift
MACRS – Modified Accelerated Cost Recovery System
MHSP – minimum hydrogen selling price
MMBtu – million British thermal units
MW - megawatt
NREL – National Renewable Energy Laboratory
ORNL – Oak Ridge National Laboratory
PFD – process flow diagram
PSA – pressure swing adsorption
Psia – pounds per square inch absolute
R&D – research and development
SCF – standard cubic feet
TIC – total installed cost
TPEC – total purchased equipment cost
TPI – total project investment
VP – vacuum pressure

1.0 Introduction

In 2003, the National Renewable Energy Laboratory performed a preliminary screening study of potential products from biomass-derived syngas (Spath and Dayton, 2003). This study showed hydrogen to be an economically feasible product, so it was used as a model product to show the process integration effects and economics of a final product from biomass gasification. In general, the analysis performed for the 2003 study was a high-level analysis that gathered material and energy balance information along with capital and operating cost data from various literature sources. In the case of hydrogen, however, NREL had previously developed two Aspen Plus models of hydrogen production via gasification. This analysis builds on one of NREL's models, the indirect gasification model. In the original model's design any excess steam was sold over the fence. In the updated model, a steam cycle produces the amount of steam required by the plant plus some electricity. Additionally, in this analysis the gas clean up and conditioning research work at NREL is also incorporated in the model.

2.0 Analysis Approach

The approach that was used in the development of the process designs and economic analysis can be seen in Figure 1. For this analysis the first step was to develop process flow diagrams (PFDs) and to use these along with literature information and research results to build an Aspen Plus model. The energy and material balance from the Aspen model were used to size equipment and determine capital and operating costs. Additionally, for this analysis, some of the capital costs were obtained from literature sources. Once the capital and operating costs are determined, the information is put into an excel spreadsheet that is set up to calculate the hydrogen selling price using a discounted cash flow rate of return analysis.

3.0 Feedstock and Plant Size

The feedstock used for this analysis is hybrid poplar wood chips delivered at 50 wt% moisture. The ultimate analysis for the feed used in this study is given in Table 1. The plant capacity is designed to be 2,000 bone dry tonne/day. The plant is considered to be an "nth" plant design (i.e., established and not a first of a kind or pioneer plant). The feedstock cost is assumed to be $30/bone dry ton (delivered) for urban wood waste, forest, and mill residues. Information from Oak Ridge National Laboratory (ORNL) suggests that the cumulative amount of biomass available at $30/dry ton is 105 million tons (http://bioenergy.ornl.gov/resourcedata/index.html).

Table 1: Ultimate Analysis of Hybrid Poplar Feed (wt%, dry basis)

Component	C	H	N	S	O	Ash
wt%, dry basis	50.88	6.04	0.17	0.09	41.90	0.92
Heating value (Btu/lb): 8,671 HHV 8,060 LHV (calculated by Aspen Plususing Boie correlation)						

Source: Craig and Mann (1996)

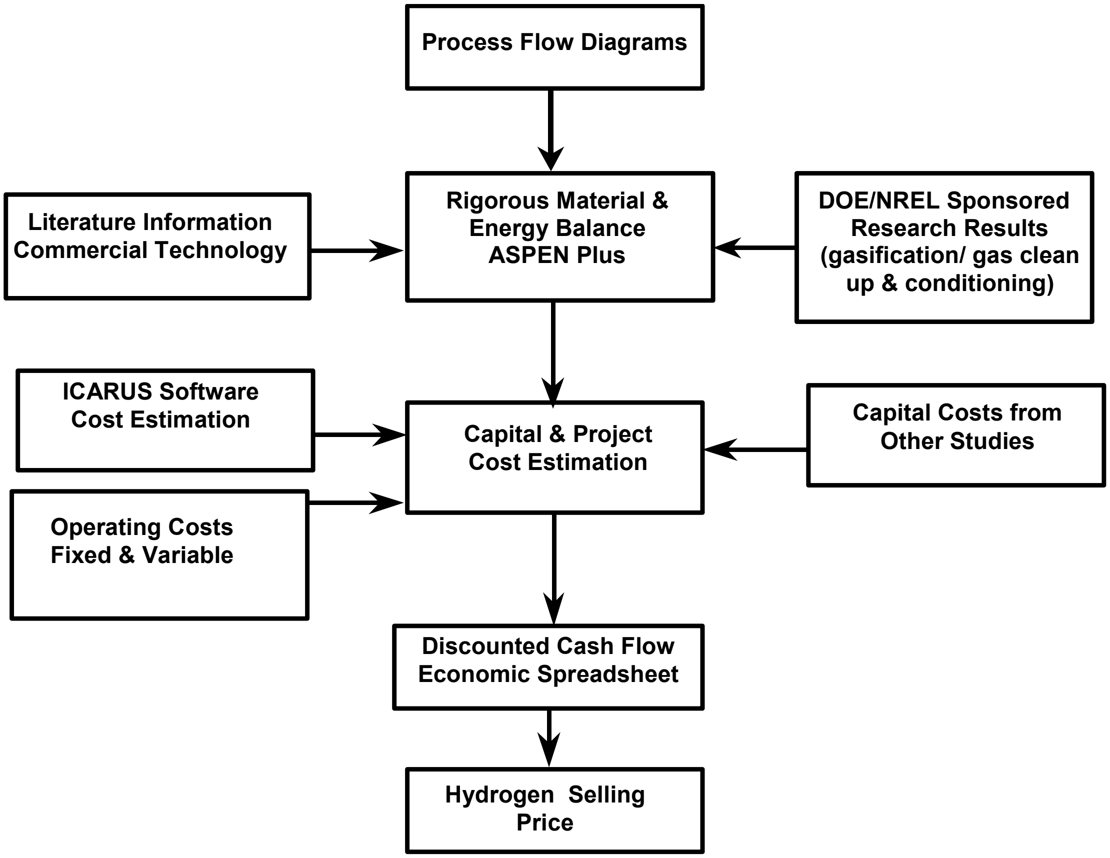

Figure 1: Approach to Process Analysis

4.0 Process Design Basis

Two process designs were examined in this study. They are based on the current operation and performance goals of the catalytic tar destruction and heteroatom removal work at NREL. The current design attempts to define today's state of the technology. The goal design is a target design that attempts to show the effect of meeting specific research and development (R&D) goals. Table 2 gives the percent conversion of various compounds whose concentrations are measured before and after NREL's tar reformer as well as the desired conversion goal (Phillips, *et al*, 2004). Each process design, both the current and goal designs, broadly consists of:

- feed handling,
- drying,
- gasification,
- gas clean up and conditioning,
- shift conversion,
- and hydrogen purification,
- integrated with a steam and power generation cycle.

There are some unit operation differences and the details of these two designs will be discussed the following sections.

Table 2: Tar Reformer Performance - % Conversion to CO & H_2

Compound	Current Design	Goal Design
Methane (CH_4)	20%	80%
Ethane (C_2H_6)	90%	99%
Ethylene (C_2H_4)	50%	90%
Tars (C_{10+})	95%	99.9%
Benzene (C_6H_6)	70%	99%
Ammonia (NH_3)*	70%	90%

* Converts to N_2 and H_2

5.0 Current Design Process Overview

A block flow diagram of the current design can be seen in Figure 2. The process flow diagrams (PFDs) for this process design are included at the end of this report in Appendix C: Current Design Process Flow Diagrams. A more detailed discussion of this process can be found in section 7.0 Current Design - Process Design, Modeling, and Costing and its subsections. First, the as-received wood is dried from 50 wt% moisture down to 12 wt% employing a rotary dryer. The dryer uses gas from the char combustor as the drying medium. Conveyors and hoppers are used to feed the wood to the low-pressure indirectly-heated entrained flow gasifier. Heat for the endothermic gasification reactions is supplied by circulating hot synthetic olivine, which is a calcined magnesium silicate (primarily Enstatite [$MgSiO_3$] Forsterite [Mg_2SiO_3], and Hematite [Fe_2O_3]) used as a sand for various applications, between the gasifier and a char combustor vessel. A small amount of MgO is added to the fresh olivine to prevent the formation of glass-like bed agglomerations that would result from biomass potassium interacting with the silicate compounds. The gasification medium is steam. The char that is formed in the gasifier is burned in the combustor to reheat the olivine. Particulate removal is performed through cyclone separators. Ash and any sand particles that are carried over end up being landfilled.

Gas clean up and conditioning consists of using a tar reformer followed by syngas cooling, compression, sulfur removal, steam methane reforming, and high and low temperature shift conversion. The tar reformer is a bubbling fluidized bed reactor. Catalyst replacement was assumed to be 1% per day of the total catalyst volume (Bain, 2004). The syngas is cooled through heat exchange with the steam cycle and additional cooling via water scrubbing. The scrubber also removes impurities such as particulates and ammonia along with any residual tars. The excess scrubber water is sent off site to a waste-water treatment facility. The syngas is compressed using a five-section centrifugal compressor with interstage cooling. The syngas exiting the gasifier contains almost 400 ppmv of H_2S, therefore sulfur removal is performed using a liquid phase oxidation process (LO-CAT®) followed by a ZnO bed. Elemental sulfur is produced and stockpiled for disposal. It is stockpiled onsite, instead of being sold or disposed of right away, because the amount produced is small and further conditioning would be required before the sulfur could be sold.

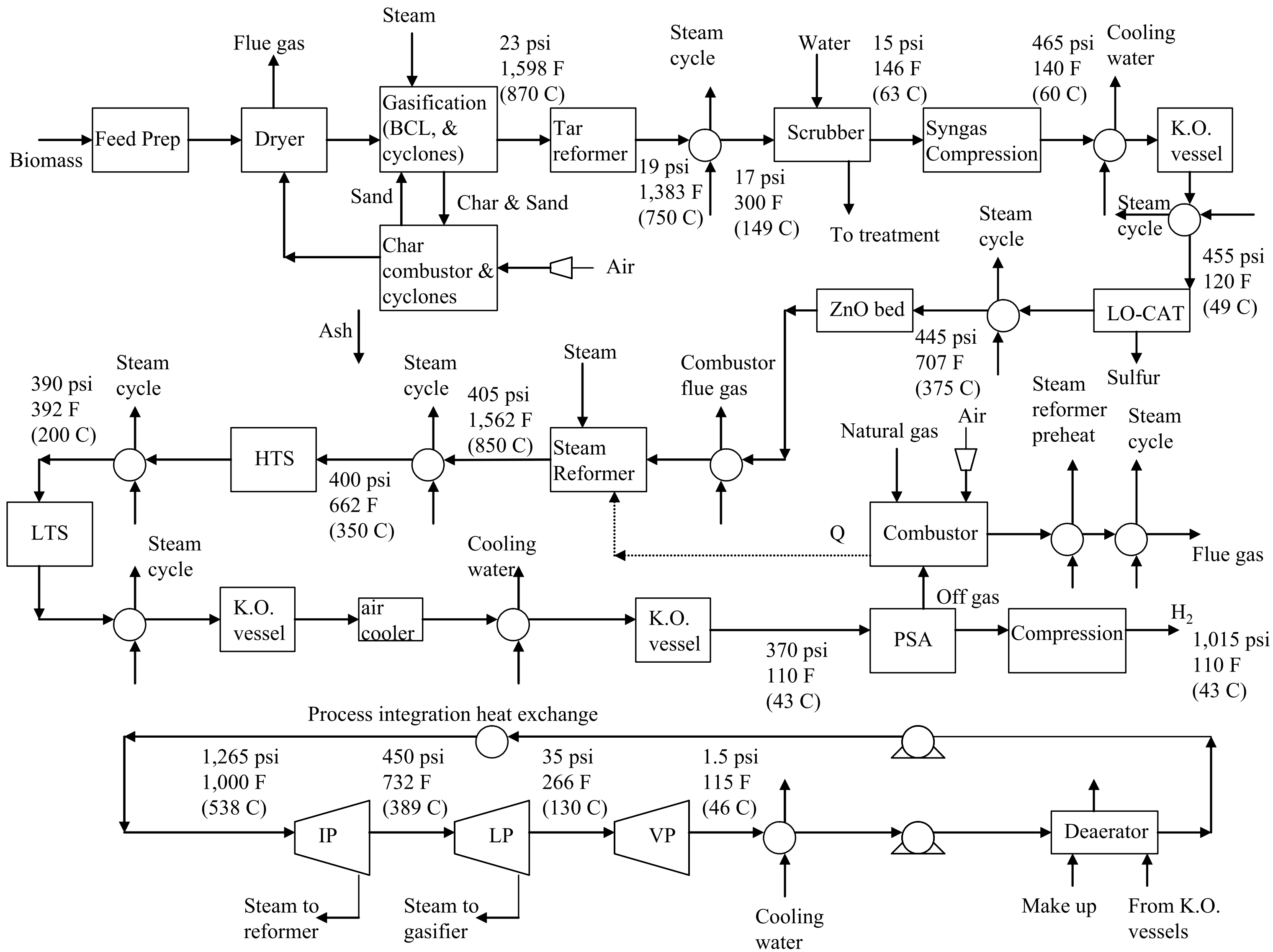

Figure 2: Block Flow Diagram of Current Design

Reforming ($C_nH_m + nH_2O \Leftrightarrow (n+m/2)H_2 + nCO$) and water-gas shift ($CO + H_2O \Leftrightarrow CO_2 + H_2$) are the main reactions in the steam reformer. The steam reformer is fueled by the pressure swing adsorption (PSA) offgas and for burner control a small amount of natural gas is added. The high temperature shift (HTS) and low temperature shift (LTS) reactors convert the majority of the CO when reacted with H_2O into CO_2 and H_2 through the water-gas shift reaction.

For purification, a pressure swing adsorption unit is used to separate the hydrogen from the other components in the shifted gas stream, mainly CO_2, and unreacted CO, CH_4, and other hydrocarbons. For a 70 mol% hydrogen PSA feed, a hydrogen recovery rate of 85% is typical with a product purity of 99.9 vol%. Finally, the hydrogen is compressed to 1,015 psia prior shipment through a pipeline.

The steam cycle produces power in addition to providing steam for the gasifier and reformer operations. The steam cycle is integrated with the biomass-to-hydrogen production process. Steam is supplied to the reformer and gasifier from the intermediate and low pressure turbine sections of the extraction steam turbine/generator, respectively. Superheated steam enters the intermediate pressure turbine at 1,000ºF and 1,265 psia and is expanded to a pressure of 450 psia. The steam then enters a low pressure turbine and is expanded to a pressure of 35 psia. Finally, the steam enters a condensing turbine and is expanded to a pressure of 1.5 psia. Preheaters, steam generators, and superheaters are integrated within the process design. The condensate from the syngas compressor and the condensate from the cooled shifted gas stream prior to the PSA are sent to the steam cycle, de-gassed, and combined with the make-up water. A pinch analysis was performed to determine the heat integration of the system.

A cooling water system is also included in the Aspen Plus model to determine the requirements of each cooling water heat exchanger within the hydrogen production system as well as the requirements of the cooling tower. The cooling water supply temperature is 90°F and the return temperature is 110°F.

6.0 Goal Design Process Overview

The goal design differs from the current design in that the tar reformer now consists of a reactor vessel and a catalyst regeneration vessel. Additionally, since the tar reformer now reforms a significant amount of the syngas methane (see Table 2), the steam reformer was eliminated from the design. The tar reforming reactor/catalyst regenerator system operates isothermally. The heat required for the tar reforming reactor/catalyst regenerator system is supplied by burning the PSA offgas along with some natural gas. The steam to carbon ratio for the shift conversion step is set at 2 mol of H_2O/mol of C. The biomass-to-hydrogen process is integrated with the steam cycle. A block flow diagram of the goal design is shown in Figure 3. Additionally, process flow diagrams (PFDs) for this process design are included at the end of this report in Appendix D: Goal Design Process Flow Diagrams and more detailed information can be found in section 12.0 Design, Modeling, and Capital Cost Changes for Goal Design.

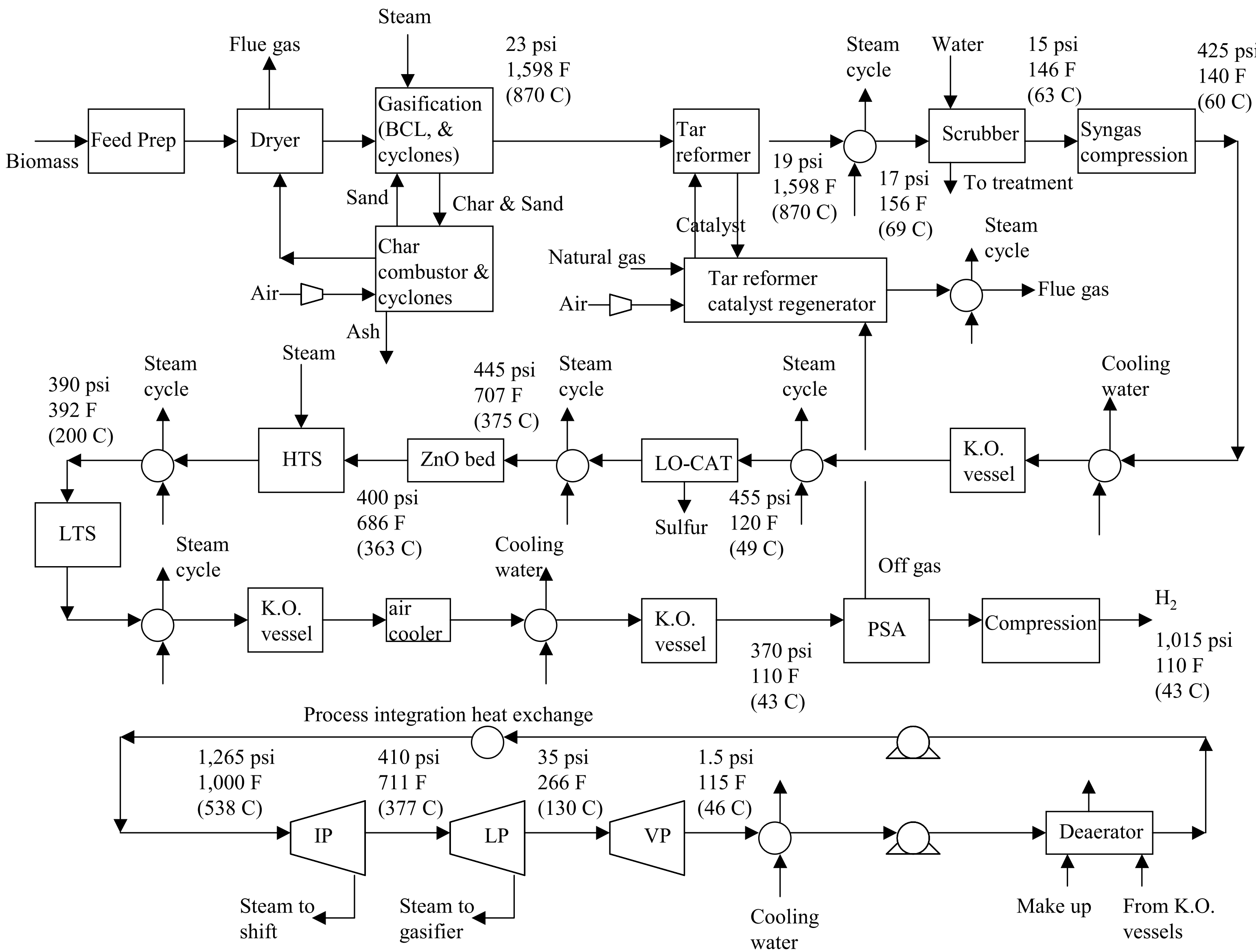

Figure 3: Block Flow Diagram of Goal Design

7.0 Current Design - Process Design, Modeling, and Costing

The following sections describe the detailed process design for the current design as outlined in section 5.0 Current Design Process Overview.

7.1 Feed Handling and Drying – Area 100

The feed handling and drying section are shown in PFD-P700-A101 and PFD-P700-A102. Wood chips are delivered to the plant primarily via trucks. However, it is envisioned that there could be some train transport. Assuming that each truck capacity is about 25 tons (Mann and Spath, 1997), this means that if the wood, at a moisture content of 50%, was delivered to the plant via truck transport only, then 176 truck deliveries per day would be required. As the trucks enter the plant they are weighed (M-101) and the wood chips are dumped into a storage pile. From the storage pile, the wood chips are conveyed (C-102) through a magnetic separator (S-101) and screened (S-102). Particles larger than 2 inches are sent through a hammer mill (T-102/M-102) for further size reduction. Front end loaders transfer the wood chips to the dryer feed bins (T-103).

Because of the large plant size there are two identical, parallel feed handling and drying trains. The wet wood chips enter each rotary biomass dryer (M-104) through a dryer feed screw conveyor (C-104). After drying the wood to a moisture content of 12 wt% with flue gas from the char combustor (R-202), the gas is sent through a cyclone (S-103) and baghouse filter (S-104) to remove any particulates prior to being emitted to the atmosphere. The stack temperature of the flue gas is set at 250°F, which is above the dew point of the gas. The stack temperature is controlled by cooling the hot flue gas (H-101) prior to entering the dryer. This heat is used to generate steam (see section 7.6 Steam System and Power Generation – Area 600). The dried biomass is then conveyed to the gasifier train (T-104/C-105).

7.2 Gasification and Tar Reforming – Area 200

From the feed handling and drying section, the dried wood enters the gasifier and tar reforming section as shown in PFD-P700-A201 and PFD-P700-A202. Because of the plant size, it is assumed that there are two gasifier trains. The gasifier (R-201) used in this analysis is a low-pressure indirectly-heated entrained flow gasifier.

The gasifier was modeled using correlations based on run data from the Battelle Columbus Laboratory (BCL) 9 tonne/day test facility. The data and correlations for the gasifier can be found in Bain (1992). The experimental runs were performed for several different wood types including Red Oak chips, Birch and Maple chips, Pine chips, sawdust, and other hard and soft wood chips. The original pilot plant data for these runs can be found in Feldmann, *et al*, (1988). The temperature range for the data is 1,280-1,857ºF and the pressure range is 2.4-14.4 psig with the majority of the data being in the 1,500-1,672ºF range.

The BCL test facility data was regressed using a polynomial function (Bain, 1992). The quadratic function ($A + B*T + C*T^2$) provides a good fit for the conversion of all of the gas components and the char. The correlations are in terms of standard cubic feet (scf) of component/lb of BDW except for the char and tar, which are in terms of lb of component/lb of BDW. Graphs of the correlations can be found in Appendix E: Graphical Correlations for Gas Components and Char. These correlations along with documentation have been programmed into a Fortran file. Aspen Plus passes the gasifier temperature to the Fortran file, the Fortran file uses the correlations to calculate the gas and char yields then elemental balances are performed for carbon, hydrogen, sulfur, and oxygen to come up with the overall material and energy balance for the gasifer. The elemental balances were put into flow charts and are included in Appendix F: Flow Charts for Gasifier Elemental Balances. The BCL model performs the elemental balances in the following order: carbon, oxygen, sulfur, and hydrogen. However, in general, the sulfur balance can be performed any time as long as it is done before the hydrogen balance. Note, when running the Aspen Plus model it is important for the user to look at the history file for errors, make any necessary changes, re-run the model, and examine the history file again when changing any of the model parameters.

Table 3 gives the resulting operating parameters, tar and char yields, and gas composition for the BCL gasifier from the Aspen Plus model.

Table 3: Gasifier Operating Parameters, Yields, and Gas Compositions

Gasifier Variable	Value	
Gasifier type	BCL	
Temperature	1,598ºF (870ºC)	
Pressure	23 psia (1.6 bar)	
Steam/bone dry feed	0.4 lb/lb	
Sand purge	0.1wt% of circulation rate	
Gas composition	mol% (wet)	mol% (dry)
H_2	12.91	23.85
CO_2	6.93	12.79
CO	22.84	42.18
H_2O	45.87	---
CH_4	8.32	15.36
C_2H_2	0.22	0.41
C_2H_4	2.35	4.35
C_2H_6	0.16	0.29
C_6H_6	0.07	0.13
tar ($C_{10}H_8$)	0.13	0.23
NH_3	0.18	0.32
H_2S	0.04	0.07
Gas yield	0.04 lb-mol of dry gas/lb bone dry feed	
Gas heating value (Btu/lb)	Wet: 4,739 HHV 4,402 LHV Dry: 7,984 HHV 7,417 LHV	
Char yield	0.22 lb/lb bone dry feed	
H_2:CO molar ratio	0.57	
Gasifier efficiency	72.1% HHV basis 71.8% LHV basis	

Note: The gasifier efficiency is defined as the combustion energy of the synthesis gas divided by the combustion energy of the biomass.

Heat for the endothermic gasification reactions is supplied by circulating a hot medium between the gasifier vessel and the char combustor. In this case the medium is synthetic olivine, a calcined magnesium silicate (primarily Enstatite [$MgSiO_3$] Forsterite [Mg_2SiO_3], and Hematite [Fe_2O_3]) used as a sand for various applications.

A small amount of MgO must be added to the fresh olivine to avoid the formation of glass-like bed agglomerations that would result from the biomass potassium interacting with the silicate compounds. The MgO titrates the potassium in the feed ash. Without MgO addition, the potassium will form glass, K_2SiO_4, with the silica in the system. K_2SiO_4 has a low melting point (~930°F) and its formation will cause the bed media to become sticky, agglomerate, and eventually defluidize. Adding MgO makes the potassium form a high melting (~2,370°F) ternary eutectic with the silica, thus sequestering it. Potassium carry over in the gasifier/combustor cyclones is also significantly reduced. The ash content of the feed is assumed to contain 0.2 wt% potassium. The MgO flow rate is set at two times the molar flow rate of potassium.

The gasification medium is steam which is supplied from the steam cycle (7.6 Steam System and Power Generation – Area 600). The steam-to-wood ratio is 0.4 lb of steam/lb of bone dry wood. This variable was tested in the sensitivity analysis. The char combustor temperature is set at 1,800°F and the gasifier temperature is obtained from the energy balance around the gasifier and combustor. The resulting gasifier temperature is 1,598ºF. The gasifier pressure is 23 psia. The olivine circulating flow rate is 27 lb of olivine/lb of bone dry wood. Fresh olivine is made up at a rate of 0.11% of the circulating rate to account for the losses from the cyclones. The combustion air is varied from 5%-20% excess air until the heat duty of the char combustor is zero. The resulting excess air turns out to be 12%.

Particulate removal is performed through cyclone separators. The majority of the olivine and char (99.9% of both) is separated in the primary gasifier cyclone (S-201) and sent to the char combustor. A secondary cyclone (S-202) removes 90% of any residual fines. The char that is formed in the gasifier is burned in the combustor to reheat the olivine. The primary combustor cyclone (S-203) separates the olivine (99.9%) from the combustion gases and the olivine is sent back to the gasifier. Ash and any sand particles that are carried over are removed in the secondary combustor cyclone (99.9% separation in S-204) followed by an electrostatic precipitator (S-205) which removes the remaining residual amount of solid particles. The sand and ash mixture is landfilled but prior to this the solids are cooled and then water is added to the sand/ash stream for conditioning to prevent the mixture from being too dusty to handle. First the ash and sand mixture is cooled to 300°F using the water cooled screw conveyor (M-201) then water is added directly to the mixture until the mixture water content is 10 wt%.

The gas from the secondary gasifier cyclone is sent to the tar reformer (R-203). In this bubbling fluidized bed reactor the following compounds are converted to CO and H_2: CH_4, C_2H_6, C_2H_4, C_6H_6, and C_{10+}; while NH_3 is converted to N_2 and H_2. In the simulation, the percent conversion of each compound is set by the conversion amount that is currently seen in the catalytic tar destruction and heteroatom removal work at NREL. Table 4 gives the conversion that has been experimentally verified from the data gathered at NREL's bench-scale thermo-catalytic conversion system and NREL's Thermochemical Pilot Process Development Unit (TCPDU) (Phillips, *et al*, 2004).

Table 4: Current Design Performance of Tar Reformer

Compound	Percent Conversion to CO & H_2
Methane (CH_4)	20%
Ethane (C_2H_6)	90%
Ethylene (C_2H_4)	50%
Tars (C_{10+})	95%
Benzene (C_6H_6)	70%
Ammonia (NH_3)*	70%

* Converts to N_2 and H_2

In the Aspen Plus simulation, the gas entering the tar reformer is at the gasifier temperature (1,598ºF) and the gas exiting the tar reformer ends up at 1,383°F. The composition of the gas from the tar reformer can be seen in Table 5. Prior to the quench step, the hot gas is cooled to 300°F with heat exchange (H-201 and H-202) that is integrated in the steam cycle (see section 7.6 Steam System and Power Generation – Area 600).

Table 5: Current Design Tar Reformer Properties and Outlet Gas Composition

Tar Reformer Variable	Value	
Tar reformer inlet temperature	1,598ºF (870ºC)	
Tar reformer outlet temperature	1,383ºF (750ºC)	
Tar reformer outlet gas composition	mol% (wet)	mol% (dry)
H_2	33.44	45.52
CO_2	16.10	21.92
CO	16.51	22.47
H_2O	26.54	---
CH_4	6.06	8.25
C_2H_2	0.10	0.14
C_2H_4	1.07	1.46
C_2H_6	0.01	0.02
C_6H_6	0.02	0.03
tar ($C_{10}H_8$)	0.01	0.01
NH_3	0.05	0.07
H_2S	0.04	0.05
N_2	0.06	0.08
Gas heating value (Btu/lb)	Wet: 4,979 HHV 4,485 LHV Dry: 6,711 HHV 6,045 LHV	
H_2:CO molar ratio	2.03	

7.3 Gas Clean Up and Compression – Area 300

After direct cooling of the syngas to a temperature of 300°F additional cooling is carried out via water scrubbing, shown in PFD-P700-A301. The scrubber also removes impurities such as particulates, ammonia, and any residual tars. The scrubbing system consists of a venturi scrubber (M-302) and quench chamber (M-301). The quench water is cooled and recirculated. The quench water flow rate is determined by adjusting the circulation rate until the exit temperature from the cooling water heat exchanger (H-301) is 110°F. The excess scrubber water is sent off site to a waste water treatment facility. This amounts to about 2 gallons per minute excess water for the 2,000 bone dry tonne/day plant. Any solids that settle out in T-301 are sent off-site for treatment as well. For modeling purposes, the water content of the sludge stream was set at 50 wt%. The quench step cools the syngas to a temperature of 140°F. The syngas is then compressed using a five-section centrifugal compressor with interstage cooling as shown in PFD-P700-A302 (K-301A/B/C/D/E, S-301, S-302A/B/C/D/E, S-303, H-302A/B/C/D/E, and H-303). The compressor was modeled such that each section has a polytropic efficiency of 78% along with intercooler temperatures of 140°F.

Sulfur compounds are the main poison of reforming catalysts. Low temperature shift catalysts are also very sensitive to sulfur. Because the syngas exiting the gasifier contains almost 400 ppmv of H_2S, a ZnO bed by itself could not be used for sulfur removal. The normal sulfur concentration at the inlet of a ZnO bed is typically 10-20 ppmv H_2S. The ZnO bed will then reduce the sulfur to less than 1 ppmv H_2S. A very low concentration of less than 1 ppmv H_2S is required for steam reforming and the LTS catalyst. Even at a concentration of 0.1 ppm the reforming catalyst can start to deactivate. Therefore, sulfur removal via a liquid phase oxidation process followed by a ZnO bed was chosen. PFD-P700-A303 shows the sulfur removal step. The LO-CAT process will remove the bulk of the sulfur but it cannot reliably reduce the sulfur concentration to the extremely low levels required by the downstream conversion steps. Therefore, two ZnO beds in series follow the LO-CAT process because the H_2S requirement is so low and a ZnO bed is a simple, relatively inexpensive piece of equipment with a known history for reducing H_2S concentrations to very low levels. Additionally, each ZnO reactor contains a layer of hydrogenation catalyst to convert organic sulfur to H_2S because it is possible that mercaptans, COS, and other sulfur compounds could be present in very small amounts in the syngas stream from the gasifier.

Although, there are several liquid phase oxidation processes available today, the LO-CAT process was selected because of its progress in minimizing catalyst degradation and its environmentally benign catalyst compared to others. LO-CAT is an iron chelate based process that consists of a venturi (M-303), absorber (M-304), oxidizer (R-301), air blower (K-302), solution circulation pump (P-303) and solution cooler (H-305). Elemental sulfur is produced and since there is such a small amount (1.6 tonne/day), it is stockpiled for eventual disposal rather than conditioned and sold. The LO-CAT process was modeled to remove the sulfur to a concentration of 10 ppm H_2S and the ZnO bed removes the remaining sulfur to a concentration of less than 1 ppm. The air flow rate for re-oxidizing the LO-CAT solution was included in the simulation and calculated based on the requirement of 2 mol of O_2 per mol of H_2S. Prior to entering the LO-CAT system the gas stream is superheated to 10°F above dew point (H-304) which in this process is equivalent to 120˚F. This degree of superheating is required for the LO-CAT system. The ZnO bed operates at higher temperatures which are needed so that the reaction (ZnO + H_2S ⇔ ZnS + H_2O) closely approaches equilibrium. Therefore, the gas stream exiting the LO-CAT process is heated to 707°F (H-306) using heat from the steam cycle (see section 7.6 Steam System and Power Generation – Area 600) prior to entering the ZnO reactors (R-302). During operation over a length of time, the reaction zone will gradually move down through the ZnO bed until the bed material finally needs to be changed out.

7.4 Reforming, Shift, and PSA – Area 400

There is a significant amount of CO, CH_4 and other hydrocarbons in the biomass derived syngas (as can be seen in Table 3), thus these components need to undergo conversion via reforming ($C_nH_m + nH_2O$ ⇔ $(n+m/2)H_2 + nCO$) and shift conversion ($CO + H_2O$ ⇔ $CO_2 + H_2$) reactions. The steam reformer is shown in PFD-P700-A401. Reforming and water-gas shift are the main reactions in the steam reformer. The reforming reaction is highly endothermic and is favored by high temperatures and low pressures. The shift reaction is exothermic and favors low temperatures and higher steam ratios. The steam reformer (R-401) is comprised of catalyst-filled tubes, surrounded by a firebox that provides the heat necessary for the endothermic reforming reaction. The main components of the reformer furnace include an air/fuel combustion system, a radiant heat transfer section, and a convection section. The radiant section supplies heat to the catalyst tubes by combusting the air/fuel mixture and the convection section recovers heat by cooling down the flue gases (H-401 and H-404). Reformer furnaces are not very efficient and only about half of the heat in the radiant section is absorbed by the furnace tubes. Generally, the feed gas flows up through the catalyst tubes but the reformer furnace can be side-, terrace-, top-, or bottom-fired (Spath and Dayton, 2003).

Steam reformers typically operate at 1,500-1,600°F and between 218-435 psia using a nickel based catalyst. In this analysis the steam reformer was simulated as an equilibrium reactor at 1,562°F with a -20°F approach temperature, an inlet pressure of 435 psia, and a steam to carbon ratio of 3 mol of H_2O/mol of C (Leiby, 1994). The approach temperature is defined as the difference between the measured outlet temperature and the temperature that would yield the measured conversion of a component at equilibrium (In this case the component is methane.). In Aspen Plus the sign of the approach temperature for this conversion step is negative but other software packages may use a different convention. In this instance, a positive sign would be erroneous resulting in a methane conversion which is higher than that obtained at equilibrium. The steam for the reformer is supplied from the steam cycle (see section 7.6 Steam System and Power Generation – Area 600). The pressure drop through the steam reformer is 30 psi. The reformer is fueled by the PSA offgas and a small amount of natural gas is added for burner control. The amount of natural gas that is added is equal to 10% of the heating value of the PSA offgas. Following the steam reformer, the HTS and LTS reactors convert the majority of the remaining CO, when reacted with H_2O, into CO_2 and H_2 through the water-gas shift reaction. PFD-P700-A402 depicts these shift reactors. The gas exiting the reformer is first cooled to 662°F (H-402) (the operating range of a HTS reactor is typically 570-840°F). The HTS (R-402) and LTS (R-403) were modeled as fixed bed equilibrium reactors with approach temperatures of 35°F and 20°F, respectively, (Leiby, 1994). In this case for the shift conversion reaction the sign convention for the approach temperature in Aspen Plus is positive. In this instance, a negative number would result in more CO being converted than is possible at equilibrium. The gas exiting the HTS reactor is cooled to 392°F (H-405 and H-406) prior to entering the LTS reactor (The LTS reactor typically operates between 350-515°F and often operates near condensation conditions.). The HTS catalyst has an iron oxide, chromium oxide basis while the major component in the LTS catalyst is copper oxide, most often in a mixture with zinc oxide (Spath and Dayton, 2003).

For purification, a PSA unit is used to separate the hydrogen from the other components in the shifted gas stream, mainly CO_2, and unreacted CO, CH_4, and other hydrocarbons. The PSA unit can be seen in PFD-P700-A403. The hydrogen purity achieved from a PSA unit can be greater than 99.99+%. Based on past conversations with industrial gas producers, the shifted gas stream must contain at least 70 mol% hydrogen before it can be economically purified in the PSA unit (Mann, 1995). Purification of streams more dilute than this decreases the product purity and recovery of hydrogen. For this analysis, the concentration of hydrogen in the shifted stream prior to the PSA is between 60-65 mol%. Therefore, part of the PSA hydrogen product stream is recycled back into the PSA feed. For a 70 mol% hydrogen PSA feed, a hydrogen recovery rate of 85% is typical with a product purity of 99.9 vol%. Prior to the PSA unit, entrained liquids (water and condensed hydrocarbons) must be removed because they will permanently damage the adsorbent, which is a mixture of activated carbon and zeolites. Cooling the product and installing a knock out drum with a mist eliminator (S-401 and S-402) prior to the PSA unit is usually sufficient. The PSA efficiency is also affected by adsorption temperature. Fewer impurities are adsorbed at higher temperatures because the equilibrium capacity of the molecular sieves decreases with increasing temperature. Therefore, the design for this analysis uses a heat exchanger integrated with the steam cycle (see section 7.6 Steam System and Power Generation – Area 600) to cool the gas down to its dew point (H-407). The stream is further cooled by an air-cooled heat exchanger (H-408) to 140°F. A cooling water heat exchanger (H-409) is then used to reduce the stream temperature to 110°F.

The minimum pressure ratio between the feed and purge gas of the PSA unit is about 4:1. The absolute pressures of the feed and purge gas are also important in regard to hydrogen recovery. The optimum feed pressure for refinery applications is in the range of 215-415 psia. The purge gas pressure is typically between 17-20 psia to obtain a high recovery of hydrogen. Hydrogen recovery is usually 85-90% at these conditions and drops to 60-80% at high purge gas pressures of 55-95 psia (Leiby, 1994). In the design for this analysis the pressure of the PSA feed gas is 360 psia and the purge gas pressure is 20 psia.

7.5 Hydrogen Compression – Area 500

Ultimately, the hydrogen is sent to a pipeline so the product hydrogen is compressed from 360 psia to 1,015 psia. This is done using a two-stage reciprocating compressor with an isentropic efficiency of 82% and interstage intercooler temperatures of 140°F each (K-501A/B, H-501A/B, S-502, H-502, and S-503). PFD-P700-A500 shows the hydrogen compression step.

7.6 Steam System and Power Generation – Area 600

The process design includes a steam cycle that produces steam via heat recovery of the hot process streams throughout the plant. Because the gasifier and reformer both require steam, power is produced from the steam cycle using an extraction steam turbine/generator (M-602). Steam is supplied to the reformer from the intermediate pressure turbine stage and to the gasifer from the low pressure turbine stage. The steam system and power generation area is shown in PFD-P700-A601, -A602, and -A603.

A condensate collection tank (T-601) gathers condensate from the syngas compressor and from the cooled shifted gas stream prior to the PSA along with the steam turbine condensate and make-up water. The total condensate stream is heated to the saturation temperature and sent to the deaerator (T-603) to de-gas any dissolved gases out of the water. The water from the deaerator is first pumped to a pressure of 1,345 psia and then pre-heated to the saturation temperature using a series of exchangers. The saturated steam is collected in the steam drum (T-604). To prevent solids build up, water must be periodically discharged from the steam drum. The blowdown rate is equal to 2% of water circulation rate. The saturated steam from the steam drum is superheated with another series of exchangers. The superheated steam temperature and pressure were set based on standard conditions given in Perry, *et al*, 1997. Superheated steam enters the intermediate pressure turbine stage at 1,000ºF and1,265 psia and is expanded to a pressure of 450 psia where a slipstream is removed and sent to the steam methane reformer. The remaining steam then enters the low pressure turbine and is expanded to a pressure of 35 psia. Here a slipstream of steam is removed and sent to the gasifier. Finally, the steam enters a condensing turbine and is expanded to a pressure of 1.5 psia. The steam is condensed in the steam turbine condenser (H-601) and re-circulated back to the condensate collection tank.

A pinch analysis was performed to determine the heat integration of the system (see section 10.0 Pinch Analysis for details). Heat integration is an important part of this thermal conversion process. Figure 4 is a drawing that shows the heat exchange network within the steam cycle. The heat duty of the various sections and the heat exchanger tag numbers are given. The figure shows where heat is exchanged between the different steps within the process and the steam cycle but it does not show the integration of the individual heat exchangers. The integration can be seen on the PFDs (Appendix C: Current Design Process Flow Diagrams). In order to close the heat balance of the

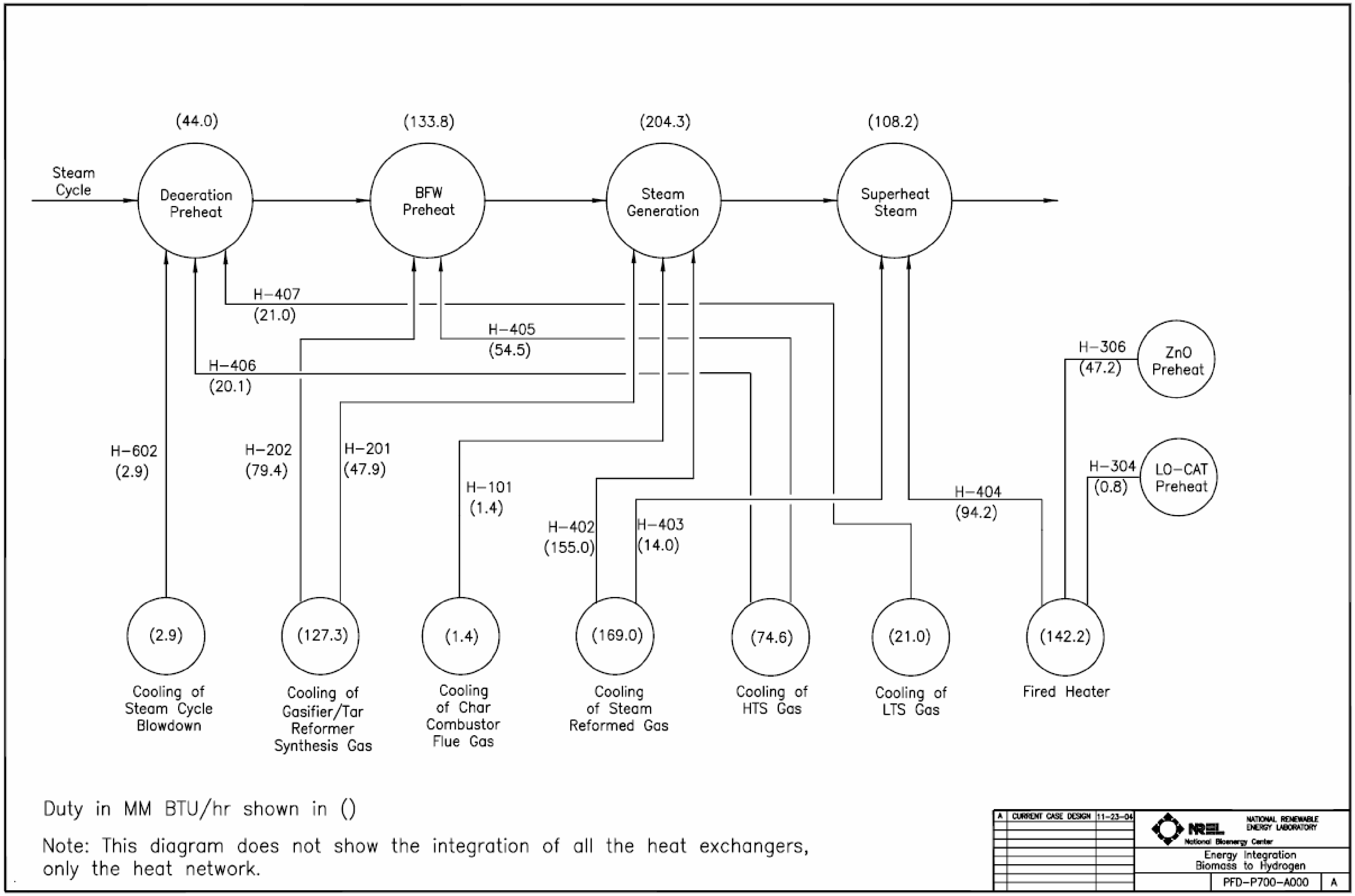

Figure 4: Current Design Heat Exchange Network within the Steam Cycle

system, the Aspen Plus model increases or decreases the water flow rate through the steam cycle until the heat balance of the system is met.

The analysis assumes that all drives for compressors, pumps, fans, etc are electric motors. Additionally, 10% excess power is added to total power requirement to account for miscellaneous usage. Table 6 contains the power requirement of the plant broken out into the different plant sections. Syngas compression accounts for the largest power use. Even though the plant produces power, it is not enough to meet the total electricity demand of the plant. Therefore, the shortage is made up from electricity that is purchased from the grid.

Table 6: Current Design Plant Power Requirement

Plant Section	Power Requirement (kW)
Feed handling & drying	742
Gasification, Tar reforming, & quench	3,636
Compression & sulfur removal	21,871
Steam methane reforming, shift, and PSA	630
Hydrogen compression	3,899
Steam system & power generation	660 required 25,583 generated
Cooling water & other utilities	1,110
Miscellaneous	3,255
Total plant power requirement	35,803
Grid electricity requirement	10,219

7.7 Cooling Water and Other Utilities – Area 700

The cooling water system is shown on PFD-P700-A701. A mechanical draft cooling tower (M-701) provides cooling water to several heat exchangers in the plant. The tower utilizes large fans to force air through circulated water. Heat is transferred from the water to the surrounding air by the transfer of sensible and latent heat. Cooling water is used in the following pieces of equipment:

- the sand/ash cooler (M-201) which cools the sand/ash mixture from the gasifier/combustor
- the quench water recirculation cooler (H-301) which cools the water used in the syngas quench step
- the water-cooled aftercooler (H-303) which follows the syngas compressor and cools the syngas after the last stage of compression
- the LO-CAT absorbent solution cooler (H-305) which cools the solution that circulates between the oxidizer and absorber vessels
- the PSA water-cooled precooler (H-409) which cools the gas in order to condense out any liquids prior to the PSA unit
- the hydrogen compressor water-cooled aftercooler (H-502) which follows the hydrogen compressor and cools the hydrogen after the last stage of compression
- the blowdown water-cooled cooler (H-603) which cools the blowdown stream

- the steam turbine condenser (H-601) which condenses the steam exiting the steam turbine

Make-up water for the cooling tower is supplied at 14.7 psia and 60°F. Water losses include evaporation, drift which is the water entrained in the discharge vapor, and blowdown. Drift losses were estimated to be 0.2% of the water supply. Evaporation losses and blowdown were calculated based on information and equations in Perry, *et al*, 1997. The cooling water supply pressure is 65 psia and the supply temperature is 90°F. The cooling water return temperature is 110°F.

An instrument air system is included to provide compressed air for both service and instruments. The instrument air system is shown on PFD-P700-A701. The system consists of an air compressor (K-701), dryer (S-701) and receiver (T-701). The instrument air is at a pressure of 115 psia, a dew point of -40°F, and oil free.

Other miscellaneous items that are taken into account in the design include:

- a firewater storage tank (T-702) and pump (P-702)
- a diesel tank (T-703) and pump (P-703) to fuel the front loaders
- an olivine truck scale with dump (M-702) and an olivine lock hopper (T-705) as well as an MgO lock hopper (T-706)
- a hydrazine storage tank (T-707) and pump (P-705)

This equipment is shown on PFD-P700-A702.

7.8 Additional Design Information

Table 7 contains some additional information used in the Aspen Plus model and biomass gasification to hydrogen production design.

Table 7: Utility and Miscellaneous Design Information

Item	Design Information
Ambient air conditions [1,2, and 3]	Pressure: 14.7 psia $T_{Dry\ Bulb}$: 90°F $T_{Wet\ Bulb}$: 80°F Composition (mol%): N_2: 75.7% O_2: 20.3% Ar: 0.9% CO_2: 0.03% H_2O: 3.1%
Pressure drop allowance	Syngas compressor intercoolers = 2 psi Heat exchangers and packed beds = 5 psi
Thermodynamics	- VLE: Redlich-Kwong-Soave EOS with Boston-Mathias modification. - Enthalpies for Non-conventional components: Boie correlation for heat of combustion, Kirov correlation for heat capacity. - Steam System: ASME Steam Tables.

(1) In Gas Processors Suppliers Association (2004), see Table 11.4 for typical design values for dry bulb and wet bulb temperature by geography. Selected values would cover summertime conditions for most of lower 48 states.
(2) In Weast (1981), see F-172 for composition of dry air. Nitrogen value adjusted slightly to force mole fraction closure using only N_2, O_2, Ar, and CO_2 as air components.

(3) In Perry, *et al*, (1997), see psychrometric chart, Figure 12-2, for moisture content of air.

8.0 Capital Costs

The following sections discuss the methods and sources for determining the capital cost of each piece of equipment within the plant. A summary of the individual equipment costs for the current design can be found in Appendix H: Current Design Summary of Individual Equipment Costs and a summary of the individual equipment costs for the goal design can be found in Appendix I: Goal Design Summary of Individual Equipment Costs.

Because the majority of the costs came from literature and Questimate® (an equipment capital cost estimating software tool by Aspen Tech) instead of vendor quotes, the purchased cost of the equipment was calculated and then cost factors were used to determine the installed equipment cost. The cost multipliers were taken from Peters and Timmerhaus, 2003. This method of cost estimation has an expected accuracy of roughly + or -30%. The factors used in determining the total installed cost (TIC) of each piece of equipment are shown in Table 8.

Table 8: Cost Factors in Determining Total Installed Equipment Costs

	% of TPEC
Total Purchased Equipment Cost (TPEC)	100
Purchased equipment installation	39
Instrumentation and controls	26
Piping	31
Electrical systems	10
Buildings (including services)	29
Yard improvements	12
Total Installed Cost (TIC)	247

The indirect costs which are the nonmanufacturing fixed-capital investment costs also need to be calculated. These costs were also determined using cost factors taken from Peters and Timmerhaus, 2003. The factors are shown in Table 9 and have been put as percentages in terms of total purchased equipment cost, total installed cost, and total project investment. The total project investment (TPI) is the sum of the total installed cost (TIC) plus the total indirect costs.

Table 9: Cost Factors for Indirect Costs

Indirect Costs	% of TPEC	% of TIC	% of TPI
Engineering	32	13	9
Construction	34	14	9
Legal and contractors fees	23	9	6
Project contingency	37	15	10
Total Indirect Costs	126	51	34

Table 10 gives the TPI results for the base case 2,000 tonne/day plant current and goal case designs. To see the detailed capital costs refer to Appendix H: Current Design Summary of Individual Equipment Costs and Appendix I: Goal Design Summary of Individual Equipment Costs.

Table 10: Current and Goal Design Base Case TPI Results

	Cost 2002 $MM	
	Current Design	**Goal Design**
Total Purchased Equipment Cost (TPEC)	41	39
Purchased equipment installation	16	15
Instrumentation and controls	11	10
Piping	13	12
Electrical systems	4	4
Buildings (including services)	12	11
Yard improvements	5	5
Total Installed Cost (TIC)	102	96
Indirect Costs		
Engineering	13	12
Construction	14	13
Legal and contractors fees	9	9
Project contingency	14	14
Total Indirect Costs	52	49
Total Project Investment (TPI)	154	144

8.1 Feed Handling, Drying, Gasification and Gas Clean Up Capital Costs

The biomass handling and drying costs as well as the gasification and gas clean up costs were obtained from several reports by others that documented detailed design and cost estimates. Some of the reports gave costs for individual pieces of equipment while others lumped the equipment costs into areas. The costs from the reports were amalgamated into (1) feedstock handling and drying and (2) gasification and clean up. Costs from those reports scaled to a 2,000 bone dry tonne/day plant are given in Table 11. The costs are divided into two types of systems: (1) a low pressure indirectly heated gasifier system using the BCL gasifier and (2) a high pressure directly heated gasifier system using the Gas Technology Institute (GTI). Table 12 gives the basic dryer and gasifier design basis for the references. The base case in this analysis uses the average feed handling and drying cost from all of the literature sources and the average gasifier and gas clean up cost for the references using the BCL gasifier. A sensitivity analysis was performed to examine the effects of these varying study costs.

Table 11: Feed Handling & Drying and Gasifier & Gas Clean Up Costs from the Literature Scaled to 2,000 tonne/day plant

Reference	Scaled Feed Handling and Drying Cost $K (2002)	BCL - Scaled Gasifier and Gas Clean Up Cost $K (2002)	GTI - Scaled Gasifier and Gas Clean Up Cost $K (2002)
Breault and Morgan (1992) [(a)]	$15,048	$15,801	---
Dravo Engineering Companies (1987) [(a)]	$14,848	$15,774	---
Weyerhaeuser, *et al,* (2000) [(a)]	$21,241	$24,063	---
Stone & Webster, *et al*, (1995) [(a)]	$25,067	---	$36,232
Wan and Malcolm (1990) [(a)]	$18,947 [(b)] $14,098 [(c)]	$11,289 [(b)] $11,109 [(c)]	---
Weyerhaeuser (1992) [(a)]	$13,468	$10,224	---
Wright and Feinberg (1993) [(a)]	$26,048 – BCL design $21,942 – GTI design	$12,318 - quench [(d)] $26,562 - HGCU [(d)]	$38,605
Craig (1994)	$13,680	---	$48,229
AVERAGE	**$18,840**	**$16,392**	**$41,071**

(a) From detailed design and cost estimates
(b) Estimated from a 200 dry ton/day plant design.
(c) Estimated from a 1,000 dry ton/day plant design.
(d) Two separate gas clean up configurations were examined for the BCL gasifier. HGCU = hot gas clean up.

Table 12: System Design Information for Gasification References

Reference	Feed Handling and Drying	BCL Gasifier and Gas Clean Up	GTI Gasifier and Gas Clean Up
Breault and Morgan (1992)	Rotary dryer	Cyclones, heat exchange & scrubber	---
Dravo Engineering Companies (1987)	Rotary drum dryer	Cyclones, heat exchange & scrubber	---
Weyerhaeuser, *et al,* (2000)	Steam dryer	Cyclones, heat exchange, tar reformer, & scrubber	---
Stone & Webster, *et al*, (1995)	Flue gas dryer	---	Cyclones, heat exchange, & tar reformer
Wan and Malcolm (1990)	Flue gas dryer	Cyclones, heat exchange & scrubber	---
Weyerhaeuser (1992)	Flue gas dryer	Cyclones, heat exchange & scrubber	---
Wright and Feinberg (1993)	Unclear	Quench system – details are not clear Tar reformer system – details are not clear	Heat exchange & solids – removal – details are not clear
Craig (1994)	Rotary drum dryer	---	Cyclones, heat exchange, & tar reformer

8.2 Other Capital Costs

The cost of reactors, heat exchangers, compressors, blowers and pumps were determined using the energy and material balance from the Aspen Plus simulation along with the costing tool Questimate. The following were the sizing criteria.

The reactors (ZnO, HTS, and LTS) were sized based on a gas hourly space velocity (GHSV), where GHSV is measured at standard temperature and pressure, 60°F and 1 atm (Fogler, 1992), and a height to diameter ratio of 2. The GHSV for the HTS and LTS reactor were set at 3,000/hr and 4,000/hr, respectively (typical values given in Kohl and Nielsen, 1997). The GHSV for each ZnO bed was set at 4,000/hr.

The surface area of each heat exchanger was calculated based on the equation $Q = U*A*\Delta Tln$ (where Q is the heat duty, U is the heat transfer coefficient, A is the exchanger surface area, and ΔTln is the log mean temperature difference). Q was taken from the Aspen Plus simulation, U was estimated from literature sources (primarily Perry, *et al*, 1997), and ΔTln was calculated using the temperatures in the Aspen Plus simulation.

The design information including flow rate, operating temperature and pressure for the blowers and compressors were all taken from the Aspen Plus simulation. The cost of the syngas compressor (K-301) includes the cost of the interstage coolers and interstage knock out (K.O.) vessels. However, the cost of the interstage coolers for the hydrogen compressor (K-501) were not included in the Questimate cost estimate. Thus, these items had to be priced out separately.

For the various pieces of equipment, the design temperature is determined to be the operating temperature plus 50°F (Walas, 1988). The design pressure is the higher of the operating pressure plus 25 psi or the operating pressure times 1.1 (Walas, 1988).

The cost of the steam reformer was based on design and cost data in Leiby (1994). The reformer capital cost was determined and scaled based on heat duty. Literature values were also used to determine the capital and operating cost of the PSA unit (Schendel, *et al*, 1983 and Leiby 1994). The cost of the PSA unit was determined based on the hydrogen production rate.

Some of the miscellaneous and balance of plant costs were scaled from information and costs in Aden, *et al*, (2002):

- cooling tower
- plant and instrument air
- steam turbine/generator/condenser package
- deaerator

Appendix G: Equipment Design Parameters and Cost References contains the design parameters and cost references for the various pieces of equipment in the plant.

9.0 Operating Costs

There are two kinds of operating costs: variable and fixed costs. The following sections discuss the operating costs for the biomass gasification to hydrogen production plant including the assumptions and values for these costs.

9.1 Variable Operating Costs

There are many variable operating costs accounted for in this analysis. The variables, information about them, and costs associated with each variable are shown in Table 13.

Table 13: Variable Operating Costs

Variable	Information and Operating Cost
Tar reformer catalyst	To determine the amount of catalyst inventory, the tar reformer was sized for a gas hourly space velocity (GHSV) of 2,000/hr based on the operation of the tar reformer at NREL's TCPDU where GHSV is measured at standard temperature and pressure (Fogler, 1992). Initial fill then a replacement of 1% per day of the total catalyst volume. Price: $4.67/lb (Leiby, 1994)
ZnO, steam reforming and shift catalyst	Initial fill then replaced every 5 years based on typical catalyst lifetime. ZnO catalyst inventory based on GHSV of 4,000/hr. Steam reformer catalyst inventory based on inventory in Leiby, 1994 and the ratio of the heat duty. Shift catalyst inventory based on GHSV of 3,000/hr for HTS and 4,000/hr for LTS (typical values given in Kohl and Nielsen, 1997). Price (all three types): $4.67/lb (Leiby, 1994)
Gasifier bed material	Synthetic olivine and MgO. Delivered to site by truck equipped with self-contained pneumatic unloading equipment. Disposal by landfill. Olivine price: $172.90/ton (Jaekel, 2004) MgO price: $365/ton (Chemical Marketing Reporter, 2004)
Solids disposal cost	Price: $18/ton (Chem Systems Report, 1994)
Electricity	Price: 4.74¢/kWh (SRI, 2003)
Natural gas	Available at required pressure or pressure can be reduced. Temperature: 60°F <u>Pipeline composition (mol%, dry) (Spath and Mann, 2000)</u>: CO_2: 0.5% N_2: 1.1% CH_4: 94.4% C_2H_6: 3.1% C_3H_8: 0.5% i-C_4H_{10}: 0.1% n-C_4H_{10}: 0.1% C_5^+: 0.2% H_2S: 0.0004% Price: $5.28/MMBtu (SRI, 2003)
Diesel fuel	Usage: 10 gallon/hr plant wide use Price: $1.00/gallon (EIA, 2003)
Chemicals	Boiler chemicals – Price: $1.4/lb (Aden *et al*, 2003) Cooling tower chemicals – Price: $1.00/lb (Aden *et al*, 2003) LO-CAT chemicals – Price: $150/tonne of sulfur produced (Graubard, 2004)
Waste Water	The waste water is sent off-site for treatment. Price: $2.07/100ft^3 (East Bay Municipal Utility District, 2004)

9.2 Fixed Operating Costs

Previous biomass gasification studies have not looked at fixed operating costs (i.e. salaries, overhead, maintenance, etc) in detail, therefore little data was available. As a result, the fixed operating costs given in Aden, *et al*, 2002 were used as a starting point to develop fixed costs for the biomass gasification-to-hydrogen production plant. Though hydrogen and ethanol production involve different processes and unit operations, it is reasonable as a first step to assume similar labor requirements because both designs are large-scale biomass conversion processes. However, this may be an area that would benefit from further examination by an engineering and consulting firm.

The fixed operating costs used in this analysis are shown in Table 14 (labor costs) and Table 15 (other fixed costs). They are shown in 2002 U.S. dollars. The following changes in base salaries and number of employees were made compared to those used in the ethanol plant design in Aden, *et al*, 2002.

- Plant manager salary raised from $80,000 to $110,000
- Shift supervisor salary raised from $37,000 to $45,000
- Lab technician salary raised from $25,000 to $35,000
- Maintenance technician salary raised from $28,000 to $40,000
- Shift operators salaries raised from $25,000 to $40,000
- Yard employees salaries raised from $20,000 to $25,000 and number reduced from 32 to 12.
- General manager position eliminated
- Clerks and secretaries salaries raised from $20,000 to $25,000 and number reduced from 5 to 3.

The number of yard employees was changed to reflect a different feedstock and feed handling system compared to Aden, *et al*, 2002. Handling baled stover obviously requires more hands-on processing when compared to a wood chip feedstock. Based on a 4-shift system, 3 yard employees were estimated to be needed, mostly to run the front end loaders. The general manager position was eliminated because a plant manager would likely be sufficient for this type of facility. Biomass gasification plants are more likely to operated by larger companies instead of operating like the dry mill ethanol model of farmer co-ops. Finally, the number of clerks and secretaries was reduced from 5 to 3. The estimate of three comes from needing 1 to handle the trucks and scales entering and leaving the facility, 1 to handle accounting matters, and 1 to answer phones, do administrative work, etc.

Table 14: Labor Costs

Position	Salary	Number	Total Cost
Plant manager	$110,000	1	$110,000
Plant engineer	$65,000	1	$65,000
Maintenance supervisor	$60,000	1	$60,000
Lab manager	$50,000	1	$50,000

Position	Salary	Number	Total Cost
Shift supervisor	$45,000	5	$225,000
Lab technician	$35,000	2	$70,000
Maintenance technician	$40,000	8	$320,000
Shift operators	$40,000	20	$800,000
Yard employees	$25,000	12	$300,000
Clerks & secretaries	$25,000	3	$75,000
Total salaries (2002 $)			$2,0800,000

Since the salaries listed above are not fully loaded (i.e. do not include benefits), a general overhead factor was used. This also covers general plant maintenance, plant security, janitorial services, communications, etc. The 2003 PEP yearbook (SRI, 2003) lists the national average loaded labor rate at $37.66/hr. Using the salaries in Table 14 above along with the 60% general overhead factor from Aden, *et al*, 2002 gave an average loaded labor rate of $30/hr. To more closely match the PEP yearbook average, the overhead factor was raised to 95%. The resulting average loaded labor rate was $36/hr. Factors for maintenance, insurance, and taxes were obtained from Peters and Timmerhaus (2003).

Table 15: Other Fixed Costs

Cost Item	Factor	Cost
General overhead	95% of total salaries	$1,976,000
Maintenance	2% of total project investment	$3,072,500
Insurance & taxes	2% of total project investment	$3,072,500

The updated salaries in Table 14 above were examined against salaries from a free salary estimation tool (BTA, 2004), which uses Bureau of Labor Statistics data and several other sources. Because the biomass analysis does not reflect a specific site in the United States, National Average Salaries for 2003 were used. With such an extensive listing of job titles in the salary estimation tool, a general position such as “clerks and secretaries” could be reflected by multiple job titles. In these instances, care was taken to examine several of the possible job titles that were applicable. A list of the job positions at the biomass-to-hydrogen production plant and the corresponding job titles in the salary estimation tool (BTA, 2004) is shown in Table 16. Overall, the salaries used in the biomass-to-hydrogen production plant design are close to the U.S. national average values given in column 4.

Table 16: Salary Comparison

Job Title in Biomass Plant	Corresponding Job Title in Salary Estimating Tool (BTA 2004)	Salary Range (17th to 67th percentile)	Average Salary (U.S. national average)	Salary used in Biomass Plant Design (see Table 14)
Plant manager	Plant manager (experience)	\$81,042-\$220,409	\$106,900	\$110,000
Plant engineer	Plant engineer	\$36,213-\$66,542	\$58,324	\$65,000
Maintenance supervisor	Maintenance crew supervisor	\$35,036-\$53,099	\$45,191	\$60,000
	Supervisor maintenance	\$34,701-\$56,097	\$47,046	
	Supervisor maintenance & custodians	\$23,087-\$45,374	\$39,924	
Lab manager	Laboratory manager	\$38,697-\$70,985	\$51,487	\$50,000
Shift supervisor	Supervisor production	\$32,008-\$51,745	\$43,395	\$45,000
Lab technician	Laboratory technician	\$25,543-\$41,005	\$34,644	\$35,000
Maintenance technician	Maintenance worker	\$27,967-\$46,754	\$39,595	\$40,000
Shift operators	Operator control room	\$33,983-\$61,362	\$49,243	\$40,000
Yard employees	Operator front end loader	\$24,805-\$39,368	\$31,123	\$25,000
Clerks & secretaries	Administrative clerk	\$19,876-\$25,610	\$26,157	\$25,000
	Secretary	\$20,643-\$31,454	\$26,534	
	Clerk general	\$15,984-\$25,610	\$22,768	

Overall, Aden, *et al*, 2002 lists fixed operating costs totaling \$7.54MM in \$2000. Using the labor indices, this equates to \$7.85MM in \$2002. On the other hand, the hydrogen design report has fixed operating costs totaling \$10.2MM in \$2002, which is \$2.35MM higher.

10.0 Pinch Analysis

A pinch analysis was performed to analyze the energy network of the biomass gasification to hydrogen production process. The pinch technology concept offers a systematic approach to optimum energy integration of the process. First temperature and enthalpy data were gathered for the "hot" process streams (i.e., those that must be cooled), "cold" process streams (i.e., those that must be heated), and utility streams such as steam, flue gas, and cooling water. The minimum approach temperature was set at 50 °F. A temperature versus enthalpy graph known as a composite curve was plotted for the hot and cold process streams. These two curves are shifted so that they touch at the pinch point. From this shifted graph, a grand composite curve is constructed which plots the enthalpy differences between the hot and cold composite curves as a function of temperature. This curve is shown in Figure 5 for the current design. This figure was used to determine the heat exchanger network of the system (Figure 4).

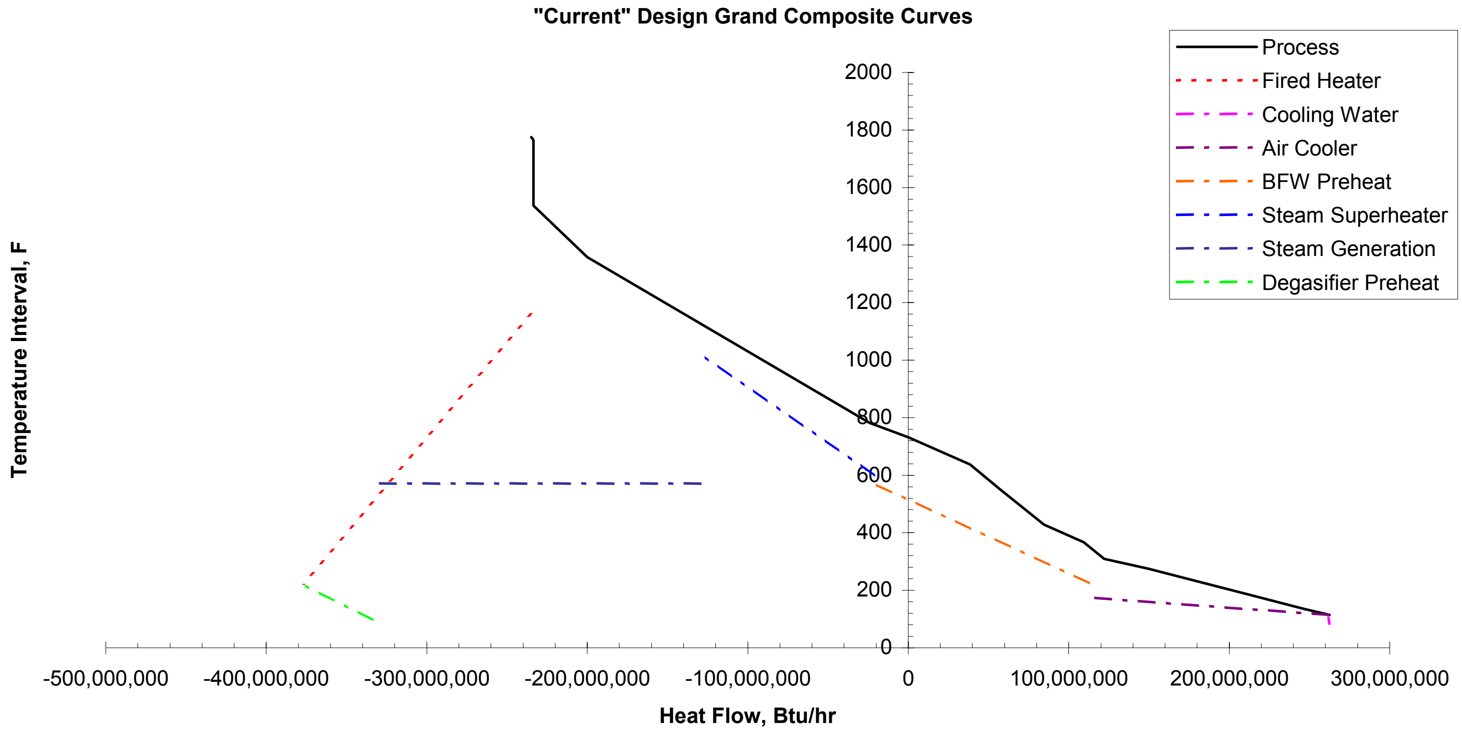

Figure 5: Current Design Grand Composite Curve

11.0 Energy Balance

Because energy integration is so important to the hydrogen production process, understanding how and where the energy is utilized and recovered is key. Detailed energy balances around the major process areas were derived using data from the Aspen Plus simulation. Comparing the process energy inputs and outputs enables the energy efficiency of the process to be quantified. Also, tracing energy transfer between process areas makes it possible to identify areas of potential improvement.

The philosophy of defining the "energy potential" of a stream is somewhat different from what was done for the ethanol process design report (Aden, *et al*, 2002). For that analysis the definition of the energy potential was based upon the higher heating values (HHVs) of each component. This HHV basis is convenient when a process is primarily made up of aqueous streams in the liquid phase. Since liquid water at the standard temperature has a zero HHV, the contributions for any liquid water is very small, especially as compared to any other combustible material also present in the stream. However, the hydrogen production process differs significantly in that most of the process streams are in the gas phase. To remove the background contributions of the water, the energy potential is instead based upon the lower heating values (LHVs) of each component.

The total energy potential for a stream has other contributions beyond that of the heating value. Other energy contributions are:

- Sensible heat effect – the stream is at a temperature (and pressure) different from that of the standard conditions at which the heating values are defined.
- Latent heat effect – one or more components in the stream are in a different phase from that at which their heating values are defined.
- Non-ideal mixing effect – any heating or cooling due to blending dissimilar components in a mixture.

The procedure for actually calculating the energy potential of a stream is also different from what was done for the ethanol process design report. When the ethanol process was analyzed the contributions for the HHVs, sensible heating effects, and the latent heat effects were directly computed and combined. The calculations of the sensible and latent heat effects were done in an approximate manner. For example, the sensible heat effect was estimated from the heat capacity at the stream's temperature, pressure, and composition; it was assumed that this heat capacity remained constant over the temperature range between the stream's temperature and the standard temperature. However, the larger the difference between the stream temperature and the standard temperature, the more likely this assumption is not accurate. Indeed, the hydrogen production process operates at such large temperatures that this would not be an accurate way to account for the sensible heat effect.

The enthalpy values reported by Aspen Plus can actually be adjusted in a fairly simple manner to reflect either an HHV or LHV basis for the energy potential. The enthalpies calculated and reported by Aspen Plus are actually based upon a heat of formation for the energy potential of a stream. So, the reported enthalpies already include the sensible, latent, and non-ideal mixing effects. If certain constants in Aspen's enthalpy expressions could be modified to be based on either the components' HHVs or LHVs instead of the heats of formation then Aspen Plus would report the desired energy potential values. However, since the constants cannot be easily changed, the reported enthalpy values were instead adjusted as part of a spreadsheet calculation. The factors used to adjust the reported enthalpies were calculated from the difference between each component's heat of combustion (LHV) and the reported pure component enthalpy at combustion conditions.

The major process energy inputs and outlets are listed in Table 17, along with their energy flowrates. Each input and output is also ratioed to the dry biomass energy entering the system. The biomass is of course the primary energy input, however other energy inputs are required. Natural gas is used as trim for the steam methane reformer, which is primarily fueled by the PSA offgas. Some electricity must be purchased from the grid to ensure that all power requirements are met. Air is also required for both the steam methane reformer as well as the char combustor, however it remains a minor energy input. Some water is used to wet the ash leaving the gasification system, however, the majority of process water is used for boiler feed water makeup and cooling water makeup. A large negative energy flow value is associated with this because it enters the process as a liquid.

The sum of these energy outlets shown in Table 17 represents greater than 97% of the energy entering the system. The difference (< 3%) is comprised of energy losses due to ambient heating effects and work (pump, compressor) efficiency losses.

Table 17: Current Design Overall Energy Analysis (LHV basis)

	Energy Flow (MMBTU/hr, LHV basis)	**Ratio to Feedstock Energy Flow**
Energy Inlets		
Wood Chip Feedstock (dry)	1480.7	1.000
Feedstock Moisture	-209.7	-0.142
Natural Gas	34.6	0.023
Air	2.4	0.002
Olivine	0.0	0.000
MgO	0.0	0.000
Water	-268.7	-0.182
Tar Reforming Catalyst	0.0	0.000
Purchased Electricity	34.9	0.024
Other	0.0	0.0
Total	**1074**	**0.725**
Energy Outlets		
Hydrogen	737.8	0.498
Cooling Tower Evaporation	26.5	0.018
Flue Gas	57.4	0.039
Sulfur	0.6	0.000
Compressor Heat	119.0	0.080
Heat from Air-cooled Exchanger	149.3	0.101

	Energy Flow (MMBTU/hr, LHV basis)	Ratio to Feedstock Energy Flow
Ash	16.0	0.011
Wastewater	-18.7	-0.013
Other	-41.9	-0.028
Total	**1046**	**0.706**

The only saleable product from this process is hydrogen, but other important energy outlets also exist. There are two sources of flue gas: the char combustor and the steam methane reformer. Together, they total about 4% of the energy in the dried biomass. Cooling tower evaporative losses, wastewater, and ash are also minor energy outlets. However, two of the larger energy outlets come from air-cooled interstage cooling of the compressors, and from the air-cooling of the shifted syngas. Together, these two heat losses represent 18% of the energy that is not recovered within the process. Some of this heat could potentially be recovered using different heat exchange equipment, however it would likely be more expensive on an overall process basis to do so.

The overall energy balance for the current design is depicted graphically in Figure 6. The energy values are listed as percentages of the dry biomass fed to the process. The 50% moisture entering the process within the wood chips has a negative value because it enters as a liquid. The same is also true for the negative values associated with cooling tower and steam cycle makeup water inputs (i.e. a latent heat "penalty").

Not all energy flows are shown within the context of this diagram. For example, the energy flows around the tar reforming and scrubbing section don't appear to balance only because various integrated small streams are not shown in Figure 6. Crude syngas (83.3%) enters the section while wastewater (-1.3%), scrubbed syngas (73.4%), and cooling tower heat (2.6%) all exits. Thus there is a difference of 8.6% which is the heat going to the steam cycle that gets redistributed throughout the process. This heat integration does not appear directly on the diagram. This is also true for many of the other process areas. The heat integration, though not shown here, is depicted in an earlier diagram (Figure 4).

It is also important to note that the 49.8% value listed for the hydrogen product should not be taken as the process efficiency. Instead, the summary sheet in Appendix A shows the hydrogen efficiency to be 45.6%. Remember that all energy inputs including electricity and natural gas must be factored into the process efficiency calculation even though these inputs are small.

For comparison, the energy balance was also calculated on a HHV basis. This is shown in Figure 7. Some of the water streams are slightly negative due to the sensible heat effect.

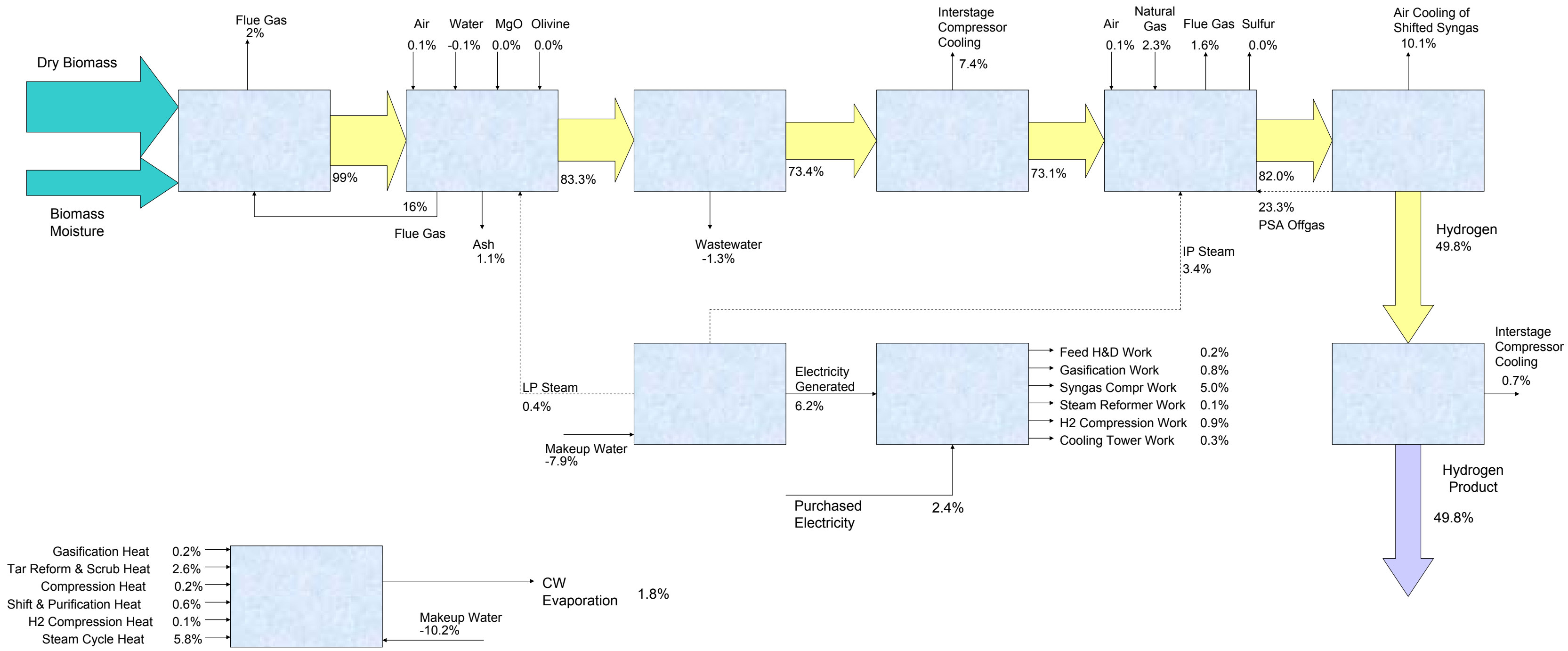

Figure 6: Current Design Process Energy Balance (LHV Basis)

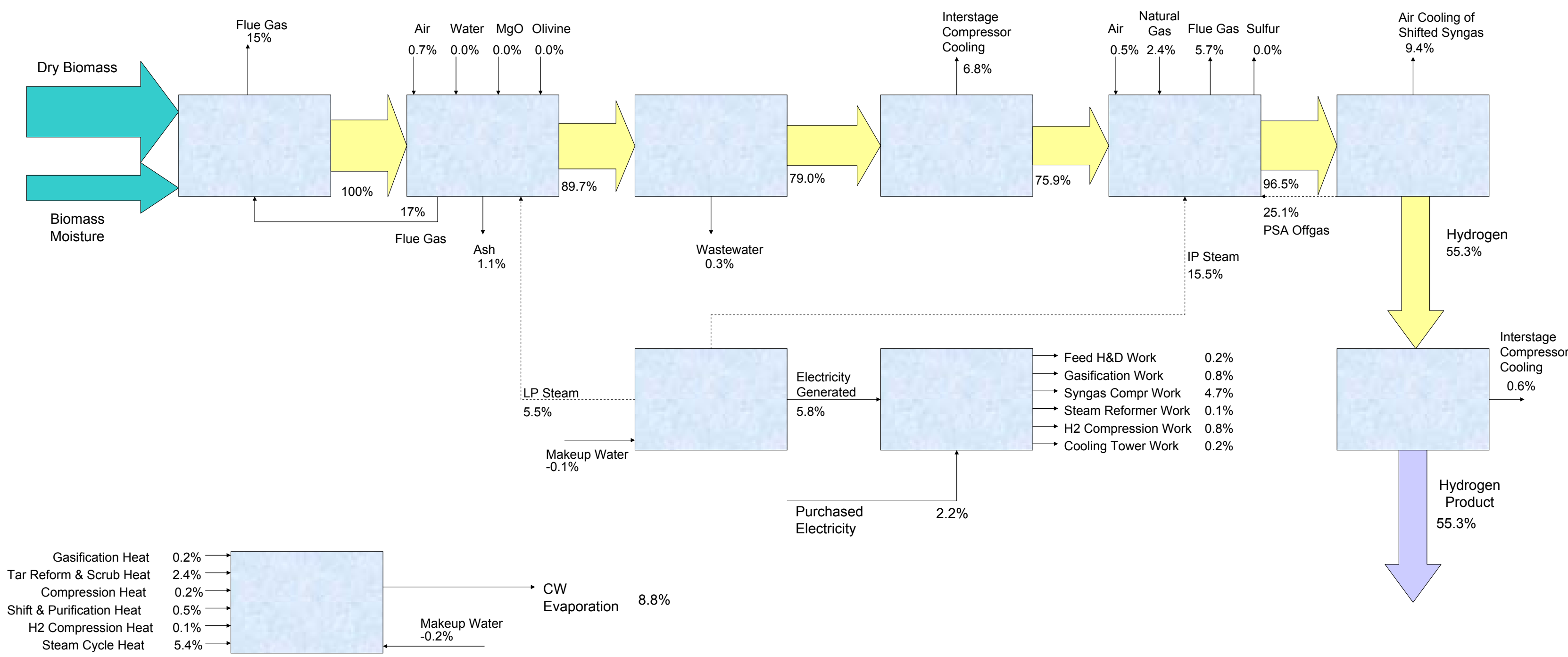

Figure 7: Current Design Process Energy Balance (HHV Basis)

12.0 Design, Modeling, and Capital Cost Changes for Goal Design

The performance goals for the catalytic tar destruction and heteroatom removal work are shown in Table 18. Because the methane conversion is much higher than that for the current design (see Table 2), the process design was changed to eliminate the steam methane reformer. See Figure 3 for the block flow diagram and Appendix D: Goal Design Process Flow Diagrams for the PFDs. The main difference in the capital costing included the deletion of the steam methane reformer cost and the addition of a catalyst regenerator system and some cyclones. The heat for the reactor/regenerator system is supplied by combusting the PSA offgas in the regenerator vessel along with natural gas in order to operate the system isothermally. A breakdown of the capital costs for the goal design can be found in Appendix I: Goal Design Summary of Individual Equipment Costs. The rolled up TPI results were given previously in Table 10.

Table 18: Goal Design Performance of Tar Reformer

Compound	**Percent Conversion to CO & H_2**
Methane (CH_4)	80%
Ethane (C_2H_6)	99%
Ethylene (C_2H_4)	90%
Tars (C_{10+})	99.9%
Benzene (C_6H_6)	99%
Ammonia (NH_3)*	90%

* Converts to N_2 and H_2

Table 19 shows the operating parameters and outlet gas composition of the tar reformer for the goal design. More methane and higher hydrocarbons are reformed producing more hydrogen and carbon monoxide. The carbon monoxide is shifted to hydrogen after the sulfur removal step.

Table 19: Goal Design Tar Reformer Properties and Outlet Gas Composition

Tar reformer Variable	Value	
Tar reformer inlet temperature	1,598ºF (870ºC)	
Tar reformer outlet temperature	1,598ºF (870ºC)	
Tar reformer outlet gas composition	mol% (wet)	mol% (dry)
H_2	41.62	53.18
CO_2	10.40	13.29
CO	24.58	31.40
H_2O	21.73	---
CH_4	1.35	1.73
C_2H_2	0.02	.02
C_2H_4	0.19	0.24
C_2H_6	0.001	0.002
C_6H_6	0.0006	0.0007
tar ($C_{10}H_8$)	0.0001	0.0001
NH_3	0.01	0.02
H_2S	0.03	0.04
N_2	0.06	0.08
Gas heating value (Btu/lb)	Wet: 5,311 HHV Dry: 6,960 HHV	4,794 LHV 6,282 LHV
H_2:CO molar ratio	1.69	

A breakdown of the power requirement for the goal design is given in Table 20. Again, this process design produces power but not enough to supply the electricity requirement of the plant.

Table 20: Goal Design Plant Power Requirement

Plant Section	Power Requirement (kW)
Feed handling & drying	742
Gasification, Tar reforming/regeneration, & quench	3,636
Compression & sulfur removal	26,058
Shift, and PSA	159
Hydrogen compression	4,190
Steam system & power generation	662 required 29,974 generated
Cooling water & other utilities	1,152
Miscellaneous	3,660
Total plant power requirement	40,259
Grid electricity requirement	10,284

The heat integration of the system was reconfigured from the current design case. The resulting heat exchange network and pinch analysis for the goal design can be seen in Figure 8 and Figure 9, respectively. Additionally, the goal design energy balance on a LHV basis can be seen in Figure 10.

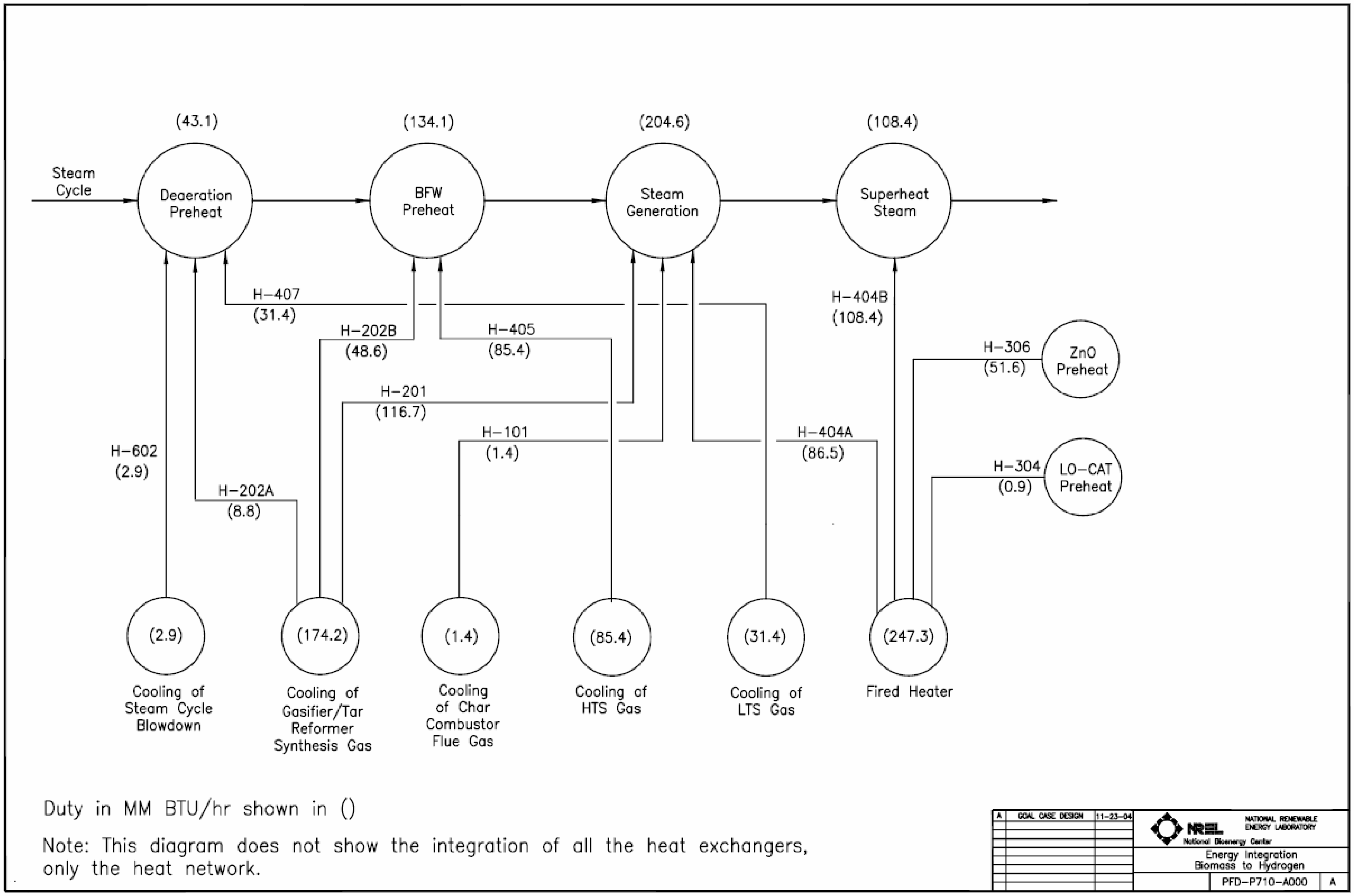

Figure 8: Goal Design Heat Exchange Network within the Steam Cycle

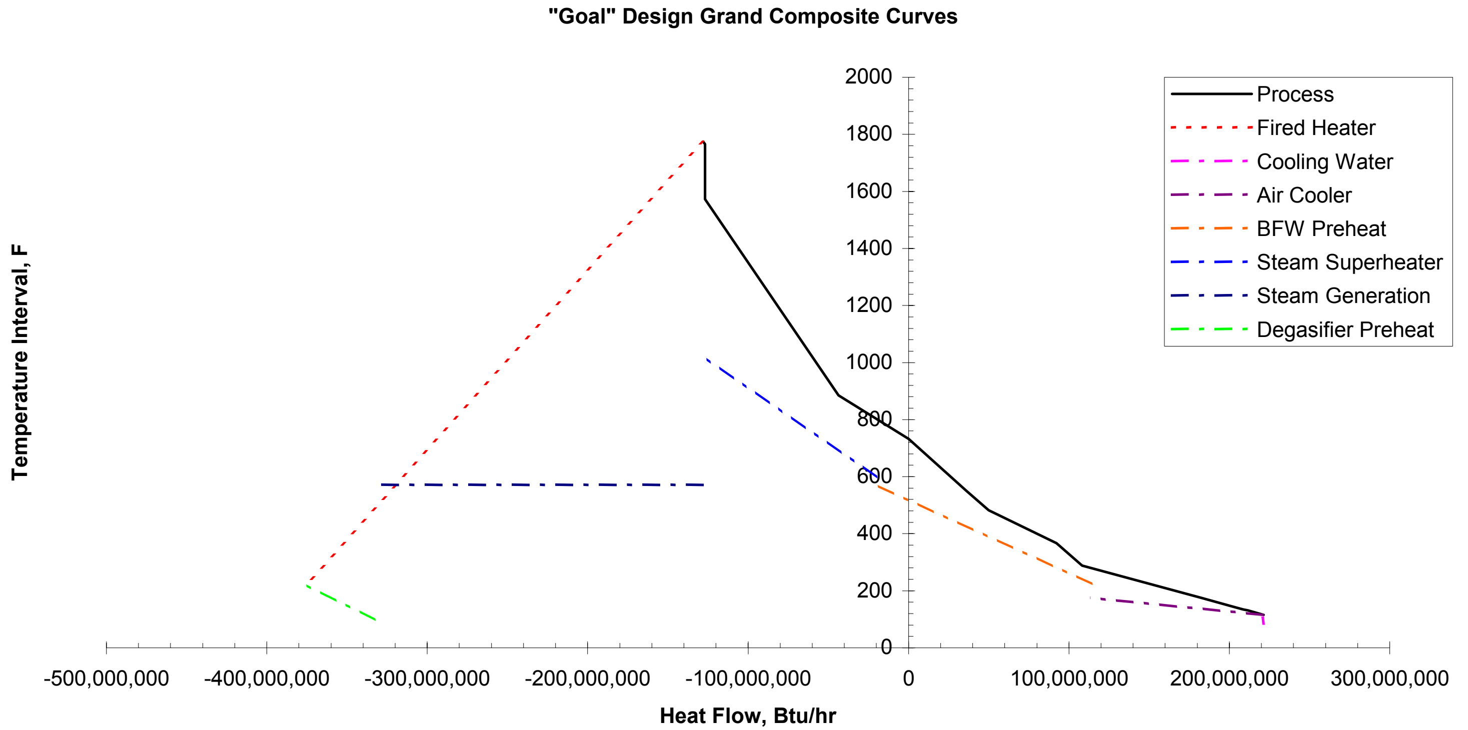

Figure 9: Goal Design Grand Composite Curve

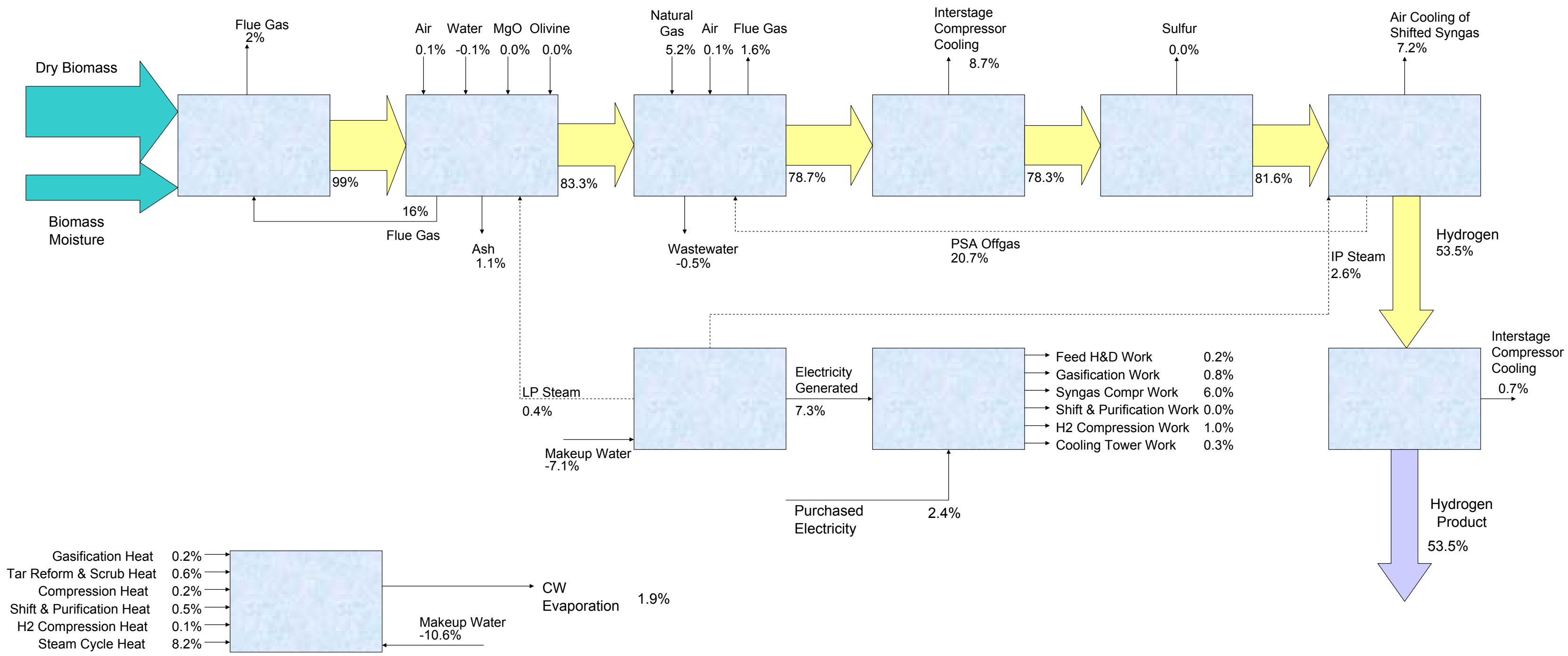

Figure 10: Goal Design Process Energy Balance (LHV Basis)

13.0 Resulting Economics of Current Design

Once the capital and operating costs have been determined, a minimum hydrogen selling price (MHSP) can be determined using a discounted cash flow rate of return analysis. The methodology used is identical to that used in Aden, *et al,* (2002). The MHSP is the selling price of hydrogen that makes the net present value of the biomass syngas to hydrogen process equal to zero with a 10% discounted cash flow rate of return over a 20 year plant life. An Excel worksheet was set up and some of the base case economic parameters used in the spreadsheet are given in Table 21. A sensitivity analysis was performed to examine the minimum hydrogen selling price for different debt/equity ratios at different internal rates of return (section 14.0 Current Design Sensitivity Analyses and section 16.0 Goal Design Sensitivity Analyses).

Table 21: Economic Parameters

Assumption	Value
Internal rate of return (after-tax)	10%
Debt/equity	0%/100%
Plant life	20 years
General plant depreciation	200% DDB
General plant recovery period	7 years
Steam plant depreciation	150% DDB
Steam plant recovery period	20 years
Construction period 1st 6 monts expenditures Next 12 months expenditures Last 12 months expenditures	2.5 years 8% 60% 32%
Start-up time Revenues Variable costs Fixed costs	6 months 50% 75% 100%
Working capital	5% of Total Capital Investment
Land	6% of Total Purchased Equipment Cost (Cost taken as an expense in the 1st construction year)

Note: The depreciation amount was determined using the same method as that documented in Aden, *et al*, 2002 using the IRS Modified Accelerated Cost Recovery System (MACRS).

The resulting minimum hydrogen selling price for the current design is $1.38/kg ($11.48/GJ, LHV) for a 2,000 bone dry tonne/day plant. A summary sheet of the capital and operating costs for the base case can be found in Appendix A: Current and Goal Base Case Summary Sheets.

Figure 11 illustrates the cost contribution to product price for feedstock, capital, and operating costs by process area for this biomass gasification to hydrogen production process. Both percentages and contribution in terms of $/kg of hydrogen are given. The feedstock cost contributes the most to the product hydrogen price (31%). This is followed by gasification, tar reforming, and quench at 20%, compression and sulfur removal also at 20%, and steam reforming, shift, and hydrogen purification at 18%. Although the system produces power, it does not produce enough to meet the plant's internal power requirements. The steam cycle generates almost 26 MW of power but the plant requires almost 36 MW of power, largely due to the syngas compression requirement. Thus 10 MW of power is purchased from the grid.

14.0 Current Design Sensitivity Analyses

Many sensitivity cases were run to examine the effects of several parameters on the current base case design Table 22 outlines the different sensitivity cases that were examined. Table 23 contains the results for the sensitivity analysis and Figure 12 shows the results in Table 23 graphically. Internal rate of return (IRR) and debt equity ratio were also examined. When a percentage of the financing is debt, the loan interest rate was set at 7.5% with a loan term of 10 years. Figure 13 is a graph showing those results and how the minimum hydrogen selling price changes with different combinations of IRR and debt/equity.

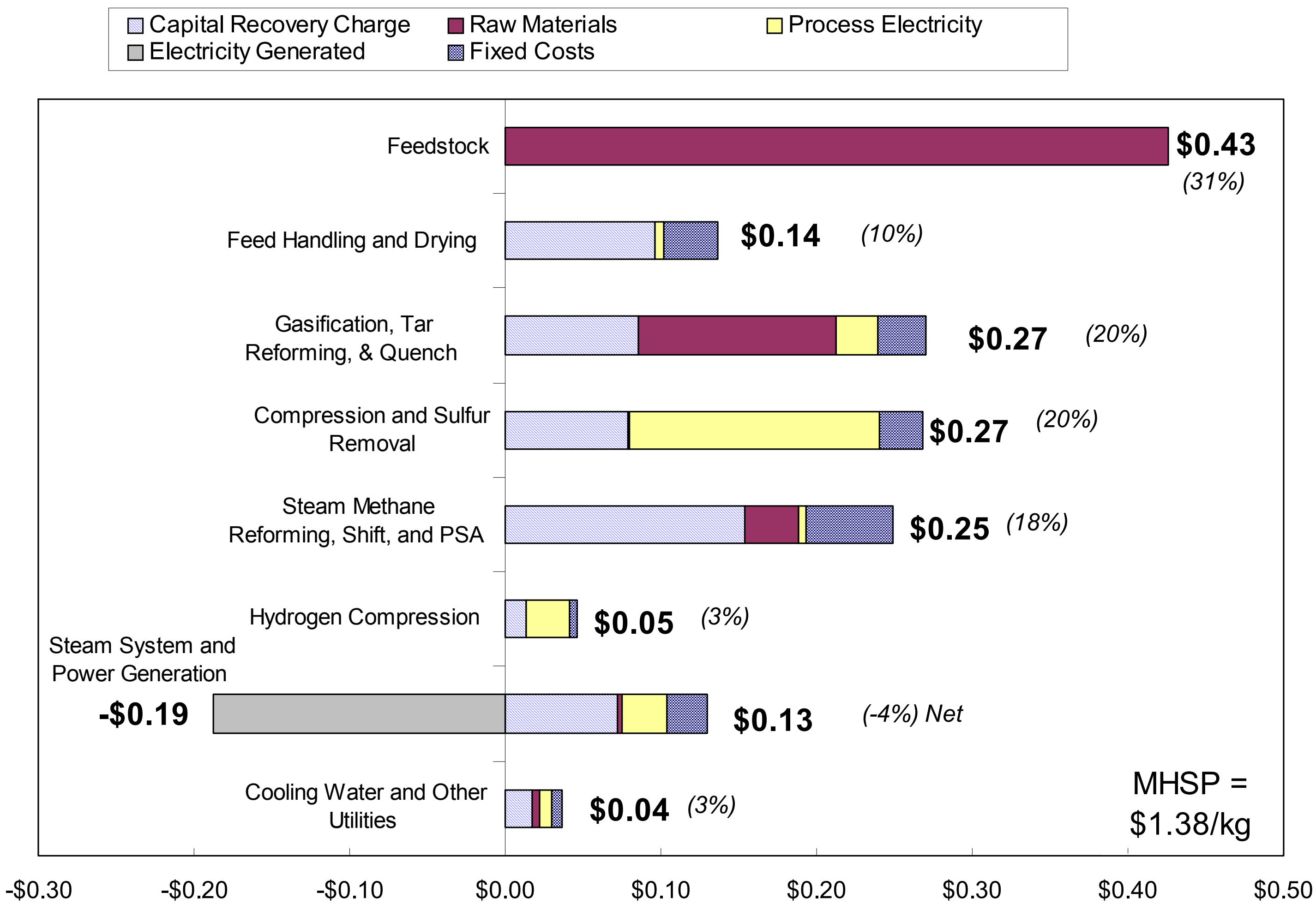

Figure 11: Current Design Base Case Cost Contribution Diagram

Table 22: Current Design - Sensitivity Analysis Cases

Letter	Sensitivity Case	Analysis Changes Made
A	Decrease feedstock cost to $0/dry ton	The feedstock cost in the DCFROR spreadsheet was changed from $30/dry ton to $0/dry ton.
B	Increase feedstock cost to $53/dry ton	The feedstock cost in the DCFROR spreadsheet was changed from $30/dry ton to $53/dry ton.
C	Lower feed moisture content of 30 wt%	The feed moisture content in the Aspen Plus model was decreased from 50 wt% to 30 wt%.
D	Less drying of biomass feed to a moisture content of 20 wt%	The wood moisture content at the dryer outlet was changed from 12 wt% to 20 wt%. The gasifier temperature dropped from 870ºC (1,598ºF) to 859ºC (1,578ºF). No additional natural gas was required to maintain the heat balance around the gasifier and combustor (enough additional char was produced at the lower gasifier temperature). The dryer cost decreased.
E	Less drying of biomass feed to a moisture content of 20 wt% and keep the gasifier temperature constant	The wood moisture content at the dryer outlet was changed from 12 wt% to 20 wt%. The olivine circulating between the gasifier and combustor had to be increased by a factor of 1.12 times the base case flow to maintain a gasifer temperature of 870ºC (1,598ºF). Natural gas at a rate of 1,709 lb/hr was added to the combustor in order to maintain the heat balance around the gasifier and combustor. The dryer cost decreased and the gasifier/combustor cost increased.
F	Less drying of biomass feed to a moisture content of 30 wt%	The wood moisture content at the dryer outlet was changed from 12 wt% to 30 wt%. The gasifier temperature dropped from 859ºC (1,598ºF) to 870ºC (1,547ºF). Natural gas at a rate of 3,417 lb/hr was added to the combustor in order to maintain the heat balance around the gasifier and combustor. The dryer cost decreased.
G	Less drying of biomass feed to a moisture content of 30 wt% and keep the gasifier temperature constant	The wood moisture content at the dryer outlet was changed from 12 wt% to 30 wt%. Olivine circulating between the gasifier and combustor increased by a factor of 1.3 times the base case flow to maintain a gasifer temperature of 870ºC (1,598ºF). Natural gas at a rate of 8,543 lb/hr was added to the combustor in order to maintain the heat balance around the gasifier and combustor. The dryer cost decreased and the gasifier/combustor cost increased.
H	No dryer	The dryer was removed from the Aspen Plus model. The olivine circulating between the gasifier and combustor had to be increased by a factor of 1.9 times the base case flow to maintain a gasifer temperature of 870ºC (1,598ºF). Natural gas at a rate of 23,920 lb/hr was added to the combustor in order to maintain the heat balance around the gasifier and combustor. The dryer cost was eliminated. The gasifier/combustor cost increased. There is a net power generation of 34 MW from the system instead of a deficiency of 10 MW which had to be purchased from the grid for the base case.
I	Lower gasifier steam:wood ratio of 0.1 and keep the gasifier temperature constant	The steam:wood ratio to the gasifier was decreased from 0.4 to 0.1. This lower rate was based on the operation of the gasifier at Burlington, Vermont during sustained operation and testing for this demonstration project (Overend, 2004). The olivine circulating between the gasifier and combustor was decreased by a factor of 0.87 times the base case rate to maintain a gasifer temperature of 870ºC (1,598ºF). The gasifier/combustor cost decreased.
J	Higher gasifier steam:wood ratio of 1	The steam:wood ratio to the gasifier was increased from 0.4 to 1. The olivine circulation rate was kept the same as the base case and thus the gasifier temperature decreased from 870ºC (1,598ºF) to 847ºC (1,557ºF). Natural gas at a rate of 1,709 lb/hr was added to the combustor in order to maintain an energy balance around the gasifier and combustor. The gasifier/combustor cost increased.

Letter	Sensitivity Case	Analysis Changes Made
K	Higher gasifier steam:wood ratio of 1 and keep the gasifier temperature constant	The steam:wood ratio to the gasifier was increased from 0.4 to 1. Typically, direct gasifiers operate at a steam:wood ratio closer to 1. However, this rate was tested here to determine the effects on the indirect gasifier system. The olivine circulating between the gasifier and combustor had to be increased by a factor of 1.25 times the base case rate to maintain a gasifer temperature of 870ºC (1,598ºF). Natural gas at a rate of 5,467 lb/hr was added to the combustor in order to maintain an energy balance around the gasifier and combustor. The gasifier/combustor cost increased.
L	No H2 recycle to PSA	The recycling of hydrogen to the PSA feed was eliminated.
M	Eliminate LTS	The LTS was removed from the Aspen Plus model. The LTS cost was eliminated.
N	Lower tar reformer catalyst replacement	The tar reformer catalyst replacement was lowered from 1 vol% to 0.5 vol%.
O	Treat waste water internally	Instead of sending the waste water stream off-site for treatment. A reverse osmosis system was installed at the plant. The waste water was cleaned and sent to the steam cycle.
P	Increase in PSA cost	There is some variability in the capital cost data for the PSA so the cost was increased by a factor of 1.6 to determine the sensitivity to this parameter. This factor was determined using two different costing methods for the PSA. One was based on the hydrogen production rate and the other was based on the inlet flow rate to the PSA.
Q	Increase in steam reforming cost	There is some variability in the capital cost data for the steam reformer so the cost was increased by a factor of 2 to determine the sensitivity to this parameter. The cost of the steam reformer was based on the duty but there could be some deviation from a standard steam methane reformer because the stream being reformed contains a low concentration of methane.
R	Increase in electricity price to 6¢/kWh	The electricity price in the DCFROR spreadsheet was changed from 4.74¢/kWh to 6¢/kWh.
S	Increase in natural gas price to $7/MMBtu	The natural gas cost in the DCFROR spreadsheet was changed from $5.28/MMBtu to $7/MMBtu.
T	Lower feed handling & drying capital cost	The feed handling and drying cost was reduced from the average cost in Table 11 to the second lowest cost in Table 11.
U	Lower gasification & clean up capital cost	The gasification and gas clean up cost was reduced from the average cost in Table 11 to the second lowest cost in Table 11.
V	Combined lower feed handling & drying and lower gasification & clean up capital cost	Both the feed handling and drying cost and the gasification and gas clean up cost were reduced to the second lowest cost in Table 11.
W	Higher feed handling & drying capital cost	The feed handling and drying cost was increased from the average cost in Table 11 to the second highest cost in Table 11.
X	Higher gasification & clean up capital cost	The gasification and gas clean up cost was increased from the average cost in Table 11 to the second highest cost in Table 11.
Y	Combined higher feed handling & drying and higher gasification & clean up capital cost	Both the feed handling and drying cost and the gasification and gas clean up cost were increased to the second highest cost in Table 11.

Table 23: Current Design - Base Case and Sensitivity Analysis Results

Letter	Sensitivity Case	Minimum Hydrogen Selling Price ($/kg)	Minimum Hydrogen Selling Price ($/GJ, LHV)
Base	*Current design - base case*	*$1.38*	*$11.48*
A	Decrease feedstock cost to $0/dry ton	$0.94	$7.86
B	Increase feedstock cost to $53/dry ton	$1.71	$14.24
C	Lower feed moisture content of 30 wt%	$1.31	$10.89
D	Less drying of biomass feed to a moisture content of 20 wt%	$1.37	$11.44
E	Less drying of biomass feed to a moisture content of 20 wt% and keep the gasifier temperature constant	$1.39	$11.59
F	Less drying of biomass feed to a moisture content of 30 wt%	$1.46	$12.20
G	Less drying of biomass feed to a moisture content of 30 wt% and keep the gasifier temperature constant	$1.50	$12.50
H	No dryer	$1.78	$14.85
I	Lower gasifier steam:wood ratio of 0.1 and keep the gasifier temperature constant	$1.30	$10.87
J	Higher gasifier steam:wood ratio of 1	$1.57	$13.07
K	Higher gasifier steam:wood ratio of 1 and keep the gasifier temperature constant	$1.58	$13.19
L	No hydrogen recycle to PSA	$1.30	$10.87
M	Eliminate LTS	$1.47	$12.23
N	Lower tar reformer catalyst replacement of 0.5 vol%	$1.35	$11.27
O	Treat waste water internally	$1.38	$11.49
P	Increase in PSA cost	$1.42	$11.82
Q	Increase in steam reforming cost	$1.45	$12.07
R	Increase in electricity price to 6¢/kWh	$1.40	$11.64
S	Increase in natural gas price to $7/MMBtu	$1.39	$11.55
T	Lower feed handling & drying capital cost	$1.35	$11.24
U	Lower gasification & clean up capital cost	$1.35	$11.22
V	Combined lower feed handling & drying and lower gasification & clean up capital cost	$1.32	$10.99
W	Higher feed handling & drying capital cost	$1.41	$11.78
X	Higher gasification & clean up capital cost	$1.42	$11.85
Y	Combined higher feed handling & drying and higher gasification & clean up capital cost	$1.46	$12.15

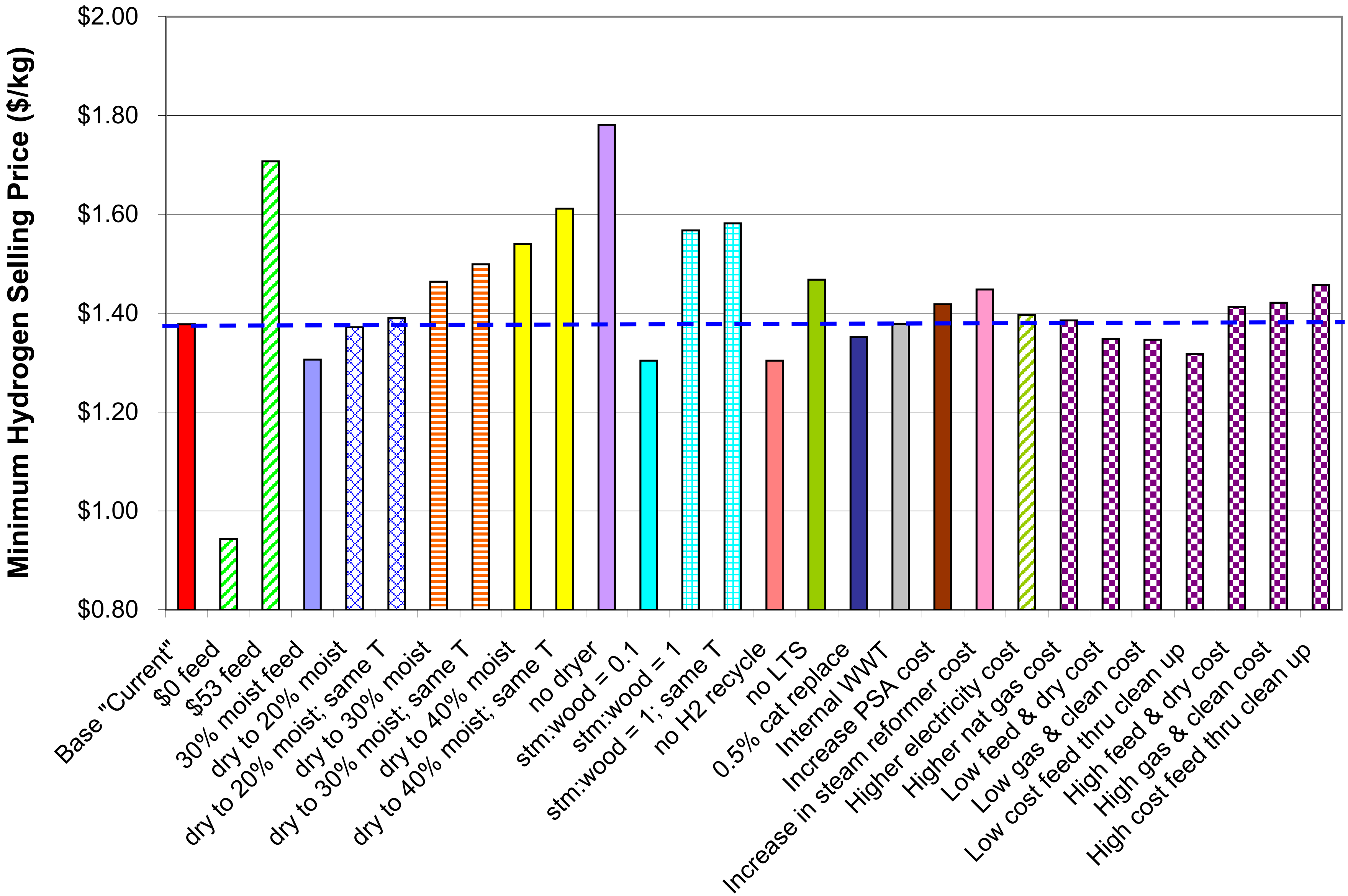

Figure 12: Current Design Sensitivity Analysis Results

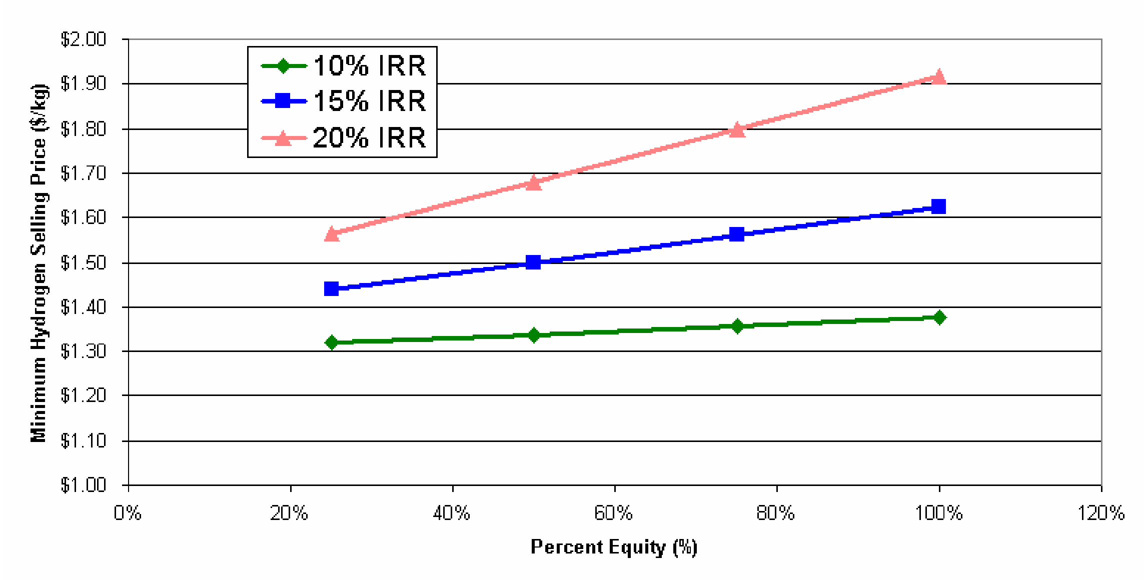

Figure 13: Effect of IRR and Debt/Equity on Current Design Base Case

Since the feedstock cost contributes a large percentage to the hydrogen selling price, the zero feedstock case (A) results in the lowest hydrogen price. Conversely, increasing the feedstock cost (B) adversely affects the minimum hydrogen selling price.

The no dryer case (H) results in the highest hydrogen selling price. In this case, eliminating the dryer eliminates the capital cost for that piece of equipment. Additionally, because there is excess high temperature heat available that would have been used for drying, this case results in more electricity being produced than consumed by the plant. However, the size and thus cost of the gasifier/combustor system increases and the amount of natural gas that must be added to the combustor is significant resulting in a hydrogen selling price that is higher than the base case.

Significantly increasing or decreasing the gasifier steam to wood ratio (I, J, and K) has a large affect on the minimum hydrogen selling price. This variable greatly affects on the heat balance of the system and the capital and operating costs.

Feeding a lower moisture feedstock (C) also affects the heat balance, thus resulting in a decrease in the hydrogen price. More heat is available for power production.

Less drying of the biomass (i.e., a higher moisture content biomass exiting the dryer) was also examined in the sensitivity analysis. Instead of drying to 12%, the biomass was dried to a moisture content of 20% in two cases (D and E) and to a moisture content of 30% in two other cases (F and G). Although less drying affects the heat balance of the system, drying to a moisture content of 20% (D and E) resulted in virtually the same hydrogen selling price as the base case. For the case where the gasifier temperature is kept constant (E), the hydrogen price does not decrease from the base case because there is an increase in operating costs (natural gas must be added to the combustor) even though the total project investment decreases slightly. For the case where the gasifier temperature is reduced (D), the hydrogen yield decreases and there is a decrease in the total project investment. However, drying to a moisture content of 30% did increase the minimum hydrogen selling price (F and G). This is due to decreased hydrogen yields and increased operating costs (natural gas) in both cases (F and G) and an increase in the total project investment for the constant gasifier temperature case (G). It should be noted that both of these cases (F and G) did generate more electricity than what was required for the plant.

A general observation can be made about the differences between sensitivity case D and E, between sensitivity case F and G, and between sensitivity case J and K. In all three of these instances lowering the gasifier temperature decreases the hydrogen yield but adding natural gas to the combustor along with increasing the olivine circulating rate will increase the gasifier temperature. However, the increase in operating cost coupled with any capital cost increases for case E, G, and K is slightly more detrimental than the lower hydrogen yield for case D, F, and J.

The case of eliminating the LTS reactor (M) was examined because often plants with PSA units will use only a HTS reactor followed by a PSA. This is because the PSA can easily remove CO and other components to produce a high purity hydrogen stream. Eliminating the LTS reactor (M) increases the hydrogen price because of a reduction in hydrogen yield that is not recovered by the increase in electricity produced. The LTS reactor is a low capital cost item. Although the PSA can easily remove CO and other components to produce a high purity hydrogen stream, in this case, it is more economical to leave the LTS reactor in.

Assuming a hydrogen recovery rate of 85% without recycling a portion of the product hydrogen to the inlet of the PSA (L) results in a higher hydrogen yield and thus a lower minimum hydrogen selling price. Although increasing the PSA cost (P) did increase the hydrogen price it did not have as large of an effect as the no hydrogen recycling case.

Increasing the steam reformer cost (Q) increased the minimum hydrogen selling price. This capital cost along with the PSA capital cost are items where vendor quotes would reduce the uncertainty in these larger capital cost items.

Because the feed handling and drying costs as well as the gasification and gas clean up costs came from cost data in other detailed studies there is a larger amount of uncertainty as to the exact costs that should be used in this process design. Therefore, several sensitivity cases were run for lower and higher capital costs for the feed handling and drying section and for the gasification and gas clean up section. Overall, decreasing the costs to the second lowest cost from the various studies (T and U) reduced the minimum hydrogen selling price but not significantly, only about 2%. Additionally, increasing the costs to the second highest cost (W and X) did not increase the hydrogen price considerably, only about 3%. A combination of increasing and decreasing the capital cost for both the feed handling and drying section and the gasification and gas clean up section was also tested (Y). This had a larger effect on the change in the minimum hydrogen selling price. The price decreased from \$1.38/kg to \$1.32/kg for the low capital cost case (V) and the price increased from \$1.38/kg to \$1.46/kg for the high capital cost case (Y).

Treating the waste water stream internally (O) had virtually no effect on the overall economics. Three cases that had very little effect on the minimum hydrogen selling price are decreasing the amount of tar reformer catalyst that must be replaced (N), increasing the electricity price (R), and increasing the natural gas price (S). This is because all of these items contribute a small amount to the overall operating cost.

15.0 Resulting Economics of Goal Design

The resulting minimum hydrogen selling price for the goal design is \$1.24/kg (\$10.34/GJ, LHV) for a 2,000 bone dry tonne/day plant. The hydrogen price decreases from the current base case design (which is \$1.38/kg or \$11.48/GJ, LHV) mainly because of an increase in the hydrogen yield. The decrease in the total project investment has some effect. A summary sheet of the capital and operating costs for the base case can be found in Appendix A: Current and Goal Base Case Summary Sheets. The cost contribution to product price for feedstock, capital, and operating costs by process area for the goal design can be seen in Figure 14. Both percentages and contribution in terms of \$/kg of hydrogen are given. Again, the feedstock cost contributes the most to the product hydrogen price (32%) and although the system produces power, it does not produce enough to meet the plant's internal power requirements. Comparing the cost contribution of the goal design (Figure 14) with that for the current design (Figure 11) shows an increase in the gasification/tar reforming/regeneration/quench bar and a decrease in the shift/PSA bar. This happens because the capital and operating costs associated with the steam methane reformer are removed. However, there are capital and operating costs associated with adding the tar catalyst regenerator.

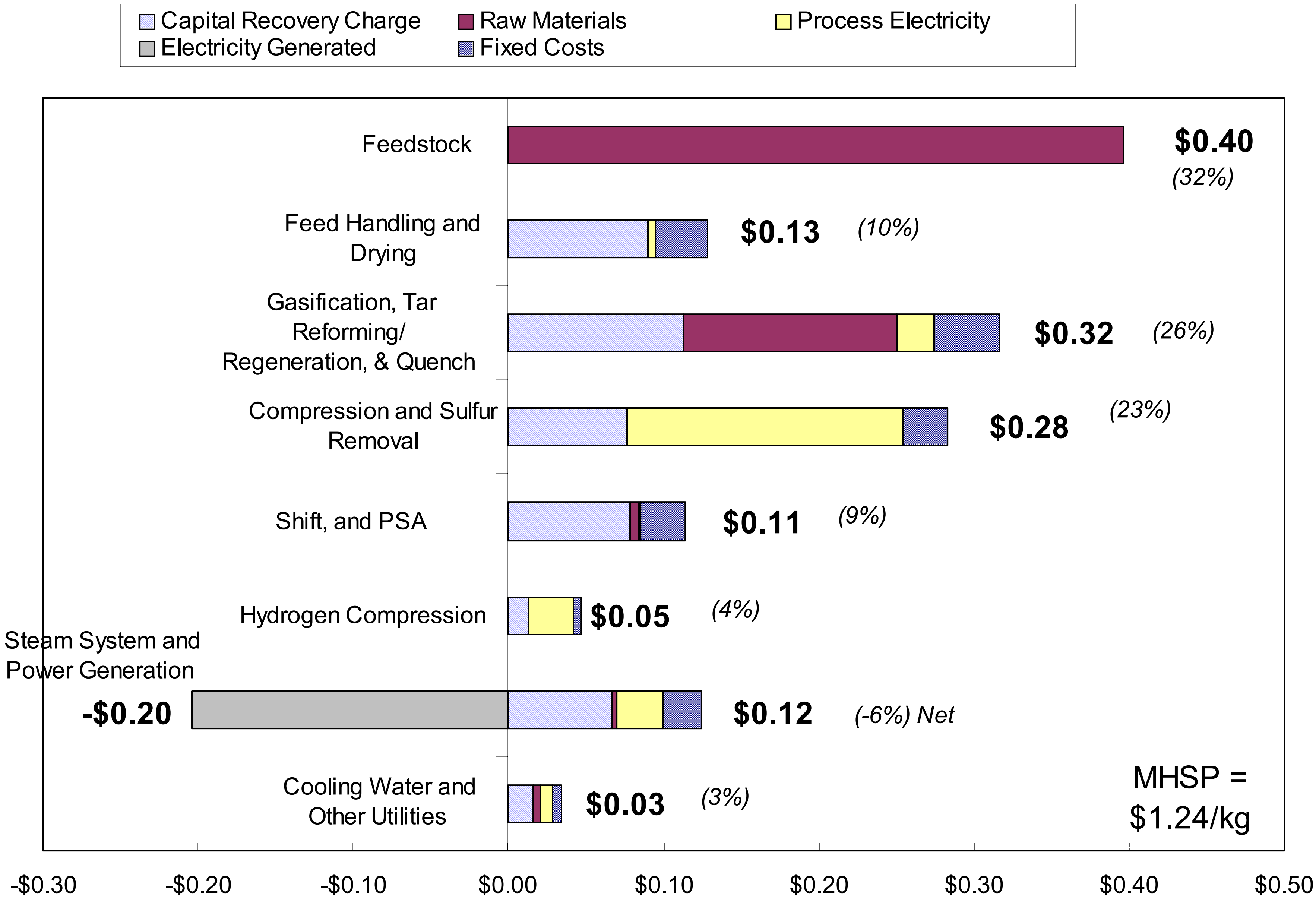

Figure 14: Goal Design Base Case Cost Contribution Diagram

16.0 Goal Design Sensitivity Analyses

Only a few of the parameters tested in the current design sensitivity analysis were tested here. Overall, the parameters tested on the current design will have a similar affect on the goal design. Since the feedstock cost has a big impact on the hydrogen price, the lower and higher feedstock costs were tested here. Because the natural gas consumption increased compared to the current design, the effect of increasing the cost of natural gas was also examined. A few of the other parameters listed above in the current design sensitivity analysis (Table 22) were also tested and are shown in Table 24. Additionally, changing the steam to carbon ratio to the shift reactors was investigated. All of the variables examined in the sensitivity analysis for the goal design are listed in Table 24 and the results are in Table 25. The results of the sensitivity analysis for the goal design are also shown in Figure 15. Internal rate of return and debt equity ratio were also examined for the goal design. Again, when a percentage of the financing is debt, the loan interest rate was set at 7.5% with a loan term of 10 years. Figure 16 shows those results.

Table 24: Goal Design – Sensitivity Analysis Cases

Letter	Sensitivity Case	Analysis Changes Made
AA	Decrease feedstock cost to \$0/dry ton	The feedstock cost in the DCFROR spreadsheet was changed from \$30/dry ton to \$0/dry ton.
BB	Increase feedstock cost to \$53/dry ton	The feedstock cost in the DCFROR spreadsheet was changed from \$30/dry ton to \$53/dry ton.
CC	Lower feed moisture content of 30 wt%	The feed moisture content in the Aspen Plus model was decreased from 50 wt% to 30 wt%.
DD	Less drying of biomass feed to a moisture content of 20 wt%	The wood moisture content at the dryer outlet was changed from 12 wt% to 20 wt%. The gasifier temperature dropped from 859ºC (1,598ºF) to 870ºC (1,578ºF). No additional natural gas was required to maintain the heat balance around the gasifier and combustor (enough additional char was produced at the lower gasifier temperature). The dryer cost decreased.
EE	No hydrogen recycle to PSA	The recycling of hydrogen to the PSA feed was eliminated.
FF	Increase in PSA cost	There is some variability in the capital cost data for the PSA so the cost was increased by a factor of 1.6 to determine the sensitivity to this parameter. This factor was determined using two different costing methods for the PSA. One was based on the hydrogen production rate and the other was based on the inlet flow rate to the PSA.
GG	Increase in natural gas price to \$7/MMBtu	The natural gas cost in the DCFROR spreadsheet was changed from \$5.28/MMBtu to \$7/MMBtu.
HH	Increase in tar reformer/catalyst regenerator system capital cost	The capital cost for the tar reformer/regenerator system was doubled making the total project investment of the goal base case design roughly the same as that for the current base case design.
II	Increase in shift steam to carbon ratio from 2 to 3	The shift steam rate in the Aspen Plus was increased from a steam:carbon ratio of 2 mol H_2O/mol of C to a value of 3.
JJ	Decrease in shift steam to carbon ratio from 2 to 1.5	The shift steam rate in the Aspen Plus was decreased from a steam:carbon ratio of 2 mol H_2O/mol of C to a value of 1.5.

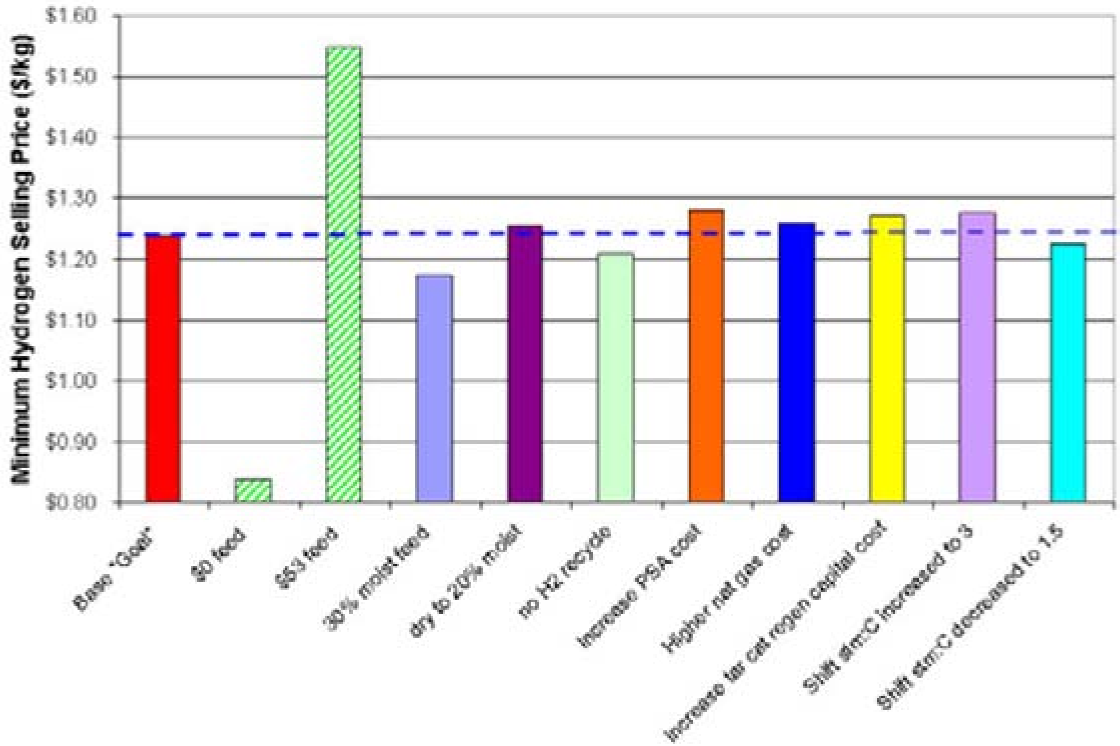

Figure 15: Goal Design Sensitivity Analysis Results

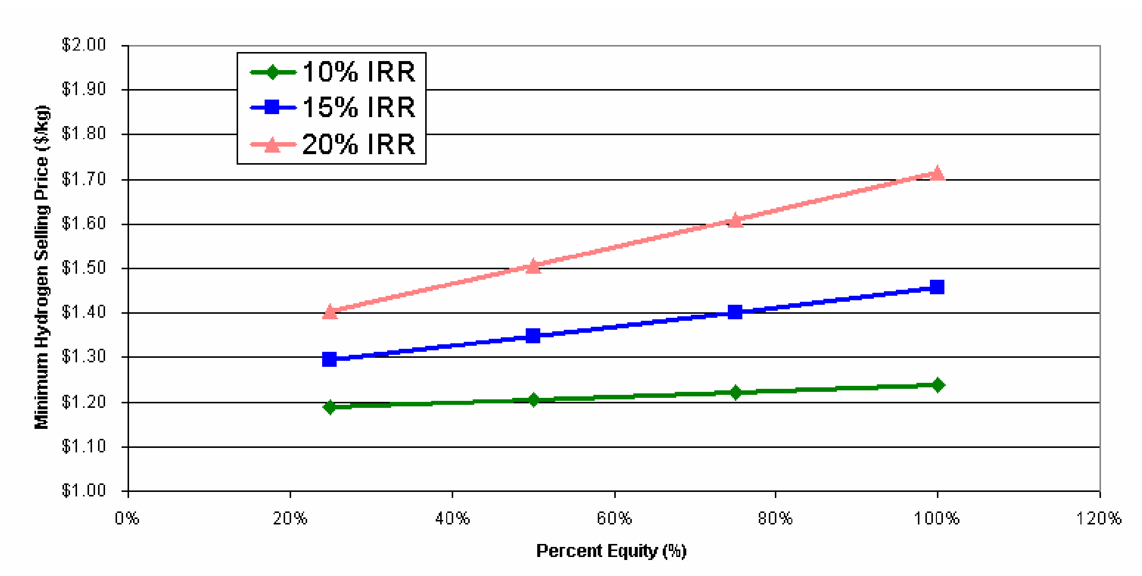

Figure 16: Effect of IRR and Debt/Equity on Goal Design Base Case

Table 25: Goal Design Base Case and Sensitivity Analysis Results

Letter	Sensitivity Case	Minimum Hydrogen Selling Price ($/kg)	Minimum Hydrogen Selling Price ($/GJ, LHV)
Base	*Goal design - base case*	*$1.24*	*$10.34*
AA	Decrease feedstock cost to $0/dry ton	$0.84	$6.97
BB	Increase feedstock cost to $53/dry ton	$1.55	$12.9
CC	Lower feed moisture content of 30%	$1.18	$9.81
DD	Less drying of biomass feed to a moisture content of 20 wt%	$1.26	$10.47
EE	No hydrogen recycle to PSA	$1.21	$10.08
FF	Increase in PSA cost	$1.28	$10.67
GG	Increase in natural gas price to $7/MMBtu	$1.26	$10.49
HH	Increase in tar reformer/catalyst regenerator system capital cost	$1.27	$10.6
II	Increase in shift steam to carbon ratio from 2 to 3	$1.28	$10.63
JJ	Decrease in shift steam to carbon ratio from 2 to 1.5	$1.22	$10.21

Even increasing the capital cost of the tar reformer/regenerator system so that the total project investment was equivalent to that of the current design (HH) resulted in a minimum hydrogen selling price that is less than the minimum hydrogen selling price for the current base case design. This is because the hydrogen yield for this design is higher.

A higher steam to carbon ratio increases the hydrogen yield but adversely affects the economics of the goal design because the operating costs increase and the total project investment goes up as well. However, there is a minimum steam to carbon ratio that the system must operate at in order to convert the CO to hydrogen ($CO + H_2O \Leftrightarrow CO_2 + H_2$).

17.0 Sensitivity to Plant Size

The plant size is another variable that was examined for both the current and goal case design. The plant size was changed in the spreadsheet from the base case size of 2,000 dry tonne/day to the desired plant size. The material and energy balances were determined by multiplying the base case values by the ratio of the plant sizes (i.e., multiplying by [the desired plant size in dry tonne/day]/[2,000 dry tonne/day]). The equipment were then scaled using the scaling exponents shown in Appendix H: Current Design Summary of Individual Equipment Costs and Appendix I: Goal Design Summary of Individual Equipment Costs (i.e., new cost = original cost * [new size/original size]^{exp}) and the minimum hydrogen selling price was recalculated. Figure 17 shows the difference in the minimum hydrogen selling price for a plant size of 500 bone dry tonnes/day to 2,000 bone dry tonnes/day. In reducing the plant size from 2,000 bone dry tonnes/day to 500, the hydrogen price increases from $1.38/kg to $1.88/kg for the current design and from $1.24/kg to $1.68/kg for the goal design. This is a 36% increase.

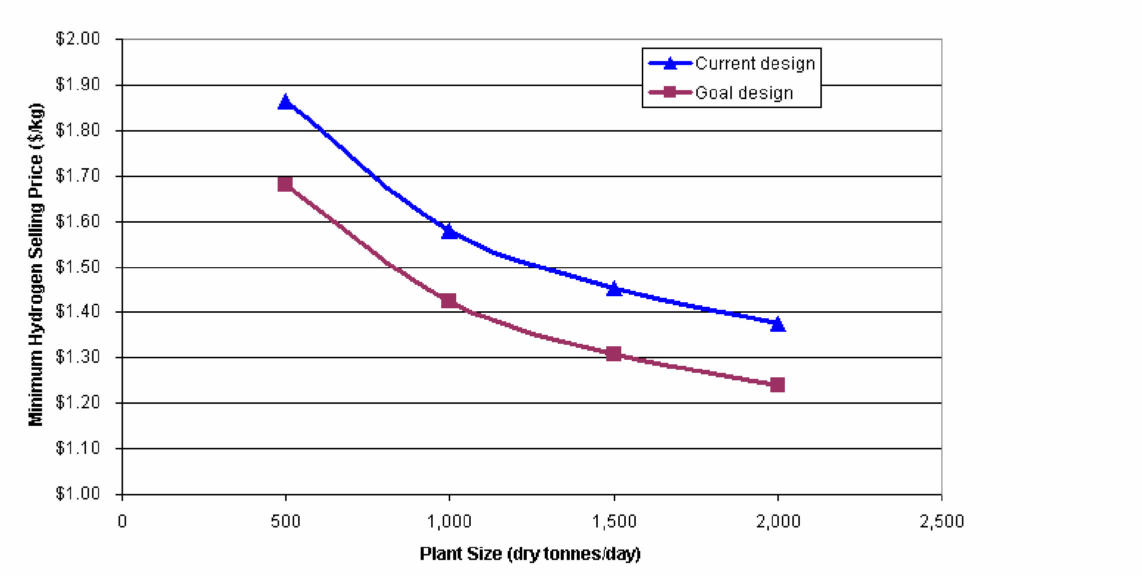

Figure 17: Effect of Plant Size on Minimum Hydrogen Selling Price

18.0 Syngas Price

As can be seen in Figure 11 and Figure 14, syngas production accounts for a significant portion of the minimum hydrogen selling price. This is also true for the synthesis of other fuel or chemical products (Spath and Dayton, 2003). As a benchmark for thermochemical conversion, the DOE Biomass Program is setting program targets based on intermediate syngas prices to track progress toward reducing the technical barriers associated with biomass gasification. Therefore, this analysis included calculations in determining both an intermediate and a stand-alone clean, reformed syngas price.

18.1 Intermediate Syngas Price

First an intermediate syngas price was determined. The value of the syngas was determined by taking a slipstream of the clean, reformed syngas and treating it as a minor co-product to the overall biomass-to-hydrogen process. The price of the syngas slipstream was determined to be the value that would maintain the MHSP equal to that of the base case hydrogen price which does not have a slipstream. This was done by taking the Aspen Plus model and separating a slipstream of clean, reformed syngas from the process, setting the hydrogen price equal to the base case cost (i.e., $1.38/kg for the current design and $1.24/kg for the goal design), and calculating the syngas price using the revised material and energy balance and thus revised capital and operating costs.

In order to calculate an intermediate syngas price, a slipstream of clean, reformed syngas from 1%-20% of the total syngas stream was examined. The heat balance was the limiting factor beyond 20%, resulting in no flow through the steam cycle beyond the steam required for gasification and reforming. A slipstream larger than this amount would require the combustion of natural gas or another fuel to raise steam. The slipstream for the current design was taken just downstream of the steam reformer (R-

401). Since the goal design eliminates the steam reformer, the slipstream for the goal design was taken just after the ZnO beds (R-302). Therefore, both of these systems are examining clean, reformed syngas. The intermediate syngas price in \$/GJ (LHV) for both designs can be seen in Figure 18. For the current design the intermediate syngas price starts out at \$6.88/GJ (\$7.25/MMBtu) for a 1% slipstream and ramps up to \$8.24/GJ (\$8.69/MMBtu) for a 20% slipstream. In the goal design the intermediate syngas price starts out at \$4.98/GJ (\$5.25/MMBtu) for a 1% slipstream and ramps up to \$6.97/GJ (\$7.35/MMBtu) for a 20% slipstream. The intermediate syngas price of the clean, reformed syngas for the integrated process should actually be considered to be the low end value at the small slipstream amount. This is the cost of the syngas for the integrated process. As the slipstream becomes larger, the price escalates quickly and then levels off thus approaching the syngas price of a stand-alone plant (see section 18.2 Stand-alone Syngas Price).

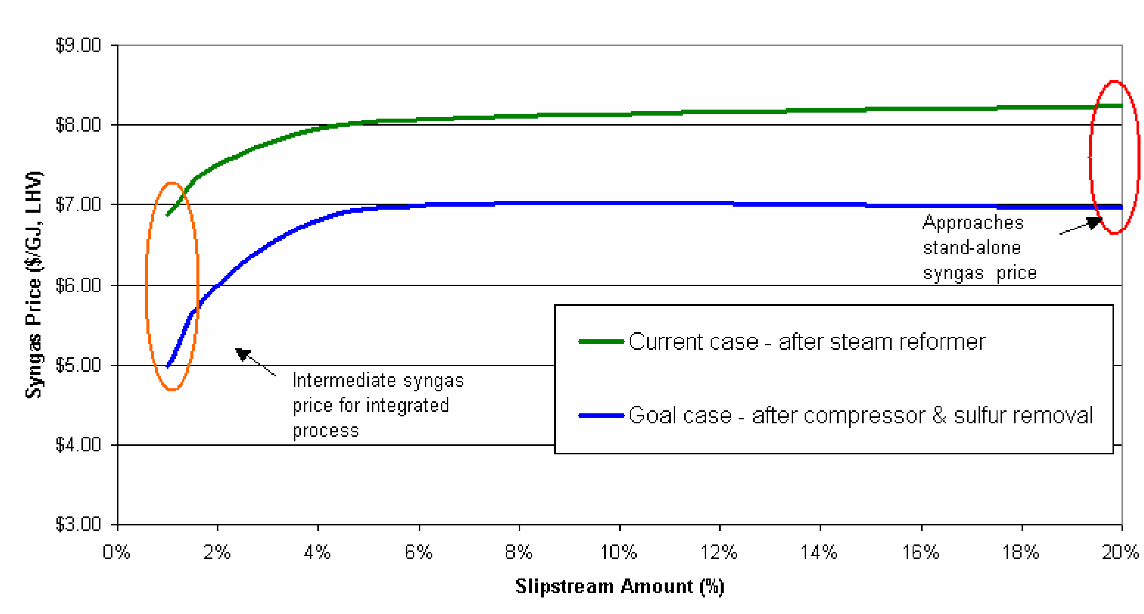

Figure 18: Intermediate Syngas Price

18.2 Stand-alone Syngas Price

Next a stand-alone syngas price was determined. For the current case this meant removing the process steps downstream of the steam reformer (shift conversion, purification, and hydrogen compression), and reconfiguring the heat balance. For the goal case this meant removing the process steps downstream of the sulfur removal step (shift conversion, purification, and hydrogen compression) and reconfiguring the heat balance. The syngas is cooled and the water is condensed from the syngas stream but no other conditioning of the syngas is done.

In the current and goal case integrated hydrogen production process designs, off gas from the PSA unit is used to fuel the steam reformer or tar regenerator, respectively, with a slight amount of natural gas used for combustion control. In the stand-alone syngas plant for the current and goal designs, only natural gas is used as fuel since the product is now syngas. The heat available within the stand-alone syngas plant is used to meet the steam demand of the system, which means steam required for gasification and for the current design, additional steam required for steam methane reforming. Some power is also produced. The resulting stand-alone syngas price for each design is given in Table 26.

Table 26: Stand-alone Syngas Price

	Current Design	Goal Design
Stand-alone syngas price (LHV)	\$8.22/GJ \$8.67/MMBtu	\$6.73/GJ \$7.10/MMBtu

For each stand-alone syngas design compared to the integrated hydrogen production plant, the total project investment decreases but the operating cost for natural gas and electricity increases. However, the natural gas and electricity operating costs for the stand-alone syngas goal design do not increase as much as those for the stand-alone syngas current design. This is because the shift conversion section has been eliminated and thus for the stand-alone goal design there is no additional steam requirement other than that for gasification.

19.0 Hydrogen Program Analysis

The results of this analysis are being used by the US Department of Energy's Hydrogen, Fuel Cells & Infrastructure Program in the standard worksheet that they have developed for their hydrogen analysis group. However, it should be noted that the hydrogen price determined from their spreadsheet will be different than ours due to their use of different economic parameters such as operating hours, feedstock cost, inflation and escalation. It should also be noted that the Hydrogen, Fuel Cells & Infrastructure Program funded a portion of this work.

20.0 Conclusions

The results of this analysis show a minimum hydrogen selling price of \$1.38/kg (\$11.48/GJ, LHV) for a 2,000 bone dry tonne/day plant for the current design and a price of \$1.24/kg (\$10.34/GJ, LHV) for the goal design. The hydrogen price decreases mainly because of an increase in the hydrogen yield. The decrease in the total project investment also has some affect. This result shows that the research at NREL in catalytic tar destruction and heteroatom removal is moving in a direction that has the potential to decrease the cost of producing clean syngas (by about \$1.5-2/GJ) and any subsequent fuel products via biomass gasification.

Since the feedstock cost contributes a large percentage to the hydrogen selling price (about 30%), this variable will always have a large impact on the economics. Overall, the sensitivity analysis shows that any parameter that significantly affects the heat balance of the system will greatly affect the minimum hydrogen selling price. For example, eliminating the dryer and adding more natural gas to the char combustor eliminates the dryer capital cost but increases operating costs and capital costs associated with the gasifier/combustor in order to maintain the heat balance around the gasifier/combustor.

Also, significantly increasing or decreasing the gasifier steam to wood ratio has a large affect on the minimum hydrogen selling price. This variable greatly affects on the heat balance of the system and the capital and operating costs. Feeding a lower moisture feedstock (the base case assumes 50% moisture in the feed) also affects the heat balance, thus resulting in a decrease in the hydrogen price.

The intermediate syngas price for the current and goal designs are $6.88/GJ ($7.25/MMBtu) and $4.98/GJ ($5.25/MMBtu), respectively. This is for clean, reformed syngas in the integrated biomass-to-hydrogen design. Stand-alone syngas plants are not being built today but for a stand-alone plant the syngas price would be $8.24/GJ ($8.69/MMBtu) for a plant based on the current design and $6.97/GJ ($7.35/MMBtu) for a plant based on the goal design. The lower intermediate syngas price shows the importance of integration within the fuels synthesis process plant.

21.0 Future Work

In addition to gas clean up and conditioning other barrier areas that could reduce the cost of fuel products from thermochemical conversion of biomass include feed handling and drying, gasification, production of different products and co-products, and process integration. Future work entails obtaining better gas clean up costs for various cleaning and conditioning configurations that will be the most beneficial for downstream conversion of biomass derived synthesis gas. Additional capital cost items where vendor information will reduce the amount of uncertainty in this analysis include a steam reformer cost for reforming synthesis gas streams particularly those containing low amounts of methane and a PSA cost for gas streams containing less than 70 mol% hydrogen. Although the capital cost information for the feed handling and gasification come from studies that have used detailed design information, specific breakdowns of the cost components as well as operating costs would improve the accuracy of the analysis. Another item that should be examined in the future from an environmental point of view as well as an economical point of view is flue gas dryers versus steam dryers. More work also needs to be done to compare indirect gasification with direct gasification to determine the most suitable and economically viable gasification system for different fuels products. Future work will also entail examining other biomass feedstocks and other products along with the integration of thermochemical and biochemical conversion processes into biorefinery concepts.

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Appendix A: Current and Goal Base Case Summary Sheets

Hydrogen Production Process Engineering Analysis

Design Report: Current Case
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.38**

		$9.62	($/GJ H2, HHV basis)
		$11.48	($/GJ H2, LHV basis)
Hydrogen Production at operating capacity (MM kg / year)	54.4	65.7	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	70.4	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$16,800,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$30,300,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$101,700,000
Indirect Costs	51,900,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$153,600,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.183
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	45%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	51.0%
Overall Plant Efficiency - LHV	45.6%

Operating Costs (cents/kg hydrogen)	
Feedstock	42.6
Natural Gas	2.8
Tar Reforming Catalyst	4.3
Other Catalysts	0.7
Olivine	7.1
Other Raw Materials	0.9
Waste Disposal	1.3
Electricity	7.5
Fixed Costs	18.7
Capital Depreciation	14.2
Average Income Tax	11.0
Average Return on Investment	26.5

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$1,500,000
Tar Cracking Catalyst	$2,400,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$500,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$10,200,000
Capital Depreciation	$7,700,000
Average Income Tax	$6,000,000
Average Return on Investment	$14,400,000

Total Plant Electricity Usage (KW)	35803
Electricity Produced Onsite (KW)	-25583
Electricity Purchased from Grid (KW)	10219
Plant Electricity Use (KWh/kg H2)	5.54
Plant Steam Use (kg steam/kg H2)	23.6

Hydrogen Production Process Engineering Analysis

Design Report: Goal Case
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.24** — $8.66 ($/GJ H2, HHV basis); $10.34 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	58.4	70.6	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	75.7	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming/Regeneration, & Quench	$23,800,000
Compression & Sulfur Removal	$16,100,000
Shift, and PSA	$16,500,000
Hydrogen Compression	$2,800,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$95,700,000
Indirect Costs	48,800,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$144,400,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.184
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	49%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	53.3%
Overall Plant Efficiency - LHV	47.8%

Operating Costs (cents/kg hydrogen)	
Feedstock	39.7
Natural Gas	5.9
Catalysts	0.7
Olivine	6.6
Other Raw Materials	0.9
Waste Disposal	1.2
Electricity	7.1
Fixed Costs	16.8
Capital Depreciation	12.3
Average Income Tax	9.8
Average Return on Investment	23.3

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$3,400,000
Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$500,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$9,800,000
Capital Depreciation	$7,200,000
Average Income Tax	$5,700,000
Average Return on Investment	$13,600,000

Total Plant Electricity Usage (KW)	40259
Electricity Produced Onsite (KW)	-29974
Electricity Purchased from Grid (KW)	10284
Plant Electricity Use (KWh/kg H2)	5.79
Plant Steam Use (kg steam/kg H2)	19.5

Appendix B: Sensitivity Summary Sheets

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity on Current Case - $0 feed cost (Case A)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg)	**$0.94**	$6.58	($/GJ H2, HHV basis)
		$7.86	($/GJ H2, LHV basis)
Hydrogen Production at operating capacity (MM kg / year)	54.4	65.7	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	70.4	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$0		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$16,800,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$30,300,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$101,700,000
Indirect Costs	51,900,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$153,600,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.181
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	45%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	51.0%
Overall Plant Efficiency - LHV	45.6%

Operating Costs (cents/kg hydrogen)	
Feedstock	0.0
Natural Gas	2.8
Tar Reforming Catalyst	4.3
Other Catalysts	0.7
Olivine	7.1
Other Raw Materials	0.8
Waste Disposal	1.3
Electricity	7.5
Fixed Costs	18.7
Capital Depreciation	14.2
Average Income Tax	11.0
Average Return on Investment	25.9

Operating Costs ($/yr)	
Feedstock	$0
Natural Gas	$1,500,000
Tar Cracking Catalyst	$2,400,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$10,200,000
Capital Depreciation	$7,700,000
Average Income Tax	$6,000,000
Average Return on Investment	$14,100,000
Total Plant Electricity Usage (KW)	35803
Electricity Produced Onsite (KW)	-25583
Electricity Purchased from Grid (KW)	10219
Plant Electricity Use (KWh/kg H2)	5.54
Plant Steam Use (kg steam/kg H2)	23.6

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity on Current Case - $53/dry ton feed cost (Case B)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.71**

$11.92	($/GJ H2, HHV basis)
$14.24	($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	54.4	65.7	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	70.4	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$53		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$16,800,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$30,300,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$101,700,000
Indirect Costs	51,900,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$153,600,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.184
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	45%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	51.0%
Overall Plant Efficiency - LHV	45.6%

Operating Costs (cents/kg hydrogen)	
Feedstock	75.3
Natural Gas	2.8
Tar Reforming Catalyst	4.3
Other Catalysts	0.7
Olivine	7.1
Other Raw Materials	0.8
Waste Disposal	1.3
Electricity	7.5
Fixed Costs	18.7
Capital Depreciation	14.2
Average Income Tax	11.1
Average Return on Investment	26.9

Operating Costs ($/yr)	
Feedstock	$40,900,000
Natural Gas	$1,500,000
Tar Cracking Catalyst	$2,400,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$10,200,000
Capital Depreciation	$7,700,000
Average Income Tax	$6,000,000
Average Return on Investment	$14,600,000
Total Plant Electricity Usage (KW)	35803
Electricity Produced Onsite (KW)	-25583
Electricity Purchased from Grid (KW)	10219
Plant Electricity Use (KWh/kg H2)	5.54
Plant Steam Use (kg steam/kg H2)	23.6

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity on Current Case - 30% moisture feedstock (Case C)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) $1.31 — $9.12 ($/GJ H2, HHV basis); $10.89 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	54.4	66	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	70.4	907	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$15,600,000
Gasification, Tar Reforming, & Quench	$20,000,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$30,400,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$16,700,000
Cooling Water and Other Utilities	$3,500,000
Total Installed Equipment Cost	$104,300,000
Indirect Costs	53,200,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$157,500,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.183
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	45%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	54.7%
Overall Plant Efficiency - LHV	49.5%

Operating Costs (cents/kg hydrogen)	
Feedstock	42.6
Natural Gas	2.8
Tar Reforming Catalyst	4.3
Other Catalysts	0.7
Olivine	7.1
Other Raw Materials	1.0
Waste Disposal	1.3
Electricity	-1.1
Fixed Costs	18.8
Capital Depreciation	14.5
Average Income Tax	11.5
Average Return on Investment	27.0

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$300,000
Tar Cracking Catalyst	$2,400,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$1,700,000
Waste Disposal	$700,000
Electricity	-$600,000
Fixed Costs	$10,200,000
Capital Depreciation	$7,900,000
Average Income Tax	$6,300,000
Average Return on Investment	$14,700,000
Total Plant Electricity Usage (KW)	36697
Electricity Produced Onsite (KW)	-38226
Electricity Purchased from Grid (KW)	-1529
Plant Electricity Use (KWh/kg H2)	5.67
Plant Steam Use (kg steam/kg H2)	23.6

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Current Case - dry to 20% moisture (lower gasifier temp) (Case D)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg)	**$1.37**	$9.57	($/GJ H2, HHV basis)
		$11.44	($/GJ H2, LHV basis)
Hydrogen Production at operating capacity (MM kg / year)	52.4	63.4	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	67.9	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$12,900,000
Gasification, Tar Reforming, & Quench	$17,800,000
Compression & Sulfur Removal	$15,300,000
Steam Methane Reforming, Shift, and PSA	$29,700,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$14,700,000
Cooling Water and Other Utilities	$3,700,000
Total Installed Equipment Cost	$96,700,000
Indirect Costs	49,300,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$145,900,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.184
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	44%
Gasifier Efficiency - HHV	70.14%
Gasifier Efficiency - LHV	69.83%
Overall Plant Efficiency - HHV	50.2%
Overall Plant Efficiency - LHV	45.0%

Operating Costs (cents/kg hydrogen)	
Feedstock	44.2
Natural Gas	2.9
Tar Reforming Catalyst	4.9
Other Catalysts	0.7
Olivine	7.3
Other Raw Materials	0.9
Waste Disposal	1.5
Electricity	4.8
Fixed Costs	18.8
Capital Depreciation	13.9
Average Income Tax	11.0
Average Return on Investment	26.1

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$1,500,000
Tar Cracking Catalyst	$2,500,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$500,000
Waste Disposal	$800,000
Electricity	$2,500,000
Fixed Costs	$9,900,000
Capital Depreciation	$7,300,000
Average Income Tax	$5,800,000
Average Return on Investment	$13,700,000
Total Plant Electricity Usage (KW)	35854
Electricity Produced Onsite (KW)	-29525
Electricity Purchased from Grid (KW)	6328
Plant Electricity Use (KWh/kg H2)	5.75
Plant Steam Use (kg steam/kg H2)	23.4

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Current Case - dry to 20% moisture with same gasifier temperature (Case E)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.39**

$9.70	($/GJ H2, HHV basis)
$11.59	($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	53.8	65.1	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	69.7	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$12,900,000
Gasification, Tar Reforming, & Quench	$19,000,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$30,200,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$15,000,000
Cooling Water and Other Utilities	$3,700,000
Total Installed Equipment Cost	$98,900,000
Indirect Costs	50,500,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$149,400,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.183
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	45%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	50.4%
Overall Plant Efficiency - LHV	45.3%

Operating Costs (cents/kg hydrogen)	
Feedstock	43.0
Natural Gas	5.8
Tar Reforming Catalyst	4.8
Other Catalysts	0.7
Olivine	8.0
Other Raw Materials	0.9
Waste Disposal	1.6
Electricity	4.6
Fixed Costs	18.6
Capital Depreciation	13.9
Average Income Tax	11.0
Average Return on Investment	26.0

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$3,100,000
Tar Cracking Catalyst	$2,600,000
Other Catalysts	$400,000
Olivine	$4,300,000
Other Raw Matl. Costs	$500,000
Waste Disposal	$800,000
Electricity	$2,500,000
Fixed Costs	$10,000,000
Capital Depreciation	$7,500,000
Average Income Tax	$5,900,000
Average Return on Investment	$14,000,000
Total Plant Electricity Usage (KW)	36588
Electricity Produced Onsite (KW)	-30442
Electricity Purchased from Grid (KW)	6146
Plant Electricity Use (KWh/kg H2)	5.71
Plant Steam Use (kg steam/kg H2)	23.2

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Current Case - dry to 30% moisture (lower gasifier temp) (Case F)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.46**

$10.22 ($/GJ H2, HHV basis)
$12.20 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	48.8	59.0	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	63.2	2,116	(dry tons / day) at operating capacity
Delivered Feedstock Cost $/Dry US Ton	$30		
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$10,400,000
Gasification, Tar Reforming, & Quench	$20,100,000
Compression & Sulfur Removal	$15,200,000
Steam Methane Reforming, Shift, and PSA	$28,700,000
Hydrogen Compression	$2,400,000
Steam System and Power Generation	$16,900,000
Cooling Water and Other Utilities	$4,200,000
Total Installed Equipment Cost	$97,900,000
Indirect Costs	49,900,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$147,700,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.185
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	41%
Gasifier Efficiency - HHV	67.08%
Gasifier Efficiency - LHV	66.84%
Overall Plant Efficiency - HHV	48.9%
Overall Plant Efficiency - LHV	44.8%

Operating Costs (cents/kg hydrogen)	
Feedstock	47.5
Natural Gas	9.5
Tar Reforming Catalyst	5.8
Other Catalysts	0.8
Olivine	7.9
Other Raw Materials	1.3
Waste Disposal	1.8
Electricity	-4.4
Fixed Costs	20.4
Capital Depreciation	15.2
Average Income Tax	12.2
Average Return on Investment	28.4

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$4,600,000
Tar Cracking Catalyst	$2,800,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$600,000
Waste Disposal	$900,000
Electricity	-$2,100,000
Fixed Costs	$10,000,000
Capital Depreciation	$7,400,000
Average Income Tax	$6,000,000
Average Return on Investment	$13,900,000
Total Plant Electricity Usage (KW)	37588
Electricity Produced Onsite (KW)	-42891
Electricity Purchased from Grid (KW)	-5303
Plant Electricity Use (KWh/kg H2)	6.47
Plant Steam Use (kg steam/kg H2)	23.5

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Current Case - dry to 30% moisture with same gasifier temperature (Case G)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.50** — $10.46 ($/GJ H2, HHV basis); $12.50 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	53.0	64.1	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	68.6	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$10,500,000
Gasification, Tar Reforming, & Quench	$23,300,000
Compression & Sulfur Removal	$15,700,000
Steam Methane Reforming, Shift, and PSA	$30,000,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$17,800,000
Cooling Water and Other Utilities	$4,300,000
Total Installed Equipment Cost	$104,200,000
Indirect Costs	53,100,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$157,300,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.185
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	44%
Gasifier Efficiency - HHV	72.02%
Gasifier Efficiency - LHV	71.66%
Overall Plant Efficiency - HHV	49.8%
Overall Plant Efficiency - LHV	45.8%

Operating Costs (cents/kg hydrogen)	
Feedstock	43.7
Natural Gas	17.8
Tar Reforming Catalyst	5.6
Other Catalysts	0.7
Olivine	9.4
Other Raw Materials	1.3
Waste Disposal	1.9
Electricity	-4.6
Fixed Costs	19.5
Capital Depreciation	14.9
Average Income Tax	12.0
Average Return on Investment	27.8

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$9,400,000
Tar Cracking Catalyst	$2,900,000
Other Catalysts	$400,000
Olivine	$5,000,000
Other Raw Matl. Costs	$700,000
Waste Disposal	$1,000,000
Electricity	-$2,500,000
Fixed Costs	$10,300,000
Capital Depreciation	$7,900,000
Average Income Tax	$6,400,000
Average Return on Investment	$14,800,000
Total Plant Electricity Usage (KW)	39856
Electricity Produced Onsite (KW)	-46002
Electricity Purchased from Grid (KW)	-6146
Plant Electricity Use (KWh/kg H2)	6.32
Plant Steam Use (kg steam/kg H2)	22.8

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity on Current Case - no dryer (Case H)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg)	**$1.78**	$12.43	($/GJ H2, HHV basis)
		$14.85	($/GJ H2, LHV basis)
Hydrogen Production at operating capacity (MM kg / year)	51.3	62	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	66.4	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$8,200,000
Gasification, Tar Reforming, & Quench	$31,900,000
Compression & Sulfur Removal	$16,200,000
Steam Methane Reforming, Shift, and PSA	$29,600,000
Hydrogen Compression	$2,500,000
Steam System and Power Generation	$23,500,000
Cooling Water and Other Utilities	$5,600,000
Total Installed Equipment Cost	$117,500,000
Indirect Costs	59,900,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$177,200,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.186
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	43%
Gasifier Efficiency - HHV	72.01%
Gasifier Efficiency - LHV	71.65%
Overall Plant Efficiency - HHV	52.0%
Overall Plant Efficiency - LHV	50.6%

Operating Costs (cents/kg hydrogen)	
Feedstock	45.2
Natural Gas	45.9
Tar Reforming Catalyst	8.0
Other Catalysts	0.7
Olivine	14.2
Other Raw Materials	2.2
Waste Disposal	3.1
Electricity	-26.8
Fixed Costs	21.5
Capital Depreciation	17.4
Average Income Tax	14.4
Average Return on Investment	32.4

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$900,000
Tar Cracking Catalyst	$4,100,000
Other Catalysts	$400,000
Olivine	$7,300,000
Other Raw Matl. Costs	$23,800,000
Waste Disposal	$1,600,000
Electricity	-$13,800,000
Fixed Costs	$11,000,000
Capital Depreciation	$8,900,000
Average Income Tax	$7,400,000
Average Return on Investment	$16,600,000
Total Plant Electricity Usage (KW)	46376
Electricity Produced Onsite (KW)	-80705
Electricity Purchased from Grid (KW)	-34329
Plant Electricity Use (KWh/kg H2)	7.60
Plant Steam Use (kg steam/kg H2)	21.8

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity on Current Case - stm:wood ratio = 0.1 (Case I)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg)	**$1.30**	$9.10	($/GJ H2, HHV basis)
		$10.87	($/GJ H2, LHV basis)
Hydrogen Production at operating capacity (MM kg / year)	56.3	68	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	72.9	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$19,300,000
Gasification, Tar Reforming, & Quench	$16,700,000
Compression & Sulfur Removal	$13,800,000
Steam Methane Reforming, Shift, and PSA	$30,900,000
Hydrogen Compression	$2,700,000
Steam System and Power Generation	$13,500,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$100,300,000
Indirect Costs	51,100,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$151,400,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.183
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	47%
Gasifier Efficiency - HHV	72.05%
Gasifier Efficiency - LHV	71.69%
Overall Plant Efficiency - HHV	52.5%
Overall Plant Efficiency - LHV	46.9%

Operating Costs (cents/kg hydrogen)	
Feedstock	41.1
Natural Gas	2.6
Tar Reforming Catalyst	3.1
Other Catalysts	0.7
Olivine	5.9
Other Raw Materials	0.7
Waste Disposal	1.1
Electricity	8.4
Fixed Costs	17.7
Capital Depreciation	13.5
Average Income Tax	10.4
Average Return on Investment	25.2

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$200,000
Tar Cracking Catalyst	$1,800,000
Other Catalysts	$400,000
Olivine	$3,300,000
Other Raw Matl. Costs	$1,600,000
Waste Disposal	$600,000
Electricity	$4,700,000
Fixed Costs	$10,000,000
Capital Depreciation	$7,600,000
Average Income Tax	$5,900,000
Average Return on Investment	$14,200,000
Total Plant Electricity Usage (KW)	34388
Electricity Produced Onsite (KW)	-22657
Electricity Purchased from Grid (KW)	11732
Plant Electricity Use (KWh/kg H2)	5.13
Plant Steam Use (kg steam/kg H2)	16.9

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity on Current Case - stm:wood ratio = 1 & lower gasifier temp (Case J)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg)	**$1.57**	$10.94	($/GJ H2, HHV basis)
		$13.07	($/GJ H2, LHV basis)
Hydrogen Production at operating capacity (MM kg / year)	48.6	58.7	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	62.9	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$19,300,000
Gasification, Tar Reforming, & Quench	$18,400,000
Compression & Sulfur Removal	$15,600,000
Steam Methane Reforming, Shift, and PSA	$28,500,000
Hydrogen Compression	$2,400,000
Steam System and Power Generation	$16,400,000
Cooling Water and Other Utilities	$4,000,000
Total Installed Equipment Cost	$104,600,000
Indirect Costs	53,300,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$157,900,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.184
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	41%
Gasifier Efficiency - HHV	68.06%
Gasifier Efficiency - LHV	67.79%
Overall Plant Efficiency - HHV	46.3%
Overall Plant Efficiency - LHV	41.7%

Operating Costs (cents/kg hydrogen)	
Feedstock	47.7
Natural Gas	6.5
Tar Reforming Catalyst	7.2
Other Catalysts	0.8
Olivine	7.9
Other Raw Materials	1.4
Waste Disposal	2.1
Electricity	2.2
Fixed Costs	21.3
Capital Depreciation	16.3
Average Income Tax	12.9
Average Return on Investment	30.5

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$3,100,000
Tar Cracking Catalyst	$3,500,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$700,000
Waste Disposal	$1,000,000
Electricity	$1,100,000
Fixed Costs	$10,400,000
Capital Depreciation	$7,900,000
Average Income Tax	$6,300,000
Average Return on Investment	$14,800,000
Total Plant Electricity Usage (KW)	37246
Electricity Produced Onsite (KW)	-34550
Electricity Purchased from Grid (KW)	2696
Plant Electricity Use (KWh/kg H2)	6.45
Plant Steam Use (kg steam/kg H2)	42.0

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity on Current Case - stm:wood ratio = 1 with same gasifier temperature (Case K)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg)	**$1.58**	$11.04	($/GJ H2, HHV basis)
		$13.19	($/GJ H2, LHV basis)
Hydrogen Production at operating capacity (MM kg / year)	52.1	63	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	67.4	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$19,300,000
Gasification, Tar Reforming, & Quench	$20,900,000
Compression & Sulfur Removal	$16,000,000
Steam Methane Reforming, Shift, and PSA	$29,700,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$17,000,000
Cooling Water and Other Utilities	$4,000,000
Total Installed Equipment Cost	$109,500,000
Indirect Costs	55,800,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$165,300,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.184
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	43%
Gasifier Efficiency - HHV	72.10%
Gasifier Efficiency - LHV	71.74%
Overall Plant Efficiency - HHV	47.2%
Overall Plant Efficiency - LHV	42.6%

Operating Costs (cents/kg hydrogen)	
Feedstock	44.5
Natural Gas	12.7
Tar Reforming Catalyst	6.9
Other Catalysts	0.7
Olivine	9.2
Other Raw Materials	1.4
Waste Disposal	2.2
Electricity	2.0
Fixed Costs	20.2
Capital Depreciation	15.9
Average Income Tax	12.6
Average Return on Investment	29.7

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$400,000
Tar Cracking Catalyst	$3,600,000
Other Catalysts	$400,000
Olivine	$4,800,000
Other Raw Matl. Costs	$6,900,000
Waste Disposal	$1,100,000
Electricity	$1,100,000
Fixed Costs	$10,500,000
Capital Depreciation	$8,300,000
Average Income Tax	$6,600,000
Average Return on Investment	$15,500,000
Total Plant Electricity Usage (KW)	38893
Electricity Produced Onsite (KW)	-36241
Electricity Purchased from Grid (KW)	2651
Plant Electricity Use (KWh/kg H2)	6.28
Plant Steam Use (kg steam/kg H2)	40.1

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity on Current Case - no H2 recycle (Case L)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) $1.30 — $9.10 ($/GJ H2, HHV basis); $10.87 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	58.6	71	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	75.9	2,116	(dry tons / day) at operating capacity
Delivered Feedstock Cost $/Dry US Ton	$30		
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$16,800,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$30,800,000
Hydrogen Compression	$2,800,000
Steam System and Power Generation	$13,000,000
Cooling Water and Other Utilities	$3,200,000
Total Installed Equipment Cost	$101,000,000
Indirect Costs	51,500,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$152,400,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.183
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	49%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	53.6%
Overall Plant Efficiency - LHV	47.7%

Operating Costs (cents/kg hydrogen)	
Feedstock	39.5
Natural Gas	2.2
Tar Reforming Catalyst	4.0
Other Catalysts	0.7
Olivine	6.6
Other Raw Materials	0.6
Waste Disposal	1.2
Electricity	10.9
Fixed Costs	17.1
Capital Depreciation	13.0
Average Income Tax	10.0
Average Return on Investment	24.5

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$200,000
Tar Cracking Catalyst	$2,400,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$1,400,000
Waste Disposal	$700,000
Electricity	$6,400,000
Fixed Costs	$10,000,000
Capital Depreciation	$7,600,000
Average Income Tax	$5,900,000
Average Return on Investment	$14,400,000
Total Plant Electricity Usage (KW)	35666
Electricity Produced Onsite (KW)	-19721
Electricity Purchased from Grid (KW)	15944
Plant Electricity Use (KWh/kg H2)	5.12
Plant Steam Use (kg steam/kg H2)	21.9

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity on Current Case - no LTS (Case M)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.47**

$10.24 ($/GJ H2, HHV basis)
$12.23 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	49.3	60	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	63.9	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$16,900,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$29,500,000
Hydrogen Compression	$2,500,000
Steam System and Power Generation	$15,500,000
Cooling Water and Other Utilities	$3,600,000
Total Installed Equipment Cost	$102,400,000
Indirect Costs	52,200,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$154,500,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.183
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	41%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	47.9%
Overall Plant Efficiency - LHV	43.0%

Operating Costs (cents/kg hydrogen)	
Feedstock	46.9
Natural Gas	3.1
Tar Reforming Catalyst	4.8
Other Catalysts	0.8
Olivine	7.8
Other Raw Materials	1.0
Waste Disposal	1.5
Electricity	3.1
Fixed Costs	20.5
Capital Depreciation	15.6
Average Income Tax	12.4
Average Return on Investment	29.4

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$300,000
Tar Cracking Catalyst	$2,400,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$1,700,000
Waste Disposal	$700,000
Electricity	$1,500,000
Fixed Costs	$10,100,000
Capital Depreciation	$7,700,000
Average Income Tax	$6,100,000
Average Return on Investment	$14,500,000
Total Plant Electricity Usage (KW)	35941
Electricity Produced Onsite (KW)	-32124
Electricity Purchased from Grid (KW)	3817
Plant Electricity Use (KWh/kg H2)	6.12
Plant Steam Use (kg steam/kg H2)	26.0

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity on Current Case - 0.5% tar reformer catalyst loss (Case N)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.35**

$9.43	($/GJ H2, HHV basis)	
$11.27	($/GJ H2, LHV basis)	

Hydrogen Production at operating capacity (MM kg / year)	54.4	66	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	70.4	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$16,800,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$30,300,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$101,700,000
Indirect Costs	51,900,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$153,600,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.183
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	45%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	51.0%
Overall Plant Efficiency - LHV	45.6%

Operating Costs (cents/kg hydrogen)	
Feedstock	42.6
Natural Gas	2.8
Tar Reforming Catalyst	2.2
Other Catalysts	0.7
Olivine	7.1
Other Raw Materials	0.8
Waste Disposal	1.3
Electricity	7.5
Fixed Costs	18.5
Capital Depreciation	14.2
Average Income Tax	11.0
Average Return on Investment	26.4

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$200,000
Tar Cracking Catalyst	$1,200,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$1,700,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$10,100,000
Capital Depreciation	$7,700,000
Average Income Tax	$6,000,000
Average Return on Investment	$14,400,000
Total Plant Electricity Usage (KW)	35803
Electricity Produced Onsite (KW)	-25583
Electricity Purchased from Grid (KW)	10219
Plant Electricity Use (KWh/kg H2)	5.54
Plant Steam Use (kg steam/kg H2)	23.6

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity on Current Case - Internal waste water treatment (Case O)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.38**

$9.62	($/GJ H2, HHV basis)
$11.49	($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	54.4	66	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	70.4	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$16,800,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$30,300,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$101,700,000
Indirect Costs	52,500,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$155,500,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.183
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	45%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	51.0%
Overall Plant Efficiency - LHV	45.7%

Operating Costs (cents/kg hydrogen)	
Feedstock	42.6
Natural Gas	2.8
Tar Reforming Catalyst	4.3
Other Catalysts	0.7
Olivine	7.1
Other Raw Materials	0.7
Waste Disposal	1.2
Electricity	7.4
Fixed Costs	18.7
Capital Depreciation	14.3
Average Income Tax	11.1
Average Return on Investment	26.8

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$200,000
Tar Cracking Catalyst	$2,400,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$1,700,000
Waste Disposal	$700,000
Electricity	$4,000,000
Fixed Costs	$10,100,000
Capital Depreciation	$7,800,000
Average Income Tax	$6,100,000
Average Return on Investment	$14,600,000
Total Plant Electricity Usage (KW)	35814
Electricity Produced Onsite (KW)	-25752
Electricity Purchased from Grid (KW)	10063
Plant Electricity Use (KWh/kg H2)	5.54
Plant Steam Use (kg steam/kg H2)	23.6

Hydrogen Production Process Engineering Analysis

Design Report: Sesitivity on Current Case - Increase PSA cost (Case P)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.42** — $9.90 ($/GJ H2, HHV basis); $11.82 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	54.4	65.7	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	70.4	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$16,800,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$37,500,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$108,900,000
Indirect Costs	55,600,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$164,400,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.182
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	45%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	51.0%
Overall Plant Efficiency - LHV	45.6%

Operating Costs (cents/kg hydrogen)	
Feedstock	42.6
Natural Gas	2.8
Tar Reforming Catalyst	4.3
Other Catalysts	0.7
Olivine	7.1
Other Raw Materials	0.8
Waste Disposal	1.3
Electricity	7.5
Fixed Costs	19.5
Capital Depreciation	15.1
Average Income Tax	11.7
Average Return on Investment	28.3

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$1,500,000
Tar Cracking Catalyst	$2,400,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$10,600,000
Capital Depreciation	$8,200,000
Average Income Tax	$6,400,000
Average Return on Investment	$15,400,000
Total Plant Electricity Usage (KW)	35803
Electricity Produced Onsite (KW)	-25583
Electricity Purchased from Grid (KW)	10219
Plant Electricity Use (KWh/kg H2)	5.54
Plant Steam Use (kg steam/kg H2)	23.6

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Current Case - Increase in steam reformer cost (Case Q)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) $1.45

Minimum Hydrogen Selling Price ($/kg)	$1.45	$10.11	($/GJ H2, HHV basis)
		$12.07	($/GJ H2, LHV basis)
Hydrogen Production at operating capacity (MM kg / year)	54.4	65.7	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	70.4	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$16,800,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$42,600,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$114,000,000
Indirect Costs	58,100,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$172,100,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.182
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	45%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	51.0%
Overall Plant Efficiency - LHV	45.6%

Operating Costs (cents/kg hydrogen)	
Feedstock	42.6
Natural Gas	2.8
Tar Reforming Catalyst	4.3
Other Catalysts	0.7
Olivine	7.1
Other Raw Materials	0.8
Waste Disposal	1.3
Electricity	7.5
Fixed Costs	20.1
Capital Depreciation	15.8
Average Income Tax	12.1
Average Return on Investment	29.5

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$1,500,000
Tar Cracking Catalyst	$2,400,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$10,900,000
Capital Depreciation	$8,600,000
Average Income Tax	$6,600,000
Average Return on Investment	$16,100,000
Total Plant Electricity Usage (KW)	35803
Electricity Produced Onsite (KW)	-25583
Electricity Purchased from Grid (KW)	10219
Plant Electricity Use (KWh/kg H2)	5.54
Plant Steam Use (kg steam/kg H2)	23.6

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Current Case - Higher Electricity Cost (Case R)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.40**

$9.75 ($/GJ H2, HHV basis)
$11.64 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	54.4	65.7	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	70.4	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$16,800,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$30,300,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$101,700,000
Indirect Costs	51,900,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$153,600,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.183
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	45%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	51.0%
Overall Plant Efficiency - LHV	45.6%

Operating Costs (cents/kg hydrogen)	
Feedstock	42.6
Natural Gas	2.8
Tar Reforming Catalyst	4.3
Other Catalysts	0.7
Olivine	7.1
Other Raw Materials	0.8
Waste Disposal	1.3
Electricity	9.5
Fixed Costs	18.7
Capital Depreciation	14.2
Average Income Tax	11.0
Average Return on Investment	26.5

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$1,500,000
Tar Cracking Catalyst	$2,400,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$5,200,000
Fixed Costs	$10,200,000
Capital Depreciation	$7,700,000
Average Income Tax	$6,000,000
Average Return on Investment	$14,400,000
Total Plant Electricity Usage (KW)	35803
Electricity Produced Onsite (KW)	-25583
Electricity Purchased from Grid (KW)	10219
Plant Electricity Use (KWh/kg H2)	5.54
Plant Steam Use (kg steam/kg H2)	23.6

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Current Case - Higher Natural Gas Cost (Case S)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.39** — $9.67 ($/GJ H2, HHV basis); $11.55 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	54.4	65.7	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	70.4	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$16,800,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$30,300,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$101,700,000
Indirect Costs	51,900,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$153,600,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.183
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	45%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	51.0%
Overall Plant Efficiency - LHV	45.6%

Operating Costs (cents/kg hydrogen)	
Feedstock	42.6
Natural Gas	3.7
Tar Reforming Catalyst	4.3
Other Catalysts	0.7
Olivine	7.1
Other Raw Materials	0.8
Waste Disposal	1.3
Electricity	7.5
Fixed Costs	18.7
Capital Depreciation	14.2
Average Income Tax	11.0
Average Return on Investment	26.5

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$2,000,000
Tar Cracking Catalyst	$2,400,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$10,200,000
Capital Depreciation	$7,700,000
Average Income Tax	$6,000,000
Average Return on Investment	$14,400,000

Total Plant Electricity Usage (KW)	35803
Electricity Produced Onsite (KW)	-25583
Electricity Purchased from Grid (KW)	10219
Plant Electricity Use (KWh/kg H2)	5.54
Plant Steam Use (kg steam/kg H2)	23.6

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Current Case - Low Feed Handling & Drying Cost (Case T)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.35**

$9.41 ($/GJ H2, HHV basis)
$11.24 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	54.4	65.7	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	70.4	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$14,200,000
Gasification, Tar Reforming, & Quench	$16,800,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$30,300,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$97,000,000
Indirect Costs	49,500,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$146,400,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.184
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	45%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	51.0%
Overall Plant Efficiency - LHV	45.6%

Operating Costs (cents/kg hydrogen)	
Feedstock	42.6
Natural Gas	2.8
Tar Reforming Catalyst	4.3
Other Catalysts	0.7
Olivine	7.1
Other Raw Materials	0.8
Waste Disposal	1.3
Electricity	7.5
Fixed Costs	18.2
Capital Depreciation	13.4
Average Income Tax	10.6
Average Return on Investment	25.4

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$1,500,000
Tar Cracking Catalyst	$2,400,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$9,900,000
Capital Depreciation	$7,300,000
Average Income Tax	$5,800,000
Average Return on Investment	$13,800,000
Total Plant Electricity Usage (KW)	35803
Electricity Produced Onsite (KW)	-25583
Electricity Purchased from Grid (KW)	10219
Plant Electricity Use (KWh/kg H2)	5.54
Plant Steam Use (kg steam/kg H2)	23.6

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Current Case - Low Gasification & Clean Up Cost (Case U)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.35** — $9.40 ($/GJ H2, HHV basis); $11.22 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	54.4	65.7	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	70.4	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$11,700,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$30,300,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$96,600,000
Indirect Costs	49,300,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$145,900,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.184
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	45%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	51.0%
Overall Plant Efficiency - LHV	45.6%

Operating Costs (cents/kg hydrogen)	
Feedstock	42.6
Natural Gas	2.8
Tar Reforming Catalyst	4.3
Other Catalysts	0.7
Olivine	7.1
Other Raw Materials	0.8
Waste Disposal	1.3
Electricity	7.5
Fixed Costs	18.2
Capital Depreciation	13.4
Average Income Tax	10.6
Average Return on Investment	25.2

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$1,500,000
Tar Cracking Catalyst	$2,400,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$9,900,000
Capital Depreciation	$7,300,000
Average Income Tax	$5,800,000
Average Return on Investment	$13,700,000
Total Plant Electricity Usage (KW)	35803
Electricity Produced Onsite (KW)	-25583
Electricity Purchased from Grid (KW)	10219
Plant Electricity Use (KWh/kg H2)	5.54
Plant Steam Use (kg steam/kg H2)	23.6

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Current Case - Low Feed Handling & Drying Cost Combined with Low Gasification & Clean Up Cost (Case V)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.32**

$9.20 ($/GJ H2, HHV basis)
$10.99 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	54.4	65.7	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	70.4	2,116	(dry tons / day) at operating capacity
Delivered Feedstock Cost $/Dry US Ton	$30		
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$14,200,000
Gasification, Tar Reforming, & Quench	$11,700,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$30,300,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$91,900,000
Indirect Costs	46,900,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$138,700,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.185
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	45%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	51.0%
Overall Plant Efficiency - LHV	45.6%

Operating Costs (cents/kg hydrogen)	
Feedstock	42.6
Natural Gas	2.8
Tar Reforming Catalyst	4.3
Other Catalysts	0.7
Olivine	7.1
Other Raw Materials	0.8
Waste Disposal	1.3
Electricity	7.5
Fixed Costs	17.6
Capital Depreciation	12.7
Average Income Tax	10.1
Average Return on Investment	24.2

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$1,500,000
Tar Cracking Catalyst	$2,400,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$9,600,000
Capital Depreciation	$6,900,000
Average Income Tax	$5,500,000
Average Return on Investment	$13,200,000
Total Plant Electricity Usage (KW)	35803
Electricity Produced Onsite (KW)	-25583
Electricity Purchased from Grid (KW)	10219
Plant Electricity Use (KWh/kg H2)	5.54
Plant Steam Use (kg steam/kg H2)	23.6

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Current Case - High Feed Handling & Drying Cost (Case W)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.41**

$9.86	($/GJ H2, HHV basis)
$11.78	($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	54.4	65.7	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	70.4	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$25,100,000
Gasification, Tar Reforming, & Quench	$16,800,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$30,300,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$107,900,000
Indirect Costs	55,100,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$163,000,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.182
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	45%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	51.0%
Overall Plant Efficiency - LHV	45.6%

Operating Costs (cents/kg hydrogen)	
Feedstock	42.6
Natural Gas	2.8
Tar Reforming Catalyst	4.3
Other Catalysts	0.7
Olivine	7.1
Other Raw Materials	0.8
Waste Disposal	1.3
Electricity	7.5
Fixed Costs	19.4
Capital Depreciation	14.9
Average Income Tax	11.6
Average Return on Investment	28.1

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$1,500,000
Tar Cracking Catalyst	$2,400,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$10,600,000
Capital Depreciation	$8,100,000
Average Income Tax	$6,300,000
Average Return on Investment	$15,300,000

Total Plant Electricity Usage (KW)	35803
Electricity Produced Onsite (KW)	-25583
Electricity Purchased from Grid (KW)	10219
Plant Electricity Use (KWh/kg H2)	5.54
Plant Steam Use (kg steam/kg H2)	23.6

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Current Case - High Gasification & Clean Up Cost (Case X)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.42**

$9.92	($/GJ H2, HHV basis)
$11.85	($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	54.4	65.7	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	70.4	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$24,500,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$30,300,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$109,400,000
Indirect Costs	55,800,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$165,200,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.182
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	45%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	51.0%
Overall Plant Efficiency - LHV	45.6%

Operating Costs (cents/kg hydrogen)	
Feedstock	42.6
Natural Gas	2.8
Tar Reforming Catalyst	4.3
Other Catalysts	0.7
Olivine	7.1
Other Raw Materials	0.8
Waste Disposal	1.3
Electricity	7.5
Fixed Costs	19.6
Capital Depreciation	15.3
Average Income Tax	11.7
Average Return on Investment	28.3

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$1,500,000
Tar Cracking Catalyst	$2,400,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$10,700,000
Capital Depreciation	$8,300,000
Average Income Tax	$6,400,000
Average Return on Investment	$15,400,000
Total Plant Electricity Usage (KW)	35803
Electricity Produced Onsite (KW)	-25583
Electricity Purchased from Grid (KW)	10219
Plant Electricity Use (KWh/kg H2)	5.54
Plant Steam Use (kg steam/kg H2)	23.6

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Current Case - High Feed Handling & Drying Cost Combined with High Gasification & Clean Up Cost (Case Y)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, Methane Reformer, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.46**

$10.17 ($/GJ H2, HHV basis)
$12.15 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	54.4	65.7	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	70.4	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$25,100,000
Gasification, Tar Reforming, & Quench	$24,500,000
Compression & Sulfur Removal	$15,500,000
Steam Methane Reforming, Shift, and PSA	$30,300,000
Hydrogen Compression	$2,600,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$115,600,000
Indirect Costs	59,000,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$174,600,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.182
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	45%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	51.0%
Overall Plant Efficiency - LHV	45.6%

Operating Costs (cents/kg hydrogen)	
Feedstock	42.6
Natural Gas	2.8
Tar Reforming Catalyst	4.3
Other Catalysts	0.7
Olivine	7.1
Other Raw Materials	0.8
Waste Disposal	1.3
Electricity	7.5
Fixed Costs	20.3
Capital Depreciation	16.0
Average Income Tax	12.3
Average Return on Investment	30.0

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$1,500,000
Tar Cracking Catalyst	$2,400,000
Other Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$11,000,000
Capital Depreciation	$8,700,000
Average Income Tax	$6,700,000
Average Return on Investment	$16,300,000
Total Plant Electricity Usage (KW)	35803
Electricity Produced Onsite (KW)	-25583
Electricity Purchased from Grid (KW)	10219
Plant Electricity Use (KWh/kg H2)	5.54
Plant Steam Use (kg steam/kg H2)	23.6

Hydrogen Production Process Engineering Analysis

Design Report: Sensitiivity on Goal Case - $0 feed cost (Case AA)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$0.84**

$5.84 ($/GJ H2, HHV basis)
$6.97 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	58.4	70.6	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	75.7	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$0		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$17,600,000
Compression & Sulfur Removal	$16,100,000
Tar Reforming Catalyst Regeneration, Shift, and PSA	$22,600,000
Hydrogen Compression	$2,800,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$95,600,000
Indirect Costs	48,800,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$144,400,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.181
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	49%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	53.3%
Overall Plant Efficiency - LHV	47.8%

Operating Costs (cents/kg hydrogen)	
Feedstock	0.0
Natural Gas	5.9
Catalysts	0.6
Olivine	6.6
Other Raw Materials	0.7
Waste Disposal	1.2
Electricity	7.1
Fixed Costs	16.8
Capital Depreciation	12.3
Average Income Tax	9.7
Average Return on Investment	22.8

Operating Costs ($/yr)	
Feedstock	$0
Natural Gas	$3,400,000
Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$9,800,000
Capital Depreciation	$7,200,000
Average Income Tax	$5,600,000
Average Return on Investment	$13,300,000
Total Plant Electricity Usage (KW)	40259
Electricity Produced Onsite (KW)	-29974
Electricity Purchased from Grid (KW)	10284
Plant Electricity Use (KWh/kg H2)	5.79
Plant Steam Use (kg steam/kg H2)	19.5

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity on Goal Case- $53/dry ton feed cost (Case BB)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.55**

$10.80 ($/GJ H2, HHV basis)
$12.90 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	58.4	70.6	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	75.7	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$53		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$17,600,000
Compression & Sulfur Removal	$16,100,000
Tar Reforming Catalyst Regeneration, Shift, and PSA	$22,600,000
Hydrogen Compression	$2,800,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$95,600,000
Indirect Costs	48,800,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$144,400,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.186
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	49%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	53.3%
Overall Plant Efficiency - LHV	47.8%

Operating Costs (cents/kg hydrogen)	
Feedstock	70.0
Natural Gas	5.9
Catalysts	0.6
Olivine	6.6
Other Raw Materials	0.7
Waste Disposal	1.2
Electricity	7.1
Fixed Costs	16.8
Capital Depreciation	12.3
Average Income Tax	9.8
Average Return on Investment	23.8

Operating Costs ($/yr)	
Feedstock	$40,900,000
Natural Gas	$3,400,000
Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$9,800,000
Capital Depreciation	$7,200,000
Average Income Tax	$5,700,000
Average Return on Investment	$13,900,000

Total Plant Electricity Usage (KW)	40259
Electricity Produced Onsite (KW)	-29974
Electricity Purchased from Grid (KW)	10284
Plant Electricity Use (KWh/kg H2)	5.79
Plant Steam Use (kg steam/kg H2)	19.5

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Goal Case - 30% moisture feedstock (Case CC)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.18** $8.22 ($/GJ H2, HHV basis)
$9.81 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	58.4	70.6	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	75.7	907	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$15,600,000
Gasification, Tar Reforming, & Quench	$21,100,000
Compression & Sulfur Removal	$16,100,000
Tar Reforming Catalyst Regeneration, Shift, and PSA	$22,600,000
Hydrogen Compression	$2,800,000
Steam System and Power Generation	$16,700,000
Cooling Water and Other Utilities	$3,500,000
Total Installed Equipment Cost	$98,400,000
Indirect Costs	50,200,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$148,700,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.184
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	49%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	57.0%
Overall Plant Efficiency - LHV	51.6%

Operating Costs (cents/kg hydrogen)	
Feedstock	39.7
Natural Gas	5.9
Catalysts	0.6
Olivine	6.6
Other Raw Materials	0.9
Waste Disposal	1.2
Electricity	-1.0
Fixed Costs	17.1
Capital Depreciation	12.7
Average Income Tax	10.2
Average Return on Investment	23.9

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$3,400,000
Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$500,000
Waste Disposal	$700,000
Electricity	-$600,000
Fixed Costs	$10,000,000
Capital Depreciation	$7,400,000
Average Income Tax	$6,000,000
Average Return on Investment	$14,000,000

Total Plant Electricity Usage (KW)	41153
Electricity Produced Onsite (KW)	-42624
Electricity Purchased from Grid (KW)	-1471
Plant Electricity Use (KWh/kg H2)	5.92
Plant Steam Use (kg steam/kg H2)	19.5

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Goal Case - dry feedstock to 20% moisture content (lower gasifier temp) (Case DD)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.26** — $8.77 ($/GJ H2, HHV basis); $10.47 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	56.8	68.6	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	73.5	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$19,200,000
Gasification, Tar Reforming, & Quench	$18,700,000
Compression & Sulfur Removal	$16,000,000
Tar Reforming Catalyst Regeneration, Shift, and PSA	$22,200,000
Hydrogen Compression	$2,700,000
Steam System and Power Generation	$14,400,000
Cooling Water and Other Utilities	$3,600,000
Total Installed Equipment Cost	$96,800,000
Indirect Costs	49,400,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$146,100,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.184
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	47%
Gasifier Efficiency - HHV	70.14%
Gasifier Efficiency - LHV	69.83%
Overall Plant Efficiency - HHV	52.9%
Overall Plant Efficiency - LHV	47.4%

Operating Costs (cents/kg hydrogen)	
Feedstock	40.8
Natural Gas	4.8
Catalysts	0.6
Olivine	6.8
Other Raw Materials	0.8
Waste Disposal	1.3
Electricity	5.8
Fixed Costs	17.4
Capital Depreciation	12.9
Average Income Tax	10.1
Average Return on Investment	24.2

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$2,700,000
Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$500,000
Waste Disposal	$700,000
Electricity	$3,300,000
Fixed Costs	$9,900,000
Capital Depreciation	$7,300,000
Average Income Tax	$5,800,000
Average Return on Investment	$13,800,000

Total Plant Electricity Usage (KW)	40276
Electricity Produced Onsite (KW)	-32052
Electricity Purchased from Grid (KW)	8224
Plant Electricity Use (KWh/kg H2)	5.96
Plant Steam Use (kg steam/kg H2)	19.2

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Goal Case - No hydrogen recycle to the PSA (Case EE)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg)	**$1.21**	$8.44	($/GJ H2, HHV basis)
		$10.08	($/GJ H2, LHV basis)
Hydrogen Production at operating capacity (MM kg / year)	61.9	74.8	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	80.2	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$17,600,000
Compression & Sulfur Removal	$16,100,000
Tar Reforming Catalyst Regeneration, Shift, and PSA	$23,100,000
Hydrogen Compression	$2,900,000
Steam System and Power Generation	$14,300,000
Cooling Water and Other Utilities	$3,500,000
Total Installed Equipment Cost	$96,400,000
Indirect Costs	49,100,000
(% of TPI)	33.7%
Total Project Investment (TPI)	$145,500,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.184
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	52%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	54.7%
Overall Plant Efficiency - LHV	49.1%

Operating Costs (cents/kg hydrogen)	
Feedstock	37.4
Natural Gas	9.1
Catalysts	0.6
Olivine	6.2
Other Raw Materials	0.7
Waste Disposal	1.1
Electricity	6.7
Fixed Costs	15.9
Capital Depreciation	11.8
Average Income Tax	9.3
Average Return on Investment	22.1

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$5,600,000
Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$9,900,000
Capital Depreciation	$7,300,000
Average Income Tax	$5,700,000
Average Return on Investment	$13,700,000

Total Plant Electricity Usage (KW)	40564
Electricity Produced Onsite (KW)	-30241
Electricity Purchased from Grid (KW)	10322
Plant Electricity Use (KWh/kg H2)	5.51
Plant Steam Use (kg steam/kg H2)	18.4

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity on Goal Case - Increase PSA cost (Case FF)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.28**

$8.94	($/GJ H2, HHV basis)
$10.67	($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	58.4	70.6	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	75.7	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$17,600,000
Compression & Sulfur Removal	$16,100,000
Tar Reforming Catalyst Regeneration, Shift, and PSA	$30,200,000
Hydrogen Compression	$2,800,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$103,200,000
Indirect Costs	52,600,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$155,800,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.183
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	49%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	53.3%
Overall Plant Efficiency - LHV	47.8%

Operating Costs (cents/kg hydrogen)	
Feedstock	39.7
Natural Gas	5.9
Catalysts	0.6
Olivine	6.6
Other Raw Materials	0.7
Waste Disposal	1.2
Electricity	7.1
Fixed Costs	17.6
Capital Depreciation	13.4
Average Income Tax	10.4
Average Return on Investment	25.0

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$3,400,000
Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$10,300,000
Capital Depreciation	$7,800,000
Average Income Tax	$6,100,000
Average Return on Investment	$14,600,000
Total Plant Electricity Usage (KW)	40259
Electricity Produced Onsite (KW)	-29974
Electricity Purchased from Grid (KW)	10284
Plant Electricity Use (KWh/kg H2)	5.79
Plant Steam Use (kg steam/kg H2)	19.5

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Goal Case - Increase in natural gas price (Case GG)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.26** — $8.78 ($/GJ H2, HHV basis); $10.49 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	58.4	70.6	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	75.7	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$17,600,000
Compression & Sulfur Removal	$16,100,000
Tar Reforming Catalyst Regeneration, Shift, and PSA	$22,600,000
Hydrogen Compression	$2,800,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$95,600,000
Indirect Costs	48,800,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$144,400,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.184
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	49%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	53.3%
Overall Plant Efficiency - LHV	47.8%

Operating Costs (cents/kg hydrogen)	
Feedstock	39.7
Natural Gas	7.8
Catalysts	0.6
Olivine	6.6
Other Raw Materials	0.7
Waste Disposal	1.2
Electricity	7.1
Fixed Costs	16.8
Capital Depreciation	12.3
Average Income Tax	9.8
Average Return on Investment	23.4

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$4,500,000
Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$9,800,000
Capital Depreciation	$7,200,000
Average Income Tax	$5,700,000
Average Return on Investment	$13,700,000
Total Plant Electricity Usage (KW)	40259
Electricity Produced Onsite (KW)	-29974
Electricity Purchased from Grid (KW)	10284
Plant Electricity Use (KWh/kg H2)	5.79
Plant Steam Use (kg steam/kg H2)	19.5

Hydrogen Production Process Engineering Analysis

Design Report: Sensivity on Goal Case - Increase in Tar Reformer/Catalyst Regeneration Cost (Case HH)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.27**

$8.88 ($/GJ H2, HHV basis)
$10.60 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	58.4	70.6	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	75.7	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$17,600,000
Compression & Sulfur Removal	$16,100,000
Tar Reforming Catalyst Regeneration, Shift, and PSA	$28,600,000
Hydrogen Compression	$2,800,000
Steam System and Power Generation	$14,200,000
Cooling Water and Other Utilities	$3,400,000
Total Installed Equipment Cost	$101,600,000
Indirect Costs	51,900,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$153,500,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.183
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	49%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	53.3%
Overall Plant Efficiency - LHV	47.8%

Operating Costs (cents/kg hydrogen)	
Feedstock	39.7
Natural Gas	5.9
Catalysts	0.6
Olivine	6.6
Other Raw Materials	0.7
Waste Disposal	1.2
Electricity	7.1
Fixed Costs	17.4
Capital Depreciation	13.2
Average Income Tax	10.3
Average Return on Investment	24.6

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$3,400,000
Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$4,100,000
Fixed Costs	$10,200,000
Capital Depreciation	$7,700,000
Average Income Tax	$6,000,000
Average Return on Investment	$14,400,000

Total Plant Electricity Usage (KW)	40259
Electricity Produced Onsite (KW)	-29974
Electricity Purchased from Grid (KW)	10284
Plant Electricity Use (KWh/kg H2)	5.79
Plant Steam Use (kg steam/kg H2)	19.5

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Goal Case - Increase steam to shift (Case II)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.28** — $8.90 ($/GJ H2, HHV basis); $10.63 ($/GJ H2, LHV basis)

Hydrogen Production at operating capacity (MM kg / year)	59.5	71.9	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	77.1	2,116	(dry tons / day) at operating capacity
Delivered Feedstock Cost $/Dry US Ton	$30		
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$17,400,000
Compression & Sulfur Removal	$16,100,000
Tar Reforming Catalyst Regeneration, Shift, and PSA	$23,500,000
Hydrogen Compression	$2,800,000
Steam System and Power Generation	$14,500,000
Cooling Water and Other Utilities	$3,100,000
Total Installed Equipment Cost	$96,300,000
Indirect Costs	49,100,000
(% of TPI)	33.7%
Total Project Investment (TPI)	$145,500,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.184
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	50%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	52.0%
Overall Plant Efficiency - LHV	46.5%

Operating Costs (cents/kg hydrogen)	
Feedstock	38.9
Natural Gas	7.6
Catalysts	0.6
Olivine	6.5
Other Raw Materials	0.6
Waste Disposal	1.2
Electricity	10.6
Fixed Costs	16.6
Capital Depreciation	12.3
Average Income Tax	9.7
Average Return on Investment	23.1

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$4,500,000
Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$6,300,000
Fixed Costs	$9,900,000
Capital Depreciation	$7,300,000
Average Income Tax	$5,800,000
Average Return on Investment	$13,700,000

Total Plant Electricity Usage (KW)	40065
Electricity Produced Onsite (KW)	-24271
Electricity Purchased from Grid (KW)	15793
Plant Electricity Use (KWh/kg H2)	5.66
Plant Steam Use (kg steam/kg H2)	23.6

Hydrogen Production Process Engineering Analysis

Design Report: Sensitivity for Goal Case- Decrease steam to shift (Case JJ)
2000 Dry Metric Tonnes Biomass per Day
BCL Gasifier, Tar Reformer, Sulfur Removal, HTS & LTS, PSA, Steam-Power Cycle
All Values in 2002$

Minimum Hydrogen Selling Price ($/kg) **$1.22** — $8.55 ($/GJ H2, HHV basis); $10.21 ($/GJ H2, LHV basis)

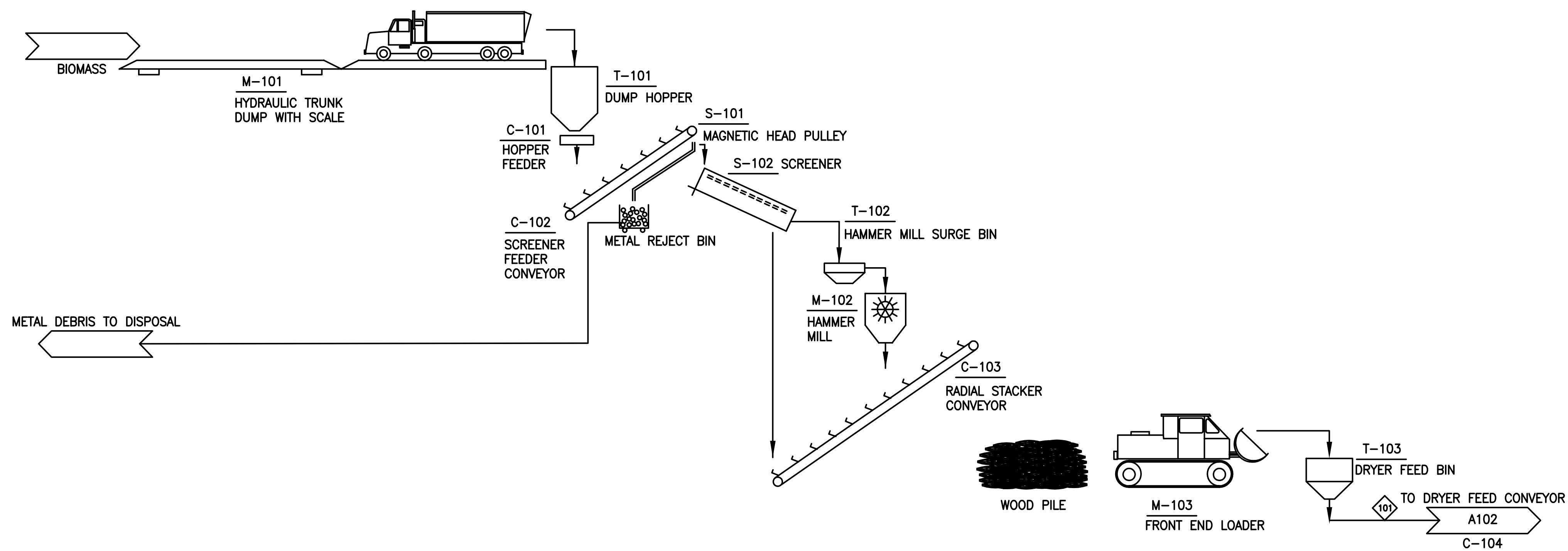
Hydrogen Production at operating capacity (MM kg / year)	56.9	68.7	(Million SCF / day)
Hydrogen Yield (kg / Dry US Ton Feedstock)	73.6	2,116	(dry tons / day)
Delivered Feedstock Cost $/Dry US Ton	$30		at operating capacity
Internal Rate of Return (After-Tax)	10%		
Equity Percent of Total Investment	100%		

Capital Costs	
Feed Handling & Drying	$18,900,000
Gasification, Tar Reforming, & Quench	$17,600,000
Compression & Sulfur Removal	$16,100,000
Tar Reforming Catalyst Regeneration, Shift, and PSA	$22,000,000
Hydrogen Compression	$2,700,000
Steam System and Power Generation	$13,900,000
Cooling Water and Other Utilities	$3,600,000
Total Installed Equipment Cost	$94,800,000
Indirect Costs	48,300,000
(% of TPI)	33.8%
Total Project Investment (TPI)	$143,100,000
Loan Rate	N/A
Term (years)	N/A
Capital Charge Factor	0.183
Maximum Yields (100% of Theoretical) based on composition	
Theoretical Hydrogen Production (MM kg/yr)	119.7
Theoretical Yield (kg/dry ton)	155.0
Current Yield (Actual/Theoretical)	48%
Gasifier Efficiency - HHV	72.14%
Gasifier Efficiency - LHV	71.78%
Overall Plant Efficiency - HHV	53.7%
Overall Plant Efficiency - LHV	48.1%

Operating Costs (cents/kg hydrogen)	
Feedstock	40.7
Natural Gas	3.2
Catalysts	0.6
Olivine	6.8
Other Raw Materials	0.8
Waste Disposal	1.2
Electricity	5.7
Fixed Costs	17.2
Capital Depreciation	12.7
Average Income Tax	9.9
Average Return on Investment	23.6

Operating Costs ($/yr)	
Feedstock	$23,200,000
Natural Gas	$1,800,000
Catalysts	$400,000
Olivine	$3,800,000
Other Raw Matl. Costs	$400,000
Waste Disposal	$700,000
Electricity	$3,300,000
Fixed Costs	$9,800,000
Capital Depreciation	$7,200,000
Average Income Tax	$5,600,000
Average Return on Investment	$13,400,000
Total Plant Electricity Usage (KW)	40210
Electricity Produced Onsite (KW)	-32059
Electricity Purchased from Grid (KW)	8151
Plant Electricity Use (KWh/kg H2)	5.95
Plant Steam Use (kg steam/kg H2)	17.8

Appendix C: Current Design Process Flow Diagrams



COMPONENT	UNITS	101
Total Flow	lb/hr	367,437
Temperature	F	60
Pressure	Psia	14.70
Vapor Fraction		0.00
Hydrogen	lb/hr	
Water	lb/hr	183,718
Carbon Monoxide	lb/hr	
Nitrogen	lb/hr	
Oxygen	lb/hr	
Argon	lb/hr	
Carbon Dioxide	lb/hr	
Hydrogen Sulfide (H2S)	lb/hr	
SO2	lb/hr	
Ammonia (NH3)	lb/hr	
NO2	lb/hr	
Methane	lb/hr	
isobutane	lb/hr	
n-butane	lb/hr	
ethane (C2H6)	lb/hr	
ethylene (C2H4)	lb/hr	
acetylene (C2H2)	lb/hr	
C3H8	lb/hr	
Pentane +	lb/hr	
Benzene (C6H6)	lb/hr	
Tar (C10H8)	lb/hr	
Carbon (Solid)	lb/hr	
Sulfur (Solid)	lb/hr	
Olivine (Solid)	lb/hr	
MgO (Solid)	lb/hr	
Ash	lb/hr	0
Char	lb/hr	0
Wood	lb/hr	183,718
Enthalpy Flow	MMBTU/hr	-1699.4
Average Density	lb/ft^3	47.44

Heat Stream No.	MM BTU/hr	Work Stream No.	HP

Eq. No.	Equipment Name	Req.	Spare	Equipment Type
C-101	Hopper Feeder	4		VIBRATING-FEEDER
C-102	Screener Feeder Conveyor	2		BELT
C-103	Radial Stacker Conveyor	2		BELT
M-101	Hydraulic Truck Dump with Scale	4		TRUCK-SCALE
M-102	Hammermill	2		
M-103	Front End Loaders	3		LOADER
S-101	Magnetic Head Pulley	2		MAGNET
S-102	Screener	2		SCREEN
T-101	Dump Hopper	4		LIVE-BTM-BIN
T-102	Hammermill Surge Bin	1		LIVE-BTM-BIN
T-103	Dryer Feed Bin	2		LIVE-BTM-BIN

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SECTION A100
FEED HANDLING & DRYING

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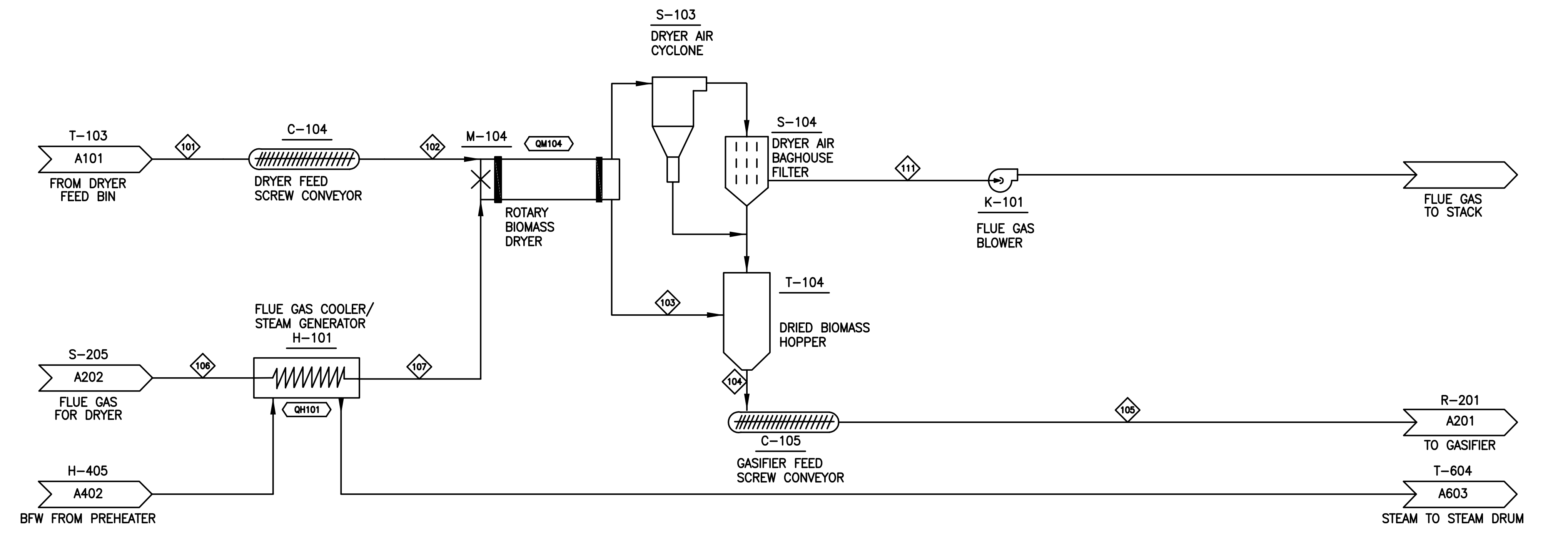

COMPONENT	UNITS	101	102	103	104	105	106	107	111
Total Flow	lb/hr	367,437	367,437	208,771	208,771	208,771	480,864	480,864	639,530
Temperature	F	60	60	220	220	220	1,800	1,791	250
Pressure	Psia	14.70	15.70	15.70	15.70	25.00	22.00	22.00	14.70
Vapor Fraction		0.00	0.00	0.00	0.00	0.00	1.00	1.00	1.00
Hydrogen	lb/hr								
Water	lb/hr	183,718	183,718	25,053	25,053	25,053	30,818	30,818	189,484
Carbon Monoxide	lb/hr								
Nitrogen	lb/hr						327,435	327,435	327,435
Oxygen	lb/hr						10,722	10,722	10,722
Argon	lb/hr						5,584	5,584	5,584
Carbon Dioxide	lb/hr						106,277	106,277	106,277
Hydrogen Sulfide (H2S)	lb/hr								
SO2	lb/hr						27	27	27
Ammonia (NH3)	lb/hr								
NO2	lb/hr								
Methane	lb/hr								
isobutane	lb/hr								
n-butane	lb/hr								
ethane (C2H6)	lb/hr								
ethylene (C2H4)	lb/hr								
acetylene (C2H2)	lb/hr								
C3H8	lb/hr								
Pentane +	lb/hr								
Benzene (C6H6)	lb/hr								
Tar (C10H8)	lb/hr								
Carbon (Solid)	lb/hr								
Sulfur (Solid)	lb/hr								
Olivine (Solid)	lb/hr								
MgO (Solid)	lb/hr								
Ash	lb/hr	0	0	0	0	0			
Char	lb/hr	0	0	0	0	0			
Wood	lb/hr	183,718	183,718	183,718	183,718	183,718			
Enthalpy Flow	MMBTU/hr	-1699.4	-1699.4	-584.6	-584.6	-584.6	-353.9	-355.3	-1470.1
Average Density	lb/ft^3	47.44	47.44	44.12	44.12	44.12	0.03	0.03	0.05

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QH101	1.37		
QM104	-208.40		

Eq. No.	Equipment Name	Req.	Spare	Equipment Type
C-104	Dryer Feed Screw Conveyor	2		SCREW
C-105	Gasifier Feed Screw Conveyor	2		SCREW
H-101	Flue Gas Cooler / Steam Generator #3	1		SHELL-TUBE
K-101	Flue Gas Blower	2		CENTRIFUGAL
M-104	Rotary Biomass Dryer	2		ROTARY-DRUM
S-103	Dryer Air Cyclone	2		GAS CYCLONE
S-104	Dryer Air Baghouse Filter	2		FABRIC-FILTER
T-104	Dried Biomass Hopper	2		VERTICAL-VESSEL

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SECTION A100
FEED HANDLING & DRYING

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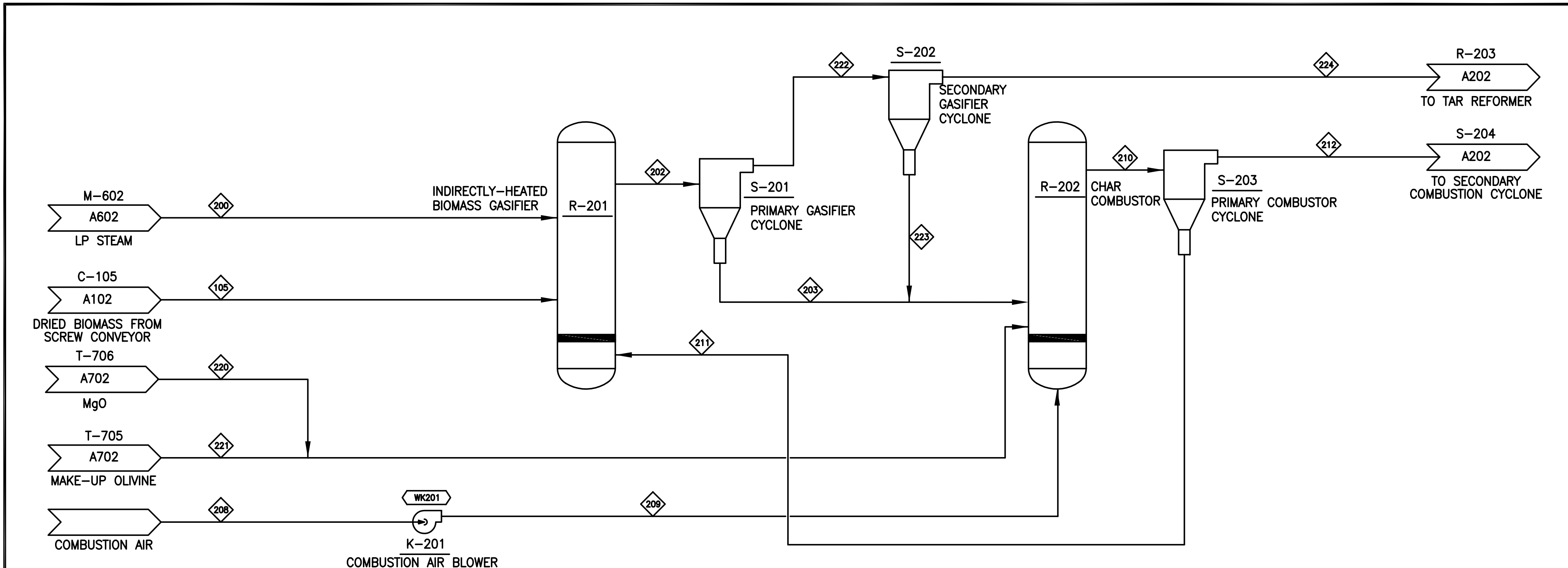

COMPONENT	UNITS	105	200	202	203	208	209	210	211	212	220	221	222	223	224
Total Flow	lb/hr	208,771	73,120	5,228,878	4,982,394	442,163	442,163	5,434,493	4,946,987	487,506	7	5,440	246,483	4,489	241,995
Temperature	F	220	260	1,598	1,598	90	185	1,800	1,800	1,800	60	60	1,598	1,598	1,598
Pressure	Psia	25.00	25.00	23.00	22.00	14.70	22.00	22.00	25.00	22.00	25.00	25.00	22.00	22.00	22.00
Vapor Fraction		0.00	1.00	1.00	0.00	1.00	1.00	1.00	0.00	1.00	0.00	0.00	1.00	0.00	1.00
Hydrogen	lb/hr			3,093									3,093		3,093
Water	lb/hr	25,053	73,120	98,172		8,627	8,627	30,818		30,818			98,172		98,172
Carbon Monoxide	lb/hr			75,994									75,994		75,994
Nitrogen	lb/hr					327,414	327,414	327,435		327,435					
Oxygen	lb/hr					100,320	100,320	10,722		10,722					
Argon	lb/hr					5,584	5,584	5,584		5,584					
Carbon Dioxide	lb/hr			36,216		218	218	106,277		106,277			36,216		36,216
Hydrogen Sulfide (H2S)	lb/hr			161									161		161
SO2	lb/hr							27		27					
Ammonia (NH3)	lb/hr			355									355		355
NO2	lb/hr														
Methane	lb/hr			15,850									15,850		15,850
isobutane	lb/hr														
n-butane	lb/hr														
ethane (C2H6)	lb/hr			565									565		565
ethylene (C2H4)	lb/hr			7,848									7,848		7,848
acetylene (C2H2)	lb/hr			683									683		683
C3H8	lb/hr														
Pentane +	lb/hr														
Benzene (C6H6)	lb/hr			640									640		640
Tar (C10H8)	lb/hr			1,919									1,919		1,919
Carbon (Solid)	lb/hr														
Sulfur (Solid)	lb/hr														
Olivine (Solid)	lb/hr			4,946,898	4,941,951			4,951,850	4,946,898	4,952	7	5,440	4,947	4,452	495
MgO (Solid)	lb/hr														
Ash	lb/hr	0		0	0			1,779	89	1,690			0	0	0
Char	lb/hr	0		40,484	40,443								40	36	4
Wood	lb/hr	183,718		0	0								0	0	0
Enthalpy Flow	MMBTU/hr	-584.6	-416.4	1192.8	1879.7	-49.3	-39.1	1842.3	2193.8	-351.6	0.0	0.0	-686.9	1.7	-688.6
Average Density	lb/ft^3	44.12	0.06	160.21	160.19	0.07	0.09	159.53	159.50	159.53	165.39	165.39	160.21	160.19	160.21

Heat Stream No.	MM BTU/hr		Work Stream No.	HP
			WK201	4023.7

Eq. No.	Equipment Name	Req.	Spare	Equipment Type
K-201	Combustion Air Blower	2		CENTRIFUGAL
R-201	Indirectly-heated Biomass Gasifier	2		VERTICAL-VESSEL
R-202	Char Combustor	2		VERTICAL-VESSEL
S-201	Primary Gasifier Cyclone	2		GAS CYCLONE
S-202	Secondary Gasifier Cyclone	2		GAS CYCLONE
S-203	Primary Combustor Cyclone	2		GAS CYCLONE

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SECTION A200
GASIFICATION & TAR REFORMING

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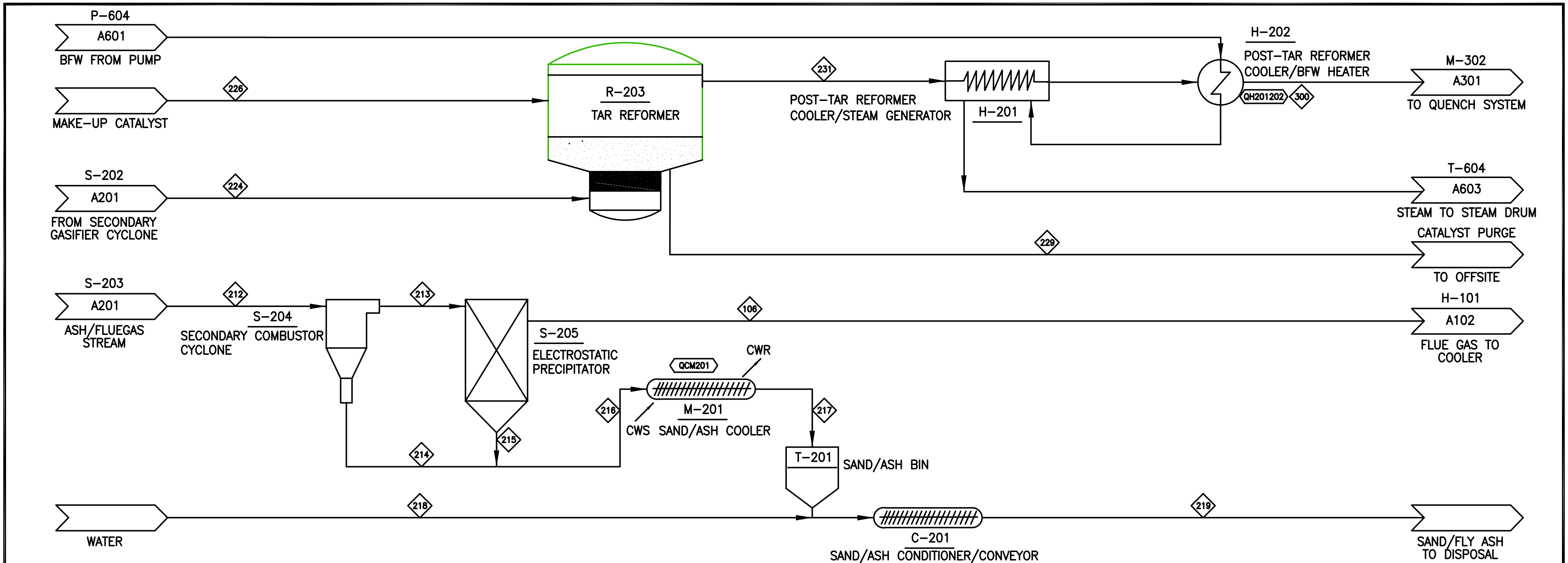

COMPONENT	UNITS	106	212	213	214	215	216	217	218	219	224	226	229	231	300
Total Flow	lb/hr	480,864	487,506	480,870	6,635	7	6,642	6,642	738	7,380	241,995	60	60	241,995	241,995
Temperature	F	1,800	1,800	1,800	1,800	1,800	1,800	300	60	208	1,598	60	1,383	1,383	300
Pressure	Psia	22.00	22.00	22.00	22.00	22.00	22.00	14.70	14.70	14.70	22.00	22.00	19.00	19.00	17.00
Vapor Fraction		1.00	1.00	1.00	0.00	0.00	0.00	0.00	0.00	0.00	1.00	0.00	0.00	1.00	1.00
Hydrogen	lb/hr										3,093			8,786	8,786
Water	lb/hr	30,818	30,818	30,818					738	738	98,172			62,329	62,329
Carbon Monoxide	lb/hr										75,994			60,264	60,264
Nitrogen	lb/hr	327,435	327,435	327,435										204	204
Oxygen	lb/hr	10,722	10,722	10,722											
Argon	lb/hr	5,584	5,584	5,584											
Carbon Dioxide	lb/hr	106,277	106,277	106,277							36,216			92,356	92,356
Hydrogen Sulfide (H2S)	lb/hr										161			161	161
SO2	lb/hr	27	27	27											
Ammonia (NH3)	lb/hr										355			106	106
NO2	lb/hr														
Methane	lb/hr										15,850			12,680	12,680
isobutane	lb/hr														
n-butane	lb/hr														
ethane (C2H6)	lb/hr										565			57	57
ethylene (C2H4)	lb/hr										7,848			3,924	3,924
acetylene (C2H2)	lb/hr										683			341	341
C3H8	lb/hr														
Pentane +	lb/hr														
Benzene (C6H6)	lb/hr										640			192	192
Tar (C10H8)	lb/hr										1,919			96	96
Carbon (Solid)	lb/hr														
Sulfur (Solid)	lb/hr														
Olivine (Solid)	lb/hr		4,952	5	4,947	5	4,952	4,952		4,952	495	60	60	495	495
MgO (Solid)	lb/hr														
Ash	lb/hr		1,690	2	1,688	2	1,690	1,690		1,690	0			0	0
Char	lb/hr										4			4	4
Wood	lb/hr										0			0	0
Enthalpy Flow	MMBTU/hr	-353.9	-351.6	-353.9	2.3	0.0	2.3	-0.3	-5.1	-5.4	-688.6	0.0	0.0	-688.6	-815.9
Average Density	lb/ft^3	0.03	159.53	159.53	159.50	159.50	159.50	164.58	47.44	209.28	160.21	165.39	160.92	160.93	164.62

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QH201202	127.28		
QCM201	2.63		

Eq. No.	Equipment Name	Req.	Spare	Equipment Type
C-201	Sand/ash Conditioner/Conveyor	1		SCREW
H-201	Post-tar Reformer Cooler / Steam Generator #1	1		SHELL-TUBE
H-202	Post-tar Reformer Cooler / BFW Preheater #2	1		SHELL-TUBE
M-201	Sand/ash Cooler	2		IMMERSED-COIL
R-203	Tar Reformer	1		VERTICAL-VESSEL
S-204	Secondary Combustor Cyclone	2		GAS CYCLONE
S-205	Electrostatic Precipitator	2		MISCELLANEOUS
T-201	Sand/ash Bin	1		FLAT-BTM-STORAGE

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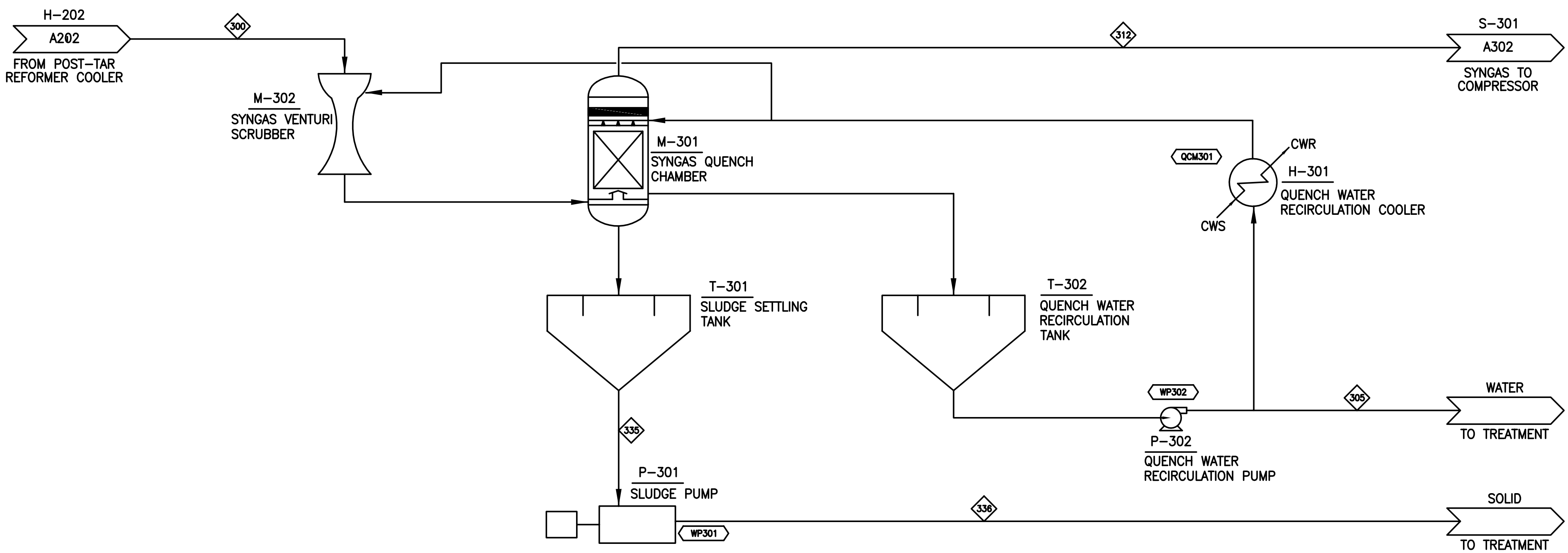

COMPONENT	UNITS	300	305	312	335	336
Total Flow	lb/hr	241,995	20,721	220,277	997	997
Temperature	F	300	140	140	140	140
Pressure	Psia	17.00	15.00	15.00	15.00	34.70
Vapor Fraction		1.00	0.00	1.00	0.00	0.00
Hydrogen	lb/hr	8,786	0	8,786		
Water	lb/hr	62,329	20,492	41,338	499	499
Carbon Monoxide	lb/hr	60,264	6	60,258		
Nitrogen	lb/hr	204	0	204		
Oxygen	lb/hr					
Argon	lb/hr					
Carbon Dioxide	lb/hr	92,356	53	92,303		
Hydrogen Sulfide (H2S)	lb/hr	161	0	161		
SO2	lb/hr					
Ammonia (NH3)	lb/hr	106	22	85		
NO2	lb/hr					
Methane	lb/hr	12,680	14	12,667		
isobutane	lb/hr					
n-butane	lb/hr					
ethane (C2H6)	lb/hr	57	0	56		
ethylene (C2H4)	lb/hr	3,924	10	3,914		
acetylene (C2H2)	lb/hr	341	1	340		
C3H8	lb/hr					
Pentane +	lb/hr					
Benzene (C6H6)	lb/hr	192	31	161		
Tar (C10H8)	lb/hr	96	92	4		
Carbon (Solid)	lb/hr					
Sulfur (Solid)	lb/hr					
Olivine (Solid)	lb/hr	495			495	495
MgO (Solid)	lb/hr					
Ash	lb/hr	0			0	0
Char	lb/hr	4			4	4
Wood	lb/hr	0			0	0
Enthalpy Flow	MMBTU/hr	-815.9	-140.0	-711.6	-3.4	-3.4
Average Density	lb/ft^3	164.62	8.81	0.04	211.02	211.02

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QCM301	37.92	WP301	0.1
		WP302	36.3

Eq. No.	Equipment Name	Req	Spare	Equipment Type
H-301	Quench Water Recirculation Cooler	1		SHELL-TUBE
M-301	Syngas Quench Chamber	1		
M-302	Syngas Venturi Scrubber	1		
P-301	Sludge Pump	1	1	CENTRIFUGAL
P-302	Quench Water Recirculation Pump	1	1	CENTRIFUGAL
T-301	Sludge Settling Tank	1		CLARIFIER
T-302	Quench Water Recirculation Tank	1		HORIZONTAL-VESSEL

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		AA 9-17-04

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SECTION A300

GAS CLEAN-UP & COMPRESSION

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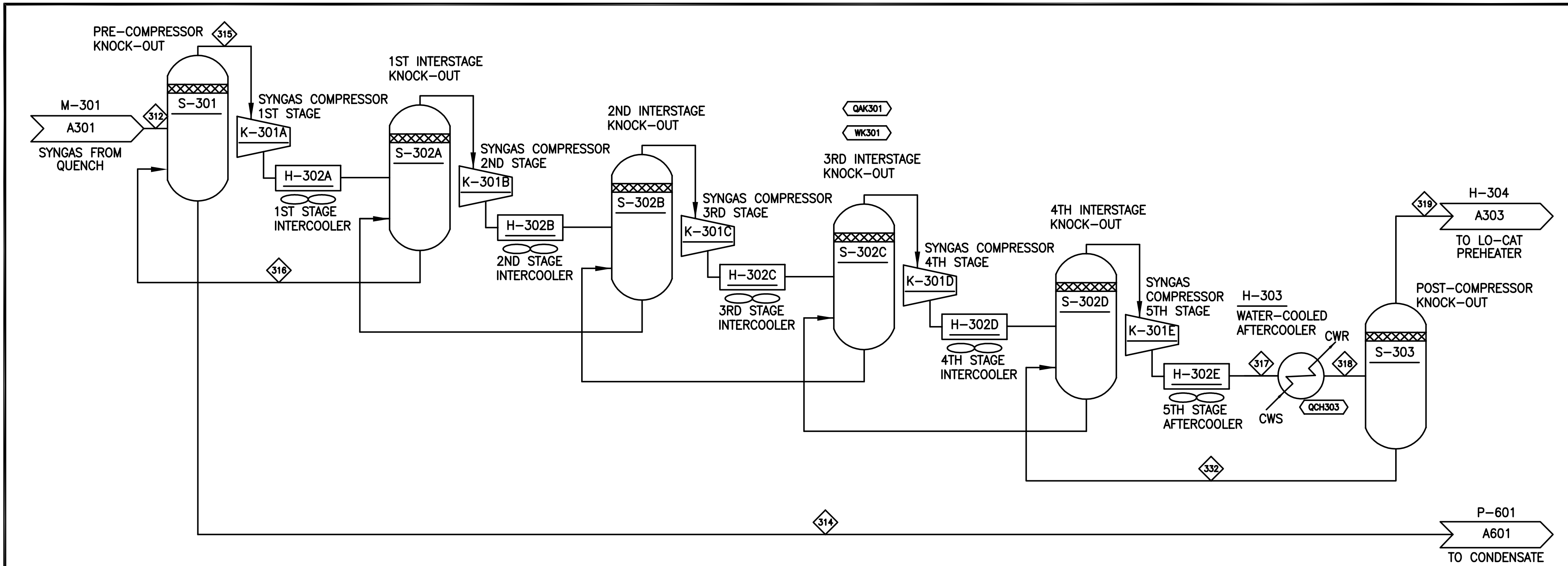

COMPONENT	UNITS	312	314	315	316	317	318	319	332
Total Flow	lb/hr	220,277	40,882	220,009	40,004	180,005	180,005	179,394	611
Temperature	F	140	146	146	140	140	110	110	110
Pressure	Psia	15.00	15.00	15.00	15.00	465.00	460.00	460.00	460.00
Vapor Fraction		1.00	0.00	1.00	0.00	1.00	1.00	1.00	0.00
Hydrogen	lb/hr	8,786	0	8,786	0	8,786	8,786	8,786	0
Water	lb/hr	41,338	40,876	41,045	39,977	1,068	1,068	461	607
Carbon Monoxide	lb/hr	60,258	0	60,258	0	60,258	60,258	60,258	0
Nitrogen	lb/hr	204	0	204	0	204	204	204	0
Oxygen	lb/hr								
Argon	lb/hr								
Carbon Dioxide	lb/hr	92,303	1	92,305	3	92,303	92,303	92,302	0
Hydrogen Sulfide (H2S)	lb/hr	161	0	161	0	161	161	161	0
SO2	lb/hr								
Ammonia (NH3)	lb/hr	85	5	107	24	83	83	79	4
NO2	lb/hr								
Methane	lb/hr	12,667	0	12,667	0	12,667	12,667	12,667	0
isobutane	lb/hr								
n-butane	lb/hr								
ethane (C2H6)	lb/hr	56	0	56	0	56	56	56	0
ethylene (C2H4)	lb/hr	3,914	0	3,914	0	3,914	3,914	3,914	0
acetylene (C2H2)	lb/hr	340	0	340	0	340	340	340	0
C3H8	lb/hr								
Pentane +	lb/hr								
Benzene (C6H6)	lb/hr	161	0	161	0	161	161	161	0
Tar (C10H8)	lb/hr	4	0	4	0	4	4	4	0
Carbon (Solid)	lb/hr								
Sulfur (Solid)	lb/hr								
Olivine (Solid)	lb/hr								
MgO (Solid)	lb/hr								
Ash	lb/hr								
Char	lb/hr								
Wood	lb/hr								
Enthalpy Flow	MMBTU/hr	-711.6	-278.7	-709.3	-272.9	-480.8	-483.8	-479.6	-4.2
Average Density	lb/ft^3	0.04	45.77	0.04	45.48	1.35	1.42	1.41	46.43

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QAK301	108.95	WK301	25373.7
QAK301A	41.13	WK301A	5541.9
QAK301B	23.30	WK301B	5410.3
QAK301C	16.66	WK301C	4894.9
QAK301D	14.64	WK301D	4863.9
QAK301E	13.22	WK301E	4662.7
QCH303	2.94		

Eq. No.	Equipment Name	Req.	Spare	Equipment Type
H-302	Syngas Compressor Intercoolers	5		AIR-COOLED EXCHANGER
H-303	Water-cooled Aftercooler	1		SHELL-TUBE
K-301	Syngas Compressor	1		CENTRIFUGAL
S-301	Pre-compressor Knock-out	1		KNOCK-OUT DRUM
S-302	Syngas Compressor Interstage Knock-outs	4		KNOCK-OUT DRUM
S-303	Post-compressor Knock-out	1		KNOCK-OUT DRUM

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GAS CLEAN-UP & COMPRESSION

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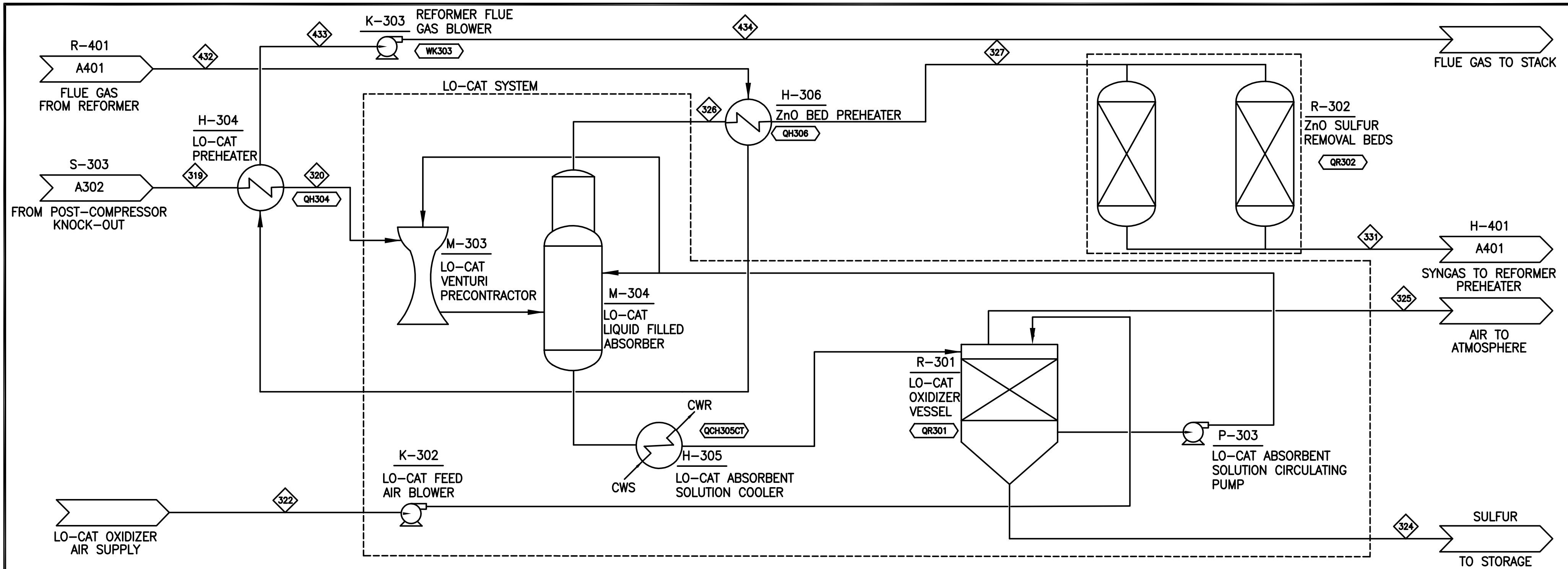

COMPONENT	UNITS	319	320	322	324	325	326	327	331	432	433	434
Total Flow	lb/hr	179,394	179,394	359	148	368	179,237	179,237	179,235	534,677	534,677	534,677
Temperature	F	110	120	90	110	110	120	707	682	1,183	250	256
Pressure	Psia	460.00	455.00	14.70	14.70	14.70	450.00	445.00	440.00	14.70	14.34	14.70
Vapor Fraction		1.00	1.00	1.00	0.00	0.71	1.00	1.00	1.00	1.00	1.00	1.00
Hydrogen	lb/hr	8,786	8,786				8,786	8,786	8,786			
Water	lb/hr	461	461	7		90	461	461	463	58,606	58,606	58,606
Carbon Monoxide	lb/hr	60,258	60,258				60,258	60,258	60,258			
Nitrogen	lb/hr	204	204	266		266	204	204	204	225,825	225,825	225,825
Oxygen	lb/hr			81		7				5,936	5,936	5,936
Argon	lb/hr			5		5				3,847	3,847	3,847
Carbon Dioxide	lb/hr	92,302	92,302	0		0	92,302	92,302	92,302	240,459	240,459	240,459
Hydrogen Sulfide (H2S)	lb/hr	161	161				3	3				
SO2	lb/hr									0	0	0
Ammonia (NH3)	lb/hr	79	79				79	79	79			
NO2	lb/hr									5	5	5
Methane	lb/hr	12,667	12,667				12,667	12,667	12,667			
isobutane	lb/hr											
n-butane	lb/hr											
ethane (C2H6)	lb/hr	56	56				56	56	56			
ethylene (C2H4)	lb/hr	3,914	3,914				3,914	3,914	3,914			
acetylene (C2H2)	lb/hr	340	340				340	340	340			
C3H8	lb/hr											
Pentane +	lb/hr											
Benzene (C6H6)	lb/hr	161	161				161	161	161			
Tar (C10H8)	lb/hr	4	4				4	4	4			
Carbon (Solid)	lb/hr											
Sulfur (Solid)	lb/hr				148							
Olivine (Solid)	lb/hr											
MgO (Solid)	lb/hr											
Ash	lb/hr											
Char	lb/hr											
Wood	lb/hr											
Enthalpy Flow	MMBTU/hr	-479.6	-478.8	0.0	0.0	-0.6	-478.8	-431.6	-433.7	-1097.9	-1240.1	-1239.2
Average Density	lb/ft^3	1.41	1.37	0.07	129.14	0.08	1.36	0.66	0.66	0.03	0.06	0.06

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QCH305CT	0.04	WK303	327.4
QH304	-0.77		
QH306	-47.21		
QR301	0.52		
QR302	2.13		

Eq. No.	Equipment Name	Req.	Spare	Equipment Type
H-304	LO-CAT Preheater	1		SHELL-TUBE
H-305	LO-CAT Absorbent Solution Cooler	1		SHELL-TUBE
H-306	ZnO Bed Preheater	1		SHELL-TUBE
K-302	LO-CAT Feed Air Blower	1		CENTRIFUGAL
K-303	Reformer Flue Gas Blower	1		CENTRIFUGAL
M-303	LO-CAT Venturi Precontactor	1		
M-304	LO-CAT Liquid-filled Absorber	1		ABSORBER
P-303	LO-CAT Absorbent Solution Circulating Pump	1	1	CENTRIFUGAL
R-301	LO-CAT Oxidizer Vessel	1		VERTICAL-VESSEL
R-302	ZnO Sulfur Removal Beds	2		VERTICAL-VESSEL

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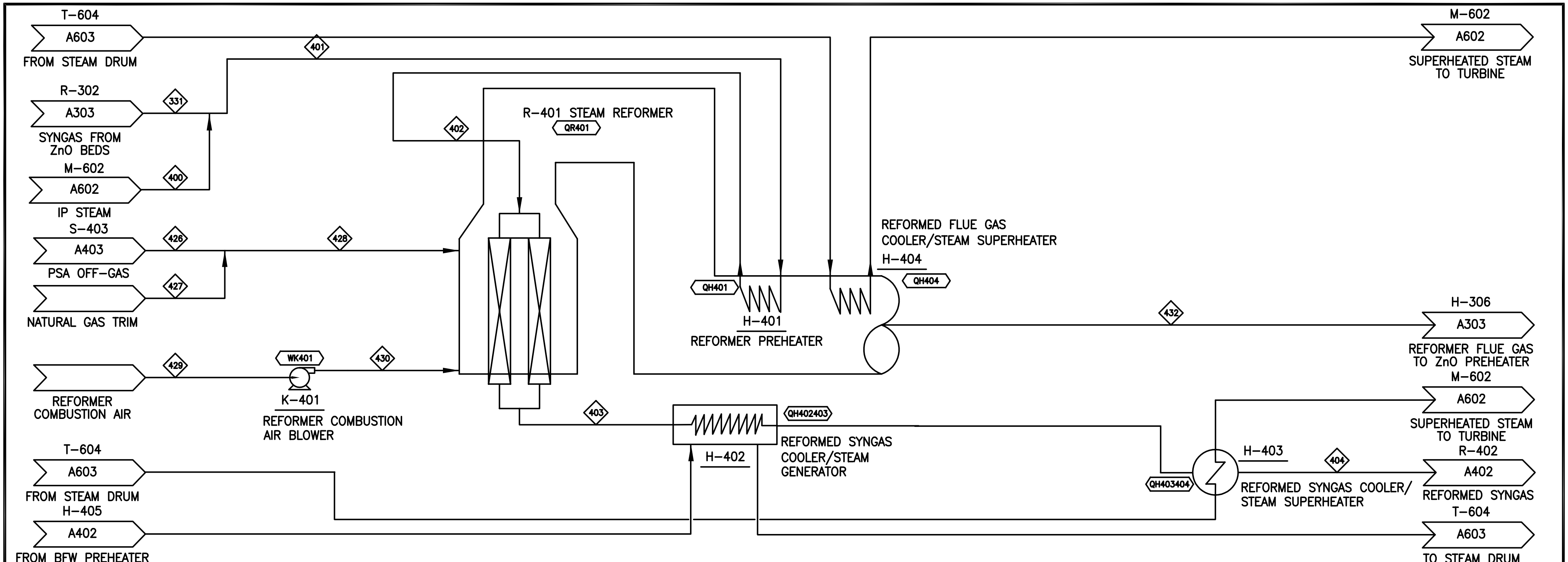

COMPONENT	UNITS	331	400	401	402	403	404	426	427	428	429	430	432
Total Flow	lb/hr	179,235	175,189	354,424	354,424	354,424	354,424	228,431	1,669	230,099	304,578	304,578	534,677
Temperature	F	682	731	700	968	1,562	662	108	60	108	90	95	1,183
Pressure	Psia	440.00	440.00	440.00	435.00	405.00	400.00	14.70	14.70	14.70	14.70	15.06	14.70
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
Hydrogen	lb/hr	8,786		8,786	8,786	13,528	13,528	3,814		3,814			
Water	lb/hr	463	175,189	175,652	175,652	153,119	153,119	828		828	5,943	5,943	58,606
Carbon Monoxide	lb/hr	60,258		60,258	60,258	65,404	65,404	2,240		2,240			
Nitrogen	lb/hr	204		204	204	260	260	260	30	290	225,534	225,534	225,825
Oxygen	lb/hr					0	0				69,104	69,104	5,936
Argon	lb/hr										3,847	3,847	3,847
Carbon Dioxide	lb/hr	92,302		92,302	92,302	115,783	115,783	214,968	21	214,990	150	150	240,459
Hydrogen Sulfide (H2S)	lb/hr								0	0			
SO2	lb/hr												0
Ammonia (NH3)	lb/hr	79		79	79	11	11	2		2			
NO2	lb/hr					0	0						5
Methane	lb/hr	12,667		12,667	12,667	6,318	6,318	6,318	1,479	7,797			
isobutane	lb/hr					0	0	0	6	6			
n-butane	lb/hr					0	0	0	6	6			
ethane (C2H6)	lb/hr	56		56	56	0	0	0	91	91			
ethylene (C2H4)	lb/hr	3,914		3,914	3,914	0	0	0		0			
acetylene (C2H2)	lb/hr	340		340	340	0	0	0		0			
C3H8	lb/hr					0	0	0	22	22			
Pentane +	lb/hr					0	0		14	14			
Benzene (C6H6)	lb/hr	161		161	161	0	0						
Tar (C10H8)	lb/hr	4		4	4	0	0						
Carbon (Solid)	lb/hr												
Sulfur (Solid)	lb/hr												
Olivine (Solid)	lb/hr												
MgO (Solid)	lb/hr												
Ash	lb/hr												
Char	lb/hr												
Wood	lb/hr												
Enthalpy Flow	MMBTU/hr	-433.7	-962.0	-1395.8	-1348.1	-1189.4	-1358.4	-846.3	-3.2	-849.5	-34.0	-33.6	-1097.9
Average Density	lb/ft^3	0.66	0.65	0.65	0.52	0.32	0.57	0.08	0.05	0.08	0.07	0.07	0.03

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QH401	-47.63	WK401	154.5
QH402403	168.99		
QH403404	-108.19		
QH404	142.19		
QR401	-158.71		

Eq. No.	Equipment Name	Req	Spare	Equipment Type
H-401	Reformer Feed Preheater	1		SHELL-TUBE
H-402	Reformed Syngas Cooler/Steam Generator #2	1		SHELL-TUBE
H-403	Reformed Syngas Cooler/Steam Superheater #1	1		SHELL-TUBE
H-404	Reformer Flue Gas Cooler/Steam Superheater #2	1		SHELL-TUBE
K-401	Reformer Combustion Air Blow er	1		CENTRIFUGAL
R-401	Steam Reformer	1		VERTICAL-VESSEL

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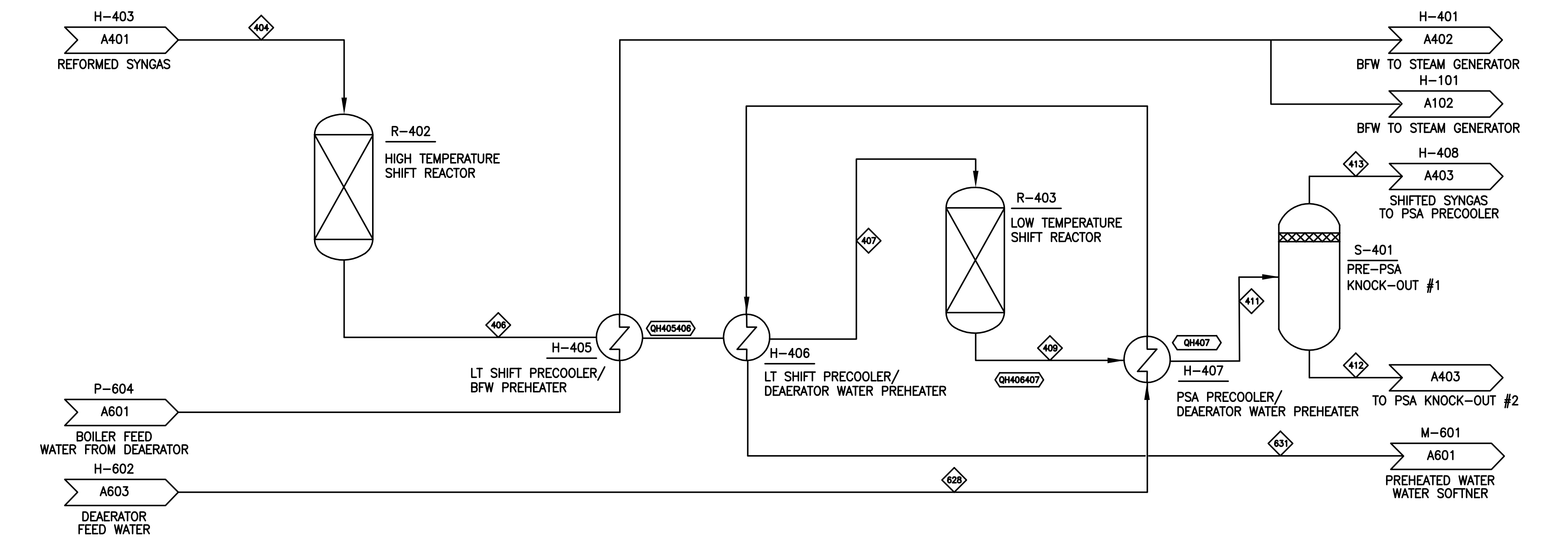

COMPONENT	UNITS	404	406	407	409	411	412	413	628	631
Total Flow	lb/hr	354,424	354,424	354,424	354,424	354,424	0	354,424	349,266	349,266
Temperature	F	662	807	392	453	334		334	101	227
Pressure	Psia	400.00	395.00	390.00	385.00	380.00	380.00	380.00	39.70	19.70
Vapor Fraction		1.00	1.00	1.00	1.00	1.00	0.00	1.00	0.00	0.00
Hydrogen	lb/hr	13,528	16,774	16,774	18,074	18,074		18,074		
Water	lb/hr	153,119	124,112	124,112	112,495	112,495		112,495	349,266	349,266
Carbon Monoxide	lb/hr	65,404	20,303	20,303	2,240	2,240		2,240		
Nitrogen	lb/hr	260	260	260	260	260		260		
Oxygen	lb/hr	0								
Argon	lb/hr									
Carbon Dioxide	lb/hr	115,783	186,646	186,646	215,025	215,025		215,025		
Hydrogen Sulfide (H2S)	lb/hr									
SO2	lb/hr									
Ammonia (NH3)	lb/hr	11	11	11	11	11		11		
NO2	lb/hr	0								
Methane	lb/hr	6,318	6,318	6,318	6,318	6,318		6,318		
isobutane	lb/hr	0	0	0	0	0		0		
n-butane	lb/hr	0	0	0	0	0		0		
ethane (C2H6)	lb/hr	0	0	0	0	0		0		
ethylene (C2H4)	lb/hr	0	0	0	0	0		0		
acetylene (C2H2)	lb/hr	0	0	0	0	0		0		
C3H8	lb/hr	0	0	0	0	0		0		
Pentane +	lb/hr	0								
Benzene (C6H6)	lb/hr	0								
Tar (C10H8)	lb/hr	0								
Carbon (Solid)	lb/hr									
Sulfur (Solid)	lb/hr									
Olivine (Solid)	lb/hr									
MgO (Solid)	lb/hr									
Ash	lb/hr									
Char	lb/hr									
Wood	lb/hr									
Enthalpy Flow	MMBTU/hr	-1358.4	-1358.4	-1433.0	-1433.0	-1454.0		-1454.0	-2375.4	-2331.4
Average Density	lb/ft^3	0.57	0.50	0.75	0.68	0.78		0.78	61.99	59.43

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QH405406	74.57		
QH406407	-44.01		
QH407	21.03		

Eq. No.	Equipment Name	Req.	Spare	Equipment Type
H-405	LT Shift Precooler/BFW Preheater #1	1		SHELL-TUBE
H-406	LT shift Precooler/Deaerator Water Preheater #1	1		SHELL-TUBE
H-407	PSA Precooler / Deaerator Water Preheater #2	1		SHELL-TUBE
R-402	High Temperature Shift Reactor	1		VERTICAL-VESSEL
R-403	Low Temperature Shift Reactor	1		VERTICAL-VESSEL
S-401	Pre-PSA Knock-out #1	1		KNOCK-OUT DRUM

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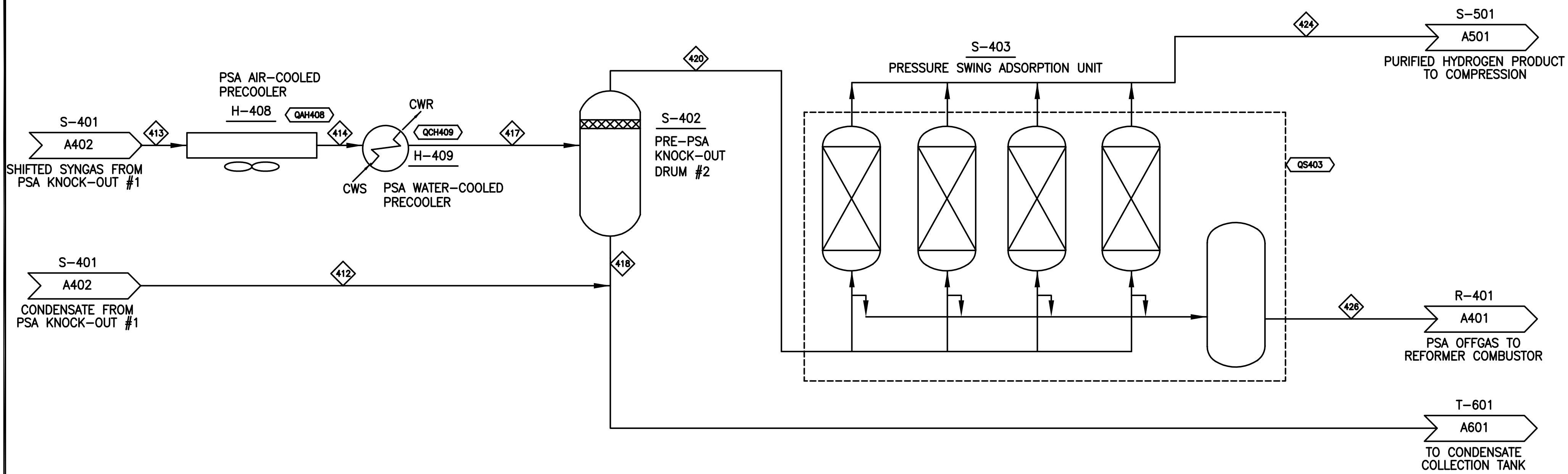

COMPONENT	UNITS	412	413	414	417	418	420	424	426
Total Flow	lb/hr	0	354,424	354,424	354,424	111,734	242,691	14,260	228,431
Temperature	F		334	140	110	110	110	109	108
Pressure	Psia	380.00	380.00	375.00	370.00	370.00	360.00	360.00	14.70
Vapor Fraction		0.00	1.00	0.70	0.70	0.00	1.00	1.00	1.00
Hydrogen	lb/hr		18,074	18,074	18,074	0	18,074	14,260	3,814
Water	lb/hr		112,495	112,495	112,495	111,667	828		828
Carbon Monoxide	lb/hr		2,240	2,240	2,240	0	2,240		2,240
Nitrogen	lb/hr		260	260	260	0	260		260
Oxygen	lb/hr								
Argon	lb/hr								
Carbon Dioxide	lb/hr		215,025	215,025	215,025	57	214,968		214,968
Hydrogen Sulfide (H2S)	lb/hr								
SO2	lb/hr								
Ammonia (NH3)	lb/hr		11	11	11	10	2		2
NO2	lb/hr								
Methane	lb/hr		6,318	6,318	6,318	0	6,318		6,318
isobutane	lb/hr		0	0	0		0		0
n-butane	lb/hr		0	0	0		0		0
ethane (C2H6)	lb/hr		0	0	0	0	0		0
ethylene (C2H4)	lb/hr		0	0	0	0	0		0
acetylene (C2H2)	lb/hr		0	0	0	0	0		0
C3H8	lb/hr		0	0	0		0		0
Pentane +	lb/hr								
Benzene (C6H6)	lb/hr								
Tar (C10H8)	lb/hr								
Carbon (Solid)	lb/hr								
Sulfur (Solid)	lb/hr								
Olivine (Solid)	lb/hr								
MgO (Solid)	lb/hr								
Ash	lb/hr								
Char	lb/hr								
Wood	lb/hr								
Enthalpy Flow	MMBTU/hr		-1454.0	-1603.3	-1611.7	-766.1	-845.6	1.6	-846.3
Average Density	lb/ft^3		0.78	1.42	1.49	46.53	1.00	0.12	0.08

Heat Stream No.	MM BTU/hr		Work Stream No.	HP
QAH408	149.28			
QCH409	8.41			
QS403	-0.93			

Eq. No.	Equipment Name	Req.	Spare	Equipment Type
H-408	PSA Air-cooled Precooler	1		AIR-COOLED EXCHANGER
H-409	PSA Water-cooled Precooler	1		SHELL-TUBE
S-402	Pre-PSA Knock-out #2	1		KNOCK-OUT DRUM
S-403	Pressure Swing Adsorption Unit	1		ABSORBER

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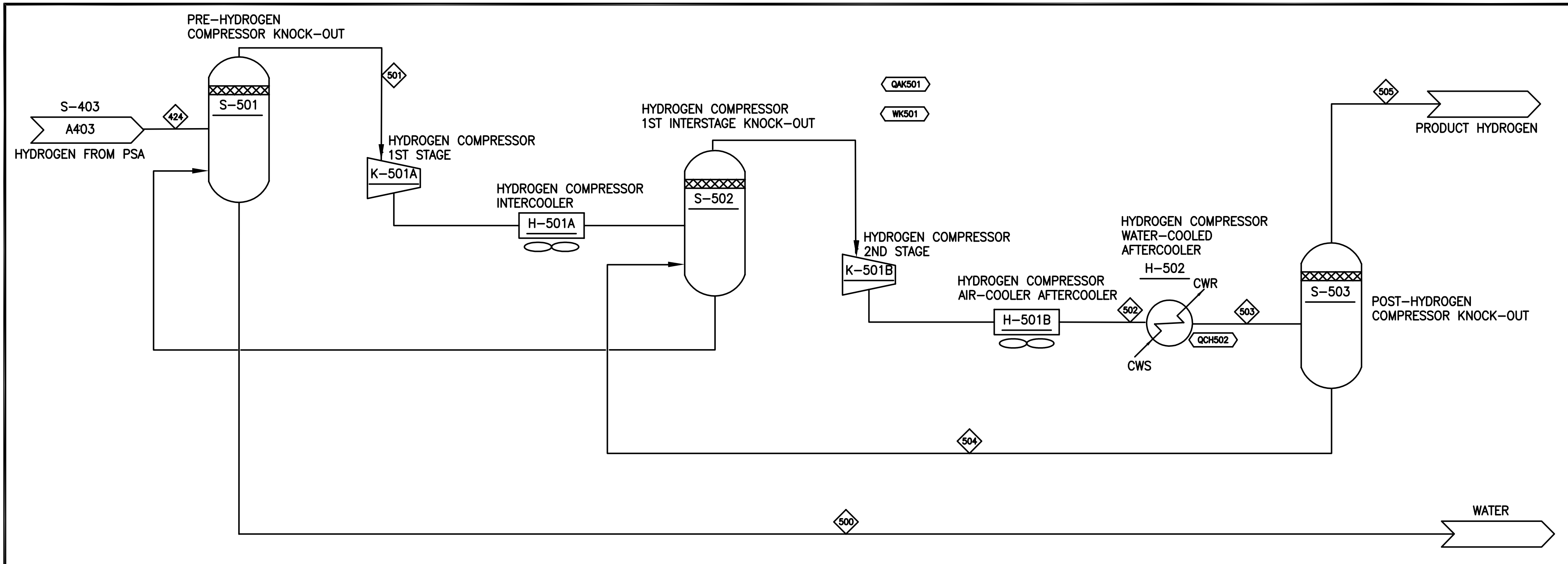

COMPONENT	UNITS	424	500	501	502	503	504	505
Total Flow	lb/hr	14,260	0	14,260	14,260	14,260	0	14,260
Temperature	F	109		109	140	110		110
Pressure	Psia	360.00	360.00	360.00	1019.70	1014.70	1014.70	1014.70
Vapor Fraction		1.00	0.00	1.00	1.00	1.00	0.00	1.00
Hydrogen	lb/hr	14,260		14,260	14,260	14,260		14,260
Water	lb/hr							
Carbon Monoxide	lb/hr							
Nitrogen	lb/hr							
Oxygen	lb/hr							
Argon	lb/hr							
Carbon Dioxide	lb/hr							
Hydrogen Sulfide (H2S)	lb/hr							
SO2	lb/hr							
Ammonia (NH3)	lb/hr							
NO2	lb/hr							
Methane	lb/hr							
isobutane	lb/hr							
n-butane	lb/hr							
ethane (C2H6)	lb/hr							
ethylene (C2H4)	lb/hr							
acetylene (C2H2)	lb/hr							
C3H8	lb/hr							
Pentane +	lb/hr							
Benzene (C6H6)	lb/hr							
Tar (C10H8)	lb/hr							
Carbon (Solid)	lb/hr							
Sulfur (Solid)	lb/hr							
Olivine (Solid)	lb/hr							
MgO (Solid)	lb/hr							
Ash	lb/hr							
Char	lb/hr							
Wood	lb/hr							
Enthalpy Flow	MMBTU/hr	1.6		1.6	3.3	1.8		1.8
Average Density	lb/ft^3	0.12		0.12	0.31	0.32		0.32

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QAK501	10.03	WK501	4593.8
QAK501A	4.04	WK501A	2203.3
QAK501B	5.98	WK501B	2390.4
QCH502	1.47		

Eq. No.	Equipment Name	Req.	Spare	Equipment Type
H-501A	Hydrogen Compressor Intercooler	1		AIR-COOLED EXCHANGER
H-501B	Hydrogen Compressor Air-cooled Aftercooler	1		AIR-COOLED EXCHANGER
H-502	Hydrogen Compressor Water-cooler Aftercooler	1		SHELL-TUBE
K-501	Hydrogen Compressor	1		RECIPROCATING
S-501	Pre-hydrogen Compressor Knock-out	1		KNOCK-OUT DRUM
S-502	Hydrogen Compressor 1st Interstage Knock-out	1		KNOCK-OUT DRUM
S-503	Post-hydrogen Compressor Knock-out	1		KNOCK-OUT DRUM

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SECTION A500
HYDROGEN COMPRESSION

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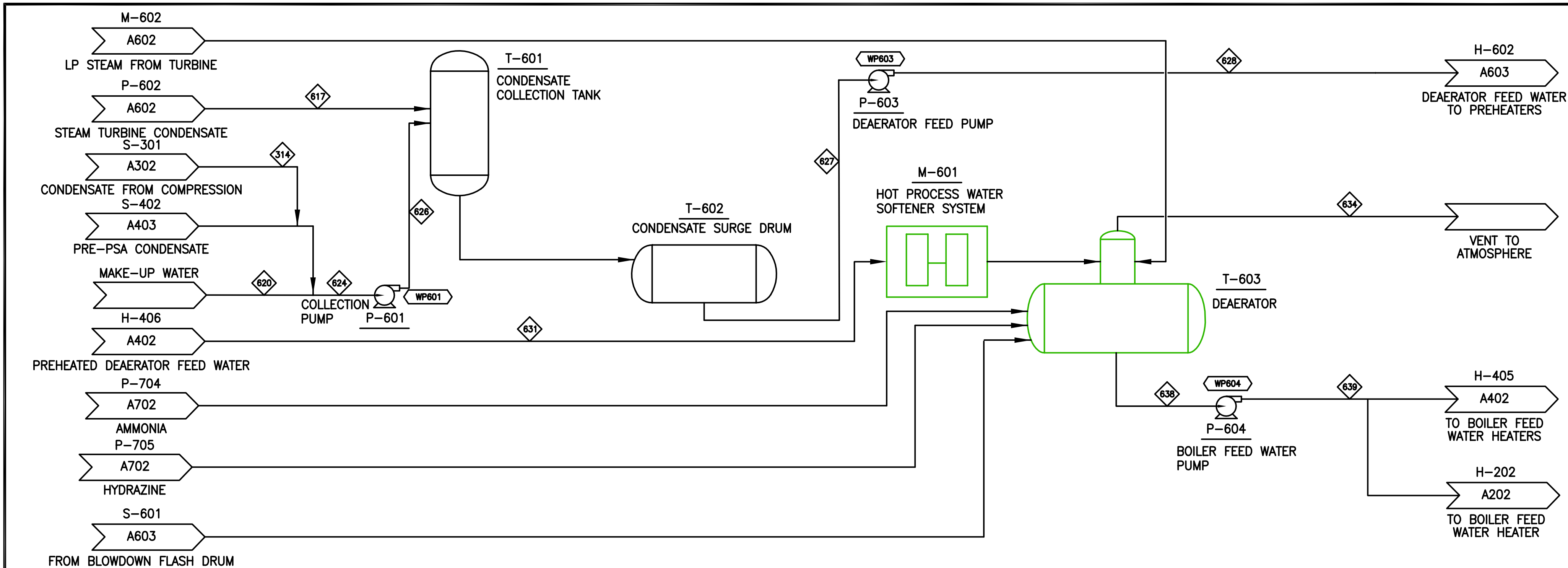

COMPONENT	UNITS	314	617	620	624	626	627	628	631	634	638	639
Total Flow	lb/hr	40,882	93,974	102,749	255,292	255,292	349,266	349,266	349,266	73	349,268	349,268
Temperature	F	146	115	60	96	96	101	101	227	120	227	230
Pressure	Psia	15.00	14.70	14.70	14.70	14.70	14.70	39.70	19.70	14.70	19.70	1344.70
Vapor Fraction		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.00	0.00	0.00
Hydrogen	lb/hr	0								0		
Water	lb/hr	40,876	93,974	102,749	255,292	255,292	349,266	349,266	349,266		349,268	349,268
Carbon Monoxide	lb/hr	0								0		
Nitrogen	lb/hr	0								0		
Oxygen	lb/hr											
Argon	lb/hr											
Carbon Dioxide	lb/hr	1								58		
Hydrogen Sulfide (H2S)	lb/hr	0								0		
SO2	lb/hr											
Ammonia (NH3)	lb/hr	5								15		
NO2	lb/hr											
Methane	lb/hr	0								0		
isobutane	lb/hr											
n-butane	lb/hr											
ethane (C2H6)	lb/hr	0								0		
ethylene (C2H4)	lb/hr	0								0		
acetylene (C2H2)	lb/hr	0								0		
C3H8	lb/hr											
Pentane +	lb/hr											
Benzene (C6H6)	lb/hr	0								0		
Tar (C10H8)	lb/hr	0								0		
Carbon (Solid)	lb/hr											
Sulfur (Solid)	lb/hr											
Olivine (Solid)	lb/hr											
MgO (Solid)	lb/hr											
Ash	lb/hr											
Char	lb/hr											
Wood	lb/hr											
Enthalpy Flow	MMBTU/hr	-278.7	-637.8	-703.0	-1737.6	-1737.6	-2375.4	-2375.4	-2331.4	-0.2	-2331.4	-2329.4
Average Density	lb/ft^3	45.77	61.79	62.37	62.05	62.05	61.98	61.99	59.43	0.08	59.43	59.64

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
		WP601	4.2
		WP603	13.8
		WP604	759.0

Eq. No.	Equipment Name	Req	Spare	Equipment Type
M-601	Hot Process Water Softener System	1		PACKAGE
P-601	Collection Pump	1	1	CENTRIFUGAL
P-603	Deaerator Feed Pump	1	1	CENTRIFUGAL
P-604	Boiler Feed Water Pump	1	1	CENTRIFUGAL
T-601	Condensate Collection Tank	1		HORIZONTAL-VESSEL
T-602	Condensate Surge Drum	1		HORIZONTAL-VESSEL
T-603	Deaerator	1		HORIZONTAL-VESSEL

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SECTION 600
STEAM SYSTEM & PWR GENERATION

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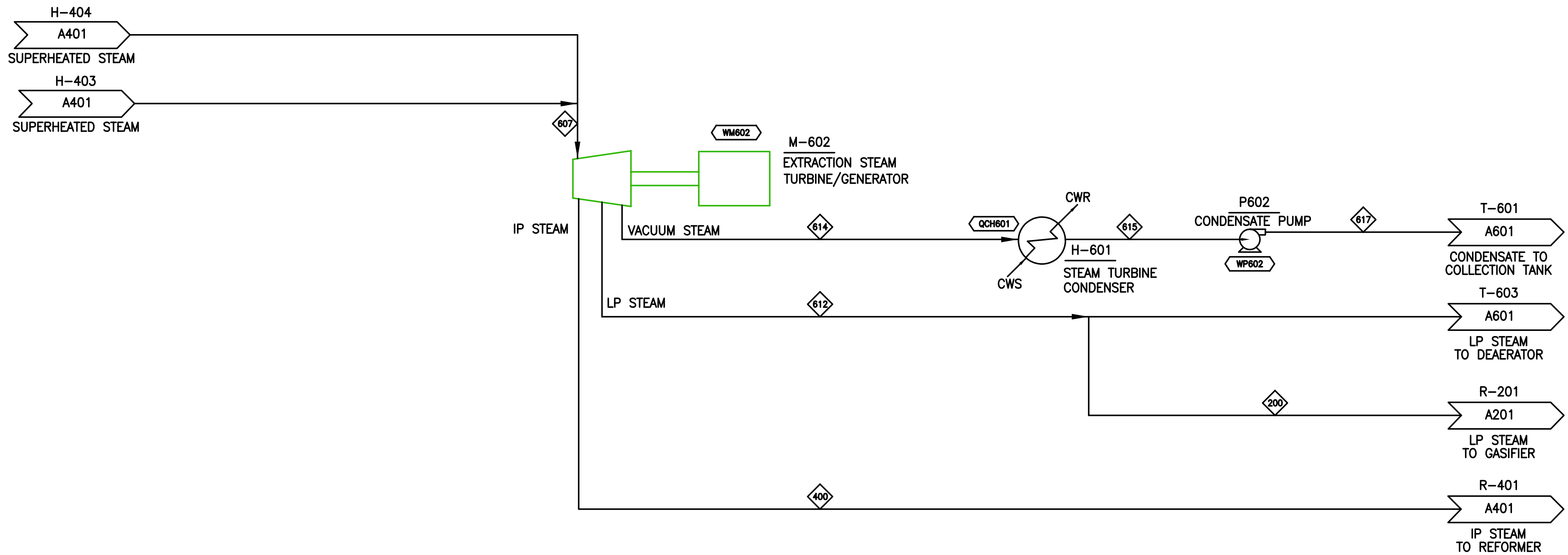

COMPONENT	UNITS	200	400	607	612	614	615	617
Total Flow	lb/hr	73,120	175,189	342,283	73,120	93,974	93,974	93,974
Temperature	F	260	731	1,000	266	115	115	115
Pressure	Psia	25.00	440.00	1264.70	35.00	1.47	1.47	14.70
Vapor Fraction		1.00	1.00	1.00	1.00	0.89	0.00	0.00
Hydrogen	lb/hr							
Water	lb/hr	73,120	175,189	342,283	73,120	93,974	93,974	93,974
Carbon Monoxide	lb/hr							
Nitrogen	lb/hr							
Oxygen	lb/hr							
Argon	lb/hr							
Carbon Dioxide	lb/hr							
Hydrogen Sulfide (H2S)	lb/hr							
SO2	lb/hr							
Ammonia (NH3)	lb/hr							
NO2	lb/hr							
Methane	lb/hr							
isobutane	lb/hr							
n-butane	lb/hr							
ethane (C2H6)	lb/hr							
ethylene (C2H4)	lb/hr							
acetylene (C2H2)	lb/hr							
C3H8	lb/hr							
Pentane +	lb/hr							
Benzene (C6H6)	lb/hr							
Tar (C10H8)	lb/hr							
Carbon (Solid)	lb/hr							
Sulfur (Solid)	lb/hr							
Olivine (Solid)	lb/hr							
MgO (Solid)	lb/hr							
Ash	lb/hr							
Char	lb/hr							
Wood	lb/hr							
Enthalpy Flow	MMBTU/hr	-416.4	-962.0	-1839.2	-416.8	-552.0	-637.8	-637.8
Average Density	lb/ft^3	0.06	0.65	1.54	0.08	0.00	61.79	61.79

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QCH601	85.81	WM602A	-16014.5
		WM602B	-13646.2
		WM602C	-6430.0
		WP602	4.2

Eq. No.	Equipment Name	Req	Spare	Equipment Type
H-601	Steam Turbine Condenser	1		SHELL-TUBE
M-602	Extraction Steam Turbine/Generator	1		STEAM-TURBINE
P-602	Condensate Pump	1	1	CENTRIFUGAL

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SECTION A600
STEAM SYSTEM & PWR GENERATION

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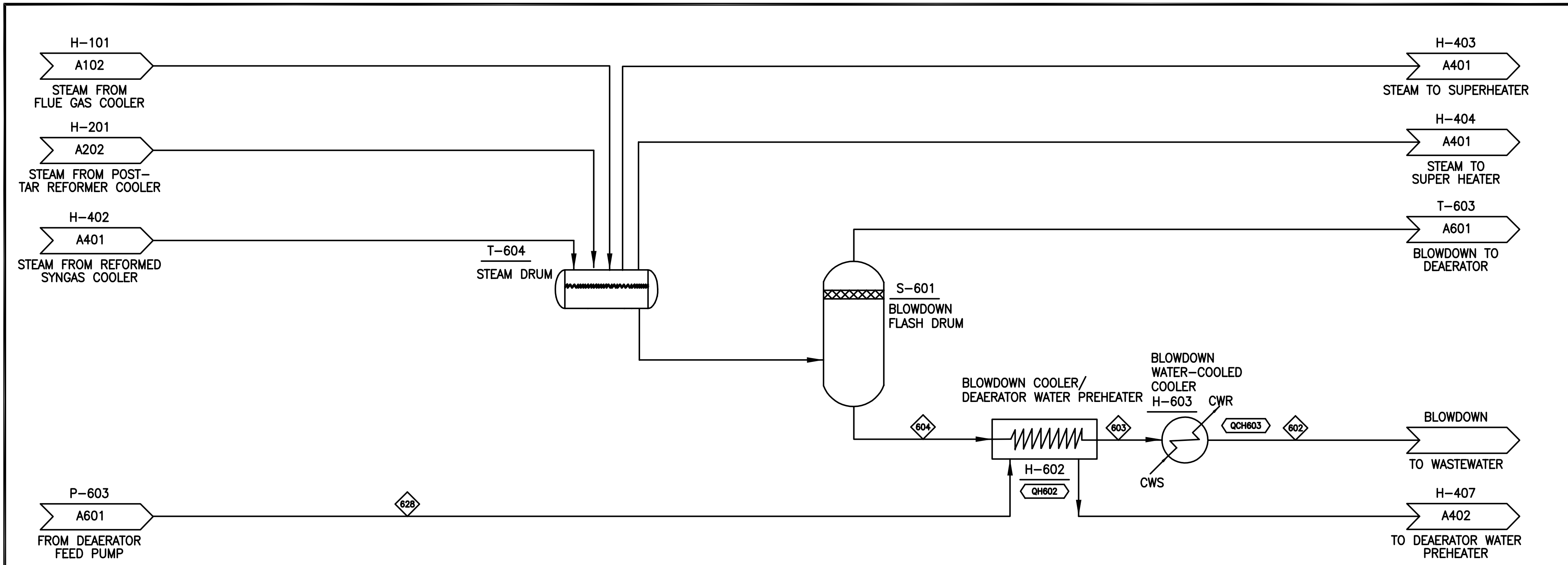

COMPONENT	UNITS	602	603	604	605	628
Total Flow	lb/hr	6,985	6,985	6,985	342,283	349,266
Temperature	F	110	200	575	575	101
Pressure	Psia	1269.70	1274.70	1279.70	1279.70	39.70
Vapor Fraction		0.00	0.00	0.00	1.00	0.00
Hydrogen	lb/hr					
Water	lb/hr	6,985	6,985	6,985	342,283	349,266
Carbon Monoxide	lb/hr					
Nitrogen	lb/hr					
Oxygen	lb/hr					
Argon	lb/hr					
Carbon Dioxide	lb/hr					
Hydrogen Sulfide (H2S)	lb/hr					
SO2	lb/hr					
Ammonia (NH3)	lb/hr					
NO2	lb/hr					
Methane	lb/hr					
isobutane	lb/hr					
n-butane	lb/hr					
ethane (C2H6)	lb/hr					
ethylene (C2H4)	lb/hr					
acetylene (C2H2)	lb/hr					
C3H8	lb/hr					
Pentane +	lb/hr					
Benzene (C6H6)	lb/hr					
Tar (C10H8)	lb/hr					
Carbon (Solid)	lb/hr					
Sulfur (Solid)	lb/hr					
Olivine (Solid)	lb/hr					
MgO (Solid)	lb/hr					
Ash	lb/hr					
Char	lb/hr					
Wood	lb/hr					
Enthalpy Flow	MMBTU/hr	-47.4	-46.8	-43.9	-1947.4	-2375.4
Average Density	lb/ft^3	62.10	60.36	44.21	2.97	61.99

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QCH603			
QH602			

Eq. No.	Equipment Name	Req	Spare	Equipment Type
H-602	Blow down Cooler / Deaerator Water Preheater	1		SHELL-TUBE
H-603	Blow down Water-cooled Cooler	1		SHELL-TUBE
S-601	Blow down Flash Drum	1		HORIZONTAL-VESSEL
T-604	Steam Drum	1		HORIZONTAL-VESSEL

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SECTION A600
STEAM SYSTEM & PWR GENERATION

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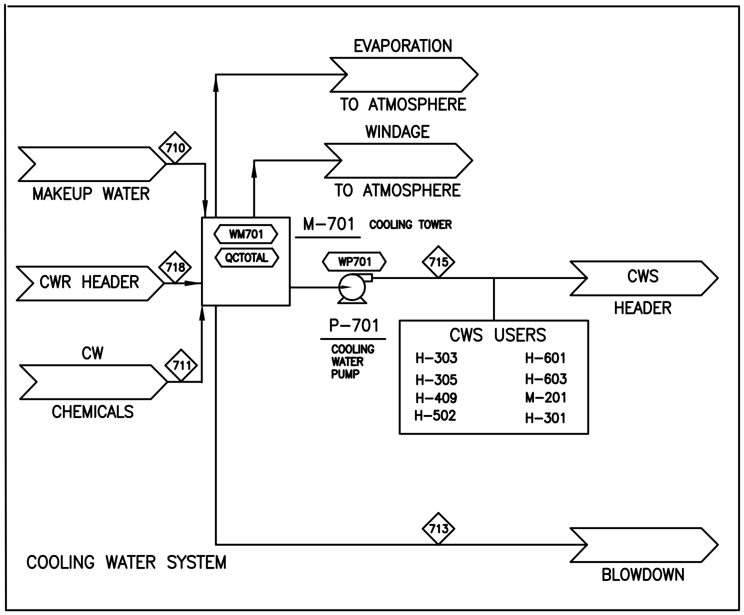

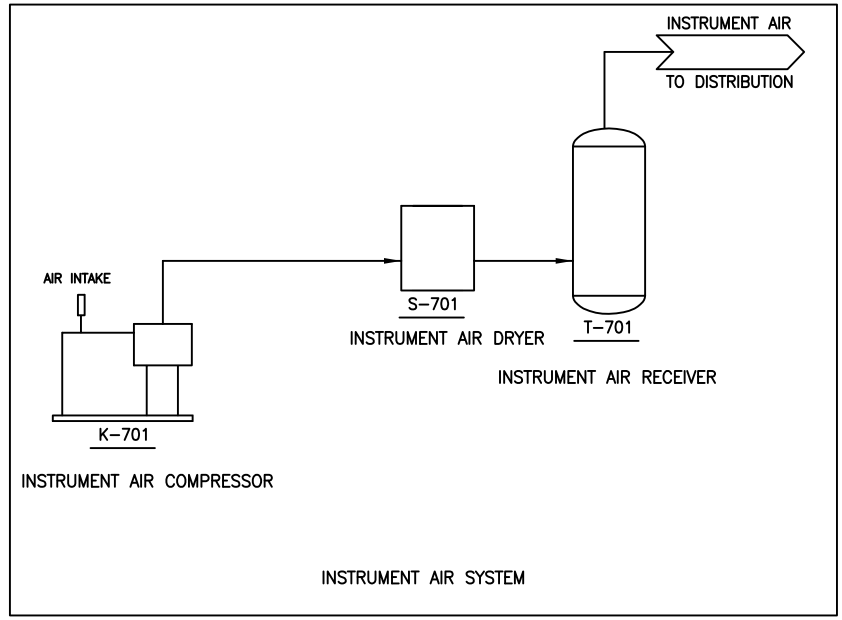

COMPONENT	UNITS	710	711	713	715	718
Total Flow	lb/hr	131,921	1	25,346	6,088,322	6,088,322
Temperature	F	60	60	90	90	110
Pressure	Psia	14.70	14.70	14.70	74.70	59.70
Vapor Fraction		0.00	0.00	0.00	0.00	0.00
Hydrogen	lb/hr					
Water	lb/hr	131,921	1	25,346	6,088,322	6,088,322
Carbon Monoxide	lb/hr					
Nitrogen	lb/hr					
Oxygen	lb/hr					
Argon	lb/hr					
Carbon Dioxide	lb/hr					
Hydrogen Sulfide (H2S)	lb/hr					
SO2	lb/hr					
Ammonia (NH3)	lb/hr					
NO2	lb/hr					
Methane	lb/hr					
isobutane	lb/hr					
n-butane	lb/hr					
ethane (C2H6)	lb/hr					
ethylene (C2H4)	lb/hr					
acetylene (C2H2)	lb/hr					
C3H8	lb/hr					
Pentane +	lb/hr					
Benzene (C6H6)	lb/hr					
Tar (C10H8)	lb/hr					
Carbon (Solid)	lb/hr					
Sulfur (Solid)	lb/hr					
Olivine (Solid)	lb/hr					
MgO (Solid)	lb/hr					
Ash	lb/hr					
Char	lb/hr					
Wood	lb/hr					
Enthalpy Flow	MMBTU/hr	-912.5	0.0	-174.4	-41900.3	-41760.5
Average Density	lb/ft^3	47.44	47.44	46.89	46.89	46.51

Heat Stream No.	MM BTU/hr		Work Stream No.	HP
QCTOTAL	139.85		WM701	653.2
			WP701	659.3

Eq. No.	Equipment Name	Req.	Spare	Equipment Type
K-701	Plant Air Compressor	2	1	RECIPROCATING
M-701	Cooling Tower System	1		INDUCED-DRAFT
P-701	Cooling Water Pump	1	1	CENTRIFUGAL
S-701	Instrument Air Dryer	1	1	PACKAGE
T-701	Plant Air Receiver	1		HORIZONTAL-VESSEL

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SECTION A700
COOLING WATER & OTHER UTILITIES

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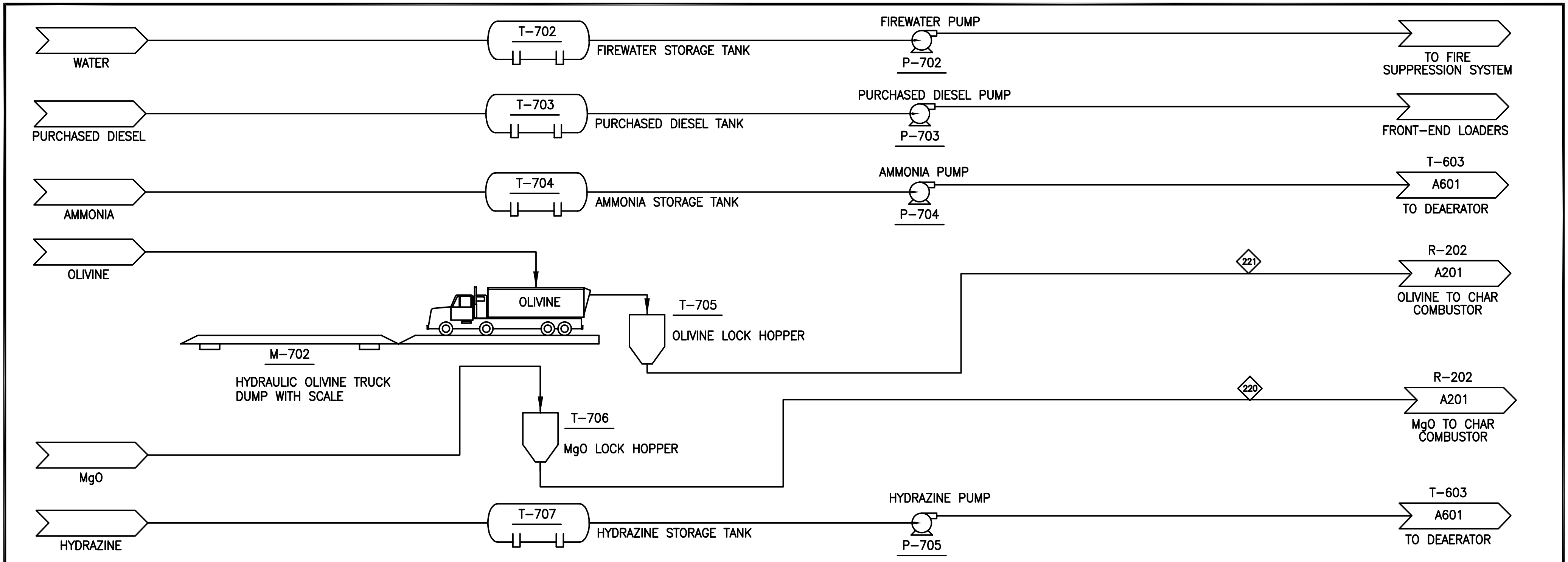

COMPONENT	UNITS	220	221
Total Flow	lb/hr	7	5,440
Temperature	F	60	60
Pressure	Psia	25.00	25.00
Vapor Fraction		0.00	0.00
Hydrogen	lb/hr		
Water	lb/hr		
Carbon Monoxide	lb/hr		
Nitrogen	lb/hr		
Oxygen	lb/hr		
Argon	lb/hr		
Carbon Dioxide	lb/hr		
Hydrogen Sulfide (H2S)	lb/hr		
SO2	lb/hr		
Ammonia (NH3)	lb/hr		
NO2	lb/hr		
Methane	lb/hr		
isobutane	lb/hr		
n-butane	lb/hr		
ethane (C2H6)	lb/hr		
ethylene (C2H4)	lb/hr		
acetylene (C2H2)	lb/hr		
C3H8	lb/hr		
Pentane +	lb/hr		
Benzene (C6H6)	lb/hr		
Tar (C10H8)	lb/hr		
Carbon (Solid)	lb/hr		
Sulfur (Solid)	lb/hr		
Olivine (Solid)	lb/hr	7	5,440
MgO (Solid)	lb/hr		
Ash	lb/hr		
Char	lb/hr		
Wood	lb/hr		
Enthalpy Flow	MMBTU/hr	0.0	0.0
Average Density	lb/ft^3	165.39	165.39

Heat Stream No.	MM BTU/hr	Work Stream No.	HP

Eq. No.	Equipment Name	Req.	Spare	Equipment Type
M-702	Hydraulic Truck Dump with Scale	1		TRUCK-SCALE
P-702	Firewater Pump	1	1	CENTRIFUGAL
P-703	Diesel Pump	1	1	CENTRIFUGAL
P-704	Ammonia Pump	1	1	CENTRIFUGAL
P-705	Hydrazine Pump	1		CENTRIFUGAL
T-702	Firewater Storage Tank	1		FLAT-BTM-STORAGE
T-703	Diesel Storage Tank	1		FLAT-BTM-STORAGE
T-704	Ammonia Storage Tank	1		HORIZONTAL-STORAGE
T-705	Olivine Lock Hopper	1		VERTICAL-VESSEL
T-706	MgO Lock Hopper	1		VERTICAL-VESSEL
T-707	Hydrazine Storage Tank	1		VERTICAL-VESSEL

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SECTION A700
COOLING WATER & OTHER UTILITIES

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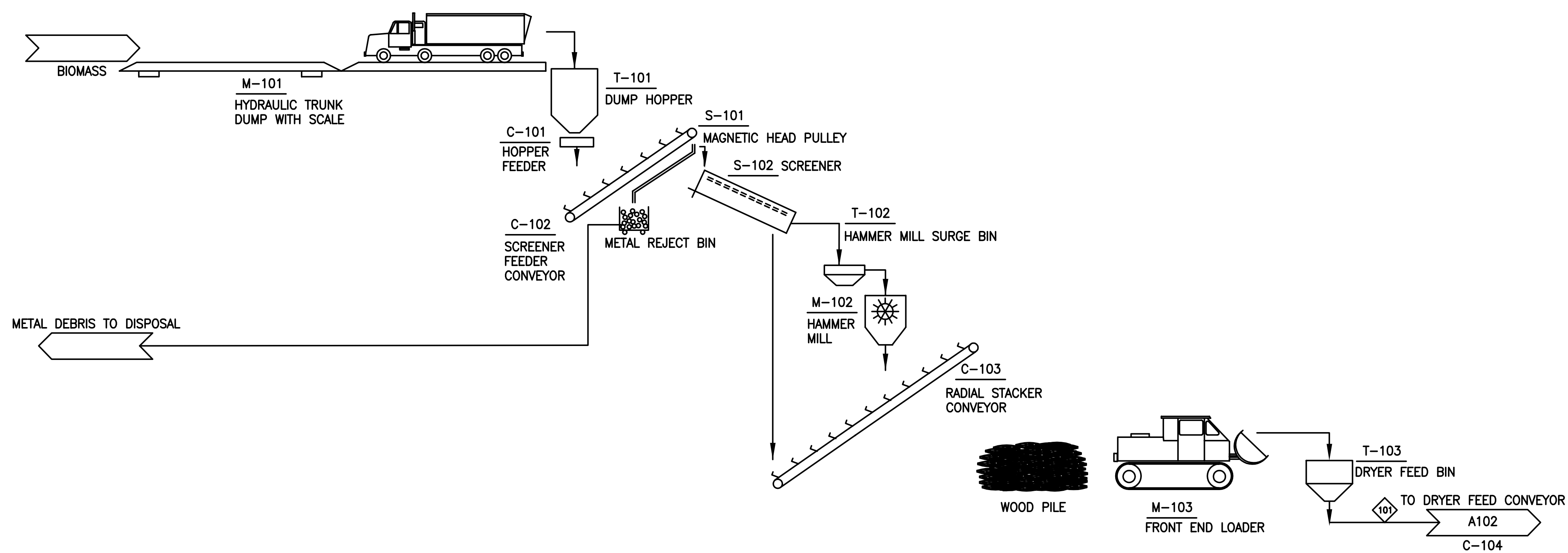
Appendix D: Goal Design Process Flow Diagrams



COMPONENT	UNITS	101
Total Flow	lb/hr	367,437
Temperature	F	60
Pressure	Psia	14.70
Vapor Fraction		0.00
Hydrogen	lb/hr	
Water	lb/hr	183,718
Carbon Monoxide	lb/hr	
Nitrogen	lb/hr	
Oxygen	lb/hr	
Argon	lb/hr	
Carbon Dioxide	lb/hr	
Hydrogen Sulfide (H2S)	lb/hr	
SO2	lb/hr	
Ammonia (NH3)	lb/hr	
NO2	lb/hr	
Methane	lb/hr	
isobutane	lb/hr	
n-butane	lb/hr	
ethane (C2H6)	lb/hr	
ethylene (C2H4)	lb/hr	
acetylene (C2H2)	lb/hr	
C3H8	lb/hr	
Pentane +	lb/hr	
Benzene (C6H6)	lb/hr	
Tar (C10H8)	lb/hr	
Carbon (Solid)	lb/hr	
Sulfur (Solid)	lb/hr	
Olivine (Solid)	lb/hr	
MgO (Solid)	lb/hr	
Ash	lb/hr	0
Char	lb/hr	0
Wood	lb/hr	183,718
Enthalpy Flow	MMBTU/hr	-1699.4
Average Density	lb/ft^3	47.44

Heat Stream No.	MM BTU/hr		Work Stream No.	HP

Eq. No.	Equipment Name	Req	Spare	Equipment Type
C-101	Hopper Feeder	4		VIBRATING-FEEDER
C-102	Screener Feeder Conveyor	2		BELT
C-103	Radial Stacker Conveyor	2		BELT
M-101	Hydraulic Truck Dump with Scale	4		TRUCK-SCALE
M-102	Hammermill	2		
M-103	Front End Loaders	3		LOADER
S-101	Magnetic Head Pulley	2		MAGNET
S-102	Screener	2		SCREEN
T-101	Dump Hopper	4		LIVE-BTM-BIN
T-102	Hammermill Surge Bin	1		LIVE-BTM-BIN
T-103	Dryer Feed Bin	2		LIVE-BTM-BIN

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SECTION A100
FEED HANDLING & DRYING

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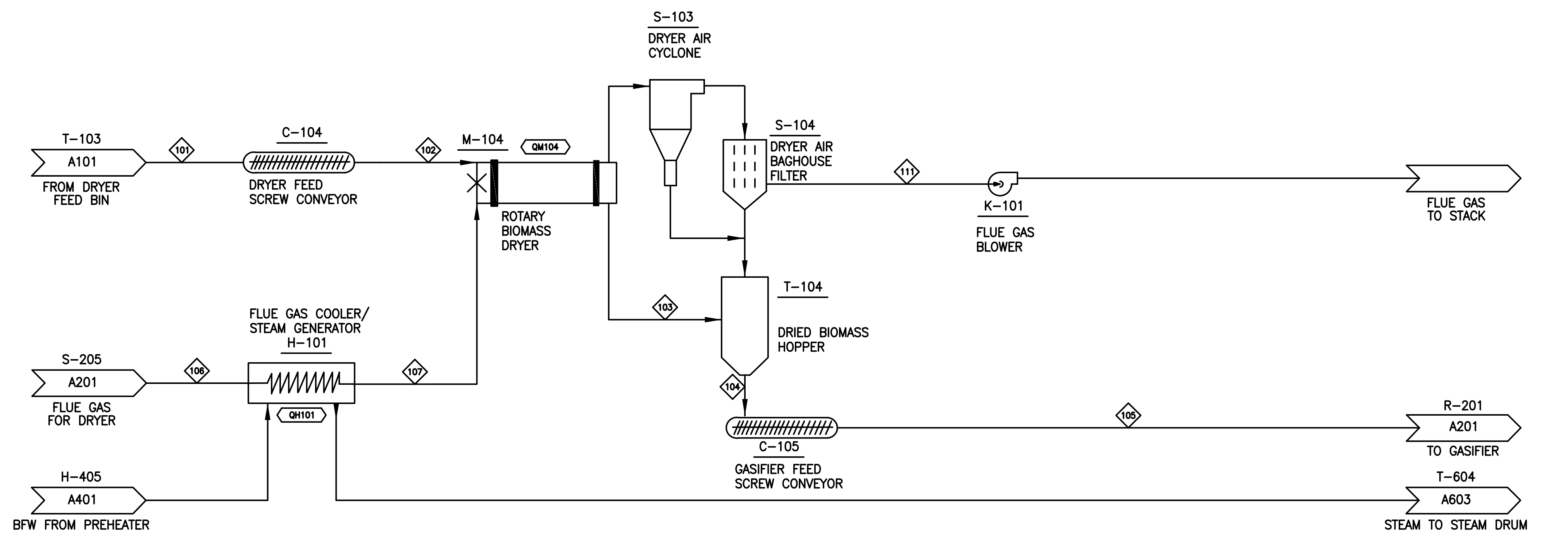

COMPONENT	UNITS	101	102	103	104	105	106	107	111
Total Flow	lb/hr	367,437	367,437	208,771	208,771	208,771	480,860	480,860	639,526
Temperature	F	60	60	220	220	220	1,800	1,791	250
Pressure	Psia	14.70	15.70	15.70	15.70	25.00	22.00	22.00	14.70
Vapor Fraction		0.00	0.00	0.00	0.00	0.00	1.00	1.00	1.00
Hydrogen	lb/hr								
Water	lb/hr	183,718	183,718	25,053	25,053	25,053	30,820	30,820	189,486
Carbon Monoxide	lb/hr								
Nitrogen	lb/hr						327,430	327,430	327,430
Oxygen	lb/hr						10,718	10,718	10,718
Argon	lb/hr						5,584	5,584	5,584
Carbon Dioxide	lb/hr						106,280	106,280	106,280
Hydrogen Sulfide (H2S)	lb/hr								
SO2	lb/hr						27	27	27
Ammonia (NH3)	lb/hr								
NO2	lb/hr								
Methane	lb/hr								
isobutane	lb/hr								
n-butane	lb/hr								
ethane (C2H6)	lb/hr								
ethylene (C2H4)	lb/hr								
acetylene (C2H2)	lb/hr								
C3H8	lb/hr								
Pentane +	lb/hr								
Benzene (C6H6)	lb/hr								
Tar (C10H8)	lb/hr								
Carbon (Solid)	lb/hr								
Sulfur (Solid)	lb/hr								
Olivine (Solid)	lb/hr								
MgO (Solid)	lb/hr								
Ash	lb/hr	0	0	0	0	0			
Char	lb/hr	0	0	0	0	0			
Wood	lb/hr	183,718	183,718	183,718	183,718	183,718			
Enthalpy Flow	MMBTU/hr	-1699.4	-1699.4	-584.6	-584.6	-584.6	-353.9	-355.3	-1470.1
Average Density	lb/ft^3	47.44	47.44	44.12	44.12	44.12	0.03	0.03	0.05

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QH101	1.37		
QM104	-208.40		

Eq. No.	Equipment Name	Req	Spare	Equipment Type
C-104	Dryer Feed Screw Conveyor	2		SCREW
C-105	Gasifier Feed Screw Conveyor	2		SCREW
H-101	Flue Gas Cooler / Steam Generator #3	1		SHELL-TUBE
K-101	Flue Gas Blower	2		CENTRIFUGAL
M-104	Rotary Biomass Dryer	2		ROTARY-DRUM
S-103	Dryer Air Cyclone	2		GAS CYCLONE
S-104	Dryer Air Baghouse Filter	2		FABRIC-FILTER
T-104	Dried Biomass Hopper	2		VERTICAL-VESSEL

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SECTION A100
FEED HANDLING & DRYING

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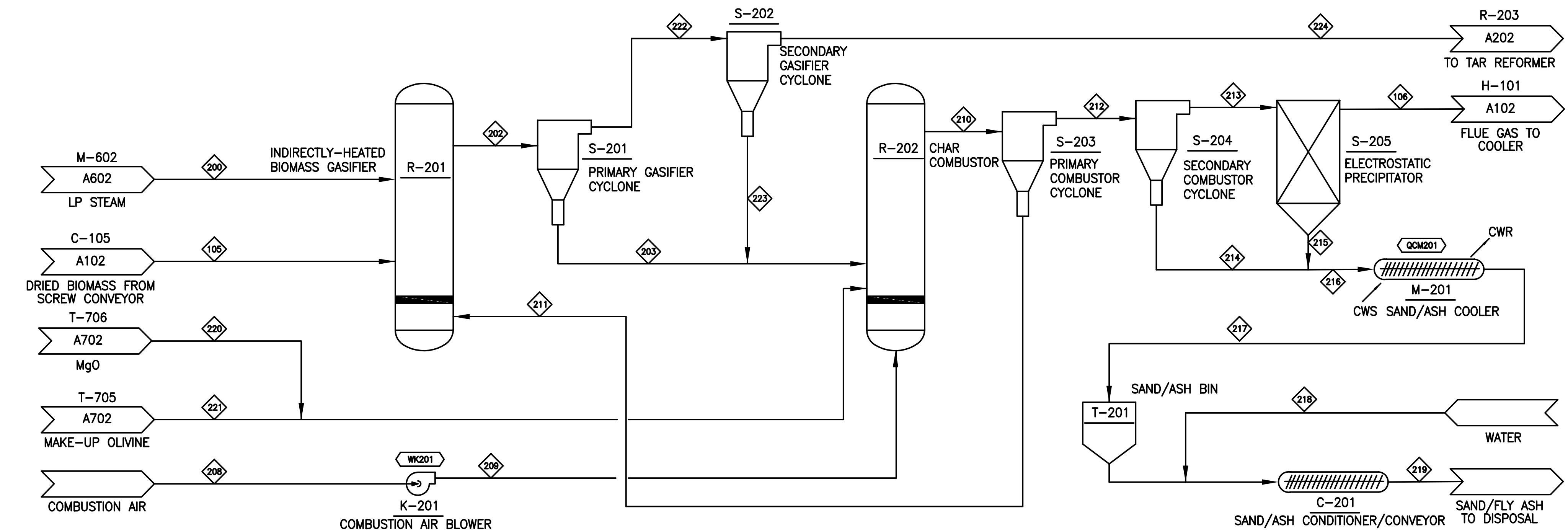

COMPONENT	UNITS	200	202	203	208	209	210	211	212	213	214	215	216	217	218	219	220	221	222	223	224
Total Flow	lb/hr	73,120	5,228,878	4,982,397	442,157	442,157	5,434,489	4,946,987	487,502	480,866	6,635	7	6,642	6,642	738	7,380	7	5,440	246,481	4,489	241,993
Temperature	F	259	1,598	1,598	90	185	1,800	1,800	1,800	1,800	1,800	1,800	1,800	300	60	208	60	60	1,598	1,598	1,598
Pressure	Psia	25.00	23.00	22.00	14.70	22.00	22.00	25.00	22.00	22.00	22.00	22.00	22.00	14.70	14.70	14.70	25.00	25.00	22.00	22.00	22.00
Vapor Fraction		1.00	1.00	0.00	1.00	1.00	1.00	0.00	1.00	1.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.00	0.00	1.00
Hydrogen	lb/hr		3,092																3,092		3,092
Water	lb/hr	73,120	98,172		8,627	8,627	30,820		30,820	30,820					738	738			98,172		98,172
Carbon Monoxide	lb/hr		75,993																75,993		75,993
Nitrogen	lb/hr				327,409	327,409	327,430		327,430	327,430											
Oxygen	lb/hr				100,319	100,319	10,718		10,718	10,718											
Argon	lb/hr				5,584	5,584	5,584		5,584	5,584											
Carbon Dioxide	lb/hr		36,216		218	218	106,280		106,280	106,280									36,216		36,216
Hydrogen Sulfide (H2S)	lb/hr		161																161		161
SO2	lb/hr						27		27	27											
Ammonia (NH3)	lb/hr		355																355		355
NO2	lb/hr																				
Methane	lb/hr		15,850																15,850		15,850
isobutane	lb/hr																				
n-butane	lb/hr																				
ethane (C2H6)	lb/hr		566																566		566
ethylene (C2H4)	lb/hr		7,848																7,848		7,848
acetylene (C2H2)	lb/hr		683																683		683
C3H8	lb/hr																				
Pentane +	lb/hr																				
Benzene (C6H6)	lb/hr		640																640		640
Tar (C10H8)	lb/hr		1,919																1,919		1,919
Carbon (Solid)	lb/hr																				
Sulfur (Solid)	lb/hr																				
Olivine (Solid)	lb/hr		4,946,898	4,941,951			4,951,850	4,946,898	4,952	5	4,947	5	4,952	4,952		4,952	7	5,440	4,947	4,452	495
MgO (Solid)	lb/hr																				
Ash	lb/hr		0	0			1,779	89	1,690	2	1,688	2	1,690	1,690		1,690			0	0	0
Char	lb/hr		40,486	40,445															40	36	4
Wood	lb/hr		0	0															0	0	0
Enthalpy Flow	MMBTU/hr	-416.5	1192.8	1879.6	-49.3	-39.1	1842.2	2193.8	-351.6	-353.9	2.3	0.0	2.3	-0.3	-5.1	-5.4	0.0	0.0	-686.9	1.7	-688.6
Average Density	lb/ft^3	0.06	160.21	160.19	0.07	0.09	159.53	159.50	159.53	159.53	159.50	159.50	159.50	164.58	47.44	209.28	165.39	165.39	160.21	160.19	160.21

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QCM201	2.63	WK201	4023.6

Eq. No.	Equipment Name	Req.	Spare	Equipment Type
C-201	Sand/ash Conditioner/Conveyor	1		SCREW
K-201	Combustion Air Blow er	2		CENTRIFUGAL
M-201	Sand/ash Cooler	2		MISCELLANEOUS
R-201	Indirectly-heated Biomass Gasifier	2		VERTICAL-VESSEL
R-202	Char Combustor	2		VERTICAL-VESSEL
S-201	Primary Gasifier Cyclone	2		GAS CYCLONE
S-202	Secondary Gasifier Cyclone	2		GAS CYCLONE
S-203	Primary Combustor Cyclone	2		GAS CYCLONE
S-204	Secondary Combustor Cyclone	2		GAS CYCLONE
S-205	Electrostatic Precipitator	2		MISCELLANEOUS
T-201	Sand/ash Bin	1		FLAT-BTM-STORAGE

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SECTION A200
GASIFICATION & TAR REFORMING

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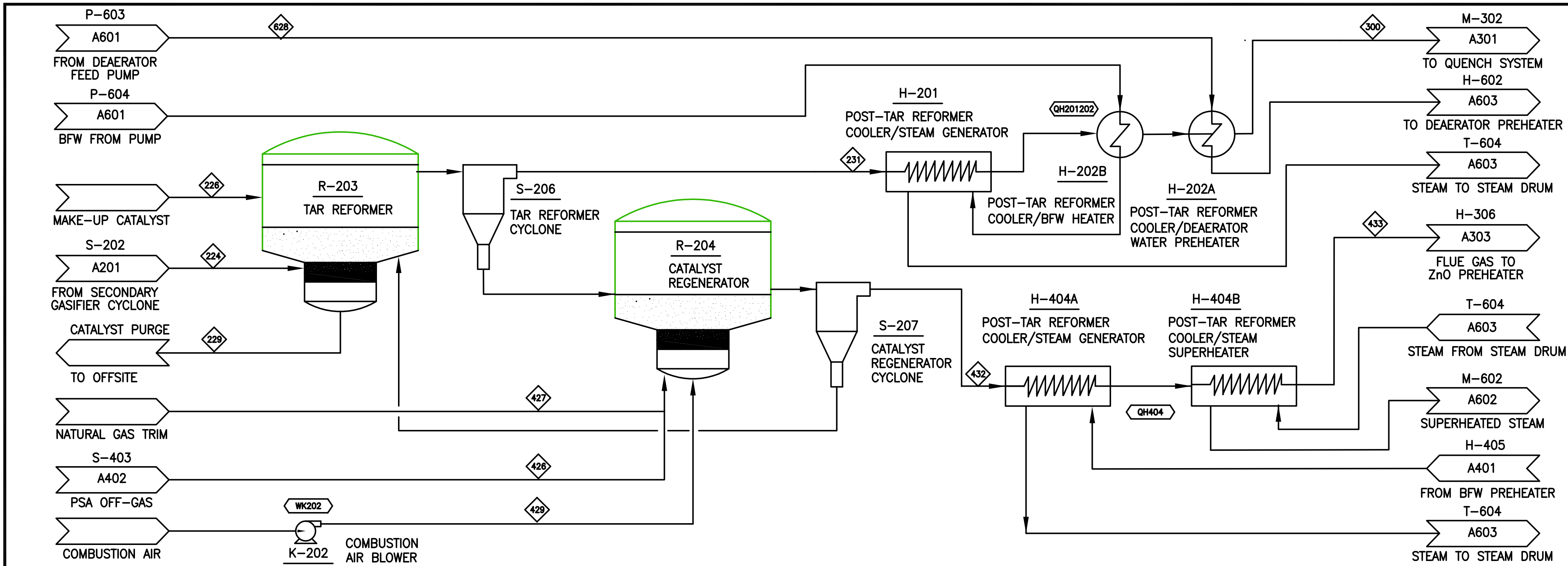

COMPONENT	UNITS	224	226	229	231	300	426	427	429	432	433	628
Total Flow	lb/hr	241,993	1	1	241,993	241,993	230,694	3,739	302,225	536,658	536,658	349,809
Temperature	F	1,598	60	1,598	1,598	156	109	60	90	1,798	250	104
Pressure	Psia	22.00	22.00	19.00	19.00	17.00	14.70	14.70	14.70	14.70	14.34	39.70
Vapor Fraction		1.00	0.00	0.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	0.00
Hydrogen	lb/hr	3,092			12,264	12,264	3,781					
Water	lb/hr	98,172			57,209	57,209	853		5,897	56,714	56,714	349,809
Carbon Monoxide	lb/hr	75,993			100,622	100,622	5,590					
Nitrogen	lb/hr				263	263	263	67	223,792	224,122	224,122	
Oxygen	lb/hr								68,570	5,411	5,411	
Argon	lb/hr								3,817	3,817	3,817	
Carbon Dioxide	lb/hr	36,216			66,902	66,902	216,168	48	149	246,576	246,576	
Hydrogen Sulfide (H2S)	lb/hr	161			161	161		0				
SO2	lb/hr									0	0	
Ammonia (NH3)	lb/hr	355			35	35	7					
NO2	lb/hr									18	18	
Methane	lb/hr	15,850			3,170	3,170	3,169	3,314				
isobutane	lb/hr							13				
n-butane	lb/hr							13				
ethane (C2H6)	lb/hr	566			6	6	6	204				
ethylene (C2H4)	lb/hr	7,848			785	785	784					
acetylene (C2H2)	lb/hr	683			68	68	68					
C3H8	lb/hr							48				
Pentane +	lb/hr							32				
Benzene (C6H6)	lb/hr	640			6	6	6					
Tar (C10H8)	lb/hr	1,919			2	2	0					
Carbon (Solid)	lb/hr											
Sulfur (Solid)	lb/hr											
Olivine (Solid)	lb/hr	495	1	1	495	495						
MgO (Solid)	lb/hr											
Ash	lb/hr	0			0	0						
Char	lb/hr	4			4	4						
Wood	lb/hr	0			0	0						
Enthalpy Flow	MMBTU/hr	-688.6	0.0	0.0	-581.5	-755.7	-849.6	-7.2	-33.7	-1005.4	-1252.7	-2378.1
Average Density	lb/ft^3	160.21	165.39	160.19	160.20	165.11	0.08	0.05	0.07	0.02	0.06	61.95

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QH201202	174.17	WK202	153.3
QH404	247.32		

Eq. No.	Equipment Name	Req	Spare	Equipment Type
H-201	Post-tar Reformer Cooler / Steam Generator #1	1		SHELL-TUBE
H-202A	Post-tar reformer cooler/Deaerator water preheater #	1		SHELL-TUBE
H-202B	Post-tar cracker cooler/BFW preheater #2	1		SHELL-TUBE
H-404A	Tar reformer flue gas cooler/steam generator #2	1		SHELL-TUBE
H-404B	Tar reformer flue gas cooler/ steam superheater #1	1		SHELL-TUBE
K-202	Regenerator Combustion Air Blower	1		CENTRIFUGAL
R-203	Tar Reformer	1		VERTICAL-VESSEL
R-204	Tar Reformer Catalyst Regenerator	1		VERTICAL-VESSEL
S-206	Tar Reformer Cyclone			GAS CYCLONE
S-207	Catalyst Regenerator Cyclone			GAS CYCLONE

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SECTION A200
GASIFICATION & TAR REFORMING

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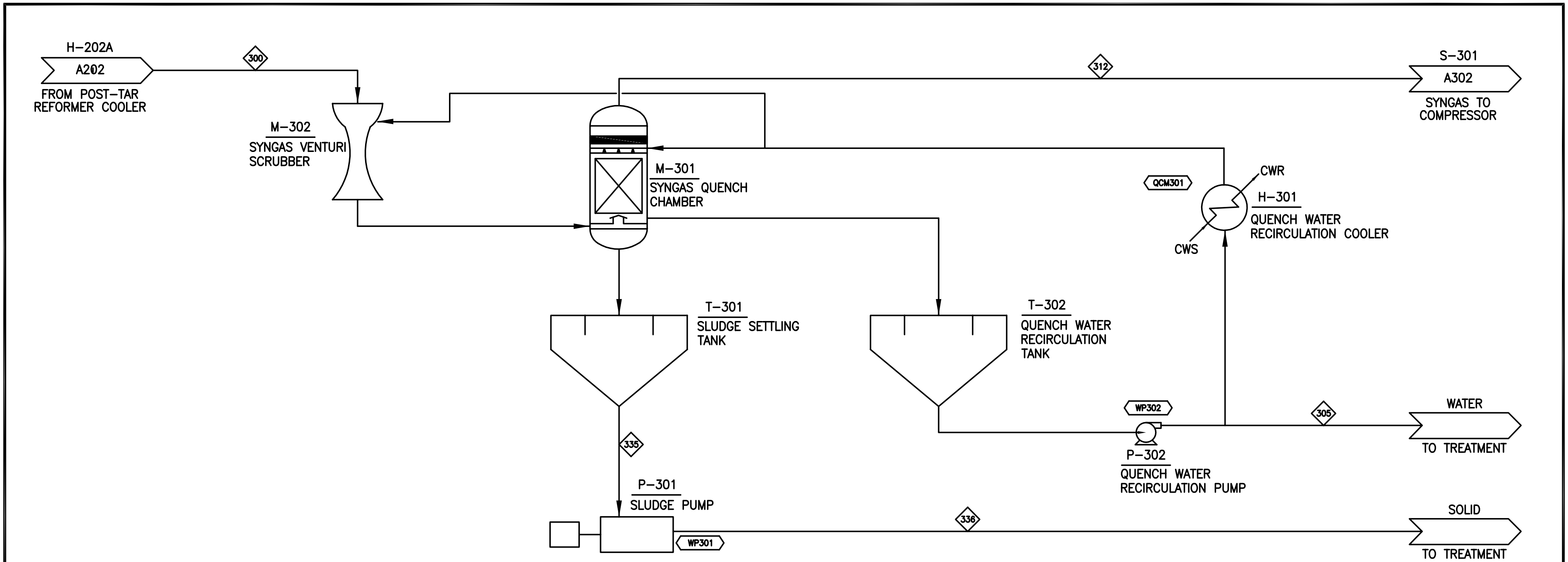

COMPONENT	UNITS	300	305	312	335	336
Total Flow	lb/hr	241,993	7,173	233,823	997	997
Temperature	F	156	140	140	140	140
Pressure	Psia	17.00	15.00	15.00	15.00	34.70
Vapor Fraction		1.00	0.00	1.00	0.00	0.00
Hydrogen	lb/hr	12,264	0	12,264		
Water	lb/hr	57,209	7,155	49,555	499	499
Carbon Monoxide	lb/hr	100,622	3	100,620		
Nitrogen	lb/hr	263	0	263		
Oxygen	lb/hr					
Argon	lb/hr					
Carbon Dioxide	lb/hr	66,902	8	66,894		
Hydrogen Sulfide (H2S)	lb/hr	161	0	161		
SO2	lb/hr					
Ammonia (NH3)	lb/hr	35	3	32		
NO2	lb/hr					
Methane	lb/hr	3,170	1	3,169		
isobutane	lb/hr					
n-butane	lb/hr					
ethane (C2H6)	lb/hr	6	0	6		
ethylene (C2H4)	lb/hr	785	1	784		
acetylene (C2H2)	lb/hr	68	0	68		
C3H8	lb/hr					
Pentane +	lb/hr					
Benzene (C6H6)	lb/hr	6	0	6		
Tar (C10H8)	lb/hr	2	2	0		
Carbon (Solid)	lb/hr					
Sulfur (Solid)	lb/hr					
Olivine (Solid)	lb/hr	495			495	495
MgO (Solid)	lb/hr					
Ash	lb/hr	0			0	0
Char	lb/hr	4			4	4
Wood	lb/hr	0			0	0
Enthalpy Flow	MMBTU/h	-755.7	-48.9	-713.4	-3.4	-3.4
Average Density	lb/ft^3	165.11	20.20	0.04	211.02	211.02

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QCM301	9.50	WP301	0.1
		WP302	10.2

Eq. No.	Equipment Name	Req	Spare	Equipment Type
H-301	Quench Water Recirculation Cooler	1		SHELL-TUBE
M-301	Syngas Quench Chamber	1		
M-302	Syngas Venturi Scrubber	1		
P-301	Sludge Pump	1	1	CENTRIFUGAL
P-302	Quench Water Recirculation Pump	1	1	CENTRIFUGAL
T-301	Sludge Settling Tank	1		CLARIFIER
T-302	Quench Water Recirculation Tank	1		HORIZONTAL-VESSEL

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SECTION A300
GAS CLEAN-UP & COMPRESSION

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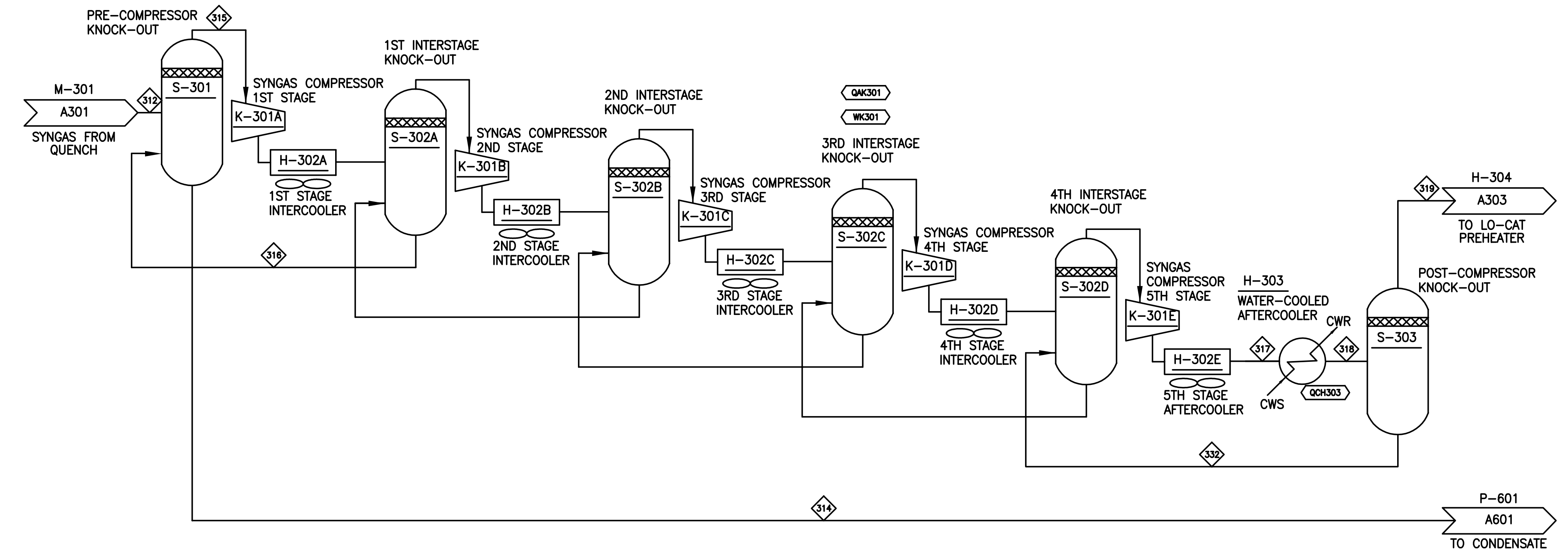

COMPONENT	UNITS	312	314	315	316	317	318	319	332
Total Flow	lb/hr	233,823	48,981	233,488	47,881	185,607	185,607	184,842	765
Temperature	F	140	147	147	140	140	110	110	110
Pressure	Psia	15.00	15.00	15.00	15.00	425.00	420.00	420.00	420.00
Vapor Fraction		1.00	0.00	1.00	0.00	1.00	1.00	1.00	0.00
Hydrogen	lb/hr	12,264	0	12,264	0	12,264	12,264	12,264	0
Water	lb/hr	49,555	48,978	49,210	47,870	1,340	1,340	577	763
Carbon Monoxide	lb/hr	100,620	0	100,620	0	100,620	100,620	100,620	0
Nitrogen	lb/hr	263	0	263	0	263	263	263	0
Oxygen	lb/hr								
Argon	lb/hr								
Carbon Dioxide	lb/hr	66,894	0	66,896	2	66,894	66,894	66,894	0
Hydrogen Sulfide (H2S)	lb/hr	161	0	161	0	161	161	161	0
SO2	lb/hr								
Ammonia (NH3)	lb/hr	32	2	41	9	32	32	30	2
NO2	lb/hr								
Methane	lb/hr	3,169	0	3,169	0	3,169	3,169	3,169	0
isobutane	lb/hr								
n-butane	lb/hr								
ethane (C2H6)	lb/hr	6	0	6	0	6	6	6	0
ethylene (C2H4)	lb/hr	784	0	784	0	784	784	784	0
acetylene (C2H2)	lb/hr	68	0	68	0	68	68	68	0
C3H8	lb/hr								
Pentane +	lb/hr								
Benzene (C6H6)	lb/hr	6	0	6	0	6	6	6	0
Tar (C10H8)	lb/hr	0	0	0	0	0	0	0	0
Carbon (Solid)	lb/hr								
Sulfur (Solid)	lb/hr								
Olivine (Solid)	lb/hr								
MgO (Solid)	lb/hr								
Ash	lb/hr								
Char	lb/hr								
Wood	lb/hr								
Enthalpy Flow	MMBTU/hr	-713.4	-333.9	-710.7	-326.7	-436.7	-440.0	-434.8	-5.2
Average Density	lb/ft^3	0.04	45.77	0.04	45.79	1.06	1.10	1.10	46.50

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QAK301	128.58	WK301	29838.9
QAK301A	48.50	WK301A	6478.0
QAK301B	27.64	WK301B	6369.0
QAK301C	19.71	WK301C	5757.7
QAK301D	17.29	WK301D	5738.0
QAK301E	15.43	WK301E	5496.1
QCH303	3.39		

Eq. No.	Equipment Name	Req	Spare	Equipment Type
H-302	Syngas Compressor Intercoolers	5		AIR-COOLED EXCHANGE
H-303	Water-cooled Aftercooler	1		SHELL-TUBE
K-301	Syngas Compressor	1		CENTRIFUGAL
S-301	Pre-compressor Knock-out	1		KNOCK-OUT DRUM
S-302	Syngas Compressor Interstage Knock-outs	4		KNOCK-OUT DRUM
S-303	Post-compressor Knock-out	1		KNOCK-OUT DRUM

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SECTION A300
GAS CLEAN-UP & COMPRESSION

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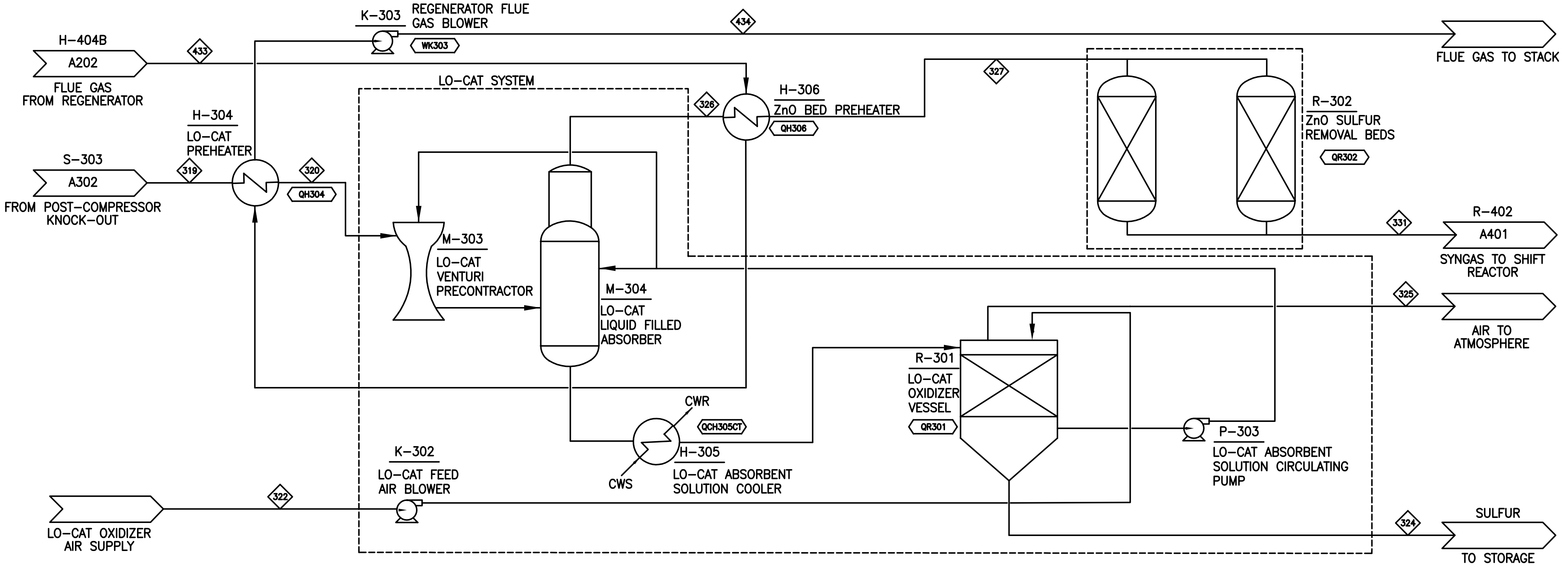

COMPONENT	UNITS	319	320	322	324	325	326	327	331	433	434
Total Flow	lb/hr	184,842	184,842	358	148	367	184,685	184,685	184,683	536,658	536,658
Temperature	F	110	120	90	110	110	120	707	682	250	256
Pressure	Psia	420.00	415.00	14.70	14.70	14.70	410.00	405.00	400.00	14.34	14.70
Vapor Fraction		1.00	1.00	1.00	0.00	0.71	1.00	1.00	1.00	1.00	1.00
Hydrogen	lb/hr	12,264	12,264				12,264	12,264	12,264		
Water	lb/hr	577	577	7		90	577	577	579	56,714	56,714
Carbon Monoxide	lb/hr	100,620	100,620				100,620	100,620	100,620		
Nitrogen	lb/hr	263	263	265		265	263	263	263	224,122	224,122
Oxygen	lb/hr			81		7				5,411	5,411
Argon	lb/hr			5		5				3,817	3,817
Carbon Dioxide	lb/hr	66,894	66,894	0		0	66,894	66,894	66,894	246,576	246,576
Hydrogen Sulfide (H2S)	lb/hr	161	161				4	4			
SO2	lb/hr									0	0
Ammonia (NH3)	lb/hr	30	30				30	30	30		
NO2	lb/hr									18	18
Methane	lb/hr	3,169	3,169				3,169	3,169	3,169		
isobutane	lb/hr										
n-butane	lb/hr										
ethane (C2H6)	lb/hr	6	6				6	6	6		
ethylene (C2H4)	lb/hr	784	784				784	784	784		
acetylene (C2H2)	lb/hr	68	68				68	68	68		
C3H8	lb/hr										
Pentane +	lb/hr										
Benzene (C6H6)	lb/hr	6	6				6	6	6		
Tar (C10H8)	lb/hr	0	0				0	0	0		
Carbon (Solid)	lb/hr										
Sulfur (Solid)	lb/hr				148						
Olivine (Solid)	lb/hr										
MgO (Solid)	lb/hr										
Ash	lb/hr										
Char	lb/hr										
Wood	lb/hr										
Enthalpy Flow	MMBTU/hr	-434.8	-433.9	0.0	0.0	-0.6	-433.9	-382.3	-384.6	-1252.7	-1251.9
Average Density	lb/ft^3	1.10	1.07	0.07	129.14	0.08	1.05	0.52	0.52	0.06	0.06

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QCH305CT	0.04	WK303	326.6
QH304	-0.86		
QH306	-51.59		
QR301	0.52		
QR302	2.28		

Eq. No.	Equipment Name	Req	Spare	Equipment Type
H-304	LO-CAT Preheater	1		SHELL-TUBE
H-305	LO-CAT Absorbent Solution Cooler	1		SHELL-TUBE
H-306	ZnO Bed Preheater	1		SHELL-TUBE
K-302	LO-CAT Feed Air Blower	1		CENTRIFUGAL
K-303	Regenerator Flue Gas Blower	1		CENTRIFUGAL
M-303	LO-CAT Venturi Precontactor	1		
M-304	LO-CAT Liquid-filled Absorber	1		ABSORBER
P-303	LO-CAT Absorbent Solution Circulating Pump	1	1	CENTRIFUGAL
R-301	LO-CAT Oxidizer Vessel	1		VERTICAL-VESSEL
R-302	ZnO Sulfur Removal Beds	2		VERTICAL-VESSEL

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GAS CLEAN-UP & COMPRESSION

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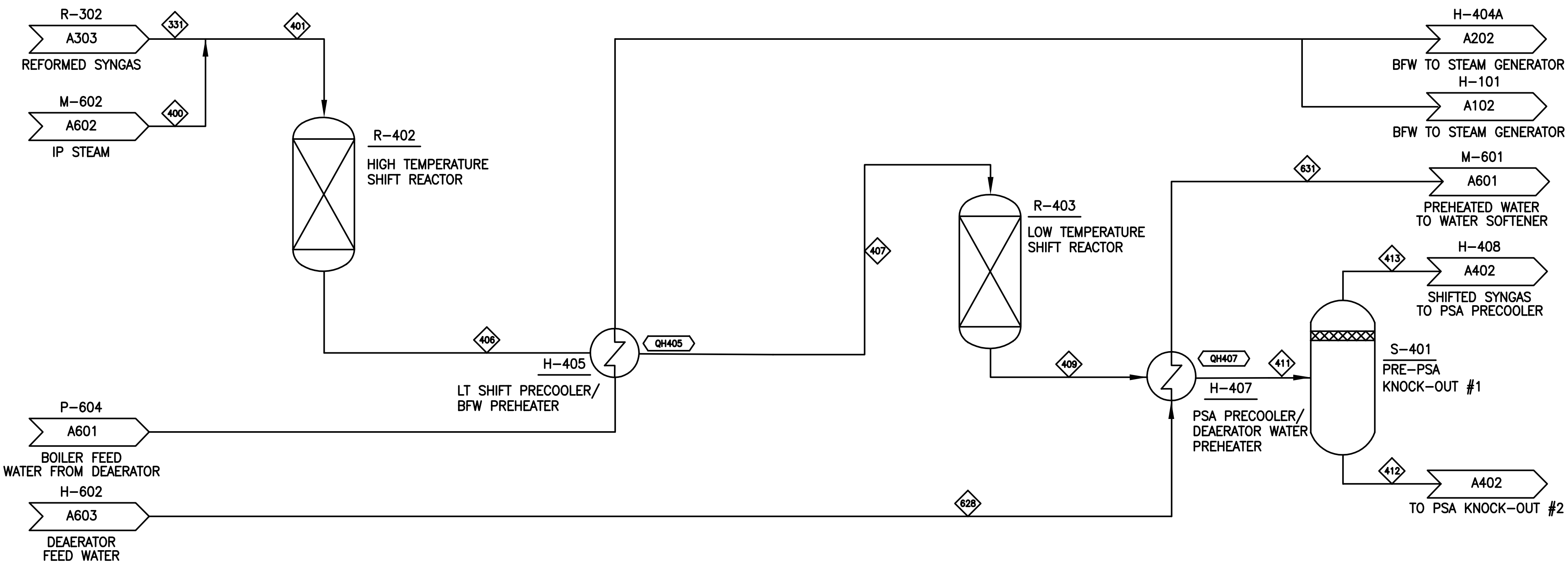

COMPONENT	UNITS	331	400	401	406	407	409	411	412	413	628	631
Total Flow	lb/hr	184,683	138,185	322,868	322,870	322,870	322,870	322,870	0	322,870	349,809	349,809
Temperature	F	682	710	686	910	392	507	313		313	104	227
Pressure	Psia	400.00	400.00	400.00	395.00	390.00	385.00	380.00	380.00	380.00	39.70	19.70
Vapor Fraction		1.00	1.00	1.00	1.00	1.00	1.00	1.00	0.00	1.00	0.00	0.00
Hydrogen	lb/hr	12,264		12,264	16,875	16,875	19,104	19,104		19,104		
Water	lb/hr	579	138,185	138,764	97,556	97,556	77,644	77,644		77,644	349,809	349,809
Carbon Monoxide	lb/hr	100,620		100,620	36,549	36,549	5,590	5,590		5,590		
Nitrogen	lb/hr	263		263	263	263	263	263		263		
Oxygen	lb/hr											
Argon	lb/hr											
Carbon Dioxide	lb/hr	66,894		66,894	167,562	167,562	216,206	216,206		216,206		
Hydrogen Sulfide (H2S)	lb/hr											
SO2	lb/hr											
Ammonia (NH3)	lb/hr	30		30	30	30	30	30		30		
NO2	lb/hr											
Methane	lb/hr	3,169		3,169	3,169	3,169	3,169	3,169		3,169		
isobutane	lb/hr											
n-butane	lb/hr											
ethane (C2H6)	lb/hr	6		6	6	6	6	6		6		
ethylene (C2H4)	lb/hr	784		784	784	784	784	784		784		
acetylene (C2H2)	lb/hr	68		68	68	68	68	68		68		
C3H8	lb/hr											
Pentane +	lb/hr											
Benzene (C6H6)	lb/hr	6		6	6	6	6	6		6		
Tar (C10H8)	lb/hr	0		0	0	0	0	0		0		
Carbon (Solid)	lb/hr											
Sulfur (Solid)	lb/hr											
Olivine (Solid)	lb/hr											
MgO (Solid)	lb/hr											
Ash	lb/hr											
Char	lb/hr											
Wood	lb/hr											
Enthalpy Flow	MMBTU/hr	-384.6	-760.1	-1144.7	-1144.7	-1230.1	-1230.1	-1261.5		-1261.5	-2378.1	-2335.0
Average Density	lb/ft^3	0.52	0.60	0.55	0.45	0.73	0.63	0.78		0.78	61.95	59.43

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QH405	85.42		
QH407	31.41		

Eq. No.	Equipment Name	Req	Spare	Equipment Type
H-405	LT Shift Precooler/BFW Preheater #1	1		SHELL-TUBE
H-407	PSA Precooler / Deaerator Water Preheater #2	1		SHELL-TUBE
R-402	High Temperature Shift Reactor	1		VERTICAL-VESSEL
R-403	Low Temperature Shift Reactor	1		VERTICAL-VESSEL
S-401	Pre-PSA Knock-out #1	1		KNOCK-OUT DRUM

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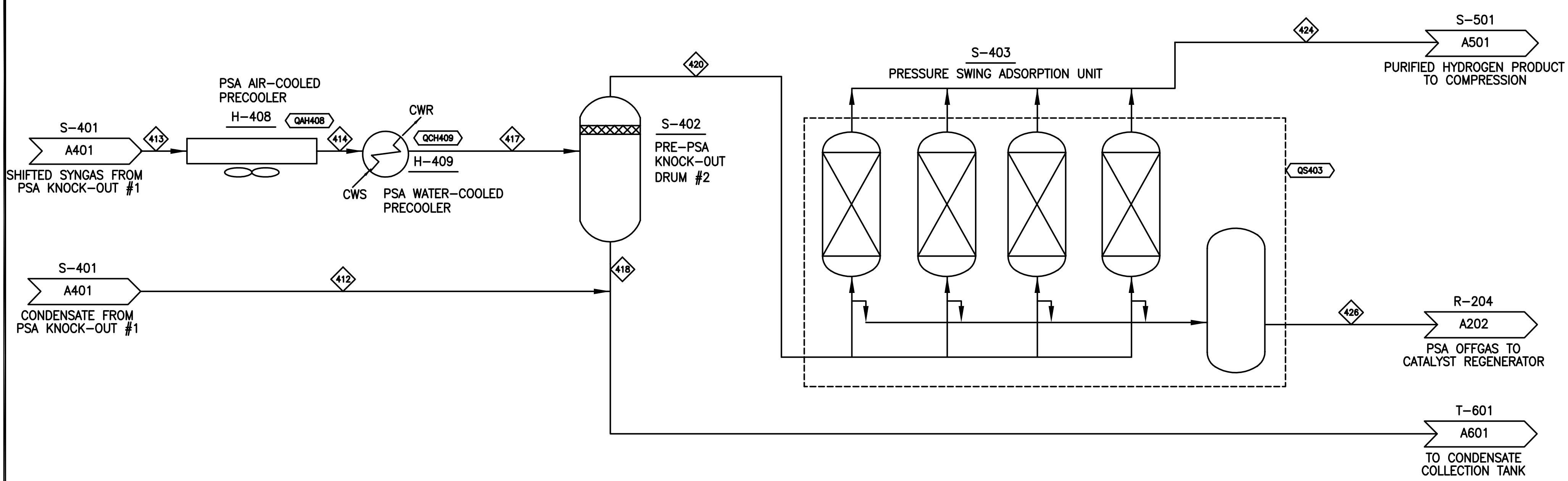

COMPONENT	UNITS	412	413	414	417	418	420	424	426
Total Flow	lb/hr	0	322,870	322,870	322,870	76,853	246,017	15,322	230,694
Temperature	F		313	140	110	110	110	109	109
Pressure	Psia	380.00	380.00	375.00	370.00	370.00	360.00	360.00	14.70
Vapor Fraction		0.00	1.00	0.78	0.78	0.00	1.00	1.00	1.00
Hydrogen	lb/hr		19,104	19,104	19,104	0	19,104	15,322	3,781
Water	lb/hr		77,644	77,644	77,644	76,791	853		853
Carbon Monoxide	lb/hr		5,590	5,590	5,590	0	5,590		5,590
Nitrogen	lb/hr		263	263	263	0	263		263
Oxygen	lb/hr								
Argon	lb/hr								
Carbon Dioxide	lb/hr		216,206	216,206	216,206	38	216,168		216,168
Hydrogen Sulfide (H2S)	lb/hr								
SO2	lb/hr								
Ammonia (NH3)	lb/hr		30	30	30	24	7		7
NO2	lb/hr								
Methane	lb/hr		3,169	3,169	3,169	0	3,169		3,169
isobutane	lb/hr								
n-butane	lb/hr								
ethane (C2H6)	lb/hr		6	6	6	0	6		6
ethylene (C2H4)	lb/hr		784	784	784	0	784		784
acetylene (C2H2)	lb/hr		68	68	68	0	68		68
C3H8	lb/hr								
Pentane +	lb/hr								
Benzene (C6H6)	lb/hr		6	6	6	0	6		6
Tar (C10H8)	lb/hr		0	0	0	0	0		0
Carbon (Solid)	lb/hr								
Sulfur (Solid)	lb/hr								
Olivine (Solid)	lb/hr								
MgO (Solid)	lb/hr								
Ash	lb/hr								
Char	lb/hr								
Wood	lb/hr								
Enthalpy Flow	MMBTU/hr		-1261.5	-1368.3	-1375.6	-526.8	-848.8	1.7	-849.6
Average Density	lb/ft^3		0.78	1.26	1.31	46.53	0.98	0.12	0.08

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QAH408	106.74		
QCH409	7.35		
QS403	-0.93		

Eq. No.	Equipment Name	Req	Spare	Equipment Type
H-408	PSA Air-cooled Precooler	1		AIR-COOLED EXCHANGE
H-409	PSA Water-cooled Precooler	1		SHELL-TUBE
S-402	Pre-PSA Knock-out #2	1		KNOCK-OUT DRUM
S-403	Pressure Swing Adsorption Unit	1		PACKAGE

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REFORMING, SHIFT & PSA

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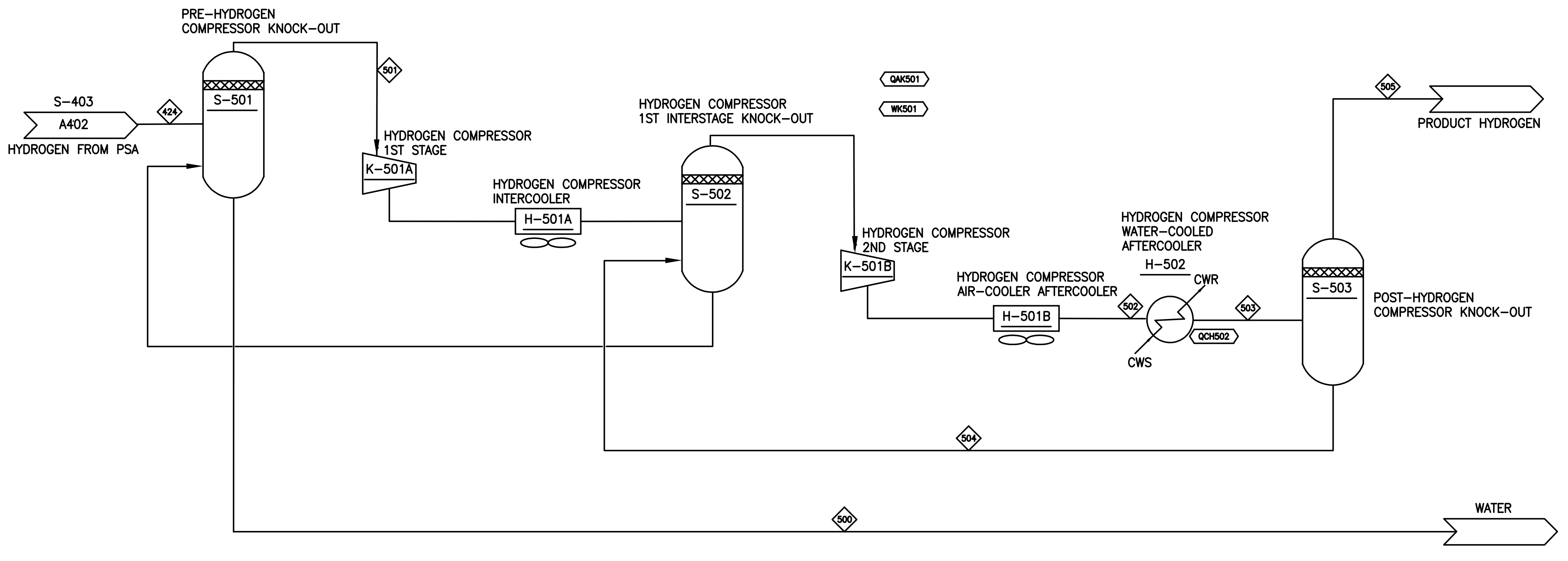

COMPONENT	UNITS	424	500	501	502	503	504	505
Total Flow	lb/hr	15,322	0	15,322	15,322	15,322	0	15,322
Temperature	F	109		109	140	110		110
Pressure	Psia	360.00	360.00	360.00	1019.70	1014.70	1014.70	1014.70
Vapor Fraction		1.00	0.00	1.00	1.00	1.00	0.00	1.00
Hydrogen	lb/hr	15,322		15,322	15,322	15,322		15,322
Water	lb/hr							
Carbon Monoxide	lb/hr							
Nitrogen	lb/hr							
Oxygen	lb/hr							
Argon	lb/hr							
Carbon Dioxide	lb/hr							
Hydrogen Sulfide (H2S)	lb/hr							
SO2	lb/hr							
Ammonia (NH3)	lb/hr							
NO2	lb/hr							
Methane	lb/hr							
isobutane	lb/hr							
n-butane	lb/hr							
ethane (C2H6)	lb/hr							
ethylene (C2H4)	lb/hr							
acetylene (C2H2)	lb/hr							
C3H8	lb/hr							
Pentane +	lb/hr							
Benzene (C6H6)	lb/hr							
Tar (C10H8)	lb/hr							
Carbon (Solid)	lb/hr							
Sulfur (Solid)	lb/hr							
Olivine (Solid)	lb/hr							
MgO (Solid)	lb/hr							
Ash	lb/hr							
Char	lb/hr							
Wood	lb/hr							
Enthalpy Flow	MMBTU/hr	1.7		1.7	3.5	1.9		1.9
Average Density	lb/ft^3	0.12		0.12	0.31	0.32		0.32

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QAK501	10.79	WK501	4936.8
QAK501A	4.36	WK501A	2368.3
QAK501B	6.43	WK501B	2568.5
QCH502	1.57		

Eq. No.	Equipment Name	Req	Spare	Equipment Type
H-501A	Hydrogen Compressor Intercooler	1		AIR-COOLED EXCHANGE
H-501B	Hydrogen Compressor Air-cooled Aftercooler	1		AIR-COOLED EXCHANGE
H-502	Hydrogen Compressor Water-cooler Aftercooler	1		SHELL-TUBE
K-501	Hydrogen Compressor	1		RECIPROCATING
S-501	Pre-hydrogen Compressor Knock-out	1		KNOCK-OUT DRUM
S-502	Hydrogen Compressor 1st Interstage Knock-out	1		KNOCK-OUT DRUM
S-503	Post-hydrogen Compressor Knock-out	1		KNOCK-OUT DRUM

VER	DESCRIPTION	DATE
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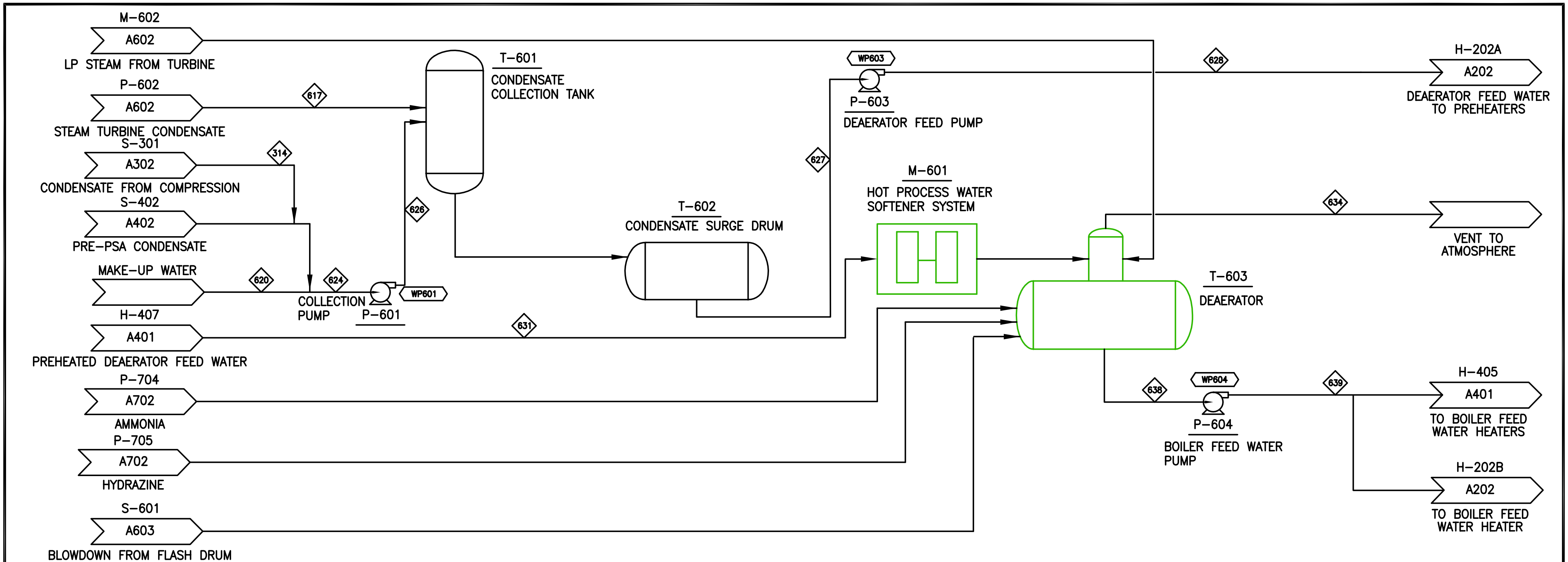
SECTION A500
HYDROGEN COMPRESSION

AA 11-17-04 | PS0410a_bhG.xls | PFD-P710-A501 | D



COMPONENT	UNITS	314	617	620	624	626	627	628	631	634	638	639
Total Flow	lb/hr	48,981	131,510	92,529	218,299	218,299	349,809	349,809	349,809	64	349,812	349,812
Temperature	F	147	115	60	97	97	104	104	227	125	227	230
Pressure	Psia	15.00	14.70	14.70	14.70	14.70	14.70	39.70	19.70	14.70	19.70	1344.70
Vapor Fraction		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.00	0.00	0.00
Hydrogen	lb/hr	0								0		
Water	lb/hr	48,978	131,510	92,529	218,299	218,299	349,809	349,809	349,809		349,812	349,812
Carbon Monoxide	lb/hr	0								0		
Nitrogen	lb/hr	0								0		
Oxygen	lb/hr											
Argon	lb/hr											
Carbon Dioxide	lb/hr	0								39		
Hydrogen Sulfide (H2S)	lb/hr	0								0		
SO2	lb/hr											
Ammonia (NH3)	lb/hr	2								26		
NO2	lb/hr											
Methane	lb/hr	0								0		
isobutane	lb/hr											
n-butane	lb/hr											
ethane (C2H6)	lb/hr	0								0		
ethylene (C2H4)	lb/hr	0								0		
acetylene (C2H2)	lb/hr	0								0		
C3H8	lb/hr											
Pentane +	lb/hr											
Benzene (C6H6)	lb/hr	0								0		
Tar (C10H8)	lb/hr	0								0		
Carbon (Solid)	lb/hr											
Sulfur (Solid)	lb/hr											
Olivine (Solid)	lb/hr											
MgO (Solid)	lb/hr											
Ash	lb/hr											
Char	lb/hr											
Wood	lb/hr											
Enthalpy Flow	MMBTU/hr	-333.9	-892.6	-633.1	-1485.5	-1485.5	-2378.1	-2378.1	-2335.0	-0.2	-2335.0	-2333.1
Average Density	lb/ft^3	45.77	61.79	62.37	62.03	62.03	61.95	61.95	59.43	0.06	59.43	59.64

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
		WP601	3.7
		WP603	13.8
		WP604	760.0

Eq. No.	Equipment Name	Req	Spare	Equipment Type
M-601	Hot Process Water Softener System	1		PACKAGE
P-601	Collection Pump	1	1	CENTRIFUGAL
P-603	Deaerator Feed Pump	1	1	CENTRIFUGAL
P-604	Boiler Feed Water Pump	1	1	CENTRIFUGAL
T-601	Condensate Collection Tank	1		HORIZONTAL-VESSEL
T-602	Condensate Surge Drum	1		HORIZONTAL-VESSEL
T-603	Deaerator	1		HORIZONTAL-VESSEL

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SECTION 600
STEAM SYSTEM & PWR GENERATION

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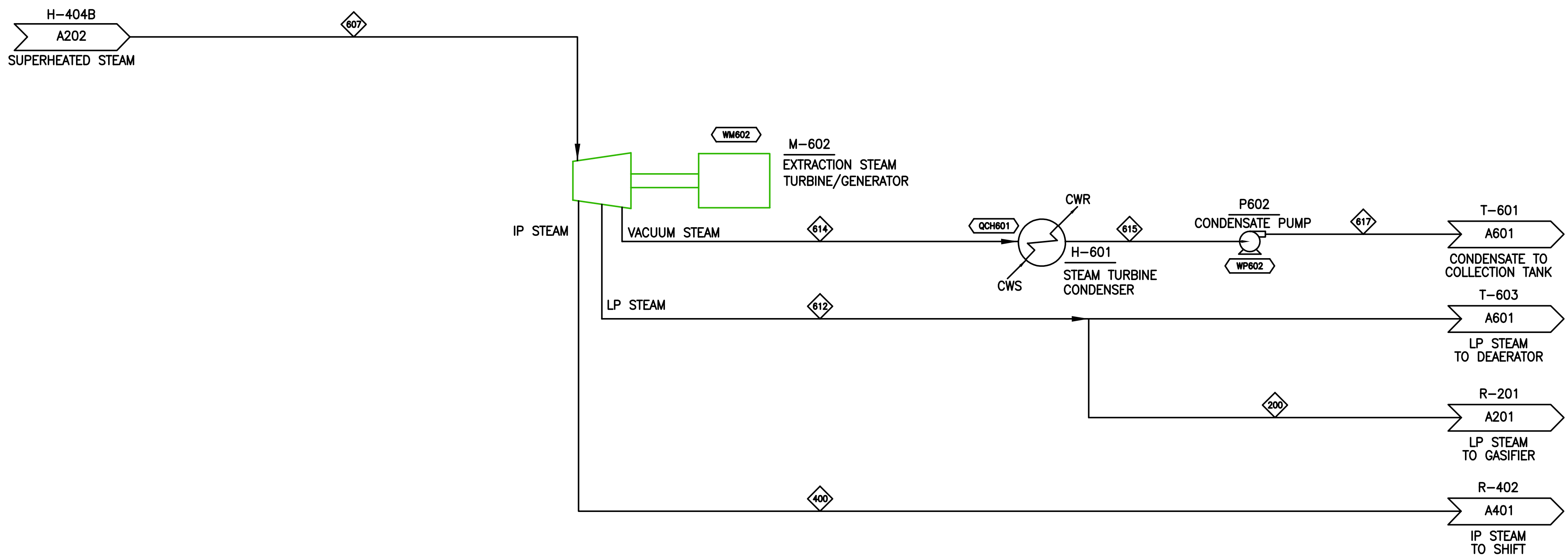

COMPONENT	UNITS	200	400	607	612	614	615	617
Total Flow	lb/hr	73,120	138,185	342,816	73,120	131,510	131,510	131,510
Temperature	F	259	710	1,000	265	115	115	115
Pressure	Psia	25.00	400.00	1264.70	35.00	1.47	1.47	14.70
Vapor Fraction		1.00	1.00	1.00	1.00	0.89	0.00	0.00
Hydrogen	lb/hr							
Water	lb/hr	73,120	138,185	342,816	73,120	131,510	131,510	131,510
Carbon Monoxide	lb/hr							
Nitrogen	lb/hr							
Oxygen	lb/hr							
Argon	lb/hr							
Carbon Dioxide	lb/hr							
Hydrogen Sulfide (H2S)	lb/hr							
SO2	lb/hr							
Ammonia (NH3)	lb/hr							
NO2	lb/hr							
Methane	lb/hr							
isobutane	lb/hr							
n-butane	lb/hr							
ethane (C2H6)	lb/hr							
ethylene (C2H4)	lb/hr							
acetylene (C2H2)	lb/hr							
C3H8	lb/hr							
Pentane +	lb/hr							
Benzene (C6H6)	lb/hr							
Tar (C10H8)	lb/hr							
Carbon (Solid)	lb/hr							
Sulfur (Solid)	lb/hr							
Olivine (Solid)	lb/hr							
MgO (Solid)	lb/hr							
Ash	lb/hr							
Char	lb/hr							
Wood	lb/hr							
Enthalpy Flow	MMBTU/hr	-416.5	-760.1	-1842.1	-416.8	-772.6	-892.6	-892.6
Average Density	lb/ft^3	0.06	0.60	1.54	0.08	0.00	61.79	61.79

Heat Stream No.	MM BTU/hr		Work Stream No.	HP
QCH601	120.05		WM602A	-17312.2
			WM602B	-15977.1
			WM602C	-8995.6
			WP602	5.5

Eq. No.	Equipment Name	Req	Spare	Equipment Type
H-601	Steam Turbine Condenser	1		SHELL-TUBE
M-602	Extraction Steam Turbine/Generator	1		STEAM-TURBINE
P-602	Condensate Pump	1	1	CENTRIFUGAL

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SECTION A600
STEAM SYSTEM & PWR GENERATION

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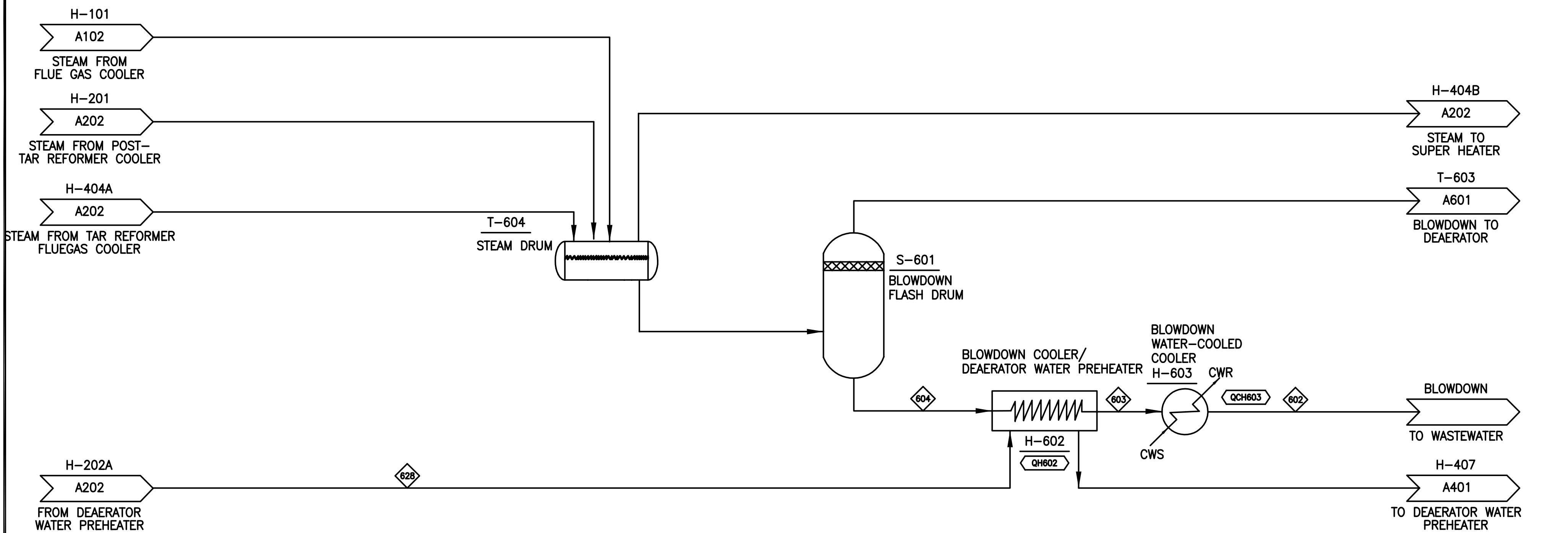

COMPONENT	UNITS	602	603	604	605	628
Total Flow	lb/hr	6,996	6,996	6,996	342,816	349,809
Temperature	F	110	200	575	575	104
Pressure	Psia	1269.70	1274.70	1279.70	1279.70	39.70
Vapor Fraction		0.00	0.00	0.00	1.00	0.00
Hydrogen	lb/hr					
Water	lb/hr	6,996	6,996	6,996	342,816	349,809
Carbon Monoxide	lb/hr					
Nitrogen	lb/hr					
Oxygen	lb/hr					
Argon	lb/hr					
Carbon Dioxide	lb/hr					
Hydrogen Sulfide (H2S)	lb/hr					
SO2	lb/hr					
Ammonia (NH3)	lb/hr					
NO2	lb/hr					
Methane	lb/hr					
isobutane	lb/hr					
n-butane	lb/hr					
ethane (C2H6)	lb/hr					
ethylene (C2H4)	lb/hr					
acetylene (C2H2)	lb/hr					
C3H8	lb/hr					
Pentane +	lb/hr					
Benzene (C6H6)	lb/hr					
Tar (C10H8)	lb/hr					
Carbon (Solid)	lb/hr					
Sulfur (Solid)	lb/hr					
Olivine (Solid)	lb/hr					
MgO (Solid)	lb/hr					
Ash	lb/hr					
Char	lb/hr					
Wood	lb/hr					
Enthalpy Flow	MMBTU/h	-47.5	-46.9	-44.0	-1950.4	-2378.1
Average Density	lb/ft^3	62.10	60.36	44.21	2.97	61.95

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QCH603	0.63		
QH602	2.88		

Eq. No.	Equipment Name	Req	Spare	Equipment Type
H-602	Blowdown Cooler / Deaerator Water Preheater	1		SHELL-TUBE
H-603	Blowdown Water-cooled Cooler	1		SHELL-TUBE
S-601	Blowdown Flash Drum	1		HORIZONTAL-VESSEL
T-604	Steam Drum	1		HORIZONTAL-VESSEL

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SECTION A600
STEAM SYSTEM & PWR GENERATION

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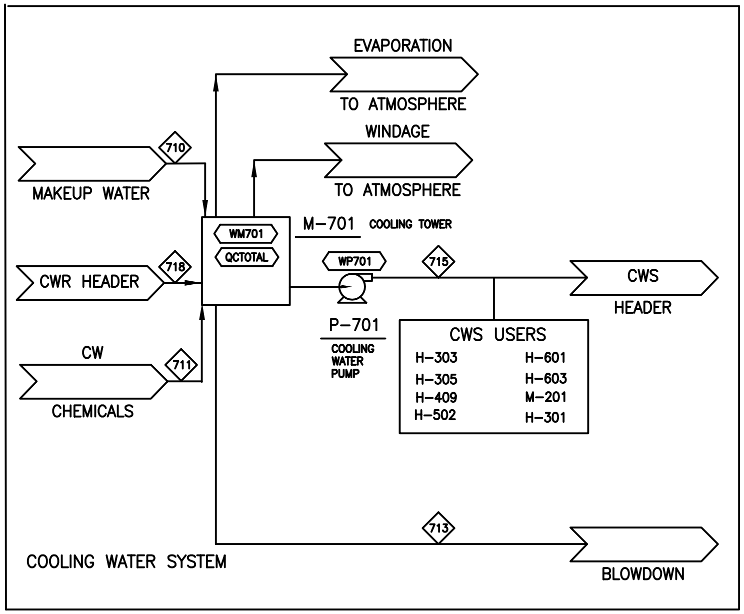

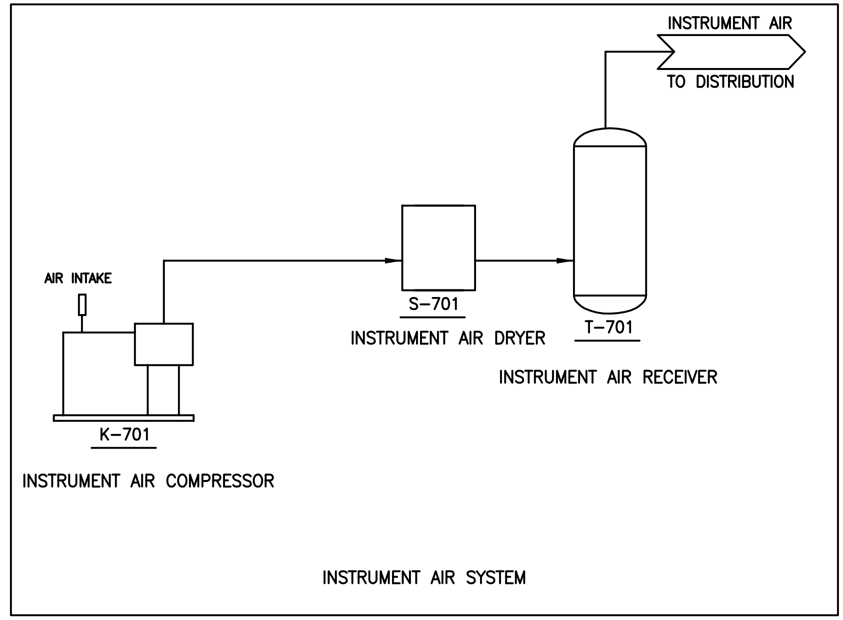

COMPONENT	UNITS	710	711	713	715	718
Total Flow	lb/hr	137,169	1	26,305	6,319,444	6,319,444
Temperature	F	60	60	90	90	110
Pressure	Psia	14.70	14.70	14.70	74.70	59.70
Vapor Fraction		0.00	0.00	0.00	0.00	0.00
Hydrogen	lb/hr					
Water	lb/hr	137,169	1	26,305	6,319,444	6,319,444
Carbon Monoxide	lb/hr					
Nitrogen	lb/hr					
Oxygen	lb/hr					
Argon	lb/hr					
Carbon Dioxide	lb/hr					
Hydrogen Sulfide (H2S)	lb/hr					
SO2	lb/hr					
Ammonia (NH3)	lb/hr					
NO2	lb/hr					
Methane	lb/hr					
isobutane	lb/hr					
n-butane	lb/hr					
ethane (C2H6)	lb/hr					
ethylene (C2H4)	lb/hr					
acetylene (C2H2)	lb/hr					
C3H8	lb/hr					
Pentane +	lb/hr					
Benzene (C6H6)	lb/hr					
Tar (C10H8)	lb/hr					
Carbon (Solid)	lb/hr					
Sulfur (Solid)	lb/hr					
Olivine (Solid)	lb/hr					
MgO (Solid)	lb/hr					
Ash	lb/hr					
Char	lb/hr					
Wood	lb/hr					
Enthalpy Flow	MMBTU/h	-948.8	0.0	-181.0	-43490.9	-43345.8
Average Density	lb/ft^3	47.44	47.44	46.89	46.89	46.51

Heat Stream No.	MM BTU/hr	Work Stream No.	HP
QCTOTAL	145.16	WM701	678.0
		WP701	684.3

Eq. No.	Equipment Name	Req	Spare	Equipment Type
K-701	Plant Air Compressor	2	1	RECIPROCATING
M-701	Cooling Tower System	1		INDUCED-DRAFT
P-701	Cooling Water Pump	1	1	CENTRIFUGAL
S-701	Instrument Air Dryer	1	1	PACKAGE
T-701	Plant Air Receiver	1		HORIZONTAL-VESSEL

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SECTION A700
COOLING WATER & OTHER UTILITIES

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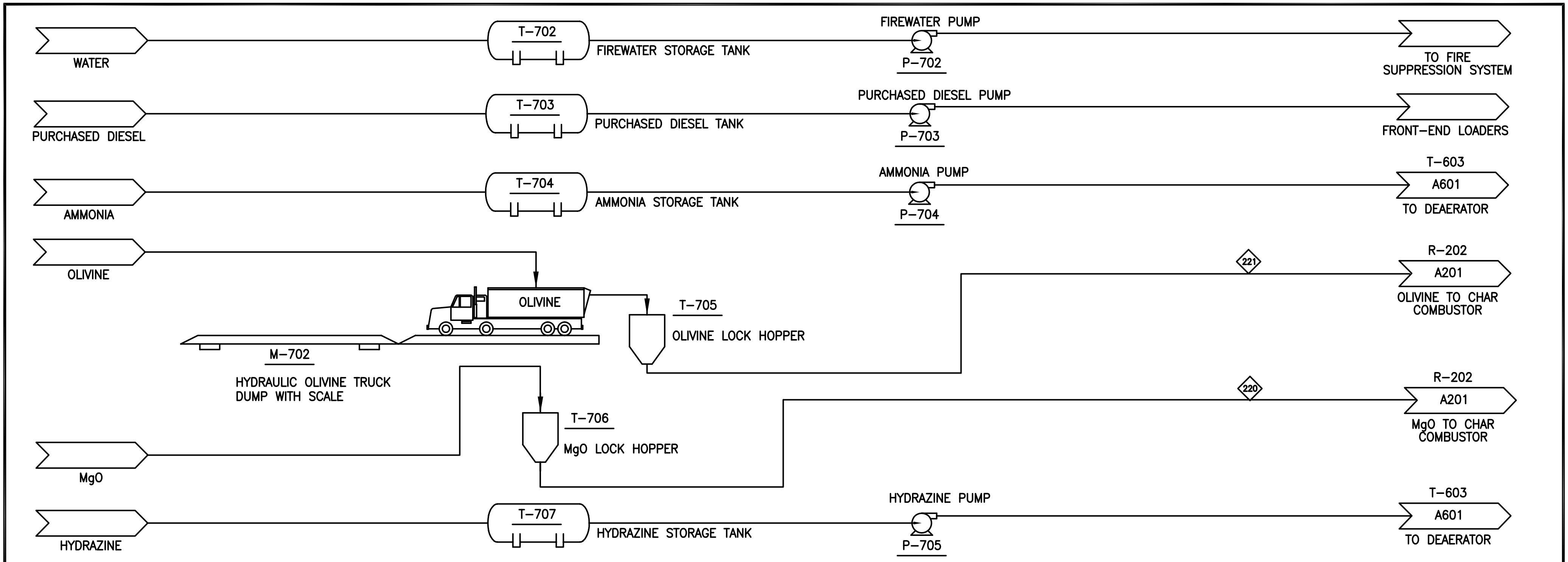

COMPONENT	UNITS	220	221
Total Flow	lb/hr	7	5,440
Temperature	F	60	60
Pressure	Psia	25.00	25.00
Vapor Fraction		0.00	0.00
Hydrogen	lb/hr		
Water	lb/hr		
Carbon Monoxide	lb/hr		
Nitrogen	lb/hr		
Oxygen	lb/hr		
Argon	lb/hr		
Carbon Dioxide	lb/hr		
Hydrogen Sulfide (H2S)	lb/hr		
SO2	lb/hr		
Ammonia (NH3)	lb/hr		
NO2	lb/hr		
Methane	lb/hr		
isobutane	lb/hr		
n-butane	lb/hr		
ethane (C2H6)	lb/hr		
ethylene (C2H4)	lb/hr		
acetylene (C2H2)	lb/hr		
C3H8	lb/hr		
Pentane +	lb/hr		
Benzene (C6H6)	lb/hr		
Tar (C10H8)	lb/hr		
Carbon (Solid)	lb/hr		
Sulfur (Solid)	lb/hr		
Olivine (Solid)	lb/hr	7	5,440
MgO (Solid)	lb/hr		
Ash	lb/hr		
Char	lb/hr		
Wood	lb/hr		
Enthalpy Flow	MMBTU/hr	0.0	0.0
Average Density	lb/ft^3	165.39	165.39

Heat Stream No.	MM BTU/hr	Work Stream No.	HP

Eq. No.	Equipment Name	Req	Spare	Equipment Type
M-702	Hydraulic Truck Dump with Scale	1		TRUCK-SCALE
P-702	Firewater Pump	1	1	CENTRIFUGAL
P-703	Diesel Pump	1	1	CENTRIFUGAL
P-704	Ammonia Pump	1	1	CENTRIFUGAL
P-705	Hydrazine Pump	1		CENTRIFUGAL
T-702	Firewater Storage Tank	1		FLAT-BTM-STORAGE
T-703	Diesel Storage Tank	1		FLAT-BTM-STORAGE
T-704	Ammonia Storage Tank	1		HORIZONTAL-STORAGE
T-705	Olivine Lock Hopper	1		VERTICAL-VESSEL
T-706	MgO Lock Hopper	1		VERTICAL-VESSEL
T-707	Hydrazine Storage Tank	1		VERTICAL-VESSEL

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SECTION A700
COOLING WATER & OTHER UTILITIES

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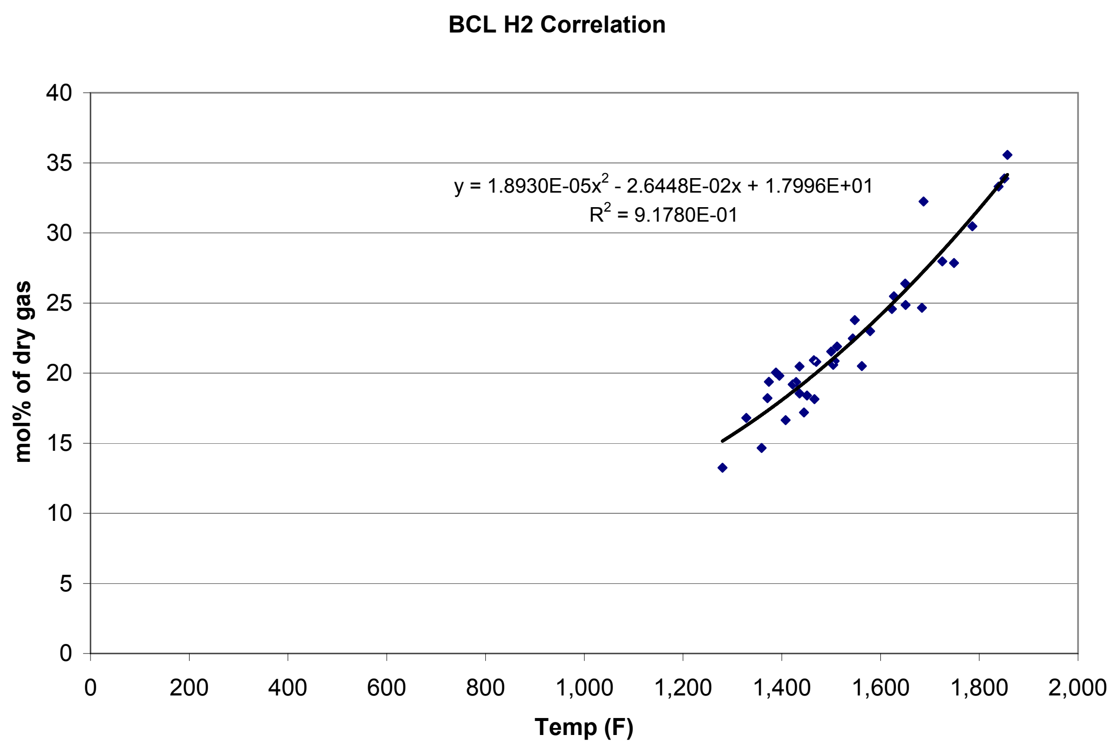
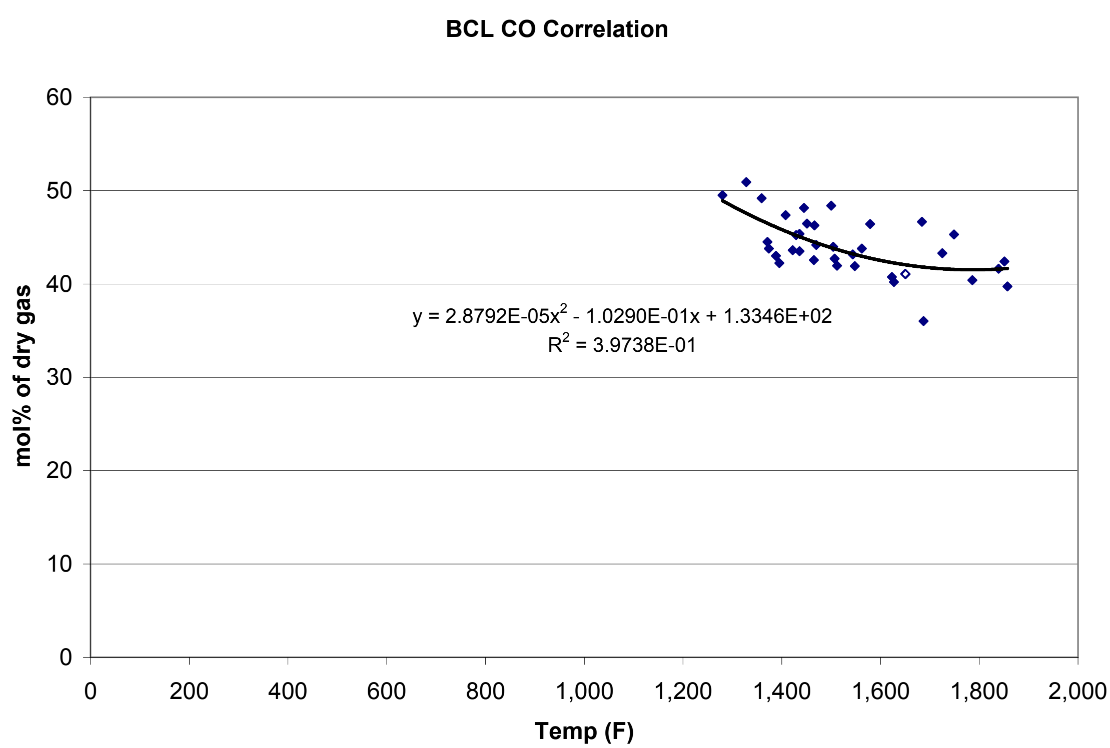
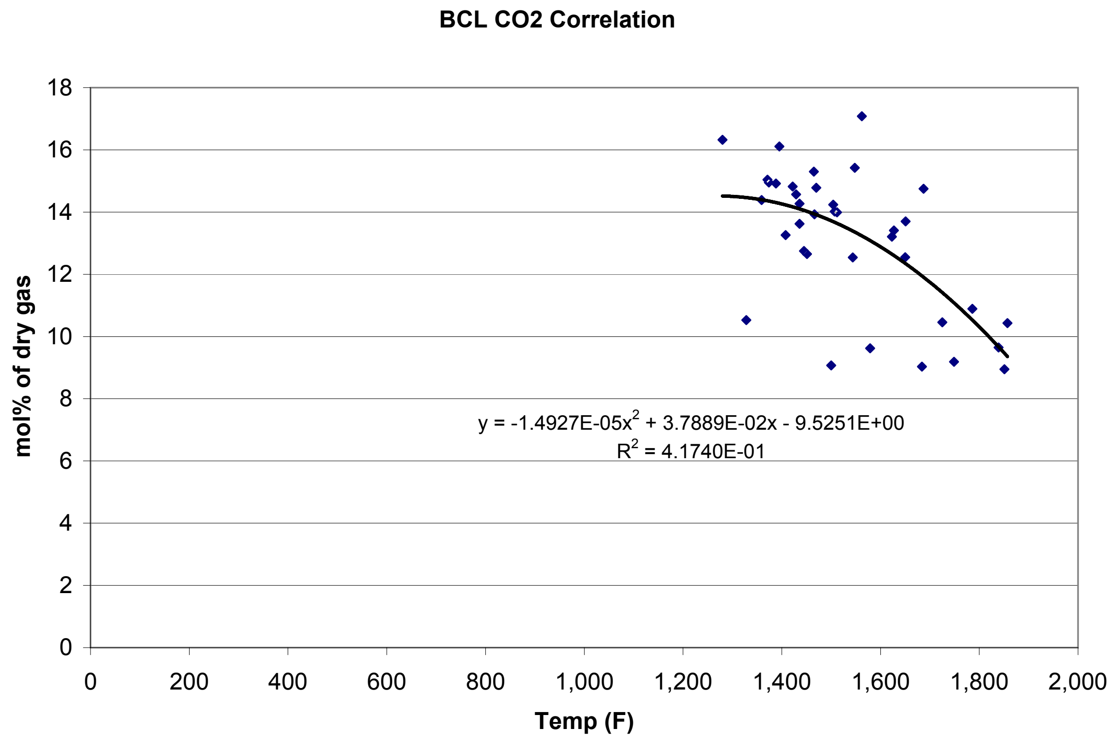
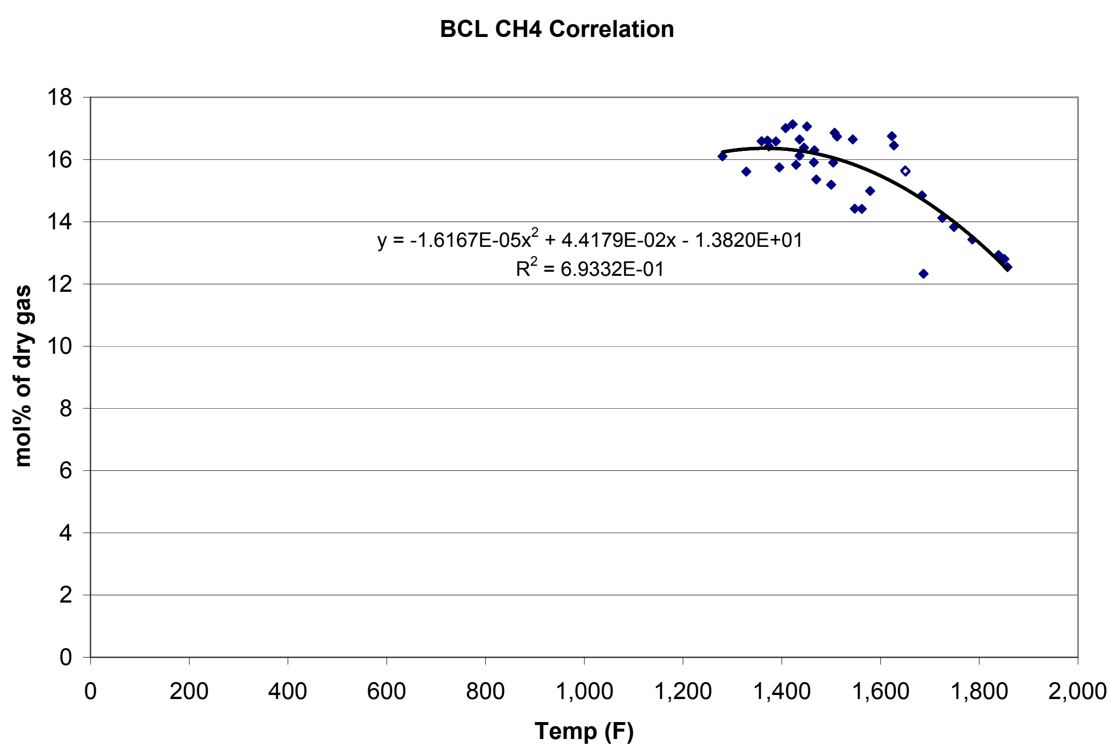
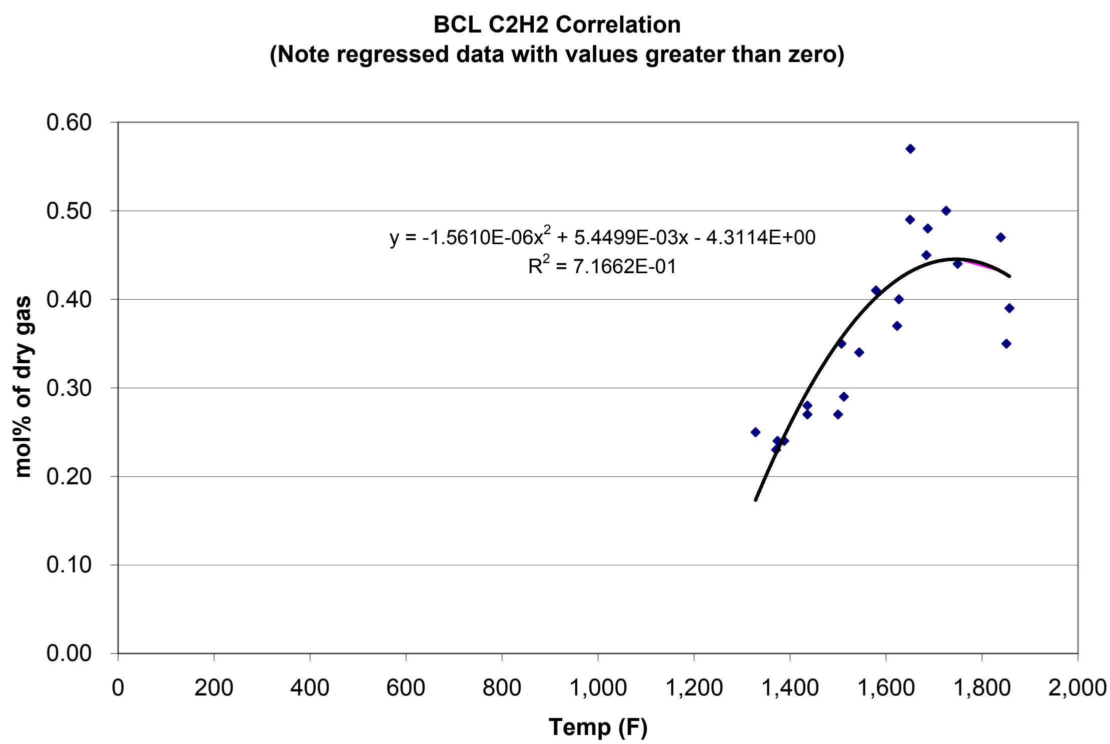
Appendix E: Graphical Correlations for Gas Components and Char

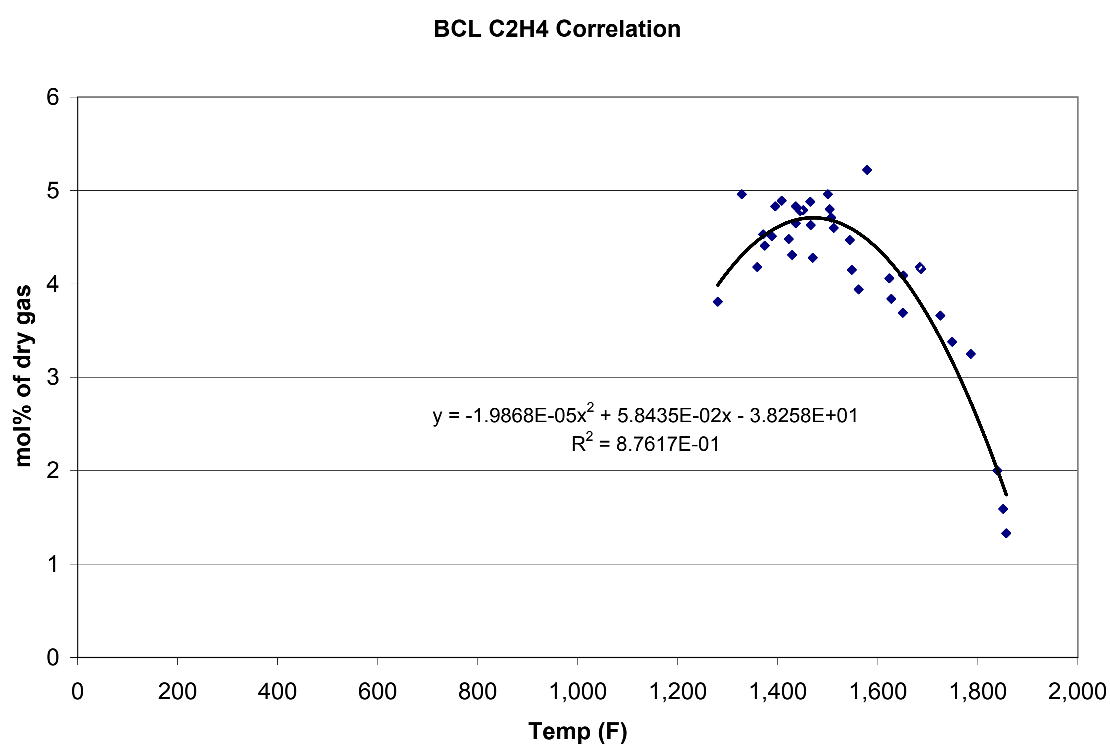

BCL H2 Correlation
y = 1.8930E-05x2 - 2.6448E-02x + 1.7996E+01
R2 = 9.1780E-01
mol% of dry gas
Temp (F)
0
5
10
15
20
25
30
35
40
0
200
400
600
800
1,000
1,200
1,400
1,600
1,800
2,000

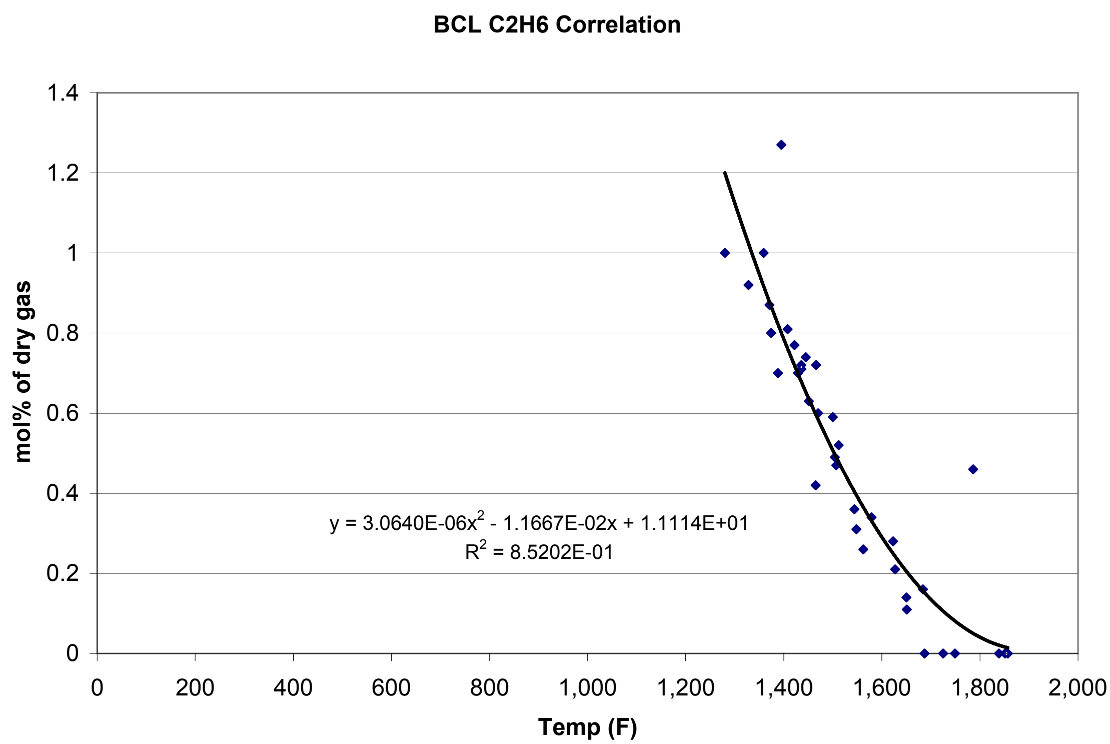

BCL CO Correlation
y = 2.8792E-05x2 - 1.0290E-01x + 1.3346E+02
R2 = 3.9738E-01
mol% of dry gas
Temp (F)
0
10
20
30
40
50
60
0
200
400
600
800
1,000
1,200
1,400
1,600
1,800
2,000


BCL CO2 Correlation
mol% of dry gas
Temp (F)
y = -1.4927E-05x2 + 3.7889E-02x - 9.5251E+00
R2 = 4.1740E-01
0
2
4
6
8
10
12
14
16
18
0
200
400
600
800
1,000
1,200
1,400
1,600
1,800
2,000

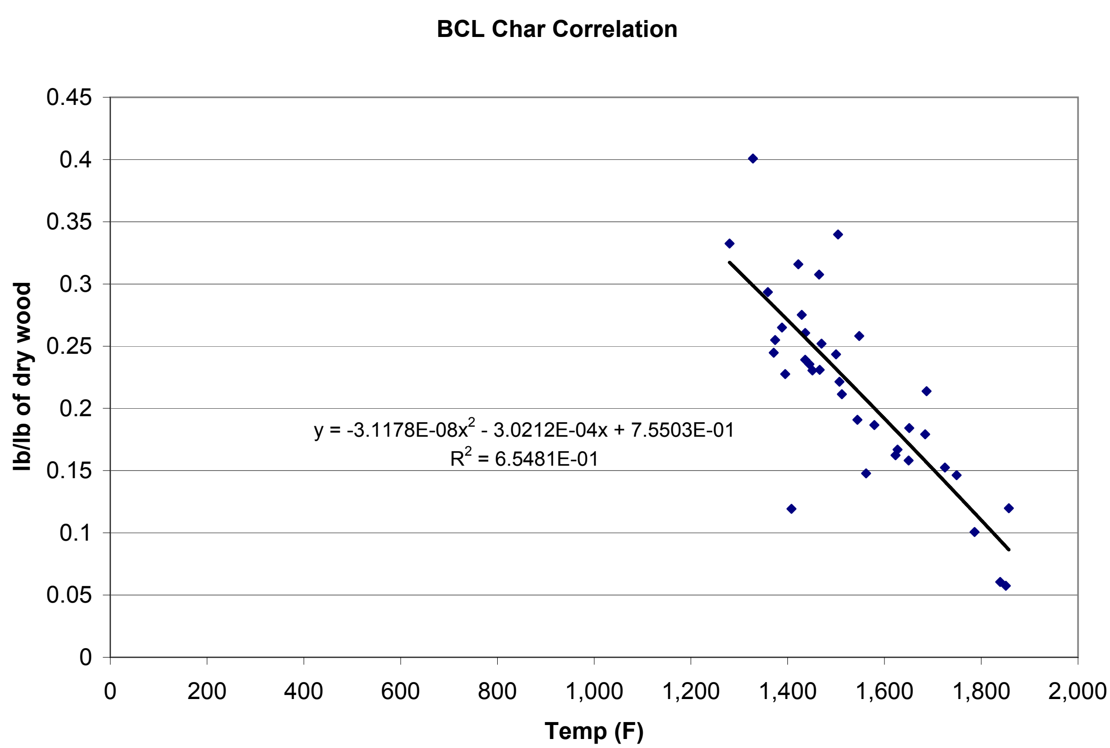

BCL CH4 Correlation
mol% of dry gas
Temp (F)
y = -1.6167E-05x2 + 4.4179E-02x - 1.3820E+01
R2 = 6.9332E-01
0
2
4
6
8
10
12
14
16
18
0
200
400
600
800
1,000
1,200
1,400
1,600
1,800
2,000


BCL C2H2 Correlation
(Note regressed data with values greater than zero)
y = -1.5610E-06x2 + 5.4499E-03x - 4.3114E+00
R2 = 7.1662E-01
mol% of dry gas
0.60
0.50
0.40
0.30
0.20
0.10
0.00
0
200
400
600
800
1,000
1,200
1,400
1,600
1,800
2,000
Temp (F)

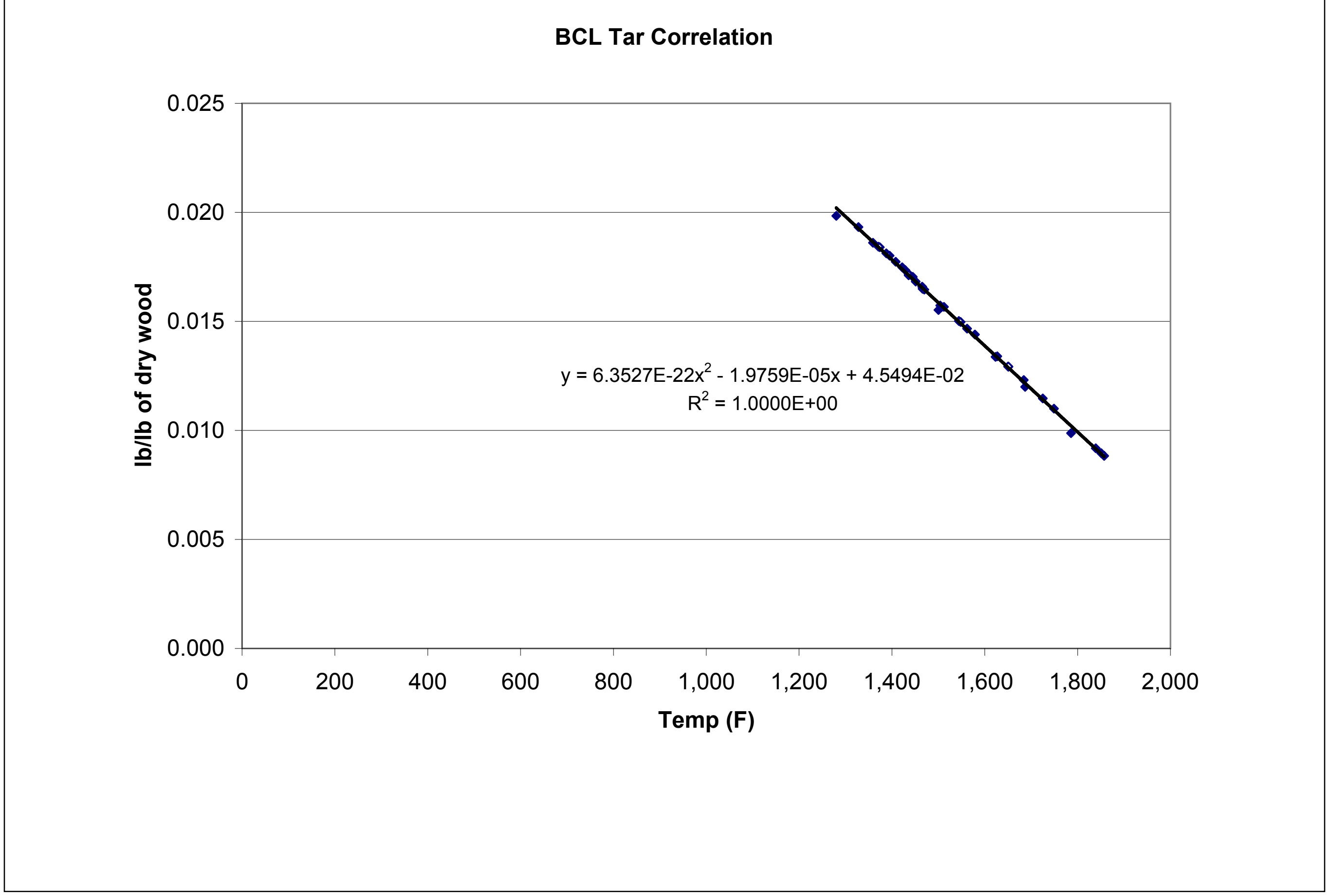

BCL C2H4 Correlation
y = -1.9868E-05x2 + 5.8435E-02x - 3.8258E+01
R2 = 8.7617E-01
mol% of dry gas
6
5
4
3
2
1
0
0
200
400
600
800
1,000
1,200
1,400
1,600
1,800
2,000
Temp (F)



BCL C2H6 Correlation
mol% of dry gas
Temp (F)
y = 3.0640E-06x2 - 1.1667E-02x + 1.1114E+01
R2 = 8.5202E-01
0
0.2
0.4
0.6
0.8
1
1.2
1.4
0
200
400
600
800
1,000
1,200
1,400
1,600
1,800
2,000



BCL Char Correlation
lb/lb of dry wood
Temp (F)
y = -3.1178E-08x2 - 3.0212E-04x + 7.5503E-01
R2 = 6.5481E-01
0
0.05
0.1
0.15
0.2
0.25
0.3
0.35
0.4
0.45
0
200
400
600
800
1,000
1,200
1,400
1,600
1,800
2,000

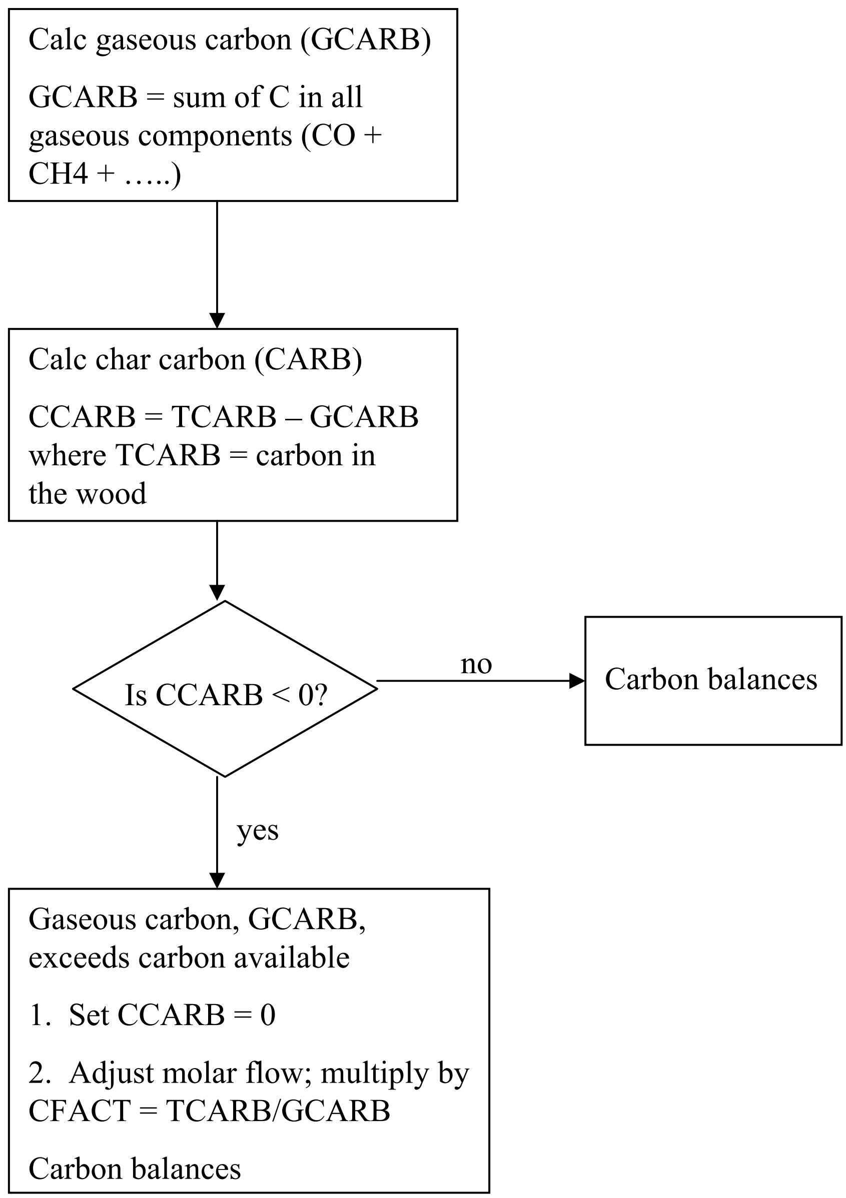

BCL Tar Correlation
0.025
0.020
0.015
0.010
0.005
0.000
lb/lb of dry wood
y = 6.3527E-22x2 - 1.9759E-05x + 4.5494E-02
R2 = 1.0000E+00
0
200
400
600
800
1,000
1,200
1,400
1,600
1,800
2,000
Temp (F)

Appendix F: Flow Charts for Gasifier Elemental Balances

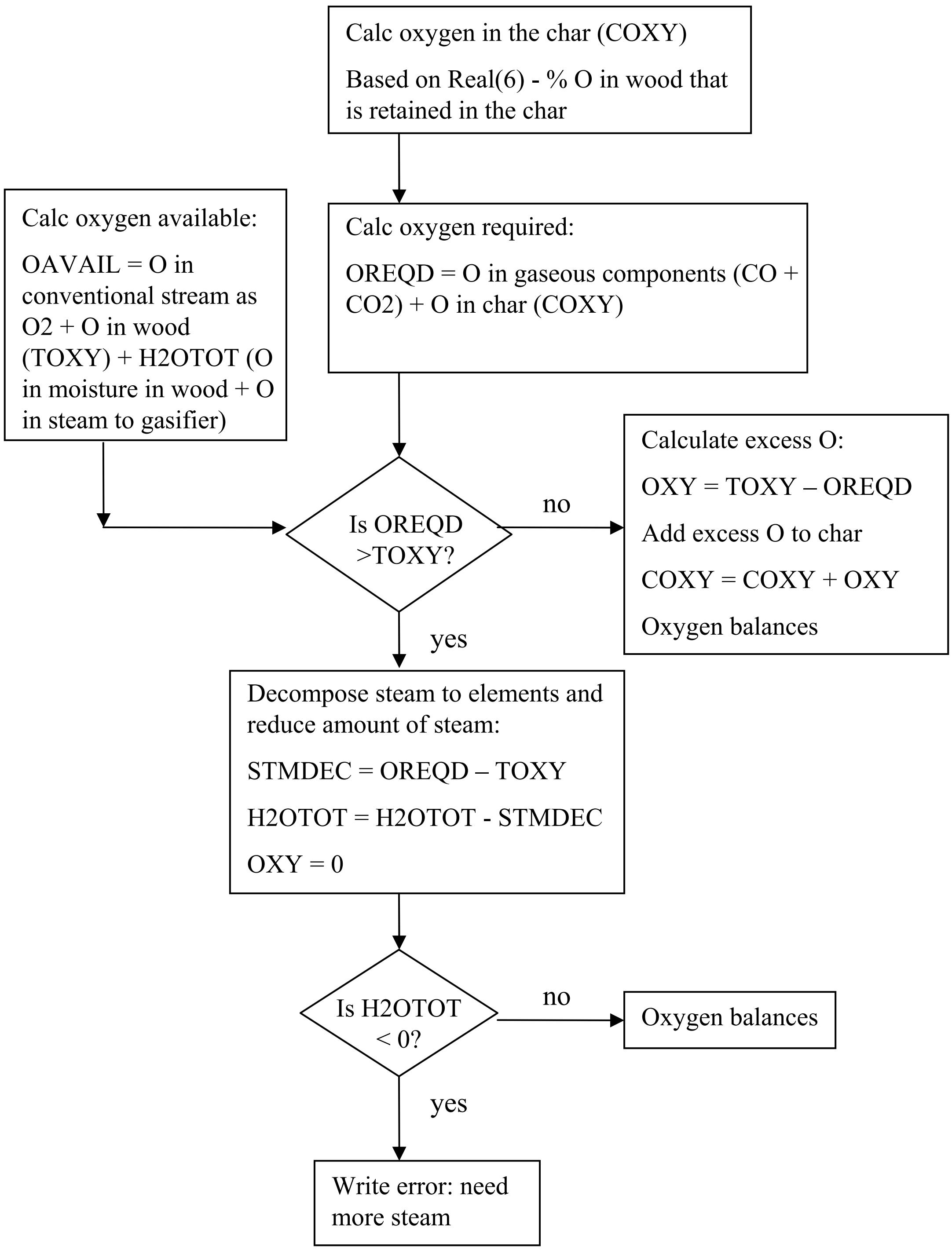
BCL model Fortran – performs balances in the following order:

1. Carbon
2. Oxygen
3. Sulfur
4. Hydrogen

BCL model - Carbon balance



BCL model - Oxygen balance



BCL model - Sulfur balance

Assume all gaseous sulfur is present as H2S AND all solid sulfur appears in the char

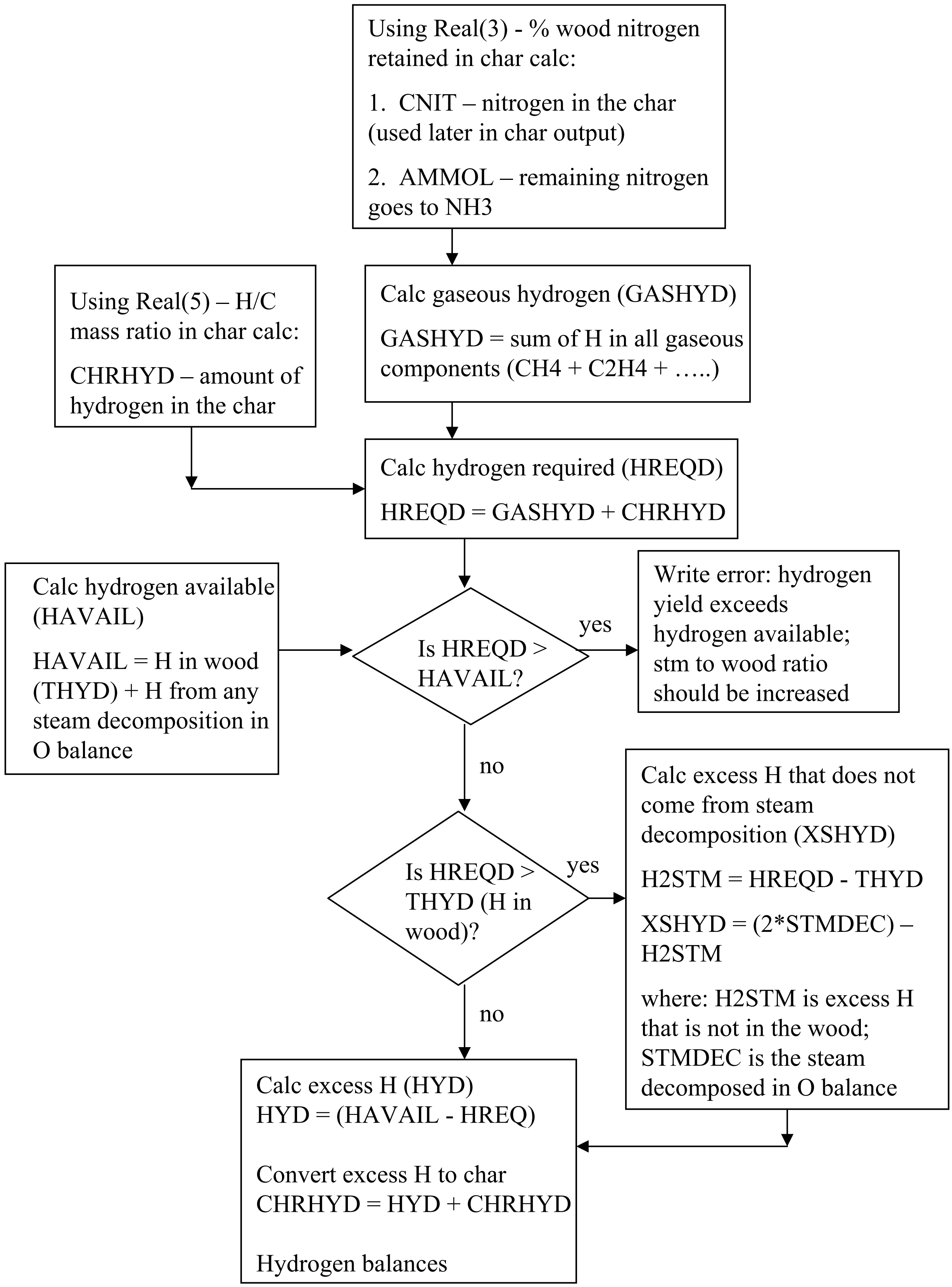
Calc H2S and sulfur in char (CSULF) using variable Real(4) - % wood sulfur retained in the char

TSULF = sulfur in the wood

H2SMOL = TSULF *(1-Real(4)/100)
CSULF = TSULF*(Real(4)/100)

Sulfur balances

BCL model - Hydrogen balance



Appendix G: Equipment Design Parameters and Cost References

EQUIPMENT_NUM	EQUIPMENT_NAME	EQUIPMENT_CATEGORY	EQUIPMENT_TYPE	EQUIPMENT_DESCRIPTION	COST_BASIS	MATERIAL_CONST
PFD-P700-A101-2						
C-101	Hopper Feeder	CONVEYOR	VIBRATING-FEEDER	Included in overall cost for feed handling & drying taken from several literature sources	LITERATURE	CS
C-102	Screener Feeder Conveyor	CONVEYOR	BELT	Included in overall cost for feed handling & drying taken from several literature sources	LITERATURE	CS
C-103	Radial Stacker Conveyor	CONVEYOR	BELT	Included in overall cost for feed handling & drying taken from several literature sources	LITERATURE	CS
C-104	Dryer Feed Screw Conveyor	CONVEYOR	SCREW	Included in overall cost for feed handling & drying taken from several literature sources	LITERATURE	CS
C-105	Gasifier Feed Screw Conveyor	CONVEYOR	SCREW	Included in overall cost for feed handling & drying taken from several literature sources	LITERATURE	316SS
H-101	Flue Gas Cooler / Steam Generator #3	HEATX	SHELL-TUBE	duty = 1.37 MMBtu/hr; LMTD = 1,220 F; U = 150 Btu/hr-ft^2-F; area = 7 ft^2; fixed TS	QUESTIMATE	CS/INCL
K-101	Flue Gas Blower	FAN	CENTRIFUGAL	Included in overall cost for feed handling & drying taken from several literature sources	LITERATURE	SS304
M-101	Hydraulic Truck Dump with Scale	SCALE	TRUCK-SCALE	Included in overall cost for feed handling & drying taken from several literature sources	LITERATURE	
M-102	Hammermill	SIZE-REDUCTION		Included in overall cost for feed handling & drying taken from several literature sources	LITERATURE	CS
M-103	Front End Loaders	VEHICLE	LOADER	Included in overall cost for feed handling & drying taken from several literature sources	LITERATURE	CS
M-104	Rotary Biomass Dryer	DRYER	ROTARY-DRUM	Included in overall cost for feed handling & drying taken from several literature sources	LITERATURE	CS
S-101	Magnetic Head Pulley	SEPARATOR	MAGNET	Included in overall cost for feed handling & drying taken from several literature sources	LITERATURE	CS
S-102	Screener	SEPARATOR	SCREEN	Included in overall cost for feed handling & drying taken from several literature sources	LITERATURE	CS
S-103	Dryer Air Cyclone	SEPARATOR	GAS CYCLONE	Included in overall cost for feed handling & drying taken from several literature sources	LITERATURE	CS
S-104	Dryer Air Baghouse Filter	SEPARATOR	FABRIC-FILTER	Included in overall cost for feed handling & drying taken from several literature sources	LITERATURE	
T-101	Dump Hopper	TANK	LIVE-BTM-BIN	Included in overall cost for feed handling & drying taken from several literature sources	LITERATURE	CS
T-102	Hammermill Surge Bin	TANK	LIVE-BTM-BIN	Included in overall cost for feed handling & drying taken from several literature sources	LITERATURE	CS
T-103	Dryer Feed Bin	TANK	LIVE-BTM-BIN	Included in overall cost for feed handling & drying taken from several literature sources	LITERATURE	CS
T-104	Dried Biomass Hopper	TANK	VERTICAL-VESSEL	Included in overall cost for feed handling & drying taken from several literature sources	LITERATURE	CS
PFD-P700-A201-2						
C-201	Sand/ash Conditioner/Conveyor	CONVEYOR	SCREW	Included in overall cost for gasification & gas clean up taken from several literature sources	LITERATURE	CS
H-201	Post-tar Reformer Cooler / Steam Generator #1	HEATX	SHELL-TUBE	duty = 47.9 MMBtu/hr; LMTD = 457; area = 698 sq ft; U = 150 Btu/hr-ft^2-F; fixed TS	ICARUS	CS/316S
H-202	Post-tar Reformer Cooler / BFW Preheater #2	HEATX	SHELL-TUBE	duty = 79.4 MMBTU; LMTD = 133 F; U = 150 Btu/hr-ft^2-F; area = 5,946 ft^2; fixed TS	QUESTIMATE	SS304CS/A214
K-201	Combustion Air Blower	FAN	CENTRIFUGAL	Included in overall cost for gasification & gas clean up taken from several literature sources	LITERATURE	CS
M-201	Sand/ash Cooler	MISCELLANEOUS	MISCELLANEOUS	Included in overall cost for gasification & gas clean up taken from several literature sources	LITERATURE	
R-201	Indirectly-heated Biomass Gasifier	REACTOR	VERTICAL-VESSEL	Included in overall cost for gasification & gas clean up taken from several literature sources	LITERATURE	CS w/refractory
R-202	Char Combustor	REACTOR	VERTICAL-VESSEL	Included in overall cost for gasification & gas clean up taken from several literature sources	LITERATURE	CS w/refractory
R-203	Tar Reformer	REACTOR	VERTICAL-VESSEL	Included in overall cost for gasification & gas clean up taken from several literature sources	LITERATURE	CS w/refractory
S-201	Primary Gasifier Cyclone	SEPARATOR	GAS CYCLONE	Included in overall cost for gasification & gas clean up taken from several literature sources	LITERATURE	CS w/refractory
S-202	Secondary Gasifier Cyclone	SEPARATOR	GAS CYCLONE	Included in overall cost for gasification & gas clean up taken from several literature sources	LITERATURE	CS w/refractory
S-203	Primary Combustor Cyclone	SEPARATOR	GAS CYCLONE	Included in overall cost for gasification & gas clean up taken from several literature sources	LITERATURE	CS w/refractory
S-204	Secondary Combustor Cyclone	SEPARATOR	GAS CYCLONE	Included in overall cost for gasification & gas clean up taken from several literature sources	LITERATURE	CS w/refractory
S-205	Electrostatic Precipitator	SEPARATOR	MISCELLANEOUS	Included in overall cost for gasification & gas clean up taken from several literature sources	LITERATURE	CS
T-201	Sand/ash Bin	TANK	FLAT-BTM-STORAGE	Included in overall cost for gasification & gas clean up taken from several literature sources	LITERATURE	CS
PFD-P700-A301-3						
H-301	Quench Water Recirculation Cooler	HEATX	SHELL-TUBE	Included in overall cost for gasification & gas clean up taken from several literature sources	LITERATURE	CS
H-302	Syngas Compressor Intercoolers	HEATX	AIR-COOLED EXCHANGER	Cost of intercoolers included in cost for syngas compressor, K-301	ICARUS	CS
H-303	Water-cooled Aftercooler	HEATX	SHELL-TUBE	duty = 2.9 MMBtu/hr; LMTD = 25F; U = 150 Btu/hr-ft^2-F; surface area = 794 ft^2; fixed TS	QUESTIMATE	SS304CS/A214
H-304	LO-CAT Preheater	HEATX	SHELL-TUBE	duty = 0.8 MMBtu/hr;LMTD = 87 F; U = 90 Btu/hr-ft^2-F; surface area = 98 ft^2; fixed TS	QUESTIMATE	A285C/CA443
H-305	LO-CAT Absorbent Solution Cooler	HEATX	SHELL-TUBE	Included in LO-CAT system cost	VENDOR	304SS
H-306	ZnO Bed Preheater	HEATX	SHELL-TUBE	duty = 47 MMBtu/hr duty; LMTD = 102 F; U = 90 Btu/hr-ft^2-F; area = 5,137 ft^2; fixed TS	QUESTIMATE	CS/A214
K-301	Syngas Compressor	COMPRESSOR	CENTRIFUGAL	gas flow rate = 70,000 CFM; 6 impellers; design outlet pressure = 465 psi; 30,000 HP; intercoolers, aftercooler, & K.O.s included	QUESTIMATE	A285C
K-302	LO-CAT Feed Air Blower	FAN	CENTRIFUGAL	Included in LO-CAT system cost	VENDOR	CS
K-303	Reformer Flue Gas Blower	FAN	CENTRIFUGAL	gas flow rate (actual) = 148,464 CFM; 327 HP	QUESTIMATE	CS
M-301	Syngas Quench Chamber	MISCELLANEOUS		Included in overall cost for gasification & gas clean up taken from several literature sources	LITERATURE	CS
M-302	Syngas Venturi Scrubber	MISCELLANEOUS		Included in overall cost for gasification & gas clean up taken from several literature sources	LITERATURE	CS
M-303	LO-CAT Venturi Precontactor	MISCELLANEOUS		Included in LO-CAT system cost	VENDOR	304SS
M-304	LO-CAT Liquid-filled Absorber	COLUMN	ABSORBER	Included in LO-CAT system cost	VENDOR	304SS
P-301	Sludge Pump	PUMP	CENTRIFUGAL	1.4 GPM; 0.053 brake HP; design pressure = 60 psia	QUESTIMATE	CS
P-302	Quench Water Recirculation Pump	PUMP	CENTRIFUGAL	Included in the cost of the gasification & gas clean up system	LITERATURE	CS
P-303	LO-CAT Absorbent Solution Circulating Pump	PUMP	CENTRIFUGAL	Included in LO-CAT system cost	VENDOR	304SS
R-301	LO-CAT Oxidizer Vessel	REACTOR	VERTICAL-VESSEL	Included in LO-CAT system cost	VENDOR	304SS
R-302	ZnO Sulfur Removal Beds	REACTOR	VERTICAL-VESSEL	6 ft diameter; 13 ft height; 427 cub ft volume; 490 psia design pressure; 757 F design temperature	QUESTIMATE	CS

EQUIPMENT_NUM	EQUIPMENT_NAME	EQUIPMENT_CATEGORY	EQUIPMENT_TYPE	EQUIPMENT_DESCRIPTION	COST_BASIS	MATERIAL_CONST
S-301	Pre-compressor Knock-out	SEPARATOR	KNOCK-OUT DRUM	18 ft diameter; 36 ft height; design pres = 40 psia; design temp = 197 F	QUESTIMATE	CS
S-302	Syngas Compressor Interstage Knock-outs	SEPARATOR	KNOCK-OUT DRUM	Cost of intercoolers K.O.s included in cost for syngas compressor, K-301	ICARUS	CS
S-303	Post-compressor Knock-out	SEPARATOR	KNOCK-OUT DRUM	7 ft. diameter; 14 ft height; design pres = 506 psia; design temp = 160 F	QUESTIMATE	CS
T-301	Sludge Settling Tank	SEPARATOR	CLARIFIER	3 ft diameter; 7 ft height; 431 gal volume;	QUESTIMATE	SS304
T-302	Quench Water Recirculation Tank	TANK	HORIZONTAL-VESSEL	Included in overall cost for gasification & gas clean up taken from several literature sources	LITERATURE	CS
PFD-P700-A401-3						
H-401	Reformer Feed Preheater	HEATX	SHELL-TUBE	duty = 47.6 MMBtu/hr; LMTD = 491 F; U = 90 Btu/hr-ft^2-F; area = 1,078 ft^2; fixed TS	ASSUMED	INCL/INCL
H-402	Reformed Syngas Cooler/Steam Generator #2	HEATX	SHELL-TUBE	duty = 155 MMBtu/hr; LMTD = 733 F; U = 150 Btu/hr-ft^2-F; area = 1,410 ft^2; fixed tube sheet	QUESTIMATE	CS/INCL
H-403	Reformed Syngas Cooler/Steam Superheater #1	HEATX	SHELL-TUBE	duty = 14 MMBtu/hr; U = 150 Btu/hr-ft^2-F; area = 983 ft^2; LMTD = 95 F	QUESTIMATE	SS316/316S
H-404	Reformer Flue Gas Cooler/Steam Superheater #2	HEATX	SHELL-TUBE	duty = 94 MMBtu/hr; LMTD = 217 F; U = 150 Btu/hr-ft^2-F; area = 2,900 ft^2; fixed TS	QUESTIMATE	CS/INCL
H-405	LT Shift Precooler/BFW Preheater #1	HEATX	SHELL-TUBE	duty = 54 MMBtu/hr; LMTD = 249 F; U = 100 Btu/hr-ft^2-F; area = 2,190 ft^2; fixed TS	QUESTIMATE	CS/A214
H-406	LT shift Precooler/Deaerator Water Preheater #1	HEATX	SHELL-TUBE	duty = 20 MMBtu/hr; LMTD = 244 F; U = 100 Btu/hr-ft^2-F; area = 823 ft^2; fixed TS	QUESTIMATE	CS/A214
H-407	PSA Precooler / Deaerator Water Preheater #2	HEATX	SHELL-TUBE	duty = 21 MMBtu/hr; LMTD = 251 F; U = 100 Btu/hr-ft^2-F; area = 858 ft^2; fixed TS	QUESTIMATE	CS/A214
H-408	PSA Air-cooled Precooler	HEATX	AIR-COOLED EXCHANGER	duty = 149 MMBtu/hr; LMTD = 103 F; U = 90 Btu/hr-ft^2-F; area = 16,117 ft^2; air cooler	QUESTIMATE	A214
H-409	PSA Water-cooled Precooler	HEATX	SHELL-TUBE	duty = 8 MMBtu/hr; LMTD = 25 F; U = 150 Btu/hr-ft^2-F; surface area = 2,274 ft^2	QUESTIMATE	A214
K-401	Reformer Combustion Air Blower	FAN	CENTRIFUGAL	gas flow rate (actual) = 70133 CFM; outlet pressure = 9.88 inches H2O	QUESTIMATE	CS
R-401	Steam Reformer	REACTOR	VERTICAL-VESSEL	heat duty = 159 MMBtu/hr	SRI	NI-CR Alloy
R-402	High Temperature Shift Reactor	REACTOR	VERTICAL-VESSEL	GHSV = 3,000/hr; H/D = 2; 12 ft diameter; 24 ft height; 400 psia op press; 807 F op temp	QUESTIMATE	316SS
R-403	Low Temperature Shift Reactor	REACTOR	VERTICAL-VESSEL	GHSV = 4,000; H/D = 2; 11 ft diameter; 22 ft height; 390 psia op press; 453 F op temp	QUESTIMATE	SS316
S-401	Pre-PSA Knock-out #1	SEPARATOR	KNOCK-OUT DRUM	H/D = 2; 12 ft diameter; 23 ft height; operating pressure = 380 psi; operating temperature = 334 F	QUESTIMATE	CS
S-402	Pre-PSA Knock-out #2	SEPARATOR	KNOCK-OUT DRUM	H/D = 2; 9 ft diameter; 17 ft height; operating pressure = 370 psi; operating temperature = 110 F	QUESTIMATE	CS
S-403	Pressure Swing Adsorption Unit	MISCELLANEOUS	PACKAGE	several beds; cost scaled from value of $0.168/SCFD of H2	LITERATURE	CS
PFD-P700-A501						
H-501A	Hydrogen Compressor Intercooler	HEATX	AIR-COOLED EXCHANGER	duty = 4 MMBtu/hr; LMTD = 61 F; U = 90 Btu/hr-ft^2-F; area = 740 ft^2; air cooler	QUESTIMATE	A214
H-501B	Hydrogen Compressor Air-cooled Aftercooler	HEATX	AIR-COOLED EXCHANGER	duty = 6 MMBtu/hr; LMTD = 77 F; U = 90 Btu/hr-ft^2-F; area = 864 sq ft.; air cooler	QUESTIMATE	A214
H-502	Hydrogen Compressor Water-cooler Aftercooler	HEATX	SHELL-TUBE	duty = 1.5 MMBtu/hr; LMTD = 25 F; U = 150 Btu/hr-ft^2-F; area = 396 ft^2; fixed TS	QUESTIMATE	A214
K-501	Hydrogen Compressor	COMPRESSOR	RECIPROCATING	gas flow rate = 2,028 actual CFM; outlet pressure = 1,020 psi	QUESTIMATE	A285C
S-501	Pre-hydrogen Compressor Knock-out	SEPARATOR	KNOCK-OUT DRUM	H/D = 2; 3 ft diam; 7 ft height; operating pressure = 360 psia; operating temperature = 109 F	QUESTIMATE	A-515
S-502	Hydrogen Compressor 1st Interstage Knock-out	SEPARATOR	KNOCK-OUT DRUM	included in the price of the hydrogen compressor (K-501)	QUESTIMATE	CS
S-503	Post-hydrogen Compressor Knock-out	SEPARATOR	KNOCK-OUT DRUM	H/D = 2; 3 ft diameter; 5 ft height;; operating pressure = 1,015 psi; operating temperature = 110 F	QUESTIMATE	A515
PFD-P700-A601-3						
H-601	Steam Turbine Condenser	HEATX	SHELL-TUBE	Included in the cost of the steam trubine/generator (M-602); condenser steam flow rate = 342,283 lb/hr	ADEN, ET. AL. 2002	
H-602	Blowdown Cooler / Deaerator Water Preheater	HEATX	SHELL-TUBE	duty = 3 MMBtu/hr; LMTD = 236 F; U = 600 Btu/hr-ft^2-F; area = 20 ft^2; pre-engineered U-tube	QUESTIMATE	A285C/CA443
H-603	Blowdown Water-cooled Cooler	HEATX	SHELL-TUBE	duty = 0.6 MMBtu/hr; LMTD = 47 F; U = 225 Btu/hr-ft^2-F; area = 60 ft^2; fixed TS	QUESTIMATE	A214
M-601	Hot Process Water Softener System	MISCELLANEOUS	PACKAGE	scaled cost to 700 gpm flow, 24" dia softener. Includes filters, chemical feeders, piping, valves	RICHARDSON	
M-602	Extraction Steam Turbine/Generator	GENERATOR	STEAM-TURBINE	25.6 MW generated; 34,308 HP	VENDOR	
P-601	Collection Pump	PUMP	CENTRIFUGAL	513 GPM; 4 brake HP; outlet pressure = 25 psia	QUESTIMATE	CS
P-602	Condensate Pump	PUMP	CENTRIFUGAL	190 GPM; 4 brake HP; outlet pressure = 25 psia	QUESTIMATE	SS304
P-603	Deaerator Feed Pump	PUMP	CENTRIFUGAL	702 GPM; 14 brake HP; outlet pressure = 40 psia	QUESTIMATE	CS
P-604	Boiler Feed Water Pump	PUMP	CENTRIFUGAL	730 GPM; 759 brake HP; outlet pressure = 1,345 psia	QUESTIMATE	CS
S-601	Blowdown Flash Drum	TANK	HORIZONTAL-VESSEL	H/D = 2; residence time = 5 min; 2 ft diameter; 4 ft height; op press = 1,280 psi; op temp = 575 F	QUESTIMATE	CS
T-601	Condensate Collection Tank	TANK	HORIZONTAL-VESSEL	residence time = 10 minutes; H/D = 2; 8 ft diameter; 17 ft height	QUESTIMATE	CS
T-602	Condensate Surge Drum	TANK	HORIZONTAL-VESSEL	residence time = 10 minutes; H/D = 2; 9 ft diameter; 17 ft height	QUESTIMATE	CS
T-603	Deaerator	TANK	HORIZONTAL-VESSEL	liquid flow rate = 348,266 lb/hr; 150 psig design pressure; 10 min residence time	VENDOR	CS;SS316
T-604	Steam Drum	TANK	HORIZONTAL-VESSEL	424 gal, 4.5' x 4'dia, 15 psig	ICARUS	CS
PFD-P700-A701-2						
K-701	Plant Air Compressor	COMPRESSOR	RECIPROCATING	450 cfm, 125 psig outlet	ICARUS	CS
M-603	Startup Boiler	MISCELLANEOUS	PACKAGE	Assume need steam requirement equal to 1/2 of steam requirement for gasifier at full rate steam rate = 36,560 lb/hr	QUESTIMATE	CS
M-701	Cooling Tower System	COOLING-TOWER	INDUCED-DRAFT	approx 16,500 gpm, 140 MMBtu/hr	DELTA-T98	FIBERGLASS
M-702	Hydraulic Truck Dump with Scale	SCALE	TRUCK-SCALE	Hydraulic Truck Dumper with Scale	VENDOR	CS
M-703	Flue Gas Stack	MISCELLANEOUS	MISCELLANEOUS	42 inch diameter; 250 deg F	QUESTIMATE	A515
P-701	Cooling Water Pump	PUMP	CENTRIFUGAL	16,188 GPM; 659 brake HP; outlet pressure 75 psi	QUESTIMATE	CS
P-702	Firewater Pump	PUMP	CENTRIFUGAL	2,500 gpm, 50 ft head	ICARUS	CS
P-703	Diesel Pump	PUMP	CENTRIFUGAL	30 gpm, 150 ft head	ICARUS	CS

EQUIPMENT_NUM	EQUIPMENT_NAME	EQUIPMENT_CATEGORY	EQUIPMENT_TYPE	EQUIPMENT_DESCRIPTION	COST_BASIS	MATERIAL_CONST
P-704	Ammonia Pump	PUMP	CENTRIFUGAL	8.5 gpm, 22 ft head	ICARUS	CS
P-705	Hydrazine Pump	PUMP	CENTRIFUGAL	5 gpm, 75 ft head	DELTA-T98	CS
S-701	Instrument Air Dryer	DRYER	PACKAGE	400 SCFM Air Dryer, -40 F Dewpoint	RICHARDSON	CS
T-701	Plant Air Receiver	TANK	HORIZONTAL-VESSEL	900 gal., 200 psig	ICARUS	CS
T-702	Firewater Storage Tank	TANK	FLAT-BTM-STORAGE	600,000 gal, 4 hr res time, 51' dia x 40' high, atmospheric	ICARUS	A285C
T-703	Diesel Storage Tank	TANK	FLAT-BTM-STORAGE	10,667 gal, 120 hr res time, 90% wv, 10' dia x 18.2' high, atmospheric	ICARUS	A285C
T-704	Ammonia Storage Tank	TANK	HORIZONTAL-STORAGE	Included in the cost of the feed handling step.	ICARUS	A515
T-705	Olivine Lock Hopper	TANK	VERTICAL-VESSEL	Included in the cost of the feed handling step.	DELTA-T98	CS
T-706	MgO Lock Hopper	TANK	VERTICAL-VESSEL	20' x 20' Bin, Tapering to 3' x 3' at Bottom. Capacity 6,345 cf, two truck loads.	DELTA-T98	CS
T-707	Hydrazine Storage Tank	TANK	VERTICAL-VESSEL	260 gal, 4.9' x 3'dia., 10psig	ICARUS	SS316

Appendix H: Current Design Summary of Individual Equipment Costs

Equipment Number	Number Required	Number Spares	Equipment Name	Scaling Stream	Scaling Stream Flow (lb/hr or btu/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2002$
C-101	4		Hopper Feeder	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
C-102	2		Screener Feeder Conveyor	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
C-103	2		Radial Stacker Conveyor	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
C-104	2		Dryer Feed Screw Conveyor	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
C-105	2		Gasifier Feed Screw Conveyor	104	208,771	208,771	1.00	Included in feed handling & drying cost (M-104)							
H-101	1		Flue Gas Cooler / Steam Generator #3	PINCH	1,369,986	1,369,986	1.00	$26,143	2002	$26,143	0.6	$26,143	2.47	$64,573	$64,573
K-101	2		Flue Gas Blower	112	639,530	639,530	1.00	Included in feed handling & drying cost (M-104)							
M-101	4		Hydraulic Truck Dump with Scale	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
M-102	2		Hammermill	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
M-103	3		Front End Loaders	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
M-104	2		Rotary Biomass Dryer	101	367,437	367,437	1.00	$3,813,728	2002	$7,627,455	0.75	$7,627,450	2.47	$18,839,801	$18,839,801
S-101	2		Magnetic Head Pulley	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
S-102	2		Screener	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
S-103	2		Dryer Air Cyclone	111	639,530	639,530	1.00	Included in feed handling & drying cost (M-104)							
S-104	2		Dryer Air Baghouse Filter	103	208,771	208,771	1.00	Included in feed handling & drying cost (M-104)							
T-101	4		Dump Hopper	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
T-102	1		Hammermill Surge Bin	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
T-103	2		Dryer Feed Bin	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
T-104	2		Dried Biomass Hopper	104	208,771	208,771	1.00	Included in feed handling & drying cost (M-104)							
A100									**Subtotal**	**$7,653,598**		**$7,653,593**		**$18,904,374**	**$18,904,374**
C-201	1		Sand/ash Conditioner/Conveyor	219	7,380	47,912,711	1.00	Included in gasification & clean up cost (R-201)							
H-201	1		Post-tar Reformer Cooler / Steam Generator #1	PINCH	47,912,711	79,370,881	1.00	$69,089	2002	$69,089	0.65	$69,089	2.47	$170,650	$170,650
H-202	1		Post-tar Reformer Cooler / BFW Preheater #2	PINCH	79,370,881	79,370,881	1.00	$99,389	2002	$99,389	0.6	$99,389	2.47	$245,491	$245,491
K-201	2		Combustion Air Blower	208	442,163	442,163	1.00	Included in gasification & clean up cost (R-201)							
M-201	2		Sand/ash Cooler	217	6,642	6,642	1.00	Included in gasification & clean up cost (R-201)							
R-201	2		Indirectly-heated Biomass Gasifier	201	5,228,880	5,228,878	1.00	$3,318,302	2002	$6,636,603	0.65	$6,636,601	2.47	$16,392,405	$16,392,405
R-202	2		Char Combustor	210	5,434,490	5,434,493	1.00	Included in gasification & clean up cost (R-201)							
R-203	1		Tar Reformer	225	241,995	241,995	1.00	Included in gasification & clean up cost (R-201)							
S-201	2		Primary Gasifier Cyclone	202	5,228,880	5,228,878	1.00	Included in gasification & clean up cost (R-201)							
S-202	2		Secondary Gasifier Cyclone	222	246,484	246,483	1.00	Included in gasification & clean up cost (R-201)							
S-203	2		Primary Combustor Cyclone	210	5,434,490	5,434,493	1.00	Included in gasification & clean up cost (R-201)							
S-204	2		Secondary Combustor Cyclone	212	487,506	487,506	1.00	Included in gasification & clean up cost (R-201)							
S-205	2		Electrostatic Precipitator	213	480,870	480,870	1.00	Included in gasification & clean up cost (R-201)							
T-201	1		Sand/ash Bin	217	6,642	6,642	1.00	Included in gasification & clean up cost (R-201)							
A200									**Subtotal**	**$6,805,081**		**$6,805,079**		**$16,808,546**	**$16,808,546**

Equipment Number	Number Required	Number Spares	Equipment Name	Scaling Stream	Scaling Stream Flow (lb/hr or btu/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2002$
H-301	1		Quench Water Recirculation Cooler	301	241,995	241,995	1.00	Included in gasification & clean up cost (R-201)							
H-302	5		Syngas Compressor Intercoolers	301	241,995	241,995	1.00	Included in the syngas compressor cost (K-301)							
H-303	1		Water-cooled Aftercooler	QCH303CT	2,938,799	2,940,165	1.00	$20,889	2002	$20,889	0.44	$20,893	2.47	$51,606	$51,606
H-304	1		LO-CAT Preheater	PINCH	770,434	770,434	1.00	$4,743	2002	$4,743	0.6	$4,743	2.47	$11,715	$11,715
H-305	1		LO-CAT Absorbent Solution Cooler	320	179,394	179,394	1.00	Included in LO-CAT oxidizer vessel cost (R-301)							
H-306	1		ZnO Bed Preheater	PINCH	47,209,942	47,209,942	1.00	$71,389	2002	$71,389	0.44	$71,389	2.47	$176,331	$176,331
K-301	1		Syngas Compressor	315	220,009	220,009	1.00	$4,817,834	2002	$4,817,834	0.8	$4,817,834	2.47	$11,900,051	$11,900,051
K-302	1		LO-CAT Feed Air Blower	322	359	359	1.00	Included in LO-CAT oxidizer vessel cost (R-301)							
K-303	1		Reformer Flue Gas Blower	434	534,677	534,677	1.00	$54,250	2002	$54,250	0.59	$54,250	2.47	$133,997	$133,997
M-301	1		Syngas Quench Chamber	301	241,496	241,995	1.00	Included in gasification & clean up cost (R-201)							
M-302	1		Syngas Venturi Scrubber	301	241,496	241,995	1.00	Included in gasification & clean up cost (R-201)							
M-303	1		LO-CAT Venturi Precontactor	323	517	517	1.00	Included in LO-CAT oxidizer vessel cost (R-301)							
M-304	1		LO-CAT Liquid-filled Absorber	320	179,394	179,394	1.00	Included in LO-CAT oxidizer vessel cost (R-301)							
P-301	1	1	Sludge Pump	336	997	997	1.00	$3,911	2002	$7,822	0.33	$7,823	2.47	$19,323	$19,323
P-302	1	1	Quench Water Recirculation Pump	307	1,272,120	1,272,123	1.00	Included in gasification & clean up cost (R-201)							
P-303	1	1	LO-CAT Absorbent Solution Circulating Pump	301	241,496	241,995	1.00	Included in LO-CAT oxidizer vessel cost (R-301)							
R-301	1		LO-CAT Oxidizer Vessel	323	517	517	1.00	$1,000,000	2002	$1,000,000	0.65	$999,653	2.47	$2,469,142	$2,469,142
R-302	2		ZnO Sulfur Removal Beds	327	179,237	179,237	1.00	$37,003	2002	$74,006	0.56	$74,006	2.47	$182,795	$182,795
S-301	1		Pre-compressor Knock-out	315	220,009	220,009	1.00	$157,277	2002	$157,277	0.6	$157,277	2.47	$388,474	$388,474
S-302	4		Syngas Compressor Interstage Knock-outs	315	220,009	220,009	1.00	Included in the syngas compressor cost (K-301)							
S-303	1		Post-compressor Knock-out	319	179,394	179,394	1.00	$40,244	2002	$40,244	0.6	$40,244	2.47	$99,403	$99,403
T-301	1		Sludge Settling Tank	302	21,718	21,718	1.00	$11,677	2002	$11,677	0.6	$11,677	2.47	$28,842	$28,842
T-302	1		Quench Water Recirculation Tank	301	241,496	241,995	1.00	Included in gasification & clean up cost (R-201)							
A300									**Subtotal**	**$6,260,131**		**$6,259,790**		**$15,461,680**	**$15,461,680**
H-401	1		Reformer Feed Preheater	QH401	47,628,665	47,628,665	1.00	$277,489	2002	$277,489	0.7	$277,489	2.47	$685,398	$685,398
H-402	1		Reformed Syngas Cooler/Steam Generator #2	PINCH	155,010,823	155,010,823	1.00	$347,989	2002	$347,989	0.6	$347,989	2.47	$859,533	$859,533
H-403	1		Reformed Syngas Cooler/Steam Superheater #1	PINCH	13,974,577	13,974,577	1.00	$92,889	2002	$92,889	0.6	$92,889	2.47	$229,436	$229,436
H-404	1		Reformer Flue Gas Cooler/Steam Superheater #2	PINCH	94,212,763	94,212,763	1.00	$196,589	2002	$196,589	0.6	$196,589	2.47	$485,575	$485,575
H-405	1		LT Shift Precooler/BFW Preheater #1	PINCH	54,476,359	54,476,359	1.00	$56,089	2002	$56,089	0.6	$56,089	2.47	$138,540	$138,540
H-406	1		LT shift Precooler/Deaerator Water Preheater #1	PINCH	20,095,131	20,095,131	1.00	$20,989	2002	$20,989	0.6	$20,989	2.47	$51,843	$51,843
H-407	1		PSA Precooler / Deaerator Water Preheater #2	PINCH	21,034,730	21,034,730	1.00	$21,089	2002	$21,089	0.6	$21,089	2.47	$52,090	$52,090
H-408	1		PSA Air-cooled Precooler	QAH408	149,281,592	149,281,592	1.00	$388,064	2002	$388,064	0.6	$388,064	2.47	$958,518	$958,518
H-409	1		PSA Water-cooled Precooler	QCH409CT	8,414,338	8,414,338	1.00	$35,689	2002	$35,689	0.44	$35,689	2.47	$88,152	$88,152
K-401	1		Reformer Combustion Air Blower	430	304,578	304,578	1.00	$35,020	2002	$35,020	0.59	$35,020	2.47	$86,499	$86,499
R-401	1		Steam Reformer	QR401	158,705,747	158,705,747	1.00	$4,965,833	2002	$4,965,833	0.7	$4,965,833	2.47	$12,265,608	$12,265,608
R-402	1		High Temperature Shift Reactor	404	354,424	354,424	1.00	$465,907	2002	$465,907	0.56	$465,907	2.47	$1,150,791	$1,150,791
R-403	1		Low Temperature Shift Reactor	407	354,424	354,424	1.00	$323,464	2002	$323,464	0.56	$323,464	2.47	$798,957	$798,957
S-401	1		Pre-PSA Knock-out #1	413	354,424	354,424	1.00	$129,979	2002	$129,979	0.6	$129,979	2.47	$321,048	$321,048
S-402	1		Pre-PSA Knock-out #2	419	242,691	242,691	1.00	$55,291	2002	$55,291	0.6	$55,291	2.47	$136,569	$136,569
S-403	1		Pressure Swing Adsorption Unit	424	14,260	14,260	1.00	$4,855,471	2002	$4,855,471	0.6	$4,855,482	2.47	$11,993,041	$11,993,041
A400									**Subtotal**	**$12,267,841**		**$12,267,853**		**$30,301,596**	**$30,301,596**

Equipment Number	Number Required	Number Spares	Equipment Name	Scaling Stream	Scaling Stream Flow (lb/hr or btu/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2002$
H-501A	1		Hydrogen Compressor Intercooler	QAK501A	4,042,813	4,042,813	1.00	$53,601	2002	$53,601	0.6	$53,601	2.47	$132,394	$132,394
H-501B	1		Hydrogen Compressor Air-cooled Aftercooler	QAK501B	5,984,714	5,984,714	1.00	$56,901	2002	$56,901	0.6	$56,901	2.47	$140,545	$140,545
H-502	1		Hydrogen Compressor Water-cooler Aftercooler	QCH502CT	1,465,277	1,465,278	1.00	$18,909	2002	$18,909	0.44	$18,909	2.47	$46,705	$46,705
K-501	1		Hydrogen Compressor	501	14,260	14,260	1.00	$914,235	2002	$914,235	0.8	$914,238	2.47	$2,258,167	$2,258,167
S-501	1		Pre-hydrogen Compressor Knock-out	501	14,260	14,260	1.00	$13,377	2002	$13,377	0.6	$13,377	2.47	$33,041	$33,041
S-502	1		Hydrogen Compressor 1st Interstage Knock-out	501	14,260	14,260	1.00	Included in the hydrogen compressor cost (K-501)							
S-503	1		Post-hydrogen Compressor Knock-out	505	14,260	14,260	1.00	$13,977	2002	$13,977	0.6	$13,977	2.47	$34,523	$34,523
A500									**Subtotal**	**$1,071,000**		**$1,071,003**		**$2,645,377**	**$2,645,377**
H-601	1		Steam Turbine Condenser	614	93,974	93,974	1.00	Included in the extraction steam trubine/generator cost (M-602)							
H-602	1		Blowdown Cooler / Deaerator Water Preheater	PINCH	2,877,029	2,877,029	1.00	$3,043	2002	$3,043	0.6	$3,043	2.47	$7,516	$7,516
H-603	1		Blowdown Water-cooled Cooler	QCH603CT	626,343	626,343	1.00	$16,143	2002	$16,143	0.44	$16,143	2.47	$39,873	$39,873
M-601	1		Hot Process Water Softener System	631	349,266	349,266	1.00	$1,031,023	1999	$1,031,023	0.82	$1,031,023	2.47	$2,546,627	$2,579,225
M-602	1		Extraction Steam Turbine/Generator	607	342,283	342,283	1.00	$4,045,870	2002	$4,045,870	0.71	$4,045,870	2.47	$9,993,300	$9,993,300
M-603	1		Startup Boiler	200	36,560	36,560	1.00	$198,351	2002	$198,351	0.6	$198,351	2.47	$489,927	$489,927
P-601	1	1	Collection Pump	625	255,292	255,292	1.00	$7,015	2002	$14,030	0.33	$14,030	2.47	$34,654	$34,654
P-602	1	1	Condensate Pump	616	93,974	93,974	1.00	$5,437	2002	$10,874	0.33	$10,874	2.47	$26,859	$26,859
P-603	1	1	Deaerator Feed Pump	628	349,266	349,266	1.00	$8,679	2002	$17,358	0.33	$17,358	2.47	$42,874	$42,874
P-604	1	1	Boiler Feed Water Pump	639	349,268	349,268	1.00	$95,660	2002	$191,320	0.33	$191,320	2.47	$472,561	$472,561
T-601	1		Condensate Collection Tank	627	349,266	349,266	1.00	$24,493	2002	$24,493	0.6	$24,493	2.47	$60,498	$60,498
T-602	1		Condensate Surge Drum	638	349,268	349,268	1.00	$28,572	2002	$28,572	0.6	$28,572	2.47	$70,573	$70,573
T-603	1		Deaerator	633	349,266	349,266	1.00	$130,721	2002	$130,721	0.72	$130,721	2.47	$322,881	$322,881
T-604	1		Steam Drum	644	349,268	349,268	1.00	$9,200	1997	$9,200	0.72	$9,200	2.47	$22,724	$23,259
S-601	1		Blowdown Flash Drum	604	6,985	6,985	1.00	$14,977	2002	$14,977	0.6	$14,977	2.47	$36,994	$36,994
A600									**Subtotal**	**$5,735,975**		**$5,735,976**		**$14,167,860**	**$14,200,994**
K-701	2	1	Plant Air Compressor	101	367,437	367,437	1.00	$32,376	2002	$97,129	0.34	$97,129	2.47	$239,908	$239,908
M-701	1		Cooling Tower System	QCTOTAL	139,850,763	139,850,763	1.00	$267,316	2002	$267,316	0.78	$267,316	2.47	$660,271	$660,271
M-702	1		Hydraulic Truck Dump with Scale	101	367,437	367,437	1.00	$80,000	1998	$80,000	0.6	$80,000	2.47	$197,600	$200,695
M-703	1		Flue Gas Stack	112	1,174,206	639,530	1.00	$51,581	2002	$51,581	1	$51,581	2.47	$127,405	$127,405
				434		534,677	The stack flow is the sum of two flow streams.								
P-701	1	1	Cooling Water Pump	715	6,088,320	6,113,668	1.00	$158,540	2002	$317,080	0.33	$317,515	2.47	$784,262	$784,262
P-702	1	1	Firewater Pump	101	367,437	367,437	1.00	$18,400	1997	$36,800	0.79	$36,800	2.47	$90,896	$93,036
P-703	1	1	Diesel Pump	101	367,437	367,437	1.00	$6,100	1997	$12,200	0.79	$12,200	2.47	$30,134	$30,843
P-704	1	1	Ammonia Pump	101	367,437	367,437	1.00	$5,000	1997	$10,000	0.79	$10,000	2.47	$24,700	$25,282
P-705	1		Hydrazine Pump	101	367,437	367,437	1.00	$5,500	1997	$5,500	0.79	$5,500	2.47	$13,585	$13,905
S-701	1	1	Instrument Air Dryer	101	367,437	367,437	1.00	$8,349	2002	$16,698	0.6	$16,698	2.47	$41,244	$41,244
T-701	1		Plant Air Receiver	101	367,437	367,437	1.00	$7,003	2002	$7,003	0.72	$7,003	2.47	$17,297	$17,297
T-702	1		Firewater Storage Tank	101	367,437	367,437	1.00	$166,100	1997	$166,100	0.51	$166,100	2.47	$410,267	$419,926
T-703	1		Diesel Storage Tank	101	367,437	367,437	1.00	$14,400	1997	$14,400	0.51	$14,400	2.47	$35,568	$36,405
T-704	1		Ammonia Storage Tank	101	367,437	367,437	1.00	$287,300	1997	$287,300	0.72	$287,300	2.47	$709,631	$726,339
T-705	1		Olivine Lock Hopper	101	367,437	367,437	1.00	Included in gasification & clean up cost (R-201)							
T-706	1		MgO Lock Hopper	101	367,437	367,437	1.00	Included in gasification & clean up cost (R-201)							
T-707	1		Hydrazine Storage Tank	101	367,437	367,437	1.00	$12,400	1997	$12,400	0.93	$12,400	2.47	$30,628	$31,349
A700									**Subtotal**	**$1,381,507**		**$1,381,942**		**$3,413,396**	**$3,416,818**

Appendix I: Goal Design Summary of Individual Equipment Costs

Equipment Number	Number Required	Number Spares	Equipment Name	Scaling Stream	Scaling Stream Flow (lb/hr or btu/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2002$
C-101	4		Hopper Feeder	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
C-102	2		Screener Feeder Conveyor	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
C-103	2		Radial Stacker Conveyor	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
C-104	2		Dryer Feed Screw Conveyor	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
C-105	2		Gasifier Feed Screw Conveyor	104	208,771	208,771	1.00	Included in feed handling & drying cost (M-104)							
H-101	1		Flue Gas Cooler / Steam Generator #3	PINCH	1,369,986	1,369,094	1.00	$26,143	2002	$26,143	0.6	$26,133	2.47	$64,548	$64,548
K-101	2		Flue Gas Blower	112	639,530	639,526	1.00	Included in feed handling & drying cost (M-104)							
M-101	4		Hydraulic Truck Dump with Scale	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
M-102	2		Hammermill	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
M-103	3		Front End Loaders	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
M-104	2		Rotary Biomass Dryer	101	367,437	367,437	1.00	$3,813,728	2002	$7,627,455	0.75	$7,627,450	2.47	$18,839,801	$18,839,801
S-101	2		Magnetic Head Pulley	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
S-102	2		Screener	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
S-103	2		Dryer Air Cyclone	111	639,530	639,526	1.00	Included in feed handling & drying cost (M-104)							
S-104	2		Dryer Air Baghouse Filter	103	208,771	208,771	1.00	Included in feed handling & drying cost (M-104)							
T-101	4		Dump Hopper	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
T-102	1		Hammermill Surge Bin	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
T-103	2		Dryer Feed Bin	101	367,437	367,437	1.00	Included in feed handling & drying cost (M-104)							
T-104	2		Dried Biomass Hopper	104	208,771	208,771	1.00	Included in feed handling & drying cost (M-104)							
A100									**Subtotal**	**$7,653,598**		**$7,653,583**		**$18,904,349**	**$18,904,349**
C-201	1		Sand/ash Conditioner/Conveyor	219	7,380	7,380	1.00	Included in gasification & clean up cost (R-201)							
H-201	1		Post-tar Reformer Cooler / Steam Generator #1	PINCH	47,912,711	116,732,109	1.00	$69,089	2002	$69,089	0.65	$69,060	2.47	$170,578	$170,578
H-202A	1		Post-tar reformer cooler/Deaerator water preheater #1	PINCH	8,807,704	8,807,704	1.00	$21,589	2002	$21,589	0.6	$21,589	2.47	$53,325	$53,325
H-202B	1		Post-tar cracker cooler/BFW preheater #2	PINCH	48,632,640	48,632,640	1.00	$429,889	2002	$429,889	0.6	$429,889	2.47	$1,061,826	$1,061,826
K-201	2		Combustion Air Blower	208	442,163	442,157	1.00	Included in gasification & clean up cost (R-201)							
K-202	1		Regenerator Combustion Air Blower	430	304,578	302,225	0.99	$35,020	2002	$35,020	0.59	$34,860	2.47	$86,104	$86,104
M-201	2		Sand/ash Cooler	217	6,642	6,642	1.00	Included in gasification & clean up cost (R-201)							
R-201	2		Indirectly-heated Biomass Gasifier	201	5,228,880	5,228,878	1.00	$3,318,302	2002	$6,636,603	0.65	$6,636,601	2.47	$16,392,405	$16,392,405
R-202	2		Char Combustor	210	5,434,490	5,434,489	1.00	Included in gasification & clean up cost (R-201)							
R-203	1		Tar Reformer	225	241,995	241,993	1.00	Included in gasification & clean up cost (R-201)							
R-204	1		Tar Reformer Catalyst Regenerator	428	234,433	234,433	1.00	$2,429,379	2002	$2,429,379	0.65	$2,429,380	2.47	$6,000,570	$6,000,570
S-201	2		Primary Gasifier Cyclone	202	5,228,880	5,228,878	1.00	Included in gasification & clean up cost (R-201)							
S-202	2		Secondary Gasifier Cyclone	222	246,484	246,481	1.00	Included in gasification & clean up cost (R-201)							
S-203	2		Primary Combustor Cyclone	210	5,434,490	5,434,489	1.00	Included in gasification & clean up cost (R-201)							
S-204	2		Secondary Combustor Cyclone	212	487,506	487,502	1.00	Included in gasification & clean up cost (R-201)							
S-205	2		Electrostatic Precipitator	213	480,870	480,866	1.00	Included in gasification & clean up cost (R-201)							
S-206	1		Tar Reformer Cyclone	225	241,995	241,993	1.00	Included in tar reformer catalyst regenerator cost							
S-207	1		Catalyst Regenerator Cyclone	428	234,433	234,433	1.00	Included in tar reformer catalyst regenerator cost							
T-201	1		Sand/ash Bin	217	6,642	6,642	1.00	Included in gasification & clean up cost (R-201)							
A200									**Subtotal**	**$9,621,569**		**$9,621,380**		**$23,764,807**	**$23,764,807**

Equipment Number	Number Required	Number Spares	Equipment Name	Scaling Stream	Scaling Stream Flow (lb/hr or btu/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2002$
H-301	1		Quench Water Recirculation Cooler	301	241,995	241,993	1.00	Included in gasification & clean up cost (R-201)							
H-302	5		Syngas Compressor Intercoolers	301	241,995	241,993	1.00	Included in the syngas compressor cost (K-301)							
H-303	1		Water-cooled Aftercooler	QCH303CT	2,938,799	3,388,287	1.15	$20,889	2002	$20,889	0.44	$22,239	2.47	$54,930	$54,930
H-304	1		LO-CAT Preheater	PINCH	770,434	858,449	1.11	$4,743	2002	$4,743	0.6	$5,061	2.47	$12,501	$12,501
H-305	1		LO-CAT Absorbent Solution Cooler	320	179,394	184,842	1.03	Included in LO-CAT oxidizer vessel cost (R-301)							
H-306	1		ZnO Bed Preheater	PINCH	47,209,942	51,594,124	1.09	$71,389	2002	$71,389	0.44	$74,234	2.47	$183,357	$183,357
K-301	1		Syngas Compressor	315	220,009	233,488	1.06	$4,817,834	2002	$4,817,834	0.8	$5,052,554	2.47	$12,479,808	$12,479,808
K-302	1		LO-CAT Feed Air Blower	322	359	358	1.00	Included in LO-CAT oxidizer vessel cost (R-301)							
K-303	1		Regenerator Flue Gas Blower	434	534,677	536,658	1.00	$54,250	2002	$54,250	0.59	$54,368	2.47	$134,290	$134,290
M-301	1		Syngas Quench Chamber	301	241,496	241,993	1.00	Included in gasification & clean up cost (R-201)							
M-302	1		Syngas Venturi Scrubber	301	241,496	241,993	1.00	Included in gasification & clean up cost (R-201)							
M-303	1		LO-CAT Venturi Precontactor	323	517	515	1.00	Included in LO-CAT oxidizer vessel cost (R-301)							
M-304	1		LO-CAT Liquid-filled Absorber	320	179,394	184,842	1.03	Included in LO-CAT oxidizer vessel cost (R-301)							
P-301	1	1	Sludge Pump	336	997	997	1.00	$3,911	2002	$7,822	0.33	$7,823	2.47	$19,323	$19,323
P-302	1	1	Quench Water Recirculation Pump	307	1,272,120	316,851	0.25	Included in gasification & clean up cost (R-201)							
P-303	1	1	LO-CAT Absorbent Solution Circulating Pump	301	241,496	241,993	1.00	Included in LO-CAT oxidizer vessel cost (R-301)							
R-301	1		LO-CAT Oxidizer Vessel	323	517	515	1.00	$1,000,000	2002	$1,000,000	0.65	$997,471	2.47	$2,463,754	$2,463,754
R-302	2		ZnO Sulfur Removal Beds	327	179,237	184,685	1.03	$37,003	2002	$74,006	0.56	$75,257	2.47	$185,885	$185,885
S-301	1		Pre-compressor Knock-out	315	220,009	233,488	1.06	$157,277	2002	$157,277	0.6	$162,989	2.47	$402,584	$402,584
S-302	4		Syngas Compressor Interstage Knock-outs	315	220,009	233,488	1.06	Included in the syngas compressor cost (K-301)							
S-303	1		Post-compressor Knock-out	319	179,394	184,842	1.03	$40,244	2002	$40,244	0.6	$40,973	2.47	$101,203	$101,203
T-301	1		Sludge Settling Tank	302	21,718	8,171	0.38	$11,677	2002	$11,677	0.6	$6,495	2.47	$16,043	$16,043
T-302	1		Quench Water Recirculation Tank	301	241,496	241,993	1.00	Included in gasification & clean up cost (R-201)							
A300									**Subtotal**	**$6,260,131**		**$6,499,465**		**$16,053,679**	**$16,053,679**
H-404A	1		Tar reformer flue gas cooler/steam generator #2	PINCH	86,510,197	86,510,197	1.00	$144,489	2002	$144,489	0.6	$144,489	2.47	$356,888	$356,888
H-404B	1		Tar reformer flue gas cooler/ steam superheater #1	PINCH	108,355,680	108,355,680	1.00	$90,889	2002	$90,889	0.6	$90,889	2.47	$224,496	$224,496
H-405	1		LT Shift Precooler/BFW Preheater #1	PINCH	54,476,359	85,423,190	1.57	$56,089	2002	$56,089	0.6	$73,468	2.47	$181,466	$181,466
H-407	1		PSA Precooler / Deaerator Water Preheater #2	PINCH	21,034,730	31,414,870	1.49	$21,089	2002	$21,089	0.6	$26,827	2.47	$66,263	$66,263
H-408	1		PSA Air-cooled Precooler	QAH408	149,281,592	106,741,857	0.72	$388,064	2002	$388,064	0.6	$317,322	2.47	$783,786	$783,786
H-409	1		PSA Water-cooled Precooler	QCH409CT	8,414,338	7,346,116	0.87	$35,689	2002	$35,689	0.44	$33,619	2.47	$83,040	$83,040
R-402	1		High Temperature Shift Reactor	404	354,424	322,868	0.91	$465,907	2002	$465,907	0.56	$442,202	2.47	$1,092,238	$1,092,238
R-403	1		Low Temperature Shift Reactor	407	354,424	322,870	0.91	$323,464	2002	$323,464	0.56	$307,007	2.47	$758,307	$758,307
S-401	1		Pre-PSA Knock-out #1	413	354,424	322,870	0.91	$129,979	2002	$129,979	0.6	$122,907	2.47	$303,580	$303,580
S-402	1		Pre-PSA Knock-out #2	419	242,691	246,017	1.01	$55,291	2002	$55,291	0.6	$55,744	2.47	$137,689	$137,689
S-403	1		Pressure Swing Adsorption Unit	424	14,260	15,322	1.07	$4,855,471	2002	$4,855,471	0.6	$5,069,390	2.47	$12,521,394	$12,521,394
A400									**Subtotal**	**$6,566,421**		**$6,683,865**		**$16,509,147**	**$16,509,147**
H-501A	1		Hydrogen Compressor Intercooler	QAK501A	4,042,813	4,356,835	1.08	$53,601	2002	$53,601	0.6	$56,062	2.47	$138,472	$138,472
H-501B	1		Hydrogen Compressor Air-cooled Aftercooler	QAK501B	5,984,714	6,430,563	1.07	$56,901	2002	$56,901	0.6	$59,408	2.47	$146,737	$146,737
H-502	1		Hydrogen Compressor Water-cooler Aftercooler	QCH502CT	1,465,277	1,574,438	1.07	$18,909	2002	$18,909	0.44	$19,516	2.47	$48,205	$48,205
K-501	1		Hydrogen Compressor	501	14,260	15,322	1.07	$914,235	2002	$914,235	0.8	$968,331	2.47	$2,391,777	$2,391,777
S-501	1		Pre-hydrogen Compressor Knock-out	501	14,260	15,322	1.07	$13,377	2002	$13,377	0.6	$13,966	2.47	$34,497	$34,497
S-502	1		Hydrogen Compressor 1st Interstage Knock-out	501	14,260	15,322	1.07	Included in the hydrogen compressor cost (K-501)							
S-503	1		Post-hydrogen Compressor Knock-out	505	14,260	15,322	1.07	$13,977	2002	$13,977	0.6	$14,593	2.47	$36,044	$36,044
A500									**Subtotal**	**$1,071,000**		**$1,131,876**		**$2,795,733**	**$2,795,733**

Equipment Number	Number Required	Number Spares	Equipment Name	Scaling Stream	Scaling Stream Flow (lb/hr or btu/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2002$
H-601	1		Steam Turbine Condenser	614	93,974	131,510	1.40	Included in the extraction steam trubine/generator cost (M-602)							
H-602	1		Blowdown Cooler / Deaerator Water Preheater	PINCH	2,877,029	2,881,506	1.00	$3,043	2002	$3,043	0.6	$3,046	2.47	$7,523	$7,523
H-603	1		Blowdown Water-cooled Cooler	QCH603CT	626,343	627,318	1.00	$16,143	2002	$16,143	0.44	$16,154	2.47	$39,901	$39,901
M-601	1		Hot Process Water Softener System	631	349,266	349,809	1.00	$1,031,023	1999	$1,031,023	0.82	$1,032,338	2.47	$2,549,875	$2,582,516
M-602	1		Extraction Steam Turbine/Generator	607	342,283	342,816	1.00	$4,045,870	2002	$4,045,870	0.71	$4,050,339	2.47	$10,004,337	$10,004,337
M-603	1		Startup Boiler	200	36,560	36,560	1.00	$198,351	2002	$198,351	0.6	$198,351	2.47	$489,927	$489,927
P-601	1	1	Collection Pump	625	255,292	218,299	0.86	$7,015	2002	$14,030	0.33	$13,324	2.47	$32,909	$32,909
P-602	1	1	Condensate Pump	616	93,974	131,510	1.40	$5,437	2002	$10,874	0.33	$12,149	2.47	$30,009	$30,009
P-603	1	1	Deaerator Feed Pump	628	349,266	349,809	1.00	$8,679	2002	$17,358	0.33	$17,367	2.47	$42,896	$42,896
P-604	1	1	Boiler Feed Water Pump	639	349,268	349,812	1.00	$95,660	2002	$191,320	0.33	$191,418	2.47	$472,803	$472,803
T-601	1		Condensate Collection Tank	627	349,266	349,809	1.00	$24,493	2002	$24,493	0.6	$24,516	2.47	$60,554	$60,554
T-602	1		Condensate Surge Drum	638	349,268	349,812	1.00	$28,572	2002	$28,572	0.6	$28,599	2.47	$70,639	$70,639
T-603	1		Deaerator	633	349,266	349,809	1.00	$130,721	2002	$130,721	0.72	$130,867	2.47	$323,242	$323,242
T-604	1		Steam Drum	644	349,268	349,812	1.00	$9,200	1997	$9,200	0.72	$9,210	2.47	$22,749	$23,285
S-601	1		Blowdown Flash Drum	604	6,985	6,996	1.00	$14,977	2002	$14,977	0.6	$14,991	2.47	$37,029	$37,029
A600									**Subtotal**	**$5,735,975**		**$5,742,670**		**$14,184,394**	**$14,217,570**
K-701	2	1	Plant Air Compressor	101	367,437	367,437	1.00	$32,376	2002	$97,129	0.34	$97,129	2.47	$239,908	$239,908
M-701	1		Cooling Tower System	QCTOTAL	139,850,763	145,159,707	1.04	$267,316	2002	$267,316	0.78	$275,199	2.47	$679,741	$679,741
M-702	1		Hydraulic Truck Dump with Scale	101	367,437	367,437	1.00	$80,000	1998	$80,000	0.6	$80,000	2.47	$197,600	$200,695
M-703	1		Flue Gas Stack	112	1,174,206	639,526	1.00	$51,581	2002	$51,581	1	$51,668	2.47	$127,620	$127,620
				434		536,658	The stack flow is the sum of two flow streams.								
P-701	1	1	Cooling Water Pump	715	6,088,320	6,319,444	1.04	$158,540	2002	$317,080	0.33	$321,003	2.47	$792,877	$792,877
P-702	1	1	Firewater Pump	101	367,437	367,437	1.00	$18,400	1997	$36,800	0.79	$36,800	2.47	$90,896	$93,036
P-703	1	1	Diesel Pump	101	367,437	367,437	1.00	$6,100	1997	$12,200	0.79	$12,200	2.47	$30,134	$30,843
P-704	1	1	Ammonia Pump	101	367,437	367,437	1.00	$5,000	1997	$10,000	0.79	$10,000	2.47	$24,700	$25,282
P-705	1		Hydrazine Pump	101	367,437	367,437	1.00	$5,500	1997	$5,500	0.79	$5,500	2.47	$13,585	$13,905
S-701	1	1	Instrument Air Dryer	101	367,437	367,437	1.00	$8,349	2002	$16,698	0.6	$16,698	2.47	$41,244	$41,244
T-701	1		Plant Air Receiver	101	367,437	367,437	1.00	$7,003	2002	$7,003	0.72	$7,003	2.47	$17,297	$17,297
T-702	1		Firewater Storage Tank	101	367,437	367,437	1.00	$166,100	1997	$166,100	0.51	$166,100	2.47	$410,267	$419,926
T-703	1		Diesel Storage Tank	101	367,437	367,437	1.00	$14,400	1997	$14,400	0.51	$14,400	2.47	$35,568	$36,405
T-704	1		Ammonia Storage Tank	101	367,437	367,437	1.00	$287,300	1997	$287,300	0.72	$287,300	2.47	$709,631	$726,339
T-705	1		Olivine Lock Hopper	101	367,437	367,437	1.00	Included in gasification & clean up cost (R-201)							
T-706	1		MgO Lock Hopper	101	367,437	367,437	1.00	Included in gasification & clean up cost (R-201)							
T-707	1		Hydrazine Storage Tank	101	367,437	367,437	1.00	$12,400	1997	$12,400	0.93	$12,400	2.47	$30,628	$31,349
A700									**Subtotal**	**$1,381,507**		**$1,393,399**		**$3,441,695**	**$3,445,118**

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14. ABSTRACT *(Maximum 200 Words)*

This analysis developed detailed process flow diagrams and an Aspen Plus® model, evaluated energy flows including a pinch analysis, obtained process equipment and operating costs, and performed an economic evaluation of two process designs based on the syngas clean up and conditioning work being performed at NREL. One design, the current design, attempts to define today's state of the technology. The other design, the goal design, is a target design that attempts to show the effect of meeting specific research goals.

15. SUBJECT TERMS

Battelle; Columbus; gasifier; energy flow; economic evaluation

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