Process Design and Economics for the Conversion of Lignocellulosic Biomass to Hydrocarbon Fuels

Fast Pyrolysis and Hydrotreating Bio-oil Pathway

November 2013

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Executive Summary

The goal of the U.S. Department of Energy's Bioenergy Technologies Office (BETO) is to enable the development of biomass technologies to:

- Reduce dependence on foreign oil
- Promote the use of diverse, domestic, and sustainable energy resource
- Establish a domestic bioenergy industry
- Reduce carbon emissions from energy production and consumption. (DOE 2013)

To meet these goals, the BETO promotes the development of liquid hydrocarbon fuels that can serve as gasoline, jet and diesel blendstocks.

This report describes a proposed thermochemical process for converting biomass into liquid transportation fuels via fast pyrolysis followed by hydroprocessing of the condensed pyrolysis oil. As such, the analysis does not reflect the current state of commercially-available technology but includes advancements that are likely, and targeted to be achieved by 2017. The purpose of this study is to quantify the economic impact of individual conversion targets to allow a focused effort towards achieving cost reductions.

The report updates a study published in 2009 (Jones 2009) using the most current publically available data from PNNL, NREL and others. The fast pyrolysis of biomass is already being commercialized, while the upgrading via bio-oil hydrotreating to transportation fuels has only been demonstrated in the laboratory and on a small engineering development scale. The pyrolysis oil upgrading via hydrotreating section is revised to incorporate the most recent improvements: a low temperature stabilizer reactor has been added ahead of the two-stage hydrotreaters described in the 2009 report. Capital and operating costs are updated to reflect the current understanding of the system.

The process presented here represents a conceptual design that considers the economics of gasoline and diesel blendstock production assuming the achievement of internal research targets for 2017 coupled with nth plant costs and financing. The assumed processing capacity is 2,205 U.S. tons (2,000 metric tons) per day of dry biomass and results in a fuel production yield of 40 gallons of gasoline per dry U.S. ton of biomass and 44 gallons of diesel per dry U.S. dry ton of biomass. Natural gas is used to generate a portion of the hydrogen needed for hydrotreating and electricity is purchased from the grid. The minimum combined gasoline and diesel blendstock fuel selling price is $3.39 per gasoline gallon equivalent (lower heating value basis) in 2011 dollars.

Figure ES-1. Summary Economic Results

Gasoline & Diesel Blendstocks from Biomass Pyrolysis and Upgrading Conceptual Design Summary

All Values in 2011$

Minimum Fuel Selling Price (MFSP)	**$3.34**	**$/gallon gasoline blendstock**
Minimum Fuel Selling Price (MFSP)	**$3.71**	**$/gallon diesel blendstock**
MFSP - GGE Basis	**$3.39**	**$/gge total blendstocks**

Gasoline Blendstock Production	28.9	million gallons/year
Diesel Blendstock Production	31.6	million gallons/year
Gasoline Blendstock Yield	39.9	gallons/dry US ton wood
Diesel Blendstock Yield	43.7	gallons/dry US ton wood
Total Blendstock Fuel Yield	10.1	million BTU/dry US ton wood
Feedstock + Handling Cost	80.00	$/dry short ton feed
Internal Rate of Return (After-Tax)	10%	
Equity Percent of Total Investment	40%	
On-Stream Factor	90%	

Capital Costs		
A100	Fast Pyrolysis & Quench	$162,000,000
A100	Heat Recovery & Filtration	$13,000,000
A200	Hydrotreating	$115,000,000
A300 - 500	Product Finishing	$19,000,000
A600	Hydrogen Generation	$69,000,000
A700	Balance of Plant	$9,000,000
Total Installed Equipment Cost		$387,000,000
Added Direct + Indirect Costs		$313,600,000
	(% of TCI)	45%
Total Capital Investment (TCI)		$700,600,000
Installed Capital/Annual Gallon Blendstocks		6
TCI/Annual Gallon Blendstocks		12
Loan Rate		8.0%
Term (years)		10
Capital Charge Factor (Computed)		0.128

Performance	
Plant Purchased Electricity (KWh/gal blendstocks)	1.27
Water Usage (gallons/gal blendstocks)	1.49
Natural Gas (scf/gal blendstocks)	20

Manufacturing Costs ($/GGE)	
Feedstock + Handling	0.92
Natural Gas	0.10
Catalysts & Chemicals	0.32
Waste Disposal	0.01
Electricity and other utilities	0.09
Credits	0.00
Fixed Costs	0.53
Capital Depreciation	0.35
Average Income Tax	0.10
Average Return on Investment	0.96
	3.39

Manufacturing Costs ($/yr)	
Feedstock + Handling	$57,900,000
Natural Gas	$6,200,000
Catalysts & Chemicals	20,400,000
Waste Disposal	$500,000
Electricity and other utilities	$5,500,000
Credits	$0
Fixed Costs	$33,600,000
Capital Depreciation	$22,200,000
Average Income Tax	$6,600,000
Average Return on Investment	$60,900,000

Abbreviations

ASTM	American Society of Testing & Materials
BETO	Bioenergy Technologies Office
BBL	Barrel (42 gallons)
BFW	Boiler Feed Water
BPSD	Barrels Per Stream Day
BTU	British Thermal Units
C&D	Construction and Demolition
CFB	Catalytic Fluidized Bed
CoMo	Cobalt Molybdenum
COD	Chemical Oxygen Demand
CWS	Cooling Water Supply
CWR	Cooling Water Return
DOE	Department of Energy
DAF	Dry and Ash Free
EISA	Energy Independence and Security Act
EPA	Environmental Protection Agency
GC	Gas Chromatography
GGE	Gasoline Gallons Equivalent
GHG	Green House Gas
GPD	Gallons Per Day
GPM	Gallons Per Minute
HCK	Hydrocracking
HDS	Hydrodesulfurization
HDO	Hydrodeoxygenation
HTS	High Temperature Shift
INL	Idaho National Laboratory
IRR	Internal Rate of Return
kWh	Kilowatt Hour
lb	Pound
LCA	Life Cycle Analysis
LHSV	Liquid Hourly Space Velocity
LHV	Lower Heating Value
LTS	Low Temperature Shift
MACRS	Modified Accelerated Cost Recovery System
MFSP	Minimum Fuels Selling Price
MJ	Mega Joule
MM	Million
mm	Millimeters
MS	Mass Spectrometry
MYPP	Multi-Year Program Plant
NiMo	Nickel Molybdenum
NOx	Nitrogen Oxides
NREL	National Renewable Energy Laboratory
PNNL	Pacific Northwest National Laboratory
PDU	Process Development Unit
PSA	Pressure Swing Adsorption
RFS2	Renewable Fuel Standard
SCF	Standard Cubic Foot
SCFD	Standard Cubic Foot per Day
SCR	Steam-to-Carbon Ratio
SOT	State of Technology
SOx	Sulfur Oxides
TAN	Total Acid Number
TEA	Techno-Economic Analysis
TIC	Total Installed Capital
Ton	U.S. ton (2000 pounds)
TPEC	Total Purchased Equipment Cost
VOC	Volatile Organic Carbon
WGS	Water Gas Shift
YR	Year

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1. Introduction

The goal of the Bioenergy Technologies Office (BETO), as stated in the Multi-Year Program Plan (MYPP), is to enable the development of biomass technologies to:

- Reduce dependence on foreign oil
- Promote the use of diverse, domestic, and sustainable energy resource
- Establish a domestic bioenergy industry
- Reduce carbon emissions from energy production and consumption (DOE 2013).

In recent years, BETO completed a number of techno-economic evaluations of both biological and thermochemical pathways for converting biomass to fuels. These evaluations, termed "design cases", provide a detailed basis for understanding the potential of various conversion technologies and help identify technical barriers where research and development could potentially lead to significant cost improvements. Consistent assumptions for items such as plant lifetimes, rates of return, and other factors were used in all cases so the various conversion pathways could be assessed on a comparative basis.

The purpose of this report is to evaluate the conversion of biomass into infrastructure-compatible hydrocarbon fuels via fast pyrolysis followed by the catalytic upgrading of condensed pyrolysis vapors. This report builds upon the 2009 design report (Jones 2009) and is updated with the most recent laboratory and commercial data. The pathway presented here represents a goal case targeting performance potentially available between now and 2017. The fast pyrolysis of biomass is already being commercialized, while the upgrading of the bio-oil to transportation fuels has only been demonstrated in the laboratory and on a small engineering development scale. As such, the analysis does not reflect the current state of commercially-available technology but includes advancements that are likely, and targeted to be achieved by 2017.

This report also provides the basis for the annual technical and economic goals leading to the overall BETO programmatic 2017 cost target of $3/gasoline gallon equivalent (gge) fuel. Research improvements will be incorporated into the models and reported in the annual state of technology assessments (Jones 2012, Jones 2013) then compared to the conversion goals set forth in this report.

1.1. Approach

The approach to developing techno-economics is similar to that employed in previous conceptual conversion pathways and associated design reports produced for BETO (Aden 2002, Spath 2005, Phillips 2007, Jones 2009, Dutta 2011, Humbird 2011). Process flow diagrams and models are based on research results from completed and ongoing research at PNNL, NREL and INL, other research organizations, as well as information from commercial vendors for mature and similar technologies. The process models are developed using commercial process flow simulation software tools such as CHEMCAD (Chemstations 2013) and Aspen Plus (Aspentech 2013). The heat and material balances from the simulation models are used to estimate the capital and operating costs. These are then assembled in a Microsoft Excel© spreadsheet employing a discounted cash flow analysis to estimate the minimum product selling price.

This approach is summarized in Figure 1:

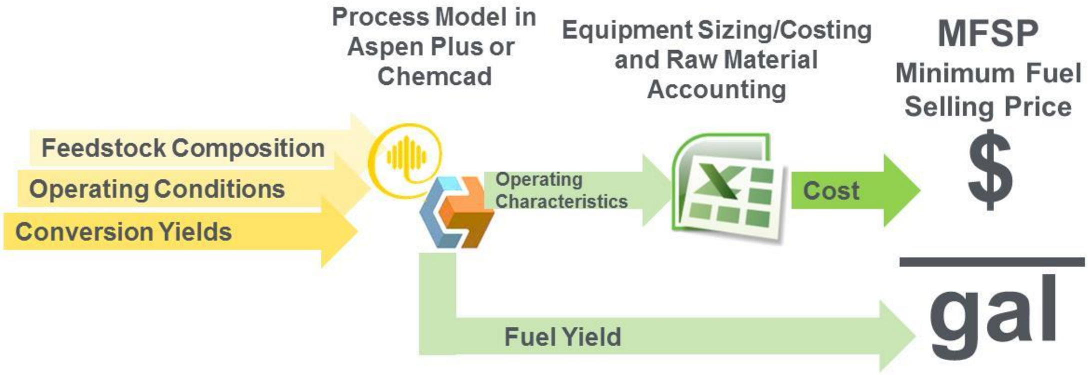


Figure 1. Techno-Economic Analysis Approach

1.2. Definition of nth Plant

A standard reference basis common to the BETO reports, known as the "nth" plant design is used. These assumptions do not account for additional first of a kind plant costs, including special financing, equipment redundancies, large contingencies and longer startup times necessary for the first few plants. For nth plant designs, it is assumed that the costs reflect a future time when the technology is mature and several plants have already been built and are operating. The specific assumptions are shown in Table 1.

Table 1. Nth Plant Assumptions

Assumption Description	**Assumed Value**
Internal rate of return	10%
Plant financing debt/equity	60% / 40% of total capital investment
Plant life	30 years
Income tax rate	35%
Interest rate for debt financing	8.0% annually
Term for debt financing	10 years
Working capital cost	5.0% of fixed capital investment (excluding land)
Depreciation schedule	7-years MACRS[1] schedule
Construction period	3 years (8% 1st yr, 60% 2nd yr, 32% 3rd yr)
Plant salvage value	No value
Startup time	6 months
Revenue and costs during startup	Revenue = 50% of normal Variable costs = 75% of normal Fixed costs = 100% of normal
On-stream factor	90% (7884 operating hours per year)

[1] Modified Accelerated Cost Recovery System

1.3. Cost Estimation Basis

All costs in this report are on a 2011 constant dollar basis. This is the current reference year that the BETO uses to facilitate comparison of various conversion technologies (DOE 2013). Indices used to convert capital and operating costs to the 2011 dollars can be found in Appendix G.

Capital costs are estimated from a variety of resources. The heat and material balances generated by the simulation software (CHEMCAD or Aspen Plus) are used to size the major pieces of equipment. These are input into Aspen Capital Cost Estimator software (reported in 1st quarter 2011 costs) to determine the installed capital cost. In addition, select data from commercial vendors, either as budgetary estimates or from their published literature are used when available.

The original cost reflects the year of the cost quote or estimate, and the scale of the equipment. All capital costs are adjusted to an annualized 2011 basis using the Chemical Engineering magazine's published indices:

$$\text{Cost in 2011 \$} = \text{Equipment cost in quote year} \times \left(\frac{\text{2011 index} = 585.7}{\text{Quote cost year index}}\right)$$

The scale is adjusted to the match the appropriate scaling term (heat exchanger area for example) by using the following expression:

$$\text{Current equipment cost} = \text{Cost at original scale} \times \left(\frac{\text{Current capacity}}{\text{Original capacity}}\right)^{n}$$

Where 'n' is the scale factor, typically, 0.6 to 0.7.

Once the equipment is scaled and adjusted to the common cost year, factors are applied to calculate the total capital investment. Individual installation factors calculated by Aspen Capital Cost Estimator are multiplied by equipment costs, unless installed costs are already available from vendors. The total direct cost is the sum of all the installed equipment costs, plus the costs for buildings, additional piping and site development (calculated as 4%, 4.5% and 10% of purchased equipment, respectively) . Indirect costs are estimated as 60% of the total installed costs. The sum of the direct and indirect costs is the total capital investment.

2. Plant Design Overview

This section describes the basis for the plant scale, feedstock type and general characteristics of the conversion units.

2.1. Feedstock and Plant Size

The feedstock is a low ash, woody biomass delivered at 30 wt% moisture. The feedstock cost includes the capital and operating costs for drying biomass to 10 wt% moisture and grinding to 2-6 mm. The ultimate analysis for the feedstock is presented in Table 2. Process heat is provided by the conversion plant to dry the biomass from 30 wt% to 10 wt% moisture.

Table 2. Feedstock and Processing Assumptions

Component	Weight % (dry basis)
Carbon	50.94
Hydrogen	6.04
Nitrogen	0.17
Sulfur	0.03
Oxygen	41.90
Ash	0.9
Heating value (Btu/lb)	8,601 HHV 7,996 LHV

The plant capacity is 2205 dry U.S. tons per day (2000 metric tons per day) of bone dry wood. Hereafter, all references to tons in this report refer to U.S. tons (2000 pounds).

2.2. Conversion Summary

A simplified block diagram of the overall design is given in Figure 2. The major processing areas are as follows:

Area 100 Fast Pyrolysis: biomass is heated to approximately 932°F (500°C) in less than two seconds at atmospheric pressure, and then rapidly cooled to stop the reaction. The cooled pyrolysis products are primarily liquid (water and organic compounds). Solid char mixed with biomass ash, and non-condensable gases are also produced.

Area 200 Hydrotreating: the fast pyrolysis liquid product or bio-oil is deoxygenated by catalytic hydrotreating at elevated pressures in an excess of hydrogen. Multiple fixed bed reactors are used with a staged increase in processing severity with each reactor in order to reduce overall coking. The upgraded oil is deoxygenated to less than 2 wt% oxygen.

Area 300 Product Separation: the wide boiling range finished oils from hydrotreating and hydrocracking are fractionated into gasoline and diesel boiling range blendstocks.

Area 400 Fuel Gas Collection: off-gas from fast pyrolysis and the upgrading sections are collected and routed to the hydrogen plant and to process heaters.

Area 500: Hydrocracking: the heavy hydrocarbon fraction from the hydrotreated oil is cracked to additional gasoline and diesel blendstocks in a reactor system similar to that used in conventional refineries.

Area 600 Steam Reforming: hydrogen for hydrotreating is produced by steam reforming of natural gas mixed with the off-gases from the hydrotreater and pyrolysis reactor.

Balance of Plant: Wastewater treatment, cooling water, tank farm, and flare are included in this section.

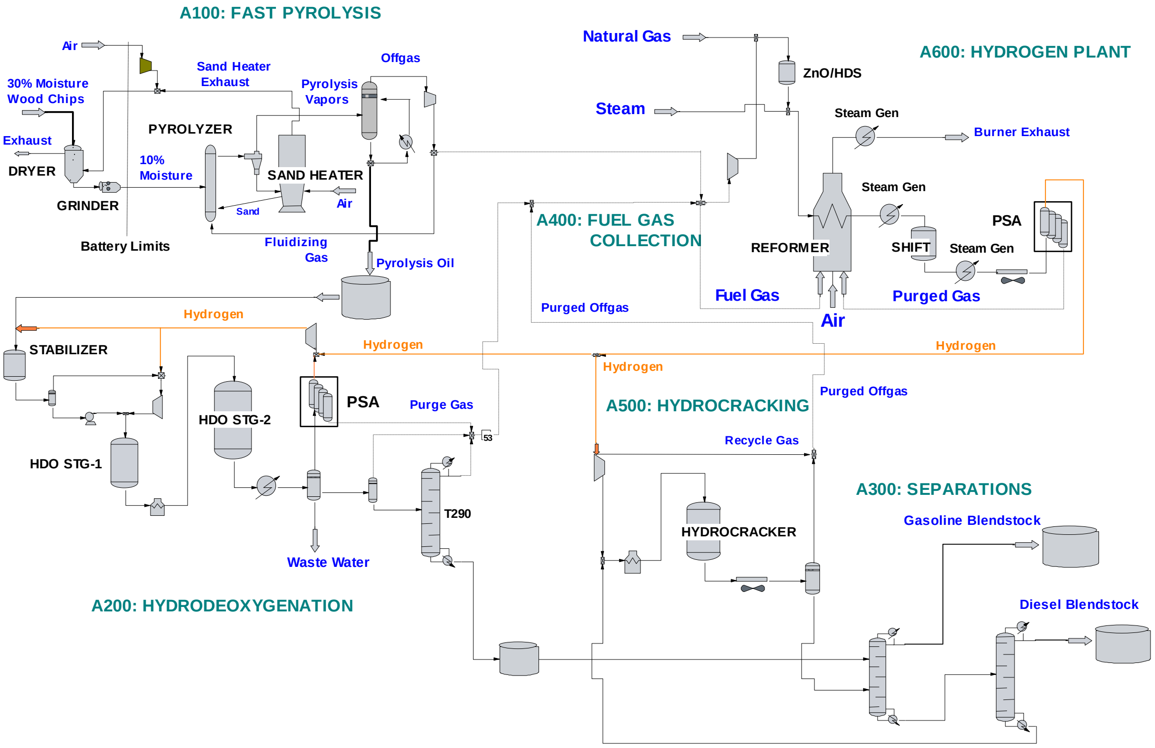


Figure 2. Simplified Flow Diagram for the Overall Process

3. Process Design and Cost Estimation

The standalone biorefinery is assumed to include feedstock reception, fast pyrolysis, hydrotreating and hydrocracking of pyrolysis oil, separation of upgraded oil to gasoline and diesel fuel blendstocks, and hydrogen generation. Process utilities include cooling water, on-site generated steam and electricity, and purchased electricity.

3.1. Feedstock Logistics, Handling and Drying

Woody feedstock is brought from the field to the conversion plant to meet a specific set of requirements. The details of the feedstock logistics and the rational for the feedstock cost and specifications are provided by the Idaho National Laboratory (INL) and are summarized in Appendix E. Forest thinnings and logging residues and construction and demolition (C&D) wastes are low cost resources to procure, but also have unfavorable quality specifications, specifically high ash content. The ash type and quantity will have an effect on the yield of fast pyrolysis oil as certain ash constituents can cause an increase in the gas production at the expense of condensable liquid (Table 3).

Table 3. The Effect of Feed Type and Ash on Fast Pyrolysis Oil Yield (Oasmaa 2010)

Feedstock 3-5mm, 7% moisture	% Ash (dry basis)	Organic Yield wt% on Dry Feed
Pine (20 kg/h PDU)	0.1wt%	64%
Pine (300 kg/h unit)	0.1wt%	62%
Forest Residues	2 – 4 wt%	46 – 55%
Straws & Hay	3 – 6 wt%	36 – 45%

Because these biomass resources are low cost, supply chains that include active ash management preprocessing unit operations can be purchased. Prior to preprocessing, certain resources can be blended to reduce the overall percentage of ash and moisture, thereby reducing the costs and severity needed to reduce the ash to in-feed specifications. Table 4 shows an example formulation.

Table 4. Costs and Specifications for Woody Feedstocks and Blends

Feedstock	Reactor Throat Feedstock Cost ($/dry ton)[2]	Formulation Fraction (%)	% Ash Delivered to Throat of Conversion Reactor
Pulp	99.38	30	0.5
Logging Residues[1]	74.83	35	1.5
Switchgrass	80.54	10	2.8
C&D Wastes	63.77	25	0.5
Formulation Totals	**80.00**	**100**	**1.1**

[1] residues do not include costs for harvest and collection; they are moved to landing while attached to the merchantable portion of the tree (for example, timber or pulpwood)

[2] includes ash mitigation

The C&D wastes are incorporated because of its low access fee cost and its low ash content. This is only an example; the actual blends will be regionally based designs that take advantage of local feedstocks and their biomass characteristics. Additionally the ability to blend feedstocks to a specification has the potential to reduce some of the risks associated with the seasonality of feedstocks.

Currently, the forest thinnings and logging residues are both reduced to 1.05% ash after a chemical pretreatment. C&D wastes will undergo a low severity pretreatment and the ash content will be reduced to 0.9%. The purpose grown pine is debarked and chipped at the landing resulting in an ash content of 0.5%. The blended material will be delivered to the conversion facility a 2 wt% ash. A further pretreatment operation will be performed on the blended material that will reduce the ash to the required 0.9%. It is important that the pretreatment operation be inexpensive or offset the costs by improving other operations, such as conversion yields, decrease grinding or pelletizing costs. There are a number of chemical pretreatment options being explored that will reduce the ash. INL is focusing on these pretreatment technologies but the goal is to meet both the $80 per short ton cost target and 0.9% ash specifications assumed in this report.

Blended woody feedstock is supplied as shown in Figure 3. Although in this report the heat for biomass drying is supplied from the conversion plant, future feedstock supplies will be dried at the depot using a local heat source.

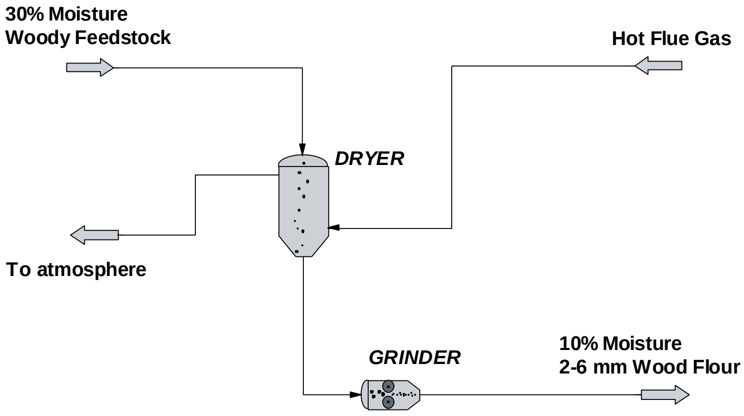


Figure 2. Feed Handling and Drying Process Flow

The feedstock is dried from 30 wt% (as received) to 10 wt% moisture content using hot flue gas from the pyrolysis reactor char combustor. The combustion exhaust is fed to the dryer at 584°F (307°C) with an oxygen content of approximately 5-10% by volume to avoid feedstock combustion (Searcy 2010). The flue gas temperature must be higher than the water dew point to avoid any corrosion and low enough to avoid combustion. The flue gas temperature is adjusted by cooling and air makeup before being fed to the dryer. After the drying step, the feedstock is ground to 2-6 mm particle size to yield sufficiently small particles, ensuring rapid reaction in the pyrolysis reactor.

3.2. Area 100: Fast Pyrolysis

Pyrolysis is the thermal decomposition of carbonaceous material in the absence of oxygen to produce solid char, gas, and a liquid product. Residence times for fast pyrolysis are on the order of seconds (Basu

2010). After rapidly quenching the pyrolysis vapors and aerosols, three product phases are formed. The yields of each phase depend on the operating conditions, reactor design and feedstock characteristics, including ash content and the relative amounts of cellulose and lignin. The products are:

Fast pyrolysis liquid: (also known as bio-oil and pyrolysis oil) is obtained after vapor condensation. Bio-oil is black or dark brown and free flowing at room temperature and typically contains less than 30% of water and hundreds of oxygenated components (Elliott 2013). It is mostly immiscible in hydrocarbon liquids and can be upgraded by hydrotreating to lower the oxygen content and decrease hydrophilicity.

Solid char: is primarily composed of carbon. It is separated from the fast pyrolysis vapors and aerosols by cyclone. This solid product can be used as fuel.

Non-condensable gas: is collected during vapor condensation. The gas is recycled internally as fluidizing gas for the fast pyrolysis reactor and/or collected for fuel use.

There are several types of pyrolysis reactor configurations:

- *Ablative pyrolysis:* Mechanical pressure is used to press biomass against a heated reactor wall, essentially melting the biomass and evaporating the resulting vapors. The two main advantages compared to other reactor types are that inert gas is not required and larger feedstock particles can be used. However, scaling is a linear function of the heat transfer area and does not benefit from the same economies of scale as the other systems. The system is also slightly more complex due to the mechanical nature of the process (IEA 2007). The PYTEC system in Germany produces approximately 4.4 tons per day of pyrolysis oil (Meier 2013).
- *Auger Pyrolysis:* This reactor uses a screw to mix a heat carrier and biomass, and is typically limited in size and potential to scale-up. Karlsruhe Institute of Technology has developed a 13.2 tons per day twin-screw mixing pyrolysis reactor (Bioliq 2013). It is designed to produce a slurry mixture of bio-oil and char that can be transported to a centralized gasification and Fischer-Tropsch plant.
- *Entrained Flow Pyrolysis:* This reactor configuration is popular for studies of thermochemical conversion kinetics and investigations of pressure effects. Feed material is typically fed into the top of the reactor, co-current with a gas stream. The flow through this configuration is assumed to approximate plug flow, with residence time controlled by the length of the heated zone. Char buildup and inefficient heat transfer can be a troublesome tradeoff for these simple and inexpensive reactors. Liquid yields are usually lower than fluid bed systems (Bridgwater 2008).
- *Rotating Cone:* This reactor combines biomass and hot sand at the bottom of the vessel and mixes the solids using a rotating cone inside the vessel. Hot pyrolysis oil vapors leave near the top of the reactor and are passed through several cyclones to remove fine particles before being condensed by direct contact with recirculating cooled oil. No carrier gas is needed, resulting in smaller downstream equipment. A 2.2 ton per day plant operating on palm oil empty fruit bunches was commissioned in Malaysia in 2006 and a 5.5 ton per day plant is in development (BTG 2013).
- *Bubbling Fluidized Bed Pyrolysis:* Here, biomass is introduced to hot sand fluidized by recirculated product gas. This technology is well understood, simple to construct, can operate on large scales, and is very efficient in transferring heat to the biomass, resulting in high liquid yields. Small biomass particle sizes are required in order to obtain high heating rates (Bridgwater 2008).

- *Circulating Fluidized Bed Pyrolysis*: Similar to bubbling fluidized beds, these configurations circulate hot sand between the pyrolysis reactor and a sand re-heater. Char or product gas and char are burned to provide heat to the sand re-heater. Circulating beds are widely used at very high throughputs in the petroleum industry and are potentially suitable for large biomass throughputs as well. Ensyn operates a 110 ton per day plant in Canada and has designed a 440 ton per day skid mounted unit (Bridgwater 2008, Ensyn 2012). Fortum announced an integrated combined heat and power plant in Finland using a circulating fluid bed pyrolyzer, where 55,000 tons per year of pyrolysis oil is expected to be produced starting in the fall of 2013 (Fortum 2013).

A circulating fluidized bed was chosen as the representative pyrolyzer for this report as this design is the most likely to be scalable to the assumed feed rate. Figure 4 shows the simplified flow diagram for the fast pyrolysis and bio-oil quenching area.

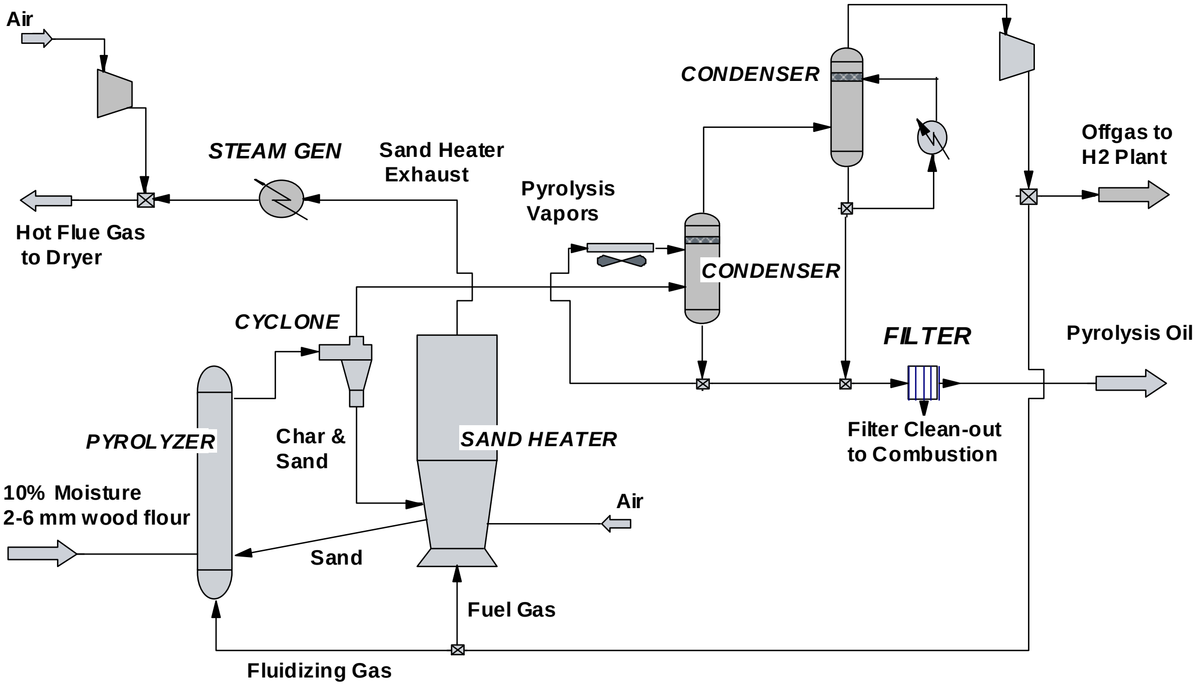


Figure 3. Fast Pyrolysis Process Flow

The dried, finely ground biomass is fed to the reactor operating at 932°F (500°C). Sand is used as a heat carrier and the residence time is less than two seconds. The biomass is converted into a mixture of vapors and char. A series of cyclones separate the sand and char from the vapors. The hot vapors are rapidly quenched in a two-stage system with previously condensed and cooled bio-oil and non-condensable gases are separated from the condensed bio-oil (Solantausta 2003, Freel 1998). In the first stage, most of the condensable products are removed using recirculated and indirect air-cooled bio-oil. The second stage operates at a lower temperature by indirect water cooling of the recirculating bio-oil. High temperature heat recovery is not included because of the likelihood of severe fouling in the recuperators (Johnson 2006). Most of the gases are recycled back to the pyrolysis reactor to assist fluidization. Char and a portion of the gas are burned to heat the circulating sand. The remaining gas product is sent to the hydrogen plant. The cooled pyrolysis oil is filtered before being processed in the hydrotreaters.

Reported char combustor temperatures are in the range of 1171°F (633°C) (Coll'Energia 2013) to 1400°F (760°C) (Trebbi 1997). The sand temperature is assumed to be the same as the char combustor operating temperature which is high enough to increase the reactor inlet temperature (dried feedstock + fluidized gas + sand) to 932°F (500°C). Silicon dioxide is used to represent sand as the heat carrier in the simulation. The sand to biomass weight ratio is reported to be in the range of 10-20 and the fluidization gas to biomass weight ratio is in the range of 2-9, with the exact conditions dictated by process conditions and the reactor design (Freel 1998, Peacocke 1997, Solantausta 2003).

Fast pyrolysis product yields have been published for various feedstock types (Oasmaa 2010). Fast pyrolysis liquid contains some solids such as residual sand and char which must be removed before hydrotreating. Filtering removes the solids as a mixture of oils and solids. A possible use for the filter retentate includes burning it for power generation or feeding it back to the fast pyrolysis reactor (Frey 2012). For this design, the bio-oil filter retentate is assumed to be burned in the sand heater. The assumptions used to determine the yields of model compounds representing pyrolysis oil are described in detail in Appendix F.

3.2.1. Fast Pyrolysis Design Basis

The model assumptions are shown in Table 5.

Table 5. Fast Pyrolysis Modeled Yields

Product	Property or compound	Base Case Assumption	Typical Values (IEA 2007)
Oil	Weight % of dry biomass	64 wt% in the reactor 62 wt% after filtration	55 – 65 wt% for low ash feed
	Carbon, wt% DAF[1]	56.61 wt%	56%
	Hydrogen, wt% DAF	6.61 wt%	6.5%
	Oxygen, wt% DAF	36.77 wt%	37.5%
	HHV, MJ/kg	16.9	17
	Density, g/ml	1.2	1.2
Reaction water	Weight % of dry biomass	12 wt%	5 – 12 wt%
Char &ash	Weight % of dry biomass	12 wt%	12 – 16 wt%
Gas	Weight % of dry biomass	12 wt%	12 – 16 wt %
	CO	46 wt% of gas	
	CO_2	43 wt% of gas	
	CH_4	6 wt% of gas	
	C_2+	5 wt% of gas	
	H_2	<1 wt% of gas	

[1] DAF = dry ash free

The design basis for the biomass dryer, fast pyrolysis reactor and char combustor are presented in Table 6. The hot air flow rate and temperature to the dryer is based on discussions with INL regarding the dryer operation.

Table 6. Area 100 Design Basis

Process	Assumption
Biomass Drying – Flue gas to dryer	
Flue gas exiting sand heater	1128°F (609°C)
Flue gas plus additional air rate	572,700 lb/h
Flue gas + added air temperature at dryer inlet	584°F (307°C)
Flue gas + added air oxygen content	10% maximum by volume
Fast Pyrolysis Reactor	
Sand to biomass ratio	14.5 wt/wt
Fluidized gas to biomass ratio	3 wt/wt
Hot Sand temperature	1128°F (609°C)
Reactor inlet temperature	932°F (500°C)
Heat loss	1% of biomass LHV
Char Combustor	
Excess oxygen	20%
Temperature	1129°F (609°C)
Heat loss	1% of biomass LHV
Cold Filter	
Cold filter bio-oil loss	2 wt% of dry biomass

3.2.2. Fast Pyrolysis Cost Estimation

The capital costs used are based on published inside battery limit costs for a 440 tons dry biomass per day commercial unit of $38 million ±40% excluding front-end material handling, utilities, offsite requirements, product storage, foundations, site preparation, buildings, spares, safety containment, fire water systems, wastewater system, startup fuel systems, emissions control, and installation (Envergent 2009). It is assumed that the n^{th} plant cost for this unit will scale to 1100 tons per day and that two 1100 ton per day units are used in parallel. A budgetary estimate for a cold filter was obtained from the Pall Corporation. Costs for the air compressor and heat recovery were estimated using Aspen Capital Cost Estimator (Aspentech 2013). These costs are detailed in Appendix B.

3.3. Area 200: Hydrotreating to Hydrocarbons

Hydrodeoxygenation can stabilize bio-oil and convert it to a conventional hydrocarbon fuel. Hydrocarbon hydrotreating to remove nitrogen and sulfur is a common and well established refinery process (Al Sabawi 2012). However, oxygen removal on the scale needed to upgrade pyrolysis oil is much less developed. Bio-oil hydrotreating involves contacting the bio-oil with a large excess of hydrogen and produces gas and two liquid fractions. The compounds in the gas product are light hydrocarbons and carbon dioxide (Elliott 2012). The liquid products consist of hydrocarbon oil and an aqueous phase, which separate easily. The aqueous phase carbon content depends upon the degree of bio-oil deoxygenation. The hydrocarbon oil yield and quality is dependent on catalyst selection, reactor configuration and hydrotreating conditions. Fixed bed reactors similar to the ones used in conventional hydroprocessing of petroleum to finished fuels can be used. An alternate reactor configuration, known as an ebullated bed, is also under development at the bench scale for this application (Elliott 2013). The ebullated bed is a liquid phase fluidization reactor using the dilute phase and the agitation of catalyst particles to eliminate the possibility of plugging the catalyst bed. The degree of the deoxygenation in the ebullated bed, however, is not expected to be as high as in the fixed bed. Fixed bed reactors are assumed in this study.

Single-stage hydrotreating has proved to be difficult, producing a heavy, tar-like product. Multi-stage processing, where mild hydrotreating is followed by more severe hydrotreating, has been found to overcome the reactivity of the bio-oil and prevent catalyst coking (Elliott 2007). Low temperature hydroprocessing can be used to pretreat the unstable bio-oil in order to reduce the most reactive oxygenated compounds before completely deoxygenating the oil under more severe hydrotreating conditions (higher temperature, lower space velocity) (Elliot 2013).

Chemical reactions during bio-oil hydrotreating are very complex. Overall, pyrolysis oil is almost completely deoxygenated by a combination of hydrodeoxygenation and decarboxylation, with oxygen removed in the form of water and carbon oxides. Example reactions are shown in Figure 5 taken from Zacher et al. (2013):

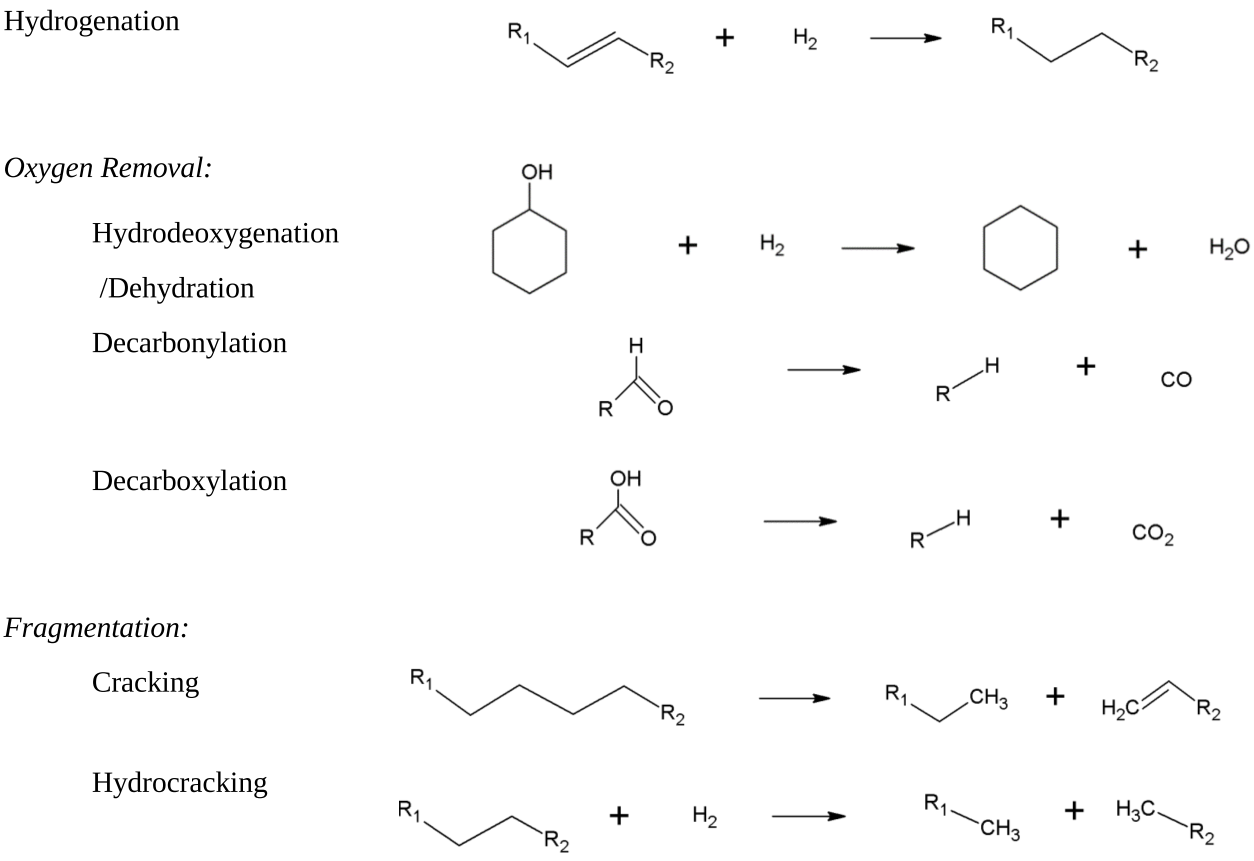


Figure 4. Examples of Hydrotreating Reactions

The reactivities of oxygenated compounds in bio-oil in the presence of catalysts are known to depend on the processing temperature. A hydrotreatment reactivity scale for various bio-oil oxygenated groups developed by E. Laurent is shown in Figure 6 (reprinted with permission from Elliott 2007).

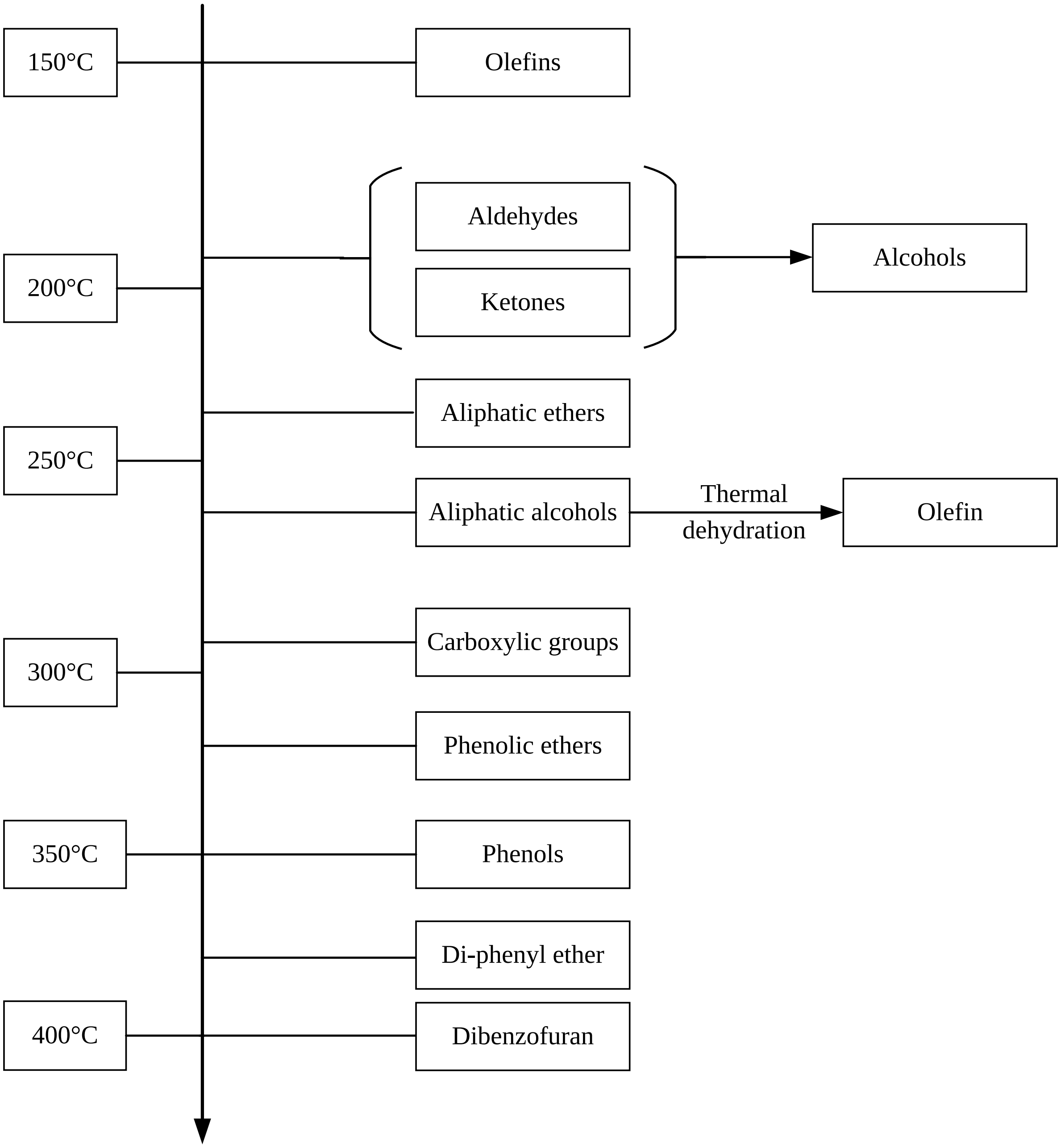


Figure 5. Reactivity Scale of Oxygenated Groups under Hydrotreatment Conditions

Proper selection of hydrotreating catalysts and optimum operating conditions can help control reaction selectivity and minimize hydrogen consumption. Several types of catalysts for bio-oil upgrading have been studied (Elliott 2007, Wang 2013). A recent study by Wang (2013) reviews hydrotreating of pyrolysis oil and model compounds; catalysts types tested are shown in Table 7.

Table 7. Catalysts for Bio-Oil Hydrotreating (Wang 2013)

Catalog	Catalysts
Mo-based sulfides	Bulk or supported MoS_2, Ni-MoS_2, Co-MoS_2
Noble metals	Supported Ru, Rh, Pd, Pt, Re, Pt-Rh, Pd-Rh, Pd-Cu, Pd-Fe, Pr-Re, Ru-Mo; Ru, Pt,Ph nanoparticles
Base metals	Supported Cu, Ni, Ni-Cu, Ni-Fe and Raney nickel
Metal phosphides	Supported Ni_2P, MoP, CoMoP, Fe_2P, WP and RuP
Other metal catalysts	Bulk Ni-Mo-B, supported nitrides (Mo_2N) and carbides, (Mo_2C); supported Mo-based oxide (MoO_2), MoO_3)
Bifunctional catalysts	Noble metal or base metal catalysts with aqueous acid including CH_3COOH, H_2PO_4, and Nafion or solid acid including HZSM-5, H-Beta, H-Y, sulfated zirconia and supported Nafion; metals including Pt, Pd and Ni supported on acid solid including HZSM-5, H-Beta, Hy and bulk acidic salt

The filtered bio-oil product from the pyrolysis unit is pumped to high pressure, then combined with compressed hydrogen and preheated. The primary objective of this section is to upgrade bio-oil to infrastructure-compatible fuels. Single-stage hydrotreating of bio-oil produces a tar-like product (Elliott 2007); in addition, the catalyst life is short as raw bio-oil is a mixture of reactive oxygenated compounds and is thermally unstable. Accordingly, the upgrading process is performed in multi-stage fashion with increased processing severity in each subsequent stage, as shown in Figure 7.

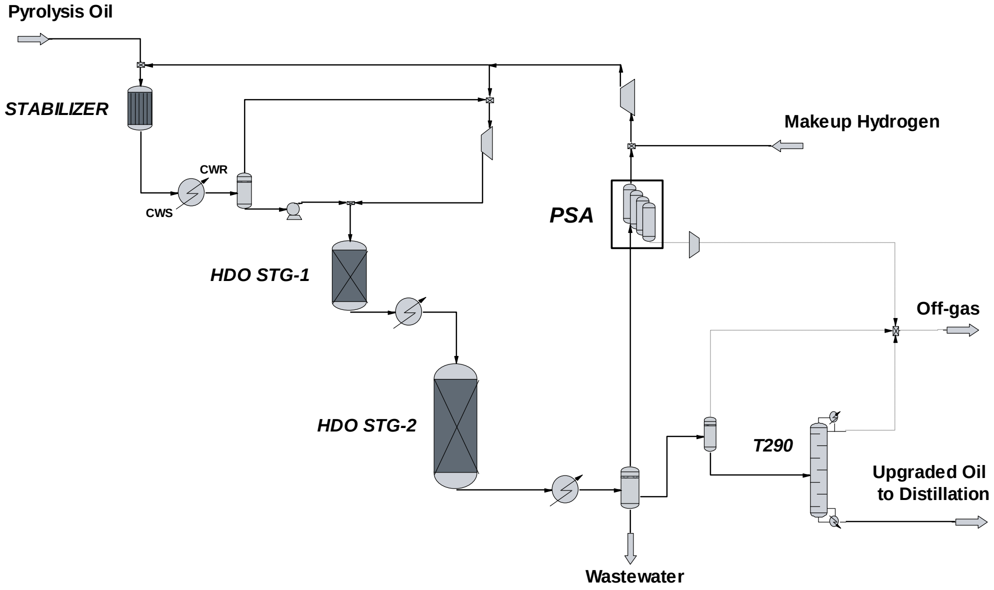


Figure 6. Bio-Oil Upgrading Process Flow

Bio-oil is first pretreated in a stabilization bed under relatively mild process conditions, 284 to 356°F (140 to180°C) and 1200 psia, followed by processing under more severe hydrotreating conditions in the 1st and 2nd stage hydrotreating reactors (Jones 2013). The 1st stage hydrotreating reactor is designed as a single bed catalytic reactor operated at 356 to 482°F (180 to 250°C) and 2000 psia. The 2nd stage hydrotreating reactor is operated at a higher temperature of 662 to 797°F (350 to 425°C).

Limited data are available to understand the conversion products from the stabilizer and the first stage reactor. Detailed product analysis is needed to understand what compounds are being formed and which are disappearing. For model development purposes, reactions and intermediate products from the stabilization bed and the 1st stage hydrotreating reactor are estimated based on the reactivity scale of oxygenated groups shown in Figure 6. At the given conditions, the stabilizer is likely to be converting carbonyls and olefins with very little hydrogen consumption and a slight exotherm. Similarly, the first stage converts more of these reactive compounds, again with low hydrogen consumption and a mild exotherm. Future work will consider the need for the first stage reactor. The second stage reactor converts the remaining oxygenated compounds to hydrocarbons and saturates some of the aromatics. The exotherm in this reactor can be large, and heat management is necessary. It is likely that this reactor will be operated in a fashion similar to hydrocrackers, where multiple beds in a single vessel are each cooled with cold hydrogen. For the 2nd stage hydrotreating reactor, product yields, gas and liquid product compositions and the hydrotreated oil distillation curves are derived from PNNL experimental data. Details of reaction conversions, product yields and process conditions of each hydrotreating reactor are summarized in Appendix C.

Products from the last hydrotreating stage are gas and two liquid fractions. The liquid fractions are an aqueous phase and stable hydrocarbon oil phase. The gas product is primarily non-condensable hydrocarbons (methane, ethane, propane, butane), carbon dioxide and excess hydrogen. The carbon dioxide concentration is smaller compared to the hydrocarbons (Elliott 2012). Excess hydrogen is assumed to be recovered by pressure swing adsorption and recycled. The separated aqueous phase product is heavier than the hydrotreated oil phase and contains very little carbon (Elliott 2012), typically less than a half percent by weight. The second stage hydrocarbon product contains less than 2 wt% oxygen. The hydrocarbon product is fractionated into gasoline blendstock, diesel blendstock and heavies. The heavy fraction is assumed to be hydrocracked in a separate reactor that is described in the next section. The product yield strongly depends on process parameters such as hydrotreating conditions, catalyst selection and target product quality (density and oxygen content).

In the studies of bio-oil hydrodeoxygenation, various catalysts, different in active phases, promoters and supports, have been well documented (Wang 2013, Zacher 2013). Sulfided molybdenum based catalysts seem to be well-known and commonly used in the modern hydrotreating processes for bio-oil. Good oil yields and near complete deoxygenation have been achieved at PNNL with such catalysts (Elliott 2013). However, this type of catalyst is not as effective in the stabilization bed, which is designed to reduce the bio-oil reactivity under mild temperatures. Ruthenium-based catalysts work well in this service because they can effectively operate at low temperature to convert highly reactive carbonyl groups to less reactive species. However, the ruthenium-based catalysts may not be suitable for the 2nd stage hydrotreating reactor because they can promote methane production at higher temperatures (Elliott 2013). Final catalyst selections are yet to be determined. In this report, ruthenium-based catalysts on carbon supports are assumed for the stabilizer and the 1st bed, and a molybdenum based catalyst for the final bed. All catalysts are assumed to have a target 1-year life before complete replacement. Pre-sulfided catalysts for the 1st and 2nd stage hydrotreating are assumed. The naturally low sulfur content of wood derived pyrolysis oil may remove sulfur from the catalyst over time. A makeup stream of sulfiding agent, such as dimethyl sulfide, may be needed to maintain the catalysts in a sulfide state. Future research will help determine if this treatment is needed for effective deoxygenation.

3.3.1. Hydrotreating to Hydrocarbons Design Basis

The main process variables for Area 200 are shown in Table 8.

Table 8. Design Basis for Area 200

Reactor	Assumption
Stabilizer	
Temperature	284 °F (140 °C)
Pressure	1200 psia
Liquid hourly space velocity	0.5 volume/h/volume catalyst
1st Stage hydrotreater	
Temperature (maximum outlet)	356 °F (180 °C)
Pressure	2000 psia
Liquid hourly space velocity	0.5 volume/h/volume catalyst
2nd Stage hydrotreater	
Temperature (maximum outlet)	770 °F (410 °C)
Pressure	2000 psia
Hydrogen partial pressure at outlet	1300 psig
Liquid hourly space velocity	0.22 volume/h/volume catalyst
Lights removal column (T290)	
Number of stages	40
Tray efficiency	0.75
Top pressure	65 psia
Other metrics	
Total liquid feed to stabilizer, bbl/day (wet)	8,800
Chemical hydrogen consumption across all hydrotreaters, scf/bbl liquid feed	3,400
Chemical hydrogen consumption across all hydrotreaters, lb/100 lb dry pyrolysis oil	5.8
Carbon in hydrotreated oil/Carbon in pyrolysis oil	0.68
Hydrotreated product specific gravity	0.85

3.3.2. Hydrotreating Cost Estimation

The capital costs for this section are estimated using Aspen Capital Cost Estimator, and have been reviewed by several external reviewers with industry experience. The pressure swing adsorption (PSA) unit is from a vendor budgetary estimate. The stabilizer is a heat exchange reactor with catalyst in the tubes and steam on the shell side. This allows the raw pyrolysis oil to be heated to the reaction temperature in the presence of hydrogen and a catalyst, thus preventing coking. The first stage reactor is a single fixed bed vessel. The second stage hydrotreater is configured as four fixed bed reactors in parallel to reduce the reactor wall thickness. This proved to be less expensive than a single vessel at the given operating conditions. The estimated installation factor for the reactors is 2.75 to account for internals such as distributors and catalyst support trays. Table 9 shows the total purchased equipment cost (TPEC) and total installed capital cost (TIC) for the hydrotreating section.

Table 9. Hydrotreating Area Capital Costs

Scaling Stream: 167,600 gpd stable oil product

Equipment	**TPEC (2011 MM$)**	**TIC (2011 MM$)**
Reactors	32	88
Compression and pumps	8	9
PSA	1	4
Heat exchange	3	8
Columns and drums	4	6
Total	48	115

3.4. Area 300: Product Recovery and Area 500: Hydrocracking

After fractionating the hydrotreated bio-oil, a heavy fraction boiling at a temperature above the final boiling point of diesel is recovered. This heavy fraction is sent to the hydrocracking reactor to be catalytically cracked to additional fuel. The hydrocracking of a heavy fraction from lignocellulosic-based hydrocarbons has not yet been demonstrated as a process separate from hydrodeoxygenation. A simplified flow diagram of hydrocracking and product separation, which forms the basis for the current design, is illustrated in Figure 8.

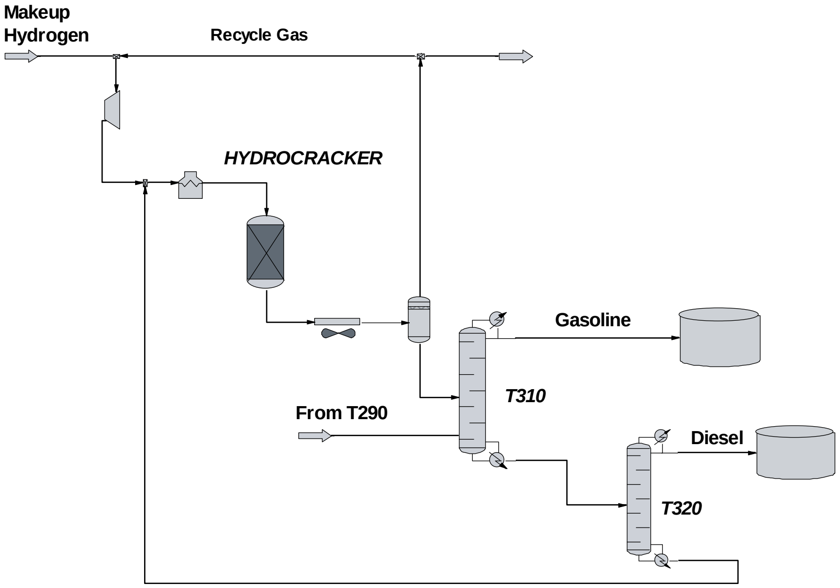


Figure 7. Hydrocracking and Product Recovery Process Flow

3.4.1. Hydrocracking and Product Separation Design Basis

The product from hydrocracking is a mixture of liquids spanning the gasoline and diesel range, and some byproduct gas. The gasoline and diesel range products are separated by distillation. These products are expected to be suitable for blending into finished fuel. More detailed characterization of these blendstocks are needed. The hydrocracking conditions are shown in Table 10.

Table 10. Areas 300 and 500 Design Basis

Reactor	Assumption
Hydrocracker	
Temperature	1380°F (750°C)
Pressure	1300 psia
Barrel/day feed	515
Naphtha splitter (T 310)	
Number of stages	30
Tray efficiency	0.75
Top pressure	25 psia
Diesel recovery column (T 320)	
Number of stages	30
Tray efficiency	0.75
Top pressure	15 psia
Other Metrics	
Gasoline + diesel specific gravity	0.79

3.4.2. Hydrocracking and Product Cost Estimation

The costs for the 550 bpd hydrocracker and associated equipment (hydrocracking reactor, heat exchangers, pumps, flash drums, and fired heater) were taken from literature (Marker 2005). The distillation columns and associated equipment (reboiler, condenser, condenser drum, and pumps) were estimated using Aspen Capital Cost Estimator.

Table 11. Hydrocracking and Product Separation Capital Costs

Scaling Stream: 550 BPSD feed		
Equipment	**TPEC (2011 MM$)**	**TIC (2011 MM$)**
Reactor & associated equipment	5	15.1
Columns and associated equipment	2.1	3.4

3.5. Area 600: Hydrogen Production

Hydrogen for hydroprocessing (hydrotreating and hydrocracking) is produced via steam reforming of process off-gases generated from fast pyrolysis and hydroprocessing. Additional natural gas is also used to obtain sufficient hydrogen production. For the base case in this study, a natural gas based hydrogen plant using a conventional multi-tube fired reactor with a commercial catalyst is assumed. A sensitivity case using a fluidized bed reformer system is also considered. Process steam is produced via heat integration with the hydrogen plant, and additional available steam is used to generate power on-site.

The base case hydrogen plant uses a conventional natural gas-based stream. An alternate case has also been developed that leverages the extensive work performed at NREL related to tar reforming of syngas. The base case is presented in this section, while the alternate scenario is presented as a sensitivity case in Section 5.1.

3.5.1. Base Case: Conventional Hydrogen Generation

Figure 9 shows the simplified flow scheme for hydrogen generation by steam reforming of natural gas (SRI International 2007, Meyers 2004, H2A 2013) and the off-gas streams from the fast pyrolysis, hydrotreating and hydrocracking processes. The off-gas by itself is insufficient to produce the amount of hydrogen required by the hydrotreaters and hydrocrackers; supplemental natural gas is required. Most of the off-gas from hydrotreating and hydrocracking is used to fire the reformer. However, a portion is compressed and mixed with pyrolysis off-gas and makeup natural gas, which is then hydrodesulfurized (HDS).

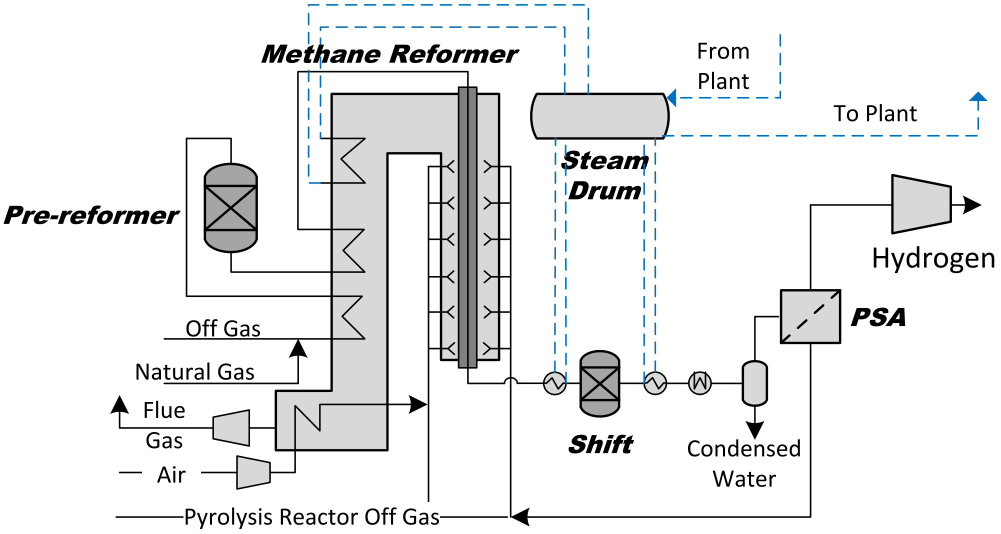


Figure 8. Hydrogen Production Process Flow

Hydrogen for the HDS unit is supplied by the off-gas stream. The gas exiting the HDS is then mixed with superheated steam and sent through an adiabatic pre-reformer to convert C_2+ compounds to methane prior to entering the main steam reformer to produce syngas. This reduces the rate of coking in the main reformer. The syngas hydrogen content is increased by high temperature shift. After condensing out the water, the hydrogen is purified by pressure swing adsorption (PSA). Off-gas from the PSA is recycled to the reformer burners.

Saturated steam and superheated steam are generated by recuperating heat from the reformer exhaust and cooling the product from the water gas shift reactor. The generated steam is used in the reformer as a reactant, and is also used for process heating, for example, in the distillation column reboilers.

3.5.2. Hydrogen Generation Design Basis

The design assumptions are shown in Table 12. Gibbs minimization reactors are used to model the pre-reforming, methane reforming and burner reactions. The reactor methane conversion of 80 mole % meets that reported by SRI (SRI International 2007).

Table 12. Area 600 Design Basis

Reactor	Assumption
Pre-reformer	
Inlet temperature	930°F (500°C)
Outlet pressure	344 psia
Steam-to-carbon ratio	2.4
Methane reformer	
Makeup natural gas	3.7 MMscf/day
Steam pressure	670 psia
Outlet temperature	1562°F (850°C)
Outlet pressure	314 psia
Burners	
Bridge wall temperature	1800°F (982°C)
Pressure	Slightly positive
Shift Reactor	
Outlet temperature	675°F (357°C)
Outlet pressure	300 psia
Approach to equilibrium	98%
PSA	
Hydrogen production rate	44.5 MM scf/day H_2
Hydrogen delivery pressure	283 psia
Hydrogen recovery	85%

3.5.3. Hydrogen Generation Cost Estimation

Capital costs for hydrogen generation are taken from the SRI 2007 Yearbook and scaled to the necessary hydrogen production rate using the SRI scale factor. The equipment includes a sulfur guard bed, pre-reformer, primary reformer with nickel catalyst, high temperature shift reactor, PSA unit, waste heat recovery producing high pressure steam and all associate outside battery limit equipment. The fixed capital investment for a 44.5 million scf/day hydrogen facility is $107 million dollars in cost year 2007. Converted to 2011 dollars and scaled to the base case hydrogen production, the capital cost at the project level (including direct and indirect costs) is $119 million dollars. This is more conservative than the capital cost from the H2A program developed for the DOE Hydrogen Program which estimated the cost at $96 million dollars (2007). However, the H2A configuration does not include a pre-reformer.

3.6. Balance of Plant

The balance of plant consists of the supporting systems for the plant: cooling water service, tank farm, wastewater treatment and flare.

Process Water

Cooling water usage is minimized through the use of air fin coolers where applicable. The process water demand is shown in Table 13. Most of the cooling tower water is used to indirectly cool the fast pyrolysis bio-oil that is recirculated in the quench system. The 2009 design case (Jones 2009) assumed that all indirect cooling of pyrolysis vapors was achieved with cooling water. For this report, cooling water demand is reduced through the combination of air cooling followed by trim cooling with water. This in turn reduces cooling water makeup.

Table 13. Process Water Demands

Source of Water Demand	**Model**
Cooling tower makeup, gpm	29
Steam reformer boiler feed water makeup, gpm	162
Total, gpm	191
Gallons water per gasoline gallon equivalent product	1.4

Power

Most of the electricity is purchase from the grid. However, some superheated steam is used as rotating equipment drivers.

Wastewater

Wastewater from the hydrotreaters typically contains less than 1 wt% carbon as shown in Figure 10. A sample of the aqueous phase from the hydrotreaters (2 wt% oxygen in organic product) at 0.55 wt% carbon was analyzed and found to have a COD of approximately 10 g/L. This carbon is likely to be mostly substituted phenolic compounds and the low concentration makes recovery uneconomical. A study by Tziotzios (2007) showed that waste water containing phenol with COD as high as 51 g/L can be effectively treated by aerobic digestion. The biochemical ethanol design report (Humbird 2011) also showed an example of using an aerobic digester to treat waste water with COD of 6 g/L before discharging the water to municipal sewer. Thus, wastewater from hydrotreating is assumed to be treated by aerobic digestion before discharge to a public water treatment facility. Capital costs for this system were taken from the original fast pyrolysis design report (Jones 2009), which compares well with scaled values from the Humbird report.

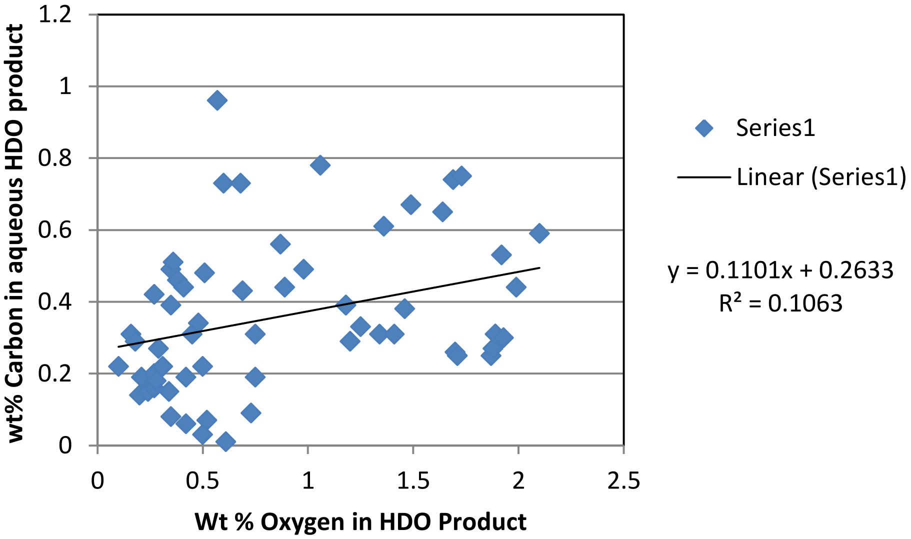


Figure 9. Carbon Content of HDO Aqueous Phase

4. Process Economics

Process economics are the combination of feedstock cost, and capital and operating costs associated with construction and operation of the conversion plant. These costs are combined in a discounted cash flow analysis to estimate the minimum fuel price needed to meet a 10% internal rate of return when the net present value is equal to zero.

4.1. Operating Costs

Table 14 lists the assumptions used to calculate the operating costs for the base case with the commercial natural gas based hydrogen plant with a pre-reformer.

Table 14. Variable Operating Cost Assumptions

Variable	**Value**	**Source**
Materials		
Feedstock, \$/dry ton	80	INL
Stabilizer & 1st Bed Catalyst, \$/lb (2011)	60	Calculated
2nd Bed Hydrotreating Catalyst , \$/lb (2007)	15.5	SRI 2007
Hydrocracking Catalyst, \$/lb (2007)	15.5	SRI 2007
Hydrogen Plant Catalysts, \$/1000 scf H_2 (2007)	3.6	SRI 2007
Sufiding Agent, \$/lb (2007)	0.799	SRI 2007
Utilities		
Natural Gas, \$/1000 scf	5.1	EIA 2011 Heywood 1998
CH_4, C_2H_6, N_2, C_3H_8, C_4H_{10}, CO_2, C_5H_{12}, C_6H_{14} LHV, BTU/lb	94%, 3.3%, 1.0%, 1.0% 0.35%, 0.3%, 0.04%, 0.01% 20,854	
Electricity, ¢/kwh	6.89	EIA 2011
Cooling tower water makeup (20 °F rise)	90 °F service 110 °F return	Assumption
Chemicals		
Boiler feed water makeup, ¢/1000 gallon	100.08	Aden 2007
Cooling tower makeup, ¢/1000 gallon	100.08	Aden 2007
Cooling tower chemicals, \$/lb	2.27	Humbird 2011
Boiler feed water chemicals, \$/lb	1.36	Humbird 2011
Ash Disposal, \$/ton	18	Phillips 2007
Wastewater disposal, \$/100 cubic feet	2.07	Aden 2007
Solids waste disposal, \$/ton	18	Aden 2007

Fixed costs are shown in Table 15. Salaries are taken from Dutta 2011 and converted to a 2011 dollar basis using US Bureau of Labor Statistics labor cost index.

Table 15. Fixed Operating Costs and Assumptions

Salaries

Position Title	**Salary (2011)**	**# Positions**	**Total Cost (2011)**
Plant Manager	161,400	1	161,400
Plant Engineer	76,800	1	76,800
Maintenance Super	62,600	1	62,600
Lab Manager	61,500	1	61,500
Shift Supervisor	52,700	5	263,400
Lab Technician	43,900	3	131,700
Maintenance Tech	43,900	16	702,5007
Shift Operators	52,700	40	2,108,000
Yard Employees	30,700	12	368,800
Clerks & Secretaries	39,500	3	118,600
Total		**83**	**4,055,000**

Other Fixed Costs

Cost Item	**Factor**	**Cost**
Benefits and general overhead	90% of total salaries	3,649,000
Maintenance	3% of fixed capital investment	21,006,000
Insurance and taxes	0.7% of fixed capital investment	4,901,000
Total Other Fixed Costs		**33,611,000**

4.2. Fixed Capital Investment

The following summarizes the purchased and installed equipment costs presented in Section 3. The capital costs are from various sources. The pyrolysis unit and associated equipment is derived from published costs for a commercial unit (Ensyn 2012). The hydrogen plant is also from published costs (SRI 2007). The majority of the remaining equipment was estimated using Aspen Capital Cost Estimator. Fixed capital costs are summarized in Table 16.

Table 16. Total Project Investment

Process Area	Total Capital	Fixed Capital Investment (2011 MM$)
100	Fast pyrolysis	279
100	Heat recovery & filtration	22
200	Hydrotreating	198
300-500	Product Finishing	32
600	Hydrogen generation	119
700	Balance of Plant	16
Total		**665**

4.3. Minimum Fuel Selling Price

The minimum fuel product selling price (MFSP) for the gasoline and diesel blendstock was determined using a discounted cash flow rate of return analysis. The methodology is identical to that used in Jones et al. (2009). The MFSP is the selling price of the fuel that makes the net present value of the process equal to zero with a 10% discounted cash flow rate of return over a 30 year plant life and 40% equity with the remainder debt financed at 8% interest for a 10 year term. The stream factor of 90% (7884 hours per year) is reasonable given the need for annual catalyst replacement. This results in a base case MFSP of $3.39/gasoline gallon equivalent (gge). This result is within the tolerance of the $3/gge programmatic target (DOE 2013). While two products are produced, (motor gasoline blendstock and diesel blendstock), they are combined and referred to as a “fuel product” as the production cost for both products are the same. A heating value of 116,090 BTU/gal is used to convert the heating value of the fuel products to a gasoline gallon equivalent basis. Section 1.2 gives the economic parameters used to calculate the MFSP. A sensitivity analysis was conducted to determine the effect of different financial and operating assumptions on the MFSP (Section 5.3).

5. Economic and Technical Sensitivities

The design case describes a single operating point for a standalone processing unit. This section investigates the production cost sensitivities to technical, financial, and market parameters. These include plant size, internal rate of return (IRR), feedstock costs, reactor conditions and product yields. Two important scenarios are highlighted: hydrogen generation methods and the effect of ash content on overall yields.

5.1. Alternate H_2 Production Scenario: Fluidized Bed Reformer

As an alternative to the conventional fixed bed reactor, steam reforming can also be carried out in a catalytic fluidized bed (CFB) reactor. The process flow diagram and heat and material balances for this alternative case of hydrogen production are given in Appendix D-1. The fluidized bed reformer is coupled with a second vessel where the catalyst is regenerated after being separated from the reformed gases. Since the reformer feed contains some of the components present in pyrolysis oil, the propensity for coking may be high. This process may be a way of mitigating any such problem. Carbon deposition on the reformer catalyst can be removed in the regenerator prior to recirculation and reused in the reformer. Consequently a pre-reformer is not necessary for this configuration.

The fluidized bed steam reformer design uses on a nickel-based catalyst developed at NREL for the reforming of biomass derived syngas containing tars and hydrocarbon species (Dutta 2011, Dutta 2013). Non-condensable gases from pyrolysis, off-gases from hydrotreating, and hydrocracking, and supplemental natural gas are combined to meet the process hydrogen requirements. In addition to the reforming reactions to produce CO and H_2 from hydrocarbons and oxygenated species, the nickel catalyst also maintains significant water gas shift activity. The overall reforming process is endothermic. Heat produced via oxidation of deposited carbon on the catalyst and combustion of supplemental process off-gases in the regenerator elevates the catalyst temperature. The hot regenerated catalyst supplies the heat required for reforming when transferred to the reformer. Further details of the reformer configuration can be found in US Patent 8,241,523 B2 (Apanel 2012).

Hot flue gases and reformed gases are used to preheat the reformer inlet and combustion air, and to generate steam. The design temperature for the reformer outlet is 1670°F (910°C). A molar steam-to-carbon ratio (SCR) of 4 was used to ensure high hydrogen efficiency.

High temperature shift (HTS) and low temperature shift (LTS) reactors downstream of the reformer are used to further shift CO and H_2O into H_2 and CO_2, without the addition of more steam during these stages. The process gas stream leaves the reformer at 1,670°F (910°C) and is cooled to 662°F (350°C) for HTS. The HTS and LTS are modeled as fixed bed equilibrium reactors using approach temperatures of 35°F and 20°F (19°C and 11°C), respectively (Spath 2005). The gas exiting the HTS reactor is cooled to 392°F (200°C) prior to entering the LTS reactor. The LTS reactor effluent process gas is cooled and process condensate is separated. Most of the process condensate is recycled to the reformer after blowdown. The gas is then pressurized to 350 psia and sent to a pressure swing adsorption (PSA) system.

The PSA unit is used to separate the hydrogen from the other components in the shifted gas stream, mainly CO_2 and other unreacted hydrocarbons. A hydrogen recovery rate of 85% is used in this design. Purge gas from the PSA system is sent to the catalyst regenerator (reformer furnace) to meet the fuel requirement. The product hydrogen at the PSA outlet is at 77 psia and 283°F (139°C) and requires further compression prior to use in the Area 200 stabilizer reactor.

Two steam levels are used. Low pressure process steam (at 74 psia and 307°F) supplying the reformer uses mostly recycled process condensate. Higher pressure (675 psia and 499°F) steam is generated using clean boiler feed water. High pressure steam is used for process heating and excess high pressure steam is superheated to 1000°F (538°C) for power generation. The heat duty available for steam generation is mainly from the reformer section. Additional heat duty from pyrolysis and bio-oil upgrading areas are also included for heat integration. A pinch analysis was performed to validate the steam production based on the available total hot duty and cold duty.

5.1.1. Fluidized Bed Reformer Design Basis

The fluidized bed reformer is designed to produce 9,841 lb/hr (44.5 MMscf/d) of pure hydrogen at 77 psia and 283°F (139°C) to meet hydrotreater demands. Table 17 summarizes the conversion assumptions at the fluidized steam reformer. These conversions were incorporated into the Aspen Plus model to simulate expected steam reformer performance.

Table 17. Design Performance for the Reforming Catalyst (Dutta 2011, Dutta 2013)

Compound	Target Conversion
Methane (CH_4)	80%
Ethane (C_2H_6)	99%
Ethylene (C_2H_4)	90%
Benzene (C_6H_6)	99%
Ammonia (NH_3)	90%
All others*	90%

* Includes oxygenated species present in reformer feed.

5.1.2. Fluidized Bed Reformer Cost Estimate

Table D-1 presents the equipment list and cost estimates for hydrogen production using a fluidized bed reformer system. All costs are projected in 2011 U.S. dollars. Capital costs were developed from a variety of sources. Capital costs for some common equipment items (e.g., tanks, pumps, vessels) are based on a recent NREL study (Dutta 2011). Cost of the fluidized bed reformer package including a dual fluidized bed loop, cyclones and ancillaries were estimated based on a prior cost estimation study (Dutta 2011, Worley 2012). Costs associated with HTS, LTS, and PSA were based on Spath (2005). Equipment costs were scaled based on material and energy balances from process simulation. Equipment costs obtained in earlier or later years are inflated or deflated to a 2011 U.S. dollars basis using the Chemical Engineering Plant Cost Indices (CEI 2013). Operating costs and utility consumption are shown in Appendix Tables D-2 and D-3, respectively.

Table 18. Comparison of Hydrogen Generation Scenarios

Cost and Consumption	Base Case with Conventional Hydrogen Generation	Base Case with CFB Hydrogen Generation
Natural gas usage, scf/gge	19.33	15.33
H_2 demand MMscfd	44.5	44.5
Electricity demand, kWh/gge	1.22	1.45
Capital, million $	665	691
MFSP, $/gge	3.35	3.45

The results are essentially identical from a MFSP standpoint and within the expected uncertainty of this work. The somewhat higher capital for the CFB scenario is mitigated by lower natural gas usage. The higher SCR ratios, higher reforming temperatures and lower pressures used in the CFB reactor allows greater conversion of methane than in the base case. Methane conversion higher than the 80% value shown in Table 17 may be possible in the CFB reformer (Dutta 2013) resulting in improved economics. Also, the addition of the LTS reactor improves the overall conversion of hydrocarbons to hydrogen. However, additional compression is needed as compared to the base case because the reformer operates at near atmospheric pressure. Additional capital cost and stream flow information for this scenario can be found in Appendix D-2.

5.2. High Ash Feedstock Scenario

The base case assumes an ash content of 0.9 wt%. This deep degree of de-ashing may not be possible as discussed in Section 3. This sensitivity considers the effects of 1.9% ash on the production costs assuming yield effects derived from Oasmaa et al. (2010) and reported in Table 3 of this report. Table 19 shows the sensitivity to the various assumptions used in the model. The lower fuel yield is somewhat offset by lower capital and lower natural gas usage. The natural gas usage is reduced as the ash content increases because significantly more off-gas is generated in the fast pyrolysis section and this gas can be used for hydrogen generation. This scenario is of interest because more biomass is available at lower cost if the ash content is increased.

Table 19. Effect of Higher Ash Biomass

Cost and Consumption	Base Case with 0.9wt% Ash	Base Case with 1.9wt% Ash
Fuel yield, gal/dry ton biomass	84	75
Natural gas usage, scf/gal	19.3	5.9
H_2 demand, MMscfd	44.5	40
TCI, million $	700	672
MFSP, $/gge	3.39	3.55

The effect of ash on pyrolysis organic yield is a combination of the amount of ash, and the type of reactive species in that ash. Further work is needed to quantify this effect. Additionally, a possible outcome that is not captured in this scenario is increased ash level in the bio-oil that is carried over into downstream equipment and the subsequent effect on catalyst maintenance.

5.3. Additional Technical and Financial Sensitivities

Figure 11 shows the sensitivity to the various market and financial assumptions used in the model.

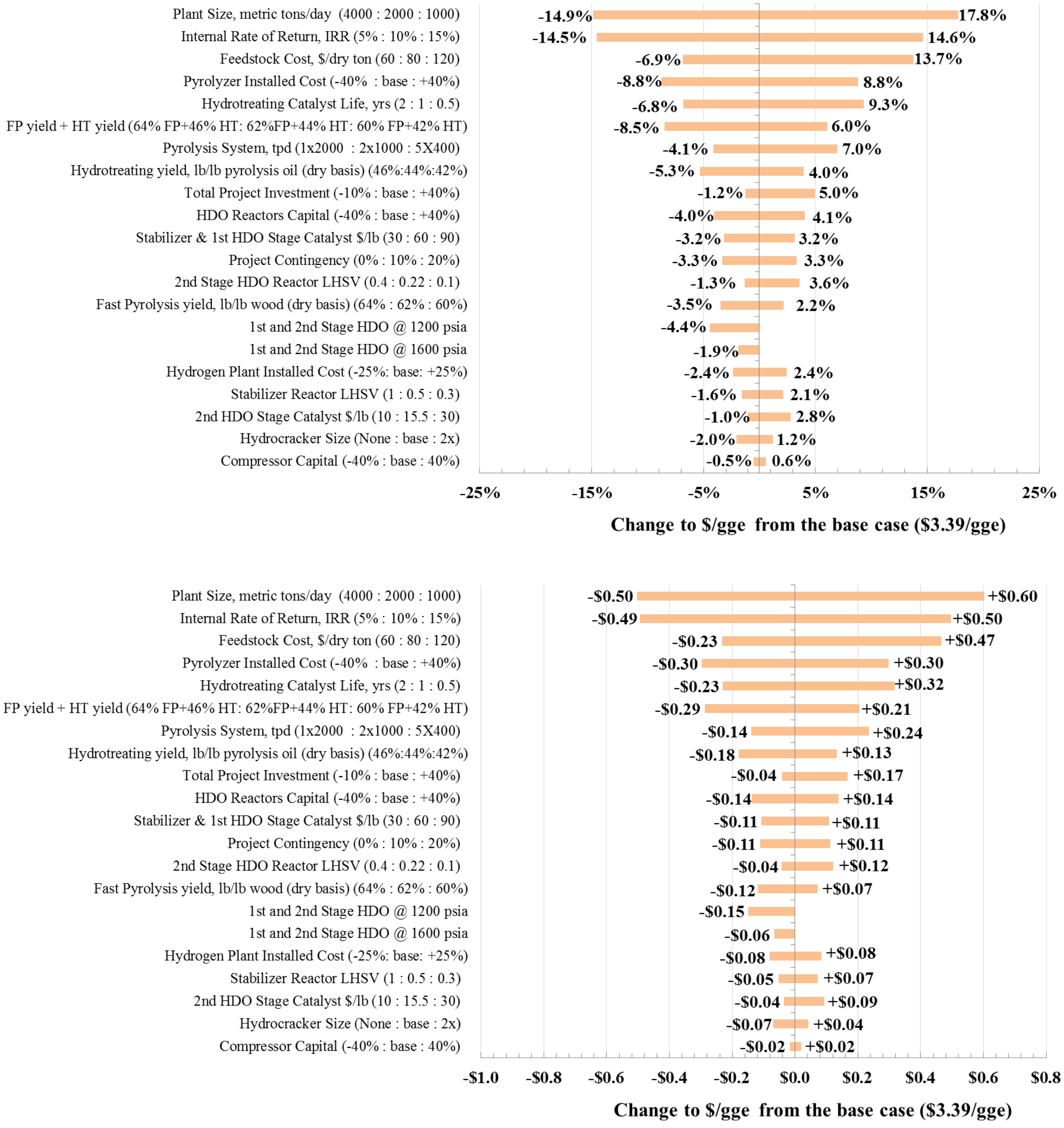


Figure 10. Sensitivity Analysis Results

Plant scale, capital investment and IRR are the largest market and financial type effects. Feedstock costs are the next largest contributor. Use of a low ash feedstock such as pine (at approximately $120/ton) increases the production cost by 14%, thus highlighting the need for blended feedstocks using lower cost materials.

Product yield is one of the most significant cost parameters. In the base case, pyrolysis oil yield and hydrotreating oil yields are 64% wt per dry wood and 44% per dry pyrolysis oil, respectively. Reduction of pyrolysis oil yield to 60% and hydrotreating oil yield to 44% will increase the selling price by 6%.

The published capital costs for the pyrolysis system (reactor, combustor, cyclones and quench equipment) are reported as ± 40% and this results in a 9% variation in the MFSP. Hydrotreater capital costs are a function of the operating temperature, pressure and liquid hourly space velocity (LHSV). The temperature and pressure set the wall thickness and the LHSV determines the vessel capacity. Reducing the first and second stage reactors pressure to 1200 psi and doubling the space velocity for the second stage results in a 7% reduction in the reactor costs and also reduces compression requirements.

Hydrotreating catalysts are also important to the overall economics of the process. The stabilizer and 1st stage reactor may use a precious metal catalyst. The annual catalyst cost for these reactors is a function of metal type, metal loading and replacement rate. The base case catalyst is assumed to be $60/lb. This price will vary with the cost of the support and the ruthernium loading. Current costs for ruthernium are at $70 per troy ounce (PGM 2013). A 50% variation in the price causes a 3% variation in the MFSP. The 2nd stage reactor may be able to use a nickel or cobalt type catalyst which is expected to be less expensive. However, this reactor also has the largest catalyst inventory, thus doubling the catalyst costs also causes a 3% increase in catalyst cost.

The time on-stream before catalyst replacement is needed has a very large effect. The base case catalyst replacement rate is assumed to be 1 year for each reactor. Increasing this to two years causes the MFSP to drop by 7%. Reducing the catalyst life to six months increases the MFSP by 9%.

6. Environmental Sustainability Metrics

In addition to setting technical and economic targets for the conversion pathways included in the MYPP (DOE 2013), BETO is working towards setting baselines and targets for environmental sustainability metrics. Sustainability is a cross-cutting element of the BETO program whose overarching goal is to "understand and promote the positive economic, social, and environmental effects and reduce the potential negative impacts of biofuels production activities" (DOE 2013). A specific goal of the sustainability element is to identify metrics and set targets for at least one conversion pathway by 2013. To reach this goal, BETO, DOE National Laboratories and others have worked together to develop an initial set of important sustainability metrics for the conversion stage of the biofuel life cycle: greenhouse gas (GHG) emissions, fossil energy consumption, fuel yield, biomass carbon-to-fuel efficiency, water consumption, and wastewater generation. Shown in Table 20 are the estimated metric values for the conversion plant for the updated design base case (low ash, conventional hydrogen generation).

Table 20. Sustainability Metrics for Fast Pyrolysis and Upgrading

Sustainability Metric	2017 Projected
GHGs (g CO_2-e/MJ fuel) – (fossil emissions; biogenic emissions)[1]	18.9; 85
Fossil Energy Consumption (MJ fossil energy/MJ fuel product)[2]	0.301
Total Fuel Yield (gal/dry ton wood; gge/dry ton wood)	84; 87
Biomass Carbon-to-Fuel Efficiency (C in fuel/C in biomass)	47%
Water Consumption (m^3/day; gal/GGE)[3]	1050; 1.4
Wastewater Generation (m^3/day; gal/GGE)[4]	932; 1.3

[1] Biogenic emissions include those contained in the char combustor exhaust, the heat from which is used in the biomass dryer (not part of the conversion plant).
[2] Fossil energy consumption does not include grinding of the feedstock prior to the pyrolysis step.
[3] Water consumption and wastewater generation include only direct use/emissions and do not include water associated with upstream production of materials and energy used at the plant.
[4] Wastewater generation includes both wastewater from hydrotreating and blowdown from the cooling towers

The GHG and fossil energy consumption metrics include both direct effects at the plant and "upstream" effects associated with the production and distribution of materials and energy for plant operations; i.e., these are the lifecycle emissions and energy usage for the conversion stage of the fuel supply chain. Water consumption and wastewater generation values in Table 20 consider only direct water inputs and wastewater generation at the plant and thus do not include water consumed or generated during production and distribution of materials and energy (e.g., thermo-electric power) used at the plant.

Metrics for other important air emissions from the conversion facility, such as criteria air pollutants (e.g., NO_x, SO_x, particulate matter, and VOCs) have not yet been estimated due to a lack of data on formation of these components during conversion processes. Through separate BETO funding, efforts are currently underway to more accurately model and measure criteria air pollutant emissions from the biorefinery, specifically those resulting from pyrolysis char and off-gas combustion at the plant. The following sections provide a more detailed description of the sustainability metrics and a discussion of the results.

6.1. Greenhouse Gas Emissions and Fossil Energy Use

BETO is developing technologies that will facilitate the volumetric requirements of the national Renewable Fuel Standard (RFS2), as legislated by the Energy Independence and Security Act (EISA) of 2007. Reducing fossil energy use and GHG emissions is integral to the RFS2. In order to be eligible for the RFS2 program, fuels made from renewable biomass must meet the renewable fuels categories definitions, which includes having lifecycle GHG emissions that are 50% (advanced biofuel) or 60% (cellulosic biofuel) less than the petroleum baseline. While RFS2 mandates the EPA to conduct its own life cycle analysis (LCA) to determine fuel qualification, it is essential that lifecycle analysis be performed during the development of these pathways in order to predict and facilitate improvement of environmental performance. In many ways, this approach is analogous to that of techno-economic analyses, allowing continual evaluation and improvement of the design throughout the technology research and development phase.

Conversion processes generate fossil-based GHG emissions stemming from fossil fuel consumption at the plant (e.g., natural gas) and from upstream production and distribution of materials and energy that are used at the plant (e.g., natural gas, grid electricity, chemicals, catalyst). Conversion GHGs consist primarily of CO_2 (95%), with the remainder being methane released during natural gas production and distribution. Conversion processes also result in direct biogenic CO_2 emissions from pyrolysis, char combustion, and oil upgrading processes, as well as off-gas reforming and combustion. The design is heat integrated and therefore any light gases that are produced at the plant are completely combusted for heat. Apart from land use change impacts for certain feedstocks, biomass is generally regarded as a carbon-neutral feedstock (i.e., carbon absorbed during growth equals carbon emitted during processing and final use), and therefore fossil energy use and emissions receive the majority of attention.

Metrics for conversion GHGs and fossil energy use are based on material and energy balances from the process models, along with information from the Ecoinvent database (2011) and U.S. LCI database (LCI 2012). Emissions and energy consumption for the U.S. average grid mix of electricity is used. The lifecycle modeling software, SimaPro (2011), is used to estimate the cumulative GHG emissions and fossil energy usage. Due to a lack of available data on catalyst manufacture, recycling and reclamation processes, this component is approximated with a zeolite product from the Ecoinvent database. Future work on development of energy and material inventory data for the catalysts is necessary to fill this data gap. It is possible that catalyst will not contribute significantly if target economic catalyst lifetimes are achieved (i.e., at least 1 year). However, estimation of the catalyst impact is necessary to be able to assert this assumption. Also, the use of precious metal or platinum group metals for catalysts may result in significantly different GHG and other metric values compared to NiMo or CoMo catalysts.

Figure 12 shows the contributing factors to the total fossil GHGs for the conversion plant. As shown, natural gas use accounts for 57% of the total emissions. Using more renewable sources of hydrogen (e.g., pyrolysis oil or adjusting gas production in pyrolysis to meet hydrogen needs) may reduce the carbon footprint for conversion, however, the technical feasibility of these options as well as the impact on economics needs to be determined.

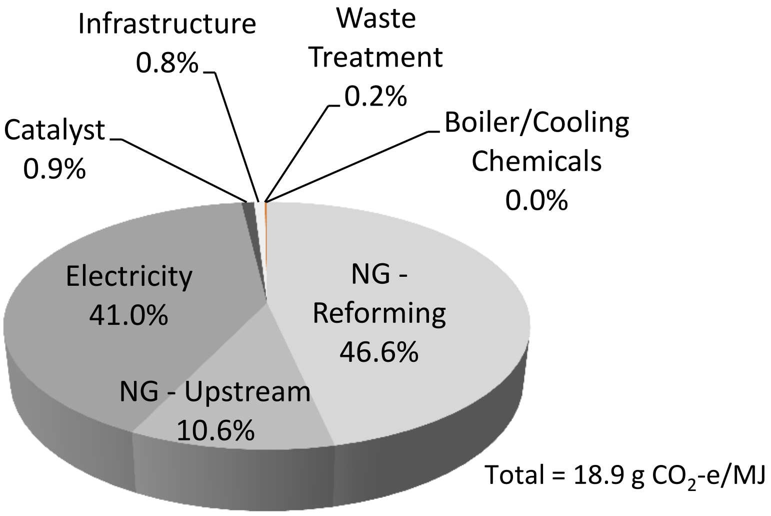


Figure 11. Contributing Processes to Conversion GHGs for Fast Pyrolysis and Upgrading

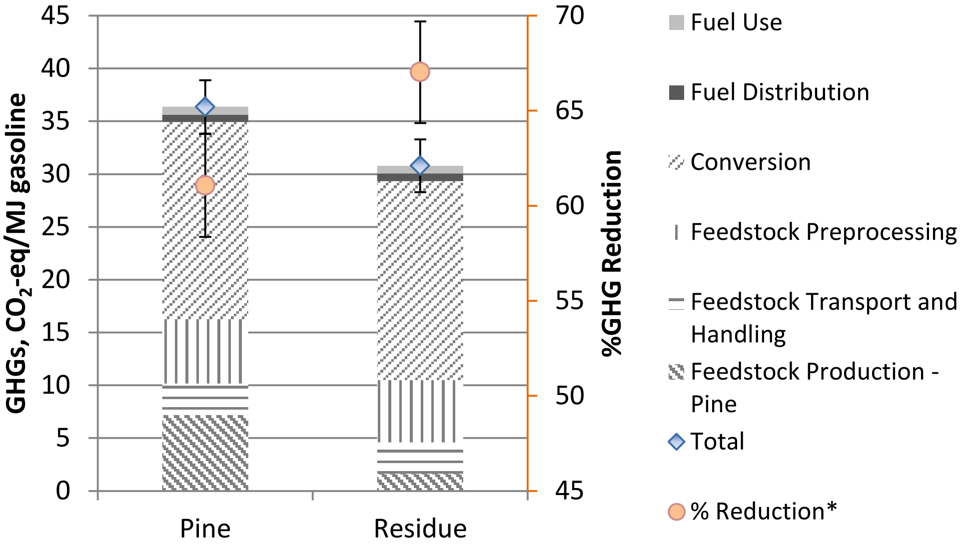
Using the conversion GHGs, the full life cycle GHG emissions were estimated for gasoline fuel from woody feedstocks using fast pyrolysis and upgrading, considering both forest residues and purpose grown southern pine feedstocks. The inventory data sources and primary assumptions for feedstock production and logistics and fuel distribution and consumption are listed in Table 21. It is important to note that the current pathway configuration assumes that the preprocessing operations are located in close proximity to the biorefinery and as such, the heat contained in the pyrolysis char combustor exhaust is used to dry the feedstock (i.e., no additional fossil energy source is used for feedstock drying). In an advanced uniform format logistics scenario where biomass is collected at regional depots for blending, biomass drying energy would need to be supplied separately. Energy consumption for hot water de-ashing is based on conventional processes used in the food industry. As the specifics of this process are highly uncertain at this time, a range of values is considered. Water makeup needed for the de-ashing process is not estimated. Direct and indirect land use change impacts of pine production are not included in the analysis.

Table 21. Data Sources and Assumptions for Fuel Life Cycle GHG Analysis

Life Cycle Process	Reference/Assumption		
Forest residue collection	U.S. LCI Database (LCI 2012)		
Forest residue logistics	Muth et al. (2013)		
Purpose grown pine production	Perlack et al. (2011); lime application from Tyree et al. (2006)		
Pine harvesting and logistics	Muth et al. (2013)		
Feedstock transportation distance	50 miles		
Wood de-ashing	<u>Low</u>	<u>Median</u>	<u>High</u>
Water circulation, gal/dry ton	450	2725	5000
Natural gas heat[1], MMBtu/dry ton	0.08	0.46	0.83
Fuel distribution and combustion	GREET model (ANL 2012)		

[1] Heat required for raising water 20 °F above ambient temperature.

Figure 13 shows the full life cycle GHGs for gasoline from fast pyrolysis and upgrading of woody feedstocks. The uncertainty bars around the total GHGs and the percent reduction estimates correspond to the range of energy values considered for the wood de-ashing process. The fuel production stage is the most GHG-intensive stage of the fuel cycle, comprising 53% and 62% of the total emissions for fuels from pine and forest residue, respectively. Total emissions for southern pine based gasoline range from 33.8 to 38.9 g CO_2-eq/MJ gasoline, equating to a reduction in GHGs of 58-64% as compared to the petroleum baseline (GREET, 93.4 g/MJ). Total emissions for forest residue based gasoline range from 28.3 to 33.3 gCO_2-eq/MJ gasoline, equating to a reduction of 64-70% as compared to the petroleum baseline. These results suggest this fuel is in the range of the EISA RFS cellulosic biofuel definition (60% GHG reduction), however, final qualification is made by the EPA based on their own analysis and determination. In addition, through collaboration with Argonne National Laboratory, energy and materials inventory from this work will be integrated into the GREET model for further analyses.



***GHG reductions relative to GREET 2005 gasoline baseline, 93.4 g CO_2-e/MJ. Qualification for RFS is ultimately determined by the EPA.**

Figure 12. Life Cycle GHG Emissions for Fuel from Fast Pyrolysis and Bio-oil Upgrading of Southern Pine and Forest Residue

(Uncertainty bars reflect range for wood de-ashing process energy, see Table 21).

6.2 Fuel Yield and Biomass Carbon-to-Fuel Efficiency

An overarching goal of BETO is to enable technologies that produce transportation biofuels in a sustainable way. Fuel yield and biomass carbon-to-fuel efficiency are important measures of natural resource utilization and are inherent to biofuel sustainability. With respect to GHGs, there is an important tradeoff that exists between the fuel yield and natural gas requirement for this pathway and others that similarly produce intermediate products requiring hydrogen-based upgrading. This tradeoff stems from the fact that as fuel yields from bio-oil hydrotreating increase, there is less off-gas available for hydrogen generation and thus, more natural gas is needed for the plant. Fuel yield and biomass carbon-to-fuel efficiency are measures of how efficient the technology is at producing liquid fuel. Therefore, when comparing fuel options, it is important to present these metrics alongside GHGs to provide a more comprehensive representation of the overall performance of the conversion process and the balance between these metrics. For example, a conversion process may have very low (or even negative) GHGs, but this may be due to low yields and a large portion of its biomass carbon going to electricity production (producing a large GHG credit).

6.3 Water Consumption and Wastewater Generation

Protection and conservation of water resources is a global concern as the potential impacts of climate change, growing population, and energy demand become increasingly evident. Energy production is a leading consumer of water within the industrial sector and the effect of biofuel production on water resources is an important sustainability concern. Consumptive water use associated with fast pyrolysis and upgrading consists of makeup water for the steam and cooling systems. Boiler feed water makeup is needed to replace steam consumed in the hydrogen plant. Fresh cooling water is needed to make-up for losses at the cooling tower (e. g, evaporation and drift) and blowdown for maintenance of the recirculating cooling system. Figure 14 shows the contributing processes to overall direct consumptive water use for the biorefinery.

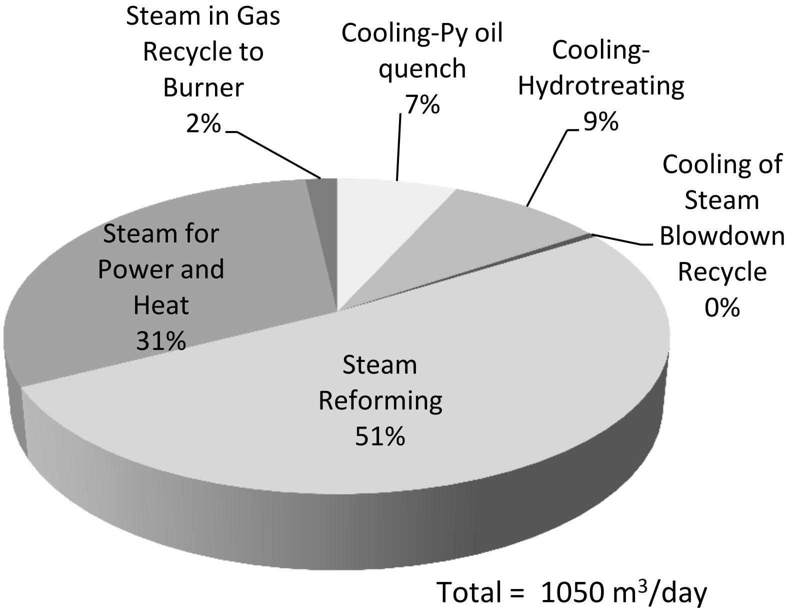


Figure 13. Contributing Processes for Consumptive Water Use for Fast Pyrolysis and Upgrading

Water consumed for production of electricity and materials used at the biorefinery (e.g., electricity, natural gas, and catalyst) and for life cycle stages upstream and downstream of conversion is not included in the analysis. As shown, boiler feed water makeup for steam production is approximately 84% of the total water consumed for the conversion processes. Approximately 60% of the steam is consumed in the reforming and water gas shift reactions for hydrogen production. About a third is used to produce power and a small amount is needed to make up for water lost in the burner off-gas feed. Cooling water makeup is approximately 16% of the total consumptive water use, which is needed primarily for the pyrolysis oil vapor quench and cooling of stabilized pyrolysis oil. It is important to note that steam drum blowdown is assumed to be recycled to the cooling loop, significantly reducing raw cooling water makeup. While the water consumption values presented in Table 20 are likely in the approximate range of an actual biorefinery using this technology, there is much uncertainty around the quench process, specifically the flow rate of oil needed to provide the direct quench and the associated temperature. In addition, bio-oil yield and composition (and associated enthalpies), fluidized gas rate for the pyrolysis reactor, and quench temperature are all inter-related variables and can have a significant effect on cooling water consumption for the quench process.

Wastewater quantity and composition are important sustainability metrics, as treated facility wastewater is ultimately discharged to streams or lakes in the region and both of these factors can affect the health of local water resources. In addition, wastewater treatment, whether located onsite or offsite at a publicly owned treatment works, or both, as is assumed in this report, requires additional energy and materials. The primary wastewater streams generated from fast pyrolysis and bio-oil upgrading processes are cooling tower blowdown and water separated from the stabilized bio-oil after hydrotreating. Water separated from the hydrotreated bio-oil stream originates from biomass moisture introduced into the system and water produced during the pyrolysis and hydrotreating reactions. This stream accounts for 60-65% of the total wastewater generated for the conversion process. Cooling system blowdown accounts for the remainder of the total plant wastewater. Efforts are underway to better characterize the hydrotreater wastewater.

6.4 Metrics for High Ash Content Scenario

Conversion stage sustainability metrics for the high ash content scenario (see Section 5.2) are presented and compared to the base case (low ash) in Table 22.

Table 22. Effect of Higher Ash on Sustainability Metrics for Fast Pyrolysis and Upgrading.

Sustainability Metric	0.9% Ash (Base)	1.9% Ash
GHGs (g CO_2-e/MJ fuel: fossil emissions; biogenic emissions)[1]	18.9; 85	11.5; 101
Fossil Energy Consumption (MJ fossil energy/MJ fuel product)[2]	0.301	0.159
Total Fuel Yield (gal/dry ton wood; gge/dry ton wood)	84; 87	75; 79
Biomass Carbon-to-Fuel Efficiency (C in fuel/C in biomass)	47%	42%
Water Consumption (m^3/day; gal/GGE)[3]	1050; 1.4	983; 1.5
Wastewater Generation (m^3/day; gal/GGE)[4]	932; 1.3	852; 1.3

[1] Biogenic emissions include those contained in the char combustor exhaust, the heat from which is used in the biomass dryer (not part of the conversion plant).
[2] Fossil energy consumption does not include grinding of the feedstock prior to the pyrolysis step.
[3] Water consumption and wastewater generation include only direct use/emissions and do not include water associated with upstream production of materials and energy used at the plant.
[4] Wastewater generation includes both wastewater from hydrotreating and blowdown from the cooling towers.

Fossil GHGs and energy consumption for the high ash case are lower than the base case by 40% and 53%, respectively. Fuel yield for the high ash case is lower than the base case by 11%. As discussed in Section 3.1, higher ash content in the feedstock can cause an increase in gas production at the expense of condensable liquid (see Table 3) from fast pyrolysis. With the assumed increase in off-gas production to the hydrogen plant, less natural gas is needed, leading to lower fossil GHGs for the conversion plant, as illustrated in Figure 15. The tradeoff between fuel yield and natural gas-based GHGs for this pathway has been shown previously (Jones 2013) and is an important consideration in designing fuels that are optimized for both economic and environmental goals

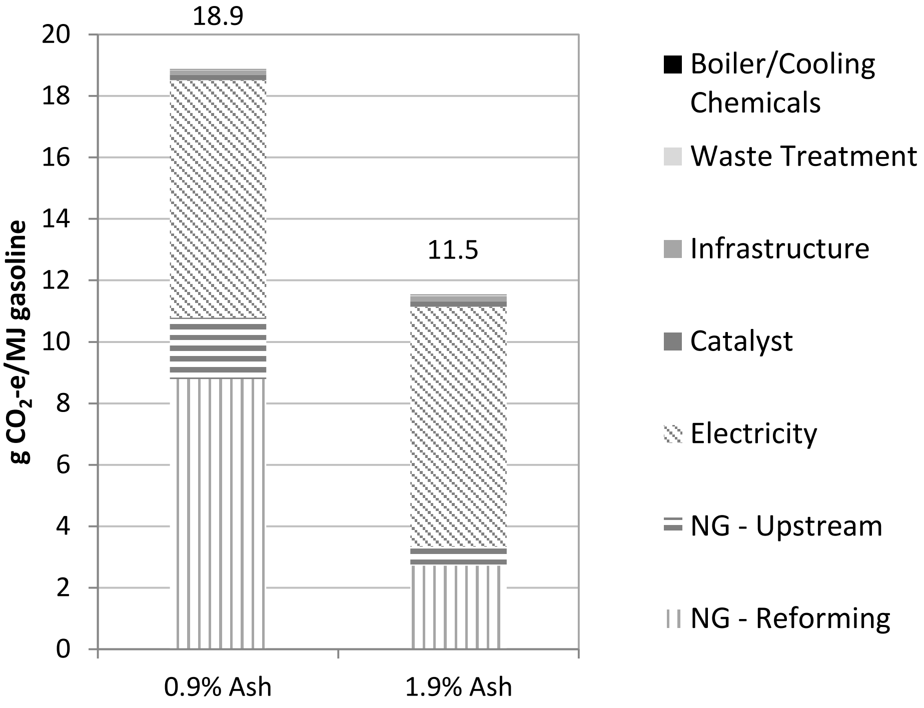


Figure 14. Effect of Biomass Ash Content on Conversion GHGs

Cooling water flowrate is lower for the high ash case because there is less pyrolysis oil product to condense than in the base case. Boiler feedwater flowrate is higher for the high ash case than the base case, with total water consumption for the two cases being equal. Lower fuel yields for the high ash case results in higher water consumption on a gallon-per-gallon basis. Daily wastewater generation is lower for the high ash case because less pyrolysis oil is being hydrotreated than in the base case, however, on a gallon-per-gallon basis, the two cases are equivalent.

Figure 16 compares the total lifecycle GHGs for southern pine based fuel for the base case and high ash case. Total emissions for the high ash case are 27.3 gCO_2-eq/MJ gasoline, corresponding to a 71% reduction from the GREET 2005 petroleum baseline. As shown, while lower yields reduce conversion GHGs (fossil), there is also an opposite but lesser effect on feedstock production GHGs. It is also important to note that because de-ashing is not used for the high ash case, feedstock preprocessing GHGs are lower than the base case.

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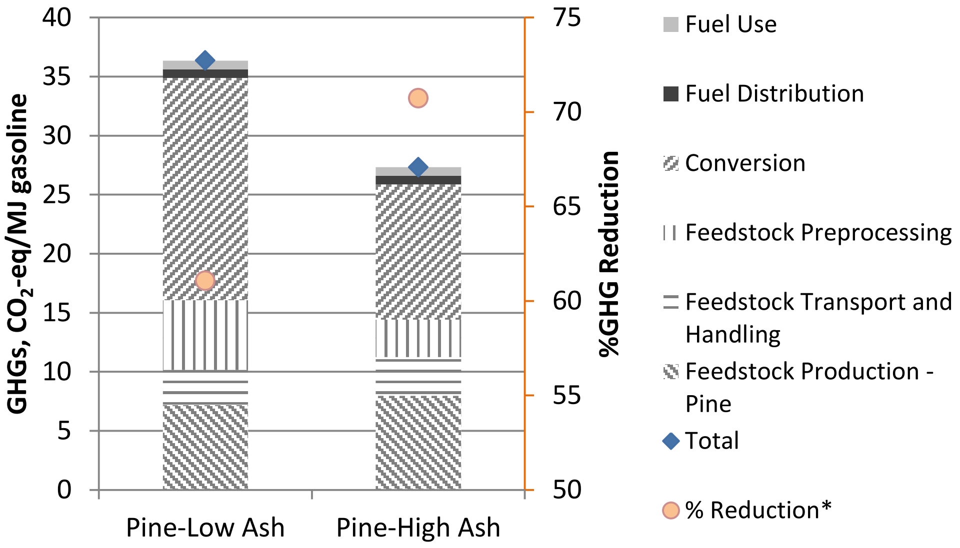

***GHG reductions relative to GREET 2005 gasoline baseline, 93.4 g CO_2-e/MJ. Qualification for RFS is ultimately determined by the EPA.**

Figure 15. Effect of Biomass Ash Content on Lifecycle GHGs (for pine feedstock)

7. Conclusions and Recommendations

This analysis shows that biomass derived liquid transportation fuels have the potential to be competitive with their fossil derived equivalents. However, there are a number of gaps that need to be addressed by future research for this pathway to be realized. Progress towards their resolution will be tracked through the annual state of technology assessments.

There is a need for a better understanding of feedstock quality and how it affects the conversion processes. It is known that ash content and volatile carbon content can have adverse effects on the yield of pyrolysis oil. Furthermore, it is not total ash content but rather the speciation and transport of the elemental ash components that must be understood in order to enable the concomitant relationship between feedstock characteristics at low cost and the optimal conversion process. An interesting technology for conversion to aid with the ash issue will be hot filtration. Ash carried with the bio-oil can also affect the life and performance of the upgrading catalysts. Research is needed to investigate mixed biomass feedstocks on the quality and yields of pyrolysis oil and upgraded oil. It would also be useful to consider blends with non-edible oil crops and to investigate biomass thermal pretreatment (time and temperature) to improve oil quality (torrefaction is just one example of this type of treatment).

Understanding the effects of catalyst use during fast pyrolysis will be the subject of another design report in FY14. It is mentioned here, because mild to medium conversion in this step could improve the life of the downstream upgrading catalysts and improve economics by requiring less severe hydrotreating and possible less expensive metallurgy as well as elimination of the stabilizer bed. An important part of this will be to understand how to capture small carbon fragments (C_4 minus) before they are hydrogenated to hydrocarbons, and how to convert them into fuel range molecules.

The upgrading steps are expensive. Eliminating one of the reactor steps and operating at pressures below 2000 psia will reduce the capital costs. Research is needed to understand what the true operating pressure limit is. More reactive catalysts are needed to allow increased space velocity and increased activity at lower temperatures. Also, catalyst life is still relatively short relative to the target life of one year. It is known that there is a loss of activity in the upgrading catalysts. Mechanisms of deactivation (carbon occlusion, migration of inorganic species) remain an opportunity waiting to be understood in order to accelerate the development of performance enhanced robust catalysts. Strategies are needed to minimize catalyst deactivation and/or develop regeneration methods. Supports such as alumina may not perform well in an aqueous environment at the given operating temperatures. While carbon supports are better able to handle aqueous environments, they also pose regeneration challenges if the means of deactivation is carbon polymerization. A better understanding of reactor heat management may also improve catalyst life and provide insight into reactor scaling.

Product quality requires further investigation. There is a need to better understand how to co-produce a jet cut that does not add additional processing steps and does not reduce the quality of the gasoline and diesel boiling range products. It is difficult to simultaneously make good octane gasoline and good cetane diesel without additional processing. Additional intermediate and final product testing is needed for the following reasons:

- Current product analysis is limited to density, ultimate analysis, Total Acid Number (TAN), distillation curves, and GC/MS analysis. There are limitations associated with each of these tests. For example, TAN analysis results include acids and phenols. Are the phenols a corrosion concern? GC/MS only identifies compounds that can be volatized.
- Analyses are currently conducted on the whole hydrotreated oil product. Similar analysis should be obtained for each product cut.
- Characterization is needed of the heavier than diesel portion that is not properly analyzed by GC/MS. Conradson Carbon Residue analysis, for example, may provide an indication of coking tendencies.
- Fuel standard analyses such as flash, octane, cetane, freeze, smoke etc. will allow determination of how to meet product blendstock specifications.

Lastly, working with refiners is key to understanding blending issues and to developing standards for fuels and fuel intermediates that could be finished in a refinery.

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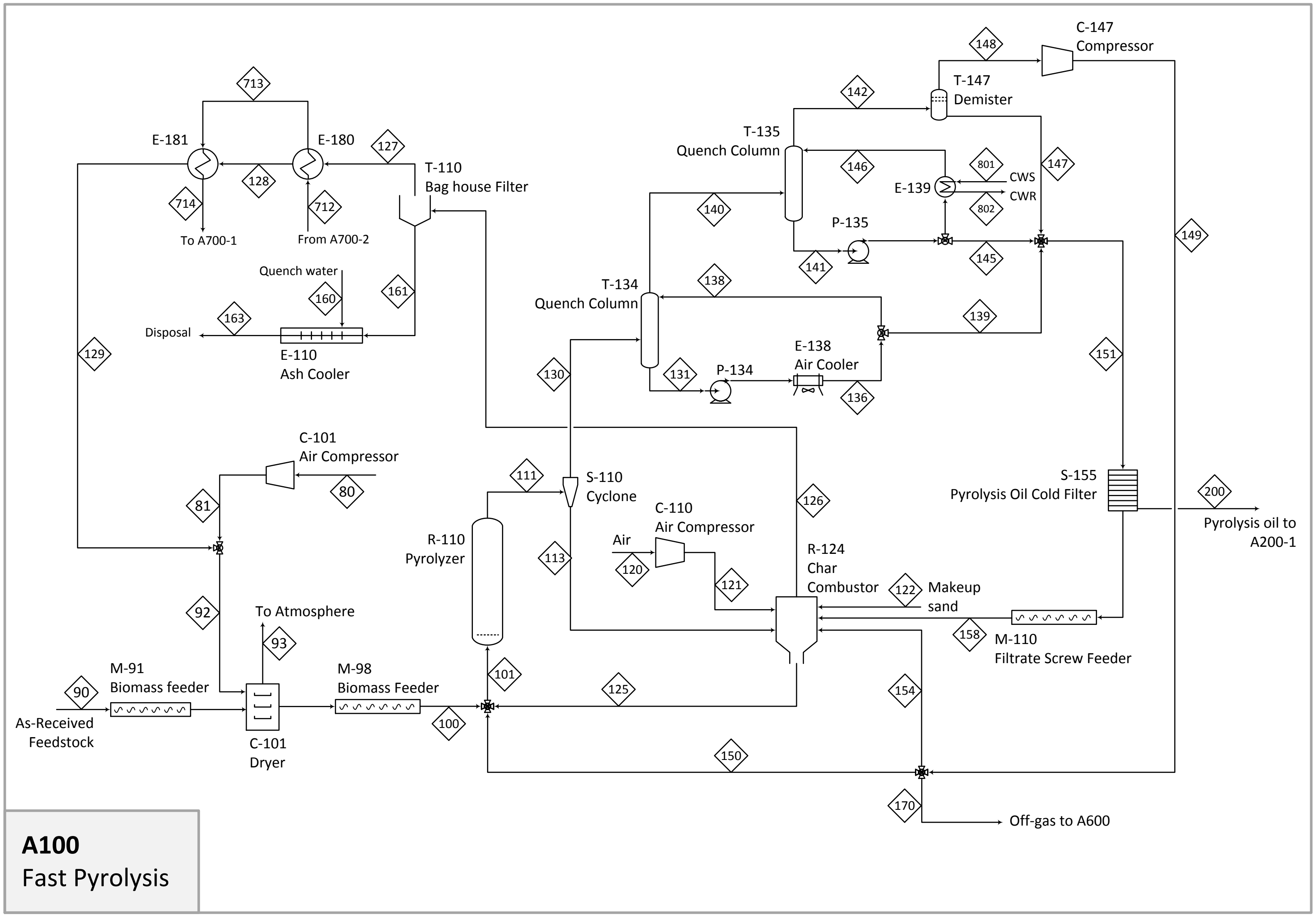
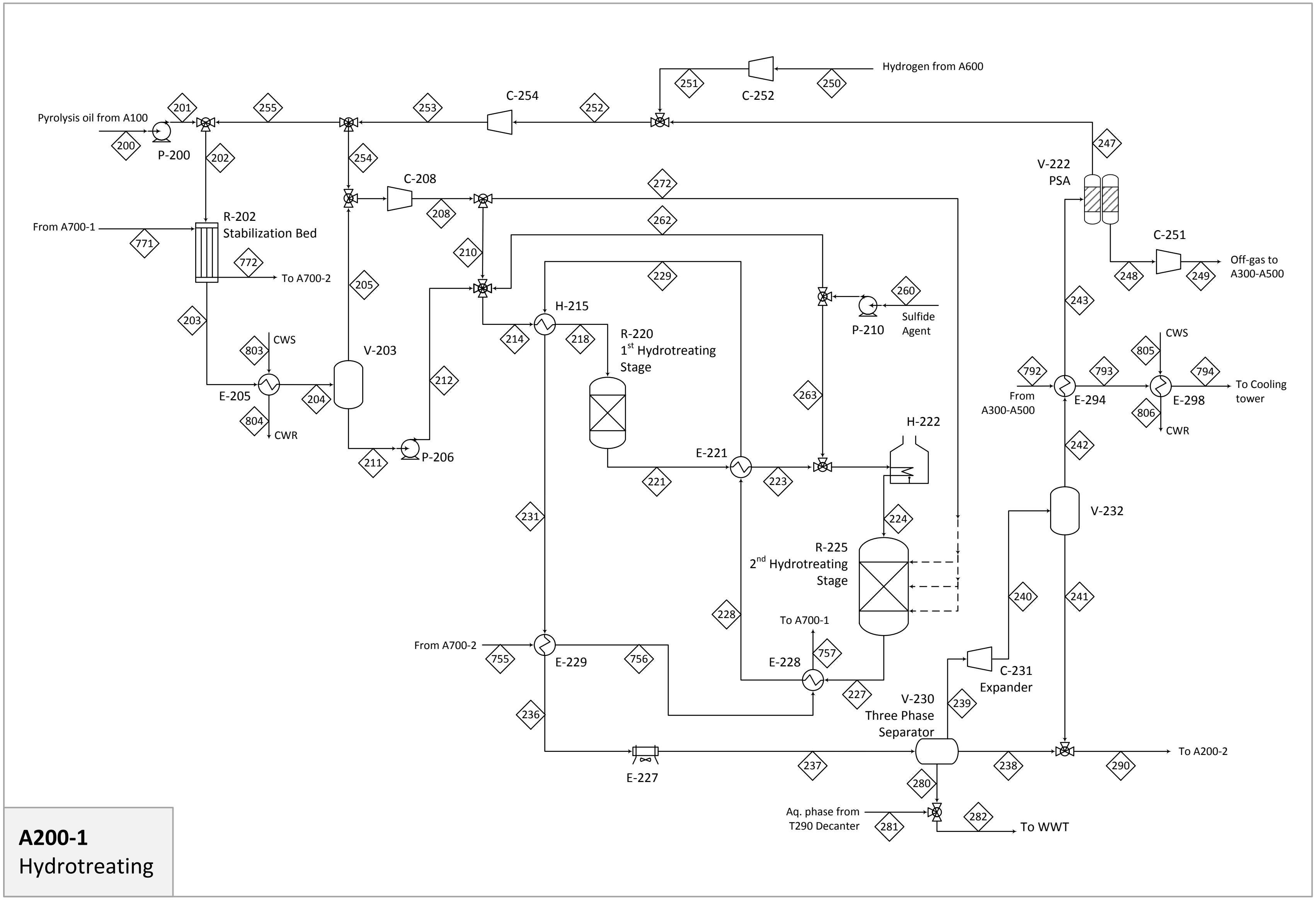
Appendix A: Base Case Heat and Material Balances



Stream No.	80	81	90	92	93	100	101	111	113	120	121	122	125	126	127	128	129	130	131	136	138	139
Temp F	90	118	68	584	161	160	931	813	813	90	163	60	1128	1128	1128	1048	800	813	158	140	140	140
Pres psia	14.7	16.7	14.7	16.7	15.7	20.7	20.7	17.7	17.7	15.0	20.7	17.7	20.7	20.7	20.7	20.7	20.7	17.7	17.7	25.0	25.0	25.0
Vapor mole frac.	1.00	1.00	0.00	1.00	1.00	0.00	1.00	1.00	1.00	1.00	1.00	0.00	0.00	1.00	1.00	1.00	1.00	1.00	0.00	0.00	0.00	0.00
Total lb/h	190000	190000	262454	572662	626345	204130	3419818	3419818	2686288	347400	347400	285	2664538	384637	382662	382662	382662	733530	2.012E+07	2.012E+07	2.000E+07	119715
Flow rates in lb/h																						
Wood	0	0	182028	0	0	182028	182028	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Water	3719	3719	78736	19234	72918	20412	22449	44495	0	6800	6800	0	0	15515	15515	15515	15515	44495	1102290	1102290	1095731	6559
Hydrogen	0	0	0	0	0	0	178	555	369	0	0	0	0	0	0	0	0	185	0	0	0	0
Oxygen	43108	43108	0	59168	59168	0	0	1446	1446	78820	78820	0	0	16060	16060	16060	16060	0	0	0	0	0
Nitrogen	140688	140688	0	398228	398228	0	0	303	303	257237	257237	0	0	257540	257540	257540	257540	0	0	0	0	0
Argon	2397	2397	0	6780	6780	0	0	0	0	4383	4383	0	0	4383	4383	4383	4383	0	0	0	0	0
Carbon monoxide	0	0	0	0	0	0	253324	263429	0	0	0	0	0	0	0	0	0	263429	550	550	547	3
Carbon dioxide	88	88	0	89151	89151	0	238706	248267	0	160	160	0	0	89063	89063	89063	89063	248267	5869	5869	5834	35
Methane	0	0	0	0	0	0	33628	34970	0	0	0	0	0	0	0	0	0	34970	104	104	104	1
Ethane	0	0	0	0	0	0	4630	4815	0	0	0	0	0	0	0	0	0	4815	75	75	74	0
Propane	0	0	0	0	0	0	114	119	0	0	0	0	0	0	0	0	0	119	4	4	4	0
N-butane	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethene	0	0	0	0	0	0	10631	11056	0	0	0	0	0	0	0	0	0	11056	168	168	167	1
Propene	0	0	0	0	0	0	7407	7704	0	0	0	0	0	0	0	0	0	7704	259	259	258	2
1-Butene	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Sulfur dioxide	0	0	0	100	100	0	0	0	0	0	0	0	0	100	100	100	100	0	0	0	0	0
Nitrogen dioxide	0	0	0	1	1	0	0	0	0	0	0	0	0	1	1	1	1	0	0	0	0	0
Acids	0	0	0	0	0	0	108	6574	0	0	0	0	0	0	0	0	0	6574	927661	927661	922141	5520
Aldehydes	0	0	0	0	0	0	0	8231	0	0	0	0	0	0	0	0	0	8231	1382390	1382390	1374165	8225
Ketones	0	0	0	0	0	0	388	9794	0	0	0	0	0	0	0	0	0	9794	1023661	1023661	1017570	6091
Alcohols	0	0	0	0	0	0	0	8231	0	0	0	0	0	0	0	0	0	8231	1383206	1383206	1374976	8230
Guaiacols	0	0	0	0	0	0	0	5879	0	0	0	0	0	0	0	0	0	5879	986741	986741	980869	5871
Low MW sugars	0	0	0	0	0	0	0	7055	0	0	0	0	0	0	0	0	0	7055	1185642	1185642	1178587	7055
High MW sugars	0	0	0	0	0	0	0	39977	0	0	0	0	0	0	0	0	0	39977	6718603	6718603	6678626	39977
Extractives	0	0	0	0	0	0	0	3527	0	0	0	0	0	0	0	0	0	3527	592490	592490	588964	3525
Low MW lignin derived A	0	0	0	0	0	0	0	12934	0	0	0	0	0	0	0	0	0	12934	2173475	2173475	2160543	12933
Low MW lignin derived B	0	0	0	0	0	0	0	2606	0	0	0	0	0	0	0	0	0	2606	349377	349377	347298	2079
High MW lignin derived A	0	0	0	0	0	0	0	10804	0	0	0	0	0	0	0	0	0	10804	1815687	1815687	1804883	10804
High MW lignin derived B	0	0	0	0	0	0	0	2352	0	0	0	0	0	0	0	0	0	2352	395212	395212	392861	2352
Nitrogen compounds	0	0	0	0	0	0	0	82	0	0	0	0	0	0	0	0	0	82	1667	1667	1657	10
Sulfur compounds	0	0	0	0	0	0	0	29	0	0	0	0	0	0	0	0	0	29	4891	4891	4862	29
Sulfur	0	0	0	0	0	0	0	50	50	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	18306	18306	0	0	0	0	0	0	0	0	0	0	0	0	0
Ash	0	0	1690	0	0	1690	1690	1690	1561	0	0	0	0	1690	0	0	0	130	21773	21773	21644	130
Sand	0	0	0	0	0	0	2664538	2664538	2664253	0	0	285	2664538	285	0	0	0	285	47925	47925	47640	285

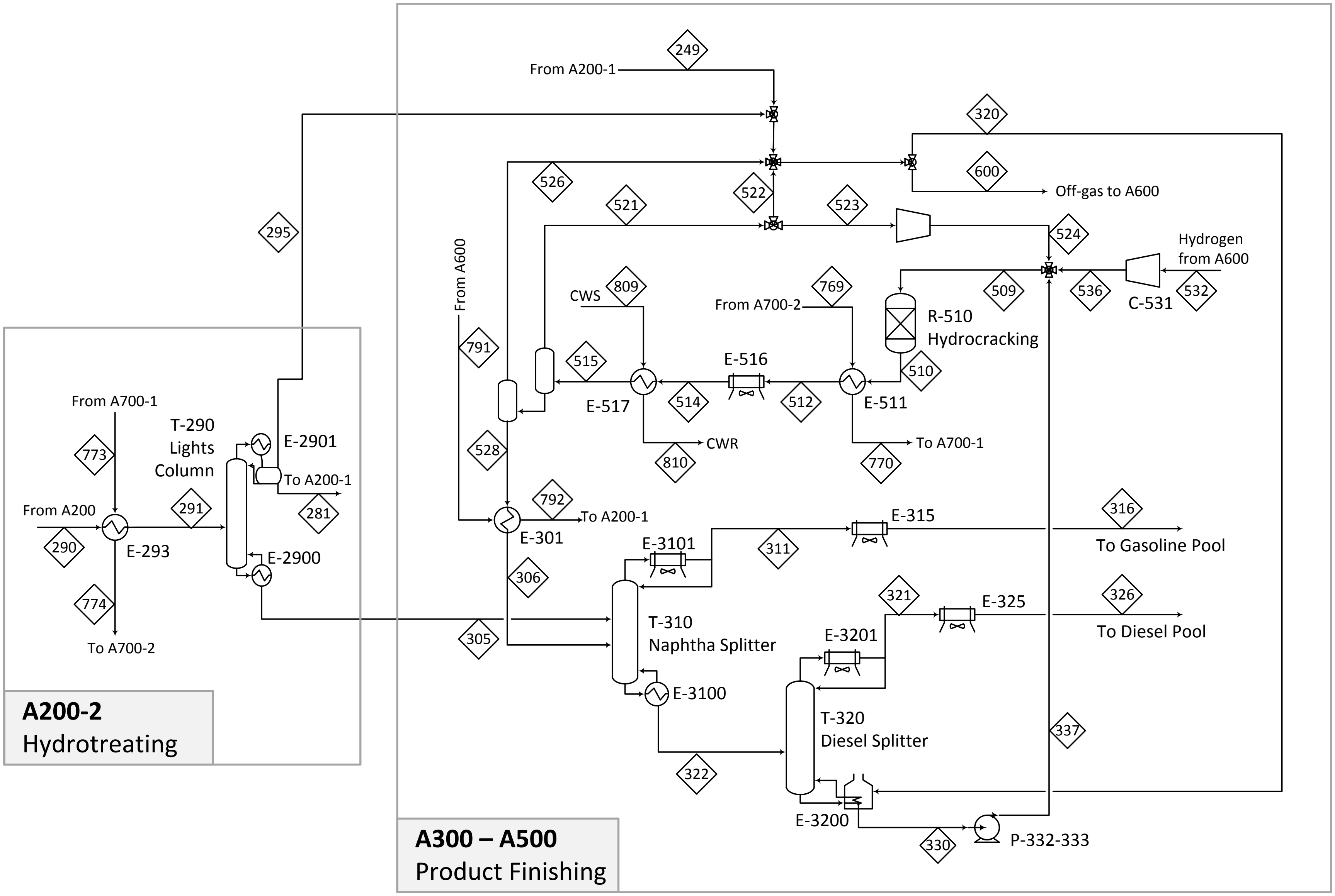
Stream No.	140	141	142	145	146	147	148	149	150	151	154	158	160	161	163	170	200	712	713	714	801	802
Temp F	158	113	113	113	110	113	113	162	162	130	162	130	60	211	211	162	130	283	498	501	90	110
Pres psia	18	18	18	35	30	18	18	23	23	18	23	18	18	18	18	23	18	685	680	675	60	55
Vapor mole frac.	1.00	0.00	1.00	0.00	0.00	0.00	1.00	1.00	1.00	0.00	1.00	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00	1.00	0.00	0.00
Total lb/h	613814	8.015E+06	598591	15222	8.000E+06	25463	573128	573128	551150	160401	9900	5302	2690	4665	4665	12078	155099	36600	36600	36600	1.060E+06	1.060E+06
Flow rates in lb/h																						
Wood	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Water	37936	6034727	26475	11461	6023266	24357	2118	2118	2037	42377	37	1294	2690	2690	2690	45	41082	36600	36600	36600	1059526	1059526
Hydrogen	185	0	185	0	0	0	185	185	178	0	3	0	0	0	0	4	0	0	0	0	0	0
Oxygen	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Nitrogen	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Argon	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbon monoxide	263426	186	263426	0	185	0	263426	263426	253324	4	4550	0	0	0	0	5551	4	0	0	0	0	0
Carbon dioxide	248233	3997	248225	8	3989	0	248225	248225	238706	43	4288	1	0	0	0	5231	41	0	0	0	0	0
Methane	34970	69	34969	0	69	0	34969	34969	33628	1	604	0	0	0	0	737	1	0	0	0	0	0
Ethane	4814	48	4814	0	48	0	4814	4814	4630	1	83	0	0	0	0	101	1	0	0	0	0	0
Propane	119	2	119	0	2	0	119	119	114	0	2	0	0	0	0	2	0	0	0	0	0	0
N-butane	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethene	11055	125	11055	0	125	0	11055	11055	10631	1	191	0	0	0	0	233	1	0	0	0	0	0
Propene	7703	122	7703	0	122	0	7703	7703	7407	2	133	0	0	0	0	162	2	0	0	0	0	0
1-Butene	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Sulfur dioxide	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Nitrogen dioxide	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Acids	1055	437572	224	831	436741	112	112	112	108	6463	2	197	0	0	0	2	6265	0	0	0	0	0
Aldehydes	5	2653	0	5	2648	0	0	0	0	8230	0	251	0	0	0	0	7979	0	0	0	0	0
Ketones	3703	1525333	806	2897	1522436	403	403	403	388	9391	7	287	0	0	0	8	9104	0	0	0	0	0
Alcohols	0	152	0	0	152	0	0	0	0	8231	0	251	0	0	0	0	7979	0	0	0	0	0
Guaiacols	8	4057	0	8	4049	0	0	0	0	5879	0	180	0	0	0	0	5699	0	0	0	0	0
Low MW sugars	0	0	0	0	0	0	0	0	0	7055	0	215	0	0	0	0	6839	0	0	0	0	0
High MW sugars	0	0	0	0	0	0	0	0	0	39977	0	1221	0	0	0	0	38756	0	0	0	0	0
Extractives	2	745	1	1	744	1	0	0	0	3527	0	108	0	0	0	0	3420	0	0	0	0	0
Low MW lignin derived A	1	641	0	1	640	0	0	0	0	12934	0	395	0	0	0	0	12539	0	0	0	0	0
Low MW lignin derived B	527	4531	519	9	4523	519	0	0	0	2606	0	80	0	0	0	0	2527	0	0	0	0	0
High MW lignin derived A	0	0	0	0	0	0	0	0	0	10804	0	330	0	0	0	0	10474	0	0	0	0	0
High MW lignin derived B	0	0	0	0	0	0	0	0	0	2352	0	72	0	0	0	0	2280	0	0	0	0	0
Nitrogen compounds	72	196	72	0	195	72	0	0	0	82	0	3	0	0	0	0	80	0	0	0	0	0
Sulfur compounds	0	67	0	0	67	0	0	0	0	29	0	1	0	0	0	0	28	0	0	0	0	0
Sulfur	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbon	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ash	0	0	0	0	0	0	0	0	0	130	0	130	0	1690	1690	0	0	0	0	0	0	0
Sand	0	0	0	0	0	0	0	0	0	285	0	285	0	285	285	0	0	0	0	0	0	0



A200-1
Hydrotreating

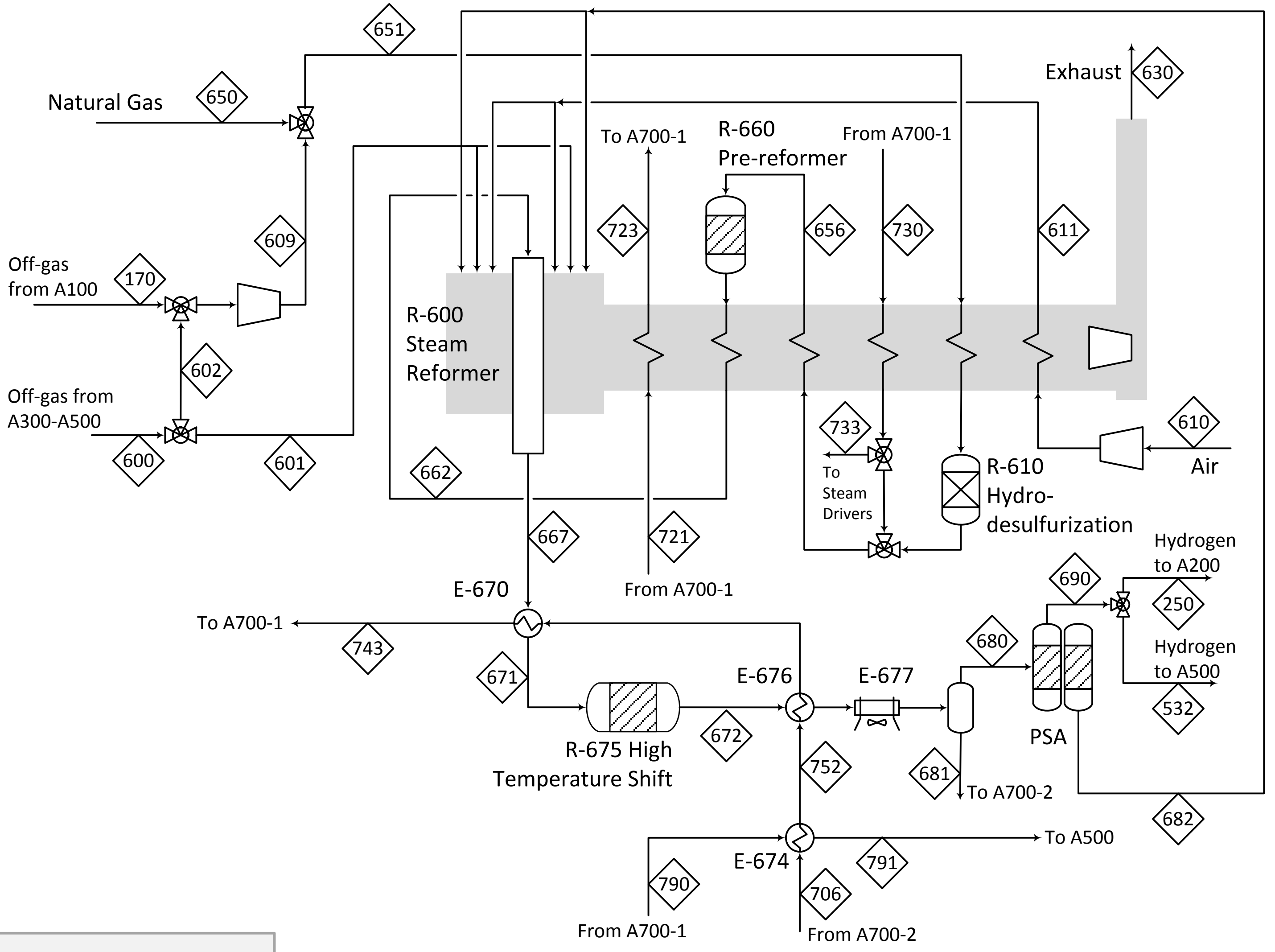
Stream No.	200	201	202	203	204	205	208	210	211	212	214	218	221	223	224	227	228	229	231	236	237	238	239	240	241	242
Temp F	130	131	132	284	105	105	150	150	105	106	115	300	348	550	550	767	580	400	352	300	150	150	150	51	51	51
Pres psia	18	1200	1200	1190	1185	1185	2065	2065	1185	2070	2060	2055	2030	2025	2010	1990	1990	1985	1985	1980	1975	1980	1980	755	755	755
Vapor mole frac.	0.00	0.00	0.53	0.55	0.53	1.00	1.00	1.00	0.00	0.00	0.59	0.61	0.63	0.90	0.90	1.00	1.00	0.91	0.83	0.78	0.74	0.00	1.00	1.00	0.00	1.00
Total lb/h	155099	155099	161729	161729	161729	6767	28453	8453	154962	154962	163428	163428	163428	163428	163466	183466	183466	183466	183466	183466	183466	53017	53505	53505	333	53172
Flow rates in lb/h																										
Water	41082	41082	41082	41082	41082	67	67	20	41016	41016	41036	41036	44690	44690	44690	77292	77292	77292	77292	77292	77292	66	397	397	331	66
Hydrogen	0	0	6630	6535	6535	6498	28185	8373	37	37	8410	8410	8153	8153	8153	21652	21652	21652	21652	21652	21652	4	21634	21634	0	21634
Carbon monoxide	4	4	4	4	4	3	3	1	0	0	1	1	1	1	1	624	624	624	624	624	624	0	624	624	0	624
Carbon dioxide	41	41	41	66	66	59	59	17	7	7	25	25	1237	1237	1237	12539	12539	12539	12539	12539	12539	22	12496	12496	1	12495
Methane	1	1	1	135	135	133	133	39	2	2	41	41	522	522	522	5965	5965	5965	5965	5965	5965	1	5961	5961	0	5961
Ethane	1	1	1	1	1	0	0	0	0	0	0	0	889	889	889	5075	5075	5075	5075	5075	5075	1	5074	5074	0	5074
Propane	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3017	3017	3017	3017	3017	3017	596	2401	2401	0	2401
N-butane	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3024	3024	3024	3024	3024	3024	1259	1752	1752	0	1752
N-pentane	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3982	3982	3982	3982	3982	3982	2623	1351	1351	0	1350
Ethene	1	1	1	1	1	1	1	0	0	0	0	0	0	0	0	1	1	1	1	1	1	0	1	1	0	1
Propene	2	2	2	2	2	1	1	0	0	0	1	1	1	1	1	2	2	2	2	2	2	0	1	1	0	1
Hydrogen sulfide	0	0	0	0	0	0	0	0	0	0	0	0	7	7	7	30	30	30	30	30	30	4	17	17	0	17
Acids	6265	6265	6265	5021	5021	1	1	0	5020	5020	5020	5020	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Aldehydes	7979	7979	7979	7979	7979	0	0	0	7979	7979	7979	7979	7979	7979	7979	0	0	0	0	0	0	0	0	0	0	0
Ketones	9104	9104	9104	7304	7304	0	0	0	7304	7304	7304	7304	0	0	0	1	1	1	1	1	1	0	0	0	0	0
Alcohols	7979	7979	7979	7979	7979	0	0	0	7979	7979	7979	7979	7979	7979	7979	0	0	0	0	0	0	0	0	0	0	0
Guaiacols	5699	5699	5699	5699	5699	0	0	0	5699	5699	5699	5699	5699	5699	5699	0	0	0	0	0	0	0	0	0	0	0
Low MW sugars	6839	6839	6839	6839	6839	0	0	0	6839	6839	6839	6839	0	0	0	0	0	0	0	0	0	0	0	0	0	0
High MW sugars	38756	38756	38756	38756	38756	0	0	0	38756	38756	38756	38756	38756	38756	38756	0	0	0	0	0	0	0	0	0	0	0
Extractives	3420	3420	3420	3249	3249	0	0	0	3249	3249	3249	3249	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Low MW lignin derived A	12539	12539	12539	11536	11536	0	0	0	11536	11536	11536	11536	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Low MW lignin derived B	2527	2527	2527	2527	2527	2	2	0	2525	2525	2525	2525	2525	2525	2525	1	1	1	1	1	1	1	0	0	0	0
High MW lignin derived A	10474	10474	10474	10474	10474	0	0	0	10474	10474	10474	10474	7855	7855	7855	0	0	0	0	0	0	0	0	0	0	0
High MW lignin derived B	2280	2280	2280	2280	2280	0	0	0	2280	2280	2280	2280	2280	2280	2280	0	0	0	0	0	0	0	0	0	0	0
Nitrogen compounds	80	80	80	80	80	1	1	0	79	79	79	79	79	79	79	80	80	80	80	80	80	79	0	0	0	0
Sulfur compounds	28	28	28	28	28	0	0	0	28	28	28	28	28	28	28	13	13	13	13	13	13	13	0	0	0	0
Acids intermediates	0	0	0	1274	1274	0	0	0	1273	1273	1274	1274	5637	5637	5637	0	0	0	0	0	0	0	0	0	0	0
Carbonyl intermediates	0	0	0	1848	1848	0	0	0	1848	1848	1848	1848	3492	3492	3492	0	0	0	0	0	0	0	0	0	0	0
Extractive intermediates	0	0	0	146	146	0	0	0	146	146	146	146	829	829	829	0	0	0	0	0	0	0	0	0	0	0
Low MW lignin compound intermediates	0	0	0	886	886	0	0	0	886	886	886	886	9681	9681	9681	0	0	0	0	0	0	0	0	0	0	0
High MW lignin compound intermediates	0	0	0	0	0	0	0	0	0	0	0	0	4302	4302	4302	0	0	0	0	0	0	0	0	0	0	0
Sugar intermediate A	0	0	0	0	0	0	0	0	0	0	0	0	2955	2955	2955	0	0	0	0	0	0	0	0	0	0	0
Sugar intermediate B	0	0	0	0	0	0	0	0	0	0	0	0	3390	3390	3390	0	0	0	0	0	0	0	0	0	0	0
Short chain n-paraffins	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	6400	6400	6400	6400	6400	6400	5361	1033	1033	0	1033
Medium chain n-paraffins	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3080	3080	3080	3080	3080	3080	3078	2	2	0	2
Iso-paraffin A	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2148	2148	2148	2148	2148	2148	1964	184	184	0	184
Iso-paraffin B	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3115	3115	3115	3115	3115	3115	3096	19	19	0	19
Cyclo C5 A	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2027	2027	2027	2027	2027	2027	1913	114	114	0	114
Cyclo C5 B	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2618	2618	2618	2618	2618	2618	2533	85	85	0	85
Cyclo C6 A	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2754	2754	2754	2754	2754	2754	2454	295	295	0	295
Cyclo C6 B	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1857	1857	1857	1857	1857	1857	1850	7	7	0	7
Bicyclos	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1699	1699	1699	1699	1699	1699	1698	1	1	0	1
Cyclo C7+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1739	1739	1739	1739	1739	1739	1735	4	4	0	4
Aromatic A	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	4024	4024	4024	4024	4024	4024	3978	43	43	0	43
Aromatic B	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1687	1687	1687	1687	1687	1687	1682	4	4	0	4
Phenanthrenes	0	0	0	0	0	0	0	0	0	0	0	0	4463	4463	4463	3247	3247	3247	3247	3247	3247	3247	0	0	0	0
Pyrenes	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3141	3141	3141	3141	3141	3141	3141	0	0	0	0
Diphenyl compounds	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	542	542	542	542	542	542	542	0	0	0	0
Indans	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	869	869	869	869	869	869	866	3	3	0	3
Indenes	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	648	648	648	648	648	648	648	0	0	0	0
1,2,3,4-Naphthalenes	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1263	1263	1263	1263	1263	1263	1263	0	0	0	0
Naphthalenes	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2173	2173	2173	2173	2173	2173	2173	0	0	0	0
Polynuclear aromatics	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2678	2678	2678	2678	2678	2678	2678	0	0	0	0
Phenols	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2458	2458	2458	2458	2458	2458	2446	1	1	1	0
Sulfide agent	0	0	0	0	0	0	0	0	0	0	13	13	0	0	38	0	0	0	0	0	0	0	0	0	0	0

Stream No.	243	247	248	249	250	251	252	253	254	255	260	262	263	272	280	281	282	290	755	756	757	771	772	792	793	794	803	804	805	806
Temp F	86	86	95	268	77	150	106	150	150	150	86	98	98	150	150	35	149	157	63	345	500	500	490	265	164	113	90	110	90	110
Pres psia	750	740	20	50	283	735	735	1200	1200	1200	15	2060	2060	2065	1980	31	31	755	685	680	675	675	670	665	665	660	60	58	60	55
Vapor mole frac.	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00	0.00	0.00	0.00	0.36	1.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total lb/h	53172	19470	33702	33702	8846	8846	28316	28316	21686	6630	50	13	38	20000	76943	397	77341	53350	75000	75000	75000	28000	28000	29700	29700	29700	1269796	1269796	76511	76511
Flow rates in lb/h																														
Water	66	0	66	66	0	0	0	0	0	0	0	0	0	47	76828	397	77226	397	75000	75000	75000	28000	28000	29700	29700	29700	1269796	1269796	76511	76511
Hydrogen	21634	19470	2163	2163	8846	8846	28316	28316	21686	6630	0	0	0	19811	13	0	13	4	0	0	0	0	0	0	0	0	0	0	0	0
Carbon monoxide	624	0	624	624	0	0	0	0	0	0	0	0	0	2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbon dioxide	12495	0	12495	12495	0	0	0	0	0	0	0	0	0	41	21	0	21	23	0	0	0	0	0	0	0	0	0	0	0	0
Methane	5961	0	5961	5961	0	0	0	0	0	0	0	0	0	93	3	0	3	1	0	0	0	0	0	0	0	0	0	0	0	0
Ethane	5074	0	5074	5074	0	0	0	0	0	0	0	0	0	0	0	0	0	1	0	0	0	0	0	0	0	0	0	0	0	0
Propane	2401	0	2401	2401	0	0	0	0	0	0	0	0	0	0	20	0	20	596	0	0	0	0	0	0	0	0	0	0	0	0
N-butane	1752	0	1752	1752	0	0	0	0	0	0	0	0	0	0	13	0	13	1260	0	0	0	0	0	0	0	0	0	0	0	0
N-pentane	1350	0	1350	1350	0	0	0	0	0	0	0	0	0	0	9	0	9	2623	0	0	0	0	0	0	0	0	0	0	0	0
Ethene	1	0	1	1	0	0	0	0	0	0	0	0	0	1	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propene	1	0	1	1	0	0	0	0	0	0	0	0	0	1	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Hydrogen sulfide	17	0	17	17	0	0	0	0	0	0	0	0	0	0	9	0	9	4	0	0	0	0	0	0	0	0	0	0	0	0
Acids	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Alcohols	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ketones	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1	0	1	0	0	0	0	0	0	0	0	0	0	0	0	0
Aldehydes	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Guaiacols	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Low MW sugars	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
High MW sugars	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Extractives	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Low MW lignin derived A	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Low MW lignin derived B	0	0	0	0	0	0	0	0	0	0	0	0	0	1	0	0	0	1	0	0	0	0	0	0	0	0	0	0	0	0
High MW lignin derived A	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
High MW lignin derived B	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Nitrogen compounds	0	0	0	0	0	0	0	0	0	0	0	0	0	1	0	0	0	79	0	0	0	0	0	0	0	0	0	0	0	0
Sulfur compounds	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	13	0	0	0	0	0	0	0	0	0	0	0	0
Acids intermediates	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbonyl intermediates	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Extractive intermediates	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Low MW lignin compound intermediates	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
High MW lignin compound intermediates	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Sugar intermediate A	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Sugar intermediate B	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Short chain n-paraffins	1033	0	1033	1033	0	0	0	0	0	0	0	0	0	0	6	0	6	5361	0	0	0	0	0	0	0	0	0	0	0	0
Medium chain n-paraffins	2	0	2	2	0	0	0	0	0	0	0	0	0	0	0	0	0	3078	0	0	0	0	0	0	0	0	0	0	0	0
Iso-paraffin A	184	0	184	184	0	0	0	0	0	0	0	0	0	0	1	0	1	1964	0	0	0	0	0	0	0	0	0	0	0	0
Iso-paraffin B	19	0	19	19	0	0	0	0	0	0	0	0	0	0	0	0	0	3096	0	0	0	0	0	0	0	0	0	0	0	0
Cyclo C5 A	114	0	114	114	0	0	0	0	0	0	0	0	0	0	1	0	1	1913	0	0	0	0	0	0	0	0	0	0	0	0
Cyclo C5 B	85	0	85	85	0	0	0	0	0	0	0	0	0	0	0	0	0	2533	0	0	0	0	0	0	0	0	0	0	0	0
Cyclo C6 A	295	0	295	295	0	0	0	0	0	0	0	0	0	0	4	0	4	2454	0	0	0	0	0	0	0	0	0	0	0	0
Cyclo C6 B	7	0	7	7	0	0	0	0	0	0	0	0	0	0	0	0	0	1850	0	0	0	0	0	0	0	0	0	0	0	0
Bicyclos	1	0	1	1	0	0	0	0	0	0	0	0	0	0	0	0	0	1698	0	0	0	0	0	0	0	0	0	0	0	0
Cyclo C7+	4	0	4	4	0	0	0	0	0	0	0	0	0	0	0	0	0	1735	0	0	0	0	0	0	0	0	0	0	0	0
Aromatic A	43	0	43	43	0	0	0	0	0	0	0	0	0	0	3	0	3	3978	0	0	0	0	0	0	0	0	0	0	0	0
Aromatic B	4	0	4	4	0	0	0	0	0	0	0	0	0	0	0	0	0	1682	0	0	0	0	0	0	0	0	0	0	0	0
Phenanthrenes	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3247	0	0	0	0	0	0	0	0	0	0	0	0
Pyrenes	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3141	0	0	0	0	0	0	0	0	0	0	0	0
Diphenyl compounds	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	542	0	0	0	0	0	0	0	0	0	0	0	0
Indans	3	0	3	3	0	0	0	0	0	0	0	0	0	0	0	0	0	866	0	0	0	0	0	0	0	0	0	0	0	0
Indenes	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	648	0	0	0	0	0	0	0	0	0	0	0	0
1,2,3,4-Naphthalenes	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1263	0	0	0	0	0	0	0	0	0	0	0	0
Naphthalenes	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2173	0	0	0	0	0	0	0	0	0	0	0	0
Polynuclear aromatics	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2678	0	0	0	0	0	0	0	0	0	0	0	0
Phenols	0	0	0	0	0	0	0	0	0	0	0	0	0	0	11	0	11	2447	0	0	0	0	0	0	0	0	0	0	0	0
Sulfide agent	0	0	0	0	0	0	0	0	0	0	50	13	38	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0


249
From A200-1
320
600
Off-gas to A600
526
522
521
523
295
524
Hydrogen
from A600
From A600
509
536
532
C-531
CWS
809
769
From A700-2
R-510
Hydrocracking
791
515
E-516
510
From A700-1
514
512
E-517
E-511
T-290
Lights
Column
E-2901
773
528
CWR
To A700-1
810
770
To A200-1
281
From A200
291
792
To A200-1
316
E-315
290
E-293
E-301
E-3101
To Gasoline Pool
E-2900
311
306
321
326
E-325
774
T-310
Naphtha Splitter
To Diesel Pool
305
To A700-2
E-3201
A200-2
Hydrotreating
E-3100
T-320
Diesel Splitter
337
322
E-3200
330
P-332-333
A300 – A500
Product Finishing

Stream No.	249	281	290	291	295	305	306	311	316	320	321	322	326	330	337	509	510	512	514
Temp F	268	35	157	300	138	336	300	149	149	258	347	400	150	678	685	382	1379	485	150
Pres psia	50	31	755	750	65	67	46	25	15	50	15	25	15	15	1295	1295	1270	1270	1260
Vapor mole frac.	1	0	0.00	0.00	1.00	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00	0.00	0.00	0.94	1.00	0.93	0.73
Total lb/h	33702	397	53350	53350	3325	49674	10560	22234	22234	1172	28400	38001	28400	9600	9600	11166	11166	11166	11166
Flow rates in lb/h																			
Water	66	397	397	397	46	0	0	0	0	4	0	0	0	0	0	0	0	0	0
Hydrogen	2163	0	4	4	4	0	0	0	0	69	0	0	0	0	0	1516	541	541	541
Carbon monoxide	624	0	0	0	0	0	0	0	0	20	0	0	0	0	0	0	0	0	0
Carbon dioxide	12495	0	23	23	23	0	0	0	0	396	0	0	0	0	0	0	0	0	0
Methane	5961	0	1	1	1	0	0	0	0	189	0	0	0	0	0	0	0	0	0
Ethane	5074	0	1	1	1	0	0	0	0	160	0	0	0	0	0	0	0	0	0
Propane	2401	0	596	596	596	0	0	0	0	95	0	0	0	0	0	0	0	0	0
N-butane	1752	0	1260	1260	1260	0	0	0	0	95	0	0	0	0	0	0	0	0	0
N-pentane	1350	0	2623	2623	1390	1233	0	1233	1233	87	0	0	0	0	0	0	0	0	0
Ethene	1	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propene	1	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Hydrogen sulfide	17	0	4	4	4	0	0	0	0	1	0	0	0	0	0	0	0	0	0
Acids	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Extractives	0	0	0	0	0	0	70	0	0	0	0	70	0	70	70	70	70	70	70
Low MW lignin derived A	0	0	0	0	0	0	252	0	0	0	0	252	0	252	252	252	252	252	252
Low MW lignin derived B	0	0	1	1	0	1	0	0	0	0	1	1	1	0	0	0	0	0	0
High MW lignin derived A	0	0	0	0	0	0	4	0	0	0	0	4	0	4	4	4	4	4	4
High MW lignin derived B	0	0	0	0	0	0	1	0	0	0	0	1	0	1	1	1	1	1	1
Nitrogen compounds	0	0	79	79	0	79	0	0	0	0	79	79	79	0	0	0	0	0	0
Sulfur compounds	0	0	13	13	0	13	224	0	0	0	13	237	13	224	224	224	224	224	224
Acids intermediates	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbonyl intermediates	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Low MW lignin compound intermediates	0	0	0	0	0	0	269	0	0	0	0	269	0	269	269	269	269	269	269
Short chain n-paraffins	1033	0	5361	5361	0	5361	0	5361	5361	33	0	0	0	0	0	0	0	0	0
Medium chain n-paraffins	2	0	3078	3078	0	3078	0	0	0	0	3078	3078	3078	0	0	0	0	0	0
Iso-paraffin A	184	0	1964	1964	0	1964	0	1964	1964	6	0	0	0	0	0	0	0	0	0
Iso-paraffin B	19	0	3096	3096	0	3096	0	1	1	1	3095	3095	3095	0	0	0	0	0	0
Cyclo C5 A	114	0	1913	1913	0	1913	0	1913	1913	4	0	0	0	0	0	0	0	0	0
Cyclo C5 B	85	0	2533	2533	0	2533	0	2530	2530	3	3	3	3	0	0	0	0	0	0
Cyclo C6 A	295	0	2454	2454	0	2454	0	2454	2454	9	0	0	0	0	0	0	0	0	0
Cyclo C6 B	7	0	1850	1850	0	1850	0	0	0	0	1850	1850	1850	0	0	0	0	0	0
Bicyclos	1	0	1698	1698	0	1698	0	0	0	0	1698	1698	1698	0	0	0	0	0	0
Cyclo C7+	4	0	1735	1735	0	1735	0	0	0	0	1735	1735	1735	0	0	0	0	0	0
Aromatic A	43	0	3978	3978	0	3978	0	592	592	1	3386	3386	3386	0	0	0	0	0	0
Aromatic B	4	0	1682	1682	0	1682	0	0	0	0	1682	1682	1682	0	0	0	0	0	0
Phenanthrenes	0	0	3247	3247	0	3247	0	0	0	0	1	3247	1	3246	3246	3246	0	0	0
Pyrenes	0	0	3141	3141	0	3141	0	0	0	0	0	3141	0	3141	3141	3141	0	0	0
Diphenyl compounds	0	0	542	542	0	542	0	0	0	0	542	542	542	0	0	0	0	0	0
Indans	3	0	866	866	0	866	0	0	0	0	866	866	866	0	0	0	0	0	0
Indenes	0	0	648	648	0	648	0	0	0	0	648	648	648	0	0	0	0	0	0
1,2,3,4-Naphthalenes	0	0	1263	1263	0	1263	0	0	0	0	1263	1263	1263	0	0	0	0	0	0
Naphthalenes	0	0	2173	2173	0	2173	0	0	0	0	2173	2173	2173	0	0	0	0	0	0
Polynuclear aromatics	0	0	2678	2678	0	2678	0	0	0	0	285	2678	285	2393	2393	2393	0	0	0
Phenols	0	0	2447	2447	0	2447	0	0	0	0	2447	2447	2447	0	0	0	0	0	0
N-paraffin-HCK product A	0	0	0	0	0	0	3492	21	21	0	3471	3471	3471	0	0	4	3498	3498	3498
Cyclo C6-HCK products	0	0	0	0	0	0	4320	4305	4305	0	14	14	14	0	0	39	4371	4371	4371
N-paraffin-HCK product B	0	0	0	0	0	0	1926	1859	1859	0	67	67	67	0	0	7	1935	1935	1935

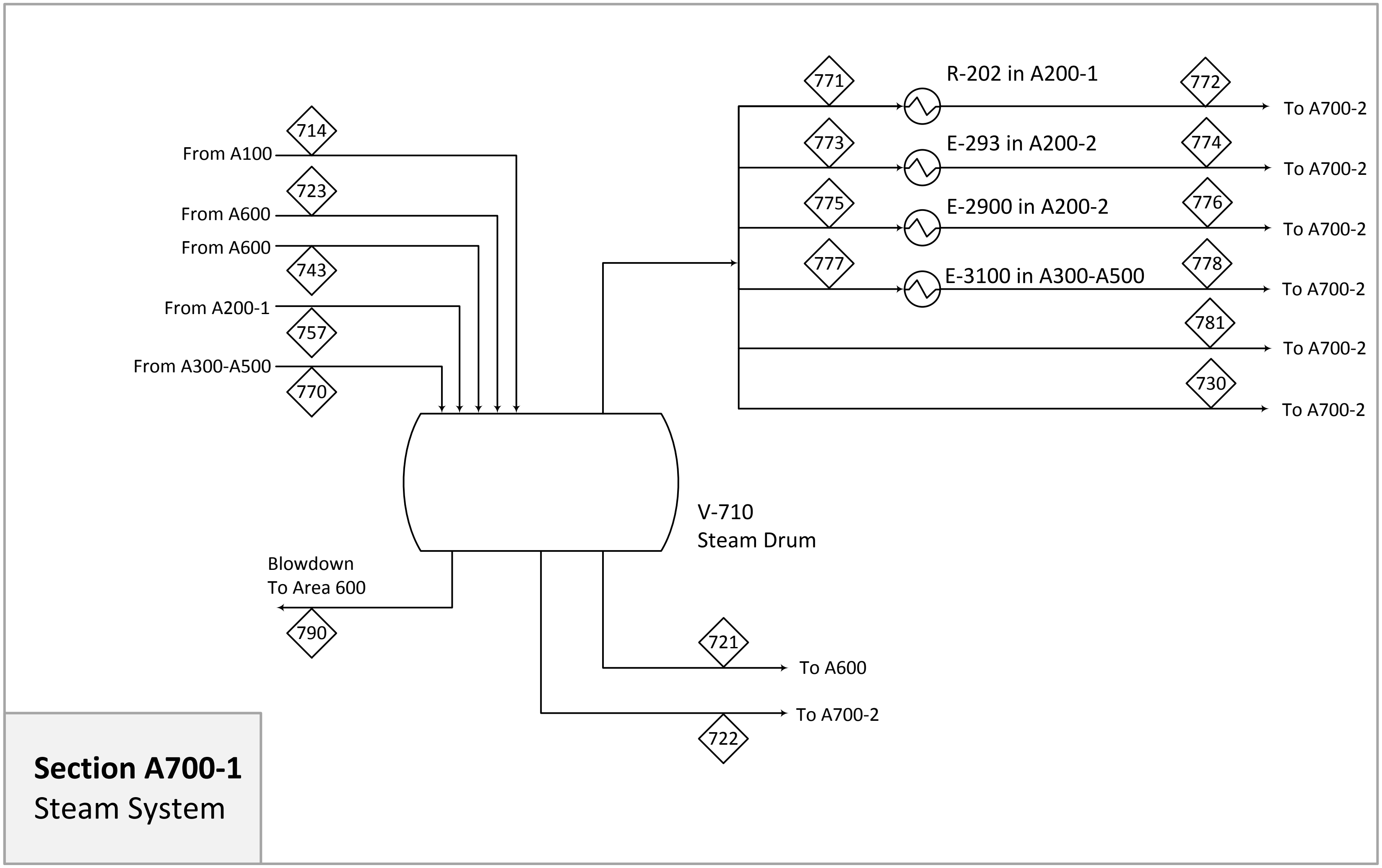
Stream No.	515	521	522	523	524	526	528	532	536	600	769	770	773	774	791	792	809	810
Temp F	110	110	110	110	117	120	120	77	95	258	283	501	500	499	300	265	90	110
Pres psia	1255	1255	1255	1255	1295	50	51	283	1295	50	685	680	675	670	670	665	60	55
Vapor mole frac.	0.73	1.00	1.00	1.00	1.00	1.00	0.00	1.00	1.00	1.00	0.00	1.00	1.00	0.01	0.00	0.00	0.00	0.00
Total lb/h	11166	581	10	571	571	25	10560	995	995	35889	9814	9814	5668	5668	29700	29700	14807	14807
Flow rates in lb/h																		
Water	0	0	0	0	0	0	0	0	0	109	9814	9814	5668	5668	29700	29700	14807	14807
Hydrogen	541	531	9	522	522	9	0	995	995	2117	0	0	0	0	0	0	0	0
Carbon monoxide	0	0	0	0	0	0	0	0	0	604	0	0	0	0	0	0	0	0
Carbon dioxide	0	0	0	0	0	0	0	0	0	12122	0	0	0	0	0	0	0	0
Methane	0	0	0	0	0	0	0	0	0	5773	0	0	0	0	0	0	0	0
Ethane	0	0	0	0	0	0	0	0	0	4914	0	0	0	0	0	0	0	0
Propane	0	0	0	0	0	0	0	0	0	2902	0	0	0	0	0	0	0	0
N-butane	0	0	0	0	0	0	0	0	0	2916	0	0	0	0	0	0	0	0
N-pentane	0	0	0	0	0	0	0	0	0	2654	0	0	0	0	0	0	0	0
Ethene	0	0	0	0	0	0	0	0	0	1	0	0	0	0	0	0	0	0
Propene	0	0	0	0	0	0	0	0	0	2	0	0	0	0	0	0	0	0
Hydrogen sulfide	0	0	0	0	0	0	0	0	0	21	0	0	0	0	0	0	0	0
Acids	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Extractives	70	0	0	0	0	0	70	0	0	0	0	0	0	0	0	0	0	0
Low MW lignin derived A	252	0	0	0	0	0	252	0	0	0	0	0	0	0	0	0	0	0
Low MW lignin derived B	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
High MW lignin derived A	4	0	0	0	0	0	4	0	0	0	0	0	0	0	0	0	0	0
High MW lignin derived B	1	0	0	0	0	0	1	0	0	0	0	0	0	0	0	0	0	0
Nitrogen compounds	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Sulfur compounds	224	0	0	0	0	0	224	0	0	0	0	0	0	0	0	0	0	0
Acids intermediates	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbonyl intermediates	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Low MW lignin compound intermediates	269	0	0	0	0	0	269	0	0	0	0	0	0	0	0	0	0	0
Short chain n-paraffins	0	0	0	0	0	0	0	0	0	1001	0	0	0	0	0	0	0	0
Medium chain n-paraffins	0	0	0	0	0	0	0	0	0	2	0	0	0	0	0	0	0	0
Iso-paraffin A	0	0	0	0	0	0	0	0	0	178	0	0	0	0	0	0	0	0
Iso-paraffin B	0	0	0	0	0	0	0	0	0	19	0	0	0	0	0	0	0	0
Cyclo C5 A	0	0	0	0	0	0	0	0	0	110	0	0	0	0	0	0	0	0
Cyclo C5 B	0	0	0	0	0	0	0	0	0	82	0	0	0	0	0	0	0	0
Cyclo C6 A	0	0	0	0	0	0	0	0	0	286	0	0	0	0	0	0	0	0
Cyclo C6 B	0	0	0	0	0	0	0	0	0	7	0	0	0	0	0	0	0	0
Bicyclos	0	0	0	0	0	0	0	0	0	1	0	0	0	0	0	0	0	0
Cyclo C7+	0	0	0	0	0	0	0	0	0	4	0	0	0	0	0	0	0	0
Aromatic A	0	0	0	0	0	0	0	0	0	42	0	0	0	0	0	0	0	0
Aromatic B	0	0	0	0	0	0	0	0	0	4	0	0	0	0	0	0	0	0
Phenanthrenes	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Pyrenes	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Diphenyl compounds	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Indans	0	0	0	0	0	0	0	0	0	3	0	0	0	0	0	0	0	0
Indenes	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
1,2,3,4-Naphthalenes	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Naphthalenes	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Polynuclear aromatics	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Phenols	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N-paraffin-HCK product A	3498	4	0	4	4	1	3492	0	0	1	0	0	0	0	0	0	0	0
Cyclo C6-HCK products	4371	40	1	39	39	12	4320	0	0	12	0	0	0	0	0	0	0	0
N-paraffin-HCK product B	1935	7	0	7	7	2	1926	0	0	2	0	0	0	0	0	0	0	0



A600
Hydrogen Production

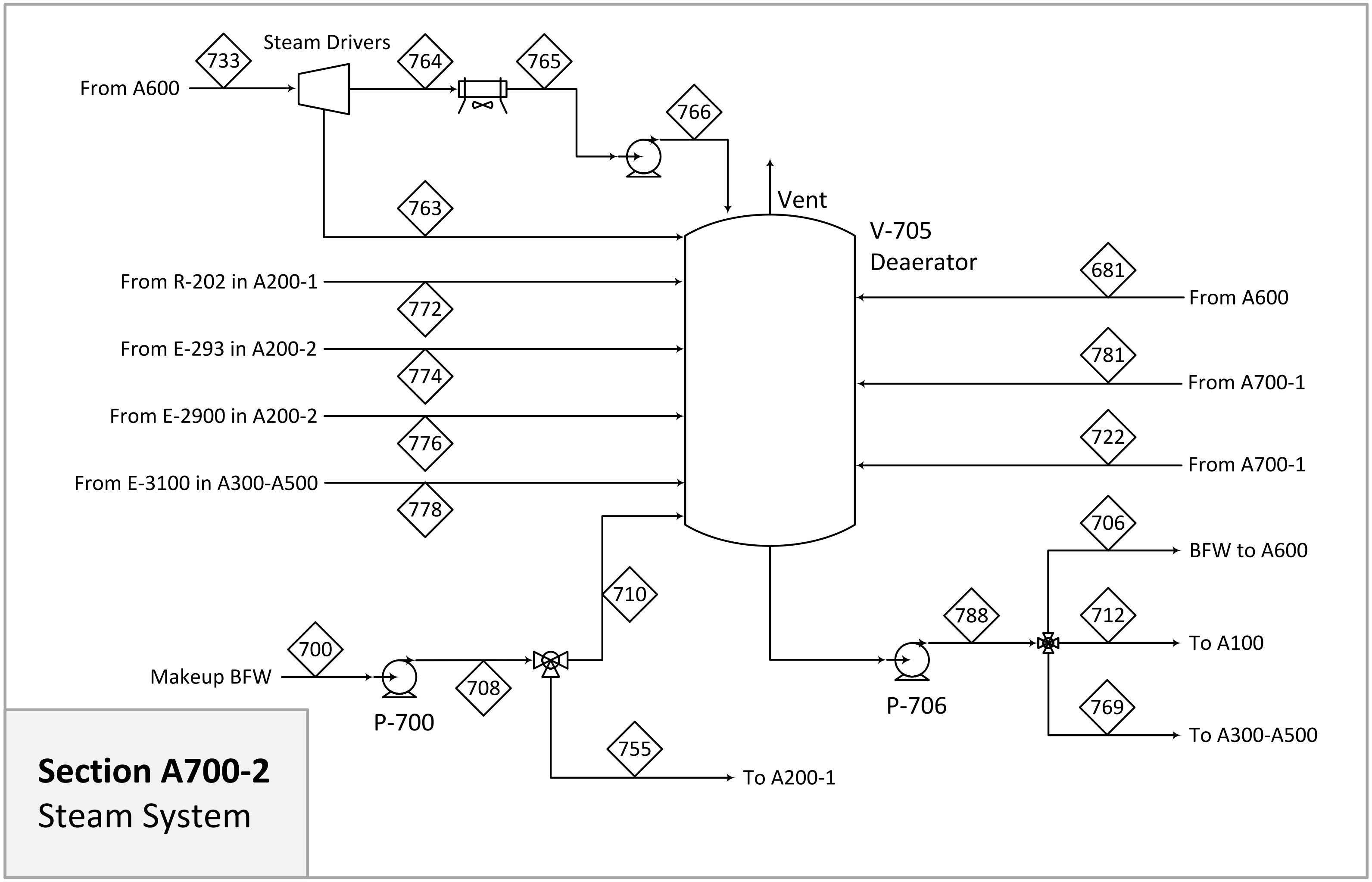
Stream No.	170	250	532	600	601	602	609	610	611	630	650	651	656
Temp F	162	77	77	258	258	258	349	90	127	324	60	304	930
Pres psia	23	283	283	50	50	50	405	15	18	16	415	375	369
Vapor mole frac.	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Total lb/h	12078	8846	995	35889	5635	30255	42333	248000	248000	344393	7000	49333	147015
Flow rates in lb/h													
Water	45	0	0	109	17	92	136	4854	4854	42118	0	136	97836
Hydrogen	4	8846	995	2117	332	1785	1789	0	0	0	0	1789	1789
Oxygen	0	0	0	0	0	0	0	56268	56268	4452	0	0	0
Nitrogen	0	0	0	0	0	0	0	183635	183635	183749	114	114	114
Argon	0	0	0	0	0	0	0	3129	3129	3129	0	0	0
Carbon monoxide	5551	0	0	604	95	509	6061	0	0	0	0	6061	6061
Carbon dioxide	5231	0	0	12122	1903	10219	15450	114	114	110943	54	15504	15504
Methane	737	0	0	5773	906	4867	5604	0	0	0	6149	11753	11753
Ethane	101	0	0	4914	772	4143	4244	0	0	0	405	4649	4649
Propane	2	0	0	2902	456	2447	2449	0	0	0	180	2629	2629
N-butane	0	0	0	2916	458	2458	2458	0	0	0	47	2506	2506
N-pentane	0	0	0	2654	417	2237	2237	0	0	0	6	2243	2243
I-butane	0	0	0	0	0	0	0	0	0	0	36	36	36
I-pentane	0	0	0	0	0	0	0	0	0	0	6	6	6
Ethene	233	0	0	1	0	1	234	0	0	0	0	234	234
Propene	162	0	0	2	0	1	164	0	0	0	0	164	164
Hydrogen sulfide	0	0	0	21	3	17	17	0	0	3	0	17	0
Acids	2	0	0	0	0	0	2	0	0	0	0	2	2
Ketones	8	0	0	0	0	0	9	0	0	0	0	9	9
Nitrogen compounds	0	0	0	0	0	0	0	0	0	0	0	0	0
Short chain n-paraffins	0	0	0	1001	157	844	844	0	0	0	4	847	847
Medium chain n-paraffins	0	0	0	2	0	2	2	0	0	0	0	2	2
Iso-paraffin A	0	0	0	178	28	150	150	0	0	0	0	150	150
Iso-paraffin B	0	0	0	19	3	16	16	0	0	0	0	16	16
Cyclo C5 A	0	0	0	110	17	93	93	0	0	0	0	93	93
Cyclo C5 B	0	0	0	82	13	69	69	0	0	0	0	69	69
Cyclo C6 A	0	0	0	286	45	241	241	0	0	0	0	241	241
Cyclo C6 B	0	0	0	7	1	6	6	0	0	0	0	6	6
Bicyclos	0	0	0	1	0	1	1	0	0	0	0	1	1
Cyclo C7+	0	0	0	4	1	3	3	0	0	0	0	3	3
Aromatic A	0	0	0	42	7	35	35	0	0	0	0	35	35
Aromatic B	0	0	0	4	1	3	3	0	0	0	0	3	3
Indans	0	0	0	3	1	3	3	0	0	0	0	3	3
Indenes	0	0	0	0	0	0	0	0	0	0	0	0	0
Naphthalenes	0	0	0	0	0	0	0	0	0	0	0	0	0
N-paraffin-HCK product A	0	0	0	1	0	1	1	0	0	0	0	1	1
Cyclo C6-HCK products	0	0	0	12	2	10	10	0	0	0	0	10	10
N-paraffin-HCK product B	0	0	0	2	0	2	2	0	0	0	0	2	2

Stream No.	662	667	671	672	680	681	682	690	706	721	723	730	733	743	752	790	791
Temp F	1110	1562	500	676	150	150	111	77	283	499	499	499	750	499	411	499	300
Pres psia	344	314	311	301	293	293	50	283	685	675	675	675	670	675	680	675	670
Vapor mole frac.	1.00	1.00	1.00	1.00	1.00	0.00	0.99	1.00	0.00	0.00	0.50	1.00	1.00	0.69	0.00	1.00	0
Total lb/h	147016	147015	147015	147016	100601	46415	90760	9841	203000	144380	144380	225650	127950	203000	203000	29700	29700
Flow rates in lb/h																	
Water	88456	65280	65280	48271	1856	46415	1856	0	203000	144380	144380	225650	127950	203000	203000	29700	29700
Hydrogen	2227	9675	9675	11578	11578	0	1737	9841	0	0	0	0	0	0	0	0	0
Oxygen	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Nitrogen	114	114	114	114	114	0	114	0	0	0	0	0	0	0	0	0	0
Argon	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbon monoxide	1245	32665	32665	6220	6220	0	6220	0	0	0	0	0	0	0	0	0	0
Carbon dioxide	30752	34377	34377	75928	75928	0	75928	0	0	0	0	0	0	0	0	0	0
Methane	24221	4905	4905	4905	4905	0	4905	0	0	0	0	0	0	0	0	0	0
Ethane	1	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propane	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N-butane	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N-pentane	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
I-butane	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
I-pentane	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethene	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propene	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Hydrogen sulfide	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Acids	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ketones	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Nitrogen compounds	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Short chain n-paraffins	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Medium chain n-paraffins	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Iso-paraffin A	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Iso-paraffin B	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Cyclo C5 A	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Cyclo C5 B	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Cyclo C6 A	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Cyclo C6 B	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Bicyclos	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Cyclo C7+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Aromatic A	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Aromatic B	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Indans	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Indenes	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Naphthalenes	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N-paraffin-HCK product A	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Cyclo C6-HCK products	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N-paraffin-HCK product B	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0



Section A700-1
Steam System

Stream No.	714	721	722	723	730	743	757	770	771	772	773	774	775	776	777	778	781	790
Temp F	501	499	499	499	499	499	499	501	499	489	499	498	499	494	499	498	499	499
Pres psia	675	675	675	675	675	675	675	680	675	670	675	670	675	675	675	670	675	675
Vapor mole frac.	1.00	0.00	0.00	0.50	1.00	0.69	0.36	1.00	1.00	0.00	1.00	0.01	1.00	0.00	1.00	0.00	1.00	0.00
Total lb/h	36600	144380	8619	144380	225650	203000	75000	9821	28000	28000	5669	5669	4316	4316	13206	13206	9261	29700
Flow rates in lb/h																		
Water	36600	144380	8619	144380	225650	203000	75000	9821	28000	28000	5669	5669	4316	4316	13206	13206	9261	29700



Stream No.	681	700	706	708	710	712	722	733	755	763	764	765	766	769	772	774	776	778	781	788
Temp F	150	60	283	63	63	283	499	750	63	482	219	150	150	283	489	498	494	498	499	283
Pres psia	293	25	685	685	685	685	675	670	685	155	17	17	70	685	670	670	675	670	675	685
Vapor mole frac.	0	0.00	0.00	0.00	0.00	0.00	0.00	1	0.00	1.00	0.98	0	0.00	0.00	0.00	0.01	0.00	0.00	1.00	0.00
Total lb/h	46415	80985	203000	80985	5985	36600	8619	127950	75000	1279	126671	126671	126671	9821	28000	5669	4316	13206	9261	249421
Flow rates in lb/h																				
Water	46415	80985	203000	80985	5985	36600	8619	127950	75000	1279	126671	126671	126671	9821	28000	5669	4316	13206	9261	249421

Appendix B: Base Case Individual Equipment Cost Summary and Specifications

Equipment Number	Number Required	Equipment Name	Scaling Stream	Original Equipment Stream Flow	New Stream Flow	stream flow units	Size Ratio	Original Equip Cost (per unit)	Base Year	COST BASIS: installed (i) or bare (b)	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 2011$	Scaled Uninstalled Cost in 2011$	source
A100	Fast Pyrolysis																	
	2	CFB Pyrolyzer, quench & auxiliaries	dry feed, mtpd	400	1,000	mtpd	2.50	$38,000,000	2009	i	$76,000,000	0.7	$144,334,981	3	$144,334,981	$161,979,303	$53,993,101	1
									Subtotal		Pyrolyzer and Auxiliaries					$161,979,303	$53,993,101	
	10	Pyrolysis oil filters	S151	6,000	38,184	gpd	6.36	$162,500	2013	b	$1,625,000	0.7	$5,935,589	1.4	$8,309,825	$8,520,771	$6,086,265	9
	1	Filtrate screw feeder	S158	5,353	5,302	lb/hr	0.99	$34,500	1Q 2011	b	$34,500	0.7	$34,270	1.28	$43,707	$44,777	$35,109	2
E180	1	Steam Gen - Char Burner Exhaust	E180 area SF	104	138	sq feet	1.33	$10,900	1Q 2011	b	$10,900	0.7	$13,306	9.73	$129,523	$132,694	$13,632	2
E181	1	BFW HTR - Char Burner Exhaust	E181 area SF	476	746	sq feet	1.57	$124,100	1Q 2011	b	$124,100	0.7	$170,016	3.9	$663,492	$679,739	$174,180	2
C101	1	Air Compressor for Char Exhaust Cooling (stm drive)	S80 acfm	0	0	MW	1.00	$2,990,900	1Q 2011	b	$2,990,900	0.7	$2,990,900	1.13	$3,380,800	$3,463,590	$3,064,142	2
									Subtotal		Filtration and Heat Recovery					$12,841,571	$9,373,328	
									Subtotal	A100	Fast Pyrolysis					$174,820,874	$63,366,429	
A200	Pyrolysis Oil Upgrading to Stable Oil																	
P200a	3	Feed Pump	S200 at half flow	125	128	gpm	1.03	$107,600	1Q 2011	b	$322,800	0.65	$328,480	1.35	$443,448	$454,308	$336,524	2
R205	1	Stabilizer	S200	2,058	2,060	liq scfh	1.00	$8,649,602	1Q 2011	b	$8,649,602	0.65	$8,655,331	2.75	$23,802,161	$24,385,037	$8,867,286	2
R205	1	1st stg Reactor	S200	2,058	2,060	liq scfh	1.00	$3,112,000	1Q 2011	b	$3,112,000	0.65	$3,114,061	2.75	$8,563,668	$8,773,379	$3,190,320	2
R215	4	2nd stg Reactors	S200	2,058	2,060	liq scfh	1.00	$4,890,600	1Q 2011	b	$19,562,400	0.65	$19,575,357	2.75	$53,832,233	$55,150,496	$20,054,726	2
P210	2	Sulfur Agent Pump	S260	90	50	lb/hr	0.56	$8,500	1Q 2011	b	$17,000	0.65	$11,602	1.79	$20,798	$21,308	$11,886	2
V203	1	Stablilizer Product Flash	S204	161,605	161,729	lb/h	1.00	$57,100	1Q 2011	b	$57,100	0.65	$57,129	3.02	$172,386	$176,608	$58,528	2
V230	1	2nd stg Three Phase Separator	S237	183,685	183,465	lb/h	1.00	$2,137,301	1Q 2011	b	$2,137,301	0.65	$2,135,643	1.40	$2,985,582	$3,058,694	$2,187,941	2
V232	1	Hydrotreating Oil Flash	S240	53,883	53,505	lb/h	0.99	$499,300	1Q 2011	b	$499,300	0.65	$497,016	1.58	$787,182	$806,459	$509,187	2
C231	1	Expander	S239	699	688	acfm	0.98	$1,157,000	1Q 2011	b	$1,157,000	0.65	$1,144,716	1.12	$1,278,030	$1,309,327	$1,172,748	2
E205	1	Stabilized Oil Cooler	duty	25	25	mmbtu/hr	1.01	$127,500	1Q 2011	b	$127,500	0.65	$128,689	5.29	$680,182	$696,838	$131,840	2
E215	1	1st Feed/Product Heat Exchanger	duty	28.47	27.70	mmbtu/hr	0.97	$285,400	1Q 2011	b	$285,400	0.65	$280,378	3.35	$940,162	$963,185	$287,244	2
E221	1	2nd Feed/Product Heat Exchanger	duty	52.81	51.72	mmbtu/hr	0.98	$1,199,000	1Q 2011	b	$1,199,000	0.65	$1,182,856	2.64	$3,121,002	$3,197,431	$1,211,822	2
E228	1	HT Oil Product Heat Recovery	duty	29.75	32.08	mmbtu/hr	1.08	$130,900	1Q 2011	b	$130,900	0.65	$137,476	6.44	$885,977	$907,673	$140,842	2
E229	1	HT Oil Product Cooler	duty	23.32	21.51	mmbtu/hr	0.92	$247,200	1Q 2011	b	$247,200	0.65	$234,581	3.86	$906,156	$928,347	$240,326	2
E294	1	PSA Feed Heater	duty	2.77	3.03	mmbtu/hr	1.09	$26,500	1Q 2011	b	$26,500	0.65	$28,091	2.34	$65,839	$67,451	$28,779	2
E298	1	Blowdown cooler	duty	1.54	1.27	mmbtu/hr	0.82	$21,400	1Q 2011	b	$21,400	0.65	$18,880	3.72	$70,282	$72,003	$19,342	2
H222	1	Fired Heater - startup	duty	8.62	8.62	mmbtu/hr	1.00	$391,700	1Q 2011	b	$391,700	0.65	$391,700	1.14	$444,935	$455,831	$401,292	2
E227	1	Air Cooler	duty	40	40	mmbtu/hr	0.99	$326,000	1Q 2011	b	$326,000	0.65	$324,001	1.29	$419,213	$429,479	$331,935	2
E293	1	Light Column Feed Heater	duty	4	4	mmbtu/hr	1.00	$19,400	1Q 2011	b	$19,400	0.65	$19,400	7.70	$149,400	$153,059	$19,875	2
E2900	1	T290 Reboiler	duty	3	3	mmbtu/hr	1.12	$29,600	1Q 2011	b	$29,600	0.65	$31,780	2.91	$92,550	$94,816	$32,559	2
E2901	1	T290 Condenser	duty	2	2	mmbtu/hr	0.94	$28,100	1Q 2011	b	$28,100	0.65	$27,064	3.34	$90,343	$92,555	$27,727	2
T290	1	Light Column	S291	53,272	53,350	lb/hr	1.00	$219,400	1Q 2011	b	$219,400	0.65	$219,610	1.46	$320,374	$328,219	$224,987	2
V290	1	T290 Reflux Drum	S299 at half flow	119	58	Std L gpm	0.49	$31,200	1Q 2011	b	$31,200	0.65	$19,653	4.25	$83,527	$85,572	$20,135	2
P206	3	Stablized Product Pump	S211	129	258	gpm	2.00	$88,400	2Q 2011	b	$265,200	0.65	$415,619	1.39	$577,710	$591,858	$425,797	2
V222	1	PSA	S243	53,484	53,172	lb/hr	0.99	$4,511,278	2013	i	$4,511,278	0.65	$4,494,160	3.00	$4,494,160	$4,052,701	$1,350,900	10
C251	1	PSA Off Gas compressor	MW	1.00	1.00	MW	0.99	$1,457,700	1Q 2011	b	$1,457,700	0.65	$1,450,850	1.12	$1,623,137	$1,662,885	$1,486,379	2
C208a,b	1	Stabilizer gas product compressor	MW	4	4	MW	1.11	$1,919,400	1Q 2011	b	$1,919,400	0.65	$2,058,678	1.09	$2,242,706	$2,297,626	$2,109,092	2
C252	1	Makeup Hydrogen Compressor	MW	2	2	MW	1.04	$1,395,800	1Q 2011	b	$1,395,800	0.65	$1,433,161	1.09	$1,562,225	$1,600,481	$1,468,256	2
C254	1	Hydrogen Compressor	MW	3	4	MW	1.12	$1,693,200	1Q 2011	b	$1,693,200	0.65	$1,817,657	1.09	$1,988,700	$2,037,400	$1,862,169	2
									Subtotal	A200	Pyrolysis Oil Upgrading to Stable Oil					$114,851,026	$48,210,444	

Equipment Number	Number Required	Equipment Name	Scaling Stream	Original Equipment Stream Flow	New Stream Flow	stream flow units	Size Ratio	Original Equip Cost (per unit)	Base Year	COST BASIS: installed (i) or bare (b)	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 2011$	Scaled Uninstalled Cost in 2011$	source
A 300 & 500	Product Separation and Hydrocracking																	
T310	1	T-310 Naphtha Splitter	S305	50,335	49,674	lb/h	0.99	$321,200	1Q 2011	b	$321,200	0.65	318,455	1.74	553,727	567,287	$326,253	2
T320	1	T-320 Product Splitter	S323	38,000	38,001	lb/h	1.00	$484,600	1Q 2011	b	$484,600	0.65	484,607	1.67	809,612	829,438	$496,474	2
E3100	1	T-310 Reboiler	DUTY	9.65	9.40	mmbtuh	0.97	$58,700	1Q 2011	b	$58,700	0.65	57,707	2.47	142,547	146,038	$59,120	2
E3200	1	T-320 Fired Reboiler	DUTY	14.60	14.00	mmbtuh	1.00	$425,300	1Q 2011	b	$425,300	0.65	413,856	1.29	533,874	546,948	$423,991	2
E3101	1	T-310 Air Cool Condenser	DUTY	10.24	10.24	mmbtuh	1.00	$255,600	1Q 2011	b	$255,600	0.65	255,600	1.46	373,176	382,314	$261,859	2
E3201	1	T-320 Air Cool Condenser	DUTY	13.89	13.52	mmbtuh	1.00	$207,100	1Q 2011	b	$207,100	0.65	203,497	1.39	282,861	289,788	$208,481	2
V310	1	T-310 Reflux Drum	wksht	83	83	gpm	1.00	$28,900	1Q 2011	b	$28,900	0.65	28,880	4.96	143,299	146,808	$29,587	2
V320	1	T-320 Reflux Drum	wksht	15	16	gpm	1.07	$14,700	1Q 2011	b	$14,700	0.65	15,367	6.72	103,284	105,813	$15,743	2
P332	2	T-310 bottom product pump	S323	16	16	gpm	1.00	$5,600	1Q 2011	b	$11,200	0.65	11,195	2.69	30,152	30,890	$11,470	2
P333	2	T-310 bottom product pump	S323	16	16	gpm	1.00	$5,700	1Q 2011	b	$11,400	0.65	11,395	4.56	51,969	53,242	$11,674	2
	1	Hydrocracker Unit :	S337	2,250	551	bpd fd	0.24	$30,000,000	1Q 2005	i	$30,000,000	0.65	12,018,172	3.00	12,018,172	15,063,222	$5,021,074	4
H508		H508 Fired heater																
R510		R510 Hydrocracker Vessel																
E513		E513 Feed/product exchanger																
E516		E516 air cooler																
E517		E517 trim cooler																
V520		V520 HP flash																
V530		V530 LP flash																
C522		Fuel Gas Compresor																
E315	1	Gasoline Product Cooler	Duty - placeholder	1.0	1.00	mmbtuh	1.00	$66,800	1Q 2011	b	$66,800	0.65	66,800	1.30	86,997	89,127	$68,436	2
E325	1	Diesel Product Cooler	Duty	3.1	2.98	mmbtuh	0.96	$185,800	1Q 2011	b	$185,800	0.65	181,093	1.37	248,637	254,726	$185,527	2
									Subtotal	A 300 & 500	Product Separation and Hydrocracking					$18,505,641	$7,119,689	
A600	Hydrogen Plant																	
	1	Steam Reformer System	S250+S501	49.0	44.5	mscfd H2	0.91	$66,875,000	May 2007	i	$66,875,000	0.65	62,781,396	1.92	62,781,396	$69,144,535	$36,012,778.46	5
									Subtotal			Hydrogen Plant			62,781,396	69,144,535	36,012,778	

Equipment Number	Number Required	Equipment Name	Scaling Stream	Original Equipment Stream Flow	New Stream Flow	stream flow units	Size Ratio	Original Equip Cost (per unit)	Base Year	COST BASIS: installed (i) or bare (b)	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 2011$	Scaled Uninstalled Cost in 2011$	source
A700	Utilities and WWT																	
	1	WWT aerobic digestion	mptd biomass	1,000	2,000	tpd	2.00	$1,554,000	2004	i	$1,554,000	0.65	2,438,487	3.00	2,438,487	3,215,268	$1,071,756	6
	1	Field Erected CTW w pumps, etc	from FLOWS	4,620	4,902	gpm	1.06	$237,700	1Q 2011	i	$237,700	0.78	248,948	1.57	248,948	255,045	$162,449	2
	1	Plant Air Compressor	mptd biomass	2,000	2,000	tpd	1.00	$32,376	2002	b	$32,376	0.34	32,376	3.00	97,129	143,803	$47,934	7
	1	Hydraulic Truck Dump with Scale	mptd biomass	2,000	2,000	tpd	1.00	$80,000	1998	b	$80,000	0.6	80,000	3.00	240,000	360,893	$120,298	7
	1	Firewater Pump	mptd biomass	2,000	2,000	tpd	1.00	$18,400	1997	b	$18,400	0.79	18,400	3.00	55,200	83,650	$27,883	7
	1	Diesel Pump	mptd biomass	2,000	2,000	tpd	1.00	$6,100	1997	b	$6,100	0.79	6,100	3.00	18,300	27,732	$9,244	7
	1	Instrument Air Dryer	mptd biomass	2,000	2,000	tpd	1.00	$8,349	2002	b	$8,349	0.6	8,349	3.00	25,047	37,083	$12,361	7
	1	Plant Air Receiver	mptd biomass	2,000	2,000	tpd	1.00	$7,003	2002	b	$7,003	0.72	7,003	3.00	21,009	31,105	$10,368	7
	1	Firewater Storage Tank	mptd biomass	2,000	2,000	tpd	1.00	$166,100	1997	b	$166,100	0.51	166,100	3.00	498,300	755,121	$251,707	7
	1	Ammonia Pump		included in Stream Reformer Cost														
	1	Hydrazine Pump		included in Stream Reformer Cost														
	1	Ammonia Storage Tank		included in Stream Reformer Cost														
	1	Hydrazine Storage Tank		included in Stream Reformer Cost														
	1	Flare		included in Stream Reformer Cost														
	1	Sulfur Agent Storage Tank	S260	242	167	gpd	0.69	$84,800	1Q 2011	b	$84,800	0.65	$66,523	2.48	$164,995	$169,036	$68,152	2
	1	Feed Storage		1,056,846	1,056,846	gallons	1.00	470,000	1Q 2005	b	$470,000	0.65	470,000	3.00	1,410,000	1,767,252	$589,084	3
	1	Product Storage		558,000	577,518	gallons	1.03	320,384	1Q 2005	b	$320,384	0.65	327,625	3.00	982,874	1,231,905	$410,635	3
	1	Product Storage		558,000	528,283	gallons	0.95	320,384	1Q 2005	b	$320,384	0.65	309,188	3.00	927,563	1,162,581	$387,527	3
									Subtotal	A700		Utilities and Auxiliaries				$9,240,474	$3,169,398	
														TOTALS		$386,562,550	$157,878,738	

Equipment Sources:

1. Envergent 2009.
2. ASPEN 7.3.2 Capital Cost Evaluator
3. Aspen Icarus 2006
4. Marker 2005.
5. SRI 2007.
6. Jones 2009.
7. Phillips 2007.
9. Budgetary Vendor Estimate
10. Budgetary Vendor Estimate

Tag	Equipment Descriptions	Type	Design Temperature and pressure	Size		Metallurgy	Cost Basis
M-110	Fitrate screw feeder			Wet feed flow, lb/h	5353	SS316	Aspen Capital Cost Estimator V7.3.2
E-181	Steam generator - Char burner exhaust	Floating head heat exchanger	Shell: 1098 F, 31 psig	Area, ft2	476	CS	Aspen Capital Cost Estimator V7.3.2
			Tube: 550 F, 726 psig	Duty, MMBtu/h	26.2		
E-180	BFW Heater - Char burner exhaust	Pre-engineered U-tube heat exchanger	Shell: 1162 F, 31 psig	Area, ft2	103	SS316	Aspen Capital Cost Estimator V7.3.2
			Tube: 801 F, 737 psig	Duty, MMBtu/h	8.7		
C-101	Air compressor for biomass drying	Reciprocating motor drive	140 F, 5 psig	Actual Inlet flowrate, ft3/min	44381	CS	Aspen Capital Cost Estimator V7.3.2
S-155	Bio-oil cold filters	Filter					Pall
R205	Stabilizer		Reaction channel: 335 F, 1509 psig	Diameter, ft	15	Reaction side: SS316 cladding	Aspen Capital Cost Estimator V7.3.2
			Steam side: 550F, 711 psig	Length, ft	60	Steam side: CS	
R205	1st stg Reactor		396 F, 2050 psig	Diameter, ft	10	SS316 cladding	Budgetary estimate from Pall
				Length, ft	52		
R215	2nd stg Reactors		820 F, 2211 psig	Diameter, ft	10	SS316 cladding	Budgetary estimate from Pall
				Length, ft	49		
V203	Stablilizer Product Flash	Vertical	155 F, 1507 psig	Diameter, ft	2.5	SS316 cladding	Aspen Capital Cost Estimator V7.3.2
		1 minute holdup		Length, ft	10.5		
V230	2nd stg Three Phase Separator	Horizontal	200 F, 2162 psig	Diameter, ft	10	SS316 cladding	Aspen Capital Cost Estimator V7.3.2
		5 minute holdup		Length, ft	49		
V232	Hydrotreating Oil Flash	Vertical, cylinder	89 F, 809 psig	Diameter, ft	7.5	SS316 cladding	Aspen Capital Cost Estimator V7.3.2
		1 minute holdup		Length, ft	36		
C231	Expander		Inlet: 200 F, 2162 psig	Actual Inlet flowrate, ft3/min	700		Aspen Capital Cost Estimator V7.3.2
			Outlet: 809 psig				
E205	Stabilization Bed Product Cooler	Fixed shell and tube	Shell: 160 F, 70 psig	Area, ft2	3860	SS316 cladding on tube side	Aspen Capital Cost Estimator V7.3.2
		heat exchanger	Tube: 334 F, 1513 psig	Duty, MMBtu/h	25.4		
E215	1st Feed/Product Heat Exchanger	Fixed shell and tube	Shell: 453 F, 2168 psig	Area, ft2	2069	SS316 cladding	Aspen Capital Cost Estimator V7.3.2
		heat exchanger	Tube: 350 F, 2255 psig	Duty, MMBtu/h	28		
E221	2nd Feed/Product Heat Exchanger	Fixed shell and tube	Shell: 650 F, 2173 psig	Area, ft2	11560	SS316 cladding	Aspen Capital Cost Estimator V7.3.2
		heat exchanger	Tube: 600 F, 2217 psig	Duty, MMBtu/h	52.2		
E228	Hydrotreated Oil Product Heat Recovery	Fixed shell and tube	Shell: 550 F, 732 psig	Area, ft2	1275	SS316 cladding on tube side	Aspen Capital Cost Estimator V7.3.2
		heat exchanger	Tube: 882 F, 2173 psig	Duty, MMBtu/h	32.2		
E229	Hydrotreated Oil Product Cooler	Fixed shell and tube	Shell: 403 F, 738 psig	Area, ft2	5468	SS316 cladding	Aspen Capital Cost Estimator V7.3.2
		heat exchanger	Tube: 405 F, 2178 psig	Duty, MMBtu/h	17.9		
E294	PSA Feed Heater	Fixed shell and tube	Shell: 136 F, 809 psig	Area, ft2	114	SS316 cladding	Aspen Capital Cost Estimator V7.3.2
		heat exchanger	Tube: 407 F, 2167 psig	Duty, MMBtu/h	3.1		
H222	Fired Heater - startup		600 F, 2211 psig	Std gas flow rate, ft3/min	81.62	CS	Aspen Capital Cost Estimator V7.3.2
				Duty, MMBtu/h	8.62		
E227	Hydrotreated Oil Air Cooler	Air fin	300 F, 150 psig	Area, ft2	6059	SS316 tube	Aspen Capital Cost Estimator V7.3.2
				Duty, MMBtu/h	40.3		
E293	Light Column Feed Heater	Fixed shell and tube	Shell: 550 F, 726 psig	Area, ft2	168	SS316 cladding on tube side	Aspen Capital Cost Estimator V7.3.2
		heat exchanger	Tube: 350 F, 809 psig	Duty, MMBtu/h	4		
E2900	T290 Reboiler	U-tube reboiler	Shell: 550 F, 726 psig	Area, ft2	190	SS316 cladding on shell side	Aspen Capital Cost Estimator V7.3.2
			Tube: 377 F, 57 psig	Duty, MMBtu/h	2.8		
E2901	T290 Condenser	Fixed shell and tube	Shell: 196 F, 55 psig	Area, ft2	417	SS316 cladding on shell side	Aspen Capital Cost Estimator V7.3.2
		heat exchanger	Tube: 160 F, 50 psig	Duty, MMBtu/h	1.8		
T290	Light Column		377 F, 77 psig	Diameter, ft	2.5	SS316 cladding	Aspen Capital Cost Estimator V7.3.2
				# of trays	40		
V290	T290 Reflux Drum	Horizontal	377 F, 57 psig	Diameter, ft	4	SS316 cladding	Aspen Capital Cost Estimator V7.3.2
		5 minute holdup		Length, ft	12		
C251	PSA Off Gas compressor	Reciprocating motor drive	Inlet: 145 F, 5 psig	Actual Inlet flowrate, ft3/min	10379	CS	Aspen Capital Cost Estimator V7.3.2
			Outlet: 60 psig				
C208a,b	Stabilizer gas product compressor	Reciprocating motor drive	Inlet: 145 F, 5 psig	Actual Inlet flowrate, ft3/min	1184	CS	Aspen Capital Cost Estimator V7.3.2
			Outlet: 2255 psig				

Tag	Equipment Descriptions	Type	Design Temperature and pressure	Size		Metallurgy	Cost Basis
C252	Makeup Hydrogen Compressor	Reciprocating motor drive	Inlet: 145 F, 5 psig	Actual Inlet flowrate, ft3/min	1458	CS	Aspen Capital Cost Estimator V7.3.2
			Outlet: 792 psig				
C254	Hydrogen Compressor	Reciprocating motor drive	Inlet: 145 F, 5 psig	Actual Inlet flowrate, ft3/min	1180	CS	Aspen Capital Cost Estimator V7.3.2
			Outlet: 1304 psig				
T310	T-310 Naphtha Splitter		448 F, 35 psig	Diameter, ft	5	SS316 cladding	Aspen Capital Cost Estimator V7.3.2
				# of trays	30		
T320	T-320 Product Splitter		731 F, 26 psig	Diameter, ft	7	SS316 cladding	Aspen Capital Cost Estimator V7.3.2
				# of trays	30		
E3100	T-310 Reboiler	U-tube reboiler	Shell: 448 F, 35 psig	Area, ft2	1030		Aspen Capital Cost Estimator V7.3.2
			Tube: 550 F, 726 psig	Duty, MMBtu/h	9.6		
E3200	T-320 Fired Reboiler		730 F, 45 psig	Std gas flow rate, ft3/min	445	CS	Aspen Capital Cost Estimator V7.3.2
				Duty, MMBtu/h	14.6		
E3101	T-310 Air Cool Condenser	Air fin	300 F, 45 psig	Area, ft2	4850	SS316 tube	Aspen Capital Cost Estimator V7.3.2
				Duty, MMBtu/h	10.3		
E3201	T-320 Air Cool Condenser	Air fin	300 F, 35 psig	Area, ft2	3849	SS316 tube	Aspen Capital Cost Estimator V7.3.2
				Duty, MMBtu/h	14.2		
V310	T-310 Reflux Drum	Horizontal	448 F, 35 psig	Diameter, ft	4	SS316 cladding	Aspen Capital Cost Estimator V7.3.2
		5 minute holdup		Length, ft	10.5		
V320	T-320 Reflux Drum	Horizontal	730 F, 25 psig	Diameter, ft	2	SS316 cladding	Aspen Capital Cost Estimator V7.3.2
		5 minute holdup		Length, ft	6.5		
E315	Gasoline Product Cooler	Air fin	300 F, 150 psig	Area, ft2	985	CS	Aspen Capital Cost Estimator V7.3.2
				Duty, MMBtu/h	1		
E325	Diesel Product Cooler	Air fin	300 F, 150 psig	Area, ft2	5137	CS	Aspen Capital Cost Estimator V7.3.2
				Duty, MMBtu/h	3		

Appendix C: Base Case Process Parameters and Operation

Feed Dryer	
Inlet biomass	
Temperature (°F)	68
Moisture content (%wt)	30
Inlet Flue gas	
Temperature (°F)	584
Oxygen content (% mole)	9.6
Outlet biomass	
Temperature (°F)	160
Moisture content (%wt)	10
Outlet Flue gas	
Dew point (°F)	139
Temperature (°F)	161
Moisture content (%wt)	12

Pyrolyzer	
Biomass moisture content (%wt)	10
Biomass particle size (mm)	2-6
Fluidized gas (wt/wt dry biomass)	14.5
Sand (wt/wt dry biomass)	3
Inlet temperature (°F)	931
Pressure (psia)	21
Heat Losses (% of biomass LHV)	1
Dry bio-oil yield (% dry biomass)	64
Water yield (% dry biomass)	12
Char yield (% dry biomass)	12
Gas yield (% dry biomass)	12

Fast Pyrolysis Oil Quench Columns	
Organic liquid recovery (%)	99.6
Cold stream temperature (°F)	
1st Quench column	140
2nd Quench column	110
Product stream temperature (°F)	
1st Quench column	158
2nd Quench column	113

Fast Pyrolysis Oil Cold Filter	
Solid removal (%)	100
Fast pyrolysis oil yield loss	
% of fast pyrolysis oil to the filter	3.055
% of dry biomass	2

Char Combustor	
Temperature (°F)	1128
Pressure (psia)	21
Air (actual/minimum for combustor)	1.20
Heat Losses (% of biomass LHV)	1

Stabilizer*	
Temperature (°F)	284
Pressure (psia)	1200
Hydrogen Consumption* (%wt, per pyrolysis oil)	0.08
H2 feed/dry oil feed (wt/wt)	0.055
LHSV (h^{-1})	0.5
Reactor volume (gallon)	76700
Number of tubes	800
Reactor length (ft)	60
Number of trains	1
Inlet	
Temperature (°F)	132
Conversion (wt%)	
Acids	19.9
Ketones	19.8
Extractives	5
Low MW lignin derived A	8
Conversion (wt%) to:	
Acids intermediates	29.5
Carbonyl intermediates	42.9
Extractive intermediates	3.4
Low MW lignin compound intermediates	20.5
Methane	3.1
Carbon dioxide	0.6
Yield (include hydrogen, wt %)	
Oil	96
Gas	4

Complete experimental data for hydrotreating and hydrocracking are not currently available. Reaction conversions, hydrotreating intermediate yields and hydrocracking yields are assumed for completing mass and energy balances.

1st Stage Hydrotreater*	
Temperature (°F)	348
Pressure (psia)	2000
Hydrogen Consumption (%wt, per pyrolysis oil	0.18
H2 feed/dry oil feed (wt/wt)	0.07
LHSV (h^{-1})	0.5
Reactor volume (gallon)	30600
Reactor length (ft)	52
Number of trains	1
Inlet	
Temperature (°F)	300
Pressure (psia)	2055
Conversion (wt%)	
Acids	80
Ketones	80
Low MW sugars	80
Extractives	80
Low MW lignin derived A	80
High MW lignin derived A	20
Hydrogen	2.3
Conversion (wt%) to:	
Acids Intermediates	11.8
Carbonyl Intermediates	4.5
Extractive Intermediates	1.9
Low MW lignin compound intermediates	23.9
High MW lignin compound intermediates	11.7
Sugar intermediate A	8
Sugar intermediate B	9.2
Phenanthrenes	12.1
Carbon dioxide	3.3
Methane	1.3
Ethane	2.4
Water	9.9
Yield (include hydrogen, wt %)	
Oil	83
Gas	10
Water	7

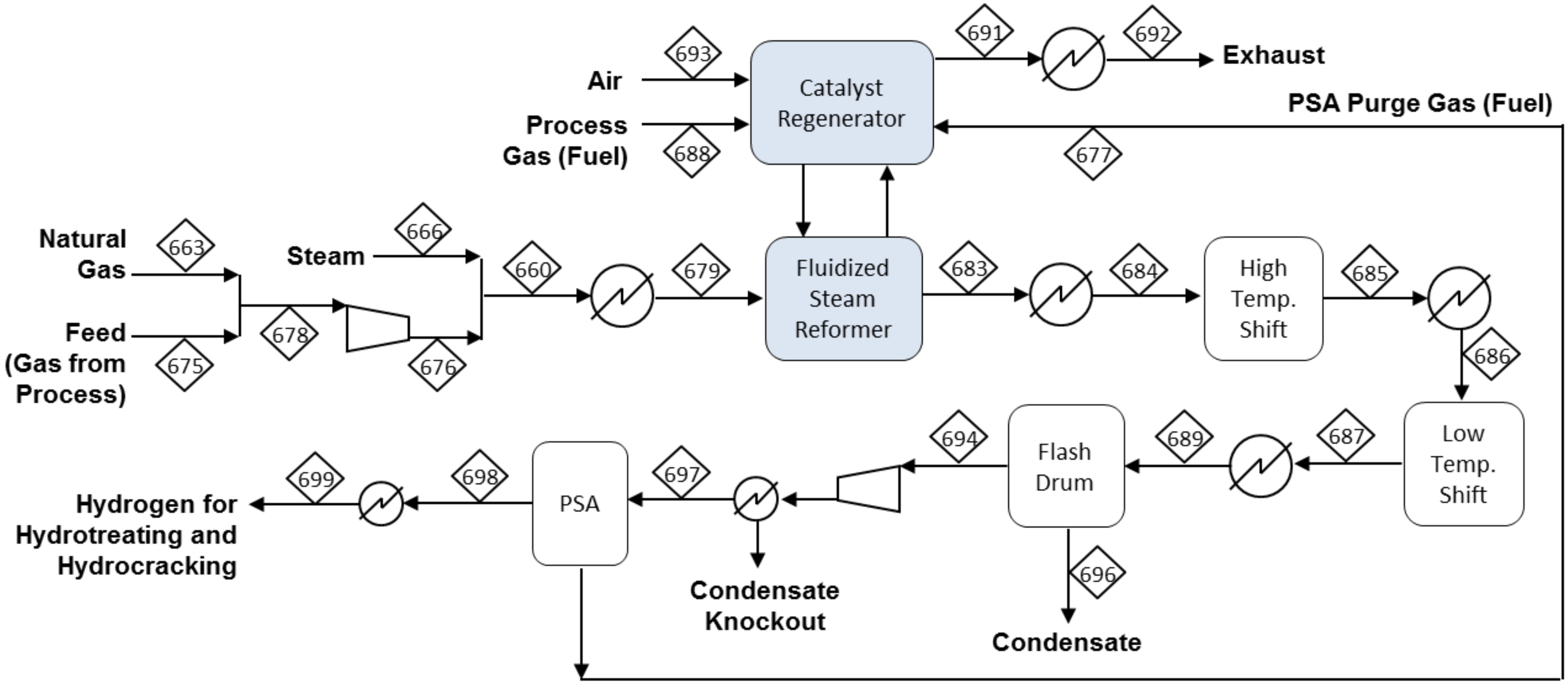
2nd Stage Hydrotreater*			
Temperature (°F)	767	Conversion (wt%) to:	
Pressure (psia)	2000	Short chain n-paraffins	5.6
Hydrogen Consumption (%wt, per pyrolysis oil)	5.54	Medium chain n-paraffins	2.7
H2 feed/dry oil feed (wt/wt)	0.21	Iso-paraffin A	1.9
LHSV (h^{-1})	0.22	Iso-paraffin B	2.7
Reactor volume	28800	Cyclo C5 A	1.8
Reactor length	49	Cyclo C5 B	2.3
Number of trains	4	Cyclo C6 A	2.4
		Cyclo C6 B	1.6
		Bicyclos	1.5
Inlet		Cyclo C7+	1.5
Temperature (°F)	550	Aromatic A	3.5
Pressure (psia)	2010	Aromatic B	1.5
		Phenanthrenes	2.8
		Pyrenes	2.8
Conversion (wt%)		Diphenyl compounds	0.5
Aldehydes	100	Indans	0.8
Alcohols	100	Indenes	0.6
Guaiacols	100	1,2,3,4-Naphthanlenes	1.1
High MW sugars	100	Naphthalenes	1.9
Low MW lignin derived B	100	Polynuclear Aromatics	2.3
High MW lignin derived A	100	Phenols	2.2
High MW lignin derived B	100	Carbon monoxide	0.5
Sulfur compounds	55	Carbon dioxide	9.9
Acids Intermediates	100	Methane	4.7
Carbonyl Intermediates	100	Ethane	3.7
Extractive Intermediates	100	Propane	2.6
Low MW lignin compound intermediates	100	Butane	2.6
High MW lignin compound intermediates	100	Pentane	3.5
Sugar intermediate A	100	Hydrogen sulfide	0.0025
Sugar intermediate B	100	Water	28.5
Phenanthrenes	27		
		Yield (include hydrogen, wt %)	
		Oil	48
		Gas	24
		Water	28

Overall Hydrotreating	
Product oxygen content, %	0.5
Hydrogen Consumption (%wt, per pyrolysis oil)	5.8
Yield (per dry pyrolysis oil, wt %)	
Oil	44
Gas	26
Water	29

Hydrocracking*	
Temperature (°F)	1382
Pressure (psia)	1270
Hydrogen Consumption (%wt, per pyrolysis oil)	0.9
H2 feed/dry oil feed (wt/wt)	0.135
Inlet	
Temperature (°F)	382
Pressure (psia)	1295
Conversion (wt%)	
Extractives	100
Low MW lignin derived A	100
High MW lignin derived A	100
High MW lignin derived B	100
Sulfur compounds	100
Extractive Intermediates	100
Low MW lignin compound intermediates	100
Conversion (wt%) to:	
N-Nonane	35.8
Methylcyclohexan	44.4
N-Octane	19.8

* Complete experimental data for hydrotreating and hydrocracking are not currently available. Reaction conversions, hydrotreating intermediate yields and hydrocracking yields are assumed for completing mass and energy balances.

Appendix D1: CFB Reformer Heat and Material Balances



Temp F	341	60	307	245	477	120	220	1600	1670	662	788	392	430	245	110	1800	685	278	110	110	120	120	77
Pres psia	74	22	74	22.7	75	35	22	69	66	61	56	51	46	22.7	30	22.7	20.7	31	30	30	350	340	283
Vapor Frac	1.00	1.00	0.98	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	0.68	1.00	1.00	1.00	1.00	0.00	0.99	1.00	1.00
Total lb/hr	169689	5550.843	125879.3	38258.84	43809.69	86980.16	43809.69	169689	169689	169689	169689	169689	169689	9708.333	169689	341278.3	341278.3	244589.8	100578.2	69110.8	96821.16	9841	9841
Flow rates in lb/hr																							
Argon	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	3085.818	3085.818	3085.818	0	0	0	0	0
Oxygen	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	9249.105	9249.105	55494.63	0	0	0	0	0
Nitrogen	90.582	90.582	0	0	90.582	90.581	90.582	90.582	90.582	90.582	90.582	90.582	90.582	0	90.582	181199.6	181199.6	181109.1	90.582	0.001	90.581	0	0
Ammonia	0	0	0	0	0	0.022	0	0	0.039	0.039	0.039	0.039	0.039	0	0.039	0	0	0	0.024	0.015	0.022	0	0
Hydrogen sulfide	16.417	0	0	16.417	16.417	16.4	16.417	16.417	16.417	16.417	16.417	16.417	16.417	4.166	16.417	0	0	0	16.404	0.013	16.4	0	0
Sulfur dioxide	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	38.659	38.659	0	0	0	0	0	0
Nitrogen dioxide	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0.087	0.087	0	0	0	0	0	0
Carbon monoxide	4909.797	0	0	4909.797	4909.797	554.691	4909.797	4909.797	27942.32	27942.32	6587.56	6587.56	554.699	1245.88	554.699	0	0	0	554.693	0.007	554.691	0	0
Carbon dioxide	13883.71	42.692	0	13841.01	13883.71	79108.88	13883.71	13883.71	36110.88	36110.88	69663.37	69663.37	79142.17	3512.212	79142.17	106982.9	106982.9	112.818	79114.51	27.658	79108.88	0	0
Hydrogen	1691.841	0	0	1691.841	1691.841	1736.647	1691.841	1691.841	9606.774	9606.774	11143.65	11143.65	11577.83	429.311	11577.83	0	0	0	11577.69	0.144	11577.65	9841	9841
Methane	10068.9	4876.221	0	5192.678	10068.9	2014.029	10068.9	10068.9	2014.074	2014.074	2014.074	2014.074	2014.074	1317.663	2014.074	0	0	0	2014.038	0.036	2014.029	0	0
Ethane	4321.405	320.862	0	4000.543	4321.405	43.202	4321.405	4321.405	43.214	43.214	43.214	43.214	43.214	1015.154	43.214	0	0	0	43.204	0.01	43.202	0	0
Propane	2459.597	142.587	0	2317.01	2459.597	245.876	2459.597	2459.597	245.959	245.959	245.959	245.959	245.959	587.951	245.959	0	0	0	245.892	0.067	245.876	0	0
Ethylene	186.737	0	0	186.737	186.737	18.668	186.737	186.737	18.674	18.674	18.674	18.674	18.674	47.385	18.674	0	0	0	18.669	0.005	18.668	0	0
Propylene	130.77	0	0	130.77	130.77	13.072	130.77	130.77	13.077	13.077	13.077	13.077	13.077	33.183	13.077	0	0	0	13.073	0.004	13.072	0	0
Water	126001.5	0	125879.3	122.187	122.187	2546.478	122.187	126001.5	92994.2	92994.2	79259.59	79259.59	75379.47	31.006	75379.47	40722.1	40722.1	4787.493	6297.689	69081.78	2546.478	0	0
Pyridine	0	0	0	0	0	0	0	0	0.001	0.001	0.001	0.001	0.001	0	0.001	0	0	0	0	0.001	0	0	0
Acids	1.882	0	0	1.882	1.882	0.005	1.882	1.882	0.188	0.188	0.188	0.188	0.188	0.478	0.188	0	0	0	0.012	0.176	0.005	0	0
Ketones	6.781	0	0	6.781	6.781	0.004	6.781	6.781	0.678	0.678	0.678	0.678	0.678	1.721	0.678	0	0	0	0.022	0.656	0.004	0	0
Nitrogen compounds	0.278	0	0	0.278	0.278	0	0.278	0.278	0	0	0	0	0	0.07	0	0	0	0	0	0	0	0	0
N-butane	2391.6	65.78	0	2325.82	2391.6	239.067	2391.6	2391.6	239.16	239.16	239.16	239.16	239.16	590.186	239.16	0	0	0	239.084	0.076	239.067	0	0
N-pentane	2126.098	9.332	0	2116.766	2126.098	212.52	2126.098	2126.098	212.61	212.61	212.61	212.61	212.61	537.138	212.61	0	0	0	212.536	0.074	212.52	0	0
Short chain n-paraffins	800.922	2.787	0	798.136	800.922	80.057	800.922	800.922	80.092	80.092	80.092	80.092	80.092	202.53	80.092	0	0	0	80.063	0.03	80.057	0	0
Medium chain n-parrafins	1.479	0	0	1.479	1.479	0.148	1.479	1.479	0.148	0.148	0.148	0.148	0.148	0.375	0.148	0	0	0	0.148	0	0.148	0	0
Decane	14.837	0	0	14.837	14.837	1.483	14.837	14.837	1.484	1.484	1.484	1.484	1.484	3.765	1.484	0	0	0	1.483	0	1.483	0	0
Iso-paraffin B	87.78	0	0	87.78	87.78	8.771	87.78	87.78	8.778	8.778	8.778	8.778	8.778	22.275	8.778	0	0	0	8.772	0.006	8.771	0	0
Cyclo C5 A	65.376	0	0	65.376	65.376	6.534	65.376	65.376	6.538	6.538	6.538	6.538	6.538	16.589	6.538	0	0	0	6.534	0.003	6.534	0	0
Cyclo C6 A	228.142	0	0	228.142	228.142	22.79	228.142	228.142	22.814	22.814	22.814	22.814	22.814	57.892	22.814	0	0	0	22.794	0.02	22.79	0	0
Diphenyl compounds	0.02	0	0	0.02	0.02	0.002	0.02	0.02	0.002	0.002	0.002	0.002	0.002	0.005	0.002	0	0	0	0.002	0	0.002	0	0
Indans	2.589	0	0	2.589	2.589	0.258	2.589	2.589	0.259	0.259	0.259	0.259	0.259	0.657	0.259	0	0	0	0.258	0.001	0.258	0	0
Polynuclear aromatics	0.009	0	0	0.009	0.009	0.001	0.009	0.009	0.001	0.001	0.001	0.001	0.001	0.002	0.001	0	0	0	0.001	0	0.001	0	0
Phenols	0.088	0	0	0.088	0.088	0.001	0.088	0.088	0.009	0.009	0.009	0.009	0.009	0.022	0.009	0	0	0	0.001	0.008	0.001	0	0
N-paraffin-HCK product A	1.04	0	0	1.04	1.04	0.104	1.04	1.04	0.104	0.104	0.104	0.104	0.104	0.264	0.104	0	0	0	0.104	0	0.104	0	0
Cyclo C6-HCK products	9.713	0	0	9.713	9.713	0.971	9.713	9.713	0.971	0.971	0.971	0.971	0.971	2.465	0.971	0	0	0	0.971	0.001	0.971	0	0
N-paraffin-HCK product B	1.62	0	0	1.62	1.62	0.162	1.62	1.62	0.162	0.162	0.162	0.162	0.162	0.411	0.162	0	0	0	0.162	0	0.162	0	0
Cyclo C6 B	5.213	0	0	5.213	5.213	0.521	5.213	5.213	0.521	0.521	0.521	0.521	0.521	1.323	0.521	0	0	0	0.521	0	0.521	0	0
Aromatic A	33.55	0	0	33.55	33.55	3.35	33.55	33.55	3.355	3.355	3.355	3.355	3.355	8.514	3.355	0	0	0	3.35	0.005	3.35	0	0
Pyrenes	0.002	0	0	0.002	0.002	0	0.002	0.002	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Iso-paraffin A	141.733	0	0	141.733	141.733	14.168	141.733	141.733	14.173	14.173	14.173	14.173	14.173	35.965	14.173	0	0	0	14.169	0.004	14.168	0	0
Indenes	0.2	0	0	0.2	0.2	0.02	0.2	0.2	0.02	0.02	0.02	0.02	0.02	0.051	0.02	0	0	0	0.02	0	0.02	0	0
Naphthalenes	0.205	0	0	0.205	0.205	0.02	0.205	0.205	0.02	0.02	0.02	0.02	0.02	0.052	0.02	0	0	0	0.02	0	0.02	0	0
Bicyclos	0.567	0	0	0.567	0.567	0.057	0.567	0.567	0.057	0.057	0.057	0.057	0.057	0.144	0.057	0	0	0	0.057	0	0.057	0	0
Cyclo C7+	2.8	0	0	2.8	2.8	0.28	2.8	2.8	0.28	0.28	0.28	0.28	0.28	0.71	0.28	0	0	0	0.28	0	0.28	0	0
Aromatic B	3.209	0	0	3.209	3.209	0.32	3.209	3.209	0.321	0.321	0.321	0.321	0.321	0.814	0.321	0	0	0	0.32	0.001	0.32	0	0
Phenanthrenes	0.008	0	0	0.008	0.008	0.001	0.008	0.008	0.001	0.001	0.001	0.001	0.001	0.002	0.001	0	0	0	0.001	0	0.001	0	0
1,2,3,4-Naphthalenes	0.009	0	0	0.009	0.009	0.001	0.009	0.009	0.001	0.001	0.001	0.001	0.001	0.002	0.001	0	0	0	0.001	0	0.001	0	0

Appendix D2: CFB Reformer Capital and Operating Costs

This section details the equipment assumptions and specifications for the alternate hydrogen plant configuration that uses a CFB reformer. Table D-1 lists the equipment and cost estimates for hydrogen production using a fluidized bed reformer system. Capital costs for some common equipment items (e.g., tanks, pumps, vessels) are based on the recent NREL study (Dutta 2011). The fluidized bed reformer package including a dual fluidized bed loop, cyclones and ancillaries was estimated from a prior cost estimation study (Dutta 2011, Worley 2012). HTS, LTS, and PSA capital were based on Spath et al. (2005).

Table D-1. Capital Cost Estimate for Fluidized Bed Reformer (2011 k$).

Equipment Description	TPEC	TIC	Source
Fluidized bed Reformer Module	7,007	16,186	(Dutta 2011)
Reformer Feed Preheater	3,049	7,626	(Dutta 2011)
PSA Air-cooled Precooler	151	373	(Spath 2005)
PSA Water-cooled Precooler	83	206	(Spath 2005)
High Temperature Shift Reactor	457	1,128	(Spath 2005)
Low Temperature Shift Reactor	317	783	(Spath 2005)
Pre-PSA Knock-out #1	124	306	(Spath 2005)
Pressure Swing Adsorption Unit	2,112	3,900	(Vendor Quote – Harris Group 2012)
PSA Compressor	10,721	19,298	(Dutta 2011)
PSA Compressor Air Cooler	291	875	(Dutta 2011)
Steam Turbine Condenser	4,147	5,806	(Dutta 2011)
Process Steam Generator	3,371	8,430	(Dutta 2011)
High Pressure Steam BFW Preheater	1,891	4,729	(Dutta 2011)
High Pressure Steam Generator	2,396	5,993	(Dutta 2011)
High Pressure Steam Superheater	1,199	2,998	(Dutta 2011)
Blowdown Cooler	46	200	(Dutta 2011)
Blowdown Cooler 2	11	47	(Dutta 2011)
Hot Process Water Softener System	1,046	1,203	(Dutta 2011)
Hot Process Water Softener System 2	1,079	1,241	(Dutta 2011)
Makeup Pump	7	31	(Dutta 2011)
Makeup Pump 2	13	63	(Dutta 2011)
Condensate Pump	18	81	(Dutta 2011)
Condensate Pump 2	15	69	(Dutta 2011)
Water Softener Pump	18	81	(Dutta 2011)
Water Softener Pump 2	15	69	(Dutta 2011)
Deaerator Feed Pump	15	63	(Dutta 2011)
Boiler Feed Water Pump	376	509	(Dutta 2011)
Boiler Feed Water Pump 2	444	601	(Dutta 2011)
Blowdown Flash Drum	37	127	(Dutta 2011)
Blowdown Flash Drum 2	27	91	(Dutta 2011)
Dearator Feed Water Tank	9	56	(Dutta 2011)
Deaerator	17	84	(Dutta 2011)
Deaerator Packed Column	6	30	(Dutta 2011)
Steam Drum	58	133	(Dutta 2011)
Steam Drum 2	48	109	(Dutta 2011)
Cooling Tower System	65	161	(Dutta 2011)
Cooling Water Pump	313	669	(Dutta 2011)
Water Make-Up Pump	7	31	(Dutta 2011)
Water Make-Up Pump 2	13	63	(Dutta 2011)
Total	**41,018**	**84,447**	

Equipment costs were scaled using the process simulation material and energy balances. Equipment costs were adjusted to the 2011 U.S. dollar basis using the Chemical Engineering Indices.

The associated variable operation costs, and utilities are shown in Tables D-2- and D-3. Unit costs are consistent with Table 29 and Appendix C of Dutta 2011. Variable operating costs are determined based on raw materials, utilities, and waste handling charges incurred only during the process operation. The Inorganic Cost Index was used to convert unit costs from 2007 dollars to 2011 dollars.

Table D-2. Variable Operating Costs.

Variable	**Unit Cost**		**Amount**		**Operating Cost**		**Reference**
	(2011$)				**(2011$)**		
Reformer Catalyst[1]	2,166	¢/ton	9.36	lb/hr	0.10	$/hr	(Dutta 2011)
ZnO & Shift Catalysts[2]	2,166	¢/ton	0	lb/hr	0.00	$/hr	(Dutta 2011)
Natural Gas	76.57	¢/MMBtu	113	MMBTU/hr	86.52	$/hr	(EIA 2013)
Boiler Feed Water Makeup	28.79	¢/ton	66,017	lb/hr	9.50	$/hr	(Dutta 2011)
Cooling Tower Water Makeup	28.79	¢/ton	21,423	lb/hr	3.08	$/hr	(Dutta 2011)
Boiler Feed Water Chemicals	555,764	¢/ton	2.20	lb/hr	6.11	$/hr	(Dutta 2011)
Cooling Tower Water Chemicals	396,975	¢/ton	0.24	lb/hr	0.48	$/hr	(Dutta 2011)
Wastewater	69.37	¢/ton	8451	lb/hr	2.93	$/hr	(Dutta 2011)
Solids Disposal Cost	3,266	¢/ton	9.36	lb/hr	0.15	$/hr	(Dutta 2011)
Electricity	5.85	¢/kWh	10,191	kW	596.17	$/hr	(Dutta 2011)

[1] Reactor space velocity: 2,000 h^{-1}, Catalyst volume: 1,767 ft^3, Required initial reformer catalyst fill: 113103 lb

[2] Required initial ZnO & WGS catalysts fill: 183453 lb. Initial fill than replaced every 5 years based on typical catalyst lifetime (Spath 2005).

Table D-3. Feed and Utilities.

Utility	**Value**	**Units**
Saturated steam for plant utilization (499ºF, 675 psia)	51,191	lb/hr
Superheated steam (1000 ºF, 660 psia) for power	116,933	lb/hr
- corresponding to power supply for steam driver with 75% efficiency	12,633	kWh
Natural gas	5,551	lb/hr
	113	MMBtu/hr (LHV)
Power consumption	10,191	kWh
Boiler feed water makeup	66,017	lb/hr
	174	GPM
Cooling tower water makeup (20 ºF rise)	21,423	lb/hr
- CWS temperature, 90 ºF	56	GPM

Appendix E: Feedstock Supply

Feedstock logistics includes operations that take place after the biomass is produced in a field or forest, but before it is introduced into a conversion process. All activities related to feedstock logistics are directed at reducing the delivered *cost* of sustainably produced feedstock, improving and preserving the *quality* of harvested feedstock to meet the needs of biorefineries, and/or expanding the *volume* of feedstock materials accessible to the bioenergy industry. Feedstock logistics efforts are primarily focused on identifying, developing, demonstrating, and validating efficient and economic systems to harvest, collect, transport, store, and preprocess raw biomass from a variety of crops to reliably deliver high quality, high volume, and affordable feedstocks to biorefineries. Some preprocessing research takes place under the conversion technology areas, while other research is funded under feedstock logistics

Sustainably supplying the required quantities of quality, affordable feedstock to the emerging biorefining industry will be achieved through a transition from logistics systems that have been designed to meet the needs of conventional agriculture and forestry systems—termed "conventional" logistics systems—to more advanced, purpose-designed systems in the 2013 to 2022 timeframe, termed "advanced" logistics systems (Hess 2009).

Existing conventional logistics systems, which adequately serve the needs of current agriculture and forestry systems, are capable of moving biomass short distances and storing it for limited periods of time. However, these conventional systems can be restrictive to the bioenergy industry. Research has focused on understanding these limitations and identifying opportunities for improvement. Examples of these limitations include an inability to address the physical and chemical variability that exists in biomass materials or to access diverse and distributed resources across the nation to satisfy the needs of an expanding biorefinery industry.

The development of advanced, infrastructure-compatible feedstocks is projected to provide a commodity-based, specification-driven system to supply the bioenergy industry. The U.S. grain commodity system serves as one example for a conceptual model for the type of feedstock handling system that will be needed to ensure biomass quality and performance in the supply chain. It also serves as a model for in-feed systems for a variety of conversion processes that envisions similar supply system infrastructure for a broad range of biomass resources, enables commodity biomass feedstocks with predictable physical and chemical characteristics, enables biomass storage over fairly long periods of time; improves transportability over relatively long distances, and enables many end uses.

As illustrated in Figure E-1, the advanced Terrestrial Feedstock Processing Supply and Logistics System is envisioned to draw in presently inaccessible and/or underused resources via local biomass preprocessing depots that format biomass into a stable, bulk, densified, and flowable material. The formatted biomass will be transported to one or more of a network of much larger supply terminals, where the material aggregated from a number of depots may be blended and/or further preprocessed to meet the specification required by each biorefinery conversion process (note that feedstock specifications may differ substantially, depending upon the actual conversion process). The advanced biomass logistics system design incrementally incorporates technology and other system improvements as the industry matures. This improved series of feedstock supply and logistics system designs will couple to, and build from, current systems and address science and engineering constraints that have been identified by rigorous sensitivity analyses as having the greatest impact on feedstock supply and logistics system efficiencies and costs. The introduction of advanced preprocessing operations and sequences, including the implementation of blending and formulation strategies, is critical to achieving feedstock cost, quality, and volume targets to meet Office 2017 and 2022 goals.

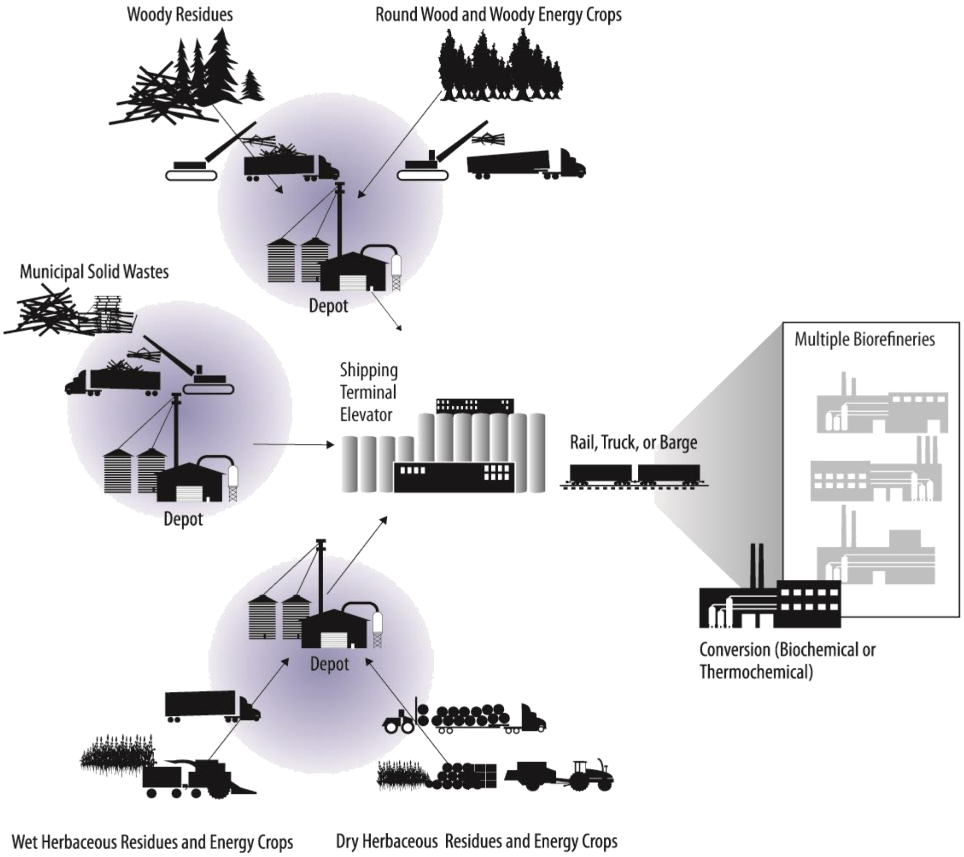


Figure E-1. Wood Biomass Feedstock Supply System.

The physical properties of raw biomass directly influence efficiency of transport and handling systems, as well as conveyance operations within the biorefinery that deliver feedstock to the conversion process reactor throat. Physical properties of feedstock, such as particle size and particle size distribution, can also significantly impact conversion process performance.

To meet longer-term goals and to support a growing bioenergy industry, research must expand its horizon beyond reliance on high-yielding biomass regions (such as the Corn Belt) to increase the amount of cellulosic feedstock available for energy production while increasing the emphasis on quality and reducing variability. One aspect of the inherent variability of biomass is illustrated by the ash content found in a wide array of corn stover samples, as shown in Table E-1. Table E-1 data were extracted from Templeton et al. (2009), where 500 samples were taken from 47 locations, including eight U.S. Corn Belt states over three harvest years.

Table E-1. Example of Variability in Corn Stover Composition.

Corn Stover Composition		
Parameter	**Minimum**	**Maximum**
Ash	0.8%	6.6%
Xylan	14.8%	22.7%
Lignin	11.2%	17.8%

In response to the variability in biomass quality, more robust biofuel conversion technologies are being developed, even though it is unlikely a single best conversion technology will be capable of handling the variability experienced within raw biomass feedstocks. Additionally current discussions are predominantly around the impact of ash on the production of bio-oils, however, the elemental ash components in bio-oil offer challenges and opportunities for the implementation of upgrading. As we move forward research on improved ash inventory and speciation will be developed in order to reach an optimal solution from either enhanced specific ash element(s) removal or improvements towards more robust conversion. Currently, eight biofuel conversion pathways have currently been identified and have active R&D programs. Other approaches to addressing the variability include blending/formulation, leaching, densification, and other preprocessing options.

The first aspect of this approach is in changing the manner in which prices for feedstocks are reported. In previous reports of feedstock availability by cost, researchers have used supply volumes available at the marginal cost of acquiring additional resources. Essentially, the simulated buyer in this instance stops purchasing when the price reaches a certain limit. In Table E-2, these volumes are reported in the column entitled “Marginal price reporting” for year 2017 with a farm gate price limit of $60/dry ton. The $60/dry ton price target reflects the cost to access biomass at the farm gate or the forest landing[1]. Reporting marginal cost limits may not reflect the true behavior of an open feedstock market. If feedstock buyers are instead targeting an average price for purchased feedstocks, they may mix more expensive resources with less expensive ones of the same type to achieve an average farm gate cost of $60/dry ton. The volumes acquired nationwide at an average price of $60/dry ton in 2017 are shown in the rightmost column of Table E-2. This does not physically change anything in the system, only the way costs and volumes are reported. However, it helps when assessing the entire supply chain to work with average costs instead of marginal costs. Note that by reporting volumes available based on average prices, the national feedstock availability increases from 298 million dry tons to 326 million dry tons in 2017, a difference of 28 million dry tons. [2]

Table E-2. Potential Increase in Volume of Biomass Available as a Result of Blending Biomass at an Example Grower Payment of $40/dry ton (KDF 2013).

National feedstock supply for 2017 at $40/dry ton		Supply (million dry ton/yr)	Supply (million dry ton/yr)
Resource		No Blending	Intra feedstock blending
Agricultural Residues	Corn Stover	13.87	22.74
	Cereal Straw	3.21	5.27
Energy Crops	Herbaceous energy crops	4.61	7.56
	Woody energy crops	0.01	0.01
Forest Resources	Pulpwood	0.02	0.06
	Logging residues and fuel treatments	101.17	164.02
	Other forestland removals	36.95	60.22
	Urban and mill wood wastes	4.56	7.79
Potential feedstock supply (Totals)		**164.4**	**267.67**

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[1] Note that this is not the same as the Grower payment. Grower payments are those made to biomass producers over and above the costs incurred for harvest, collection, storage, preprocessing, and transport. The Office models the grower payment based on anticipated biomass demand. For more information on grower payments, see: Robert Perlack, Bryce Stokes et al., “U.S. Billion-Ton Update: Biomass Supply for a Bioenergy and Bioproducts Industry,” Oak Ridge National Laboratory, ORNL/TM-2011/224, 2011, http://www1.eere.energy.gov/biomass/pdfs/billion_ton_update.pdf.

[2] Volumes derived from data available at https://www.bioenergykdf.net/. Please refer to the Knowledge Discovery Framework for a more expansive list of grower payments and the associated quantity of biomass.

The second aspect important to consider for a nationwide bioenergy industry is the combination of cost, volume, and quality necessary for economic energy production. Quality targets can have a large impact on whether or not a particular feedstock is cost effective. One aspect of the inherent spatial and temporal variability of biomass resource quality is illustrated by the ash, xylan, and glucan distributions for Midwestern corn stover in Figure E-2. In response to the variability in biomass quality, a variety of more robust biofuel conversion technologies are being developed, even though it is unlikely a single best conversion technology will be capable of handling the variability experienced within raw biomass feedstocks. Other approaches to addressing the variability include blending/formulation, leaching, densification, and other preprocessing options.

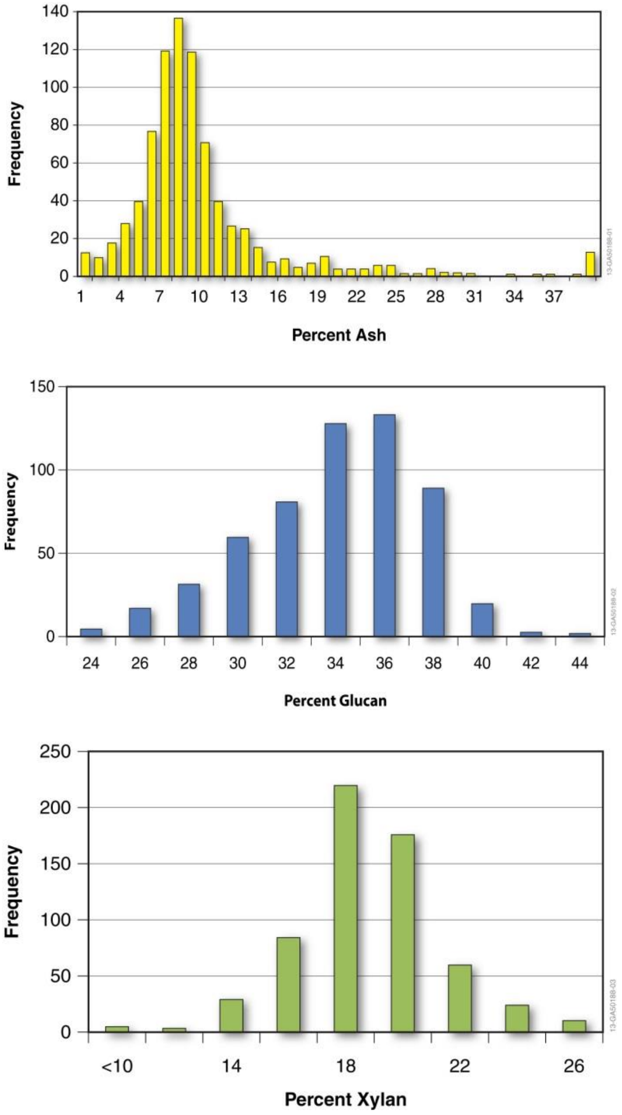

Figure E-2. Example of the Spatial and Temporal Variability of Corn Stover Characteristics such as % Ash Content, % Glucan Content and % Xylan.

Data extracted from the Biomass R&D Library

By combining analyses using average farm gate price assumptions with quality information obtained from the Biomass R&D Library, gains in the projected volumes available at cost and biorefinery specifications are being realized by transitioning to a blended feedstock approach. Feedstock blending allows a theoretical biorefinery to collect less of any one feedstock and thus move down the cost vs. supply curve, paying a lower average price for each feedstock. Not that this does not change the supply vs. cost curves for each resource, but it instead describes a system where purchasers are using a combination of least-cost resources and blending them to reach the bioenergy application's desired specification.

With blended feedstocks, biomass quality is a key aspect to consider when analyzing cost and volume availability. Formulating a designed feedstock through blending and other pre-processing logistical methods allows low cost and typically low quality biomass to be blended with biomass of higher cost and typically higher quality to achieve the specifications at the in-feed of a conversion facility. The use of low cost biomass allows the supply chain to implement additional preprocessing technologies that actively control feedstock quality, while also bringing more biomass into the system. This analysis and design approach is being called the "least-cost formulation" strategy.

Along with farm gate price, volume, and quality, a key consideration is feedstock logistics cost, which includes costs of harvest, collection, preprocessing, transportation, and storage. There are a variety of factors that influence logistics cost, including feedstock type, biomass yield per unit of land area, preprocessing strategies required to achieve conversion specifications, storage requirements, transportation distance, and dry matter losses throughout the supply system. Research is directed at reducing these logistics costs, moving towards the gradual attainment of Office targets by 2017.

Regional biomass production supply curves from the US Billion Ton Update[3] were used to identify "high impact[4]" scenarios to develop and demonstrate the 2017 feedstock supply chain scenarios at the $80/dry ton target. A scenario below focuses on a Southeastern US scenario delivering a formulated feedstock for fast pyrolysis conversion pathway that produces a bio-oil. It is important to note that the baseline scenarios are regionally focused to establish feedstock costs and technical performance requirements of the supply chain; however, the supply chain technologies and designs are broadly applicable across the US.

Using a least-cost formulation spatial tool developed at INL, a high impact formulation based on forest thinning, logging residues, and plantation-grown loblolly pine creates an opportunity to meet the feedstock cost and quality targets for the bio-oil conversion pathway. Forest thinning and logging residues are low cost resources to procure, but often have unfavorable qualities, specifically high ash content. Money saved in procuring these biomass resources (which are lower cost to procure than, say, pulpwood trees) can be applied in the supply chain to address quality issues, such as through active ash management. The identified formulation is shown in (Table E-3). The C&D wastes are incorporated because of its low access fee cost and its low ash content. This is only an example; the actual blends will be regionally based designs that take advantage of local feedstocks and their biomass characteristics. Additionally the ability to blend feedstocks to a specification has the potential to reduce some of the risks associated with the seasonality of feedstocks.

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[3] Robert Perlack, Bryce Stokes, et al., "U.S. Billion-Ton Update: Biomass Supply for a Bioenergy and Bioproducts Industry," Oak Ridge National Laboratory, ORNL/TM-2011/224, 2011, http://www1.eere.energy.gov/biomass/pdfs/billion_ton_update.pdf.

[4] To be considered high impact, the feedstock must be domestically available and have the agronomically and ecologically sustainable ultimate availability potential to produce at least 1 billion gallons per year of an acceptable biofuel.

Currently, the forest thinnings and logging residues are both reduced to 1.05% ash after a chemical pretreatment. C&D wastes will undergo a low severity pre-treatment and the ash content will be reduced to 0.9%. The purpose grown pine is debarked and chipped at the landing resulting in an ash content of 0.5%. The blended material will be delivered to the conversion facility a 2 wt% ash. A further pre-treatment operation will be performed on the blended material that will reduce the ash to the required 0.9%. It is important that the pretreatment operation be inexpensive or offset the costs by improving other operations, such as conversion yields, decrease grinding or pelletizing costs. There are a number of chemical pretreatment options being explored that will reduce the ash. INL is focusing on these pretreatment technologies but the goal is to meet both the $80 per short ton cost target and 0.9% ash specifications assumed in this report. When blended, the complete formulation meets both the cost and feedstock quality targets.

Table E-3. List of Costs and Specifications for Woody Feedstocks and Blends for Thermochemical Conversion

Feedstock	Reactor Throat Feedstock Cost ($/dry ton)[2]	Formulation Fraction (%)	% Ash Delivered to Throat of Conversion Reactor
Pulp	99.38	30	0.5
Logging Residues[1]	74.83	35	1.5
Switchgrass	80.54	10	2.8
C&D Wastes	63.77	25	0.5
Formulation Totals	**80.00**	**100**	**1.1**

[1] residues do not include costs for harvest and collection; they are moved to landing while attached to the merchantable portion of the tree (for example, timber or pulpwood)

[2] includes ash mitigation

The supply system design for the Southeast blends the three feedstocks (thinnings, residues, and pulpwood). Initial harvest, collection, and preprocessing (i.e., de-barking and size reduction) occur at the landing for these cost estimates while secondary preprocessing, storage, and handling occur within the gates of the biorefinery for this cost estimate. At the landing all feedstock are comminuted, however only pulpwood includes a debarking process. For pulpwood, preprocessing involves a grinder and a dryer. For thinnings and residues, preprocessing includes a chip cleaning operation as well as a dryer and grinder. Note that many potential design elements may be incorporated to meet cost targets, including densification (including high-moisture densification), ash reduction, fractionation, thermal and hydrothermal treatments, and other developing technologies.

Preliminary results on feedstock supply chains delivering material to cellulosic sugar, bio-oil, and syngas conversion pathways demonstrate that with the ability to blend multiple feedstocks and include some preprocessing operations it is possible to acquire high volumes of material, reduce feedstock variability to meet biorefinery in-feed specifications and meet the required $80/dry ton cost of material to the throat of the biorefinery. Much more research needs to be done both on the performance of blended material as well as the blending strategies themselves, as well as other technologies incorporated into advanced designs. In addition, current pre-conversion systems are expensive, and reducing those costs is a significant barrier to meeting future cost targets. However, there is ongoing research at the National Laboratories, universities, and industry to address these challenges.

Appendix F: Compound Selection

Bio-oil Composition

Pyrolysis oil or bio-oil is a complex mixture comprising of hundreds of compounds. Only up to 40% of the compounds can be observed by gas chromatography-mass spectroscopy (GC/MS) because bio-oil typically contains a large portion of lignin and carbohydrate derived components which are nonvolatile and undetectable by GC/MS (Mullen 2009). A combination of solvent fractionation, element (C,H,N) analysis and GC chromatography is more effective for chemical characterization of the whole bio-oil (Kai Sipilä 1998). For example, pine derived bio-oil can be extracted and divided into 6 fractions using water, ether, hexane and dichloromethane (DCM). Compound types and functional groups in different fractions analyzed by GC/MS, Py-GC/MS and CHN analysis (Venderbosch 2010) are shown in Table F-1.

Table F-1. Example Chemical Types in Pine Bio-oil

Fraction/Chemical groups	**Compound types**	**Wt% (wet basis)** Venderbosch 2010	**Wt% (wet basis),** Oasmaa 2010	**Wt% (dry basis),** Oasmaa 2010
Water solubles		75-85		
Acids, alcohols	Small acids, small alcohols	5-10	6.5 Total Acids = 4.3 Alcohols = 2.2	8.5 Total Acids = 4.3 Alcohols = 2.2
Ether-solubles	Catechols, syringols, guaiacols, aldehydes, ketones, furans and pyrans	5-15	15.4	20.3
Ether-insolubles	Sugars	30-40	34.4	45.3
Water	Water	20-30	23.9	-
Water insolubles		15-25		
Hexane-solubles	Extractives (High MW compounds with functional groups such as acids, alcohols)	2-6	4.35	5.7
DCM solubles	Stilbenes, Low MW lignin degraded compounds	5-10	13.4	17.7
DCM insolubles	High MW lignin degraded compounds	2-10	1.95	2.6

Bio-oil Model Compounds

A mixture of model compounds in Table F-2 is used to represent bio-oil. The model compounds were selected based on the analysis of wood derived bio-oil published in research literature. Compounds in the Aspen Plus and CHEMCAD databases were preferred over user defined compounds for which chemical and physical properties must be inserted in the simulation, and rely on limited data. The number of model compounds representing bio-oil was also minimized to reduce complexity of the model. However, key properties of the model compound mixture such as C,H,O,N,S element analysis, density, and heating value were matched with the bio-oil for thermodynamic consistency with experimental results.

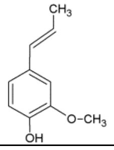
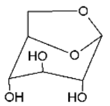
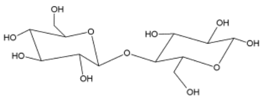
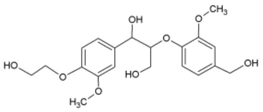
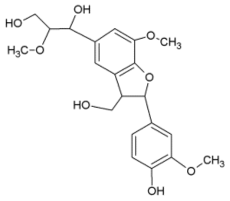
Table F-2. Model Compounds Used to Represent Bio-oil

Compound group	Model compound	Formula	% wt (dry basis)	Reference suggesting compounds, chemical structures or functional groups
Water soluble				
Acids	Crotonic acid	$C_4H_6O_2$	5.57	Thangalazhy-Gopakumar 2010
Alcohols	1,4-Benzenediol	$C_6H_6O_2$	6.97	Branca 2003; Garcia-Perez 2008; Djokic 2012
Ketones	Hydroxyacetone	$C_3H_6O_2$	8.29	Djokic 2012; Garcia-Perez 2008
Aldehydes	3-Methoxy-4-Hydroxybenzaldehyde	$C_8H_8O_3$	6.97	Djokic 2012; Garcia-Perez 2008; Liu 2012; Thangalazhy-Gopakumar 2010
Guaiacols	Isoeugenol	$C_{10}H_{12}O_2$	4.98	Branca 2003; DeSisto 2010; Thangalazhy-Gopakumar 2010; Kim 2011
Low MW sugars	Levoglucosan	$C_6H_{10}O_5$	5.97	Branca 2003; DeSisto 2010; Garcia-Perez 2008; Venderbosch 2010
High MW sugars	Cellobiose	$C_{12}H_{22}O_{11}$	33.86	Radlein in Bridgewater 2002
Water insoluble				
Low MW lignin derived compound A	Dimethoxy stilbene	$C_{16}H_{16}O_2$	10.95	Venderbosch 2010
Low MW lignin derived compound B	Dibenzofuran (representing diphenyl compounds)	$C_{12}H_8O$	2.21	Bayerbach 2006
Extractives	Dehydroabietic acid	$C_{20}H_{28}O_2$	2.99	DeSisto 2010; Venderbosch 2010
High MW lignin derived compound A	Oligomeric compounds with β-O-4 bond*	$C_{20}H_{26}O_8$	9.15	β-O-4 bond is the major type of linkage in wood derived lignin (Chu 2013)
High MW lignin derived compound B	Phenylcoumaran compounds*	$C_{21}H_{26}O_8$	1.99	Bayerbach 2006
Nitrogen compounds	2,4,6-trimethylpyridine	$C_8H_{11}N$	0.070	Pyridines are found as nitrogen containing compounds in bio-oil (Diebold in Bridgwater 1999 volume 1)
Sulfur compounds	Dibenzothiophene	$C_{12}H_8S$	0.025	Garcia-Perez 2008

*Standard name is not available. Molecular structure is shown in Table F-3 and drawn based on the properties and functional groups suggested in the literature [Bayerbach 2006; Chu 2013].

Most of the model compounds are available in the software databases, however, several are user defined compounds: levoglucosan, high MW sugars and low MW lignin derived compounds and high MW lignin. The properties are based on experimental data if available, or estimated by using the UNIFAC contribution group method and ACDlabs software. Properties of the user defined compounds are listed in Table F-3.

Table F-3. Molecular Structures, and Physical and Chemical Properties for User Defined Compounds

Compound	Formula and Molecular Structure	Properties							
		Estimated by ACDlabs software, unless noted otherwise.		Estimated by using the UNIFAC contribution group method, unless noted otherwise.					
		Tbp, F	Coefficients for vapor pressure equation[1]	Tc, F	Pc, psia	Vc, ft^3/ lbmol	Tm, F	ΔH_f, Btu/ lbmol	ΔG_f, Btu/ lbmol
Isoeugenol	C10H12O2 	507.2 (Ref 1)	A = 50.75 B = -9430.29 C = -3.46 D = -1.17E-6 E = 2 (Ref 2)	987.5	476.4	7.8	57.2 (Ref 3)	-88516.7	-16887.1
Levoglucosan	C6H10O5 	772.8	A = 149.33 B = -18972.6 C = -16.826 D = -4.57E-9 E = 2 (Ref 2)	1056	668.9	3.0	99.5 (Ref 2)	-412740 (Ref 5)	-255625
Cellobiose	C12H22O11 	1234.2	A = 53.01 B = -19999.5 C = -4.4294 D = -1.5E-7 E = 2	1281.5	340	6.8	462.2 (Ref 4)	-958499 (Ref 5)	-503968
High MW lignin derived compound A (Oligomeric compounds with β-O-4 bond)	C20H26O8 	1198.4	A = 81.08 B = -20000 C = -7.036 D = 2.279E-7 E = 2	1482.2	241.9	15.9	218.3	-498806	-297590
High MW lignin derived compound B (Phenyl-coumaran compounds)	C21H26O8 	1172.4	A = 85.84 B = -20000 C = -7.72 D = 3.84E-7 E = 2	1468.2	243.7	16.0	252.3	-496040	-284159

[1] $Vp = \exp[A + B/T + C\ln(T) + DT^E]$ When Vp = vapor pressure (Pa), T = temperature (K)

Ref 1 CRC 2013
Ref 2 Dortmund 2013
Ref 3 Chemicalland 2013
Ref 4 Chemical book 2013
Ref 5 Domalski 1972

The solid char from fast pyrolysis primarily consists of carbon. The char compositions are calculated by different in the design and are comparable to the experimental results as presented in Table F-4. The char heating value is calculated by using this equation (H.H. Schobert, Lignites of North America, pp 507):

$$\text{HHV (Btu/lb)} = 146.58C + 568.78H - 51.53(O+N) - 6.58Ash + 29.45S$$

Table F-4. Characteristics of Solid Char from Fast Pyrolysis

From	DeSisto 2010 Pyrolysis of pine sawdust (fluidized bed)			Kim 2012 Fast pyrolysis of pitch pine (fluidized bed)			Gregoire 1995 Vortex reactor, wood chips	Base Case
	400°C	**500°C**	**600°C**	**300°C**	**400°C**	**500°C**	**500°C**	**500°C**
Ash (%)	1.45	1.69	3.51	4.5 ± 0.33	7.9 ± 0.54	7.7 ± 0.80	2.87	7.67
C, %	70.9	74.1	83.4	63.9 ± 2.04	70.7 ± 2.33	90.5 ± 1.13	81.8	83.03
H, %	4.7	4.4	3.1	5.4 ± 0.34	3.4 ± 0.09	2.5 ± 0.10	3.7	1.14
N, %	<0.5	<0.5	<0.5	0.3 ± 0.03	0.4 ± 0.05	0.3 ± 0.04	0.10	1.37
O, %	23.0	21.5	10.0	30.4 ± 1.73	25.5 ± 2.47	6.7 ± 1.20	11.5	6.56
O:C	0.320	0.290	0.120				0.141	0.08
Total	100.6	102.2	100.5				99.87	100
HHV (MJ/kg)	28.5	28.05	29.91					28.81

Hydrotreated Oil Model Compounds

The following describes the criteria for choosing the compounds used to represent hydrotreated bio-oil. Hydrotreating of bio-oil occurs in three stages; stabilization bed, 1st stage hydrotreating reactor and 2nd stage hydrotreating reactor. The hydrotreating reactions of bio-oil are very complex. Published experimental data were generated from different reactor configurations and a wide range of process conditions.

Tables F-5, shows the mixture of model compounds used to represent the hydrotreated oil from the stabilizer and the first reactor. The intermediates in the upgrading of pyrolysis oil are selected based on the reactivity scale of oxygenated groups in Figure 6 (Elliott 2007) and the study by Vispute 2009.
The compounds and properties of the upgraded oil from the second reactor are shown in Tables F-6 and F-7. GC/MS data were used to guide the selection. Figure F-1 shows experimentally derived distillation curves that were used to help select the final fuel compounds based on their boiling points.

Table F-5. Model Compounds Representing Mild Hydrotreating Products

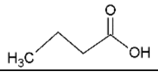
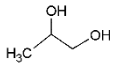
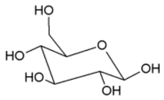
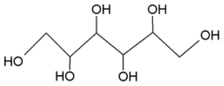
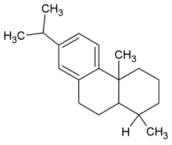
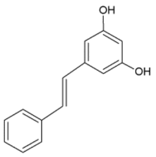
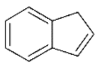
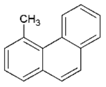
Compound	Formula	Molecular structure	Hydrotreating intermediates from
N-Butyric Acid	C4H8O2		Small acids
1,2-Propanediol	C3H8O2		Small aldehydes, alcohols, ketones
Glucose	C6H12O6		Sugars (Vispute 2009)
Sorbitol	C6H14O6		Sugars (Vispute 2009)
HT dehydroabietic acid	C19H28		Hydrogenation of extractive compounds
Pinosylvin	C14H12O2		Low MW derived lignin compounds
Indene	C9H8		High MW derived lignin compounds
4-methylphenanthrene	C15H12		High MW derived lignin compounds

Table F-6. Upgraded Oil Model Compounds

Compound group	Model compound	Formula	% wt in upgraded oil (dry basis)
Normal paraffins	Hexane	C6H14	10.61%
	Dodecane	C12H26	6.11%
	Pentane	C18H38	3.88%
Iso-paraffins	3-Methylhexane	C7H16	3.89%
	4-methylnonane	C10H22	6.15%
Cyclopentanes	Cyclopentane, ethyl	C7H14	3.79%
	1-methyl-1-ethylcyclopentane	C8H16	5.03%
Cyclohexanes	Cyclohexane	C6H12	4.86%
	Cylcohexane, butyl-	C10H20	3.67%
	1,1-Bicyclohexyl	C12H22	3.37%
Cyclo C7+	1,3-dimethyladamantane	C12H20	3.44%
Aromatics	o-xylene	C8H10	7.90%
	Benzene, 1-ethenyl-4-ethyl-	C10H12	3.34%
Heavies	4-methylphenanthrene	C15H12	6.45%
	Pyrene	C16H10	6.24%
Diphenyl	1,2-Diphenylethane	C14H14	1.08%
Indanes and indenes	Indane	C9H10	1.72%
	1H-Indene, 1,2,3-trimethyl-	C12H14	1.29%
1,2,3,4 Naphthalene	1-n-Hexyl-1,2,3,4-Tetrahydronaphthalene	C16H24	2.51%
Naphthalenes	Naphthalene, 2,7-dimethyl	C12H12	4.31%
PNAs	Naphthalene, 1-phenyl-	C16H12	5.32%
Oxygenates	5-Methyl-2-(1-methylethyl)phenol	C10H14O	4.86%
Nitrogen compounds	2,4,6-trimethylpyridine	C8H11N	0.157%
Sulfur compounds	Dibenzothiophene	C12H8S	0.025%

Table F-7. Hydrotreated Oil Element Balances and Properties

Characteristic	Elliott 2013	Elliott 2012	Base case
Moisture content, wt%	0.1	0.01-0.03	-
Carbon, wt%	87.5	83.5-87.3	87.7
Hydrogen, wt%	12	11.2-13.6	11.8
Oxygen, wt%	0.5	0.2-0.3	0.52
Nitrogen, wt%	<0.05	<0.05	0.02
Sulfur, wt%	<0.005	<0.005-0.014	0.004
HHV, MJ/kg	45	N/A	44.3
Density, g/ml	0.84	0.76-0.87	0.848

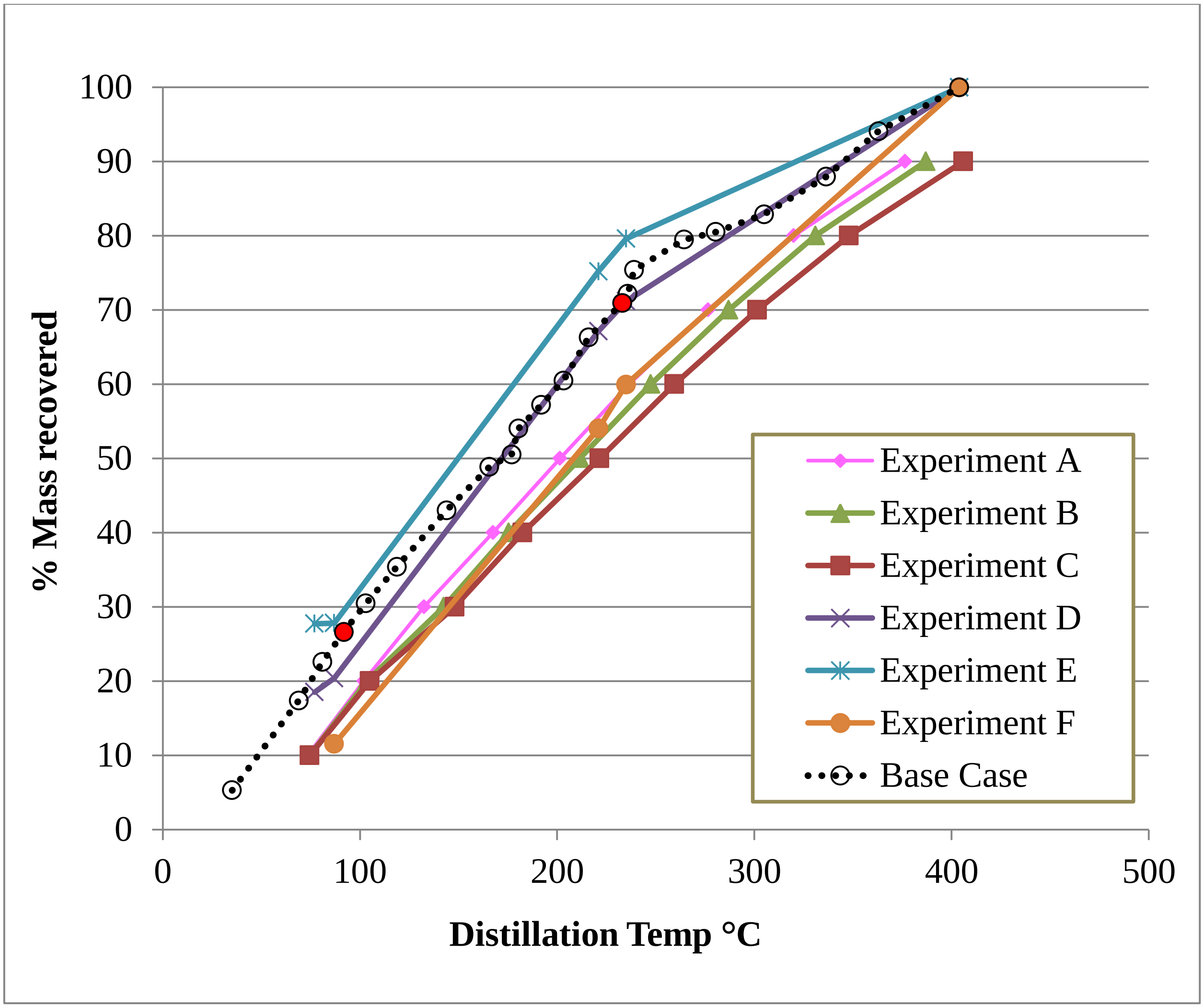


Figure F-1. Hydrotreated Oil Distillation Curve by ASTM D2887

Each model compound in the process simulation is used to represent a group of compounds having similar physical or chemical properties. Table F-8 lists of all model compounds and their generic names used in this work.

Table F-8. Model Compounds and Their Generic Names

Generic names	Model compounds used to represent groups of similar compounds
1,2,3,4-Naphthalenes	6-Hexyltetralin
Acids	Crotonic acid
Acids Intermediates	N-Butyric acid
Alcohols	1,4-Benzenediol
Aldehydes	3-Methoxy-4-Hydroxybenzaldehyde
Aromatic A	O-Xylene
Aromatic B	Benzene, 1-ethenyl-4-ethyl-
Ash	Ash
Bicyclos	1,1-Bicyclohexyl
Carbon	Carbon
Carbonyl intermediates	1,2-Propanediol
Extractive intermediates	HT Dehydroabietic acid (see Table F-5)
Cyclo C5 A	Ethyl-cyclopentane
Cyclo C5 B	1-methyl-1-ethylcyclopentane
Cyclo C6 A	Cyclohexane
Cyclo C6 B	N-Butylcyclohexane
Cyclo C6-HCK product	Methyl-cyclohexane
Cyclo C7+	1,3-Dimethyladamantane
Diphenyl compounds	1,2-Diphenylethane
Extractives	Dehydroabietic acid
Guaiacols	Guaiacol
High MW lignin compound intermediates	Indene
High MW lignin derived A	High MW lignin derived compound A (see Table F-3)
High MW lignin derived B	High MW lignin derived compound B (see Table F-3)
High MW sugars	Cellobiose (see Table F-3)
Indans	Indan
Indenes	1H-Indene, 1,2,3-trimethyl-
Iso-paraffin A	3-Methylhexane
Iso-paraffin B	4-Methylnonane
Ketones	Acetol
Low MW lignin compound intermediates	Pinosylvin
Low MW lignin derived B	1,2-Diphenylethane
Low MW sugars	Levoglucosan
Medium chain n-paraffins	N-Dodecane
Naphthalenes	2,7 Dimethylnapthalene
Nirogen compounds	Trimethyl Pyridine
N-paraffin-HCK product A	N-Nonane
N-paraffin-HCK product B	N-Octane
Phenanthrenes	4-Methylphenanthrene
Phenols	Thymol
Polynuclear aromatics	Naphthalene, 1-phenyl-
Pyrenes	Pyrene
Sand	Silicon Dioxide
Short chain n-paraffins	N-Hexane
Sugar intermediate A	Alpha-D-Glucose
Sugar intermediate B	Sorbitol
Sulfur agent	Dimethyl Sulfide
Sulfur compounds	Dibenzofuran

Appendix G: Indices

Table G-1 Labor Indices

Source: Bureau of Labor Statistics
Series ID: CEU3232500008 Chemicals
Average Hourly Earnings of Production Workers
Current indices at http://data.bls.gov/cgi-bin/srgate

YEAR	INDEX	YEAR	INDEX
2002	17.97	2007	19.55
2003	18.50	2008	19.50
2004	19.17	2009	20.30
2005	19.67	2010	21.07
2006	19.60	2011	21.46

Table G-2 Capital Cost Indices

Chemical Engineering Magazine, CEI annual index
2012 and 2013 regressed from previous year data (DOE 2013)

YEAR	INDEX	YEAR	INDEX
1990	357.6	2002	395.6
1991	361.3	2003	402.0
1992	358.2	2004	444.2
1993	359.2	2005	468.2
1994	368.1	2006	499.6
1995	381.1	2007	525.4
1996	381.7	2008	575.4
1997	386.5	2009	521.9
1998	389.5	2010	550.8
1999	390.6	2011	585.7
2000	394.1	*2012*	*617.6*
2001	394.3	*2013*	*649.5*
2000	394.1		

Table G-3 Inorganic Chemical Indices

Source: SRI International Chemical Economics Handbook, Economic Environment of the Chemical Industry, September 2006
Current indices at https://www.sriconsulting.com/CEH/Private/EECI/EECI.pdf

YEAR	INDEX	YEAR	INDEX
1990	123.6	2001	158.4
1991	125.6	2002	157.3
1992	125.9	2003	164.6
1993	128.2	2004	172.8
1994	132.1	2005	187.3
1995	139.5	2006	196.8
1996	142.1	2007	203.3
1997	147.1	2008	228.2
1998	148.7	2009	224.7
1999	149.7	2010	233.7
2000	156.7	2011	249.3

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