Process Design and Economics for the Conversion of Lignocellulosic Biomass to Hydrocarbons via Indirect Liquefaction

Thermochemical Research Pathway to High-Octane Gasoline Blendstock Through Methanol/Dimethyl Ether Intermediates

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Executive Summary

This report was developed as part of the U.S. Department of Energy's Bioenergy Technologies Office's (BETO's) efforts to enable the development of technologies for the production of infrastructure-compatible, cost-competitive liquid hydrocarbon fuels from lignocellulosic biomass feedstocks. The research funded by BETO is designed to advance the state of technology of biomass feedstock supply and logistics, conversion, and overall system sustainability. It is expected that these research improvements will be made within the 2022 timeframe. As part of their involvement in this research and development effort, the National Renewable Energy Laboratory and the Pacific Northwest National Laboratory investigate the economics of conversion pathways through the development of conceptual biorefinery process models and techno-economic analysis models.

This report describes in detail one potential conversion process for the production of high-octane gasoline blendstock via indirect liquefaction of biomass. The processing steps of this pathway include the conversion of biomass to synthesis gas or syngas via indirect gasification, gas cleanup, catalytic conversion of syngas to methanol intermediate, methanol dehydration to dimethyl ether (DME), and catalytic conversion of DME to high-octane, gasoline-range hydrocarbon blendstock product. The conversion process configuration leverages technologies previously advanced by research funded by BETO and demonstrated in 2012 with the production of mixed alcohols from biomass. Biomass-derived syngas cleanup via reforming of tars and other hydrocarbons is one of the key technology advancements realized as part of this prior research and 2012 demonstrations. The process described in this report evaluates a new technology area for the downstream utilization of clean biomass-derived syngas for the production of high-octane hydrocarbon products through methanol and DME intermediates. In this process, methanol undergoes dehydration to DME, which is subsequently converted via homologation reactions to high-octane, gasoline-range hydrocarbon products.

The process configuration presented in this design report possesses similarities to methanol-to-gasoline (MTG) technologies such as the process licensed by ExxonMobil Research and Engineering. However, the technology presented herein differs from conventional MTG processes with respect to the quality of hydrocarbons produced and the severity of process conditions. The conventional MTG process generates a product mix with high concentrations of aromatic hydrocarbon compounds. The product mixture possesses an average octane number ((RON+MON)/2) of approximately 87 and yield of gasoline blendstock is limited by production of light hydrocarbon coproducts and coke formation. In contrast, the chemistry proposed herein and supported by preliminary experimental results produces a gasoline blendstock with high concentration of iso-paraffinic compounds with an expected average octane number of >93 and low concentrations of aromatics. Additionally, coke formation has not been observed during initial experiments, although some coke formation is expected with extended on-stream operations. The reduced coking potential can result in higher yields and improved carbon utilization. With biomass feedstock being the single most expensive component of the process, higher yields per ton of biomass can make a significant positive impact on the process economics. With respect to the process conditions, the hydrocarbon synthesis reactor in the ExxonMobil MTG process operates in the 650°F–950°F range and is limited by reactor configuration. MTG reactor operations must be controlled to limit per-pass conversions to prevent runaway in a fixed-bed, shell and tube reactor, or a reactor system must be designed to

iii

utilize a fluidized bed reactor for efficient heat management to allow high single-pass conversions. Conversely, the hydrocarbon synthesis reactor conditions for the design presented here are less severe, operating in the temperature range of 350°F–450°F. In addition, the reactor system can operate at higher per-pass conversions using multiple fixed-bed reactors with inter-bed cooling. By operating at lower temperature, coke formation is significantly reduced. Thus, catalyst bed regeneration can be less frequent, which results in the potential to reduce capital and operating costs.

The conceptual design presented here considers the economics of high-octane gasoline blendstock production, under the assumption of achieving the described product selectivities and conversions through near-term research efforts. The design features a processing capacity of 2,205 U.S. tons (2,000 metric tonnes) of dry biomass per day and a gasoline yield of 64.9 gallons per dry U.S. ton of feedstock. The gasoline selling price corresponding to this design (2022 target) is $3.25 per gallon of blendstock (or $3.41/gallon of gasoline equivalent) in 2011 dollars, assuming a 30-year plant life and 40% equity financing with a 10% internal rate of return and the remaining 60% debt financed at 8% interest. A summary of the techno-economic analysis results for this pathway is presented in the following table.

Process Engineering Analysis for High Octane Gasoline via Indirect Gasification and Methanol Intermediate
(2022 Target Case)

2,000 Dry Metric Tonnes Biomass per Day
Indirect Gasifier, Tar Reformer, Sulfur Removal, Methanol Synthesis, Hydrocarbon Synthesis on Beta-Zeolite Catalyst, Fuel Purification, Steam-Power Cycle
All Values in 2011 US$

Gasoline Minimum Fuel Selling Price (MFSP)	**$3.25 per Gallon**
MFSP (Gasoline-Equivalent Basis)	**$3.41 per GGE**
Contributions: Feedstock Costs	1.232 per Gallon
Operating Costs & Credits	0.545 per Gallon
Capital Charges & Taxes	1.471 per Gallon
Gasoline Production at Operating Capacity	47.0 MM Gallons per Year
Gasoline Product Yield	64.9 Gallons per Dry US Ton Feedstock
Delivered Feedstock Cost	$80.00 per Dry US Ton

Capital Costs

Feed Handling & Drying	$200,000
Gasification	$50,500,000
Gas Cleanup	$70,700,000
Methanol Synthesis	$37,300,000
Methanol Conditioning	$3,400,000
DME & Hydrocarbons Conversion	$38,700,000
Gasoline Separations	$5,400,000
Steam System & Power Generation	$38,100,000
Cooling Water & Other Utilities	$7,000,000
Total Installed Equipment Cost (TIC)	$251,300,000
ISBL (Areas A100 to A500, A1400, A1500)	$206,100,000
OSBL (Areas A600, A700)	$45,100,000
Other Direct Costs	8,200,000
(% of ISBL)	4.0%
Total Direct Costs (TDC)	259,500,000
Indirect Costs	155,700,000
(% of TDC)	60.0%
Land Purchase Cost	1,600,000
Working Capital	20,800,000
Total Capital Investment (TCI)	437,500,000
Installed Equipment Cost per Annual Gallon	$4.81
Total Capital Investment per Annual Gallon	$8.37
Debt Financing (% of Investment)	60.0%
Loan Interest Rate	8.0%
Loan Term (years)	10.0
Equity Financing (% of Investment)	40.0%
Internal Rate of Return (After-Tax)	10.0%
Plant Operating Hours per year	7,884
On-Stream Percentage	90.0%

Process Efficiency

Gasifier Efficiency - HHV %	72.9
Gasifier Efficiency - LHV %	72.5
Efficiency to Gasoline - HHV %	45.3
Efficiency to Gasoline - LHV %	45.0
Overall Plant Efficiency - HHV %	45.3
Overall Plant Efficiency - LHV %	45.0

Annual Operating Costs

Feedstock	$57,900,000
Natural Gas	$0
Catalysts	$10,700,000
Olivine	$500,000
Other Raw Matl. Costs	$1,400,000
Waste Disposal	$600,000
Electricity Transfer Charge	$0
Electricity	$0
Fixed Costs	$20,600,000
Coproduct credits	$0
Capital Depreciation	$13,800,000
Average Income Tax	$8,100,000
Average Return on Investment	$47,200,000

Operating Costs per Product	(¢/MMBtu)	(¢/Gal)
Feedstock	1112.6	123.2
Natural Gas	0.0	0.0
Catalysts	48.5	5.4
Olivine	10.3	1.1
Other Raw Materials	26.4	2.9
Waste Disposal	10.8	1.2
Electricity Transfer	0.0	0.0
Electricity	0.0	0.0
Fixed Costs	396.5	43.9
Coproduct credits	0.0	0.0
Capital Depreciation	265.0	29.4
Average Income Tax	156.5	17.3
Average Return on Investment	906.8	100.4
Total (Plant Gate Price)	2933.4	324.9

Power Balance	(KW)	(hp)
Total Plant Power Consumption	36,875	49,451
Power Generated Onsite	36,879	49,456
Power Imported from Grid	0	0
Power Exported to Grid	4	5

Power Generation	(KW)	(hp)
Steam Turbine Generators	35,344	47,397
Process Gas Turboexpander(s)	1,535	2,059

Sustainability Metrics

Plant Electricity Consumption (KWh/ Gal Gasoline)	6.2
Gasification & Reforming Steam (lb / Gal Gasoline)	21.8
Water Consumption (Gal Water / Gal Gasoline)	1.7
Carbon Conversion Efficiency (C in Fuel/C in Feedstock)	31.2%
Fossil GHG Emissions (g CO_{2-e}/MJ Fuel)	0.6
Fossil Enegy Consumption (MJ Fossil Energy/MJ Fuel)	0.006

Feedstock Rate and Cost

Feed Rate	Dry Tonnes / Day	2,000
	Dry US Tons / Day	2,205
Feedstock Cost	$ / Dry Ton	$80.00
	$ / Moisture & Ash Free Ton	$80.74

v

Acknowledgements

We thank the U.S. Department of Energy's Bioenergy Technologies Office for funding and supporting this work, and the Idaho National Laboratory for providing feedstock logistics information. We also thank the reviewers of this report for their valuable comments. Reviewers are listed below by company/organization names. Note that "review" does not imply endorsement of the analysis presented by either individuals or companies/organizations. Reviewer comments on the draft version of the report are presented in Appendix K along with National Renewable Energy Laboratory authors' responses.

List of reviewers

Argonne National Laboratory	Felix Adom
	Jennifer Dunn
The Dow Chemical Company	Brien Stears
Gas Technology Institute	Rick Knight
Haldor Topsoe, Inc.	Anders Olsen
	Henrik Rasmussen
	Esben Sorensen
	Henrik Udesen
	Niels Udengaard
Idaho National Laboratory	Jake Jacobson
National Renewable Energy Laboratory	Mark Davis
	Richard Bain (retired)
Pacific Northwest National Laboratory	Rob Dagle
	Sue Jones
	Karthi Ramasamy
Phillips 66	Alex Platon
Renewable Energy Institute International	Dennis Schuetzle
Sundrop Fuels, Inc.	Andrew Argo
VTT Technical Research Centre of Finland	Ilkka Hannula

Table of Contents

List of Figures

List of Tables

List of Acronyms

AGR	acid gas removal
ASTM	American Society of Testing and Materials
bbl	barrel = 42 gallons
BCL	Battelle Columbus Laboratory
BETO	Bioenergy Technologies Office
BFW	boiler feed water
BLM	Biomass Logistics Model
BPD	barrels per day
Btu	British thermal unit
C&D	construction and demolition
CFM	cubic feet per minute
CO_{2-e}	carbon dioxide equivalent
DCFROR	discounted cash flow rate of return
DEPG	dimethyl ether of polyethylene glycol
DME	dimethyl ether
DOE	U.S. Department of Energy
EDI	electrodeionization
EIA	Energy Information Administration
EISA	Energy Independence and Security Act
FCC	fluid catalytic cracker
FCI	fixed capital investment
FT	Fischer-Tropsch
GGE	gallon gasoline equivalent
GHG	greenhouse gas
GHSV	gas hourly space velocity
GJ	gigajoule
gpm	gallons per minute
HHV	higher heating value
HMB	hexamethylbenzene
hp	horsepower
IDL	indirect liquefaction
INL	Idaho National Laboratory
IPE	Icarus Process Evaluator (software from Aspen)
IRR	internal rate of return
ISBL	inside battery limits
kWh	kilowatt-hour
LCA	life-cycle assessment
LHV	lower heating value
LPG	liquefied petroleum gas
MAF	moisture and ash free
MDEA	methyl diethanolamine
MFSP	minimum fuel selling price
MM	million
MON	motor octane number
MTBE	methyl-tertiary-butyl ether
MTG	methanol-to-gasoline
MTPD	metric tonnes per day
MW	megawatts
NREL	National Renewable Energy Laboratory
OSBL	outside battery limits
Pa	pascal
PFD	process flow diagram
PGP	plant gate price
PNNL	Pacific Northwest National Laboratory
ppmv	parts per million by volume
ppmw	parts per million by weight
POX	partial oxidation
PSA	pressure swing adsorption
psia	pounds per square inch (absolute)
psig	pounds per square inch (gauge)
R&D	research and development
RFS	Renewable Fuel Standard
RO	reverse osmosis
RON	research octane number
SCR	steam-to-carbon ratio
SOT	state of technology
TCI	total capital investment
TDC	total direct cost
TCPDU	thermochemical conversion pilot demonstration unit
TEA	techno-economic analysis
TIC	total installed cost
TPD	short (U.S.) tons per day
TPEC	total purchased equipment cost
WGS	water-gas-shift
WWT	wastewater treatment

1 Introduction

1.1 Motivation and Background

There is a current emphasis on the conversion of biomass to cost-competitive, liquid hydrocarbon fuels via direct and indirect liquefaction pathways. This report summarizes the analysis performed by the National Renewable Energy Laboratory (NREL) and Pacific Northwest National Laboratory (PNNL) in support of the indirect liquefaction (IDL) platform to identify opportunities and research gaps for conversion pathways of interest to the Bioenergy Technologies Office (BETO). The fuel production pathway analyzed in this report is syngas to methanol followed by the synthesis of high-octane gasoline. This modeled process includes the following highlights:

- Produces a high-octane, gasoline-range, hydrocarbon product.

- Leverages advances in syngas production and clean-up technologies developed for the thermochemical pathway from biomass to mixed alcohols, demonstrated in 2012.

- Operates at lower severity conditions (lower operating temperature and pressure) relative to traditional methanol-to-gasoline (MTG) processes.

- Potentially combines homologation and conversion of recycled C_4- olefins and iso-paraffins to larger products in a single reactor system and does not require a separate alkylation step. This can reduce the overall capital cost relative to MTG processes if alkylation of C_3 and C_4 olefins is included in the process.

- Possesses high selectivity to C_7 (mostly 2,2,3-trimethylbutane and 2,2,3-trimethyl-but-1-ene) and C_4 paraffins and olefins, which can be recycled to maximize yield of gasoline blendstock product.

- Produces low aromatic-content product and has a low coke formation rate. Benefits associated with a low aromatic-content gasoline product include higher volume expansion (due to lower density) and improved fuel quality relative to fuel specification (resulting in higher blending potentials and premium sale price).

- Utilizes metal-modified beta-zeolite catalysts for hydrocarbon synthesis that can easily be scaled for commercial processes.

Important development needs for this pathway include improved understanding of the lifetime, stability, and maintenance of the catalysts, demonstration of high yields at larger scales, selectivity tuning for the desired product slate, and reduction of costs through process consolidation and integration. Overall, this report captures envisioned process performances to be achieved for cost-competiveness by 2022. Intermediate research targets leading up to 2022 were developed subsequently for publication in BETO's Multi-Year Program Plan, using key assumptions in this report as the foundation; those metrics will be updated annually with research realities, and necessary deviations from the current design basis.

While the current conceptual process configuration possesses similarities to conventional MTG processes, it differs from the conventional MTG process in two key aspects: (1) the quality of the hydrocarbon products and (2) the severity of hydrocarbon synthesis process conditions.

1

Hydrocarbon Products: The ExxonMobil MTG process produces a mix of gasoline-range, hydrocarbon compounds with a high concentration of aromatics with byproducts durene (1,2,4,5-tetramethylbenzene), light hydrocarbon coproduct, and coke. The gasoline product mixture has a motor octane number (MON) of ~83, a research octane number (RON) of ~92, and an average octane number ((RON+MON)/2) of approximately 87 [1]. Fuel standards are changing to limit the amount of aromatic compounds allowed in gasoline [2], which means a highly aromatic bio-gasoline blendstock could experience limited blending potential and lower relative product values. The chemistry proposed herein produces a low-aromatic, branched-paraffin product that has a MON of 90+, a RON of 95+, and an expected average octane number ((RON+MON)/2) of greater than 93 [3]. In addition, coke formation has not been observed during initial experiments, although some coke formation is expected given the chemistry and the assumption of multi-year catalyst cycles for the process design. The reduced coking potential relative to conventional MTG can result in higher yields. With biomass feedstock being the single most expensive component of the process, higher yields per ton of biomass can have a significant positive impact on the process economics.

Hydrocarbon Synthesis Process Conditions: The hydrocarbon-producing reactor in the ExxonMobil MTG process operates in the range 650°F–950°F and the yield is limited by reactor configuration. Operations are limited to lower per-pass conversions to prevent runaways in a fixed-bed shell and tube exchange reactor or fluidized bed reactors are specified in the process design for efficient heat management to allow high single-pass conversions. Conversely, for the process configuration presented in this design report, the reactions occur at 350°F–450°F and can be operated at high per-pass conversions using reactors with inter-stage cooling. Additionally, by operating at the lower temperature, coke formation is reduced, resulting in less frequent catalyst bed regenerations and the potential for reduced capital and operating costs.

In order to assess the economic viability and potential of this conversion process, NREL, PNNL, and its partners have developed a process model with economic projections to estimate the minimum fuel selling price (MFSP). The MFSP can be used by policymakers and BETO to help refine research objectives and associated performance targets that will be necessary to produce cost-competitive fuels from biomass. Techno-economic analysis also facilitates direct biomass conversion research by examining the sensitivity of the MFSP to process alternatives and projected research advances, as well as to track research progress via state of technology (SOT) assessments.

This report builds upon a previous design reports released in 2011 [4, 5], written by engineers from NREL with engineering support provided by Harris Group, Inc. For the current report, NREL and PNNL worked with Harris Group Inc. to provide engineering support through estimates and reviews of the equipment and raw material cost assumptions used in the process design and economic analysis. Thus, the economics of this conceptual process assumes the best available equipment and raw material costs and an "n[th]-plant" project cost structure and financing.

Because the heat and power requirements of the process cannot be met through char combustion and combustion of available fuel gas and other process off-gases alone, it was decided that some raw syngas from indirect gasifier could be diverted from liquid fuel production for heat and power production. This option makes the design energy self-sufficient; however, it also lowers

2

the overall product yield. In designs that result in excess heat or power production, energy in the form of electricity will be exported to the grid for a coproduct credit.

The analysis presented here also includes consideration of the life-cycle implications of the baseline process model, by tracking sustainability metrics for the modeled biorefinery, including greenhouse gas (GHG) emissions, non-renewable energy or fossil energy consumption, and consumptive water use.

1.2 Techno-Economic Analysis Approach

The general approach used in the process design, process model, and economic analysis is depicted in Figure 1. This design report includes information from completed and ongoing research at NREL and other national laboratories, detailed reviews of commercially available technologies, process modeling using Aspen Plus process modeling software, equipment cost estimation through vendor quotes and Aspen Icarus software, and discounted cash flow analysis. The techno-economic analysis ultimately provides a MFSP based on the financial assumptions in Table 1.
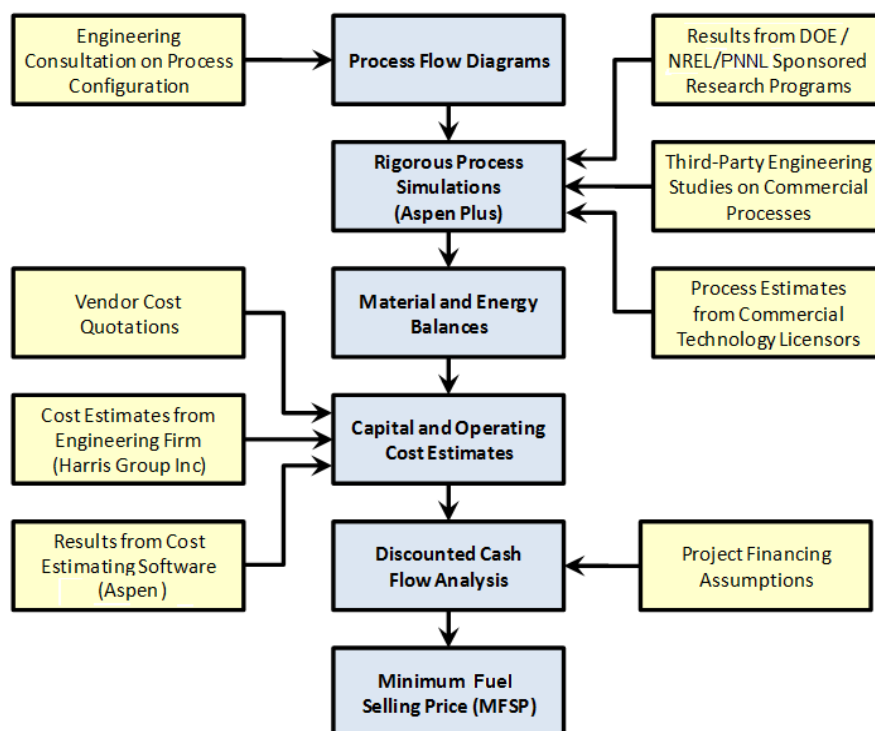
Figure 1. Approach for process design and techno-economic analysis

Aspen Plus Version 7.2 was used to develop mass and energy balances for the process. The plant operations are separated into nine major process areas as follows:

- Area 100: Feed handling and preparation
- Area 200: Gasification

3

- Area 300: Synthesis gas cleanup and syngas compression

- Area 400: Acid gas removal and methanol synthesis

- Area 500: Methanol conditioning

- Area 1400: Methanol to high-octane gasoline synthesis

- Area 1500: Product recovery

- Area 600: Steam system and power generation

- Area 700: Cooling water and other utilities.

1.3 Definition of nth-Plant Economics

The techno-economic analysis reported here uses n^{th}-plant economics. The key theoretical assumption associated with n^{th}-plant economics is that several plants using the same technology have already been built and are operating. In other words, the assumption reflects a future in which a successful industry has been established with many operating plants. Because the techno-economic model is a tool used primarily for (1) studying new process technologies or (2) comparing integrated schemes in order to comment on their relative economic impact, it is prudent to ignore artificial inflation of project costs associated with risk financing, longer startups, equipment overdesign, and other costs associated with pioneer plants, as these overshadow the real economic impact of advances in conversion science or process engineering research. At the very least, n^{th}-plant economics should help to provide justification and support for early technology adopters and pioneer plants.

Because equipment costs in this design report have been estimated explicitly, the nth-plant assumptions apply primarily to the factored cost model used to determine the total capital investment from the purchased equipment cost and to the assumptions applied for plant financing. The nth-plant assumption also applies to operating parameters, such as process uptime and start-up time. A summary of the nth-plant assumptions applied in this report are listed in Table 1. These financial assumptions are consistent with assumptions used for other economic analyses done for BETO, with some deviations to reflect the uniqueness of this process.

4

Table 1. Summary of nth-Plant Assumptions for Techno-Economic Analysis

Description of Assumption	Assumed Value
Internal rate of return (IRR)	10%
Plant financing by equity/debt	40%/60% 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 purchase cost)
Depreciation schedule[a]	7-year MACRS schedule [6]
Construction period (spending schedule)	3 years (8% Y1, 60% Y2, 32% Y3)
Plant salvage value	No value
Start-up time	6 months
Revenue and costs during startup	Revenue = 50% of normal
	Variable costs = 75% of normal
	Fixed costs = 100% of normal
On-stream percentage after startup	90% (7,884 operating hours per year)

[a] Capital depreciation is computed according to the IRS Modified Accelerated Cost Recovery System (MACRS). Because the plant described here is not a net exporter of electricity, the steam plant and power generation equipment are not depreciated over a 20-year recovery period, according to the IRS. The whole plant capital is depreciated over a 7-year recovery period.

1.4 Estimation of Capital Costs

Capital costs were estimated using a variety of resources. For sub-processes that utilize well-developed technologies and can be purchased as modular packages (e.g., acid gas removal system), an overall package cost was used instead of the sum of costs for individual pieces of equipment. Costs for common process equipment (e.g., tanks, drums, pumps, and simple heat exchangers) were estimated largely based on costs reported in the earlier design reports [4, 5] which in turn were estimated using Aspen IPE 2006.5 costing software. Unit operations that are specific to intermediate and fuel production (e.g., gasifier, tar reformer, methanol synthesis reactor, dimethyl ether (DME) synthesis reactor, and high-octane gasoline synthesis reactor) were priced through vendor quotations. It is noteworthy that capital cost estimates for pre-commercial technologies like the gasifier and tar reformer were derived from current vendor quotes for first-of-its-kind fabrications, which should be higher than costs for mature nth-plant technologies. Some of the unit operations have been discussed in the previous design reports.

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

$$\text{Scale-Up Equipment Cost} \; = \; \text{Base Equipment Cost} \left(\frac{\text{Scale-Up Capacity}}{\text{Base Capacity}} \right)^n \qquad \text{(Eq. 1)}$$

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

- Vendors' estimates of scaling exponent or inference from vendor quotes when multiple quotes were available for equipment of various processing capacities.

- Development of correlations by multiple estimates from Aspen IPE software.

- Standard reference from published sources such as Garrett [7], Peters, Timmerhaus, and West [8], and Perry et al. [9].

When cost data were not available in 2011 dollars, costs were adjusted with Chemical Engineering's (CE) Plant Cost Index [10] using Equation 2:

$$\text{Corrected Equipment Cost} \; = \; \text{Base Equipment Cost} \left(\frac{\text{2011 Cost Index Value}}{\text{Base Year Cost Index Value}} \right) \qquad \text{(Eq. 2)}$$

The CE indices used in this study are listed and plotted in Figure 2. The index data show a sharp increase after 2003 due to increases in global steel demand and a dip in 2009 due to the global recession. An intermediate 2014 index was used for 2014 corresponding to roughly the time when most of the recent cost quotes were obtained.

Year	Index	Year	Index
1990	357.6	2003	402.0
1991	361.3	2004	444.2
1992	358.2	2005	468.2
1993	359.2	2006	499.6
1994	368.1	2007	525.4
1995	381.1	2008	575.4
1996	381.7	2009	521.9
1997	386.5	2010	550.8
1998	389.5	2011	585.7
1999	390.6	2012	584.6
2000	394.1	2013	567.3
2001	394.3	2014	573.9
2002	395.6	(Mid-Year)	
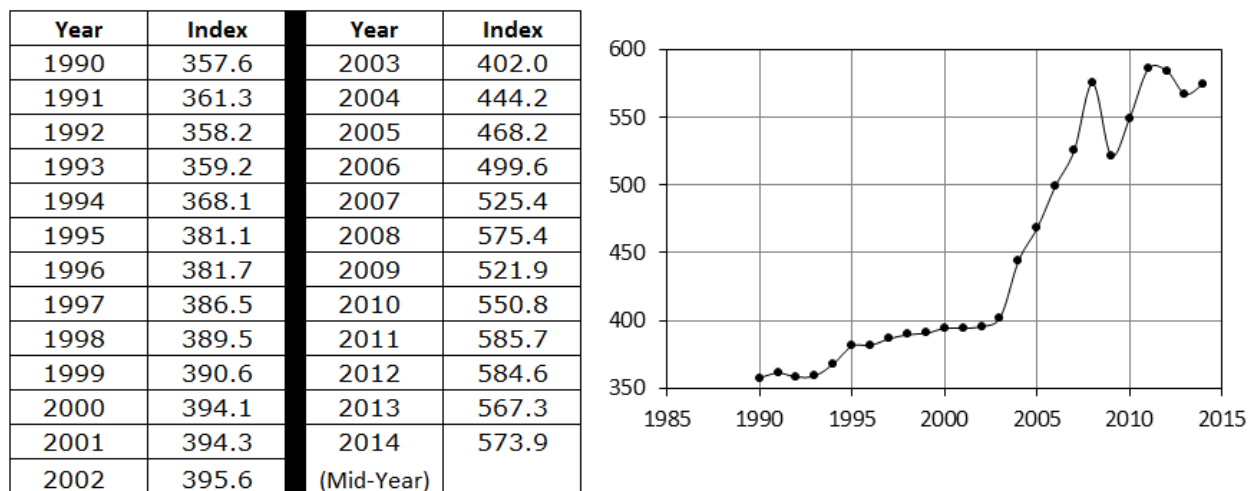
Figure 2. *Chemical Engineering's* **Plant Cost Index data**

6

Once the total purchased equipment costs (TPEC) were determined, scaled, and time-corrected, an equipment installation factor was applied to estimate the total installed cost (TIC) for the equipment or process unit including associated piping, instrumentation and controls, electrical systems, buildings, yard improvements, and direct labor per Equation 3, where $f_{Installation}$ is the installation factor:

$$Total\ Installed\ Cost\ (TIC)\ =\ f_{Installation}\ *\ Total\ Purchased\ Equipment\ Cost\ (TPEC) \qquad \text{(Eq. 3)}$$

Specific installation factors from vendor quotes or published data were applied to equipment or process units for the estimation of the TIC. When specific data was not available for equipment or package units, a default installation factor of 2.47 was applied to determine TIC. This value is based on the direct cost factor method presented by Peters, Timmerhaus, and West [8] for a solid-fluid processing plant. Details for TPEC, installation factors, and TIC values for each process area are presented in Section 4. Harris Group Inc. prepared a detailed list of direct costs included in the equipment installation factors.

Once the scaled TICs were determined, overhead and contingency factors were applied to determine a total capital investment (TCI) cost. The TCI, along with plant operating expenses, serves as the basis for the discounted cash flow analysis. The discounted cash flow analysis then yields the MFSP. On the per gallon gasoline equivalent (GGE) basis, MFSP is the primary metric by which the current high-octane gasoline conversion process can be compared to alternate fuel production designs.

2 Plant Design Basis

2.1 Feedstock and Plant Size

The dry basis elemental composition of the feedstock, shown in Table 2, is identical to the previous NREL design report [4]. The composition was originally assumed to come from pulpwood. Recent feedstock logistics work at the Idaho National Laboratory (INL) suggests the use of blended material may be required to meet a cost target of $80/dry U.S. ton while still meeting these specifications [11]. Additionally, INL has also looked into a logistics solution based on distributed biomass preprocessing depots. The feedstock preprocessing costs reduction can be achieved by blending high-quality feedstocks with marginal quality feedstocks [12]. The information sources for Table 2 indicate marginal differences in the ultimate analysis between hybrid poplar, used in the 2007 report [13], and pine, used as the basis for INL's feedstock model, namely, the Biomass Logistics Model (BLM). For the purpose of this report, it is assumed that any blended material provided to meet this feedstock elemental composition will not adversely affect gasification conversion efficiencies. Ongoing studies being conducted jointly by INL, NREL, and PNNL will provide experimental evidence of the impact of blended feedstocks on fast pyrolysis and gasification processes. Future TEA will be modified to reflect conversion impacts inferred from such studies.

Table 2. Ultimate Analysis of Woody Biomass Feedstock

Component	Weight % (Dry Basis [14, 15])
Carbon	50.94
Hydrogen	6.04
Nitrogen	0.17
Sulfur	0.03
Oxygen	41.90
Ash	0.92
Heating value[a] (Btu/lb)	8,601 HHV
	7,996 LHV

[a] Calculated using the Aspen Plus Boie correlation.

The current feedstock moisture specification is 10 wt%, unlike the previous studies [4, 16] where a moisture content of 30 wt% was assumed at the plant gate. The cost of the material provided to the gasification reactor is $80/dry U.S. ton [11]. This includes all feedstock logistics including the delivery of a nominal particle size of 0.25 inches into the reactors.

The design capacity for this study is 2,000 dry metric tonnes per day (2,205 dry tons per day), which matches that of previous design reports [16–19]. With an expected 7,884 operating hours per year (90% on-stream factor/availability), the annual feedstock requirement is approximately 657,000 dry metric tonnes per year (724,000 dry tons per year). The assumed on-stream factor allows approximately 36 days of planned and unplanned downtime per year.

The delivered feedstock cost was estimated using INL's model. This cost of $80 per dry ton includes all capital costs, operating costs, and dry matter losses associated with feed delivery, drying, and handling as determined by INL. The only additional equipment included with the plant costs is a cross-flow dryer, which was included for warming up the feedstock using flue gas and allowing for drying contingency during the wet weather.

It should be noted that the term "ash" is used throughout the text to refer to both mineral matter contained in the feedstock as well as ash produced after combustion of residual biomass solids in the conversion process.

Impacts of plant size and feedstock cost on the MFSP are included in the sensitivity analysis in Section 0.

2.2 Process Overview

A block flow diagram of the current design is shown in Figure 3. Detailed process flow diagrams (PFDs) with material and energy balances are provided in Appendix D.

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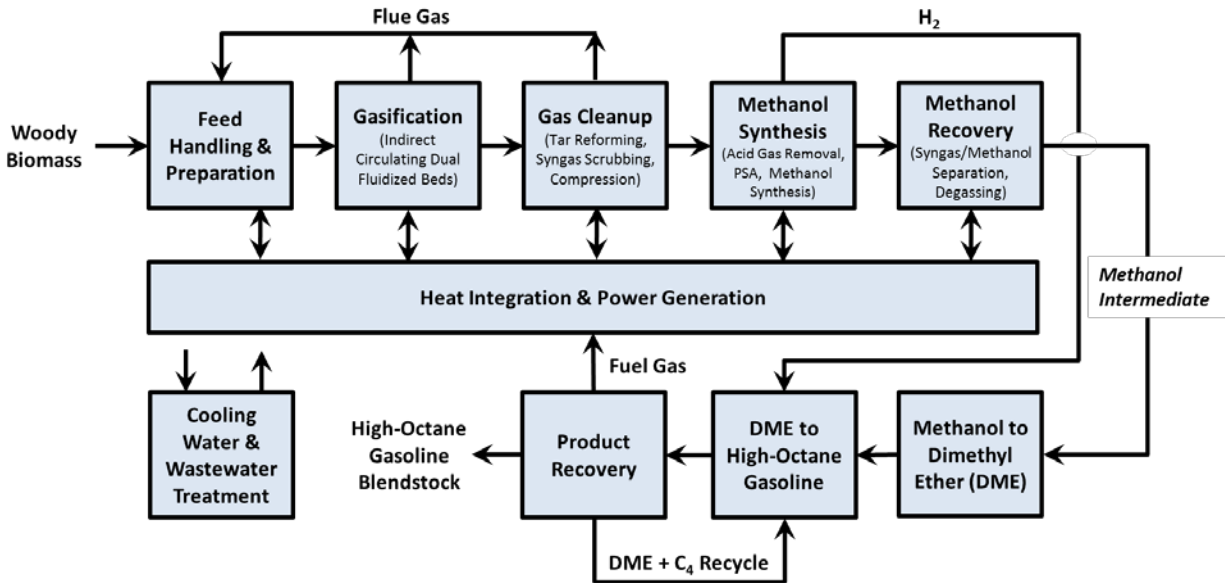
Figure 3. Simplified flow diagram for the thermochemical process

The proposed high-octane gasoline production process includes the following steps, with the detailed operating conditions provided in the corresponding sections:

- *Feed handling and preparation.* The modeled biomass feedstock has an ultimate analysis shown in Table 2. The assumed moisture content is 10 wt%, with an ash content of <1% and nominally sized to 2 mm for the gasifier. A cross-flow dryer is included in the system to allow warming up of the feed prior to feeding in to the reactor, using process waste heat. This also allows for contingencies during wet weather when additional feed drying may be necessary. The feedstock is delivered at $80/dry U.S. ton.

- *Gasification.* Biomass is indirectly gasified. Heat for the gasification reactions is supplied by circulating synthetic olivine[1] sand that is preheated in a char combustor and fed to the gasifier. Conveyors and hoppers feed biomass to the low-pressure (18 psig) entrained flow gasifier. Steam is injected into the gasifier to stabilize the flow of biomass and olivine through the gasifier. Within the gasifier, biomass thermally deconstructs at 1,598°F (870°C) to a mixture of syngas components (such as CO, H_2, CO_2, and CH_4), tars, and solid char containing residual carbon from the biomass and coke deposited on the olivine. Cyclones at the exit of the gasifier separate the char and olivine from the syngas. The solids flow to the char combustor where the char is burned in air in a fluidized bed, resulting in olivine temperatures greater than 1,800°F (982°C). The hot olivine and residual ash is carried out of the combustor by the combustion gases and separated using a pair of cyclones. The first cyclone captures olivine while the second cyclone captures ash and olivine fines. Hot olivine flows back into the gasifier,

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[1] Calcined magnesium silicate, primarily enstatite ($MgSiO_3$), forsterite (Mg_2SiO_3), and hematite (Fe_2O_3). A small amount of magnesium oxide (MgO) is added to the fresh olivine to prevent the formation of glass-like agglomerations formed through biomass potassium interacting with silicate compounds.

9

completing the gasification loop. The hot flue gas from the char combustor is utilized for heat recovery and feedstock preheating. Ash and olivine fines are cooled, moistened to minimize dust, and removed as waste.

- *Synthesis gas cleanup and syngas compression.* Syngas cleanup is defined in this report as reforming of tars, methane, and other hydrocarbons followed by cooling, quench, and scrubbing of the syngas for downstream operations. The water-gas shift reaction also occurs in the reformer. Tars, methane, and light hydrocarbons are reformed to syngas in a circulating, fluidized, solid catalyst system, complete with reforming and regeneration operations in separate beds. The summary of the hydrocarbon reforming conversions can be found in Section 3.3. The syngas is reacted with tar reforming catalyst in an entrained flow reactor at (1,670°F/910°C) and at a gas hourly space velocity of approximately 2,500 h^{-1}. The catalyst is then separated from the effluent syngas in a cyclone. From the cyclone, the spent catalyst flows to the catalyst regenerator vessel where residual coke from the reforming reactions is removed from the catalyst by combustion. The hot catalyst is separated from the combustion flue gas in the regenerator cyclone and flows back to the tar reformer reactor to provide the energy necessary for the reforming reactions. Additional syngas and unreacted gases from the methanol synthesis reactor are also combusted in the regenerator to provide all the heat necessary for the endothermic reforming reactions. A similar approach is documented in a patent application by Rentech [20]. The hot reformed syngas is cooled through heat exchange with other process streams and scrubbed with water to remove persistent impurities like particulates, ammonia, halides, and recalcitrant tars. Scrubber water is purged and treated continuously at an on-site wastewater treatment facility. After heat recovery, the remaining low-quality heat in the flue gas from the catalyst regenerator is utilized for feedstock preheating. After quenching and removing any condensable material and solids, the low-pressure cooled scrubbed syngas is compressed using a three-stage centrifugal compressor with inter-stage cooling where the pressure is increased to approximately 430 psia (30 bar).

- *Acid gas removal and methanol synthesis.* The compressed fresh syngas (430 psia/30 bar) enters an amine-based acid gas enrichment unit and a Merichem LO-CAT sulfur recovery unit for removal of the CO_2 and H_2S. The recovered H_2S-rich acid gas stream is routed to the Merichem LO-CAT sulfur recovery unit where H_2S is converted to elemental sulfur and stored for disposal. The remaining CO_2 is vented to the atmosphere. After the acid gas removal step, the cleaned syngas is then split into two streams. The smaller stream of the cleaned syngas (about 6%) is sent to a pressure swing adsorption (PSA) system where hydrogen is separated for hydrocarbon synthesis in the methanol to high-octane gasoline area. The majority of the cleaned and conditioned syngas is further compressed to 735 psia (51 bar) for the methanol synthesis; the syngas is converted to methanol in a tubular, fixed-bed reactor containing a copper/zinc oxide/alumina catalyst. The vapor-phase product from the methanol synthesis reactor must be cooled to recover the methanol and to allow unconverted syngas and any inert gaseous species (CO_2, CH_4) to be recycled or purged. This is accomplished with a series of heat exchangers, including air cooler and water cooling. The mixture of methanol and unconverted syngas is cooled through heat exchange with the steam cycle and other process streams. The methanol is separated by condensing it away from the unconverted syngas. Unconverted syngas is recycled back to

10

the methanol synthesis reactor inlet. Heat must be removed from the methanol synthesis reactors because the synthesis reaction is exothermic. Temperature control and heat removal from the exothermic reactor is accomplished by steam production on the shell-side of the tubular reactor. The steam temperature and pressure can be maintained and controlled by back-pressure control at the outlet of the steam drum.

- *Methanol conditioning.* The methanol leaving the reactor has been condensed at elevated pressure and has absorbed a sizeable quantity of gas (mostly CO_2). Once the crude methanol stream is reduced to lower temperature (110 °F/43.3°C) and pressure (98 psia/6.8 bar), it is sent to a distillation column to de-gas the methanol. The methanol intermediate is sent to storage for upgrading to gasoline.

- *Methanol to high-octane gasoline synthesis.* This area of the process includes the conversion of methanol to DME and the subsequent conversion of DME to high-octane hydrocarbons. Hydrocarbon formation from DME is accomplished in two four-stage packed bed reactors containing metal modified beta-zeolite (H-BEA) catalyst. The yields from the process are heavily weighted toward branched C_7 molecules, with octane ratings greater than 100. Temperature control and heat removal from the hydrocarbon synthesis reactors is accomplished by using multiple adiabatic reactors in series with inter-stage cooling. The heat is recovered as low pressure steam. It is assumed that while two reactors are in DME-to-hydrocarbons service, the other reactor is in coke-burn/catalyst regeneration. The catalyst regenerator burns carbon (coke) deposits off the catalyst particles, regenerating the catalyst activity and providing heat for steam generation.

- *Product recovery.* This process area consists of two distillation columns. The water-free crude hydrocarbon product is first sent to the first distillation column (Deethanizer) where liquid C_{4+} and gasoline-range hydrocarbons are separated from the light ends (C_{3-}) and unconverted DME. DME is recycled to the hydrocarbon synthesis reactor and the light gas stream (i.e., C_{3-} or fuel gas) is sent to the fuel combustor in Area 300. The Deethanizer bottom product is subsequently sent to the second column (Debutanizer) where C_4 is separated from gasoline-range hydrocarbons. The Debutanizer overhead is then recycled to the hydrocarbon synthesis reactors in Area 1400. The bottoms of the Debutanizer are the high-octane hydrocarbons which are cooled and then stored for sale as high-octane gasoline.

- *Heat and power generation.* A conventional steam cycle produces steam for direct injection into the gasifier and reformer. Indirect steam is used for heating associated with distillation and the acid gas strippers in the acid gas removal system. Electricity is generated using three steam turbines, with intermediate reheat, to meet the demands of the plant. As mentioned earlier in the methanol synthesis section, additional electricity is generated by sending a portion of the pressurized unreacted gases from the methanol synthesis reactor through a turbo-expander. The majority of electricity is used for syngas compression. The steam cycle is integrated with the biomass conversion process. Preheaters, steam generators, and super-heaters are integrated within the process design to generate the steam from boiler feed water (BFW). Process condensate is recycled to the steam cycle, de-gassed, treated, and combined with makeup water.

- *Cooling water and other utilities.* A cooling water system is included in the Aspen Plus model to determine the requirements of each cooling water heat exchanger within the

11

biomass conversion process as well as makeup water and power requirements. The design also includes additional water optimization; process condensate is treated and combined with the cooling water makeup to reduce fresh water consumption.

2.3 Aspen Plus Model

An Aspen Plus Version 7.2 simulation was used as the basis for this report. The gaseous and liquid components that are important for the process were included as distinct molecular species using Aspen Technology's component properties database. The biomass feedstock, ash, and char components were modeled as non-conventional components. The thermochemical conversion of biomass to high-octane gasoline design requires processing of three different phases of matter (solid, liquid, and gas phases). Therefore, no single physical property package was sufficient to describe the entire plant from feed to products. Different property packages were used within the Aspen Plus simulation to more accurately represent chemical component behaviors in specific process areas. The Redlich–Kwong–Soave equation of state with Boston-Mathias modifications (RKS-BM) was used throughout much of the process simulation for high-temperature, high-pressure gas and liquid phase behavior. The non-random, two-liquid option with Redlich-Kwong equation of state (NRTL-RK) was used for methanol separation calculations. The American Society of Engineers (ASME) 1967 steam table correlations were used for the steam cycle calculations. The ELECNRTL package was used to model the electrolyte species in the quench water system.

3 Process Design and Cost Estimation

The process design for the high-octane gasoline production facility leverages advances in syngas production and clean-up technologies developed for the thermochemical mixed alcohols pathway, specifically around gasification and tar reforming processes [4], as well as methanol synthesis and gasoline separation from the NREL MTG design report [5]. For the conversion of DME to high-octane hydrocarbons, the selectivity and conversion of each compound is set to match the technical targets established under the research program for this technology.

The process design broadly consists of the following areas:

- Area 100: Feed handling and preparation
- Area 200: Gasification
- Area 300: Synthesis gas cleanup and syngas compression
- Area 400: Acid gas removal and methanol synthesis
- Area 500: Methanol conditioning
- Area 1400: Methanol to high-octane gasoline synthesis
- Area 1500: Product recovery
- Area 600: Steam system and power generation
- Area 700: Cooling water and other utilities.

12

The following sections present process overviews, design bases, and installed equipment cost estimations for each process area of the plant. Heat exchanger equipment sizes and costs are discussed separately in Section 3.9.

3.1 Area 100: Feed Handling and Preparation

The following section presents an overview of the feed handling facilities. The basis for the system design is discussed, and cost estimates for the feedstock are provided based on capital requirements and operating costs of the facilities. The information provided is based on work done at INL.

Area 100 Overview

This area of the process accommodates the delivery of biomass feedstock from the field to the biorefinery, short-term on-site storage, and the preprocessing/preparation of the feedstock for conversion in the gasifier. The current design is based upon a specific blended biomass feedstock—pulpwood (45%), wood residues (32%), switchgrass (3%), and construction and demolition waste (20%). A simplified flow diagram for the updated woody biomass feedstock supply system is provided in Figure 4. Details on the feedstock supply system design and logistics for conversion of lignocellulosic biomass to hydrocarbon fuels can be found in the recent design report published by INL [11]. Process flow diagrams (PFD) for the feed handling and preparation area are shown in PFD-61713-A101 and PFD-61713-A102 in Appendix D.
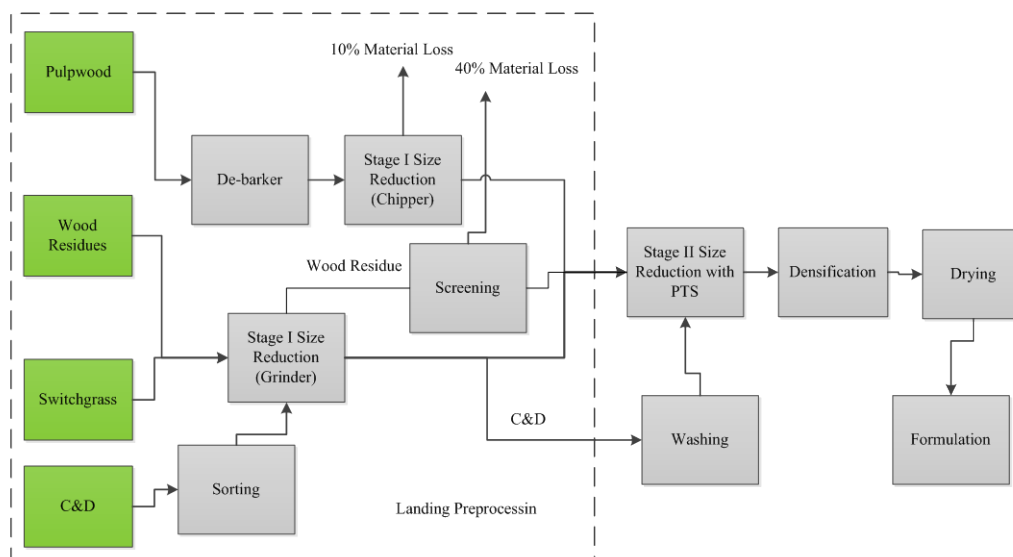
Figure 4. Material flow that incorporates many improvements in preprocessing, including pneumatics, fractional milling, high-moisture densification, and formulation/blending (Courtesy INL [11])

At the biorefinery, the trucks are weighed and a truck dumper unloads the truck contents into a hopper. A dust collection system is present during unloading to prevent excessive dust accumulation and to limit dry matter losses. The chips are cleaned using an electromagnet, moisture is monitored, and material is passed over a vibratory conveyor to remove any excess

13

dirt/ash that may still be present. The material is then conveyed into a pile using a circular stacker and an overpile reclaimer. A five-day supply of material is stored at the refinery. The pile, which acts as a queue, is on an asphalt pad with a long grate running through it that allows material to flow out of the pile. A front end loader continuously pushes material onto the grate to maintain flow. A conveyor located at the bottom of the pit under the grate conveys the biomass into a cross-flow dryer, where the biomass is preheated using hot flue gas from the biomass conversion processes. From the dryer, the biomass is conveyed into a metering bin where it is fed into the conversion process.

Area 100 Design Basis

The cross-flow dryer is newly added equipment in the feed handling area that was not included in the earlier design [4]. The main objective of this equipment is to warm the biomass feed (utilizing the heat from flue gas) from ambient temperature (60°F) to 220°F at the inlet to the gasifier in order to maximize the efficiency of the gasifier. The cross-flow dryer will also take care of any seasonal wetness.

The moisture content of the material is 10 wt%, with a nominal particle size of 0.25 inches. Based on prior research and process demonstrations, the biomass gasifier is capable of processing larger feedstock particle sizes (up to 2 inches) [4]. The feedstock basis from INL is considered as a level of conservatism for the purposes of this theoretical biorefinery design.

The conceptual plant is modeled to consume 2,000 metric tonnes per day (2,205 tons per day). Assuming biomass delivery by truck with a per truck capacity of 25 tons [21] of biomass at 10 wt% moisture content, approximately 97 truck deliveries per day would be required to operate the plant at design capacity of 2,000 metric tonnes per day (2,205 tons per day).

Area 100 Equipment Cost Estimations

INL, in partnership with other national labs, has developed a woody supply system design and modeled the costs using a woody biomass supply system model. The model is continuously refined through input from industry, academia, and national labs to improve the accuracy of the model parameters and output. Future model iterations will identify key sensitivities in the system in terms of costs, material flows, and equipment performance parameters.

Table 3 shows the costs associated with the operations in the supply system (Figure 4). The data presented include unit operation costs for the design case scenario for the supply of blended woody biomass to a gasification process. All values are presented in units of 2011 U.S. dollars per dry U.S. ton of feedstock.

14

Table 3. Unit Cost Breakdown for Woody Biomass Feedstock (in 2011 dollars) [11]

Cost Element *Formulation Contribution*	Pulpwood *45%*	Wood Residues *32%*	Switchgrass *3%*	Construction and Demolition Waste *20%*	Blend
Grower payment/access cost	25.00	26.35	19.67	8.15	**21.90**
Harvest and collection	22.24	0.00	15.41	–	**10.47**
Landing preprocessing	12.17	8.73	0.00	9.85	**10.24**
Transportation	10.89	3.33	4.50	6.87	**7.52**
Preprocessing	23.97	23.97	19.70	28.12	**24.67**
Storage	3.23	3.23	5.50	3.23	**3.30**
Handling	1.90	1.90	1.90	1.90	**1.90**
Total Delivered Feedstock Cost, $/dry U.S. ton	99.49	67.51	66.68	58.12	**80.00**

The feedstock logistic cost of blended biomass, as determined from the INL's BLM, is $58.10/dry U.S. ton (short ton). When the grower payment of $21.90/dry U.S. ton is added, the total delivered feedstock cost through the gasifier inlet is $80.00/dry U.S. ton. This figure includes ownership costs and operating costs.

Equipment lists for the feed handling and preparation area, as well as other pertinent information, are provided in the appendices. It should be pointed out that all purchased and installed equipment costs shown for this area are for the cross-flow dryer. The costs associated with other equipment in the area, such as hopper feeder, hammermill, and screener feeder conveyor, are not included for the plant economics because all of these capital costs are included in the delivered feedstock cost.

3.2 Area 200: Gasification

The following section presents an overview, basis for design, and cost estimates for construction of the gasification facilities. The section is developed based on the NREL mixed alcohols design report [4].

Area 200 Overview

This area of the process converts a mixture of dried feedstock (10 wt% moisture) and steam to syngas and char. Heat is provided indirectly through circulation of olivine (heat transfer media) that is heated by the combustion of char produced in the gasifier. Steam is injected into the gasifier to serve as a fluidizing medium and reactant at the high gasifier temperatures. A simplified flow diagram of the gasification process area is provided in Figure 5.

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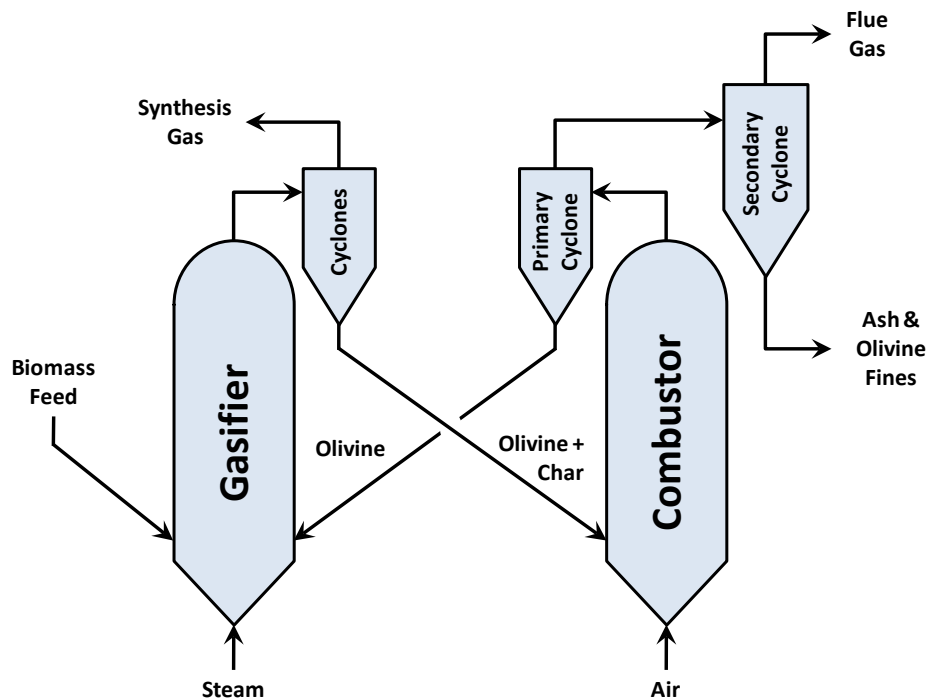
Figure 5. Simplified process flow diagram of gasification area

After going through the feed handling and drying area, the dried blended biomass enters the gasification area, as shown in PFD-61713-A201 in Appendix D. The design includes two parallel gasifier trains, each with a biomass processing capacity of 1,000 dry metric tonnes per day. The gasifier (R-201) used in this analysis is a low-pressure, indirectly heated, circulating fluidized bed (CFB) gasifier. The gasifier output was modeled using correlations based on run data from the Battelle Columbus Laboratory (BCL) 9 tonne/day test facility. These correlations are provided for reference in Appendix E.

Heat for the gasification reactions is supplied by circulating a hot medium between the gasifier vessel and the char combustor (R-202). In this case the medium is synthetic olivine—calcined magnesium silicate consisting of enstatite ($MgSiO_3$), forsterite (Mg_2SiO_3), and hematite (Fe_2O_3). A small amount of magnesium oxide (MgO) is added to the fresh olivine. The MgO rejects the potassium present in biomass as ash by forming a high melting (~2,370°F/1,300°C) ternary eutectic with the silica, thus sequestering it. Without MgO addition, the potassium will form glass (K_2SiO_4) by interacting with the silica in the system. K_2SiO_4 has a low melting point (~930°F/500°C) and its formation would cause the bed media to become sticky, agglomerate, and eventually defluidize. Potassium carry-over from the gasifier to downstream processes is also significantly reduced with the addition of MgO. The biomass ash is assumed to contain 0.2 wt% potassium. The MgO molar flow rate is set at 2 times the molar flow rate of potassium. With an expected potassium inlet flow rate of 0.087 lb-moles per hour, the estimated MgO requirement is 0.174 lb-moles per hour or 7.0 lb per hour.

The fluidization medium in the gasifier is steam supplied from the steam cycle (Area 600). The steam-to-feed ratio is set to 0.4 lb of steam per lb of bone dry biomass feedstock [22]. The gasifier is designed to operate at 18 psig to allow for pressure drops in subsequent equipment so

16

that no further compression is required until after gas cleanup and quench. The average olivine circulation rate is approximately 27 lb of olivine per lb of bone dry wood. Fresh olivine is added to compensate for losses from the gasifier and combustor cyclones; the net consumption of olivine is 543 lb/hr.

The gasifier and char combustor will remain in heat balance by producing (gasifier) and burning (char combustor) enough char to satisfy the energy requirements of the operation. The amount of char formed in the gasifier is an inverse function of temperature. Unless supplemental fuel is provided to the char combustor for temperature control, the gasifier and combustor reach thermal equilibrium based on the amount of char formed in the gasifier. If the gasifier temperature is lower than the equilibrium temperature, then more char is formed and more heat is generated by char combustion. This results in more heat transfer from the combustor to the gasifier, thus bringing the gasifier temperature up toward equilibrium. Although the gasifier, char combustor, and associated piping will be insulated to maximize thermal efficiency, some heat losses will occur. Two percent (2%) of the lower heating value of the feed biomass (dry basis) is assumed to be lost from the system. The resulting gasifier temperature is 1,580°F (860°C) and the char combustor temperature is 1,775°F (968°C). The char combustor is operated with 20% excess air (based on the stoichiometric requirement) to ensure complete oxidation of char components. The composition of the outlet gas from the gasifier is shown in Table 4.

Olivine is recovered from the raw syngas exiting the gasifier using two stages of cyclone separators. Nearly all of the olivine and char (99.9% of both) are separated in the primary gasifier cyclone (S-201) and fed by gravity to the char combustor. A secondary cyclone (S-202) removes 90% of residual fines.

The combustion flue gas and olivine from the char combustor are sent to a cyclone separation system similar to the gasifier cyclone. A primary combustor cyclone (S-203) separates the olivine (99.9%) from the combustion gases and the olivine is gravity-fed back to the gasifier. The flue gas from the primary cyclone is sent through a secondary cyclone (S-204) and then utilized for heat recovery before it is mixed with flue gas from the tar reformer area. The combined flue gas stream is utilized for feedstock drying and finally enters the flue gas scrubber for removal of residual particulate matter and regulated chemical species such as SO_2.

The mixture of olivine, ash, and catalyst fines from the secondary flue gas cyclone is indirectly cooled to 300°F in the water-cooled screw conveyor (M-201). Water is added directly to the solids for further cooling and dust prevention. The solids are disposed as landfill.

Pressure balance between the gasifier and char combustor in reliably controlling the circulation of olivine through the system is critical. The pressure balance ensures that olivine circulation is maintained from the gasifier to the char combustor and back to the gasifier (not vice versa). The pressure balance in such systems is usually controlled by two slide valves (or L-valve loop seal or J-leg loop seal), one controlling the flow of spent olivine from the gasifier cyclones to the char combustor and the other controlling the flow of regenerated olivine from the primary char combustor cyclone to the gasifier. The pressure balance for these types of processes is addressed during the detailed design phase of a project.

Table 4. Gasifier Operating Parameters, Gas Compositions, and Efficiencies

Gasifier Operating Parameters	Value	
Temperature	1,598°F (870°C)	
Pressure	33 psia (227.5 kPa)	
Gasifier Outlet Gas Composition	**mol % (wet)**	**mol % (dry)**
H_2	13.2	23.9
CO_2	7.08	12.8
CO	23.3	42.2
H_2O	44.6	--
CH_4	8.50	15.4
C_2H_2	0.23	0.41
C_2H_4	2.41	4.35
C_2H_6	0.16	0.29
C_6H_6	0.07	0.13
Tar ($C_{10}H_8$)	0.13	0.23
NH_3	0.18	0.32
H_2S	0.04	0.07
H_2:CO molar ratio	0.57	
Gasifier Efficiency	72.9% HHV[a] basis	
	72.5% LHV[b] basis	

[a] Higher heating value. [b] Lower heating value.

Area 200 Design Basis

The design basis for the gasification area is presented in Table 5. This basis reflects the overall plant capacity of 2,000 dry metric tonnes per day (2,205 dry tons per day).

Table 5. Design Basis for Gasification Area

Parallel gasification trains		2
Biomass feed rate per train (dry basis)		1,000 metric tonnes per day
Biomass moisture content to gasifier		10 wt%
	Gasifier	Char combustor
Operating pressure (psia)	33.0	29.0
Operating temperature (°F)	1,598	1,798

Area 200 Equipment Cost Estimations

Capital costs for the equipment in this area were estimated by Taylor Biomass Energy for a 500 metric tonnes per day (MTPD) unit. The estimate included the cost for a thermal tar reformer, which will be discussed in the following section. While the Taylor Biomass Energy base design

18

is for a 500 MTPD modular unit, the company supports the feasibility of scaling the unit to a 1,000 MTPD capacity using a scaling exponent of 0.6 as shown below.

1,000 MTPD equipment cost = 500 MTPD equipment cost x $(1,000/500)^{0.6}$

1,000 MTPD equipment cost = $9.7 MM x 1.5157 = $14.7 MM

Because the total capacity of the thermochemical plant design is 2,000 MTPD, two parallel 1,000 MTPD units will be incorporated in the design. The cost estimate from Taylor Biomass Energy is summarized in Table 6 along with scaled costs for two parallel 1,000 MTPD units. Taylor Biomass Energy's quotation for the 500 MTPD unit is more representative of a first-of-its-kind plant than an nth-plant, suggesting that the capital cost estimate is conservative. However, the design presented in this report has minor variations from the Taylor Biomass Energy design [23]. It is assumed that additional costs for these variations are adequately covered by the conservatism in the Taylor Biomass Energy quotation, as n^{th}-plant costs are expected to be 10% to 15% lower than costs for first-of-a-kind predecessors. Because the quote also includes gas clean-up equipment (thermal cracker or tar reformer), two-thirds of the total equipment costs is allocated to the two 1,000 MTPD gasifiers and the remaining one-third is allocated to the tar reformer (cost details provided in Appendix A). The installation factor applied to calculate TIC was determined from the Taylor Biomass Energy quotation.

Table 6. Estimates for Gasification and Tar Reforming Facilities

	Taylor Biomass Energy 500 MTPD Single Train	Scaled 1,000 MTPD Single Train	Two Scaled Parallel 1,000 MTPD Trains	
	(Aug 2010)	(Aug 2010)	(Aug 2010)	(2011)
Dry feedstock capacity (metric tonnes per day)	500	1,000	2,000	2,000
Total purchased equipment cost (TPEC)	$9.7 MM	$14.7 MM	$29.4 MM	$31.3 MM
Equipment installation factor	2.31	2.31	2.31	2.31
Total installed cost (TIC)	$22.4 MM	$33.9 MM	$67.9 MM	$72.3 MM
TIC allocated to gasifier (2/3)				**$48.2 MM**
TIC allocated to tar reformer (1/3)				**$24.1 MM**

The Taylor Biomass Energy system includes the following equipment:

- Biomass feeding system (excluding storage, pre-treatment, and transport equipment)
- Gasification process reactors (gasification, gas conditioning, and combustion)
- All cyclones and interconnecting ducting
- Combustion air blower
- Combustion air preheat exchanger(s)
- Start-up equipment

19

- Syngas polishing scrubber (final cleanup and particulate removal)

- Ash discharge system

- Emergency flare system

- Process instrumentation

- Olivine and magnesium oxide makeup system

- Inert gas system.

Table 7 presents the anticipated equipment requirements and costs for the gasification area. The gasifier cost presented in this table reflects the two-thirds cost assumption explained above.

Table 7. Cost Estimate for Gasification Area

Equipment Description	TPEC (2011 $k)	TIC (2011 $k)
Gasifier (ancillaries included)	20,895	48,219
Heat integration	895	2,254
Area A200 subtotal	**21,789**	**50,472**

3.3 Area 300: Synthesis Gas Cleanup and Syngas Compression

The following section presents an overview, basis for design, and cost estimates for construction of the gas clean-up area.

Area 300 Overview

In Area 300, the methane, tars, and other hydrocarbons in the syngas are reformed to additional CO and H_2. Particulates are removed by scrubbing. The cleaned syngas is subsequently compressed.

Figure 6 presents a simplified process flow diagram for the gas clean-up area of the plant, which includes the following major steps:

- Tars, methane, and other hydrocarbons in the syngas are reformed to CO and H_2.

- Particulates and other contaminants are removed by a quenching and scrubbing process.

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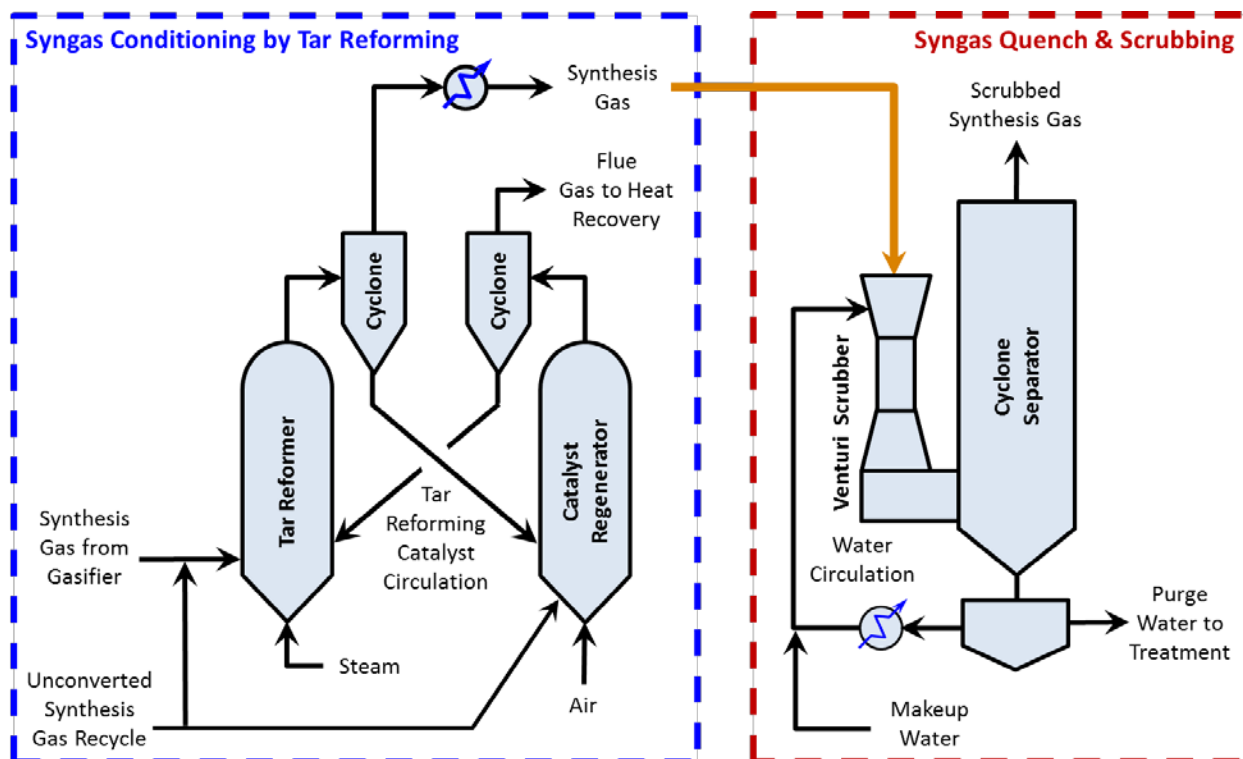
Figure 6. Simplified process flow diagram of gas clean-up area

The type and the extent of cleanup are dictated by the requirements of the methanol synthesis catalyst and other downstream equipment. The syngas from the secondary gasifier cyclone is sent to the catalytic tar reformer (R-303 in PFD-61713-A301, Appendix D), which is an entrained-flow, fluidized catalytic reactor. In the reformer, methane, tars, and other hydrocarbons are converted to CO and H_2, and NH_3 is converted to N_2 and H_2. Because the reforming catalyst has significant water-gas-shift (WGS) activity, the syngas is assumed to exit the reformer at water-gas-shift equilibrium. This extent of WGS is not necessarily the best scenario for making methanol because it also produces 1 mole of CO_2 for every mole of H_2 made. As discussed in Section 3.4, the CO_2 in the syngas must be removed to achieve the specified level of 5 vol % at the methanol synthesis reactor. A molar steam-to-carbon ratio (SCR) of 4.4 was used with the resulting H_2:CO ratio of 2:1. The carbon in hydrocarbons and tars are included for the SCR calculation (i.e., SCR = [H_2O]/[Total C in hydrocarbons and tars]). The current SCR simultaneously produces desirable H_2:CO ratio and mitigates coke formation.

In the Aspen Plus simulation, the conversion of each compound is set to match targets that have been demonstrated [24, 25] and are summarized in Table 8. Similar to the gasification area, the heat for tar conversion is supplied by circulating catalyst (Ni/Mg/K supported on alumina) [26] between the tar reformer reactor and the catalyst regenerator. The catalyst regenerator burns carbon (coke) deposits off the catalyst particles, regenerating the catalyst activity and providing heat for the tar reforming reactions. The concentration of coke precursors in the syngas feed to the tar reformer will vary depending on feedstock type and conversion in the gasifier. Because the rate of coke formation in the tar reformer is not sufficient to sustain the endothermic reforming reactions, supplemental combustion gases are supplied to the catalyst regenerator from various process areas.

21

Air is supplied to the catalyst regenerator by the combustion air blower (K-305) to provide the oxygen for combustion. Continuous operation has been demonstrated with simulated syngas (including some H_2S) using NREL's reforming catalyst [20]. The raw syngas from the gasifier and supplemental steam serve as fluidization media for the catalyst in the reformer. Supplemental steam is required in the base case design, as the steam present in the syngas from the gasifier alone is not sufficient to maintain a target molar steam-to-carbon ratio (SCR) of 4.4.

The spent catalyst from the tar reformer is separated from the syngas in a two-stage cyclone separator (S-306) and flows to the catalyst regenerator (R-301A). The base tar reformer operating temperature for this design is 1,670°F (910°C) at the reactor outlet. This temperature is based on experimental results using NREL's reforming catalyst, which showed that the catalyst maintains high reforming activity in the presence of biomass-derived syngas at 1,697°F (925°C). The lower temperature threshold of the high-performance model has not yet been determined. The model can be modified to maintain the reformer at 1,697°F instead of 1,670°F without significant changes in yields.

The combustion flue gas and catalyst from the regenerator are sent to a cyclone separator system. A two-stage cyclone (S-307) separates the regenerated catalyst from the combustion gases and the catalyst flows back to the tar reformer. The flue gas is then utilized for heat recovery before it is mixed with flue gas from the gasification area. The combined flue gas stream is used for the feed preheating system and then sent to a flue gas scrubber.

Hot syngas from the tar reformer is cooled to 140°F (60°C) through heat exchangers, which are used to heat other process streams. A syngas scrubbing system removes impurities such as particulates, ammonia, halides, and residual tars. As shown in PFD-61713-A302 in Appendix D, the scrubbing system consists of a venturi scrubber (M-302), cyclone separator (M-301), and a quench water circulation system. The quench water is contained in a closed recirculation loop with heat rejected to the cooling tower through the quench water cooler (H-301). The quench water circulation rate is set to an L/G (liquid flow to actual volumetric gas flow) ratio equal to 1 liter/m^3. This ratio is based on a study of venturi scrubbers that suggests operating conditions for near-optimal performance [27]. The syngas exits the quench system at a temperature of 130°F (54°C). A continuous purge and fresh water makeup is maintained in the circulation loop at approximately 3 gpm to control the solids level in the circulating water. The purge stream is sent to a clarifier to separate the solids from the water. The solids are routed for disposal and the water, which includes process condensate, is recycled to the reverse osmosis (RO) purification system in the base case design. The current version of the design includes water optimization; the water is treated and used for flue gas scrubbing and the remainder is mixed with the cooling water makeup stream (shown in PFD-61713-A701 in Appendix D).

After quenching and removing any condensable material and solids, the scrubbed syngas is compressed to 430 psia (30 bar) using a three-stage centrifugal compressor with inter-stage cooling (K-310 in PFD-61713-A303 in Appendix D). The compressor was modeled such that each section has a polytropic efficiency of 78% to 87% (based on vendor quotes), assumed mechanical efficiency of 98%, and intercooler outlet temperatures of 110°F (43°C). Each inter-stage cooler includes water, air, and process stream exchangers. These exchangers are deliberately oversized to minimize pressure drops to an average of 6 psi (0.41 bar) per inter-stage (i.e., an average of 2 psi pressure drop in each of the three inter-stage coolers). Minimizing

pressure drop in the coolers will minimize electrical consumption as the compressors represent the most significant electrical load on the system, consuming approximately 19 megawatts (MW) of electricity in the base case.

Area 300 Design Basis

Table 8 summarizes the hydrocarbon conversions during continuous operation.

Table 8. Performance of Reforming Catalyst [24, 25]

Compound	Conversion
Methane (CH_4)	80 %
Ethane (C_2H_6)	99 %
Ethylene (C_2H_4)	90 %
Tars (C_{10+})	99 %
Benzene (C_6H_6)	99 %
Ammonia (NH_3)	90 %

These conversions were incorporated into the Aspen Plus model to simulate expected tar reformer performance. The compositions of the reformer inlet and outlet streams are presented in Table 9.

Table 9. Tar Reformer Conditions and Gas Compositions for the Base Case

	Inlet		Outlet	
Temperature, °F (°C)	1,598 (870)		1,670 (910)	
Pressure, psia	29		26	
H_2:CO molar ratio	0.63		2.11	
(H_2-CO_2)/(CO+CO_2)	0.25		0.93	
Gas composition (mole %)	Wet	Dry	Wet	Dry
Hydrogen (H_2)	11.55	25.75	38.26	55.54
Carbon dioxide (CO_2)	5.50	12.26	11.01	15.98
Carbon monoxide (CO)	18.42	41.05	18.17	26.37
Water (H_2O)	55.12	---	31.12	---
Methane (CH_4)	6.86	15.29	1.16	1.68
Acetylene (C_2H_2)	0.17	0.39	146 ppmv	212 ppmv
Ethylene (C_2H_4)	1.86	4.13	0.16	0.23
Ethane (C_2H_6)	0.12	0.27	10 ppmv	15 ppmv
Benzene (C_6H_6)	526 ppmv	0.12	4 ppmv	6 ppmv
Tars ($C_{10}H_8$)	961 ppmv	0.21	1 ppmv	1.18 ppmv
Ammonia (NH_3)	0.13	0.30	114 ppmv	165 ppmv
Hydrogen sulfide (H_2S)	304 ppmv	677 ppmv	256 ppmv	372 ppmv
Methanol (CH_3OH)	353 ppmv	786 ppmv	0 ppmv	0 ppmv
Nitrogen (N_2)	386 ppmv	861 ppmv	837 ppmv	1215 ppmv

Area 300 Equipment Cost Estimations

Table 10 presents the equipment list and cost estimates for the gas clean-up area of the plant. The tar reformer cost presented in this table reflects the one-third cost assumption based on Taylor Biomass Energy's cost estimate for gasification and gas cleanup, as discussed in detail in Section 3.2. Equipment costs in the table were obtained from quotations by Taylor Biomass Energy (tar reformer) and TurboSonic (syngas quench and scrubbing system).

Table 10. Cost Estimate for Gas Clean-up Area

Equipment Description	TPEC (2011 $k)	TIC (2011 $k)
Tar reformer and syngas scrubber	10,447	24,109
Syngas scrubber	2,431	3,009
Synthesis gas compressor	17,105	31,040
Heat integration	6,391	12,523
Area A300 subtotal	**36,374**	**70,681**

3.4 Area 400: Acid Gas Removal and Methanol Synthesis

The following section presents an overview, basis for design, and cost estimates for construction of the methanol synthesis reactor, syngas and recycle compressors, turbo-expander, and acid gas removal facilities. Figure 7 illustrates a simplified process flow for the methanol synthesis area.

Area 400 Overview

The compressed fresh syngas (430 psia/30 bar) enters an amine-based acid gas removal (AGR) unit for removal of the CO_2 and H_2S and subsequently enters the methanol synthesis reactor. The recovered H_2S-rich acid gas stream is routed to the Merichem LO-CAT sulfur recovery unit (shown in PFD-61713-A401, PFD-61713-A402, PFD-61713-A403, and PFD-61713-A404).

The acid gases removed in the amine scrubber are stripped to regenerate the sorbent and sent through a sulfur removal operation using a liquid phase oxidation process. The amine acid gas removal (AGR) and LO-CAT processes work together to remove most of the sulfur and CO_2 to the levels (i.e., sulfur free and $CO_2 \leq 5\%$) desired for the copper/zinc oxide/alumina catalyst. We assumed that 100 ppmv of the sulfur level could be achieved with the AGR removal system because it is also suitable to H_2S removal. However, to ensure the methanol synthesis catalyst is protected from sulfur poisoning, a zinc oxide (ZnO) guard bed (S-405 shown in PFD-61713-A401) is also employed which can remove sulfur (in the form of H_2S) down to undetectable levels [28].

Although there are several liquid-phase oxidation processes for H_2S removal and conversion available today, the LO-CAT process was selected because of its progress in minimizing catalyst degradation and for its environmentally benign catalyst. LO-CAT is an iron chelate-based process that consists of a venturi precontactor, liquid-filled absorber, air-blown oxidizer, air blower, solution circulation pump, and solution cooler. The air flow rate for re-oxidizing the LO-CAT solution was included in the simulation and calculated based on the requirement of two moles O_2 per mole H_2S. Prior to entering the LO-CAT system, the gas stream is superheated in

24

the preheater to 10°F above its dew point, which in this process is equivalent to 148°F (64.4°C). This degree of superheating is required for the LO-CAT system. The LO-CAT process was modeled to remove the H_2S to a concentration of 10 ppmv, which is the permissible exposure limit [29] in the CO_2 vent effluent from the amine scrubber. The CO_2 from the LO-CAT unit is vented to the atmosphere. The H_2S is reduced to elemental sulfur and stockpiled for disposal. Table 11 summarizes the AGR design parameters.

The cleaned and conditioned syngas is split into two streams. The smaller stream of the cleaned syngas (about 6%) is sent to a pressure swing adsorption (PSA) system (S-445). The PSA unit is used to separate the hydrogen from the other components in the syngas stream for the high-octane hydrocarbon synthesis in Area 1400. A hydrogen recovery rate of 85% is used in this design. Purge gas from the PSA system, along with the purged unconverted syngas from the methanol synthesis reactor, is sent to the tar reformer and fuel combustor in Area 300. The product hydrogen at the PSA outlet is at 410 psia (28 bar) and 110°F (43.3°C).

The majority of the cleaned and conditioned syngas (94%) is further compressed (K-410) to 735 psia (51 bar) and mixed with unconverted recycled syngas and preheated to 437°F (225°C) before entering the methanol synthesis reactor. The methanol synthesis reactor (R-490) is a fixed bed reactor containing a copper/zinc oxide/alumina catalyst. Table 12 lists the methanol synthesis conditions for a typical copper/zinc oxide/alumina catalyst given in Bartholomew [29]. The reactor operating conditions assumed here are 482°F (250°C), 730 psia (5.0 MPa), space velocity of 8,000 hr^{-1}, and single-pass CO conversion of 46%. The literature notes that CO_2 concentrations of up to 8 vol % can improve productivity to methanol [30]. Excessive levels of CO_2 decrease conversion. A value of 5 vol % CO_2 was chosen for the base case design. The operating pressure to make methanol is significantly lower than that for making mixed alcohols (730 psia/50 bar versus 1,000–2,000 psia/69–138 bar) [18]. The temperature is comparable to other synthesis reactions.

The methanol reactions are fast, and equilibrium is quickly achieved. The associated equilibrium reactions modeled in the Aspen model are:

$$CO + 2\ H_2 \leftrightarrow CH_3OH \qquad \text{(Eq. 4)}$$
$$3\ H_2 + CO_2 \leftrightarrow CH_3OH + H_2O \qquad \text{(Eq. 5)}$$
$$CO + H_2O \leftrightarrow H_2 + CO_2 \qquad \text{(Eq. 6)}$$
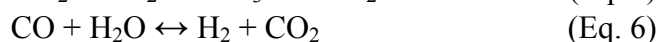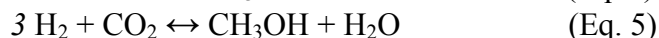
The heat released during the reaction is a serious concern, because even short excursions of only a few degrees can irreversibly damage the catalyst. Heat must be removed from the reactors because the synthesis reaction is exothermic. Temperature is controlled by steam generation on the shell side of the reactor. Boiler feed water was assumed to be cross-exchanged within the reactor to generate steam for the process. Steam is generated at 476 psia (32.8 bar) and 462°F (239°C) and maintained at these conditions by back-pressure control at the outlet of the steam drum (T-605).
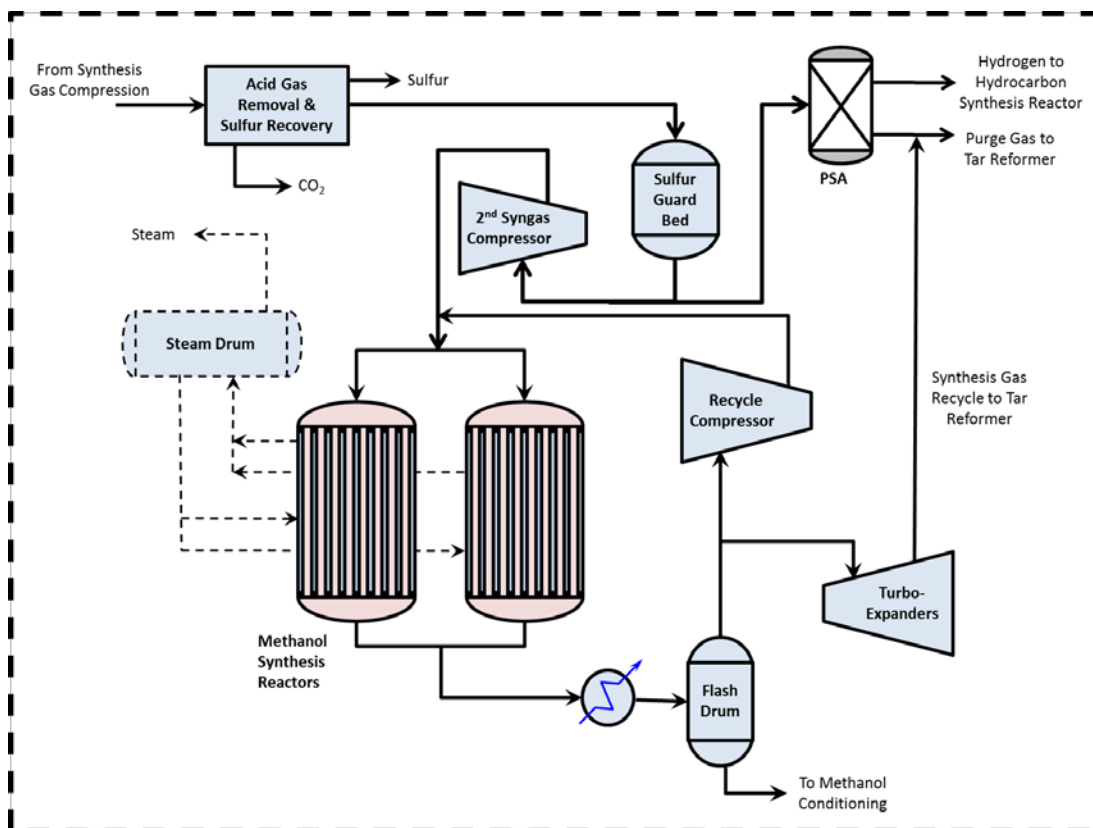
Figure 7. Simplified process flow diagram of methanol synthesis area

The vapor-phase product from the methanol synthesis reactor (R-490) must be cooled to recover the methanol and to allow unconverted syngas and any inert gaseous species (CO_2, CH_4) to be recycled or purged. The reactor effluent, a mixture of methanol and unconverted syngas, is cooled from 482°F (250°C) to 140°F (60°C) through a series of heat exchangers, including heat exchange with the steam cycle and other process streams and air coolers. The liquid methanol is recovered by condensing it and separating the liquids from the residual syngas. Cooling water is used to further lower the temperature to 110°F (43.3°C), a temperature at which a majority of the liquid methanol condenses and is separated in a knock-out vessel (S-471). About 8,208 lb/h (18,057 kg/h) of methanol, or 13% of the total methanol, is not recovered from the product stream at this temperature. Almost 95 wt% of the unconverted syngas is recycled back to the entrance of the synthesis reactor [31], with 5% of the recycle gas being purged and sent to the tar reformer to prevent excessive buildup of inert gases in the synthesis loop and expanded through a turbo-expander (K-412) to recover power (~2 MW).

Area 400 Design Basis

The current AGR system design is based on a detailed analysis of the methyl diethanolamine (MDEA) system removing acid gases from NREL biomass-derived syngas conducted by Nexant Inc. (San Francisco, CA) [32]. The design basis for the AGR system is presented in Table 11.

26

Table 11. Acid Gas Removal Design Parameters

Acid Gas Removal Parameter	Value
Amine used	Methyl diethanolamine (MDEA)
Amine concentration	50 wt%
Amine circ. rate	2,877 gpm (10,888 L/min)
Amine temp. @ absorber	110°F (43.3°C)
Absorber pressure	735 psia (5.1 MPa)
Stripper condenser temperature	212°F (100°C)
Stripper reboiler temperature	230°F (110°C)
Stripper pressure	65 psia (449 kPa)
Stripper reboiler duty	159 MMBtu/h (167 GJ/h)
Stripper condenser duty	39 MMBtu/h (41 GJ/h)
Amine cooler duty	26 MMBtu/h (28 GJ/h)
Heat duty per pound CO_2 removed	3,029 Btu/lb (7 MJ/kg)
CO_2 removed	77,196 lb/h (35,089 kg/h)

The process conditions for the methanol synthesis reactor system are presented in Table 12.

Table 12. Process Conditions for Methanol Synthesis

Parameter	Conditions Used in Process Design & Aspen Model
Temperature, °F (°C)	482 (250)
Pressure, MPa (psia)	5.0 (730)
H_2:CO ratio	2.8
CO_2 concentration (mol %)	5.0%
Sulfur concentration (ppmv)	0
Stoichiometric number -	
$(H_2-CO_2)/(CO+CO_2)$	2.04
$H_2/(2CO+3CO_2)$	1.02

Area 400 Equipment Cost Estimations

The estimated purchased equipment costs and total installed costs for the methanol synthesis area were gathered from various sources including technology licensors, industrial suppliers, published literature, and Aspen Icarus Process Evaluator 2006.5 computer software. Table 13 presents the equipment list and cost estimates for the methanol synthesis area of the plant.

Table 13. Cost Estimate for Methanol Synthesis Area

Equipment Description	TPEC (2011 $k)	TIC (2011 $k)
Methanol synthesis reactor	3,215	8,360
Amine acid gas removal (AGR) system	3,487	9,100
LO-CAT sulfur recovery system	3,918	5,295
Unconverted syngas recycle compressor	2,114	5,221
Syngas 2nd compressor	759	1,875
ZnO sulfur removal beds	84	208
Pressure swing adsorption unit	22	54
Heat integration	355	1,082
Others	2,414	6,061
Area A400 subtotal	**16,368**	**37,256**

3.5 Area 500: Methanol Conditioning

The following section presents an overview, basis for design, and cost estimates for construction of the methanol conditioning area. Process flow diagrams for the methanol conditioning area are shown in PFD-61713-A501 and PFD-61713-A502 in Appendix D.

Area 500 Overview

The liquid methanol containing 2 vol % water from Area 400 is sent to Area 500 where it is de-compressed and de-gassed, as illustrated in Figure 8. The methanol is still at elevated pressure (567 psia/39 bar) at this point in the process, resulting in a significant quantity of gas being absorbed in the methanol stream as it leaves the synthesis section of the process. These gases are removed from the methanol at this stage of the process and are then recycled to the reformer inlet. The crude liquid methanol stream is cooled to 110°F (43°C) (H-504B) and subsequently is sent to a distillation column (D-500) to de-gas the methanol. The de-gassed methanol product is sent to a storage tank for short-term surge buffering between the methanol synthesis and Area 1400 (i.e., methanol to DME and DME to hydrocarbon synthesis) sections of the plant.

The composition of this methanol product is shown in Table 14. After the de-gassing step, the product is 95.8% methanol, with the remainder being mainly water.

Table 14. Composition of Crude Methanol Intermediate in Model

	NREL Model (vol %)
Methanol	95.8%
CO_2	1.3%
H_2O	1.9%
Others	1.0%
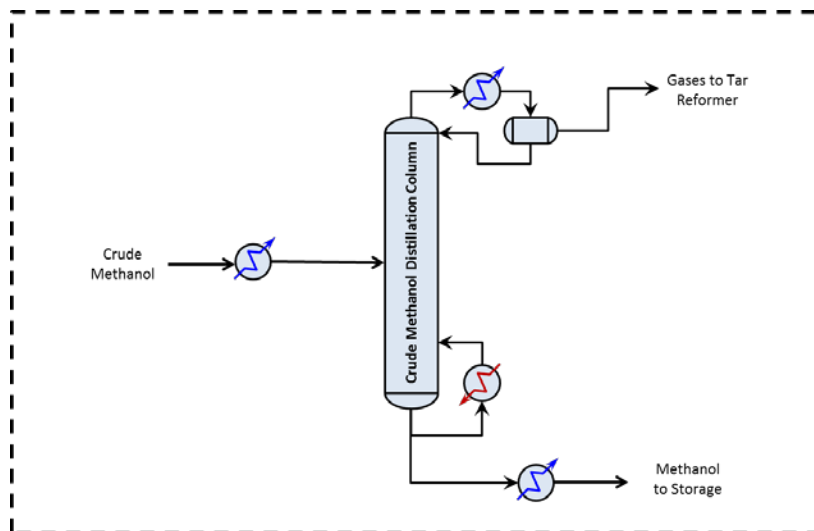
Figure 8. Simplified process flow diagram of methanol conditioning area

Area 500 Design Basis

Crude liquid methanol from the methanol synthesis reactor (95.8 wt% methanol) is fed to a distillation column after being cooled. The methanol distillation column operates in a mode that removes dissolved gases (mostly CO_2) to the overhead. The removed gas stream is sent to the tar reformer while the de-gassed methanol product is cooled to 145°F (62.8°C) and then sent to a storage tank. The column is modeled with rigorous vapor-liquid equilibrium calculations in Aspen Plus using the RADFRAC model.

Area 500 Equipment Cost Estimations

Table 15 presents the equipment list and cost estimates for the methanol conditioning area of the plant.

Table 15. Cost Estimate for Methanol Conditioning Area

Equipment Description	TPEC (2011 $k)	TIC (2011 $k)
Methanol degassing column (ancillaries included)	451	1,115
Heat integration	172	783
Others	661	1,508
Area A500 subtotal	**1,258**	**3,406**

3.6 Area 1400: Methanol to High-Octane Gasoline Synthesis

This section presents an overview, basis for design, and cost estimates for construction of the high-octane hydrocarbon synthesis process. Process flow diagrams for Area 1400 are shown in PFD-61713-A1401, PFD-61713-A1402, PFD-61713-A1403, and PFD-61713-A1404 in Appendix D.

29

Area 1400 Overview

This area of the process includes the conversion of methanol to DME and the subsequent conversion of DME to high-octane hydrocarbons. Figure 9 depicts a simplified process flow for the area.

From storage, methanol is sent to an adiabatic packed bed reactor (R-1424) of approximately 2,000 ft^3 in volume to convert the methanol to DME on commercially available gamma alumina (γ-Al$_2$O$_3$) catalyst at 482°F (250°C) and 140 psia (9.7 bar). Among various alumina-based catalysts, the γ-Al$_2$O$_3$ catalyst generally exhibits the best catalytic performance and hydrothermal stability for vapor-phase dehydration of methanol to DME [33]. The catalytic vapor-phase dehydration of methanol to DME is an exothermic reaction. The reactor heat is recovered with an intercooler for steam generation to allow for the adiabatic temperature rise with a targeted maximum reactor temperature of 482°F (250°C). DME is assumed to exit the methanol-to-DME reactor in equilibrium with methanol at the reactor exit temperature (88.5% conversion of methanol). The stoichiometric reaction for DME synthesis is summarized in Equation 7:

$$2\ CH_3OH \leftrightarrow CH_3OCH_3 + H_2O \qquad\qquad \text{(Eq. 7)}$$
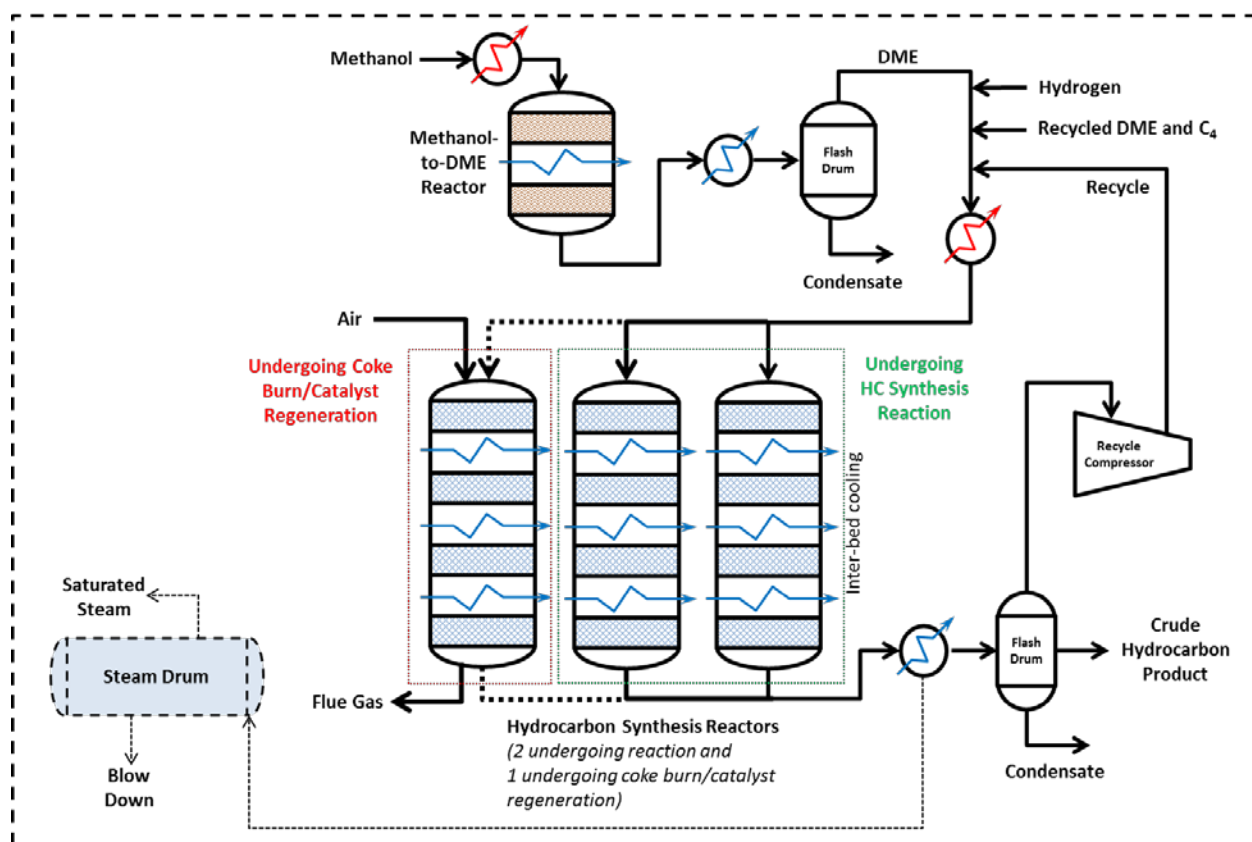
Note: Detailed process flow diagram for hydrocarbon synthesis reactors is shown in PFD-61713-A1402 in Appendix D.

Figure 9. Simplified flow diagram of methanol to hydrocarbons synthesis area

Table 16 lists the reaction conditions and yields for the methanol-to-DME process. DME and water are the only products. The effluent from the DME reactor is cooled to 250°F (121°C) to remove approximately 85% of water and subsequently sent to hydrocarbon synthesis reactors.

30

Table 16. Methanol-to-DME Reaction Conditions and Yields

Temperature	°F (°C)
Reactor inlet	437 (225)
Reactor outlet	482 (250)
Pressure	**psia (bar)**
Reactor inlet	145 (10)
Reactor outlet	140 (9.7)
Yield	**wt%**
DME	71
Water	29
Total	100

DME reactor (R-1424) effluent is sent to the four-stage adiabatic packed bed reactors (R-1410C) containing metal modified beta-zeolite (H-BEA) catalyst. High-octane gasoline is produced by the combination of acid-catalyzed homologation of DME and methylation of olefins. It has been demonstrated that the H-BEA catalyst exhibits some of the highest productivities and selectivities among zeolites studied for the homologation of DME to branched paraffins and olefins [3, 34, 35]. DME is preferentially converted to C_4 and C_7 hydrocarbons. The predominant isomers for isobutyl species are isobutane and isobutene. Triptyl species (2,2,3-trimethylbutane and 2,2,3-trimethyl-but-1-ene) make up about 75% of all C_7 species. Rates increase but selectivities transition to an aromatic-dominated product with increasing temperature (especially above 250°C). In this model, DME is converted to gasoline-range hydrocarbons at a maximum temperature of 450°F (232°C). In the Aspen Plus simulation, the selectivity and conversion of each compound is set to match targets that are believed to be attainable through near-term research efforts, as summarized in Table 17. With the current metal modified beta-zeolite catalyst, the C_4 products are recycled to the reactors for methylation of olefins to larger hydrocarbons. Unconverted DME is also recycled to the reactors for additional homologation. The single-pass conversion of DME is about 40% and the resulting overall DME conversion (including recycle) is 92.5%. Hydrogen addition is considered for the current target process assessment and has the purpose of reducing aromatic byproducts. Hexamethylbenzene (HMB) is used as a proxy for coke and heavy aromatic deposits on the catalyst in this analysis, given the literature precedent for its formation in this process [34]. The inclusion of HMB is also used to satisfy the overall stoichiometry of the reactions and is summarized in Equation 8 [34]:
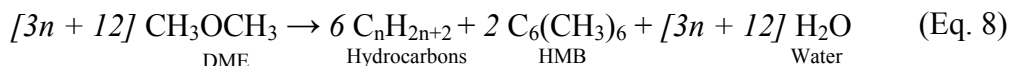
$$\textit{[3n + 12]} \underset{\text{DME}}{CH_3OCH_3} \rightarrow 6 \underset{\text{Hydrocarbons}}{C_nH_{2n+2}} + 2 \underset{\text{HMB}}{C_6(CH_3)_6} + \textit{[3n + 12]} \underset{\text{Water}}{H_2O} \qquad \text{(Eq. 8)}$$

Temperature control and heat removal from the hydrocarbon synthesis reactors is accomplished using adiabatic reactors, each with four fixed catalyst beds and inter-bed cooling (R-1410C). The heat is captured as low pressure steam (330°F/166°C and 103 psia). Water is supplied from process water condensate from this area and steam flows to the reformer in Area 300.

Catalyst deactivation is assumed to result from the formation of unsaturated residues like HMB. It has been reported that HMB is a precursor to coke formation, which can lead to reversible catalyst deactivation on beta-zeolite catalyst under methylation conditions [36]. It is assumed that while two reactors are in DME-to-hydrocarbons service, a third reactor is in coke-burn/catalyst regeneration. The catalyst regenerator burns coke deposits off the catalyst particles,

31

regenerating the catalyst activity and providing heat for steam generation. The flue gas is then utilized for heat recovery before it is mixed with flue gas from the char combustor in the gasification area (Area 200) and from the fuel combustor (R-301A) in the syngas clean-up area (Area 300). To protect the structural integrity of the zeolite catalyst, temperature is regulated with either air flow or recycled flue gas, and heat generated during catalyst regeneration is removed by steam generation.

Table 17. Target and Design Basis Parameters for DME to Hydrocarbons Synthesis Reactor

Technical Target Parameter			
Inlet temperature, °F (°C)			350 (177)
Inlet pressure, psia (bar)			129 (8.9)
Single-pass DME conversion			40%
Overall DME conversion			93%
%Trimethylbutane(ene) in C_7 fraction			75%
Space velocity, L/kg-cat/h			692

Carbon Number	Carbon Selectivity	Species	Species Selectivity
C_1	1.0%	Methane (CH_4)	100%
C_2	0.5%	Ethane (C_2H_6)	60%
		Ethene (C_2H_4)	40%
C_3	1.2%	Propane (C_3H_8)	99%
		Propene (C_3H_6)	1%
C_4	3.7%	Methylpropane (C_4H_{10})	96%
		N-Butane (C_4H_{10})	1%
		2-Methylpropene (C_4H_8)	2%
		But-1-ene (C_4H_8)	1%
C_5	5.0%	2-Methylbutane (C_5H_{12})	57%
		2-Methylbutene (C_5H_{10})	43%
C_6	6.0%	2-Methylpentane (C_6H_{14})	81%
		2-Methylpentene (C_6H_{12})	19%
C_7	39.8%	2,2,3-Trimethylbutane (C_7H_{16})	65%
		Heptane (C_7H_{16})	12%
		2,2,3-Trimethylbutane (C_7H_{14})	7%
		Hept-1-ene (C_7H_{14})	12%
		2-Methylhexane (C7H16)	5%
C_8	41.3%	2,2,3,3-Tetramethylbutane (C_8H_{18})	16%
		2,4,4-Trimethyl-1-pentene (C_8H_{16})	4%
		2,2,4-Trimethylpentane (C8H18)	81%
C_9	1.0%	Nonane (C_9H_{20})	80%
		Non-1-ene (C_9H_{18})	20%
Aromatics	0.5%	Hexamethylbenzene ($C_6(CH_3)_6$)	100%
Total	100%		

Note: The species presented in this table are model compounds used for modeling the process and bulk properties of gasoline product. However, actual isomers from experiments and research may vary from the model compounds.

Table 18 lists the reaction conditions and yields for the DME to hydrocarbons process. The effluent from the hydrocarbon synthesis reactor effluent is cooled to 375°F (191°C) to fully utilize all the available process water condensate in the area for reformer steam. The effluent gases are further cooled and then flashed at 250°F (121°C) through heat exchangers, which are used to heat other process streams. Liquid hydrocarbon products and water are recovered by additional condensation to 110°F (43.3°C) via air coolers and trim coolers and sent for purification in Area 1500. Process water condensate contains trace amounts of various hydrocarbons and is used for reformer steam.

Table 18. DME to Hydrocarbons Reaction Conditions and Yields

Temperature	°F (°C)
Reactor inlet	350 (177)
Reactor outlet	438 (225)
Pressure	**psia (bar)**
Reactor inlet	130 (9.0)
Reactor outlet	90 (6.2)
DME Conversions	
Single-pass	40%
Overall (with recycle)	93%
Reactor Outlet	**wt%**
Hydrocarbons	88%
Water	12%
Total	100%
Crude Hydrocarbon Product	**wt%**
Light gas	0.4%
Propane	2.0%
Propylene	0.02%
Isobutane	16%
n-Butane	0.3%
Butenes	0.4%
C5+ gasoline	81%
Total	100%

Area 1400 Design Basis

DME is produced from methanol on gamma alumina (γ-Al_2O_3) catalyst. Table 19 summarizes the design basis for the DME synthesis reactor system. The reaction is equilibrium limited and controlled. The corresponding single-pass methanol conversion is 88.5% at 482°F (250°C) and 140 psia.

The design basis for the high-octane hydrocarbon synthesis reactor system is presented in Table 17. The C_7 selectivity is assumed to be 38%, with 75% of all C_{7s} assumed to be trimethylbutanes(enes). Reactor design parameters are presented in Table 20. The sizing of the hydrocarbon synthesis reactor is calculated based on the space velocity of 692 L/kg-cat/h.

Catalytic homologation is an exothermic reaction. It is imperative to manage adiabatic temperature rise during the high-octane hydrocarbon synthesis. The current design uses multiple

This report is available at no cost from the National Renewable Energy Laboratory (NREL) at www.nrel.gov/publications.

stage adiabatic packed bed reactors with inter-stage cooling to control temperature and remove reactor duty. The heat is captured as low pressure steam (330°F/166°C and 103 psia). Boiling water cooled shell and tube reactors were also considered, but were found to be infeasible due to low reactor duty-to-volume ratio.

Table 19. Design Basis and Technical Target Parameters for DME Synthesis Reactor

Design Basis Parameters	Inlet	Outlet
Temperature, °F (°C)	437 (225)	482 (250)
Pressure, psia (bar)	145 (10)	140 (9.7)
Methanol conversion, %		88
DME Productivity, kg/kg-cat/h		0.35
Inlet Gas Composition		
Methanol, mol%		98
Water, mol%		2
Total		100
Reactor Design Parameters		
Reactor diameter, ft		8.88
Reactor length (max.), ft		48
Catalyst bed volume, ft^3		2006
Number of reactors		1
Intercooling		Yes
Gas hourly space velocity, hour^{-1}		601
Reactor pressure drop, psi		5
Reactor construction material		321 SS

There are three reactors for this reactor design configuration, as shown in Figure 9, with the detailed process flow diagram shown in PFD-61713-A1402 in Appendix D. Hydrocarbon synthesis reactions take place in two reactors while coke burn and catalyst regeneration is allowed to be carried out for the third reactor. The reactors are rotated through conversion and regeneration services with a lead-lag strategy (lead reactor train to coke burn, regenerated reactor train to lag, lag to lead). It is assumed that the catalyst has a two-year replacement cycle.

Table 20. Reactor Design Parameters for DME to Hydrocarbons Synthesis Reactor

Design Basis Process Parameters	
Minimum inlet temperature, °F (°C)	350 (177)
Maximum out temperature, °F (°C)	438 (225)
Operating pressure, psia (bar)	90–130 (6.2–9.0)
Space velocity, m^3/kg-cat/h	0.69
Inlet gas composition (1st reactor):	
DME, mol%	32.9%
Methanol, mol%	3.3%
H$_2$O, mol%	3.8%
H$_2$, mol%	3.7%
Methane, mol%	23.3%
Isobutane, mol%	19.0%
Other hydrocarbons, mol%	14.1%

Table 21 presents the equipment list and estimated purchased equipment costs and total installed costs for the methanol to high-octane hydrocarbons synthesis area of the plant. Preliminary reactor costs were furnished by Harris Group. All other equipment was priced using cost basis from the earlier design reports [4, 5].

Table 21. Cost Estimate for Methanol to High-Octane Gasoline Synthesis Area

Equipment Description	TPEC (2011 $k)	TIC (2011 $k)
DME synthesis reactor and effluent cooling/flash	1,331	3,010
Hydrocarbons synthesis reactor and effluent cooling/flash	2,578	6,560
Hydrocarbons synthesis reactor intercoolers	3,862	10,404
Hydrocarbon synthesis reactor recycle compressor	7,860	14,147
Methanol intermediate pump and storage tank	42	103
Coke burn/catalyst regeneration air blower	144	163
Heat integration	2,166	4,331
Area A1400 subtotal	**17,982**	**38,718**

3.7 Area 1500: Product Recovery

This section presents an overview, basis for design, and cost estimates for construction of the high-octane gasoline separation process.

Area 1500 Overview

Area 1500 recovers the crude high-octane hydrocarbons from Area 1400 into gasoline and fuel gas. It also recycles the C_4 products and unconverted DME to the reactors in Area 1400 for additional homologation to larger hydrocarbons. The process schematic for the crude gasoline separation area is shown in Figure 10 with the corresponding process flow diagrams shown in PFD-61713-A1501 and PFD-61713-A1502 in Appendix D. The design used in this model is noticeably simpler than the current separations employed in hydrocarbon recovery process in petroleum refineries.

Distillation is used to separate the final products and is accomplished in two distillation columns. The first, called the Deethanizer (D-1503), separates and removes the lighter hydrocarbons (C_3-) from the crude gasoline stream. The overhead from this column contains about 99% of C_3- hydrocarbons and DME. The fuel gas (C_3-) from the overhead of the Deethanizer (D-1503) is sent to the tar reformer in Area 300. DME is recovered and recycled to the hydrocarbon synthesis reactor in Area 1400.
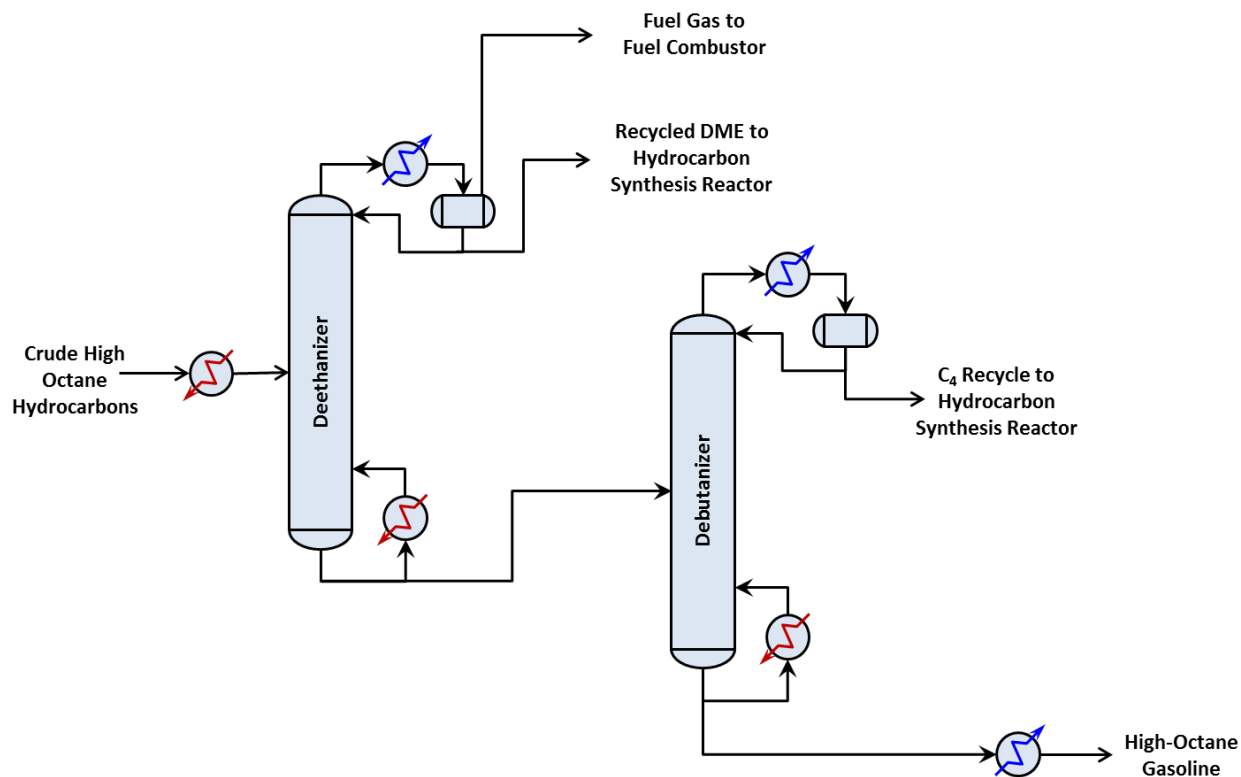
Figure 10. Process schematic of product separation area

The bottoms product from the Deethanizer (D-1503) is sent to the second column called Debutanizer (D-1504) to remove C_{4s}. The column is operated to remove most of the butanes and butenes from the high-octane gasoline-range hydrocarbons (C_{5+}). Once C_{4+} materials are removed, the high-octane gasoline is sent from the Debutanizer to product storage for sales.

The overhead product from the Debutanizer (C_{4+}) is recycled to the hydrocarbon synthesis reactor (R-1410C). To mitigate the buildup of butanes/butenes, a portion (5%) of the recycle stream is purged. The purged stream is mixed with the fuel gas from Deethanizer (D-1503); the fuel gas is sent to the fuel combustor in Area 300 for heat generation. Table 22 shows the finished fuel product distribution. The final composition of the high-octane gasoline from the current methanol to hydrocarbons process is compared with those from a typical conventional gasoline and the reported composition for MTG gasoline [5], as presented in Figure 11. Gasoline produced from this design contains only paraffins (87 wt%) and olefins (13 wt%). The high-octane number of the gasoline is attributed to its high concentration of branched paraffins (99% of the total paraffins). Furthermore, as opposed to MTG, no durene is formed in this high-octane gasoline process. In this design, hydrogen is added to suppress formation of coke and aromatic byproducts (coke pre-cursors), represented by the model-compound HMB. Coke/HMB is removed from the catalyst surface by coke combustion with air during off-line catalyst regeneration, as depicted in Figure 9. A comparative summary of the characteristics of the pathway from biomass feedstock to high-octane gasoline blendstock relative to those of the traditional MTG process is presented in Appendix H.

Table 22. Finished Fuel Products

Finished Fuel Products	wt%
Gasoline	95%
Fuel gas	5%
Total	100%
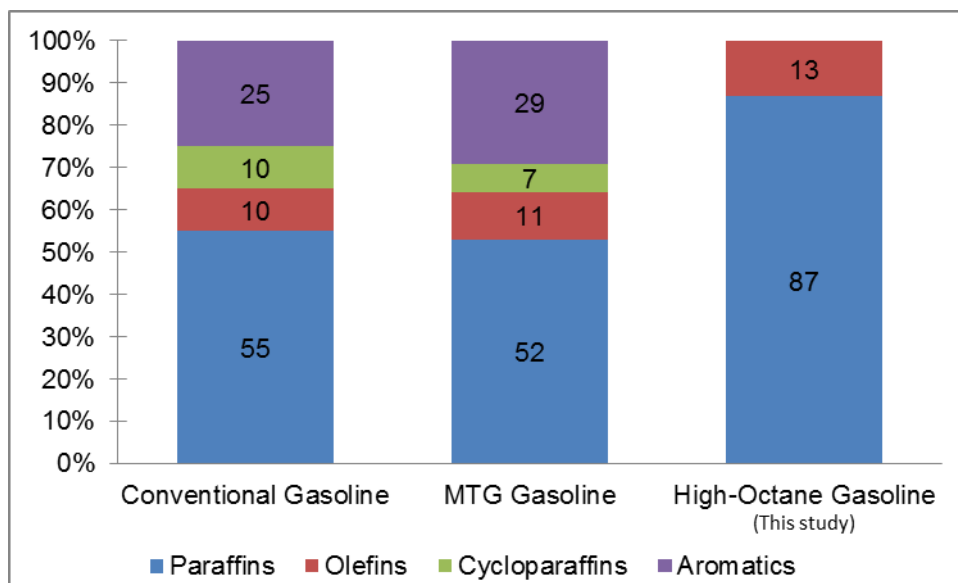
Note: For high-octane gasoline, over 98% of the paraffins and olefins are branched.

Figure 11. Comparison of gasoline composition (in wt%)

Area 1500 Design Basis

All distillation columns in the gasoline separation area are modeled with rigorous vapor-liquid equilibrium calculations in Aspen using the RADFRAC model. The associated design basis and assumptions are summarized in Table 23.

Table 23. Design Basis Parameters for Product Recovery Area

Distillation Column Design Parameters

Columns	Deethanizer	Debutanizer
Recovery in overhead	99% DME	99% (nC$_{4-}$)
Recover in bottoms	99% (C$_{4+}$)	99.7% (iC$_{5+}$)
Column diameter	6.0 ft	3.5 ft
Number of theoretical trays	38	38
Tray efficiency	60%	60%
Number of actual trays	63	63
Type of trays	Valve	Valve
Tower and trays material	304SS	304SS

Heat Exchanger and Vessel Design Parameters

Column overhead condenser cooling media	Cooling water
Column reboiler heating media	Process heat
Column overhead accumulator sizing	10-minute liquid inventory 3 vapor space factor
Product rundown cooling media	Process + cooling water

Area 1500 Equipment Cost Estimations

Table 24 presents the equipment list and cost estimates for the gasoline separation area of the plant. Column costs were provided from vendors. All other equipment was priced using cost basis from the earlier design reports [4, 5], which in turn were gathered from various sources including technology licensors, industrial suppliers, published literature, and Aspen Icarus Process Evaluator 2006.5 computer software.

Table 24. Cost Estimate for Product Recovery Area

Equipment Description	TPEC (2011 $k)	TIC (2011 $k)
Deethanizer column (ancillaries included)	502	1,239
Debutanizer column (ancillaries included)	470	1,161
Crude hydrocarbons pump	13	31
High-octane gasoline product water cooler	24	105
High-octane gasoline product pump	10	25
Heat integration	207	639
High-octane product storage tank	1,197	2,213
Area A1500 subtotal	**2,423**	**5,413**

3.8 Area 600: Steam System and Power Generation

The following section presents an overview, basis for design, and cost estimates for construction of the steam system and power generation facilities.

Area 600 Overview

The present design includes a steam cycle that produces steam through recovery of heat from the hot process streams throughout the plant. Operations requiring steam include gasification,

reforming, acid gas removal, LO-CAT preheater, methanol distillation, and high-octane hydrocarbons separation distillation columns. Steam is directly injected into the gasifier and the reformer. The rest of the plant heat demands are provided by indirect heat exchange of steam with process streams that have condensate return loops.

Power for plant operations is produced from the steam cycle using a three-stage steam turbine (M-602) with intermediate reheat to increase electricity production. The steam turbine efficiencies are assumed to be 75% and the generator mechanical efficiencies are assumed to be 97%. The first stage turbine reduces the steam pressure from 466 psia to 176 psia. The second stage turbine reduces the pressure from 176 psia to 65 psia. The third stage turbine drops the pressure from 65 psia to 3.5 psia. The steam inlets to all three turbine stages are preheated to 1,000°F. Slip-streams of low pressure steam are extracted at 65 psia from the second stage for the uses described above. Power is also produced from the process turbo-expander system in Area 400. The plant energy balance is controlled by the rate of syngas combustion in the reformer catalyst regenerator. The energy from syngas combustion is recovered from the flue gas and contributes to the conversion to electrical energy in the steam turbine generators. The rate of syngas combustion is controlled to generate only enough electricity to supply the plant, minimizing electricity exported to the grid.

Low pressure (3.5 psia) steam exhaust from the turbine is cooled by exchanging heat with other process streams and is condensed in an air-cooled condenser. While an air-cooled exchanger is considerably more expensive than a water cooler, it has the advantage of reducing water consumption through evaporation in the cooling towers and was preferred in this design for environmental sustainability considerations.

A condensate collection tank (T-606) gathers condensate returns from indirect heat exchange along with the steam turbine condensate. The condensate from the collection tank is treated in the hot condensate polishing unit, mixed with treated makeup water, and sent to the condensate surge tank. From the surge tank, the water is pumped, heated to the saturation temperature, and sent to the deaerator (T-603) to remove any dissolved gases from the water. The water from the deaerator (T-603) is pumped to a pressure of 486 psia and then preheated to its saturation (bubble point) temperature using a series of exchangers. Steam is generated by heat exchange with flue gas and on the shell side of the methanol synthesis reactor (R-490). This saturated steam is collected in the steam drum (T-605). To prevent buildup of solids, water is periodically discharged from the steam drum (also known as blowdown). Saturated steam from the steam drum is superheated with another series of exchangers to 1,000°F (538°C) before it goes to the steam turbine (M-602).

The PFDs (PFD-61713-A601, PFD-61713-A602, and PFD-61713-A603) of the integration of the individual heat exchangers are presented in Appendix D. The Aspen Plus model increases or decreases the boiler feed water flow rate through the steam cycle in order to close the heat balance. The control system in the plant will function in a similar fashion by controlling the flow rate of BFW makeup to the steam system in order to maintain levels in the de-aerator (T-603) and steam drum (T-605).

All compressors, pumps and fans utilize electric motor drives and are modeled and quoted accordingly. An allowance of 1 MW of excess power is specified to account for miscellaneous usage and general electric needs (such as lights and computers).

Area 600 Design Basis

Table 25 presents a summary of the power requirements for each area of the plant. As all necessary power is generated in the plant, these power requirements define the basis for design of the steam system and power generation facilities.

Table 25. Plant Power Requirements

Plant Area	Generated (kW)	Consumed (kW)
Feed handling and preheating[a]		742
Gasification		6,359
Tar reforming and gas cleanup, syngas compressor		19,529
Methanol synthesis, recycle compressor, turbo-expander, 2nd syngas compressor	1,535	5,669
Methanol conditioning		2
Methanol to high-octane hydrocarbons conversion		2,406
Gasoline separation		35
Steam system and power generation	35,344	774
Cooling water and other utilities		359
Miscellaneous		1,000
Balance to grid		4
Plant totals	**36,879**	**36,879**

a Power costs for all feed handling, and for drying outside the plant, are included in the feedstock cost provided by INL. Energy for cross-flow drying inside the plant is provided by process heat from flue gas.

Plant power demands and power production rates are designed to be balanced. Therefore, no significant excess power is claimed for byproduct credits (sale to the grid) or bought from the grid. The plant design was conducted with an aim of energy self-sufficiency, and this was accomplished by burning various on-board generated hydrocarbons—including a small portion of unreformed syngas from Area 200, the unconverted syngas purged from the methanol synthesis reactor and hydrocarbon synthesis recycle loops in Areas 400 and 1400, respectively, and fuel gas from Area 1500—in the tar reformer fuel combustor (Area 300).

Area 600 Equipment Cost Estimations

Table 26 presents the equipment list and cost estimates for the steam system and power generation area of the plant.

Table 26. Cost Estimate for Steam System and Power Generation Area

Equipment Description	TPEC (2011 $k)	TIC (2011 $k)
Water treatment system (RO and EDI units)	2,499	2,962
Boiler feed water/steam/condensate handling	1,413	3,115
Extraction steam turbines/generators	13,216	23,789
Steam turbine condenser	5,623	8,244
Area A600 subtotal	**22,751**	**38,110**

3.9 Area 700: Cooling Water and Other Utilities

The following section presents an overview, basis for design, and cost estimates for construction of the cooling water facilities and other plant utility equipment. Process flow diagrams of the area are shown in PFD-61713-A701, PFD-61713-A702, and PFD-61713-A703 in Appendix D.

Area 700 Overview

For the cooling water and other utilities systems, a mechanical draft cooling tower provides cooling water to several heat exchangers in the plant. The tower utilizes fans to force air through circulated water.

An instrument air system is included to provide compressed air for both service and instruments. The system consists of an air compressor, dryer, and receiver. The instrument air is delivered at a pressure of 115 psia and a moisture dew point of -40°F (-40°C), and it is oil free.

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

- Firewater storage tank and firewater pump for fire emergencies
- Diesel tank and diesel pump to fuel the front loaders
- Olivine truck dump with scale and olivine/MgO loading system
- Tar reformer catalyst loading system
- Ammonia storage tank, ammonia pump, BFW chemical storage tank, and BFW chemical pump for BFW treatment
- Caustic (NaOH) storage tank and pump.

Area 700 Design Basis

Table 27 summarizes the design basis parameters for the plant's cooling water system.

41

Table 27. Design Basis for Cooling Water System

Cooling Water System Design Parameters	
Cooling water supply temperature (T_S)	90°F
Cooling water return temperature (T_R)	110°F
Cooling water supply rate (F_S)	10,967 gpm
Losses from evaporation (F_{EVAP})	$0.00085 * F_S * (T_R - T_S)$
Losses from drift	$0.002 * F_S$
Cycles of concentration (C)	5
Required blowdown	$F_{EVAP}/(C − 1)$
Cooling water makeup pressure	14.7 psia
Cooling water makeup temperature	60°F

Water losses from the cooling water system include evaporation, drift (water entrained in the cooling tower exhaust air), and tower basin blowdown. Drift losses were estimated to be 0.2% of the cooling water supply rate. Evaporation losses and blowdown were calculated based on information and equations in Perry et al. [9] (as shown in Table 27).

Area 700 Cost Estimation

Table 28. Cost Estimate for Cooling Water and Other Utilities Area presents the equipment list and cost estimates for the cooling water and other utilities area of the plant.

Table 28. Cost Estimate for Cooling Water and Other Utilities Area

Equipment Description	TPEC (2011 $k)	TIC (2011 $k)
Plant/instrument air systems	314	588
Cooling water system	515	1,141
Flue gas scrubber and stack	832	1,850
Chemicals storage	370	725
Miscellaneous utility systems	1,066	2,671
Area A700 subtotal	**3,096**	**6,975**

3.10 Additional Process Design Information

Table 29 contains additional information used in the Aspen Plus model and design basis.

42

Table 29. Utility and Miscellaneous Design Information

Item	Design Information
Ambient air conditions [a,b,c]	Pressure: 14.7 psia $T_{dry\ bulb}$: 90°F $T_{wet\ bulb}$: 80°F <u>Air composition (mol %):</u> N_2: 75.7% O_2: 20.3% Ar: 0.9% CO_2: 0.03% H_2O: 3.1%
Pressure drop allowance	Syngas compressor intercoolers = 2 psi per exchanger Air preheat exchangers = 1 psi Heat exchangers = 5 psi

[a] In the GPSA *Engineering Data Book* [37], see Table 11.4 for typical design values for dry bulb and wet bulb temperature by geography. Selected values would cover summertime conditions for most lower 48 states.
[b] In Weast [38], see F-172 for composition of dry air. Nitrogen value adjusted slightly to force mole fraction closure using only N_2, O_2, Ar, and CO_2 as air components.
[c] In Perry et al. [9], see psychrometric chart, Figure 12-2, for moisture content of air.

3.11 Pinch Analysis

A pinch analysis was performed to design the heat exchange network of the biomass gasification to high-octane gasoline production process. The concept of pinch technology offers a systematic approach for optimizing energy integration in process design. Temperature and heat flow (Q) data were gathered for the hot process streams (streams cooled in the process), cold process streams (streams heated in the process), and utility streams like steam, flue gas, and cooling water. A composite curve, a graphical presentation of stream temperature and heat flow (Q) data, was constructed for the hot and cold process streams. These two curves are shifted so that they touch at the pinch point. From this shifted graph, a grand composite curve is constructed that plots the difference in heat flow between the hot and cold composite curves as a function of temperature.

The total hot duty equals the total cold duty, as the Aspen Plus model is designed to adjust the water flow rate through the steam cycle until the heat balance in the system is closed. The minimum vertical distance between the curves is ΔT_{min}, which is theoretically the smallest approach temperature in the exchanger network. For this design, the pinch occurs at ~ Q = 19.2 MMBtu/h where the upper and lower pinch temperatures are 194.2°F and 172.8°F, respectively, resulting in ΔT_{min} of 21.4°F.

There are a multitude of options for heat integration network design, depending on objectives and available capital. In general, as efficiency is increased in an exchanger network design, the required heat transfer area increases and therefore capital cost increases. The objective of this network design is to lower energy consumption and boost product yields, keeping in mind that very low ΔT can lead to impractically large exchangers and high capital costs. Figure 12 shows the cumulative hot and cold pinch curves after they were simplified by removing dedicated steam generation (from the methanol synthesis reaction), indirect steam use systems (distillation reboilers and acid gas removal), and water- and air-cooling utilities. Heat exchanger network was not done for the current draft report. A specific network was designed based on this philosophy and the heating and cooling curves shown in Figure 12. The heat exchangers are listed in the equipment list in Appendix A, with the scaling stream entries listed as "Pinch." The exchangers in the heat integration network are also shown on the process flow diagrams in Appendix D, with

43

details of each exchanger listed in Appendix J. The total installed cost for the heat exchangers associated with the heat integration network is approximately $26MM. Aspen Icarus Process Evaluator is the source of the equipment cost and installation estimates for the heat integration network.
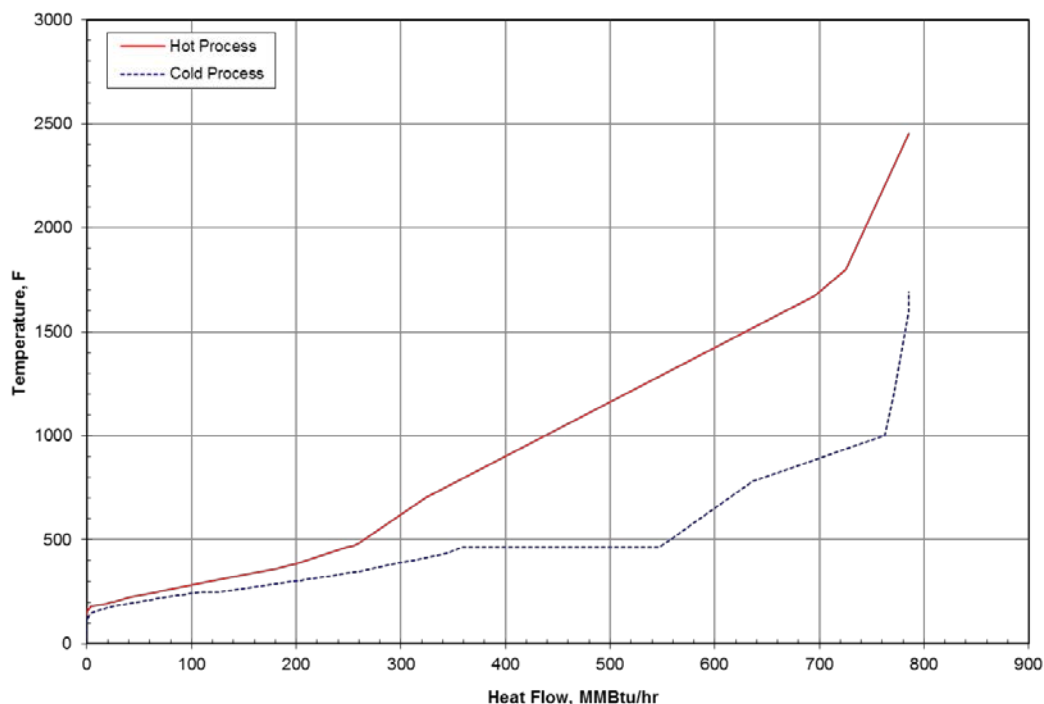
Figure 12. Pinch analysis composite curve

3.12 Energy Balance

Energy integration is important to the overall economics and efficiency of the process. Therefore a detailed understanding of how and where the energy is utilized and recovered is required. Detailed energy balances around the major process areas were derived using data from the Aspen Plus simulation. Comparing the process energy inputs and outputs allows the energy efficiency of the process to be quantified. Tracing energy transfer between process areas also makes it possible to identify potential improvements to the energy efficiency.

This design for thermochemical conversion of cellulosic biomass was constructed with the goal of energy self-sufficiency. Natural gas inputs that could be used to fire the char combustor and fuel combustor have been eliminated. Instead, fuel gas and a slip-stream of unreformed syngas and a portion of the unreacted syngas from the methanol synthesis reactor are used to meet the fuel demand. The downside to this is a decrease in methanol intermediate and final fuel yield. Because no electricity is purchased from the grid, the integrated combined heat and power system must supply all steam and electricity needed by the plant. A negligible amount of electricity is sold as a coproduct. The only saleable product is high-octane gasoline.

44

The distribution of the total energy content in dry biomass (LHV basis) is shown in Figure 13. The total energy in the dry portion of the biomass feed (0% moisture) is approximately 1,470 MMBtu/h. Figure 13 shows that approximately 45% of the LHV is recovered in gasoline-range product. A significant amount (31%) is lost through air-cooled exchangers, and the rest (24%) is lost because of moisture in the feed and other water input to the process, as well as electrical and thermal losses.

Further optimization of the heat integration network can be evaluated. However, it is our expectation that further heat integration for this process design will have marginal impact on the MFSP and should therefore be addressed only if there is a detailed plant design.
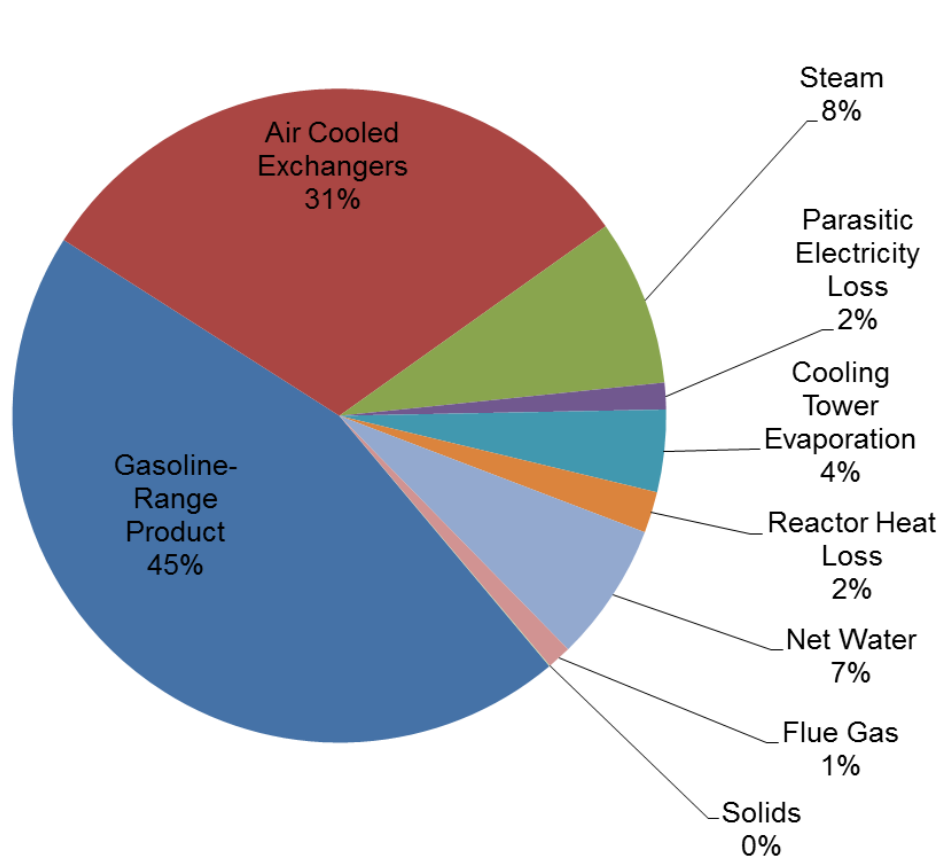
Figure 13. Overall energy analysis (dry biomass LHV basis)

3.13 Environmental Sustainability Metrics

This section covers the environmental sustainability assessment for the high-octane gasoline production process.

3.13.1 Environmental Sustainability Overview

In addition to setting technical and economic targets for the conversion pathways included in the Multi-Year Program Plan, BETO is working toward 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" [39]. 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. Table 30 lists the material and energy flow data for the modeled conversion process. Based on these data, the sustainability metrics are estimated for the conversion plant, as listed in Table 31.

The GHGs and fossil energy consumption metrics include both direct effects at the plant and indirect effects associated with the production and distribution of materials and energy for plant operations, i.e., the life-cycle emissions and energy usage for the conversion stage of the fuel supply chain. Water consumption and wastewater generation values in Table 31 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. While Table 30 provides a preliminary estimate of SO_2 and NO_2, efforts are currently underway to provide a more accurate assessment of these and other criteria air pollutant emissions from the biorefinery. The following sections provide a description of each sustainability metric and a discussion of the results.

Table 30. Material and Energy Flows for the High-Octane Gasoline Conversion Process

Products	Production Rate		
	lb/hr	gal/hr	MMBtu/hr (LHV)
High-Octane Gasoline	34,916	5,966	661
Byproducts			
Sulfur	119		
Excess electricity	101	hp	

Resource Consumption	Flow Rate	Per Gal Gasoline	Per MMBtu
	lb/hr	lb/gal	lb/MMBtu
Blended woody biomass (wet)	204,131	34	309
Blended woody biomass (dry)	183,718	31	278
Magnesium oxide (MgO)	7.0	1.2.E-03	1.1.E-02
Fresh olivine	543	9.1.E-02	0.82
Tar reformer catalyst	8.8	1.5E-03	1.3E-02
Methanol synthesis catalyst	5.2	8.7E-04	7.9E-03
DME catalyst	6.7	1.1E-03	1.0E-02
Beta zeolite catalyst	31	5.3E-03	4.8E-02
Zinc oxide catalyst	2	4.1E-04	3.7E-03
Cooling tower water makeup	1,986	0.33	3.0
Boiler feed water makeup	81,722	14	124
Dimethyl disulfide (DMDS)	2.1	3.5E-04	3.2E-03
Amine (MDEA) makeup	3.9	6.5E-04	5.9E-03
LO-CAT chemicals	119	2.0E-02	0.18
Boiler feed water chemicals	2.9	4.8E-04	4.33E-03
Cooling tower chemicals	0.51	8.6E-05	7.74E-04
No. 2 diesel fuel	69	1.2E-02	0.10

Waste Streams	lb/hr	lb/gal	lb/MMBtu
Sand and ash purge	2433	0.41	3.68
Tar reformer catalyst	8.8	1.E-03	1.3E-02
Scrubber solids	9.4	1.6E-03	1.4E-02
Wastewater	11,337	1.90	17.16

Air Emissions	lb/hr	lb/gal	lb/MMBtu
CO2 (biogenic)	235,909	40	357
CH4	0.0.E+00	0.0.E+00	0.0.E+00
CO	0.0.E+00	0.0.E+00	0.0.E+00
NO2	129	2.2.E-02	0.20
SO2	47	7.9.E-03	7.2.E-02
H2O	69,755	11.69	106
H2S	0.0.E+00	0.0.E+00	0.0.E+00

Table 31. Summary of Environmental Sustainability Metrics for High-Octane Gasoline Conversion Process

GHGs (g CO_{2-e}/MJ fuel) – (fossil emissions, excluding biogenic CO_2)	0.60
Fossil Energy Consumption (MJ fossil energy/MJ fuel product)	6.0E-03
Total Fuel Yield (gal/dry ton feedstock)	64.9
Biomass Carbon-to-Fuel Efficiency (C in fuel/C in biomass)	31.2%
Water Consumption (m^3/day; gal/GGE)[a]	911; 1.76
Wastewater Generation (m^3/day; gal/GGE)	123; 0.24

[a] 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.

3.13.2 Greenhouse Gas Emissions and Fossil Energy Use

BETO is developing technologies that will facilitate the volumetric requirements of the national Renewable Fuel Standard (RFS) [40], as legislated by the Energy Independence and Security Act (EISA) of 2007 [41]. 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 include having life-cycle GHG emissions that are 50% (advanced biofuel) or 60% (cellulosic biofuel) less than the petroleum baseline. While RFS mandates the EPA to conduct its own life-cycle assessment (LCA) to determine fuel qualification, it is essential that life-cycle 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 the techno-economic analyses, allowing continual evaluation and improvement of the design throughout the technology research and development phase.

Conversion processes can generate fossil-based GHG emissions stemming from direct use of fossil fuels 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). 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. The current high-octane gasoline production process does not require any direct non-renewable energy inputs at the plant. As mentioned in Section 3.12, this design for thermochemical conversion of cellulosic biomass was constructed with the goal of energy self-sufficiency. Natural gas inputs that could be used to fire the char combustor and fuel combustor have been eliminated. Instead, a slip-stream of unreformed syngas and a portion of the unreacted syngas from the methanol synthesis reactor are used to meet the fuel demand. Thus all direct CO_2 emissions from the plant, including from char combustion, fuel combustion, and beta-zeolite catalyst coke removal, are biogenic CO_2.

The material and energy flows for the modeled conversion process (Table 30), along with information from the Ecoinvent database [42] and U.S. Life-Cycle Inventory (LCI) database [43], are used to estimate GHGs and fossil energy consumption with the life-cycle modeling software [44]. GHG basis values for natural gas (NG) and the U.S. average electricity mix from the grid were applied consistently with the basis from Argonne National Laboratory's GREET

model software; this includes GHG emission factors of 3.25 kg $CO_{2\text{-}e}$/kg NG (LHV) and 0.65 kg $CO_{2\text{-}e}$/kWh purchased electricity [45]. Likewise, the basis from GREET was also applied for non-renewable (fossil) energy consumption values of 52.1 MJ/kg (LHV) natural gas and 7.5 MJ/kWh purchased electricity [45]. 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 conversion catalysts is necessary to fill this data gap.

Figure 14 shows the total life-cycle GHG emissions for the conversion plant and the contribution of individual conversion inputs. As the conversion plant is essentially balanced for zero net electricity and does not require natural gas for heat generation, the net GHG emissions associated with the conversion stage are very low (0.6 g $CO_{2\text{-}e}$/MJ).
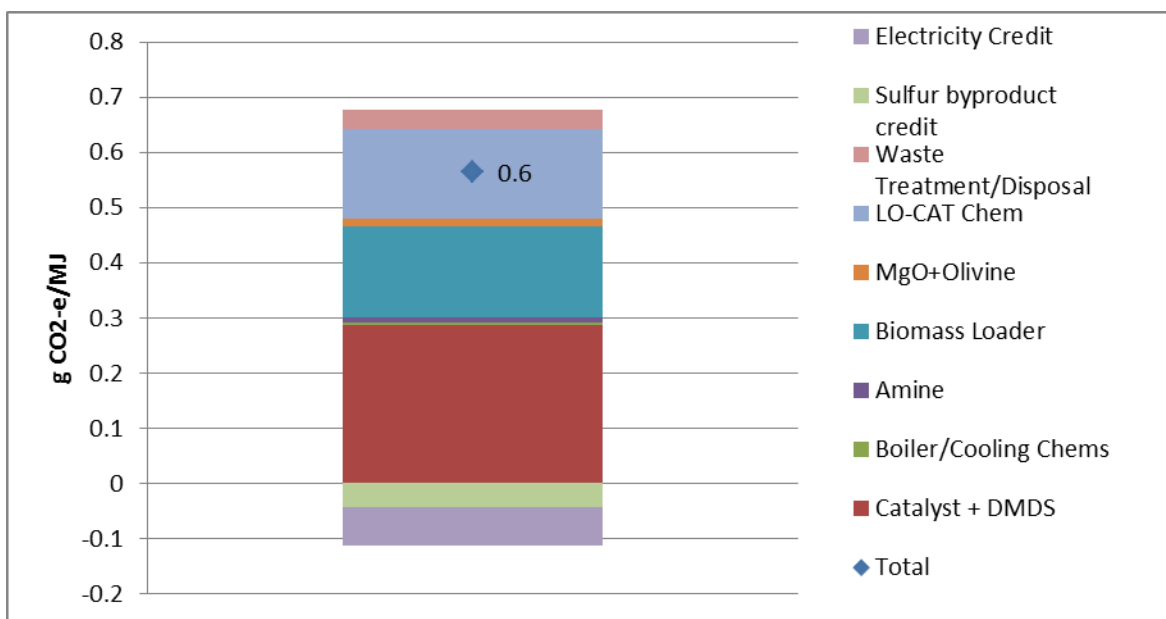
Figure 14. Contributing processes to conversion GHGs for biomass to high-octane gasoline

The analysis was expanded to estimate the full life-cycle GHG emissions for fuels produced from the projected 2017 model feedstock, a blend of pulpwood (45%), wood residues (32%), switchgrass (3%), and construction and demolition waste (20%) [11]. Full life-cycle GHGs for gasoline from the blended feedstock are estimated at 14.6 g $CO_{2\text{-}e}$/MJ which corresponds to an 84% GHG reduction relative to the 2005 petroleum gasoline baseline GHGs (93.08 g $CO_{2\text{-}e}$/MJ) [46]. The underlying assumptions and detailed results for the full life-cycle GHG analysis are given in Appendix F. While this analysis provides a preliminary estimate of GHG reduction associated with this biofuel, it is important to note that the EPA conducts its own LCA to make final determination of fuel qualification under the RFS2. In addition, Argonne National Laboratory will provide more detailed LCA for this pathway in the future.

3.13.3 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. 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). The carbon conversion efficiency of the design case is estimated to be 31.2% (Table 31) which is comparable to that achieved by the thermochemical ethanol process (31%) [25], but is noticeably higher than that from the biological conversion of the sugars to hydrocarbons process (26%) [19].

3.13.4 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 the high-octane gasoline production consists of makeup water for the steam and cooling systems. Boiler feed water makeup is needed to replace steam consumed in the reformer. 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.

Water consumed for production of materials used at the biorefinery (e.g., catalysts) and for life-cycle stages upstream and downstream of conversion is not included in the analysis. Boiler feed water makeup for steam production is approximately 68% of the total water consumed for the conversion processes. Approximately 12% of the total steam is consumed in the reforming area. Cooling water makeup is approximately 2% of the total water use. It is important to note that steam drum blowdown, cooling tower blowdown, and process water condensate are recycled for use for both steam generation and cooling, significantly reducing raw cooling water makeup. The water consumption for the current high-octane gasoline production is 1.55 gal/GGE, which is better than the crude oil refining to gasoline (2.8–6.6 gal/GGE) [47].

4 Process Economics

The first steps of the economic analysis are to determine the TCI, variable operating costs, and fixed operating costs associated with constructing and operating the plant. With these estimated costs, a discounted cash flow analysis is then used to determine the production cost of high-octane gasoline for a project having a net present value of zero at an assumed internal rate of return (IRR). This section describes the cost areas and the necessary assumptions required to complete the discounted cash flow analysis.

50

4.1 Total Capital Investment (TCI)

The following discussion summarizes the purchased and installed equipment costs presented in Section 3 of the design report and describes how the total installed costs (TIC) for the plant serve as the basis for determining the total capital investment (TCI).

As previously discussed, the capital costs for each plant area are based on data from various sources including NREL design reports, technology licensors, industry equipment suppliers, cost estimates from Harris Group Inc., published literature, and Aspen Icarus Process Evaluator 2006.5 cost estimating software. The purchased costs for the equipment and installation factors are used to determine the installed equipment cost. Table 32 presents a summary of total purchased equipment costs (TPEC), installation factors ($f_{installation}$), and TIC for each area of the plant, the bases for which are discussed in further detail in Section 0.

Table 32. Summary of Total Purchased Equipment Costs, Installation Factors, andTotal Installed Costs (2011$)

Process Area	TPEC (MM$)	$f_{Installation}$	TIC (MM$)
A100[a]	$0.1	2.00	$0.2
A200	$21.8	2.32	$50.5
A300	$36.4	1.94	$70.7
A400	$16.4	2.28	$37.3
A500	$1.3	2.65	$3.4
A1400	$18.0	2.15	$38.7
A1500	$2.4	2.23	$5.4
A600	$22.8	1.68	$38.1
A700	$3.1	2.25	$7.0
ISBL (Areas A100-A500, A1400, A1500)	$96.3	2.14	$206.1
OSBL (Areas A600, A700)	$25.8	1.74	$45.1
Total	**$122.2**	**2.06**	**$251.2**

[a] Investment costs for feed handling and drying are included in the per-unit woody feedstock price. Only costs associated with cross-flow dryers are included here.

The indirect costs (non-manufacturing fixed-capital investment costs) are estimated using factors based on the total direct cost (TDC). The factors are summarized in Table 33 as percentages of TPEC, total direct cost (TDC), and fixed capital investment (FCI), which is equal to the sum of TDC and total indirect costs. With the calculated total installed cost and assumptions for indirect costs and working capital, the FCI and TCI are calculated. Table 34 presents a summary of these calculations.

51

Table 33. Cost Factors for Indirect Costs

Indirect Costs	% of TPEC	% of TDC*	% of FCI*
Prorated expenses	21.2	10.0	6.3
Home office and construction fees	42.5	20.0	12.5
Field expenses	21.2	10.0	6.3
Project contingency	21.2	10.0	6.3
Other costs (start-up and permits)	21.2	10.0	6.3
Total indirect costs	**127.4**	**60.0**	**37.5**
Working capital			**5.0**

* Percentages of TDC and FCI exclude land purchase cost.

Table 34. Project Cost Worksheet

Total purchased equipment cost (TPEC)		**$122,168,355**
Installation factor		2.056
Total installed cost (TIC)		**$251,231,890**
Other direct costs		
Site development	4.0% of ISBL	$8,245,850
Total direct costs (TDC)		**$259,477,740**
Indirect costs	% of TDC (ex Land)	
Prorated expenses	10.0%	$25,947,774
Home office and construction fees	20.0%	$51,895,548
Field expenses	10.0%	$25,947,774
Project contingency	10.0%	$25,947,774
Other costs (start-up and permits)	10.0%	$25,947,774
Total indirect costs	**60.0%**	**$155,686,644**
Fixed capital investment (FCI)		**$415,164,385**
Land (not depreciated)		$1,610,000
Working capital	5.0% of FCI (ex Land)	$20,758,219
Total capital investment (TCI)		**$437,532,604**
TCI / TPEC		3.581
FCI Lang Factor = FCI / ISBL TPEC		4.310
TCI Lang Factor = TCI / ISBL TPEC		4.542

4.2 Variable Operating Costs

There are two types of operating costs—variable and fixed. The following sections present assumptions and calculation approaches to determine the variable and fixed operating costs associated with thermochemical ethanol production. The basis for variable operating costs considered in this analysis is summarized in Table 35. A detailed summary of variable plant operating costs is presented in Appendix B.

Table 35. Summary of Variable Operating Costs

Variable	Information and Operating Cost (cost year in parentheses)
Gasifier bed material	Synthetic olivine and MgO. Initial fill then a replacement rate of 0.01 wt% of circulation or 7.2 wt% per day of total inventory. Delivered to site by truck with self-contained pneumatic unloading equipment. Disposal by landfill. Olivine price: $172.90/ton (2004) [48]. MgO price: $365/ton (2004) [49].
Tar reformer catalyst	To determine the amount of catalyst inventory, the tar reformer was sized for a gas hourly space velocity (GHSV) of 2,476/h based on the operation of the tar reformer at NREL's TCPDU. GHSV is measured at standard temperature and pressure. Initial fill then a replacement rate of 0.15 wt% of catalyst inventory per day. Price: $17.64/lb (2007) based on NREL calculations using metals pricing and costs for manufacturing processes.
Methanol synthesis catalyst	Initial fill then replaced every 4 years based on expected catalyst lifetime. Catalyst inventory based on GHSV of 8,000/h. Price: $9.69/lb (2011) [5].
DME synthesis catalyst	Initial fill then replaced every 4 years based on expected catalyst lifetime. Catalyst inventory based on GHSV of 605/h. Price: $10.30/lb (2011) [5].
Beta-zeolite catalyst	Initial fill then replaced every 2 years based on expected catalyst lifetime. Catalyst inventory based on total gasoline productivity of 0.11 kg/kg-cat/h. Price: $30.80/lb (2011) based on cost survey from manufacturers.
Zinc oxide catalyst	Initial fill then replaced every 3 years based on expected catalyst lifetime. Catalyst inventory based on GHSV of 4,280/h. Price: $4.67/lb (2005) [50].
Solids disposal	Price: $36/tonne (1998) [8].
Diesel fuel	Usage: 10 gal/h plant-wide use. 2012 price projection: $21.29/MMBtu (2009) [44] = $2.86/gal at 0.85 specific gravity [(corrected to 2007 currency by EIA index (where 2007 = 106.2, 2009 = 109.8)].
Water makeup	Price: $0.22/tonne (2001) [8] = $0.20/ton.
Chemicals	Boiler feed water chemicals—Price: $1.40/lb (1991) [18]. Cooling tower chemicals—Price: $1.00/lb (1999) [18]. LO-CAT chemicals—Price: $408/ton sulfur produced (2009) from NREL/Harris Group Inc. estimates based on other projects. Selective amine makeup—Price: $16.94/million lb acid gas removed (2010).
Wastewater	Most wastewater is cleaned using an RO system and recycled. The balance of the wastewater is sent to off-site treatment facility. Price: $0.53/tonne (2001) [8].

Note: Costs shown in Table 35 were updated to 2011 dollars using an inorganic cost index [51].
Values used: 1991 = 125.6; 1998 = 148.7; 1999 = 149.7; 2001 = 158.4; 2004 = 172.8; 2007 = 203.3; 2010 = 233.9.; 2011 = 249.3; 2014 = 280.2 (extrapolated)..

4.3 Fixed Operating Costs

This section considers the plant's fixed operating costs, which include employee salaries and benefits, overhead, plant maintenance costs, insurance, and taxes (other than income taxes). The fixed operating costs used in this analysis are shown in Table 36 (labor costs) and Table 37 (other fixed costs).

Table 36. Salary Costs for Plant Employees

Position Title	Salary (2011)	Number of Positions	Total Cost (2011)
Plant manager	$161,362	1	$161,362
Plant engineer	$76,839	1	$76,839
Maintenance supervisor	$62,569	1	$62,569
Laboratory manager	$61,471	1	$61,471
Shift supervisor	$52,690	5	$263,448
Lab technician	$43,908	2	$87,816
Maintenance technician	$43,908	16	$702,527
Shift operators	$43,908	20	$878,159
Yard employees	$30,736	12	$368,827
Clerks and secretaries	$39,517	3	$118,551
Total Salaries (2011)			**$2,781,567**

Note: Labor costs are indexed, if necessary, to values from the U.S. Bureau of Labor Statistics (http://data.bls.gov/cgi-bin/srgate CEU3232500008)

Because the salaries listed above do not include benefits, a general overhead factor is applied. The factors for other fixed costs in Table 37 also include general plant maintenance, plant security, janitorial services, communications, etc. A detailed summary of fixed plant operating costs is presented in Appendix B.

Table 37. Other Fixed Operating Costs

Cost Item	Factor	Cost
Benefits and general overhead	90.0% of total salaries	$2,503,411
Maintenance	3.0% of fixed capital investment (FCI*)	$12,454,932
Insurance and taxes	0.7% of fixed capital investment (FCI*)	$2,906,151
Total Other Fixed Operating Costs (2011)		**$17,864,493**

* Percentages of FCI exclude land purchase cost.

54

4.4 Minimum Fuel Selling Price

Once the capital and operating costs are determined, a discounted cash flow rate of return (DCFROR) calculation was performed to determine the minimum fuel selling price (MFSP) that meets the economic parameter using the general methodology that is identical to those applied in previous design reports [4, 5, 16, 19]. The base case economic parameters used in this analysis were presented in Section 1.3 and summarized in Table 1. The MFSP value represents the minimum selling price of high-octane gasoline assuming a 30-year plant life and 40% equity financing with 10% internal rate of return and the remaining 60% debt financed at 8% interest.

The total energy production is used to first determine the plant gate price (PGP) on a cost per energy basis (i.e., $/MMBtu) using all products in the calculation. The lower heating value (LHV) of the individual products was then used to calculate the volumetric cost of the fuels and the per-kilowatt-hour cost of electricity. This approach allocates a proportional fraction of the capital and operating costs for the plant to each of the main products. High-octane gasoline essentially is the only product for this study. The resulting base case MFSP for high-octane gasoline is determined to be $3.25 per gallon of gasoline blendstock (or $3.41/GGE) in 2011 U.S. dollars. Note that the baseline used in the GGE calculation is obtained from GREET, 116,090 BTU/gal (LHV) for gasoline blendstock [52]; the LHV for the high-octane gasoline produced in this work is 111,560 BTU/gal (LHV).

The value of a gasoline blendstock is a function of its properties (i.e. octane value). High-octane blendstocks are likely to carry a premium price relative to the finished fuel value (as illustrated in Appendix I). The potential premium value of the high-octane gasoline blendstock from the process described in this design report is not taken into consideration in MFSP determination.

A sensitivity analysis was performed to examine the variability of the MFSP of high-octane gasoline as a function of changes in various process parameters, raw material costs, and financial assumptions. The results of the sensitivity analysis are discussed in Section 5.

5 Process Economics and Sensitivity Analyses

The cost of high-octane gasoline as determined in the previous section was derived using technology that has been developed and demonstrated or is currently being developed as part of the DOE's biomass research programs, which includes collaboration with industrial partners. The combination of all process, market, and financial targets, as well as other assumptions presented in this design report, result in a minimum fuel selling price of $3.25 per gallon gasoline ($3.41/GGE). A summary of the costs contributing to the total high-octane selling price is presented in Figure 15 and tabulated in Appendix B.
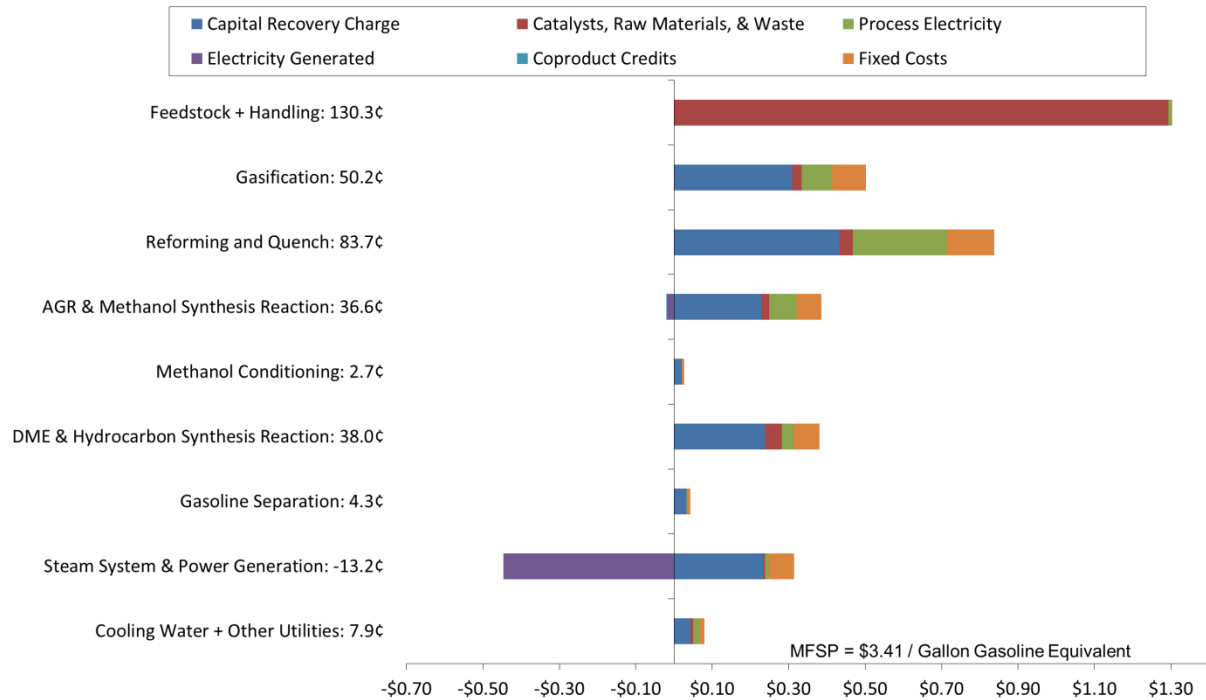
Figure 15. Cost contribution details for high-octane gasoline from each process area for the 2022 target case

This cost contribution chart shows coproduct credits from the methanol synthesis area (A400), for electricity from the syngas expansion, and for electricity from the steam system and power generation area (A600). However, the process was adjusted so that the electricity generated is balanced by the electricity required by all other areas, and thus there is no net credit for electricity generation.

The total cost of high-octane gasoline production, as indicated by the MFSP, is determined from the combination of various assumptions, both technical and financial. Financial and market assumptions include financing criteria like internal rate of return for equity investors and interest on debt. In addition, potential variability in equipment design, installation, and construction costs may impact economics. Although the assumptions and estimated plant capital costs are reasonable, it is important to consider the impact of deviations from base case (2022 target case) values. Sensitivity analysis results for project parameters identified by subject matter experts as critical factors in determining an accurate MFSP are discussed here. The results for the sensitivity analysis are summarized in Figure 16. The data are organized by potential impact of parameter deviation on MFSP value (highest to lowest). The sensitivity scenarios represent a deviation in a single project parameter with all other parameters remaining constant at base case values.

Note that each sensitivity scenario in Figure 16 has an associated deviation value from the base case. If a deviation bar is not visible, then the impact on MFSP is negligible. All sensitivity scenarios are discussed further in the following sections. The case numbers in the text refer to the numbers shown in the labels of Figure 16.

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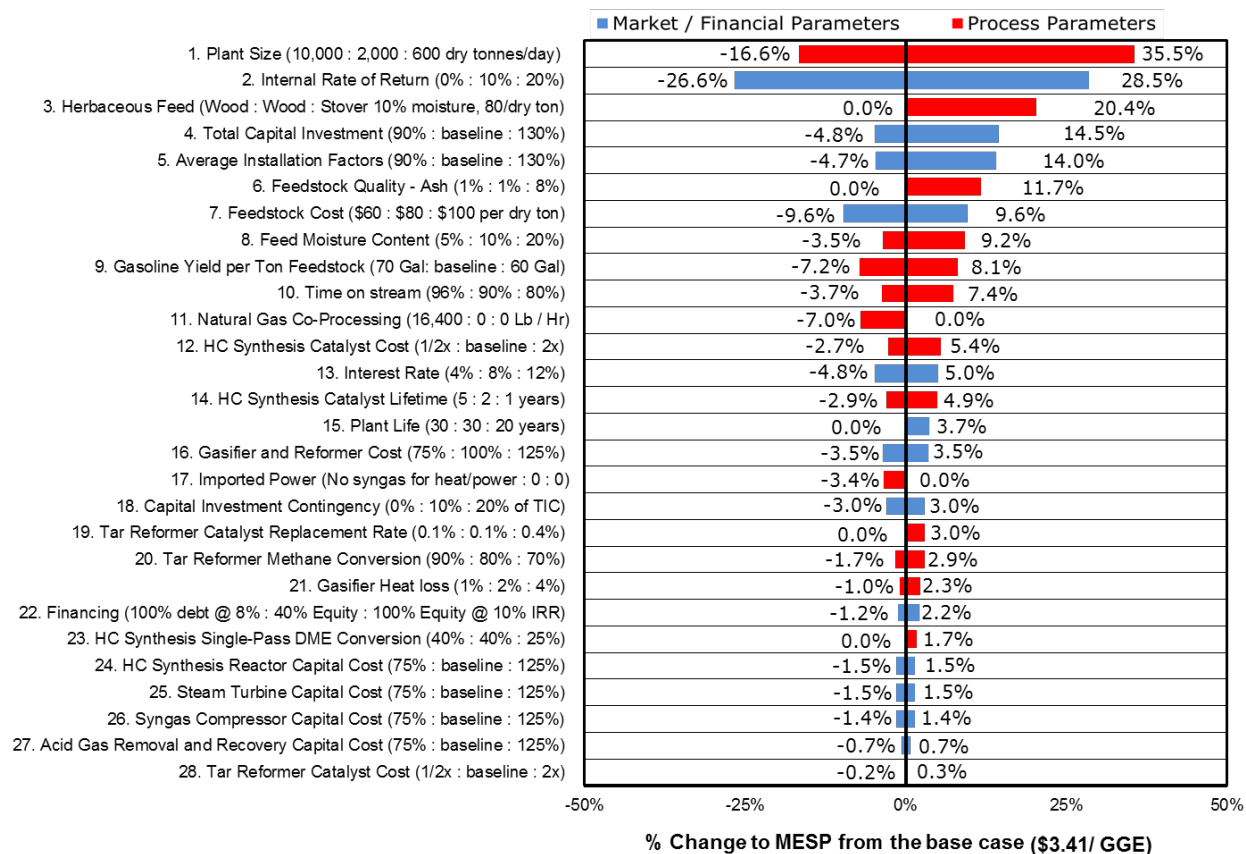
Figure 16. Results of sensitivity analyses

Deviation from base case financial and market parameters can have significant impacts on the MFSP. The base case financial assumptions are shown in Table 1. Case 1 in Figure 16 shows the possible savings realized by economies of scale from changes to the plant size. It is assumed in the plant size sensitivity scenario that the same technology is utilized for all plant sizes. A major constraint to larger plant sizes will be significantly higher feedstock logistics costs. However, the economies of scale can potentially be enabled by an advanced logistic supply system [12]. An increase in required internal rate of return (IRR) for the project from 10% to 20% (Case 2 in Figure 16) would increase the MFSP from $3.41 to $4.38 per GGE (an increase of 28.5%).

Although the plant design basis specifies blended biomass feedstock with mostly woody biomass from pulpwood and forest residues, there is potential for other feedstocks to be introduced as feed, based on future biomass availability. The high ash, low carbon, and high oxygen content in corn stover results in a 20.4% increase in the MFSP (Case 3). The negative impacts of higher ash and high moisture contents are quantified in Cases 6 and 8, respectively. Product yields decrease for higher moisture feedstocks because more energy is used for drying, resulting in higher MFSP; an increase in moisture content from 10% to 20% results in a 9.2% increase in the MFSP. Similarly, an increase in the ash content from 1% to 8% raises the MFSP by 11.7%. The impacts of changes in the feedstock cost are shown in Case 7.

57

The impact of biomass particle size on the MFSP is not considered here. It has been reported that char formation increases with increasing particle size but decreases with increasing temperature [53]. However, at gasification temperature above 1,472°F (800°C), particle size exhibited little impact on char formation. As the gasifier of the current design operates at 1,617°F (880.5°C), the amount of char produced is expected to be similar with different particle size. Furthermore, with little variation in char production, the heat generated from char combustion remains unaffected. Consequently, the particle size is expected to have little impact on the process economics.

The yield of the high-octane gasoline blendstock exhibits a direct impact on MFSP, as shown in Case 9. An increase in the yield from the baseline (65 gal/dry U.S. ton) to 70 gallons leads to a 7.2% decrease in MFSP. Likewise, decreasing the yield from 65 gallons to 60 gallons per dry U.S. ton results in an increase of 8.1% in the MFSP.

The impact of changes to the capital costs of the gasifier and reformer is shown in Case 16. Variability in TCI can also have a major impact on MFSP. Conceptual designs like this generally result in accuracy of ±10% to 30% for capital investment estimation [8]. Applying a variability range of −10% to +30% to a TCI sensitivity (Case 4) results in an MFSP range of $3.24 to $3.90 per GGE (−4.8% to +14.5%).

An interest rate increase (Case 13) from the base value of 8% to 12% increases MFSP to $3.58 per GGE (an increase of about 5.0%). Successful R&D and demonstration projects would reduce investment and lending risks and ease the expected rate of return to investors and minimize loan interest rates, reducing MFSP toward the base case value.

Extending the hydrocarbon synthesis catalyst lifetime (Case 14) from the base case of two years and reducing the tar reformer catalyst cost from the base case value (Case 28) will also reduce the MFSP. Case 23 shows that decreasing the single-pass DME conversion to high-octane hydrocarbons from 40% to 25% leads to a mere 1.7 % increase in MFSP.

Increased availability of low cost natural gas facilitates the development of combined biomass and natural gas to liquids technology. Co-conversion of biomass with natural gas can simultaneously increase fuel yields and reduce fuel production costs provided that the lifecycle GHG thresholds specified in EISA [46] are not violated. Case 11 in Figure 16 shows that co-processing natural gas at the lifecycle GHG threshold limit (i.e., 60% GHG emissions reduction relative to the petroleum baseline) decreases the MFSP by 7% ($3.17/GGE).

Because the heat and power requirements of the process cannot be met through char combustion and combustion of available fuel gas and other process off-gases alone, some raw syngas from indirect gasifier is diverted from liquid fuel production for heat and power production. Although this option makes the design energy self-sufficient, it also lowers the overall product yield. Case 17 is a sensitivity case without utilizing raw syngas for plant heat and power; electricity import is required. Importing electricity in lieu of combusting syngas improves both fuel yield and carbon-to-fuel efficiency by 7%, leading to lower MFSP, $3.29/GGE, or 3.4% lower than the base case. The corresponding GHG emissions reduction relative to petroleum gasoline benchmark is 82%.

58

6 Conclusions

This report summarizes the results of a conceptual process design with detailed mass and energy balances in Aspen Plus, economic analysis, and environmental sustainability analysis for the production of high-octane gasoline from biomass via a methanol intermediate. The analysis shows that biomass-derived high-octane gasoline from a thermochemical conversion process has the potential to be cost-competitive with gasoline. The gasoline equivalent price by lower heating value is $3.41 per gallon with underlying assumptions and catalyst performance targets outlined in this report. This price does not include retail taxes, tax credits, or costs for distribution, blending, and marketing. This analysis was made based on the nth-plant assumption and it is likely that the first such plants will be associated with higher costs. While the 2022 target case with underlying assumptions project MFSP is below $3.50/GGE, further improvements and cost reductions to $3/GGE are possible, as shown in the sensitivity analysis. Significant impacts identified in the sensitivity cases include economies of larger scales, lower feedstock costs, and higher product yields. The sensitivity analysis also shows that there is less than ±20% prediction uncertainty in MFSP resulting from a ±30% uncertainty in the total capital cost estimation. While there is always a chance of large escalations in capital costs, acquisition of recent estimates and vendor quotes for most of the equipment reduces the probability of gross over- or under-estimation of costs. Most of the unit operations used in the plant design are commercially available. Gasification and tar reforming have been demonstrated at NREL's thermochemical pilot plant. The DME to high-octane hydrocarbons synthesis is the area that is least developed and most in need of further research. The net GHG emissions associated with the conversion stage are very low (0.6 g CO_{2-e}/MJ) because the conversion plant has no net electricity import requirement and does not require any fossil energy input. Preliminary full life-cycle assessment (from feedstock production to vehicle operation) showed that the overall life-cycle GHG emissions reduction exceeds the 60% reduction criteria relative to the 2005 petroleum gasoline baseline.

7 Future Work

The following discussion summarizes future work associated with the methanol to high-octane gasoline process pathway.

- Develop and validate DME to hydrocarbons kinetic and hydrodynamic models. In order to rigorously model a process using this technology, kinetics are required for the estimation of product composition and heat generation based on feed composition (including recycle).

- Develop catalysts that minimize the production of HMB byproducts. Catalyst development is needed to move from the current state of technology to the assumed performance used for this process design. This includes strategies for recycling aromatic products to the reactor or other parts of the process, reducing the rate of aromatic formation, or a combination thereof.

- Demonstrate current design assumptions on larger-scale systems. The DME to hydrocarbon synthesis data utilized in this design are largely based on bench-scale information [3, 34]. There is significant uncertainty in assuming that comparable yields can be achieved on a commercial scale with similar operating conditions. The design

59

basis for sizing the reactor and associated catalyst volume must be confirmed by further bench- and pilot-scale demonstrations.

- Understand DME to hydrocarbon catalyst life and stability more fully. The catalyst lifetime, maintenance, stability, and optimal regeneration protocols will have impacts on the process design and operating costs. Testing with the upgrading intermediate (methanol in this case) and a detailed characterization of catalyst performance and deactivation modes are needed. A better fundamental understanding of both chemical conversion and poisoning mechanisms may also lead to the development of better catalysts that are resistant to contaminants and poisons.

- Investigate the consolidation of reaction steps. Capital and operation costs could potentially be reduced if methanol and DME are coproduced in the same reactor. It is recommended that industry users of the combined methanol/DME reaction be consulted if possible.

- Validate the feasibility of using some recycle streams. The preliminary reactor system design includes the recycle of C_3- and C_4- olefins and iso-paraffins to the reactor inlet with conversion to larger products near 100%. This assumption must be validated using reactor data from actual or simulated recycle.

- Develop catalysts to enable hydrogen incorporation. Catalysts must be developed to incorporate H_2 into reaction products and reduce aromatic production to improve yields.

- Explore integration of process streams with on-site hydrogen production. The economic and operability aspects of an on-site hydrogen plant must also be considered. Off-gas conditioning may be needed for hydrogen plant use. There is a need to better characterize the gas composition and any required clean-up or design modifications to the hydrogen plant (i.e., the addition of pre-reformers or hydrogenation steps prior to reforming).

- Validate reactor designs and refine the cost basis. The methanol to hydrocarbons reactor system cost estimates are based on estimates from the Aspen Icarus Process Evaluator (IPE). The design basis for sizing the reactor and associated catalyst volume must be confirmed by further bench- and pilot-scale demonstrations.

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Appendix A. Individual Equipment Cost Summary

Equipment Number	Number Required	Number Spares	No. Req. Variable	Equipment Name	Scaling Stream	Scaling Stream Flow (lb/hr or btu/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2011$	Scaled Uninstalled Cost in 2011$
Area A100: Feed Handling & Preparation																	
C-101	4			Hopper Feeder	STRM.A100.101	262,454	204,131	0.78	$0	2007	$0	0.75	$0	2.47	$0	$0	$0
C-102	2			Screener Feeder Conveyor	STRM.A100.101	262,454	204,131	0.78	$0	2007	$0	0.75	$0	2.47	$0	$0	$0
C-103	2			Radial Stacker Conveyor	STRM.A100.101	262,454	204,131	0.78	$0	2007	$0	0.75	$0	2.47	$0	$0	$0
C-104	2			Dryer Feed Screw Conveyor	STRM.A100.101	262,454	204,131	0.78	$0	2007	$0	0.75	$0	2.47	$0	$0	$0
C-105	2			Gasifier Feed Screw Conveyor	STRM.A100.104	204,131	204,131	1.00	$0	2007	$0	0.75	$0	2.47	$0	$0	$0
K-101	2			Flue Gas Blower	STRM.112	977,060	833,704	0.85	$0	2007	$0	0.75	$0	2.47	$0	$0	$0
K-102	2			CO2 Booster Blower	STRM.112	262,454	833,704	3.18	$0	2007	$0	0.75	$0	2.47	$0	$0	$0
M-101	4			Hydraulic Truck Dump with Scale	STRM.A100.101	262,454	204,131	0.78	$0	2007	$0	0.75	$0	2.47	$0	$0	$0
M-102	2			Hammermill	STRM.A100.101	262,454	204,131	0.78	$0	2007	$0	0.75	$0	2.47	$0	$0	$0
M-103	3			Front End Loaders	STRM.A100.101	262,454	204,131	0.78	$0	2007	$0	0.75	$0	2.47	$0	$0	$0
S-101	2			Magnetic Head Pulley	STRM.A100.101	262,454	204,131	0.78	$0	2007	$0	0.75	$0	2.47	$0	$0	$0
S-102	2			Vibratory Conveyor	STRM.A100.101	262,454	204,131	0.78	$0	2007	$0	0.75	$0	2.47	$0	$0	$0
T-101	4			Dump Hopper	STRM.A100.101	262,454	204,131	0.78	$0	2007	$0	0.75	$0	2.47	$0	$0	$0
T-102	1			Hammermill Surge Bin	STRM.A100.101	262,454	204,131	0.78	$0	2007	$0	0.75	$0	2.47	$0	$0	$0
T-103	2			Dryer Feed Bin	STRM.A100.101	262,454	204,131	0.78	$0	2007	$0	0.75	$0	2.47	$0	$0	$0
T-104	2			Dried Biomass Hopper	STRM.A100.104	204,131	204,131	1.00	$0	2007	$0	0.75	$0	2.47	$0	$0	$0
T-105	2			Lock Hopper	STRM.A100.104	204,131	204,131	1.00	$0	2007	$0	0.75	$0	2.47	$0	$0	$0
T-106	2			Feed Hopper	STRM.A100.104	204,131	204,131	1.00	$0	2007	$0	0.75	$0	2.47	$0	$0	$0
M-105	2			Cross Flow Pellet Dryer	STRM.A100.101	204,132	204,131	1.00	$50,000	2011	$100,000	0.60	$100,000	2.00	$200,000	$200,000	$100,000
Area A100 Subtotal											$100,000		$100,000	2.00	$200,000	$200,000	$100,000

Equipment Number	Number Required	Number Spares	No. Req. Variable	Equipment Name	Scaling Stream	Scaling Stream Flow (lb/hr or btu/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2011$	Scaled Uninstalled Cost in 2011$
Area A200: Gasification																	
C-201	1			Sand / Ash Conditioner / Conveyor	STRM.A200.219	2,430	2,433	1.00	$0	2002	$0	0.65	$0	2.47	$0	$0	$0
HP-201	1			Char Combustor Flue Gas / Hydrocarbon Synthesis Reactor Feed Exchanger	Pinch	785	785	1.00	$173,400	2010	$173,400	0.65	$173,398	2.13	$368,795	$393,090	$184,821
HP-202	1			Char Combustor Flue Gas / DME Reactor Feed Exchanger	Pinch	785	785	1.00	$127,500	2010	$127,500	0.65	$127,498	2.44	$311,296	$331,803	$135,898
HP-203	1			Char Combustor Flue Gas / Methanol Synthesis Reactor Feed Exchanger	Pinch	785	785	1.00	$77,700	2010	$77,700	0.65	$77,699	3.31	$257,196	$274,140	$82,818
HP-204	1			Char Combustor Flue Gas / BFW Preheat Exchanger	Pinch	785	785	1.00	$44,300	2010	$44,300	0.65	$44,299	4.23	$187,197	$199,530	$47,218
HP-205	1			Char Combustor Flue Gas Steam Superheater	Pinch	785	785	1.00	$328,500	2010	$328,500	0.65	$328,495	2.16	$708,390	$755,057	$350,136
HP-206	1			Char Combustor Flue Gas Steam Generator	Pinch	785	785	1.00	$87,900	2010	$87,900	0.65	$87,899	3.20	$281,596	$300,147	$93,689
K-202	2			Char Combustor Air Blower	STRM.A200.A200CC.208	8,624	473,382	54.89	$0	2010	$0	0.65	$0	2.47	$0	$0	$0
M-201	2			Sand / Ash Cooler	STRM.A200.A200CC.217	2,428	2,190	0.90	$0	2010	$0	0.65	$0	2.47	$0	$0	$0
R-201	2			Indirectly-Heated Biomass Gasifier	STRM.A200.201	500	1,000	2.00	$6,466,667	2010	$12,933,333	0.60	$19,603,268	2.31	$45,238,310	$48,218,522	$20,894,693
R-202	2			Char Combustor	STRM.A200.A200CC.210	5,452,890	5,514,285	1.01	$0	2010	$0	0.65	$0	2.47	$0	$0	$0
S-201	2			Primary Gasifier Cyclone	STRM.A200.202	5,237,857	5,277,340	1.01	$0	2010	$0	0.65	$0	2.47	$0	$0	$0
S-202	2			Secondary Gasifier Cyclone	STRM.A200.222	254,544	242,338	0.95	$0	2010	$0	0.65	$0	2.47	$0	$0	$0
S-203	2			Primary Combustor Cyclone	STRM.A200.A200CC.210	5,452,889	5,514,285	1.01	$0	2010	$0	0.65	$0	2.47	$0	$0	$0
S-204	2			Char Combustor Secondary Cyclone	STRM.A200.A200CC.212	502,930	514,196	1.02	$0	2010	$0	0.65	$0	2.47	$0	$0	$0
T-201	1			Sand / Ash Bin	STRM.A200.A200CC.217	2,430	2,190	0.90	$0	2010	$0	0.65	$0	2.47	$0	$0	$0
Area A200 Subtotal											$13,772,633		$20,442,556	2.32	$47,352,780	$50,472,290	$21,789,272

65

Area A300: Syngas Cleanup and Compression

Equipment Number	Number Required	Number Spares	No. Req. Variable	Equipment Name	Scaling Stream	Scaling Stream Flow (lb/hr or btu/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2011$	Scaled Uninstalled Cost in 2011$
HP-301	1			Compressor Interstage / Methanol Synthesis Reactor Feed Exchanger	Pinch	785	785	1.00	$217,400	2010	$217,400	0.65	$217,397	2.26	$490,493	$522,806	$231,719
HP-302	1			Compressor Interstage / BFW Preheat Exchanger	Pinch	785	785	1.00	$18,500	2010	$18,500	0.65	$18,500	5.23	$96,699	$103,069	$19,718
HP-303	1			Compressor Interstage / Methanol Synthesis Reactor Feed Exchanger	Pinch	785	785	1.00	$212,300	2010	$212,300	0.65	$212,297	2.30	$488,693	$520,887	$226,283
HP-304	1			Compressor Interstage / BFW Preheat Exchanger	Pinch	785	785	1.00	$30,900	2010	$30,900	0.65	$30,900	3.65	$112,798	$120,229	$32,935
HP-305	1			Tar Reformer Effluent / Catalyst Regenerator Air Preheat Exchanger	Pinch	785	785	1.00	$78,100	2010	$78,100	0.65	$78,099	2.88	$224,697	$239,499	$83,244
HP-306	1			Tar Reformer Effluent / Hydrocarbon Synthesis Reactor Feed Exchanger	Pinch	785	785	1.00	$1,819,100	2010	$1,819,100	0.65	$1,819,075	1.76	$3,203,855	$3,414,919	$1,938,912
HP-307	1			Tar Reformer Effluent / Char Combustor Air Preheat Exchanger	Pinch	785	785	1.00	$213,200	2010	$213,200	0.65	$213,197	2.03	$433,494	$462,052	$227,242
HP-308	1			Tar Reformer Effluent Steam Generator	Pinch	785	785	1.00	$608,500	2010	$608,500	0.65	$608,491	1.70	$1,032,186	$1,100,184	$648,578
HP-309	1			Catalyst Regenerator Flue Gas / Steam Turbine Generator Inlet (Stage 2) Exchanger	Pinch	785	785	1.00	$234,500	2010	$234,500	0.65	$234,497	2.28	$534,593	$569,810	$249,945
HP-310	1			Catalyst Regenerator Flue Gas / Steam Turbine Generator Inlet (Stage 3) Exchanger	Pinch	785	785	1.00	$147,900	2010	$147,900	0.65	$147,898	2.88	$426,594	$454,697	$157,641
HP-311	1			Catalyst Regenerator Flue Gas / Purge Gas Expander Inlet Exchanger	Pinch	785	785	1.00	$101,300	2010	$101,300	0.65	$101,299	2.95	$299,296	$319,013	$107,972
HP-312	1			Catalyst Regenerator Flue Gas Steam Superheater	Pinch	785	785	1.00	$1,030,700	2010	$1,030,700	0.65	$1,030,686	1.68	$1,730,476	$1,844,476	$1,098,585
HP-313	1			Catalyst Regenerator Flue Gas / Gasifier Steam Preheat Exchanger	Pinch	785	785	1.00	$74,400	2010	$74,400	0.65	$74,399	3.34	$248,297	$264,654	$79,300
HP-314	1			Catalyst Regenerator Flue Gas Steam Superheater	Pinch	785	785	1.00	$256,100	2010	$256,100	0.65	$256,096	2.48	$634,791	$676,610	$272,968
HP-315	1			Catalyst Regenerator Flue Gas / Syngas Recycle Exchanger	Pinch	785	785	1.00	$736,400	2010	$736,400	0.65	$736,390	1.58	$1,165,784	$1,242,583	$784,902
HP-316	1			Tar Reformer Effluent / BFW Preheat Exchanger	Pinch	785	785	1.00	$45,500	2010	$45,500	0.65	$45,499	3.71	$168,898	$180,024	$48,497
HP-317	1			Steam Turbine Generator Exhaust / Catalyst Regenerator Air Preheat Exchanger	Pinch	785	785	1.00	$99,400	2010	$99,400	0.65	$99,399	2.50	$248,097	$264,441	$105,947
HP-318	1			Tar Reformer Effluent / BFW Preheat Exchanger	Pinch	785	785	1.00	$12,300	2010	$12,300	0.65	$12,300	5.38	$66,199	$70,560	$13,110
HP-319	1			Tar Reformer Effluent / Catalyst Regenerator Air Preheat Exchanger	Pinch	785	785	1.00	$59,400	2010	$59,400	0.65	$59,399	2.41	$143,210	$152,644	$63,312
H-301	1			Quench Water Recirculation Cooler	HEAT.A300.A300Q.QCM301	10,000,000	6,831,373	0.68		2007		0.65		3.96	$0	$0	$0
H-301B	1			Post-reformer Air Cooler	HEAT.A300.A300Q.QAH301B	37,400,000	78,679,450	2.10	$318,900	2010	$318,900	0.65	$517,128	1.20	$620,243	$661,103	$551,195
K-305	1			Combustion Air Blower	WORK.A300.A300TR.A300FC.WK305	1,926	991	0.51	$2,469,748	2007	$2,469,748	0.65	$1,602,993	1.13	$1,817,278	$2,025,847	$1,786,968
M-300	1			Combustor Fuel Pre-Mix	STRM.A300.A300TR.325A	363,102	285,994	0.79	$0	2010	$0	0.65	$0	2.47	$0	$0	$0
M-300H	1			Tar Reformer Inlet Mix	STRM.A300.A300TR.325A	363,102	285,994	0.79	$0	2010	$0	0.65	$0	2.47	$0	$0	$0
M-301	1			Syngas Cyclone Separator	STRM.A300.A300Q.301	489,600	285,985	0.58	$0	2010	$0	0.65	$0	2.47	$0	$0	$0
M-302	1			Syngas Venturi Scrubber	STRM.A300.A300Q.301	489,600	285,985	0.58	$0	2010	$0	0.65	$0	2.47	$0	$0	$0
R-301A	2			Tar Reformer Catalyst Regenerator	STRM.A100.101	500	204,131	408.26	$0	2010	$0	0.60	$0	2.31	$0	$0	$0
R-303	2			Tar Reformer	STRM.A100.101	500	1,000	2.00	$3,233,333	2010	$6,466,667	0.60	$9,801,634	2.31	$22,619,155	$24,109,261	$10,447,347
S-306	1			Tar Reformer 2-Stage Cyclone	STRM.A300.A300TR.325A	500	285,994	571.99	$0	2010	$0	0.60	$0	2.31	$0	$0	$0
S-307	1			Catalyst Regenerator 2-Stage Cyclone	STRM.A300.A300TR.325A	500	285,994	571.99	$0	2010	$0	0.60	$0	2.31	$0	$0	$0
P-302	1	1		Quench Water Recirculation Pump	STRM.A300.A300Q.307	504,844	339,702	0.67	$12,510	2007	$25,020	0.30	$22,216	4.12	$91,458	$101,954	$24,766
P-301	1	1		Sludge Pump	STRM.A300.A300Q.335	2,127	101	0.05	$5,253	2007	$10,506	-	$10,506	4.21	$44,200	$49,273	$11,712
P-314	1	1		Condensate Pump	STRM.A300.A300Q.305C	2,127	75,638	35.57	$5,253	2007	$10,506	-	$10,506	4.21	$44,200	$49,273	$11,712
T-301	1			Sludge Settling Tank	STRM.A300.A300Q.302	2,127	75,837	35.66	$9,605	2007	$9,605	-	$9,605	3.99	$38,300	$42,696	$10,707
T-302	1			Quench Water Recirculation Tank	STRM.A300.A300Q.307	504,844	285,985	0.57	$44,005	2007	$44,005	0.65	$30,414	2.33	$70,912	$79,050	$33,905
K-310	1			Synthesis Gas Compressor	WORK.A300.A300Q.A300CMPR.WK310	78,374	24,464	0.31	$38,716,300	2010	$38,716,300	0.80	$15,253,587	1.80	$27,456,457	$29,265,235	$16,258,464
H-310AAC	1			1st Stage Air Intercooler	HEAT.A300.A300Q.A300CMPR.QAK-310A	37,400,000	22,892,924	0.61	$318,900	2010	$318,900	0.65	$231,790	1.20	$278,008	$296,323	$247,059
H-310BAC	1			2nd Stage Air Intercooler	HEAT.A300.A300Q.A300CMPR.QAK-310B	36,000,000	4,081,052	0.11	$249,600	2010	$249,600	0.65	$60,625	1.20	$72,704	$77,494	$64,619
H-310CAC	1			3rd Stage Air Intercooler	HEAT.A300.A300Q.A300CMPR.QAK-310C	32,600,000	6,456,872	0.20	$219,300	2010	$219,300	0.65	$76,553	1.23	$93,779	$99,957	$81,596
H-310DAC	1			4th Stage Air Intercooler	HEAT.A300.A300Q.A300CMPR.QAK-310D	38,600,000	4,283,256	0.11	$231,900	2010	$231,900	0.65	$55,548	1.30	$72,486	$77,262	$59,208
H-310ACW	1			1st Stage Water Intercooler	HEAT.A300.A300Q.A300CMPR.QKCW310A	11,100,000	5,143,866	0.46	$128,506	2007	$128,506	0.65	$77,947	2.21	$172,108	$191,861	$86,893
H-310BCW	1			2nd Stage Water Intercooler	HEAT.A300.A300Q.A300CMPR.QKCW310B	6,400,000	2,748,720	0.43	$70,035	2007	$70,035	0.65	$40,433	2.78	$112,524	$125,439	$45,073
H-310CCW	1			3rd Stage Water Intercooler	HEAT.A300.A300Q.A300CMPR.QKCW310C	4,400,000	2,790,066	0.63	$49,835	2007	$49,835	0.65	$37,063	3.18	$117,971	$131,511	$41,317
H-310DCW	1			4th Stage Water Intercooler	HEAT.A300.A300Q.A300CMPR.QKCW310D	3,300,000	2,075,487	0.63	$45,735	2007	$45,735	0.65	$33,833	3.79	$128,194	$142,907	$37,716
S-315	1			Pre-Compressor Knock-out	STRM.A300.A300Q.A300CMPR.313	431,712	212,665	0.49	$72,472	2007	$72,472	0.65	$45,740	3.63	$165,990	$185,040	$50,990
S-310A	1			1st Interstage KO Drum	STRM.A300.A300Q.A300CMPR.313	431,712	212,665	0.49	$72,472	2007	$72,472	0.65	$45,740	3.63	$165,990	$185,040	$50,990
S-310B	1			2nd Interstage KO Drum	STRM.A300.A300Q.A300CMPR.313	431,712	212,665	0.49	$51,137	2007	$51,137	0.65	$32,275	3.21	$103,570	$115,457	$35,979
S-310C	1			3rd Interstage KO Drum	STRM.A300.A300Q.A300CMPR.313	431,712	212,665	0.49	$47,835	2007	$47,835	0.65	$30,191	2.87	$86,782	$96,742	$33,656
P-315	1	1		Syngas Compressor Condensate Pump	STRM.A300.A300Q.A300CMPR.315L	2,127	21,830	10.26	$5,253	2007	$10,506	-	$10,506	4.21	$44,200	$49,273	$11,712
Area A300 Subtotal										**Subtotal**	**$55,635,390**		**$34,032,648**	**1.94**	**$66,165,656**	**$70,681,155**	**$36,374,390**

Area A400: Acid Gas Removal & Methanol Synthesis

Equipment Number	Number Required	Number Spares	No. Req. Variable	Equipment Name	Scaling Stream	Scaling Stream Flow (lb/hr or btu/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2011$	Scaled Uninstalled Cost in 2011$
HP-401	1			Steam Blowdown / Purge Gas Expander Inlet Exchanger	Pinch	785	785	1.00	$23,600	2010	$23,600	0.65	$23,600	4.18	$98,699	$105,201	$25,154
HP-402	1			Steam Turbine Generator Exhaust / BFW Preheat Exchanger	Pinch	785	785	1.00	$49,900	2010	$49,900	0.65	$49,899	3.48	$173,498	$184,927	$53,187
HP-403	1			Methanol Synthesis Reactor Effluent / Catalyst Regenerator Air Preheat Exchanger	Pinch	785	785	1.00	$55,000	2010	$55,000	0.65	$54,999	3.73	$204,997	$218,502	$58,622
HP-404	1			Methanol Synthesis Reactor Effluent / BFW Preheat Exchanger	Pinch	785	785	1.00	$57,300	2010	$57,300	0.65	$57,299	3.21	$183,897	$196,012	$61,074
HP-405	1			Methanol Synthesis Reactor Effluent / BFW Preheat Exchanger	Pinch	785	785	1.00	$147,300	2010	$147,300	0.65	$147,298	2.40	$354,195	$377,529	$157,002
H-411B	1			Post-Reactor Syngas Air Cooler	HEAT.A400.QAH411B	33,800,000	22,313,708	0.66	$289,106	2007	$289,106	0.65	$220,716	3.01	$664,118	$740,338	$246,047
H-413	1			Syngas Air Cooler	HEAT.A400.QAH413	33,800,000	32,726,046	0.97	$289,106	2007	$289,106	0.65	$283,101	3.01	$851,833	$949,597	$315,593
H-414	1			Methanol Condenser	HEAT.A400.QCH414	77,554,944	10,597,738	0.14	$262,922	2007	$262,922	0.44	$109,520	1.88	$205,525	$229,113	$122,089
K-412	1			Purge Gas Expander Stage 1	WORK.A400.WK412	-10,058	-2,166	0.22	$3,300,000	2007	$3,300,000	0.80	$966,026	1.80	$1,738,846	$1,938,413	$1,076,896
K-414	1			Unreacted Syngas Recycle Compressor	WORK.A400.WK414	678	3,294	4.86	$403,122	2002	$403,122	0.80	$1,427,736	2.47	$3,526,509	$5,221,123	$2,113,815
K-410	1			Syngas 2nd compressor	WORK.A400.A400CMPR.WK410	10,617	3,514	0.33	$1,241,523	2002	$1,241,523	0.80	$512,660	2.47	$1,266,270	$1,874,758	$759,011
R-490	4			Methanol Synthesis Reactor	STRM.A400.A400SGR.499A	12.5	10.9	0.87	$872,300	2011	$3,489,200	0.60	$3,215,386	2.60	$8,360,005	$8,360,005	$3,215,386
S-414	1			Methanol Flash Drum	STRM.A400.473A	6,985	311,834	44.64	$20,604	2007	$20,604	0.60	$201,282	5.22	$1,050,178	$1,170,706	$224,384
S-471	1			Methanol Condensation Knock-out	STRM.A400.471	142,038	349,903	2.46	$70,865	2007	$70,865	0.60	$121,719	3.12	$380,108	$423,733	$135,688
M-405	1			LO-CAT Sulfur Recovery System	STRM.424	93	119	1.27	$2,917,500	2009	$2,917,500	0.75	$3,491,257	1.35	$4,717,835	$5,294,569	$3,918,048
S-410	1			Amine Acid Gas Removal (AGR) System	MOLE.A400.A400AGR.A400AGR.431	17,942	11,114	0.62	$4,449,561	2009	$4,449,561	0.56	$3,106,801	2.61	$8,108,749	$9,100,009	$3,486,593
S-405	2			ZnO Sulfur Removal Beds (Methanol)	STRM.A400.470C	179,237	112,425	0.63	$37,003	2002	$74,006	0.56	$56,994	2.47	$140,776	$208,424	$84,382
S-445	1			Pressure Swing Adsorption Unit	MOLE.A400.478B	5,204	627	0.12	$975,000	2013	$975,000	0.60	$273,766	1.90	$520,155	$537,025	$282,645
T-480	1			Methanol Surge Tank	STRM.A400.475B	500,400	92,533	0.18	$28,505	2007	$28,505	0.65	$9,516	6.83	$64,995	$72,454	$10,608
P-480	1			Methanol Surge Tank Pump	STRM.A400.475B	40,894	92,533	2.26	$7,500	1997	$7,500	0.79	$14,296	2.47	$35,312	$53,512	$21,665
Area A400 Subtotal											**Subtotal $18,151,620**		**$14,343,872**	**2.28**	**$32,646,498**	**$37,255,951**	**$16,367,891**

Area A500: Methanol Conditioning

Equipment Number	Number Required	Number Spares	No. Req. Variable	Equipment Name	Scaling Stream	Scaling Stream Flow (lb/hr or btu/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2011$	Scaled Uninstalled Cost in 2011$
D-500	1			Methanol Degassing Column	STRM.A500.505	42,702	92,528	2.17	$236,400	2007	$236,400	0.68	$399,954	2.47	$987,886	$1,101,265	$445,856
HP-501	1			Char Combustor Flue Gas / Methanol Degassing Column Reboiler Exchanger	Pinch	785	785	1.00	$28,000	2010	$28,000	0.65	$28,000	4.58	$128,098	$136,537	$29,844
HP-502	1			DME Reactor Effluent / Methanol Degassing Column Reboiler Exchanger	Pinch	785	785	1.00	$33,200	2010	$33,200	0.65	$33,200	4.05	$134,498	$143,359	$35,387
HP-503	1			Methanol Intermediate / BFW Preheat Exchanger	Pinch	785	785	1.00	$15,500	2010	$15,500	0.65	$15,500	5.22	$80,899	$86,228	$16,521
HP-504	1			Methanol Intermediate / Light Hydrocarbon Recycle Exchanger	Pinch	785	785	1.00	$20,900	2010	$20,900	0.65	$20,900	4.88	$101,899	$108,611	$22,277
HP-505	1			Methanol Intermediate / Catalyst Regenerator Air Preheat Exchanger	Pinch	785	785	1.00	$23,900	2010	$23,900	0.65	$23,900	4.30	$102,799	$109,571	$25,474
HP-506	1			Methanol Intermediate / Purge Gas Expander Inlet Exchanger	Pinch	785	785	1.00	$21,800	2010	$21,800	0.65	$21,800	4.14	$90,299	$96,247	$23,236
HP-507	1			Methanol Intermediate / AGR Acid Gas Exchanger	Pinch	785	785	1.00	$18,300	2010	$18,300	0.65	$18,300	5.27	$96,499	$102,856	$19,505
H-504B	1			Cooler	HEAT.A500.QCH504B	77,554,944	3,177,293	0.04	$338,016	2002	$338,016	0.44	$82,873	2.47	$204,697	$303,061	$122,697
H-593	1			Methanol Product Finishing Cooler	HEAT.A500.QCH593	381,671	387,274	1.01	$16,780	2007	$16,780	0.65	$16,940	4.32	$73,190	$81,590	$18,884
P-500B	1	1		D-500 Bottoms Pump	VBOT500	349,266	3,135	0.01	$8,679	2002	$17,358	0.33	$3,665	2.47	$9,052	$13,401	$5,426
T-592	1			Methanol Product Storage Tank	STRM.592	40,894	87,710	2.14	$165,800	1997	$255,986	0.51	$378,301	2.47	$790,301	$1,123,123	$519,743
Area A500 Subtotal											**Subtotal $1,026,140**		**$1,043,330**	**2.68**	**$2,800,115**	**$3,405,850**	**$1,284,849**

Area A600: Steam System & Power Generation

Equipment Number	Number Required	Number Spares	No. Req. Variable	Equipment Name	Scaling Stream	Scaling Stream Flow (lb/hr or btu/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2011$	Scaled Uninstalled Cost in 2011$
H-601	1			Steam Turbine Condenser	HEAT.A600.A600ST.QAH601	1	4	4.00	$1,300,000	2010	$1,300,000	1.00	$5,200,000	1.40	$7,280,000	$7,759,592	$5,542,566
M-601A	1			BFW Reverse Osmosis (RO) Unit	STRM.A600.A600BF.631	300,240	410,302	1.37	$1,865,000	2010	$1,865,000	0.70	$2,320,723	1.15	$2,668,831	$2,844,649	$2,473,608
M-601B	1			BFW Electrodeionization (EDI) Unit	STRM.A600.A600BF.631	300,240	410,302	1.37	$0	2010	$0	0.70	$0	1.15	$0	$0	$0
M-601C	1			Hot Condensate Polishing Unit	STRM.A600.616	400,439	177,601	0.44	$0	2010	$0	0.70	$0	1.15	$0	$0	$0
M-602A	1			Extraction Steam Turbine / Generator Stage 1	WTURB1	-40,418	-15,980	0.40	$7,700,000	2010	$7,700,000	0.70	$4,021,564	1.80	$7,238,815	$7,715,694	$4,286,497
M-602B	1			Extraction Steam Turbine / Generator Stage 2	WTURB2	-40,418	-16,437	0.41	$7,700,000	2010	$7,700,000	0.70	$4,101,666	1.80	$7,382,998	$7,869,376	$4,371,875
M-602C	1			Extraction Steam Turbine / Generator Stage 3	WTURB3	-40,418	-17,443	0.43	$7,700,000	2010	$7,700,000	0.70	$4,275,866	1.80	$7,696,558	$8,203,593	$4,557,551
M-603	1			Startup Boiler	STRM.A200.200	36,560	73,120	2.00	$198,351	2002	$198,351	0.60	$300,644	2.47	$742,590	$1,099,431	$445,114
M-604	1			Brine Recovery Reverse Osmosis (RO) Unit	STRM.H2O-MGMT.TOROBRIN	300,240	45,338	0.15	$0	2010	$0	0.70	$0	1.15	$0	$0	$0
P-601	1	1		Make-Up Pump	STRM.A600.A600WTR.618	80,411	124,430	1.55	$6,528	2007	$13,056	0.30	$14,883	4.72	$70,218	$78,277	$16,591
P-602	1	1		Condensate Pump	STRM.A600.633	247,010	410,302	1.66	$9,810	2007	$19,620	0.30	$22,846	4.61	$105,276	$117,358	$25,468
P-605	1	1		Condensate Collection Pump	STRM.A600.633	93,974	410,302	4.37	$5,437	2002	$10,874	0.33	$17,686	2.47	$43,684	$64,675	$26,184
P-603	1	1		EDI Pump	STRM.A600.633	247,010	410,302	1.66	$9,810	2007	$19,620	0.30	$22,846	4.61	$105,276	$117,358	$25,468
P-604	1	1		Boiler Feed Water Pump	STRM.A600.633	494,622	410,302	0.83	$304,578	2007	$609,156	0.30	$575,940	1.35	$779,247	$868,681	$642,041
P-661	1	1		Intercooler Charge Pump	STRM.A600.A600BL.661A1	93,974	58,184	0.62	$10,874	2002	$10,874	0.33	$9,283	2.47	$22,929	$33,947	$13,744
S-601	1			Blowdown Flash Drum No. 1	STRM.A600.604	9,892	4,088	0.41	$47,205	2007	$47,205	0.65	$26,579	3.42	$90,768	$101,185	$29,630
S-661	1			Blowdown Flash Drum No. 2	STRM.A600.A600BL.662A	9,892	1,164	0.12	$47,205	2007	$47,205	0.65	$11,745	3.42	$40,108	$44,712	$13,093
T-601	1			Condensate Collection Tank	STRM.A600.A600BF.627	500,400	410,302	0.82	$28,505	2007	$28,505	0.65	$25,054	6.83	$171,121	$190,760	$27,930
T-608	1			Condensate Collection Tank	STRM.A600.A600BL.661A1	500,400	58,184	0.12	$28,505	2007	$28,505	0.65	$7,039	6.83	$48,074	$53,591	$7,846
T-602	1			Condensate Surge Tank	STRM.A600.A600BF.627	500,400	410,302	0.82	$27,704	2007	$27,704	0.65	$24,350	6.51	$158,569	$176,768	$27,145
T-603	1			Deaerator	STRM.A600.633	494,619	410,302	0.83	$53,299	2007	$53,299	0.65	$47,202	5.07	$239,220	$266,675	$52,619
T-603A	1			Deaerator Packed Column	STRM.A600.633	494,619	410,302	0.83	$18,405	2007	$18,405	0.65	$16,300	5.18	$84,399	$94,086	$18,170
T-606	1			Steam Turbine Condenstate Tank	STRM.A600.A600ST.615	500,400	177,601	0.35	$28,505	2007	$28,505	0.65	$14,538	6.83	$99,295	$110,691	$16,207
T-605	1			Steam Drum No. 1	STRM.A600.644	494,622	408,807	0.83	$104,100	2007	$104,100	0.65	$91,973	2.28	$209,514	$233,560	$102,529
T-607	1			Steam Drum No. 2	STRM.A600.A600BL.661A	494,622	58,184	0.12	$104,100	2007	$104,100	0.65	$25,900	2.28	$59,000	$65,771	$28,872
Area A600 Subtotal										**Subtotal**	**$27,634,084**		**$21,174,626**	**1.67**	**$35,336,489**	**$38,110,430**	**$22,750,748**

Area A700: Cooling Water & Other Utilities

Equipment Number	Number Required	Number Spares	No. Req. Variable	Equipment Name	Scaling Stream	Scaling Stream Flow (lb/hr or btu/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2011$	Scaled Uninstalled Cost in 2011$
K-701	2	1		Plant Air Compressor	STRM.A100.101	262,454	204,131	0.78	$87,922	2007	$263,766	0.30	$244,611	1.57	$383,656	$427,688	$272,685
M-701	1			Cooling Tower System	STRM.A700.715	7,506,000	2,556,084	0.34	$260,852	2010	$260,852	0.78	$112,586	2.47	$278,087	$296,407	$120,003
M-702	1			Hydraulic Truck Dump with Scale	STRM.A100.101	367,437	204,131	0.56	$80,000	1998	$80,000	0.60	$56,225	2.47	$138,875	$208,829	$84,546
M-703	1			Flue Gas Scrubber	STRM.112	489,600	833,704	1.70	$436,250	2010	$436,250	0.65	$616,590	2.47	$1,522,976	$1,623,307	$657,209
M-704	1			Flue Gas Stack	STRM.112	939,119	833,704	0.89	$169,187	2007	$169,187	0.65	$156,587	1.30	$203,431	$226,779	$174,559
M-705	1			Clarifier		220	220		$96,221	2007	$96,221	0.65	$96,221	2.41	$232,300	$258,961	$107,264
M-706	1			Belt Press		1	1		$135,000	2010	$135,000	0.65	$135,000	2.47	$333,450	$355,417	$143,894
M-707	1			Sand Filter		265	265		$120,400	2010	$120,400	0.65	$120,400	2.47	$297,388	$316,979	$128,332
M-708	1			Emergency Flare	STRM.112	262,454	833,704	3.18	$0	2010	$0	-	$0	2.47	$0	$0	$0
M-709	1			Cooling Tower Reverse Osmosis (RO) Unit	STRM.A100.101	300,240	204,131	0.68	$0	2010	$0	0.70	$0	1.15	$0	$0	$0
M-1590	1			Product Loading Rack	STRM.A100.101	262,454	204,131	0.78	$0	2010	$0	-	$0	2.47	$0	$0	$0
P-701	1	1		Cooling Water Pump	STRM.A700.715	7,001,377	2,556,084	0.37	$239,375	2007	$478,750	0.30	$353,856	2.14	$757,307	$844,223	$394,468
P-702	1	1		Firewater Pump	STRM.A100.101	262,454	204,131	0.78	$23,043	2007	$46,086	0.30	$42,739	3.70	$158,025	$176,162	$47,644
P-703	1	1		Diesel Pump	STRM.A100.101	262,454	204,131	0.78	$3,842	2007	$7,684	0.30	$7,126	5.39	$38,393	$42,800	$7,944
P-704	1	1		Ammonia Pump	STRM.A100.101	262,454	204,131	0.78	$3,842	2007	$7,684	0.30	$7,126	5.21	$37,095	$41,353	$7,944
P-705	1	1		Caustic Pump		1	1		$4,906	2007	$9,812	0.30	$9,812	4.30	$42,200	$47,043	$10,938
P-707	1	1		BFW Chemical Pump	STRM.A100.101	262,454	204,131	0.78	$3,842	2007	$7,684	0.30	$7,126	5.21	$37,095	$41,353	$7,944
P-708	1	1		Flue Gas Scrubber Circulation Pump	STRM.112	489,600	833,704	1.70	$12,510	2007	$25,020	0.30	$29,352	4.12	$120,834	$134,702	$32,721
P-709	1	1		Slurry Pump		1	1		$3,242	2007	$6,484	0.30	$6,484	6.63	$43,000	$47,935	$7,228
P-710	1			Clarifier Overflow Pump	STRM.A700.715	7,001,377	2,556,084	0.37	$239,375	2007	$239,375	0.30	$176,928	2.14	$378,653	$422,111	$197,234
S-701	1	1		Instrument Air Dryer	STRM.A100.101	262,454	204,131	0.78	$8,349	2002	$16,698	0.60	$14,361	2.47	$35,471	$52,516	$21,262
T-701	1			Plant Air Receiver	STRM.A100.101	262,454	204,131	0.78	$21,005	2007	$21,005	0.65	$17,839	5.44	$97,074	$108,215	$19,887
T-702	1			Firewater Storage Tank	STRM.A100.101	262,454	204,131	0.78	$229,900	2007	$229,900	0.65	$195,252	1.46	$285,447	$318,207	$217,661
T-703	1			Purchased Diesel Storage Tank	STRM.A100.101	262,454	204,131	0.78	$104,674	2007	$104,674	0.65	$88,899	1.35	$119,665	$133,399	$99,102
T-704	1			Ammonia Storage Tank	STRM.A100.101	262,454	204,131	0.78	$15,704	2007	$15,704	0.65	$13,337	5.39	$71,850	$80,096	$14,868
T-705	1			Caustic Storage Tank		1	1		$16,005	2007	$16,005	0.65	$16,005	3.01	$48,200	$53,732	$17,842
T-706	1			Olivine / MgO Loading System	STRM.A100.101	262,454	204,131	0.78	$0	2010	$0	-	$0	2.00	$0	$0	$0
T-707	1			Tar Reformer Catalyst Loading System	STRM.A100.101	262,454	204,131	0.78	$235,000	2010	$235,000	-	$235,000	2.00	$470,000	$500,963	$250,481
T-708	1			BFW Chemical Storage Tank	STRM.A100.101	262,454	204,131	0.78	$22,004	2007	$22,004	0.65	$18,688	6.70	$125,186	$139,553	$20,833
T-710	1			Clarifier Overflow Tank		1	1	1.00	$14,205	2007	$14,205	0.65	$14,205	2.41	$34,300	$38,237	$15,835
T-709	1			Slurry Tank		1	1		$14,205	2007	$14,205	0.65	$14,205	2.41	$34,300	$38,237	$15,835
Area A700 Subtotal										**Subtotal**	**$3,079,655**		**$2,806,559**	**2.25**	**$6,324,259**	**$6,975,203**	**$3,096,161**

68

Area A1400: Methanol to High Octane Gasoline Synthesis

Equipment Number	Number Required	Number Spares	No. Req. Variable	Equipment Name	Scaling Stream	Scaling Stream Flow (lb/hr or btu/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2011$	Scaled Uninstalled Cost in 2011$
HP-1401	1			Unconverted DME Recycle / DME Reactor Feed Exchanger	Pinch	785	785	1.00	$559,900	2010	$559,900	0.65	$559,892	1.87	$1,046,885	$1,115,852	$596,777
HP-1402	1			Hydrocarbon Synthesis Reactor Effluent / Catalyst Regenerator Air Preheat Exchanger	Pinch	785	785	1.00	$146,900	2010	$146,900	0.65	$146,898	2.32	$340,495	$362,926	$156,575
HP-1403	1			Hydrocarbon Synthesis Reactor Effluent / DME Reactor Feed Exchanger	Pinch	785	785	1.00	$1,204,370	2010	$1,204,370	0.65	$1,204,353	1.84	$2,218,669	$2,364,831	$1,283,694
HP-1404	1			Steam Blowdown / Unconverted DME Purge	Pinch	785	785	1.00	$15,800	2010	$15,800	0.65	$15,800	5.66	$89,399	$95,288	$16,841
HP-1405	1			Unconverted DME Recycle / Deethanizer Feed Preheat Exchanger	Pinch	785	785	1.00	$15,500	2010	$15,500	0.65	$15,500	5.22	$80,899	$86,228	$16,521
HP-1406	1			Unconverted DME Recycle / DME Reactor Feed Exchanger	Pinch	785	785	1.00	$75,200	2010	$75,200	0.65	$75,199	2.74	$206,197	$219,781	$80,153
HP-1407	1			Unconverted DME Recycle / Light Hydrocarbon Recycle Exchanger	Pinch	785	785	1.00	$14,800	2010	$14,800	0.65	$14,800	5.47	$80,899	$86,228	$15,775
P-1000	1	1		Methanol Intermediate Pump	STRM.A1400.1000	40,894	87,710	2.14	$7,500	1997	$15,000	0.79	$27,409	2.47	$67,700	$102,592	$41,535
R-1424	1			DME Synthesis Reactor	STRM.A1400.A1400MTD.1411	153,521	195,754	1.28	$665,078	2014	$665,078	0.57	$763,893	1.87	$1,428,479	$1,455,314	$778,243
H-1424	1			DME Reactor Intercooler	HEAT.A1400.A1400MTD.QR1424	127,700,000	4,378,721	0.03	$2,353,181	2007	$2,353,181	0.65	$262,728	2.78	$731,173	$815,090	$292,882
H-1471	1			Post DME Reactor Air Cooler	HEAT.A1400.A1400MTG.QAH1471	33,800,000	18,415,691	0.54	$289,106	2007	$289,106	0.65	$194,820	3.01	$586,200	$653,478	$217,179
S-1470	1			DME Reactor Effluent KO Drum	STRM.A1400.A1400MTG.1470A	431,712	87,710	0.20	$108,700	2007	$108,700	0.65	$38,578	2.00	$77,333	$86,208	$43,005
P-1470	1			Condensate Pump	STRM.A1400.A1400MTG.1400WW3	93,974	18,580	0.20	$5,437	2002	$5,437	0.33	$3,185	2.47	$7,866	$11,646	$4,715
R-1410C	12			Hydrocarbon Synthesis Reactor	STRM.A1400.A1400MTG.1485	977,200	695,146	0.71	$139,920	2011	$1,679,040	0.60	$1,368,727	2.20	$3,011,199	$3,011,199	$1,368,727
H-1410E	3			Hydrocarbon Synthesis Reactor 1st Stage Intercooler	HEAT.A1400.A1400MTG.QH1410E	127,700,000	9,089,353	0.07	$2,353,181	2007	$7,059,543	0.65	$1,267,065	2.21	$2,797,680	$3,118,769	$1,412,486
H-1410F	3			Hydrocarbon Synthesis Reactor 2nd Stage Intercooler	HEAT.A1400.A1400MTG.QH1410F	127,700,000	7,779,572	0.06	$2,353,181	2007	$7,059,543	0.65	$1,145,180	2.78	$3,187,036	$3,552,811	$1,276,612
H-1410J	3			Hydrocarbon Synthesis Reactor 3rd Stage Intercooler	HEAT.A1400.A1400MTG.QH1410J	127,700,000	6,827,149	0.05	$2,353,181	2007	$7,059,543	0.65	$1,051,981	3.18	$3,348,457	$3,732,758	$1,172,717
H-1410C	1			Hydrocarbon Synthesis Reactor Effluent Cooler	HEAT.A1400.A1400MTG.QH1410C	127,700,000	10,400,093	0.08	$2,353,181	2007	$2,353,181	0.65	$461,005	3.18	$1,467,378	$1,635,789	$513,914
H-1413	1			Hydrocarbon Synthesis Reactor Effluent Air Cooler	HEAT.A1400.A1400CT.QAH1413	33,800,000	36,880,908	1.09	$289,106	2007	$289,106	0.65	$305,972	3.01	$920,650	$1,026,313	$341,089
H-1414	1			Hydrocarbon Synthesis Reactor Effluent Water Cooler	HEAT.A1400.A1400CT.QCH1414	15,600,000	8,259,600	0.53	$161,435	2007	$161,435	0.65	$106,780	3.24	$345,538	$385,195	$119,035
K-1472	1			Hydrocarbon Synthesis Reactor Recycle Compressor	WORK.A1400.WK1472	233	2,699	11.58	$993,485	2007	$993,485	0.80	$7,050,343	1.80	$12,690,618	$14,147,116	$7,859,509
K-1410	1			Coke Burn / Cat Regeneration Air Blower	WORK.A1400.A1400MTG.WK-CBURN	1,926	20	0.01	$2,469,748	2007	$2,469,748	0.65	$128,888	1.13	$146,117	$162,887	$143,680
S-1472	1			Reactor Effluent KO Drum 1st Stage	STRM.A1400.A1400CT.1472	431,712	372,153	0.86	$108,700	2007	$108,700	0.65	$98,701	2.00	$197,857	$220,564	$110,029
S-1473	1			Reactor Effluent KO Drum 2nd Stage	STRM.A1400.A1400CT.1472	431,712	372,153	0.86	$108,700	2007	$108,700	0.65	$98,701	2.00	$197,857	$220,564	$110,029
T-1473	1			Condensate Surge Tank	STRM.A1400.A1400CT.1476WW	500,400	30,110	0.06	$28,505	2007	$28,505	0.65	$4,587	6.83	$31,329	$34,924	$5,113
P-1473	1			Condensate Surge Tank Pump	STRM.A1400.A1400CT.1476WW	93,974	30,110	0.32	$5,437	2002	$5,437	0.33	$3,735	2.47	$9,224	$13,657	$5,529
Area A1400 Subtotal										**Subtotal**	**$34,844,938**		**$16,414,719**	**2.15**	**$35,313,132**	**$38,718,009**	**$17,982,363**

Area A1500: Product Recovery

Equipment Number	Number Required	Number Spares	No. Req. Variable	Equipment Name	Scaling Stream	Scaling Stream Flow (lb/hr or btu/hr)	New Stream Flow	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2011$	Scaled Uninstalled Cost in 2011$
HP-1501	1			High-Octane Gasoline Blendstock Product / DME Reactor Feed Exchanger	Pinch	785	785	1.00	$27,800	2010	$27,800	0.65	$27,800	3.94	$109,598	$116,819	$29,631
HP-1502	1			High-Octane Gasoline Blendstock Product / Deethanizer Feed Preheat Exchanger	Pinch	785	785	1.00	$15,300	2010	$15,300	0.65	$15,300	5.27	$80,699	$86,015	$16,308
HP-1503	1			Tar Reformer Effluent / Debutanizer Reboiler Exchanger	Pinch	785	785	1.00	$18,500	2010	$18,500	0.65	$18,500	5.23	$96,699	$103,069	$19,718
HP-1504	1			Char Combustor Flue Gas / Deethanizer Reboiler Exchanger	Pinch	785	785	1.00	$132,500	2010	$132,500	0.65	$132,498	2.36	$312,796	$333,402	$141,227
D-1503	1			Deethanizer Column	STRM.A1500.A1504.1507	42,702	46,621	1.09	$236,400	2007	$236,400	0.68	$250,946	2.47	$619,836	$690,975	$279,747
D-1504	1			Debutanizer Column	STRM.A1500.A1504.1509B	43,158	41,942	0.97	$236,400	2007	$236,400	0.68	$231,851	2.47	$572,672	$638,398	$258,461
H-1511	1			High Octane Gasoline Product Water Cooler	HEAT.A1500.A1504.QCH1511	381,671	572,026	1.50	$16,780	2007	$16,780	0.65	$21,828	4.32	$94,311	$105,135	$24,333
P-1503	1			Crude Hydrocarbons Pump	STRM.A1500.A1504.1503	40,894	46,621	1.14	$7,500	1997	$7,500	0.79	$8,318	2.47	$20,546	$31,136	$12,606
P-1503B	1			Dethanizer Reboiler Pump	STRM.A1500.A1504.1509B	40,894	41,942	1.03	$7,500	1997	$7,500	0.79	$7,652	2.47	$18,899	$28,640	$11,595
P-1504B	1			Debutanizer Reboiler Pump	STRM.A1500.A1504.D1504CAL.STRMADJ	40,894	30,452	0.74	$7,500	1997	$7,500	0.79	$5,942	2.47	$14,676	$22,240	$9,004
P-1503C	1			Deethanizer Reflux Pump	STRM.A1500.A1504.1508	40,894	3,512	0.09	$7,500	1997	$7,500	0.79	$1,079	2.47	$2,664	$4,037	$1,635
P-1504C	1			Debutanizer Reflux Pump	STRM.A1500.A1504.1510	40,894	7,040	0.17	$7,500	1997	$7,500	0.79	$1,868	2.47	$4,615	$6,993	$2,831
H-1503C	1			Deethanizer Column Condenser	HEAT.A1500.A1504.QCH-DC2	77,554,944	10,607,261	0.14	$338,016	2002	$338,016	0.44	$140,856	2.47	$347,914	$515,099	$208,542
H-1504C	1			Debutanizer Column Condenser	HEAT.A1500.A1504.QCH-DC4	77,554,944	3,306,673	0.04	$338,016	2002	$338,016	0.44	$84,341	2.47	$208,323	$308,430	$124,871
H-1504D	1			Debutanizer Overhead Product Condenser	HEAT.A1500.A1504.D1504OCA.QCH1504A	77,554,944	946,102	0.01	$338,016	2002	$338,016	0.44	$48,632	2.47	$120,121	$177,843	$72,001
P-1555	1			Debutanizer Overhead Product Pump	STRM.A1500.A1504.1510	40,894	7,040	0.17	$7,500	1997	$7,500	0.79	$1,868	2.47	$4,615	$6,993	$2,831
P-1592	1			Gasoline Product Pump	STRM.A1500.1592	40,894	34,902	0.85	$7,500	1997	$7,500	0.79	$6,618	2.47	$16,346	$24,770	$10,028
T-1592	1	1		High Octane Gasoline Product Storage Tank	STRM.A1500.1592	51,427	34,902	0.68	$690,900	2007	$1,381,800	0.65	$1,074,045	1.85	$1,985,172	$2,213,010	$1,197,313
Area A1500 Subtotal										**Subtotal**	**$3,132,028**		**$2,079,941**	**2.23**	**$4,630,502**	**$5,413,003**	**$2,422,681**

Appendix B. Discounted Cash Flow Rate of Return (DCFROR) and Operating Cost Summary

Process Engineering Analysis for High Octane Gasoline via Indirect Gasification and Methanol Intermediate (2022 Target Case)

2,000 Dry Metric Tonnes Biomass per Day
Indirect Gasifier, Tar Reformer, Sulfur Removal, Methanol Synthesis, Hydrocarbon Synthesis on Beta-Zeolite Catalyst, Fuel Purification, Steam-Power Cycle
All Values in 2011 US$

Gasoline Minimum Fuel Selling Price (MFSP)	**$3.25 per Gallon**
MFSP (Gasoline-Equivalent Basis)	**$3.41 per GGE**
Contributions: Feedstock Costs	1.232 per Gallon
Operating Costs & Credits	0.545 per Gallon
Capital Charges & Taxes	1.471 per Gallon

Gasoline Production at Operating Capacity	47.0 MM Gallons per Year
Gasoline Product Yield	64.9 Gallons per Dry US Ton Feedstock

Delivered Feedstock Cost $80.00 per Dry US Ton

Capital Costs

Feed Handling & Drying	$200,000
Gasification	$50,500,000
Gas Cleanup	$70,700,000
Methanol Synthesis	$37,300,000
Methanol Conditioning	$3,400,000
DME & Hydrocarbons Conversion	$38,700,000
Gasoline Separations	$5,400,000
Steam System & Power Generation	$38,100,000
Cooling Water & Other Utilities	$7,000,000
Total Installed Equipment Cost (TIC)	$251,300,000
ISBL (Areas A100 to A500, A1400, A1500)	$206,100,000
OSBL (Areas A600, A700)	$45,100,000
Other Direct Costs	8,200,000
(% of ISBL)	4.0%
Total Direct Costs (TDC)	259,500,000
Indirect Costs	155,700,000
(% of TDC)	60.0%
Land Purchase Cost	1,600,000
Working Capital	20,800,000
Total Capital Investment (TCI)	437,500,000
Installed Equipment Cost per Annual Gallon	$4.81
Total Capital Investment per Annual Gallon	$8.37
Debt Financing (% of Investment)	60.0%
Loan Interest Rate	8.0%
Loan Term (years)	10.0
Equity Financing (% of Investment)	40.0%
Internal Rate of Return (After-Tax)	10.0%
Plant Operating Hours per year	7,884
On-Stream Percentage	90.0%

Process Efficiency

Gasifier Efficiency - HHV %	72.9
Gasifier Efficiency - LHV %	72.5
Efficiency to Gasoline - HHV %	45.3
Efficiency to Gasoline - LHV %	45.0
Overall Plant Efficiency - HHV %	45.3
Overall Plant Efficiency - LHV %	45.0

Annual Operating Costs

Feedstock	$57,900,000
Natural Gas	$0
Catalysts	$10,700,000
Olivine	$500,000
Other Raw Matl. Costs	$1,400,000
Waste Disposal	$600,000
Electricity Transfer Charge	$0
Electricity	$0
Fixed Costs	$20,600,000
Coproduct credits	$0
Capital Depreciation	$13,800,000
Average Income Tax	$8,100,000
Average Return on Investment	$47,200,000

Operating Costs per Product	(¢/MMBtu)	(¢/Gal)
Feedstock	1112.6	123.2
Natural Gas	0.0	0.0
Catalysts	48.5	5.4
Olivine	10.3	1.1
Other Raw Materials	26.4	2.9
Waste Disposal	10.8	1.2
Electricity Transfer	0.0	0.0
Electricity	0.0	0.0
Fixed Costs	396.5	43.9
Coproduct credits	0.0	0.0
Capital Depreciation	265.0	29.4
Average Income Tax	156.5	17.3
Average Return on Investment	906.8	100.4
Total (Plant Gate Price)	2933.4	324.9

Power Balance	(KW)	(hp)
Total Plant Power Consumption	36,875	49,451
Power Generated Onsite	36,879	49,456
Power Imported from Grid	0	0
Power Exported to Grid	4	5

Power Generation	(KW)	(hp)
Steam Turbine Generators	35,344	47,397
Process Gas Turboexpander(s)	1,535	2,059

Sustainability Metrics

Plant Electricity Consumption (KWh/ Gal Gasoline)	6.2
Gasification & Reforming Steam (lb / Gal Gasoline)	21.8
Water Consumption (Gal Water / Gal Gasoline)	1.7
Carbon Conversion Efficiency (C in Fuel/C in Feedstock)	31.2%
Fossil GHG Emissions (g CO_{2-e}/MJ Fuel)	0.6
Fossil Enegy Consumption (MJ Fossil Energy/MJ Fuel)	0.006

Feedstock Rate and Cost

Feed Rate	Dry Tonnes / Day	2,000
	Dry US Tons / Day	2,205
Feedstock Cost	$ / Dry Ton	$80.00
	$ / Moisture & Ash Free Ton	$80.74

70

Discounted Cash Flow Rate of Return Worksheet

Year		-2	-1	0	1	2	3	4	5	6	7	8
Fixed Capital Investment		$14,895,260	$99,639,452	$53,141,041								
Working Capital				$20,758,219								
Loan Payment					$37,123,042	$37,123,042	$37,123,042	$37,123,042	$37,123,042	$37,123,042	$37,123,042	$37,123,042
Loan Interest Payment		$1,594,231	$13,550,966	$19,927,890	$19,927,890	$18,552,278	$17,066,617	$15,462,103	$13,729,228	$11,857,723	$9,836,498	$7,653,574
Loan Principal		$19,927,890	$169,387,069	$249,098,631	$231,903,480	$213,332,717	$193,276,292	$171,615,354	$148,221,541	$122,956,222	$95,669,679	$66,200,211
Fuel Sales					$114,563,921	$152,751,895	$152,751,895	$152,751,895	$152,751,895	$152,751,895	$152,751,895	$152,751,895
By-Product Credit					$14,023	$18,697	$18,697	$18,697	$18,697	$18,697	$18,697	$18,697
Total Annual Sales					$114,577,944	$152,770,592	$152,770,592	$152,770,592	$152,770,592	$152,770,592	$152,770,592	$152,770,592
Annual Manufacturing Cost												
Raw Materials					$50,695,237	$57,937,413	$57,937,413	$57,937,413	$57,937,413	$57,937,413	$57,937,413	$57,937,413
Zinc Oxide Catalyst					$362,410	$0	$0	$362,410	$0	$0	$362,410	$0
Tar Reforming Catalyst					$4,488,798	$0	$0	$0	$0	$0	$0	$0
Methanol Synthesis Catalyst					$1,491,991	$0	$0	$1,491,991	$0	$0	$1,491,991	$0
DME Conversion Catalyst					$1,927,022	$0	$0	$1,927,022	$0	$0	$1,927,022	$0
Beta-Zeolite Catalyst					$15,584,732	$0	$15,584,732	$0	$15,584,732	$0	$15,584,732	$0
Other Variable Costs					$3,502,322	$3,976,808	$3,976,808	$3,976,808	$3,976,808	$3,976,808	$3,976,808	$3,976,808
Fixed Operating Costs					$20,646,060	$20,646,060	$20,646,060	$20,646,060	$20,646,060	$20,646,060	$20,646,060	$20,646,060
Total Product Cost					$98,698,572	$82,560,282	$98,145,013	$86,341,704	$98,145,013	$82,560,282	$101,926,436	$82,560,282
Depreciation					$59,326,991	$101,673,758	$72,612,251	$51,854,032	$37,074,180	$37,032,663	$37,074,180	$18,516,332
Net Revenue					($63,375,509)	($50,015,726)	($35,053,290)	($887,247)	$3,822,171	$21,319,924	$3,933,479	$44,040,404
Losses Forward						($63,375,509)	($113,391,235)	($148,444,525)	($149,331,772)	($145,509,601)	($124,189,678)	($120,256,199)
Taxable Income					($63,375,509)	($113,391,235)	($148,444,525)	($149,331,772)	($145,509,601)	($124,189,678)	($120,256,199)	($76,215,795)
Income Tax					$0	$0	$0	$0	$0	$0	$0	$0
Annual Cash Income					($21,243,669)	$33,087,268	$17,502,537	$29,305,846	$17,502,537	$33,087,268	$13,721,115	$33,087,268
Discount Factor		1.210	1.100	1.000	0.909	0.826	0.751	0.683	0.621	0.564	0.513	0.467
Annual Present Value	$237,007,005				($19,312,427)	$27,344,850	$13,149,915	$20,016,287	$10,867,698	$18,676,900	$7,041,101	$15,435,455
Total Capital Investment + Interest		$19,952,285	$124,509,460	$93,827,151								
Net Present Worth				$0								

71

Discounted Cash Flow Rate of Return Worksheet

Year	9	10	11	12	13	14	15	16	17	18	19	20
Fixed Capital Investment												
Working Capital												
Loan Payment	$37,123,042	$37,123,042	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$5,296,017	$2,749,855	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$34,373,187	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$152,751,895	$152,751,895	$152,751,895	$152,751,895	$152,751,895	$152,751,895	$152,751,895	$152,751,895	$152,751,895	$152,751,895	$152,751,895	$152,751,895
By-Product Credit	$18,697	$18,697	$18,697	$18,697	$18,697	$18,697	$18,697	$18,697	$18,697	$18,697	$18,697	$18,697
Total Annual Sales	$152,770,592	$152,770,592	$152,770,592	$152,770,592	$152,770,592	$152,770,592	$152,770,592	$152,770,592	$152,770,592	$152,770,592	$152,770,592	$152,770,592
Annual Manufacturing Cost												
Raw Materials	$57,937,413	$57,937,413	$57,937,413	$57,937,413	$57,937,413	$57,937,413	$57,937,413	$57,937,413	$57,937,413	$57,937,413	$57,937,413	$57,937,413
Zinc Oxide Catalyst	$0	$362,410	$0	$0	$362,410	$0	$0	$362,410	$0	$0	$362,410	$0
Tar Reforming Catalyst	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Methanol Synthesis Catalyst	$0	$1,491,991	$0	$0	$1,491,991	$0	$0	$1,491,991	$0	$0	$1,491,991	$0
DME Conversion Catalyst	$0	$1,927,022	$0	$0	$1,927,022	$0	$0	$1,927,022	$0	$0	$1,927,022	$0
Beta-Zeolite Catalyst	$15,584,732	$0	$15,584,732	$0	$15,584,732	$0	$15,584,732	$0	$15,584,732	$0	$15,584,732	$0
Other Variable Costs	$3,976,808	$3,976,808	$3,976,808	$3,976,808	$3,976,808	$3,976,808	$3,976,808	$3,976,808	$3,976,808	$3,976,808	$3,976,808	$3,976,808
Fixed Operating Costs	$20,646,060	$20,646,060	$20,646,060	$20,646,060	$20,646,060	$20,646,060	$20,646,060	$20,646,060	$20,646,060	$20,646,060	$20,646,060	$20,646,060
Total Product Cost	$98,145,013	$86,341,704	$98,145,013	$82,560,282	$101,926,436	$82,560,282	$98,145,013	$86,341,704	$98,145,013	$82,560,282	$101,926,436	$82,560,282
Depreciation	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue	$49,329,562	$63,679,033	$54,625,579	$70,210,310	$50,844,156	$70,210,310	$54,625,579	$66,428,888	$54,625,579	$70,210,310	$50,844,156	$70,210,310
Losses Forward	($76,215,795)	($26,886,233)	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	($26,886,233)	$36,792,800	$54,625,579	$70,210,310	$50,844,156	$70,210,310	$54,625,579	$66,428,888	$54,625,579	$70,210,310	$50,844,156	$70,210,310
Income Tax	$0	$12,877,480	$19,118,952	$24,573,609	$17,795,455	$24,573,609	$19,118,952	$23,250,111	$19,118,952	$24,573,609	$17,795,455	$24,573,609
Annual Cash Income	$17,502,537	$16,428,366	$35,506,626	$45,636,702	$33,048,701	$45,636,702	$35,506,626	$43,178,777	$35,506,626	$45,636,702	$33,048,701	$45,636,702
Discount Factor	0.424	0.386	0.350	0.319	0.290	0.263	0.239	0.218	0.198	0.180	0.164	0.149
Annual Present Value	$7,422,784	$6,333,846	$12,444,856	$14,541,260	$9,573,032	$12,017,570	$8,500,004	$9,396,960	$7,024,797	$8,208,162	$5,403,727	$6,783,605
Total Capital Investment + Interest												
Net Present Worth												

72

Discounted Cash Flow Rate of Return Worksheet

Year	21	22	23	24	25	26	27	28	29	30
Fixed Capital Investment										
Working Capital										($22,368,219)
Loan Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$152,751,895	$152,751,895	$152,751,895	$152,751,895	$152,751,895	$152,751,895	$152,751,895	$152,751,895	$152,751,895	$152,751,895
By-Product Credit	$18,697	$18,697	$18,697	$18,697	$18,697	$18,697	$18,697	$18,697	$18,697	$18,697
Total Annual Sales	$152,770,592	$152,770,592	$152,770,592	$152,770,592	$152,770,592	$152,770,592	$152,770,592	$152,770,592	$152,770,592	$152,770,592
Annual Manufacturing Cost										
Raw Materials	$57,937,413	$57,937,413	$57,937,413	$57,937,413	$57,937,413	$57,937,413	$57,937,413	$57,937,413	$57,937,413	$57,937,413
Zinc Oxide Catalyst	$0	$362,410	$0	$0	$362,410	$0	$0	$362,410	$0	$0
Tar Reforming Catalyst	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Methanol Synthesis Catalyst	$0	$1,491,991	$0	$0	$1,491,991	$0	$0	$1,491,991	$0	$0
DME Conversion Catalyst	$0	$1,927,022	$0	$0	$1,927,022	$0	$0	$1,927,022	$0	$0
Beta-Zeolite Catalyst	$15,584,732	$0	$15,584,732	$0	$15,584,732	$0	$15,584,732	$0	$15,584,732	$0
Other Variable Costs	$3,976,808	$3,976,808	$3,976,808	$3,976,808	$3,976,808	$3,976,808	$3,976,808	$3,976,808	$3,976,808	$3,976,808
Fixed Operating Costs	$20,646,060	$20,646,060	$20,646,060	$20,646,060	$20,646,060	$20,646,060	$20,646,060	$20,646,060	$20,646,060	$20,646,060
Total Product Cost	$98,145,013	$86,341,704	$98,145,013	$82,560,282	$101,926,436	$82,560,282	$98,145,013	$86,341,704	$98,145,013	$82,560,282
Depreciation	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue	$54,625,579	$66,428,888	$54,625,579	$70,210,310	$50,844,156	$70,210,310	$54,625,579	$66,428,888	$54,625,579	$70,210,310
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$54,625,579	$66,428,888	$54,625,579	$70,210,310	$50,844,156	$70,210,310	$54,625,579	$66,428,888	$54,625,579	$70,210,310
Income Tax	$19,118,952	$23,250,111	$19,118,952	$24,573,609	$17,795,455	$24,573,609	$19,118,952	$23,250,111	$19,118,952	$24,573,609
Annual Cash Income	$35,506,626	$43,178,777	$35,506,626	$45,636,702	$33,048,701	$45,636,702	$35,506,626	$43,178,777	$35,506,626	$45,636,702
Discount Factor	0.135	0.123	0.112	0.102	0.092	0.084	0.076	0.069	0.063	0.057
Annual Present Value	$4,798,031	$5,304,339	$3,965,315	$4,633,293	$3,050,263	$3,829,168	$2,708,363	$2,994,161	$2,238,317	$2,615,373
Total Capital Investment + Interest										($1,281,890)
Net Present Worth										

73

Base Case Operating Cost Summary

Costing Code	Raw Material	Stream No.	kg/hr (or kW)	lb/hr (or HP)	Quoted Price (cents/ton, or cents/kWh)	Year of Price Quote	2000 Cost (cents / ton)	2000 Cost ($/lb)	$/hour	MM$/yr (2011)	Cents/MMBtu (2011)
Variable Operating Costs											
F-4	Woody Biomass	STRM.100	92,592.59	204,131.48	7,200.00	2011	4,525.63	0.02	4,619.12	57.94	1,112.61
R-12	Magnesium Oxide (MgO)	STRM.220	3.16	6.97	36,500.00	2004	33,099.25	0.17	1.15	0.01	0.28
R-13	Fresh Olivine	STRM.221	246.34	543.08	17,290.00	2004	15,679.07	0.08	42.57	0.53	10.26
S-33B	Natural Gas for Gasifier	STRM.206	0.00	0.00	23,885.27	2011	15,013.33	0.08	0.00	0.00	0.00
R-14	Tar Reformer Catalyst	STRM.A300.A300TR.326	4.00	8.82	3,528,849.94	2007	2,719,974.35	13.60	120.01	1.51	28.91
S-28	LO-CAT chemicals	STRM.424	53.79	118.58	160,150.00	2009	111,684.49	0.56	66.22	0.83	15.95
S-33A	Natural Gas for Reformer	STRM.A300.A300TR.341	0.05	0.10	23,885.27	2011	15,013.33	0.08	0.01	0.00	0.00
S-44	Dimethyl Disulfide (DMDS)		0.94	2.08	593,480.93	2009	413,878.34	2.07	4.31	0.05	1.04
S-45	Amine (MDEA) Make-Up	MOLE.A400.A400AGR.A400AGR.431	1.75	3.86	193,477.29	2009	134,926.08	0.67	2.61	0.03	0.63
S-25	Boiler Feed Water Makeup	STRM.618	37,076.05	81,738.59	19.96	2001	19.74	0.00	8.07	0.10	1.94
S-21	Boiler Chemicals	STRM.635	1.30	2.86	280,000.00	1991	349,331.21	1.75	5.00	0.06	1.20
S-22	Cooling Tower Chems	STRM.711	0.23	0.51	200,000.00	1999	209,352.04	1.05	0.54	0.01	0.13
S-24	Cooling Tower Makeup	STRM.710	898.88	1,981.69	19.96	2001	19.74	0.00	0.20	0.00	0.05
S-27	No. 2 Diesel Fuel		31.39	69.20	80,588.65	2007	62,116.29	0.31	21.49	0.27	5.18
S-151	Alkylation Unit Sulfuric Acid		-	-	9,000.00	2009	6,276.37	0.03	-	-	-
S-152	Alkylation Unit Caustic Soda		-	-	15,000.00	2009	10,460.61	0.05	-	-	-
	Inter-plant electricity	WORK.WMTG	0.07	0.10	5.01	2007			0.00	0.00	0.00
	Subtotal								4,891.29	61.35	1,178.17
	Waste Streams										
W-1	Sand and Ash Purge	STRM.219	1,103.78	2,433.42	3,265.90	1998	3,441.60	0.02	41.87	0.53	10.09
W-2	Tar Reformer Catalyst Disposal	STRM.A300.A300TR.329	4.00	8.82	3,265.90	1998	3,441.60	0.02	0.15	0.00	0.04
W-1A	Scrubber Solids	STRM.H2O-MGMT.SCRB-SLD	4.30	9.47	3,265.90	1998	3,441.60	0.02	0.16	0.00	0.04
W-3	WWT Cost	STRM.H2O-MGMT.BRINEWWT	5,140.82	11,333.55	48.08	2001	47.57	0.00	2.70	0.03	0.65
	Subtotal								44.88	0.56	10.81
	By-Product Credits										
	Sulfur	STRM.424	53.79	118.58	4,000.00	2011	2,514.24	0.01	1.49	0.02	0.36
	Subtotal								1.49	0.02	0.36
	Total Variable Operating Costs								4,934.68	61.896	1,188.62

Fixed Operating Costs								
		Salary	Salary quote year	2011 Salary	# Positions	2011 Cost		2011 Hourly Wage
	Plant Manager	$147,000	2007	$161,362	1	$161,362		$77.58
	Plant Engineer	$70,000	2007	$76,839	1	$76,839		$36.94
	Maintenance Supr	$57,000	2007	$62,569	1	$62,569		$30.08
	Lab Manager	$56,000	2007	$61,471	1	$61,471		$29.55
	Shift Supervisor	$48,000	2007	$52,690	5	$263,448		$25.33
	Lab Technician	$40,000	2007	$43,908	2	$87,816		$21.11
	Maintenance Tech	$40,000	2007	$43,908	16	$702,527		$21.11
	Shift Operators	$40,000	2007	$43,908	20	$878,159		$21.11
	Yard Employees	$28,000	2007	$30,736	12	$368,827		$14.78
	Clerks & Secretaries	$36,000	2007	$39,517	3	$118,551		$19.00

Total Salaries					62.00	2,781,567			2.782	53.42
							Avg Salary (w/ Benefits)			
	Overhead and Benefits	% of Labor & Supervison = 90.0%		$2,503,411			$40.98	per Hour	2.503	48.07
	Maintenance	% of FCI = 3.0%		$12,454,932			$85,242	per Year	12.455	239.18
	Insurance & Taxes	% of FCI = 0.7%		$2,906,151					2.906	55.81
Total Fixed Operating Costs									20.646	396.48

								MM$/yr (2011)	Cents/MMBtu (2011)
Total Operating Costs								82.542	1,585.10

Appendix C. Process Parameters and Operation Summary

Base Case Process Parameters and Operation Summary

Energy Efficiencies

Gasifier Efficiency - HHV %	72.9
Gasifier Efficiency - LHV %	72.5
Efficiency to Gasoline - HHV %	45.3
Efficiency to Gasoline - LHV %	45.0
Overall Plant Efficiency - HHV %	45.3
Overall Plant Efficiency - LHV %	45.0

Dryer

Inlet:	
Temperature (°F)	60.0
Moisture Content (wt%)	10.0%
Outlet:	
Temperature (°F)	219.7
Moisture Content (wt%)	10.0%
Inlet Flue Gas (°F)	409
Outlet Flue Gas (°F)	331.7
Dew Point Flue Gas (°F)	137.9
Difference	193.8

Gasifier

Temperature (°F)	1,598
Pressure (psia)	33.0
H_2:CO Molar Ratio After Gasifier	0.57
Methane (vol%)	8.5%
Benzene (vol%)	0.07%
Tar (wt%)	0.92%
Tar (g/Nm³)	9.8
Char (wt%)	14.6%
H_2S (ppm)	407
Residual Heat (MBtu/hr)	14,667
Raw Dry Syngas Yield (lb/lb dry feed)	0.78

Char Combustor

Temperature (°F)	1,798
Pressure (psia)	28.9
Ratio Actual:Minimum air for combustion	1.20
Residual Heat (MMBtu/hr)	14.7

Syngas Usage

To Reformer (lb/hr)	221,320
To Char Combustor (lb/hr)	0
To Fuel System (lb/hr)	15,668

Fuel System

Additional fuel (lb/hr)	0
Raw Syngas (lb/hr)	15,668
Unconverted Syngas (lb/hr)	11,468
Into Reformer (°F)	3,575
Out of Reformer (°F)	2,452

Tar Reformer

Inlet Volumetric Flow (MMscf/hr)	5.55
Space Velocity (hr⁻¹)	2,476
Reactor Volume (ft³)	2,243
Catalyst Replacement (%inventory/day)	0.15%
Inlet:	
Temperature (°F)	1,598
Carbon as CO (mol%)	50.6%
Carbon as tar (ppmv)	35,071
H_2:CO Ratio (mole)	0.63
Steam-to-Carbon Ratio	4.42
Reformer Conversions:	
CO_2 --> CO	-137.4%
Methane --> CO	80.0%
Ethane --> CO	99.0%
Ethylene --> CO	90.0%
Benzene --> CO	99.0%
Tar --> CO	99.9%
Ammonia --> CO	90.0%
Outlet:	
Temperature (°F)	1,670
Carbon as CO (mol%)	59.2%
Carbon as tar (ppmv)	113
H_2:CO Ratio (mole)	2.11
Methane (vol%)	1.2%
Benzene (ppmv)	4.4
Tars (ppmv)	5
Tars (g/Nm³)	0.02
H_2S (ppm)	256
NH_3 (ppm)	114

Quench

Benzene (ppmv)	3.0
Tars (ppmv)	3
Tars (g/Nm³)	0.01
H_2S (ppm)	369
NH_3 (ppm)	142

Acid Gas Removal

Inlet:	
CO_2 (mol/hr)	1,771
CO_2 (mol%)	16%
H_2S (mol/hr)	4
H_2S (ppmv)	369
Outlet:	
CO_2 (mol/hr)	18
CO_2 (mol%)	0.2%
Fraction CO_2 removed	99%
H_2S (mol/hr)	0
H_2S (ppmv)	43
Fraction H_2S removed	90%

Methanol Synthesis

Syngas from Conditioning (lb/hr)	105,402
Recycled from initial flash tank (lb/hr)	244,501
Total	349,903
At Reactor Inlet	
Conditioned Syngas H_2:CO Ratio	2.11
Recycled Gas H_2:CO Ratio	3.36
At Reactor Inlet	
Temperature (°F)	437
Pressure (psia)	730
H_2:CO Molar Ratio	2.75
CO_2 (mol %)	5.0%
Methane (mol%)	12.2%
H_2O (wt%)	0.1%
Inlet Volumetric Flow (MMscf/hr)	10.9
Space Velocity (hr⁻¹)	8,000
Reactor/Catalyst Volume (ft³)	1,367
CO Conversion - Overall	94.4%
CO Conversion - Single-pass	45.8%
Conversion To:	
CO_2	-0.5%
Methane	0.0%
Ethane	0.0%
Methanol	46.3%
C3	0.0%
C4	0.0%
C5+	0.0%
Total	45.8%
Selectivity (CO_2 Free)	
Methanol	100.0%
Hydrocarbons	0.0%
At Reactor Outlet	
Temperature (°F)	482
Pressure (psia)	580
CO_2 (mol%)	6.1%
Methane (mol%)	0.0%
H_2O (wt%)	0.30%
Methanol Productivity (kg/kg-cat/hr)	3.84
Methanol Productivity (kg/L-cat/hr)	5.55

Electricity

Electricity Produced (kWh)	290,754,674
Electricity Produced (kWh/gal gasoline)	6.18
Electricity Used (kWh)	290,725,326
Electricity Used (kWh/gal gasoline)	6.18

DME Synthesis

Methanol to DME Reactor (ln/hr)	87,710
At Reactor Inlet	
Temperature (°F)	437
Pressure (psia)	145
Methanol (mol%)	98.0%
H_2O (wt%)	1.2%
Inlet Volumetric Flow (MMscf/hr)	1.1
Space Velocity (hr⁻¹)	601
Reactor Volume (ft³)	1,755
Methanol Conversion - Overall	88.5%
Methanol Conversion - Single-pass	88.5%
Conversion To:	
DME	44.2%
H_2O	44.2%
Total	88.5%
At Reactor Outlet	
Temperature (°F)	482
Pressure (psia)	140
Yield	
DME (wt%)	71%
Water (wt%)	29%
Total	100%
DME Productivity (kg/kg-cat/hr)	0.35

Water Usage

Boiler Feed Water Makeup (MM lb/yr)	644
Cooling Water Makeup (MM lb/yr)	16
Total Makeup Water (MM gal/yr)	79
Water Usage (gal/gal gasoline)	1.69

Crude Hydrocarbon Product

Light gas	0%
Propane	2%
Propylene	0%
Isobutane	16%
n-Butane	0%
Butenes	0%
C5+ gasoline	81%
Total	100%

Finished Fuel Products

Gasoline (wt%)	95%
LPG (wt%)	0%
Fuel gas (wt%)	5%
Total	100%

Hydrocarbon Synthesis

DME Feed (lb/hr)	69,130
C4 Recycle (lb/hr)	6,688
Reactor Recycle (lb/hr)	292,468
H_2 Feed (lb/hr)	698
Total	368,984
At Reactor Inlet	
Temperature (°F)	350
Pressure (psia)	129
Inlet gas composition:	
DME (mol%)	32.9%
Methanol (mol%)	3.3%
H_2O (mol%)	3.8%
H_2 (mol%)	3.7%
Methane (mol%)	23.3%
Isobutane (mol%)	19.0%
Other hydrocarbons (mol%)	14.1%
Total	100.0%
Inlet Volumetric Flow (MMscf/hr)	3.7
Space Velocity (hr⁻¹)	622
Reactor Volume (ft³)	5,893
DME Conversion - Overall	92.5%
DME Conversion - Single-pass	39.6%
DME to hydrocarbons (C selectivity):	
C1	1.0%
C2	0.5%
C3	1.2%
C4	3.7%
C5	5.0%
C6	6.0%
C7	39.8%
C8	41.3%
C9	1.0%
HMB	0.5%
Total	100.0%
At Reactor Outlet	
Temperature (°F)	438
Pressure (psia)	95
Yield	
Hydrocarbons (wt%)	88%
Water (wt%)	12%
Total	100%
C_{5+} Selectivity	93.1%
Total Gasoline (C_{5+}) Productivity (kg/kg-cat/hr)	0.11
Total HC (C_1-C_6) Productivity (kg/kg-cat/hr)	0.11

Appendix D. Process Flow Diagrams (PFDs) and Material and Energy Balances

M-101
HYDRAULIC TRUCK
DUMP WITH SCALE

C-101
HOPPER FEEDER

T-101
DUMP HOPPER

C-102
SCREENER FEEDER
CONVEYOR

S-101
MAGNETIC
HEAD PULLEY

S-102
VIBRATORY
CONVEYOR

C-103
RADIAL STACKER
CONVEYOR

M-103
FRONT END
LOADERS

T-103
DRYER FEED
BIN

BIOMASS

M-101

T-101

C-101

C-102

S-101

S-102

C-103

CHIP PILE

M-103

T-103

101

BIOMASS
TO C-104
PFD-61713-A102

METAL DEBRIS TO
DISPOSAL

NOTE: FEEDSTOCK HANDLING COSTS ARE INCLUDED IN THE TARGET FEEDSTOCK COST PROVIDED BY THE IDAHO NATIONAL
LABORATORY. ONLY A CROSS FLOW DRYER USING PROCESS FLUE GASES FOR FEED PREHEAT IS INCLUDED IN THE CONVERSION COST.

Harris Group

NREL
NATIONAL RENEWABLE ENERGY LABORATORY

THERMOCHEMICAL CONVERSION OF BIOMASS
TO FUELS AND CHEMICALS
GASIFICATION

PROCESS FLOW DIAGRAM
AREA 100: FEED HANDLING

Rev.	Date	By	Description		
0	9/26/14	JAR	ISSUED FOR DESIGN REPORT		

Project No: 30482.00

Drawing: PFD-61713-A101

Rev.: 0

PFD-61713-A101

COMPONENT	UNITS	101	
Total Flow	lb/hr	204,131	
Temperature	F	60	
Pressure	Psia	14.7	
Vapor Fraction		-	
Hydrogen (H2)	lb/hr	-	
Carbon Monoxide (CO)	lb/hr	-	
Nitrogen (N2)	lb/hr	-	
Oxygen (O2)	lb/hr	-	
Argon (Ar)	lb/hr	-	
Carbon Dioxide (CO2)	lb/hr	-	
Water (H2O)	lb/hr	20,413	
Hydrogen Sulfide (H2S)	lb/hr	-	
Sulfur dioxide (SO2)	lb/hr	-	
Ammonia (NH3)	lb/hr	-	
Nitrogen dioxide (NO2)	lb/hr	-	
Methane (CH4)	lb/hr	-	
Ethane (C2H6)	lb/hr	-	
Ethylene (C2H4)	lb/hr	-	
Acetylene (C2H2)	lb/hr	-	
Propane (C3H8)	lb/hr	-	
Propene (C3H6)	lb/hr	-	
I-Butane (C4H10)	lb/hr	-	
N-Butane (C4H10)	lb/hr	-	
Butenes (C4H8)	lb/hr	-	
Pentanes (C5H12)	lb/hr	-	
Pentenes (C5H10)	lb/hr	-	
Hexanes (C6H14)	lb/hr	-	
Hexenes (C6H12)	lb/hr	-	
Benzene (C6H6)	lb/hr	-	
Heptanes (C7H16)	lb/hr	-	
Triptane	lb/hr	-	
Heptenes (C7H14)	lb/hr	-	
Triptene	lb/hr	-	
Octanes (C8H18)	lb/hr	-	
Octenes (C8H16)	lb/hr	-	
N-Nonanes (C9H20)	lb/hr	-	
N-Nonenes (C9H18)	lb/hr	-	
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	
Tar (C10H8)	lb/hr	-	
Carbon (Solid)	lb/hr	-	
Sulfur (Solid)	lb/hr	-	
Olivine + MgO (Solid)	lb/hr	-	
Methanol (CH4O)	lb/hr	-	
Dimethyl ether (C2H6O)	lb/hr	-	
MDEA (C5H13NO2)	lb/hr	-	
Ash	lb/hr	0	
Char	lb/hr	0	
Wood	lb/hr	183,718	
Enthalpy Flow	MMBTU/hr	(570)	

78

C-104
DRYER FEED SCREW
CONVEYOR

K-102
CO2 BOOSTER
BLOWER

M-105
CROSS FLOW
PELLET DRYER

T-104
DRIED BIOMASS
HOPPER

T-105
LOCK
HOPPER

T-106
FEED
HOPPER

M-102
HAMMERMILL

T-102
HAMMERMILL
SURGE BIN

K-101
FLUE GAS
BLOWER

C-105
GASIFIER FEED
SCREW CONVEYOR

BIOMASS
FROM T-103
PFD-61713-A101

FLUE GAS
FROM HP-311/312/313/314/315
PFD-61713-A305

CAT REGEN FLUE GAS
FROM R-1410C
PFD-61713-A1402

FLUE GAS
FROM HP-201
PFD-61713-A202

CO2
FROM M-405
PFD-61713-A404

CO2
TO T-706/T-707
PFD-61713-A702

FLUE GAS
TO M-703
PFD-61713-A701

MILLED BIOMASS
TO R-201
PFD-61713-A201

NOTE: FEEDSTOCK HANDLING COSTS ARE INCLUDED IN THE TARGET FEEDSTOCK COST PROVIDED BY THE IDAHO NATIONAL LABORATORY. ONLY A CROSS FLOW DRYER USING PROCESS FLUE GASES FOR FEED PREHEAT IS INCLUDED IN THE CONVERSION COST.

Plot date: 1/29/2015   J:\1A - Process Solutions\30482 NREL Thermochemical Conv - Design-Model\Drawings\ 30482 Gasification PFDs 61713 150128.vsd

PFD-61713-A102

COMPONENT	UNITS	101	102	103	104	105	107	108	110	333	Flue-Gas	287	112
Total Flow	lb/hr	204,131	204,131	204,131	204,131	204,131	803,707	803,707	803,707	288,332	3,369	512,006	833,704
Temperature	F	60	60	220	220	220	409	332	332	702	230	230	164
Pressure	Psia	14.7	15.7	15.7	15.7	35.0	17.4	16.4	16.4	17.4	19.7	26.8	15.7
Vapor Fraction		-	-	-	-	-	1	1	1	1	1	1	1
Hydrogen (H2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Carbon Monoxide (CO)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Nitrogen (N2)	lb/hr	-	-	-	-	-	539,563	539,563	539,563	186,648	2,364	350,552	539,563
Oxygen (O2)	lb/hr	-	-	-	-	-	26,152	26,152	26,152	8,104	147	17,901	26,152
Argon (Ar)	lb/hr	-	-	-	-	-	9,198	9,198	9,198	3,179	40	5,979	9,198
Carbon Dioxide (CO2)	lb/hr	-	-	-	-	-	158,793	158,793	158,793	52,037	579	106,177	158,793
Water (H2O)	lb/hr	20,413	20,413	20,413	20,413	20,413	69,821	69,821	69,821	38,211	240	31,371	99,821
Hydrogen Sulfide (H2S)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-	47	47	47	20	-	27	47
Ammonia (NH3)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-	129	129	129	129	-	0	129
Methane (CH4)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Ethane (C2H6)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Ethylene (C2H4)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Acetylene (C2H2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Propane (C3H8)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Propene (C3H6)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
I-Butane (C4H10)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
N-Butane (C4H10)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Butenes (C4H8)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Pentanes (C5H12)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Pentenes (C5H10)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Heptanes (C7H16)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Triptane	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Triptene	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Octanes (C8H18)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Octenes (C8H16)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
N-Nonanes (C9H20)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Tar (C10H8)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Carbon (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	-	-	-	3	3	3	3	-	-	-
Methanol (CH4O)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Dimethyl ether (C2H6O)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-
Ash	lb/hr	0	0	0	0	0	-	-	-	-	-	-	-
Char	lb/hr	0	0	0	0	0	0	0	0	0	-	-	-
Wood	lb/hr	183,718	183,718	183,718	183,718	183,718	-	-	-	-	-	-	-
Enthalpy Flow	MMBTU/hr	(570)	(570)	(553)	(553)	(553)	(944)	(961)	(961)	(371)	(3)	(570)	(1,168)

80

Process Flow Diagram — Area 200: Indirect Gasifier

Equipment labels:

- **K-202** CHAR COMBUSTOR AIR BLOWER
- **R-201** INDIRECTLY-HEATED BIOMASS GASIFIER
- **S-201** PRIMARY GASIFIER CYCLONE
- **S-202** SECONDARY GASIFIER CYCLONE
- **R-202** CHAR COMBUSTOR
- **S-203** PRIMARY COMBUSTOR CYCLONE
- **S-204** CHAR COMBUSTOR SECONDARY CYCLONE
- **M-201** SAND/ASH COOLER
- **T-201** SAND/ASH BIN
- **C-201** SAND/ASH CONDITIONER CONVEYOR

Stream / destination labels:

- 225A — SYNGAS TO M-300H PFD-61713-A301
- 225F — SYNGAS TO M-300 PFD-61713-A301
- 286 — FLUE GAS TO HP-205/HP-206 PFD-61713-A202
- 105 — MILLED BIOMASS FROM C-105 PFD-61713-A102
- 200B — SUPERHEATED STEAM FROM HP-313 PFD-61713-A305
- 221 — OLIVINE FROM T-706 PFD-61713-A702
- 220 — MGO FROM T-706 PFD-61713-A702
- 218 — BRINE FROM M-604 PFD-61713-A601
- 219 — SAND, ASH, CATALYST FINES TO DISPOSAL
- 208 — COMBUSTION AIR
- 209A — COMBUSTION AIR TO HP-307 PFD-61713-A304
- 209B — COMBUSTION AIR FROM HP-307 PFD-61713-A304
- 202, 222, 223, 203, 210, 212, 211, 217, WK202, QCM201

Title block:

THERMOCHEMICAL CONVERSION OF BIOMASS TO FUELS AND CHEMICALS GASIFICATION

PROCESS FLOW DIAGRAM
AREA 200: INDIRECT GASIFIER

Project No: 30482.00 Drawing: PFD-61713-A201 Rev: 1

Rev.	Date	By	Description
1	12/5/14	DMS	ADDED HX NETWORK
0	9/26/14	JAR	ISSUED FOR DESIGN REPORT

Harris Group

NREL — NATIONAL RENEWABLE ENERGY LABORATORY

Plot date: 1/29/2015   J:\1\A - Process Solutions\30482 NREL Thermochemical Conv - Design-Model\Drawings\ 30482 Gasification PFDs 61713 150128.vsd

PFD-61713-A201

COMPONENT	UNITS	105	200B	221	220	208	209A	209B	202	203	210	211	212	217	218	219	222	225A	225F	286
Total Flow	lb/hr	204,131	73,120	543	7	473,382	473,382	473,382	5,277,340	5,035,002	5,514,285	5,000,089	514,196	2,190	243	2,433	242,338	221,320	15,668	512,006
Temperature	F	220	500	60	60	90	258	400	1,598	-	1,798	-	1,798	-	60	206	1,598	1,598	1,598	1,798
Pressure	Psia	35.0	35.0	31.0	31.0	14.7	29.0	28.9	33.0	32	29	35	26.9	14.7	14.7	14.7	31.8	31.4	31.4	26.9
Vapor Fraction		-	1	1	1	1	1	1	1	-	1	-	1	-	-	-	1	1	1	1
Hydrogen (H2)	lb/hr	-	-	-	-	-	-	-	3,097	-	-	-	-	-	-	-	3,097	2,892	205	-
Carbon Monoxide (CO)	lb/hr	-	-	-	-	-	-	-	76,043	-	-	-	-	-	-	-	76,043	71,015	5,027	-
Nitrogen (N2)	lb/hr	-	-	-	-	350,531	350,531	350,531	-	-	350,552	-	350,552	-	-	-	-	-	-	350,552
Oxygen (O2)	lb/hr	-	-	-	-	107,403	107,403	107,403	-	-	17,901	-	17,901	-	-	-	-	-	-	17,901
Argon (Ar)	lb/hr	-	-	-	-	5,979	5,979	5,979	-	-	5,979	-	5,979	-	-	-	-	-	-	5,979
Carbon Dioxide (CO2)	lb/hr	-	-	-	-	233	233	233	36,232	-	106,177	-	106,177	-	-	-	36,232	33,837	2,395	106,177
Water (H2O)	lb/hr	20,413	73,120	-	-	9,236	9,236	9,236	93,533	-	31,371	-	31,371	-	243	243	93,533	87,349	6,184	31,371
Hydrogen Sulfide (H2S)	lb/hr	-	-	0	0	-	-	-	161	-	-	-	-	-	-	-	161	151	11	-
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-	-	-	-	-	27	-	27	-	-	-	-	-	-	27
Ammonia (NH3)	lb/hr	-	-	-	-	-	-	-	355	-	-	-	-	-	-	-	355	331	23	-
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-	-	-	-	-	0	-	0	-	-	-	-	-	-	0
Methane (CH4)	lb/hr	-	-	-	-	-	-	-	15,859	-	-	-	-	-	-	-	15,859	14,811	1,048	-
Ethane (C2H6)	lb/hr	-	-	-	-	-	-	-	564	-	-	-	-	-	-	-	564	527	37	-
Ethylene (C2H4)	lb/hr	-	-	-	-	-	-	-	7,850	-	-	-	-	-	-	-	7,850	7,331	519	-
Acetylene (C2H2)	lb/hr	-	-	-	-	-	-	-	684	-	-	-	-	-	-	-	684	638	45	-
Propane (C3H8)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Propene (C3H6)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
I-Butane (C4H10)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
N-Butane (C4H10)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Butenes (C4H8)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Pentanes (C5H12)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Pentenes (C5H10)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	-	-	-	-	-	-	-	639	-	-	-	-	-	-	-	639	597	42	-
Heptanes (C7H16)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Triptane	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Triptene	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Octanes (C8H18)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Octenes (C8H16)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
N-Nonanes (C9H20)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Tar (C10H8)	lb/hr	-	-	-	-	-	-	-	1,918	-	-	-	-	-	-	-	1,918	1,791	127	-
Carbon (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	543	7	-	-	-	5,000,000	4,995,000	5,000,500	5,000,000	500	500	-	500	5,000	47	3	-
Methanol (CH4O)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Dimethyl ether (C2H6O)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ash	lb/hr	0	-	-	-	-	-	-	0	0	1,779	89	1,690	1,690	-	1,690	0	0	0	-
Char	lb/hr	0	-	-	-	-	-	-	40,406	40,002	0	0	0	0	-	0	404	4	0	-
Wood	lb/hr	183,718	-	-	-	-	-	-	0	0	-	-	-	-	-	-	0	0	0	-
Enthalpy Flow	MMBTU/hr	(553)	(408)	(0)	(0)	(53)	(33)	(17)	1,237	1,901	1,871	2,214	(342)	(0)	(2)	(2)	(664)	(621)	(44)	(343)

82

HP-205
CHAR COMBUSTOR
FLUE GAS STEAM
SUPERHEATER

HP-206
CHAR COMBUSTOR
FLUE GAS STEAM
GENERATOR

HP-202
CHAR COMBUSTOR FLUE
GAS/DME REACTOR FEED
EXCHANGER

HP-203
CHAR COMBUSTOR FLUE
GAS/METHANOL SYNTHESIS
REACTOR FEED EXCHANGER

HP-204
CHAR COMBUSTOR FLUE
GAS /BOILER FEEDWATER
PREHEAT EXCHANGER

HP-201
CHAR COMBUSTOR FLUE GAS/
HYDROCARBON SYNTHESIS
REACTOR FEED EXCHANGER

SATURATED STEAM
FROM T-605
PFD-61713-A603

SUPERHEATED STEAM
TO M-602A
PFD-61713-A603

DME REACTOR FEED
FROM HP-1403
PFD-61713-A1404

HP-202
QHP202

1411 DME REACTOR FEED
TO R-1424
PFD-61713-A1401

METH SYN FEED
FROM HP-301/HP-303
PFD-61713-A303

HC SYNTHESIS FEED
FROM
PFD-61713-A1402

HP-205
QHP205

FLUE GAS
FROM S-204
PFD-61713-A201

286

FLUE GAS
FROM HP-1503
PFD-61713-A1501

HP-201
QHP201

287

FLUE GAS
TO M-105
PFD-61713-A102

FLUE GAS
TO HP-501
PFD-61713-A501

HP-203
QHP203

485 METH SYN FEED
TO R-490
PFD-61713-A401

HC SYNTHESIS FEED
TO R-1410C
PFD-61713-A1402

SATURATED STEAM
TO T-605
PFD-61713-A603

BOILER FEEDWATER
FROM HP-405
PFD-61713-A402

HP-206
QHP206

HP-204
QHP204

BOILER FEEDWATER
TO M-603
PFD-61713-A602

BOILER FEEDWATER
TO R-490
PFD-61713-A401

BOILER FEEDWATER
TO HP-308
PFD-61713-A304

Plot date: 1/29/2015   J:\1A - Process Solutions\30482 NREL Thermochemical Conv - Design-Model\Drawings\ 30482 Gasification PFDs 61713 150128.vsd

PFD-61713-A202

COMPONENT	UNITS	286	1,411	485	287
Total Flow	lb/hr	512,006	87,710	349,903	512,006
Temperature	F	1,798	437	437	230
Pressure	Psia	26.9	145.0	730.0	26.8
Vapor Fraction		1	1	1	1
Hydrogen (H2)	lb/hr	-	-	33,613	-
Carbon Monoxide (CO)	lb/hr	-	-	169,605	-
Nitrogen (N2)	lb/hr	350,552	-	7,377	350,552
Oxygen (O2)	lb/hr	17,901	-	-	17,901
Argon (Ar)	lb/hr	5,979	-	-	5,979
Carbon Dioxide (CO2)	lb/hr	106,177	0	62,958	106,177
Water (H2O)	lb/hr	31,371	1,010	490	31,371
Hydrogen Sulfide (H2S)	lb/hr	-	-	-	-
Sulfur dioxide (SO2)	lb/hr	27	-	-	27
Ammonia (NH3)	lb/hr	-	0	109	-
Nitrogen dioxide (NO2)	lb/hr	0	-	-	0
Methane (CH4)	lb/hr	-	-	56,032	-
Ethane (C2H6)	lb/hr	-	0	76	-
Ethylene (C2H4)	lb/hr	-	0	11,051	-
Acetylene (C2H2)	lb/hr	-	0	796	-
Propane (C3H8)	lb/hr	-	0	0	-
Propene (C3H6)	lb/hr	-	-	-	-
I-Butane (C4H10)	lb/hr	-	0	0	-
N-Butane (C4H10)	lb/hr	-	0	0	-
Butenes (C4H8)	lb/hr	-	-	-	-
Pentanes (C5H12)	lb/hr	-	0	0	-
Pentenes (C5H10)	lb/hr	-	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-	-
Benzene (C6H6)	lb/hr	-	3	3	-
Heptanes (C7H16)	lb/hr	-	-	-	-
Triptane	lb/hr	-	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-	-
Triptene	lb/hr	-	-	-	-
Octanes (C8H18)	lb/hr	-	-	-	-
Octenes (C8H16)	lb/hr	-	-	-	-
N-Nonanes (C9H20)	lb/hr	-	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-
Tar (C10H8)	lb/hr	-	0	0	-
Carbon (Solid)	lb/hr	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	-	-
Methanol (CH4O)	lb/hr	-	86,697	7,794	-
Dimethyl ether (C2H6O)	lb/hr	-	-	-	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-
Ash	lb/hr	-	-	-	-
Char	lb/hr	-	-	-	-
Wood	lb/hr	-	-	-	-
Enthalpy Flow	MMBTU/hr	(343)	(229)	(576)	(570)

PROCESS FLOW DIAGRAM
AREA 300: TAR REFORMING

THERMOCHEMICAL CONVERSION OF BIOMASS TO FUELS AND CHEMICALS GASIFICATION

Equipment labels:

- K-305 COMBUSTION AIR BLOWER
- M-300H TAR REFORMER INLET MIXER
- M-300 COMBUSTOR FUEL PRE-MIXER
- R-303 TAR REFORMER
- S-306 TAR REFORMER 2-STAGE CYCLONE
- R-301A TAR REFORMER CATALYST REGENERATOR
- S-307 CATALYST REGENERATOR 2-STAGE CYCLONE

Stream labels:

- 330 SYNGAS TO HP-308 PFD-61713-A304
- 332 FLUE GAS TO HP-309/HP-310 PFD-61713-A305
- 326 MAKE-UP CATALYST FROM T-707 PFD-61713-702
- 662A SUPERHEATED STEAM FROM HP-312 PFD-61713-A305
- 225A SYNGAS FROM S-202 PFD-61713-A201
- 325A
- 385T RECYCLE GAS TO HP-315 PFD-61713-A305
- 370 RECYCLE GAS FROM HP-315 PFD-61713-A305
- 426 METH SYN OFFGAS FROM K-412/S-445 PFD-61713-A401
- 385F
- 513 METH COL OFFGAS FROM D-500 PFD-61713-A501
- 1483 HC OFFGAS FROM HP-1404 PFD-61713-A1403
- 328
- 225F SYNGAS FROM S-202 PFD-61713-A201
- 329 CATALYST PURGE TO DISPOSAL
- 330B
- 330A
- COMBUSTION AIR
- WK305
- REGEN AIR TO HP-317/HP-319 PFD-61713-A304
- REGEN AIR TO HP-505 PFD-61713-A502
- REGEN AIR FROM HP-305 PFD-61713-A304
- REGEN AIR FROM HP-403 PFD-61713-A402
- REGEN AIR FROM HP-1402 PFD-61713-A1404

Title block:

Rev.	Date	By	Description
1	12/5/14	DMS	ADDED HX NETWORK
0	9/26/14	JAR	ISSUED FOR DESIGN REPORT

Project No: 30482.00
Drawing: PFD-61713-A301
Rev: 1

Harris Group

PFD-61713-A301

COMPONENT	UNITS	326	662A	225A	370	385T	426	513	1483	225F	385F	328	325A	330	332	329	330A	330B
Total Flow	lb/hr	9	57,020	221,320	7,645	7,645	19,113	4,819	4,649	15,668	11,468	59	285,994	285,985	288,332	251,729	50	285,985
Temperature	F	-	1,600	1,598	1,600	507	507	217	173	1,598	507	-	1,598	1,670	2,452	90	-	1,670
Pressure	Psia	35.0	102.6	31.4	29	34	34	98.0	69.5	31.4	33.7	25.7	28.7	25.7	17.5	14.7	25.7	25.7
Vapor Fraction		-	1	1	1	1	1	1	1	1	1	-	1	1	1	1	-	1
Hydrogen (H2)	lb/hr	-	-	2,892	494	494	1,236	11	0	205	741	-	3,386	13,303	-	-	-	13,303
Carbon Monoxide (CO)	lb/hr	-	-	71,015	4,004	4,004	10,011	75	-	5,027	6,007	-	75,020	87,773	-	-	-	87,773
Nitrogen (N2)	lb/hr	-	-	-	157	157	393	11	-	-	236	-	157	404	186,648	186,401	-	404
Oxygen (O2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	8,104	57,113	-	-
Argon (Ar)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	3,179	3,179	-	-
Carbon Dioxide (CO2)	lb/hr	-	-	33,837	1,357	1,357	3,393	1,639	-	2,395	2,036	-	35,194	83,544	52,037	124	-	83,544
Water (H2O)	lb/hr	-	57,020	87,349	14	14	36	8	27	6,184	22	-	144,384	96,689	38,211	4,912	-	96,689
Hydrogen Sulfide (H2S)	lb/hr	-	-	151	0	0	0	-	-	11	0	-	151	151	-	-	-	151
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	20	-	-	-
Ammonia (NH3)	lb/hr	-	-	331	2	2	6	21	0	23	4	-	334	33	-	-	-	33
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	129	-	-	-
Methane (CH4)	lb/hr	-	-	14,811	1,196	1,196	2,989	193	421	1,048	1,794	-	16,006	3,201	-	-	-	3,201
Ethane (C2H6)	lb/hr	-	-	527	2	2	4	1	118	37	2	-	529	5	-	-	-	5
Ethylene (C2H4)	lb/hr	-	-	7,331	236	236	591	156	74	519	355	-	7,568	757	-	-	-	757
Acetylene (C2H2)	lb/hr	-	-	638	17	17	43	22	0	45	26	-	655	66	-	-	-	66
Propane (C3H8)	lb/hr	-	-	-	0	0	0	0	463	-	0	-	0	0	-	-	-	0
Propene (C3H6)	lb/hr	-	-	-	-	-	-	-	5	-	-	-	-	-	-	-	-	-
I-Butane (C4H10)	lb/hr	-	-	-	0	0	0	0	1,356	-	0	-	0	0	-	-	-	0
N-Butane (C4H10)	lb/hr	-	-	-	0	0	0	0	17	-	0	-	0	0	-	-	-	0
Butenes (C4H8)	lb/hr	-	-	-	-	-	-	-	34	-	-	-	-	-	-	-	-	-
Pentanes (C5H12)	lb/hr	-	-	-	0	0	0	0	49	-	0	-	0	0	-	-	-	0
Pentenes (C5H10)	lb/hr	-	-	-	-	-	-	-	33	-	-	-	-	-	-	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-	-	-	-	-	30	-	-	-	-	-	-	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-	-	-	-	-	6	-	-	-	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	-	-	597	0	0	0	0	0	42	0	-	597	6	-	-	-	6
Heptanes (C7H16)	lb/hr	-	-	-	-	-	-	-	12	-	-	-	-	-	-	-	-	-
Triptane	lb/hr	-	-	-	-	-	-	-	85	-	-	-	-	-	-	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-	-	-	-	-	8	-	-	-	-	-	-	-	-	-
Triptene	lb/hr	-	-	-	-	-	-	-	10	-	-	-	-	-	-	-	-	-
Octanes (C8H18)	lb/hr	-	-	-	-	-	-	-	37	-	-	-	-	-	-	-	-	-
Octenes (C8H16)	lb/hr	-	-	-	-	-	-	-	3	-	-	-	-	-	-	-	-	-
N-Nonanes (C9H20)	lb/hr	-	-	-	-	-	-	-	0	-	-	-	-	-	-	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-	-	-	-	-	0	-	-	-	-	-	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Tar (C10H8)	lb/hr	-	-	1,791	0	0	0	0	0	127	0	-	1,791	2	-	-	-	2
Carbon (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	9	-	47	-	-	-	-	-	3	-	56	56	47	3	-	47	47
Methanol (CH4O)	lb/hr	-	-	-	164	164	411	2,681	-	-	246	-	164	0	-	-	-	0
Dimethyl ether (C2H6O)	lb/hr	-	-	-	-	-	-	-	1,859	-	-	-	-	-	-	-	-	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ash	lb/hr	-	-	0	-	-	-	-	-	0	-	0	0	0	-	-	0	0
Char	lb/hr	-	-	4	-	-	-	-	-	0	-	4	4	4	0	-	4	4
Wood	lb/hr	-	-	0	-	-	-	-	-	0	-	0	0	0	-	-	0	0
Enthalpy Flow	MMBTU/hr	(0)	(286)	(621)	(8)	(13)	(32)	(14)	(6)	(44)	(19)	0	(915)	(803)	(210)	(28)	0	(803)

86

Process Flow Diagram — Area 300: Syngas Quench

Equipment labels:

Tag	Description
H-301B	POST-REFORMER AIR COOLER
T-302	QUENCH WATER RECIRCULATION TANK
M-301	SYNGAS CYCLONE SEPARATOR
P-302	QUENCH WATER RECIRCULATION PUMP
M-302	SYNGAS VENTURI SCRUBBER
H-301	QUENCH WATER RECIRCULATION COOLER
T-301	SLUDGE SETTLING TANK
P-314	CONDENSATE PUMP
P-301	SLUDGE PUMP

Stream/connection labels:

- 315 — SYNGAS TO K-310 PFD-61713-A303
- 330C — SYNGAS FROM HP-316 PFD-61713-A304
- 331
- 432 — WATER MAKE-UP FROM S-410 PFD-61713-A403
- 314 — SCRUBBER COND TO M-705 PFD-61713-A701
- 336 — SOLIDS TO T-709 PFD-61713-A701
- 315SCRUB — CONDENSATE FROM P-315 PFD-61713-A303
- QAH301B
- WP302
- QCH301
- CWR / CWS
- WP314
- WP301

Harris Group
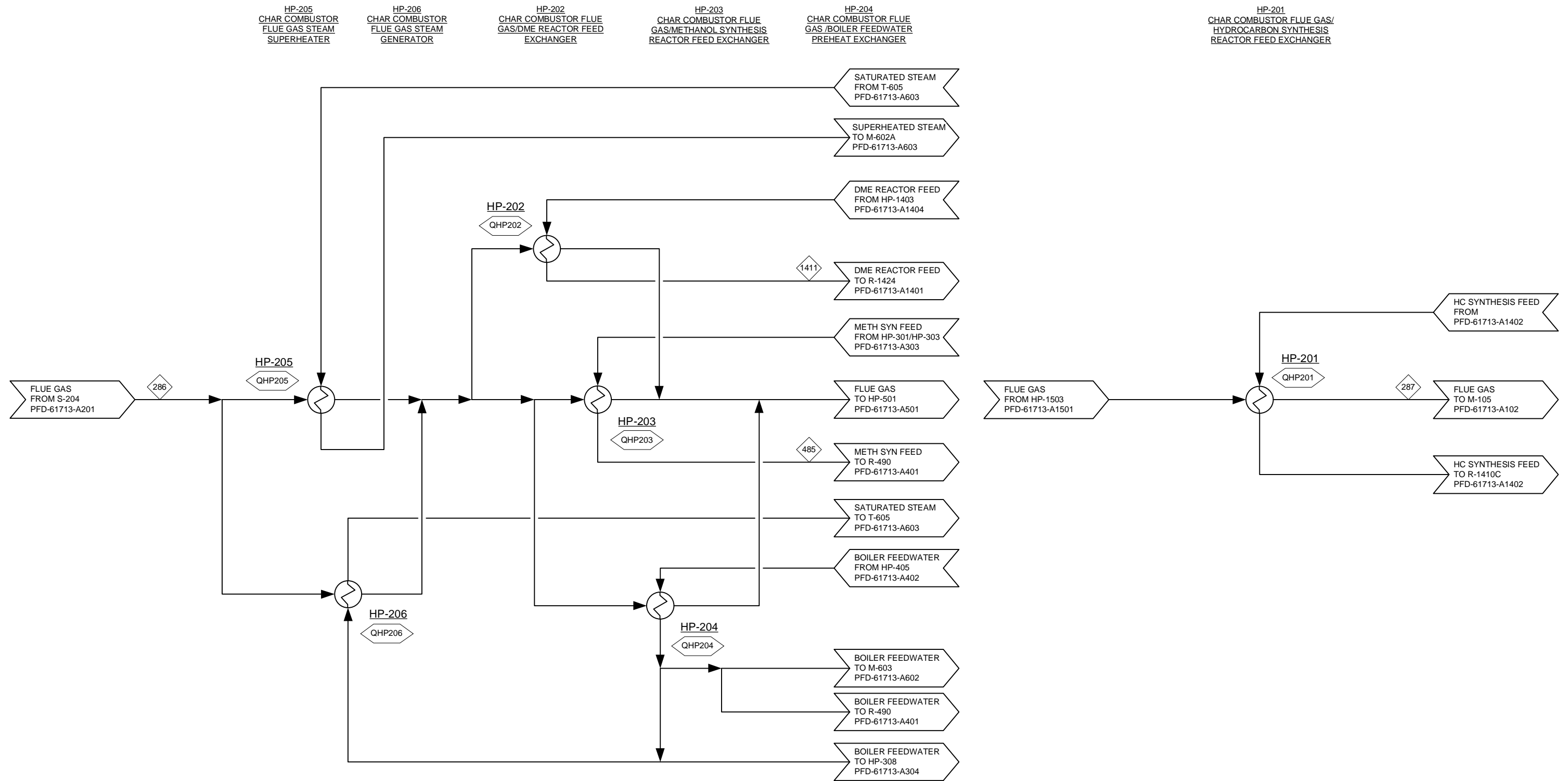
NREL — NATIONAL RENEWABLE ENERGY LABORATORY

THERMOCHEMICAL CONVERSION OF BIOMASS TO FUELS AND CHEMICALS GASIFICATION

PROCESS FLOW DIAGRAM AREA 300: SYNGAS QUENCH

Rev.	Date	By	Description
1	12/5/14	DMS	ADDED HX NETWORK
0	9/26/14	JAR	ISSUED FOR DESIGN REPORT

Project No: 30482.00   Drawing: PFD-61713-A302   Rev: 1

PFD-61713-A302

COMPONENT	UNITS	330C	331	432	315	315SCRUB	314	336
Total Flow	lb/hr	285,985	190,834	1,015	212,665	1,501	75,638	101
Temperature	F	179	110	110	137	136	130	130
Pressure	Psia	24.7	430.0	420.0	22.7	35.0	35.0	35
Vapor Fraction		1	1	-	1	-	-	-
Hydrogen (H2)	lb/hr	13,303	13,303	0	13,303	0	0	-
Carbon Monoxide (CO)	lb/hr	87,773	87,728	0	87,728	0	0	-
Nitrogen (N2)	lb/hr	404	404	0	404	0	0	-
Oxygen (O2)	lb/hr	-	-	-	-	-	-	-
Argon (Ar)	lb/hr	-	-	-	-	-	-	-
Carbon Dioxide (CO2)	lb/hr	83,544	83,531	0	83,531	0	1	-
Water (H2O)	lb/hr	96,689	1,686	1,013	23,514	1,501	75,632	50
Hydrogen Sulfide (H2S)	lb/hr	151	150	0	150	0	0	-
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	33	29	2	30	0	5	-
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-	-	-
Methane (CH4)	lb/hr	3,201	3,183	0	3,183	0	0	-
Ethane (C2H6)	lb/hr	5	5	0	5	0	0	-
Ethylene (C2H4)	lb/hr	757	747	0	747	0	0	-
Acetylene (C2H2)	lb/hr	66	65	0	65	0	0	-
Propane (C3H8)	lb/hr	0	0	0	0	0	0	-
Propene (C3H6)	lb/hr	-	-	-	-	-	-	-
I-Butane (C4H10)	lb/hr	0	0	0	0	0	0	-
N-Butane (C4H10)	lb/hr	0	0	0	0	0	0	-
Butenes (C4H8)	lb/hr	-	-	-	-	-	-	-
Pentanes (C5H12)	lb/hr	0	0	0	0	0	0	-
Pentenes (C5H10)	lb/hr	-	-	-	-	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-	-	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	6	3	0	3	0	0	-
Heptanes (C7H16)	lb/hr	-	-	-	-	-	-	-
Triptane	lb/hr	-	-	-	-	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-	-	-	-	-
Triptene	lb/hr	-	-	-	-	-	-	-
Octanes (C8H18)	lb/hr	-	-	-	-	-	-	-
Octenes (C8H16)	lb/hr	-	-	-	-	-	-	-
N-Nonanes (C9H20)	lb/hr	-	-	-	-	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-	-	-
Tar (C10H8)	lb/hr	2	0	0	0	0	0	-
Carbon (Solid)	lb/hr	-	-	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	47	-	-	-	-	-	47
Methanol (CH4O)	lb/hr	0	0	0	0	0	0	-
Dimethyl ether (C2H6O)	lb/hr	-	-	-	-	-	-	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-	-	-
Ash	lb/hr	0	-	-	-	-	-	0
Char	lb/hr	4	-	-	-	-	-	4
Wood	lb/hr	0	-	-	-	-	-	0
Enthalpy Flow	MMBTU/hr	(1,026)	(484)	(7)	(606)	(10)	(517)	(0)

88

Equipment labels (top of diagram):

- **HP-301** COMPRESSOR INTERSTAGE/ METHANOL SYNTHESIS REACTOR FEED EXCHANGER
- **HP-302** COMPRESSOR INTERSTAGE/ BOILER FEEDWATER PREHEAT EXCHANGER
- **K-310** SYNTHESIS GAS COMPRESSOR
- **H-310AAC** 1ST STAGE AIR INTERCOOLER
- **H-310ACW** 1ST STAGE WATER INTERCOOLER
- **H-310BAC** 2ND STAGE AIR INTERCOOLER
- **H-310BCW** 2ND STAGE WATER INTERCOOLER
- **H-310CAC** 3RD STAGE AIR INTERCOOLER
- **H-310CCW** 3RD STAGE WATER INTERCOOLER
- **HP-303** COMPRESSOR INTERSTAGE/ METHANOL SYNTHESIS REACTOR FEED EXCHANGER
- **HP-304** COMPRESSOR INTERSTAGE/ BOILER FEEDWATER PREHEAT EXCHANGER
- **H-310DAC** 4TH STAGE AIR INTERCOOLER
- **H-310DCW** 4TH STAGE WATER INTERCOOLER
- **P-315** SYNGAS COMPRESSOR CONDENSATE PUMP

Stream connections:

- METH SYN FEED FROM K-410/K-414 PFD-61713-A401 (470)
- BOILER FEEDWATER FROM M-601A PFD-61713-A601
- METH SYN FEED TO HP-203 PFD-61713-A202
- COMPRESSED SYNGAS TO S-405/S-410 PFD-61713-A401 (331)
- BOILER FEEDWATER TO HP-503 PFD-61713-A502
- SYNGAS FROM M-301 PFD-61713-A302 (315)
- CONDENSATE TO M-701 PFD-61713-A701 (315)
- CONDENSATE TO T-302 PFD-61713-A302 (315SCRUB)

Tags: HP-301 QHP301; HP-302 QHP302; HP-303 QHP303; HP-304 QHP304; H-310AAC QH310AAC; H-310ACW QH310ACW; H-310BAC QH310BAC; H-310BCW QH310BCW; H-310CAC QH310CAC; H-310CCW QH310CCW; H-310DAC QH310DAC; H-310DCW QH310DCW; K-310 WK-310; P-315 WP315; 315L; 315P; CWR; CWS

Rev.	Date	By	Description
1	12/5/14	DMS	ADDED HX NETWORK
0	9/26/14	JAR	ISSUED FOR DESIGN REPORT

THERMOCHEMICAL CONVERSION OF BIOMASS TO FUELS AND CHEMICALS GASIFICATION

PROCESS FLOW DIAGRAM
AREA 300: HIGH PRESSURE SYNGAS COMPRESSION

Project No. 30482.00   Drawing PFD-61713-A303   Rev. 1

PFD-61713-A303

COMPONENT	UNITS	315	470	331	315L	315P	315	315SCRUB
Total Flow	lb/hr	212,665	349,903	190,834	21,830	21,830	20,329	1,501
Temperature	F	137	173	110	136	136	136	136
Pressure	Psia	22.7	735.0	430.0	22.7	35.0	35.0	35.0
Vapor Fraction		1	1	1	-	-	-	-
Hydrogen (H2)	lb/hr	13,303	33,613	13,303	0	0	0	0
Carbon Monoxide (CO)	lb/hr	87,728	169,605	87,728	0	0	0	0
Nitrogen (N2)	lb/hr	404	7,377	404	0	0	0	0
Oxygen (O2)	lb/hr	-	-	-	-	-	-	-
Argon (Ar)	lb/hr	-	-	-	-	-	-	-
Carbon Dioxide (CO2)	lb/hr	83,531	62,958	83,531	0	0	0	0
Water (H2O)	lb/hr	23,514	490	1,686	21,828	21,828	20,327	1,501
Hydrogen Sulfide (H2S)	lb/hr	150	-	150	0	0	0	0
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	30	109	29	2	2	1	0
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-	-	-
Methane (CH4)	lb/hr	3,183	56,032	3,183	0	0	0	0
Ethane (C2H6)	lb/hr	5	76	5	0	0	0	0
Ethylene (C2H4)	lb/hr	747	11,051	747	0	0	0	0
Acetylene (C2H2)	lb/hr	65	796	65	0	0	0	0
Propane (C3H8)	lb/hr	0	0	0	0	0	0	0
Propene (C3H6)	lb/hr	-	-	-	-	-	-	-
I-Butane (C4H10)	lb/hr	0	0	0	0	0	0	0
N-Butane (C4H10)	lb/hr	0	0	0	0	0	0	0
Butenes (C4H8)	lb/hr	-	-	-	-	-	-	-
Pentanes (C5H12)	lb/hr	0	0	0	0	0	0	0
Pentenes (C5H10)	lb/hr	-	-	-	-	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-	-	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	3	3	3	0	0	0	0
Heptanes (C7H16)	lb/hr	-	-	-	-	-	-	-
Triptane	lb/hr	-	-	-	-	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-	-	-	-	-
Triptene	lb/hr	-	-	-	-	-	-	-
Octanes (C8H18)	lb/hr	-	-	-	-	-	-	-
Octenes (C8H16)	lb/hr	-	-	-	-	-	-	-
N-Nonanes (C9H20)	lb/hr	-	-	-	-	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-	-	-
Tar (C10H8)	lb/hr	0	0	0	0	0	0	0
Carbon (Solid)	lb/hr	-	-	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	-	-	-	-	-
Methanol (CH4O)	lb/hr	0	7,794	0	0	0	0	0
Dimethyl ether (C2H6O)	lb/hr	-	-	-	-	-	-	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-	-	-
Ash	lb/hr	-	-	-	-	-	-	-
Char	lb/hr	-	-	-	-	-	-	-
Wood	lb/hr	-	-	-	-	-	-	-
Enthalpy Flow	MMBTU/hr	(606)	(635)	(484)	(149)	(149)	(139)	(10)

90
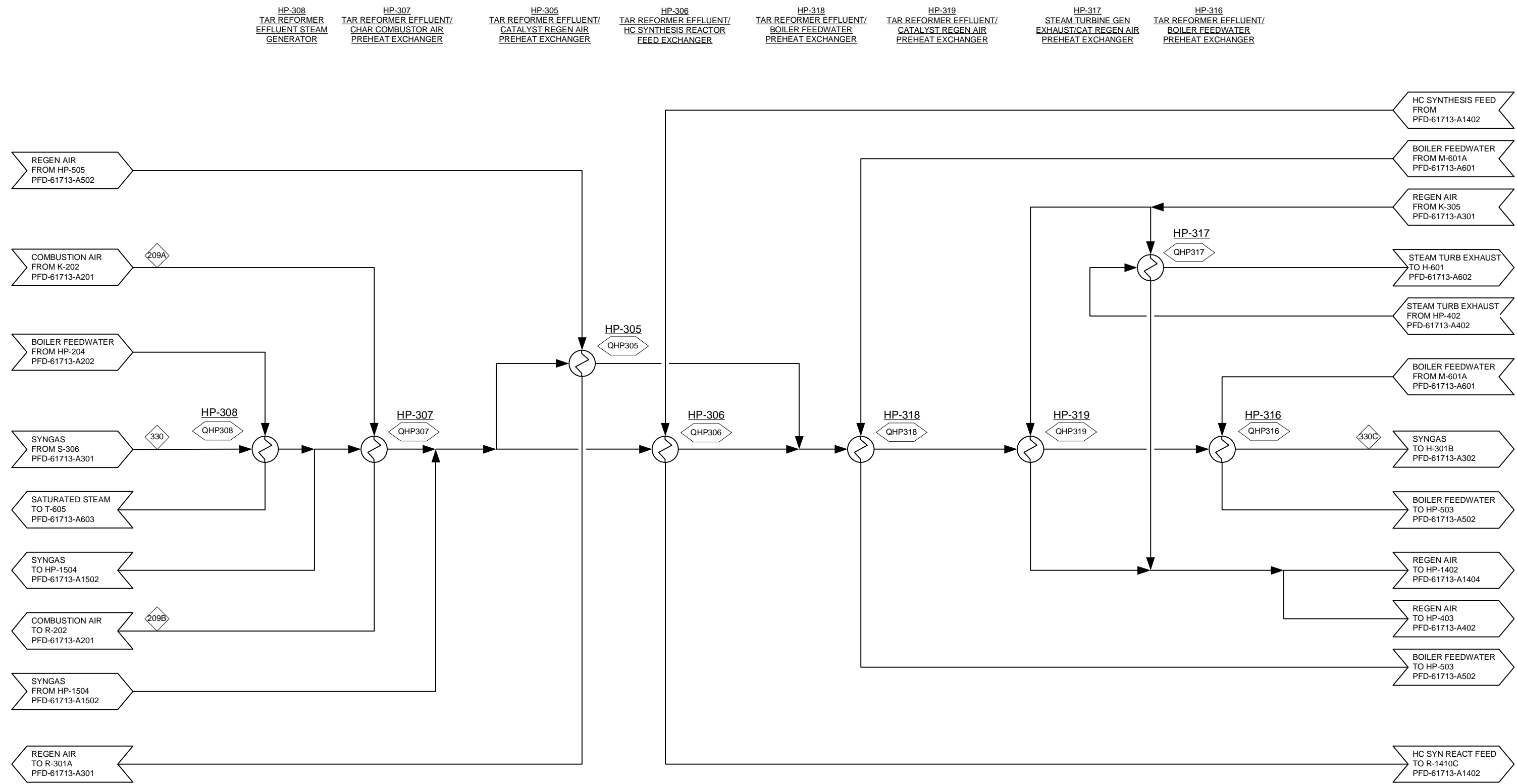
HP-308
TAR REFORMER
EFFLUENT STEAM
GENERATOR

HP-307
TAR REFORMER EFFLUENT/
CHAR COMBUSTOR AIR
PREHEAT EXCHANGER

HP-305
TAR REFORMER EFFLUENT/
CATALYST REGEN AIR
PREHEAT EXCHANGER

HP-306
TAR REFORMER EFFLUENT/
HC SYNTHESIS REACTOR
FEED EXCHANGER

HP-318
TAR REFORMER EFFLUENT/
BOILER FEEDWATER
PREHEAT EXCHANGER

HP-319
TAR REFORMER EFFLUENT/
CATALYST REGEN AIR
PREHEAT EXCHANGER

HP-317
STEAM TURBINE GEN
EXHAUST/CAT REGEN AIR
PREHEAT EXCHANGER

HP-316
TAR REFORMER EFFLUENT/
BOILER FEEDWATER
PREHEAT EXCHANGER

HC SYNTHESIS FEED
FROM
PFD-61713-A1402

REGEN AIR
FROM HP-505
PFD-61713-A502

BOILER FEEDWATER
FROM M-601A
PFD-61713-A601

REGEN AIR
FROM K-305
PFD-61713-A301

HP-317
QHP317

COMBUSTION AIR
FROM K-202
PFD-61713-A201

209A

STEAM TURB EXHAUST
TO H-601
PFD-61713-A602

STEAM TURB EXHAUST
FROM HP-402
PFD-61713-A402

HP-305
QHP305

BOILER FEEDWATER
FROM HP-204
PFD-61713-A202

BOILER FEEDWATER
FROM M-601A
PFD-61713-A601

HP-308
QHP308

HP-307
QHP307

HP-306
QHP306

HP-318
QHP318

HP-319
QHP319

HP-316
QHP316

SYNGAS
FROM S-306
PFD-61713-A301

330

330C

SYNGAS
TO H-301B
PFD-61713-A302

SATURATED STEAM
TO T-605
PFD-61713-A603

BOILER FEEDWATER
TO HP-503
PFD-61713-A502

SYNGAS
TO HP-1504
PFD-61713-A1502

REGEN AIR
TO HP-1402
PFD-61713-A1404

COMBUSTION AIR
TO R-202
PFD-61713-A201

209B

REGEN AIR
TO HP-403
PFD-61713-A402

SYNGAS
FROM HP-1504
PFD-61713-A1502

BOILER FEEDWATER
TO HP-503
PFD-61713-A502

REGEN AIR
TO R-301A
PFD-61713-A301

HC SYN REACT FEED
TO R-1410C
PFD-61713-A1402

Pbt date: 1/29/2015   J:\1A - Process Solutions\30482 NREL Thermochemical Conv - Design-Model\Drawings\ 30482 Gasification PFDs 61713 150128.vsd
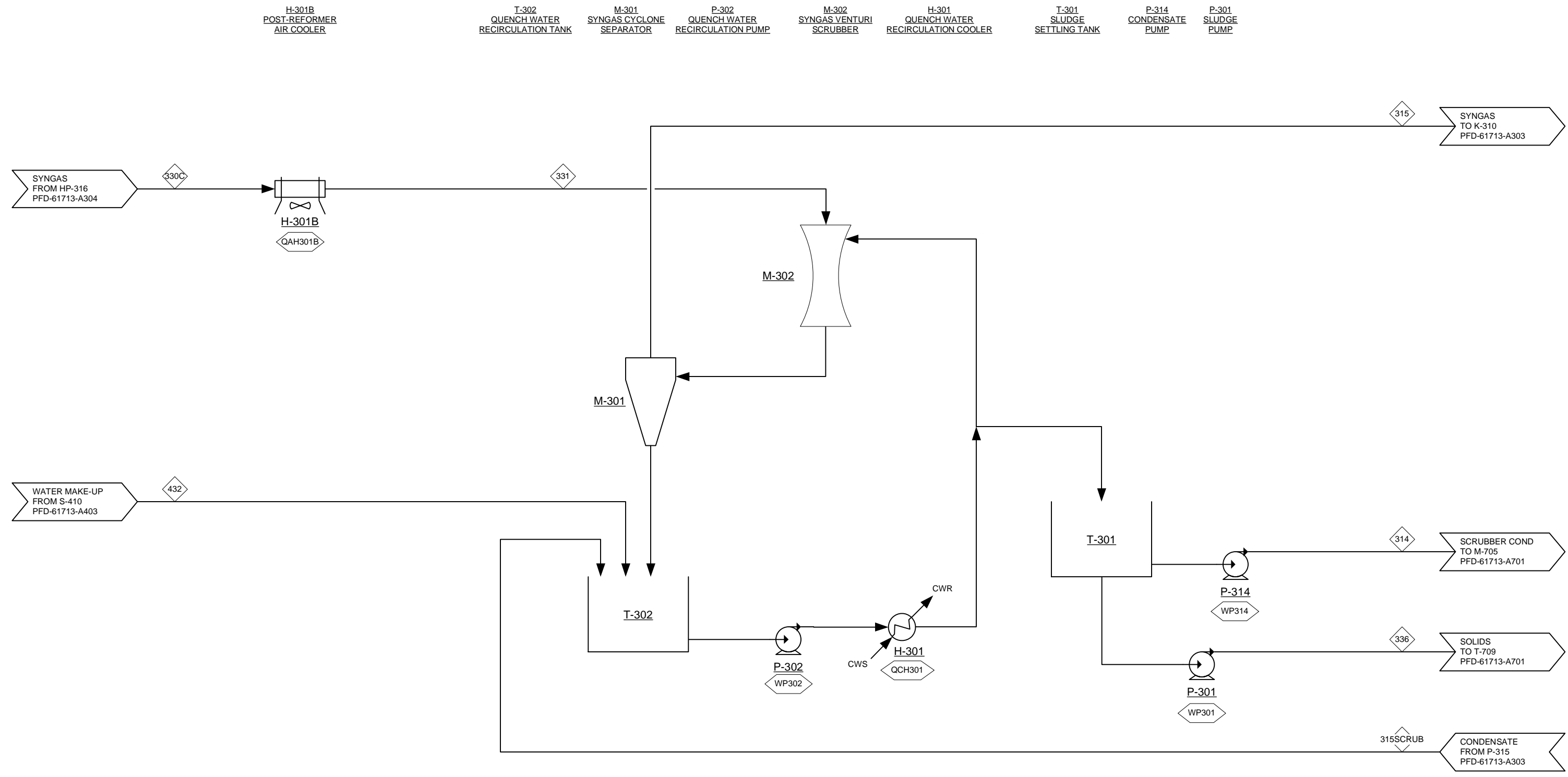

THERMOCHEMICAL CONVERSION OF BIOMASS
TO FUELS AND CHEMICALS
GASIFICATION

PROCESS FLOW DIAGRAM
AREA 300: A300 HEAT INTEGRATION

NATIONAL RENEWABLE ENERGY LABORATORY

Harris Group

Rev.	Date	By	Description	Project No.	Drawing	Rev:
0	12/5/14	DMS	ADDED HX NETWORK	30482.00	PFD-61713-A304	0

PFD-61713-A304

COMPONENT	UNITS	209A	330	209B	330C
Total Flow	lb/hr	473,382	285,985	473,382	285,985
Temperature	F	258	1,670	400	179
Pressure	Psia	29.0	25.7	28.9	24.7
Vapor Fraction		1	1	1	1
Hydrogen (H2)	lb/hr	-	13,303	-	13,303
Carbon Monoxide (CO)	lb/hr	-	87,773	-	87,773
Nitrogen (N2)	lb/hr	350,531	404	350,531	404
Oxygen (O2)	lb/hr	107,403	-	107,403	-
Argon (Ar)	lb/hr	5,979	-	5,979	-
Carbon Dioxide (CO2)	lb/hr	233	83,544	233	83,544
Water (H2O)	lb/hr	9,236	96,689	9,236	96,689
Hydrogen Sulfide (H2S)	lb/hr	-	151	-	151
Sulfur dioxide (SO2)	lb/hr	-	-	-	-
Ammonia (NH3)	lb/hr	-	33	-	33
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-
Methane (CH4)	lb/hr	-	3,201	-	3,201
Ethane (C2H6)	lb/hr	-	5	-	5
Ethylene (C2H4)	lb/hr	-	757	-	757
Acetylene (C2H2)	lb/hr	-	66	-	66
Propane (C3H8)	lb/hr	-	0	-	0
Propene (C3H6)	lb/hr	-	-	-	-
I-Butane (C4H10)	lb/hr	-	0	-	0
N-Butane (C4H10)	lb/hr	-	0	-	0
Butenes (C4H8)	lb/hr	-	-	-	-
Pentanes (C5H12)	lb/hr	-	0	-	0
Pentenes (C5H10)	lb/hr	-	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-	-
Benzene (C6H6)	lb/hr	-	6	-	6
Heptanes (C7H16)	lb/hr	-	-	-	-
Triptane	lb/hr	-	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-	-
Triptene	lb/hr	-	-	-	-
Octanes (C8H18)	lb/hr	-	-	-	-
Octenes (C8H16)	lb/hr	-	-	-	-
N-Nonanes (C9H20)	lb/hr	-	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-
Tar (C10H8)	lb/hr	-	2	-	2
Carbon (Solid)	lb/hr	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	47	-	47
Methanol (CH4O)	lb/hr	-	0	-	0
Dimethyl ether (C2H6O)	lb/hr	-	-	-	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-
Ash	lb/hr	-	0	-	0
Char	lb/hr	-	4	-	4
Wood	lb/hr	-	0	-	0
Enthalpy Flow	MMBTU/hr	(33)	(803)	(17)	(1,026)

92

Process Flow Diagram — Area 300: A300 Heat Integration

HP-309
CAT REGEN FLUE GAS/
STEAM TURB GEN INLET
(STAGE 2) EXCHANGER

HP-310
CAT REGEN FLUE GAS/
STEAM TURB GEN INLET
(STAGE 3) EXCHANGER

HP-311
CAT REGEN FLUE GAS/
PURGE GAS EXPANDER
INLET EXCHANGER

HP-312
CAT REGEN FLUE GAS
STEAM SUPERHEATER

HP-313
CAT REGEN FLUE GAS/
GASIFIER STEAM
PREHEAT EXCHANGER

HP-314
CAT REGEN FLUE GAS
STEAM SUPERHEATER

HP-315
CAT REGEN FLUE GAS/
SYNGAS RECYCLE
EXCHANGER

HP-311
QHP311

HP-312
QHP312

HP-313
QHP313

HP-314
QHP314

HP-315
QHP315

HP-309
QHP309

HP-310
QHP310

PURGE GAS
FROM HP-401
PFD-61713-A402

498 — PURGE GAS
TO K-412
PFD-61713-A401

662 — STEAM
FROM T-607
PFD-61713-A603

610 — STEAM
FROM M-602B
PFD-61713-A602

662A — SUPERHEATED STEAM
TO M-300H
PFD-61713-A301

609A — STEAM
FROM M-602A
PFD-61713-A602

200 — STEAM
FROM M-602B/M-603
PFD-61713-A602

332 — FLUE GAS
FROM S-307
PFD-61713-A301

333 — FLUE GAS
TO M-105
PFD-61713-A102

200B — SUPERHEATED STEAM
TO R-201
PFD-61713-A201

602C — STEAM
TO M-602B
PFD-61713-A602

SATURATED STEAM
FROM T-605
PFD-61713-A603

611C — STEAM
TO M-602C
PFD-61713-A602

SUPERHEATED STEAM
TO M-602A
PFD-61713-A603

385T — RECYCLE GAS
FROM K-412/S-445
PFD-61713-A301

370 — RECYCLE GAS
TO M-300H
PFD-61713-A301
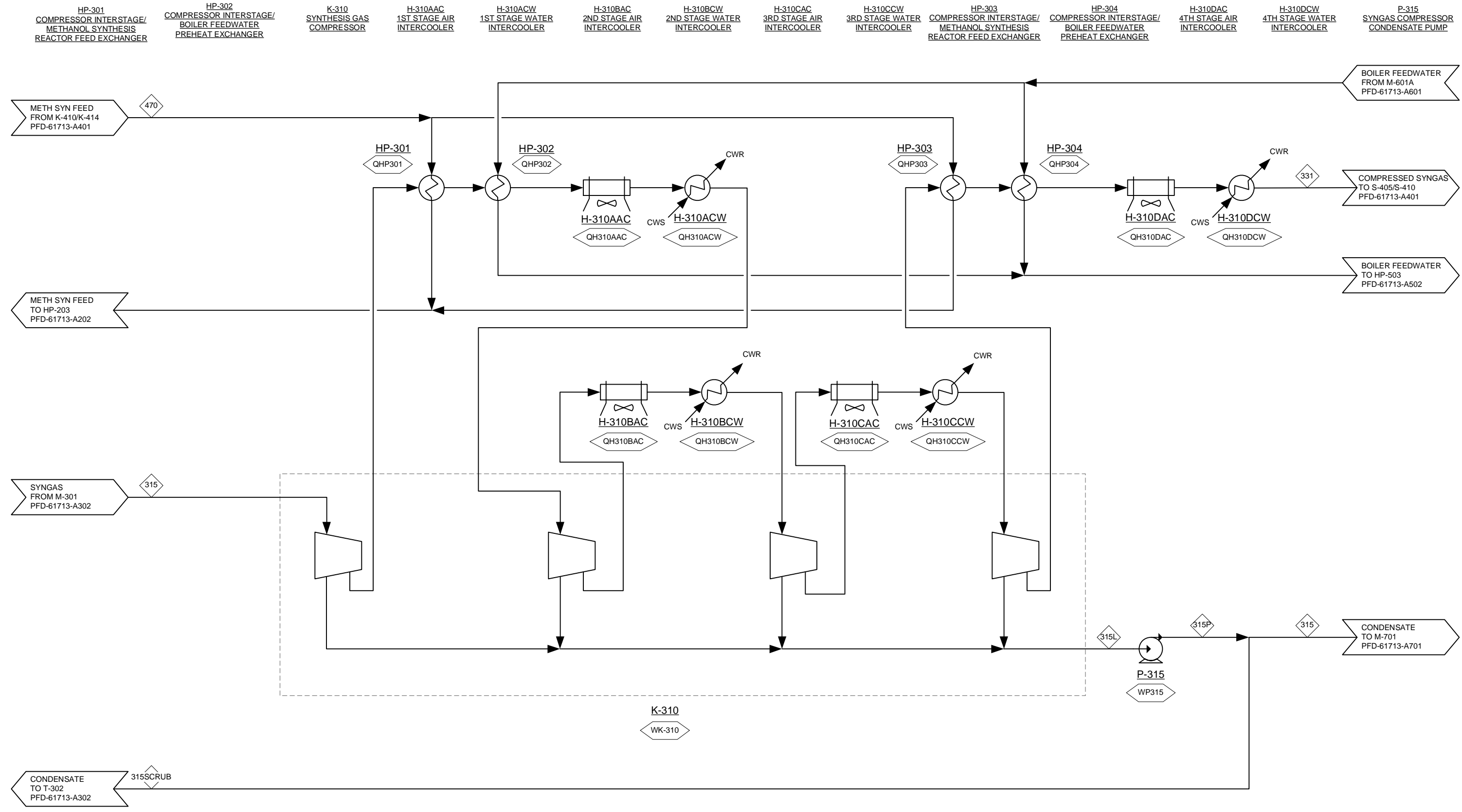
THERMOCHEMICAL CONVERSION OF BIOMASS
TO FUELS AND CHEMICALS
GASIFICATION

PROCESS FLOW DIAGRAM
AREA 300: A300 HEAT INTEGRATION

Harris Group

NREL
NATIONAL RENEWABLE ENERGY LABORATORY

Rev.	Date	By	Description
0	12/5/14	DMS	ADDED HX NETWORK

Project No: 30482.00   Drawing: PFD-61713-A305   Rev: 0

PFD-61713-A305

COMPONENT	UNITS	610	609A	332	602C	611C	498	662	662A	200	333	200B	385T	370
Total Flow	lb/hr	404,719	404,719	288,332	404,719	177,601	12,873	57,020	57,020	73,120	288,332	73,120	7,645	7,645
Temperature	F	784	783	2,452	1,000	1,000	1,200	330	1,600	782	702	500	507	1,600
Pressure	Psia	64.7	176.0	17.5	174.0	64.7	419.5	103.1	102.6	35.0	17.4	35.0	33.7	28.7
Vapor Fraction		1	1	1	1	1	1	1	1	1	1	1	1	1
Hydrogen (H2)	lb/hr	-	-	-	-	-	1,113	-	-	-	-	-	494	494
Carbon Monoxide (CO)	lb/hr	-	-	-	-	-	4,596	-	-	-	-	-	4,004	4,004
Nitrogen (N2)	lb/hr	-	-	186,648	-	-	368	-	-	-	186,648	-	157	157
Oxygen (O2)	lb/hr	-	-	8,104	-	-	-	-	-	-	8,104	-	-	-
Argon (Ar)	lb/hr	-	-	3,179	-	-	-	-	-	-	3,179	-	-	-
Carbon Dioxide (CO2)	lb/hr	-	-	52,037	-	-	3,000	-	-	-	52,037	-	1,357	1,357
Water (H2O)	lb/hr	404,719	404,719	38,211	404,719	177,601	1	57,020	57,020	73,120	38,211	73,120	14	14
Hydrogen Sulfide (H2S)	lb/hr	-	-	-	-	-	0	-	-	-	-	-	0	0
Sulfur dioxide (SO2)	lb/hr	-	-	20	-	-	-	-	-	-	20	-	-	-
Ammonia (NH3)	lb/hr	-	-	-	-	-	4	-	-	-	-	-	2	2
Nitrogen dioxide (NO2)	lb/hr	-	-	129	-	-	-	-	-	-	129	-	-	-
Methane (CH4)	lb/hr	-	-	-	-	-	2,793	-	-	-	-	-	1,196	1,196
Ethane (C2H6)	lb/hr	-	-	-	-	-	4	-	-	-	-	-	2	2
Ethylene (C2H4)	lb/hr	-	-	-	-	-	545	-	-	-	-	-	236	236
Acetylene (C2H2)	lb/hr	-	-	-	-	-	39	-	-	-	-	-	17	17
Propane (C3H8)	lb/hr	-	-	-	-	-	0	-	-	-	-	-	0	0
Propene (C3H6)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
I-Butane (C4H10)	lb/hr	-	-	-	-	-	0	-	-	-	-	-	0	0
N-Butane (C4H10)	lb/hr	-	-	-	-	-	0	-	-	-	-	-	0	0
Butenes (C4H8)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Pentanes (C5H12)	lb/hr	-	-	-	-	-	0	-	-	-	-	-	0	0
Pentenes (C5H10)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	-	-	-	-	-	0	-	-	-	-	-	0	0
Heptanes (C7H16)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Triptane	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Triptene	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Octanes (C8H18)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Octenes (C8H16)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
N-Nonanes (C9H20)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Tar (C10H8)	lb/hr	-	-	-	-	-	0	-	-	-	-	-	0	0
Carbon (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	3	-	-	-	-	-	-	3	-	-	-
Methanol (CH4O)	lb/hr	-	-	-	-	-	411	-	-	-	-	-	164	164
Dimethyl ether (C2H6O)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Ash	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Char	lb/hr	-	-	0	-	-	-	-	-	-	0	-	-	-
Wood	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Enthalpy Flow	MMBTU/hr	(2,205)	(2,207)	(210)	(2,162)	(948)	(16)	(324)	(286)	(398)	(371)	(408)	(13)	(8)

Equipment list (column headers):
- S-405 ZNO SULFUR REMOVAL BEDS
- K-412 PURGE GAS EXPANDER
- S-445 PRESSURE SWING ADSORPTION UNIT
- K-414 UNREACTED SYNGAS RECYCLE COMPRESSOR
- K-410 SYNGAS 2ND COMPRESSOR
- S-414 METHANOL FLASH DRUM
- H-414 METHANOL CONDENSER
- T-480 METHANOL SURGE TANK
- P-480 METHANOL SURGE TANK PUMP
- H-413 SYNGAS AIR COOLER
- S-471 METHANOL CONDENSATION KNOCK-OUT
- R-490 METHANOL SYNTHESIS REACTOR
- H-411B POST-REACTOR SYNGAS AIR COOLER

Stream/connection labels:
- 331 — COMPRESSED SYNGAS FROM H-310DCW PFD-61713-A303
- 470B
- 470A — PURIFIED SYNGAS FROM S-410 PFD-61713-A403
- 470C
- 462A
- 478B
- 470D
- S-405
- S-445
- K-410 WK410
- 470 — METH SYN FEED TO HP-301/HP-303 PFD-61713-A303
- METH SYN FEED FROM HP-203 PFD-61713-A202
- 485
- 466A
- 483
- K-414 WH414
- 478
- 482
- 478D
- 478E
- 426 — METH SYN OFFGAS TO HP-315/M-300 PFD-61713-A301
- 426A
- K-412 WK412
- 498 — PURGE GAS FROM HP-311 PFD-61713-A305
- PURGE GAS TO HP-506 PFD-61713-A502 — 416Z
- S-414
- 473A
- 475
- CWR
- CWS
- H-414 QCH414
- 473
- H-413 QAH413
- 471V
- S-471
- 472
- 411B — METH SYN EFFLUENT FROM HP-403 PFD-61713-A402
- METH SYN EFFLUENT TO HP-405 PFD-61713-A402 — 471
- H-411B QAH411
- 471C
- T-480
- 503 — METHANOL TO H-504B PFD-61713-A501
- P-480
- 470TOAGR — COMPRESSED SYNGAS TO S-410 PFD-61713-A403
- 478C — HYDROGEN TO HP-201/HP-306 PFD-61713-A1402
- STEAM TO T-605 PFD-61713-A603
- R-490
- BOILER FEEDWATER FROM HP-204 PFD-61713-A202

Drawing notes (left margin):
Plot date: 1/29/2015
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Title block:

Harris Group	NREL NATIONAL RENEWABLE ENERGY LABORATORY	THERMOCHEMICAL CONVERSION OF BIOMASS TO FUELS AND CHEMICALS GASIFICATION

PROCESS FLOW DIAGRAM
AREA 400: METHANOL SYNTHESIS

Rev.	Date	By	Description
1	12/5/14	DMS	ADDED HX NETWORK
0	9/26/14	JAR	ISSUED FOR DESIGN REPORT

Project No: 30482.00
Drawing: PFD-61713-A401
Rev: 1

PFD-61713-A401

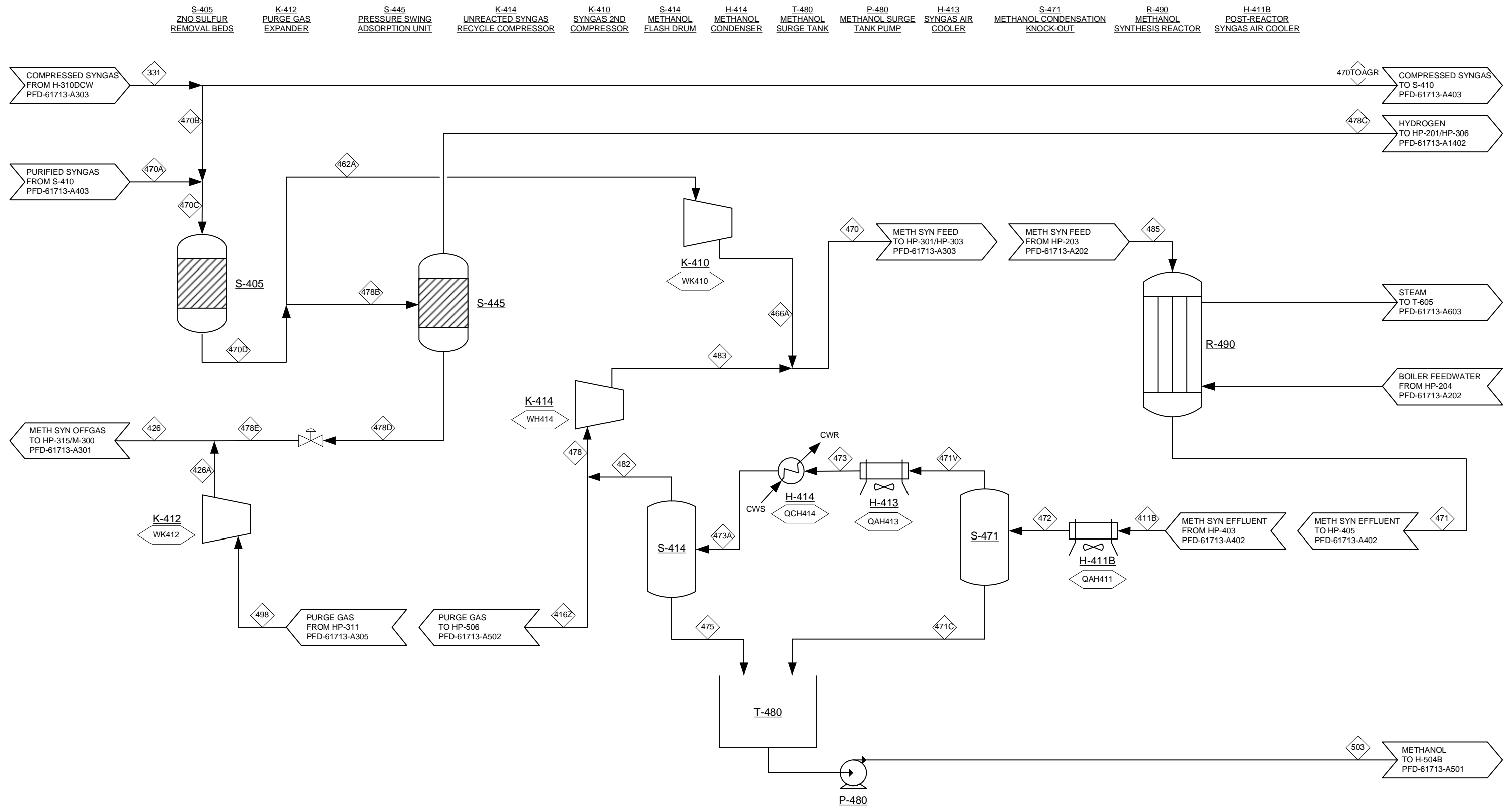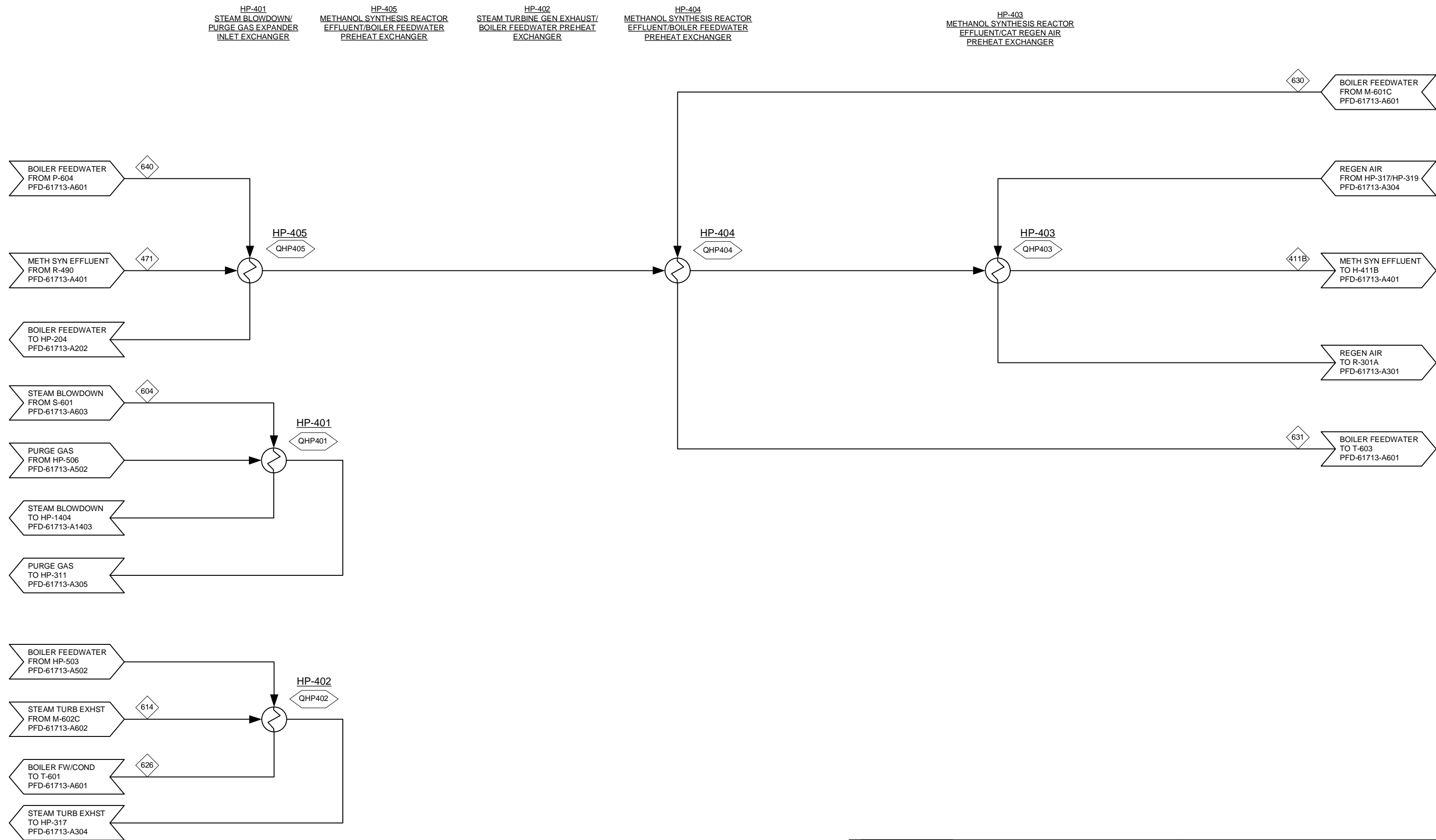| COMPONENT	UNITS	331	470TOAGR	470A	470B	470C	470D	478B	478C	478D	478E	426	462A	466A	470	485	471	411B	472	471V	473	473A	482	478	416Z	483	498	475	471C	503
Total Flow	lb/hr	190,834	178,068	99,659	12,767	112,425	112,401	6,938	698	6,240	6,240	19,113	105,463	105,402	349,903	349,903	349,903	311,834	311,834	311,834	257,370	244,501	12,873	244,501	12,873	244,501	12,873	54,465	38,069	92,528
Temperature	F	110	112	110	112	110	110	110	110	110	98	507	110	185	173	437	482	225	200	200	140	110	110	108	168	1,200	110	200	148	148
Pressure	Psia	430.0	425.0	415.0	425.0	415.0	410.0	410.0	410.0	410.0	135.0	33.7	410.0	735.0	735.0	730.0	580.0	575.0	574.0	573.0	572.0	567.0	565.0	565.0	424.5	735.0	419.5	565.0	573.0	565.0
Vapor Fraction		1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	-	-	-
Hydrogen (H2)	lb/hr	13,303	12,413	12,413	890	13,303	13,303	821	698	123	123	1,236	12,482	12,482	33,613	33,613	22,254	22,254	22,254	22,248	22,248	22,248	22,243	21,131	1,113	21,131	1,113	5	6	11
Carbon Monoxide (CO)	lb/hr	87,728	81,859	81,859	5,869	87,728	87,728	5,415	-	5,415	5,415	10,011	82,313	82,313	169,605	169,605	91,963	91,963	91,963	91,923	91,923	91,923	91,886	87,292	4,596	87,292	4,596	37	40	75
Nitrogen (N2)	lb/hr	404	377	377	27	404	404	25	-	25	25	393	379	379	7,377	7,377	7,377	7,377	7,377	7,371	7,371	7,371	7,365	6,997	368	6,997	368	6	5	11
Oxygen (O2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Argon (Ar)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Carbon Dioxide (CO2)	lb/hr	83,531	77,943	779	5,588	6,368	6,368	393	-	393	393	3,393	5,974	5,974	62,958	62,958	61,623	61,623	61,623	61,133	61,133	61,133	59,982	56,983	3,000	56,983	3,000	1,151	490	1,639
Water (H2O)	lb/hr	1,686	1,573	456	113	569	569	35	-	35	35	36	534	472	490	490	1,036	1,036	1,036	410	410	410	18	18	1	18	1	392	626	1,018
Hydrogen Sulfide (H2S)	lb/hr	150	140	14	10	24	-	-	-	-	-	0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	0	-	0
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	29	27	25	2	27	27	2	-	2	2	6	25	25	109	109	109	109	109	104	104	104	88	84	4	84	4	16	5	21
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Methane (CH4)	lb/hr	3,183	2,970	2,970	213	3,183	3,183	196	-	196	196	2,989	2,986	2,986	56,032	56,032	56,032	56,032	56,032	55,946	55,946	55,946	55,838	53,046	2,793	53,046	2,793	108	86	193
Ethane (C2H6)	lb/hr	5	5	5	0	5	5	0	-	0	0	4	5	5	76	76	76	76	76	75	75	75	74	71	4	71	4	1	0	1
Ethylene (C2H4)	lb/hr	747	697	697	50	747	747	46	-	46	46	591	701	701	11,051	11,051	11,051	11,051	11,051	10,993	10,993	10,993	10,895	10,350	545	10,350	545	98	58	156
Acetylene (C2H2)	lb/hr	65	60	60	4	65	65	4	-	4	4	43	61	61	796	796	796	796	796	789	789	789	774	735	39	735	39	15	7	22
Propane (C3H8)	lb/hr	0	0	0	0	0	0	0	-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Propene (C3H6)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
I-Butane (C4H10)	lb/hr	0	0	0	0	0	0	0	-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N-Butane (C4H10)	lb/hr	0	0	0	0	0	0	0	-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Butenes (C4H8)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Pentanes (C5H12)	lb/hr	0	0	0	0	0	0	0	-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Pentenes (C5H10)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	3	3	3	0	3	3	0	-	0	0	3	3	3	3	3	3	3	3	2	2	2	0	0	0	0	0	1	1	3
Heptanes (C7H16)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Triptane	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Triptene	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Octanes (C8H18)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Octenes (C8H16)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
N-Nonanes (C9H20)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Tar (C10H8)	lb/hr	0	0	0	0	0	0	0	-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbon (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Methanol (CH4O)	lb/hr	0	0	0	0	0	0	0	-	0	0	411	0	0	7,794	7,794	97,583	97,583	97,583	60,840	60,840	60,840	8,204	7,794	411	7,794	411	52,635	36,744	89,378
Dimethyl ether (C2H6O)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ash	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Char	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Wood	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Enthalpy Flow	MMBTU/hr	(484)	(452)	(148)	(32)	(180)	(180)	(11)	0	(11)	(11)	(32)	(169)	(163)	(635)	(576)	(686)	(739)	(761)	(640)	(672)	(683)	(506)	(480)	(25)	(472)	(16)	(177)	(122)	(299)

HP-401
STEAM BLOWDOWN/
PURGE GAS EXPANDER
INLET EXCHANGER

HP-405
METHANOL SYNTHESIS REACTOR
EFFLUENT/BOILER FEEDWATER
PREHEAT EXCHANGER

HP-402
STEAM TURBINE GEN EXHAUST/
BOILER FEEDWATER PREHEAT
EXCHANGER

HP-404
METHANOL SYNTHESIS REACTOR
EFFLUENT/BOILER FEEDWATER
PREHEAT EXCHANGER

HP-403
METHANOL SYNTHESIS REACTOR
EFFLUENT/CAT REGEN AIR
PREHEAT EXCHANGER

630 BOILER FEEDWATER
FROM M-601C
PFD-61713-A601

BOILER FEEDWATER
FROM P-604
PFD-61713-A601 640

REGEN AIR
FROM HP-317/HP-319
PFD-61713-A304

HP-405
QHP405

HP-404
QHP404

HP-403
QHP403

METH SYN EFFLUENT
FROM R-490
PFD-61713-A401 471

411B METH SYN EFFLUENT
TO H-411B
PFD-61713-A401

BOILER FEEDWATER
TO HP-204
PFD-61713-A202

REGEN AIR
TO R-301A
PFD-61713-A301

STEAM BLOWDOWN
FROM S-601
PFD-61713-A603 604

HP-401
QHP401

631 BOILER FEEDWATER
TO T-603
PFD-61713-A601

PURGE GAS
FROM HP-506
PFD-61713-A502

STEAM BLOWDOWN
TO HP-1404
PFD-61713-A1403

PURGE GAS
TO HP-311
PFD-61713-A305

BOILER FEEDWATER
FROM HP-503
PFD-61713-A502

HP-402
QHP402

STEAM TURB EXHST
FROM M-602C
PFD-61713-A602 614

BOILER FW/COND
TO T-601
PFD-61713-A601 626

STEAM TURB EXHST
TO HP-317
PFD-61713-A304

				THERMOCHEMICAL CONVERSION OF BIOMASS TO FUELS AND CHEMICALS GASIFICATION		
				PROCESS FLOW DIAGRAM AREA 400: A400 HEAT INTEGRATION		
0	12/5/14	DMS	ADDED HX NETWORK	Project No: 30482.00	Drawing PFD-61713-A402	Rev. 0
Rev.	Date	By	Description			

Harris Group

NREL
NATIONAL RENEWABLE ENERGY LABORATORY

PFD-61713-A402

COMPONENT	UNITS	640	471	604	614	626	630	411B	631
Total Flow	lb/hr	410,305	349,903	4,088	177,601	415,415	410,302	349,903	410,302
Temperature	F	233	482	462	472	237	226	225	232
Pressure	Psia	485.9	580.0	475.9	3.5	19.1	24.7	575.0	21.7
Vapor Fraction		-	1	-	1	-	-	1	-
Hydrogen (H2)	lb/hr	-	22,254	-	-	-	-	22,254	-
Carbon Monoxide (CO)	lb/hr	-	91,963	-	-	-	-	91,963	-
Nitrogen (N2)	lb/hr	-	7,377	-	-	-	-	7,377	-
Oxygen (O2)	lb/hr	-	-	-	-	-	-	-	-
Argon (Ar)	lb/hr	-	-	-	-	-	-	-	-
Carbon Dioxide (CO2)	lb/hr	-	61,623	-	-	-	-	61,623	-
Water (H2O)	lb/hr	410,305	1,036	4,088	177,601	415,415	410,302	1,036	410,302
Hydrogen Sulfide (H2S)	lb/hr	-	-	-	-	-	-	-	-
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	-	109	-	-	-	-	109	-
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-	-	-	-
Methane (CH4)	lb/hr	-	56,032	-	-	-	-	56,032	-
Ethane (C2H6)	lb/hr	-	76	-	-	-	-	76	-
Ethylene (C2H4)	lb/hr	-	11,051	-	-	-	-	11,051	-
Acetylene (C2H2)	lb/hr	-	796	-	-	-	-	796	-
Propane (C3H8)	lb/hr	-	0	-	-	-	-	0	-
Propene (C3H6)	lb/hr	-	-	-	-	-	-	-	-
I-Butane (C4H10)	lb/hr	-	0	-	-	-	-	0	-
N-Butane (C4H10)	lb/hr	-	0	-	-	-	-	0	-
Butenes (C4H8)	lb/hr	-	-	-	-	-	-	-	-
Pentanes (C5H12)	lb/hr	-	0	-	-	-	-	0	-
Pentenes (C5H10)	lb/hr	-	-	-	-	-	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-	-	-	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	-	3	-	-	-	-	3	-
Heptanes (C7H16)	lb/hr	-	-	-	-	-	-	-	-
Triptane	lb/hr	-	-	-	-	-	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-	-	-	-	-	-
Triptene	lb/hr	-	-	-	-	-	-	-	-
Octanes (C8H18)	lb/hr	-	-	-	-	-	-	-	-
Octenes (C8H16)	lb/hr	-	-	-	-	-	-	-	-
N-Nonanes (C9H20)	lb/hr	-	-	-	-	-	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-	-	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-	-	-	-
Tar (C10H8)	lb/hr	-	0	-	-	-	-	0	-
Carbon (Solid)	lb/hr	-	-	-	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	-	-	-	-	-	-
Methanol (CH4O)	lb/hr	-	97,583	-	-	-	-	97,583	-
Dimethyl ether (C2H6O)	lb/hr	-	-	-	-	-	-	-	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-	-	-	-
Ash	lb/hr	-	-	-	-	-	-	-	-
Char	lb/hr	-	-	-	-	-	-	-	-
Wood	lb/hr	-	-	-	-	-	-	-	-
Enthalpy Flow	MMBTU/hr	(2,736)	(686)	(26)	(994)	(2,769)	(2,739)	(739)	(2,737)

S-410
AMINE ACID GAS
REMOVAL SYSTEM

AMINE MAKEUP

470A → PURIFIED SYNGAS
TO S-405
PFD-61713-A401

FLASH DRUM

ABSORBER

432 → WATER MAKE-UP
TO T-302
PFD-61713-A302

CONDENSATE PUMP

419 → ACID GAS
TO HP-507
PFD-61713-A502

OVERHEAD CONDENSER

REFLUX DRUM

COMPRESSED SYNGAS
FROM
PFD-61713-A401

470TOAGR →

LEAN SURGE VESSEL

LEAN COOLER

RICH/LEAN EXCHANGER

REGENERATOR

REFLUX PUMP

LEAN AMINE PUMP

450 → STEAM
FROM M-603/M-602B
PFD-61713-A602

REBOILER

LEAN AMINE PUMP

CONDENSATE
TO
PFD-61713-A601

FLASH DRUM

OFF-GAS

RICH AMINE PUMP

S-410

① QAGR

WAGR

① UTILITY REQUIREMENTS SHOWN FOR S-410 ALSO INCLUDE UTILITY REQUIREMENTS FOR M-405

Harris Group

NREL
NATIONAL RENEWABLE ENERGY LABORATORY

THERMOCHEMICAL CONVERSION OF BIOMASS
TO FUELS AND CHEMICALS
GASIFICATION

PROCESS FLOW DIAGRAM
AREA 400: ACID GAS REMOVAL

Rev.	Date	By	Description
2	12/12/14	JAR	UPDATED EQ TITLES
1	12/5/14	DMS	ADDED HX NETWORK
0	9/26/14	JAR	ISSUED FOR DESIGN REPORT

Project No: 30482.00   Drawing: PFD-61713-A403   Rev: 2

Plot date: 1/29/2015   J:\1A - Process Solutions\30482 NREL Thermochemical Conv - Design-Model\Drawings\ 30482 Gasification PFDs 61713 150128.vsd
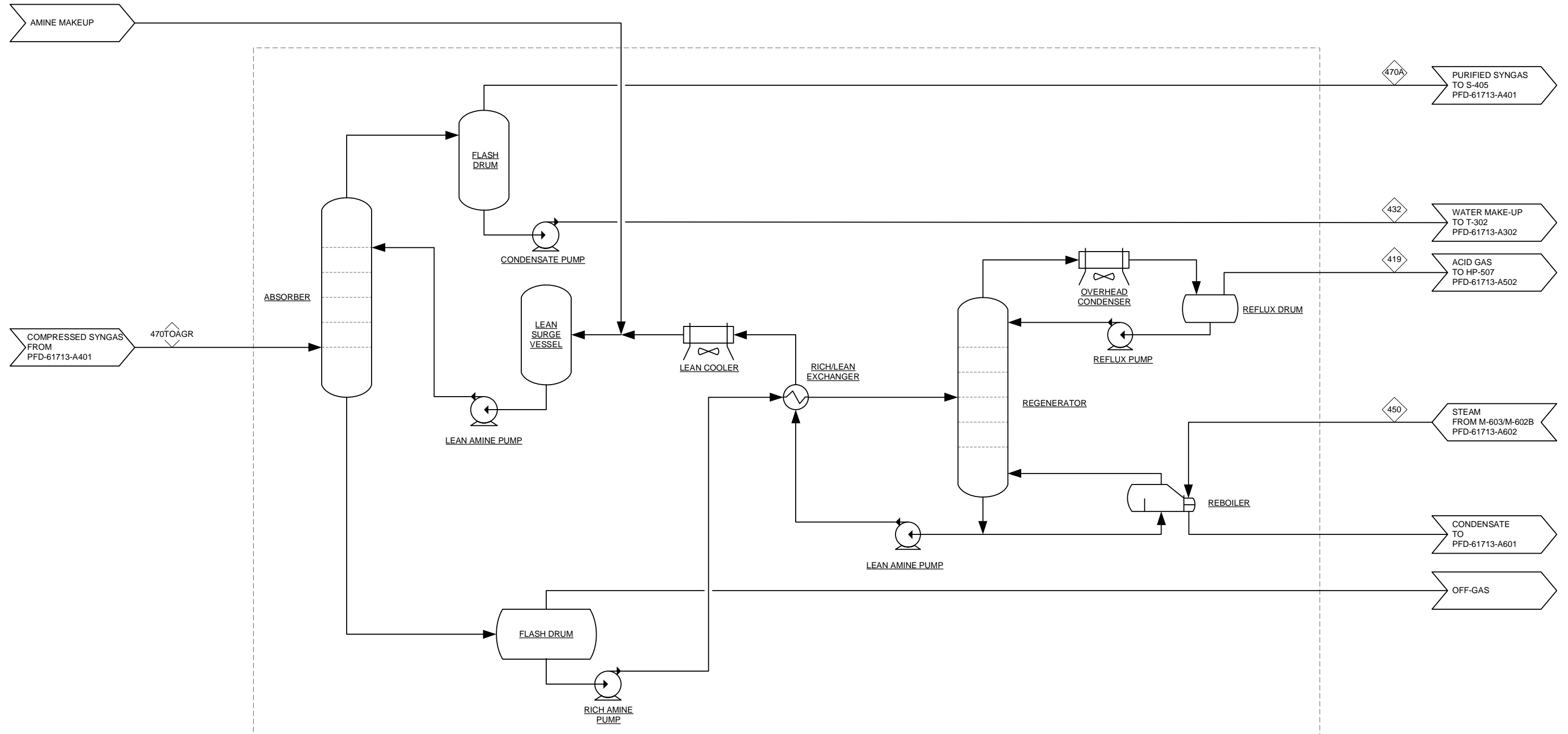
PFD-61713-A403

COMPONENT	UNITS	470TOAGR	470A	432	419	450
Total Flow	lb/hr	178,068	99,659	1,015	77,290	153,812
Temperature	F	112	110	110	110	395
Pressure	Psia	425.0	415.0	420.0	415.0	64.7
Vapor Fraction		1	1	-	1	1
Hydrogen (H2)	lb/hr	12,413	12,413	0	-	-
Carbon Monoxide (CO)	lb/hr	81,859	81,859	0	-	-
Nitrogen (N2)	lb/hr	377	377	0	-	-
Oxygen (O2)	lb/hr	-	-	-	-	-
Argon (Ar)	lb/hr	-	-	-	-	-
Carbon Dioxide (CO2)	lb/hr	77,943	779	0	77,163	-
Water (H2O)	lb/hr	1,573	456	1,013	-	153,812
Hydrogen Sulfide (H2S)	lb/hr	140	14	0	127	-
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-
Ammonia (NH3)	lb/hr	27	25	2	-	-
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-
Methane (CH4)	lb/hr	2,970	2,970	0	-	-
Ethane (C2H6)	lb/hr	5	5	0	-	-
Ethylene (C2H4)	lb/hr	697	697	0	-	-
Acetylene (C2H2)	lb/hr	60	60	0	-	-
Propane (C3H8)	lb/hr	0	0	0	-	-
Propene (C3H6)	lb/hr	-	-	-	-	-
I-Butane (C4H10)	lb/hr	0	0	0	-	-
N-Butane (C4H10)	lb/hr	0	0	0	-	-
Butenes (C4H8)	lb/hr	-	-	-	-	-
Pentanes (C5H12)	lb/hr	0	0	0	-	-
Pentenes (C5H10)	lb/hr	-	-	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-	-	-
Benzene (C6H6)	lb/hr	3	3	0	-	-
Heptanes (C7H16)	lb/hr	-	-	-	-	-
Triptane	lb/hr	-	-	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-	-	-
Triptene	lb/hr	-	-	-	-	-
Octanes (C8H18)	lb/hr	-	-	-	-	-
Octenes (C8H16)	lb/hr	-	-	-	-	-
N-Nonanes (C9H20)	lb/hr	-	-	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-
Tar (C10H8)	lb/hr	0	0	0	-	-
Carbon (Solid)	lb/hr	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	-	-	-
Methanol (CH4O)	lb/hr	0	0	0	-	-
Dimethyl ether (C2H6O)	lb/hr	-	-	-	-	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-
Ash	lb/hr	-	-	-	-	-
Char	lb/hr	-	-	-	-	-
Wood	lb/hr	-	-	-	-	-
Enthalpy Flow	MMBTU/hr	(452)	(148)	(7)	(297)	(868)

100

M-405
LO-CAT SULFUR
RECOVERY SYSTEM

484VENT

CO2
TO K-102
PFD-61713-A102

ACID GAS
FROM HP-507
PFD-61713-A502

420

LO-CAT VENTURI
PRECONTRACTOR

LO-CAT LIQUID FILLED
ABSORBER

ATM

LO-CAT OXIDIZER
VESSEL

LO-CAT ABSORBENT
SOLUTION
CIRCULATING PUMP

LO-CAT ABSORBENT
SOLUTION COOLER

424

SULFUR TO STORAGE

AIR

422

LO-CAT FEED AIR
BLOWER
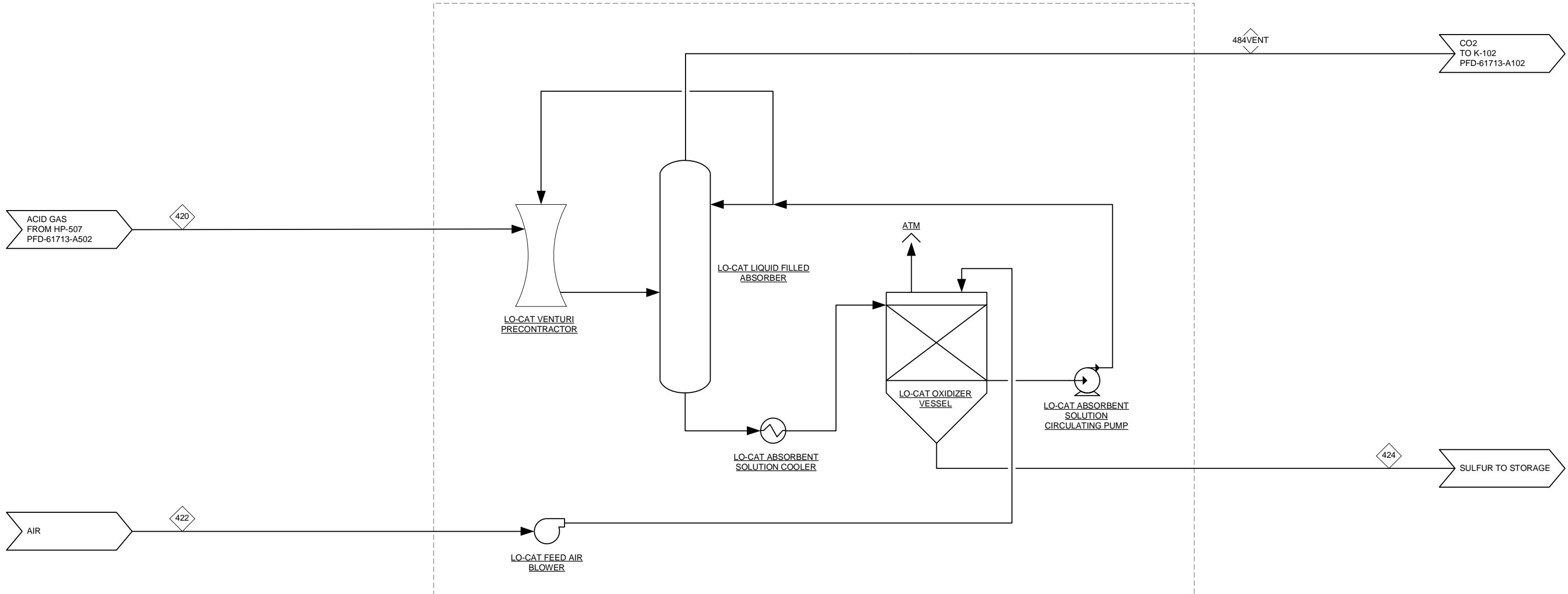
M-405　①

①　UTILITY REQUIREMENTS FOR M-405 INCLUDED IN UTILITY
REQUIREMENTS SHOWN FOR S-410

PFD-61713-A404

COMPONENT	UNITS	420	422	484VENT	424
Total Flow	lb/hr	77,290	287	77,164	119
Temperature	F	149	90	148	-
Pressure	Psia	20.7	14.7	15.7	14.7
Vapor Fraction		1	1	1	-
Hydrogen (H2)	lb/hr	-	-	-	-
Carbon Monoxide (CO)	lb/hr	-	-	-	-
Nitrogen (N2)	lb/hr	-	212	-	-
Oxygen (O2)	lb/hr	-	65	-	-
Argon (Ar)	lb/hr	-	4	-	-
Carbon Dioxide (CO2)	lb/hr	77,163	0	77,163	-
Water (H2O)	lb/hr	-	6	-	-
Hydrogen Sulfide (H2S)	lb/hr	127	-	1	-
Sulfur dioxide (SO2)	lb/hr	-	-	-	-
Ammonia (NH3)	lb/hr	-	-	-	-
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-
Methane (CH4)	lb/hr	-	-	-	-
Ethane (C2H6)	lb/hr	-	-	-	-
Ethylene (C2H4)	lb/hr	-	-	-	-
Acetylene (C2H2)	lb/hr	-	-	-	-
Propane (C3H8)	lb/hr	-	-	-	-
Propene (C3H6)	lb/hr	-	-	-	-
I-Butane (C4H10)	lb/hr	-	-	-	-
N-Butane (C4H10)	lb/hr	-	-	-	-
Butenes (C4H8)	lb/hr	-	-	-	-
Pentanes (C5H12)	lb/hr	-	-	-	-
Pentenes (C5H10)	lb/hr	-	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-	-
Benzene (C6H6)	lb/hr	-	-	-	-
Heptanes (C7H16)	lb/hr	-	-	-	-
Triptane	lb/hr	-	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-	-
Triptene	lb/hr	-	-	-	-
Octanes (C8H18)	lb/hr	-	-	-	-
Octenes (C8H16)	lb/hr	-	-	-	-
N-Nonanes (C9H20)	lb/hr	-	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-
Tar (C10H8)	lb/hr	-	-	-	-
Carbon (Solid)	lb/hr	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	119
Olivine + MgO (Solid)	lb/hr	-	-	-	-
Methanol (CH4O)	lb/hr	-	-	-	-
Dimethyl ether (C2H6O)	lb/hr	-	-	-	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-
Ash	lb/hr	-	-	-	-
Char	lb/hr	-	-	-	-
Wood	lb/hr	-	-	-	-
Enthalpy Flow	MMBTU/hr	(296)	(0)	(296)	0

102

H-504B
EXPANDED METHANOL
COOLER B

D-500
METHANOL
DEGASSING COLUMN

P-500B
D-500 BOTTOMS
PUMP

HP-502
DME REACTOR
EFFLUENT/METHANOL
DEGASSING COLUMN
REBOILER EXCHANGER

HP-501
CHAR COMBUSTOR
FLUE GAS/METHANOL
DEGASSING COLUMN
REBOILER EXCHANGER

| 513 | METH COL OFFGAS TO M-300 PFD-61713-A301 |

| 1470 | DME FROM R-1424 PFD-61713-A1401 |

METHANOL
FROM P-480
PFD-61713-A401 — 503 — CWR — CWS — H-504B — QCH504B — 505 — D-500

HP-502
QHP502

| 1470A | DME TO H-1471 PFD-61713-A1402 |

| | FLUE GAS TO HP-1503 PFD-61713-A1501 |

HP-501
QHP501

| 505C | METHANOL TO HP-503 PFD-61713-A502 |

P-500B

FLUE GAS
FROM HP-202/HP-203/HP-204
PFD-61713-A202

			THERMOCHEMICAL CONVERSION OF BIOMASS TO FUELS AND CHEMICALS GASIFICATION
		NREL NATIONAL RENEWABLE ENERGY LABORATORY	
			PROCESS FLOW DIAGRAM AREA 500: METHANOL CONDITIONING
1	12/5/14	DMS	ADDED HX NETWORK
0	9/26/14	JAR	ISSUED FOR DESIGN REPORT
Rev.	Date	By	Description

Harris Group

Project No: 30482.00   Drawing: PFD-61713-A501   Rev: 1

PFD-61713-A501

COMPONENT	UNITS	503	505	513	505C	1,470	1470A
Total Flow	lb/hr	92,528	92,528	4,819	87,710	87,710	87,710
Temperature	F	148	110	217	253	482	296
Pressure	Psia	565.0	560.0	98.0	98.0	140.0	135.0
Vapor Fraction		-	0	1	-	1	1
Hydrogen (H2)	lb/hr	11	11	11	-	-	-
Carbon Monoxide (CO)	lb/hr	75	75	75	-	-	-
Nitrogen (N2)	lb/hr	11	11	11	-	-	-
Oxygen (O2)	lb/hr	-	-	-	-	-	-
Argon (Ar)	lb/hr	-	-	-	-	-	-
Carbon Dioxide (CO2)	lb/hr	1,639	1,639	1,639	0	-	-
Water (H2O)	lb/hr	1,018	1,018	8	1,010	22,569	22,569
Hydrogen Sulfide (H2S)	lb/hr	-	-	-	-	-	-
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	21	21	21	0	0	0
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-	-
Methane (CH4)	lb/hr	193	193	193	0	-	-
Ethane (C2H6)	lb/hr	1	1	1	0	0	0
Ethylene (C2H4)	lb/hr	156	156	156	0	0	0
Acetylene (C2H2)	lb/hr	22	22	22	0	0	0
Propane (C3H8)	lb/hr	0	0	0	0	0	0
Propene (C3H6)	lb/hr	-	-	-	-	-	-
I-Butane (C4H10)	lb/hr	0	0	0	0	0	0
N-Butane (C4H10)	lb/hr	0	0	0	0	0	0
Butenes (C4H8)	lb/hr	-	-	-	-	-	-
Pentanes (C5H12)	lb/hr	0	0	0	0	0	0
Pentenes (C5H10)	lb/hr	-	-	-	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	3	3	0	3	3	3
Heptanes (C7H16)	lb/hr	-	-	-	-	-	-
Triptane	lb/hr	-	-	-	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-	-	-	-
Triptene	lb/hr	-	-	-	-	-	-
Octanes (C8H18)	lb/hr	-	-	-	-	-	-
Octenes (C8H16)	lb/hr	-	-	-	-	-	-
N-Nonanes (C9H20)	lb/hr	-	-	-	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-	-
Tar (C10H8)	lb/hr	0	0	0	0	0	0
Carbon (Solid)	lb/hr	-	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	-	-	-	-
Methanol (CH4O)	lb/hr	89,378	89,378	2,681	86,697	10,008	10,008
Dimethyl ether (C2H6O)	lb/hr	-	-	-	-	55,131	55,131
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-	-
Ash	lb/hr	-	-	-	-	-	-
Char	lb/hr	-	-	-	-	-	-
Wood	lb/hr	-	-	-	-	-	-
Enthalpy Flow	MMBTU/hr	(299)	(302)	(14)	(271)	(237)	(245)

104

Process Flow Diagram — Area 500: A500 Heat Integration

HP-503
METHANOL INTERMEDIATE/BOILER FEEDWATER PREHEAT EXCHANGER

HP-504
METHANOL INTERMEDIATE/LIGHT HC RECYCLE EXCHANGER

HP-505
METHANOL INTERMEDIATE/ CAT REGEN AIR PREHEAT EXCHANGER

HP-506
METHANOL INTERMEDIATE/ PURGE GAS EXPANDER INLET EXCHANGER

HP-507
METHANOL INTERMEDIATE/ AGR ACID GAS EXCHANGER

H-593
METHANOL PRODUCT FINISHING COOLER

LIGHT HC
FROM P-1555
PFD-61713-A1502

1510A

BOILER FEEDWATER
FROM HP-302/HP-304
PFD-61713-A303

BOILER FEEDWATER
FROM HP-316
PFD-61713-A304

BOILER FEEDWATER
FROM HP-318
PFD-61713-A304

METHANOL
FROM P-500B
PFD-61713-A501

505C

BOILER FEEDWATER
TO HP-402
PFD-61713-A402

LIGHT HC
TO HP-1407
PFD-61713-A1404

HP-503
QHP503

HP-504
QHP504

HP-505
QHP505

HP-506
QHP506

HP-507
QHP507

H-593
QH593

CWR

CWS

REGEN AIR
FROM K-305
PFD-61713-A301

REGEN AIR
TO HP-305
PFD-61713-A304

PURGE GAS
FROM S-414
PFD-61713-A401

416Z

METHANOL
TO T-1401
PFD-61713-A1401

592

PURGE GAS
TO HP-401
PFD-61713-A402

ACID GAS
FROM S-410
PFD-61713-A403

419

ACID GAS
TO M-405
PFD-61713-A404

420
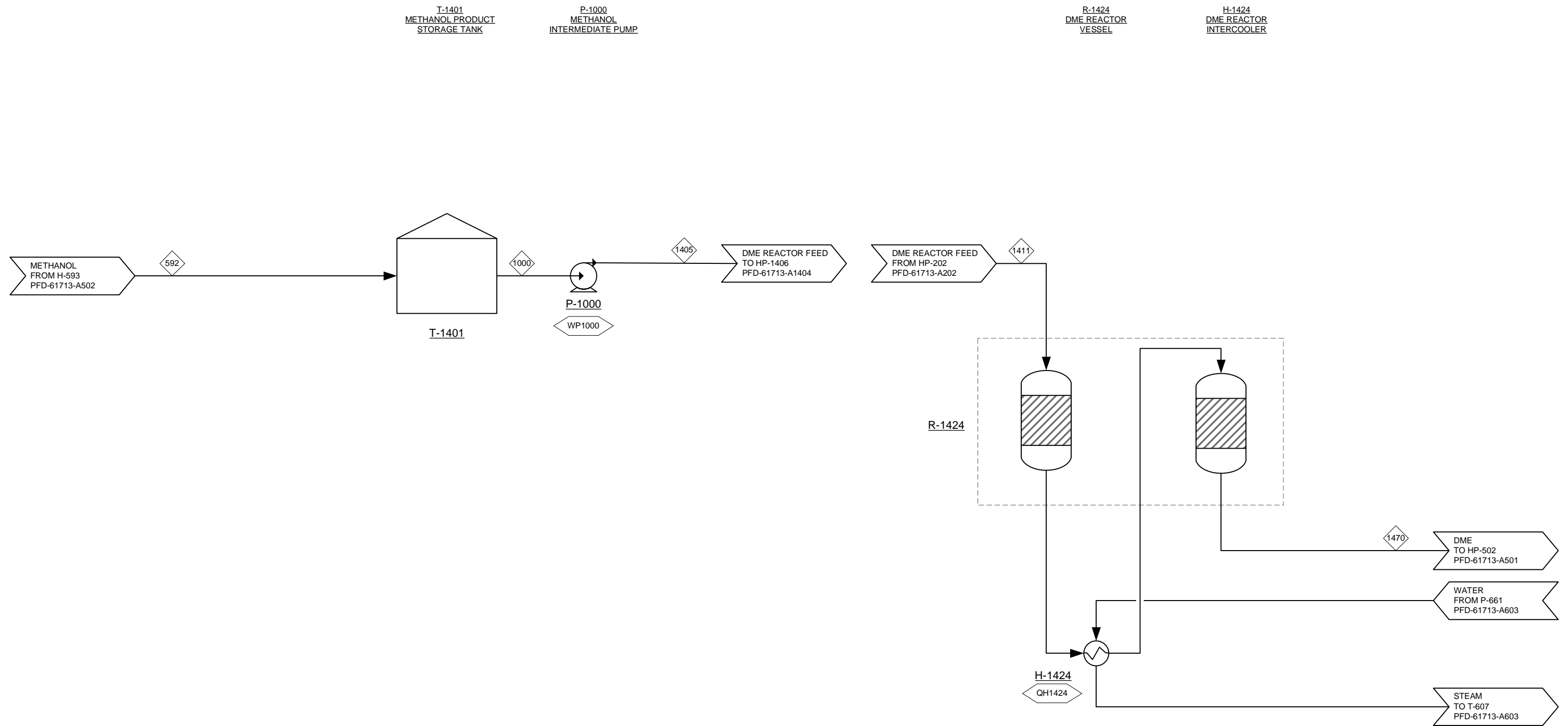
THERMOCHEMICAL CONVERSION OF BIOMASS TO FUELS AND CHEMICALS GASIFICATION

PROCESS FLOW DIAGRAM
AREA 500: A500 HEAT INTEGRATION

Harris Group

Rev.	Date	By	Description
0	12/5/14	DMS	ADDED HX NETWORK

Project No: 30482.00

Drawing: PFD-61713-A502

Rev: 0

PFD-61713-A502

COMPONENT	UNITS	1510A	505C	416Z	592	419	420
Total Flow	lb/hr	7,040	87,710	12,873	87,710	77,290	77,290
Temperature	F	128	253	108	145	110	149
Pressure	Psia	152.0	98.0	424.5	14.7	415.0	20.7
Vapor Fraction		-	-	1	-	1	1
Hydrogen (H2)	lb/hr	-	-	1,113	-	-	-
Carbon Monoxide (CO)	lb/hr	-	-	4,596	-	-	-
Nitrogen (N2)	lb/hr	-	-	368	-	-	-
Oxygen (O2)	lb/hr	-	-	-	-	-	-
Argon (Ar)	lb/hr	-	-	-	-	-	-
Carbon Dioxide (CO2)	lb/hr	-	0	3,000	0	77,163	77,163
Water (H2O)	lb/hr	44	1,010	1	1,010	-	-
Hydrogen Sulfide (H2S)	lb/hr	-	-	0	-	127	127
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	-	0	4	0	-	-
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-	-
Methane (CH4)	lb/hr	-	0	2,793	0	-	-
Ethane (C2H6)	lb/hr	-	0	4	0	-	-
Ethylene (C2H4)	lb/hr	-	0	545	0	-	-
Acetylene (C2H2)	lb/hr	-	0	39	0	-	-
Propane (C3H8)	lb/hr	0	0	0	0	-	-
Propene (C3H6)	lb/hr	0	-	-	-	-	-
I-Butane (C4H10)	lb/hr	6,662	0	0	0	-	-
N-Butane (C4H10)	lb/hr	106	0	0	0	-	-
Butenes (C4H8)	lb/hr	186	-	-	-	-	-
Pentanes (C5H12)	lb/hr	3	0	0	0	-	-
Pentenes (C5H10)	lb/hr	0	-	-	-	-	-
Hexanes (C6H14)	lb/hr	0	-	-	-	-	-
Hexenes (C6H12)	lb/hr	0	-	-	-	-	-
Benzene (C6H6)	lb/hr	0	3	0	3	-	-
Heptanes (C7H16)	lb/hr	0	-	-	-	-	-
Triptane	lb/hr	0	-	-	-	-	-
Heptenes (C7H14)	lb/hr	0	-	-	-	-	-
Triptene	lb/hr	0	-	-	-	-	-
Octanes (C8H18)	lb/hr	0	-	-	-	-	-
Octenes (C8H16)	lb/hr	0	-	-	-	-	-
N-Nonanes (C9H20)	lb/hr	0	-	-	-	-	-
N-Nonenes (C9H18)	lb/hr	0	-	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-	-
Tar (C10H8)	lb/hr	-	0	0	0	-	-
Carbon (Solid)	lb/hr	-	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	-	-	-	-
Methanol (CH4O)	lb/hr	-	86,697	411	86,697	-	-
Dimethyl ether (C2H6O)	lb/hr	38	-	-	-	-	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-	-
Ash	lb/hr	-	-	-	-	-	-
Char	lb/hr	-	-	-	-	-	-
Wood	lb/hr	-	-	-	-	-	-
Enthalpy Flow	MMBTU/hr	(8)	(271)	(25)	(280)	(297)	(296)

106

T-1401
METHANOL PRODUCT
STORAGE TANK

P-1000
METHANOL
INTERMEDIATE PUMP

R-1424
DME REACTOR
VESSEL

H-1424
DME REACTOR
INTERCOOLER

METHANOL
FROM H-593
PFD-61713-A502

⟨592⟩

T-1401

⟨1000⟩

P-1000

WP1000

⟨1405⟩

DME REACTOR FEED
TO HP-1406
PFD-61713-A1404

DME REACTOR FEED
FROM HP-202
PFD-61713-A202

⟨1411⟩

R-1424

⟨1470⟩

DME
TO HP-502
PFD-61713-A501

⟨1470⟩

WATER
FROM P-661
PFD-61713-A603

H-1424

QH1424

STEAM
TO T-607
PFD-61713-A603

Plot date: 1/29/2015   J:\1A - Process Solutions\30482 NREL Thermochemical Conv - Design-Model\Drawings\ 30482 Gasification PFDs 61713 150128.vsd

			THERMOCHEMICAL CONVERSION OF BIOMASS TO FUELS AND CHEMICALS GASIFICATION		
			PROCESS FLOW DIAGRAM AREA 1400: HP DME REFORMER		
1	12/5/14	DMS	ADDED HX NETWORK	Project No:	Drawing
0	9/26/14	JAR	ISSUED FOR DESIGN REPORT	30482.00	PFD-61713-A1401	Rev: 1
Rev.	Date	By	Description		

NREL
NATIONAL RENEWABLE ENERGY LABORATORY

Harris Group

PFD-61713-A1401

COMPONENT	UNITS	592	1000	1405	1411	1470
Total Flow	lb/hr	87,710	87,710	87,710	87,710	87,710
Temperature	F	145	150	150	437	482
Pressure	Psia	14.7	15.0	150.0	145.0	140.0
Vapor Fraction		-	-	-	1	1
Hydrogen (H2)	lb/hr	-	-	-	-	-
Carbon Monoxide (CO)	lb/hr	-	-	-	-	-
Nitrogen (N2)	lb/hr	-	-	-	-	-
Oxygen (O2)	lb/hr	-	-	-	-	-
Argon (Ar)	lb/hr	-	-	-	-	-
Carbon Dioxide (CO2)	lb/hr	0	0	0	0	-
Water (H2O)	lb/hr	1,010	1,010	1,010	1,010	22,569
Hydrogen Sulfide (H2S)	lb/hr	-	-	-	-	-
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-
Ammonia (NH3)	lb/hr	0	0	0	0	0
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-
Methane (CH4)	lb/hr	0	-	-	-	-
Ethane (C2H6)	lb/hr	0	0	0	0	0
Ethylene (C2H4)	lb/hr	0	0	0	0	0
Acetylene (C2H2)	lb/hr	0	0	0	0	0
Propane (C3H8)	lb/hr	0	0	0	0	0
Propene (C3H6)	lb/hr	-	-	-	-	-
I-Butane (C4H10)	lb/hr	0	0	0	0	0
N-Butane (C4H10)	lb/hr	0	0	0	0	0
Butenes (C4H8)	lb/hr	-	-	-	-	-
Pentanes (C5H12)	lb/hr	0	0	0	0	0
Pentenes (C5H10)	lb/hr	-	-	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-	-	-
Benzene (C6H6)	lb/hr	3	3	3	3	3
Heptanes (C7H16)	lb/hr	-	-	-	-	-
Triptane	lb/hr	-	-	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-	-	-
Triptene	lb/hr	-	-	-	-	-
Octanes (C8H18)	lb/hr	-	-	-	-	-
Octenes (C8H16)	lb/hr	-	-	-	-	-
N-Nonanes (C9H20)	lb/hr	-	-	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-
Tar (C10H8)	lb/hr	0	0	0	0	0
Carbon (Solid)	lb/hr	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	-	-	-
Methanol (CH4O)	lb/hr	86,697	86,697	86,697	86,697	10,008
Dimethyl ether (C2H6O)	lb/hr	-	-	-	-	55,131
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-
Ash	lb/hr	-	-	-	-	-
Char	lb/hr	-	-	-	-	-
Wood	lb/hr	-	-	-	-	-
Enthalpy Flow	MMBTU/hr	(280)	(280)	(280)	(229)	(237)

Equipment titles (top):

- K-1410 — COKE BURN/CAT REGENERATION AIR BLOWER
- H-1471 — POST DME REACTOR AIR COOLER
- S-1470 — DME REACTOR EFFLUENT KO DRUM
- P-1470 — CONDENSATE PUMP
- R-1410C — HYDROCARBON SYNTHESIS REACTOR VESSEL
- H-1410E — HYDROCARBON SYNTHESIS REACTOR 1ST STAGE INTERCOOLER
- H-1410F — HYDROCARBON SYNTHESIS REACTOR 2ND STAGE INTERCOOLER
- H-1410J — HYDROCARBON SYNTHESIS REACTOR 3RD STAGE INTERCOOLER
- H-1410C — HYDROCARBON SYNTEHSIS REACTOR EFFLUENT COOLER

Stream labels:

- DME FROM P-1503C PFD-61712-A1501
- HC OFFGAS FROM K-1472 PFD-61713-A1403
- HYDROGEN FROM S-445 PFD-61713-A401
- HC SYNTHESIS FEED TO HP-201 PFD-61713-A202
- HC SYNTHESIS FEED TO HP-306 PFD-61713-A304
- HC SYNTHESIS FEED FROM HP-201 PFD-61713-A202
- HC SYNTHESIS FEED FROM HP-306 PFD-61713-A304
- LIGHT HC RECYCLE FROM PFD-61713-A1404
- CAT REGEN FLUE GAS TO M-105 PFD-61713-A102
- DME FROM HP-502 PFD-61713-A501
- COMBUSTION AIR
- HYDROCARBONS TO HP-1402 PFD-61713-A1404
- WATER FROM P-661 PFD-61713-A603
- STEAM TO T-607 PFD-61713-A603
- CONDENSATE TO T-608 PFD-61713-A603

Stream tags: 1508A, 1474B, 1470B, 478C, 470V, 1470A, 1470A1, 1485, 1510D, FLUE-GAS, AIR1, AIR2, 1471, 1400WW3

Equipment tags: H-1471 QAH1471, S-1470, P-1470, K-1410 WK1410, H-1410E QH1410E, H-1410F QH1410F, H-1410J QH1410J, H-1410C QH1410C

NOTE: THERE ARE A TOTAL OF THREE (3) REACTOR TRAINS IN THE PROCESS CONFIGURATION. DURING NORMAL OPERATION, TWO (2) TRAINS ARE IN HC SYNTHESIS SERVICE AND ONE (1) TRAIN IS IN CATALYST REGENERATION. DURING CATALYST REGENERATION, H-1410C/E/F/J SERVE AS COOLERS FOR THE COMBUSTION FLUE GAS.

PFD-61713-A1402

COMPONENT	UNITS	1508A	1474B	478C	1470B	1470V	AIR1	AIR2	1470A	1470A1	1400WW3	1510D	1485	1471	FLUE-GAS
Total Flow	lb/hr	3,337	292,468	698	372,331	69,130	3,192	3,192	87,710	87,710	18,580	6,688	372,331	372,153	3,369
Temperature	F	110	168	110	179	267	90	153	296	250	267	173	350	375	230
Pressure	Psia	200.0	135.0	410.0	134.0	134.0	14.7	19.7	135.0	134.0	134.0	147.0	129.0	90.0	19.7
Vapor Fraction		-	1	1	1	1	1	1	1	1	-	1	1	1	1
Hydrogen (H2)	lb/hr	0	4	698	712	-	-	-	-	-	-	-	712	4	-
Carbon Monoxide (CO)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Nitrogen (N2)	lb/hr	-	-	-	-	-	2,364	2,364	-	-	-	-	-	-	2,364
Oxygen (O2)	lb/hr	-	-	-	-	-	724	724	-	-	-	-	-	-	147
Argon (Ar)	lb/hr	-	-	-	-	-	40	40	-	-	-	-	-	-	40
Carbon Dioxide (CO2)	lb/hr	-	-	-	-	-	1	1	-	-	-	-	-	-	579
Water (H2O)	lb/hr	0	2,499	-	6,530	3,989	62	62	22,569	22,569	18,580	42	6,530	32,989	240
Hydrogen Sulfide (H2S)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	0	0	-	0	0	-	-	0	0	-	-	0	0	-
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Methane (CH4)	lb/hr	10	35,874	-	35,884	-	-	-	-	-	-	-	35,884	36,306	-
Ethane (C2H6)	lb/hr	32	7,590	-	7,623	0	-	-	0	0	-	-	7,623	7,741	-
Ethylene (C2H4)	lb/hr	12	5,103	-	5,116	0	-	-	0	0	-	-	5,116	5,189	-
Acetylene (C2H2)	lb/hr	0	0	-	0	0	-	-	0	0	-	-	0	0	-
Propane (C3H8)	lb/hr	473	23,132	-	23,605	0	-	-	0	0	-	0	23,605	24,068	-
Propene (C3H6)	lb/hr	5	279	-	284	-	-	-	-	-	-	0	284	289	-
I-Butane (C4H10)	lb/hr	54	99,866	-	106,249	0	-	-	0	0	-	6,329	106,249	107,606	-
N-Butane (C4H10)	lb/hr	0	1,167	-	1,268	0	-	-	0	0	-	101	1,268	1,286	-
Butenes (C4H8)	lb/hr	0	2,412	-	2,590	-	-	-	-	-	-	177	2,590	2,624	-
Pentanes (C5H12)	lb/hr	0	4,827	-	4,830	0	-	-	0	0	-	3	4,830	5,910	-
Pentenes (C5H10)	lb/hr	0	3,236	-	3,237	-	-	-	-	-	-	0	3,237	4,028	-
Hexanes (C6H14)	lb/hr	0	3,017	-	3,017	-	-	-	-	-	-	0	3,017	4,840	-
Hexenes (C6H12)	lb/hr	0	587	-	587	-	-	-	-	-	-	0	587	1,015	-
Benzene (C6H6)	lb/hr	0	3	-	5	3	-	-	3	3	-	0	5	5	-
Heptanes (C7H16)	lb/hr	0	1,211	-	1,211	-	-	-	-	-	-	0	1,211	3,684	-
Triptane	lb/hr	0	8,429	-	8,429	-	-	-	-	-	-	0	8,429	18,139	-
Heptenes (C7H14)	lb/hr	0	838	-	838	-	-	-	-	-	-	0	838	2,587	-
Triptene	lb/hr	0	1,009	-	1,009	-	-	-	-	-	-	0	1,009	1,989	-
Octanes (C8H18)	lb/hr	0	3,699	-	3,699	-	-	-	-	-	-	0	3,699	18,563	-
Octenes (C8H16)	lb/hr	0	250	-	250	-	-	-	-	-	-	0	250	839	-
N-Nonanes (C9H20)	lb/hr	0	18	-	18	-	-	-	-	-	-	0	18	317	-
N-Nonenes (C9H18)	lb/hr	0	5	-	5	-	-	-	-	-	-	0	5	79	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Tar (C10H8)	lb/hr	0	0	-	0	0	-	-	0	0	-	-	0	0	-
Carbon (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Methanol (CH4O)	lb/hr	-	-	-	10,008	10,008	-	-	10,008	10,008	-	-	10,008	-	-
Dimethyl ether (C2H6O)	lb/hr	2,749	87,413	-	145,329	55,131	-	-	55,131	55,131	-	36	145,329	92,057	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ash	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Char	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Wood	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Enthalpy Flow	MMBTU/hr	(6)	(377)	0	(529)	(140)	(0)	(0)	(245)	(264)	(123)	(7)	(498)	(555)	(3)

110

S-1472
REACTOR EFFLUENT
KO DRUM 1ST STAGE

H-1413
HYDROCARBON AIR
COOLER

H-1414
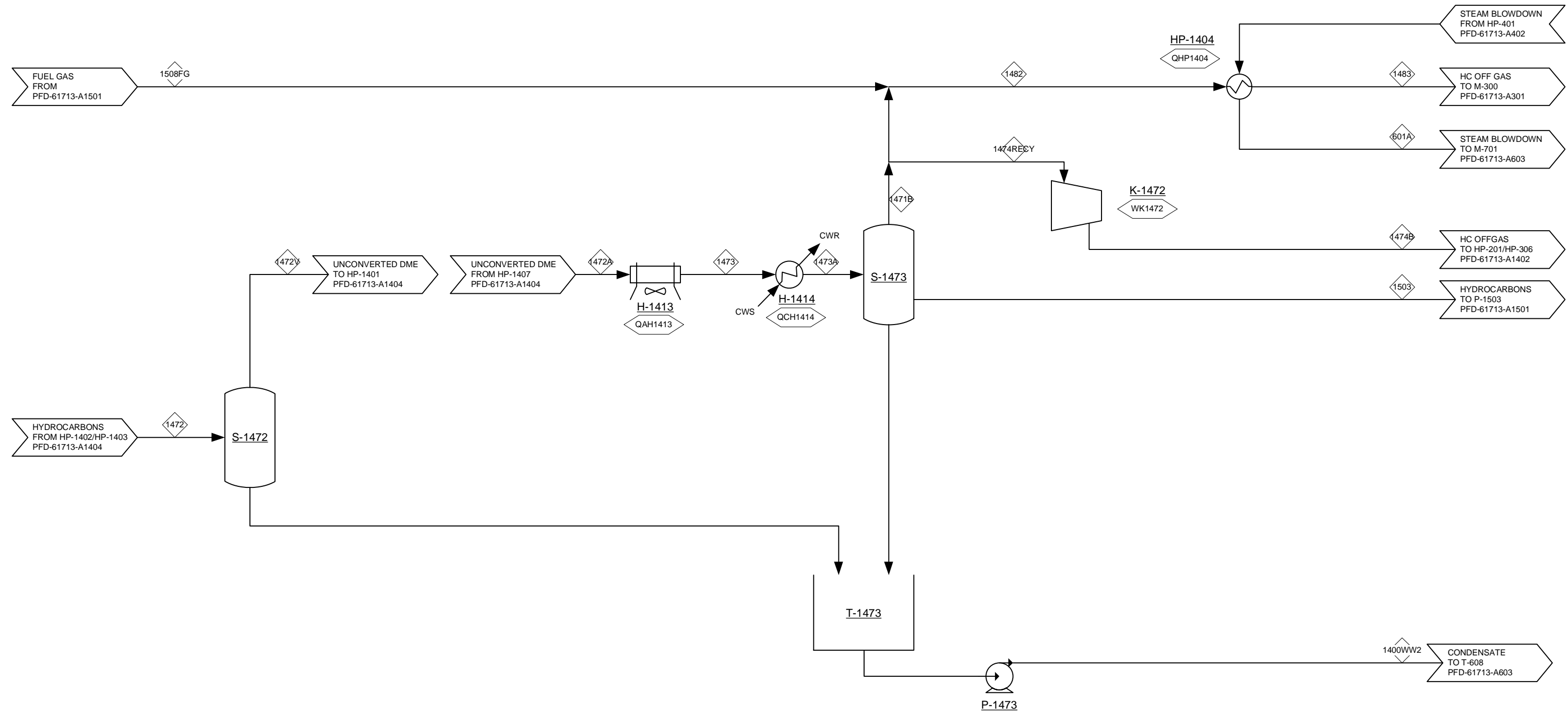HYDROCARBON
WATER COOLER

S-1473
REACTOR EFFLUENT KO
DRUM 2ND STAGE

T-1473
CONDENSATE SURGE
TANK

P-1473
CONDENSATE SURGE
TANK PUMP

K-1472
HYDROCARBON SYNTHESIS
REACTOR RECYCLE COMPRESSOR
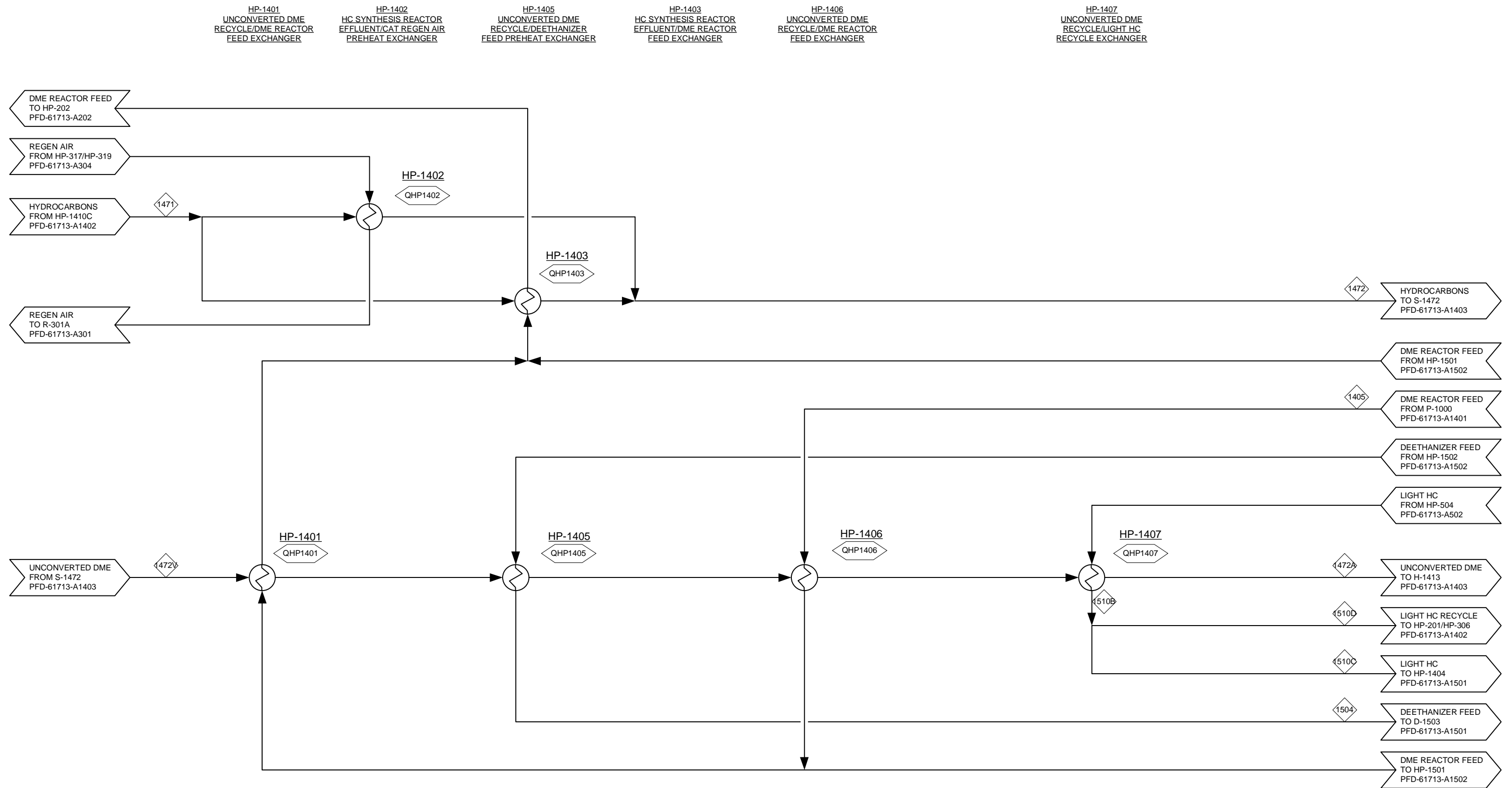
HP-1404
STEAM BLOWDOWN/
UNCONVERTED DME PURGE

HP-1404
QHP1404

STEAM BLOWDOWN
FROM HP-401
PFD-61713-A402

FUEL GAS
FROM
PFD-61713-A1501

1508FG

1482

1483

HC OFF GAS
TO M-300
PFD-61713-A301

601A

STEAM BLOWDOWN
TO M-701
PFD-61713-A603

1474RECY

K-1472
WK1472

1471B

UNCONVERTED DME
TO HP-1401
PFD-61713-A1404

472V

UNCONVERTED DME
FROM HP-1407
PFD-61713-A1404

472A

1473

H-1413
QAH1413

CWS

CWR

H-1414
QCH1414

473A

S-1473

1474B

HC OFFGAS
TO HP-201/HP-306
PFD-61713-A1402

1503

HYDROCARBONS
TO P-1503
PFD-61713-A1501

HYDROCARBONS
FROM HP-1402/HP-1403
PFD-61713-A1404

1472

S-1472

T-1473

1400WW2

CONDENSATE
TO T-608
PFD-61713-A603

P-1473

THERMOCHEMICAL CONVERSION OF BIOMASS
TO FUELS AND CHEMICALS
GASIFICATION

PROCESS FLOW DIAGRAM
AREA 1400: POST REACTOR COOLING

Rev.	Date	By	Description
1	12/5/14	DMS	ADDED HX NETWORK
0	9/26/14	JAR	ISSUED FOR DESIGN REPORT

Project No: 30482.00
Drawing: PFD-61713-A1403
Rev: 1

Harris Group

PFD-61713-A1403

COMPONENT	UNITS	1508FG	1482	1483	601A	1472	1472V	1472A	1473	1473A	1471B	1474RECY	1474B	1503	1400WW2
Total Flow	lb/hr	1,694	4,649	4,649	4,088	372,153	372,153	372,153	372,153	372,153	295,423	292,468	292,468	46,621	30,110
Temperature	F	102	96	173	111	250	250	180	140	110	110	110	168	110	168
Pressure	Psia	147.0	72.0	69.5	14.7	85.0	85.0	80.0	79.0	74.0	72.0	72.0	135.0	72.0	72.0
Vapor Fraction		1	1	1	-	1	1	1	1	1	1	1	1	-	-
Hydrogen (H2)	lb/hr	0	0	0	-	4	4	4	4	4	4	4	4	0	-
Carbon Monoxide (CO)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Nitrogen (N2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Oxygen (O2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Argon (Ar)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Carbon Dioxide (CO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Water (H2O)	lb/hr	2	27	27	4,088	32,989	32,989	32,989	32,989	32,989	2,525	2,499	2,499	355	30,110
Hydrogen Sulfide (H2S)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	0	0	0	-	0	0	0	0	0	0	0	0	0	-
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Methane (CH4)	lb/hr	59	421	421	-	36,306	36,306	36,306	36,306	36,306	36,236	35,874	35,874	69	-
Ethane (C2H6)	lb/hr	42	118	118	-	7,741	7,741	7,741	7,741	7,741	7,667	7,590	7,590	74	-
Ethylene (C2H4)	lb/hr	22	74	74	-	5,189	5,189	5,189	5,189	5,189	5,155	5,103	5,103	34	-
Acetylene (C2H2)	lb/hr	0	0	0	-	0	0	0	0	0	0	0	0	0	-
Propane (C3H8)	lb/hr	229	463	463	-	24,068	24,068	24,068	24,068	24,068	23,365	23,132	23,132	703	-
Propene (C3H6)	lb/hr	3	5	5	-	289	289	289	289	289	282	279	279	7	-
I-Butane (C4H10)	lb/hr	347	1,356	1,356	-	107,606	107,606	107,606	107,606	107,606	100,875	99,866	99,866	6,731	-
N-Butane (C4H10)	lb/hr	5	17	17	-	1,286	1,286	1,286	1,286	1,286	1,179	1,167	1,167	107	-
Butenes (C4H8)	lb/hr	9	34	34	-	2,624	2,624	2,624	2,624	2,624	2,437	2,412	2,412	187	-
Pentanes (C5H12)	lb/hr	0	49	49	-	5,910	5,910	5,910	5,910	5,910	4,876	4,827	4,827	1,034	-
Pentenes (C5H10)	lb/hr	0	33	33	-	4,028	4,028	4,028	4,028	4,028	3,269	3,236	3,236	759	-
Hexanes (C6H14)	lb/hr	0	30	30	-	4,840	4,840	4,840	4,840	4,840	3,047	3,017	3,017	1,793	-
Hexenes (C6H12)	lb/hr	0	6	6	-	1,015	1,015	1,015	1,015	1,015	592	587	587	423	-
Benzene (C6H6)	lb/hr	0	0	0	-	5	5	5	5	5	3	3	3	3	-
Heptanes (C7H16)	lb/hr	0	12	12	-	3,684	3,684	3,684	3,684	3,684	1,223	1,211	1,211	2,461	-
Triptane	lb/hr	0	85	85	-	18,139	18,139	18,139	18,139	18,139	8,514	8,429	8,429	9,625	-
Heptenes (C7H14)	lb/hr	0	8	8	-	2,587	2,587	2,587	2,587	2,587	846	838	838	1,741	-
Triptene	lb/hr	0	10	10	-	1,989	1,989	1,989	1,989	1,989	1,019	1,009	1,009	969	-
Octanes (C8H18)	lb/hr	0	37	37	-	18,563	18,563	18,563	18,563	18,563	3,737	3,699	3,699	14,826	-
Octenes (C8H16)	lb/hr	0	3	3	-	839	839	839	839	839	252	250	250	587	-
N-Nonanes (C9H20)	lb/hr	0	0	0	-	317	317	317	317	317	18	18	18	299	-
N-Nonenes (C9H18)	lb/hr	0	0	0	-	79	79	79	79	79	5	5	5	74	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Tar (C10H8)	lb/hr	0	0	0	-	0	0	0	0	0	0	0	0	0	-
Carbon (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Methanol (CH4O)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Dimethyl ether (C2H6O)	lb/hr	976	1,859	1,859	-	92,057	92,057	92,057	92,057	92,057	88,296	87,413	87,413	3,761	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ash	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Char	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Wood	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Enthalpy Flow	MMBTU/hr	(2)	(6)	(6)	(28)	(580)	(580)	(592)	(629)	(637)	(387)	(383)	(377)	(49)	(205)

HP-1401
UNCONVERTED DME
RECYCLE/DME REACTOR
FEED EXCHANGER

HP-1402
HC SYNTHESIS REACTOR
EFFLUENT/CAT REGEN AIR
PREHEAT EXCHANGER

HP-1405
UNCONVERTED DME
RECYCLE/DEETHANIZER
FEED PREHEAT EXCHANGER

HP-1403
HC SYNTHESIS REACTOR
EFFLUENT/DME REACTOR
FEED EXCHANGER

HP-1406
UNCONVERTED DME
RECYCLE/DME REACTOR
FEED EXCHANGER

HP-1407
UNCONVERTED DME
RECYCLE/LIGHT HC
RECYCLE EXCHANGER

DME REACTOR FEED
TO HP-202
PFD-61713-A202

REGEN AIR
FROM HP-317/HP-319
PFD-61713-A304

HYDROCARBONS
FROM HP-1410C
PFD-61713-A1402

1471

HP-1402
QHP1402

HP-1403
QHP1403

REGEN AIR
TO R-301A
PFD-61713-A301

1472

HYDROCARBONS
TO S-1472
PFD-61713-A1403

DME REACTOR FEED
FROM HP-1501
PFD-61713-A1502

1405

DME REACTOR FEED
FROM P-1000
PFD-61713-A1401

DEETHANIZER FEED
FROM HP-1502
PFD-61713-A1502

LIGHT HC
FROM HP-504
PFD-61713-A502

HP-1401
QHP1401

HP-1405
QHP1405

HP-1406
QHP1406

HP-1407
QHP1407

UNCONVERTED DME
FROM S-1472
PFD-61713-A1403

1472

1472A

UNCONVERTED DME
TO H-1413
PFD-61713-A1403

510D

LIGHT HC RECYCLE
TO HP-201/HP-306
PFD-61713-A1402

510C

LIGHT HC
TO HP-1404
PFD-61713-A1501

1504

DEETHANIZER FEED
TO D-1503
PFD-61713-A1501

DME REACTOR FEED
TO HP-1501
PFD-61713-A1502

Harris Group

NREL
NATIONAL RENEWABLE ENERGY LABORATORY

THERMOCHEMICAL CONVERSION OF BIOMASS
TO FUELS AND CHEMICALS
GASIFICATION

PROCESS FLOW DIAGRAM
AREA 1400: A1400 HEAT INTEGRATION

Rev.	Date	By	Description
0	12/5/14	DMS	ADDED HX NETWORK

Project No: 30482.00   Drawing: PFD-61713-A1404   Rev: 0

PFD-61713-A1404

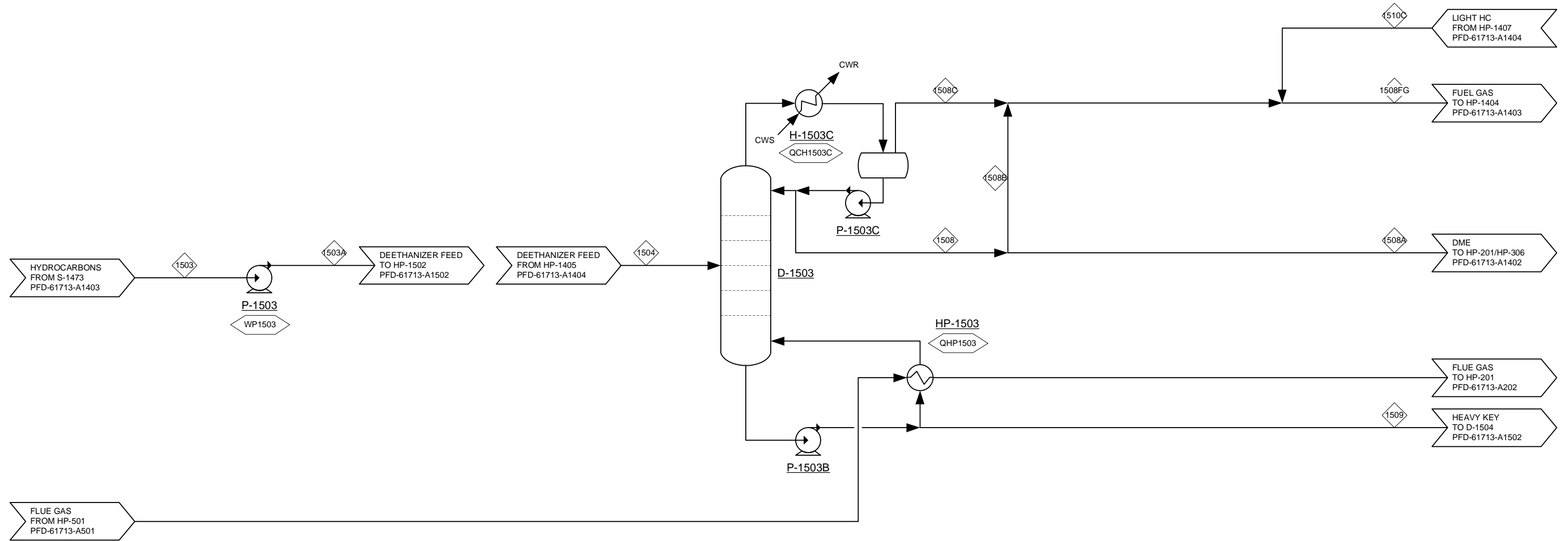| COMPONENT	UNITS	1471	1472V	1472	1405	1472A	1510D	1510C	1504
Total Flow	lb/hr	372,153	372,153	372,153	87,710	372,153	6,688	352	46,621
Temperature	F	375	250	250	150	180	173	173	173
Pressure	Psia	90.0	85.0	85.0	150.0	80.0	147.0	147.0	312.2
Vapor Fraction		1	1	1	-	1	1	1	-
Hydrogen (H2)	lb/hr	4	4	4	-	4	-	-	0
Carbon Monoxide (CO)	lb/hr	-	-	-	-	-	-	-	-
Nitrogen (N2)	lb/hr	-	-	-	-	-	-	-	-
Oxygen (O2)	lb/hr	-	-	-	-	-	-	-	-
Argon (Ar)	lb/hr	-	-	-	-	-	-	-	-
Carbon Dioxide (CO2)	lb/hr	-	-	-	0	-	-	-	-
Water (H2O)	lb/hr	32,989	32,989	32,989	1,010	32,989	42	2	355
Hydrogen Sulfide (H2S)	lb/hr	-	-	-	-	-	-	-	-
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	0	0	0	0	0	-	-	0
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-	-	-	-
Methane (CH4)	lb/hr	36,306	36,306	36,306	-	36,306	-	-	69
Ethane (C2H6)	lb/hr	7,741	7,741	7,741	0	7,741	-	-	74
Ethylene (C2H4)	lb/hr	5,189	5,189	5,189	0	5,189	-	-	34
Acetylene (C2H2)	lb/hr	0	0	0	0	0	-	-	0
Propane (C3H8)	lb/hr	24,068	24,068	24,068	0	24,068	0	0	703
Propene (C3H6)	lb/hr	289	289	289	-	289	0	0	7
I-Butane (C4H10)	lb/hr	107,606	107,606	107,606	0	107,606	6,329	333	6,731
N-Butane (C4H10)	lb/hr	1,286	1,286	1,286	0	1,286	101	5	107
Butenes (C4H8)	lb/hr	2,624	2,624	2,624	-	2,624	177	9	187
Pentanes (C5H12)	lb/hr	5,910	5,910	5,910	0	5,910	3	0	1,034
Pentenes (C5H10)	lb/hr	4,028	4,028	4,028	-	4,028	0	0	759
Hexanes (C6H14)	lb/hr	4,840	4,840	4,840	-	4,840	0	0	1,793
Hexenes (C6H12)	lb/hr	1,015	1,015	1,015	-	1,015	0	0	423
Benzene (C6H6)	lb/hr	5	5	5	3	5	0	0	3
Heptanes (C7H16)	lb/hr	3,684	3,684	3,684	-	3,684	0	0	2,461
Triptane	lb/hr	18,139	18,139	18,139	-	18,139	0	0	9,625
Heptenes (C7H14)	lb/hr	2,587	2,587	2,587	-	2,587	0	0	1,741
Triptene	lb/hr	1,989	1,989	1,989	-	1,989	0	0	969
Octanes (C8H18)	lb/hr	18,563	18,563	18,563	-	18,563	0	0	14,826
Octenes (C8H16)	lb/hr	839	839	839	-	839	0	0	587
N-Nonanes (C9H20)	lb/hr	317	317	317	-	317	0	0	299
N-Nonenes (C9H18)	lb/hr	79	79	79	-	79	0	0	74
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-	-	-	-
Tar (C10H8)	lb/hr	0	0	0	0	0	-	-	0
Carbon (Solid)	lb/hr	-	-	-	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	-	-	-	-	-	-
Methanol (CH4O)	lb/hr	-	-	-	86,697	-	-	-	-
Dimethyl ether (C2H6O)	lb/hr	92,057	92,057	92,057	-	92,057	36	2	3,761
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-	-	-	-
Ash	lb/hr	-	-	-	-	-	-	-	-
Char	lb/hr	-	-	-	-	-	-	-	-
Wood	lb/hr	-	-	-	-	-	-	-	-
Enthalpy Flow	MMBTU/hr	(555)	(580)	(580)	(280)	(592)	(7)	(0)	(48)

114

P-1503
CRUDE
HYDROCARBONS PUMP

D-1503
DEETHANIZER
COLUMN

H-1503C
DEETHANIZER COLUMN
CONDENSER

P-1503C
DEETHANIZER
REFLUX PUMP

P-1503B
DEETHANIZER
REBOILER PUMP

HP-1503
CHAR COMBUSTOR FLUE
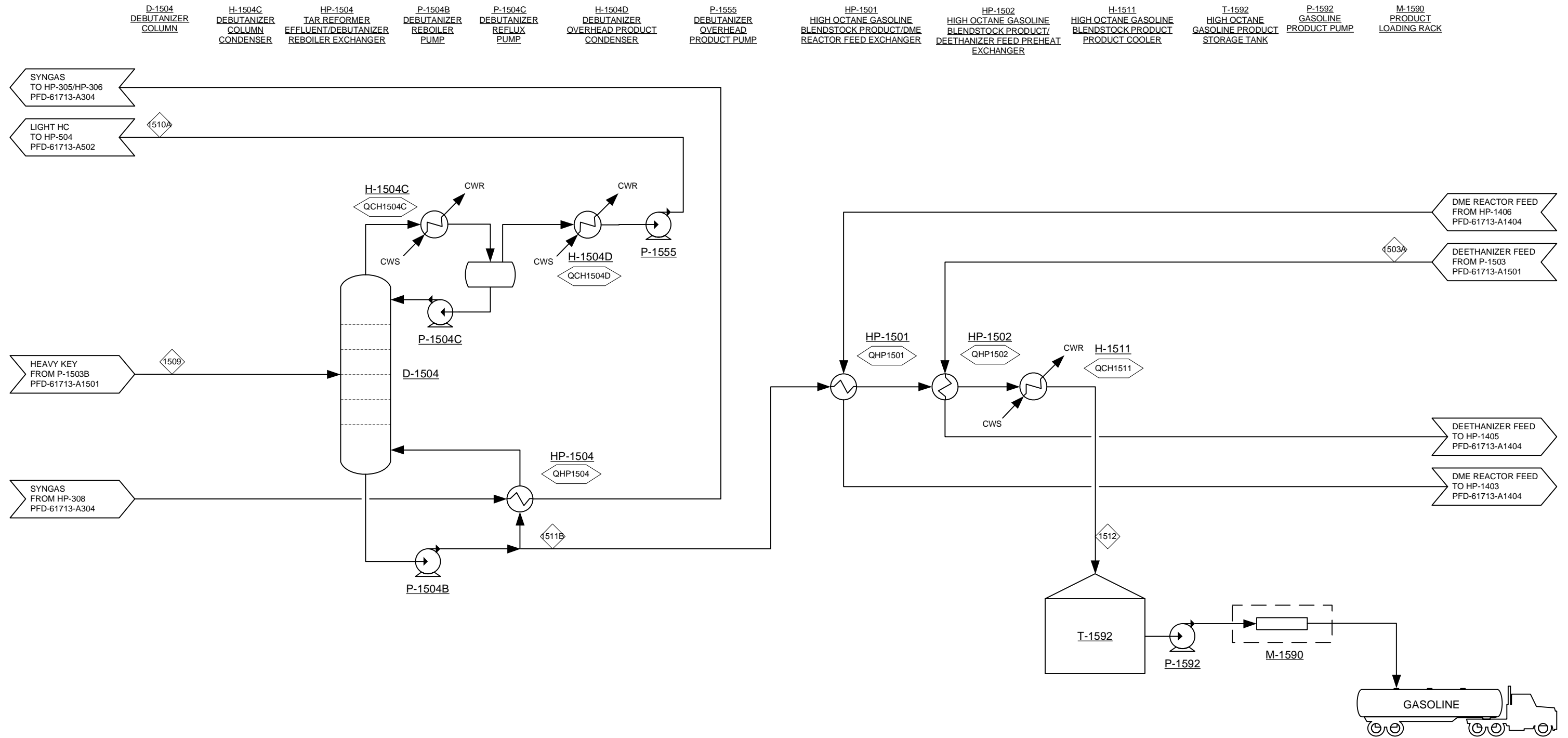GAS/DEETHANIZER
REBOILER EXCHANGER

1510C
LIGHT HC
FROM HP-1407
PFD-61713-A1404

CWR

1508C

1508FG
FUEL GAS
TO HP-1404
PFD-61713-A1403

CWS

H-1503C
QCH1503C

1508B

P-1503C

1508

1508A
DME
TO HP-201/HP-306
PFD-61713-A1402

HYDROCARBONS
FROM S-1473
PFD-61713-A1403

1503

1503A

DEETHANIZER FEED
TO HP-1502
PFD-61713-A1502

DEETHANIZER FEED
FROM HP-1405
PFD-61713-A1404

1504

D-1503

P-1503
WP1503

HP-1503
QHP1503

FLUE GAS
TO HP-201
PFD-61713-A202

FLUE GAS
FROM HP-501
PFD-61713-A501

1509
HEAVY KEY
TO D-1504
PFD-61713-A1502

P-1503B

					THERMOCHEMICAL CONVERSION OF BIOMASS TO FUELS AND CHEMICALS GASIFICATION
					PROCESS FLOW DIAGRAM AREA 1500: DEETHANIZER
1	12/5/14	DMS	ADDED HX NETWORK		
0	9/26/14	JAR	ISSUED FOR DESIGN REPORT	Project No: 30482.00	Drawing PFD-61713-A1501 Rev: 1
Rev.	Date	By	Description		

NREL
NATIONAL RENEWABLE ENERGY LABORATORY

Harris Group

PFD-61713-A1501

COMPONENT	UNITS	1503	1503A	1504	1509	1508	1508A	1508B	1508C	1508FG	1510C
Total Flow	lb/hr	46,621	46,621	46,621	41,942	3,512	3,337	176	1,167	1,694	1,167
Temperature	F	110	113	173	306	110	110	110	110	102	110
Pressure	Psia	72.0	314.7	312.2	200.0	200.0	200.0	200.0	200.0	147.0	200.0
Vapor Fraction		-	-	-	-	-	-	-	1	1	1
Hydrogen (H2)	lb/hr	0	0	0	0	0	0	0	0	0	0
Carbon Monoxide (CO)	lb/hr	-	-	-	-	-	-	-	-	-	-
Nitrogen (N2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Oxygen (O2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Argon (Ar)	lb/hr	-	-	-	-	-	-	-	-	-	-
Carbon Dioxide (CO2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Water (H2O)	lb/hr	355	355	355	355	0	0	0	0	2	0
Hydrogen Sulfide (H2S)	lb/hr	-	-	-	-	-	-	-	-	-	-
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	0	0	0	0	0	0	0	0	0	0
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Methane (CH4)	lb/hr	69	69	69	0	11	10	1	58	59	58
Ethane (C2H6)	lb/hr	74	74	74	0	34	32	2	40	42	40
Ethylene (C2H4)	lb/hr	34	34	34	0	13	12	1	21	22	21
Acetylene (C2H2)	lb/hr	0	0	0	-	0	0	0	0	0	0
Propane (C3H8)	lb/hr	703	703	703	0	498	473	25	204	229	204
Propene (C3H6)	lb/hr	7	7	7	0	5	5	0	2	3	2
I-Butane (C4H10)	lb/hr	6,731	6,731	6,731	6,663	56	54	3	11	347	11
N-Butane (C4H10)	lb/hr	107	107	107	107	0	0	0	0	5	0
Butenes (C4H8)	lb/hr	187	187	187	187	0	0	0	0	9	0
Pentanes (C5H12)	lb/hr	1,034	1,034	1,034	1,034	0	0	0	0	0	0
Pentenes (C5H10)	lb/hr	759	759	759	759	0	0	0	0	0	0
Hexanes (C6H14)	lb/hr	1,793	1,793	1,793	1,793	0	0	0	0	0	0
Hexenes (C6H12)	lb/hr	423	423	423	423	0	0	0	0	0	0
Benzene (C6H6)	lb/hr	3	3	3	3	0	0	0	0	0	0
Heptanes (C7H16)	lb/hr	2,461	2,461	2,461	2,461	0	0	0	0	0	0
Triptane	lb/hr	9,625	9,625	9,625	9,625	0	0	0	0	0	0
Heptenes (C7H14)	lb/hr	1,741	1,741	1,741	1,741	0	0	0	0	0	0
Triptene	lb/hr	969	969	969	969	0	0	0	0	0	0
Octanes (C8H18)	lb/hr	14,826	14,826	14,826	14,826	0	0	0	0	0	0
Octenes (C8H16)	lb/hr	587	587	587	587	0	0	0	0	0	0
N-Nonanes (C9H20)	lb/hr	299	299	299	299	0	0	0	0	0	0
N-Nonenes (C9H18)	lb/hr	74	74	74	74	0	0	0	0	0	0
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-	-	-	-	-	-
Tar (C10H8)	lb/hr	0	0	0	0	0	0	0	-	0	-
Carbon (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-
Methanol (CH4O)	lb/hr	-	-	-	-	-	-	-	-	-	-
Dimethyl ether (C2H6O)	lb/hr	3,761	3,761	3,761	38	2,894	2,749	145	829	976	829
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Ash	lb/hr	-	-	-	-	-	-	-	-	-	-
Char	lb/hr	-	-	-	-	-	-	-	-	-	-
Wood	lb/hr	-	-	-	-	-	-	-	-	-	-
Enthalpy Flow	MMBTU/hr	(49)	(49)	(48)	(36)	(6)	(6)	(0)	(2)	(2)	(2)

116

Equipment Labels

Tag	Description
D-1504	DEBUTANIZER COLUMN
H-1504C	DEBUTANIZER COLUMN CONDENSER
HP-1504	TAR REFORMER EFFLUENT/DEBUTANIZER REBOILER EXCHANGER
P-1504B	DEBUTANIZER REBOILER PUMP
P-1504C	DEBUTANIZER REFLUX PUMP
H-1504D	DEBUTANIZER OVERHEAD PRODUCT CONDENSER
P-1555	DEBUTANIZER OVERHEAD PRODUCT PUMP
HP-1501	HIGH OCTANE GASOLINE BLENDSTOCK PRODUCT/DME REACTOR FEED EXCHANGER
HP-1502	HIGH OCTANE GASOLINE BLENDSTOCK PRODUCT/ DEETHANIZER FEED PREHEAT EXCHANGER
H-1511	HIGH OCTANE GASOLINE BLENDSTOCK PRODUCT PRODUCT COOLER
T-1592	HIGH OCTANE GASOLINE PRODUCT STORAGE TANK
P-1592	GASOLINE PRODUCT PUMP
M-1590	PRODUCT LOADING RACK

Stream Connections (left side)

SYNGAS TO HP-305/HP-306 PFD-61713-A304

LIGHT HC TO HP-504 PFD-61713-A502 — 1510A

HEAVY KEY FROM P-1503B PFD-61713-A1501 — 1509

SYNGAS FROM HP-308 PFD-61713-A304

Stream Connections (right side)

DME REACTOR FEED FROM HP-1406 PFD-61713-A1404

DEETHANIZER FEED FROM P-1503 PFD-61713-A1501 — 1503A

DEETHANIZER FEED TO HP-1405 PFD-61713-A1404

DME REACTOR FEED TO HP-1403 PFD-61713-A1404

Process Stream / Equipment Tags

H-1504C — QCH1504C — CWR — CWS
H-1504D — QCH1504D — CWR — CWS
P-1555
P-1504C
P-1504B
HP-1504 — QHP1504 — 1511B
HP-1501 — QHP1501
HP-1502 — QHP1502 — CWR — CWS
H-1511 — QCH1511
1512
T-1592
P-1592
M-1590
GASOLINE

				THERMOCHEMICAL CONVERSION OF BIOMASS TO FUELS AND CHEMICALS GASIFICATION
				PROCESS FLOW DIAGRAM AREA 1500: DEBUTANIZER
1	12/5/14	DMS	ADDED HX NETWORK	
0	9/26/14	JAR	ISSUED FOR DESIGN REPORT	Project No: 30482.00 — Drawing: PFD-61713-A1502 — Rev: 1
Rev.	Date	By	Description	

Harris Group

NREL — NATIONAL RENEWABLE ENERGY LABORATORY

PFD-61713-A1502

COMPONENT	UNITS	1509	1510A	1511B	1503A
Total Flow	lb/hr	41,942	7,040	34,902	46,621
Temperature	F	306	128	336	113
Pressure	Psia	200.0	152.0	120.0	314.7
Vapor Fraction		-	-	-	-
Hydrogen (H2)	lb/hr	0	-	-	0
Carbon Monoxide (CO)	lb/hr	-	-	-	-
Nitrogen (N2)	lb/hr	-	-	-	-
Oxygen (O2)	lb/hr	-	-	-	-
Argon (Ar)	lb/hr	-	-	-	-
Carbon Dioxide (CO2)	lb/hr	-	-	-	-
Water (H2O)	lb/hr	355	44	310	355
Hydrogen Sulfide (H2S)	lb/hr	-	-	-	-
Sulfur dioxide (SO2)	lb/hr	-	-	-	-
Ammonia (NH3)	lb/hr	0	-	-	0
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-
Methane (CH4)	lb/hr	0	-	-	69
Ethane (C2H6)	lb/hr	0	-	-	74
Ethylene (C2H4)	lb/hr	0	-	-	34
Acetylene (C2H2)	lb/hr	-	-	-	0
Propane (C3H8)	lb/hr	0	0	0	703
Propene (C3H6)	lb/hr	0	0	0	7
I-Butane (C4H10)	lb/hr	6,663	6,662	1	6,731
N-Butane (C4H10)	lb/hr	107	106	1	107
Butenes (C4H8)	lb/hr	187	186	0	187
Pentanes (C5H12)	lb/hr	1,034	3	1,031	1,034
Pentenes (C5H10)	lb/hr	759	0	759	759
Hexanes (C6H14)	lb/hr	1,793	0	1,793	1,793
Hexenes (C6H12)	lb/hr	423	0	423	423
Benzene (C6H6)	lb/hr	3	0	3	3
Heptanes (C7H16)	lb/hr	2,461	0	2,461	2,461
Triptane	lb/hr	9,625	0	9,625	9,625
Heptenes (C7H14)	lb/hr	1,741	0	1,741	1,741
Triptene	lb/hr	969	0	969	969
Octanes (C8H18)	lb/hr	14,826	0	14,826	14,826
Octenes (C8H16)	lb/hr	587	0	587	587
N-Nonanes (C9H20)	lb/hr	299	0	299	299
N-Nonenes (C9H18)	lb/hr	74	0	74	74
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-
Tar (C10H8)	lb/hr	0	-	0	0
Carbon (Solid)	lb/hr	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	-	-
Methanol (CH4O)	lb/hr	-	-	-	-
Dimethyl ether (C2H6O)	lb/hr	38	38	0	3,761
MDEA (C5H13NO2)	lb/hr	-	-	-	-
Ash	lb/hr	-	-	-	-
Char	lb/hr	-	-	-	-
Wood	lb/hr	-	-	-	-
Enthalpy Flow	MMBTU/hr	(36)	(8)	(28)	(49)

118

Process Flow Diagram labels:

- P-601 MAKE-UP PUMP
- M-601A BOILER FEEDWATER REVERSE OSMOSIS UNIT
- T-601 CONDENSATE COLLECTION TANK
- P-605 CONDENSATE COLLECTION PUMP
- T-602 CONDENSATE SURGE TANK
- M-604 BRINE RECOVERY RO UNIT
- P-603 EDI PUMP
- M-601B BOILER FEEDWATER ELECTRODEIONIZATION UNIT
- M-601C HOT CONDENSATE POLISHING UNIT
- T-603A DEAERATOR PACKED COLUMN
- T-603 DEAERATOR
- P-604 BOILER FEEDWATER PUMP

Stream labels:
- BOILER FW/COND FROM HP-402 PFD-61713-A402 — 626
- ATM
- T-601
- 627
- P-605
- T-602
- P-603 / WP603
- EDI / M-601B
- M-601C — 630
- BOILER FEEDWATER TO HP-404 PFD-61713-A402
- BOILER FEEDWATER FROM HP-404 PFD-61713-A402 — 631
- T-603A / T-603
- ATM — 632
- FLASH STEAM FROM S-601/S-661 PFD-61713-A603
- AMMONIA FROM P-704 PFD-61713-A702
- BFW CHEMICALS FROM P-707 PFD-61713-A702
- 633 — P-604 / WP604
- 640 — BOILER FEEDWATER TO HP-405 PFD-61713-A402
- CONDENSATE FROM S-410 PFD-61713-A403
- CONDENSATE FROM P-602 PFD-61713-A602 — 616
- MAKE-UP WATER FROM DISTRIBUTION PFD-61713-A701 — 618
- M-601A / RO
- P-601 / WP601
- 661 — MAKEUP WATER TO T-608 PFD-61713-A603
- M-604 / RO
- RO BRINE FROM M-709 PFD-61713-A701
- RECYCLED BRINE TO M-701 PFD-61713-A701
- BRINE TO C-201 PFD-61713-A201 — 218
- BOILER FEEDWATER TO HP-302/HP-304 PFD-61713-A303
- BOILER FEEDWATER TO HP-316 PFD-61713-A304
- BOILER FEEDWATER TO HP-318 PFD-61713-A304

Left margin text: J:\1A - Process Solutions\30482 NREL Thermochemical Conv - Design-Model\Drawings\30482 Gasification PFDs 61713 150128.vsd
Plot date: 1/29/2015

Title block:
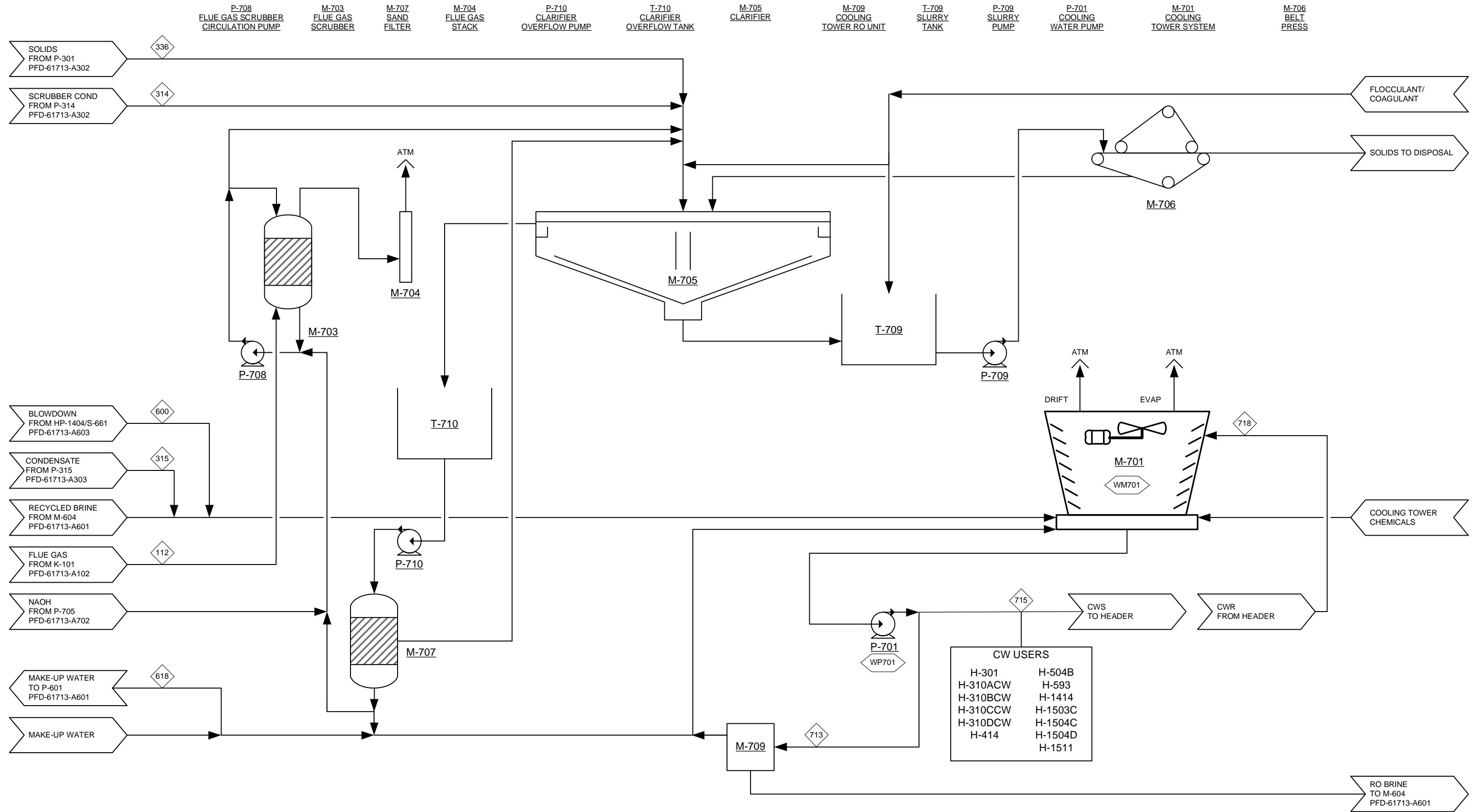
THERMOCHEMICAL CONVERSION OF BIOMASS TO FUELS AND CHEMICALS GASIFICATION

PROCESS FLOW DIAGRAM
AREA 600: STEAM SYSTEM & POWER GENERATION

Rev.	Date	By	Description
1	12/5/14	DMS	ADDED HX NETWORK
0	9/26/14	JAR	ISSUED FOR DESIGN REPORT

Project No: 30482.00   Drawing: PFD-61713-A601   Rev: 1

Harris Group
NREL NATIONAL RENEWABLE ENERGY LABORATORY

PFD-61713-A601

COMPONENT	UNITS	616	618	626	627	630	631	633	640	661	218
Total Flow	lb/hr	177,601	81,739	415,415	410,302	410,302	410,302	410,302	410,305	9,494	243
Temperature	F	148	60	237	226	226	232	232	233	71	60
Pressure	Psia	24.7	14.7	19.1	19.1	24.7	21.7	21.7	485.9	14.7	14.7
Vapor Fraction		-	-	-	-	-	-	-	-	-	-
Hydrogen (H2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Carbon Monoxide (CO)	lb/hr	-	-	-	-	-	-	-	-	-	-
Nitrogen (N2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Oxygen (O2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Argon (Ar)	lb/hr	-	-	-	-	-	-	-	-	-	-
Carbon Dioxide (CO2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Water (H2O)	lb/hr	177,601	81,739	415,415	410,302	410,302	410,302	410,302	410,305	9,494	243
Hydrogen Sulfide (H2S)	lb/hr	-	-	-	-	-	-	-	-	-	-
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	-	-	-	-	-	-	-	-	-	-
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Methane (CH4)	lb/hr	-	-	-	-	-	-	-	-	-	-
Ethane (C2H6)	lb/hr	-	-	-	-	-	-	-	-	-	-
Ethylene (C2H4)	lb/hr	-	-	-	-	-	-	-	-	-	-
Acetylene (C2H2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Propane (C3H8)	lb/hr	-	-	-	-	-	-	-	-	-	-
Propene (C3H6)	lb/hr	-	-	-	-	-	-	-	-	-	-
I-Butane (C4H10)	lb/hr	-	-	-	-	-	-	-	-	-	-
N-Butane (C4H10)	lb/hr	-	-	-	-	-	-	-	-	-	-
Butenes (C4H8)	lb/hr	-	-	-	-	-	-	-	-	-	-
Pentanes (C5H12)	lb/hr	-	-	-	-	-	-	-	-	-	-
Pentenes (C5H10)	lb/hr	-	-	-	-	-	-	-	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-	-	-	-	-	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-	-	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	-	-	-	-	-	-	-	-	-	-
Heptanes (C7H16)	lb/hr	-	-	-	-	-	-	-	-	-	-
Triptane	lb/hr	-	-	-	-	-	-	-	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-	-	-	-	-	-	-	-
Triptene	lb/hr	-	-	-	-	-	-	-	-	-	-
Octanes (C8H18)	lb/hr	-	-	-	-	-	-	-	-	-	-
Octenes (C8H16)	lb/hr										
N-Nonanes (C9H20)	lb/hr	-	-	-	-	-	-	-	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-	-	-	-	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-	-	-	-	-	-
Tar (C10H8)	lb/hr	-	-	-	-	-	-	-	-	-	-
Carbon (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-
Methanol (CH4O)	lb/hr	-	-	-	-	-	-	-	-	-	-
Dimethyl ether (C2H6O)	lb/hr	-	-	-	-	-	-	-	-	-	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Ash	lb/hr	-	-	-	-	-	-	-	-	-	-
Char	lb/hr	-	-	-	-	-	-	-	-	-	-
Wood	lb/hr	-	-	-	-	-	-	-	-	-	-
Enthalpy Flow	MMBTU/hr	(1,200)	(559)	(2,769)	(2,739)	(2,739)	(2,737)	(2,737)	(2,736)	(65)	(2)

120

Process Flow Diagram — Area 600: Steam System & Power Generation

Equipment labels:

- M-602A — EXTRACTION STEAM TURBINE/GENERATOR STAGE 1
- H-601 — STEAM TURBINE CONDENSER
- M-602B — EXTRACTION STEAM TURBINE/GENERATOR STAGE 2
- M-603 — STARTUP BOILER
- T-606 — STEAM TURBINE CONDENSATE TANK
- P-602 — CONDENSATE PUMP
- M-602C — EXTRACTION STEAM TURBINE/GENERATOR STAGE 3

Stream labels:

- BOILER FEEDWATER FROM HP-204 PFD-61713-A202
- M-603
- STEAM TO S-410 PFD-61713-A403 (450)
- STEAM TO HP-313 PFD-61713-A305 (200)
- SUPERHEATED STEAM FROM HP-205/HP-314 PFD-61713-A603 (607)
- (609A)
- STEAM TO HP-309 PFD-61713-A305
- STEAM FROM HP-309 PFD-61713-A305
- (602C)
- (610)
- STEAM TO HP-310 PFD-61713-A305
- STEAM FROM HP-310 PFD-61713-A305
- (611C)
- STEAM TURB EXHST TO HP-402 PFD-61713-A402 (614)
- M-602A
- M-602B
- M-602C
- WM602
- STEAM TURB EXHST FROM HP-317 PFD-61713-A304
- (615)
- T-606
- H-601 / QAH601
- P-602 / WP602
- (616)
- CONDENSATE TO PFD-61713-A601

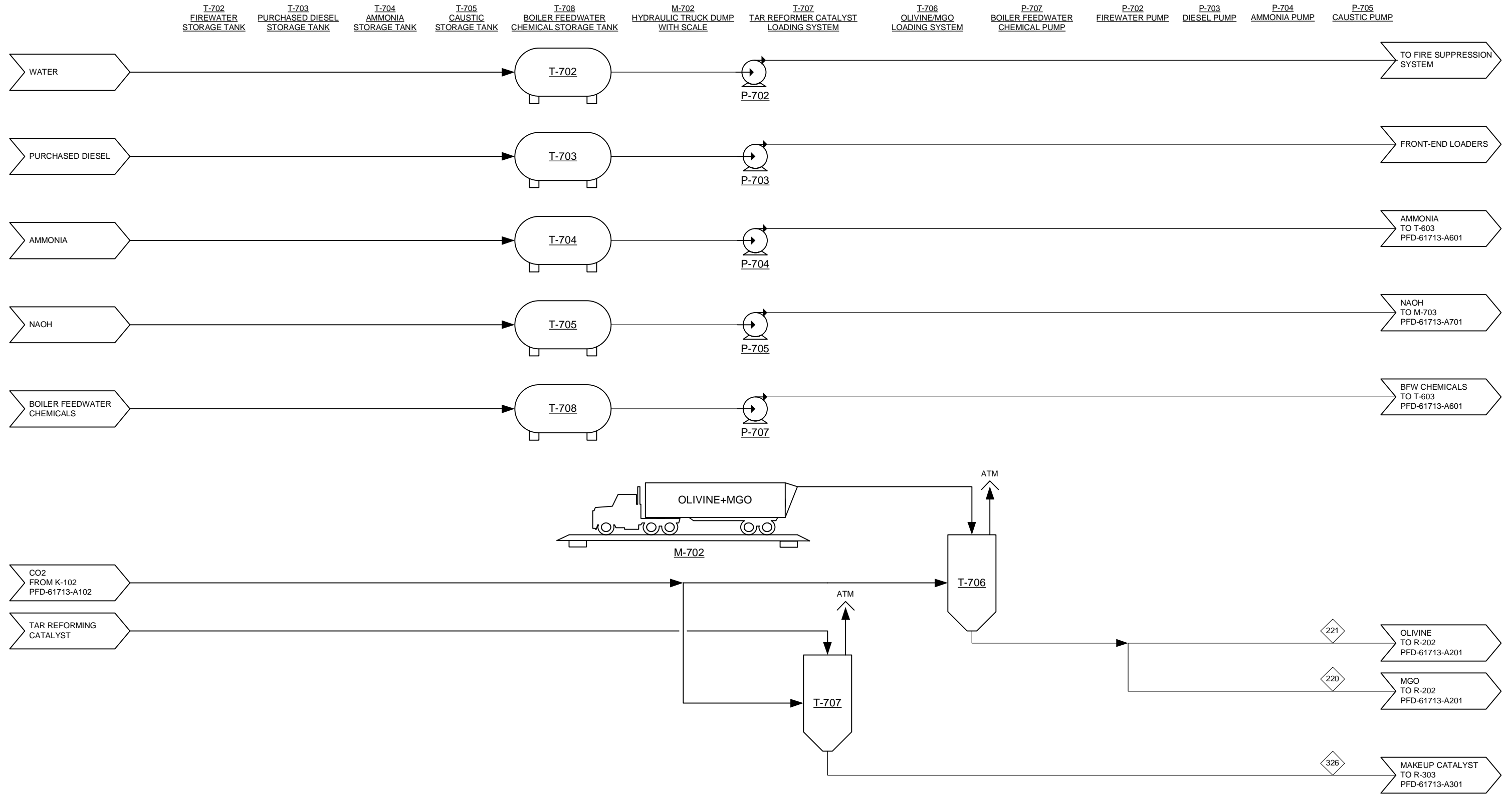| Rev.	Date	By	Description
1	12/5/14	DMS	ADDED HX NETWORK
0	9/26/14	JAR	ISSUED FOR DESIGN REPORT

THERMOCHEMICAL CONVERSION OF BIOMASS TO FUELS AND CHEMICALS GASIFICATION

PROCESS FLOW DIAGRAM
AREA 600: STEAM SYSTEM & POWER GENERATION

Project No: 30482.00

Drawing: PFD-61713-A602

Rev: 1

PFD-61713-A602

COMPONENT	UNITS	607	609A	602C	610	611C	614	615	616	450	200
Total Flow	lb/hr	404,719	404,719	404,719	404,719	177,601	177,601	177,601	177,601	153,812	73,120
Temperature	F	1,000	783	1,000	784	1,000	472	148	148	784	782
Pressure	Psia	465.9	176.0	174.0	64.7	64.7	3.5	3.5	24.7	64.7	35.0
Vapor Fraction		1	1	1	1	1	1	-	-	1	1
Hydrogen (H2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Carbon Monoxide (CO)	lb/hr	-	-	-	-	-	-	-	-	-	-
Nitrogen (N2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Oxygen (O2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Argon (Ar)	lb/hr	-	-	-	-	-	-	-	-	-	-
Carbon Dioxide (CO2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Water (H2O)	lb/hr	404,719	404,719	404,719	404,719	177,601	177,601	177,601	177,601	153,812	73,120
Hydrogen Sulfide (H2S)	lb/hr	-	-	-	-	-	-	-	-	-	-
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	-	-	-	-	-	-	-	-	-	-
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Methane (CH4)	lb/hr	-	-	-	-	-	-	-	-	-	-
Ethane (C2H6)	lb/hr	-	-	-	-	-	-	-	-	-	-
Ethylene (C2H4)	lb/hr	-	-	-	-	-	-	-	-	-	-
Acetylene (C2H2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Propane (C3H8)	lb/hr	-	-	-	-	-	-	-	-	-	-
Propene (C3H6)	lb/hr	-	-	-	-	-	-	-	-	-	-
I-Butane (C4H10)	lb/hr	-	-	-	-	-	-	-	-	-	-
N-Butane (C4H10)	lb/hr	-	-	-	-	-	-	-	-	-	-
Butenes (C4H8)	lb/hr	-	-	-	-	-	-	-	-	-	-
Pentanes (C5H12)	lb/hr	-	-	-	-	-	-	-	-	-	-
Pentenes (C5H10)	lb/hr	-	-	-	-	-	-	-	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-	-	-	-	-	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-	-	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	-	-	-	-	-	-	-	-	-	-
Heptanes (C7H16)	lb/hr	-	-	-	-	-	-	-	-	-	-
Triptane	lb/hr	-	-	-	-	-	-	-	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-	-	-	-	-	-	-	-
Triptene	lb/hr	-	-	-	-	-	-	-	-	-	-
Octanes (C8H18)	lb/hr	-	-	-	-	-	-	-	-	-	-
Octenes (C8H16)	lb/hr										
N-Nonanes (C9H20)	lb/hr	-	-	-	-	-	-	-	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-	-	-	-	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-	-	-	-	-	-
Tar (C10H8)	lb/hr	-	-	-	-	-	-	-	-	-	-
Carbon (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-
Methanol (CH4O)	lb/hr	-	-	-	-	-	-	-	-	-	-
Dimethyl ether (C2H6O)	lb/hr	-	-	-	-	-	-	-	-	-	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-
Ash	lb/hr	-	-	-	-	-	-	-	-	-	-
Char	lb/hr	-	-	-	-	-	-	-	-	-	-
Wood	lb/hr	-	-	-	-	-	-	-	-	-	-
Enthalpy Flow	MMBTU/hr	(2,165)	(2,207)	(2,162)	(2,205)	(948)	(994)	(1,200)	(1,200)	(838)	(398)

122

T-608
CONDENSATE
COLLECTION TANK

T-605
STEAM
DRUM

P-661
INTERCOOLER
CHARGE PUMP

S-601
BLOWDOWN
FLASH DRUM

T-607
STEAM DRUM

S-661
BLOWDOWN
FLASH DRUM

SATURATED STEAM
FROM HP-206
PFD-61713-A202

SATURATED STEAM
FROM HP-308
PFD-61713-A304

STEAM
FROM R-490
PFD-61713-A401

644

T-605

606

SATURATED STEAM
TO HP-205
PFD-61713-A202

SATURATED STEAM
TO HP-314
PFD-61713-A305

SUPERHEATED STEAM
FROM HP-205
PFD-61713-A202

SUPERHEATED STEAM
FROM HP-314
PFD-61713-A305

607

SUPERHEATED STEAM
TO M-602A
PFD-61713-A602

FLASH STEAM
TO T-603A
PFD-61713-A601

S-601

604

STEAM BLOWDOWN
TO HP-401
PFD-61713-A402

STEAM BLOWDOWN
FROM HP-1404
PFD-61713-A1403

601A

600

BLOWDOWN
TO M-701
PFD-61713-A701

662

STEAM
TO HP-312
PFD-61713-A305

MAKEUP WATER
FROM M-601A
PFD-61713-A601

661

CONDENSATE
FROM P-1470
PFD-61713-A1402

1400WW3

CONDENSATE
FROM P-1473
PFD-61713-A1403

1400WW2

T-608

P-661

WP661

660A

WATER
TO H-1424
PFD-61713-A1401

WATER
TO H-1410C/E/F/J
PFD-61713-A1402

STEAM
FROM H-1424
PFD-61713-A1401

STEAM
FROM H-1410C/E/F/J
PFD-61713-A1402

661A

T-607

S-661

601B

THERMOCHEMICAL CONVERSION OF BIOMASS
TO FUELS AND CHEMICALS
GASIFICATION

PROCESS FLOW DIAGRAM
AREA 600: STEAM SYSTEM & POWER GENERATION

1	12/5/14	DMS	ADDED HX NETWORK	
0	9/26/14	JAR	ISSUED FOR DESIGN REPORT	
Rev.	Date	By	Description	

Project No. 30482.00

Drawing PFD-61713-A603

Rev. 1

PFD-61713-A603

COMPONENT	UNITS	644	606	607	604	601A	600	661A	662	601B	661	1400WW3	1400WW2	660A
Total Flow	lb/hr	408,807	404,719	404,719	4,088	4,088	5,252	58,184	57,020	1,164	9,494	18,580	30,110	58,184
Temperature	F	462	460	1,000	462	111	134	330	330	212	71	267	168	154
Pressure	Psia	475.9	465.9	465.9	475.9	14.7	14.7	103.1	103.1	14.7	14.7	134.0	72.0	103.6
Vapor Fraction		1	1	1	-	-	-	1	1	-	-	-	-	-
Hydrogen (H2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Carbon Monoxide (CO)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Nitrogen (N2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Oxygen (O2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Argon (Ar)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Carbon Dioxide (CO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Water (H2O)	lb/hr	408,807	404,719	404,719	4,088	4,088	5,252	58,184	57,020	1,164	9,494	18,580	30,110	58,184
Hydrogen Sulfide (H2S)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Methane (CH4)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Ethane (C2H6)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Ethylene (C2H4)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Acetylene (C2H2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Propane (C3H8)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Propene (C3H6)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
I-Butane (C4H10)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
N-Butane (C4H10)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Butenes (C4H8)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Pentanes (C5H12)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Pentenes (C5H10)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Heptanes (C7H16)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Triptane	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Triptene	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Octanes (C8H18)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Octenes (C8H16)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
N-Nonanes (C9H20)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Tar (C10H8)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Carbon (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Methanol (CH4O)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Dimethyl ether (C2H6O)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Ash	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Char	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Wood	lb/hr	-	-	-	-	-	-	-	-	-	-	-	-	-
Enthalpy Flow	MMBTU/hr	(2,319)	(2,293)	(2,165)	(26)	(28)	(36)	(332)	(324)	(8)	(65)	(123)	(205)	(393)

124

Process Flow Diagram — Area 700: Cooling Water & Other Utilities

Equipment labels (top):

Tag	Description
P-708	FLUE GAS SCRUBBER CIRCULATION PUMP
M-703	FLUE GAS SCRUBBER
M-707	SAND FILTER
M-704	FLUE GAS STACK
P-710	CLARIFIER OVERFLOW PUMP
T-710	CLARIFIER OVERFLOW TANK
M-705	CLARIFIER
M-709	COOLING TOWER RO UNIT
T-709	SLURRY TANK
P-709	SLURRY PUMP
P-701	COOLING WATER PUMP
M-701	COOLING TOWER SYSTEM
M-706	BELT PRESS

Stream connections (left side):

- 336 — SOLIDS FROM P-301 PFD-61713-A302
- 314 — SCRUBBER COND FROM P-314 PFD-61713-A302
- 600 — BLOWDOWN FROM HP-1404/S-661 PFD-61713-A603
- 315 — CONDENSATE FROM P-315 PFD-61713-A303
- RECYCLED BRINE FROM M-604 PFD-61713-A601
- 112 — FLUE GAS FROM K-101 PFD-61713-A102
- NAOH FROM P-705 PFD-61713-A702
- 618 — MAKE-UP WATER TO P-601 PFD-61713-A601
- MAKE-UP WATER

Stream connections (right side):

- FLOCCULANT/ COAGULANT
- SOLIDS TO DISPOSAL
- COOLING TOWER CHEMICALS
- CWS TO HEADER
- CWR FROM HEADER
- RO BRINE TO M-604 PFD-61713-A601

Other labels: ATM, DRIFT, EVAP, WM701, WP701, 718, 715, 713

CW USERS

H-301	H-504B
H-310ACW	H-593
H-310BCW	H-1414
H-310CCW	H-1503C
H-310DCW	H-1504C
H-414	H-1504D
	H-1511

PFD-61713-A701

COMPONENT	UNITS	336	314	112	600	315	618	713	715	718
Total Flow	lb/hr	101	75,638	30,000	5,252	20,329	81,739	11,291	2,556,084	2,556,084
Temperature	F	130	130	66	134	136	60	90	90	110
Pressure	Psia	34.7	35.0	15.7	14.7	35.0	14.7	14.7	74.7	59.7
Vapor Fraction		-	-	-	-	-	-	-	-	-
Hydrogen (H2)	lb/hr	-	0	-	-	0	-	-	-	-
Carbon Monoxide (CO)	lb/hr	-	0	-	-	0	-	-	-	-
Nitrogen (N2)	lb/hr	-	0	-	-	0	-	-	-	-
Oxygen (O2)	lb/hr	-	-	-	-	-	-	-	-	-
Argon (Ar)	lb/hr	-	-	-	-	-	-	-	-	-
Carbon Dioxide (CO2)	lb/hr	-	1	-	-	0	-	-	-	-
Water (H2O)	lb/hr	50	75,632	30,000	5,252	20,327	81,739	11,291	2,556,084	2,556,084
Hydrogen Sulfide (H2S)	lb/hr	-	0	-	-	0	-	-	-	-
Sulfur dioxide (SO2)	lb/hr	-	-	-	-	-	-	-	-	-
Ammonia (NH3)	lb/hr	-	5	-	-	1	-	-	-	-
Nitrogen dioxide (NO2)	lb/hr	-	-	-	-	-	-	-	-	-
Methane (CH4)	lb/hr	-	0	-	-	0	-	-	-	-
Ethane (C2H6)	lb/hr	-	0	-	-	0	-	-	-	-
Ethylene (C2H4)	lb/hr	-	0	-	-	0	-	-	-	-
Acetylene (C2H2)	lb/hr	-	0	-	-	0	-	-	-	-
Propane (C3H8)	lb/hr	-	0	-	-	0	-	-	-	-
Propene (C3H6)	lb/hr	-	-	-	-	-	-	-	-	-
I-Butane (C4H10)	lb/hr	-	0	-	-	0	-	-	-	-
N-Butane (C4H10)	lb/hr	-	0	-	-	0	-	-	-	-
Butenes (C4H8)	lb/hr	-	-	-	-	-	-	-	-	-
Pentanes (C5H12)	lb/hr	-	0	-	-	0	-	-	-	-
Pentenes (C5H10)	lb/hr	-	-	-	-	-	-	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-	-	-	-	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-	-	-	-	-	-	-
Benzene (C6H6)	lb/hr	-	0	-	-	0	-	-	-	-
Heptanes (C7H16)	lb/hr	-	-	-	-	-	-	-	-	-
Triptane	lb/hr	-	-	-	-	-	-	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-	-	-	-	-	-	-
Triptene	lb/hr	-	-	-	-	-	-	-	-	-
Octanes (C8H18)	lb/hr	-	-	-	-	-	-	-	-	-
Octenes (C8H16)	lb/hr	-	-	-	-	-	-	-	-	-
N-Nonanes (C9H20)	lb/hr	-	-	-	-	-	-	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-	-	-	-	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-	-	-	-	-	-	-
Tar (C10H8)	lb/hr	-	0	-	-	0	-	-	-	-
Carbon (Solid)	lb/hr	-	-	-	-	-	-	-	-	-
Sulfur (Solid)	lb/hr	-	-	-	-	-	-	-	-	-
Olivine + MgO (Solid)	lb/hr	47	-	-	-	-	-	-	-	-
Methanol (CH4O)	lb/hr	-	0	-	-	0	-	-	-	-
Dimethyl ether (C2H6O)	lb/hr	-	-	-	-	-	-	-	-	-
MDEA (C5H13NO2)	lb/hr	-	-	-	-	-	-	-	-	-
Ash	lb/hr	0	-	-	-	-	-	-	-	-
Char	lb/hr	4	-	-	-	-	-	-	-	-
Wood	lb/hr	0	-	-	-	-	-	-	-	-
Enthalpy Flow	MMBTU/hr	(0)	(517)	(205)	(36)	(139)	(559)	(78)	(17,591)	(17,532)

126

Column headers (equipment tags):

| T-702 FIREWATER STORAGE TANK | T-703 PURCHASED DIESEL STORAGE TANK | T-704 AMMONIA STORAGE TANK | T-705 CAUSTIC STORAGE TANK | T-708 BOILER FEEDWATER CHEMICAL STORAGE TANK | M-702 HYDRAULIC TRUCK DUMP WITH SCALE | T-707 TAR REFORMER CATALYST LOADING SYSTEM | T-706 OLIVINE/MGO LOADING SYSTEM | P-707 BOILER FEEDWATER CHEMICAL PUMP | P-702 FIREWATER PUMP | P-703 DIESEL PUMP | P-704 AMMONIA PUMP | P-705 CAUSTIC PUMP |

WATER → T-702 → P-702 → TO FIRE SUPPRESSION SYSTEM

PURCHASED DIESEL → T-703 → P-703 → FRONT-END LOADERS

AMMONIA → T-704 → P-704 → AMMONIA TO T-603 PFD-61713-A601

NAOH → T-705 → P-705 → NAOH TO M-703 PFD-61713-A701

BOILER FEEDWATER CHEMICALS → T-708 → P-707 → BFW CHEMICALS TO T-603 PFD-61713-A601

OLIVINE+MGO — M-702

ATM

CO2 FROM K-102 PFD-61713-A102

TAR REFORMING CATALYST

ATM

T-706

T-707

221 OLIVINE TO R-202 PFD-61713-A201

220 MGO TO R-202 PFD-61713-A201

326 MAKEUP CATALYST TO R-303 PFD-61713-A301

THERMOCHEMICAL CONVERSION OF BIOMASS TO FUELS AND CHEMICALS GASIFICATION

PROCESS FLOW DIAGRAM AREA 700: MATERIALS STORAGE

Rev.	Date	By	Description
1	12/5/14	DMS	ADDED HX NETWORK
0	9/26/14	JAR	ISSUED FOR DESIGN REPORT

Project No: 30482.00
Drawing: PFD-61713-A702
Rev: 1

Harris Group

NREL NATIONAL RENEWABLE ENERGY LABORATORY

PFD-61713-A702

COMPONENT	UNITS	221	220	326
Total Flow	lb/hr	543	7	9
Temperature	F	60	60	-
Pressure	Psia	31.0	31.0	35.0
Vapor Fraction		1	1	-
Hydrogen (H2)	lb/hr	-	-	-
Carbon Monoxide (CO)	lb/hr	-	-	-
Nitrogen (N2)	lb/hr	-	-	-
Oxygen (O2)	lb/hr	-	-	-
Argon (Ar)	lb/hr	-	-	-
Carbon Dioxide (CO2)	lb/hr	-	-	-
Water (H2O)	lb/hr	-	-	-
Hydrogen Sulfide (H2S)	lb/hr	0	0	-
Sulfur dioxide (SO2)	lb/hr	-	-	-
Ammonia (NH3)	lb/hr	-	-	-
Nitrogen dioxide (NO2)	lb/hr	-	-	-
Methane (CH4)	lb/hr	-	-	-
Ethane (C2H6)	lb/hr	-	-	-
Ethylene (C2H4)	lb/hr	-	-	-
Acetylene (C2H2)	lb/hr	-	-	-
Propane (C3H8)	lb/hr	-	-	-
Propene (C3H6)	lb/hr	-	-	-
I-Butane (C4H10)	lb/hr	-	-	-
N-Butane (C4H10)	lb/hr	-	-	-
Butenes (C4H8)	lb/hr	-	-	-
Pentanes (C5H12)	lb/hr	-	-	-
Pentenes (C5H10)	lb/hr	-	-	-
Hexanes (C6H14)	lb/hr	-	-	-
Hexenes (C6H12)	lb/hr	-	-	-
Benzene (C6H6)	lb/hr	-	-	-
Heptanes (C7H16)	lb/hr	-	-	-
Triptane	lb/hr	-	-	-
Heptenes (C7H14)	lb/hr	-	-	-
Triptene	lb/hr	-	-	-
Octanes (C8H18)	lb/hr	-	-	-
Octenes (C8H16)	lb/hr	-	-	-
N-Nonanes (C9H20)	lb/hr	-	-	-
N-Nonenes (C9H18)	lb/hr	-	-	-
Hexamethylbenzene (C6(CH3)6)	lb/hr	-	-	-
Tar (C10H8)	lb/hr	-	-	-
Carbon (Solid)	lb/hr	-	-	-
Sulfur (Solid)	lb/hr	-	-	-
Olivine + MgO (Solid)	lb/hr	543	7	9
Methanol (CH4O)	lb/hr	-	-	-
Dimethyl ether (C2H6O)	lb/hr	-	-	-
MDEA (C5H13NO2)	lb/hr	-	-	-
Ash	lb/hr	-	-	-
Char	lb/hr	-	-	-
Wood	lb/hr	-	-	-
Enthalpy Flow	MMBTU/hr	(0)	(0)	(0)

128

M-708
EMERGENCY
FLARE

K-701
PLANT AIR
COMPRESSOR

S-701
INSTRUMENT
AIR DRYER

T-701
PLANT AIR
RECEIVER

ATM

INSTRUMENT AIR
TO DISTRIBUTION

S-701   T-701

K-701

PLANT AIR
TO DISTRIBUTION

EMERGENCY FLARE
HEADER

M-708

EMERGENCY FLARE SYSTEM

PLANT AND INSTRUMENT AIR SYSTEM

				THERMOCHEMICAL CONVERSION OF BIOMASS TO FUELS AND CHEMICALS GASIFICATION
				PROCESS FLOW DIAGRAM AREA 700: AIR AND FLARE UTILITIES
0	9/26/14	JAR	ISSUED FOR DESIGN REPORT	Project No: 30482.00 Drawing: PFD-61713-A703 Rev: 0
Rev.	Date	By	Description	

Harris Group

NREL
NATIONAL RENEWABLE ENERGY LABORATORY

Heat Exchanger Duty Summary

Process Flow Diagram (PFD)	Equip ID	Heat Exchanger Service	Duty ID	Duty (MMBTU/Hr)
Area A100: Feed Handling & Preparation				
PFD-61713-A102	M-105	Cross Flow Pellet Dryer	QM105	16.5
Area A200: Gasification				
PFD-61713-A201	M-201	Sand / Ash Cooler	QCM201	0.9
PFD-61713-A202	HP-201	Char Combustor Flue Gas / Hydrocarbon Synthesis Reactor Feed Exchanger	QHP201	3.0
	HP-202	Char Combustor Flue Gas / DME Reactor Feed Exchanger	QHP202	15.6
	HP-203	Char Combustor Flue Gas / Methanol Synthesis Reactor Feed Exchanger	QHP203	24.9
	HP-204	Char Combustor Flue Gas / BFW Preheat Exchanger	QHP204	51.4
	HP-205	Char Combustor Flue Gas Steam Superheater	QHP205	87.6
	HP-206	Char Combustor Flue Gas Steam Generator	QHP206	22.2
Area A300: Syngas Cleanup and Compression				
PFD-61713-A302	H-301	Quench Water Recirculation Cooler	QCH301	6.8
	H-301B	Post-reformer Air Cooler	QAH301B	78.7
PFD-61713-A303	HP-301	Compressor Interstage / Methanol Synthesis Reactor Feed Exchanger	QHP301	19.0
	HP-302	Compressor Interstage / BFW Preheat Exchanger	QHP302	1.5
	HP-303	Compressor Interstage / Methanol Synthesis Reactor Feed Exchanger	QHP303	15.3
	H-310AAC	1st Stage Air Intercooler	QH310AAC	22.9
	H-310BAC	2nd Stage Air Intercooler	QH310BAC	4.1
	H-310CAC	3rd Stage Air Intercooler	QH310CAC	6.5
	H-310DAC	4th Stage Air Intercooler	QH310DAC	4.3
	H-310ACW	1st Stage Water Intercooler	QH310ACW	5.1
	H-310BCW	2nd Stage Water Intercooler	QH310BCW	2.7
	H-310CCW	3rd Stage Water Intercooler	QH310CCW	2.8
	H-310DCW	4th Stage Water Intercooler	QH310DCW	2.1
PFD-61713-A304	HP-305	Tar Reformer Effluent / Catalyst Regenerator Air Preheat Exchanger	QHP305	2.0
	HP-306	Tar Reformer Effluent / Hydrocarbon Synthesis Reactor Feed Exchanger	QHP306	28.9
	HP-307	Tar Reformer Effluent / Char Combustor Air Preheat Exchanger	QHP307	16.7
	HP-308	Tar Reformer Effluent Steam Generator	QHP308	164.6
	HP-316	Tar Reformer Effluent / BFW Preheat Exchanger	QHP316	6.3
	HP-317	Steam Turbine Generator Exhaust / Catalyst Regenerator Air Preheat Exchanger	QHP317	1.2
	HP-318	Tar Reformer Effluent / BFW Preheat Exchanger	QHP318	0.3
	HP-319	Tar Reformer Effluent / Catalyst Regenerator Air Preheat Exchanger	QHP319	0.7
PFD-61713-A305	HP-309	Catalyst Regenerator Flue Gas / Steam Turbine Generator Inlet (Stage 2) Exchanger	QHP309	45.4
	HP-310	Catalyst Regenerator Flue Gas / Steam Turbine Generator Inlet (Stage 3) Exchanger	QHP310	19.5
	HP-311	Catalyst Regenerator Flue Gas / Purge Gas Expander Inlet Exchanger	QHP311	7.9
	HP-312	Catalyst Regenerator Flue Gas Steam Superheater	QHP312	38.3
	HP-313	Catalyst Regenerator Flue Gas / Gasifier Steam Preheat Exchanger	QHP313	4.0
	HP-314	Catalyst Regenerator Flue Gas Steam Superheater	QHP314	40.5
	HP-315	Catalyst Regenerator Flue Gas / Syngas Recycle Exchanger	QHP315	5.1
Area A400: Acid Gas Removal & Methanol Synthesis				
PFD-61713-A401	H-411B	Post-Reactor Syngas Air Cooler	QAH411	22.3
	H-413	Syngas Air Cooler	QAH413	32.7
	H-414	Methanol Condenser	QCH414	10.6
PFD-61713-A402	HP-401	Steam Blowdown / Purge Gas Expander Inlet Exchanger	QHP401	1.1
	HP-402	Steam Turbine Generator Exhaust / BFW Preheat Exchanger	QHP402	22.9
	HP-403	Methanol Synthesis Reactor Effluent / Catalyst Regenerator Air Preheat Exchanger	QHP403	2.0
	HP-404	Methanol Synthesis Reactor Effluent / BFW Preheat Exchanger	QHP404	2.8
	HP-405	Methanol Synthesis Reactor Effluent / BFW Preheat Exchanger	QHP405	48.5
Area A500: Methanol Conditioning				
PFD-61713-A501	HP-501	Char Combustor Flue Gas / Methanol Degassing Column Reboiler Exchanger	QHP501	9.1
	HP-502	DME Reactor Effluent / Methanol Degassing Column Reboiler Exchanger	QHP502	7.7
	H-504B	Expanded Methanol Cooler B	QCH504B	3.2
PFD-61713-A502	HP-503	Methanol Intermediate / BFW Preheat Exchanger	QHP503	4.9
	HP-504	Methanol Intermediate / Light Hydrocarbon Recycle Exchanger	QHP504	0.9
	HP-505	Methanol Intermediate / Catalyst Regenerator Air Preheat Exchanger	QHP505	0.7
	HP-506	Methanol Intermediate / Purge Gas Expander Inlet Exchanger	QHP506	0.5
	HP-507	Methanol Intermediate / AGR Acid Gas Exchanger	QHP507	1.3
	H-593	Methanol Product Finishing Cooler	QCH593	0.4
Area A1400: Methanol to High Octane Gasoline Synthesis				
PFD-61713-A1401	H-1424	DME Reactor Intercooler	QH1424	8.8
PFD-61713-A1402	H-1471	Post DME Reactor Air Cooler	QAH1471	18.4
	H-1410C	Hydrocarbon Synthesis Reactor Effluent Cooler	QH1410C	10.4
	H-1410E	Hydrocarbon Synthesis Reactor 1st Stage Intercooler	QH1410E	18.2
	H-1410F	Hydrocarbon Synthesis Reactor 2nd Stage Intercooler	QH1410F	15.6
	H-1410J	Hydrocarbon Synthesis Reactor 3rd Stage Intercooler	QH1410J	13.7
PFD-61713-A1403	HP-1404	Steam Blowdown / Unconverted DME Purge	QHP1404	0.2
	H-1413	Hydrocarbon Synthesis Reactor Effluent Air Cooler	QAH1413	36.9
	H-1414	Hydrocarbon Synthesis Reactor Effluent Water Cooler	QCH1414	8.3
PFD-61713-A1404	HP-1401	Unconverted DME Recycle / DME Reactor Feed Exchanger	QHP1401	10.1
	HP-1402	Hydrocarbon Synthesis Reactor Effluent / Catalyst Regenerator Air Preheat Exchanger	QHP1402	3.9
	HP-1403	Hydrocarbon Synthesis Reactor Effluent / DME Reactor Feed Exchanger	QHP1403	20.4
	HP-1405	Unconverted DME Recycle / Deethanizer Feed Preheat Exchanger	QHP1405	0.5
	HP-1406	Unconverted DME Recycle / DME Reactor Feed Exchanger	QHP1406	1.8
	HP-1407	Unconverted DME Recycle / Light Hydrocarbon Recycle Exchanger	QHP1407	0.1
Area A1500: Product Recovery				
PFD-61713-A1501	HP-1503	Tar Reformer Effluent / Debutanizer Reboiler Exchanger	HP-1503	4.1
	H-1503C	Deethanizer Column Condenser	QCH1503C	10.6
PFD-61713-A1502	HP-1501	High-Octane Gasoline Blendstock Product / DME Reactor Feed Exchanger	QHP1501	3.2
	HP-1502	High-Octane Gasoline Blendstock Product / Deethanizer Feed Preheat Exchanger	QHP1502	1.1
	HP-1504	Char Combustor Flue Gas / Deethanizer Reboiler Exchanger	QHP1504	13.5
	H-1504C	Debutanizer Column Condenser	QCH1504C	3.3
	H-1504D	Debutanizer Overhead Product Condenser	QCH1504D	0.9
	H-1511	High Octane Gasoline Product Water Cooler	QCH1511	0.6
Area A600: Steam System & Power Generation				
PFD-61713-A602	H-601	Steam Turbine Condenser	QAH601	181.9

130

Power Consumption and Generation Summaries with Contributions from Major Rotating Equipment

POWER CONSUMPTION SUMMARY

Process Flow Diagram (PFD)	Equip ID	Work / Power Consumption Service	Work ID	Power (HP)
Area A100: Feed Handling & Preparation			**A100 Total**	**995**
Note: Power requirements for this process area are captured in the delievered feedstock costs assessed by Idaho National Laboratory				
Area A200: Gasification			**A200 Total**	**8,528**
PFD-61713-A201	K-202	Char Combustor Air Blower	WK202	7,684
Area A300: Syngas Cleanup and Compression			**A300 Total**	**26,189**
PFD-61713-A301	K-305	Combustion Air Blower	WK305	991
PFD-61713-A302	P-301	Sludge Pump	WP301	0
	P-302	Quench Water Recirculation Pump	WP302	11
	P-314	Condensate Pump	WP314	2
PFD-61713-A303	K-310	Synthesis Gas Compressor	WK310	24,464
	P-315	Syngas Compressor Condensate Pump	WP315	1
Area A400: Acid Gas Removal & Methanol Synthesis			**A400 Total**	**7,602**
PFD-61713-A401	K-410	Syngas 2nd compressor	WK410	3,514
	K-414	Unreacted Syngas Recycle Compressor	WK414	3,294
PFD-61713-A403/A404	S-410/M-405	Amine Acid Gas Removal System / LO-CAT Sulfur Recovery System	WAGR	472
Area A500: Methanol Conditioning			**A500 Total**	**3**
Area A1400: Methanol to High Octane Gasoline Synthesis			**A1400 Total**	**3,227**
PFD-61713-A1401	P-1000	Methanol Intermediate Pump	WP1000	29
PFD-61713-A1402	K-1410	Coke Burn / Cat Regeneration Air Blower	WK1410	20
PFD-61713-A1403	K-1472	Hydrocarbon Synthesis Reactor Recycle Compressor	WK1472	2,699
Area A1500: Product Recovery			**A1500 Total**	**48**
PFD-61713-A1501	P-1503	Crude Hydrocarbons Pump	WP1503	38
Area A600: Steam System & Power Generation			**A600 Total**	**1,038**
PFD-61713-A601	P-601	Make-Up Pump	WP601	2
	P-603	EDI Pump	WP603	14
	P-604	Boiler Feed Water Pump	WP604	341
PFD-61713-A602	P-602	Condensate Pump	WP602	7
PFD-61713-A603	P-661	Intercooler Charge Pump	WP661	11
Area A700: Cooling Water & Other Utilities			**A700 Total**	**481**
PFD-61713-A701	M-701	Cooling Tower System	WM701	144
	P-701	Cooling Water Pump	WP701	277
Allocation for Miscellaneous In-Plant Power Demands			**Misc. Total**	**1,341**
TOTAL IN-PLANT POWER CONSUMPTION				**49,451**

POWER GENERATION SUMMARY

Process Flow Diagram (PFD)	Equip ID	Work / Power Consumption Service	Work ID	Power (HP)
Area A400: Acid Gas Removal & Methanol Synthesis			**A400 Total**	**(2,059)**
PFD-61713-A401	K-412	Purge Gas Expander Stage 1	WK412	(2,059)
Area A600: Steam System & Power Generation			**A600 Total**	**(47,397)**
PFD-61713-A602	M-602 A/B/C	Extraction Steam Turbine / Generator Stages 1 - 3	WM602	(47,397)
TOTAL IN-PLANT POWER GENERATION				**(49,456)**

131

Appendix E. Synthesis Gas and Char Correlations for Indirect Gasifier

The gasifier is modeled by correlations based on data from the Battelle Columbus Laboratory (BCL) 9 tonne per day test facility. The data and original correlations for the gasifier can be found in Bain [1]. Experimental runs were performed for several different wood types including red oak, birch, maple, and pine chips, sawdust, and other hard and soft wood chips. The original pilot plant data for these runs can be found in Feldmann et al. [2]. The temperature range for the data is 1,280°F to 1,857°F and the pressure range is 2.4 to 14.4 psig; the majority of the data are in the range of 1,500°F to 1,672°F.

The BCL test facility's gas production data was correlated to gasifier temperature with a quadratic function in the following form:

$$X = a + bT + cT^2$$

where the temperature, T, is in units of °F. The coefficients a, b, and c, as well as the units for the correlated variable, are shown in Table E-1. Even though there is a correlation for the char formation, it is not used; instead the amount and elemental analysis for the char is determined by mass differences between the produced syngas and the converted biomass.

Table E-1. Coefficients for Indirect Gasifier Correlations

Variable	a	b	c	Units
Dry syngas	28.993	-0.043325	0.000020966	scf gas/lb MAF wood[a]
CO	133.46	-0.102900	0.000028792	mole % dry gas
CO_2	-9.5251	0.037889	-0.000014927	mole % dry gas
CH_4	-13.82	0.044179	-0.000016167	mole % dry gas
C_2H_4	-38.258	0.058435	-0.000019868	mole % dry gas
C_2H_6	11.114	-0.011667	0.000003064	mole % dry gas
H_2	17.996	-0.026448	0.000018930	mole % dry gas
C_2H_2	-4.3114	0.0054499	-0.000001561	mole % dry gas
Tar	0.045494	-0.000019759		lb/lb dry wood

[a] Scf = standard cubic feet. The standard conditions are 1 atm pressure and 60°F temperature.

The following general procedure is used for the gasifier production:

- A gasifier temperature T is assumed.

- The mass and molar amounts of carbon, hydrogen, oxygen, sulfur, nitrogen, and ash (as a pseudo-element) are determined from the biomass's ultimate analysis.

- The amount of syngas and its composition is determined from the gasifier correlations.

- The amount of carbon in the syngas and tar is determined. Residual carbon is parsed in the char.

- The amount of oxygen in the syngas is determined. A minimum amount of oxygen is required to be parsed to the char (4% of biomass oxygen). If there is a deficit of oxygen, then the associated water is decomposed to make sure that this amount of oxygen is parsed to the char; if there is excess oxygen, then it is parsed to the char without decomposing hydrogen.

- A set amount of sulfur is parsed to the char (8.3%). All remaining sulfur is set as H_2S in the syngas.

- A set amount of nitrogen is parsed to the char (6.6%). All remaining nitrogen is set as NH_3 in the syngas.

- The amount of hydrogen in the syngas (including tar, H_2S, NH_3, and decomposed water) is determined. All remaining hydrogen is parsed to the char.

- All ash is parsed to the char.

- The heat of formation of the char is estimated from the resulting ultimate analysis from this elemental material balance.

- The gasifier temperature is adjusted so that there is no net heat for an adiabatic reaction.

The syngas amount and composition will be dependent upon the biomass composition and the gasifier temperature. As an example, the resulting syngas composition for the woody biomass used in this design report can be seen in Figure E-1. Note from this figure that the amount of char decreases with increasing temperature and that the water does not start to decompose until high temperatures (here at 1,650°F and higher).
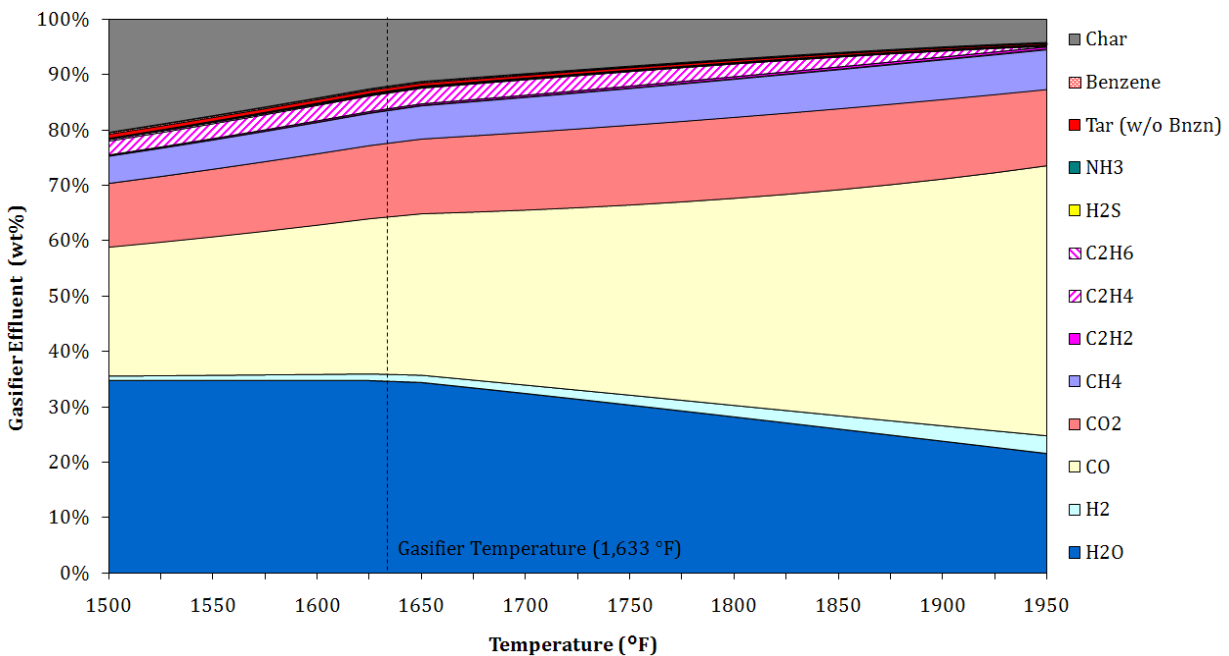
Figure E-1. Syngas composition for woody biomass used in design report

Appendix E References

1. Bain, R. *Material and Energy Balances for Methanol from Biomass Using Biomass Gasifiers*. NREL/TP-510-17098. Golden, CO: National Renewable Energy Laboratory, 1992. Available from: http://www.nrel.gov/docs/legosti/old/17098.pdf.
2. Feldmann, H.F.; Paisley, M.A.; Apelbaum, H.R.; Taylor, D.R. *Conversion of Forest Residues to a Methane-Rich Gas in a High-Throughput Gasifier*. PNL-6570. Columbus, OH: Battelle Columbus Division for the Pacific Northwest Laboratory, 1988. Available from: http://www.osti.gov/scitech/biblio/7150257.

Appendix F. Assumptions and Analysis for Full Life-Cycle GHG Emissions

To provide an initial estimate of life-cycle fuel GHGs associated with this conversion pathway, a preliminary life-cycle GHG analysis was performed for fuel produced via the high-octane gasoline pathway using the model 2017 blended feedstock (45% pulpwood, 32% wood residues, 3% switchgrass, and 20% construction and demolition waste). The data and sources used for each of the life-cycle stages are described in the following paragraphs, along with the results of the analysis.

Feedstock Crop Establishment and Growing

For the production of the purpose-grown bioenergy crops, pulpwood and switchgrass, emissions calculations from the GREET 2014 model [1] were aggregated and used in the SimaPro fuel life-cycle model [1]. Poplar was used to represent pulpwood in the feedstock blend. Switchgrass and poplar farming emissions were taken from the "EtOH" worksheet in the GREET model. Information on the development of GREET crop production parameters from Wang et al. was used to subtract the harvesting energy (which is accounted for in the next stage) from the GREET model total farming energy in order to calculate emissions for only the establishment and growing phase of crop production [2]. Table F-1 shows the total farming energy assumed in GREET (including harvesting) and the estimated energy value for establishment and growing operations ("Farming Energy Excluding Harvesting" in Table F-1) for this analysis.

Table F-1. Farming Energy Values From GREET and Modified for This Analysis

	GREET Farming Energy (including harvesting), Btu/dry Ton	Farming Energy Excluding Harvesting Estimated from GREET, Btu/dry Ton
Switchgrass*	201,995 (92.8% diesel;7.2% electricity)	56,870 (100% diesel)
Poplar	219,200 (100% diesel)	20,496 (100% diesel)

*Values include 14% dry matter loss.

The overall emissions from crop establishment and growing operations including fertilizer and herbicide application for switchgrass and poplar estimated from the GREET model data are 107,434 g CO_{2-e}/dry ton and 40,056 g CO_{2-e}/dry ton, respectively. For the production of the remaining feedstocks, wood residues, and construction/demolition waste, it is assumed that these are strictly waste products. As such, their life cycle begins at the waste pile and they do not receive any energy or GHG burdens from the upstream processes from which they are generated.

Feedstock Harvesting, Preprocessing, and Transportation

Jacobson et al. has estimated the GHG emissions associated with the logistics required for production of a conversion reactor-ready 2017 design case feedstock blend at 75.91 kg CO_{2-e}/dry U.S. ton [3]. This value includes energy and emissions associated with harvesting of purpose-grown crops (pulpwood and switchgrass), collection of wastes (wood residues and construction and demolition [C&D] waste), preprocessing and handling required at the landing, centralized feedstock processing facility (depot) and the biorefinery, and transportation of biomass from the

landing to the depot and from the depot to the biorefinery (see Figure 4, Section 3.1). About 70% of the feedstock logistics GHGs stems from preprocessing of the feedstocks [3]. The 2017 feedstock supply chain design incorporates several research advances that will significantly reduce energy use and cost compared to the current state of technology. More detail on the individual process steps and associated energy requirements for the feedstock supply system design can be found in the recent design report published by INL [3]. The aggregated value of 75.91 kg CO_{2-e}/dry U.S. ton of conversion plant feedstock was used in the SimaPro model for calculation of the full fuel life-cycle emissions.

Conversion

The inventory for the conversion stage of the life cycle is listed in Table 30. Assumptions for the conversion stage processes are given in Section 3.13.

Fuel Distribution and Consumption

Emissions associated with fuel distribution to the end user (fuel transportation and operation of storage tanks and fueling stations) are modeled using an EcoInvent database [4] process ("petrol, unleaded, at refinery/kg/RER/U"). Emissions of biogenic methane and N_2O from combustion of gasoline and diesel fuel in a vehicle are adapted from GREET [1].

Life-Cycle GHG Results

Figure F-1 shows the full life-cycle GHGs estimated for gasoline produced from this technology pathway. Emissions are 14.6 g CO_{2-e}/MJ, corresponding to a GHG reduction of 84% as compared to the 2005 petroleum gasoline baseline of 93.08 g CO_{2-e}/MJ [5]. This preliminary analysis suggests that this fuel meets the EISA RFS cellulosic biofuel definition (60% GHG reduction); however, final qualification is made by the EPA based on its own analysis and determination. In addition, through collaboration with Argonne National Laboratory, conversion energy and materials inventory from this work will be integrated into the GREET model for more detailed life-cycle analyses.

136

	2017 INL Blend
■ Fuel Use	0.49
■ Fuel Distribution	0.72
■ Conversion	0.57
■ Harvest, Collection and Preprocessing	10.0
■ Crop Establishment and Growing	2.8
Total	14.6
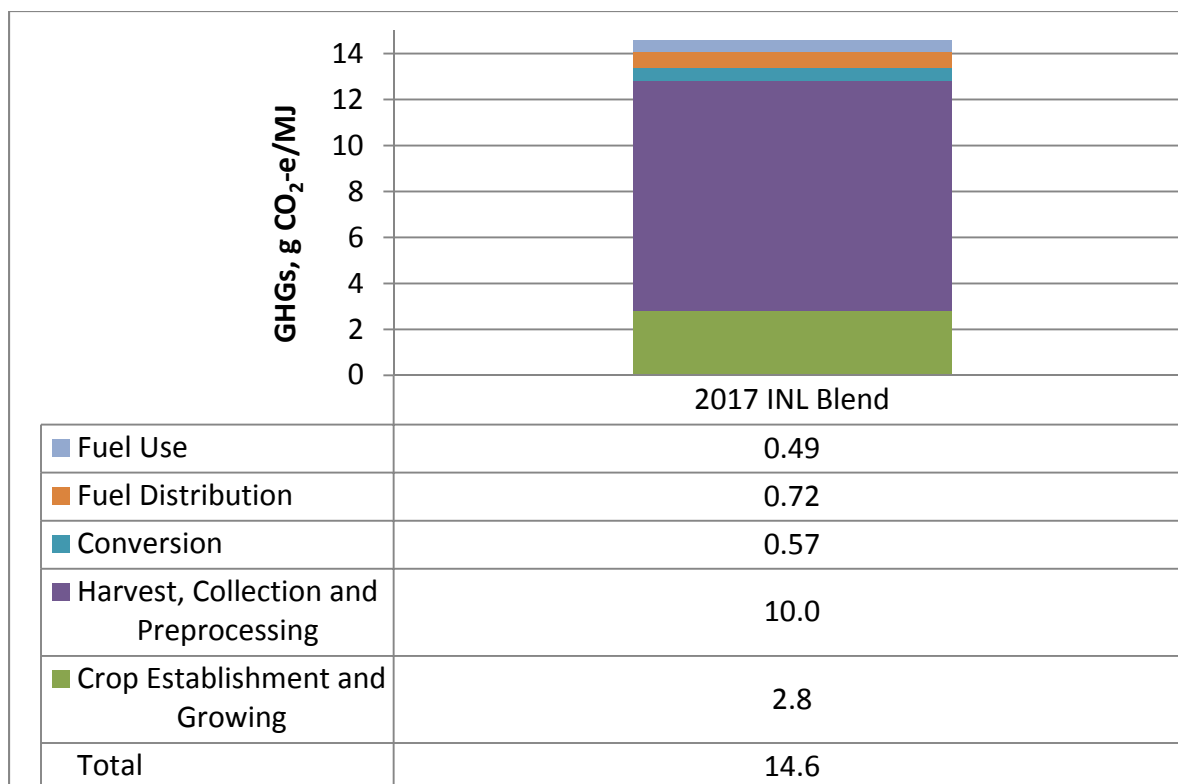
Figure F-1. Life-cycle GHG emissions for high-octane gasoline fuel from the INL 2017 blended feedstock (45% pulpwood, 32% wood residues, 3% switchgrass, and 20% construction and demolition waste) [3]

Co-Conversion of Biomass With Natural Gas

Co-conversion of biomass with natural gas can simultaneously increase fuel yields and reduce fuel production costs provided that the life-cycle GHG thresholds specified in EISA [6] is not violated. Figure F-2 illustrates that co-processing natural gas at the life-cycle GHG threshold limit (i.e., 60% GHG emissions reduction relative to the petroleum baseline) decreases the MFSP by 7%, from $3.41/GGE to $3.17/GGE. Note that Argonne National Laboratory will provide more detailed LCA for this pathway in the future, which may lead to slight variation of the results presented in Figure F-2.

137
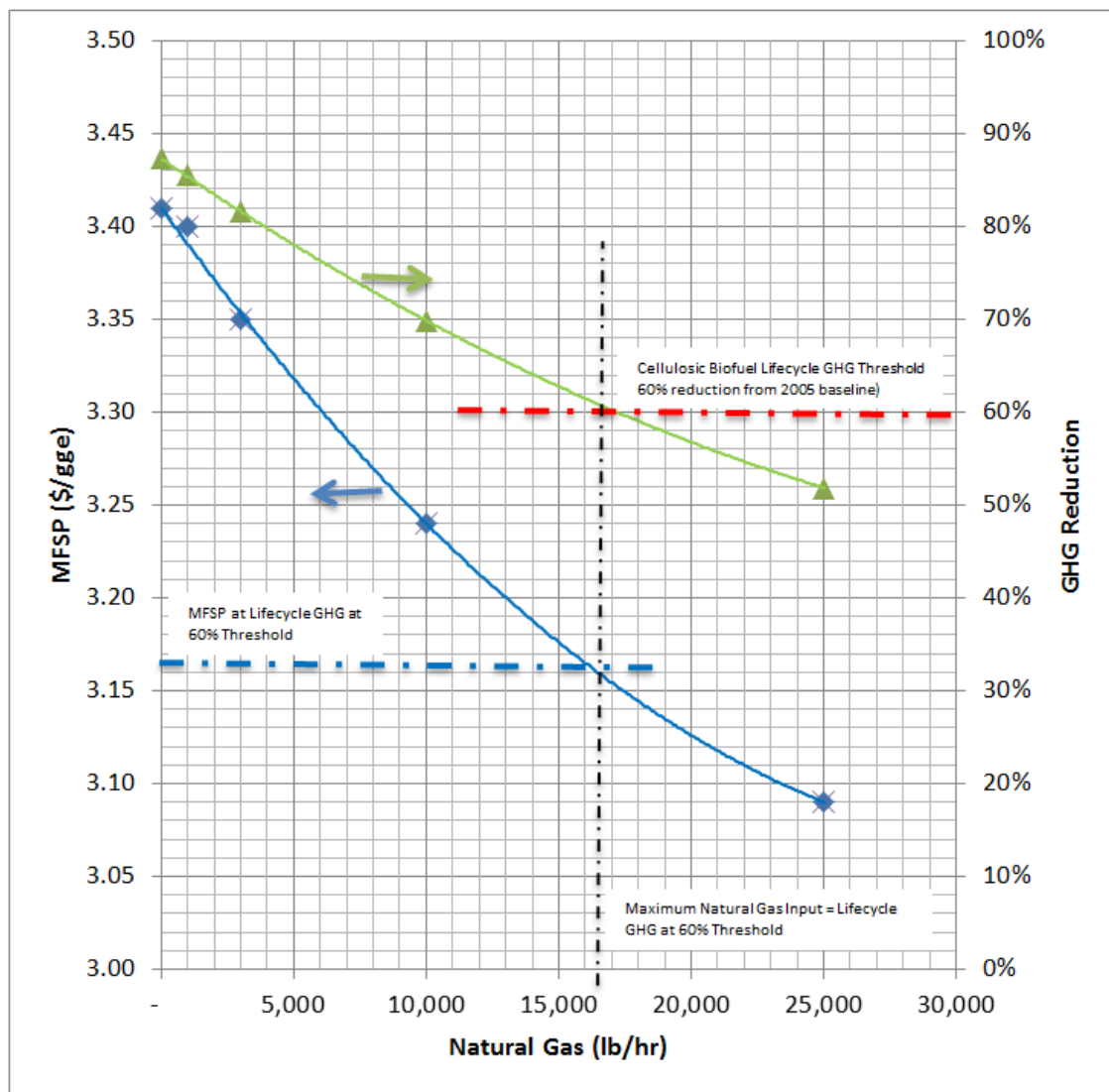
Figure F-2. MFSP and life-cycle GHG emissions for high-octane gasoline fuel production as a function of natural gas co-processing with biomass feedstock

Appendix F References

1. "The Greenhouse Gases, Regulated Emissions, and Energy use in Transportation (GREET) Model." Argonne, IL: Argonne National Laboratory, 2012. Accessed on July 1, 2013: http://greet.es.anl.gov.

2. Wang, Z.; Dunn, J.; Han, J.; Wang, M. *Material and Energy Flows in the Production of Cellulosic Feedstocks for Biofuels for the GREET Model.* ANL/ESD-13/9. Argonne, IL: Argonne National Laboratory, 2013. Available from: https://greet.es.anl.gov/files/feedstocks-13.

3. Jacobson, J.; Roni, M.S.; Cafferty, K.; Kenney, K.; Searcy, E.; Hansen, J. *Feedstock and Conversion Supply System Design and Analysis – "The Feedstock Logistics Design Case."* INL/EXT-14-33227. Idaho Falls, ID: Idaho National Laboratory, 2014. Available fromn: http://www.osti.gov/servlets/purl/1169237/.

4. "Ecoinvent, v.2.2." Duebendorf, Switzerland: Swiss Center for Life Cycle Inventories, 2010.

5. U.S. EPA. "Fuel-Specific Lifecycle Greenhouse Gas Emissions Results." Docket # EPA-HQ-OAR-2005-0161-3173. Washington, DC: U.S. Environmental Protection Agency, 2010. Available from: http://www.regulations.gov/#!documentDetail;D=EPA-HQ-OAR-2005-0161-3173.

6. Government Printing Office. "Supplemental Determination for Renewable Fuels Produced Under the Final RFS2 Program From Grain Sorghum." *Federal Register* (77: 242), December 17, 2012; pp. 74592–74607. Available from: http://www.gpo.gov/fdsys/pkg/FR-2012-12-17/pdf/2012-30100.pdf.

Appendix G. Fuel-Grade Gasoline Specifications

Specifications Contained in ASTM D 4814 Standard Specification for Gasoline
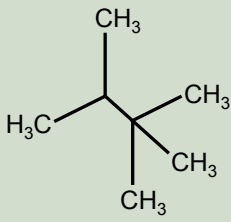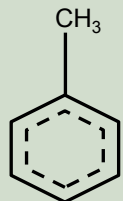
Property	ASTM Test Method	Spec UOM	Specification
Lead Content	D 3341 or D 5059	g/L, max	0.013
Silver Strip Corrosion Test	D 4814 Annex A1, 3h at 100°C	max	1
Copper Strip Corrosion	D 130, 3h at 50°C	max	No. 1
Solvent-Washed Gum Content	D 381	mg/100 mL, max	5
Sulfur	D 1266, D 2623, D 4294 or D 5453	mass%, max	0.008
Oxidation Stability	D 525	minutes, min	240
Water Tolerance	D 6422		Varies with region and season
Vapor Lock Protection Class Requirements	D 5188 or D 2533(for oxygenated containing fuels)	°C, min with V/L =20	Varies with region and season (between 54-35)
Vapor Pressure	D 4953, D 5190, D 5191, D 5482, or D 6378	kPa, max	Varies with region and season
Distillation Class Requirements	D 86		Varies with region and season

Appendix H. Differentiating the High-Octane Gasoline Blendstock Pathway From MTG

This appendix presents a comparative summary of the characteristics of the pathway from woody biomass to high-octane gasoline blendstock relative to those of the traditional methanol-to-gasoline process.

Process Attribute	High-Octane Gasoline Pathway	Methanol-to-Gasoline (MTG)	Impact on Techno-Economic Analysis
Molecular structures favored in synthesis reactions Example compound specific gravity	Branched paraffins CH_3 H_3C — C — CH_3 (with CH_3 and CH_3) Triptane 0.70	Aromatics CH_3 (benzene ring) Toluene 0.87	High-octane product rich in branched paraffins, similar to a refinery alkylate; H-saturation decreases density, increasing product volume.
Hydrocarbon synthesis catalyst	Beta-Zeolite (12-membered rings)	ZSM-5 (10-membered rings)	Different pore sizes and structures result in different compound selectivities.
Octane number of gasoline-range product	RON: 95+ MON: 90+	RON: 92 MON: 83	Octane number increases value of product as a finished fuel blendstock.
Selectivity of C_5+product	C_5+ product only (67.1 gal/Ton)	~ 85% C_5+ (55.1 gal/Ton)	High selectivity to primary (premium quality) product maximizes overall product value.
Severity of synthesis operating conditions	350–450 Deg. F 130 PSIA	650–950 Deg. F 315 PSIA	The lower severity operating conditions result in lower capital and operating costs relative to MTG.
Coke formation	Coke formation is minimized by hydrogen addition and selectivity to branched paraffins rather than aromatics.	High propensity for coke formation due to aromatic coke pre-cursors.	Minimizing coke formation helps to maximize product yield/carbon efficiency and maximizes catalyst regeneration and replacement cycles.

141

Appendix H References

1. Hindman, M. "Methanol to Gasoline (MTG) Technology: An Alternative for Liquid Fuel Production." Presented at World CTL Conference, April 13-16, 2010.
2. ExxonMobil Research and Engineering. *Methanol to Gasoline (MTG): Production of Clean Gasoline from Coal.* Available from: http://www.exxonmobil.com/Apps/RefiningTechnologies/files/sellsheet_09_mtg_brochure.pdf

Appendix I. Analysis of Premium Properties of High-Octane Gasoline Blendstock

This appendix presents the results of a simple analysis intended to show (1) the potential premium value of the high-octane gasoline blendstock from the process described in this design report and (2) a blending strategy for the blendstock based on the modeled distillation curve.

In the plot below, the estimated octane value (x-axis) and minimum fuel selling price (y-axis) are compared with octane values and wholesale prices of gasoline finished fuel blends and common gasoline blendstocks (with varying octane values). Based on the data in Figure I-1, the high-octane gasoline blendstock is at least cost-competitive with fossil blendstocks of comparable octane numbers at ~92 (RON+MON)/2 and $3.25 per gallon. In addition, the product has potential to demonstrate higher blended octane benefit, approaching 110 (RON+MON)/2. Further explanation of the two different data points for the high-octane gasoline blendstock is provided below. Analytical testing on actual products and blends with other gasoline components will serve to validate the analysis results.
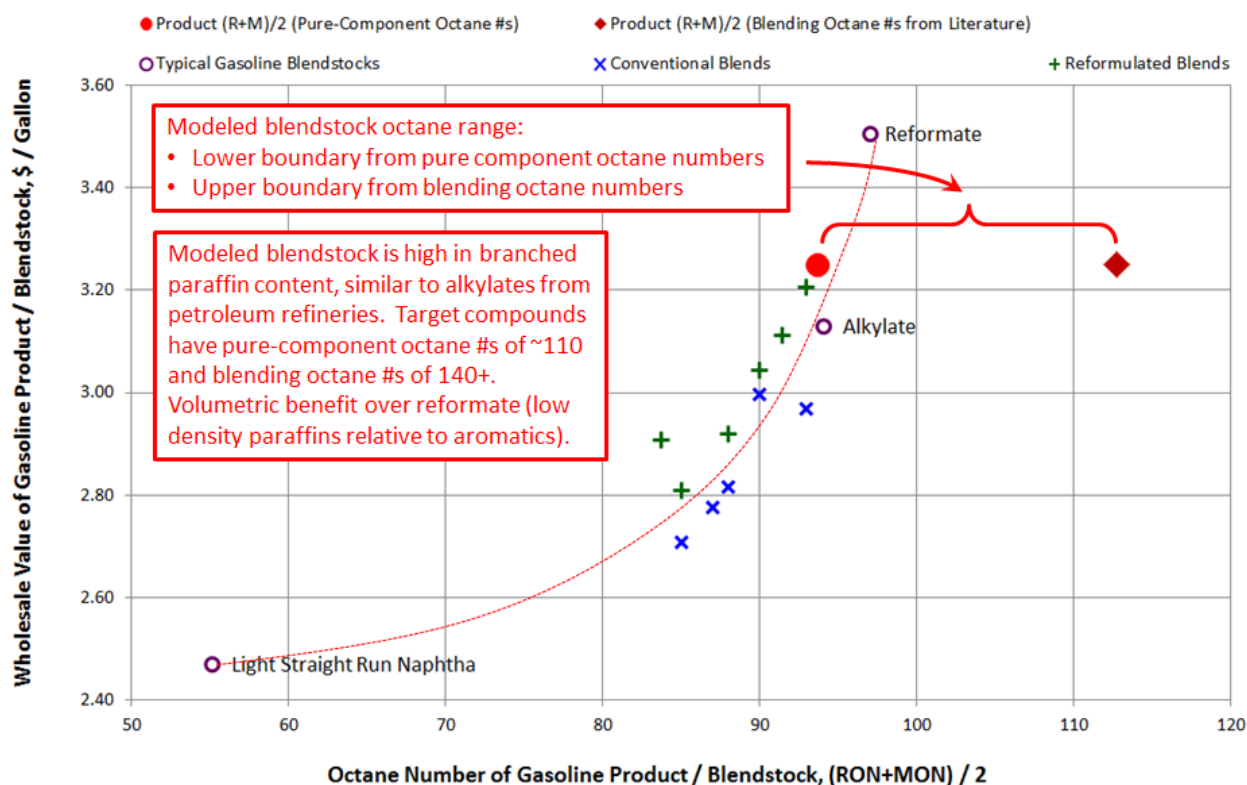
Figure I-1. Comparison of octane values and wholesale prices

There are two data points on the plot representing the high-octane gasoline blendstock. The first data point, labeled "Product (R+M)/2 (Pure-Component Octane #s)", represents the calculated weighted average ACTUAL octane value (RON+MON)/2 based on the molar composition of the high-octane gasoline blendstock product and published octane values for ACTUAL pure components.

143

The second data point, labeled "Product (R+M)/2 (Blending Octane #s from Literature)", represents the calculated weighted average BLENDING octane value (RON+MON)/2 based on the molar composition of the high-octane gasoline blendstock product and published octane values for BLENDING pure components.

Refer to Table I-1 for a simple example of blend calculations based on a three-component mixture of n-heptane, iso-octane (2,2,4-trimethyl pentane), and cyclopentane. Simple calculations of the weighted average ACTUAL RON (research octane number) and MON (motor octane number) yield an estimated 66.6 octane value (RON+MON)/2 for this blend. However, octane numbers do not blend linearly in practice and, as a result, analytical results for this particular blend show significantly higher (RON+MON)/2 results (approximately 76.2). The fact that the resulting blend octane number is higher than the calculated value based on ACTUAL pure-component values suggests that the BLENDING octane value of one of the components is greater than its ACTUAL value due to mixing interactions between various components in the blend. For the purposes of assessing BLENDING octane values, the octane numbers (RON and MON) for n-heptane and iso-octane (2,2,4-trimethyl pentane) are defined to remain constant at 0 and 100, respectively [1,2]. Analytical testing of 20% mixtures of various pure components with 32% n-heptane and 48% iso-octane, result in the BLENDING RON and MON values for each component. BLENDING octane values from literature [1,2] were applied to calculate the potential blending octane value for the high-octane gasoline blendstock of 110 (RON+MON)/2 shown in Figure I-1. The pectroluem-derived blendstock and finished fuel pricing data utilized for the figure are from EIA [3] and Platts [4]. As mentioned previously, analytical testing on actual products and blends with other gasoline components is required to validate these results.

Table I-1. Simple Example of (RON+MON)/2 Calculations Using Both ACTUAL and BLENDING Octane Values for Components in the Blend

Compound	Mole %	Actual RON	Actual MON	Blending RON	Blending MON
n-heptane	32.0%	0.0	0.0	0.0	0.0
iso-octane	48.0%	100.0	100.0	100.0	100.0
cyclopentane	20.0%	101.0	85.0	141.0	141.0
Total / Weighted Average	**100.0%**	**68.2**	**65.0**	**76.2**	**76.2**
(RON + MON) / 2			**66.6**		**76.2**

The plot presented in Figure I-2 below shows the modeled distillation curve for the high-octane gasoline blendstock compared with several seasonal average finished gasoline blends. The plot is intended to identify a simple blending strategy based solely on the distillation curve for the blendstock. Based on this simple analysis, the blending strategy for the blendstock is likely to include two major steps—(1) blend lighter naphtha blendstock to meet the maximum 10 vol % of 158°F in the finished fuel and (2) blend a heavier naphtha blendstock to meet minimum end point of 300°F in the finished fuel. The sources of quality information utilized for Figure I-2 are Colonial Pipeline Company [5] and Chevron [6].

144
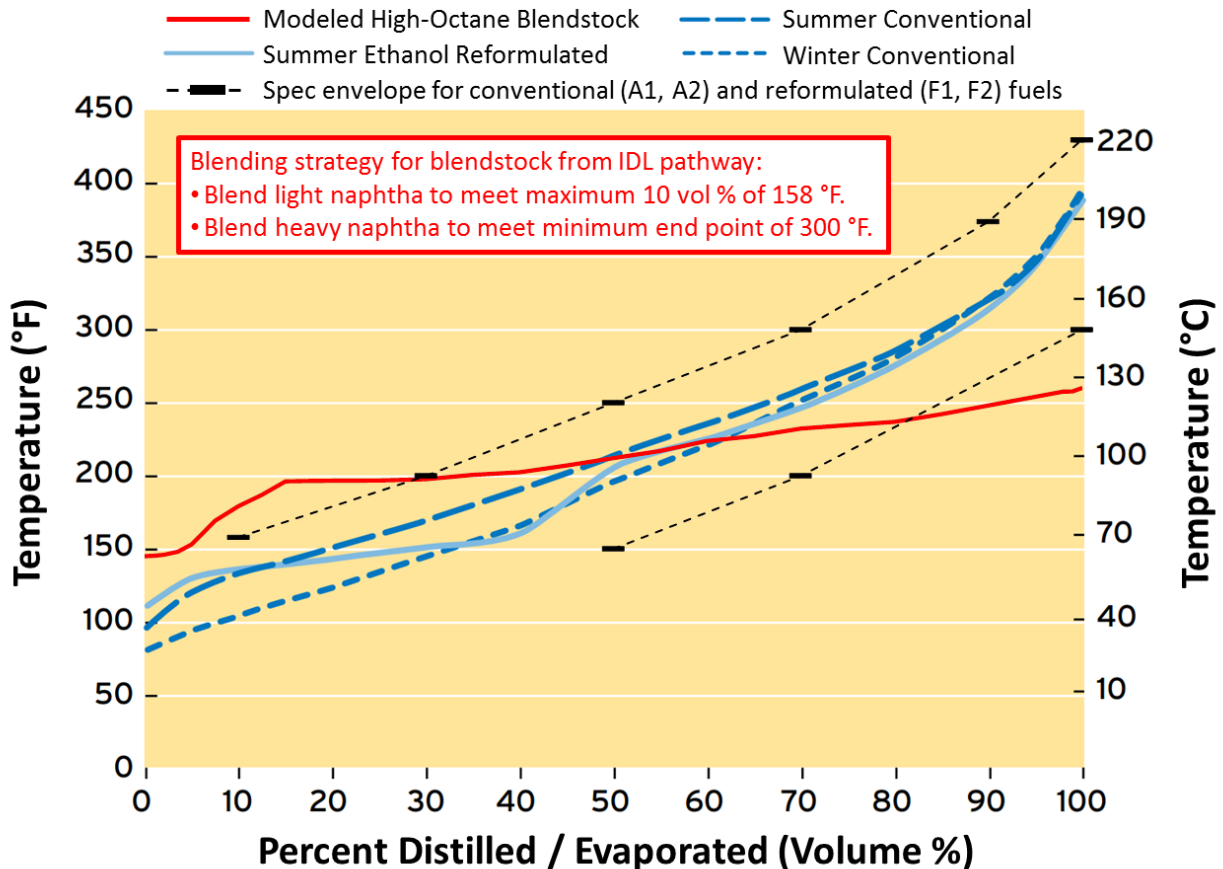
Sources: (1) Colonial Pipeline Company Product Codes and Specifications, (2) Chevron Motor Gasolines Technical Review

Figure I-2. Distillation curve comparison between target blendstock and finished gasolines

Appendix I References

1. Hobson, G.D., ed. *Modern Petroleum Technology.* 5th edition, Part II. New York, NY: Wiley, 1984.
2. "FCC Octane MON versus RON." Refining Online. Available from: http://www.refiningonline.com/engelhardkb/crep/TCR4_29.htm
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4. "Oilgram Price Report." Platts, July 11, 2013. Accessed on August 28, 2014: http://www.platts.com/products/oilgram-price-report.
5. "Product Codes and Specifications." Colonial Pipeline Company, March 2013. Available from: http://www.colpipe.com/docs/default-source/product-specs/product-specifications.pdf.
6. *Motor Gasolines Technical Review.* San Ramon, CA: Chevron Corporation, 2009. Available from: http://www.chevron.com/documents/pdf/MotorGasTechReview.pdf.

Appendix J. Pinch Analysis and Heat Integration System Design

Equip ID	Duty ID	Service	Duty	Hot In T	Hot Out T	Cold In T	Cold Out T	LMTD	U*	A	TPEC	TIC
			(MMBTU/Hr)	(Deg F)	(Deg F)	(Deg F)	(Deg F)	(Deg F)	(Btu/Hr-ft2-F)	(ft2)	(2011 $)	(2011 $)
HP-201	QHP201	Char Combustor Flue Gas / Hydrocarbon Synthesis Reactor Feed Exchanger	3.0	250.7	230.0	179.5	243.7	22.0	9.0	15,137	184,823	393,096
HP-202	QHP202	Char Combustor Flue Gas / DME Reactor Feed Exchanger	15.6	1,040.1	406.7	353.3	437.0	226.7	10.0	6,900	135,899	331,808
HP-203	QHP203	Char Combustor Flue Gas / Methanol Synthesis Reactor Feed Exchanger	24.9	1,040.1	406.7	325.9	437.0	259.8	30.0	3,191	82,819	274,144
HP-204	QHP204	Char Combustor Flue Gas / BFW Preheat Exchanger	51.4	1,040.1	406.7	345.5	464.1	229.6	105.0	2,130	47,218	199,532
HP-205	QHP205	Char Combustor Flue Gas Steam Superheater	87.6	1,797.6	1,040.1	459.8	1,000.0	683.2	25.0	5,129	350,141	755,068
HP-206	QHP206	Char Combustor Flue Gas Steam Generator	22.2	1,797.6	1,040.1	462.0	462.0	904.6	25.0	982	93,691	300,151
HP-301	QHP301	Compressor Interstage / Methanol Synthesis Reactor Feed Exchanger	19.0	383.9	194.2	172.8	325.9	36.7	35.0	14,774	231,722	522,813
HP-302	QHP302	Compressor Interstage / BFW Preheat Exchanger	1.5	194.2	179.0	148.1	172.8	25.9	105.0	550	19,719	103,070
HP-303	QHP303	Compressor Interstage / Methanol Synthesis Reactor Feed Exchanger	15.3	361.7	194.2	172.8	325.9	28.0	45.0	12,163	226,286	520,895
HP-304	QHP304	Compressor Interstage / BFW Preheat Exchanger	2.2	194.2	170.0	148.1	172.8	21.7	120.0	842	32,936	120,231
HP-305	QHP305	Tar Reformer Effluent / Catalyst Regenerator Air Preheat Exchanger	2.0	404.9	194.2	172.8	253.4	66.5	5.0	6,018	83,245	239,503
HP-306	QHP306	Tar Reformer Effluent / Hydrocarbon Synthesis Reactor Feed Exchanger	28.9	404.9	194.2	179.5	385.4	17.0	9.0	189,053	1,938,939	3,414,967
HP-307	QHP307	Tar Reformer Effluent / Char Combustor Air Preheat Exchanger	16.7	546.5	404.9	258.3	400.0	146.5	5.0	22,735	227,245	462,058
HP-308	QHP308	Tar Reformer Effluent Steam Generator	164.6	1,670.0	546.5	462.0	462.0	422.4	25.0	15,590	648,587	1,100,199
HP-309	QHP309	Catalyst Regenerator Flue Gas / Steam Turbine Generator Inlet (Stage 2) Exchanger	45.4	2,452.0	1,746.2	782.8	1,000.0	1,191.0	10.0	3,809	249,948	569,818
HP-310	QHP310	Catalyst Regenerator Flue Gas / Steam Turbine Generator Inlet (Stage 3) Exchanger	19.5	2,452.0	1,746.2	784.4	1,000.0	1,190.1	7.0	2,339	157,643	454,704
HP-311	QHP311	Catalyst Regenerator Flue Gas / Purge Gas Expander Inlet Exchanger	7.9	1,746.2	701.9	302.4	1,200.0	469.1	20.0	847	107,973	319,017
HP-312	QHP312	Catalyst Regenerator Flue Gas Steam Superheater	38.3	1,746.2	701.9	330.0	1,600.0	241.7	25.0	6,341	1,098,601	1,844,502
HP-313	QHP313	Catalyst Regenerator Flue Gas / Gasifier Steam Preheat Exchanger	4.0	1,746.2	701.9	382.7	500.0	680.6	6.0	991	79,301	264,658
HP-314	QHP314	Catalyst Regenerator Flue Gas Steam Superheater	40.5	1,746.2	701.9	459.8	1,000.0	447.8	25.0	3,616	272,971	676,619
HP-315	QHP315	Catalyst Regenerator Flue Gas / Syngas Recycle Exchanger	5.1	1,746.2	701.9	507.0	1,600.0	169.4	6.0	5,067	784,913	1,242,601
HP-316	QHP316	Tar Reformer Effluent / BFW Preheat Exchanger	6.3	186.6	179.0	148.1	172.8	21.2	105.0	2,824	48,497	180,027
HP-317	QHP317	Steam Turbine Generator Exhaust / Catalyst Regenerator Air Preheat Exchanger	1.2	194.2	180.0	130.9	172.8	33.3	4.0	8,676	105,948	264,444
HP-318	QHP318	Tar Reformer Effluent / BFW Preheat Exchanger	0.3	194.2	191.9	148.1	172.8	31.3	105.0	90	13,110	70,561
HP-319	QHP319	Tar Reformer Effluent / Catalyst Regenerator Air Preheat Exchanger	0.7	191.9	186.6	130.9	172.8	34.2	5.0	4,105	63,313	152,647
HP-401	QHP401	Steam Blowdown / Purge Gas Expander Inlet Exchanger	1.1	462.0	194.2	172.8	302.4	68.8	40.0	417	25,155	105,202
HP-402	QHP402	Steam Turbine Generator Exhaust / BFW Preheat Exchanger	22.9	472.4	194.2	184.6	237.1	70.6	100.0	3,242	53,187	184,930
HP-403	QHP403	Methanol Synthesis Reactor Effluent / Catalyst Regenerator Air Preheat Exchanger	2.0	234.8	225.2	172.8	226.5	23.9	30.0	2,784	58,623	218,505
HP-404	QHP404	Methanol Synthesis Reactor Effluent / BFW Preheat Exchanger	2.8	248.2	234.8	225.6	232.3	12.3	125.0	1,810	61,075	196,015
HP-405	QHP405	Methanol Synthesis Reactor Effluent / BFW Preheat Exchanger	48.5	482.0	248.2	233.4	345.5	54.8	125.0	7,081	157,004	377,534
HP-501	QHP501	Char Combustor Flue Gas / Methanol Degassing Column Reboiler Exchanger	9.1	406.7	343.6	249.6	249.8	122.8	50.0	1,490	29,845	136,539
HP-502	QHP502	DME Reactor Effluent / Methanol Degassing Column Reboiler Exchanger	7.7	482.0	295.9	249.6	249.8	115.2	55.0	1,221	35,387	143,361
HP-503	QHP503	Methanol Intermediate / BFW Preheat Exchanger	4.9	252.6	310.9	172.8	184.6	98.9	160.0	311	16,521	86,230
HP-504	QHP504	Methanol Intermediate / Light Hydrocarbon Recycle Exchanger	0.9	194.2	182.5	152.9	172.8	25.3	70.0	527	22,277	108,613
HP-505	QHP505	Methanol Intermediate / Catalyst Regenerator Air Preheat Exchanger	0.7	182.5	150.0	130.9	172.8	13.9	65.0	808	25,474	109,572
HP-506	QHP506	Methanol Intermediate / Purge Gas Expander Inlet Exchanger	0.5	182.5	150.0	108.0	172.8	22.0	80.0	287	23,236	96,249
HP-507	QHP507	Methanol Intermediate / AGR Acid Gas Exchanger	1.3	182.5	150.0	109.9	140.0	41.3	65.0	481	19,506	102,857
HP-1401	QHP1401	Unconverted DME Recycle / DME Reactor Feed Exchanger	10.1	250.0	194.2	172.8	244.1	12.1	15.0	55,724	596,785	1,115,868
HP-1402	QHP1402	Hydrocarbon Synthesis Reactor Effluent / Catalyst Regenerator Air Preheat Exchanger	3.9	375.0	250.0	237.3	300.0	35.1	9.0	12,317	156,577	362,931
HP-1403	QHP1403	Hydrocarbon Synthesis Reactor Effluent / DME Reactor Feed Exchanger	20.4	375.0	250.0	244.1	353.3	12.1	15.0	112,138	1,283,712	2,364,864
HP-1404	QHP1404	Steam Blowdown / Unconverted DME Purge	0.2	194.2	155.4	95.6	172.8	37.4	25.0	170	16,841	95,289
HP-1405	QHP1405	Unconverted DME Recycle / Deethanizer Feed Preheat Exchanger	0.5	194.2	191.1	153.3	172.8	28.8	65.0	279	16,521	86,230
HP-1406	QHP1406	Unconverted DME Recycle / DME Reactor Feed Exchanger	1.8	191.1	180.5	150.4	172.8	23.7	15.0	4,941	80,154	219,784
HP-1407	QHP1407	Unconverted DME Recycle / Light Hydrocarbon Recycle Exchanger	0.1	180.5	179.8	128.5	152.9	38.3	15.0	205	15,775	86,230
HP-1501	QHP1501	High-Octane Gasoline Blendstock Product / DME Reactor Feed Exchanger	3.2	335.8	194.2	172.8	244.1	48.3	70.0	960	29,631	116,820
HP-1502	QHP1502	High-Octane Gasoline Blendstock Product / Deethanizer Feed Preheat Exchanger	1.1	194.2	140.0	112.7	153.3	33.6	120.0	271	16,308	86,016
HP-1503	QHP1503	Tar Reformer Effluent / Debutanizer Reboiler Exchanger	4.1	546.5	404.9	308.7	335.8	146.0	50.0	561	19,719	103,070
HP-1504	QHP1504	Char Combustor Flue Gas / Deethanizer Reboiler Exchanger	13.5	343.6	250.7	240.3	306.1	21.1	50.0	12,734	141,229	333,407
Total			785.4							554,656	10,186,032	21,613,249

147

Appendix K. Reviewer Comments and NREL Responses on Draft Design Report

This appendix presents a summary of reviewer comments from the draft version of the report, which was issued for peer review in August 2014. Each comment from the peer review panel is summarized below, followed by responses from the authors of the design report. The responses describe any actions taken to address the comment in the final draft of the report. The reviewer comments and corresponding responses are organized by process area.

Area 100: Feed Handling and Preparation

1. **Comment:** The yield will be dependent upon the amount of inorganic materials present in the feedstock. Was this yield based upon dry wood?

   Response: Feedstock quality/composition is a key factor that directly impacts the fuel yield. The product yield value of 64.9 gallons per ton is based on dry wood. Additional details on the feedstock composition are shown in Table 2 and a summary of the feedstock blend formulation and cost analysis is presented in Table 3.

2. **Comment:** Suggest specifying % blended feedstock formulation in Section 2.1 (p. 7).

   Response: The blended feedstock formulation and target delivered feedstock price for the design case are presented in Table 3 in Section 3.1.

3. **Comment:** Question on how much of construction and demolition waste is available annually (p. 12).

   Response: Details on the feedstock supply system and blended feedstock approach are included in the recent design report published by Idaho National Laboratory [1]. The report states that, in a 2009 report (EPA530-R-09-002), the Environmental Protection Agency (EPA) estimated approximately 170 million tons of construction and demolition waste was generated in 2003 in the United States. The waste was sent to an EPA-estimated 1,900 C&D landfills.

4. **Comment:** Based on data from EcoInvent? GREET also has data for forest residue and BETO is supporting analysis that will see pine included in GREET at end of FY15. I hope you will be able to use these data because they will be quite specific to the US.

   Response: The feedstock production data for residue and pine are not based on EcoInvent data. They are based upon Johnson et al. 2012 [2] for logging residue and Searcy and Hess 2010 [3] for southern pine (specified in Appendix F).

5. **Comment:** Can you provide an example or two of which equipment is lumped with the delivered feedstock cost? This passage is a little confusing at present (p. 14).

   Response: All capital and operating costs for feedstock grower payment, harvesting and collection, transportation, storage, handling, and pre-processing up to the gasification reactor are included in the $80 per ton of delivered feedstock with the exception of the capital costs for the cross-flow dryer and the plant-generated heat required for preheating the biomass.

Refer to the table below for a complete list of capital items included in delivered feedstock cost and the capital required for the biorefinery facility.

Area A100: Feed Handling & Preparation			
Equipment #	Number Required	Equipment Name	Source of Capital Cost Coverage
C-101	4	Hopper Feeder	Delivered Feedstock Cost
C-102	2	Screener Feeder Conveyor	Delivered Feedstock Cost
C-103	2	Radial Stacker Conveyor	Delivered Feedstock Cost
C-104	2	Dryer Feed Screw Conveyor	Delivered Feedstock Cost
C-105	2	Gasifier Feed Screw Conveyor	Delivered Feedstock Cost
K-101	2	Flue Gas Blower	Delivered Feedstock Cost
K-102	2	CO2 Booster Blower	Delivered Feedstock Cost
M-101	4	Hydraulic Truck Dump with Scale	Delivered Feedstock Cost
M-102	2	Hammermill	Delivered Feedstock Cost
M-103	3	Front End Loaders	Delivered Feedstock Cost
S-101	2	Magnetic Head Pulley	Delivered Feedstock Cost
S-102	2	Vibratory Conveyor	Delivered Feedstock Cost
T-101	4	Dump Hopper	Delivered Feedstock Cost
T-102	1	Hammermill Surge Bin	Delivered Feedstock Cost
T-103	2	Dryer Feed Bin	Delivered Feedstock Cost
T-104	2	Dried Biomass Hopper	Delivered Feedstock Cost
T-105	2	Lock Hopper	Delivered Feedstock Cost
T-106	2	Feed Hopper	Delivered Feedstock Cost
M-105	2	Cross-Flow Pellet Dryer	Biorefinery Capital

6. **Comment:** Can also look at treatment of forest residue in GREET model. See https://greet.es.anl.gov/publication-feedstocks-13

   Response: Information on logging residues are from Johnson et al. 2012 [2]. It is the understanding of the authors that the forest residue data in GREET assumes a mix of logging residue and thinnings, the results of which are consistent with results from logging residues.

7. **Comment:** Is this (10% moisture content in dry, pre-heated feedstock) a result of using dry demo wood etc (and only 45 % of wet pulp wood) in the mix (p. 7 Section 2.1)?

   Response: The moisture content of the biomass at the inlet to the gasifier is 10% due to the drying steps required during the pre-processing and blending of various woody biomass resources (refer to Table 3 in Section 3.1) to produce a uniform feedstock for the conversion process. Idaho National Laboratory presents the details on the feedstock supply system design and economics in their referenced report [1].

8. **Comment:** It was mentioned earlier that no drying is required, so this actually means feedstock preheating (p. 9 Section 2.2)? If so, remove "feedstock drying" from this paragraph.

   Response: Yes, the reviewer is correct. The moisture content of the feedstock upstream of the gasifier is 10%, requiring no additional drying prior to entering the conversion process. Therefore, the text will be changed to "feedstock preheating."

9. **Comment:** Feed handling & drying capital of $200K is very optimistically low (Page iv).

149

Response: All capital and operating costs for feedstock grower payment, harvesting and collection, transportation, storage, handling, and pre-processing up to the gasification reactor are included in the $80 per ton of delivered feedstock with the exception of the capital costs for the cross-flow dryer and the plant-generated heat required for preheating the biomass. Refer to the table in Response #5 for a complete list of capital items included in delivered feedstock cost and the capital required for the biorefinery facility.

10. **Comment:** What is the feedstock form to gasification (Page 8)?

    Response: The blend of four biomass materials (pulpwood, wood residues, switchgrass, and construction and demolition waste) are dried and pre-processed into pellets (1/4-inch minus or 4-6 mm) prior to delivery to the biorefinery.

11. **Comment:** Area 100 design basis – how much drying occurs in the cross-flow dryer (Page 13)? You will need emissions controls on the effluent gas from the dryer.

    Response: Since the receiving feedstock moisture is already at 10 wt% and satisfies the feedstock moisture content specification for the gasifier, no additional drying after in-plant handling is required. However, the flue gas utilized for preheating the biomass for gasification is routed to the flue gas scrubber (M-703) prior to release to the atmosphere to ensure removal/recovery of biomass particulates.

12. **Comment:** The process description includes "Densification" but there is no description of what that means. Although it may be detailed in the INL report, it would be useful to say a few words about it here. (Page 13)

    Response: As depicted in Figure 4, "densification" is one of the steps in the feedstock preprocessing designed by researchers at Idaho National Laboratory. Though feedstock logistics and preprocessing is not the focus of this design report and research effort, the authors will consider adding content on the preprocessing steps including densification in the final version of the design report. Details on the feedstock preprocessing design can be found in the recent design report published by INL [1].

13. **Comment:** In Table 3, does "Preprocessing" include "Densification"? (Page 14)

    Response: Yes, per Response #12, densification of the biomass feedstocks are included in the preprocessing steps prior to feedstock reaching the conversion facility plant gate.

14. **Comment:** It would be useful to add a few words about the biomass feeding method. There is some very informative detail about the two-reactor system with regard to heat management, but nothing about biomass feeding. I would recommend at least a brief description of the method. (Page 15 / 19)

    Response: The biomass feeding system design is expected to be consistent to that described in the "Biomass Gasification Technology Assessment" report by Worley and Yale (published by NREL) [4] (http://www.nrel.gov/docs/fy13osti/57085.pdf). Specifically, Section 9 and Appendices E-4, G-4, and H-4 provide information on the feeding system for low-pressure (<150 psi) thermochemical operations. Further details of the feeding system, as envisioned by INL as part of their feedstock supply and logistics system, are provided in the publication

by Jacobson et al [1]. With INL, the authors will develop some discussion on the feeding system for inclusion in the final draft of the report.

Area 200: Gasification

15. **Comment:** We believe that a more realistic capital cost for the gasification section is approximately $120 MM.

    Response: The total capital investment, after applying installation costs and other factored cost assumptions, is approximately $100MM. This is for indirect gasification without an oxygen plant in 2011$. These capital estimates for major process areas are consistent with other gasification-based biofuels pathway assessments from NREL [5, 6, 7] and a detailed gasification cost assessment completed by Harris Group Inc. [4]. Also, note that feed handling and drying costs are not included in the biorefinery plant design as they are captured in the total delivered cost of the feedstock to the biorefinery gate. Sensitivity analysis on gasifier and reformer capital costs is included in Section 5.

16. **Comment:** The cost associated with the gasification and feed handling areas of the plant seem very low compared to others, the gas clean up section for example (p. iv).

    Response: The capital cost for the feed handling and drying area includes only the cost associated with the cross-flow dryer for preheating the biomass feedstock. All other feedstock handling costs are included in the delivered cost of the supplied feedstock. There is no oxygen plant or rotary dryer included in this process.

    The total capital investment, after applying installation costs and other factored cost assumptions, is approximately $100MM for indirect gasification without an oxygen plant in 2011$. The capital estimates for the major process areas of the plant (gasification, gas clean-up, major rotating equipment, high-pressure vessels, etc.) are developed in collaboration with an engineering firm, Harris Group Inc., for this design report. In addition, the capital estimates for these major process areas are consistent with other gasification-based biofuels pathway assessments from NREL [5, 6, 7] and a detailed gasification cost assessment completed by Harris Group Inc. [4].

17. **Comment:** We asked a catalyst industry contact to review the GREET catalyst module documentation, which includes olivine because olivine is also used in the 2011 thermochemical (mixed alcohols) design report and we are adding that pathway to GREET.

    The reviewer had this to say about olivine: olivine is not a good choice for a heat transfer medium. There are plenty of good, cheap, commercial materials available here in the USA, such as mullite and alpha alumina.

    Also, approximately 2/3 of olivine comes from Norway.

    Response: The authors appreciate the concerns regarding olivine for the gasification heat transfer media. However, the authors are not in a position to easily change the gasification basis for this design report. The basis for syngas production in this design report will continue to leverage the demonstrated gasification/syngas cleanup basis from the 2011 mixed alcohols design report [7] and subsequent demonstration in 2012 [5, 6]. The focus of the research and development associated with the high-octane gasoline blendstock pathway is

151

hydrocarbon synthesis rather than syngas production. However, the authors will consider other materials for biomass gasifier heat transfer media for future state of technology assessments if experimental information for the successful utilization of such materials is available in the public forum.

18. **Comment:** The fluidization medium in the gasifier is steam supplied from the steam cycle. The steam-to-feed ratio is set to 0.4 lb of steam per lb of bone dry biomass feedstock. What's the rationale for this ratio?

    Response: The steam-to-feed ratio in the gasifier is within the range of experimental test results showing stable gasifier operation from the Battelle Columbus Laboratories indirect gasifier [8].

19. **Comment:** What is net consumption of olivine?

    Response: For the base design case, the net consumption of olivine is 543 lb/hr.

20. **Comment:** General questions on operating conditions for the gasifier (pressure, temperature, flow regime) in the process summary section (p. 9 Section 2.2).

    Response: The authors will include the gasifier operating conditions in Section 2.2 per the reviewer's comment/request.

21. **Comment:** Comment that the char combustor is not really downstream of the gasifier because it is part of the same reactor system (p. 15 Section 3.2).

    Response: Good point. The authors have corrected the text.

22. **Comment:** Gasifier exit gas composition (and Table 4) – reference the experimental basis for the operating conditions and the outlet gas composition (Pages 16, 17).

    Response: The experimental basis for the gasifier is consistent with the 2011 mixed alcohols design report [7] and 2012 demonstration [5]. The basis is also presented in Appendix E of the high-octane gasoline blendstock report (Appendix E. Synthesis Gas and Char Correlations for Indirect Gasifier).

Area 300: Synthesis Gas Cleanup and Syngas Compression

23. **Comment:** The contaminant levels in the syngas (e.g. NH_3 and H_2S) are very high. As a result, the catalysts will be de-activated in a short period of time.

    Response: H_2S concentration is reduced down to trace levels via acid gas removal (AGR) and sulfur guard bed. NH_3 concentration is reduced to trace levels by the wet scrubber following the steam reforming step.

24. **Comment:** What is the expected lifetime of the reforming catalyst?

    Response: The tar reformer is an entrained-flow fluidized reactor and the catalyst is continuously regenerated on-board (refer to Figure 6). The spent catalyst is sent to the adjacent catalyst regenerator and the regenerated catalyst is sent back to the reformer reactor. Like other circulating fluidized bed systems (fluid catalytic cracking operations), the catalyst activity is maintained in the tar reformer system by routine purge of spent/deactivated

152

catalyst and makeup of fresh material. The current makeup rate, based on information from industrial partners, is set at 0.15% of the total catalyst inventory [5].

25. **Comment:** Question on if there is a separate water-gas shift (WGS) stage (p. 9 Section 2.2)?

    Response: There is not a dedicated water-gas-shift reactor included in the process design. The reforming catalyst has sufficient WGS activity at the tar reformer outlet conditions. Although a separate WGS would have the added benefit of favorable equilibrium toward hydrogen production, we have sufficient hydrogen from the reformer in this case without having additional WGS equipment. The off-gases from the PSA, as the reviewer is aware, are reused within the process.

26. **Comment:** What is the assumed conversion for the mentioned components (tars, methane, and light hydrocarbons) in the tar reformer operation (p. 9 Section 2.2)? I see they are given in Table 8, but for a first time reader this would be a good place to mention them.

    Response: The process descriptions outlined in Section 2.2 are intended to present an introduction of the overall process design to the reader. However, many of the details on the process basis, operating conditions, etc., are discussed later in the report under the "Process Design and Cost Estimation" sections. The tar conversion performance targets are summarized in Table 8 (shown below for simple reference) [5, 6].

Compound	Conversion
Methane (CH_4)	80 %
Ethane (C_2H_6)	99 %
Ethylene (C_2H_4)	90 %
Tars (C_{10+})	99 %
Benzene (C_6H_6)	99 %
Ammonia (NH_3)	90 %

27. **Comment:** Does this (molar steam to carbon) ratio also include the carbon in hydrocarbons and tars (p. 21 Section 3.3)? Please provide an equation to shown how defined.

    Response: Yes, the steam to carbon ratio here represents the molar ratio of H_2O relative to the total carbon in hydrocarbons and tars.

    Steam to Carbon Ratio = [H_2O] / [Total C in hydrocarbons and tars]

28. **Comment:** Was the steam to carbon ratio chosen to achieve the desired H_2:CO ratio, rather than to facilitate smooth reformer operation? Discuss the minimum requirement to maintain successful reformer operation.

    Response: The steam to carbon ratio was chosen to simultaneously achieve (1) the H_2:CO ratio target of 2:1 and (2) to minimize coking propensity.

29. **Comment:** Is steam generated from the recovered heat from the reformer syngas (p. 9 Section 2.2 and p. 21, Section 3.3)?

    Response: The hot reformed syngas is cooled through heat exchange with other process streams, which may include exchangers in boiler feed water preheat, steam generation, and

153

steam superheating. The authors are working on the design of the final heat exchanger network, which will be included in the final published design report.

30. **Comment:** Tars are modeled as naphthalene?

    Response: Yes, the tars are modeled as naphthalene ($C_{10}H_8$).

31. **Comment:** What is the experimental basis for the tar reformer (Page 19)?

    Response: The experimental basis for the tar reformer is the demonstrated performance results presented by NREL in 2012 [5, 6].

32. **Comment:** Has the low SCR been experimentally demonstrated at believable scale (Page 21, paragraph 1)?

    Response: Based on reviewer comments, the authors have increased the SCR 4.4 to simultaneously achieve (1) the H_2:CO ratio target of 2:1 and (2) to minimize coking propensity. Steam to Carbon Ratio = [H_2O] / [Total C in hydrocarbons and tars].

33. **Comment:** Reference 22, 23 reforming catalyst data not available (Page 22).

    Response: Reforming catalyst information can be found in the demonstration report from Dutta et al. [5], which is located here: http://pubs.acs.org/doi/ipdf/10.1021/ie402045q.

34. **Comment:** At the bottom of the page, you say "… the syngas is assumed to exit the reformer at water-gas shift equilibrium." Is there any test data to support that or is it just an assumption based on calculations? (Page 20)

    Response: There is not a dedicated water-gas shift reactor included in the process design. The reforming catalyst has sufficient WGS activity at the tar reformer outlet conditions. Although a separate WGS would have the added benefit of favorable equilibrium towards hydrogen production, we have sufficient hydrogen from the reformer in this case without having additional WGS equipment. The off-gases from the PSA are reused within the process. The authors have conducted sensitivity cases to assess the impact of approach to WGS-equilibrium at the outlet of the tar reformer. The analysis shows that, within reasonable equilibrium approach temperatures (<100 °F), this assumption has negligible impact to the plant design and economics at the tar reformer operating conditions.

35. **Comment:** In the 3rd paragraph, you say, "Continuous operation has been demonstrated with simulated syngas … using NREL's reforming catalyst." I am curious whether that testing was done with a fluidized bed as described for the commercial process, or in a fixed bed? (Page 21)

    Response: In 2012, NREL demonstrated an integrated process for the production of mixed alcohols from biomass via gasification at the pilot facility in Golden, Colorado. The demonstrated process utilized a fixed-bed tar reforming reactor and the fluidized bed tar reforming process performance was demonstrated separately at the pilot scale by Rentech. These performance demonstrations are documented in publications [5, 9] and presentations [6]. Depending on research progress, a future demonstration of integrated performance for this pathway may be conducted within the 2022 timeframe.

154

Area 400: Acid Gas Removal and Methanol Synthesis

36. **Comment:** Contamination of the syngas with the amine may poison the catalysts over time.

    Response: Many commercial fixed-bed processes utilize amine scrubbing systems to remove H_2S from vapor streams prior to reintroducing the scrubbed gas to the reactor. For example, hydroprocessing operations utilize amine scrubbers in the recycle gas circuit. The AGR operates at 110°F, which is significantly lower than the boiling point of amine (477°F for MDEA) and higher pressure (above 735 psia). Carry-over of amines by vapor entrainment is typically mitigated during detailed equipment design by velocity control and demisters installations to protect downstream operations.

37. **Comment:** H_2 / CO should be about 2.1 / 2.0 molar ratio.

    Response: Raw syngas from gasifier contains substantial amounts of hydrocarbons, including tars and lighter hydrocarbons (C_1–C_3). The tar reformer serves to convert the hydrocarbon species to syngas. Downstream of the syngas cleanup operations, the H_2:CO ratio of clean syngas is approximately 2.08.

38. **Comment:** We wonder if you are confusing the CO_2 inlet concentration to the methanol reactor with that of the make-up stream. Bartholomew may have written that 8 % inlet to the methanol reactor is often seen, but 8% in the make-up gas is much too high (p.25).

    Response: Based on the reviewer comment, the authors have adjusted the CO_2 inlet concentration from 8 mole % down to 5 mole %.

39. **Comment:** The cost of the AGR equals that of the methanol reactor. This is not realistic and confirms that the design of the methanol reactor system is far from optimal (p.27).

    Response: The AGR cost does not include the cost associated with the syngas compression, which is required to achieve sufficient operating pressures for efficient AGR performance. Lowering the CO_2 inlet concentration from 8 to 5 mole % (per the previous comment and associated response) will also result in lower methanol reactor cost and higher AGR cost.

40. **Comment:** Given what you know about the technology, is it likely 100% removal possible? If 100% removal not achieved, do you have a sense of what happens to the catalyst lifetime?

    Response: In the form of H_2S, sulfur can be removed with zinc oxide (ZnO) catalyst down to undetectable levels [10]. If H_2S breakthrough occurs from the zinc oxide bed, the catalyst deactivation rate will increase due to sulfur poisoning by irreversible blocking of active catalyst sites.

41. **Comment:** Is the LO-CAT process the same as the AGR in this case? Might want to make that clear at the start of the paragraph.

    Response: The acid gas removal unit and LO-CAT process are distinct operations in the plant configuration. The AGR is an amine process that utilizes methyldiethanolamine for removing H_2S and CO_2 from the syngas. The H_2S and CO_2 from the AGR, often called acid gas, feeds the LO-CAT process. The LO-CAT process then converts the sulfur in H_2S to innocuous elemental sulfur.

155

42. **Comment:** Question on if TIGAS is one of the industry users of the combined methanol/DME reaction that should be consulted if possible.

    Response: Yes, based on published information on the Haldor Topsoe TIGAS technology [11, 12], the process configuration utilizes a syngas to methanol/DME conversion step, which does not require methanol purification and recovery as an intermediate. This configuration clearly possesses advantages over the multi-step conversion of syngas to methanol to DME. However, until a technical basis for this approach is shared in the public domain, the authors plan to maintain the current multi-step process as the base design.

43. **Comment:** The excerpt "The mixture of methanol and unconverted syngas is cooled through heat exchange with the steam cycle and other process streams" (p. 10) gives an impression of a separate cooling stage after the reactor.

    Response: Based on the reviewer comment, the authors will provide additional content for clarification. As described in Section 3.4, the vapor-phase product from the methanol synthesis reactor must be cooled to recover the methanol and to allow unconverted syngas and any inert gaseous species (CO_2, CH_4) to be recovered for recycle and purge. This is accomplished with a series of heat exchangers, including air cooler and water cooling.

44. **Comment:** The excerpt "The mixture of methanol and unconverted syngas is cooled through heat exchange with the steam cycle and other process streams" (p. 10) gives an impression of a boiling-water reactor.

    Response: Heat must be removed from the methanol synthesis reactors because the synthesis reaction is exothermic. Heat is removed by steam production on the shell-side of the tubular reactor. The control of steam temperature and pressure is maintained by back-pressure control at the outlet of the steam drum.

45. **Comment:** There doesn't seem to be more detailed discussion of the PSA in the report: Does the configuration include a guard bed, or is a "sour-PSA" assumed instead where H_2S is separated to purge as other non-H_2 components? The purge gas is at near ambient pressure and needs to be compressed before feeding to tar reformer? Why the acid gases are not separated BEFORE splitting syngas to PSA? The hydrogen mole fraction is only 55 % at PSA inlet, which must be at the lower end of feasible separation. Also the costs of PSA are likely the scale with the purge gas volume making CO_2 separation before PSA advisable. It would also significantly increase purge gas heating value making it more valuable feed for combustion. Please discuss these issues somewhere in the report.

    Response: The authors have modified the design based on the reviewer comment. The PSA in this design will not be a sour service. Acid gas compounds (CO_2 and H_2S) are now removed upstream of the PSA via an acid gas removal unit and sulfur guard bed.

46. **Comment:** This goes against my previous understanding that "amine processes are particularly applicable where acid gas partial pressures are low", while physical solvent processes should be considered when the "partial pressure of the acid gas in the feed is greater than 50 psi." (quotations from GPSA Engineering Data Book, section 21). Please explain.

156

Response: The authors agree that capacity of scrubbing amines (MDEA) for CO_2 recovery increases with increasing CO_2 partial pressure. Refer to the plot below showing relative pick-up capacities for 50 wt% aqueous MDEA solution and DEPG (primary component in Selexol) physical solvent. Based on published values for MDEA CO_2 loading and CO_2 solubility in DEPG [10], the point at which the physical solvent becomes more efficient than the amine is at much higher CO_2 partial pressures (>500 psi).
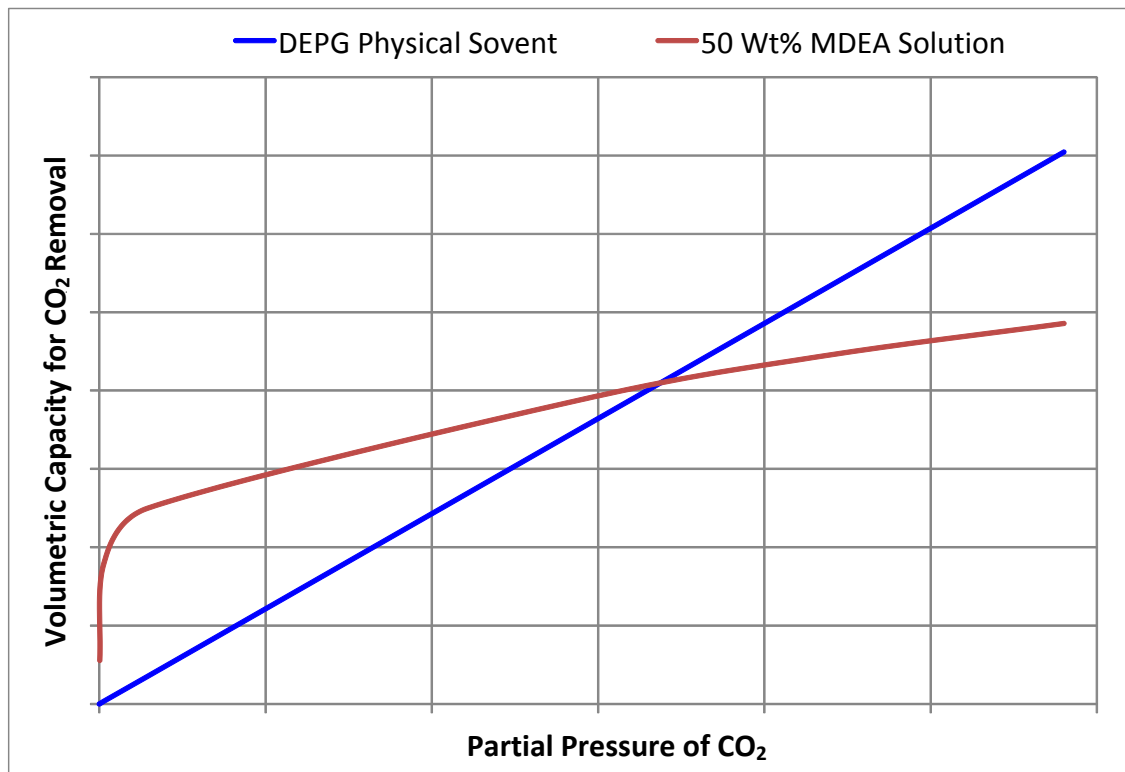
47. **Comment:** Why is syngas compressed directly to methanol converter pressure (p. 24, Section 3.4)? The acid gas removal should work well at lower pressures than this (e.g. 20 – 30 bar) and then the final compression could be carried out for gas that has less CO_2 for compression savings.

    Response: Per reviewer comments, the design has been modified to split syngas compression steps. Initial syngas compression will increase syngas pressure up to AGR operating pressure of approximately 400 psi. The final syngas compression up to methanol synthesis operating pressure is carried out after the acid gas removal steps.

48. **Comment:** Referring to the statement "concentration of 10 ppmv, which is the permissible exposure limit" (p. 24, Section 3.4). This sounds interesting, but the source seems to be a bit shady? Please add a better source to support this assumption as it has very important implications for real life plant designs.

    Response: The authors recognize that permissible emissions concentrations in biorefinery point sources are not the same as permissible worker exposure limits. However, the LO-CAT process is capable of removing H_2S to concentrations of less than 10 ppmv in the CO_2 vent [10, 13]. This concentration is consistent with other best available control technologies

157

(BACT) for sulfur removal. It is also important to note that construction of an actual biorefinery will require environmental permits specific for the detailed plant design, emissions profile, and location of the facility.

The authors posed the following question to Merichem, the licensors of the LO-CAT technology, "in your experience licensing LO-CAT units, do the operating permits typically require an incinerator or thermal oxidizer on the CO_2-rich off-gas to convert trace H_2S to SO_2 prior to venting to the atmosphere?" If Merichem responds that H_2S conversion to SO_2 is a typical add-on, then we will address this in the plant design by including the capital and operating costs for a thermal oxidizer for the vent gas. The authors believe this will be the least costly option as routing the stream to the char combustor or tar reformer catalyst regenerator would require additional compression equipment.

Gary Nagl of Merichem Company provided the following response on September 24, 2014: *In response to your question concerning incineration of the LO-CAT flue gas, in most cases the answer is incineration is not required. For treating biogas we employ the flow scheme illustrated below. The biogas is contacted with the LO-CAT solution in liquid full absorber where the H_2S in the gas stream is generally reduced to <5 ppm. The spent solution, which contains no H_2S, is sent to the oxidizer where it is contact with air. The flue gas from the oxidizer will contain no H_2S. However, depending on the composition of the biogas there may be components, which are water soluble such as BTX, which will be stripped out in the solution and leave with the oxidizer flue gas. For these components each case must be evaluated to determine if incineration is required. If you go to our website (www.merichem.com), on the first page on the bottom under "Resources" you will find many articles describing the LO-CAT process.*

Based on the response from Merichem Company, the authors are not planning to include an incinerator or thermal oxidizer in the plant design for this service at this point. However, the authors agree to note that the actual plant design requirements for the LO-CAT vent gas will depend on the location of biorefinery facilities and applicable permitting requirements.

158
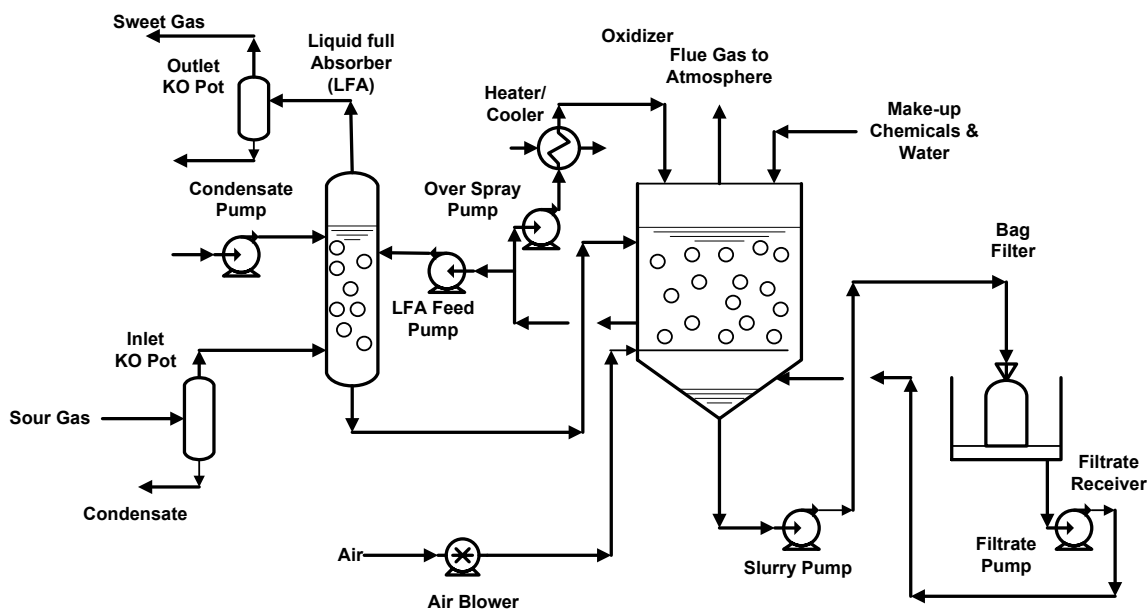
49. **Comment:** Include methanol reaction equations (p. 25, Section 3.4).

    Response: The equilibrium reactions for methanol modeled in Aspen Plus are listed below. Per the reviewer comment, the authors will include the reactions in the final report.

    $$CO + 2\,H_2 \leftrightarrow CH_3OH$$

    $$3\,H_2 + CO_2 \leftrightarrow CH_3OH + H_2O$$

    $$CO + H_2O \leftrightarrow H_2 + CO_2$$
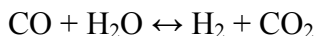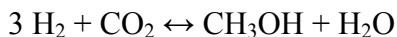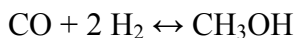
50. **Comment:** Where does the heat come from for heating the methanol recycle purge stream prior to expansion (through the turbo-expander).

    Response: The stream is heated with combustion flue gas from the char combustor or tar reformer combustor. The authors are working on the design of the final heat exchanger network, which will be included in the final published design report.

51. **Comment:** I was under the impression that although methyldiethanolamine (MDEA) have significant selectivity for H_2S removal, something else needs to be used for CO_2 removal.

    Response: The selectivity of MDEA for preferential H_2S pick-up versus CO_2 makes it an attractive amine for applications where acid gas streams highly-concentrated in H_2S are desired. For example, a facility that is hydraulically limited in sulfur plant capacity will limit the amount of CO_2 sent to the sulfur plant in the acid gas streams because it takes up hydraulic capacity and may limit the processing rate of the plant. However, this does not suggest that MDEA is not adequate for bulk CO_2 removal for dilute H_2S streams. MDEA is also an attractive option for bulk CO_2 removal based on an analysis from Bryan Research & Engineering, Inc. and Propak Systems, Inc. [14].

52. **Comment:** Comment on the $[(H_2 - CO_2) / (CO + CO_2)]$ ratio: Ideally, this should equal 2.03 for methanol synthesis. Why using 1.41 here?

    Response: Based on reviewer comments, the CO_2 concentration in the methanol reactor feed was lowered from 8 vol % to 5 vol % to achieve the target $[(H_2 - CO_2) / (CO + CO_2)]$ ratio of 2.04. The authors have incorporated this change in the Aspen Plus model and will reflect the design modification in the final report.

53. **Comment:** Acid gas removal section – what are the impurity levels to the MeOH synthesis catalyst and their basis? What are the vendor specs? (Page 10)

    Response: The methanol synthesis reactor feed contains 5% CO_2 and is sulfur free (trace levels). In addition to the AGR/LO-CAT system, a ZnO sulfur guard bed is also employed, which can remove sulfur (in the form of H_2S) down to undetectable levels [15]. The literature notes that CO_2 concentrations of up to 8 vol % can improve productivity to methanol [10, 16].

54. **Comment:** This may be too big a change to consider, but I wondered why you combined acid gas removal with methanol synthesis in Area 400. Consider that one of your process recommendations on p. 60 is to consider combining MeOH and DME synthesis in a single reactor (as Topsoe does in TIGAS). That would be much easier to evaluate if the AGR unit was a separate Area. (Page 24)

159

Response: Based on published information on the Haldor Topsoe TIGAS technology [11, 12], the process configuration utilizes a syngas to methanol/DME conversion step, which does not require methanol purification and recovery as an intermediate. This configuration clearly possesses advantages over the multi-step conversion of syngas to methanol to DME. However, until a technical basis for this approach is shared in the public domain, the authors plan to maintain the current multi-step process as the base design. Therefore, we closely associated the AGR as a necessary part of methanol production and did not consider this as a separate area within our simulation.

Area 500: Methanol Conditioning

55. **Comment:** Methanol conditioning is listed with a negative cost contribution in Figure 15. It should probably be positive, which means that the total cost may be slightly ($0.10) higher than reported (p.57 - Figure 15).

    Response: The authors have updated the figure to appropriately reflect the positive cost contribution of methanol conditioning.

56. **Comment:** Question on the source of heat for crude methanol product heating prior to turbo-expander / turbine generator (p. 10 Section 2.2)?

    Response: Heating the crude methanol stream is no longer required as the authors have modified the design to remove the turbo-expander generator.

57. **Comment:** Question on whether or not water is removed at the methanol conditioning step (p. 10 Section 2.2)?

    Response: The crude methanol contains approximately 2 vol % water at this point in the process. Water is removed from the process downstream of the methanol conversion to dimethyl ether via dehydration.

Area 1400: Methanol to High-Octane Gasoline Synthesis

58. **Comment:** Quantify what is meant by "low" in the context of coke formation. Any level of coke formation can significantly reduce the lifetime of reformers and reactors.

    Response: Coke and aromatic byproducts are represented by the accumulation of the model-compound, hexamethylbenzene. HMB selectivity is set at 0.5%, based on published values and research results to date, and is removed from the catalyst surface by coke combustion with air during off-line catalyst regeneration.

59. **Comment:** Has this catalyst already been produced and validated at the multi-ton production level?

    Response: All catalysts utilized in the process are commercially available with the exception of the metal-modified beta-zeolite catalyst for hydrocarbon synthesis. Beta-zeolite catalysts in general are commercially available materials. However, the major focus of this research effort is modifying beta-zeolites to maximize yields and quality of desired products.

60. **Comment:** The conversion of DME to gasoline is "set to match targets that are believed to be attainable…." This assumption can lead to huge errors on the final result (p.12).

160

Response: The purpose of the design report is to present a conceptual design case based on targets associated with the DOE-BETO funded research and development for the process. Subsequent TEAs for this process (state of technology assessments) will track the progress towards cost-competitive production of hydrocarbon fuel blendstocks and provide feedback on the direction of research. In order to advance the technology, NREL researchers are performing R&D on catalyst development and other aspects of process optimization, aiming to achieve the conversion targets in the near future. NREL has used this approach in the past for demonstrating (at the pilot scale) technology for the cost-competitive production of ethanol [5, 6, 7].

61. **Comment:** The inlet temperature to the DME reactor is governed by the delta-T and a maximum constraint on temperature. However, the resulting inlet temperature of 157 °C (315 °F) gives nearly zero catalyst activity, so this is not realistic (p.29).

    Response: Based on the reviewer comment, the authors will increase the inlet temperature from 157°C (315°F) to 225°C (437°F). This change will ensure the inlet temperature and subsequent delta-T from the exothermic reactions are sufficient to meet the required conversion.

62. **Comment:** Hydrogen is added to suppress formation of durene formation in the reactor loop. Figure 9 does not show any purge, which seems optimistic (p.30).

    Response: As opposed to the ExxonMobil MTG process, the current process does not produce durene. However, hydrogen is added to suppress formation of coke and aromatic byproducts (coke pre-cursors), represented by the model-compound, hexamethylbenzene. Coke/HMB is removed from the catalyst surface by coke combustion with air during off-line catalyst regeneration. Based on the reviewer comment, the authors will improve the clarity of this aspect of the process configuration in Figure 9.

63. **Comment:** 2 bar pressure drop for the gasoline reactor which comprises four beds and three inter-stage coolers seems to be on the low side as a certain minimum bed height is required to ensure even plug flow through the reactors. A detailed analysis would be required to determine whether this is realistic or not (p.32).

    Response: Based on the reviewer comment, the authors will increase the pressure drop per bed to 10 psi (0.7 bar) and the overall pressure drop to 40 psi (2.8 bar). This pressure drop profile is consistent with other fixed, multi-bed reactor systems such as hydrocrackers. The authors also agree with the comment on detailed analysis and recognize that the detailed reactor design should include contingency for increasing pressure drop during the course of longer-term operation.

64. **Comment:** It is assumed that the gasoline catalyst has a three year life-time. This assumption would need to be validated. A shorter life will be more realistic (p.33).

    Response: Based on the reviewer comment, the catalyst life-time assumed for the analysis will be decreased from 3 years to 2 years. The catalyst activity is maintained with *in-situ* catalyst regenerations between the 2-year replacement intervals.

161

65. **Comment:** You state clearly that large uncertainties are associated with the catalyst and reactor configuration assumed for the gasoline synthesis. This fact casts doubt on the feasibility of the entire project (p.60).

   Response: While the techno-economic analysis reported here assumes n^{th}-plant (mature technology) economics (refer to Section 1.3 for details), the technology is largely still at its nascent stage, specifically in the area of high-octane hydrocarbons synthesis from DME. As mentioned in Response #60, the design case presents a target scenario based on the research goals. The design report is then utilized as the "measuring stick" for comparing annual state of technology assessments to track progress and guide research direction.

66. **Comment:** The potential for reduced coking would not enhance carbon utilization substantially. The quantity of carbon formed from coke is quite small compared to the amount of carbon fed from the CO contained in the syngas feed. Rather, the real benefit has to do with catalyst regeneration implications (e.g., a longer run time before regeneration would be needed, etc.). This would certainly be economically beneficial indeed (i.e., more stable catalyst with mitigated regeneration demand).

   Response: The authors definitely agree with the reviewer comment on impacts from coke formation. Suppression of coke and coke-precursor formation rates is a major focus of the research and development effort.

67. **Comment:** Per your reference(s), the ExxonMobil MTG commercial reactors were fixed bed adiabatic with inter-bed cooling between the DME and MTG reactors. Lurgi and Mobil piloted shell and tube reactors, but it didn't go anywhere.

   Response: Thank you for this great information as it further justifies the specification of fixed bed reactors with inter-bed cooling for the hydrocarbon synthesis reactor rather than a shell and tube design.

68. **Comment:** Figure 9 is a misleading picture – had to look at the equipment list and PFD to understand that these are separate reactors with HX between each one. Strongly urge you to redraw this as it colors how one understands the operation and catalyst regeneration.

   Response: The authors will improve the accuracy and consistency of the simplified flow diagram in Figure 9 relative to detailed process flow diagrams in Appendix D.

69. **Comment:** An additional paragraph describing any present experience with HBEA catalyst activity maintenance would be useful, even if just a short summary of literature data. I am concerned about its low hydrothermal stability, especially during coke combustion. Strict temperature control through air dilution will be necessary which significantly lengthens the regeneration cycle and adds to costs. Also of concern are its sensitivity to basic nitrogen poisoning from amine carry-over, to leached metals from the MeOH synthesis catalyst and from the equipment construction materials (iron, alloying metals, silicon). Also, with the DME recycle, contaminants will tend to accumulate to the point where they affect the catalyst or equipment. Again, any literature pointers would help better define this proposed technology.

   Response: The general response to the reviewer's comment is that the research effort has not yet identified activity and stability characteristics for commercial design. The major focus

areas of this research project include (1) developing and demonstrating a catalyst formulation that delivers high selectivity to the desired, high-octane product (2) determining the impacts of catalyst performance resulting from deactivation by accumulation of coke and other potential poisons; (3) identifying the required frequency of catalyst regenerations to maintain sufficient on-stream activity; (4) understanding the impact of *in-situ* coke-burn regenerations on long-term catalyst activity maintenance; and (5) identifying the optimal catalyst replacement cycle. NREL will present progress in these research areas in annual state of technology assessments to track progress and guide research direction. NREL is currently discussing a research plan with Argonne National Laboratory to evaluate activity maintenance through simulated redox cycles.

Regarding your comment on temperature control during coke-burn regenerations, the temperature control system would likely possess similar capabilities to that designed for a cyclic-regeneration naphtha reformer. Temperature control (panel-mount controlloer or distribted control system) would be monitored by bed thermocouples and oxygen analyzers. Air flow rate is controlled by temperature feedback to minimize risk of catalyst damage by excess temperatures/sintering. Details on the catalyst characteristics will inform the analysis on optimal reactor design and regeneration control strategy.

Regarding the concerns for amine carry-over and other poisoning, many commercial fixed-bed processes utilize amine scrubbing systems to remove H_2S from vapor streams prior to reintroducing the scrubbed gas to the reactor. For example, hydroprocessing operations utilize amine scrubbers in the recycle gas circuit. The AGR operates at 110°F, which is significantly lower than the boiling point of amine (477°F for MDEA) and results in low vapor pressure (<0.1 mm Hg for MDEA), minimizing losses by partial pressure of amine in the treated gas. Carry-over of amines by vapor entrainment is typically mitigated during detailed equipment design by velocity control and demister installations to protect downstream operations. In addition, the research team activates the catalyst by flowing air (similar to regeneration) and it completely removes the nitrogen from the catalyst. Therefore, the authors suspect any poisoning would be reversible as with coke. As for catalyst contamination from upstream reactors, equipment, piping, and DME catalyst metals, the research team intends to perform future process validations and demonstrations on a configuration that includes recycle loops in the hydrocarbon synthesis system. Issues with upstream contamination will be identified during the research effort.

70. **Comment:** If no durene is formed in this process, then it should be mentioned as it's a clear advantage over MTG. However, if durene is formed, please discuss how to treat it.

    Response: No durene is formed in this process, which is mentioned in the text. The clear (anticipated) advantage over MTG is also pointed out. However, the authors will include additional content on this benefit in the final report if possible.

71. **Comment:** Neither Iglesia group MTG catalyst publication (refs 2, 32) are suitable for design (Page 2).

    Response: The authors agree that the technical information available in published literature at this time is not sufficient for the design of a commercial facility. However, the purpose of the design report is to present a conceptual design case based on targets associated with the

163

DOE-BETO funded research and development for the process. Subsequent TEAs for this process (state of technology assessments) will track the progress towards cost-competitive production of hydrocarbon fuel blendstocks and provide feedback on the direction of research. In order to advance the technology, NREL researchers are performing R&D on catalyst development and other aspects of process optimization, aiming to achieve the conversion targets in the near future. NREL has used this approach in the past for demonstrating (at the pilot scale) technology for the cost-competitive production of ethanol [5, 6, 7].

72. **Comment:** MeOH to high octane gasoline synthesis – does DME to hydrocarbons produce any byproduct oxygenates? How does selectivity/activity vary with catalyst deactivation? How does selectivity/activity vary with catalyst regeneration (Page 10)?

Response: No oxygenates are produced in the base design assumptions. Selectivity/activity of the catalyst will clearly deteriorate with catalyst deactivation. However, the magnitude of the impact is not yet well understood. The impacts of catalyst performance with (1) deactivation by accumulation of coke and other potential poisons and (2) *in-situ* coke-burn regenerations, is a major focus area of the DOE-BETO funded research effort. NREL will present progress in this research area in annual state of technology assessments to track progress and guide research direction.

73. **Comment:** DME to hydrocarbons reactor – how long are reactors on stream before regeneration? Are reported yields average values over catalyst performance? –multiple catalyst regenerations? –state basis. What byproducts are formed by DME to hydrocarbons catalyst? (Page 32)

Response: Coke and aromatic byproducts can be formed in the reactor. Hydrogen is added to suppress formation of coke and aromatic byproducts, represented in the Aspen Plus model as hexamethylbenzene. HMB and other sources of coke deposition are removed during catalyst regeneration by *in-situ* combustion with air. The research effort has not yet identified the required frequency of catalysts regenerations to date. The impacts of catalyst performance with (1) deactivation by accumulation of coke and other potential poisons and (2) *in-situ* coke-burn regenerations, is a major focus area of the DOE-BETO funded research effort. NREL will present progress in this research area in annual state of technology assessments to track progress and guide research direction.

74. **Comment:** There is no discussion of how the gasoline catalyst behavior changes (distribution of hydrocarbon types) as it ages, which is something we have observed. There may not be a clear answer to this, but it could have implications for the product composition. (Page 31)

Response: Coke and aromatic byproducts can be formed in the reactor. Hydrogen is added to suppress formation of coke and aromatic byproducts, represented in the Aspen Plus model as hexamethylbenzene. HMB and other sources of coke deposition are removed during catalyst regeneration by *in-situ* combustion with air. The research effort has not yet identified the required frequency of catalysts regenerations to date. The impacts of catalyst performance with (1) deactivation by accumulation of coke and other potential poisons and (2) *in-situ* coke-burn regenerations, is a major focus area of the DOE-BETO funded research effort.

NREL will present progress in this research area in annual state of technology assessments to track progress and guide research direction.

Area 1500: Product Recovery

75. **Comment:** Referring to the product separation area of the plant (p. 36 / Section 3.7), is the configuration comparable to conventional MTG, except for a simpler C4 / C5+ split?

    Response: The product separation operations are consistent with the ExxonMobil MTG design [17] and Haldor Topsoe TIGAS design [11] through the deethanizer and debutanizer. The ExxonMobil design includes additional downstream separation of the gasoline product into light and heavy fractions for the purpose of durene conversion. However, durene is not an expected byproduct in the high-octane gasoline blendstock pathway specified here. The Haldor Topsoe process also includes downstream upgrading of the gasoline product.

Areas 600 and 700: Steam System, Power Generation, Cooling Water, and Other Utilities

76. **Comment:** This (utilizing raw syngas for plant heat and power) is an interesting design choice that has potentially significant implications to plant performance. A discussion on why such decision was made instead of acquiring electricity from the power grid would be a welcomed addition here (p. 2 Section 1.1).

    Response: The design constraint that biorefinery facilities will be energy self-sufficient for steady state operation is consistent with assumptions for other DOE-BETO funded analyses from NREL and PNNL. Though economics may improve by purchasing electricity and natural gas for plant operations, the long-term sustainability of the processes will suffer. The consistent assumption allows for easy comparison (a level playing field) among the various technology platforms without adjustments. The impact of imported grid power and natural gas for fuel/co-processing biomass is captured in the Sensitivity Analysis in Section 5. Specifically, the authors are focusing on the following sensitivity scenarios for fossil energy imports to the biorefinery:

    - Using natural gas import for coproduction of syngas and with biomass to maximize product yield.

    - Using power import, develop a case with 0 syngas flow to the fuel combustor.

    In general, the GHG reduction criteria (60% reduction relative to gasoline benchmark) will be a limiting factor for fossil-derived electricity and natural gas import. The electricity import impact on GHGs may be significantly lower for European countries moving quickly towards renewable electricity.

77. **Comment:** What is included in the condensate cleanup process (Page 11)?

    Response: The condensate cleanup processing steps are described in Area 600 and PFD-A601. The process condensate is collected, pumped to an electro-deionization unit (M-601B), processed in a hot condensate polishing unit (M-601C), heated, and sent to the deaerator/packed column (T603 / T603A) prior to steam production.

78. **Comment:** Can the reverse osmosis (RO) purification system (syngas scrubber purge) handle the residual tars that accumulate to scrubbing water (p. 22, Section 3.3)? Please discuss.

   Response: Refer to PFD-A-301 and PFD-A701 for details on the treatment processes downstream of the syngas venture scrubber. The purge water from the scrubber circulation loop is first sent to the sludge settling tank (T-301). Following the settling tank, the purge is processed in the clarifier (M-705), belt press (M-706), and sand filter (M-707) prior to entering the reverse osmosis process. The design configuration for the scrubber purge water is recommended by Harris Group Inc.

General

79. **Comment:** Has there been enough work carried out to validate the individual unit processes at the pilot and / or demonstration scale? Has this process been demonstrated as a fully integrated system?

   Response: In 2012, NREL demonstrated an integrated process for the production of mixed alcohols from biomass via gasification at the pilot facility in Golden, Colorado. The demonstrated process utilized a fixed-bed tar reforming reactor and the fluidized bed tar reforming process performance was demonstrated separately at the pilot scale by Rentech. These performance demonstrations are documented in publications [5, 9] and presentation [6]. If DOE-BETO continues to support research and development efforts for the process presented in this design report, NREL will demonstrate the integrated performance as part of a suite of hydrocarbon pathway demonstrations expected by 2022.

80. **Comment:** There is not enough information presented to determine if the energy and carbon conversion efficiencies are reasonable.

   Response: The key sustainability metrics associated with this process, including carbon conversion efficiency, are summarized in Table 31. The table below summarizes energy and carbon efficiency values for the high-octane gasoline blendstock pathway and other biomass to fuels pathways for comparison.

Process/Pathway	Energy Efficiency (% LHV)	Carbon Efficiency (%)	Appendix K References
High-Octane Gasoline Blendstock via Gasification	45.0	31.2	
Mixed Alcohols via Gasification	45.7	32.1	[5] [7]
Ethanol via Biological Conversion	47.5	27.1–30.2	[18] [19]
Hydrocarbons via Fischer-Tropsch	50.5	34.2	[20] [21] [22] [23] [24] [25]
Hydrocarbons via Methanol-to-Gasoline (MTG)	48.0	33.6	[12] [26] [27]

81. **Comment:** A comprehensive risk assessment analysis for this proposed process needs to be included.

   Response: The techno-economic analysis reported here uses "n[th]-plant" economic and operating assumptions (refer to Section 1.3 for additional details). The key assumption

associated with the "n[th]-plant" is that the facility is constructed and operated in a mature industry where several (n) plants using the same technology are constructed and operating. In other words, the assumption reflects a future in which a successful industry has been established with many operating plants.

The goal of this report is to identify the impacts of research on a conceptual scaled up process. In the past we have leveraged similar reports to derive pioneer plant costs through categorization of risks in various aspects of the plant. We have not included a pioneer plant assessment in this design report, though we do not rule out such studies after further progress in the research.

82. **Comment:** As based upon the composition of the gasoline product, it would not be expected to meet RVP specifications.

    Response: The estimated Redi vapor pressure (RVP) for the high-octane gasoline blendstock is less than 5 psi, relative to typical RVP specifications for finished gasoline fuels in the range of 7–15 psi (depending on location and season). In addition, the product from this process is a gasoline blendstock rather than a standalone finished fuel. The blender is ultimately responsible for meeting final specifications and using high-quality blendstocks to their economic advantage for finished fuels through blending of multiple blendstocks with varying properties.

83. **Comment:** As based upon the composition of the gasoline product, the RON/MON would be expected to be lower than 90.

    Response: Based on initial assessment of the modeled products and published values for pure component octane numbers and blending octane numbers, the calculated weighted average octane value (average of RON and MON) for the gasoline blendstock is between 92 (weighted average of pure component octane numbers) and 110 (weighted average of blending octane numbers). The octane value of the product will be confirmed by analysis consistent with appropriate ASTM methods. Based on the reviewer comment, the authors will include an octane-based comparison of the high-octane product with various gasoline grades and typical blendstocks with varying octane values in the final version of the design report (refer to Response #85 for an example of such additional content).

84. **Comment:** We believe that the capital and O&M cost estimates are low by 50% or more.

    Response: The capital estimates for the major process areas of the plant (gasification, gas clean-up, major rotating equipment, high-pressure vessels, etc.) are developed in collaboration with an engineering firm, Harris Group Inc., for this design report. In addition, the capital estimates for these major process areas are consistent with other gasification-based biofuels pathway assessments from NREL [5, 6, 7] and a detailed gasification cost assessment completed by Harris Group Inc. [4]. The impact of higher capital costs for the project is captured in the Sensitivity Analysis in Section 5. Increasing the total capital investment for the plant by 30% increases the minimum fuel selling price from $3.41 per GGE to $3.90 per GGE.

85. **Comment:** In conclusion, this integrated process is much too complex for economical and efficient operation as a distributed system.

Response: The high-octane blendstock pathway is similar in complexity to the commercial process for converting syngas to hydrocarbon fuel blendstocks licensed by ExxonMobil Research and Engineering [17, 26] and to other biomass to fuel pathways under development such as the TIGAS process by Haldor Topsoe and partners [11].

86. **Comment:** The current U.S. selling price of gasoline in the $2.60 – 2.70 / gallon range.

    Response: The process technology is evaluated based on the DOE-BETO cost target for 2022 of $3.00 per GGE . In addition, the value of a gasoline blendstock is a function of its properties (i.e., octane value). High-octane blendstocks are likely to carry a premium price relative to the finished fuel value. Refer to the octane comparisons presented in the figure below, which will be included in the final version of the design report.
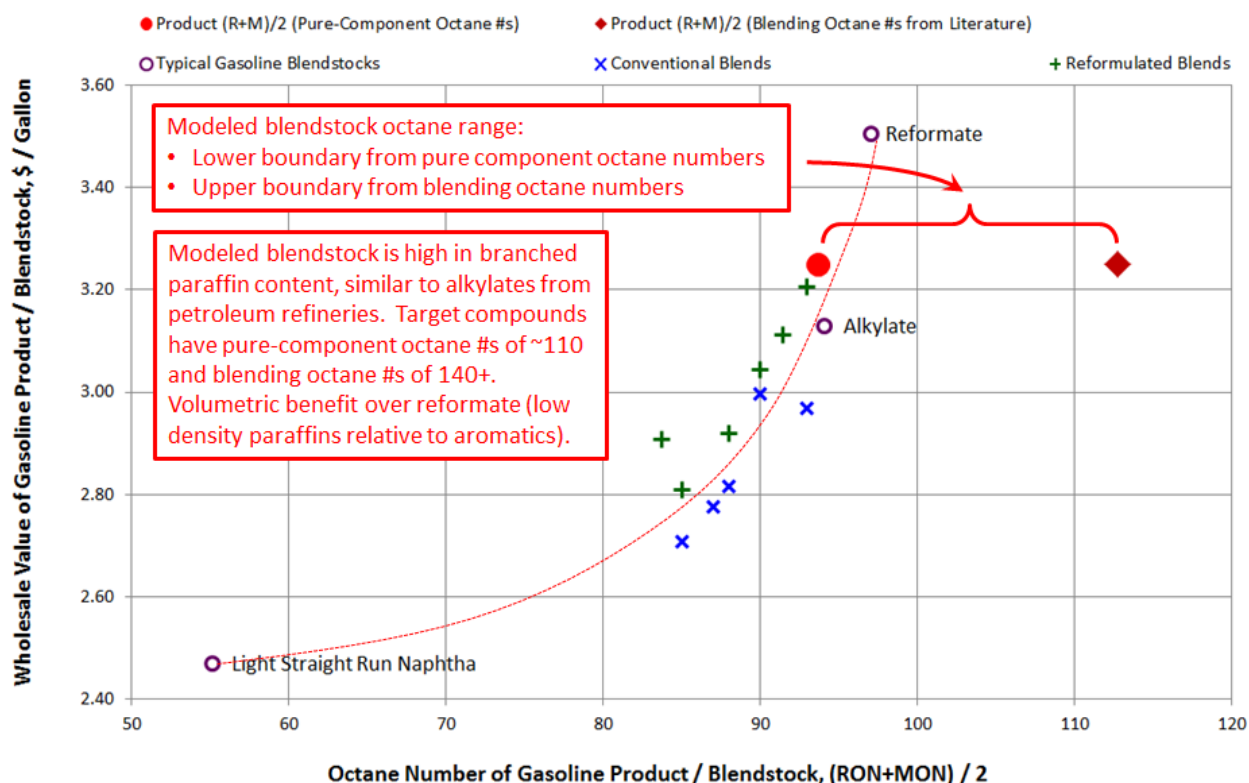
87. **Comment**: A 30 year life is too long – it should be no more than 20 years.

    Response: The impact of shorter plant life on project economics is captured in the Sensitivity Analysis in Section 5. Decreasing the plant life from 30 years to 20 years increases the minimum fuel selling price from $3.41 per GGE to $3.53 per GGE. The plant life of 30 years assumption for the nth-plant analysis is consistent with assumptions for other DOE-BETO funded analyses from NREL and PNNL.

88. **Comment:** Process guarantees will be needed to get debt financing. Process guarantees can be quite expensive for the deployment of a first plant.

    Response: Process guarantees have historically not been included in nth-plant economic analyses. The authors recognize that performance guarantees are critical in reducing risk to

secure financing for pioneer plants. The authors will consider inclusion of financial risk-mitigating measures in future analyses of pioneer plant costs.

89. **Comment:** The (maintenance cost) estimate is too low. As based upon the complexity of the integrated process, 8% is more realistic.

    Response: The annual maintenance cost assumption for the nth-plant analysis is 3% of the fixed capital investment (FCI), which is consistent with factors applied in other DOE-BETO funded analyses from NREL and PNNL and published resources for factored costs [28]. In addition to the 3% of FCI allocation for maintenance costs, the fixed operating cost calculations include labor, overhead, and benefits for 16 full-time maintenance technicians and a maintenance supervisor.

90. **Comment:** Is it possible to make composition units consistent (e.g. all mole %) instead of using ppmv for some constituents? This comment applies to a few other tables.

    Response: The authors recognize that this presentation may be confusing. However, it makes sense to present species with trace concentrations in molar parts-per-million rather than in mole %. This is common practice for identifying concentrations of trace compounds.

91. **Comment:** If at all possible, can you provide catalyst compositions in Table 30 (even if in text earlier)?

    Response: The authors have listed below best estimates for catalyst compositions based on availability of published information. However, there are instances where the compositions are proprietary, which is the case for the metal-modified beta-zeolite catalyst for hydrocarbon synthesis.

Catalyst	Composition (wt%)
Magnesium oxide (MgO)	60.3 Mg + 39.7 O
Olivine	25.4 Mg + 14.6 Fe + 18.3 Si + 41.7 O
Tar reformer catalyst	47.6 Al + 42.4 O + 10.0 (SiO_2, MgO, CaO, BaO, Na_2O) [7]
Methanol synthesis catalyst	52.0 Cu + 20.0 Zn + 5.3 Al + 22.7 O [8]
DME catalyst (Al_2O_3)	52.9 Al + 47.1 O
Beta zeolite catalyst	50.9 Al + 1.8 Si + 47.3 O [9]
Zinc oxide catalyst (ZnO)	80.3 Zn + 19.7 O

92. **Comment:** From what I can tell, the product from this analysis is not shown in Figure 11. MTG composition is cited as from a 2011 report. Can you include composition of this product on this figure?

    Response: The product from this pathway is commonly referred to in the text as "high-octane gasoline" or "high-octane gasoline blendstock." Figure 11 shows that the high octane

169

gasoline blendstock from this pathway contains much higher concentrations of branched paraffins and much lower concentrations of aromatics relative to traditional MTG product.

In the final version of the report, the authors intend to include additional comparisons between the high-octane gasoline blendstock and fossil-derived blendstocks/finished fuels on the basis of octane value. An example is provided in Response # 85.

93. **Comment:** For these three categories of chemicals, can you provide specific chemical identities (LO-CAT, boiler feed, and cooling tower).

    Response: Process chemicals for LO-CAT, boiler feed water treatment, and cooling tower water treatment are commercially available. The authors have listed below best estimates for compositions of the makeup chemicals based on availability of published information.

Process Raw Material(s)	Description of Materials
LO-CAT Chemicals	Primary chemical consumptions are chelated iron and caustic [10]. Specific ratios are not known for the proprietary process.
Boiler Feed Water Treatment Chemicals	Chemicals utilized for boiler feed water and/or steam condensate treatment are an oxygen scavenger like sodium sulfite (Na_2SO_3) or hydrazine (N_2H_4) and a corrosion inhibitor/neutralizing amine like morpholine (C_4H_9NO) [29].
Cooling Tower Treatment Chemicals	Chemicals utilized for cooling tower water treatment generally contain proprietary combinations of corrosion inhibitor, oxygen scavenger, pH control, and biological fouling mitigation. Typical compounds used in treatment programs are phosphates, azoles, copolymers, and zinc [29].

94. **Comment:** Can compare water consumption values to May Wu's recent analysis in her technical report. https://greet.es.anl.gov/publication-consumptive-water

    Response: Water consumption is compared with the suggested reference. An incorrect reference number in the text is likely causing some confusion. The correct reference is below. The authors will correct the reference number for the final report.

    Wu, M.; Mintz, M.; Wang, M.; Arora, S. *Consumptive Water Use in the Production of Ethanol and Petroleum Gasoline*. ANL/ESD/09-1. Argonne, IL: Center for Transportation Research, Energy Systems Division, Argonne National Laboratory, January 2009. http://www.transportation.anl.gov/pdfs/AF/557.pdf.

95. **Comment:** In Figure 15, it seems like you might want to have catalyst cost separate from feedstock because the text says feedstock is the driver. Would be helpful to see that catalyst is only a small (if that is the case) portion of the red bar in the Feedstock and Handling category.

    Response: The authors believe the cost categories presented in the figure are sufficient to show the major impacts to economics. There are no catalysts utilized in the feedstock and handling section of the plant. Therefore, the red bar associated with this process area represents feedstock raw material costs only.

96. **Comment:** Suggestion to use up-to-date values for the electric grid mix. We update these in GREET every year based on EIA.

   Response: The authors will update the GHG basis value for the U.S. average electricity mix from the grid for consistency with the basis from Argonne National Laboratory's GREET model software. The authors will also apply the basis from GREET for consumption of other non-renewable (fossil) materials.

97. **Comment:** Referring to Table F-3. biomass we use as 0.29%. The "other" category in GREET includes hydro, geothermal, wind, solar PV. As this table stands, that would include the share from biomass. ANL can provide the authors with the split for the "renewable" category if GREET values will be adopted.

   Response: The authors appreciate the information and will update the electricity grid mix distribution table to accurately reflect the GHG emission factor mentioned on Comment #17.

98. **Comment:** Are upstream impacts based on European or US scenario? E.g., is fuel consumed in T&D based on European fuel production or US fuel production? Should be using US data if you are not currently.

   Response: The process used is based on an EcoInvent process for petroleum fuel consumed in transmission and distribution (T&D) based on European fuel production. Being such a minor component in the fuel cycle, the authors believe a change from European to U.S. fuel mix would result in negligible impact the results. However, if the reviewer can provide the emissions factor for T&D used in GREET, the authors will integrate the factors for the final version of the report.

99. **Comment:** DOE-BETO asked that MFSP be shown as $/gge for the pyrolysis design case.

   Response: The authors have tried to provide MFSP values as both $/gallon of high-octane blendstock product and $/gasoline equivalent wherever possible/appropriate.

100. **Comment:** The work appears premised on a target manufacturing price of $3/gal by 2022, which needs to be stated clearly in the executive summary; also, all the major future technical development assumptions made in support of achieving this target need to be stated explicitly upfront. I believe this would help interested parties better target their near-term research and development efforts.

   Response: This is an excellent recommendation. The authors will include additional information to explain how the technical and cost targets presented in the design report relate to the BETO programmatic goals and state of technology metrics reported in BETO's Multi-Year Program Plan (MYPP).

101. **Comment:** Is downtime (the reciprocal on-stream time) accounted for as zero operating costs, or does it include fixed operating costs that are normally being paid during both scheduled and unscheduled downtime? The MFSP doesn't appear very sensitive to the on-stream factor (at least subjectively). It may help explicitly stating how downtime is accounted for.

   Response: Yes, downtime is equal to the difference between calendar time and on-stream time. Fixed operating costs are incurred in full whether or not the plant is producing at full

171

capacity. The fixed costs include labor, overhead and benefits, taxes and insurance, and plant maintenance costs. However, maintenance costs are based on a factor (3% of fixed capital investment) and do not increase with increasing downtime. In reality, increased downtime equates to increased maintenance/turnaround costs. The sensitivity scenarios for MFSP as a function of on-stream factor (on-stream time/calendar time), with constant maintenance costs, are summarized below.

On-Stream Factor	MFSP ($ / GGE)	% Δ from Base Case
0.80 (High Downtime)	3.66	+ 7.4%
0.90 (Base Case)	3.41	
0.96 (Low Downtime)	3.28	− 3.7%

102. **Comment:** Comments on total capital investment related calculations. This section is hard to follow. Please add an equation to illustrate how the costs add up.

Response: The authors believe the confusion in Table 33 may be due to the indirect costs percentages based on TPEC, TDC, and FCI. The basis for the indirect cost calculations are the percentages of total direct costs (summing to 70%). The indirect cost factors are consistent with assumptions for other DOE-BETO funded analyses from NREL and PNNL. The values were recommended and agreed upon through the peer review process. The additional percentages presented in Table 33 (TPEC and FCI) are presented for comparisons with other published sources [28]. Per reviewer request, equations are provided below.

TPEC = Total Purchased Equipment Costs

TIC = Total Installed Costs = TPEC * Installation Factor

TDC = Total Direct Costs = TIC + Site Development Costs

IDC = Indirect Costs = 70% of TDC

FCI = Fixed Capital Investment = TDC + IDC

TCI = Total Capital Investment = FCI + Land Costs + Working Capital (5% of FCI)

103. **Comment:** Please discuss where the factors (in Table 33, p. 52) come from and why they were chosen to be used in this analysis.

Response: The indirect cost factors are consistent with assumptions for other DOE-BETO funded analyses from NREL and PNNL. The values were recommended and agreed upon through the peer review process. Seventy percent of total direct costs (TDC) correspond to approximately 40% of fixed capital investment (FCI) for this design, which is consistent with the 16% to 61% of FCI range presented by Peters, Timmerhaus and West [28].

104. **Comment:** Considering that "the potential to be cost-competitive with gasoline" seems to be the main finding of this report, it would need additional clarifications and discussion on what is meant by this. The $3.19/gge equals 119 $/bbl of crude (assuming 15 $/bbl refining margin) which is close to current long-term price projections, but not under. In addition, it

172

might be worthwhile to mention that this analysis was made based on Nth plant assumptions and it is likely that first such plants will be associated with higher costs.
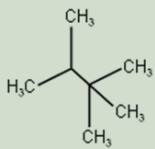
Response: These are excellent suggestions. The authors will incorporate additional discussion around this conclusion to address these points in Section 6 (Conclusions).

105. **Comment:** Suggest adding GTG LCA GHG reduction or equivalent performance metric to Sustainability Metrics section (Page v).

Response: GHG and fossil energy consumption are added to the Sustainability Metrics in the analysis summary presented on page iv of the report.

106. **Comment:** Differences between ExxonMobil MTG and MTG process…suggest add catalysts to list (Page 1).

Response: In this section, the authors are highlighting the key difference between the ExxonMobil MTG and the current process, namely the hydrocarbons produced and process conditions. However, the authors agree that a comparison between MTG and the high-octane gasoline blendstock pathway would be useful and will work to incorporate a more detailed comparison in the final report (perhaps as an appendix). An example of a possible comparative summary is provided below.

Process Attribute	High-Octane Gasoline Pathway Target	Methanol to Gasoline (MTG) Pathway	Impact on Techno-Economic Analysis
Molecular structures favored in synthesis reactions Example Compound Specific Gravity	Branched paraffins Triptane 0.70	Aromatics Toluene 0.87	High octane product rich in branched paraffins, similar to a refinery alkylate. H-saturation decreases density, increasing product volume.
Hydrocarbon synthesis catalyst	Beta-Zeolite (12-membered rings)	ZSM-5 (10-membered rings)	Different pore sizes and structures result in different compound selectivities.
Octane number of gasoline-range product	RON: 95+ MON: 90+	RON: 92 MON: 83	Octane number increases value of product as a finished fuel blendstock.
Selectivity of C_5+product	C_5+ product only (~65 Gal / Ton)	~ 85% C_5+ (~55 Gal / Ton)	High selectivity to primary (premium quality) product maximizes overall product value.
Severity of synthesis operating conditions	350 – 450 Deg. F 130 PSIA	650 – 950 Deg. F 315 PSIA	The lower severity operating conditions result in lower capital and operating costs relative to MTG.
Coke formation	Coke formation is minimized by hydrogen addition and selectivity to branched paraffins rather than aromatics.	High propensity for coke formation due to aromatic coke pre-cursors.	Minimizing coke formation helps to maximize product yield / carbon efficiency and maximizes catalyst regeneration and replacement cycles.

107. **Comment:** Initial catalyst/chemicals/product storage TIC of $725K seems low for this process that includes several multi-bed catalytic processes (Page 42, Table 28).

Response: The category in Table 28 "catalyst/chemicals/product storage" will be renamed "chemicals storage." The fixed bed catalyst storage requirements are likely warehouse locations (indoor or covered) to accommodate significant numbers of "super-sacks" of

173

extrudates. Site development costs ($8MM+) are allocated in the economic assessment for such requirements.

108. **Comment:** Good listing of assumptions around catalysts and chemicals. What of these assumed catalysts lifetimes/attrition rates are based on credible experimental data? -Would be useful to include some rationale (Page 54, Table 35).

Response: The lifetime of each catalyst is specified in Table 35. Among all the catalysts, only reformer catalyst is replaced continuously. The tar reformer catalyst loss/purge rate per day is set at 0.15% of the catalyst inventory and the same amount of fresh catalyst is supplemented. As part of the 2012 demonstration, the catalyst loss rate was evaluated separately at the pilot scale by Rentech. Reforming catalyst loss rates are discussed in the demonstration report from Dutta et al. [5], which is located here:http://pubs.acs.org/doi/ipdf/10.1021/ie402045q

109. **Comment:** Not finding a process yield sensitivity…earlier it is stated that yield is key to economics (Page 58, Figure 16)

Response: The authors will include sensitivity scenario(s) for product yield/selectivity in the final version of the report.

110. **Comment:** What premium can be expected for high octane gasoline blendstock (Page 57)?

Response: The process technology is evaluated based on the DOE-BETO cost target for 2022 of $3.00 per GGE . In addition, the value of a gasoline blendstock is a function of its properties (i.e., octane value). High-octane blendstocks are likely to carry a premium price relative to the finished fuel value. Refer to the octane comparisons presented in the figure of Response #86, which will be included in the final version of the design report. In addition, the research effort must validate the actual blending benefits for the high-octane blendstock product.

111. **Comment:** Under "Hydrocarbons produced", you say fuel standards are changing to limit the amount of aromatics in gasoline. It would be useful to elaborate a bit on what those changes are likely to be. (Page 1)

Response: This statement is specific to benzene and is based on the EPA regulatory announcement from February 2007 [30], which defines the regulations for Mobile Source Air Toxics (MSAT). MSAT requires U.S. refiners to meet (1) an annual average benzene content of 0.62 vol % in gasoline (reformulated and conventional) and (2) a maximum average benzene standard of 1.3 vol %. This regulation puts pressure on refiners to continue to meet gasoline specifications, but limit blending of high-octane blendstocks like reformate, which contain substantial amounts of benzene. High-octane blendstocks containing relatively high concentrations of paraffins rather than aromatics would then be desirable to refiners/finished fuel blenders.

112. **Comment:** Just below Figure 3 the sentence reads, "The proposed thermochemical ethanol production process …" which I believe should read, "The proposed thermochemical gasoline production process …" (Page 8)

Response: The authors have corrected this error and updated the text to state "The proposed high-octane gasoline production process…"

113. **Comment:** In section 2.3 Aspen Plus Model, the 4th sentence reads, "The thermochemical ethanol design …" which should read, "The thermochemical gasoline design …" (Page 11)

Response: The authors have corrected this error and updated the text to state "The thermochemical conversion of biomass to high-octane gasoline design…"

114. **Comment:** Very tiny formatting issue: in the 3rd paragraph, the "-" sign for C3-hydrocarbons is not subscripted in the 1st and 3rd instances. Chemists notice these things. (Page 36)

Response: The authors have addressed the formatting issue.

115. **Comment:** At first glance, I thought "Paraffins" in Figure 11 denoted only n-paraffins. It might be good to note that it includes both n- and i-paraffins. Even better would be to show those separately in the bars, since you stress the importance of iso-paraffins for high octane value. (Page 37)

Response: This is a good suggestion. Based on the reviewer comment, the authors will make the note in the figure that paraffins include both n- and i- paraffins or design the figure to separately identify n- and i- paraffins.

116. **Comment:** In the paragraph just ahead of Table 28, part of the sentence is in a smaller font. (Page 42)

Response: The authors have addressed the font discrepancy.

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