



Process Design and Economics for the Conversion of Lignocellulosic Biomass to Hydrocarbons: Dilute-Acid and Enzymatic Deconstruction of Biomass to Sugars and Catalytic Conversion of Sugars to Hydrocarbons

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Executive Summary

The U.S. Department of Energy (DOE) promotes the production of an array of liquid fuels and fuel blendstocks from lignocellulosic biomass feedstocks by funding fundamental and applied research that advances the state of technology in biomass collection, conversion, and sustainability. As part of its involvement in this program, the National Renewable Energy Laboratory (NREL) investigates the conceptual production economics of these fuels.

Over the past decade, NREL conducted a campaign to quantify the economic implications associated with observed and future targeted performance for the biochemical conversion of corn stover to ethanol through techno-economic modeling. This effort served to set “state of technology” benchmarks and to guide research and development by setting cost targets and tracking progress towards final achievement of these targets in 2012. Beginning in 2013, NREL began transitioning from the singular focus on ethanol to a broad slate of products and conversion pathways, ultimately to establish similar benchmarking and targeting efforts. The first product from this expanded focus was a technical report documenting a conceptual process pathway for biological (i.e., microbial) conversion of cellulosic sugars to hydrocarbon fuels. The analysis in this design report expands from this initial baseline to consider an approach via *catalytic* conversion of sugars and other hydrolysate components to hydrocarbons.

This report describes one potential conversion process to hydrocarbon products by way of catalytic conversion of lignocellulosic-derived hydrolysate. This model leverages expertise established over time in biomass deconstruction and process integration research at NREL, while adding in new technology areas for sugar purification and catalysis. The overarching process design converts biomass to diesel- and naphtha-range fuels using dilute-acid pretreatment, enzymatic saccharification, purifications, and catalytic conversion focused on deoxygenating and oligomerizing biomass hydrolysates. Ancillary areas—feed handling, hydrolysate conditioning and upgrading (hydrotreating) to final fuel products, wastewater treatment, lignin combustion, and utilities—are also included in the design. Detailed material and energy balances and capital and operating costs for this baseline process are also documented.

This techno-economic analysis models a production cost for cellulosic hydrocarbon biofuels that can be considered as a baseline to assess the competitiveness and market potential for the technology. It can also be used to quantify the economic impact of individual conversion performance targets and prioritize them in terms of their potential to reduce cost. The analysis presented here also includes consideration of the life-cycle implications of the process model by tracking sustainability metrics for the modeled biorefinery, including greenhouse gas emissions, fossil energy demand, and water consumption.

Building on prior design report practices, NREL, together with Harris Group, performed a feasibility-level analysis for a plausible catalytic conversion process to meet an intermediate DOE cost goal of \$5/gallon gasoline equivalent (GGE) in a timeframe prior to 2022. The modeled biorefinery processes 2,205 dry tons biomass/day and achieves a fuel selling price of \$4.05/GGE in 2011-year dollars as determined by modeled conversion targets and “*n*th-plant” project costs and financing, associated with a total fuel yield of 78.3 GGE/dry ton when assuming hydrogen is purchased from off-site production. Alternatively, if hydrogen is instead produced *in situ* by diverting a fraction (41%) of hydrolysate carbon towards reforming

reactions, fuel yield drops to 45.3 GGE/dry ton and selling price increases to \$5.48/GGE. If instead hydrogen is produced by diverting a fraction of biomass to a gasification process to produce and subsequently refine syngas to hydrogen, such a scenario would require 36% of the available biomass and would translate to a fuel yield of 50.1 GGE/dry ton and selling price of \$4.95/GGE. Preliminary assessment of sustainability metrics associated with these three hydrogen scenarios indicates a more favorable greenhouse gas profile for the *in situ* and gasification cases as it avoids the use of natural gas for off-site hydrogen production. However, additional benefits are likely to be seen for a case including on-site hydrogen production via steam methane reforming (rather than off-site as in the base case evaluated here), given additional integration advantages that may be brought about through heat/power integration and carbon recycles which were beyond the scope of this effort; in any case, a full well-to-wheel life-cycle analysis is required to better understand greenhouse gas implications for the fully integrated technology pathway. Finally, the report includes a brief discussion on improvements needed to achieve a final 2022 DOE target of \$3/GGE moving forward, focused on coproducts from lignin.

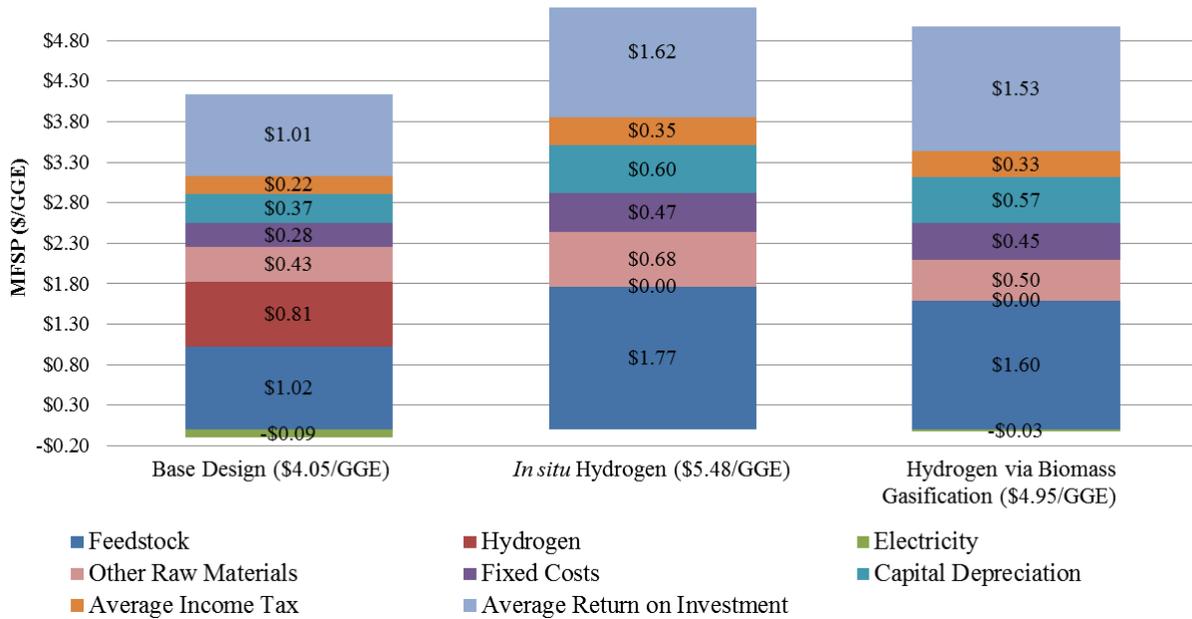


Figure ES-1. Minimum fuel selling price results for external (purchased) hydrogen base design versus internal hydrogen production via *in situ* reforming of hydrolysate and biomass gasification

Biomass Derived Hydrolysate Catalysis to Hydrocarbons Process Engineering Analysis

Dilute Acid Pretreatment, Enzymatic Hydrolysis, Hydrolysate Catalytic Conversion, Hydrotreating to HC
All Values in 2011\$

Minimum Fuel Selling Price (MFSP): \$4.05 /GGE

Contributions:	Feedstock	\$1.02 /GGE
	Enzymes	\$0.21 /GGE
	Non-Enzyme Conversion	\$2.81 /GGE
	HC Production	56.7 MM GGE/yr
	HC Yield	78.3 GGE/dry U.S. ton feedstock
Carbon Retention Efficiency for Hydrolysate Catalysis		0.856 carbon in HC product/carbon in sugars
	Feedstock + Handling Cost	\$80.00 /dry U.S. ton feedstock
	Internal Rate of Return (After-Tax)	10%
	Equity Percent of Total Investment	40%

Capital Costs	Manufacturing Costs (cents/GGE)
Pretreatment	Feedstock + Handling 102.18
\$51,400,000	Sulfuric Acid 3.43
Neutralization/Conditioning	Ammonia (pretreatment conditioning) 2.37
\$2,200,000	Caustic (deacetylation pretreatment) 3.59
Enzymatic Hydrolysis, Hydrolysate Conditioning & Purification	Glucose (enzyme production) 12.01
\$68,800,000	Hydrogen 80.63
On-site Enzyme Production	Catalysts for Conversion Reactors 10.94
\$12,400,000	Other Raw Materials 8.49
Catalytic Conversion & Upgrading	Waste Disposal 2.44
\$82,900,000	Net Electricity -9.46
Wastewater Treatment	Fixed Costs 28.36
\$47,700,000	Capital Depreciation 36.85
Storage	Average Income Tax 21.64
\$4,800,000	Average Return on Investment 101.30
Boiler/Turbogenerator	
\$76,300,000	
Utilities	
\$7,000,000	
Total Installed Equipment Cost	Manufacturing Costs (\$/yr)
\$353,500,000	Feedstock + Handling \$57,900,000
Added Direct + Indirect Costs	Sulfuric Acid \$1,900,000
\$306,100,000	Ammonia (pretreatment conditioning) \$1,300,000
(% of TCI)	Caustic (deacetylation pretreatment) \$2,000,000
46%	Glucose (enzyme production) \$6,800,000
Total Capital Investment (TCI)	Hydrogen \$45,700,000
\$659,600,000	Catalysts for Conversion Reactors \$6,200,000
	Other Raw Materials \$4,800,000
Installed Equipment Cost/Annual Gallon	Waste Disposal \$1,400,000
\$6.23	Net Electricity -\$5,400,000
Total Capital Investment/Annual Gallon	Fixed Costs \$16,100,000
\$11.63	Capital Depreciation \$20,900,000
Loan Rate	Average Income Tax \$12,300,000
8.0%	Average Return on Investment \$57,400,000
Term (years)	
10	
Capital Charge Factor (computed)	
0.137	
Carbon Retention Efficiencies:	Specific Operating Conditions
From Sugar Catalytic Reaction (Fuel C / Sugar C)	Total H2 Reaction Consumption (wt% of hydrolysate) 6.5%
85.6%	Excess Electricity (KWh/GGE) 2
From Biomass (Fuel C / Biomass C)	Plant Electricity Use (KWh/GGE) 5
45.3%	Enzyme Loading (mg/g cell) 10
From Total C (Fuel C / Biomass + Natural Gas C) ^a	Saccharification Time (days) 3.5
35.7%	

^a Includes carbon attributed to natural gas for external SMR H2 production

Figure ES-2. Economic summary for catalytic conversion base case process (purchased H₂)

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Acronyms

APR	aqueous phase reforming	MFSP	minimum fuel selling price
BETO	Bioenergy Technologies Office	MM	million
BFW	boiler feed water	MVR	mechanical vapor recompression
CIP	clean-in-place	MYPP	BETO's Multi-Year Program Plan
COD	chemical oxygen demand	NPV	net present value
CSL	corn steep liquor	NREL	National Renewable Energy Laboratory
DAP	diammonium phosphate	OSBL	outside battery limits (of the plant)
DB	declining balance	PFD	process flow diagram
DCFROR	discounted cash flow rate of return	P&ID	pipng and instrumentation design
DOE	U.S. Department of Energy	PNNL	Pacific Northwest National Laboratory
FCI	fixed capital investment	PSA	pressure swing adsorption
FGD	flue gas desulfurization	R&D	research and development
GBT	gravity belt thickener	RDB	renewable diesel blendstock
GGE	gallon gasoline equivalent	RFS	Renewable Fuel Standard
GHG	greenhouse gas	RO	reverse osmosis
H/D	height-to-diameter ratio	SCF	standard cubic feet
HDO	hydrodeoxygenation	SMR	steam-methane reforming
HHV	higher heating value	SOT	annual State of Technology case
HMF	5-hydroxymethyl furfural	TAN	total acid number
INL	Idaho National Laboratory	TCI	total capital investment
IS	insoluble solids	TDC	total direct cost
IRR	internal rate of return	TEA	techno-economic analysis
ISBL	inside battery limits (of the plant)	TS	total solids
IX	Ion Exchange	WHSV	weight hourly space velocity
LCA	life cycle assessment	WWT	wastewater treatment
LCI	life cycle inventory		
LHSV	liquid hour space velocity		
LHV	lower heating value		
MACRS	IRS Modified Accelerated Cost Recovery System		
MESP	minimum ethanol selling price		

1 Introduction

1.1 Background and Motivation

The U.S. Department of Energy (DOE) Bioenergy Technologies Office (BETO) promotes the production of liquid fuels from lignocellulosic feedstocks by sponsoring programs in fundamental and applied research that aim to advance the state of biomass conversion technology. These programs include laboratory research to develop improved cellulose hydrolysis enzymes and metabolic conversion microorganisms through synthetic biology, catalyst development and testing, chemical and mechanical pretreatment work, detailed engineering studies of potential processes, and construction of pilot-scale demonstration and production facilities. This research is conducted by national laboratories, universities, and private industry in conjunction with engineering and construction companies.

To support the DOE program, the National Renewable Energy Laboratory (NREL) investigates the process design and economics of cellulosic biofuel manufacturing in order to develop a plant gate price for fuels and fuel blendstocks based on process and plant design assumptions consistent with applicable best practices in engineering, construction, and operation. This plant gate price is referred to as the *minimum fuel selling price* or MFSP. The MFSP can be used by DOE to assess the cost-competitiveness and market penetration potential of a given cellulosic biofuel technology pathway in comparison with petroleum-derived fuels and established biofuel technologies such as starch- or sugar-based ethanol.

The techno-economic analysis (TEA) effort at NREL also helps to direct biomass conversion research by examining the sensitivity of the MFSP to process alternatives and research advances. Proposed research and its anticipated results can be translated into a new MFSP that can be compared to the benchmark case documented in this report. Such comparison helps to quantify the economic impact of core research targets at NREL, and elsewhere, and to track progress toward meeting competitive cost targets. It also allows DOE to make more informed decisions about research proposals that claim to reduce MFSP, and to better quantify trade-offs between process metrics (yields, chemical inputs, carbon efficiencies), economics (MFSP), and sustainability (greenhouse gas, fossil energy, water profiles) among process options for a given technology pathway or amongst multiple pathways.

For more than 10 years, NREL has developed design case models and associated reports [1, 2] that documented process and cost targets for ethanol production from cellulosic feedstocks via biochemical conversion (e.g., deconstruction to monomeric sugars followed by fermentation to ethanol), based on the best understanding of the technology and equipment costs at the time. As understanding evolved on process and economic metrics, models were refined and re-benchmarked relative to updated targets. The final update to the ethanol design targets was published in 2011 (referred to hereafter as the “2011 design report”) [2] and established an MFSP cost goal of \$2.15/gal ethanol to be achieved in 2012 based on performance data from pilot-scale demonstration runs conducted at NREL. This goal was subsequently achieved through 2012 pilot plant trials, with the pertinent experimental data input to the model, which calculated an ethanol MFSP of \$2.15/gal for a commercial-scale “*n*th-plant” facility processing 2,000 dry metric tons/day of corn stover [3-5]. Further context on the history of NREL’s ethanol program, including individual year-by-year modeled production costs, may be found in Tao et al. and Davis [4, 6].

While further room for improvement exists to continue reducing ethanol costs, the achievement of the 2012 minimum ethanol selling price (MESP) target represented the culmination and closeout of NREL's primary charter to demonstrate technology improvements required to de-risk the technology in support of moving it toward commercialization, and to provide a benchmark for private industry to expand upon. Moving forward, in 2013 DOE-BETO began transitioning its core platform focus from ethanol to hydrocarbon fuel and blendstock products (fungible/infrastructure-compatible fuels), including a new pathway MFSP cost target of \$3/gallon gasoline equivalent (GGE) in 2022 for the biochemical (sugar conversion) platform [7]. Within the biochemical platform, two options were established to consider for sugar conversion to hydrocarbon fuels, namely biological (microbial "fermentation") conversion and catalytic conversion. The biological conversion pathway was first modeled, analyzed, and documented in a design report published in 2013 (hereafter referred to as the "2013 design report") [8], which presented a modeled MFSP of \$5.10/GGE and hydrocarbon fuel yield of 45 GGE/ton as an intermediate target case prior to 2022, with additional insight provided for a path forward to ultimately meet the final 2022 target of \$3/GGE.

As described in the 2013 design report, a key benefit of the biological conversion approach is the ability to produce specific, easily-characterized and potentially high-value hydrocarbon products afforded by metabolic engineering. However, a drawback of the pathway lies in the challenging carbon efficiency and fuel yields inherent to metabolic conversion of sugars (rejecting oxygen as CO₂), as well as challenges for large-scale aerobic bioconversion. Building from the basis established for the 2013 design report, the focus of this report is to similarly document a plausible pathway model for the catalytic conversion of cellulosic sugars to hydrocarbon blendstock fuels, both to enable achieving similar future cost targets, as well as to compare and contrast key differences in this technology pathway relative to the biological pathway. The overall scope of the report focuses on model details attributed to achieving intermediate MFSP costs of \$5/GGE or less prior to 2022, but with additional brief insight provided for a path forward to ultimately meet the final 2022 target of \$3/GGE.

1.2 Process Overview

The process modeled in this report uses preprocessing (deacetylation) and cocurrent dilute-acid pretreatment of lignocellulosic biomass feedstock (primarily corn stover), followed by enzymatic hydrolysis (saccharification) of the remaining cellulose, followed by hydrolysate conditioning and catalytic conversion/upgrading of the resulting hydrolysate soluble carbon components to naphtha- and diesel-range fuel products. The process design also includes feedstock handling and storage, product upgrading, wastewater treatment (WWT), lignin combustion, product storage, and utilities. The process is divided into nine areas (Figure 1).

Area 100: Feed handling. The feedstock, in this case milled corn stover blended with switchgrass, is delivered to the feed handling area from a uniform-format feedstock supply system. Only minimum storage and feed handling are required. From there, the biomass is conveyed to the pretreatment reactor (Area 200).

Area 200: Pretreatment and conditioning. In this area, the biomass is processed in an alkaline deacetylation step to solubilize and remove acetate as well as portions of ash, lignin, and other components. It is then drained and treated with dilute sulfuric acid catalyst at a high temperature for a short time to liberate the hemicellulose sugars and break down the biomass for enzymatic

hydrolysis. Ammonia is then added to the whole pretreated slurry to raise its pH to approximately 5 for enzymatic hydrolysis.

Area 300: Enzymatic hydrolysis, hydrolysate conditioning, and purification. Enzymatic hydrolysis is initiated in a high-solids continuous reactor using a cellulase enzyme produced on site. The partially hydrolyzed slurry is batched to one of several parallel reactors. Hydrolysis is completed in the batch reactor with a total time of 3.5 days between the continuous and batch steps. The slurry is then fed to a vacuum filter press to remove insoluble solids (primarily lignin) and the solids fraction exiting the filter press is sent to the boiler (Area 800). The hydrolysate is then concentrated in a mechanical vapor recompression evaporator. A polishing filter is used to remove any residual insoluble particulates carried over from the filter press and any solids that may have precipitated in the evaporator. Finally, the filtered and concentrated hydrolysate is treated by ion exchange to remove soluble cations and anions prior to catalytic upgrading.

Area 400: Cellulase enzyme production. An on-site enzyme production section was maintained in this design, consistent with the details provided in the 2011 ethanol design report. Purchased glucose (corn syrup) is the primary carbon source for enzyme production. Media preparation involves a step in which a portion of the glucose is converted to sophorose to induce cellulase production. An enzyme-producing fungus (modeled after *Trichoderma reesei*) is grown aerobically in fed-batch bioreactors. The whole fermentation broth, containing the secreted enzyme, is fed to Area 300 to carry out enzymatic hydrolysis.

Area 500: Catalytic conversion and upgrading. The catalytic upgrading stage begins after the hydrolysate stream is concentrated and purified in Area 300. The catalytic conversion process consists of four stages: hydrogenation, aqueous phase reforming (APR), condensation and oligomerization, and hydrotreating. Each stage consists of packed-bed reactor vessels. Hydrogen is added to the reactors in each stage which operate at varying process conditions and have varying catalyst composition. The goal of these successive catalytic steps is to remove oxygen or “de-functionalize” carbohydrates and other carbon components and oligomerize them to primarily diesel-range hydrocarbons.

Area 600: Wastewater treatment. Wastewater streams are treated by anaerobic and aerobic digestion. The methane-rich biogas from anaerobic digestion is sent to the boiler (Area 800), where sludge from the digesters is also burned for the production of steam and power. The treated water is suitable for recycling and is returned to the process.

Area 700: Storage. This area provides bulk storage for chemicals used and produced in the process, including sulfuric acid, ammonia, nutrients, water, and products.

Area 800: Combustor, boiler, and turbogenerator. The solids from the filter press and WWT are combusted along with the biogas from anaerobic digestion and the process off-gases to generate high-pressure steam for electricity production and process heat. Most of the process steam demand is in the pretreatment reactor. The boiler produces excess steam beyond facility heating demands that is converted to electricity for use in the plant.

Area 900: Utilities. This area includes a cooling water system, chilled water system, process water manifold, and power systems.

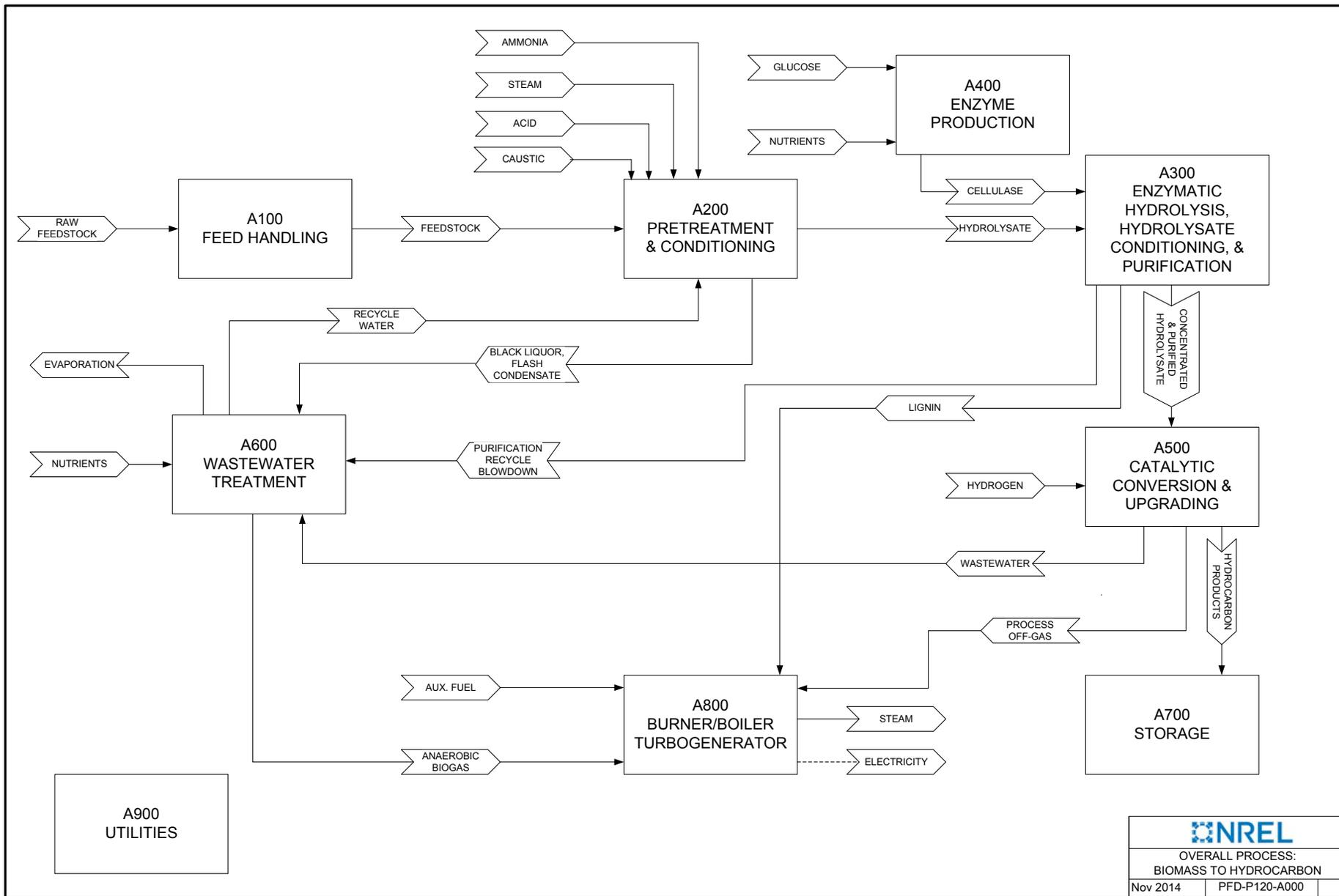


Figure 1. Simplified flow diagram of the overall process. (Key streams only. See Appendix E for more detailed schematic and process flow diagrams.)

1.3 Techno-Economic Analysis Approach

Figure 2 describes the engineering approach used for modeling the conversion of biomass to biofuels, including process design, process modeling, and economic analysis. This approach was largely followed for this study as well, albeit under a condensed timeline and with additional external inputs from literature, patents, consultants, and vendors, for areas of the process in which NREL does not yet have sufficient in-house data (primarily the purification and catalytic conversion steps). As such, this report is less prescriptive in some sections than are previous reports, because of the early stage of understanding for new areas of the process and the somewhat more preliminary nature of the associated models.

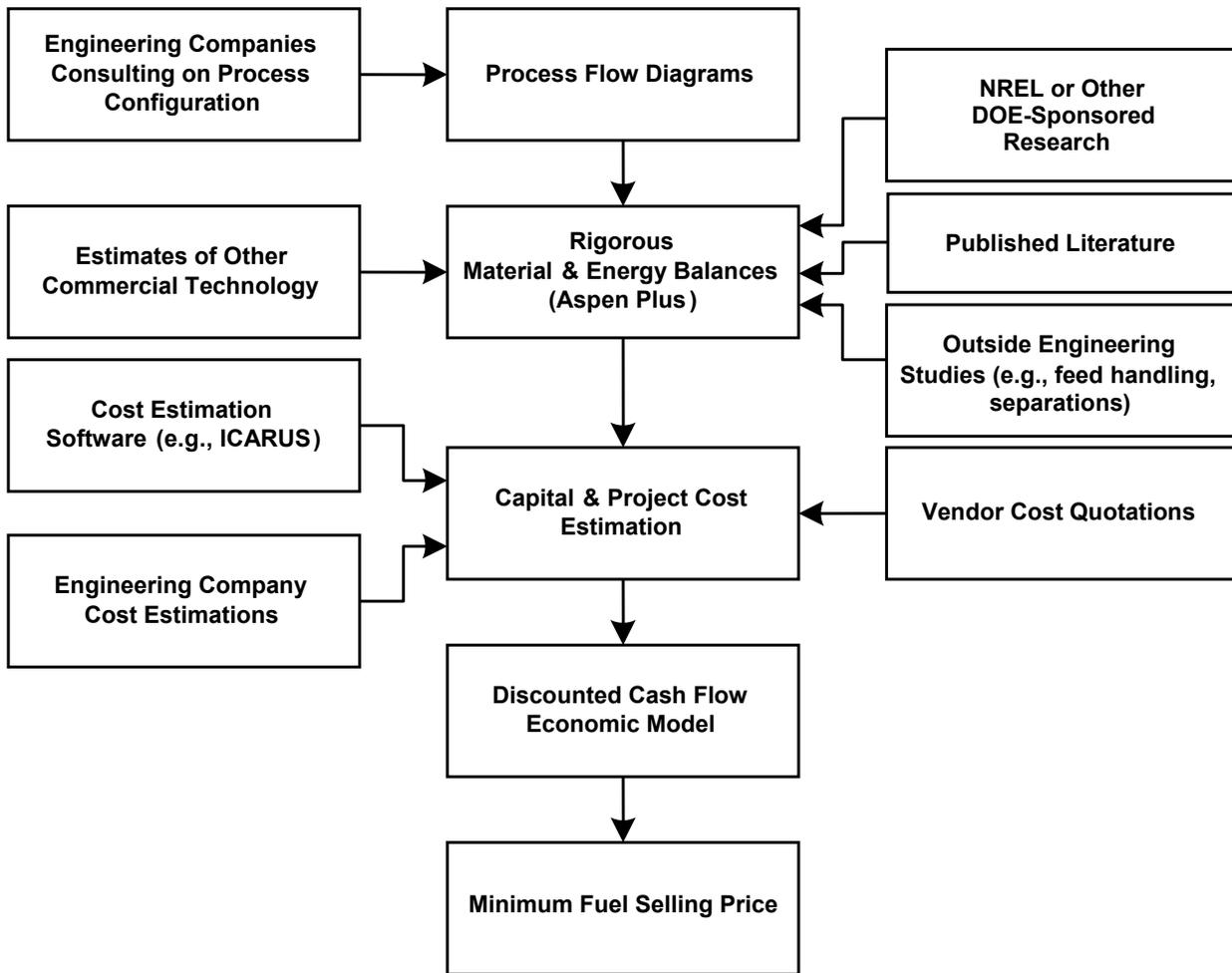


Figure 2. NREL's approach to process design and economic analysis

Starting from the general process flow diagram shown in Figure 1 and the more detailed diagrams contained in Appendix E, a process simulation is developed using Aspen Plus software [9]. This process model computes thermodynamically rigorous material and energy balances for each unit operation in this conceptual biorefinery.

The material and energy balance data from the Aspen simulation are next used to assist in determining the number and size of capital equipment items. As process conditions and flows change, baseline equipment costs are automatically adjusted in an Excel spreadsheet using a scaling factor. These baseline costs come from vendor quotes (a favored procedure for larger or nonstandard unit operations and packaged or skid-mounted subsystems) or from Harris Group's historical cost data (for secondary equipment such as tanks, pumps, and heat exchangers). Harris Group provided updated design and cost estimates based largely on vendor budgetary quotations for critical new or modified areas of the current process model, primarily hydrolysate polishing filtration, ion exchange purification, and catalytic conversion operations, as "new" operations relative to NREL's prior 2013 design report focused on biological conversion. Final equipment costs for this report are tabulated in Appendix A.

Once equipment costs are determined, direct and indirect overhead cost factors (e.g., installation costs and project contingency) are applied to determine a feasibility-level estimate of total capital investment (TCI) in 2011 U.S. dollars. The TCI, along with the plant operating expenses (also developed using flow rates from the Aspen model), is used in a discounted cash flow rate of return (DCFROR) analysis to determine a plant gate price for total hydrocarbon fuel for a given discount rate. In this model, the fuel product is primarily refined "renewable diesel blendstock" (RDB) with a lesser fraction of naphtha. The plant gate price is also called the minimum fuel selling price (MFSP, in \$/GGE), required to obtain a net present value (NPV) of zero for a 10% internal rate of return (IRR) after taxes, associated with an n^{th} -plant model (discussed in the next section).

The product of the analysis described above is a techno-economic model that estimates a product price for a pre-commercial process. The resultant MFSP is unique for the set of process conditions simulated, and it should be emphasized that some amount of uncertainty always exists around these chosen conditions, as well as around the assumptions made for capital and raw material costs. Without a detailed understanding of the basis behind it, the computed MFSP carries a risk of being taken out of context. While the MFSP can be used to assess the marketplace competitiveness of a given process, it is best suited for comparing technological variations against one another or for performing sensitivity analyses that indicate where economic or process performance improvements are needed.

1.4 About n^{th} -Plant Assumptions

The techno-economic analysis reported here uses what are known as " n^{th} -plant" economics. The key assumption implied by n^{th} -plant economics is that our analysis does not describe a pioneer plant; instead, it assumes several plants using the same technology have already been built and are operating. In other words, it reflects a mature future in which a successful industry of n plants has been established. Because the techno-economic model is primarily a tool for studying new process technologies or integration schemes in order to comment on their comparative economic impact, n^{th} -plant analysis avoids artificial inflation of project costs associated with risk financing, longer startups, equipment overdesign, and other costs associated with first-of-a-kind or pioneer plants, lest these overshadow the economic impact of research advances in conversion or process integration. At the same time, NREL also continues to work on quantifying economic factors associated with first-of-a-kind implementation. At the very least, these n^{th} -plant economics should help to provide justification and support for early technology adopters and pioneer plants.

The n^{th} -plant assumptions in the present model apply primarily to the factored cost model used to determine the total capital investment from the purchased equipment cost and to the choices made in plant financing. The n^{th} -plant assumption also applies to some operating parameters, such as process

uptime of 90%. These assumptions were agreed upon by BETO and reflect standard metrics applied universally across multiple national laboratory TEA models to allow for a consistent basis for analysis. It should be emphasized, however, that these assumptions carry a degree of uncertainty and are subject to refinement.

1.5 About the NREL Aspen Model

While Aspen Plus can be thermodynamically rigorous, such detail is not always warranted in the simulation, whether for lack of data or introduction of additional complexity for little gain in accuracy. Some unit operations, particularly solid-liquid separation and hydrolysate purification (polishing filtration and deionization), were modeled with a fixed performance determined by experimental testing or by standard engineering practices confirmed with vendors. Catalytic reactors were modeled using published values from patent literature for carbon yields to product “classes” and carbon number ranges (discussed further in Section 3.5) rather than rigorous kinetics or rate expressions, dictated by a dearth of publicly available guiding data and the granularity of the data which is available. This simple stoichiometric model still satisfies mass and energy balances, and corresponds closely with published yield targets.

The Aspen Plus simulation uses component physical properties internal to the software, as well as property data developed at NREL or from the literature [10, 11]. Similar to the 2011 and 2013 models, the current model does not rely on external property databanks and minimizes the number of custom-defined components within reason. A discussion of components and properties used is given in Appendix D.

2 Design Basis and Conventions

2.1 Plant Size

The plant size in the present design is the same as that used in prior designs: 2,205 dry U.S. ton/day (2,000 metric tonne/day). With an expected 7,880 operating hours per year (90% uptime), the annual feedstock requirement is 724,000 dry U.S. ton/year. The present model maintains consistency with the 2013 design report basis and assumes that feedstock is delivered under a uniform-format logistics system, which includes a blended feedstock consisting of multi-pass corn stover, single-pass corn stover, and switchgrass, to enable achieving cost and composition requirements set for 2017 targets by Idaho National Laboratory (INL) [12]. This blending strategy assumes that the formulation of multiple feedstocks will perform as well as, or better than, a singular feedstock, while allowing for ultimately achieving stipulated feedstock cost targets. There are research plans for FY 2014 to test various blended feedstocks in multiple technology research pathways. Under the new uniform-format system, opportunities may exist to justify a larger facility scale with an increased feed rate to the biorefinery, thus allowing for economy of scale advantages; however, the specifics are not yet defined and we retain the 2,000 dry tonne/day basis here.

2.2 Feedstock Composition

The conversion target feedstock composition remains consistent with what was assumed in the 2011 and 2013 design cases; however, the biomass *supply* composition is updated relative to the original 2011 case. As noted by INL, the previous requirements for feedstocks assumed in the 2011-era ethanol models included volume and cost targets only. There were no quality requirements; further analysis indicated a mismatch on biomass characteristics and conversion in-feed specifications. The update applied in the 2013 design case, and maintained here, improves on the 2011 design basis and the associated original feedstock price of \$58.50/dry ton (2007\$), which included a more uncertain grower payment, to a more reasonable target at an increased price of \$80/dry ton, including grower payment (2011\$). This new price is more appropriate for a large commodity scale going beyond a “niche market” price.

As has been described in prior design reports, the feedstock composition (see Table 1) plays a critical role in overall process design and economics. Of particular importance are carbohydrate components (cellulose and hemicellulose), as well as lignin, acetate, and ash, given the modifications being made to the pretreatment strategy such as the use of deacetylation as well as high sensitivity to impurity components such as ash and metals in the catalytic reactor section of this design. The blended uniform-format feedstock composition assumed here for purposes of future design case targets is shown below, with supporting details (in the context of corn stover compositional variability) described in the 2011 ethanol report [2]. Also consistent with prior design cases, the moisture content for the delivered feedstock is 20%.

When converting the analytical composition to components used in the Aspen Plus model, the nonstructural component fractions from the compositional analysis were combined under “extractives.” The extractives component is assumed to be primarily organic, with an average composition of CH₂O, and consists primarily of sugars, sugar alcohols, and organic acids, as well as some nonstructural inorganics [13]. The presence of extractives in the biomass depends on the time of harvest and, in part, to how much microbial degradation of the material occurs after harvest; the amount of extractives in a given sample may therefore be indicative of its age. Additionally, where the

mass balance did not sum to 100%, the extractives component was used to close it by difference. Sucrose is another extractive component, but it is measured separately in laboratory analysis and has been added as a separate feedstock component in the present design. The amount of sucrose present in the biomass is dependent on harvesting and handling practices. In pretreatment, this sucrose is assumed to be fully hydrolyzed to glucose and fructose. The labile fructose is further converted to degradation products in pretreatment, but the less reactive glucose resists degradation [14].

Table 1. Delivered Feedstock Composition Assumed in the Present Design

Component	Composition (dry wt %)
Glucan	35.1
Xylan	19.5
Lignin	15.8
Ash	4.9
Acetate ^a	1.8
Protein	3.1
Extractives	14.7
Arabinan	2.4
Galactan	1.4
Mannan	0.6
Sucrose	0.8
<i>Total structural carbohydrate</i>	<i>59.0</i>
<i>Total structural carbohydrate + sucrose</i>	<i>59.8</i>
<i>Moisture (bulk wt %)</i>	<i>20.0</i>

^a Represents acetyl groups present in the hemicellulose polymer; converted to acetic acid in pretreatment.

In the present design, the lignin fraction will also become essentially of equal economic importance as the carbohydrates because of the need to improve overall carbon efficiency to product(s) in order to ultimately achieve the 2022 \$3/GGE target. This is described further in Section 5.4.

2.3 Updated Modeling Basis

Similar to other biofuel pathways recently published in design reports under DOE-BETO efforts [8, 15], a number of key modeling assumptions are fixed constant to allow for comparable modeling outputs. These parameters include:

1. Facility start-up time: 0.5 year (6 months)
2. Cost-year dollar basis: 2011 dollars (all cost results presented here will be in 2011 dollars)
3. Facility on-stream time: 90% (7,884 hours/year); note, this is still intended to represent an n^{th} -plant facility, but allows for slightly more operational downtime in part to reflect the increased process complexity of the new pathway models considered for hydrocarbon production, relative to the prior 2011 ethanol basis of 96% on-stream time.

Additionally, the feedstock cost for the biochemical pathway is updated from a previous basis of \$58.50/dry ton (2007\$) as utilized in the 2011 design case to \$80/dry ton (2011\$) for the 2013 biological case and present catalytic design target cases, to reflect a more commercially relevant pricing model at a larger national scale (described in more detail in the 2013 report).

2.4 Design Report Conventions

2.4.1 Units

The Aspen Plus model described in this report is, by legacy, based on the set of units required by Aspen for specifying custom component properties: kg, kmol, atm, and °C for materials, and MMkcal (Gcal) for energy. Values in this report that were drawn directly from the Aspen model therefore tend to be reported in these units. Harris Group preferred to use U.S. standards (lb, BTU, °F, gal, etc.) when communicating with equipment vendors. Therefore, equipment specifications tend to be cited in these U.S. units.

Note that in the present report, certain quantities (e.g., yields and costs) are computed and reported in terms of “tons.” To avoid ambiguity, *tonne* will denote a metric tonne (1,000 kg) and *ton* will denote a short or U.S. ton (2,000 lb). In general, the U.S. ton is the standard for this document.

2.4.2 Total Solids Loading

The process described here converts a solid feedstock (corn stover blended with switchgrass) into liquid fuel products (diesel/RDB + naphtha). Most material streams in the process therefore have a solid fraction and a liquid fraction. The relative amount of solids in a given stream is called its “solids loading.” Total solids (TS) loading is defined as the total weight percent of soluble solids (e.g., sugars and salts) and insoluble solids (e.g., cellulose and lignin) in a given material stream. Where useful, the TS loading and the insoluble solids (IS) loading will be reported together. Note that in our convention, sulfuric acid, acetic acid, and ammonia are not considered soluble solids, but ammonium acetate and ammonium sulfate are. Therefore, around some unit operations, e.g., hydrolysate conditioning, TS loading is not a conserved quantity.

2.4.3 Product Yield and Heating Value

The results from this analysis are reported primarily in terms of energy yields in gallons- gasoline equivalent: \$/GGE, GGE/yr, GGE/ton, etc. This includes energy yield in both the diesel (RDB) and naphtha product cuts from hydrotreating distillation, rather than yield of a single product with the other treated as a coproduct; this is done to maintain focus on *total fuel (energy)* yield from hydrolysate carbon, to avoid introducing unnecessary subjectivity in over-interpreting the granularity of available public data regarding absolute yields to diesel- versus naphtha-range products from catalytic conversion/upgrading. Additionally, as modeled, the naphtha product is a non-trivial fraction of total product yield on a volume basis (approximately 20 vol % of total fuel produced), and as such is more appropriate to consider the summative energy-equivalent (GGE) yield for both products, as dramatic sensitivity to MFSP could be introduced when assigning naphtha a coproduct value; this is consistent with other recently-published TEA work [16]. Lower heating values (LHVs) for both diesel- and naphtha-range products were calculated by the Aspen model, and correspond to similar values for standard petroleum-equivalent products [17]. To translate to a GGE basis, a conventional gasoline heating value of 116,090 BTU/gal (LHV basis) was applied [17].

3 Process Design and Cost Estimation Details

3.1 Area 100: Feedstock Storage and Handling

3.1.1 Overview

Area 100 processes incoming biomass feedstock. Herbaceous biomass is the preferred feedstock for a biochemical conversion pathway focused on biomass deconstruction to monomeric sugars and other carbohydrates. The present design maintains all feedstock assumptions as consistent with NREL's 2013 biological hydrocarbon design report [8] and is summarized here. The design assumes a blend of corn stover and switchgrass delivered according to the specifications detailed in the INL design report for the uniform-format feedstock supply system [18]. In this envisioned design, biomass is stored in a central depot and is preprocessed, densified, stabilized, and homogenized to a degree before delivery, such that the biorefinery receives feedstock with known, uniform-format specifications including particle size, ash content, and moisture content. All analysis and cost implications associated with Area 100, and with feedstock logistics in general, are outside the scope of the present effort focusing on the conversion biorefinery, but are summarized here to convey information attributed to the updated feedstock cost of \$80/dry ton delivered to Area 200.

To address the high cost of biomass procurement, lower cost resources could be supplemented from other regions beyond the U.S. Midwest, but this strategy would increase transportation costs. Another strategy to reduce the cost of obtaining biomass is to “blend” or use numerous types of feedstocks within a region to design an aggregated or formulated feedstock. This strategy assumes that the formulation of multiple feedstocks will perform as well as or better than a singular feedstock. As illustrated in Figure 3, the advanced Terrestrial Feedstock Processing Supply and Logistics System is envisioned to draw in presently inaccessible and/or underused resources via local biomass preprocessing depots that format biomass into a stable, bulk, densified, and flowable material. The formatted biomass will be transported to one or more networks of much larger supply terminals, where the material aggregated from a number of depots may be blended and/or further preprocessed to meet the specification required by each biorefinery conversion process. The advanced biomass logistics system design incrementally incorporates technology and other system improvements as the industry matures. The introduction of advanced preprocessing operations and sequences, including the implementation of blending and formulation strategies, and preprocessing strategies, is critical to achieving feedstock cost, quality, and volume targets to meet BETO 2017 and 2022 goals. In supporting this effort, partners at NREL and INL are working to understand the impact that blended feedstocks have on operability and convertibility through the various conversion unit operations, to ultimately demonstrate that the targets stipulated in this design report are achievable for such blended feedstock formulations rather than merely a single corn stover feedstock.

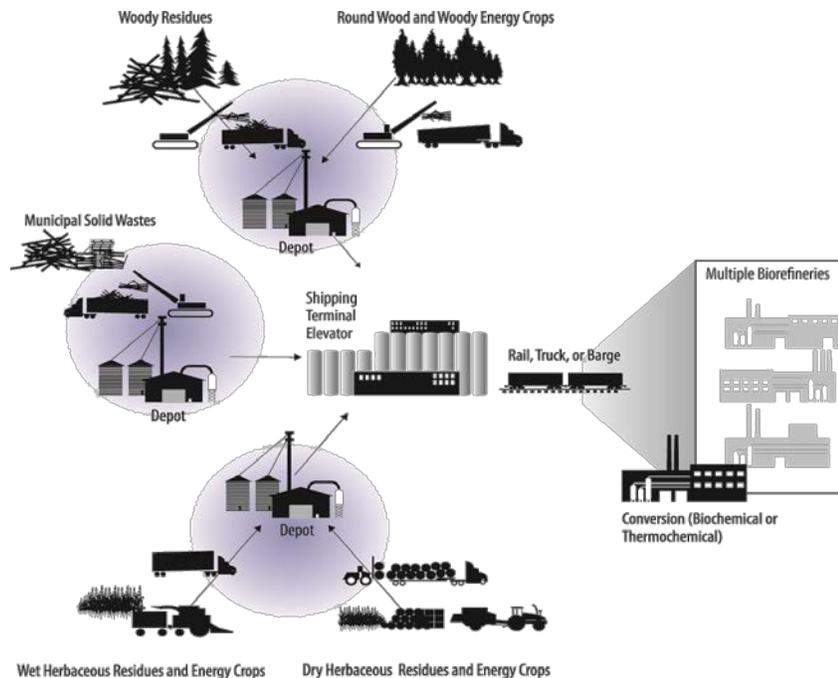


Figure 3. The depot concept behind the advanced Terrestrial Feedstock Processing Supply and Logistics System. Courtesy INL.

Blending strategies could also improve overall feedstock quality to help meet biofuel quality specifications and conversion in-feed requirements. Quality targets can have a large impact on whether or not a particular feedstock is cost effective. Inherent spatial and temporal variability of biomass resource quality translates to variability in biomass compositional attributes including ash, xylan, glucan, and lignin. Formulation allows the use of low-cost, but typically low-quality, biomass blended with higher cost and quality biomass to achieve minimum conversion process quality specifications. One particularly important consideration in the present analysis is that catalytic conversion processes are more sensitive than biological conversion processes to the amount, type, and variability of ash (part of which is dependent on different feedstock sources and types). It is known that metal ions, silica, and other species present in herbaceous biomass ash may poison, deactivate, or otherwise precipitate and clog catalyst systems and related equipment [19]. In the base analysis, to protect catalysts a fraction of the ash delivered to the biorefinery (see Table 1) is assumed to be removed up front during pretreatment deacetylation (Area 200), and the remainder removed during hydrolysate purification (Area 300). However, additional opportunities may exist to reduce ash content earlier on in the process through other logistics preprocessing mechanisms. This will be a continued subject of further analysis in collaboration with partners at INL.

3.1.2 Design Basis

In the new uniform-format feedstock supply system, feedstock would be stored in a satellite depot location with delivery to the biorefinery occurring 6 days per week by truck, or possibly by rail. At the depot, material would be milled to a mean size of 0.15–0.25 in. (with a high content of fines), and then densified into pellet format in order to achieve a mean bulk density of 30-40 lb/ft³ to maximize the biomass load per trailer. The incoming uniform-format feedstock (corn stover/switchgrass blend) is assumed to have 20% moisture when it reaches the biorefinery; this is representative of a mixture of field-dried material having less than 15% moisture and co-harvested material having greater than 20%

moisture. Because the preprocessing operation is designed to lose very little dry matter and does not include any rinsing of biomass, it is assumed that the biomass composition discussed in Section 2.2 is valid for the delivered material.

The as-received feedstock requirement for a plant is 2,750 U.S. ton/day (104,200 kg/h; 229,700 lb/h) including moisture. In the projected design, refinery receiving operates on the same schedule as the biomass depot: 24 hours per day, 6 days per week. Incoming trucks are weighed by electronic scale and unloaded using a whole-truck dumper capable of a 7–10 minute unloading time. The dumpers empty into dedicated hoppers that meter the biomass to a series of conveyors. The conveyors carry material from the truck tipper to short-term storage. The minimum receiving rate is 250 ton/h to maintain 110 ton/h of continuous processing. In order to process 250 ton/h, and assuming a relatively constant flow of trucks, a pair of scales (one inbound and one outbound) and two truck dumpers are required.

On-site storage is kept to a minimum of 72 hours to allow for a weekend buffer. Open piles are not favored due to concerns of fire, rodent infestation, and moisture degradation. Instead, the unloaded feedstock is stored in concrete domes. Two domes (each with a 36-hour capacity) are required so that one can be loaded while the other empties to the conversion process.

Conveyors connect the storage domes to the feedstock receiving bins on the pretreatment reactor in Area 200. A dust collection system integrated with the conveyors and domes handles airborne particles released during the unloading and conveying processes. No dry matter is assumed lost in Area 100.

3.1.3 Cost Estimation

Consistent with the 2013 biological hydrocarbon design report, the feedstock cost assumed in this report is \$80/dry ton (2011\$) attributed to the new uniform-format logistics system described above. This cost comes from the Multi-Year Program Plan (MYPP) published by DOE-BETO [7]. This is a rolled-up cost which includes both grower payment as well as logistics considerations for all collection, processing, storage, and transportation costs between the field and the receiving bin on the pretreatment reactor (e.g., Table B-2 of the referenced MYPP document). It should be stressed that these costs are future (2017–2022) DOE research targets, similarly to the conversion performance targets used in the Aspen Plus model.

3.2 Area 200: Pretreatment and Conditioning

3.2.1 Overview

Pretreatment of biomass for biofuel production is a crucial step. Its primary role is to disrupt the matrix of polymeric compounds that are physically and chemically bonded within lignocellulosic biomass cell wall structures, including cellulose microfibrils, lignin, and hemicellulose. Pretreatment has a significant impact not only on enzymatic hydrolysis, conversion, and downstream processing, but also on overall process economics and sustainability [20]. Acid pretreatments tend to hydrolyze significant fractions of hemicellulose and small fractions of cellulose.

Figure 4 shows a simplified flow diagram of the pretreatment area that uses a two-step process (deacetylation and dilute-acid pretreatment). Milled biomass (corn stover and switchgrass blend) is conveyed to the pretreatment area and treated first with dilute sodium hydroxide (for deacetylation), followed by sulfuric acid at moderate to high temperature (see Table 2) to liberate the hemicellulose

sugars and break down the biomass for enzymatic hydrolysis. Solids exiting the alkaline deacetylation step are sent on to dilute-acid pretreatment. Alkaline pretreatments break bonds between lignin and carbohydrates and disrupt lignin structure, making carbohydrates more accessible to enzymatic hydrolysis [20]. The dilute-acid pretreatment process converts most of the hemicellulose carbohydrates in the feedstock to soluble sugars (xylose, arabinose, galactose, and mannose) by hydrolysis reactions. This is followed by ammonia conditioning to raise the pH of the whole pretreated slurry from near 1 to 5, which is then sent to enzymatic hydrolysis.

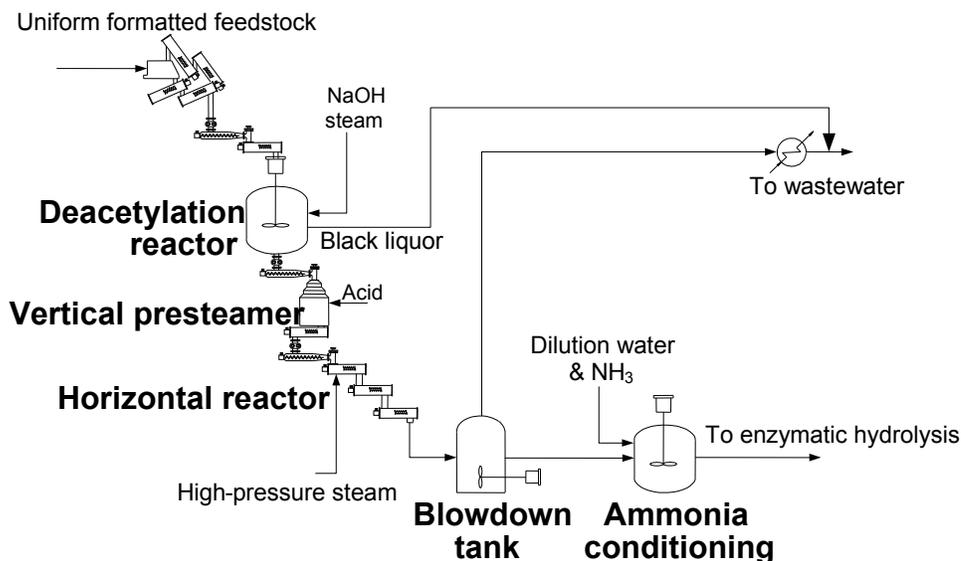


Figure 4. Simplified flow diagram of the deacetylation, pretreatment, and conditioning process

3.2.2 Design Basis

The model used in this report employs the same design and yield assumptions as the 2013 design case [8] and is summarized below. Deacetylation has been used to improve conversion of oligomeric xylan to monomeric xylose, with one study showing a decrease in oligomeric xylose yield from 21%–23% for non-deacetylated corn stover to 7%–10% for deacetylated corn stover [21]. Biological upgrading of sugars to fuel can benefit from the conversion of oligomers to monomeric sugar due to the metabolic limits of microorganisms; however, research suggests that catalytic conversion of biomass may also utilize oligomers equally to monomers and that they may even be desirable [19]. The current base case design is identical to the 2013 case for two reasons: (1) it provides consistency between design reports and (2) there is a lack of public data from which to build a detailed, redesigned pretreatment system that targets oligomer production. Additionally, maintaining the deacetylation step is useful in the present design as it also removes a significant fraction of ash up front, which is a problematic component that must be removed to protect conversion reactor catalysts. There are potential opportunities for cost improvements by targeting oligomers, including lower cost reactor metallurgy, fewer chemical inputs to pretreatment, and lower conditioning and purification costs as discussed further in Section 5.2.

The deacetylation step uses a sodium hydroxide soaking process at 80°C for 1 hour, with a NaOH loading of 17 mg/g dry biomass (see Table 2). The deacetylated material is dewatered by draining through screens at the bottom of the deacetylation reactor. A 316 stainless steel vertical pressure vessel

is used for the deacetylation tank. Three identical batch reactors are used with square trough live-bottoms consisting of four 24-inch horizontal discharge screws each. Each tank volume is 44,000 gal with a total cycle time of 2.4 hours. The drained liquor, often referred to as “black liquor,” contains 20%–25% of the original dry biomass constituent material, including water-soluble extractives (100% solubilization), soluble ash constituents (75% ash solubilization), 20% of the lignin, 2% of the xylan, 50% of the sucrose, and 88% of the acetate that was originally present in the feedstock (dry basis). This liquor stream is pumped to wastewater treatment (A600) in the baseline process design. The remaining biomass solids are discharged from the deacetylation reactor and transported to the acid pretreatment system. Moving forward, there may be opportunities to identify conditions which optimize deacetylation more towards ash removal than acetate removal, as acetate may be utilized as an additional carbon source in the downstream catalytic conversion operations in this case. Additionally, more room for further investigation exists regarding the implications that solubilized lignin components have on performance of the wastewater treatment organisms, which may become problematic at a certain threshold of soluble lignin removal.

The acid pretreatment reactor system was described in detail in the 2011 ethanol design report [2]. The system includes a feedstock receiving system, followed by a vertical vessel with a long residence time for steam heating and potential acid impregnation of the biomass, followed by the horizontal pretreatment reactor, which operates at a higher pressure and a short residence time. The 26-inch plug screw feeder is a rugged, high-compression screw device designed to form a pressure-tight plug of material through axial compression. Dilute sulfuric acid is metered at the discharge spool of each plug screw feeder. Feedstock drops from the plug screw discharge into a mixing and heating screw, which discharges the feedstock into the top of the presteamer. High-pressure steam is injected into this vessel to maintain temperature, while hot water is added at this point to control the pretreatment effluent at 30 wt % TS; at this concentration, the material is not easily pumpable and is primarily transported between unit operations within Area 200 using transfer conveyors.

The current model assumes operation of the presteamer at 100°C such that no significant hydrolysis reactions occur in this section. It can be used if additional hold up is required for acid hydrolysis. Feedstock flows downward through the vertical presteamer with uniform temperature throughout, discharging through a dual screw outlet device to two plug screw feeders. The plug screw feeders meter feedstock to the horizontal pretreatment reactor, with a horizontal reactor configuration chosen to allow tighter residence time distribution control than with a vertical reactor.

Upon deacetylation and solid-liquid separation by draining the black liquor, the actual sulfuric acid loading present in the horizontal acid pretreatment reactor is only 9 mg/g dry feedstock. The reactor pressure is held at the bubble point for the mixture. Heat losses from the reactor are not accounted for in the energy balance calculations. The residence time in the pretreatment reactor is nominally 5 minutes. The effective sulfuric acid concentration in the pretreatment reactor (estimated at 0.3–0.4 wt % after dilution by condensing steam is accounted for) may allow for the use of lower-cost metallurgies in the reaction zone (such as 904L or other duplex stainless alloys) instead of Incoloy-clad carbon steel, which could reduce pretreatment equipment costs. However, relevant corrosion data under these conditions are not available, so in the model, Incoloy–825 cladding is still conservatively assumed for the material of construction (consistent with the 2011 and 2013 design reports). The potential impact of lower-cost metallurgy and other pretreatment reactor cost savings is addressed in the Sensitivity Analysis section.

Additional acid is used to pre-impregnate the deacetylated biomass (the cost of this usage is included in the purchased sulfuric acid cost in the techno-economic model), but much of this acid is removed in pilot plant operation upon dewatering of the impregnated biomass and does not enter the pretreatment reactor. Acid pre-impregnation and its associated dewatering step are not explicitly modeled in our design, but it is envisioned that these could take place in the vertical presteamer, similar to the rationale in the 2011 design report [2]. The pretreatment reactor operating conditions are summarized in Table 2, and are maintained consistently with the 2013 design report targets.

Table 2. Deacetylation and Pretreatment Conditions Applied in this Design

Parameter (DA = deacetylation; PT = pretreatment)	Current design
DA Temperature	80 °C
DA Residence time	1 hour
DA Caustic (NaOH) loading	17 mg/g dry biomass
PT Total sulfuric acid loading present in pretreatment reactor	9 mg/g dry biomass
PT Residence time	5 minutes
PT Temperature	160°C
PT Pressure	5.5 atm
PT Total solids loading	30 wt %

Table 3 summarizes the reactions and conversions that take place in pretreatment. Glucan contained in the hemicellulose is converted to glucose along with a small portion of the cellulose. Minor hemicellulose carbohydrates (arabinan and galactan) are assumed to have the same reactions and conversions as xylan. The xylan-to-xylose conversion is a total hydrolysis value, which also may include an enzymatic component that will be discussed later. The sucrose reaction to HMF and glucose reflects 100% hydrolysis of sucrose to fructose and glucose, followed by complete degradation of the fructose to HMF.

Table 3. Pretreatment Hydrolysis Reactions and Assumed Conversions

Reaction	Reactant	% Converted to Product
$(\text{Glucan})_n + n \text{H}_2\text{O} \rightarrow n \text{Glucose}$	Glucan	9.9%
$(\text{Glucan})_n + n \text{H}_2\text{O} \rightarrow n \text{Glucose Oligomer}^a$	Glucan	0.3%
$(\text{Glucan})_n \rightarrow n \text{HMF} + 2n \text{H}_2\text{O}$	Glucan	0.3%
$\text{Sucrose} \rightarrow \text{HMF} + \text{Glucose} + 2 \text{H}_2\text{O}$	Sucrose	100%
$(\text{Xylan})_n + n \text{H}_2\text{O} \rightarrow n \text{Xylose}$	Xylan	90.0%
$(\text{Xylan})_n + m \text{H}_2\text{O} \rightarrow m \text{Xylose Oligomer}^a$	Xylan	2.4%
$(\text{Xylan})_n \rightarrow n \text{Furfural} + 2n \text{H}_2\text{O}$	Xylan	5.0%
$\text{Acetate} \rightarrow \text{Acetic Acid}$	Acetate	100%
$(\text{Lignin})_n \rightarrow n \text{Soluble Lignin}$	Lignin	5.0%

^a Sugar oligomers are considered soluble and convertible via downstream catalytic conversion.

Table 4 shows demonstrated performance results achieved during NREL’s 2012 pilot-scale demonstration runs, as well as the 2011 ethanol report assumptions and the present pathway conversion targets (i.e., those used in the current Aspen model) for the pretreatment area. The present model target is 90% conversion of xylan to monomeric xylose with 5% loss to degradation products.

Table 4. Research Status and 2017 Targets in the Pretreatment Area

	2011 Design Report [2]	2012 State of Technology [6]	Present Design Targets
Pretreatment			
Solids loading (wt %)	30%	30%	30%
Xylan conversion to xylose (%)	90%	82%	90%
Xylan conversion to furfural (%)	5%	6%	5%
Conditioning			
Ammonia loading (g/L of hydrolysate)	4.8	1.6	1.6
Hydrolysate solid-liquid separation	no	no	no
Xylose sugar loss (% entering conditioning)	1%	0%	0%
Glucose sugar loss (% entering conditioning)	0%	0%	0%

The pretreatment reactor is discharged to a flash tank. The pressure of the flash is controlled to keep the temperature at 100°C (212°F). The flash is condensed and routed to WWT (Area 600), containing more than 30% of the furfural along with other volatile organics from pretreatment. While this stream could potentially be routed to the conversion reactors (Area 500) as an additional source of carbon, it would first require a separation step to remove a large amount of water at added cost, as sending the full flash stream to the conversion section would greatly re-dilute the hydrolysate stream fed to conversion by more than doubling the total amount of water; thus, the model assumes it is instead routed to wastewater treatment. After the flash, the hydrolysate whole slurry containing 30% TS and 16% IS is sent to conditioning, namely neutralization by ammonia in stoichiometric quantities. Ammonia gas is mixed into dilution water to raise the hydrolysate pH to 5. The residence time for neutralization is 30 minutes and the dilution cools the slurry to 75°C (167°F). The slurry is diluted with water to slightly greater than 20 wt % TS to ensure miscibility through enzymatic hydrolysis and conversion. The composition of the stream at this point, and other major points throughout the process, is shown in the stream table information in Appendix E. The material from conditioning is conveyed to a saccharification storage tank.

3.2.3 Cost Estimation

Each deacetylation tank was quoted by Harris Group at \$890,000 for this design, including reactor tanks and conveyors. The acid pretreatment reactor design and cost basis was updated in the 2013 design report to allow for a more robust cost estimate with higher flexibility in operational variables. Namely, an evaluation of pretreatment reactor pricing in terms of metallurgy, reactor conditions, and reaction residence time was conducted by Harris Group based on input from a vendor. In addition to the base case Incoloy-clad metallurgy used as the baseline assumption, a less costly duplex stainless steel option was considered as an alternative. Further details behind the cost study may be found in the 2013 report, but in summary, the effort established correlation equations that allow for capital cost estimation of the pretreatment reactor system as a function of residence time for the two reactor metallurgy choices:

$$\text{Total Equipment Cost (2011\$ MM); Incoloy-825} = 16.4 + 7.4 \times m \quad (1)$$

$$\text{Total Equipment Cost (2011\$ MM); Duplex steel} = 16.8 + 3.9 \times m \quad (2)$$

Where, m = integer (minutes of resident time/3.3)

The above equations are based on an original design basis of 1,516 dry tonne/day biomass feed rate to the pretreatment operation (reflecting fractional biomass solubilization upstream during deacetylation from an initial feed rate of 2,000 dry tonne/day), and may be scaled to a different basis following standard scaling factor adjustments as discussed in Section 4.2. For the conditions used in this design, which are milder than earlier designs as employed in the 2011 ethanol report and other related work under the ethanol program, the recommended metallurgy from the vendor continues to be Incoloy-825. However, further opportunity exists to continue pursuing more mild pretreatment conditions, particularly for this catalytic conversion pathway, given the ability to utilize sugar oligomers rather than strictly monomers as noted above. This may translate to lower acid and caustic neutralization demands (lower operating costs as well as lower ammonium salt removal demands during downstream hydrolysate cleanup), as well as allowing for the use of the lower-cost metallurgy presented above. The total installed capital cost for the pretreatment system is approximately \$46.8 MM (2011\$) including feedstock receiving bin, pre-steaming, pressurized heating, reaction, and flash cooling, with another \$4.6 MM for the deacetylation equipment. The reactor schematic remains consistent with the configuration detailed in the 2011 and 2013 reports.

Area 200 contributes roughly \$0.44/GGE to the MFSP, including deacetylation, acid pretreatment, and conditioning. About 55% of this is attributed to capital cost, of which deacetylation and acid pretreatment equipment accounts for 96% of total capital expenses. Given the capital-intensive nature of the acid pretreatment step combined with the problematic ammonium sulfate salts introduced in the process which must be subsequently removed for downstream catalyst protection, an alternative pretreatment operation based on mechanical disc refining is considered later in the Sensitivity Analysis section. This operation eliminates the need for steam heating as well as acid and caustic chemicals, and may enable production of cleaner hydrolysate sugars; however, it also incurs a substantial power demand as well as potentially requiring the addition of hemicellulase enzymes, as it does not significantly hydrolyze hemicellulose to sugars like acid pretreatment does. This alternative will also be investigated in the future for biological conversion pathways. Further discussion on this option is presented in Section 5.2.

3.3 Area 300: Enzymatic Hydrolysis, Hydrolysate Conditioning, and Purification

3.3.1 Overview

In this process area, cellulose is converted to glucose using cellulase enzymes. This process is known as enzymatic saccharification or enzymatic hydrolysis. A cellulase enzyme preparation is a mixture of enzymes that work together to break down cellulose fibers into cellobiose and soluble gluco-oligomers and ultimately into glucose monomers. The resulting glucose and other sugars hydrolyzed from hemicellulose during pretreatment are then filtered to remove insoluble solids, concentrated, and purified by microfiltration and ion exchange.

The enzymatic hydrolysis portion of this process area follows the same fundamental process schematic described in the 2011 ethanol report and 2013 biological hydrocarbon report [2, 8], namely, carrying out hydrolysis at elevated temperature (48°C) so the reaction proceeds faster and requires lower enzyme loading, with hydrolysis split between two sequential reactions: a high-solids continuous flow reactor followed by batch hydrolysis in a stirred tank.

The hydrolysate is then treated to remove problematic soluble and insoluble material, and to reduce water content of the stream prior to catalytic upgrading steps. Figure 5 presents a schematic diagram of the purification and concentration steps used in this design, which were established based on recommendations from Harris Group and equipment vendors. First, insoluble lignin and other solids are removed from hydrolysate using vacuum belt filtration with a high wash ratio to minimize soluble sugar/carbon losses carried over into the solids fraction. The washed lignin-rich solids fraction is routed to the boiler for combustion. The hydrolysate is then concentrated in a mechanical vapor recompression evaporator under mild vacuum. A polishing filter (microfiltration) is used to remove any residual particulates carried over from the belt filter and solids that may have precipitated in the evaporator to protect downstream catalysts. Finally, the filtered and concentrated hydrolysate is treated by ion exchange to remove soluble cations and anions that may otherwise deactivate catalysts, before sending to catalytic upgrading.

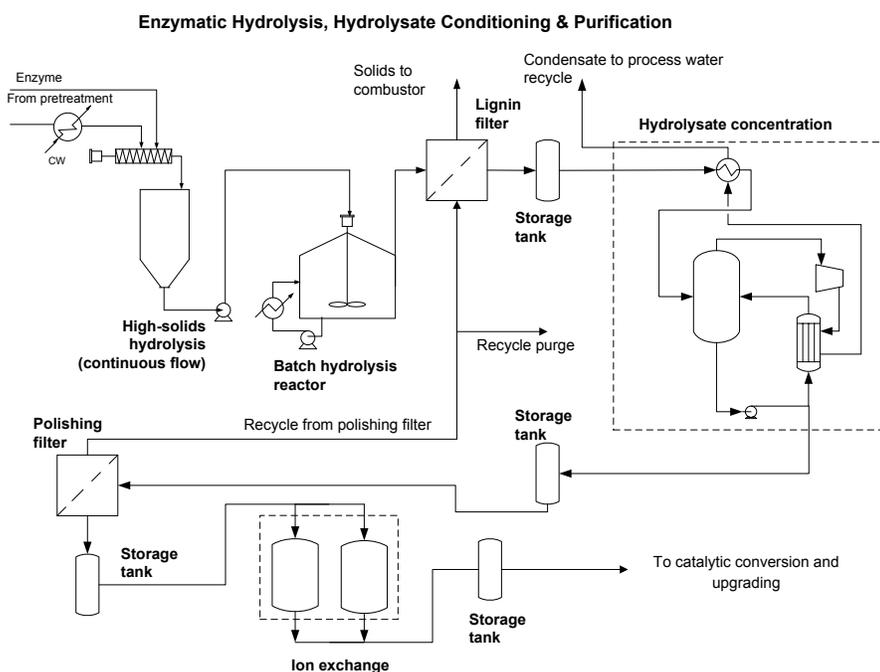


Figure 5. Simplified flow diagram of the hydrolysis, hydrolysate purification, and concentration unit operations utilized in Area 300

3.3.2 Design Basis

The process and design basis for enzymatic hydrolysis is the same as described in the 2011 ethanol report and 2013 biological hydrocarbon report, thus will only be briefly summarized here. In short, enzymatic hydrolysis is initiated in a continuous, high-solids vertical tower reactor with the slurry flowing down the reactor by gravity; this first step is required as the feed material at 20% solids (or more) is not pumpable until the cellulose has been partially hydrolyzed. After mixing in the cellulase enzyme, the total solids loading into the continuous column reactor is 20 wt% (12.2 wt% insoluble) and the temperature is 48°C (118°F). The residence time in the continuous reactor is 24 hours. After this point, the slurry is pumpable and is batched to one of six 1 MM gal vessels (950,000 gal working volume), where enzymatic hydrolysis continues for another 60 hours. The batch reactors are agitated and temperature controlled at 48°C using a pump-around loop with cooling water heat exchange.

The amount of enzyme used (the enzyme loading) is determined based on the amount of cellulose present in the hydrolysate and the specific activity of the enzyme. Consistent with the 2013 biological hydrocarbon design report [8], enzyme loading is targeted at 10 mg enzyme protein/g cellulose with a 90% conversion target of cellulose to glucose. This represents a roughly 50% reduction in enzyme loading and concomitant 12% increase in cellulose-to-glucose conversion relative to NREL’s demonstrated 2012 state of technology pilot-scale trials, whose best-case performance runs demonstrated 78%–82% cellulose to glucose conversion at 19 mg/g enzyme loading, or 89% cellulose to glucose at 26 mg/g loading [3]. Given ongoing advancements in enzyme preparations, discussions with enzyme manufacturers, and preliminary NREL data, utilizing new enzyme packages suggest the target loading of 10 mg/g is not an unreasonable projection within several years; however, an additional risk is posed in the new shift towards blended biomass feedstocks rather than a single feedstock (corn stover), which may prove more recalcitrant to enzymatic hydrolysis. In addition, new enzyme cocktails with increasing xylanase activity may become available to allow for a greater degree of xylan-to-xylose conversion during enzymatic hydrolysis, taking some of the burden off the pretreatment section to achieve extremely high levels of xylan deconstruction to xylose (potentially allowing milder pretreatments at reduced cost). Furthermore, this particular technology pathway does not require breaking carbohydrates down all the way to monomeric sugars, but also may catalytically convert soluble sugar oligomers; thus, potentially reducing cost burdens on the biomass deconstruction operations even further. The target design conditions for enzymatic hydrolysis are summarized in Table 5.

Table 5. Enzymatic Hydrolysis Conditions

Temperature	48°C (118°F)
Initial solids loading	20 wt % total solids (12.2% insoluble/7.8% soluble)
Residence time	3.5 days total (84 h)
Number and size of continuous vessels	8 @ 950 m ³ (250,000 gal) each
Number and size of batch vessels	6 @ 3,600 m ³ (950,000 gal) each
Cellulase loading	10 mg protein/g cellulose

The reactions and conversions taking place during enzymatic hydrolysis are listed in Table 6. As noted above, although the use of more advanced enzymes may achieve some xylan conversion during enzymatic hydrolysis, for modeling purposes, all xylan conversion is assumed to be achieved upstream during dilute-acid pretreatment.

Table 6. Enzymatic Hydrolysis Reactions and Assumed Conversions

Reaction	Reactant	% Converted to Product
$(\text{Glucan})_n \rightarrow n \text{ Glucose Oligomer}$	Glucan	4%
$(\text{Glucan})_n + \frac{1}{2}n \text{ H}_2\text{O} \rightarrow \frac{1}{2}n \text{ Cellobiose}$	Glucan	1.2%
$(\text{Glucan})_n + n \text{ H}_2\text{O} \rightarrow n \text{ Glucose}$	Glucan	90%
$\text{Cellobiose} + \text{H}_2\text{O} \rightarrow 2 \text{ Glucose}$	Cellobiose	100%

Once biomass deconstruction is complete, the hydrolysate is processed through a series of conditioning steps to purify and concentrate the sugars and other soluble carbon components prior to conversion. Some of these operations are consistent with those utilized in NREL’s 2013 biological design report (namely lignin filtration and hydrolysate concentration), but additional purification steps are added in this case, given more stringent purity requirements dictated by catalyst tolerance.

Lignin is the largest source of insoluble solid residue from the process, along with smaller amounts of insoluble material from feedstock deconstruction. Insoluble solids can plug catalytic reactors and are primarily separated from the liquid portion of the hydrolysate by horizontal vacuum belt filters. This operation includes a wash step to recover soluble sugars carried over into the solids fraction. The washed, lignin-rich solids are sent to the boiler for combustion. This step is similar to the process described in the 2013 design report [8] and the vacuum belt filter specifications are summarized in Table 7. The design of the belt filter was based on extrapolating from preliminary NREL data as well as input from Harris Group and vendors. The vacuum belt filter was chosen because it was shown to accommodate a larger solids processing capacity, thus requiring lower total filtration area, relative to pressure filtration. Further details on trade-offs between pressure filtration, vacuum filtration, and centrifugation options for solids separation are discussed in the 2013 design report.

As discussed in the 2013 report, the vacuum filter is expected to provide the most cost-efficient processing operation for hydrolysate solids separation, while maintaining high sugar recovery (up to 99% at a wash ratio of 2.5 L water/L of liquor remaining in the filter cake). Although this is a relatively high wash ratio and incurs an associated cost penalty, the stipulation for 99% sugar recovery is also fairly aggressive, and not yet well understood in the context of separating solids in post-saccharification slurries, which is more challenging than separating prior to enzymatic hydrolysis. The economic impact incurred by varying the assumption around sugar recovery is considered in the Sensitivity Analysis section. Alternatively, to achieve sufficient solids separation in this step after 90% hydrolysis of cellulose to glucose, additional filtration area could plausibly be required relative to the baseline assumption (depending on the ability of the cake to be filtered). The cost impact associated with varying the vacuum filter capital cost, attributed to varying the total filtration area, is also investigated in the Sensitivity Analysis section. Another constraint also imposed on the belt filter in this design that was not present in the 2013 process is the addition of a recycle stream from a new downstream microfiltration step (discussed below), consisting of particle fines as small as 0.1 microns. This could translate to more challenging operability of the belt filter, including more frequent removal of filter cake and potentially increased replacement maintenance costs for the belt itself. While this issue requires experimental validation to better understand, for modeling purposes this is addressed by assuming a 50% purge rate of the microfilter retentate recycle (discussed below), coupled with a sensitivity analysis considering the addition of periodic belt replacement costs beyond standard maintenance factors otherwise assumed (see Section 5.2).

Table 7. Vacuum Belt Filter Specifications for Hydrolysate Solids Separation

Number of stages	2
Wash ratio	2.5:1 (L water : L liquor in filter cake)
Capacity (insoluble solids basis)	20 kg/hr-m ² (IS basis)
Soluble sugar recovery	99.0%
Maximum unit size	170 m ²
Power demand	490 KW (660 hp) per unit

The clarified filtrate is then pumped from the belt filter to hydrolysate concentration, consisting of a four-stage mechanical vapor recompression (MVR) evaporation system, which produces a concentrated sugar stream of 46% (monomeric) sugar, 50% water, and 4% other dissolved solids. The evaporator design and cost basis is consistent with the 2013 design report [8]. MVR evaporators were considered to be preferable over standard thermal (i.e., steam-driven) evaporators given preliminary

heat and power balance information, where MVR evaporators are primarily electricity-driven rather than steam-driven. Four effects are required to concentrate the sugars from 14% to 46% (i.e., reduce water content to 50%). To avoid the possibility of sugar degradation at high temperatures [22], the evaporators are assumed to be operated under slight vacuum to keep the maximum temperature below 80°C (176°F). Even so, some fraction of monomeric sugars could revert back to disaccharides or higher monomers across the concentration step, but this is not anticipated to be a concern here as this pathway is assumed to utilize oligomers as an additional carbon source in the conversion step. The vapor exiting the evaporation system is nearly all water (approximately 99.9% purity) and can be recycled directly to the process water manifold after being condensed by water cooling. This has an added benefit of reducing the total water flow rate to the WWT section (A600), which reduces cost for WWT units that are dependent on the hydraulic flow rate. However, it also incurs a loss of convertible volatile carbon, primarily furfural, constituting the majority of the remainder of the vaporized stream. Recovery methods may be plausible to recapture the vaporized carbon species for subsequent conversion in the catalytic step, however the relatively low flow rate of such species in the vapor product (310 kg/hr) likely does not justify the added cost for such a system. The vendor-provided evaporation system design specifications are summarized in Table 8.

The monomeric sugar concentration of the filtrate stream from the horizontal belt filter is calculated by the model to be 14% by weight. The solids product stream from the horizontal vacuum belt filter has a total solids content of 36% and is conveyed to the boiler. A filtered hydrolysate storage tank provides 20 minutes of intermediate storage of filtrate from the belt filter. A concentrated hydrolysate storage tank, which provides a residence time of 20 minutes, is located downstream of the evaporator.

Table 8. Hydrolysate Evaporator Specifications

Feed sugar concentration	14 wt %
Product sugar concentration	46 wt % (50% water, 4% other solubles)
Maximum operating temperature	80°C
Evaporator technology	MVR
Number of effects	4
Electricity usage	3,600 KW (4,830 hp)
Steam usage (low-pressure steam)	1,550 kg/hr (0.8 MMkcal/hr)

Beyond the insoluble solids separation (lignin filtration) and hydrolysate concentration steps, which were also utilized in the 2013 design report and whose design and cost assumptions were derived from the 2013 basis, additional hydrolysate purification steps are utilized in this design, given more stringent specifications for the downstream catalytic operations. First, a polishing filter is included to protect the downstream catalytic conversion reactors from fouling by residual insoluble solids that were not removed in the vacuum belt filtration step. The polishing filter consists of four parallel crossflow microfiltration skids with a pore size of 0.1 microns and other specifications as listed in Table 9. This is a conservative specification since hydrotreating reactors, which are expected to operate similarly to the catalytic conversion reactors, typically have suspended particulate size limits between 1 and 70 microns [23, 24]. In this case, a considerably smaller particle size for microfilter design was selected to err on the conservative side, given the early level of understanding for the catalytic reactors required here. At an assumed maximum particle size of 50 microns in the feed to the microfiltration step (based on typical specifications in the upstream belt filter press operation as furnished by Harris Group), the vendor suggested that a moderate increase in the pore size of the polishing filter would have a minimal impact on the price of the system.

The microfilters were specified to allow for a soluble solids recovery rate of approximately 90% in the filtrate. To retain as much sugar as possible in the process, the retentate, which includes the remaining 10% of sugars and other convertible species, is recycled back to the vacuum belt filtration feed. However, as the retentate contains particle fines down to the 0.1 micron size, recycling this stream in its entirety could alter the particle size distribution on the belt filter and cause operability issues either by fouling the belt and requiring frequent replacement, requiring short cycle times, or fouling the microfilter units, due to a buildup of particle fines in the recycle loop. To mitigate this concern, 50% of the retentate recycle stream is removed as a purge and routed to wastewater treatment and the remaining 50% reaches the belt filter step for a subsequent pass. Thus, at a 99% targeted sugar recovery across the vacuum belt filters as discussed above, the overall net recovery of sugars and other soluble/convertible carbon species from the initial hydrolysate is still reasonably high at 93.7%, but lower than the 2013 design case reflecting the increased loss of soluble carbon attributed to more stringent purification demands. A polished hydrolysate storage tank is located downstream and provides a residence time of 20 minutes.

Table 9. Polishing Filter Specifications

<i>Parameter</i>	<i>Value</i>	<i>Units</i>
Total feed flow rate	111,930	kg/hr
Total feed flow rate	356	gpm (1,348 L/min)
Total solids concentration	50%	wt %
Soluble solids flow rate	55,585	kg/hr
Feed max particle size	50	micron
Feed total suspended solids (TSS)	706	ppm by wt.
Filtrate max particle size	0.1	micron
Filter type	Microfiltration	
Filter recovery rate of soluble solids	90	% wt.
Filter metallurgy	316L stainless steel	

Next, the polished filtrate is routed to ion exchange for further purification. This is an important step as certain soluble solids from process- or feedstock-related sources can poison catalysts and diminish their performance. The two expected sources of catalyst poisons are chemicals and enzymes used in biomass deconstruction, and components of the biomass feedstock itself. Sulfur from the acid used in pretreatment has been shown to poison metal-doped catalysts like those used in hydrogenation, Fisher-Tropsch synthesis, and methanation [25]. Nitrogen present in enzymes and in the feedstock can also pose a potential issue. Ammonia (e.g., ammonium salts produced during the pretreatment neutralization step) or organic bases like amines poison acidic catalysts like those used in this design. Other inorganic anions and cations can poison catalysts as well. For example; nickel, lead, vanadium, and zinc are poisons for hydrotreating catalysts [25], as are other metals more common to biomass, such as alkali metals. Additionally, ash components, such as silica, can accumulate and potentially plug fixed bed catalytic reactors or foul heat exchangers. Any halogen ions like chloride and fluoride can also corrode process equipment [19].

A separate-bed ion exchange system is utilized to remove a range of ionic species. Two resin bed trains are included—one for anions and one for cations, with pertinent specifications listed in Table 10. There are two units installed in parallel to allow for regeneration of the resin, which is expected to be required every 17 hours. Resin is regenerated with acid and caustic, whose estimated usage requirements are listed in Table 10. While specific components expected to be most problematic for

downstream catalysts and their concentration limits were considered proprietary and were not made available, it is expected that a separate-bed system utilizing both anionic and cationic resins will be sufficiently flexible to remove a wide variety of ionic impurities as required for downstream catalyst protection. A deionized sugar storage tank is located downstream and provides a residence time of 20 minutes.

Table 10. Ion Exchange System Specifications

<i>Parameter</i>	<i>Value</i>	<i>Units</i>
Total feed flow rate	100,766	kg/h
Soluble solids flow rate	50,127	kg/hr
Bed type	Separate-bed	
Cation	Strong acid cation	
Anion	Weak base anion	
Time between regenerations	17.4	hours
Caustic (NaOH) demand for regeneration	12.8	g/kg total soluble solids feed
Acid (HCl) demand for regeneration	22.4	g/kg total soluble solids feed

3.3.3 Cost Estimation

The design and cost basis assumptions for all enzymatic hydrolysis equipment were left unchanged from the 2011 and 2013 design cases [2, 8], namely empty towers for the continuous hydrolysis reactor based on a vendor quotation for flat-bottomed plug-flow reactors with a 10:1 height to diameter ratio, as well as 1 MM-gal batch hydrolysis reactors and agitators. The material of construction for enzymatic hydrolysis equipment is 304SS. Similar to the 2013 design case, six batch hydrolysis vessels are utilized in the present model versus eight vessels in the 2011 ethanol report, due to a reduction in total volumetric throughput into the hydrolysis section associated with the addition of up-front deacetylation to solubilize and remove a portion of non-convertible biomass components.

The vacuum belt filter design and cost information was provided by a vendor, who provided a design basis of 170 m² maximum filtration area per individual filter, with filter sizing based on a filtration rate of 20 kg/h m² based on insoluble solids. The cost includes a two-stage countercurrent water wash step at a wash ratio of 2.5 L/L liquor in the filter cake. A filtered hydrolysate storage tank with a residence time of 20 minutes is included to provide intermediate storage for filtrate from the belt filter. The sugar concentration (evaporation) equipment was also based on vendor-provided design and cost estimates. The system utilizes MVR evaporators in a series of four effects to concentrate the filtered hydrolysate from 14 wt % to 46 wt % (reducing water content to 50%). A concentrated sugar storage tank with a residence time of 20 minutes is included downstream of the evaporator. The vacuum belt filter and the hydrolysate concentration systems as well as associated storage tanks are consistent with the 2013 design report basis, with costs scaled here according to system throughputs in the present model.

The polishing filtration system includes four parallel skid microfiltration systems, including CIP and backwash. Installed equipment cost is \$3.4 MM total for the four skids. The ion exchange system uses a separate-bed configuration, with strong acid cation and weak base anion resins. Estimated installed equipment cost is \$10.0 MM including the resins. Capital costs for the key operations utilized for the hydrolysate purification and concentration steps are summarized in Table 11.

Table 11. Installed Equipment Costs for Purification and Concentration

<i>Equipment</i>	<i>Number of Units</i>	<i>Installed Cost</i>
Vacuum Belt Filter	5	\$18,331,000
MVR Evaporator System	1	\$16,778,000
Polishing Filter	4	\$3,361,000
Ion Exchange	1 ^a	\$10,025,000

^a Ion exchange “system” consists of parallel cation and anion resin beds, with parallel redundancy provided for resin regeneration.

Although a TEA assessment on the cost of sugar production (as the “final” product) is beyond the intended scope of this effort, an earlier analysis for drawing the product boundary after sugar production/concentration was presented in NREL’s 2011 ethanol design report and the reader is referred there for that information [2]. To summarize, the 2011 report evaluated a “minimum sugar selling price” associated with the front-end operations for lignocellulosic biomass deconstruction to sugars, followed by lignin/insoluble solids removal and optional sugar concentration. The resulting sugar selling price was estimated to be 11.6 cents/lb for dilute sugars (127 g/L total sugar concentration) or 14.3 cents/lb if including concentration via evaporation (487 g/L total sugar concentration); as noted in the ethanol design report, these prices were in 2007-year dollars. Updating to 2011 dollars, the same sugar selling price would be 12.7 cents/lb or 16.0 cents/lb for the dilute and concentrated cases respectively. These values provide some insight into what the intermediate sugar costs could look like in the present work, although the present design presented here contains a number of process deviations including (a) the addition of deacetylation, (b) revised pretreatment reactor costs now as a function of residence time, (c) additional sugar purification steps beyond merely lignin separation (namely microfiltration and ion exchange), and (d) differences in design basis assumptions including higher feedstock costs and lower targeted enzyme loadings.

3.4 Area 400: Cellulase Enzyme Production

3.4.1 Overview

This process area produces cellulase enzyme that is used in Area 300 to hydrolyze cellulose into glucose. Cellulase refers to a mixture of enzymes (catalytic proteins) that includes: (1) endoglucanases, which attack randomly within the cellulose fiber, reducing polymer chain length rapidly; (2) exoglucanases, which attack the ends of highly crystalline cellulose fibers; and (3) β -glucosidase, which hydrolyzes the small cellulose fragments (cellobiose, a glucose dimer) to glucose. Cellulase is produced industrially using (among other microorganisms) *T. reesei*, a filamentous fungus that secretes high levels of cellulase enzymes when grown aerobically in the presence of cellulose or other cellulase inducers.

The present analysis maintains consistency with the fundamental assumptions for enzyme production and cost estimation as detailed in the 2011 ethanol report and 2013 biological hydrocarbon report, most importantly, the use of on-site enzyme production rather than a purchased-enzyme model. As stipulated in prior reports, we again note that by including an on-site enzyme production section, NREL and DOE are not making a judgment about whether or not the cellulosic biofuel industry should align to this mode of enzyme distribution. Rather, the model on-site enzyme section is intended to improve transparency in determining the true cost of cellulase enzymes for large-scale cellulosic biofuel production.

Consistent with earlier design cases, the present design considers submerged aerobic cultivation (“aerobic fermentation”) of a *T. reesei*-like fungus on a feedstock of glucose and fresh water. Producing cellulase enzymes directly with glucose is not straightforward, because the absence of cellulose does not encourage the microorganism to secrete cellulase enzymes. Here, we have assumed a media preparation step where a small fraction of glucose is converted to sophorose, a powerful inducer of cellulase, using a small amount of the cellulase enzyme itself. When grown on this substrate, *T. reesei* has been shown to productively secrete cellulase [26]. Using glucose as the substrate is not necessarily more expensive than using hydrolysate slurry from pretreatment because the expected enzyme titer is significantly higher with glucose, reducing the capital and utility costs compared with using hydrolysate. The whole broth product is transferred to the hydrolysis tanks without an intermediate enzyme isolation step. Figure 6 is a simplified flow diagram of the enzyme production section.

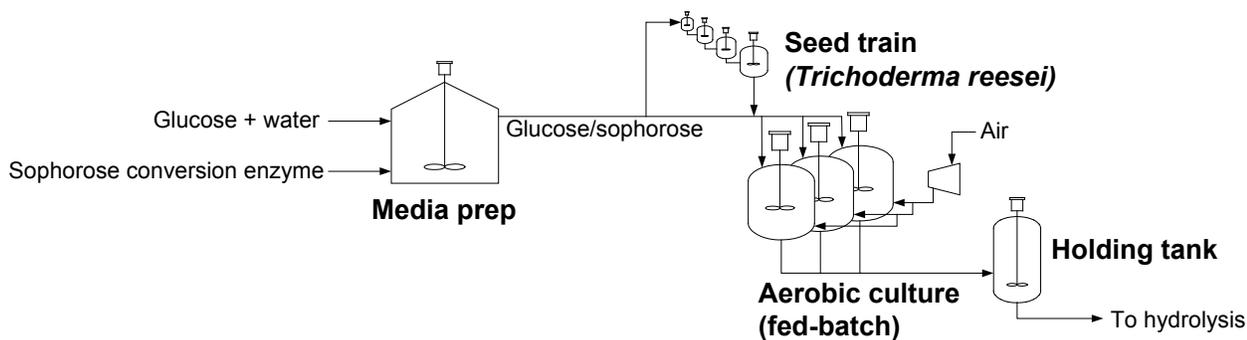


Figure 6. Simplified flow diagram of the enzyme production process

3.4.2 Design Basis

The design for the enzyme production operations is described in detail in the 2011 ethanol report and will not be repeated in such detail here. In summary, the design for the system was based in part on the claims in U.S. Patent 4,762,788 (Example 5) [27, 28] in combination with a number of reasonable assumptions to develop a rudimentary process. The key assumptions used in the current design are summarized in Table 12.

Table 12. Area 400 Guiding Design Basis Assumptions

Parameter	Assumption
Protein loading to enzymatic hydrolysis	10 mg protein/g cellulose
Reactor size	300,000 L @ 80% final working volume
Enzyme titer at harvest	50 g/L
Mass yield of enzyme from glucose	0.24 kg enzyme/kg glucose
Enzyme production cycle time	120 h online, 48 h offline, 168 h total

The assumed cellulase loading to enzymatic hydrolysis is reduced in the present pathway model to 10 mg of enzyme protein per g of cellulose as also assumed in the 2013 biological conversion report, down from 20 mg/g in the 2011 ethanol report. The rationale for this decision (keeping in mind this is a target for 2017–2022) is described above in Section 3.3.2. Here, “protein” refers to the total amount of high molecular weight protein in the enzyme broth as determined by assay; not all of this protein is active cellulase. The total protein demand was thus calculated to be 260 kg/h (570 lb/h). An additional 10% is produced to account for the slipstream provided to the media preparation tank to make the

glucose/sophorose mixture. The size of the cellulase production vessels was set at 300 m³ (80,000 gal) with a height-to-diameter ratio (H/D) of 2. Fermentation is assumed to be a fed-batch process starting at 50% working volume and ending at 80%. Over one week, each bioreactor will see a 24-hour cell growth period, a 96-hour protein production period, and a 48-hour offline period for draining, cleaning, and refilling. With a one-week total cycle time and the production parameters listed in Table 12, one bioreactor is capable of producing 12,000 kg of protein in a week, or 71.4 kg/h (157 lb/h). The equivalent enzyme volumetric productivity is 0.30 g protein/L-h. Five reactors were therefore required to deliver the 260 kg/h of protein needed for enzymatic hydrolysis.

The reactors are loaded initially with the glucose/sophorose carbon source and nutrients, including corn steep liquor (CSL), ammonia, and SO₂. After the initial cell growth period, additional substrate is added to maintain protein production. The bioreactors are sparged with compressed and cooled air and corn oil is added as antifoam. The reactors are temperature-controlled by chilled water flowing through internal coils.

Aeration and agitation requirements for the production bioreactors are functions of oxygen transfer rate and oxygen uptake rate (OTR and OUR), which have been discussed in detail in the 2011 ethanol report and again in the 2013 report. Thus, the details and governing principles for these metrics will not be repeated again here. In the modeled bioreactors, the reaction stoichiometry balances the reactions of substrate, oxygen, ammonia, and SO₂ to cell mass and enzyme (plus CO₂ and water) using an elemental composition for commercial cellulase provided by Novozymes [28]. The composition of cell mass was taken as the average of a generic cell mass composition [29] and the enzyme composition, with the assumption that cell mass includes some unreleased protein.



In the production bioreactors, it is assumed that 90% of the carbon source is converted via the protein reaction and 10% is converted via the cell mass reaction. In the seed reactors, 85% of the carbon is converted via the cell mass reaction and 5% via the protein reaction, with 10% unreacted. This represents an overall molar selectivity of glucose to 31% protein, 4% cell mass, and 65% CO₂, yielding 0.24 kg enzyme protein/kg glucose. The average aeration rate to the bioreactors remains the same as in the 2011 and 2013 models at 0.83 standard m³/s (including two days of offline time), as does the average impeller power *P* at 260 KW (350 hp), with a maximum of 600 KW (800 hp). The final specifications for the enzyme production reactors are shown in Table 13.

Table 13. Specifications of the Enzyme Production Bioreactors

Total volume	300 m ³ /80,000 gal
Maximum working volume	80%
H/D ratio	2
Height	11.5 m
Diameter	5.75 m
Operating pressure	1 atm
Operating temperature	28°C (82°F)
Material	316SS
Agitator	800 hp
Total electricity demand per kg protein (air compressors, agitators, chillers, pumps)	9 KWh/kg

Four trains of three seed fermentors provide inoculum to the main enzyme production bioreactors. Each vessel in the seed trains is run batchwise on the same substrate as the production vessels. Air is also sparged through each of the seed vessels, which are cooled with chilled water. The seed bioreactors are each sized at 10% of the next bioreactor volume, i.e., 0.3 m³, 3 m³, and 30 m³. The aeration demand is assumed to be 10% of the production aeration rate. Four trains were chosen because each production fermentor has a total cycle time of 7 days; each seed fermentor should have a cycle time of 2 days (including cleaning and sterilization) to get through the cell growth phase only.

Like the oxygen uptake rate, the glucose demand is also computed stoichiometrically from the required protein production rate. Ammonia and SO₂ are fed to the reactors stoichiometrically and CSL, trace nutrients, and antifoam (corn oil) are added to the substrate based on flow rate. The required nutrient concentrations are based on Schell et al. [30] and remain the same as presented in prior design reports. Glucose, the carbon source for cell mass and protein, is the most significant enzyme production expense in this model. The cost for glucose was maintained consistently with the 2011 ethanol design case (after adjusting to 2011\$), but may be revisited moving forward as it may err slightly on the conservative side. Electricity also remains a significant contributor due to the power requirements of air injection, agitation, and refrigeration. Total electricity demand for these operations is shown in Table 13.

3.4.3 Cost Estimation

The cost estimation for all equipment in A400 was left unchanged from the basis values provided in the 2011 ethanol report and maintained in the 2013 case. Most equipment in this area is stainless steel. The air compressor and some of the nutrient delivery equipment items are specified as carbon steel. Quotes for the production bioreactors, internal cooling coils, production agitators and motors, skid-mounted seed fermentors, and air compressor were provided by vendors through Harris Group, which developed costs for the pumps in this area using their historical database.

Not included in the enzyme production model are any costs for concentration, stabilization, or transportation of the enzyme to the plant. At the very least, one expects to have to pay licensing fees for the cellulase production microorganism, but these costs are not included because we lack information on what they might be. Additionally, while the design and cost assumptions for the enzyme production vessels themselves were left unchanged from those described in the 2011 ethanol report, the 300,000 L vessel size could potentially be on the large side for realistic commercial operation of a highly viscous system such as that for *T. reesei*. The implications for variations in assumed vessel size (e.g., capital cost) are considered in the Sensitivity Analysis section. The enzyme production system is also sized strictly to provide the amount of enzyme required for the hydrolysis step (e.g., is not oversized to accommodate occasional poor-yielding runs). A reasonable backup measure for this risk would be to either over-design the enzyme system or to add supplemental purchased enzyme capacity. These strategies are not considered here for consistency with the *n*th-plant approach taken elsewhere (e.g., avoiding over-design of equipment operations), but may be considered in future analyses exercising this design model.

Based on the economics of the on-site enzyme section described above, the predicted cost of enzymes to the facility is \$0.21/GGE of total fuel product. As enzyme loading (10 mg/g) and production design/cost assumptions were maintained identically with the 2013 biological conversion design case, the resulting enzyme cost on a per-kg basis is the same, e.g., \$5.38/kg protein, with consistent cost contribution breakdowns as presented in the 2013 report (and thus, will not be shown again here).

However, the enzyme cost allocation to overall fuel MFSP is lower at \$0.21/GGE in this design relative to the 2013 biological conversion case (\$0.37/GGE), given the higher fuel yields calculated in this case (discussed further in Section 4).

The caveat discussed in the 2011 and 2013 reports bears repeating, that the enzyme cost contribution modeled here is lower than one would expect for an enzyme preparation purchased from a separate, non-adjacent production facility. Transportation of the enzyme to the biorefinery facility could add a non-trivial amount to enzyme costs, even if formulation costs could be avoided. Furthermore, by lumping the enzyme production equipment in with the biorefinery, some key items are inherently shared, e.g., the land and buildings, cooling tower, and utilities infrastructure. Overhead and fixed costs, especially labor and management, would also be higher for a standalone facility. Eliminating the shared aspects between the enzyme unit and the biorefinery could easily add another \$0.05/GGE of product to the enzyme contribution. Additionally, an external enzyme production facility would probably demand a higher rate of return than the 10% IRR assumed for the biorefinery plant because it is a higher-risk and lower-volume business.

Including all these extra costs would likely bring the total enzyme cost contribution in line with numbers cited publicly (on the order of \$0.50/gal for ethanol under today's loading near 20 mg/g as demonstrated in NREL's 2012 State of Technology [SOT] work, which could translate to \$0.25–\$0.50/GGE of fuel products in this pathway for 10–20 mg/g loading, respectively). Such differences between off-site and on-site production make an argument for eventual integration and on-site production. However, in the near term it is more likely that cellulosic biorefineries (whether targeting sugars, ethanol, or hydrocarbons) will purchase enzyme from an external supplier with an organization dedicated to improving enzyme performance and reducing costs.

A400 systems and costs in this design were originally designed for performance specifications that are suitable for microbial conversion of biomass sugars. Most of those conversion processes are intended to use a microbe that primarily metabolizes monosaccharides. Therefore, there are opportunities to optimize biomass deconstruction steps, pretreatment, and enzymatic hydrolysis for catalytic conversion processes. One area for potential cost savings in the current model is that the catalytic conversion process may tolerate soluble oligomeric sugars [19]. This was not explicitly modeled due to a lack of publicly available data. However, costs could potentially be lowered by decreasing enzyme loading or by using lower performance/lower cost enzyme packages.

3.5 Area 500: Catalytic Conversion and Upgrading

3.5.1 Overview

The catalytic conversion process begins after the hydrolysate stream from pretreatment and enzymatic hydrolysis is processed through upstream filtration, concentration, and purification, leaving a relatively pure stream of water and convertible carbon species. The purified and concentrated hydrolysate is converted to fuel/blendstock products (primarily RDB with a lesser amount of naphtha) through four catalytic reactor stages. Figure 7 presents a high level schematic of the catalytic conversion and upgrading stages employed. The process begins with a set of two reforming reactors allocated into hydrogenation and aqueous phase reforming steps, followed by condensation and oligomerization, and a final hydrotreating step. The goal of these successive catalytic steps is to remove oxygen or “de-functionalize” carbohydrates and oligomerize them to primarily diesel-range hydrocarbons.

Each of the catalytic reactors shares a number of similarities. All of the reactors operate at elevated temperature and pressure, up to 350°C and 1,050 psig in the presence of added hydrogen, with reactions occurring in the liquid and gas phases [31]. Reactors in the base case model described in this section are supplied with purchased hydrogen from an external source to maximize carbon yield to fuels. *Ex situ* hydrogen is assumed to be produced by off-site steam-methane reforming (SMR) from natural gas. Hydrogen from SMR exhibits a large contribution to operating expenses as well as carrying sustainability implications, as will be shown later. Making use of externally supplied hydrogen, the catalytic conversion/upgrading process leads to very good overall carbon efficiency to fuels, roughly 85%, relative to starting carbon in the purified hydrolysate (discussed further in Section 5.1). However, the fossil CO₂ penalty attributed to natural gas-based SMR hydrogen production may detrimentally impact the overall life cycle of the process given the resulting amount of hydrogen import required. Information from patent literature has described alternative systems to produce hydrogen *in situ* from aqueous phase reforming of biomass hydrolysates [32, 33]. Alternatively, a fraction of the biomass may also be diverted and gasified to produce hydrogen from syngas [34]. Hydrogen via *in situ* or gasification production can improve the sustainability profile for the process relative to SMR-derived hydrogen, but at the cost of fuel yield. While external SMR-based hydrogen is considered for the base case, the *in situ* and gasification hydrogen alternatives are presented as important comparative cases in the Sensitivity Analysis section (see Section 5.2).

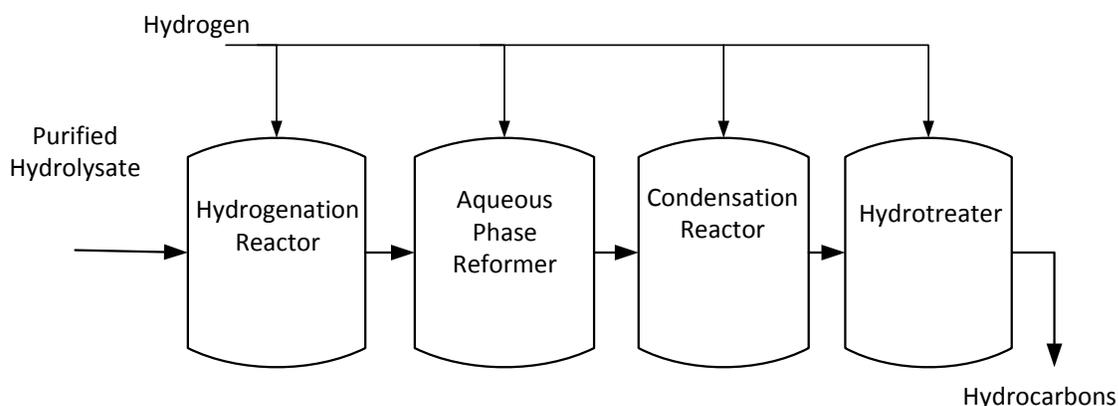


Figure 7. A simplified block diagram of overall catalytic conversion and upgrading process stages

3.5.2 Design Basis

Acid catalysis is a widely used approach for deconstructing biomass, as well as upgrading process intermediates. Section 3.2 described the well-known approach of dilute-acid pretreatment to solubilize hemicellulose. Dilute-acid pretreatment is one example of aqueous-phase, acid-catalyzed processing of biomass. The acid is in solution in pretreatment. In contrast, solid-phase acid catalysts are used in the upgrading process described here with the reactants in aqueous solution. Acid catalysts are attractive due to their ability to deoxygenate biomass-derived compounds like carbohydrates [35]. As an example, consider the acid-catalyzed conversion of glucose to HMF. Three oxygen atoms are removed from the carbohydrate in this reaction.



The acidity of solid catalysts can be much greater than mineral acids in solution and is dependent on factors such as catalyst composition, surface topology, and interaction with reaction solvents [36, 37]. Detailed understanding of the interaction between catalyst active sites, reactants, and water remains incomplete [35]. However, there have been a number of journal articles and patents published on the effectiveness of solid-phase acid catalyst for the aqueous-phase conversion of lignocellulose (primarily based on bench-scale experimental studies with clean commodity sugar substrates) which help to form the basis for the current design targets [31-33, 35, 38-51]. Aside from patent information, literature or other data in the public domain remains scarce on catalytic conversion of sugars to fuels, particularly in the context of cellulosic hydrolysate. As such, details and results presented in this analysis are limited to a similar level of detail as the underlying data presented in the cited patent literature, e.g., discussing yields based on modeled carbon conversions to component “classes” (alcohols, ketones, cyclic mono-oxygenates, etc.) as well as carbon number ranges and overall carbon efficiencies as described in patents, rather than rigorous kinetics or rate expressions as more comprehensive datasets might allow for.

As designed and costed by Harris Group based on patent information (details discussed below), the catalytic conversion process of hydrolysate to hydrocarbons consists of four stages (Figure 7): hydrogenation, aqueous phase reforming, condensation, and hydrotreating. Each stage consists of packed-bed pressure vessels, clad in 317L SS, and includes internals that support the catalyst and distribute the process fluid. Hydrogen is added to the reactors which operate at differing process conditions and catalyst compositions for each reaction step. The composition of products from these reactors is dependent on the composition of the feed stream, the reaction temperature and pressure, space velocity, and catalyst type. As modeled, element balance closure is maintained for all conversion reactors using stoichiometric reactions based on yields presented in literature, with model convergence based on the use of the NRTL property package given the presence of oxygenated/polar components. As discussed in A300, catalyst performance can be sensitive to a number of impurities that can poison the catalyst bed or reduce the performance of the reactors; primary examples include sulfur, cations, anions, solids, and coke [36]. Purification and filtration described in A300 is intended to minimize the amount of catalyst poisons and the selected operating conditions are assumed to minimize the formation of coke in order to maximize catalyst lifetimes and minimize frequency of catalyst regeneration.

This discussion begins with a focus on the first three catalytic conversion steps (hydrogenation through condensation), followed by the final hydrotreating operation. A schematic diagram depicting the envisioned design and stream routing for these first conversion steps is shown in Figure 8. It should be noted, however, that while this basis was used to assist in developing equipment costing by Harris Group, some details regarding heat exchange and stream routing were different in the Aspen model, and are discussed below.

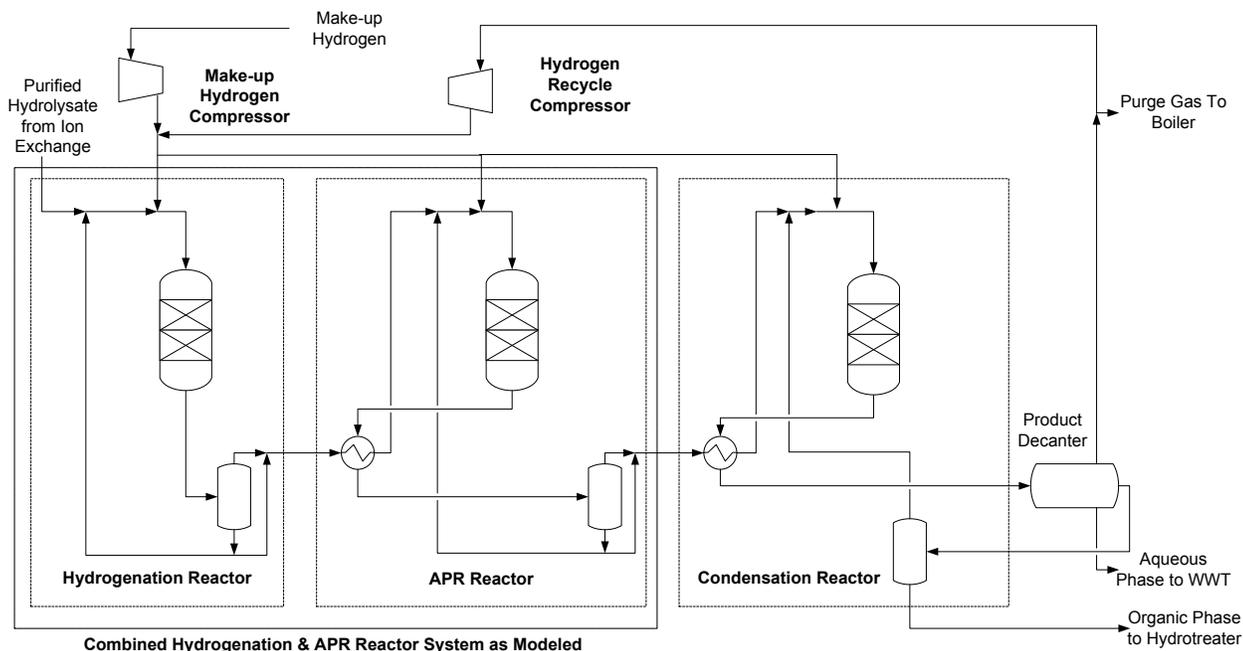


Figure 8: An overview of the three catalytic reactor stages used to upgrade purified biomass hydrolysate to longer chain oxygenates and hydrocarbons, as envisioned for commercial production and utilized for equipment design/costing purposes. (The intermediate products from this stage are sent to the hydrotreater for final conversion into deoxygenated diesel- and naphtha-range fuels.)

Hydrogenation and Aqueous Phase Reforming

In the hydrogenation step, the feed stream of purified, concentrated hydrolysate is mixed with fresh and recycled hydrogen from the process, and then heated to the inlet reactor operating temperature of 205°C [33]. Hydrogenation reduces sugars to sugar alcohols (e.g., sorbitol, xylitol) or other polyols (e.g., glycerol, ethylene, or propylene glycol) [31, 32]. Most biomass-derived sugars are aldoses, primarily glucose and xylose, although ketoses (e.g., fructose) may be found in some feedstocks. These monosaccharides undergo mutarotation to predominantly form pyranose and furanose ring structures in solution. These ring structures are less reactive than the open, linear form of sugars [52]. Hydrogenation converts sugars to their corresponding linear sugar alcohol form which has been shown to be more reactive during APR [31, 32, 53]. Catalyst formulations for the hydrogenation of sugars, organic acids, furans, and ketones often include rhenium and/or ruthenium on titania, carbon, or Raney nickel support and may include a variety of transition metal promoters [48, 54-59].

Following hydrogenation reactions, aqueous phase reforming is used to reform oxygenated hydrocarbons (i.e., polyhydric alcohol or other materials) in water to produce hydrogen, carbon dioxide, light alkanes, and oxygenates [32, 33, 41, 42, 44, 60, 61]. APR reactions tend to cleave C-C and C-O bonds. APR catalysts may be multi-functional where acidic supports catalyze dehydration reactions of oxygenated compounds and hydrogenation reactions are catalyzed by metallic sites incorporated into the catalysts [31, 32, 39, 40, 46, 49]. Examples of APR catalysts include: tungstated, titania, or sulfated zirconia; acidic or silica-alumina zeolites; or heteropolyacids alone or doped with metals including palladium (Pd), platinum, copper (Cu), cobalt, ruthenium (Ru), chromium, nickel, or silver.

One example of an APR reaction is the reforming of sorbitol. APR of sorbitol begins with the formation of an aldehyde [33]. The aldehyde reacts with water to form H₂, carbon monoxide (CO), and a smaller polyol which can react with hydrogen to form other compounds. The CO can react with water to form CO₂ and H₂ by the water-gas shift reaction [32]. On a high level, this general example is summarized by the following stoichiometry: $C_6H_{14}O_6 \rightarrow C_6H_{12}O_6 + H_2 + 6H_2O \rightarrow 6CO_2 + 13H_2$. The products from the APR reactor are a variety of oxygenates including alcohols, ketones, aldehydes, furans, diols, triols, and organic acids which may be upgraded by condensation reactions in the next reactor stage. Other organic compounds in biomass hydrolysates (e.g., organic acids, furans, and solubilized lignin) may follow similar reactions to cleaving of C-C or C-O bonds, hydrogenation, rearrangement, and dehydration [31]. Methanation and Fisher-Tropsch reactions also occur in APR to produce light hydrocarbons. Off-gases containing light hydrocarbons are separated downstream and are sent to the boiler by way of a purge split stream from the gas recycle line, which is useful for providing heat and power to the plant but incurs some loss of potential fuel-range carbon. The outlet stream from the APR is further heated and pumped to the condensation reactor.

The hydrogenation and APR reactors were designed and priced as separate units, both given their different functionalities (although different catalysts could also be utilized in different beds of a single reactor depending on catalyst and support types), as well as for better control over heat management in a commercial configuration. However, for modeling purposes these two steps were combined into a single set of reactions and yields. This choice was dictated by the limited degree of data availability in publications and patent literature from which the model was established. As data from patent literature was only available for a single reaction step taking sugars to partially deoxygenated APR products [33], the conditions and yields employed in the model were set accordingly and thus were not modeled independently with consideration of internal recycle flows or intermediate heat exchange. The hydrogen consumption of the two reactors was combined, as was the conversion of biomass sugars and other hydrolysate carbon compounds to oxygenates through the reactors. The hydrogenation reaction is a relatively mild step intended to chemically reduce sugars prior to reforming, thus could be viewed as a catalytic “preconditioning” step that could also take place in the first catalyst beds of a combined reactor.

Additionally, the heat integration as conducted in the Aspen model utilizes a feed-effluent exchanger to heat the hydrogenation/APR feed stream up to reaction temperature by cross-exchange with the condensation reactor effluent (rather than a feed-effluent exchanger only around the condensation reactor, as indicated in Figure 8). The APR effluent is then further heated to 262°C utilizing high-pressure steam generated from the steam system (A800), such that the condensation effluent temperature reaches the targeted 300°C [33]. The condensation effluent is then further cooled via air- and water-cooling to a temperature of 30°C to maximize fuel-range carbon recovery into the liquid phase and minimize losses into the off-gas. Thus, the general process concept is similar to the equipment costing/design basis presented in Figure 8, but with small differences in stream and heat exchanger placement to best match the available public data for use in establishing the model.

Table 14 lists the composition of the feed stream into the catalytic reactor block (exiting hydrolysate purification) starting with the hydrogenation/APR reactor system. All carbon-containing components shown in the table were assumed to be convertible to fuels via the catalytic steps employed [19], highlighting a key benefit of this pathway in the ability of the catalyst to utilize additional carbon rather than strictly monomeric sugars, including oligomers, degradation products (furfurals), and solubilized lignin produced from upstream deacetylation. Of course, this benefit comes with the trade-

off of increased costs for more stringent hydrolysate purification requirements to remove solids, ash, and ionic species than is necessary for biological conversion. As modeled, monomeric and oligomeric sugars were converted stoichiometrically to a number of representative components (CO₂ and oxygenates) discussed below. Other minor components, including soluble lignin, HMF, and furfural, were modeled using simplifying assumptions. In this case, minor components were assumed to be converted to a single representative component (e.g., HMF was converted to a 6-carbon oxygenate) for simplicity in maintaining carbon balance closures; while some loss of carbon from such minor components to side reactions and other products is likely, the overall impact on yields or costs would be marginal as these minor components represent 2% of all convertible carbon in the feed stream to the conversion section. The total modeled flow of hydrolysate feed is 99,620 kg/hr with 49% of that as convertible species.

Table 14. Mass Flow of Feed Stream Components into the Catalytic Reactor Section. (All carbon containing components listed below were assumed to be convertible across the catalytic reactor system and account for 49% of the total mass flow with the balance as water.)

<i>Component</i>	<i>Mass Flow (kg/h)</i>
Water	50,639
Total Convertible Components	48,981
Glucose	27,754
Galactose	1,106
Mannose	464
Xylose	15,289
Arabinose	1,882
Cellobiose	311
Gluco-oligomers	1,066
Galacto-oligomers	27
Manno-oligomers	11
Xylo-oligomers	359
Arabino-oligomers	45
Soluble lignin	492
5-Hydroxymethyl furfural	170
Furfural	6
Total Mass Flow	99,620

The hydrogenation and APR reactor operating conditions employed in the model are listed in Table 15. The operating temperature and pressure for the single reaction step as modeled was set at 254°C and 1,050 psig based on the most relevant set of conditions specified in patent literature [33]. Additionally, the total hydrogen rate in the gas stream (sum of fresh and recycle hydrogen streams) was set at a molar ratio of 7.9 mol H₂/mol reactor feed, based on the same patent data. As the fresh makeup hydrogen is assumed in the base case to be produced from a nearby off-site SMR plant, the makeup hydrogen was assumed to be delivered to the facility at a typical SMR product pressure of 18 atm (265 psia), and the makeup hydrogen compressor was designed accordingly based on estimates provided by Harris Group and a vendor. Weight hourly space velocity (WHSV) was used as the guiding basis for reactor design as employed by Harris Group. Although today’s “state of technology” experimental baseline for hydrogenation/APR was presented in the referenced patent literature as 0.7 hr⁻¹ [33], the target case for this value was increased to 1.2 hr⁻¹ for the hydrogenation reactor and 1.0 hr⁻¹ for the APR reactor, based on guidance from Harris Group, the vendor, and communication with Harris Group’s industrial contacts. This value translates to a total reactor catalyst bed volume of 2,325 ft³ for the hydrogenation step and 1,441 ft³ for the APR step. As a general note, WHSV values are

targeted to increase by roughly 40-70% relative to benchmarks demonstrated in the above-cited patent literature (from 0.7 to 1.0-1.2 hr⁻¹), coupled with targeted catalyst lifetimes of 1-2 years across all three conversion reactor steps (see Table 20); while these represent considerable R&D improvements over currently demonstrated performance which has primarily been on clean commodity sugar substrates, they are in line with typical values demonstrated in the petroleum industry. Additionally, WHSV alone was not seen to exhibit strong impact on overall MFSP, as demonstrated by the sensitivity analysis below in Figure 15.

Table 15. Hydrogenation and APR Reaction Conditions

<i>Overall Modeling Parameters</i>	<i>Operating Condition</i>	<i>Units</i>
Total Feed Stream Flow Rate	99,620	kg/h (modeled)
Convertible Species in Feed Stream	49%	(modeled)
Hydrogen Feed Molar Ratio	7.9	mol hydrogen/ mol convertible feed
Operating Temperature ^a	205-254 (inlet-outlet)	°C
Operating Pressure ^a	1,050	psig
Hydrogenation Reactor Parameters ^b		
WHSV	1.2	hr ⁻¹
Total Catalyst Bed Volume	2,325	ft ³
Catalyst Type	Precious metal on carbon support	
APR Reactor Parameters ^b		
WHSV	1.0	hr ⁻¹
Total Catalyst Bed Volume	1,441	ft ³
Catalyst Type	Base metal on mixed oxide support	

^a Process conditions for single hydrogenation/APR reactor as modeled

^b Design specifications for separate hydrogenation/APR reactors as designed and costed

As noted above, the combined yields through the modeled hydrogenation and APR reactors were set based on patent literature data [33]. For this step of the process, the model made use of stoichiometric reactions to target representative components within each class of molecules, primarily less than C₇ carbon chain length. The modeled yields from the hydrogenation/APR reaction steps are summarized in Table 16 alongside the original patent data guidance, and show close agreement with targeted yields. In some instances yields fluctuated slightly due to the stoichiometric reactions selected as well as the composition of the recycle gas stream co-fed to the hydrogenation/APR reactors along with the hydrolysate material.

Table 16. Product Carbon Yields from the Hydrogenation and APR Reactors. (Data from recent patent literature and the current model results are provided for comparison [33].)

<i>Component</i>	<i>Patent Data (% of feed C)^a</i>	<i>Model Results (% of feed C)</i>
Carbon Dioxide	2	2
Alkanes	2	2
Total Mono-oxygenates	62	64
<i>Alcohols</i>	15	14
<i>Ketones</i>	16	17
<i>Cyclic Ethers</i>	17	18
<i>Cyclic Mono-oxygenates</i>	15	14
<i>Organic Acids</i>	2	2
Di-oxygenates	9	9
Poly-oxygenates	3	3
Unknown Aqueous	17	20

^aBlommel, P., et al., US20120198760 A1, Method and systems for making distillate fuels from biomass. 2012, Virent Inc. Madison, WI: United States. Table 9.

Condensation

In the condensation processing step, the intermediates from the previous APR step are reacted to form longer continuous carbon chains through C-C bond forming reactions. Several types of reactions occur in this single catalytic step in the presence of water, including dehydration, oligomerization, cyclization, aromatization, and hydrogenation. The average carbon chain length increases from less than 6 in the feed to include a substantial amount of molecules within the C₈–C₂₄ range most suitable for diesel fuel. Products of this step include normal and iso-paraffins, olefins, ketones, aromatics, and cycloparaffins [32, 33].

The APR product stream is heated and fed into the condensation reactor. APR, as designed in the current model, tends to produce oxygenates with shorter carbon chain lengths [33], primarily less than C₇ given similar carbon chain lengths of the starting carbohydrate and other (minor) convertible components. Aldol condensation is a useful route to couple light oxygenates to form higher molecular weight compounds, as is the focus here for a diesel-predominant production process (e.g., RDB targeting carbon chain lengths in the C₁₀–C₂₄ range). Aldol condensation can be catalyzed by solid acid or solid base catalysts. However, basic catalysts can be poisoned by organic acids in the biomass hydrolysate feed stream. Acid catalysts are not affected by organic acids and have an advantage when processing biomass hydrolysates [35, 62]. Examples of acidic condensation catalysts include zirconia, alumina, silica, aluminasilicates, aluminum phosphates, silica-alumina phosphates alone or doped with metals including Ga, Ru, Cu, or Pd [32, 35].

In the model the hydrogenation/APR product stream is first further heated to 262°C, calculated by the Aspen model to correspond to a target condensation reactor product temperature of 300°C [33]. This duty is supplied by high-pressure steam generated in the steam section (A800) and removed from the steam loop prior to being sent through the first turbine stage. The output from the condensation reactor is cooled to 141°C by cross-exchange with the hydrogenation/APR feed stream, then further cooled by air- and water-cooling to a final temperature of 30°C to recover light naphtha-range components, and then flashed in a three-phase separator. The flash stream containing hydrogen, CO₂, and light hydrocarbons is routed to the recycle gas compressor for recycle through the system, with a purge slipstream removed to mitigate buildup of produced CO₂. The purge split is set to achieve a minimum 90 mol % hydrogen purity after combining with the fresh hydrogen makeup stream (which is sent through a separate multi-stage makeup compressor prior to combining with the recycle gas and

feeding to the hydrogenation/APR feed stream at 1,050 psig). The purge stream is routed to the boiler. The aqueous phase contains a fraction of miscible or partially miscible carbon components along with water from the hydrolysate feed, as well as additional produced water as a product of oxygen rejection, and is sent to wastewater treatment. The organic phase product stream contains mostly C₆₊ hydrocarbons and mono-oxygenates which are further processed by hydrotreating.

The condensation reactor operating conditions are listed in Table 17. The operating temperature and pressures were set based on the same above-referenced patent literature source as utilized for the hydrogenation/APR step [33], namely 300°C and 900 psig. While the majority (80%) of the combined fresh/recycle gas hydrogen stream is fed to the hydrogenation/APR step, 20% of the stream is split and routed to the condensation step following the hydrogen molar feed ratio stipulated in the same patent experimental basis of 1.9 mol H₂/mol feed to condensation (e.g., product stream exiting APR). Additional hydrogen is present in the APR product stream, thus the total hydrogen molar ratio is higher than this basis. Similar to the WHSV values discussed above for the hydrogenation/APR steps, the WHSV for the condensation step was targeted to increase from a “state of technology” experimental baseline of 0.7 hr⁻¹ to a projected value of 1.2 hr⁻¹ based on guidance from Harris Group, the vendor, and Harris Group’s industry contacts. This translates to a total reactor catalyst bed volume of 2,059 ft³.

Table 17. Condensation Reactor Conditions

<i>Condensation Parameters</i>	<i>Operating Condition</i>	<i>Units</i>
Hydrogen Molar Ratio	1.9	mol hydrogen/ mol condensation reactor feed
Operating Temperature	262-300 (inlet-outlet)	°C
Operating Pressure	900	psig
WHSV	1.2	hr ⁻¹
Total Catalyst Bed Volume	2,059	ft ³
Catalyst Type	Precious metal on mixed oxide support	

Similar to the hydrogenation/APR reaction, the condensation reaction was modeled based on stoichiometric equations to target yields presented in literature. Table 18 lists the relative yield split of intermediate products from the condensation reactor. The modeled results are contrasted with data from the same recent patent literature as above [33]. The products are listed in broad groups by carbon number, and consist largely of mono-oxygenates and deoxygenated hydrocarbons. The modeled yields from condensation correspond closely with targets in patent literature information [33], with remaining carbon closure allocated to CO₂ (4% of convertible feed carbon in the hydrolysate sent to hydrogenation/APR). The total modeled flow rate of the condensation reactor product stream is 349,850 kg/hr. Of this rate, a large fraction (71%) is flashed into the gas phase with the majority recycled back to the hydrogenation/APR inlet, and roughly 1% split off as a purge stream. Such a large recycle flow translates to considerable power demand for the recycle gas compressor to accommodate the 150 psi (10.2 atm) pressure drop across the recycle loop, at nearly twice the power consumption of the makeup compressor (more details discussed in Section 3.9). The purge stream flow rate is 2,630 kg/hr, and consists primarily of CO₂ (roughly 53 wt %), hydrogen (20%), off-gases (2%), and light naphtha-range components (25%) whose loss detracts from ultimate fuel yields, but are unavoidable unless more complex and costly recovery schemes are employed. The aqueous product fraction comprises roughly 22% of the total condensation product stream, and contains slightly more than 1% carbon species with marginal to high water miscibility including light (carbon numbers 1–5) organic

acids, alcohols, ketones, cyclic ethers, di- and poly-oxygenates, and other aqueous-partitioning components (e.g., the “unknown aqueous” fraction as listed in Table 16 exiting the condensation reactor). Finally, roughly 6.5% or 22,874 kg/hr of the total condensation product stream is recovered as the organic liquid product phase and is routed to hydrotreating.

Table 18. Overall Product Carbon Yields from the Condensation Reactor. (Data from recent patent literature and the current model results are provided for comparison [33].)

Products by Carbon Number	Patent Data (% of feed C to hydrogenation/APR) ^a	Model Results (% of feed C to hydrogenation/APR)
C ₁ -C ₇	23%	22%
C ₈ -C ₁₄	50%	50%
C ₁₅ -C ₂₄	23%	23%
C ₂₄ +	1%	1%

^aBlommel, P., et al., US20120198760 A1, Method and systems for making distillate fuels from biomass. 2012, Virent Inc. Madison, WI: United States. Figure 6.

Hydrotreating

The organic liquid fraction exiting the condensation reactor system is routed to hydrotreating as a final catalytic upgrading step to complete the deoxygenation to fuel-range hydrocarbons. The design assumptions and operating conditions for this step were set to be consistent with the hydrotreating section described in NREL’s 2013 design report, given their origins based on vendor-supplied information for a similar concept focused on deoxygenation to primarily diesel-range fuel/blend stock products [8]. While it is recognized that a number of compositional details differ for the feed material to hydrotreating¹, the associated design and cost information provided for the hydrotreating facility in the 2013 design report [8] is still viewed as the most credible basis available to continue utilizing in this case, and is still largely appropriate for use here with minor adjustments discussed below. While full details for the hydrotreating facility are provided in the associated 2013 design report, the process and key differences will be summarized here.

For the 2013 fatty acid upgrading design, Harris Group consulted with a number of vendors, including catalyst suppliers and process licensors, to obtain cost estimates for a hydrotreating facility that may be expected to process a moderately oxygenated (10–15 wt % oxygen content) feed into deoxygenated diesel-range components. While a number of cost estimates were provided from vendors for such a facility, associated operational specifics were not available without confidentiality restrictions, thus operating conditions such as temperature, pressure, and hydrogen partial pressure were based on Harris Group estimates combined with literature values. The primary literature basis utilized was a report jointly authored by UOP, NREL, Pacific Northwest National Laboratory (PNNL), and Michigan Technological University [63]. This report described relatively mild hydrotreating temperature and pressure conditions at 350°C and 500 psig, but a high hydrogen feed ratio at 6,000 standard cubic feet/barrel of feed (this translates to a hydrogen partial pressure of approximately 430 psig at reactor inlet), where high hydrogen feed is necessary to control the heat generation (exotherm) incurred for deoxygenation. A reasonable liquid hourly space velocity (LHSV) associated with the original fatty

¹ The Davis et al. 2013 design report was based on hydrotreating free fatty acid material with a high TAN number and potentially higher oxygen content, versus a cleaner feed with lower impurity levels that may look more similar to standard petroleum hydrotreater feedstocks in the present design.

relatively low oxygen content of the feed stream attributed to the mixture of hydrocarbons and mono-oxygenates noted above, the amount of oxygen removal and associated hydrogen demand is still relatively low, even for the 100% HDO assumption. There would be some sustainability trade-offs between higher hydrogen consumption (HDO pathway rejecting oxygen as H₂O) versus lower carbon yields to products (decarboxylation/decarbonylation pathway rejecting oxygen as CO₂ and CO), but such trade-offs are expected to be minor given the low feed oxygen content as modeled for this case. The liquid product is fractionated in a distillation column (including heat and cooling demand considerations in overall heat balances), yielding cuts of diesel- and naphtha-range products, as well as a light overhead gas stream routed to the boiler. The relative carbon yields to both the diesel and naphtha streams are reported in Table 19; however as discussed previously, given the lack of information specifics in patent literature and to avoid unnecessary subjectivity in interpreting this information in a way that could introduce additional uncertainty into the economic results, the primary focus for reported yields is on overall total fuel yields, i.e. both diesel and naphtha, translated to energy or carbon outputs. This said, the analysis does include considerations for capital and energy (heating/cooling) costs to fractionate the product into these cuts. The primary diesel product cut exiting the distillation column was also seen to exhibit a boiling point curve that closely resembles information presented publicly for this technology pathway, also resembling standard petroleum diesel [64]. The operating assumptions discussed above for the hydrotreating section, as well as modeled yields and overall hydrogen consumed for all catalytic conversion steps (starting from the APR feed), are summarized below in Table 19.

Table 19. Summary of Hydrotreating Process Assumptions and Modeled Yields for Overall Process

Hydrotreater feed rate	191,610 gal/day (725 m ³ /day)
Average reaction temperature	660 °F (350 °C)
Operating pressure	525 psig (37 atm) at outlet
Hydrogen partial pressure	430 psig (30 atm)/6,000 SCF/bbl
Reactor LHSV	~1.2 hr ⁻¹ (estimated) ^a
Hydrogen loss across purge split	5%
Overall carbon yield to fuel-range products from hydrotreater vs. APR feed components	85.6% (172,637 GGE/day)
<i>C yield to RDB vs. APR feed components</i>	69.4% (139,962 GGE/day)
<i>C yield to naphtha vs. APR feed components</i>	16.2% (32,675 GGE/day)
Total H ₂ consumption across all catalytic conversion steps, wt % of APR feed components	6.5% (reaction only; 7.9% total H ₂ makeup) ^b
Total makeup hydrogen rate	38.2 MM SCFD (1.0 MM Nm ³ /day)

^a Reactor LHSV is not used in the present model. (Total hydrotreater facility cost was used based on vendor feedback, which implicitly includes hydrotreater sizing which was not provided by vendor.)

^b Value is for consumption across reaction steps only (sum of hydrogenation/APR, condensation, and hydrotreating), not including additional losses due to purge streams, etc. (increases to 7.9% of APR feed components for total H₂ makeup rate).

3.5.3 Cost Estimation

Harris Group developed capital and operating cost estimates for the unit operations utilized in the catalytic reactor scheme as presented in Figure 8, i.e. the hydrogenation, APR, and condensation reactor steps, as well as supporting equipment including compressors and heat exchangers. Capital costs for each new piece of major equipment are based on budgetary pricing from vendors, adjusted to 2011 dollars. Each catalytic reactor in this section (excluding the hydrotreating section) is clad with 3/16 inch 317SS for the material of construction. In general, higher combinations of temperature and

hydrogen partial pressure require higher metallurgy. In the base process design, the reactors are clad with 317L stainless steel for corrosion protection.

Table 20 provides a summary of the catalytic conversion reactor designs. Similar to the constraints imposed on inclusion of reactor details around the hydrotreating operation as presented in the 2013 design report and maintained in this case (which vendors typically consider to be proprietary), limited details were made available to NREL attributed to the underlying cost estimates for the hydrogenation, APR, and condensation reactors. Summarizing on a high level, overall reactor sizing and associated catalyst demands are dictated by reactor WHSV values for each of the three steps. As discussed above, based on guidance from Harris Group as well as expertise provided to Harris Group by the vendor and other industry consultants, reactor WHSV values were targeted to increase from the experimental baseline of 0.7 hr^{-1} provided in patent literature [33] to aspirational projections of $1\text{--}1.2 \text{ hr}^{-1}$ for the three reactor steps. This translates to total reactor catalyst bed volumes of $2,325 \text{ ft}^3$, $1,441 \text{ ft}^3$, and $2,059 \text{ ft}^3$ for the hydrogenation, APR, and condensation steps respectively, based on catalyst loading densities as shown in Table 20. Based on a 33% allowance for reactor internals per the design furnished by Harris Group, this equates to a total (internal) reactor volume of $3,470 \text{ ft}^3$, $2,151 \text{ ft}^3$, and $3,073 \text{ ft}^3$ for the respective steps. The resulting installed capital cost estimates for the three steps are approximately \$18.3 MM, \$4.7 MM, and \$6.1 MM (\$29.1 MM in 2011 dollars for the total conversion system reactors), which includes a 6% allowance factor for operations which were not costed explicitly by Harris Group, namely pumps and separation vessels. Additionally, Harris Group (with guidance from vendors) furnished cost estimates for the key heat exchangers employed in the model, including the feed-effluent exchanger, condensation feed heater, and condensation product coolers; as well as both the makeup and recycle hydrogen compressors, based on modeled stream conditions. These items translate to a total installed cost of \$23.7 MM.

In addition to reactor design and capital cost estimates for the three conversion steps prior to hydrotreating, Harris Group also furnished estimates for catalyst cost, lifetime, and time between regenerations, the net sum of which translates to amortized annual catalyst operating expenses for a given overall catalyst demand dictated by reactor WHSV targets discussed above. Again, similar to limitations on reactor design details provided by the vendor, specific catalyst composition details were considered proprietary and were not provided to NREL, but are consistent with performance expected to be commensurate with the above-cited patent literature basis which formed the majority of the modeling assumptions here [32, 33]. In summary, Table 20 presents amortized catalyst operating costs for the hydrogenation, APR, and condensation reactors including the annual catalyst consumption and cost per kilogram. The hydrogenation reactors are initially filled with 40,818 kg of precious metal on carbon support catalyst at a cost of \$121/kg (2011\$). The estimated catalyst lifetime is 1 year with regeneration at 6 months, as a targeted n^{th} -plant projection for anticipated target performance based on guidance from Harris Group and Harris Group's industry contacts. Being based on a carbon support, it is assumed that the catalyst is regenerated by hot hydrogen stripping techniques as utilized in the petroleum industry for catalysts which cannot be regenerated by standard combustion regeneration [65-67]. The APR and condensation reactors are filled with 48,981 kg and 40,818 kg of catalyst respectively, each with projected target 2-year lifetimes with regeneration every 12 months. As noted above, the 1-2 year targeted catalyst lifetimes for all three reactor steps correspond with standard catalyst lifetimes for established refining technologies in the petroleum industry. The APR catalyst is a base metal on mixed oxide support at a cost of \$16/kg (2011\$). The condensation catalyst is a precious metal on mixed oxide support catalyst with a cost of \$43/kg (2011\$). The catalyst consumption amortized to an annual basis for each reactor is based on the catalyst loading, time between

regenerations, and total lifetime for each catalyst, and translates to annual operating costs of \$4.9 MM, \$0.4 MM, and \$0.9 MM for the three respective reactors.

Table 20. Catalytic Reactor Design and Cost Summary for Hydrogenation, APR, and Condensation Reactors

<i>Parameter</i>	<i>Units</i>	<i>Hydrogenation</i>	<i>APR</i>	<i>Condensation</i>
Cladding material		317SS	317SS	317SS
Weight hourly space velocity	hr ⁻¹	1.2	1.0	1.2
Catalyst loading (total)	kg	40,818	48,981	40,818
Catalyst bulk density	g/mL	0.62	1.2	0.70
Total catalyst bed volume	ft ³	2,325	1,441	2,059
Reactor internal allowance	%	33%	33%	33%
Total reactor (internal) volume	ft ³	3,470	2,151	3,073
Reactor internals		Inlet diffuser, inlet distribution tray, catalyst support grid		
Catalyst lifetime	Years	1	2	2
Time between regenerations	Years	0.5	1	1
Catalyst cost (2011\$)	\$/kg	121	16	43
Initial catalyst charge cost (2011\$)	\$MM	4.9	0.8	1.8
Amortized annual catalyst cost (2011\$)	\$/year	4,940,000	390,000	880,000

As presented in NREL’s 2013 design report, the hydrotreating facility cost is an “inside battery limits” (ISBL) cost estimate, including costs for reactors, makeup and recycle gas compressors, fired heater, separation vessels, and distillation [8]. Additionally, while catalyst replacement cost was not explicitly specified for the hydrotreating section, one vendor noted that the catalyst would be an insignificant part of the total cost estimate for this particular system, at a typical catalyst lifetime of 2 years (standard for petroleum hydroprocessing) and the given feed specification. This guidance was based on a likely more difficult feed in the 2013 design basis, including higher oxygen content, higher total acid number (TAN), and higher level of impurities; thus, this assumption would be even more accurate here as catalyst costs would likely be more marginal in this design than the 2013 basis.

Based on a number of separate vendor-supplied cost estimates, combined with Harris Group’s expertise on the subject as well as typical industry costs for petroleum hydroprocessing, an original hydrotreater cost estimate of \$23 MM was used (installed cost) in the 2013 design case, which translates to \$30 MM installed in the present analysis scaling to the increased hydrotreater feed rate modeled here. The uncertainty range associated with this cost estimate was quoted as -50%/+100%, and thus is most appropriately viewed as an order-of-magnitude cost for processing this type of mildly to moderately oxygenated feedstock. The overall cost impact associated with this given range of uncertainty in the hydrotreating cost estimate is evaluated in Section 5.2, Cost Sensitivity Analysis. Even with such a relatively high margin of uncertainty in the hydrotreating cost estimate, we feel it is more prudent to leverage this information than attempt a cost estimate on our own (e.g., using cost estimation software) as the vendor estimates bring valuable insight as to important considerations such as reactor design, number of stages, and operating conditions.

3.6 Area 600: Wastewater Treatment

3.6.1 Overview

All the wastewater generated in the conversion process is sent to the wastewater treatment system in Area 600. The treated water is assumed clean and fully reusable by the process, which reduces both the fresh makeup water requirement and discharge to the environment. This WWT design adopted the same design used in the 2013 biological hydrocarbon design report using the design from Harris Group and Brown and Caldwell (a WWT technology vendor). Each unit was costed out on an individual basis with associated scaling factors both on hydraulic rate and chemical oxygen demand (COD) load. The simplified flow diagram is shown in Figure 10.

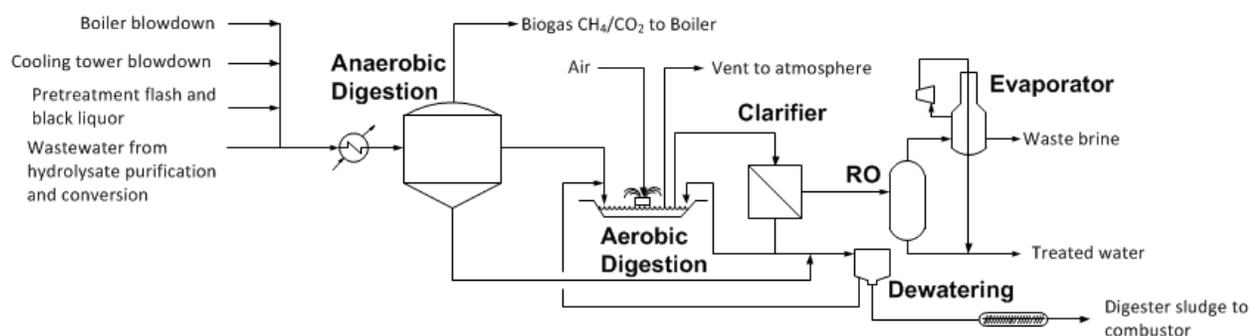


Figure 10. Simplified flow diagram of the WWT process

Condensed pretreatment flash vapor, black liquor from deacetylation, boiler blowdown, cooling tower blowdown, and the wastewater streams from hydrolysate purification (i.e., ion exchange backwash) and conversion operations (i.e., aqueous product streams from the condensation reactor flash and hydrotreater produced water) are mixed together. The combined wastewater stream is processed in A600 and consists of anaerobic and aerobic digestion, membrane filtration, reverse osmosis, evaporation, dewatering, and gravity belt thickening. Anaerobic digestion produces a biogas stream that is rich in methane, so it is fed to the combustor. The aerobic system utilizes a membrane bioreactor which is essentially an aeration tank followed by an ultrafiltration membrane system with biomass recycle. The effluent of the membrane system is of sufficient quality to feed a reverse osmosis system directly (i.e., without additional filtration). The dewatered sludge and biogas are sent to the boiler in A800. The dewatering process for the anaerobic sludge is centrifugation. The aerobic sludge is thickened by gravity belt thickeners (GBTs) prior to centrifugation in order to reduce the number of centrifuges required. In general, centrifugation provides higher solids capture and a drier biomass cake compared to other dewatering systems thus reducing the size of the biomass burner system. It also provides a margin of safety against changes in sludge quality which could impact dewatering processes relying upon filtration (e.g., belt filter press). Reverse osmosis is required to remove dissolved salts so that cleaned water can be recycled to upstream processes while mitigating salt accumulation.

This process produces a relatively clean water stream that can be reused in the process. The water stream is routed to the common process water header (A900) to be used in other process areas, such as dilution water in pretreatment or enzymatic hydrolysis.

3.6.2 Design Basis

The WWT system as designed by Harris Group and Brown and Caldwell includes anaerobic digestion, aerobic digestion, sludge dewatering, reverse osmosis, evaporation, and centrifugation. Evaporated water from the sugar concentration step is of high purity (> 99.5% water), and thus is not routed to A600 for further cleanup but is recycled directly for process water use. This helps reduce the total hydraulic flow and cost of a number of WWT unit operations. The deacetylation black liquor, condensed pretreatment flash vapor, ion exchange backwash, condensation/upgrading reactor aqueous product phases, boiler blowdown, and cooling tower blowdown are mixed together and pumped through a heat exchanger and cooled to 35°C before being sent to the anaerobic digester.

The total COD entering anaerobic digestion in the current design case is approximately 100 g/L (20,000 kg/h), shown in Table 21. This COD concentration represents a reduction from the 2013 design report (120 g/L), primarily because of improved carbon efficiency in the present catalytic conversion pathway and less residual unconverted carbon (COD) ultimately reaching this section, although total volume flow is similar. Along with COD, total (suspended + soluble) solids and suspended solids have also decreased in concentration in the wastewater feed stream again due to less unconverted material ultimately reaching wastewater treatment. Total solids in the wastewater feed include unconverted sugars, extractives, solubilized lignin, salts, and insoluble solids such as ash, lignin, and cellulose. While the COD concentration has decreased, the wastewater material in this case has the potential to include more problematic and potentially toxic components for the microbes, including light oxygenates and hydrocarbons produced in the conversion section; given the lack of operational data for treatment of wastewater generated in this process, the assumptions for the WWT system were left unchanged for the time being. Low-rate bulk volume fermentors (BVF reactors) are used as the anaerobic digester vessels, capable of removing 89% of incoming COD. 85% of the COD is converted to biogas (methane and carbon dioxide) and 5% is converted to cell mass. Methane is produced from the organic matter at 281 g/kg COD removed. Carbon dioxide is also produced at a nearly equimolar rate to methane, such that the biogas from the digester is 51% CH₄/49% CO₂ on a dry molar basis; this ratio is lower than typical digester biogas at 60%–70% CH₄, but is consistent with the vendor suggestions for such an application. Cell mass is produced at a yield of 58 g cell mass per kg COD removed. Biogas blowers pull a negative pressure to remove the produced gas.

The preferred use of the biogas is in the on-site combustor for combined heat and power benefits. However, economic factors may drive the sale of the gas via pipelines. In either case, an emergency biogas flare is required and designed into the process to burn the biogas in case of a process upset in the combustor. Sulfate is converted to hydrogen sulfide (H₂S) and leaves the system in the biogas. Since the COD-to-sulfate ratio is approximately 2,000, the methane-producing organisms will out-compete the sulfate-reducing organisms, so upstream removal of sulfate is unnecessary. The H₂S and any other sulfur species in the biogas are converted to SO₂ downstream in the combustor, which is mitigated using flue gas desulfurization (FGD) (Area 800).

Table 21. WWT System Design Basis

	2013 WWT design (biological conversion)	This design (catalytic conversion)
Hydraulic load	1.6 MMgal/d	1.3 MMgal/d
Wastewater inlet temperature	60°C	60°C
Total COD	120 g/L	100 g/L
Total solids	130 g/L	120 g/L
Total suspended solids	27 g/L	17 g/L
Total Kjeldahl N	724 mg/L	710 mg/L

The liquid from the anaerobic digester is pumped to the aerobic activated-sludge basin with surface aerators. In the concrete and steel basin, 96% of the remaining soluble organic matter is removed, with 74% producing water and carbon dioxide and 22% forming cell mass. With the 89% COD reduction in anaerobic digestion followed by 96% reduction of the remaining COD in aerobic digestion, the total COD reduction is 99.6%. The fully digested material is pumped to a membrane bioreactor for clarification. The membrane unit removes additional COD along with colloidal particles (especially silica). Coarse- and ultra-filtration upstream of the membrane unit separate the aerobic biomass sludge. The sludge is pumped to a dewatering device to reduce the sludge volume and water content to the boiler. A gravity belt thickener applied first produces sludge with a solid content of 4%. The centrifuge units are then periodically operated to provide a solids capture rate of 95% and a cake concentration of 20%. The centrate flow is recycled with polymer addition to the activated-sludge aerobic system. Most of the dewatered sludge is recycled to the aeration basin to maintain a high cell mass loading. The remaining portion is removed to a holding tank, where it is mixed with sludge from anaerobic digestion. Dewatered sludge is then routed to the combustor in A800. The sludge solid is primarily cell mass and unconverted biomass components. The treated water is pumped to a reverse osmosis (RO) membrane system for salt removal. RO produces brine stream containing primarily sodium nitrate along with all remaining ions and organics. The brine is further concentrated in a mechanical-vapor-recompression evaporator to 50% solids and the condensate is also recycled to the process. We do not further process the brine in our model at this time, and it is assumed to be waste.

3.6.3 Cost Estimation

The cost estimation using the 2013 design basis allows for greater detail to adjust individual unit costs (relative to prior estimates utilized in the 2011 case) either based on hydraulic flow rate or by COD, thus resulting in improved accuracy (e.g., to accommodate differences in COD concentration even at similar total hydraulic flow rate). The anaerobic system is scaled based on total COD and the installed cost in this design is reduced to \$24 MM due to lower COD feed rate as discussed above. For the aerobic digestion step, liquid volume per basin is 4.2 MM gal with mean cell residence time of 15 days. The installed capital cost of the aeration basin is estimated at \$10.2 MM here, including capital costs for the surface aerators and air blowers. The membrane bioreactor, dewatering units, reverse osmosis system, and evaporators are also scaled based on hydraulic flow rate, contributing another \$13.7 MM capital costs. Influent Total Kjeldahl Nitrogen (TKN) is estimated at 710 mg-N/L in this design by the Aspen Plus simulation, and it is low enough that a nitrification step is not needed (similar to the 2013 case as well). This is due to the shift toward increasingly more mild pretreatment conditions and acid loading (with subsequent lower ammonia neutralization demand), consistent with the 2013 report. Instead, supplemental ammonia is needed to make up for a slight nitrogen *deficit* now

present in the model to meet the optimum nitrogen demand (922 mg-N/L) for microorganisms to metabolize organic compounds in the aerobic digestion unit. The resulting ammonia demand in the aerobic basin is calculated to be 109 kg/h. In addition, polymer is added to the centrifuge to increase dewatering efficiency, specified at 9 lb polymer per ton total solids feeding into the aeration basin. The model estimates a polymer loading at 2 kg/h, with operating expenses discussed in more detail in Section 4.

3.7 Area 700: Product and Feed Chemical Storage

3.7.1 Overview

This portion of the plant provides bulk storage for process chemicals and the fuel products. The chemicals stored in this area include ammonia, CSL, sulfuric acid, sodium hydroxide, and purchased glucose for enzymes. Water for fire suppression is also stored here.

3.7.2 Design Basis

Table 22 shows the major storage requirements for the present design. The design and cost bases for all tanks maintained in the current design were left unchanged relative to the 2013 basis, and are scaled to a new cost if needed. The only exception is the product storage tanks, whose total storage volume is increased to 1.2 MM gal for the diesel (RDB) product given a higher RDB production rate relative to the 2013 basis, as well as the addition of a second product storage tank for the naphtha product cut, sized at 300,000 gallons; both tanks are sized for seven days of storage. Other supplemental tanks and pumps were left unchanged relative to the 2011 ethanol design basis, with more details provided in the 2011 report.

3.7.3 Cost Estimation

The costs for the A700 storage section were left unchanged relative to the 2011 ethanol and subsequent 2013 biological hydrocarbon design basis (with the exception of the main product tanks as noted above), and costs are scaled according to new material flow rates to estimate new tank prices if applicable.

Table 22. Storage Requirements

Material	Size (size basis maintained from 2013 biological pathway design report)
Diesel product	Sufficient to contain 7 days of production: 1 carbon steel tank @1,200,000 gal
Naphtha product	Sufficient to contain 7 days of production: 1 carbon steel tank @ 300,000 gal
Sulfuric acid (93%)	1 carbon steel tank @ 12,600 gal
Caustic (as pure)	Duplicated from sulfuric acid tank (includes caustic for deacetylation + IX regeneration)
Ammonia	2 SA-516-70 tanks @ 28,000 gal: ammonia is stored anhydrous at 250 psig
HCl (IX regeneration)	Duplicated from sulfuric acid tank
Fire water	4 hours of fire suppression @ 2,500 gpm: 1 glass-lined carbon steel tank @ 600,000 gal
Corn steep liquor	1 glass-lined carbon steel tank @ 70,000 gal
Glucose syrup for A400	Duplicated from CSL tank: 1 glass-lined carbon steel tank @ 70,000 gal
Diammonium phosphate (DAP)	1 SS304 tank @ 12,800 gal

3.8 Area 800: Combustor, Boiler, and Turbogenerator

3.8.1 Overview

The purpose of the combustor, boiler, and turbogenerator subsystem is to burn various organic byproduct streams to produce steam and electricity. Combustible byproducts include insoluble lignin and the unconverted cellulose and hemicellulose from the feedstock, biogas from anaerobic digestion, biomass sludge from WWT, and off-gases from the conversion/upgrading operations. Burning these byproduct streams to generate steam and electricity allows the plant to be self-sufficient in heating energy (“thermal-neutral”), reduces solid waste disposal costs, and minimizes the amount of electricity that must be purchased from the grid. Sulfur oxide (SO_x) emissions are mitigated from the combustor via flue-gas desulfurization.

The fuel streams are fed to a combustor capable of handling the wet solids. A fan moves air into the combustion chamber. Treated water enters the heat exchanger circuit in the combustor and is boiled and superheated to high-pressure steam. A multi-stage turbine and generator are used to generate electricity. Steam is extracted from the turbine at two different pressures for use in the process, and additional saturated steam is also extracted prior to the superheater. In the final stage of the turbine, the remaining steam is taken down to a vacuum and condensed with cooling water for maximum energy conversion. The condensate is returned to the boiler feed water system along with condensate from the various process heat exchangers. The steam turbine turns a generator that produces AC electricity to supply users in the plant. After considering all facility power demands, a net power surplus remains which is exported to the grid for coproduct revenue.

3.8.2 Design Basis

For the A800 section equipment, the fundamental design and cost basis assumptions as provided by a boiler vendor employed in the 2011 ethanol design report were maintained in the present model. Thus, details of the boiler and turbogenerator design will not be repeated here, but can be found in the 2011 report. To briefly summarize, the combustor system features a live-bottom grated fuel bin to ensure drying and complete combustion of the wet solid fuel, with a boiler efficiency near 80%. Flue gas from the combustor preheats the entering combustion air then enters a spray dryer absorber for flue gas desulfurization (FGD) (with the majority of sulfur entering the boiler system in the form of ammonium sulfate salt introduced upstream during pretreatment and conditioning operations). Ash and residual lime and calcium sulfate from the spray dryer absorber are removed in a baghouse and disposed to a landfill. Finally, the boiler raises steam at 875 psig, a portion of which (18%) is split off for use in satisfying high-temperature steam utility heat demands in the conversion/upgrading process (namely, heating the feed stream to the condensation reactor as well as the hydrotreater distillation column reboiler). The remainder is superheated and then sent through a multi-stage steam turbine with two extraction ports and a final condenser.

Similar to the 2013 biological design case, a number of process parameters are somewhat different for the boiler and turbogenerator system than the original design basis established in the 2011 report. Primarily, this includes a relatively high moisture content of 60% in the combined solid feed to the combustor versus 44% in the original 2011 ethanol model. This is primarily due to the lignin separation step, which is now placed upstream of the conversion step. This necessitates the use of a solids wash step to recover carryover losses of sugars and other soluble carbon components into the solids material from the filter press. This wash step adds more water to the lignin stream, which diverts more heat toward drying and reduces the amount of heat generated in the combustor. Compared to the

2013 biological design case, this process sends less combustible material to the boiler by way of the liquid sludge stream from WWT (due in turn to higher utilization of carbon in the hydrolysate including sugar oligomers, solubilized lignin, and furfurals), but also produces more combustible offgas through the conversion and upgrading process, thus resulting in similar overall heat generation from the boiler (146 versus 143 MMkcal/hr in the present design and the 2013 biological design respectively, after accounting for boiler heat losses). This translates to a similar amount of total steam generated as the biological design case, however as noted above the present design diverts nearly 20% of the high-pressure steam immediately away from the turbine system for use in high-temperature utility heat demands required for the “thermochemical” catalytic operations which was not required in the low-temperature bioconversion pathway. Consequently, a lower amount of steam is sent through the turbine stages, which reduces the net power generation from the turbine by roughly 13% (from 53.1 MW to 46.4 MW). Aside from the added steam heating demands for the conversion/upgrading section, remaining steam demands stay consistent with the 2013 case, namely high-pressure steam extracted after the first turbine stage for injection to the pretreatment reactor as well as low-pressure steam extracted after the second turbine stage to provide minor supplemental heating to the MVR evaporation unit.

After the intermediate extraction point, the remainder of the steam is sent through a final turbine stage to vacuum pressure, and is condensed at -13 psig (0.1 atm) and pumped back to the boiler. The turbine generator efficiency is assumed to be 85% as in prior designs. The resulting power generation is 46 MW. The process requires 35 MW of power, resulting in a net power surplus of 11 MW which is exported to the grid as a coproduct.

3.8.3 Cost Estimation

The cost basis for the A800 equipment remains the same as described in the 2011 and 2013 reports. Namely, the combustor/boiler system is based on a vendor quote and includes the boiler feed water preheater, FGD spray dryer, and baghouse. For the baghouse, bag replacement appears as a periodic charge in the cash flow worksheet. The turbogenerator is also based on a vendor quote for an industrial generator which remains suitable for this project. Harris Group also obtained quotes from a third vendor for support equipment including the deaerator, chemical injection system, tanks, and pumps.

3.9 Area 900: Utilities

3.9.1 Overview

Area 900 contains the utilities required by the biorefinery facility (except for steam, which is provided by Area 800). Area 900 tracks cooling water, chilled water, plant and instrument air, process water, and the clean-in-place (CIP) system. In the model, Area 900 also tracks the electricity usage throughout the plant.

The process water manifold in Area 900 mixes fresh water with treated wastewater and condensate from the sugar evaporation system (assumed suitable for all plant users) and provides this water at a constant pressure to the facility. Water is provided to the cellulase production unit, boiler and cooling tower makeup, the CIP system, and the wash for the lignin filter press. Fresh water is also mixed with some internally-recycled water for dilution before pretreatment and enzymatic hydrolysis. The plant and instrument air systems provide compressed air for general use (pneumatic tools and cleanup) and instrument operation. Larger users of compressed air, namely the cellulase system, have their own

compressors specified. The CIP system provides hot cleaning and sterilization chemicals to hydrolysis and the enzyme production section.

3.9.2 Design Basis

The cooling water system is designed for a 28°C supply temperature with a 9°C temperature rise in coolers throughout the facility. This is an assumed average rise; the actual cooling water rises across each exchanger are not explicitly modeled in Aspen. The primary cooling water users in this process are listed in Table 23. The percentage of cooling duty contributed by each user is shown in Figure 11. Compared to the 2013 biological pathway design basis, the total cooling water demand has decreased by 14% from 130 to 112 MMkcal/h, due primarily to reduced steam turbine condenser duty for a smaller steam flow reaching the condensing turbine (noted above in Section 3.8), as well as reduced chiller duty and elimination of a “bioreactor air cooler” which had been required in the 2013 biological pathway utilizing aerated bioreactors. This is discussed and analyzed in further detail in Section 5.

Table 23. Cooling Water Users

Hydrolysate flash condenser	Condenses residual pretreatment flash vapor before it enters WWT. Most of the vapor is condensed by heat exchange with the incoming dilution water and boiler feed water.
Hydrolysis cooler	Cools the pretreated slurry to enzymatic hydrolysis temperature (48°C).
Evaporator condensate cooler	Provides cooling of the sugar evaporator condensate.
Enzyme air compressor cooler	Uses cooling water to cool the compressed air used in enzyme production.
Conversion/upgrading trim coolers	Trim cooling duty is used in the catalytic conversion/upgrading area, including distillation column condenser duty, hydrotreating cooling duty, and cooling to reduce both naphtha and diesel products to 43°C.
Condensation rx product cooler	Cools the product stream from the catalytic condensation reactor stage prior to flashing.
Turbine condenser	Condenses the steam turbine exhaust at a vacuum.
Chiller condenser	The chilled-water loop requires cooling water to condense the refrigerant. The cooling water duty to M-908 is set equal to the total load on the chilled-water loop.

As was the case in prior design models, the largest user of cooling water continues to be the condensing turbine. Aspen computes the cooling tower evaporation rate based on a temperature drop from 37°C to 28°C. It was assumed that windage would be 0.005% of the total flow to the tower. The tower blowdown was assumed to be 0.15% of the flow leaving the tower basin.

Chilled water is provided by two 2,350-ton Trane centrifugal chillers. Per the chiller spec sheet, the compressor electricity demand for the chiller was estimated at 0.56 KW/ton of refrigeration. The cooling water demand for the chiller system was assumed to be equal to the heat removed in the chilled-water loop. The chiller provides cooling to the enzyme production bioreactor in Area 400. The total chiller duty has decreased considerably relative to the 2011 and 2013 cases, which both required chilled water to maintain bioreactor operating temperatures and have now been replaced with high-temperature catalytic reactors.

Fresh water is assumed to enter the facility at 13°C and is used to cool the wastewater entering Area 600 to digestion temperature before entering the process water tank. The fresh water is mixed with the treated wastewater effluent and the evaporation condensate in the process water tank (T-914) and then

split several ways. Clean water must be provided to biomass dilution in the front end, to the cellulase production unit, to the lignin filter press wash step, and to the boiler and cooling tower water makeup. The process water tank is designed for an 8-hour residence time. The process water pump (P-914) pumps water from the tank into the facility and is designed to handle 1.5 times the process water flow requirement.

The plant and instrument air systems provide compressed air for pneumatic tools and cleanup and instrument operation (not including major air demands such as enzyme bioreactors). The plant air compressor is sized for 400 standard cubic feet per minute (SCFM) at 125 psig. An instrument air dryer and surge tank were designed to provide clean dry air at a consistent pressure to the instrument air system. The surge tank was sized at 3,800 gal.

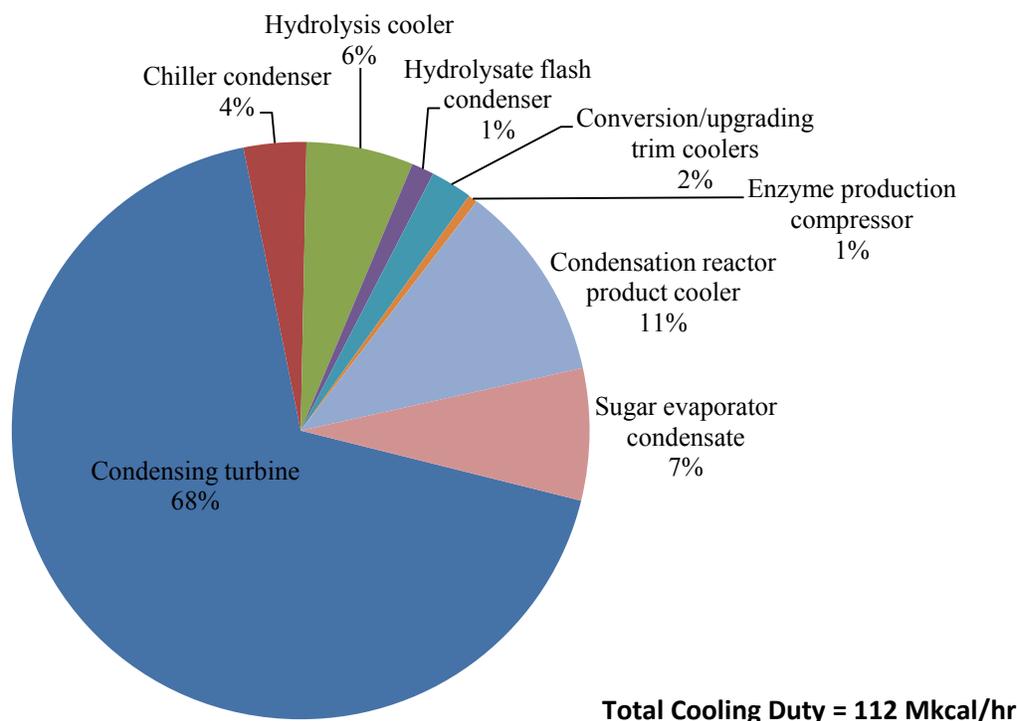


Figure 11. Cooling water heat duty distribution between major users

About 76% of the electricity generated by the boiler in Area 800 is used throughout the plant to power pumps, agitators, compressors, etc. The surplus is sold to the grid for credit. The distribution of total plant power utilization among all areas is shown in Figure 12. The conversion/upgrading process area contributes the most to overall facility power demand, requiring nearly 9 MW of power. This is largely attributed to high compressor power demand for the recycle gas compressor and to a lesser degree the makeup hydrogen compressor, given the high hydrogen demand for the process coupled with relatively high operating pressures (up to 1,050 psig for the conversion step). Note that the cost of the power required by Area 100 is already assumed to be included in the feedstock cost but must be subtracted from the plant's electricity balance. This is reflected in the economics by an operating cost credit equal to this amount of electricity.

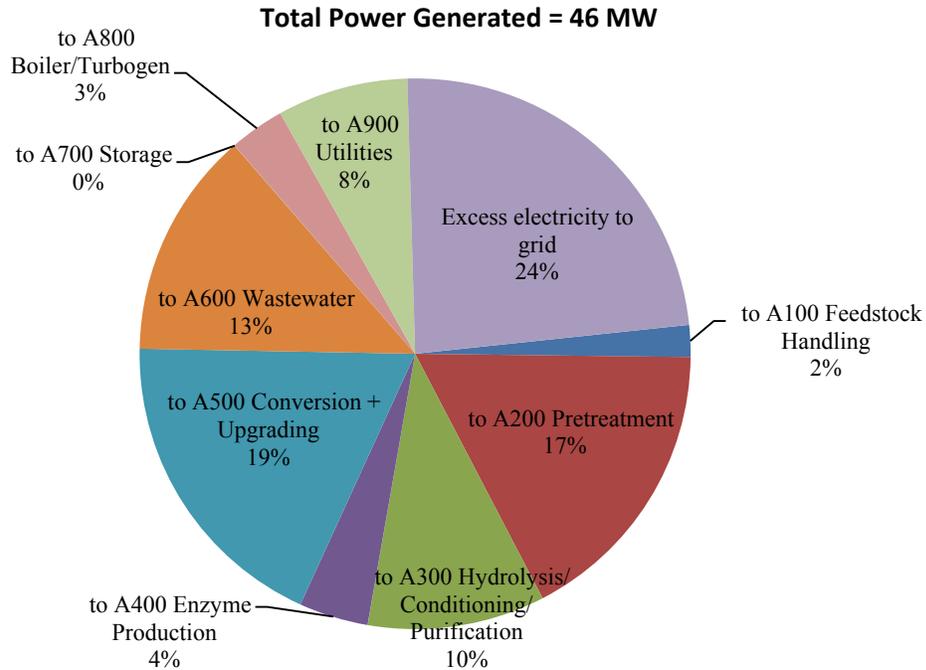


Figure 12. Distribution of plant electricity utilization by process area

3.9.3 Cost Estimation

All cost estimates for the utility equipment in A900 were maintained consistent with the basis values used in the 2011 report. To summarize, the cooling tower was based on a cost estimate from a vendor for a fiberglass cooling tower capable of handling 44,000 gpm; this cost is scaled to the increased cooling water throughput estimated here. Harris Group provided estimated costs for the cooling water circulation and makeup pumps from their historical database. The material of construction for the cooling water piping is carbon steel. The cost for the chiller came from a recent quote for a similarly-sized system. Harris Group had also used their historical database to estimate costs for the remaining equipment: the process water tank and pump; the plant/instrument air compressor, dryer, and surge tank; and the CIP system.

4 Process Economics

The ultimate purpose for developing such a detailed process design, simulation model, and cost estimate is to determine the economics of biofuel production. This information is used either as an absolute cost to assess the product's potential in the marketplace or as a relative cost that can be used to guide research by examining the change in production cost associated with a process modification or other core research activity.

The total capital investment (TCI) is first computed from the total equipment cost. Next, variable and fixed operating costs are determined. With these costs, we use a discounted cash flow analysis to determine the minimum fuel selling price required to obtain a zero net present value (NPV) with a fixed internal rate of return (IRR). This section summarizes the assumptions made in completing the discounted cash flow analysis, with more details and supporting description available in the 2011 ethanol design report [2] and 2013 biological hydrocarbon design report [8] for assumptions that are unchanged. Our analysis does not take into account any policy factors such as subsidies, mandates, or carbon credits, because these would be purely speculative. The purpose of this analysis is to demonstrate the process requirements needed to achieve specific DOE cost targets which are set from the top down, and to demonstrate how the technology pathway described here is able to achieve such targets on its own merits (through bottom-up TEA modeling) and, if it cannot, to give policymakers a sense of the magnitude of incentive required to make it so.

4.1 About Cost-Year Indices

The cost-year of 2011 was chosen for this analysis to provide more updated and relevant cost output information relative to the 2007-year basis, which had been consistently utilized for a number of years in prior analyses [2, 68]. This new basis is being applied consistently across all DOE-BETO platforms for which similar “design case target” reports are being established during 2013–2014 efforts, and it is expected that performance goals and TEA outputs will remain in 2011\$ through 2017 to permit comparison of future feedstocks, conversion technologies, and other alternative scenarios. However, the present equipment costs were obtained in 2012, 2013, or 2014 dollars for new pieces of key equipment which were added or replaced in the current design model (as part of the subcontract with Harris Group), and in 2009 or 2010 dollars for unit operations that were previously provided by Harris Group and vendors in support of NREL's 2011 ethanol design report and 2013 biological design report, and are still maintained in the present model. Cost-years for chemicals range from 1999 to 2012.

The methods used for determining MFSP in another year's dollar value and for scaling capital, operating, and labor cost estimates to a desired target year, remain the same as described in the previous NREL design reports [2, 8, 69]. Thus, the details will not be repeated here, but will be summarized briefly. Capital costs provided in a year other than 2011\$ were adjusted using the Plant Cost Index from *Chemical Engineering Magazine* [70] to a common basis year of 2011. The final cost index for a given year is generally not made available until the spring of the following year. Therefore, for the small number of equipment items that were quoted in 2014\$, we assumed the same Plant Cost Index value from 2013 (all cost quotes which fall in this category were provided in the first half of 2014). Similarly, for chemical costs we used the Industrial Inorganic Chemical Index from SRI Consulting [71]. Employee salaries were maintained from 2009\$ and were scaled using the labor indices provided by the U.S. Department of Labor Bureau of Labor Statistics [72]. The general formula for cost-year dollar back-casting is:

$$2011 \text{ Cost} = (\text{Base Cost}) \left(\frac{2011 \text{ Cost Index}}{\text{Base Year Index}} \right)$$

To maintain consistency with the 2013 and 2011 reports, capital costs and material prices for items which were unchanged in the present analysis were not modified except to update to 2011 dollars. The only new “feed materials” that were added to the present design that were not required in the previous 2013 biological hydrocarbon case were chemicals and catalysts used in the sugar purification and catalytic conversion steps. The electricity export price, taken to be the average wholesale price as determined by the North American Electric Reliability Corporation, was left unchanged from the 2013 report at \$0.0572/KWh [73-74].

4.2 Total Capital Investment

Section 3 of this report describes the details of the conceptual process design and how the purchased cost of the equipment was determined. The next step is to determine the installed cost of that equipment. The installation cost can be determined by performing a detailed study of everything required to install the necessary equipment and make it operational (e.g., foundation, piping, and wiring). This type of detail is not warranted at this level of process development, and a factored approach in which multipliers are applied to the purchased equipment cost is considered satisfactory. The methodology and rationale for applying unit-level installation costs remain the same as described in the 2011 and 2013 reports, and again further detail can be found there that will not be repeated here.

In summary, each type of equipment utilizes a different installation factor to scale the given direct equipment purchased cost to a final installed cost, with these factors generally varying between 1.5 and 3.0. A complete list of the equipment is provided in Appendix A, along with equipment purchased and installed costs. As described in the 2013 report, a number of recent cost estimates furnished by Harris Group utilized budgetary vendor quotes for system “package” costs, whereby a given unit operation and all of its supporting equipment were quoted under one price; the installation factor for such packages can be relatively low because much of the engineering is already included in the price. Additionally, equipment designed as a pre-fabricated skid generally has a lower construction cost. Also, components that are more highly machined or have higher metallurgy tend to have a higher purchase price relative to the cost of installation than a less sophisticated component, and therefore have a lower installation factor.

The purchased cost for a given component reflects a baseline equipment size. As changes are made to the process, the equipment size required may be different than what was originally designed and costed. Instead of re-costing in detail, an exponential scaling expression was used:

$$\text{New Cost} = (\text{Base Cost}) \left(\frac{\text{New Size}}{\text{Base Size}} \right)^n$$

In this equation, the scaling exponent n varies depending on the type of equipment to reflect economy-of-scale dependencies (more detail on reasonable scaling values for different types of equipment is provided in the 2011 ethanol report [2]). The basis for scaling is typically some characteristic of the equipment related to production capacity, such as flow or heat duty. Some equipment does not follow such a scaling-factor approach, namely when the capacity for a given operation is exceeded and requires multiple units in parallel, thus losing economy of scale benefits which are captured in the exponential expression above.

Once the total equipment cost has been determined in the year of interest, several other direct and indirect costs were added to determine the total capital investment (TCI). Site development and warehouse costs are based on the inside-battery-limits (ISBL) equipment costs (Areas 200, 300, 400, and 500) and are considered part of the total direct cost (TDC). Beyond ISBL Areas 200–500, the other process areas are considered outside battery limits (OSBL), including Areas 100 (rolled up into feedstock costs) and 600-900. Project contingency, field expenses, home office engineering and construction activities, and other costs related to construction are computed relative to the TDC and give the fixed capital investment (FCI) when summed. The sum of FCI and the working capital for the project is the TCI. Table 24 summarizes these categories and additional factors. The values assumed for each respective factor were maintained consistently with those discussed in the 2011 ethanol report [2] as well as the 2013 biological hydrocarbon report [8].

Table 24. Additional Costs for Determining TCI

Item	Description	Amount
Additional direct costs		
Warehouse	On-site storage of equipment and supplies.	4% of ISBL ^a
Site development	Includes fencing, curbing, parking lot, roads, well drainage, rail system, soil borings, and general paving. This factor allows for minimum site development assuming a clear site with no unusual problems such as right-of-way, difficult land clearing, or unusual environmental problems.	9% of ISBL
Additional piping	To connect ISBL equipment to storage and utilities outside battery limits.	4.5% of ISBL
Indirect costs		
Prorateable costs	This includes fringe benefits, burdens, and insurance of the construction contractor.	10% of total direct cost (TDC)
Field expenses	Consumables, small tool and equipment rental, field services, temporary construction facilities, and field construction supervision.	10% of TDC
Home office and construction	Engineering plus incidentals, purchasing, and construction.	20% of TDC
Project contingency	Extra cash on hand for unforeseen issues during construction.	10% of TDC
Other costs	Start-up and commissioning costs. Land, rights-of-way, permits, surveys, and fees. Piling, soil compaction/dewatering, unusual foundations. Sales, use, and other taxes. Freight, insurance in transit, and import duties on equipment, piping, steel, instrumentation, etc. Overtime pay during construction. Field insurance. Project team. Transportation equipment, bulk shipping containers, plant vehicles, etc.	10% of TDC

^a ISBL = installed cost of equipment inside battery limits (A200,300,400,500).

Table 25. Project Cost Worksheet Including TDC and TCI

Process Area		Purchased Cost	Installed Cost
Area 100: Feedstock Storage and Handling ^a		\$ 15,800,000	\$ 26,900,000
Area 200: Pretreatment		\$ 33,900,000	\$ 51,400,000
Area 200: Neutralization		\$ 1,100,000	\$ 2,200,000
Area 300: Enzymatic Hydrolysis, Hydrolysate Conditioning & Purification		\$ 52,700,000	\$ 68,800,000
Area 400: Enzyme Production		\$ 7,300,000	\$ 12,400,000
Area 500: Catalytic Conversion & Upgrading		\$ 39,600,000	\$ 82,900,000
Area 600: Wastewater ^b		\$ 34,400,000	\$ 47,700,000
Area 700: Storage		\$ 2,800,000	\$ 4,800,000
Area 800: Boiler		\$ 42,200,000	\$ 76,300,000
Area 900: Utilities		\$ 4,000,000	\$ 7,000,000
Totals (Excl. Area 100)		\$ 217,900,000	\$ 353,500,000
Warehouse	4.0%	of ISBL	\$ 8,700,000
Site Development	9.0%	of ISBL	\$ 19,600,000
Additional Piping	4.5%	of ISBL	\$ 9,800,000
Total Direct Costs (TDC)			\$ 391,500,000
Prorateable Expenses	10.0%	of TDC	\$ 39,200,000
Field Expenses	10.0%	of TDC	\$ 39,200,000
Home Office & Construction Fee	20.0%	of TDC	\$ 78,300,000
Project Contingency	10.0%	of TDC	\$ 39,200,000
Other Costs (Start-Up, Permits, etc.)	10.0%	of TDC	\$ 39,200,000
Total Indirect Costs			\$ 234,900,000
Fixed Capital Investment (FCI)			\$ 626,500,000
Land			\$ 1,800,000
Working Capital	5.0%	of FCI	\$ 31,300,000
Total Capital Investment (TCI)			\$ 659,600,000
Lang Factor (TCI/Purchased Equip Cost)			3.4
TCI per Annual Gallon Gasoline Equivalent 2011 Dollars			\$11.63/GGE
			BC1411A

^a Feedstock handling not included in this calculation.

^b Area 600 not included in Lang Factor.

4.3 Variable Operating Costs

Variable operating costs, which include raw materials, waste handling charges, and byproduct credits, are incurred only when the process is operating. Quantities of raw materials used and wastes produced were determined using the Aspen material balance. Table 26 documents the costs and sources of chemicals used in the process and Table 27 summarizes the variable costs on a per-year and per-GGE basis. As noted above, the cost basis for all material costs in the present model, which were also used in the 2013 biological hydrocarbon model, were left unchanged.

Hydrogen was assumed here to be purchased as a product from standard natural gas-derived steam methane reforming (SMR) consistent with the 2013 report which utilized hydrogen for hydrotreating fatty acids. However, the hydrogen demands required for upgrading the fatty acid material in the 2013

case are substantially lower than total hydrogen demands in the present pathway, so two alternative on-site hydrogen production options are considered as alternate cases using *in situ* hydrogen production from a fraction of the hydrolysate carbon, as well as gasification of a fraction of the biomass to syngas and subsequently to purified hydrogen. The implications for cost and sustainability metrics associated with these alternative scenarios were evaluated to compare with externally supplied hydrogen sourced from natural gas; results for the alternate hydrogen cases are presented in Section 5. For the base case, the purchased hydrogen price was set based on a recent DOE Hydrogen Program report, which presented a current price for natural gas-based SMR hydrogen of \$1.57/kg (assumed in 2012\$) associated with a natural gas price of \$4/MM BTU [75], which was applied consistently following the 2013 biological hydrocarbon report. The overall cost sensitivity to the assumed hydrogen price is presented in the Sensitivity Analysis section, using a reasonable maximum and minimum hydrogen cost range as presented in the referenced DOE report, namely \$1.10–\$2.00/kg associated with a natural gas price range of \$2–\$7.37/MM BTU, respectively (a larger hydrogen price range is also considered separately in Section 5). Prices for all the raw materials are summarized below.

Table 26. Chemical Costs and Sources [8]

Component	Cost (2011\$)	Source
Biomass uniform-format feedstock	\$0.0320/lb	2013 MYPP, \$80/dry ton @ 20% moisture [7]
Sulfuric acid, 93%	\$0.0499/lb	Basic Chemical of Omaha via Harris Group
Ammonia	\$0.2496/lb	Terra Industries via Harris Group
Acid (HCl) for IX regeneration	\$0.0489/lb	Harris Group
Caustic (NaOH) for IX regeneration	\$0.1601/lb	Harris Group
Corn steep liquor	\$0.0316/lb	Corn Products via Harris Group
Corn oil (antifoam)	\$0.6077/lb	SRI Chemical Economics Handbook (CEH)
Glucose	\$0.3230/lb	USDA ERS [76]
SO ₂	\$0.1690/lb	SRI Chemical Economics Handbook (CEH)
Enzyme nutrients	\$0.4570/lb	SRI CEH (See 2011 design report for details)
Hydrogen	\$0.6838/lb	DOE report, SMR H ₂ @ \$4/MM BTU NG [75]
APR-1 catalyst (hydrogenation)	\$54.8730/lb	Harris Group
APR-2 catalyst (APR reactor)	\$7.2626/lb	Harris Group
Condensation catalyst	\$19.3670/lb	Harris Group
Caustic for WWT	\$0.0832/lb	Brown and Caldwell 2011 WWT design [77]
Polymer for WWT	\$2.4806/lb	Brown and Caldwell 2012 WWT design [78]
Lime	\$0.1109/lb	Harris Group Report 2013
Boiler chemicals	\$2.7788/lb	2002 Design Report [69]
Cooling tower chemicals	\$1.6653/lb	2002 Design Report [69]
Fresh water	\$0.0001/lb	Peters & Timmerhaus [79]

Table 27. Variable Operating Costs

Process Area	Stream Description	Usage (kg/hr)	Usage (lb/hr)	Cost (\$/ton)	\$/hour	MM\$/yr (2011\$)	Cent/GGE (2011\$)
Raw Materials							
N/A	Feedstock	104,167	229,688	64.00	7,350.00	57.95	102.18
A200	Sulfuric Acid, 93%	2,240	4,940	99.85	246.62	1.94	3.43
	Caustic (as pure)	1,406	3,101	166.42	258.02	2.03	3.59
	Ammonia	310	683	499.27	170.55	1.34	2.37
A300	HCl for IX Regeneration	1,122	2,474	97.86	121.04	0.95	1.68
	Caustic for IX Regeneration	640	1,410	320.28	225.82	1.78	3.14
A400	Glucose	1,213	2,675	645.94	863.86	6.81	12.01
	Corn Steep Liquor	83	182	97.86	8.90	0.07	0.12
	Corn oil	7	15	1,215.42	8.98	0.07	0.12
	Ammonia	58	127	499.27	31.75	0.25	0.44
	Host Nutrients	34	75	913.94	34.05	0.27	0.47
	Sulfur Dioxide	8	18	388.08	3.07	0.02	0.04
A500	Hydrogen	3,847	8,483	1,367.51	5,800.17	45.73	80.63
	APR-1 Catalyst ^a	5	11	109,746.10	626.42	4.94	8.71
	APR-2 Catalyst ^a	3	7	14,525.22	49.75	0.39	0.69
	Condensation Catalyst ^a	3	6	38,733.92	110.54	0.87	1.54
A600	Ammonia	109	239	499.27	59.78	0.47	0.83
	Polymer	2	5	4,961.25	11.16	0.09	0.16
A800	Boiler Chems	0.2	0.5	5,557.64	1.42	0.01	0.02
	FGD Lime	180	397	221.90	44.04	0.35	0.61
A900	Cooling Tower Chems	3	6	3,330.66	10.09	0.08	0.14
	Makeup Water	159,026	350,653	0.29	50.51	0.40	0.70
	Subtotal				16,086.56	126.83	223.63
Waste Disposal							
A800	Disposal of Ash	4,506	9,936	35.39	175.84	1.39	2.44
	Subtotal				175.84	1.39	2.44
By-Products and Credits							
	Grid Electricity	11,045	KW	\$0.0572/KWh	631.43	4.98	8.78
	Area 100 Electricity	859	KW	\$0.0572/KWh	49.13	0.39	0.68
	Subtotal				680.57	5.37	9.46
Total Variable Operating Costs					15,582	122.85	216.62

^aCatalyst usage amortized to kg/hr-basis for consistency with rest of table. APR-1 = "hydrogenation" reactor, APR-2 = "APR" reactor

4.4 Fixed Operating Costs

Fixed operating costs are generally incurred in full whether or not the plant is producing at full capacity. These costs include labor and various overhead items. The assumptions on fixed operating costs were maintained consistently from the 2011 design basis (after updating to 2011\$), which in turn were based in large part on NREL's 2002 ethanol design report [69] and/or Peters and Timmerhaus [79].

Table 28 shows the recommended number of employees and associated salaries. The number of employees was estimated by considering the likely degree of automation for each area and adding a reasonable number of management and support employees. Details behind the assumed number of employees and associated salaries are provided in the 2011 ethanol report. Because the model feedstock is predominately corn stover, salaries were estimated for rural regions of the U.S. Midwest (Iowa, Missouri, etc.). These estimates may vary depending on location. While beyond the scope of this analysis, some economy of scale advantages could further be gained with respect to labor costs by considering multiple units at the same site rather than a single stand-alone unit, such that some

overlapping positions could be shared (i.e. one plant manager); however, such savings would likely be marginal in comparison to overall facility costs.

Table 28. Fixed Operating Costs

Position	2011 Salary	# Required	2011 Cost	MM\$/yr (2011\$)	Cent/GGE(2011\$)
Labor and supervision					
Plant Manager	155400	1	\$155,000		
Plant Engineer	74000	2	\$148,000		
Maintenance Supr	60257	1	\$60,000		
Maintenance Tech	42286	12	\$507,000		
Lab Manager	59200	1	\$59,000		
Lab Technician	42286	2	\$85,000		
Lab Tech-Enzyme	42286	2	\$85,000		
Shift Supervisor	50743	4	\$203,000		
Shift Operators	42286	20	\$846,000		
Shift Oper-Enzyme	42286	8	\$338,000		
Yard Employees	29600	4	\$118,000		
Clerks & Secretaries	38057	3	\$114,000		
Total Salaries			\$2,719,000	2.72	4.79
Labor Burden (90%)			\$2,447,000	2.45	4.31
Other overhead					
Maintenance	3.0%	of ISBL	\$6,530,000	6.53	11.51
Property Insurance	0.7%	of FCI	\$4,385,000	4.39	7.73
Total fixed operating costs				16.08	28.36

A 90% labor burden is applied to the salary total and covers items such as safety, general engineering, general plant maintenance, payroll overhead (including benefits), plant security, janitorial and similar services, phone, light, heat, and plant communications. The 90% estimate is the median of the general overhead range suggested in the 2008 PEP Yearbook produced by SRI Consulting [71]. Annual maintenance materials were estimated as 3% of the installed ISBL capital cost and property insurance and local property tax were estimated as 0.7% of the fixed capital investment, based on the 1994 Chem Systems report described in NREL’s 2011 ethanol report. These factors are all consistent with those used in the 2011 and 2013 design reports.

4.5 Discounted Cash Flow Analysis and the Minimum Selling Price of Fuel

4.5.1 Discount Rate

For this analysis, the discount rate (which is also the IRR in this analysis) was set to 10% and the plant lifetime was set to 30 years. The discount rate was also used in previous design reports and was based on the recommendation in Short et al. [80] on how to perform economic evaluations of renewable energy technologies for DOE. His view was that, “In the absence of statistical data on discount rates used by industrial, transportation and commercial investors for investments with risks similar to those of conservation and renewable energy investments, it is recommended that an after tax discount rate of 10%...be used.”

4.5.2 Equity Financing

For this analysis, it was assumed that the plant would be 40% equity financed. The terms of the loan were established at 8% interest for 10 years. The principal is taken out in stages over the 3-year construction period. Interest on the loan is paid during this period, but principal is not paid back (this is another n^{th} -plant assumption, which says that this cash flow comes from the parent company until the plant starts up). This is all consistent with the assumptions used in the 2013 biological hydrocarbon and the 2011 ethanol reports. Figure 13 illustrates the sensitivity of MFSP to the percentage of equity financing and the after-tax discount rate (the IRR).

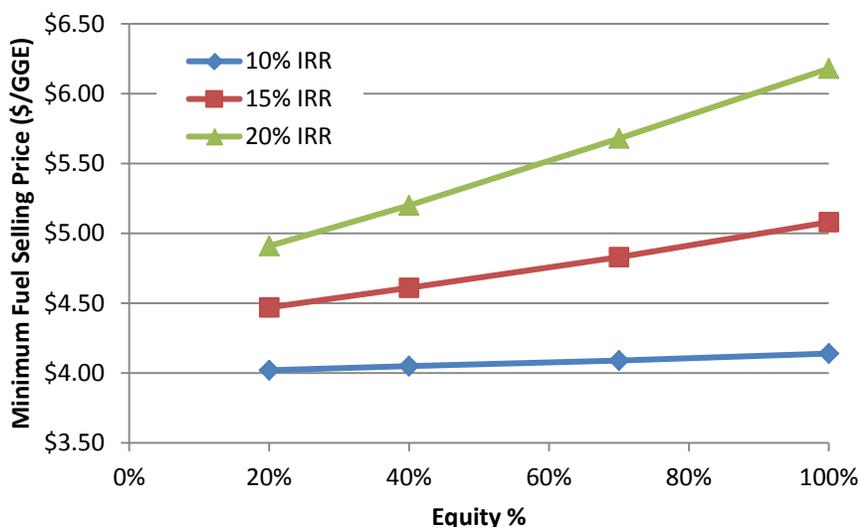


Figure 13. Sensitivity of MFSP to IRR and % equity. (8% interest on a 10-year loan.)

4.5.3 Depreciation

To determine the capital depreciation amount for the calculation of federal taxes to be paid, we used the IRS Modified Accelerated Cost Recovery System (MACRS). Within the MACRS system is the General Depreciation System (GDS), which allows both the 200% and 150% declining balance (DB) methods of depreciation. This offers the shortest recovery period and the largest tax deductions. According to IRS publication 946 [81], a cellulosic biorefinery plant would fall under Asset Class 49.5, “Waste Reduction and Resource Recovery Plants.” This class uses a 7-year recovery period, not including the steam plant equipment, which has a 20-year recovery period (Asset Class 49.13). IRS publication 946 contains a special provision for cellulosic biofuels plants that allows them to write off 50% of the capital investment in the first year. This was not implemented in our cost model because it does not ultimately affect the MFSP (although the provision affects the cash flow in the first few years of the analysis, it does not change the year in which the plant goes into the black and must start paying taxes).

4.5.4 Taxes

The federal corporate tax rate used in our analysis is 35%. Income tax is averaged over the plant life and that average is calculated on a per-GGE basis. The amount of income tax to be paid by a potential fuel producer varies annually due to changes in the volume of product produced and the allowable depreciation deduction. In fact, no income tax is paid in the first eight years of operation because the

depreciation and loan interest deductions are greater than the net income. State taxes are not considered, primarily because the location of the plant has not been determined and tax rates vary from state to state (from 0% to 12%).

4.5.5 Construction Time

The construction time is important to the cash flow analysis because no income is earned during construction, but huge sums of money are being expended. Construction time assumptions were left unchanged from the 2011 and 2013 design basis assumptions. Perry and Green [82] indicate that small projects (less than \$10 million investment) can be constructed in fewer than 18 months and that larger projects can take up to 42 months. An overview of petroleum refining economics indicates that large refineries (on the order of \$1.5 billion investment) can be constructed in 24 months [83]. Certainly this facility is much smaller than a petroleum refinery, so using a construction time of 24 months fits within these references, although an important difference between this type of facility and a refinery is the large number of field-erected vessels. These are constructed on-site and have a longer construction time than if the tanks were delivered finished. Table 29 summarizes the schedule for construction and the cash flow during that time. Twelve months are added before construction for planning and engineering.

Table 29. Construction Activities and Cash Flow

Project Start Month	Project End Month	Activity Description	% of Project Cost
0	12	Project plan and schedule established; conceptual and basic design engineering, permitting completed. Major equipment bid packages issued, engineering started on selected sub-packages, P&IDs complete, preliminary plant and equipment arrangements complete.	8%
12	24	All detailed engineering including foundations, structure, piping, electrical, site, etc., complete; all equipment and instrument components purchased and delivered; all site grading, drainage, sewers, rail, fire pond, foundation, and major structural installation complete; 80% of all major process equipment set (all except longest-lead items), all field fabricated tanks built, and the majority of piping and electrical materials procured.	60%
24	36	Complete process equipment setting, piping, and instrumentation installation complete; all electrical wiring complete; all building finishing and plumbing complete; all landscaping complete; pre-commissioning complete; and commissioning, start-up, and initial performance test complete.	32%
TOTAL			100%

Note: The above assumes no utility or process equipment orders placed prior to month seven. Expenditures are based on a typical 60 MMgal/yr grain-to-ethanol facility.

4.5.6 Start-Up Time

Perry and Green [82] indicate that for a moderately complex plant, startup should be about 25% of the construction time, or 6 months in this case. Similar to 2013 design basis, a start-up time of 6 months under an n^{th} plant assumption was assumed in this analysis. The start-up period is not completely wasted, however. We expect that an average of 50% production could be achieved during that period while incurring 75% of variable expenses and 100% of fixed expenses.

4.5.7 Working Capital

Peters and Timmerhaus [79] define working capital as money available to cover (1) raw materials and supplies in inventory, (2) finished product in storage, (3) accounts receivable, (4) cash on hand for monthly payments such as wages and maintenance supplies, (5) accounts payable, and (6) taxes

payable. The present analysis applies the same basis for working capital as was used in the 2011 ethanol report, namely 5% of FCI; this translates to a slightly higher working capital cost value given the increased direct capital costs and FCI in the present model.

Table 30 summarizes the parameters used in the discounted cash flow analysis. Using these parameters, plus the cost information in Table 25, Table 27, and Table 28, the resulting MFSP of total fuel products is \$4.05/GGE (2011\$), including both naphtha- and diesel-range product cuts adjusted by heating values (calculated in the Aspen model) to gasoline-equivalents. Table 31 summarizes the yields and conversion costs for the present design. According to the methodology of Cran [84], the expected accuracy of the overall TCI analysis is $\pm 25\%$ (although some specific pieces of equipment carry a higher degree of uncertainty in underlying cost quotations as identified above). If we apply this uncertainty to the TCI, the impact on the cost of total fuel is $\pm \$0.40/\text{GGE}$. The complete discounted cash flow summary worksheet is shown in Appendix B. The MFSP can be further broken down into the cost of each process area. Figure 14 illustrates the contribution to the overall cost by process area and capital, operations, and fixed costs (the bar for feedstock plus handling reflects the single feedstock cost of \$80.00/dry U.S. ton delivered to pretreatment and has not been decomposed).

Table 30. Discounted Cash Flow Analysis Parameters

Plant life	30 years
Discount rate	10%
General plant depreciation	200% declining balance (DB)
General plant recovery period	7 years
Steam plant depreciation	150% DB
Steam plant recovery period	20 years
Federal tax rate	35%
Financing	40% equity
Loan terms	10-year loan at 8% APR
Construction period	3 years
First 12 months' expenditures	8%
Next 12 months' expenditures	60%
Last 12 months' expenditures	32%
Working capital	5% of fixed capital investment
Start-up time	6 months
Revenues during startup	50%
Variable costs incurred during startup	75%
Fixed costs incurred during startup	100%

Table 31. Summary of Yields, Rates, and Conversion Costs

Feedstock rate	2,205 dry U.S. ton/day
Online time	7,884 h/yr (90% online factor)
Total fuel yield	78.3 GGE/dry U.S. ton feedstock
Total fuel production rate	56.7 MM GGE/yr
<i>Diesel-range production rate</i>	46.0 MM GGE/yr
<i>Naphtha-range production rate</i>	10.7 MM GGE/yr
Total equipment cost	\$354 MM
Total capital investment (TCI)	\$660 MM
TCI per annual gallon	\$11.63/GGE
Minimum Fuel Selling Price	\$4.05/GGE
Feedstock contribution	\$1.02/GGE
Enzyme contribution	\$0.21/GGE
Non-enzyme conversion contribution	\$2.81/GGE

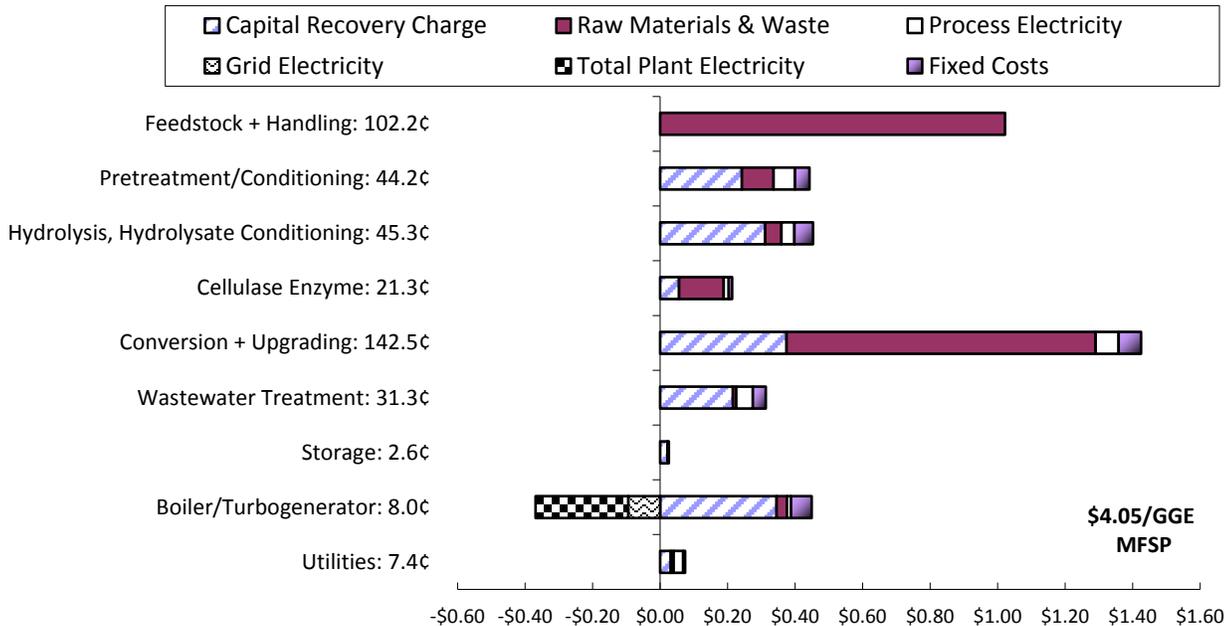


Figure 14. Cost contribution details from each process area. (Per GGE total fuel products.)

Figure 14 highlights a number of clear differences relative to NREL’s 2013 design report for biological conversion of sugars to hydrocarbons. Primarily, it presents a rare case where feedstock is *not* the primary cost driver at \$80/ton (translating to a cost contribution of \$1.02/GGE to overall MFSP), but is surpassed by the conversion/upgrading section contributing \$1.43/GGE or 35% to overall MFSP. In turn, this is driven by purchased hydrogen cost, contributing \$0.81/GGE attributed to the total hydrogen demand of 7.9 wt % (net, including purge losses) relative to the total convertible component feed rate to the conversion system. Although purchased hydrogen is clearly a key cost driver, because it is derived from low-cost external natural gas it is ultimately more economical to

source hydrogen this way than to produce it *in situ* via reforming a fraction of biomass hydrolysate or by diverting a fraction of biomass feedstock to gasification to produce hydrogen from syngas, and taking a yield reduction penalty accordingly (i.e., sourcing hydrogen from biomass which is much more costly than natural gas). Details for this trade-off are presented in Section 5.2.2. As a point of reference, the 2013 biological design case required hydrogen at a rate that equated to \$0.08/GGE contribution to overall MFSP, as hydrogen was only required during the hydrotreating step to upgrade fatty acids to RDB but not in the production of the fatty acid intermediate itself from sugars. Or, to state another way, in the 2013 biological case, oxygen from biomass carbohydrates is rejected as CO₂ inherent to biological metabolism (resulting in lower carbon retention to fuels), while in the present design, oxygen is primarily rejected as water with the addition of hydrogen (resulting in higher carbon-to-fuel yield but higher hydrogen demand). Indeed, overall fuel yield is roughly 78 GGE/ton in the present pathway associated with 85.6% carbon retention efficiency from hydrolysate, compared with 45 GGE/ton and 49.5% carbon efficiency from sugars to product in the 2013 biological case.

Beyond costs for conversion and feedstock, the next highest cost drivers are hydrolysis and hydrolysate conditioning/concentration as well as pretreatment/neutralization, each contributing a similar amount to MFSP at \$0.45/GGE and \$0.44/GGE respectively. Both process areas are driven primarily by capital expenses, in turn associated with the hydrolysate purification operations and the pretreatment reactor respectively. Similar to the 2013 case, enzyme costs are not a large burden on the overall MFSP, at the sixth highest cost driver of the nine process areas considered. This result is driven by the low enzyme loading target of 10 mg/g cellulose; alternative enzyme loadings are considered in the Sensitivity Analysis section.

While the analysis presented here demonstrates a favorable result for both fuel yields and economics, with the base case MFSP of \$4.05/GGE well on the way to the final 2022 target of \$3/GGE, it may do so at the expense of sustainability given the large amount of natural gas-derived hydrogen utilized for this case. A full well-to-wheel life cycle assessment (LCA) evaluation of the process (outside the scope of this work) must be conducted to understand the overall greenhouse gas profile of the process and associated implications for Renewable Fuel Standard (RFS) qualifications [85], but to frame the outermost boundaries for the options, the alternatives based on producing 100% of the required hydrogen internally via *in situ* reforming of hydrolysate or routing a fraction of biomass to a gasification train are considered below in Section 5.2.2. A plausible scenario could in fact fall between the boundaries, e.g., sourcing as much hydrogen as possible from external SMR-derived natural gas to still qualify for RFS criteria, while sourcing the remainder internally by an alternative route such as *in situ* or gasification. However, it is acknowledged that the purchased SMR hydrogen base case likely leaves more room for optimization if an alternative scenario were considered for an on-site integrated SMR plant, which would allow for heat/power integration opportunities as well as benefits associated with routing light gases produced elsewhere to the SMR rather than burning them in the boiler as in the present design for off-site SMR production. Considering any of the hydrogen scenarios, further improvements would be required to ultimately reach final 2022 cost goals of \$3/GGE; a brief discussion on paths toward achieving this final goal is presented in Section 5.4.

5 Analysis and Discussion

5.1 Carbon and Water Balances

Table 32 shows the overall flow of carbon inputs and outputs, with a carbon balance closure very near unity. As shown in the table, roughly 99% of all carbon in the process enters in the biomass feed, with small amounts of additional carbon coming from glucose for enzyme production. Given the low target enzyme loading (10 mg/g cellulose), the glucose contribution to total input carbon flow was only 1.3%.

Table 32. Biorefinery Overall Carbon Balance

Stream	Carbon Flow (kmol/h)	% of Carbon Flow
Carbon inlets		
Biomass feedstock	3,117	99%
Glucose	40	1%
A400 nutrients	<1	0%
Total	3,158	100%
Carbon outlets		
Diesel-range (RDB) product	1,145	36%
Naphtha-range product	267	8%
Combustion exhaust	1,629	52%
WWT brine	19	1%
Aerobic lagoons	58	2%
A400 vent	26	1%
Combustor ash	8	0%
Hydrotreater furnace flue gas	3	0%
Total	3,155	100%

Of the carbon inputs to the process, a total of 44.7% of carbon atoms leave as “total fuel products” (either naphtha- or diesel-range hydrocarbons) after hydrotreating and product distillation. The major exit point for the remainder of the carbon is primarily the combustor stack (52%). We expect the amount of carbon exiting in the combustion exhaust to be rather large because most byproducts (lignin, other unconverted sugars, etc.) of this process are burned to form CO₂ with a small amount of CO₂ also vented from the aerobic basin in the WWT section.

In addition to the carbohydrate components of the biomass (both monomeric and oligomeric sugars), other components such as furfural, HMF, solubilized lignin, and acetate may also be convertible by the catalytic process (although furfural and HMF themselves are degradation products from starting carbohydrates). Of the carbon in the feed stream, 1,857 kmol/h (59.6%) are carbohydrates (glucan, xylan, arabinan, mannan, galactan, and sucrose); 690 kmol/h (22.2%) are lignin, and the remaining 570 kmol/hr (18.2%) are the sum of acetate, extractives, and protein. Theoretically all the components with carbon could plausibly be converted to hydrocarbons using appropriate catalysts and reaction conditions. In the case of lignin, this would first require solubilizing the lignin which is possible for a fraction of lignin but unlikely to be practical for 100% of the starting lignin in the biomass feed. According to Table 32, 1,412 kmol/h of the feed stream carbon is actually converted by the catalytic

process, so the actual carbon yield is 45.3% of the biomass feedstock, without considering the carbon footprint associated with externally supplied hydrogen. When also including carbon in off-site natural gas SMR associated with the required hydrogen utilization for catalytic conversion/upgrading, overall carbon yield decreases to 35.7% based on a translation of 0.44 mol natural gas feed to SMR per mol H₂ produced [86]. The carbon retention efficiency is 85.6%, relative to carbon in the hydrolysate feed stream to the catalytic reactor system (after lignin/solids separation and hydrolysate concentration/purification). Further improvements to carbon efficiency from biomass feedstock could be achieved through improving pretreatment and enzymatic hydrolysis technologies to solubilize more carbon components, such as lignin, while minimizing loss of extractives and carbohydrates.

The overall flow of water throughout the model is presented below in Table 33. The primary point of entry for dedicated process water demand is the “makeup water” stream, shown in the current model to be 159,026 kg/h or 5.8 gal/GGE hydrocarbon product. On an absolute flowrate basis, this is similar to the 2011 ethanol report demand of 147,100 kg/h (5.4 gal/gal or 8.2 gal/GGE) but considerably lower than the 2013 biological hydrocarbon report demand of 217,100 kg/h (14.4 gal/gal or 13.7gal/GGE). This is primarily due to a lower cooling water usage and associated lower cooling tower losses (typically the largest consumer of facility process water), due in turn to a lower amount of steam sent through the steam turbine condenser (typically the largest contributor to overall cooling water demands). Specifically, total cooling water demand in the present design is 112 MMkcal/h, compared to 97 MMkcal/h in the 2011 ethanol design and 130 MMkcal/h in the 2013 biological hydrocarbon design. Furthermore, a substantial amount of water is generated in the catalytic conversion reactions of the present design model (associated with high hydrogen consumption rejecting oxygen as water), which ultimately is routed to wastewater treatment and made available for subsequent recycle to the process. Finally, on a per-GGE basis, water consumption also appears lower than the 2011 and 2013 designs given the much higher product yield in this case (approximately 78 GGE/ton in this model, versus 52 GGE/ton in the 2011 ethanol report or 45 GGE/ton in the 2013 biological hydrocarbon case).

Table 33. Biorefinery Water Balance

Water Balance					
Inputs	kg/hr	gal/GGE	Outputs	kg/hr	gal/GGE
Moisture in feedstock	20,833	0.76	Water in fuel product	5	0.00
Water in glucose syrup	214	0.01	Cooling tower evap.	158,773	5.82
Water in raw chemicals	198	0.01	Stripped in enzyme aeration	388	0.01
Generated in enzyme prodn.	648	0.00	Consumed in WWT	332	0.01
Generated in combustor	29,086	1.07	Consumed in pretreatment	2,394	0.09
Enzyme air intake	320	0.01	Consumed in enz. hydr.	2,640	0.10
Generated in APR & Condensation	26,297	0.96	WWT evap.	4,535	0.17
WWT air intake	4,595	0.17	WWT brine	3,362	0.12
Combustion air intake	6,225	0.23	Combustor stack	71,871	2.64
Makeup water	159,026	5.83	Boiler blowdown vent	2,705	0.10
			Upgrading flue gas	241	0.01
Sum of Inputs	247,443	9.08	Sum of Outputs	247,247	9.07

Similar to the 2013 case, a vacuum filter press is utilized after hydrolysis to remove lignin and other insoluble solids, which again requires a wash step to recover lost sugars; this wash step results in a lignin material with considerably higher moisture content, and thus water loss, as described previously in the 2013 report. Additional water enters the process via the feedstock and raw chemicals, both as free water and as “potential” water, i.e., the combustion product of lignin and unconverted sugars. Water is also consumed in pretreatment and hydrolysis reactions, and these are accounted for as well.

As noted above, utilities are responsible for a large majority of the water loss, with the cooling tower evaporation rate nearly matching the makeup water demand (leading to an “apparent” makeup water demand of approximately 100% allocated to cooling tower evaporation alone). The remainder of water makeup is met by the large generation during catalysis, or put differently, the required process water makeup rate would be higher if the generation during the conversion step were lower. Table 34 expresses the individual cooling water users in terms of their water loss responsibility in gal/GGE.

Table 34. Individual Contributors to Cooling Water Evaporation

Cooling Water User	MMKcal/hr	% of duty	gal/GGE
Condensing turbine	76.3	67.9%	3.96
Chiller condenser (enzyme prod.)	3.9	3.5%	0.20
Hydrolysis cooler	6.7	6.0%	0.35
Hydrolysate flash condenser	1.4	1.3%	0.07
Enzyme production compressor	0.6	0.5%	0.03
Conversion/upgrading trim coolers	2.6	2.3%	0.13
Condensation reactor effluent cooler	12.5	11.1%	0.65
Sugar evaporator condensate cooler	8.3	7.4%	0.43
Total	112.3	100.0%	5.82

The largest cooling water user is the condensing turbine, which is responsible for 3.96 gal/GGE of water demand (68% of cooling water duty). A condenser on the turbine allows the steam to discharge at a vacuum (about 0.1 atm) for maximum conversion of compressive energy to electricity. With the condensing turbine in place, the process generates a significant amount of electricity to supply all users in the plant, leaving roughly 11 MW of excess electricity exported to the grid as a coproduct. Without the turbine condenser’s ability to let the steam down to such low pressure (and temperature), the total amount of generated electricity would decrease and could potentially lead to a net power deficit that would have to be purchased from the grid. As noted above, the amount of heat that must be removed from the condensing turbine is less than in the 2013 biological design (76.3 versus 86.8 MMkcal/hr), due to higher heating steam demands for this system given the larger temperature swings in the back-end conversion steps; particularly an added demand for high-pressure steam that must be removed from the steam system prior to being sent through any of the turbine stages.

As discussed in the 2013 report, one potential possibility to reduce the biorefinery water demand could be to employ direct air-cooled condenser (ACC) units on the steam turbine exhaust. In contrast to standard fin-fan wet evaporative cooling, an ACC unit is a completely closed-circuit heat exchange system in which the turbine steam is condensed on the inside of the tubes as a result of colder ambient air flowing across the outside of the finned tube surfaces. Because the heat rejection is achieved using only air, no water is consumed in the condensing process while achieving lower process temperatures compared to standard fin-fan air coolers. Further study is warranted to confirm that replacing the

water-cooled steam turbine condenser with an air-cooled condenser (either in full or in part) to reduce consumptive water use is technically and economically feasible.

5.2 Cost Sensitivity Analysis

5.2.1 Single-Point Sensitivity Analysis

A single-point sensitivity was performed on the Aspen model using the variables and limits shown in Table 35. The baseline for all variables used in the design case is described previously in this report. Reasonable minima and maxima for each variable were chosen to understand and quantify the resulting cost impact on overall MFSP. Each variable was changed to its maximum and minimum value with all other factors held constant. The sensitivity on MFSP is displayed as a tornado chart in Figure 15.

Table 35. Assumptions Varied in the Sensitivity Analysis

	Assumption Name	Min	Baseline	Max
Pretreatment	Deacetylation xylan loss	0%	2%	10%
	PT reactor metallurgy	Stainless steel	High alloy	
	PT residence time	2	5	10
	PT acid loading (mg/g)	5	9	20
	PT temperature °C	150	158	170
	PT xylan to xylose	80%	90%	95%
	PT xylan to furfural	3%	5%	8%
	PT glucan to glucose	6%	10%	12%
	Disc refining option	Disc refining	Acid pretreatment	-
Enz Hydrolysis + Conditioning	EH % solids	17.5%	20%	25%
	EH cellulose to glucose	75%	90%	95%
	EH enzyme loading mg/g	5	10	20
	EH time (d)	2	3.5	6
	Sugar loss in S/L separation	-	1%	5%
	S/L separation capex	-50%	-	+50%
	Belt filter replacement	-	Included in maintenance factor	Replace every other year
	Microfilter recycle purge	0%	50%	100%
	Microfilter retentate loss	5%	10%	15%
Ion exchange capital	-25%	-	+50%	
Enz Production	A400 capex	-50%	-	+50%
Catalytic Upgrading	Catalytic upgrading capital	-50%	-	+100%
	Carbon yield from sugars	0.7	0.86	-
	WHSV-APR-1 (hydrogenation)	0.7	1.2	2
	WHSV-APR-2 (APR reactor)	0.7	1	2
	WHSV-condensation	0.7	1.2	2
	Total H2 feed molar ratio	6	9.8	15
	H2 price (\$/kg)	1.1	1.57	2
	Catalyst price	-20%	-	+20%
Capital	Total capital investment	-25%	-	+25%

The metrics of carbon efficiency from hydrolysate, uncertainty in total capital investment resulting from the factored approach we used ($\pm 25\%$), as well as the quoted catalytic conversion capital cost uncertainty range, have the largest impact on MFSP; additionally, glucose yield from hydrolysis and hydrogen purchase price also weigh heavily on costs. The critical role that hydrogen and the associated hydrogen price plays is noteworthy to emphasize, given the large amount of hydrogen import required in the base case. While the single-point sensitivity to hydrogen price is intended to capture a reasonable range of expected costs in the context of U.S. natural gas price fluctuations over recent years (i.e., \$2–\$7.4/MM BTU as discussed previously), hydrogen price variances are further expanded over a larger range of values as shown in Figure 16 to consider potential implications for higher natural gas prices. The upper end of prices shown in Figure 16 corresponds to other countries (such as Japan) with significantly higher natural gas prices (\$14–17/MM BTU) than the United States [87], which has fluctuated closer to \$4/MM BTU in recent years [88]. The associated correlation between hydrogen prices and natural gas prices is based on the above-cited H2A report for current SMR technology [75].

Beyond drivers associated with capital costs, hydrogen cost, and overall carbon yields to fuels, cellulose-to-monomer glucose yield and enzyme loading play a less critical role in overall cost impacts on a cost-per-GGE basis than in the 2013 biological hydrocarbon report, but still carry a higher cost impact over the remainder of the parameters evaluated here. Additionally, pretreatment and hydrolysis operations may have room to be further tailored to produce more oligomeric sugars and less monomers (leading to potential cost savings for deconstruction operations) given the ability of the catalyst to also convert oligomer components beyond strictly monomers. This scenario was not considered here, but would be an alternative consideration to the “cellulose-to-monomeric glucose” evaluation. Conversion parameters factor heavily into production cost, because these parameters not only affect product final yield, but also affect capital and other operating costs. We note that this is an imperfect comparison because the TCI and other cost-related bars represent an actual uncertainty while the other process-related bars are more accurately called a risk, in the sense that they represent a range of results that might be expected in future pilot-scale demonstration runs.

The majority of the remaining parameters are self-explanatory; however, a brief discussion is warranted on the inclusion of “pretreatment disc refining” as an alternative option. In recent years, NREL has begun investigating mechanical disc refining as a means to improve digestibility while maintaining high process yields [21, 89]. Mechanical refining is a relatively established technology that historically has been used in the pulp and paper industry to improve fiber strength but also has been found to be an attractive approach to biomass deconstruction processes as a means to increase accessibility of the cellulose in biomass and improve process sugar yields [90]. More recently, a new process technology has been investigated using deacetylation followed by disc refining only, eliminating the acid pretreatment step. In this process, biomass is first subjected to a deacetylation step where about 70%–80% of the acetyl groups in the biomass are removed by a dilute alkali solution heated to approximately 80°C. The deacetylated biomass is mechanically refined through a single stage of disc refining. The refined biomass is then hydrolyzed using cellulase and hemicellulase enzymes. The process leaves relatively high purity sugar syrups that can then be biologically or catalytically upgraded to produce biofuels and chemicals; this may be advantageous in the present pathway as it avoids introduction of acid and caustic (neutralization) chemicals, thus reducing ammonium salt formation and subsequent removal.

As a cursory feasibility analysis, disc refining (specifically deacetylation and disc refining or DDR) is considered here as an alternative to acid pretreatment. For a 2,000 dry ton biomass facility, four 54-inch double disc refiners were quoted with motor, surge bins, and feed and discharge conveyors. The selected disc refiners are 54 inches in diameter with a maximum 5,000 connected horsepower and a no-load power demand of 453 hp each. Based on NREL test data for corn stover refining, we assumed a specific refining energy of 200 KWh/bone dry ton of biomass. Thus, while minimizing operating expenses by way of acid and caustic neutralization chemicals relative to the dilute-acid pretreatment base case, the alternative utilizing disc refining requires considerably more electrical energy. No change in cellulose or hemicellulose sugar yields were assumed, thus this analysis strictly considers trade-offs in capital and operating cost estimates. Although cost impact for replacing dilute-acid pretreatment with disc refining (Figure 15) was not found to be significant, additional benefits may be seen that were not included here; namely, production of “cleaner” sugars with lower ammonium salt levels (thus likely lower purification demands through ion exchange in this pathway), as well as expected lower production levels of components known to be fermentation inhibitors, such as furfurals and acetic acid [90-92] (likely a benefit for biological conversion pathways, although such species are assumed to be convertible in the present catalytic design).

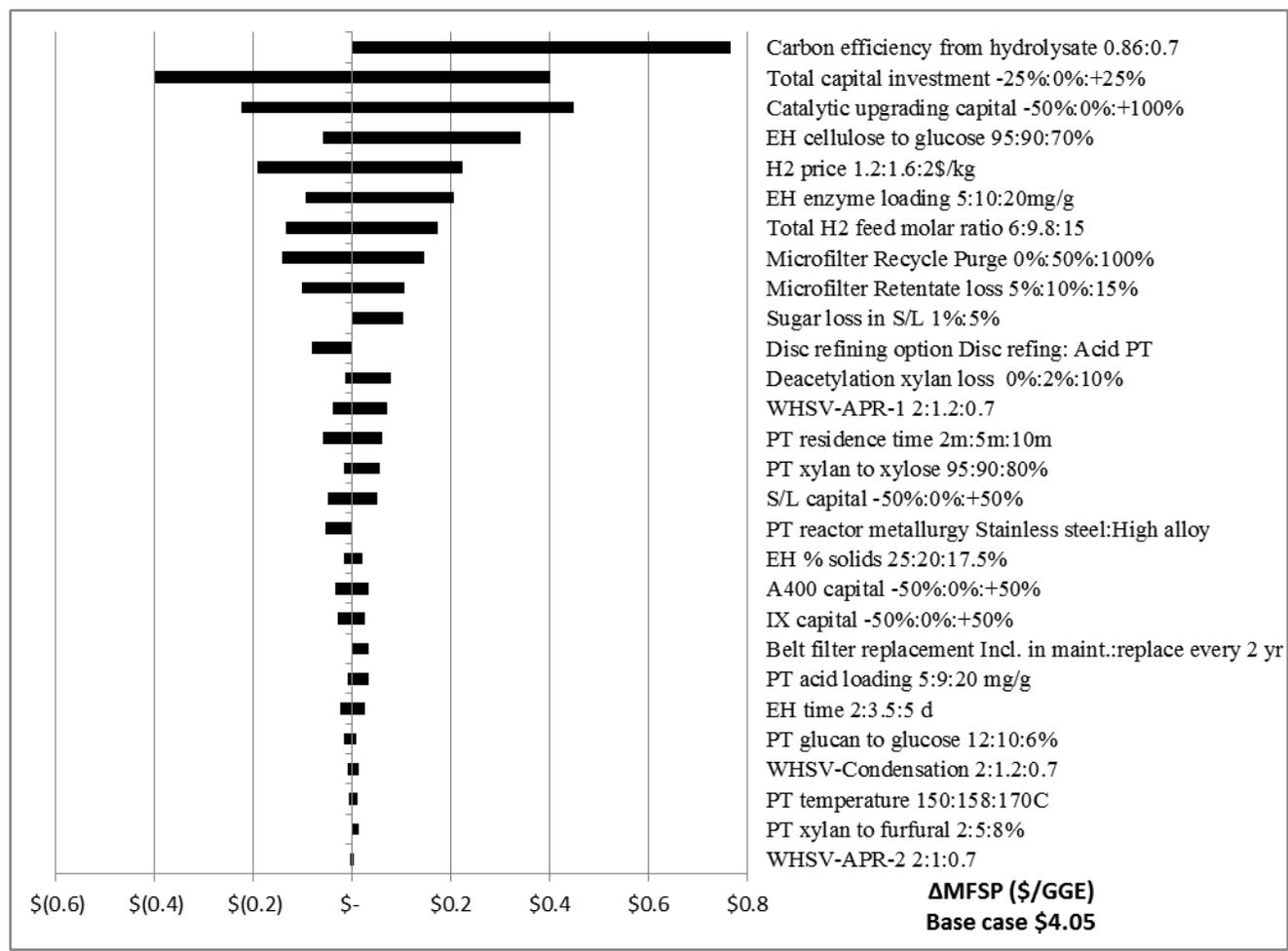


Figure 15. Single-point sensitivity tornado chart for purchased hydrogen base case

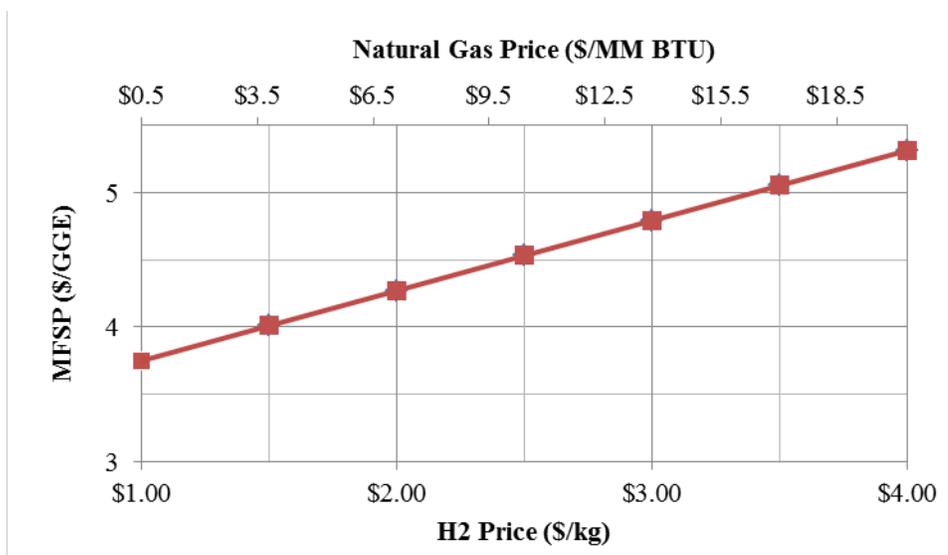


Figure 16. MFSP sensitivity scan as a function of hydrogen price and correlated SMR natural gas prices

5.2.2 Evaluation of Alternative Hydrogen Sourcing Options

The carbon yield from hydrolysate through finished fuels in the base case design is very high at 85.6%, attributed to the use of externally-supplied SMR hydrogen. However, as mentioned previously, the greenhouse gas and fossil energy penalties attributed to natural gas SMR hydrogen production may detrimentally impact the overall life cycle of the process, given the resulting amount of hydrogen import required. There are a large number of alternative methods to source hydrogen from more sustainable sources including electrolysis of water, solar-thermal and photoelectrochemical water splitting, high-temperature electrolysis coupled with nuclear energy, or biological pathways [93]; while most of these are expected to increase hydrogen costs considerably over standard SMR production (today), they are noteworthy to mention as future opportunities for improving sustainability for a large hydrogen consuming process such as this. Additionally, two other options also exist that may be more sensible in this context by offering the potential to leverage the biorefinery infrastructure, namely *in situ* hydrogen production from the produced sugars, or gasification of a fraction of biomass feed. Both of these approaches tap into the potential to use a fraction of the biomass feedstock resource otherwise devoted entirely to fuel production in the base case purchased hydrogen scenario. To preserve consistency across modeled hydrogen scenarios, both alternative cases considered here maintain a total biomass feed rate to the biorefinery of 2,000 dry tonne/day.

In Situ Hydrogen Production

In situ hydrogen production offers the potential to improve sustainability over natural gas-derived hydrogen, but at the cost of fuel yield. While external SMR-based hydrogen is considered for the base case, the *in situ* hydrogen approach is presented here as one alternative to frame the issue from a high-level cost and sustainability standpoint. Information from patent literature has described alternative systems to produce hydrogen *in situ* from aqueous phase reforming of biomass hydrolysates [32, 33]; a key advantage of APR is that it enables flexibility in choosing the source of hydrogen supplied to the plant.

Early published work on APR focused on generation of hydrogen from renewable biomass feedstocks with the implication that it would be a more viable process than electrolysis of water, gasification, steam-reforming of bio-oils, or enzymatic processes [31]. The initial focus was to produce hydrogen from biomass-derived material for use in fuel cells [94]. Information from patent literature has suggested that hydrogen produced by APR (“APR-H₂”) can also be integrated into a catalytic sugar reforming process like the base case model in this report [32, 33]. The selectivity of APR reactions to hydrogen can be improved by using reduced substrates like sugar alcohols (for example, the basic reaction from sorbitol: C₆H₁₄O₆ + 6 H₂O ↔ 13 H₂ + 6 CO₂) [31] and therefore APR-H₂ could use a stream diverted from the hydrogenation reactor feed (i.e., first considering reduction of glucose to sorbitol: C₆H₁₂O₆ + H₂ → C₆H₁₄O₆) [31, 95-98]. *In situ* hydrogen by APR has other potential advantages over SMR. APR-H₂ production takes place at lower temperature (less than 400°C) versus SMR which takes place at 400°C to 700°C [94]. SMR requires steam whereas APR-H₂ production takes place in the liquid phase eliminating the energy penalty due the heat of vaporization of water.

A fraction of the hydrolysate after enzymatic hydrolysis, sugar concentration, and cleanup is used to produce hydrogen for sugar catalytic conversion to fuels. Of course, the more hydrolysate is diverted to produce hydrogen, the lower the amount of hydrogen then required to satisfy the decreased amount of hydrocarbons produced from the lower remaining amount of available sugars (carbon), leading to an optimum split. In order to provide the required hydrogen demands for hydrogenation/APR, condensation, and hydrotreating reactions to hydrocarbon fuels, approximately 41% of the initial purified hydrolysate material is required, thereby reducing carbon and overall fuel yields accordingly by a similar fraction relative to the purchased hydrogen base case (shown further below in Table 38). This yield optimization is based on assuming stoichiometric conversion of all pertinent hydrolysate species to hydrogen and CO₂ using simplified reforming reactions as shown in Table 36; thus, this presents a likely “best-case” scenario based on complete conversion of all hydrolysate components and does not consider possible subsequent recombination of H₂ and CO₂ to alkanes and water or other hydrogen yield losses [31].

Table 36. Governing Stoichiometry Assumed for Components in Hydrolysate to *in situ* Hydrogen Production (“APR H₂” Train)

Reaction
GLUCOSE + 6 H ₂ O --> 12 H ₂ + 6 CO ₂
XYLOSE + 5 H ₂ O --> 10 H ₂ + 5 CO ₂
GALACTOSE + 6 H ₂ O --> 12 H ₂ + 6 CO ₂
MANNOSE + 6 H ₂ O --> 12 H ₂ + 6 CO ₂
ARABINOSE + 5 H ₂ O --> 10 H ₂ + 5 CO ₂
GLUCOSE OLIGOSACCHARIDE + 7 H ₂ O --> 12 H ₂ + 6 CO ₂
XYLOSE OLIGOSACCHARIDE + 6 H ₂ O --> 10 H ₂ + 5 CO ₂
GALACTOSE OLIGOSACCHARIDE + 7 H ₂ O --> 12 H ₂ + 6 CO ₂
MANNOSE OLIGOSACCHARIDE + 7 H ₂ O --> 12 H ₂ + 6 CO ₂
ARABINOSE OLIGOSACCHARIDE + 6 H ₂ O --> 10 H ₂ + 5 CO ₂
CELLOBIOSE + 13 H ₂ O --> 24 H ₂ + 12 CO ₂
SOLUBLE LIGNIN + 13 H ₂ O --> 17 H ₂ + 8 CO ₂
HMF + 9 H ₂ O --> 12 H ₂ + 6 CO ₂
FURFURAL + 8 H ₂ O --> 10 H ₂ + 5 CO ₂

To model the *in situ* process train, the diverted hydrolysate stream is first heated to near-reaction temperature in a feed-effluent heat exchanger (by cross-exchange with the APR-H₂ reactor effluent), then heated further using high pressure steam from the A800 steam system. The reaction takes place at 227°C and 30 atm (420 psig) [31], and the reactor product stream is then further cooled to 30°C to knock out water (routed to wastewater treatment). Hydrogen is purified by a PSA system to remove CO₂, then compressed to 1,050 psig and routed to the hydrocarbon catalytic reactor train. The design and cost estimates for the PSA unit were adopted from the 2013 biological design report utilizing zeolite-based sorbent with 95% hydrogen recovery. The PSA tailgas, containing mainly CO₂ with 5% H₂ loss, is combined with the purge gas stream from the hydrocarbon catalytic train and routed to the boiler (A800). The APR-H₂ reactor capital cost and operating expenses for annualized catalyst cost (and associated catalyst composition) were set consistently with the hydrocarbon train APR reactor (“APR-2,” i.e., the second of the three individual conversion reactors quoted by Harris Group and the vendor).

Biomass Gasification Hydrogen Production

The second alternative hydrogen sourcing option considered is to gasify a fraction of biomass supplied to the biorefinery in a separate gasification process train operated in parallel to the primary sugar/fuel train which otherwise maintains all process assumptions consistently with the purchased hydrogen base case. In this scenario, the overall biorefinery feedstock rate remains fixed at 2,000 dry tonne/day, but the “split” towards hydrogen production occurs further upstream in the process (i.e. at the biorefinery plant gate) relative to the *in situ* pathway which instead takes the split from the produced sugar stream. However, another difference is that the gasification process is more favorable towards woody biomass feedstocks as the standard for design rather than herbaceous feedstocks such as corn stover, due to lower ash and higher carbon content for woody feedstocks as preferential for thermochemical processes such as gasification [68]. As such, the gasification scenario employs a dual-feedstock basis delivered to the biorefinery, with the same herbaceous blend utilized for the sugar/fuel train (see Section 2.2) and a woody feedstock blend utilized for the gasification/hydrogen train, with the sum of the two feedstocks fixed at 2,000 dry tonne/day.

The gasification train model was constructed based primarily on the process and design assumptions as documented in NREL’s 2011 thermochemical ethanol design report [68]. This includes the use of indirect steam gasification coupled with a fluidized bed char combustor with olivine circulating between the two operations as a heat transfer medium, followed by syngas cleanup based on catalytic tar reforming and then cooling, quench, and scrubbing of the syngas to remove impurities such as particulates and ammonia. The tar reformer is designed similar to a small-scale fluidized catalytic cracker (FCC) unit, whereby tars, methane, and light hydrocarbons are reformed to additional syngas over a catalyst, with continuous regeneration of the catalyst by combustion of the deposited coke. The syngas is then compressed in a multi-stage compressor to a pressure of 29 atm (430 psig). From this point, a number of additional unit operations were added for hydrogen production and recovery which were not part of the thermochemical ethanol design, but rather were based on the biomass gasification-to-hydrogen design report published by NREL in 2005 [34]; this includes removing sulfur species in a zinc oxide (ZnO) guard bed, increasing hydrogen yield through catalytic high-temperature shift (HTS) and low-temperature shift (LTS) reactors (i.e. following the water gas shift reaction $H_2O + CO \rightarrow H_2 + CO_2$), and purifying the hydrogen via a PSA unit. All capital and operating costs for the process are based on the above-referenced NREL thermochemical ethanol and hydrogen reports for the respective unit operations. The reader is referred to these reports for more information on process and design

details beyond the brief high-level summary provided here. Key operational and yield parameters for the gasification/hydrogen production train are summarized in Table 37.

Table 37: Key Operational and Yield Parameters for Gasification/Hydrogen Process Train

<i>Overall Modeling Parameters</i>	<i>Value</i>
Delivered feedstock moisture content	10%
Gasifier temperature	870 °C
Char combustor temperature	981 °C
Tar reformer temperature	910 °C
Tar reformer catalyst type	Ni/Mg/K/Al ₂ O ₃
H ₂ :CO molar ratio across tar reformer	0.57 (in) – 1.61 (out)
<i>Tar reformer component conversions:</i>	
Methane (CH ₄)	80%
Ethane (C ₂ H ₆)	99%
Ethylene (C ₂ H ₄)	90%
Tars (C ₁₀₊)	99%
Benzene (C ₆ H ₆)	99%
Ammonia (NH ₃)	90%
Syngas compressor outlet pressure	29.3 atm
H ₂ S concentration across ZnO guard bed	414 ppm (in) – 0 ppm (out)
HTS temperature	468 °C
LTS temperature	285 °C
Catalyst used in shift reactors	Copper-zinc oxide
PSA H ₂ recovery	85%
H ₂ product stream purity	99.99%
Overall hydrogen yield across H₂ process train	82.0 kg/dry tonne biomass

Beyond unit operation design and cost assumptions taken from the above-referenced thermochemical ethanol and biomass-to-hydrogen design reports, assumptions pertaining to woody feedstock delivery to the gasifier train were updated to maintain consistency with NREL’s most recent modeling efforts making use of this feedstock blend, namely NREL’s 2014 indirect gasification of biomass-to-high octane gasoline design report [99]. Based on recent updates from partners at INL, the feedstock delivery details documented in this new 2014 report stipulate a feedstock cost of \$80/dry ton delivered to the throat of the gasifier system, associated with a delivered moisture content of 10% after upstream preprocessing; at this moisture content, the new 2014 design also makes use of a new cross-flow heater prior to the gasifier in place of a biomass dryer that had been used in prior designs to accommodate higher-moisture feedstocks. Similar to the herbaceous feedstock blend on the biochemical process train (discussed above in Section 3.1), the woody feed to the gasifier process train is also based on a blending strategy that combines pulpwood, wood residues, construction and demolition waste, and switchgrass [100]. Based on the modeled hydrogen yield of 82.0 kg/dry metric tonne biomass to the gasifier (Table 37), this scenario requires allocating 720 dry tonne/day of woody biomass to the gasifier train (36% the total 2,000 dry tonne/day available for the overall system) to produce the required amount of hydrogen, leaving 1,280 dry tonne/day of herbaceous feedstock available for the primary sugar/fuel process train, thereby reducing fuel yields by 36% relative to the purchased hydrogen base case; see Table 38.

Model Results

Results for fuel and carbon yields for the *in situ* and gasification scenarios for hydrogen sourcing are presented in Table 38, alongside the purchased hydrogen base case.

Table 38. Comparison of Fuel and Carbon Yields between Base Design and Alternative Hydrogen Cases

	Base Case	<i>In situ</i> Hydrogen	Biomass Gasification
Hydrogen source	purchased	41% split of hydrolysate feed	36% split of total biomass feed
Fuel yield (GGE/ton total biomass)	78.3	45.3	50.1
Carbon efficiency to fuels from total hydrolysate ^a	85.6%	49.5%	85.6%
Carbon efficiency to fuels from total biomass ^a	45.3%	26.2%	27.7%

^a Note, calculations do not include carbon from natural gas for purchased hydrogen case (reduces carbon yield down to 35.7%); additionally, calculations are based on total starting hydrolysate carbon (and biomass); actual carbon efficiency of the *in situ* hydrocarbon train from the fraction of hydrolysate sent to this train is equal to the purchased H₂ case.

As over 41% of the hydrolysate stream is diverted to produce hydrogen for the *in situ* pathway, carbon retention efficiency from total hydrolysate carbon is reduced from 85.6% to 49.5%, and the carbon efficiency from biomass feedstock is reduced from 45.3% to 26.2% (this only considers total initial carbon in hydrolysate and biomass respectively, excluding carbon in natural gas used to produce H₂ in the “purchased” scenario). Correspondingly, total fuel yield decreases from 78.3 GGE/ton in the *ex situ* base case to 45.3 GGE/ton in the *in situ* scenario. While this saves on purchased hydrogen cost, the yield penalty outweighs this benefit and results in an increase in MFSP from \$4.05 to \$5.48/GGE, shown in Figure 17. Likewise, the gasification scenario requires 36% of the biomass originally available for the *ex situ* base case, reducing fuel yield to 50.1 GGE/ton of total biomass which corresponds to a carbon efficiency from total biomass to fuels of 27.7% (including biomass diverted to the gasifier train). Although the gasification case is estimated to be slightly more capital cost-intensive than the *in situ* case (\$345 MM versus \$331 MM total installed costs), the higher fuel yield for this scenario ultimately translates to a roughly 10% lower MFSP at \$4.95/GGE, although still markedly higher than the \$4.05/GGE *ex situ* base case. Put differently, both of these alternative hydrogen cases replace hydrogen sourced from low-cost natural gas with much higher-cost biomass. However, as noted above, a key rationale for such alternative scenarios is the ensuing sustainability benefits for eliminating natural gas SMR demands. While a full well-to-wheel LCA of the system is beyond the scope of this analysis, sustainability metrics associated strictly with the modeled biorefinery conversion process are reported in the next section for all three hydrogen scenarios. Cost breakdowns for all three scenarios are provided in Figure 17, with more details for the two alternate cases shown in Figures 18-19.

It is noteworthy that the *in situ* scenario coincides closely with NREL’s 2013 biological hydrocarbon design case, both in terms of cost and yields. Namely, the 2013 design basis presented modeled fuel (RDB) yields of 45.4 GGE/ton and 26.2% carbon retention efficiency from biomass to finished fuels [8] compared to 45.3 GGE/ton and 26.2% carbon retention efficiency from biomass carbon in the present *in situ* case. Likewise, the MFSP for the 2013 design was \$5.10/GGE, compared to \$5.48/GGE in the *in situ* case here. **Thus, the case for *in situ* hydrogen production coupled with catalytic conversion of remaining hydrolysate to fuels presents a more comparable basis relative to the 2013 biological conversion base case,** both in terms of TEA results and the process concept itself being self-sufficient in how the fuel product is produced across the conversion step.

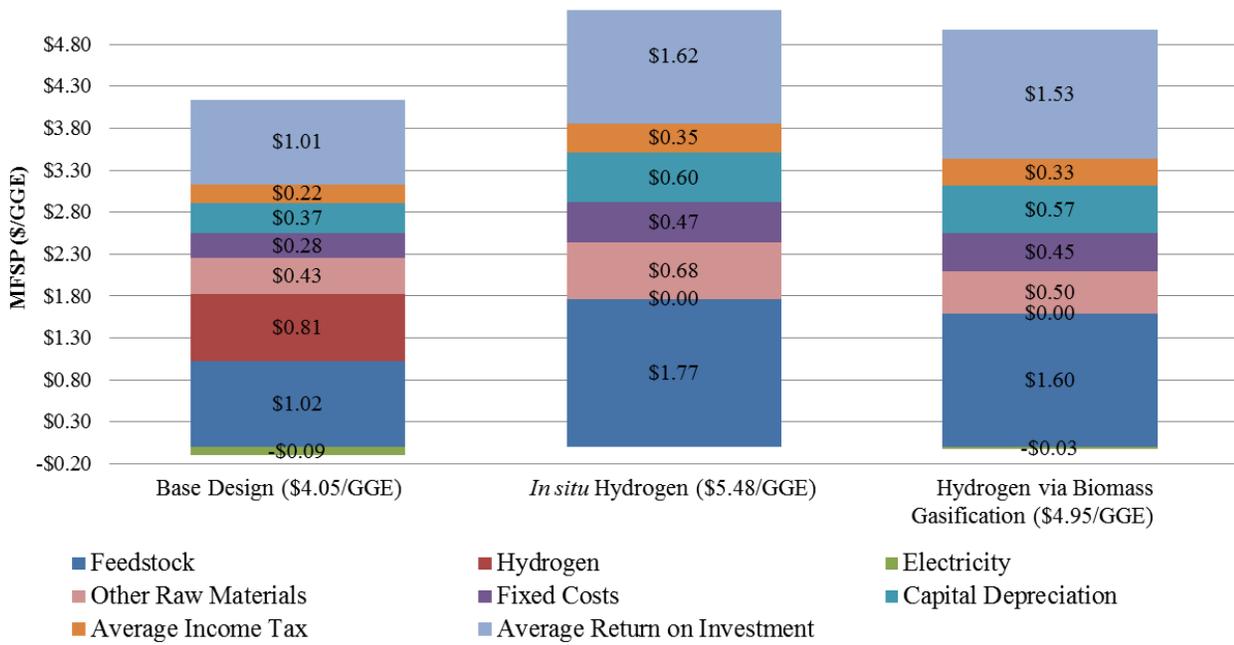


Figure 17. Cost comparison of the *ex situ* hydrogen base case with *in situ* and gasification hydrogen alternatives

Biomass Derived Hydrolysate Catalysis to Hydrocarbons Process Engineering Analysis

Dilute Acid Pretreatment, Enzymatic Hydrolysis, Hydrolysate Catalytic Conversion/Upgrading, *in situ* Hydrogen
All Values in 2011\$

Minimum Fuel Selling Price (MFSP): \$5.48 /GGE

Contributions:	Feedstock	\$1.77 /GGE
	Enzymes	\$0.37 /GGE
	Non-Enzyme Conversion	\$3.34 /GGE
	HC Production	32.8 MM GGE/yr
	HC Yield	45.3 GGE/dry U.S. ton feedstock
Carbon Retention Efficiency for Hydrolysate Catalysis		0.495 carbon in HC product/carbon in sugars
	Feedstock + Handling Cost	\$80.00 /dry U.S. ton feedstock
	Internal Rate of Return (After-Tax)	10%
	Equity Percent of Total Investment	40%

Capital Costs		Manufacturing Costs (cents/GGE)	
Pretreatment	\$51,400,000	Feedstock + Handling	176.6
Neutralization/Conditioning	\$2,200,000	Sulfuric Acid	5.9
Enzymatic Hydrolysis, Hydrolysate Conditioning & Purification	\$68,800,000	Ammonia (pretreatment conditioning)	4.1
On-site Enzyme Production	\$12,400,000	Caustic (deacetylation pretreatment)	6.2
Catalytic Conversion & Upgrading	\$71,300,000	Glucose (enzyme production)	20.8
Wastewater Treatment	\$45,400,000	Hydrogen	0.0
Storage	\$3,200,000	Catalysts for Conversion Reactors	11.6
Boiler/Turbogenerator	\$70,100,000	Other Raw Materials	14.8
Utilities	\$6,500,000	Waste Disposal	4.2
Total Installed Equipment Cost	\$331,300,000	Net Electricity	0.1
		Fixed Costs	47.1
Added Direct + Indirect Costs	\$287,600,000	Capital Depreciation	59.7
(% of TCI)	46%	Average Income Tax	34.7
		Average Return on Investment	162.0
Total Capital Investment (TCI)	\$618,900,000		
		Manufacturing Costs (\$/yr)	
Installed Equipment Cost/Annual Gallon	\$10.10	Feedstock + Handling	\$57,900,000
Total Capital Investment/Annual Gallon	\$18.86	Sulfuric Acid	\$1,900,000
		Ammonia (pretreatment conditioning)	\$1,300,000
Loan Rate	8.0%	Caustic (deacetylation pretreatment)	\$2,000,000
Term (years)	10	Glucose (enzyme production)	\$6,800,000
Capital Charge Factor (computed)	0.136	Hydrogen	\$0
		Catalysts for Conversion Reactors	\$3,800,000
Carbon Retention Efficiencies:		Other Raw Materials	\$4,900,000
From Sugar Catalytic Reaction (Fuel C / Sugar C)	49.5%	Waste Disposal	\$1,400,000
From Biomass (Fuel C / Biomass C)	26.2%	Net Electricity	\$0
		Fixed Costs	\$15,500,000
		Capital Depreciation	\$19,600,000
		Average Income Tax	\$11,400,000
		Average Return on Investment	\$53,100,000
		Specific Operating Conditions	
		Hydrolysate diversion to H2 production train	41.3%
		Purchased Electricity (KWh/GGE)	0.01
		Plant Electricity Use (KWh/GGE)	8
		Enzyme Loading (mg/g cell)	10
		Saccharification Time (days)	3.5

Figure 18: Economic and process summary results for *in situ* H₂ scenario

Biomass Derived Hydrolysate Catalysis to Hydrocarbons Process Engineering Analysis

Dilute Acid Pretreatment, Enzymatic Hydrolysis, Hydrolysate Catalytic Conversion/Upgrading, Hydrogen via Biomass Gasification

All Values in 2011\$

Minimum Fuel Selling Price (MFSP): \$4.95 /GGE

Contributions:	Feedstock	\$1.60 /GGE
	Enzymes	\$0.22 /GGE
	Non-Enzyme Conversion	\$3.13 /GGE
	HC Production	36.3 MM GGE/yr
	HC Yield	50.1 GGE/dry U.S. ton feedstock
Carbon Retention Efficiency for Hydrolysate Catalysis		0.856 carbon in HC product/carbon in sugars
Feedstock + Handling Cost		\$80.00 /dry U.S. ton feedstock
Internal Rate of Return (After-Tax)		10%
Equity Percent of Total Investment		40%

Capital Costs	
Pretreatment	\$39,100,000
Neutralization/Conditioning	\$1,700,000
Enzymatic Hydrolysis, Hydrolysate Conditioning & Purification	\$46,500,000
On-site Enzyme Production	\$8,400,000
Catalytic Conversion & Upgrading	\$64,500,000
Gasification to Hydrogen	\$69,600,000
Wastewater Treatment	\$36,000,000
Storage	\$3,600,000
Boiler/Turbogenerator	\$69,400,000
Utilities	\$6,300,000
Total Installed Equipment Cost	\$345,100,000
Added Direct + Indirect Costs (% of TCI)	\$304,200,000 47%
Total Capital Investment (TCI)	\$649,300,000
Installed Equipment Cost/Annual Gallon	\$9.51
Total Capital Investment/Annual Gallon	\$17.88
Loan Rate	8.0%
Term (years)	10
Capital Charge Factor (computed)	0.136
Carbon Retention Efficiencies:	
From Sugar Catalytic Reaction (Fuel C / Sugar C)	85.6%
From Biomass (Fuel C / Biomass C)	27.7%

Manufacturing Costs (cents/GGE)	
Feedstock + Handling	159.61
Sulfuric Acid	4.66
Ammonia (pretreatment conditioning)	2.37
Caustic (deacetylation pretreatment)	3.59
Glucose (enzyme production)	12.00
Hydrogen	0.00
Catalyst/Chemical Costs for Gasification Train	4.19
Catalysts for Conversion Reactors	10.93
Other Raw Materials	9.62
Waste Disposal	2.44
Net Electricity	-2.65
Fixed Costs	45.11
Capital Depreciation	56.74
Average Income Tax	32.80
Average Return on Investment	153.28

Manufacturing Costs (\$/yr)	
Feedstock + Handling	\$57,900,000
Sulfuric Acid	\$1,700,000
Ammonia (pretreatment conditioning)	\$900,000
Caustic (deacetylation pretreatment)	\$1,300,000
Glucose (enzyme production)	\$4,400,000
Hydrogen	\$0
Catalyst/Chemical Costs for Gasification Train	\$1,500,000
Catalysts for Conversion Reactors	\$4,000,000
Other Raw Materials	\$3,500,000
Waste Disposal	\$900,000
Net Electricity	-\$1,000,000
Fixed Costs	\$16,400,000
Capital Depreciation	\$20,600,000
Average Income Tax	\$11,900,000
Average Return on Investment	\$55,600,000

Specific Operating Conditions	
Total H2 Reaction Consumption (wt% of hydrolysate)	6.5%
Excess Electricity (KWh/GGE)	0.5
Plant Electricity Use (KWh/GGE)	8.8
Enzyme Loading (mg/g cell)	10
Saccharification Time (days)	3.5

Figure 19: Economic and process summary results for H₂ via biomass gasification

5.3 Sustainability Metrics

An important aspect of evaluating biofuel processes is the quantification of life cycle resource consumption and environmental emissions. Life cycle assessment (LCA) provides a framework from which the environmental sustainability of a given process may be quantified and assessed. This section presents the salient sustainability metrics of the current conceptual process at the conversion stage. Direct biorefinery emissions (i.e., CO₂, NO₂ and SO₂), water consumption, and other process-related metrics were derived from the conversion process model described above.

SimaPro v.8.0.2 software [101] was used to develop and link units quantifying life cycle impacts as previously documented by Hsu et al. [102]. Greenhouse gas (GHG) basis values for purchased hydrogen (derived via natural gas steam methane reforming) 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 11.2 kg CO_{2e}/kg H₂ and 0.65 kg CO_{2e}/KWh purchased electricity [103]. Likewise, the basis from GREET was also applied for non-renewable (fossil) consumption values of 183 MJ/kg H₂ and 7.46 MJ/KWh purchased electricity [103]. All other process input/output inventory estimates were derived from Ecoinvent v.2.2 [104] and the U.S. Life Cycle Inventory (LCI) [105] processes to fill the data gaps, similar to the methods applied in NREL's 2013 biological conversion design report. The Ecoinvent processes were modified to reflect U.S. conditions and the U.S. LCI processes were adapted to account for embodied emissions and fossil energy usage. The material and energy flows of the conversion step capture the impacts of input raw materials, and outputs, such as emissions, wastes, and coproducts as predicted by the process model, and are shown in Table 39. This table and subsequent discussion of sustainability metrics consider the base case model utilizing purchased hydrogen and the alternative cases with internal hydrogen generation via *in situ* and gasification routes. Additionally, the LCI data presented below in Table 39 is translated to an average hourly basis, i.e., dividing the annual rates by the operating factor of 7,884 hours/year. In some cases the hourly rates for certain items, such as catalyst usage, do not reflect an actual hourly consumption but rather are amortized as such for purposes of evaluating the resulting sustainability metrics on a consistent basis.

Table 39. Input and Output Data Associated with the Modeled Conversion Facility

	<i>Purchased Hydrogen</i>		<i>In Situ Hydrogen</i>		<i>Gasification Hydrogen</i>	
Products	Production Rate		Production Rate		Production Rate	
	kg/hr	MJ/hr (LHV)	kg/hr	MJ/hr (LHV)	kg/hr	MJ/hr (LHV)
Total Hydrocarbon Fuel	20,047	874,291	11,601	505,353	12,833	559,715
Electricity Export (Import)	11,045	KW	(914)	KW	1,588	KW
Resource Consumption ^a	Flow Rate	Per MJ Fuel	Flow Rate	Per MJ Fuel	Flow Rate	Per MJ Fuel
	kg/hr	kg/MJ	kg/hr	kg/MJ	kg/hr	kg/MJ
Biomass Feedstock	104,167	1.2E-01	104,167	2.1E-01	99,998	2.0E-01
Sulfuric Acid (93%)	2,240	2.6E-03	2,240	4.4E-03	1,950	3.5E-03
Caustic (NaOH as pure)	1,406	1.6E-03	1,406	2.8E-03	900	1.6E-03
Ammonia	310	3.5E-04	310	6.1E-04	198	3.5E-04
HCl for IX Regeneration	1,122	1.3E-03	1,122	2.2E-03	718	1.3E-03
Caustic for IX regeneration	640	7.3E-04	640	1.3E-03	409	7.3E-04
Imported Hydrogen	3,847	4.4E-03	NA ^c	NA ^c	NA ^c	NA ^c
Total APR Catalysts	8	9.5E-06	6	1.2E-05	5	8.9E-06
Condensation Catalyst	3	3.0E-06	2	4.0E-06	2	3.6E-06
Polymer	2	2.1E-06	2	3.6E-06	1	1.8E-06
Ammonia (WWT)	109	1.2E-04	128	2.5E-04	104	1.9E-04
Boiler Chemicals	2.3E-01	2.6E-07	2.3E-01	4.5E-07	2.3E-01	4.0E-07
FGD Lime	180	2.1E-04	184	3.6E-04	115	2.1E-04
Cooling Tower Chemicals	2	2.8E-06	2	4.8E-06	2	4.3E-06
Makeup Water	159,026	1.6E-01	154,837	3.1E-01	199,495	3.8E-01
Air Demand	568,809	0.7	542,388	1.1	472,942	8.4E-01
Glucose	1,213	1.4E-03	1,213	2.4E-03	776	1.4E-03
Corn Steep Liquor	83	9.4E-05	83	1.6E-04	53	9.4E-05
Corn Oil	7	7.7E-06	7	1.3E-05	4	7.7E-06
Host Nutrients	34	3.9E-05	34	6.7E-05	22	3.9E-05
Sulfur Dioxide	8	9.4E-06	8	1.6E-05	5	9.4E-06
Ammonia	58	6.6E-05	58	6.6E-05	37	6.6E-05
Gasifier Train Chemicals	NA	NA	NA	NA	95	1.7E-04
Air Emissions	kg/hr	kg/MJ	kg/hr	kg/MJ	kg/hr	kg/MJ
Water (H ₂ O)	260,437	3.0E-01	230,219	4.6E-01	232,737	4.2E-01
Nitrogen (N ₂)	429,467	4.9E-01	421,070	8.3E-01	362,830	6.5E-01
Oxygen (O ₂)	66,379	8.5E-02	64,369	1.3E-01	46,721	8.3E-02
Carbon Dioxide (CO ₂) ^b	74,741	7.6E-02	101,595	2.0E-01	103,914	1.9E-01
Methane (CH ₄)	1.6	1.8E-06	1.4	2.7E-06	1.4	2.5E-06
Nitrogen Dioxide (NO ₂)	53	6.0E-05	53	1.0E-04	42	7.5E-05
Carbon Monoxide (CO)	53	6.0E-05	53	1.0E-04	34	6.0E-05
Sulfur Dioxide (SO ₂)	11	1.2E-05	11	2.1E-05	12	2.1E-05
Waste Streams	kg/hr	kg/MJ	kg/hr	kg/MJ	kg/hr	kg/MJ
Disposal of Ash/Spent Catalyst	4,506	5.2E-03	4,512	8.9E-03	3,442	6.2E-03
Wastewater (brine)	6,720	7.7E-03	6,356	1.3E-02	4,305	7.7E-03

^a See discussion above for details on chemical inputs listed here.

^b Direct carbon dioxide (CO₂) emission is 100% biogenic CO₂.

^c Hydrogen provided internally via *in situ* or gasification generation.

Details of contributions to GHG emissions and fossil energy consumption at the conversion stage are presented in Table 40, which corresponds to the information in Table 39. GHG emissions associated with the conversion stage for the purchased hydrogen base case, as well as the alternative *in situ* and gasification cases, are estimated to be 6.0 kg CO_{2-eq}/GGE, 1.9 kg CO_{2-eq}/GGE, and 0.9 kg CO_{2-eq}/GGE respectively. The higher GHG profile associated with the purchased hydrogen case is largely attributed to the relatively large hydrogen import, derived from natural gas via external SMR processing (corresponding to a fossil GHG emissions footprint of 11.2 kg CO_{2e}/kg H₂). Alternatively, the gasification scenario for internal hydrogen generation results in a lower GHG profile than the *in situ* scenario primarily due to the reduction in feed rate to the sugar/fuel process train which has a higher GHG footprint due to chemicals, enzyme usage, etc. relative to the gasifier train for a given biomass input; additionally, the gasification scenario also results in a marginally higher fuel yield and also maintains a small electricity export, while the *in situ* scenario requires a small net power import. Direct CO₂ emissions from the conversion facility itself (shown in Table 39) consist of only biogenic CO₂ (i.e., CO₂ absorbed from the atmosphere during feedstock growth). With its biomass origin, biogenic CO₂ does not contribute to the increase of GHG in the atmosphere [106] and is not considered in the IPCC global warming methodology [107]. Hence, the contributions to GHG at the conversion stage are solely from the fossil GHG profile associated with the underlying processes (e.g., material inputs/outputs to and from the facility to support process operations). For the purchased hydrogen base case, beyond external hydrogen it is evident that underlying processes associated with sodium hydroxide (NaOH), HCl, and enzyme production contribute the most to the overall GHG emissions estimate, while excess electricity exported to the grid (largely displacing coal-derived electricity) provides a considerable credit to reduce overall GHG emissions.

Table 40 also shows the fossil energy consumption for the conversion process. Fossil energy consumption associated with the purchased, *in situ*, and gasification hydrogen scenarios are estimated to be 99.3 MJ/GGE, 23.8 MJ/GGE, and 12.2 MJ/GGE respectively. Similar to the GHG emissions, the difference in total fossil energy consumption for the purchased versus internal hydrogen cases is largely driven by hydrogen sourced from natural gas in the purchased hydrogen case; as this is itself a fossil fuel, the resulting fossil energy profile for the purchased hydrogen scenario is higher. Similarly, the gasification scenario results in roughly 50% reduction in fossil energy demand relative to the *in situ* case due to the diversion of a portion of the biomass feedstock away from the more chemical-intensive sugar/fuel process train to the less-intensive gasification train, coupled with slightly higher fuel yields and lower power demand for the gasification scenario.

Table 40. Conversion Process GHG Emissions and Fossil Energy Consumption per GGE Total Fuel Product

H ₂ Sourcing:	GHG Emission			Fossil Energy Input		
	kg CO _{2-eq} /GGE			MJ/GGE		
	Purchased	<i>In situ</i>	Gasification	Purchased	<i>In situ</i>	Gasification
Direct refinery emission	-	-	-	-	-	-
Sulfuric acid	0.03	0.04	0.04	0.36	0.62	0.49
Ammonia	0.09	0.16	0.09	1.80	3.11	1.79
Sodium hydroxide	0.28	0.48	0.28	2.97	5.13	2.97
HCl for IX regeneration	0.23	0.40	0.23	3.07	5.30	3.07
Caustic for IX regeneration	0.09	0.16	0.09	1.37	2.37	1.37
Hydrogen	6.00	NA ^a	NA ^a	97.98	NA ^a	NA ^a
APR/condensation catalysts	0.01	0.01	0.00	0.09	0.12	0.07
Enzyme production	0.19	0.34	0.21	2.09	3.62	2.39
Biorefinery infrastructure	0.01	0.02	0.02	0.14	0.18	0.17
Electricity	-1.00	0.14	-0.23	-11.46	1.64	-2.57
Cooling water chemicals	6.0E-04	1.0E-03	9.4E-04	0.01	0.01	0.01
WWT chemicals	0.03	0.07	0.05	0.64	1.30	0.95
Boiler chemicals	0.02	0.04	0.02	0.11	0.19	0.11
Waste disposal	4.8E-03	0.01	0.01	0.12	0.21	0.15
Gasifier chemicals	NA	NA	0.09	NA	NA	1.25
Total	5.98	1.86	0.91	99.29	23.79	12.21

^a Hydrogen provided internally via *in situ* or gasification generation.

Table 41 summarizes the key sustainability metrics for the conversion process across the three hydrogen sourcing scenarios. On an energy basis, the GHG emissions at the conversion stage for the purchased, *in situ*, and gasification hydrogen cases are 49.2 g CO_{2-eq}/MJ, 15.3 g CO_{2-eq}/MJ, and 7.5 g CO_{2-eq}/MJ respectively. Similarly, the fossil energy consumption for the three scenarios is 0.82 MJ/MJ, 0.20 MJ/MJ, and 0.10 MJ/MJ respectively. In addition to GHG emissions and fossil energy consumption, water consumption (i.e., net water use during the biorefinery operation), total fuel efficiency (yield), and carbon-to-fuel efficiency are also summarized again in Table 41. Biorefinery net water consumption includes, but is not limited to, water that is incorporated into the product and evaporation directly from process operations or indirectly from cooling and heating processes (e.g., cooling tower evaporative losses). By this definition, net water consumption for the modeled biorefinery is 5.8 gal/GGE total fuels for the purchased hydrogen case (discussed above in Section 5.1), increasing to 9.8 gal/GGE for the *in situ* case and 11.4 gal/GGE for the gasification case, which is primarily attributed to lower fuel yields for both alternative hydrogen scenarios but also to higher overall water demand for the gasification case (water demand for the *in situ* case is similar to the purchased hydrogen basis). As discussed previously, total modeled fuel yields are 78.3, 45.3, and 50.1 GGE/dry ton for the purchased, *in situ*, and gasification hydrogen scenarios respectively, associated with a total biomass carbon-to-fuel retention efficiency of 45.3%, 26.2%, and 27.7% respectively. For the purchased hydrogen base case, when including the carbon inherent to natural gas for SMR hydrogen production (0.44 mol/mol H₂ [75]), overall carbon efficiency decreases to 35.7%.

Table 41. Summary of Sustainability Metrics for the Biorefinery Process

	<i>Purchased H₂</i>	<i>In situ H₂</i>	<i>Gasification H₂</i>
GHG emissions (g CO _{2e} /MJ)	49.2	15.3	7.5
Water consumption (m ³ /day)	3817	3716	4788
Water consumption (gal/GGE fuel)	5.8	9.8	11.4
Total fuel yield (GGE/dry ton)	78.3	45.3	50.1
Carbon-to-fuel efficiency (C in fuel/C in biomass) ^a	45%	26%	28%
Net fossil energy consumption (MJ/MJ)	0.82	0.20	0.10

^a Carbon efficiency is defined here as the ratio of carbon in primary fuel products (RDB + naphtha) relative to carbon in biomass feed; when also including carbon in natural gas for SMR H₂ production, the purchased hydrogen case decreases to 36%.

Based on this analysis, the key sustainability metrics for GHG and fossil energy profiles are notably influenced in this technology pathway by the choice for hydrogen sourcing. Although the purchased hydrogen case results in substantially higher fuel yields (which benefits sustainability as it does economics by ultimately amortizing inputs over a larger denominator), this benefit is outweighed by the larger sustainability penalty attributed to fossil-derived hydrogen. However, an important caveat is noted here that the purchased hydrogen case assumes off-site SMR hydrogen production and thereby does not allow for realizing any potential benefits gained by on-site integration. An on-site SMR unit integrated with the rest of the facility is likely to improve the sustainability profile for the purchased hydrogen case, as it would allow for additional heat and power integration opportunities, as well as offsetting a portion of the required natural gas by routing light gases produced in conversion/upgrading to the reformer unit rather than the combustor as is done currently. Such an option should be evaluated moving forward. A complete well-to-wheel LCA evaluation of all scenarios is required to more fully understand the sustainability implications for the full supply chain based on this technology pathway, such as how the results may weigh into qualification for credits under the Renewable Fuel Standard (RFS) set by the U.S. Environmental Protection Agency (EPA) [108]. Such an analysis is outside the scope of this work; however, this preliminary analysis around the conversion stage suggests an optimum could exist between the two extremes. Namely, a commercial facility utilizing this technology would be likely to import purchased hydrogen up to the point of reaching a threshold for GHG limits allowed by RFS requirements, given the low cost of natural gas (and thereby hydrogen) and ensuing economic benefits for MFSP as presented above. The remaining hydrogen deficit could then be made up via *in situ* hydrogen production from a fraction of hydrolysate carbon and/or process off-gases, or from gasifying a portion of the incoming biomass feedstock, or other renewable sources. This trade-off between economics and sustainability is depicted in Figure 20, which presents the boundary conditions for the hydrogen scenarios considered here, and assumes linear interpolation between these boundaries (taking the averages of the two internal hydrogen scenarios for illustrative purposes). It is recognized that the system may not behave linearly between these boundaries, but a linear interpolation is assumed strictly to illustrate the opposing trends between the two metrics and the possibility of an optimum existing in between. Again, a different set of tradeoffs may be observed for a process utilizing on-site SMR hydrogen production integrated with the rest of the facility, which could plausibly eliminate any RFS constraints altogether without needing to include alternative hydrogen sourcing options. This will be an important area for further analysis moving forward, to better understand and quantify such details.

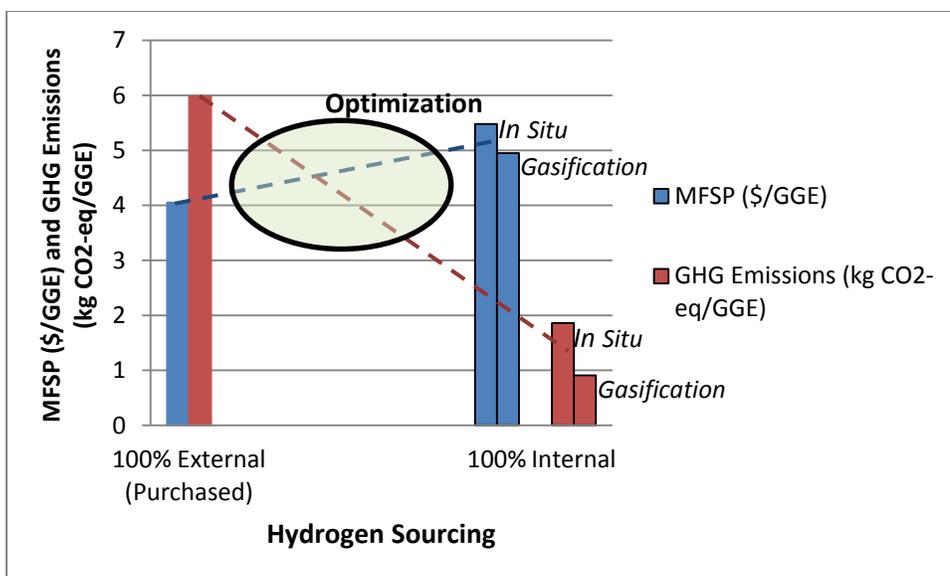


Figure 20. Trade-offs depicting conceptual optimization between economics (MFSP) and sustainability (fossil GHG emissions) based on hydrogen sourcing as modeled here (illustrative only). (Purchased hydrogen via external natural gas SMR versus internal production via *in situ* or gasification scenarios.)

5.4 Paths Forward to \$3/GGE

The modeled overall fuel selling price for this conversion pathway is projected to range between \$4.05–\$5.48/GGE as an n^{th} -plant target case, depending on the mode of hydrogen sourcing (purchased versus *in situ* hydrogen respectively, with biomass gasification falling within this range at \$4.95/GGE). While it is somewhat speculative to assign a specific target year by which process performance will have been shown publicly to increase to a level consistent with these cost targets (due to a large lack of publicly available data from which to draw such conclusions around technology progress), it is expected that the targets documented here would be achievable *prior to year 2022*. Similar to the base case results presented for the 2013 biological conversion pathway, this still leaves further room for improvement required to ultimately achieve final DOE-BETO cost goals of \$3/GGE by 2022. A number of possibilities to allow for such improvements are briefly discussed here.

Moving beyond the process and economic performance metrics described here, the primary means to further improve MFSP is either to further reduce costs or to increase yields (i.e., overall carbon efficiency from biomass to useful products). Additional cost reductions are possible for this technology pathway that have not been considered in this analysis. For example, further reducing sugar costs and optimizing process integration will continue to be important areas for further R&D and may be achieved through the use of alternative or milder pretreatment options and/or improved enzyme performance (higher conversion yields and/or lower enzyme doses or cost), including the incorporation of new enzyme classes and enzymatic hydrolysis mechanisms [109]. Furthermore, the flexibility of the catalytic conversion operations towards utilizing a variety of carbon components beyond just monomeric sugars, including oligomers, sugar degradation products, and solubilized lignin, is a key benefit that may have more opportunity to be leveraged than what was considered here; namely, pursuing more synergistic integration to further tailor biomass deconstruction steps to produce less costly “sugar” (hydrolysate) consisting of more oligomers and less monomers at reduced chemical and enzyme demands [19].

Aside from cost reductions, further opportunities also exist for further improving fuel yields, primarily by way of solubilizing additional lignin up front during deacetylation for subsequent catalytic conversion and upgrading. However, the majority of solubilized lignin and other solubilized components (e.g., acetate and extractives) from deacetylation partition into the separate liquor product stream, which is then routed directly to the boiler in the base case. This highlights the primary opportunity for ultimately achieving \$3/GGE goals, which is to utilize additional lignin and potentially other carbon components in this separate liquor stream for conversion to value-added coproducts. The possibility for utilization of lignin for conversion to coproducts was investigated in detail in the 2013 biological pathway design report, including process modeling scenarios and consideration of economic and sustainability implications, and will not be repeated here, but will be briefly summarized. The reader is encouraged to refer to the 2013 report for additional detail and context for the lignin-to-coproduct pathway analysis.

As described in the 2013 design report, there are a number of possible approaches for isolating lignin for subsequent conversion to coproducts, including: (1) deconstructing and solubilizing lignin up front through more severe alkaline pretreatment operations than the deacetylation base case; (2) fractionating biomass into lignin and carbohydrate fractions utilizing organic solvents with an acid catalyst; and (3) isolating and solubilizing lignin after it has passed through carbohydrate deconstruction steps and separated out from soluble hydrolysate, e.g., via belt filtration as employed in the base case design. The 2013 report focused on the first option, utilizing alkaline pretreatment for purposes of conducting a feasibility TEA assessment for subsequent upgrading to commodity chemicals or fuels. The alkaline pretreatment operation was modeled to take place at 130°C with a caustic (NaOH) loading of roughly 40 mg/g dry biomass and a small amount of added anthraquinone to minimize carbohydrate losses into the soluble liquor stream. Under these conditions, 80% of the lignin was assumed to be solubilized for illustrative purposes, and subsequently converted to one of four example coproducts (among many other coproduct options): (1) 1,3-butadiene, (2) 1,4-butanediol, (3) adipic acid, and (4) cyclohexane. All of these coproducts were deemed reasonable examples in light of sufficient market volumes (greater than 1 MM tons/year world market volume) so as to not risk flooding a particular product market from a small number of biorefinery facilities. The economic analysis included consideration of all pertinent steps, including lignin solubilization (alkaline pretreatment), conversion to the given product, and product purification to more than 95% purity.

The 2013 analysis showed that a number of the selected coproduct options supported achieving or even surpassing \$3/GGE MFSP targets, based on a starting baseline MFSP of \$5.10/GGE prior to including a lignin coproduct train (i.e. solely focused on bioconversion of sugars to hydrocarbons). Namely, adipic acid and 1,4-butanediol both supported achievement of \$3/GGE at 60%–80% conversion of available lignin to coproduct. The results of the TEA assessment are repeated below in Figure 21. Although no new analysis was conducted here to repeat this study, this figure is included below to illustrate that a similar result would be expected in this pathway. Specifically, given that the incoming biomass feedstock composition remains the same as the assumptions employed in the 2013 analysis for the purchased and *in situ* hydrogen scenarios, both of these cases would track in parallel to the original cost curves presented in Figure 21 for the biological pathway in terms of economic implications for achieving ultimate \$3/GGE goals, but shifted up by \$0.38/GGE for the *in situ* case (\$5.48/GGE versus \$5.10/GGE for the 2013 biological case), or shifted down by \$1.05/GGE for the *ex situ* purchased hydrogen case (\$4.05/GGE versus \$5.10/GGE). This concept is highlighted in Figure 21 by overlaying parallel lines for two of the example coproducts, shifted by these respective

differences; note, the original curves for the 2013 biological case originate at \$5.57/GGE rather than \$5.10/GGE reflecting the increased capital costs for adding a lignin coproduct train prior to introducing any coproduct yields (i.e. 0% conversion of lignin to coproducts). This suggests a lower lignin conversion around 40% would be required for the purchased hydrogen scenario to ultimately achieve \$3/GGE for either the adipic acid or 1,4-butanediol coproducts; alternatively, the required lignin conversion would increase to 80-90% for the *in situ* hydrogen scenario to support achieving \$3/GGE. The biomass gasification case would fall between these example boundaries.

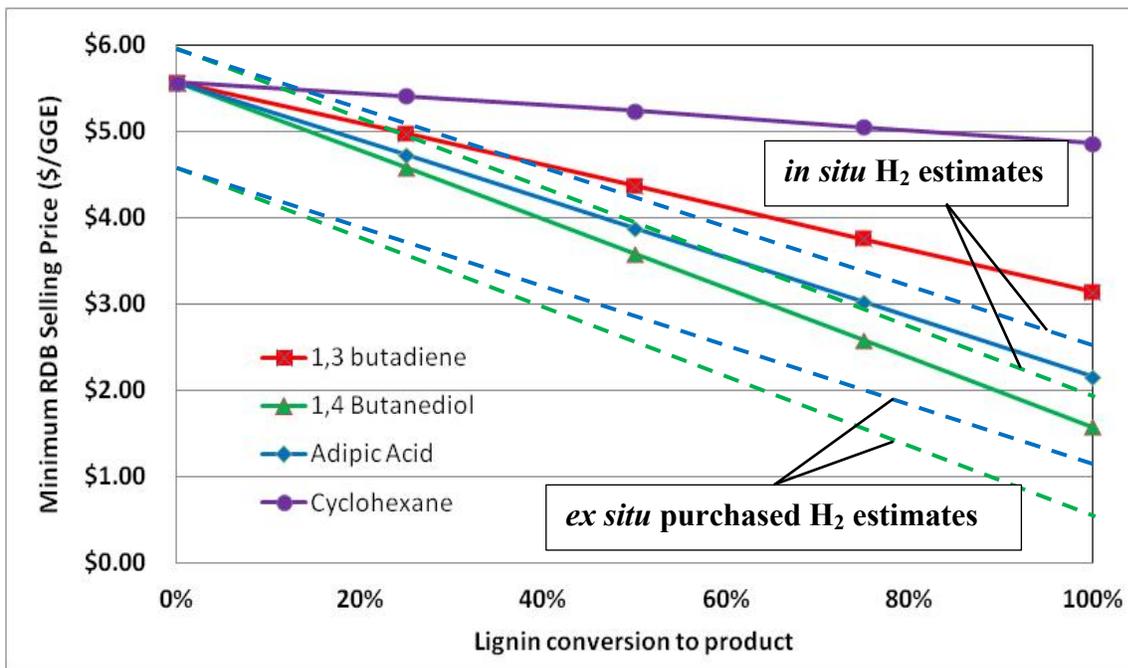


Figure 21. NREL 2013 design report analysis of lignin-to-coproduct scenarios [8]. (Minimum fuel selling price as a function of lignin conversion to selected coproducts, excerpted from 2013 report. Note: 0% point (\$5.57/GGE) was higher than 2013 base case (\$5.10/GGE) given the initial cost premium of adding a new lignin conversion train without any coproduct yield. Dashed-green (1,4-butanediol) and blue (adipic acid) lines are overlaid here to approximate *ex situ* purchased hydrogen scenario and *in situ* hydrogen scenario for the present catalytic conversion pathway analysis.)

Beyond economic considerations for coproduct pathways, it is also critical to understand trade-offs against sustainability metrics (including greenhouse gas and fossil energy profiles) for pursuing such a strategy rather than lignin combustion. NREL’s 2013 design report also included quantification of sustainability metrics attributed to the above coproduct scenarios from lignin and found that the 1,4-butanediol and adipic acid coproduct options fared better than the base case for lignin combustion in terms of greenhouse gas emissions, while the other two coproducts fared worse. A similar result would be expected here; however, it is more important to reiterate the primary point, which is that a much larger array of coproduct options are possible than merely the four options discussed above, and some (but not all) coproducts are likely to support ultimately achieving \$3/GGE cost targets for the integrated biorefinery, and promising options must be weighed against considerations for sustainability as well as market volume capacity. These scenarios and associated economic, sustainability, and market drivers will continue to be investigated for both the biological and catalytic conversion pathways moving forward.

6 Concluding Remarks

6.1 Summary

This report establishes a plausible case for achieving an intermediate cost goal prior to 2022 of approximately \$4–\$5/GGE of upgraded renewable diesel- and naphtha-range blendstocks via a “hybrid” biochemical/catalytic process, based on a single set of process R&D metrics and associated cost estimates run across two scenarios for hydrogen sourcing. Namely, the pathway discussed here assumes continued improvements are made in the biomass deconstruction areas including deacetylation, dilute-acid pretreatment, and enzymatic hydrolysis, beyond performance demonstrated during NREL’s 2012 State of Technology trials. This includes demonstrating high removal of acetate, extractives, ash, and other non-convertible or otherwise problematic components at low loss of carbohydrates under reasonable conditions during deacetylation, as well as high sugar yields under mild conditions with low chemical and enzyme loadings during pretreatment and hydrolysis. These front-end unit operations leverage NREL expertise and prior work in these areas around biomass deconstruction, and will continue to be important areas of future research for continued improvement.

In addition to biomass deconstruction to sugars, the design also sets targets for sugar and other hydrolysate carbon upgrading to fuels, primarily carbon yield (85.6% carbon retention efficiency from hydrolysate carbon to fuels exiting hydrotreating), hydrogen consumption for catalytic reactions (6.5 wt % relative to hydrolysate components converted to fuels), and reactor WHSVs for conversion-stage operations (1–1.2 hr⁻¹ across three conversion reactor stages). These metrics apply equally to the performance of the catalytic reaction stages whether sourcing hydrogen externally via off-site natural gas SMR (*ex situ*), or internally by (a) diverting a fraction of hydrolysate through separate reforming reactions (*in situ*) or (b) diverting a fraction of delivered biomass feedstock to a separate gasification process train to produce hydrogen for the fuel production train. The primary difference is that while the *in situ* route maintains the same carbon efficiency and reaction hydrogen consumption per input rate of hydrolysate carbon converted to fuels, the amount of carbon converted to fuels is roughly 40% lower given the requirement to divert 41% of the hydrolysate stream to the hydrogen production train to satisfy all hydrogen requirements for both the conversion fuel train as well as the hydrotreating step; similarly, the gasification train requires diverting 36% of the total available biomass towards syngas production for this purpose. While the above conversion process targets are derived from patent literature data with reasonable extrapolations made as to ultimately achievable performance in the future, it is subjective to speculate at what point these targets may be achieved given the lack of publication data and granularity of patent data available today (particularly related to catalytic conversion of lignocellulosic hydrolysate, rather than commodity sugars). At a minimum, it is anticipated these targets would be achievable prior to year 2022 and thus would represent an “intermediate” achievement case on the way to ultimate cost goals of \$3/GGE by 2022. To further improve costs towards this ultimate \$3/GGE goal, additional improvements would be required by way of improved process integration with lower costs for biomass deconstruction to sugars, as well as increasing carbon yield beyond what is achievable from only the carbohydrate fraction (e.g., pursuing opportunities for value-added coproducts from lignin).

The end result of the techno-economic analysis was a predicted minimum fuel selling price (MFSP) of \$4.05/GGE (2011\$) at a final upgraded total fuel yield of 78.3 GGE/dry ton of biomass or overall carbon efficiency from biomass to fuels of 45.3% (35.7% when also considering the carbon implicit for natural gas-derived hydrogen) for the “base case” pathway, assuming purchase of off-site hydrogen. This reflects a \$1.02/GGE contribution from feedstock, a \$0.21/GGE contribution from

enzymes, and a \$2.81/GGE contribution from the remainder of the conversion process (including \$0.81/GGE attributed to hydrogen costs). Although this case results in favorable fuel yields and MFSP results, it does so at the expense of substantial natural gas-derived hydrogen import, which carries sustainability implications. To begin understanding cost and sustainability drivers associated with hydrogen sourcing, cursory models were also run for *in situ* hydrogen production directly from a fraction of biomass hydrolysate, as well as gasification of a fraction of incoming biomass feedstock. Given the requirement for 41% diversion of hydrolysate carbon to *in situ* hydrogen, the resulting MFSP cost for the *in situ* approach was found to be \$5.48/GGE with a fuel yield of 45.3 GGE/ton or overall carbon efficiency from biomass of 26.2%. As the gasification scenario required a slightly lower diversion (36%) of biomass feedstock for production of hydrogen from syngas, fuel and carbon yields were slightly higher than the *in situ* scenario at 50.1 GGE/ton and 27.7% overall carbon yield, translating to an MFSP cost of \$4.95/GGE. These modeled selling prices are strictly representative of *nth*-plant assumptions regarding biorefinery design, operation, and financing, and are not intended to reflect first-of-a-kind or early-entry commercial facilities. Consistent with the biological conversion pathway discussed in the 2013 report, the technology pathway described here carries a distinct advantage of producing a mixture of components which may be directly utilized by the existing fuel infrastructure and blended into existing fuel markets (namely diesel and naphtha in this case).

Beyond the detailed modeling and analysis conducted to establish the \$4–\$5/GGE intermediate targets for this pathway, this report also reiterates viable improvements to ultimately achieve the \$3/GGE target by 2022, based on lignin conversion to value-added coproducts as originally detailed in the 2013 report. While the analysis for lignin coproduct pathways was not repeated here, the baseline MFSP costs from this work suggest it is reasonable to conclude that lignin deconstruction and conversion to a number of possible coproduct options still exhibits the potential to ultimately achieve the \$3/GGE target at least similarly to the 2013 biological conversion pathway (when comparing to the *in situ* hydrogen scenario here), if not more easily (when considering the *ex situ* purchased hydrogen scenario).

In addition to providing an economic analysis, this report also considers sustainability metrics for the baseline purchased hydrogen and alternative (*in situ* and gasification) hydrogen scenarios by tracking and reporting on greenhouse gas emissions, fossil energy demand, and direct consumptive water use at the modeled biorefinery facility. The sustainability analysis for the purchased hydrogen scenario indicates a GHG profile of 6.0 kg CO_{2-eq}/GGE, fossil energy consumption of 99.3 MJ/GGE, and water demand of 5.8 gal/GGE. In comparison, the *in situ* hydrogen scenario results in a GHG profile reducing to 1.9 kg CO_{2-eq}/GGE, fossil energy consumption of 23.8 MJ/GGE, and water demand of 9.8 gal/GGE, while the gasification scenario is estimated at a GHG profile of 0.9 kg CO_{2-eq}/GGE, fossil energy consumption of 12.2 MJ/GGE, and water demand of 11.4 gal/GGE. While this is strictly associated with the biorefinery conversion process and does not represent a full well-to-wheel LCA analysis (which is required to more completely understand the implications for these scenarios on overall sustainability), it does indicate notable differences between external versus internal hydrogen sourcing and highlights the key benefit for considering the alternative internal hydrogen options. As discussed previously, a plausible case for commercial-scale production could fall between the two extremes, where a facility would import the maximum amount of natural gas-derived hydrogen as practical given restrictions imposed by applicable policy guidelines, while producing the rest internally either via hydrolysate carbon reforming and/or biomass gasification; however, additional sustainability benefits could also be gained by considering on-site SMR hydrogen production which could alter these trade-offs. Given these important results in both cost and sustainability implications for hydrogen

sourcing, as well as additional trade-offs that would become apparent when also introducing coproduction of value-added products from lignin (or other potential carbon sources), NREL will continue to evaluate these scenarios as further experimental data becomes available.

Similar to the 2013 biological pathway, it is worthwhile to reiterate that cellulosic ethanol production via biochemical processing is supported by decades of research and process understanding, both at NREL and elsewhere, thus carrying a level of certainty and credibility when making target projections for biochemical ethanol TEA modeling, as in the case of the 2011 and even 2002 ethanol design reports. In contrast, catalytic hydrocarbon production via aqueous phase reforming of sugars is a much newer and more novel approach to biofuel production, with fairly limited understanding of technology status and progress offered by information in the public domain, particularly in the context of cellulosic hydrolysate conversion. Thus, the absolute cost values established here, as well as the timeframe required to achieve these outcomes, inherently carry a somewhat higher degree of uncertainty given the nascent stage of research as presented in the public domain.

6.2 Future Work

Moving forward, to ultimately achieve cost goals as well as reduce uncertainty in key areas for the catalytic conversion pathway, a number of important bottlenecks, uncertainties, and areas for further development are summarized below. These points are repeated from a recently published technical memo for catalytic hydrocarbon production [19]:

- **Optimize upstream biomass deconstruction processes for downstream upgrading.** Catalytic conversion routes allow for more flexibility in the hydrolysate composition that can be utilized for fuel production. Unlike biological routes which require primarily monomeric sugars for fuel production, catalytic routes can utilize oligomeric species, carbohydrate degradation products, and solubilized lignin intermediates. To help reduce the overall costs and maximize biomass carbon utilization, future work is needed to develop opportunities to lower downstream upgrading costs by tailoring the biomass deconstruction processes for catalytic upgrading. This includes investigating alternative deconstruction routes that maximize the desirable slate of components and minimize the cost of hydrolysate purification and conditioning steps for catalytic upgrading. One such option could include an alternative approach based on dehydrating sugars to produce furfurals which may phase separate out of the aqueous stream, thereby bypassing multiple hydrolysate cleanup steps. Opportunities for biomass preprocessing to remove components (i.e., ash) should also be explored as a means to improve catalytic performance and minimize hydrolysate conditioning costs.
- **Minimize cost of hydrolysate conditioning steps.** Alternative, lower cost hydrolysate conditioning steps, including solid/liquid separations and improved purification process steps, should be investigated. This includes developing a better understanding for particle size requirements associated with the catalytic conversion step and associated implications on optimizing separations designs and costs with increased certainty.
- **Maximize carbon selectivity towards desired fuel products.** Catalysts could be designed to have increased selectivity to hydrocarbon fuel products and to minimize side- or byproduct steps. An understanding of catalyst fundamentals is needed to design better catalysts and understand catalyst selectivity. Rational design and development of catalysts could be accomplished using a combination of computational and experimental techniques.

- **Improve catalyst lifetime and durability.** Understanding catalyst deactivation mechanisms for hydrolysate poisoning will be critical to improving catalyst lifetimes and minimizing processing costs. Using biomass-derived hydrolysates can help identify problematic components. Determining the impact of lignin deconstruction products on catalyst lifetimes as a function of deconstruction routes will improve process integration and carbon efficiency for fuel production. A better fundamental understanding of both chemical conversion and poisoning mechanisms may help to design better catalysts that are resistant to contaminants and poisons. Additionally, optimizing catalysts and conditions to improve catalyst WHSV and reduce reactor volumes will also be important.
- **Pursue opportunities for lignin utilization.** Catalytic processes are more tolerant to the introduction of a range of carbon species than biological systems are. For these catalytic processes, lignin utilization has great potential for improving total hydrocarbon product yields and increasing overall carbon efficiency for biomass conversion compared to pathways focused on carbohydrate carbon conversion alone. The key challenges for lignin usage are difficulty solubilizing the components and avoiding side reactions that form more recalcitrant compounds. Development and demonstration of a catalyst that can effectively convert lignin products without severe deactivation can significantly improve overall economic viability. Furthermore, parallel conversion pathways from lignin to other value-added coproducts will also be an important consideration for ultimately achieving \$3/GGE production costs for the integrated pathway.
- **Expand analysis on hydrogen sourcing tradeoffs.** While the analysis presented here considered economic and sustainability implications for a number of possible hydrogen sourcing scenarios strictly within the context of the biorefinery facility alone, this work could be expanded upon to conduct a more thorough LCA analysis for the full well-to-wheel supply chain. This would enable a more complete understanding of TEA and LCA tradeoffs associated with hydrogen sourcing, in light of the large amount of hydrogen consumed in the catalytic steps for this process. This expanded modeling work should also consider added benefits that could be brought about by considering a fully integrated on-site SMR hydrogen plant, as this would allow for additional optimization and recycle opportunities beyond model cases that assume hydrogen import from off-site.

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Appendix A. Individual Equipment Costs Summary

The following table shows abbreviated specifications, purchased cost, and installed cost for each piece of equipment in this process design. Although each piece of equipment has its own line, many were quoted as part of a package so their scaling calculations are not shown. NREL and Harris Group would like to acknowledge the equipment vendors who assisted us with the cost estimation effort for this design report.

Mechanical Equipment List: Advanced Biofuels Engineering Design: Dilute Acid Pretreatment, Enzymatic Hydrolysis, Hydrolysate Catalytic Conversion, Hydrotreating to HC							Scaled Installed Costs											
EQPT NO	EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	\$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year	
							11/11/2014											
C- 101	Transfer Conveyor	160 MTPH ea., enclosed, 60 in. x 65 ft.	20 hp	CS	2	\$5,397,000	2009	\$5,397,000	STRM.101	94697	kg/hr	0.60	1.7	104167	1.10	\$5,714,628	\$6,413,216	\$10,902,467
C- 102	High Angle Transfer Conveyor	160 MTPH ea., enclosed, 72 in. wide	50 hp	CS	2	INCLUDED												
C- 103	Reversing Load-in Conveyor	320 MT / hr, enclosed, 84 in. wide	20 hp	CS	1	INCLUDED												
C- 104	Dome Reclaim System	100 MTPH ea.	45 KW	CS	2	\$3,046,000	2009	\$3,046,000	STRM.101	94697	kg/hr	0.60	1.7	104167	1.10	\$3,225,265	\$3,619,540	\$6,153,218
C- 105	Reclaim Conveyor	100 MTPH ea., enclosed, 48 in. x 125	10 hp	CS	2	INCLUDED												
C- 106	High Angle Transfer Conveyor	100 MTPH, enclosed, 72 in. wide	20 hp	CS	1	INCLUDED												
C- 107	Elevated Transfer Conveyor	100 MTPH, enclosed, 48 in. x 200 ft.	10 hp	CS	1	INCLUDED												
C- 108	Process Feed Conveyor	70 MTPH ea., enclosed, 42 in. x 25 ft.	5 hp ea.	CS	1	INCLUDED												
M- 101	Truck Scale	10' x 70', 200,000 lb		CONCRETE	2	\$110,000	2009	\$110,000	STRM.101	94697	kg/hr	0.60	1.7	104167	1.10	\$116,474	\$130,712	\$222,211
M- 102	Truck Dumper	70' x 55 ton x 63 degree	2 x 50 hp	CS	2	\$484,000	2009	\$484,000	STRM.101	94697	kg/hr	0.60	1.7	104167	1.10	\$512,485	\$575,134	\$977,727
M- 103	Truck Dumper Hopper	3500 cu.ft. hopper w/ drag chain	50 hp	CS	2	\$502,000	2009	\$502,000	STRM.101	94697	kg/hr	0.60	1.7	104167	1.10	\$531,544	\$596,523	\$1,014,089
M- 104	Concrete Feedstock Storage	98 ft. dia., 160 ft. high., 4000 MT		CONCRETE	2	\$3,500,000	2009	\$3,500,000	STRM.101	94697	kg/hr	0.60	1.7	104167	1.10	\$3,705,984	\$4,159,025	\$7,070,342
M- 105	Belt Scale	Scale plus processor		CS	2	\$10,790	2009	\$10,790	STRM.101	94697	kg/hr	0.60	1.7	104167	1.10	\$11,425	\$12,822	\$21,797
M- 106	Dust Collection System	8500 ACFM	25 hp	CS	6	\$279,900	2009	\$279,900	STRM.101	94697	kg/hr	0.60	1.7	104167	1.10	\$296,373	\$332,603	\$565,425
Area 100 Totals															\$14,114,178	\$15,839,574	\$26,927,276	

EQPT NO	EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	\$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
Mechanical Equipment List							Scaled Installed Costs											
S- 201	In-line Sulfuric Acid Mixer	Kynar lined - 600 gpm H2O - 5 gpm		SS304	1	\$6,000	2009	\$6,000	strm.a200.214	136260	kg/hr	0.50	1.0	806	0.01	\$462	\$518	\$518
A- 203	Blowdown Tank Agitator	Side-mounted, 3 x 75 hp. (170 KW)	170 KW	316LSS	1	INCLUDED												
A- 204	Flash Tank Agitator	Side-mounted, 3 x 75 hp. (170 KW)	170 KW	316LSS	3	\$90,000	2009	\$90,000	strm.a200.254	252891	kg/hr	0.50	1.5	216084	0.85	\$83,193	\$93,363	\$140,045
A- 209	Ammonia Addition Tank		10 hp	SS	1	\$21,900	2009	\$21,900	strm.a200.228	410369	kg/hr	0.50	1.5	324669	0.79	\$19,479	\$21,861	\$32,791
A- 210	Ammonia Static Mixer			SS	1	\$5,000	2009	\$5,000	strm.a200.275	157478	kg/hr	0.50	1.0	108584	0.69	\$4,152	\$4,659	\$4,659
C- 201	Transfer Conveyor	800 mm x 7600 mm (2'-8" x 25')	60 hp ea	316LSS	2	INCLUDED												
C- 202	Distribution Conveyor	800 mm x 7600 mm (2'-8" x 25')	60 hp ea	316LSS	2	INCLUDED												
C- 203	Overfeed Conveyor	800 mm x 7600 mm (2'-8" x 25')	60 hp ea	316LSS	4	INCLUDED												
C- 204	Pressurized Heating Screw	2500 mm x 9500 mm (8' x 31')	100 hp	Dup. 2205	1	INCLUDED												
C- 205	Pressurized Pre-heater	850 mm x 3500 mm (2'-10" x 12')	50 hp	Dup. 2205	2	INCLUDED												
C- 206	Pressurized Transport - No. 1	900 mm x 3500 mm (3' x 12')	25 hp	Incoloy 825	1	INCLUDED												
C- 207	Pressurized Transport - No. 2	1200 mm x 3500 mm (4' x 12')	50 hp	Incoloy 825	1	INCLUDED												
H- 201	Pretreatment Water Heater	29.9 MMBTU		304SS	1	\$92,000	2010	\$92,000	Heat.A200.QH201		-8 Gcal/hr	0.70	2.2	-11	1.41	\$117,262	\$124,692	\$274,323
H- 244	Waste Vapor Condenser	Copied H-201		304SS	1	\$34,000	2009	\$34,000	Heat.A200.QH244		2 Gcal/hr	0.70	2.2	1	0.77	\$28,403	\$31,876	\$70,126
M- 201	Doffing Roll Storage Bins	60 cu. m. (2100 cu.ft.) with conveyors/scrapers	54 hp ea	304/316SS	2	INCLUDED												
M- 202	Pin Drum Feeder		7.5 hp	316LSS	2	INCLUDED												
M- 203	Plug Screw Feeder		1510 hp	316LSS	2	INCLUDED												
M- 204	Prehydrolysis/Vertical	16' x 62' - 10 min. residence time		Dup. 2205	1	INCLUDED												
M- 205	Pin Drum Feeder		7.5 hp	Incoloy 825	2	INCLUDED												
M- 206	Plug Screw Feeder		1510 hp	Incoloy 825	2	INCLUDED												
M- 207	Pretreatment Reactor	3,000 mm x 10,300 mm each - scaled	135 hp	Incoloy 825	5	\$16,400,000	2013	\$31,200,000	SCIS.a200.211b	63166	kg/hr	0.60	1.5	62942	1.00	\$31,133,709	\$31,192,291	\$46,788,436
P- 201	Sulfuric Acid Pump	9 gpm, 245 FT TDH		316SS	1	\$8,000	2009	\$8,000	strm.710	3720	kg/hr	0.80	2.3	2240	0.60	\$5,332	\$5,984	\$13,763
P- 203	Blowdown Tank Discharge	1900 gpm, 150 FT TDH	125	316SS	1	\$25,635	2010	\$25,635	strm.a200.222	292407	kg/hr	0.80	2.3	226960	0.78	\$20,932	\$22,258	\$51,193
P- 204	Flash Tank Discharge Pump	900 gpm, 150 FT TDH	75	316SS	1	\$30,000	2009	\$30,000	strm.a200.254	204390	kg/hr	0.80	2.3	216084	1.06	\$31,366	\$35,200	\$80,960
P- 209	Hydrolyzate Pump	1771 gpm, 150 FT TDH	100	316SS	1	\$22,500	2009	\$22,500	strm.a200.228	402194	kg/hr	0.80	2.3	324669	0.81	\$18,958	\$21,275	\$48,933
T- 201	Sulfuric Acid Tank	12,800 gal, 24hr residence time		PLASTIC	1	\$6,210	2010	\$6,210	strm.710	1981	kg/hr	0.70	3.0	2240	1.13	\$6,768	\$7,197	\$21,591
T- 203	Blowdown Tank	23' x 48' - 25 min. - 110,000 gal.		SS316	1	INCLUDED												
T- 204	Flash Tank	23' x 48' - 110,000 gal.		SS316	1	\$511,000	2009	\$511,000	strm.a200.223	264116	kg/hr	0.70	2.0	226960	0.86	\$459,546	\$515,724	\$1,031,447
T- 209	Ammonia Addition Tank	118,000 gal, 1hr residence time		SS304	1	\$236,000	2009	\$236,000	strm.a200.228	410369	kg/hr	0.70	2.0	324669	0.79	\$200,308	\$224,795	\$449,590
T- 225	Deacetylation Reactor	14' x 30' vessel, quad discharge screw	15hp ea	SS316	3	\$780,000	2013	\$780,000	strm.a200.211a	277167	kg/hr	0.70	1.7	278573	1.01	\$2,348,303	\$2,352,721	\$3,999,626
	Deacetylation Reactor	Feed and discharge drag conveyors	40 hp	SS316	3	\$110,000	2013	\$110,000	strm.a200.211a	277167	kg/hr	0.80	1.7	278573	1.01	\$331,338	\$331,962	\$564,335
Area 200 Totals																\$34,809,510	\$34,986,375	\$53,572,336

EQPT NO	EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	\$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year	
Mechanical Equipment List							Scaled Installed Costs												
A- 300	Enzymatic Hydrolysis Reactor		30 hp	SS304	1	\$52,500	2009	\$52,500	inumsscf	1	ea	1.00	1.5	6	6.00	\$315,000	\$353,507	\$530,261	
A- 306	Enzymatic Hydrolysis Storage		20 hp	SS304	2	\$68,300	2009	\$68,300	strm.a300.306a	425878	kg/hr	0.50	1.5	331610	0.78	\$60,269	\$67,636	\$101,455	
A- 308	Enzyme-Hydrolysate Mixer	Inline mixer 1673 gpm	100 hp	SS316	1	\$109,000	2009	\$109,000	strm.a300.t310fd	379938	kg/hr	0.50	1.7	331610	0.87	\$101,832	\$114,281	\$194,277	
F- 300	Enzymatic Hydrolysis Reactor	1,000,000 gallon ea		304SS	12	\$10,128,000	2009	\$10,128,000	inumsscf	12	ea	1.00	1.5	6	0.50	\$5,064,000	\$5,683,052	\$8,524,578	
H- 301	Hydrolyzate Cooler	Plate & frame 32.5 MMBTU/hr		SS 304	1	\$85,000	2010	\$85,000	heat.a300.qc301	8	Gcal/hr	0.70	2.2	7	0.82	\$74,086	\$78,781	\$173,317	
P- 306	Hydrolyzate Storage Transfer	2152 gpm, 171 FT TDH	125 hp	316SS	1	\$26,800	2009	\$26,800	strm.a300.306a	488719	kg/hr	0.80	2.3	331610	0.68	\$19,651	\$22,054	\$50,723	
P- 310	Saccharification Transfer	352 gpm, 150 FT TDH	20	316SS	5	\$47,200	2009	\$47,200	strm.a300.306	421776	kg/hr	0.80	2.3	331610	0.79	\$38,938	\$43,699	\$100,507	
T- 306 A	Enzymatic Hydrolysis Storage Tank	1,200,000 gallon		316SS	1	\$1,317,325	2011	\$1,317,325	strm.a300.306a	328984	kg/hr	0.70	1.8	331610	1.01	\$1,324,678	\$1,324,678	\$2,384,421	
T- 310	Saccharification Tank	250,000 gal each - 19' dia. x 120' tall		304SS	8	\$3,840,000	2009	\$3,840,000	strm.a300.306a	421776	kg/hr	0.70	2.0	331610	0.79	\$3,244,992	\$3,641,678	\$7,283,356	
S- 205	Lignin Filter (after enzymatic hydrolysis)	(4) 170 m2 horizontal belt filters	660 hp ea	304SS	5	\$2,152,500	2013	\$2,152,500	scis.a200.solids	328984	kg/hr	1.00	1.7	15801	1.00	\$10,762,500	\$10,782,751	\$17,230,677	
E- 303	Sugar Concentration	Mechanical vapor recompression	3600 KW	304SS	1	\$6,370,000	2013	\$6,370,000	strm.sugmix.301SUG	244084	kg/hr	0.70	2.0	360732	1.48	\$8,373,199	\$8,388,954	\$16,777,909	
F- 320	Polishing Filter	Ceramic microfiltration		316L	4	\$440,000	2014	\$440,000	STRM.REFINE.101	53,204	kg/hr	0.90	1.8	55,664	1.05	\$1,833,058	\$1,867,169	\$3,360,905	
T- 321	Polished Hydrolysate Storage Tank	8500 gal - 20 min residence time		SS	1	\$168,000	2011	\$168,000	STRM.REFINE.101	76,712	kg/hr	0.70	1.8	111,930	1.46	\$218,860	\$218,860	\$393,948	
W- 330	Ion Exchange	Strong acid cation/weak base anion			1	\$5,250,000	2014	\$5,250,000	STRM.REFINE.101	53,204	kg/hr	0.90	1.8	55,664	1.05	\$5,467,930	\$5,569,681	\$10,025,426	
T- 331	Deionized Sugar Storage Tank	8500 gal - 20 min residence time		SS	1	\$168,000	2011	\$168,000	STRM.REFINE.101	76,712	kg/hr	0.70	1.8	111,930	1.46	\$218,860	\$218,860	\$393,948	
T- 303	Concentrated Sugar Storage	5,500 gallons - 20 min residence time	20 hp	SS	1	\$168,000	2011	\$168,000	scis.a200.solids	76712	kg/hr	0.70	1.8	15801	0.21	\$55,588	\$55,588	\$100,059	
T- 305	Filtered Hydrolysate Storage	30,000 gallons - 20 min residence	20 hp	SS	1	\$51,600	2011	\$51,600	STRM.a200.S205.S205LI	348690	kg/hr	0.70	1.8	360732	1.03	\$52,841	\$52,841	\$95,114	
Area 300 Totals																\$51,476,284	\$52,734,070	\$68,820,879	

EQPT NO	EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	\$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year	
Mechanical Equipment List							Scaled Installed Costs												
A- 400	Cellulase Fermentor Agitators		800.0	SS316		\$580,000	2009	\$580,000	CLVESSEL	1	ea	1.00	1.5	5	5.00	\$2,900,000	\$3,254,512	\$4,881,769	
A- 401	Cellulase Fermentor Agitators		0.75hp	SS316		\$3,420	2009	\$3,420	ICLSEED	1	ea	1.00	1.5	4	4.00	\$13,680	\$15,352	\$23,028	
A- 402	Cellulase Fermentor Agitators		8 hp	SS316		\$63,000	2009	\$63,000	ICLSEED	1	ea	1.00	1.5	4	4.00	\$252,000	\$282,806	\$424,209	
A- 403	Cellulase Fermentor Agitators		80 hp	SS316		\$11,000	2009	\$11,000	ICLSEED	1	ea	1.00	1.5	4	4.00	\$44,000	\$49,379	\$74,068	
A- 406	Cellulase Nutrient Mix Tank		3 hp	CS	1	\$4,800	2009	\$4,800	strm.a400.416	174	kg/hr	0.50	1.6	112	0.65	\$3,857	\$4,329	\$6,926	
A- 410	Cellulase Hold Tank Agitator		10 hp	SS316	1	\$26,900	2009	\$26,900	strm.422	10930	kg/hr	0.50	1.5	6941	0.64	\$21,437	\$24,058	\$36,087	
F- 400	Cellulase Fermentor	80,000 gal, 1 atm, 28 °C, internal coil		SS316		\$400,500	2009	\$400,500	CLVESSEL	1	ea	1.00	2.0	5	5.00	\$2,002,500	\$2,247,297	\$4,494,594	
F- 401	1st Cellulase Seed Fermentor	80 gallon skid complete - \$46,000 ea		304SS		\$46,000	2009	\$46,000	ICLSEED	1	ea	1.00	1.8	4	4.00	\$184,000	\$206,493	\$371,688	
F- 402	2nd Cellulase Seed Fermentor	800 gallon skid complete - \$57,500 ea		304SS		\$57,500	2009	\$57,500	ICLSEED	1	ea	1.00	1.8	4	4.00	\$230,000	\$258,116	\$464,610	
F- 403	3rd Cellulase Seed Fermentor	8,000 gallon skid complete - \$95,400		304SS		\$95,400	2009	\$95,400	ICLSEED	1	ea	1.00	1.8	4	4.00	\$381,600	\$428,249	\$770,848	
H- 400	Cellulase Fermentation Cooler	Cooling coil included with cellulase		304SS		INCLUDED													
H- 405	Media Prep Tank Cooler	Cooling coil included with media prep tank		304SS	1	INCLUDED													
M- 401	Fermenter Air Compressor	8000 SCFM @ 16 psig		CS	2	\$350,000	2009	\$350,000	strm.a400.450	33168	kg/hr	0.60	1.6	16346	0.49	\$228,920	\$256,904	\$411,047	
P- 400	Cellulase Transfer Pump	59 gpm, 100 FT, TDH SIZE 2X1-10C	3	316SS	1	\$7,357	2010	\$7,357	strm.a400.420	13399	kg/hr	0.80	2.3	6941	0.52	\$4,347	\$4,623	\$10,632	
P- 401	Cellulase Seed Pump	3 gpm, 100 FT TDH SIZE 2X1-10	2	316SS	4	\$29,972	2010	\$29,972	strm.a400.409	681	kg/hr	0.80	2.3	386	0.57	\$19,030	\$20,236	\$46,543	
P- 406	Cellulase Nutrient Transfer	Gear Pump 2 gpm, 100 FT	1	316SS	1	\$1,500	2009	\$1,500	strm.a400.416	454	kg/hr	0.80	2.3	112	0.25	\$491	\$551	\$1,267	
P- 410	Cellulase Feed Pump	Gear Pump	1	316SS	1	\$1,500	2009	\$5,700	strm.a400.422	18168	kg/hr	0.80	2.3	6941	0.38	\$2,640	\$2,963	\$6,814	
P- 420	Anti-foam Pump	Gear Pump 2 gpm, 100 FT	1	316SS	1	\$1,500	2009	\$1,500	strm.a400.444	11	kg/hr	0.80	2.3	6.7	0.63	\$1,039	\$1,166	\$2,683	
T- 406	Cellulase Nutrient Mix Tank	HDPE, 8,000 gal		HDPE	1	\$9,000	2010	\$9,000	strm.a400.416	224	kg/hr	0.70	3.0	112	0.50	\$5,553	\$5,905	\$17,714	
T- 410	Cellulase Hold Tank	80,000 gal		304SS	1	\$248,070	2009	\$248,070	strm.a400.422	10930	kg/hr	0.70	1.8	6941	0.64	\$180,532	\$202,602	\$364,683	
Area 400 Totals																\$6,475,627	\$7,265,541	\$12,409,209	

EQPT NO	EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	\$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year	
Mechanical Equipment List									Scaled Installed Costs										
R- 510	APR-1 Reactor System		Note 1	317L Clad		\$7,960,000	2014	\$7,960,000	STRM.REFINE.101c	53,204	kg/hr	0.37	2.2	48,981	0.92	\$7,720,111	\$7,863,773	\$18,338,318	
R- 520	APR-2 Reactor System		Note 1	317L Clad		\$2,044,000	2014	\$2,044,000	STRM.REFINE.101c	53,204	kg/hr	0.37	2.2	48,981	0.92	\$1,982,400	\$2,019,290	\$4,708,985	
R- 530	Condensation Reactor System		Note 1	317L Clad		\$2,645,000	2014	\$2,645,000	STRM.REFINE.101c	53,204	kg/hr	0.37	2.2	48,981	0.92	\$2,565,288	\$2,613,025	\$6,093,574	
	Conversion FE HX	Shell and tube		317 SS	1	\$948,600	2011	\$948,600	heat.REFINE.17	79.2	Gcal/hr	0.70	2.2	74	0.94	\$906,592	\$906,592	\$1,994,503	
	Condensation Feed Heater	Plate and shell		Duplex SS	1	\$908,250	2014	\$908,250	heat.REFINE.QHB20	17	Gcal/hr	0.70	2.2	15	0.85	\$811,106	\$826,199	\$1,817,639	
	Condensation Effluent Economizer	Plate and shell		Duplex SS	1	\$1,668,850	2014	\$1,668,850	heat.REFINE.DHOGEFC	4	Gcal/hr	0.70	2.2	4	0.96	\$1,625,343	\$1,655,589	\$3,642,295	
	Condensation Effluent Air Cooler	Air cooler		317 SS	1	\$1,028,500	2011	\$1,028,500	heat.REFINE.QACCATCL	45	Gcal/hr	0.70	2.8	42	0.94	\$982,809	\$982,809	\$2,751,865	
	Condensation Effluent Water Cooler	Shell and tube		317 SS	1	\$1,405,000	2011	\$1,405,000	heat.REFINE.25	13	Gcal/hr	0.70	2.2	12	0.93	\$1,339,247	\$1,339,247	\$2,946,342	
	MU Compressor	Reciprocating compressor	3162 hp		1	\$2,930,000	2014	\$2,930,000	work.REFINE.10	2,359	KW	0.87	1.9	2620	1.11	\$3,210,187	\$3,269,925	\$6,212,857	
	Recycle Compressor	Centrifugal compressor	6300 hp		1	\$2,900,000	2014	\$2,900,000	work.REFINE.16	4,700	KW	0.78	1.4	5060	1.08	\$3,071,665	\$3,128,825	\$4,380,355	
S- 570	Hydrotreating Facility		610 hp		1	\$11,500,000	2013	\$11,500,000	strm.REFINE.HRT.OUT-	13528	kg/hr	0.50	2.0	22874	1.69	\$14,953,898	\$14,982,036	\$29,964,072	
																REFINE Totals	\$39,168,648	\$39,587,309	\$82,850,805

EQPT NO	EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	\$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year	
Mechanical Equipment List									Scaled Installed Costs										
T- 606	Anaerobic System	(4) 27 MG reactors, 8 recycle pumps,	795 hp	Concrete	4	\$25,800,000	2012	\$25,800,000	COD	27211	kg/hr	0.60	1.1	19955	0.73	\$21,419,106	\$21,459,408	\$23,819,943	
	Biogas Emergency Flare				4	INCLUDED	2012	INCLUDED											
T- 608	Aeration Basin	Concrete and steel, not installed cost		Concrete	3	\$4,908,054	2012	\$4,908,054	Hydraulic flow	2.7	MGD	0.60	2.1	1.3	0.47	\$3,108,842	\$3,114,692	\$6,447,412	
	Pump - Submersible, Anaerobic Feed	2500 gpm ea	15 hp ea	CS	2	INCLUDED	2012	INCLUDED											
	Pump - Centrifugal, Aeration	852 gpm ea	45 hp	CS	4	INCLUDED	2012	INCLUDED											
	Aeration Grid	Full floor aeration grid		CS	1	INCLUDED	2012	INCLUDED											
	Caustic Feed System		1.5 hp	CS	4	\$20,000	2012	\$20,000	COD	27211	kg/hr	0.60	3.0	19955	0.73	\$16,604	\$16,635	\$49,906	
	Blowers	15000 SCFM @ 10.3 psig ea	1000 hp	CS	9	\$2,070,000	2012	\$2,070,000	COD	27211	kg/hr	0.60	2.0	19955	0.73	\$1,718,510	\$1,721,743	\$3,443,486	
	Surface Aerators		100 hp	CS	3	\$150,000	2012	\$150,000	Hydraulic flow	2.7	MGD	0.60	2.7	1.3	0.47	\$95,012	\$95,191	\$254,161	
R- 609	Membrane Bioreactor	Includes membrane, CIP, scour	85 hp ea	CS	1	\$4,898,500	2012	\$4,898,500	Hydraulic flow	2.7	MGD	1.00	1.6	1.3	0.47	\$2,288,471	\$2,292,777	\$3,760,154	
	Pump, Centrifugal, MBR, RAS		160 hp	CS	6	INCLUDED	2012	INCLUDED											
S- 612	Gravity Belt Thickeners	2m presses	48hp	CS	3	\$750,000	2012	\$750,000	COD	27211	kg/hr	0.60	1.6	19955	0.73	\$622,648	\$623,820	\$991,874	
S- 611	Centrifuge		165 hp	CS	1	\$686,800	2012	\$686,800	COD	27211	kg/hr	0.60	2.7	19955	0.73	\$570,180	\$571,253	\$1,536,670	
	Pump, Centrifugal, Centrifuge	105 gpm	15hp	CS	2	INCLUDED	2012	INCLUDED											
	Pump, Submersible, Centrate	100 gpm	10 hp ea	CS	2	INCLUDED	2012	INCLUDED											
	Dewatering Polymer Addition	9.8 gph neat polymer	1 hp ea	CS	2	INCLUDED	2012	INCLUDED											
	Conveyor		10 hp ea	CS	1	\$7,000	2012	\$7,000	COD	27211	kg/hr	0.60	2.9	19955	0.73	\$5,811	\$5,822	\$16,652	
S- 610	Reverse Osmosis			CS	7	\$2,450,000	2012	\$2,450,000	Hydraulic flow	2.7	MGD	1.00	1.8	1.3	0.47	\$1,144,586	\$1,146,739	\$2,006,794	
M- 640	Evaporator	368 gpm	1480 hp	Titanium	1	\$5,000,000	2012	\$5,000,000	Hydraulic flow	2.7	MGD	0.60	1.6	1.3	0.47	\$3,167,082	\$3,173,041	\$5,108,597	
	Ammonia Addition System	0.63 gpm	4.5 hp	CS	4	\$195,200	2012	\$195,200	COD	27211	kg/hr	0.60	1.5	19955	0.73	\$162,055	\$162,360	\$250,034	
																Area 600 Totals	\$34,318,906	\$34,383,482	\$47,685,681

EQPT NO	EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	\$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
Mechanical Equipment List						Scaled Installed Costs												
A- 720	CSL Storage Tank Agitator		10 hp	SS304	1	\$21,200	2009	\$21,200	strm.735	1393	kg/hr	0.50	1.5	83	0.06	\$5,160	\$5,791	\$8,686
A- 760	DAP Makeup Tank Agitator		5.5 hp	SS304	1	\$9,800	2009	\$9,800	strm.755	163	kg/hr	0.50	1.5	0	0.00	\$0	\$0	\$0
C- 755	DAP Bulk Bag Unloader	Super sack unloader			1	\$30,000	2009	\$30,000	strm.755	163	kg/hr	0.60	1.7	0	0.00	\$0	\$0	\$0
P- 701	Product Pump	150 gpm, 112 FT TDH	5.0	CS	2	\$9,200	2009	\$9,200	strm.703	22681	kg/hr	0.80	3.1	16226	0.72	\$7,038	\$7,898	\$24,484
P- 703	Sulfuric Acid Pump	5 gpm, 150 FT TDH SIZE 2X1-10	0.5	SS316	1	\$7,493	2010	\$7,493	strm.710	1981	kg/hr	0.80	2.3	2240	1.13	\$8,268	\$8,792	\$20,221
P- 704	Firewater Pump	2500 gpm, 150 FT TDH	125.0	CS	1	\$15,000	2009	\$15,000	strm.a700.713	8343	kg/hr	0.80	3.1	5968	0.72	\$11,474	\$12,877	\$39,917
P- 720	CSL Pump	8 gpm, 80 FT TDH	0.5	CS	1	\$3,000	2009	\$3,000	strm.735	1393	kg/hr	0.80	3.1	83	0.06	\$313	\$351	\$1,088
P- 760	DAP Pump	2 gpm, 100 FT TDH	0.5	CS	1	\$3,000	2009	\$3,000	strm.755	163	kg/hr	0.80	3.1	0	0.00	\$0	\$0	\$0
T- 701	Product Storage Tank	1.2 M gal., 7 day storage, floating roof		A285C	1	\$1,146,219	2009	\$1,146,219	strm.703	18145	kg/hr	0.70	1.7	16226	0.89	\$1,059,967	\$1,189,543	\$2,022,223
T- 702	Product Storage Tank	300,000 gal., 7 day storage, floating roof		A285C	1	\$352,790	2009	\$352,790	strm.REFINE.C8-MIX	4536	kg/hr	0.70	1.7	3821	0.84	\$312,840	\$351,083	\$596,841
T- 703	Sulfuric Acid Storage Tank	12,600 gal, 12' dia x15' H		SS	1	\$96,000	2010	\$96,000	strm.710	1981	kg/hr	0.70	1.5	2240	1.13	\$104,630	\$111,260	\$166,890
	Caustic Storage Tank	12,600 gal, 12' dia x15' H		SS	1	\$96,000	2011	\$96,000	strm.a200.2	1981	kg/hr	0.70	1.5	2046	1.03	\$98,187	\$98,187	\$147,281
	Caustic Storage Tank	12,600 gal, 12' dia x15' H		SS	1	\$96,000	2011	\$96,000	strm.a200.2	1981	kg/hr	0.70	1.5	1122	0.57	\$64,478	\$64,478	\$96,717
T- 704	Firewater Storage Tank	600,000 gal - 4 hrs @ 2500 gpm		Glass lined	1	\$803,000	2009	\$803,000	strm.a700.713	8343	kg/hr	0.70	1.7	5968	0.72	\$635,160	\$712,805	\$1,211,769
T- 706	Ammonia Storage Tank	28,000 gal		SA- 516-70	2	\$196,000	2010	\$196,000	strm.717	1171	kg/hr	0.70	2.0	368	0.31	\$87,090	\$92,608	\$185,217
T- 720	CSL Storage Tank	70,000 gal		Glass lined	1	\$70,000	2009	\$70,000	strm.735	1393	kg/hr	0.70	2.6	83	0.06	\$9,681	\$10,865	\$28,248
	Glucose Storage Tank	70,000 gal		Glass lined	1	\$70,000	2009	\$70,000	strm.a400.401	1393	kg/hr	0.70	2.6	1427	1.02	\$71,195	\$79,898	\$207,736
T- 755	DAP Bulk Bag Holder	Super sack holder			1	INCLUDED												
T- 760	DAP Make-up Tank	12,800 gal		SS304	1	\$102,000	2009	\$102,000	strm.717	1615	kg/hr	0.70	1.8	368	0.23	\$36,189	\$40,613	\$73,104
Area 700 Totals																\$2,511,670	\$2,787,049	\$4,830,422

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Mechanical Equipment List									Scaled Installed Costs									
H- 801	Burner Combustion Air	INCLUDED			1	INCLUDED												
H- 811	BFW Preheater	INCLUDED			1	INCLUDED												
H- 812	Pretreatment/BFW Heat Recovery	9.4 MM BTU/hr		SS304	1	\$41,000	2009	\$41,000	heat.QH812	-2	Gcal/hr	0.70	2.2	-2	0.86	\$36,776	\$41,272	\$90,797
M- 802	Air Intake Fan	INCLUDED				INCLUDED												
M- 803	Boiler	525,000 lb/hr @ 900 psig		CS	1	\$28,550,000	2010	\$28,550,000	strm.a800.a810.813C	238203	kg/hr	0.60	1.8	246134	1.03	\$29,116,607	\$30,961,505	\$55,730,709
M- 804	Combustion Gas Baghouse	Baghouse, spray dryer scrubber, flues/ducting			1	\$11,000,000	2013	\$11,000,000	strm.a800.a810.813C	238203	kg/hr	0.60	1.8	0	0.00	\$0	\$0	\$0
M- 811	Turbine/Generator	23.6 KW, 2 extractions			1	\$9,500,000	2010	\$9,500,000	work.wtotal	-42200	KW	0.60	1.8	-46446	1.10	\$10,062,505	\$10,700,089	\$19,260,161
M- 820	Hot Process Water Softener				1	\$78,000	2010	\$78,000	strm.a800.a810.813C	235803	kg/hr	0.60	1.8	246134	1.04	\$80,033	\$85,104	\$153,187
M- 830	Amine Addition Pkg.				1	\$40,000	2010	\$40,000	strm.a800.a810.813C	235803	kg/hr	0.00	1.8	246134	1.04	\$40,000	\$42,534	\$76,562
M- 832	Ammonia Addition Pkg.				1	INCLUDED												
M- 834	Phosphate Addition Pkg.				1	INCLUDED												
P- 804	Condensate Pump			SS316	2	INCLUDED												
P- 811	Turbine Condensate Pump			SS304	2	INCLUDED												
P- 824	Deaerator Feed Pump			SS304	2	INCLUDED												
P- 826	BFW Pump			SS316	5	INCLUDED												
P- 828	Blowdown Pump			CS	2	INCLUDED												
P- 830	Amine Transfer Pump			CS	1	INCLUDED												
T- 804	Condensate Collection Tank			A285C	1	INCLUDED												
T- 824	Condensate Surge Drum			SS304	1	INCLUDED												
T- 826	Deaerator	Tray type		CS,SS316	1	\$305,000	2010	\$305,000	strm.a800.a810.813C	235803	kg/hr	0.60	3.0	246134	1.04	\$312,949	\$332,778	\$998,334
T- 828	Blowdown Flash Drum			CS	1	INCLUDED												
T- 830	Amine Drum			SS316	1	INCLUDED												
Area 800 Totals																\$39,648,870	\$42,163,282	\$76,309,750

EQPT NO	EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	\$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
Mechanical Equipment List									Scaled Installed Costs									
M- 902	Cooling Tower System	44,200 gpm	750 hp	FIBERGLASS	1	\$1,375,000	2010	\$1,375,000	strm.a900.945	10037820	kg/hr	0.60	1.5	13741966	1.37	\$1,660,154	\$1,765,345	\$2,648,018
M- 904	Plant Air Compressor	400 SCFM@125 psig	150 hp		1	\$28,000	2010	\$28,000	DRY101	83333	kg/hr	0.60	1.6	83333	1.00	\$28,000	\$29,774	\$47,639
M- 908	Chilled Water Package	2 x 2350 tons (14.2 MM kcal/hr)	3400 hp		1	\$1,275,750	2010	\$1,275,750	heat.a900.qchwop	14	Gcal/hr	0.60	1.6	4	0.28	\$589,025	\$626,347	\$1,002,155
M- 910	CIP System	100,000 GAL		SS304/SS31	1	\$421,000	2009	\$421,000	strm.a900.914	63	kg/hr	0.60	1.8	145	2.30	\$694,222	\$779,088	\$1,402,358
P- 902	Cooling Water Pump	16,120 gpm, 100 FT TDH SIZE 20X20-28	500.0	CS	3	\$283,671	2010	\$283,671	strm.a900.945	10982556	kg/hr	0.80	3.1	13741966	1.25	\$339,384	\$360,888	\$1,118,753
P- 912	Makeup Water Pump	685 gpm, 75 FT TDH SIZE 6X4-13	20.0	CS	1	\$6,864	2010	\$6,864	strm.a900.904	155564	kg/hr	0.80	3.1	341548	2.20	\$12,877	\$13,693	\$42,448
P- 914	Process Water Circulating	2285 gpm, 75 FT TDH SIZE 8X6-13	75.0	CS	1	\$15,292	2010	\$15,292	strm.a900.905	518924	kg/hr	0.80	3.1	590356	1.14	\$16,954	\$18,028	\$55,888
S- 904	Instrument Air Dryer	670 SCFM - cycling type		CS	1	\$15,000	2009	\$15,000	DRY101	83333	kg/hr	0.60	1.8	83333	1.00	\$15,000	\$16,834	\$30,301
T- 904	Plant Air Receiver	3800 gal - 72" x 228" vertical		CS	1	\$16,000	2009	\$16,000	DRY101	83333	kg/hr	0.60	3.1	83333	1.00	\$16,000	\$17,956	\$55,664
T- 914	Process Water Tank No. 1	250,000 gal		CS	1	\$250,000	2009	\$250,000	strm.a900.905	451555	kg/hr	0.70	1.7	590356	1.31	\$301,594	\$338,462	\$575,386
Area 900 Totals																\$3,673,210	\$3,966,415	\$6,978,608

Appendix B. Discounted Cash Flow Rate of Return Worksheet

Year	-2	-1	0	1	2	3	4	5	6
Fixed Capital Investment	\$20,047,207	\$150,354,051	\$80,188,827						
Land	\$1,848,000								
Working Capital			\$31,323,761						
Loan Payment				\$56,017,968	\$56,017,968	\$56,017,968	\$56,017,968	\$56,017,968	\$56,017,968
Loan Interest Payment	\$2,405,665	\$20,448,151	\$30,070,810	\$30,070,810	\$27,995,037	\$25,753,203	\$23,332,022	\$20,717,146	\$17,893,080
Loan Principal	\$30,070,810	\$255,601,886	\$375,885,127	\$349,937,969	\$321,915,038	\$291,650,273	\$258,964,326	\$223,663,504	\$185,538,616
Fuel Sales				\$172,159,887	\$229,546,516	\$229,546,516	\$229,546,516	\$229,546,516	\$229,546,516
Byproduct Credit				\$4,024,206	\$5,365,608	\$5,365,608	\$5,365,608	\$5,365,608	\$5,365,608
Total Annual Sales				\$176,184,093	\$234,912,124	\$234,912,124	\$234,912,124	\$234,912,124	\$234,912,124
Annual Manufacturing Cost									
Feedstock				\$43,460,550	\$57,947,400	\$57,947,400	\$57,947,400	\$57,947,400	\$57,947,400
Baghouse Bags				\$519,687					\$519,687
Initial Catalyst Charge				\$7,466,161					
Other Variable Costs				\$61,482,169	\$70,265,336	\$70,265,336	\$70,265,336	\$70,265,336	\$70,265,336
Fixed Operating Costs				\$16,080,969	\$16,080,969	\$16,080,969	\$16,080,969	\$16,080,969	\$16,080,969
Total Product Cost				\$129,009,536	\$144,293,705	\$144,293,705	\$144,293,705	\$144,293,705	\$144,813,392
Annual Depreciation									
General Plant Writedown				14%	24.49%	17.49%	12.49%	8.93%	8.92%
Depreciation Charge				\$78,618,644	\$134,735,521	\$96,223,939	\$68,715,666	\$49,129,776	\$49,074,759
Remaining Value				\$471,546,817	\$336,811,295	\$240,587,356	\$171,871,690	\$122,741,914	\$73,667,155
Steam Plant Writedown				3.75%	7.22%	6.68%	6.18%	5.71%	5.29%
Depreciation Charge				\$2,861,616	\$5,508,801	\$5,095,202	\$4,713,653	\$4,359,576	\$4,032,970
Remaining Value				\$73,448,134	\$67,939,333	\$62,844,131	\$58,130,478	\$53,770,902	\$49,737,932
Net Revenue				(\$64,376,514)	(\$77,620,941)	(\$36,453,926)	(\$6,142,923)	\$16,411,921	\$19,097,922
Losses Forward				(\$64,376,514)	(\$141,997,455)	(\$178,451,381)	(\$178,451,381)	(\$184,594,304)	(\$168,182,383)
Taxable Income				(\$64,376,514)	(\$141,997,455)	(\$178,451,381)	(\$184,594,304)	(\$168,182,383)	(\$149,084,461)
Income Tax				\$0	\$0	\$0	\$0	\$0	\$0
Annual Cash Income				-\$8,843,412	\$34,600,450	\$34,600,450	\$34,600,450	\$34,600,450	\$34,080,764
Discount Factor		1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830	0.6209
Annual Present Value	\$356,968,848			-\$8,039,465	\$28,595,413	\$25,995,830	\$23,632,573	\$21,484,157	\$19,237,703
Total Capital Investment + Interest	\$29,404,055	\$187,882,422	\$141,583,398						
Net Present Worth			\$0						

Year	7	8	9	10	11	12	13	14	15
Fixed Capital Investment									
Land									
Working Capital									
Loan Payment	\$56,017,968	\$56,017,968	\$56,017,968	\$56,017,968	\$0	\$0	\$0	\$0	\$0
Loan Interest Payment	\$14,843,089	\$11,549,099	\$7,991,589	\$4,149,479	\$0	\$0	\$0	\$0	\$0
Loan Principal	\$144,363,737	\$99,894,868	\$51,868,489	\$0	\$0	\$0	\$0	\$0	\$0
Fuel Sales	\$229,546,516	\$229,546,516	\$229,546,516	\$229,546,516	\$229,546,516	\$229,546,516	\$229,546,516	\$229,546,516	\$229,546,516
By-Product Credit	\$5,365,608	\$5,365,608	\$5,365,608	\$5,365,608	\$5,365,608	\$5,365,608	\$5,365,608	\$5,365,608	\$5,365,608
Total Annual Sales	\$234,912,124	\$234,912,124	\$234,912,124	\$234,912,124	\$234,912,124	\$234,912,124	\$234,912,124	\$234,912,124	\$234,912,124
Annual Manufacturing Cost									
Feedstock	\$57,947,400	\$57,947,400	\$57,947,400	\$57,947,400	\$57,947,400	\$57,947,400	\$57,947,400	\$57,947,400	\$57,947,400
Baghouse Bags					\$519,687				
Other Variable Costs	\$70,265,336	\$70,265,336	\$70,265,336	\$70,265,336	\$70,265,336	\$70,265,336	\$70,265,336	\$70,265,336	\$70,265,336
Fixed Operating Costs	\$16,080,969	\$16,080,969	\$16,080,969	\$16,080,969	\$16,080,969	\$16,080,969	\$16,080,969	\$16,080,969	\$16,080,969
Total Product Cost	\$144,293,705	\$144,293,705	\$144,293,705	\$144,293,705	\$144,813,392	\$144,293,705	\$144,293,705	\$144,293,705	\$144,293,705
Annual Depreciation									
General Plant Writedown	8.93%	4.46%							
Depreciation Charge	\$49,129,776	\$24,537,380							
Remaining Value	\$24,537,380	\$0							
Steam Plant Writedown	4.89%	4.52%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%
Depreciation Charge	\$3,730,021	\$3,450,727	\$3,404,941	\$3,404,178	\$3,404,941	\$3,404,178	\$3,404,941	\$3,404,178	\$3,404,941
Remaining Value	\$46,007,911	\$42,557,184	\$39,152,243	\$35,748,065	\$32,343,124	\$28,938,946	\$25,534,005	\$22,129,827	\$18,724,886
Net Revenue	\$22,915,533	\$51,081,213	\$79,221,888	\$83,064,761	\$86,693,791	\$87,214,240	\$87,213,477	\$87,214,240	\$87,213,477
Losses Forward	(\$149,084,461)	(\$126,168,928)	(\$75,087,715)	\$0	\$0	\$0	\$0	\$0	\$0
Taxable Income	(\$126,168,928)	(\$75,087,715)	\$4,134,173	\$83,064,761	\$86,693,791	\$87,214,240	\$87,213,477	\$87,214,240	\$87,213,477
Income Tax	\$0	\$0	\$1,446,960	\$29,072,666	\$30,342,827	\$30,524,984	\$30,524,717	\$30,524,984	\$30,524,717
Annual Cash Income	\$34,600,450	\$34,600,450	\$33,153,490	\$5,527,784	\$59,755,905	\$60,093,434	\$60,093,701	\$60,093,434	\$60,093,701
Discount Factor	0.5132	0.4665	0.4241	0.3855	0.3505	0.3186	0.2897	0.2633	0.2394
Annual Present Value	\$17,755,502	\$16,141,365	\$14,060,316	\$2,131,200	\$20,944,080	\$19,147,620	\$17,407,005	\$15,824,479	\$14,385,954
Total Capital Investment + Interest									
Net Present Worth									

Year	16	17	18	19	20	21	22	23	24
Fixed Capital Investment									
Land									
Working Capital									
Loan Payment	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0
Loan Interest Payment	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0
Loan Principal	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0
Fuel Sales	\$229,546,516	\$229,546,516	\$229,546,516	\$229,546,516	\$229,546,516	\$229,546,516	\$229,546,516	\$229,546,516	\$229,546,516
By-Product Credit	\$5,365,608	\$5,365,608	\$5,365,608	\$5,365,608	\$5,365,608	\$5,365,608	\$5,365,608	\$5,365,608	\$5,365,608
Total Annual Sales	\$234,912,124	\$234,912,124	\$234,912,124	\$234,912,124	\$234,912,124	\$234,912,124	\$234,912,124	\$234,912,124	\$234,912,124
Annual Manufacturing Cost									
Feedstock	\$57,947,400	\$57,947,400	\$57,947,400	\$57,947,400	\$57,947,400	\$57,947,400	\$57,947,400	\$57,947,400	\$57,947,400
	\$519,687					\$519,687			
Baghouse Bags									
Other Variable Costs	\$70,265,336	\$70,265,336	\$70,265,336	\$70,265,336	\$70,265,336	\$70,265,336	\$70,265,336	\$70,265,336	\$70,265,336
Fixed Operating Costs	\$16,080,969	\$16,080,969	\$16,080,969	\$16,080,969	\$16,080,969	\$16,080,969	\$16,080,969	\$16,080,969	\$16,080,969
Total Product Cost	\$144,813,392	\$144,293,705	\$144,293,705	\$144,293,705	\$144,293,705	\$144,813,392	\$144,293,705	\$144,293,705	\$144,293,705
Annual Depreciation									
General Plant Writedown									
Depreciation Charge									
Remaining Value									
Steam Plant Writedown	4.46%	4.46%	4.46%	4.46%	4.46%	2.23%			
Depreciation Charge	\$3,404,178	\$3,404,941	\$3,404,178	\$3,404,941	\$3,404,178	\$1,702,471			
Remaining Value	\$15,320,708	\$11,915,767	\$8,511,590	\$5,106,648	\$1,702,471	\$0			
Net Revenue	\$86,694,554	\$87,213,477	\$87,214,240	\$87,213,477	\$87,214,240	\$88,396,261	\$90,618,418	\$90,618,418	\$90,618,418
Losses Forward	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0
Taxable Income	\$86,694,554	\$87,213,477	\$87,214,240	\$87,213,477	\$87,214,240	\$88,396,261	\$90,618,418	\$90,618,418	\$90,618,418
Income Tax	\$30,343,094	\$30,524,717	\$30,524,984	\$30,524,717	\$30,524,984	\$30,938,691	\$31,716,446	\$31,716,446	\$31,716,446
Annual Cash Income	\$59,755,638	\$60,093,701	\$60,093,434	\$60,093,701	\$60,093,434	\$59,160,040	\$58,901,972	\$58,901,972	\$58,901,972
Discount Factor	0.2176	0.1978	0.1799	0.1635	0.1486	0.1351	0.1228	0.1117	0.1015
Annual Present Value	\$13,004,568	\$11,889,218	\$10,808,332	\$9,825,800	\$8,932,506	\$7,994,330	\$7,235,870	\$6,578,064	\$5,980,058
Total Capital Investment + Interest									
Net Present Worth									

Year	25	26	27	28	29	30
Fixed Capital Investment						
Land						(\$1,848,000)
Working Capital						(\$31,323,761)
Loan Payment	\$0	\$0	\$0	\$0	\$0	\$0
Loan Interest Payment	\$0	\$0	\$0	\$0	\$0	\$0
Loan Principal	\$0	\$0	\$0	\$0	\$0	\$0
Fuel Sales	\$229,546,516	\$229,546,516	\$229,546,516	\$229,546,516	\$229,546,516	\$229,546,516
By-Product Credit	\$5,365,608	\$5,365,608	\$5,365,608	\$5,365,608	\$5,365,608	\$5,365,608
Total Annual Sales	\$234,912,124	\$234,912,124	\$234,912,124	\$234,912,124	\$234,912,124	\$234,912,124
Annual Manufacturing Cost						
Feedstock	\$57,947,400	\$57,947,400	\$57,947,400	\$57,947,400	\$57,947,400	\$57,947,400
		\$519,687				
Baghouse Bags						
Other Variable Costs	\$70,265,336	\$70,265,336	\$70,265,336	\$70,265,336	\$70,265,336	\$70,265,336
Fixed Operating Costs	\$16,080,969	\$16,080,969	\$16,080,969	\$16,080,969	\$16,080,969	\$16,080,969
Total Product Cost	\$144,293,705	\$144,813,392	\$144,293,705	\$144,293,705	\$144,293,705	\$144,293,705
Annual Depreciation						
General Plant Writedown						
Depreciation Charge						
Remaining Value						
Steam Plant Writedown						
Depreciation Charge						
Remaining Value						
Net Revenue	\$90,618,418	\$90,098,732	\$90,618,418	\$90,618,418	\$90,618,418	\$90,618,418
Losses Forward	\$0	\$0	\$0	\$0	\$0	\$0
Taxable Income	\$90,618,418	\$90,098,732	\$90,618,418	\$90,618,418	\$90,618,418	\$90,618,418
Income Tax	\$31,716,446	\$31,534,556	\$31,716,446	\$31,716,446	\$31,716,446	\$31,716,446
Annual Cash Income	\$58,901,972	\$58,564,176	\$58,901,972	\$58,901,972	\$58,901,972	\$58,901,972
Discount Factor	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
Annual Present Value	\$5,436,416	\$4,913,854	\$4,492,906	\$4,084,460	\$3,713,145	\$3,375,587
Total Capital Investment + Interest						(\$1,901,026)
Net Present Worth						

Appendix D. Aspen Properties

The table below is a list of the components used in the Aspen model. Previous versions of the model used custom property databanks created at NREL. Where possible in the new model, these components have been replaced with components from Aspen's native databanks. Property definitions for the few remaining custom components were moved into the model itself (i.e., inside the simulation file) and are discussed here.

The component CSL (corn steep liquor) used in previous models was removed. CSL streams are now modeled as 50% water, 25% protein, and 25% lactic acid [110].

Component	Property	Quantity	Units	Reference
H2O	-	-	-	Native Aspen component
GLUCOSE	-	-	-	Native Aspen component (dextrose)
GALACTOS	-	-	-	Duplicate of GLUCOSE
MANNOSE	-	-	-	Duplicate of GLUCOSE
XYLOSE	DHFORM	-216752.65	cal/mol	Native Aspen component (d-xylose) with DHFORM specified (5/6 of GLUCOSE DHFORM)
ARABINOS	-	-	-	Duplicate of XYLOSE
CELLOB	-	-	-	Cellobiose. Used native Aspen component sucrose
SUCROSE	-	-	-	Native Aspen component
GLUCOLIG	MW	162.1424		Glucose oligomers. Most properties from GLUCOSE; MW is GLUCOSE minus H2O
	DHFORM	-192875.34	cal/mol	Back-calculated to match ΔH_c of CELLULOS
GALAOLIG	-	-	-	Galactose oligomers. Duplicate of GLUCOLIG
MANOLIG	-	-	-	Mannose oligomers. Duplicate of GLUCOLIG
XYLOLIG	MW	132.11612		Xylose oligomers. Most properties from XYLOSE; MW is XYLOSE minus H2O
	DHFORM	-149412.58	cal/mol	Back-calculated to match ΔH_c of XYLAN
ARABOLIG	-	-	-	Arabinose oligomers. Duplicate of XYLOLIG
EXTRACT	-	-	-	Organic extractives. Duplicate of GLUCOSE
LGNSOL	-	-	-	Solubilized lignin. Native Aspen component vanillin (see note at LIGNIN)

Component	Property	Quantity	Units	Reference
HMF	MW	126.11		(5-hydroxymethylfurfural) Properties for HMF were estimated within Aspen using NIST TDE routines. Specify molecular structure, MW, TB and DHFORM. From [90].
	TB	532.7	K	
	DHFORM	-79774.53	cal/mol	
	DHVLWT-1	80550000	J/kmol	
	TC	731.012	K	NIST TDE
	PC	5235810	Pa	
	OMEGA	0.99364671		
	VC	0.3425	m ³ /kmol	
	RKTZRA	0.198177974		
FURFURAL	-	-	-	Native Aspen component
AACID	-	-	-	Native Aspen component (acetic acid)
LACID	-	-	-	Native Aspen component (lactic acid)
XYLITOL	-	-	-	Native Aspen component
GLYCEROL	-	-	-	Native Aspen component
SUCCACID	-	-	-	Native Aspen component (Succinic acid)
NH3	-	-	-	Native Aspen component
H2SO4	-	-	-	Native Aspen component
NH4SO4	-	-	-	Native Aspen component (ammonium sulfate)
NH4ACET	PLXANT/1	-1.00E+20	atm	Native Aspen component (ammonium acetate) forced non-volatile
DAP	-	-	-	Native Aspen component (diammonium phosphate)
HNO3	-	-	-	Native Aspen component
NANO3	-	-	-	Native Aspen component
NAOH	-	-	-	Native Aspen component
CNUTR	-	-	-	Cellulase nutrient mix. Duplicate of glucose
WNUTR	-	-	-	WWT nutrient mix. Duplicate of glucose
DENAT	-	-	-	Denaturant. Native Aspen component (n-heptane)

Component	Property	Quantity	Units	Reference
OIL	-	-	-	Corn oil antifoam. Native Aspen component (oleic acid)
O2	-	-	-	Native Aspen component
N2	-	-	-	Native Aspen component
NO	-	-	-	Native Aspen component
NO2	-	-	-	Native Aspen component
CO	-	-	-	Native Aspen component
CO2	-	-	-	Native Aspen component
CH4	-	-	-	Native Aspen component
H2S	-	-	-	Native Aspen component
SO2	-	-	-	Native Aspen component
CELLULOS	DHSFRM	-233200.06	cal/mol	Native Aspen component with specified heat of formation; back-calculated
GALACTAN	-	-	-	Duplicate of CELLULOS
MANNAN	-	-	-	Duplicate of CELLULOS
XYLAN	Formula	C ₅ H ₈ O ₄ (monomer)		
	MW	132.117		
	DHSFRM	-182099.93	cal/mol	From [22]. Assumes the ratio of ΔH_c of glucose:xylose is the same for cellulose:xylan.
ARABINAN	-	-	-	Duplicate of xylan
LIGNIN	-	-	-	Used native Aspen component vanillin (C ₈ H ₈ O ₃). The HHV of this compound (-23,906 BTU/kg) is very close to what we previously assumed for lignin as a custom component (-24,206)
ACETATE	-	-	-	Used native Aspen component acetic acid
PROTEIN	Formula	CH _{1.57} O _{0.31} N _{0.29} S _{0.007}		Wheat gliadin [91]
	MW	22.8396		
	DHSFRM	-17618	cal/mol	From literature value of gliadin ΔH_c [92]
	PLXANT/1	-1.00E+20	atm	Forces non-volatility

Component	Property	Quantity	Units	Reference
ASH	-	-	-	Native Aspen component CaO
ENZYME	Formula	CH _{1.59} O _{0.42} N _{0.24} S _{0.01}		Provided by Novozymes [55]
	MW	24.0156		
	DHSFRM	-17618	cal/mol	copied from PROTEIN
DENZ	-	-	-	Denatured enzyme. Duplicate of enzyme
ZYMO	Formula	CH _{1.8} O _{0.5} N _{0.2}		Z. mobilis cell mass. Average composition of several microorganisms outlined in [93]
	MW	24.6264		
	DHSFRM	-31169.39	cal/mol	From [22] after [94]
TRICHO	Formula	CH _{1.645} O _{0.445} N _{0.205} S _{0.005}		T. reesei cell mass. Average of generic cell mass [56] and the ENZYME composition
	MW	23.8204		
	DHSFRM	-23200.01	cal/mol	Copied BIOMASS
BIOMASS	Formula	CH _{1.64} O _{0.39} N _{0.23} S _{0.0035}		WWT sludge.
	MW	23.238		
	DHSFRM	-23200.01	cal/mol	From [22] after [94]
TAR	-	-	-	Modeled as solid xylose
LIME	-	-	-	Aspen native component (calcium hydroxide)
CASO4	-	-	-	Aspen native component
C	-	-	-	Aspen native component (graphite carbon)
NAPHTHA	-	C4-C9		Group C4-C9 components
DIESEL	-	C10-C30+		Group C10-C30+ components
HYDROGEN	Formula	H ₂		Aspen native component

Appendix E. Process Flow Diagrams

High-level stream table information from Aspen Plus modeling output follows, for key streams associated with each process operation area for the base case model (external purchased hydrogen). This is followed by high-level PFDs for the associated process areas. Space for stream tables was limited; below is a key to lumped components. As the stream table information focuses primarily on the high-level overall process and does not include every individual modeled stream within each process area, mass balance closure around a given unit area may not be 100%.

Other sugars (SS)	arabinose, mannose, galactose, cellobiose, sucrose
Sugar oligomers (SS)	oligomers of glucose, xylose, and all minor sugars
Organic soluble solids (SS)	ammonium acetate, solubilized lignin, organic extractives, lactic acid, cellulase and WWT chemicals
Inorganic soluble solids (SS)	ammonium sulfate, diammonium phosphate, sodium hydroxide, nitric acid, sodium nitrate
Furfurals	furfural, HMF
Other organics	glycerol, corn oil, sorbitol, succinic acid, xylitol, other organic components; including alcohols, ketones, aldehydes, furans, diols, triols, cyclic mono-oxygenates, polyoxygenates, organic acids, propane
CO/SO _x /NO _x /H ₂ S	NO, NO ₂ , SO ₂ , CO, H ₂ S
Other struct. carbohydr. (IS)	arabinan, mannan, galactan
Protein (IS)	corn protein, cellulase, denatured cellulase
Cell mass (IS)	WWT sludge, <i>T. reesei</i> from cellulase production
Other insoluble solids (IS)	tar, ash, carbon, lime, gypsum

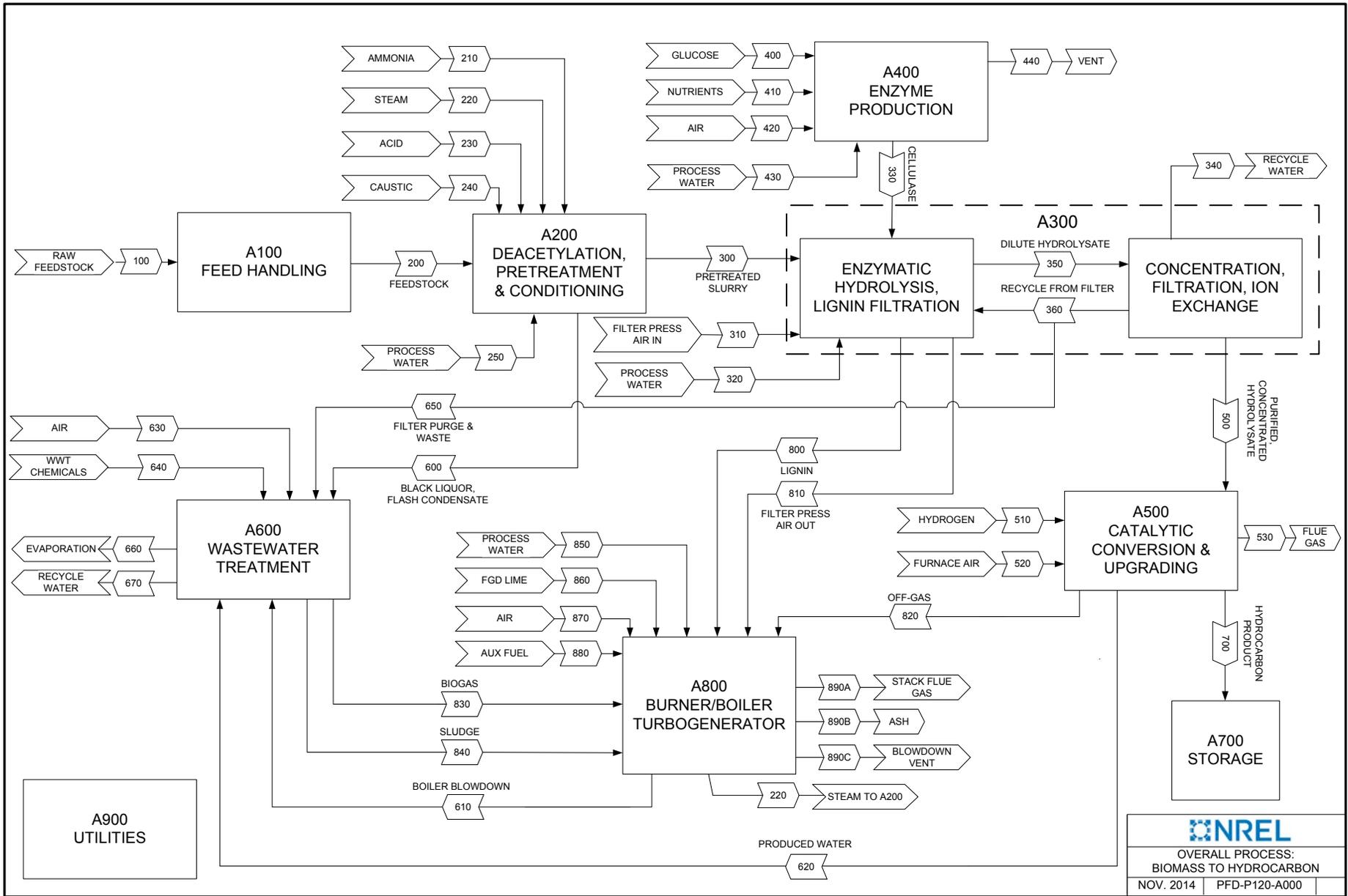


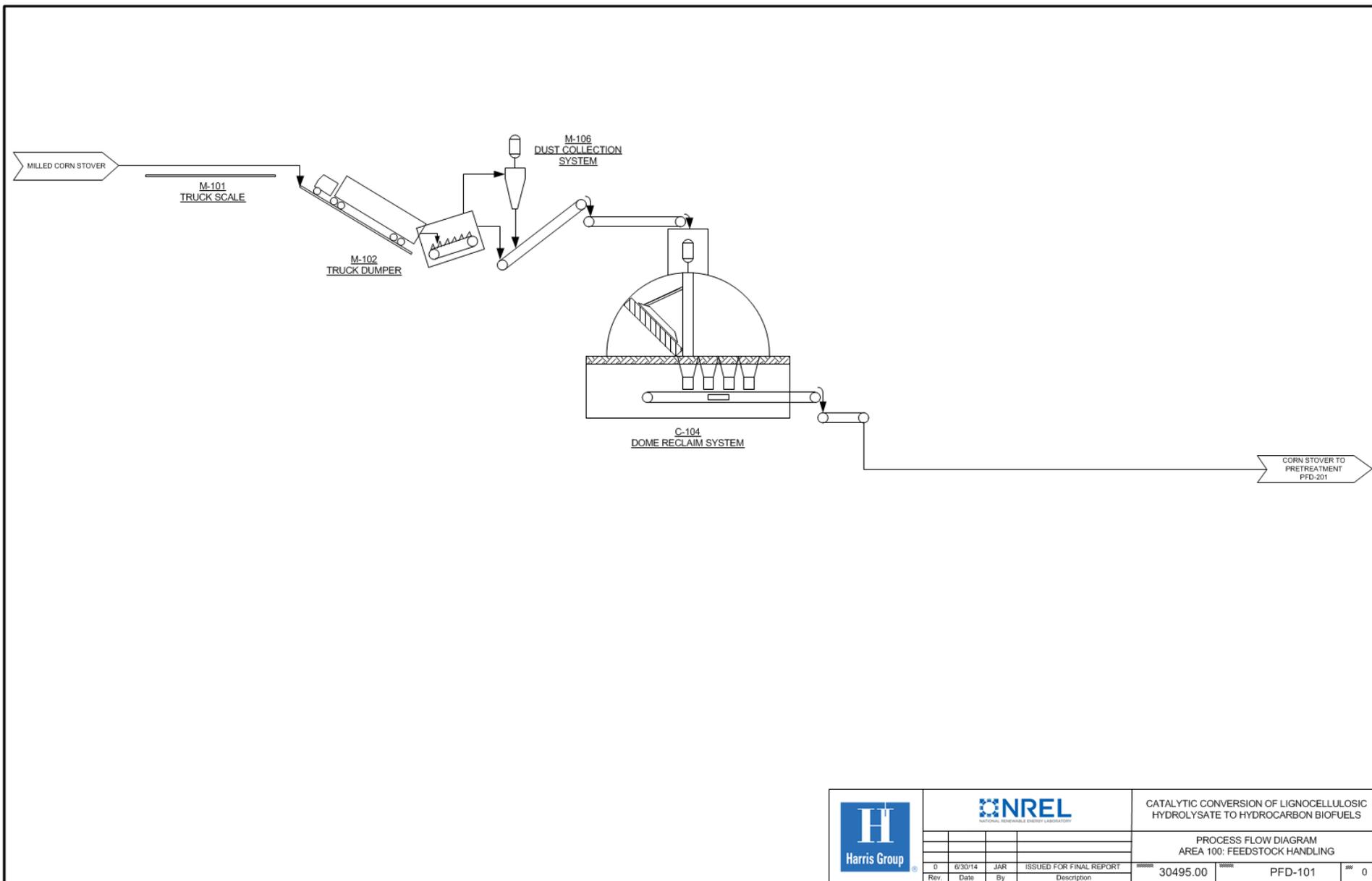
Figure 22. Overall process schematic for Aspen stream tables

Aspen Plus mass balance information for key stream tables:

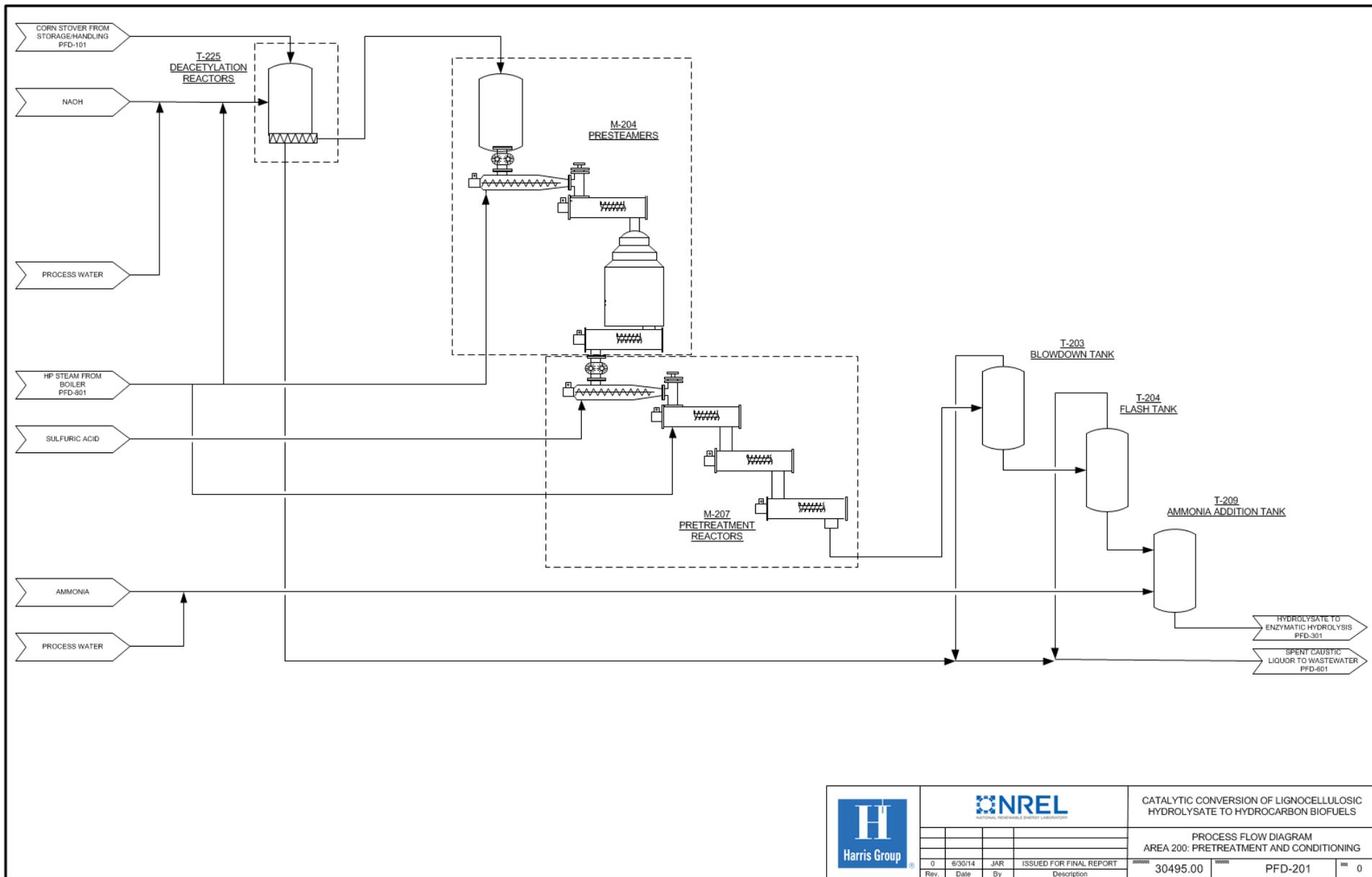
COMPONENT	UNITS	100	200	210	220	230	240	250	300	310	320	330	340	350	360	400
Total Flow	kg/hr	104,167	104,167	310	23,888	806	1,406	304,369	324,669	163	67,177	6,941	248,807	360,733	5,587	1,427
Insoluble Solids	%	67.7%	67.7%	0.0%	0.0%	0.0%	0.0%	0.0%	12.4%	0.0%	0.0%	5.3%	0.0%	0.0%	0.7%	0.0%
Soluble Solids	%	12.3%	12.3%	0.0%	0.0%	0.0%	100.0%	0.0%	7.9%	0.0%	0.0%	0.3%	0.0%	15.4%	48.5%	85.0%
Temperature	°C	25	25	20	536	20	20	99	73	25	20	29	69	43	64	28
Pressure	atm	1.0	1.0	9.0	26.0	3.4	1.0	15.0	7.8	1.0	1.0	1.7	2.0	1.0	72.4	1.0
Vapor Fraction		0.0	0.0	0.0	2.0	0.0	0.0	0.0	0.0	1.0	0.0	0.0	0.0	0.0	0.0	0.0
Fuel-range hydrocarbons	kg/hr															
Water	kg/hr	20,833	20,833		23,888	56		304,369	258,110		67,177	6,549	248,493	304,755	2,819	214
Glucose (SS)	kg/hr								3,381					30,838	1,542	1,213
Xylose (SS)	kg/hr								16,310					16,988	849	
Other sugars (SS)	kg/hr	642	642						3,682					4,181	209	
Sugar Oligomers (SS)	kg/hr								559					1,674	84	
Organic Soluble Solids (SS)	kg/hr	12,208	12,208						749			20	0.3	772	28	
Inorganic Soluble Solids (SS)	kg/hr						1,406		1,011					923		
Ammonia	kg/hr			310								0.1	0.1	0.1		
Acetic Acid	kg/hr															
Sulfuric Acid	kg/hr					750										
Furfurals	kg/hr								504				313	509	10	
Other Organics	kg/hr											7		13	7	
Carbon Dioxide	kg/hr											0.4	0.4	0.4		
Methane	kg/hr															
H2	kg/hr															
O2	kg/hr									38						
N2	kg/hr									125		0.1	0.1	0.1		
CO/SOX/NOX/H2S	kg/hr															
Cellulose (IS)	kg/hr	29,205	29,205						26,226					6	3	
Xylan (IS)	kg/hr	16,273	16,273						415					2	1	
Other Struct. Carbohydr. (IS)	kg/hr	3,675	3,675						96					0.5	0.2	
Acetate (IS)	kg/hr	1,508	1,508													
Lignin (IS)	kg/hr	13,132	13,132						9,980					50	25	
Protein (IS)	kg/hr	2,583	2,583						2,583			311		15	7	
Cell Mass (IS)	kg/hr											54		0.3	0.1	
Other Insoluble Solids (IS)	kg/hr	4,108	4,108						1,064					5	3	

COMPONENT	UNITS	410	420	430	440	500	510	520	530	600	610	620	630	640	650	660
Total Flow	kg/hr	189	16,346	5,729	16,750	99,620	3,847	1,082	1,234	110,277	4,907	78,130	234,604	111	6,722	234,472
Insoluble Solids	%	10.9%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	4.3%	0.0%	0.0%	0.0%	0.0%	0.6%	0.0%
Soluble Solids	%	28.7%	0.0%	0.0%	0.0%	49.0%	0.0%	0.0%	0.0%	15.0%	0.0%	0.0%	0.0%	1.8%	57.4%	0.0%
Temperature	°C	144	25	33	56	64	63	32	380	85	100	-34	25	0	77	25
Pressure	atm	28.4	1.0	5.0	2.0	72.4	19.0	1.0	1.2	1.0	5.4	-36.4	1.0	0.0	-5.1	1.0
Vapor Fraction		3.0	1.0	0.0	2.0	0.0	2.0	1.0	1.0	0.0	0.0	0.0	1.0	0.0	0.0	1.0
Fuel-range hydrocarbons	kg/hr															
Water	kg/hr	41	320	5,729	388	50,639			241	88,642	4,907	77,202	4,595		2,808	4,535
Glucose (SS)	kg/hr					27,754									1,542	
Xylose (SS)	kg/hr					15,289									849	
Other sugars (SS)	kg/hr					3,763				321					209	
Sugar Oligomers (SS)	kg/hr					1,507									84	
Organic Soluble Solids (SS)	kg/hr	54				492				14,835				2	252	
Inorganic Soluble Solids (SS)	kg/hr									1,406					923	0.2
Ammonia	kg/hr	58			0.2									109		0.2
Acetic Acid	kg/hr									7		13				
Sulfuric Acid	kg/hr															
Furfurals	kg/hr					176				332					10	
Other Organics	kg/hr	7										916			7	270
Carbon Dioxide	kg/hr				1,195				148							1,813
Methane	kg/hr															1.6
H2	kg/hr						3,847									
O2	kg/hr		3,733		2,873			252	15				53,573			51,418
N2	kg/hr		12,293		12,293			830	830				176,436			176,434
CO/SOX/NOX/H2S	kg/hr	8														0.5
Cellulose (IS)	kg/hr														3	
Xylan (IS)	kg/hr									325					1	
Other Struct. Carbohydr. (IS)	kg/hr														0.2	
Acetate (IS)	kg/hr									1,327						
Lignin (IS)	kg/hr														25	
Protein (IS)	kg/hr	21													7	
Cell Mass (IS)	kg/hr														0.1	
Other Insoluble Solids (IS)	kg/hr									3,081					3	

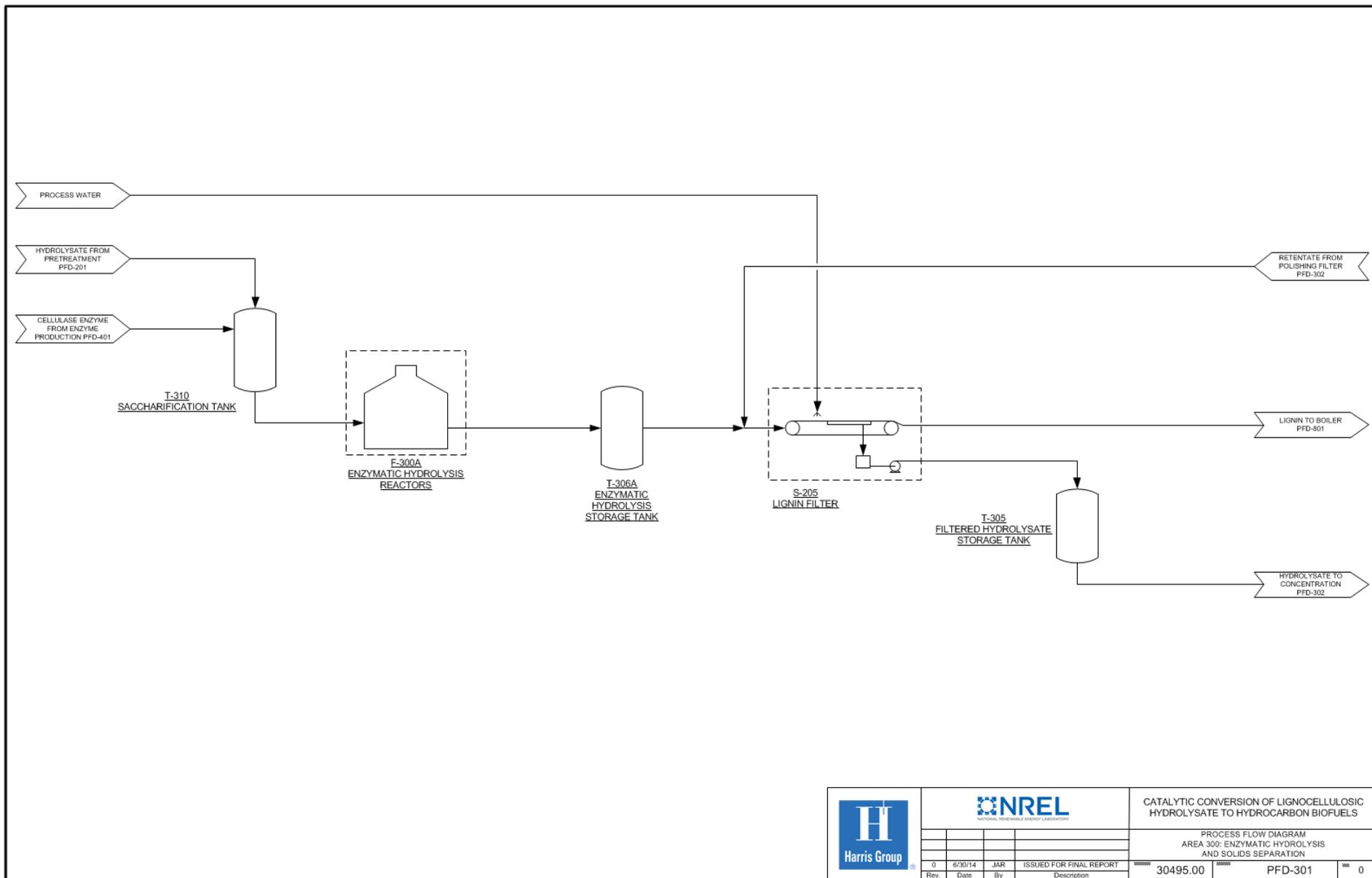
COMPONENT	UNITS	670	700	800	810	820	830	840	850	860	870	880	890a	890b	890C
Total Flow	kg/hr	182,522	20,047	43,639	167	5,132	19,459	13,138	33,769	180	317,859	0	395,788	4,506	2,705
Insoluble Solids	%	0.0%	0.0%	36.0%	0.0%	0.0%	0.0%	36.6%	0.0%	100.0%	0.0%	0.0%	0.0%	100.0%	0.0%
Soluble Solids	%	0.0%	0.0%	1.5%	0.0%	0.0%	0.0%	3.0%	0.0%	0.0%	0.0%	0.0%	0.1%	0.0%	0.0%
Temperature	°C	27	31	29	29	33	37	30	66	25	25	25	155		100
Pressure	atm	1.0	2.0	1.0	1.0	1.5	1.0	2.0	10.4	1.0	1.0	1.0	1.0	1.0	1.0
Vapor Fraction		0.0	0.0	0.0	1.0	1.0	1.0	0.0	0.0		1.0	1.0	1.0		1.0
Fuel-range hydrocarbons	kg/hr		19,985			1,353							1		
Water	kg/hr	182,522	5	27,256	4	72	624	7,883	33,769		6,225		71,871		2,705
Glucose (SS)	kg/hr			311				2							
Xylose (SS)	kg/hr			172				1							
Other sugars (SS)	kg/hr			42				2					0		
Sugar Oligomers (SS)	kg/hr			17				0.4							
Organic Soluble Solids (SS)	kg/hr			25				162							
Inorganic Soluble Solids (SS)	kg/hr			88				234					223		
Ammonia	kg/hr						1	0.1					1		
Acetic Acid	kg/hr														
Sulfuric Acid	kg/hr							23							
Furfurals	kg/hr			5			1	0.3							
Other Organics	kg/hr		44	0.1		403	216	20					0.1		
Carbon Dioxide	kg/hr		13			2,666	13,516	1					71,585		
Methane	kg/hr						5,097								
H2	kg/hr					638							6		
O2	kg/hr	0.1		0.2	38		1				72,585		12,074		
N2	kg/hr	0.2		0.4	125		2	0.1			239,049		239,911		
CO/SOX/NOX/H2S	kg/hr												116		
Cellulose (IS)	kg/hr			1,256				3							
Xylan (IS)	kg/hr			414				327							
Other Struct. Carbohydr. (IS)	kg/hr			95				0.2							
Acetate (IS)	kg/hr														
Lignin (IS)	kg/hr			9,955				25							
Protein (IS)	kg/hr			2,887				5							
Cell Mass (IS)	kg/hr			54				1,366							
Other Insoluble Solids (IS)	kg/hr			1,061				3,084		180				4,506	



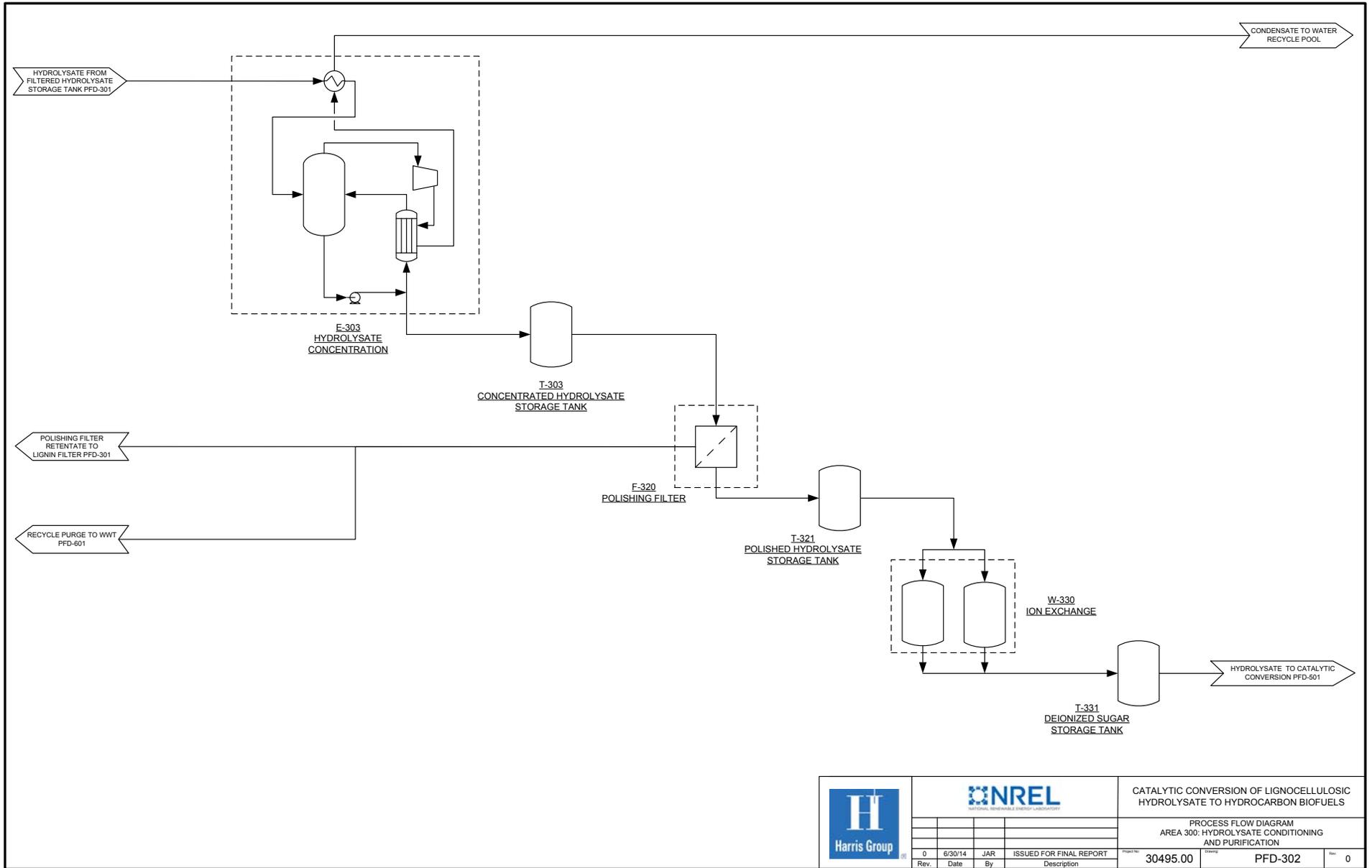
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0	6/30/14	JAR	ISSUED FOR FINAL REPORT	30495.00	PFD-101	0
Rev.	Date	By	Description			



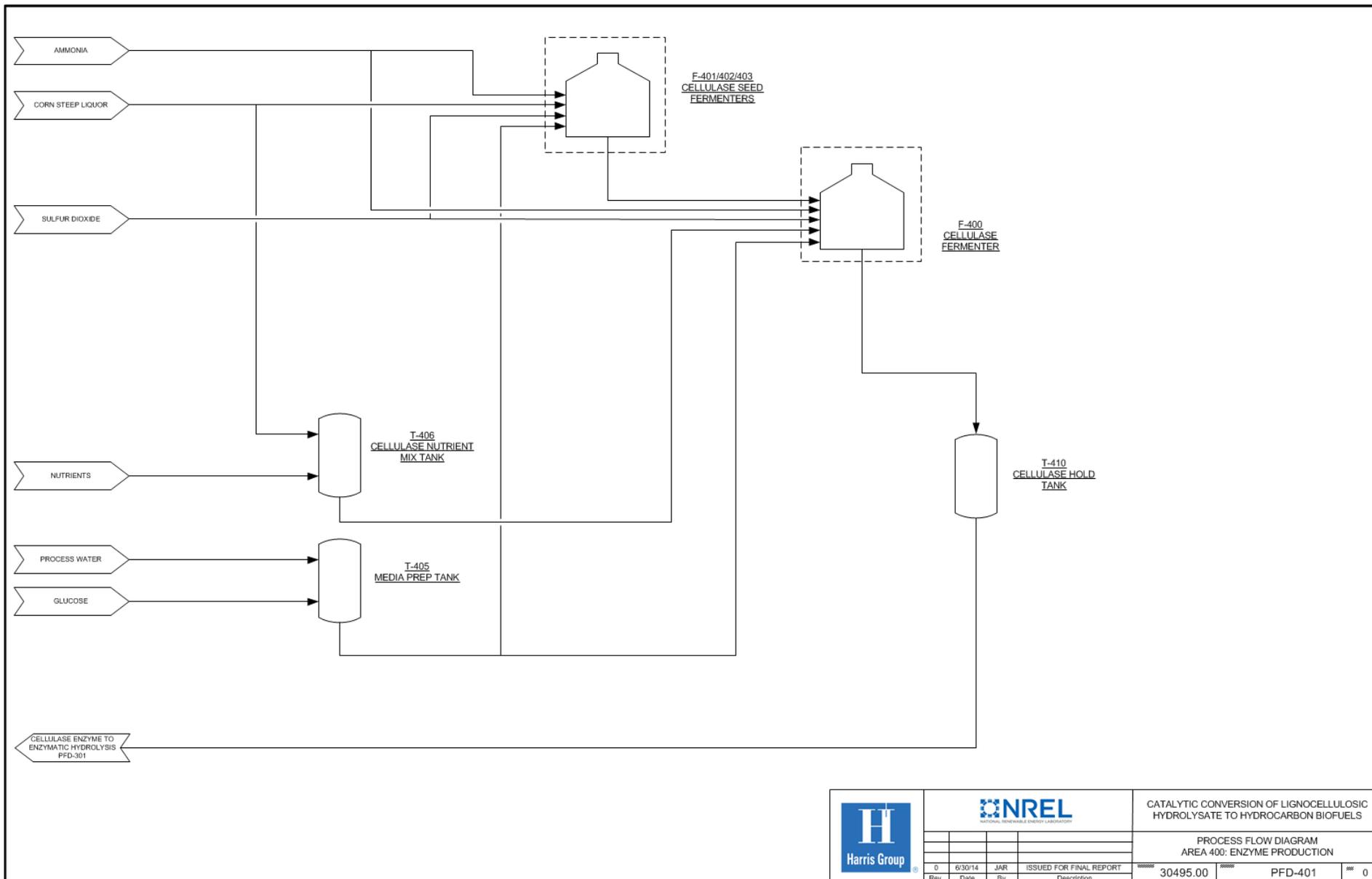
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	PROCESS FLOW DIAGRAM AREA 200: PRETREATMENT AND CONDITIONING					
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Rev.	Date	By	Description			



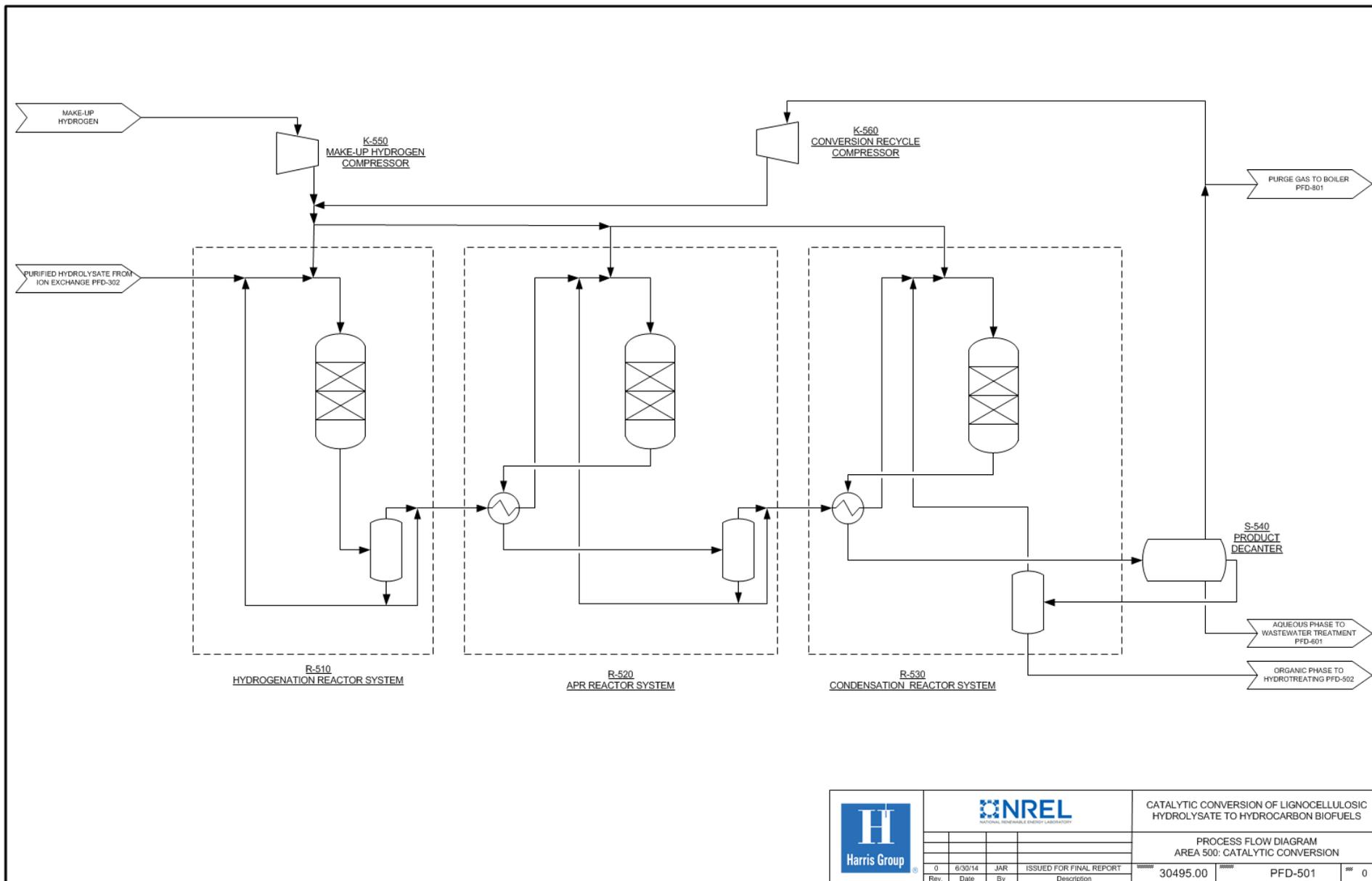
			CATALYTIC CONVERSION OF LIGNOCELLULOSIC HYDROLYSATE TO HYDROCARBON BIOFUELS	
	PROCESS FLOW DIAGRAM AREA 300: ENZYMIC HYDROLYSIS AND SOLIDS SEPARATION			
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Rtv.	Date	By	Description	PFD-301



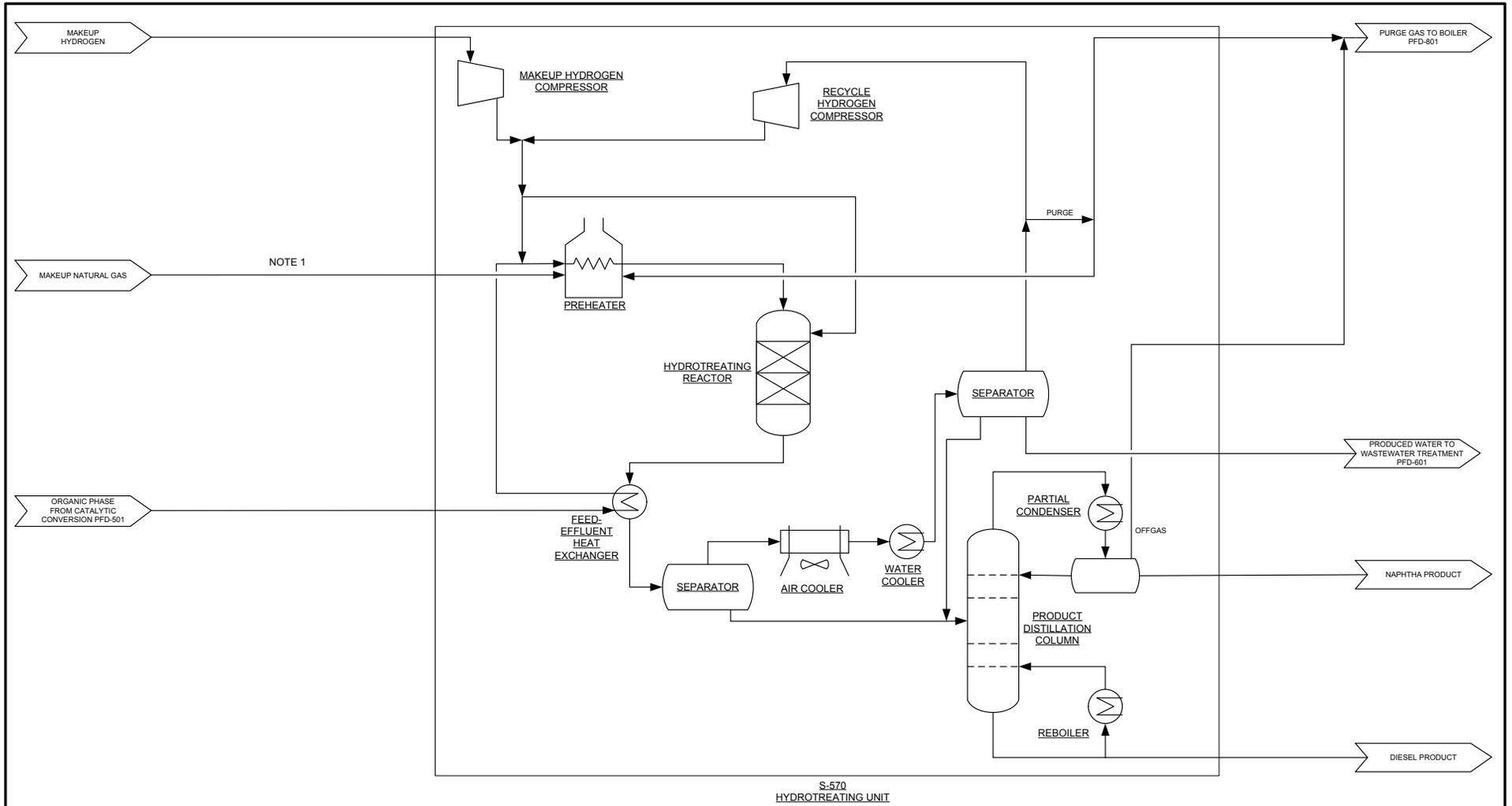
			CATALYTIC CONVERSION OF LIGNOCELLULOSIC HYDROLYSATE TO HYDROCARBON BIOFUELS		
	PROCESS FLOW DIAGRAM AREA 300: HYDROLYSATE CONDITIONING AND PURIFICATION				
0	6/30/14	JAR	ISSUED FOR FINAL REPORT	30495.00	PFD-302
Rev.	Date	By	Description	Quantity	Rev



				CATALYTIC CONVERSION OF LIGNOCELLULOSIC HYDROLYSATE TO HYDROCARBON BIOFUELS		
	PROCESS FLOW DIAGRAM AREA 400: ENZYME PRODUCTION					
0	6/30/14	JAR	ISSUED FOR FINAL REPORT	30495.00	PFD-401	0
Rev	Date	By	Description			

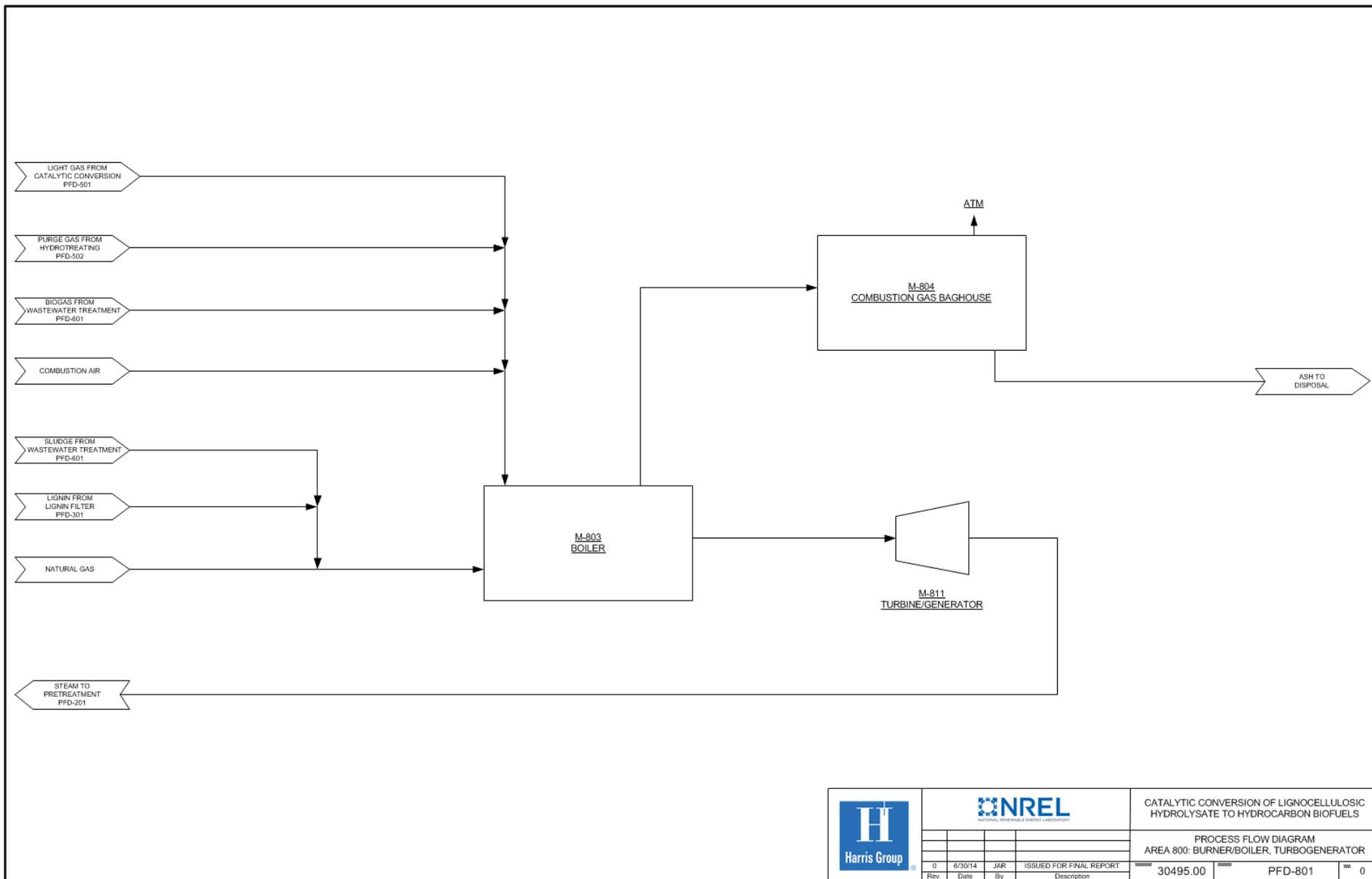


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	PROCESS FLOW DIAGRAM AREA 500: CATALYTIC CONVERSION					
0	6/30/14	JAR	ISSUED FOR FINAL REPORT	30495.00	PFD-501	0
Rev.	Date	By	Description			



NOTES:
1. NATURAL GAS FOR STARTUP AND AS NEEDED

				CATALYTIC CONVERSION OF LIGNOCELLULOSIC HYDROLYSATE TO HYDROCARBON BIOFUELS		
	PROCESS FLOW DIAGRAM AREA 500: UPGRADING					
0	6/30/14	JAR	ISSUED FOR FINAL REPORT	Project No.	30495.00	Process
Rev.	Date	By	Description			PFD-502
						Rev 0



				CATALYTIC CONVERSION OF LIGNOCELLULOSIC HYDROLYSATE TO HYDROCARBON BIOFUELS		
	PROCESS FLOW DIAGRAM AREA 800: BURNER/BOILER, TURBOGENERATOR					
0	6/30/14	JAR	ISSUED FOR FINAL REPORT	30495.00	PFD-801	0
Rev	Date	By	Description			