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Technical Report
NREL/TP-5100-62368
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Executive Summary

The U.S. Department of Energy (DOE) promotes the production of a range of liquid fuels and fuel blendstocks from biomass feedstocks by funding fundamental and applied research that advances the state of technology in biomass production, 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. This includes fuel pathways from lignocellulosic (terrestrial) biomass, as well as from algal (aquatic) biomass systems.

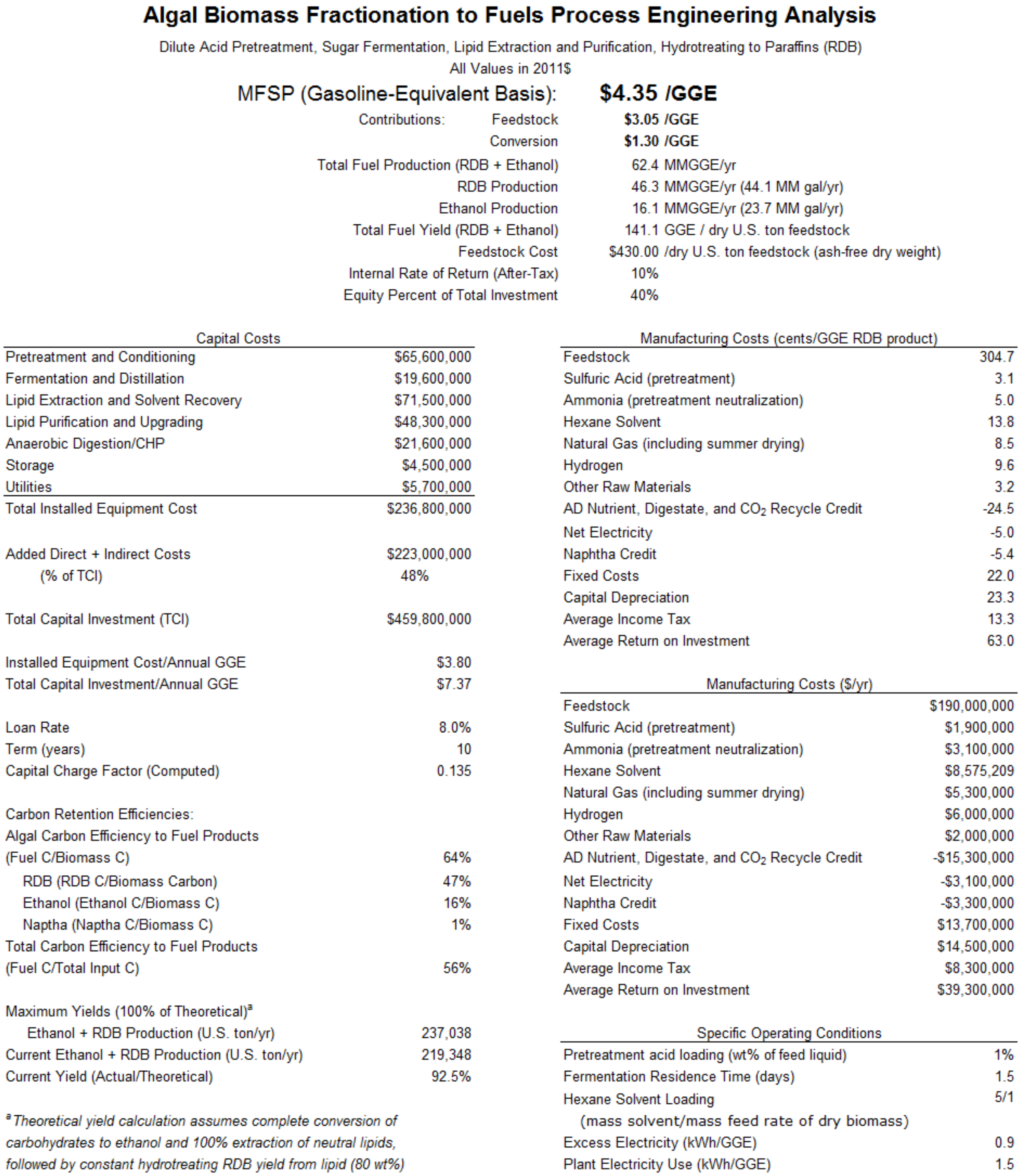
Over the past decade, NREL conducted a campaign to quantify the economic implications associated with observed and aspirational performance for the conversion of corn stover to ethanol through techno-economic modeling. This effort served two important purposes: (1) to establish a benchmark representing the initial "state of technology" at the time, and (2) to set goals for near-term R&D and cost targets, as well as to track progress toward achieving these targets by periodically updating the models based on the latest research improvements. 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. One of these pathways is the conversion of algal biomass to fuels via extraction of lipids (and potentially other components), termed the "algal lipid upgrading" or ALU pathway.

This report describes in detail one potential ALU approach based on a biochemical processing strategy to selectively recover and convert select algal biomass components to fuels, namely carbohydrates to ethanol and lipids to a renewable diesel blendstock (RDB) product. The overarching process design converts algal biomass delivered from upstream cultivation and dewatering (outside the present scope) to ethanol, RDB, and minor coproducts, using dilute-acid pretreatment, fermentation, lipid extraction, and hydrotreating. Ancillary areas—anaerobic digestion of spent algal residues, combined heat and power generation, 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 case study techno-economic model provides a production cost for the fuel products that can be used to gauge the technology potential and to quantify critical cost drivers. The analysis presented here also includes consideration of life-cycle implications by tracking environmental sustainability metrics for the modeled conversion step, including greenhouse gas emissions, fossil energy demand, and consumptive water use.

Following similar methods for design report modeling as conducted over recent years, NREL, supported by Harris Group Inc., performed a feasibility-level analysis for a plausible conversion pathway to meet a cost goal below $5/gallon gasoline equivalent (GGE). The modeled facility processes an average of 1,339 dry ash-free ton biomass/day and achieves an overall fuel selling price of $4.35/GGE in 2011-dollars as determined by modeled conversion targets and "n^{th}-plant" project costs and financing, associated with a net fuel yield of 141 GGE/dry ton. These values are attributed to a high (41%) lipid, late-harvest biomass scenario, which implies aggressive future targets in achieving simultaneously high cultivation productivity rates during upstream biomass growth, implicit in the year 2022 target projections stipulated in this design. Consequently, an alternative scenario was also considered based on lower (27%) lipid, earlier-harvest biomass, which resulted in fractionally higher costs at $5.04/GGE associated with a lower overall fuel yield of 116 GGE/dry ton. As this alternate case is based on a lipid content

more typically achievable today, a realistic future target case may in fact fall between these two points as an optimum between aspirational goals for algal growth rate and lipid content.

The analysis also indicates that economics are influenced strongly by feedstock costs, contributing 70% to overall selling price at $430/ton. The report reiterates prior findings that seasonality plays a unique and critical role in algal systems, with high variation in upstream algal cultivation translating to quantified tradeoffs in cost and sustainability metrics depending on how the facility is designed to accommodate such variability. In addition, the report includes a high-level discussion on improvements needed to achieve an ultimate target of $3/GGE moving forward, including unique advantages inherent to this particular processing strategy in enabling numerous alternative product or coproduct routes.



Algal Biomass Fractionation to Fuels Process Engineering Analysis

Dilute Acid Pretreatment, Sugar Fermentation, Lipid Extraction and Purification, Hydrotreating to Paraffins (RDB)
All Values in 2011$

MFSP (Gasoline-Equivalent Basis):		**$4.35 /GGE**
Contributions:	Feedstock	**$3.05 /GGE**
	Conversion	**$1.30 /GGE**
Total Fuel Production (RDB + Ethanol)		62.4 MMGGE/yr
RDB Production		46.3 MMGGE/yr (44.1 MM gal/yr)
Ethanol Production		16.1 MMGGE/yr (23.7 MM gal/yr)
Total Fuel Yield (RDB + Ethanol)		141.1 GGE / dry U.S. ton feedstock
Feedstock Cost		$430.00 /dry U.S. ton feedstock (ash-free dry weight)
Internal Rate of Return (After-Tax)		10%
Equity Percent of Total Investment		40%

Capital Costs	
Pretreatment and Conditioning	$65,600,000
Fermentation and Distillation	$19,600,000
Lipid Extraction and Solvent Recovery	$71,500,000
Lipid Purification and Upgrading	$48,300,000
Anaerobic Digestion/CHP	$21,600,000
Storage	$4,500,000
Utilities	$5,700,000
Total Installed Equipment Cost	$236,800,000
Added Direct + Indirect Costs	$223,000,000
(% of TCI)	48%
Total Capital Investment (TCI)	$459,800,000
Installed Equipment Cost/Annual GGE	$3.80
Total Capital Investment/Annual GGE	$7.37
Loan Rate	8.0%
Term (years)	10
Capital Charge Factor (Computed)	0.135
Carbon Retention Efficiencies:	
Algal Carbon Efficiency to Fuel Products	
(Fuel C/Biomass C)	64%
RDB (RDB C/Biomass Carbon)	47%
Ethanol (Ethanol C/Biomass C)	16%
Naptha (Naptha C/Biomass C)	1%
Total Carbon Efficiency to Fuel Products	
(Fuel C/Total Input C)	56%
Maximum Yields (100% of Theoretical)[a]	
Ethanol + RDB Production (U.S. ton/yr)	237,038
Current Ethanol + RDB Production (U.S. ton/yr)	219,348
Current Yield (Actual/Theoretical)	92.5%

[a] *Theoretical yield calculation assumes complete conversion of carbohydrates to ethanol and 100% extraction of neutral lipids, followed by constant hydrotreating RDB yield from lipid (80 wt%)*

Manufacturing Costs (cents/GGE RDB product)	
Feedstock	304.7
Sulfuric Acid (pretreatment)	3.1
Ammonia (pretreatment neutralization)	5.0
Hexane Solvent	13.8
Natural Gas (including summer drying)	8.5
Hydrogen	9.6
Other Raw Materials	3.2
AD Nutrient, Digestate, and CO_2 Recycle Credit	-24.5
Net Electricity	-5.0
Naphtha Credit	-5.4
Fixed Costs	22.0
Capital Depreciation	23.3
Average Income Tax	13.3
Average Return on Investment	63.0

Manufacturing Costs ($/yr)	
Feedstock	$190,000,000
Sulfuric Acid (pretreatment)	$1,900,000
Ammonia (pretreatment neutralization)	$3,100,000
Hexane Solvent	$8,575,209
Natural Gas (including summer drying)	$5,300,000
Hydrogen	$6,000,000
Other Raw Materials	$2,000,000
AD Nutrient, Digestate, and CO_2 Recycle Credit	-$15,300,000
Net Electricity	-$3,100,000
Naphtha Credit	-$3,300,000
Fixed Costs	$13,700,000
Capital Depreciation	$14,500,000
Average Income Tax	$8,300,000
Average Return on Investment	$39,300,000

Specific Operating Conditions	
Pretreatment acid loading (wt% of feed liquid)	1%
Fermentation Residence Time (days)	1.5
Hexane Solvent Loading (mass solvent/mass feed rate of dry biomass)	5/1
Excess Electricity (kWh/GGE)	0.9
Plant Electricity Use (kWh/GGE)	1.5

Figure ES-1. Economic summary for ALU pathway

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Acronyms

AD	anaerobic digestion
AFDW	ash-free dry weight
ALU	algal lipid upgrading
ANL	Argonne National Laboratory
ATP3	Algae Testbed Public-Private Partnership
BETO	Bioenergy Technologies Office
BFW	boiler feed water
BGY	billion gallons per year
CAPEX	capital expenditure
CHP	combined heat and power
CIP	clean-in-place
COD	chemical oxygen demand
CSL	corn steep liquor
DAP	diammonium phosphate
DB	declining balance
DCFROR	discounted cash flow rate of return
DCO	decarboxylation
DOE	U.S. Department of Energy
FAME	fatty acid methyl ester
FCI	fixed capital investment
FFA	free fatty acid
GGE	gallon gasoline equivalent
GHG	greenhouse gas
HCSD	high-carbohydrate (mid-harvest) *Scenedesmus*
HDO	hydrodeoxygenation
HHV	higher heating value
HLSD	high-lipid (late harvest) *Scenedesmus*
HMF	5-hydroxymethyl furfural
HRSG	heat recovery steam generator
HRT	hydraulic retention time
HTL	hydrothermal liquefaction
IPCC	Intergovernmental Panel on Climate Change
IRR	internal rate of return
ISBL	inside battery limits (of the plant)
LCA	life-cycle assessment
LHV	lower heating value
LHSV	liquid hourly space velocity
MACRS	IRS Modified Accelerated Cost Recovery System
MFSP	minimum fuel selling price
MM	million (e.g., MMBTU or $__MM)
MYPP	DOE-BETO's Multi-Year Program Plan
NG	natural gas
NPV	net present value
NREL	National Renewable Energy Laboratory
OPEX	operating expenditure
OSBL	outside battery limits
PFD	process flow diagram
PNNL	Pacific Northwest National Laboratory
PSA	pressure swing adsorption
R&D	research and development
RA	resource assessment
RDB	renewable diesel blendstock
SABC	Sustainable Algal Biofuels Consortium
SCFM	standard cubic feet per minute
SMR	steam methane reforming
SOT	annual State of Technology case
TAN	total acid number
TCI	total capital investment
TDC	total direct cost
TEA	techno-economic analysis
TS	total solids
VOC	volatile organic compound
WWT	wastewater treatment

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 and algal feedstocks by sponsoring programs in fundamental and applied research that aim to advance the state of technology spanning the supply chain from biomass production through processing and conversion to fuels. Within the algae platform, these programs include laboratory research to improve biological characteristics (e.g., algal cultivation productivity, biomass composition, strain robustness) through screening and synthetic biology; engineering studies of potential systems for growth, dewatering, and conversion technologies; improvement of laboratory analytical capabilities to accurately characterize feed and product materials; and support for biomass production test-bed and processing demonstration facilities. This research is conducted by national laboratories, universities, and private industry, both individually and through multi-organization consortia partnerships.

To support the DOE program, the National Renewable Energy Laboratory (NREL) investigates the process design and economics of numerous biofuel manufacturing pathways in order to develop an absolute 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 biofuel technology in comparison with petroleum-derived fuels and established biofuel pathways such as starch- or sugar-based ethanol, vegetable oil-based biodiesel, etc.

The techno-economic analysis 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 established benchmark cases. Such comparison helps to quantify the economic impact of core research targets 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 costs.

For more than 10 years, NREL has developed design case models and associated reports that document process and cost targets focused on *cellulosic* (terrestrial) biofuel pathways, focused first on ethanol production and more recently on hydrocarbon biofuel production [1-6], based on the best understanding of the technology and equipment costs at the time. Over the past four years, DOE-BETO expanded its focus to include *algal* biofuel pathways in light of renewed industry interest in these pathways and the promising potential across metrics such as high biomass yields, flexible product slates, and the ability to utilize low-value land and water resources that need not necessarily compete with food-based crops [7]. For one of the first analysis activities under this renewed focus, DOE chartered a collaborative “Harmonization Initiative” which brought together modeling partners from NREL, Argonne National Laboratory (ANL), and Pacific Northwest National Laboratory (PNNL) to harmonize their conceptual models around techno-economic analysis (TEA), life-cycle assessment (LCA), and resource assessment (RA) respectively, such that the results from each model would carry the same implications being based on consistent inputs and assumptions. This effort included a workshop to serve as a vetting process for the respective collaborators to present the details of their models to a group of expert stakeholders in industry, academia, and other national laboratories, in order to begin validating or improving key modeling assumptions. This ultimately resulted in the publication of a harmonization report (referred to as such hereafter) documenting model details and the resulting near-term cost, sustainability, and resource implications for production of 5 billion gallons per year (BGY) of renewable diesel at the national scale spread across a large consortium of individual unit farms. The harmonization scenario was based on an “algal lipid extraction and upgrading” (ALU) process focused on extracting algal lipids and routing all remaining residual material to anaerobic digestion. Details of the 2012 “ALU harmonization report” may be found in [8] and will not be repeated here, save but to emphasize the key result from the analysis, which was that algae as a feedstock possess a unique challenge given high seasonal variability in cultivation productivity rates, varying up to tenfold between summer and winter; such variations exhibit important impacts on TEA and LCA evaluations and must be considered in such analyses.

The 2012 ALU harmonization effort was subsequently leveraged in a number of ways, first by DOE-BETO in translating the results to an initial baseline and projecting out-year process and cost improvement goals [9], and next by being expanded on in 2013 to repeat a similar harmonization focused on hydrothermal liquefaction (HTL) conversion of the algal biomass [10]. The newer 2013 harmonization scenarios were extrapolated out to a design report focused on year 2022 targets for the algal HTL conversion pathway [11] (referred to hereafter as the “HTL design report”), which established a modeled MFSP target of $4.49/gallon gasoline equivalent (GGE) for the pathway. The present report presents a comparative TEA analysis for an alternative approach to achieving a similar MFSP cost, by way of a targeted process to fractionate algal biomass and selectively convert the major biomass constituents to fuel products, specifically carbohydrates to ethanol and lipids to renewable diesel blendstock (RDB). In contrast to HTL conversion, which is a thermochemical-based approach to convert the whole algal biomass feedstock to a hydrocarbon (primarily diesel-range) intermediate fuel precursor, the fractionation process considered here is based on biochemical conversion principles whereby specific components of the biomass are selectively targeted to specific fuel products via biological and chemical/physical pathways.

As with any technology pathway, there are strengths and challenges for both HTL processing as well as the present fractionation processing approach. In the case of HTL, the technology translates to high carbon-retention efficiency to fuels by way of converting biomass carbon beyond only lipids, but to do so requires operating at high pressure (approximately 3,000 psia as documented in [11]) and as with any thermochemical processing technique, does not allow for stringent control over heteroatom partitioning (namely nitrogen and phosphorus) into the organic biocrude phase amongst other product phases. Early studies have indicated that low-level blends of HTL biocrude on the order of 10% or less co-processed with petroleum feedstocks allow for more mild operating conditions similar to standard petroleum hydrotreating [12]. In the case of the biochemical fractionation approach considered here, our analysis finds that this approach also allows for high carbon efficiencies by fractionating and then converting both carbohydrates and lipids, which constitute a large fraction of the biomass for mid- to late-stage harvesting practices; however, this pathway also relies more heavily on lipid content to maximize fuel energy yields, which translates to more challenging algal productivity growth rates during upstream cultivation. Additionally, conditions required for release of fermentable sugars may vary with species and upstream cultivation conditions. Key benefits for this fractionation process may be summarized as follows:

- Uses established and proven technologies for fractionation (dilute acid pretreatment/hydrolysis), carbohydrate utilization (fermentation), and spent residual biomass utilization (anaerobic digestion), with all aspects of the process operating at reasonable conditions; maximum operating pressure for this system is 70 psia, except for the hydrotreater which operates at 465 psia as typical for mild hydroprocessing.
- Is highly amenable to fermentative conversion of sugars to a range of biological products (ethanol or hydrocarbons), as algal biomass typically contains low levels of pentose sugars that are not readily metabolized by most organisms without genetic engineering [13]. The algal biomass is also less recalcitrant to biochemical deconstruction than lignocellulosic feedstock pathways, and produces relatively clean sugars.
- Does not require thermal evaporation (drying) for any aspect of the conversion process, with efficacy of all unit operations experimentally demonstrated based on a starting 20% biomass solids content delivered to fractionation [13].
- Allows for a high degree of flexibility in product/coproduct options, namely conversion of sugars to a wide range of fuels or chemicals, as well as leaving a potentially high-value protein residue that may be converted to additional coproducts.
- Selectively converts specific components to specific, easily-characterized products with mild subsequent upgrading demands (hydrotreating at approximately 500 psig, in the case of the lipid product).

The focus of this report is to document a plausible pathway model for conversion of algal carbohydrates and lipids to fuel and blendstock products, with high fractional energy yield to hydrocarbon products (e.g., renewable diesel) supplemented by additional energy yield to ethanol as a representative fermentative product from sugars; primarily to demonstrate a means to achieve modeled MFSP costs under $5/GGE by 2022, but with additional insight provided towards a path forward to ultimately reduce costs to $3/GGE. This analysis leverages the decades of experience that NREL has established in biochemical conversion research, primarily with respect to dilute acid biomass pretreatment to hydrolyze carbohydrates as well as fermentation of sugars. Likewise, the work also continues NREL's practice of consulting with vendors through the assistance of an engineering company to assist in design and cost estimation for critical unit operations. For the present report, NREL worked with Harris Group to provide engineering support primarily for new unit operations that had not been previously used in other cellulosic technology designs (such as extraction and algal lipid cleanup/hydrotreating), while also leveraging vendor-supplied cost estimates supplied by Harris Group for existing unit operations (such as dilute acid pretreatment, fermentation, distillation, and anaerobic digestion) [3, 6]. Thus the economics of this conceptual process uses the best available equipment and raw material costs and an "*n*th-plant" project cost structure and financing. The projected 2022 *n*th-plant MFSP computed in this report is $4.35/GGE ($4.57/gal RDB and $2.95/gal ethanol) in 2011-year dollars.

Similar to caveats noted in prior NREL design reports, we stress that this design report serves to describe a *single, feasible* conversion process and to transparently document the assumptions and details that went into its design. This report is not meant to provide an exhaustive survey of process alternatives or cost-sensitivity analyses. Furthermore, it is important to note that algal biofuels as a technology platform are in a more nascent stage of development and understanding than other more-established biofuel pathways (e.g., from terrestrial biomass), and continue to suffer from a dearth of available public data on processing performance at sufficient scale. While the key inputs assumed in this analysis are extrapolated from data gathered through experimental work conducted at NREL and the recently-completed Sustainable Algal Biofuels Consortium (SABC) [14], further experimental work and scale-up is required to reduce model uncertainties. Moving forward, as the science and technology progresses for less well-studied areas of the process such as lipid extraction or alternative conversion routes of sugars to hydrocarbon products, the process models and economic tools developed for this report may be updated in a similar fashion as prior NREL design report iterations have evolved.

1.2 Process Overview

The process described here uses cocurrent dilute-acid pretreatment of algal biomass delivered after upstream dewatering to 20 wt% solids (outside the scope of this analysis), followed by whole-slurry fermentation of the resulting monomeric sugars to ethanol, followed by distillation and solvent extraction of the stillage to recover lipids (primarily fatty acid-based lipids with inclusion of polar lipid impurities). The process design also includes lipid product purification, product upgrading (hydrotreating) to straight-chain paraffin blend stocks, anaerobic digestion and combined heat and power (CHP) generation, product storage, and required utilities. The process is divided into seven areas (see Figure 1).

- *Area 100: Pretreatment and conditioning.* In this area, the algal biomass is combined with steam and treated with dilute sulfuric acid catalyst at a high temperature for a short time to hydrolyze the glucan carbohydrates to monomeric sugars and make the remaining biomass more amenable to lipid extraction. Ammonia is then added to the whole pretreated slurry to raise its pH to ~5 for fermentation.

- *Area 200: Fermentation and distillation.* After flashing the pretreatment hydrolysate back to 20% total solids (TS), the slurry is cooled and inoculated with the fermenting organism *Saccharomyces cerevisiae*. Enzymatic hydrolysis is not required, as hydrolysis (by acid) was already accomplished in Area 100. Fermentation proceeds in batch mode for 1.5 days, after which point most of the sugars (primarily the hexose sugars glucose and mannose) are converted to ethanol. The resulting dilute ethanol broth is distilled to near-azeotropic concentration and then purified to 99.5% using vapor-phase molecular sieve adsorption. The beer stillage containing all residual solids is routed to lipid extraction; while water from the rectification column is sent to the water recycle pool. CO_2 vented from the fermentation step is recycled to upstream algal cultivation ponds.

- *Area 300: Lipid extraction and solvent recovery.* The distillation stillage is fed to a countercurrent liquid-liquid extraction column along with hexane solvent required at a 5:1 mass ratio (solvent versus dry algal solids in the stillage feed). The light product phase containing primarily solvent and oil is routed to a distillation stripping column to recover the solvent, with marginal solvent losses to the bottoms product as well as the extraction aqueous phase. The resulting distillation bottoms product is composed of >99.5% oil (neutral and polar lipids) and is sent to lipid purification to remove polar lipids and other impurities. The aqueous phase is routed to anaerobic digestion.

- *Area 400: Product purification and upgrading.* The recovered oil product is sent to a series of purification steps consisting of degumming, demetallization, and bleaching to remove phospholipids and other polar lipids, metals, salts, and other impurities with the addition of phosphoric acid, wash water, silica, and clay. The purified oil, consisting primarily of fatty acid based lipids, is sent to product upgrading in an on-site hydrotreating facility (including reactors, fresh and recycle gas compressors, flash columns, and product fractionation, utilizing purchased hydrogen). The primary product from the hydrotreating section is a diesel-range paraffinic product suitable as a diesel blendstock (RDB) with a small naphtha coproduct. The hydrotreating section also includes a PSA unit in the recycle gas loop to remove CO_2 generated during decarboxylation.
- *Area 500: Anaerobic digestion/CHP.* The residual raffinate product stream from extraction is combined with the oil cleanup waste stream and sent to anaerobic digestion, primarily utilized as a means to reclaim carbon via biogas production as well as enable nutrient recycle coproduct credits intended to be routed back to upstream algal cultivation (outside the scope of this model). The methane-rich biogas is combusted in a gas turbine to generate electricity, which is produced in excess of facility power demands and sold to the grid. Flue gas heat is recovered by generating steam to satisfy process and utility steam demands. A small amount of supplemental natural gas is co-fed with the biogas to the turbines to meet facility heat balances. The digester effluent water contains nitrogen and phosphorus nutrients which garner additional coproduct revenues, as does the solids digestate cake which is sold for the nitrogen content as a land-application fertilizer.
- *Area 600: Storage.* This area provides bulk storage for diesel, naphtha, and ethanol products, as well as firewater and chemical inputs.
- *Area 700: Utilities.* This area includes a cooling water system, chilled water system, process water manifold, and power systems. The steam generation units and steam utility loops are also included in this section.

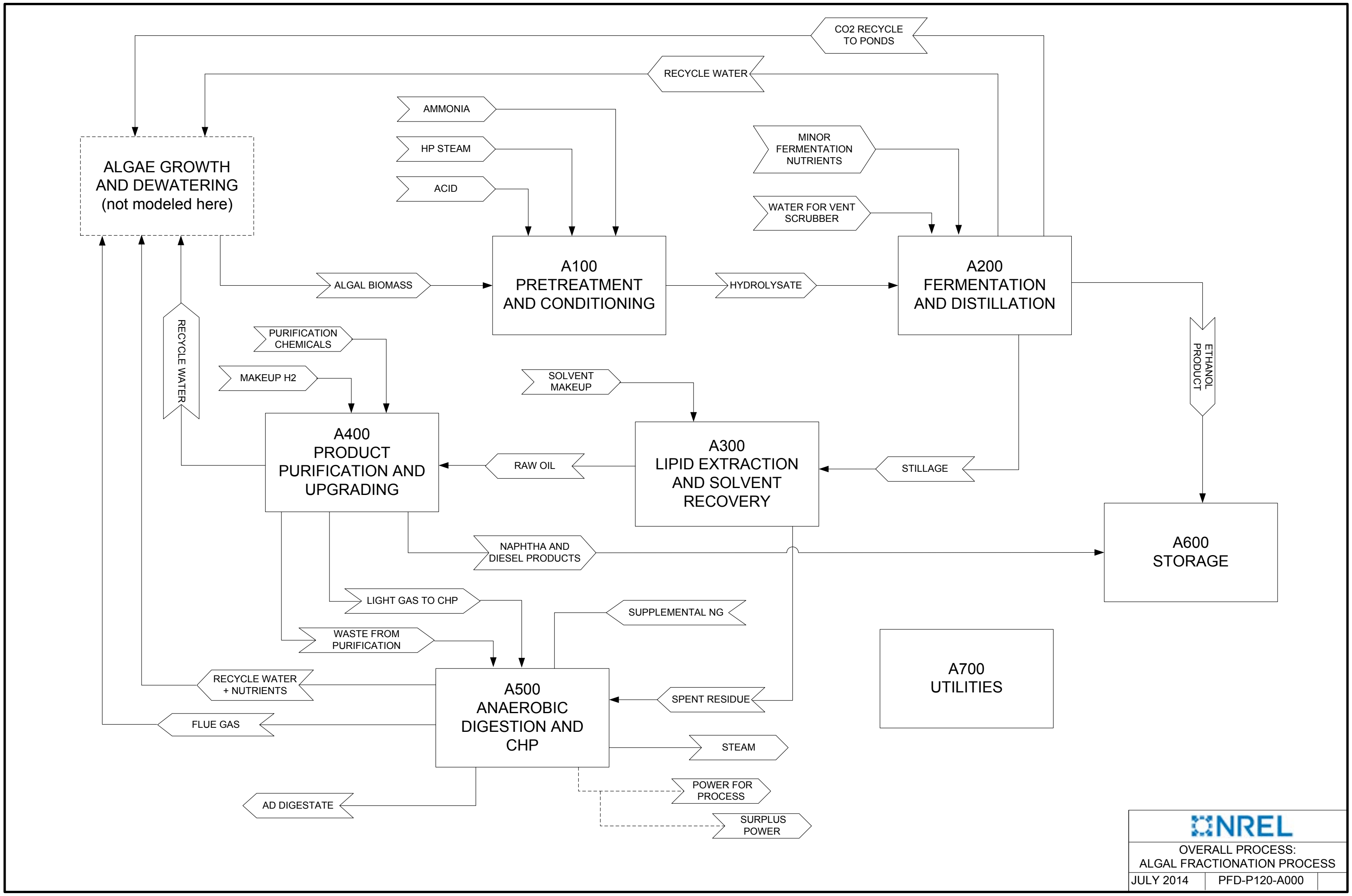


Figure 1. Simplified flow diagram of the overall process. (Key streams only; see Appendix D for more detailed diagram and PFDs.)

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 extrapolations around areas with more limited supporting experimental data to date (primarily the extraction step as well as operability and yield assumptions related to anaerobic digestion of a material with elevated nitrogen, discussed later). As such, this report is less prescriptive in some sections than previous reports due to the early stage of understanding for this novel process pathway and somewhat more preliminary nature of the associated models.

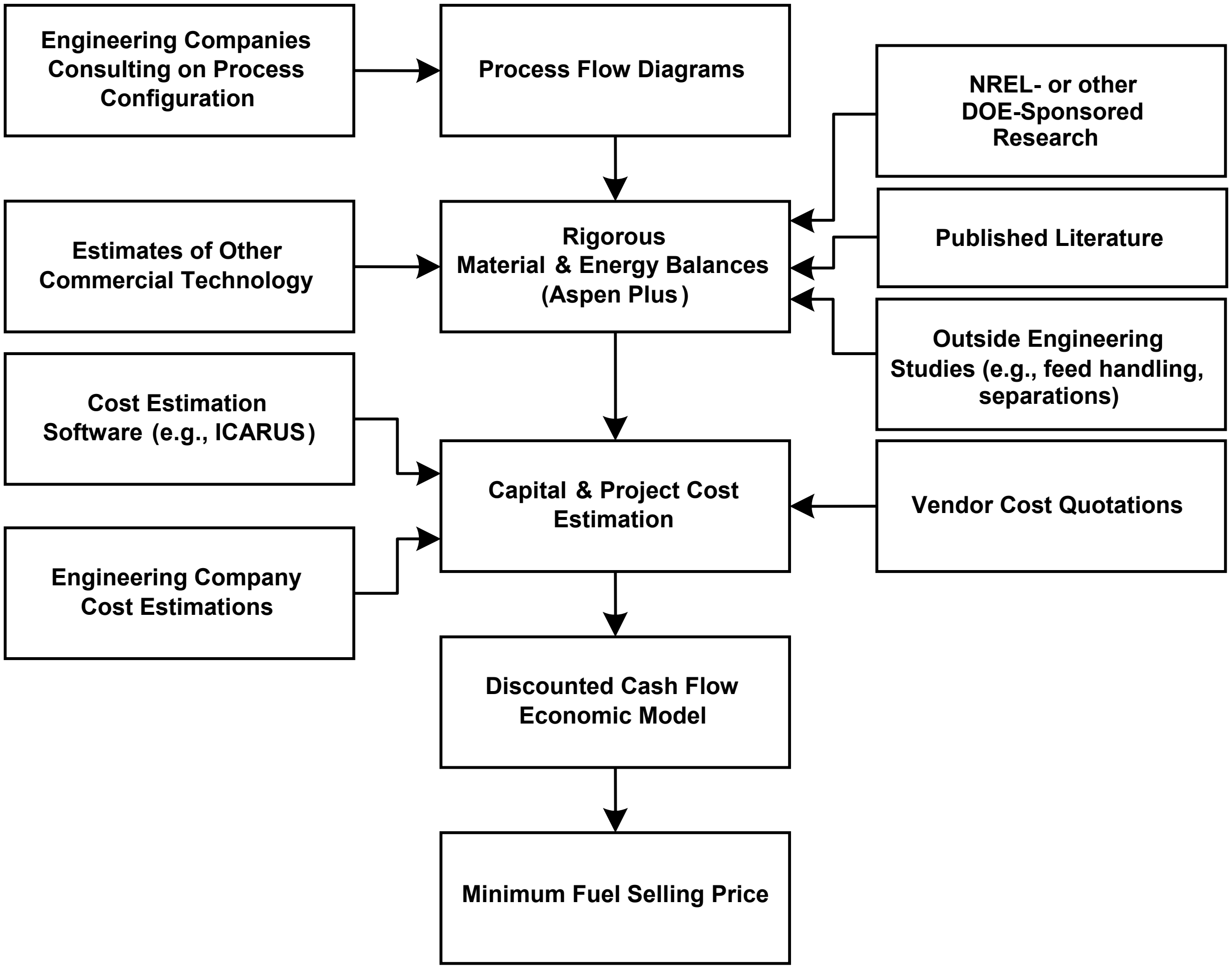


Figure 2. NREL's approach to process design and economic analysis

Starting from the general process flow diagram (PFD) shown in Figure 2 and the more detailed PFDs contained in Appendix D, a process simulation is developed using Aspen Plus software [15]. This process model computes thermodynamically rigorous material and energy balances for each unit operation in this conceptual integrated conversion process. 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 cost estimates come from vendor quotes typically based on AACE Class 4 estimates (a favored procedure for larger or non-standard unit operations and packaged or skid-mounted subsystems) or from Harris Group's historical cost database (for secondary equipment such as tanks, pumps, and heat exchangers). Harris Group provided new design and cost estimates based on vendor quotations for critical areas of the process model for which similar unit operations had not previously been considered in other analyses, including a liquid-liquid extraction system, solvent distillation, and oil cleanup operations. Other process steps, such as dilute acid pretreatment (fractionation) and neutralization, fermentation, ethanol purification, anaerobic digestion, and hydrotreating, were based on recently updated Harris Group estimates from NREL's 2011 cellulosic ethanol design report and 2013 biological hydrocarbon design report [3,6], referred to as such hereafter. 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 total capital investment (TCI). 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 the refined renewable diesel blendstock and ethanol products, combined together based on energy content of each stream. This 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.

The product of the analysis described above is a techno-economic model that reasonably 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 a certain percentage 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 absolute computed MFSP carries a risk of being taken out of context. While the MFSP can be used to assess the marketplace competiveness 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, *nth*-plant analysis avoids artificial inflation of project costs associated with risk financing, longer start-ups, equipment overdesign, and other costs associated with first-of-a-kind or pioneer plants, lest these overshadow the real 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 NREL and DOE for this report and reflect our best estimates at the time of publication. 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, namely lipid extraction and oil cleanup, were modeled with a fixed set of inputs determined by expected performance for the given flowrates and stream compositions confirmed with vendors. Bioreactors were modeled using experimentally-determined conversions of specific reactions (e.g., glucose to ethanol) rather than using rigorous kinetics or rate expressions. This simple stoichiometric model still satisfies mass and energy balances.

The Aspen Plus simulation uses component physical properties internal to the software as well as property data developed at NREL or from the literature [16,17]. Similar to other recent NREL models, the current model does not rely on external property databanks and minimizes the number of custom-defined components within reason. A discussion of the components and properties used is given in Appendix C.

2 Design Basis and Conventions

2.1 Plant Size: Feedstock Logistics and Cost

Similar to the recently-released algal HTL design report [11], the scope of this analysis begins with dewatered algal biomass feedstock delivered at 20 wt% solids, thus does not include upstream biomass production and dewatering, but instead assumes a rolled-up feedstock cost that accounts for all upstream activities. As noted above and discussed at length in recent harmonization publications [8, 10], regardless of conversion method, all integrated systems based on algal biomass processing are constrained by high cultivation variability, which is both location- and season-dependent [18, 19]. This variability must be considered in TEA and LCA analysis, as it carries non-trivial impacts on results given the requirements for systems engineering to accommodate this variability. Specifically, PNNL's resource assessment modeling under recent harmonization efforts identified suitable candidate locations that are envisioned to provide an optimum between high cultivation productivity and low net water consumption (after considering evaporation and precipitation at a given location), based on algal biomass cultivation in large open raceway ponds. For both generic and specific strains (namely a *Chlorella* strain titled NAABB 2412 or DOE 1412 [10, 20]), the RA model predicted swings in cultivation productivity up to 10:1 between summer and winter seasons, with the most recent models averaging out slightly more than 5:1 between these peak and minimum seasons as an average across all selected sites across the U.S. Gulf Coast. While it is possible, and in fact preferential, for a particular site and algal strain to fluctuate over a lower range of extremes across the course of a year, the same degree of variability near 5:1 was assumed in this analysis, to maintain consistency with the HTL design report [11].

The analysis assumes an annual average feed rate of 1,339 dry U.S. ton/day (1,215 metric tonne/day, ash-free dry weight (AFDW) basis) of algal biomass delivered to the conversion facility from upstream dewatering. This represents a roughly twofold increase over the 2013 HTL harmonization baseline of 650 ton/day (AFDW overall average) which corresponded to an annual average 14.6 g/m^2/day cultivation productivity rate [10] and thus implies an annual average productivity target near 30 g/m^2/day if all other upstream process and facility size assumptions were identical to the previously published work. Likewise, the feedstock cost is assumed at $430/ton (AFDW basis) as consistent with the HTL design report, which is based in turn on DOE's most recent MYPP report that targets a 2022 cost for algal biomass at $430/ton after cultivation (contributing $340/ton) and dewatering (contributing $87/ton) [9]. This cost target also is predicated on an algal cultivation productivity increasing to 30 g/m^2/day as well as a number of cost reductions for cultivation and dewatering relative to the "2010 baseline" as presented in the referenced report. Specifically, the MYPP document presents a 2010 baseline cost of $1,091/ton of algal biomass, based on the 2012 harmonization report assumptions for initial benchmark cultivation and dewatering performance. Relative to this benchmark, the 2022 target scenario stipulated that algal cultivation productivity increases from 13.2 to 30 g/m^2/day and removes the use of costly plastic pond liners, while reducing dewatering costs by 50% per amount of harvested material processed.

When the 1,339 ton/day annual average feed rate basis is combined with seasonal fluctuations as presented in Davis et al. [10], the resulting seasonal feed rates to the conversion facility are shown in Figure 3. Two cases are considered: one where the facility is sized to the peak summer flowrate and remains partially under-utilized for the remainder of the year, and a second where the amount of material produced in excess of the second-highest season (spring) is diverted, dried, and stored to be used in the winter.

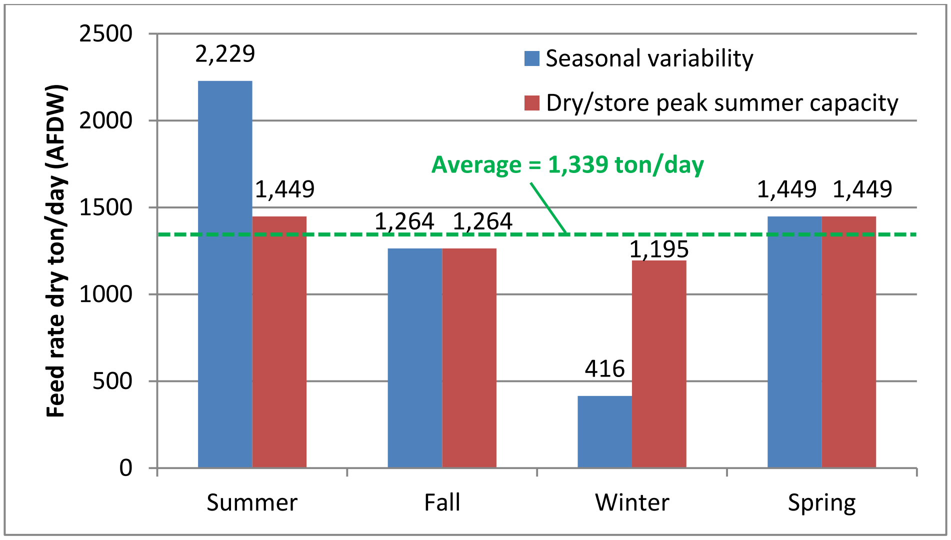


Figure 3. Seasonal and annual average feed rates to conversion facility. (Full seasonal variability without (blue bar) and with (red bar) excess peak capacity storage for use in winter.)

As shown in Figure 3, if the facility were designed to accommodate the raw peak summertime capacity, installed capital costs would be considerably under-utilized for the remainder of the year, particularly in the winter season which would also imply a required turndown capacity for the equipment of roughly 80% (5.4:1 variation) in instances where only a single unit operation is used rather than multiple units in parallel. In contrast, if excess summertime capacity were diverted, dried, and stored for later use in the winter, the capital utilization turndown factor would decrease to a significantly lower 18% (1.2:1) variation. Given more favorable operability as well as modeled MFSP costs (presented later), the latter approach to divert the excess material for use in the winter was selected as the base case for the analysis, which is consistent with the HTL design report approach [11]. Still, the practice of excess summer drying adds both capital costs for drying equipment as well as operating costs for natural gas (which also brings sustainability implications), and also adds risks of stored biomass spoilage that may require additional mitigation measures. Additionally, there is also a risk that the biomass could behave differently after being dried and reconstituted, but this is not yet known. Design and cost implications associated with this approach are discussed in Section 3.1.

2.2 Feedstock Composition

The algal biomass composition is based on recent NREL analytical characterizations [13], which were conducted for two separate fresh water algal strains over three harvesting time points as originally established under the SABC consortium [14]. Specifically, these were *Scenedesmus* sp. and *Chlorella* sp. provided by Arizona State University from harvests taken in early-, mid-, and late-cultivation stages, corresponding to low-, mid-, and high-nitrogen starvation states, and also to high-protein, high-carbohydrate, and high-lipid content biomass, respectively. By timing the harvest, biomass of different composition was obtained in a controlled fashion [21]. Cultivation time after reaching nutrient deplete conditions depended on final target biomass composition desired, which, depending on season, typically was 3 to 5 days for high carbohydrate (midpoint harvest) biomass and 6 to 9 days for high lipid (late harvest) biomass. High protein (early harvest) biomass was obtained by harvesting prior to nutrient depletion. Although this analysis focuses on fresh water media, the use of brackish or saline water strains may also be plausible for this technology pathway, although the efficacy for this option has not been experimentally investigated; at a minimum, biomass ash content would be considerably higher for saline species which could translate to increased lipid purification demands following the extraction step, and operation of anaerobic digestion could become challenging at high salt levels.

More favorable cultivation and processing performance as well as higher theoretical yield potential favored focusing on the *Scenedesmus* strain for TEA investigation; thus we consider this strain for both the mid- and late-harvest time points, but exclude the early-harvest case as it does not possess substantial levels of fuel precursor components. An abbreviated table for these specific cases is presented below (Table 1) showing both elemental and component compositions, excerpted from a more complete table for all strain/harvest combinations documented in analysis by Laurens et al. [13]. While only the mid- and late-harvest points are considered for modeling purposes, the measured early-harvest composition is also included to demonstrate the clear shift in biomass composition that occurs over the harvest regime originally studied. The measured shift in composition of the biomass across the early, mid and late stages of cultivation is considered typical for nutrient limited biomass with carbohydrate, protein and lipid content values consistent with earlier reports for *Chlorella* and *Scenedesmus* sp. [22-24]. The ash content reduced from 6.7% to < 3% between the early and late-harvest points. The protein content is inversely related to the lipid content and was found to drop rapidly from 34% in the early harvest biomass to 8.9% in the late harvest biomass, while lipid content increased from ~7% to 41%. The nitrogen content underpins the protein content with a consistent relationship of a nitrogen-to-protein conversion factor and previous reports have shown a protein content of <10% for nutrient limited biomass [22]. In Table 1 and hereafter, “HCSD” represents high-carbohydrate (mid-harvest) *Scenedesmus*, and “HLSD” represents high-lipid (late harvest) *Scenedesmus*. Likewise, “HPSD” represents high-protein (early-harvest) *Scenedesmus*, but is only included below in the original compositional analysis data and not considered further in this TEA evaluation.

Table 1. Algal Biomass Elemental and Component Composition for Early-Harvest (HPSD), Mid-Harvest (HCSD) and Late-Harvest (HLSD) *Scenedesmus*. (Left columns are raw data from experimental measurements, right columns are adjusted values to achieve 100% mass closure for modeling purposes. Excerpted from [13]. HPSD case was not considered for modeling purposes, but is included below to highlight the shift in biomass composition over time)

	Scenedesmus sp. (measured)			*Scenedesmus* sp. (input to model)	
Wt% composition (dry basis)	Early (HPSD)	Mid (HCSD)	Late (HLSD)	Mid (HCSD)	Late (HLSD)
C	49.3	52.9	56.3	54.0	57.4
H	7.1	8.0	8.5	8.2	8.7
N	8.3	1.8	1.6	1.8	1.6
O	28.7	34.8	31.2	35.5	31.8
S	0.0	0.2	0.2	0.2	0.2
P	0.97	0.22	0.20	0.22	0.20
Total [a]	94.4	97.9	98.0	100.0	100.0
Ash	6.7	2.3	2.1	2.4	2.1
Fermentable carbohydrates	24.3	46.3	37.9	47.8	38.2
-Starch	6.9	*12.2*	*8.1*	*47.8* [c]	*38.2* [c]
-Non-starch glucose	6.8	*22.6*	*18*	-	-
-Mannose	7.2	*11.5*	*11.8*	-	-
Other carbohydrates [b]	3.4	1.6	1.3	5.0	3.9
Protein	34.5	12.8	8.9	13.2	9.0
Lipids (Fuel-relevant lipids as FAME)	6.6	26.5	40.9	27.4	41.2
Non-fuel polar lipid impurities	ND [d]	ND	ND	2.7	2.1
Cell mass	ND	ND	ND	1.6	3.5
Total	75.5	89.5	91.1	100.0	100.0

[a] Measured element balance on dry weight basis; adjusted to AFDW basis for model

[b] Non-fermentable carbohydrates

[c] For modeling purposes, all fermentable carbohydrates are assumed to start in polymeric starch glucan form with subsequent hydrolysis to monomeric hexose sugar

[d] ND = not determined

As shown in Table 1, a number of adjustments were made to the baseline raw experimental composition values for use in the models. First, elemental compositions were adjusted to 100% mass closure on an ash-free dry weight basis. For component compositions, initial measured mass balance closures of the HCSD and HLSD cases were near 90% (including ash). Rather than adjusting all components evenly to close the balance to 100%, specific components were adjusted first based on the level of certainty in analytical measurements. Specifically, the routine analytical method used to determine lipid content based on total fatty acids did not account for the non-fatty acid polar portion of the lipids or for hydrocarbon-based compounds that are present in the biomass. The non-fuel polar fraction of lipids can contribute a sizeable amount to total lipid mass and consist of for example, the polar headgroups of lipids (e.g. phosphatidic acid of phospholipids, digalactosyl moieties on glycolipids, etc.). Literature indicates that the non-fuel polar lipid fraction can particularly make up a considerable portion of the lipids in early-harvest nutrient replete conditions, with a drastic reduction in polar lipids as the lipid content increases under nutrient limited cultivation scenarios (e.g. moving towards mid-harvest HCSD and further to late-harvest HLSD cases) [25, 26]. This guidance was applied in adding a non-fuel “polar lipid impurity” fraction to the biomass composition as shown in Table 1, assumed to constitute 10% of the remaining fatty acid lipid fraction in the HCSD case and dropping to 5% of this fraction in the HLSD case. Intact lipid composition for these biomass samples is required to definitively measure the polar lipid fractions. Of the "polar lipid impurities" fraction, a portion is assumed to be extracted with the rest of the lipids and requires subsequent removal in a lipid cleanup step, discussed below in Section 3.4.

Other, hydrocarbon-based lipid species such as straight and branched-chain hydrocarbons, sterols and isoprenoid-based compounds were also found to be present in the biomass but do not convert to fatty acid methyl esters (FAME) when transesterified according to NREL published methods [27]); thus, the lipid fraction shown in Table 1 is based on *fatty acid lipids as FAME*. The additional non-fatty acid hydrocarbon-based compounds would not be counted in such a measurement but are still likely to be extracted in a hexane-based lipid extraction protocol, and ultimately upgraded to additional fuels through hydrotreating. It is possible that these compounds add a small fraction to total "fuel potential lipids", but for the purpose of this analysis this fraction will conservatively not be considered as it was not characterized in the compositional data (Table 1) utilized as inputs to the model. It does, however, provide a small margin of "safety" in the targeted lipid extraction efficiency yield discussed in Section 3, as a given extraction yield of FAME lipids would be equivalent to a slightly lower extraction yield of *total* (FAME + non-FAME but still convertible) lipids.

Following the "polar lipid impurity" fraction, the next highest level of uncertainty is around non-fermentable carbohydrates, which includes species such as uronic acids, amino sugars, deoxy sugars, etc.; thus, this fraction was conservatively tripled relative to the baseline analytical value for the modeling basis. Finally, a small 1.5-3.5% fraction (HCSD and HLSD respectively) was added to account for other non-characterized cell mass such as nucleic acids based on prior internal analysis. These three additions increased component mass closures to 97%-99%. To reach 100% mass closure, all components were then subsequently adjusted evenly.

While a large fraction of fermentable carbohydrates were characterized as non-starch glucose and mannoses carbohydrates, the specific nature of these non-starch carbohydrates is not yet known; thus for modeling purposes the fermentable carbohydrate fraction was assumed to initially start off in polymeric glucan form with all fermentable carbohydrates modeled as starch. This allows the model to calculate heat of reaction for a specified glucan carbohydrate and subsequently estimate the steam demand and overall pretreatment reactor heat balance required to yield the stipulated monomeric sugar release.

Although we conduct TEA modeling for both the HCSD and HLSD cases (with results presented in Section 5.2), the HLSD biomass serves as the official base-case for subsequent discussion and focus for process and economic results. This case was selected due to its higher net fuel potential (driven by substantially higher lipid content) and resulting superior economics, and in keeping with the focus of this technology pathway being "algal *lipid* extraction and upgrading" (ALU). However, it is recognized that this case requires longer cultivation time to reach nutrient deprivation and associated high lipid levels, which must be considered in understanding ultimately achievable biomass growth rates and associated feedstock costs from the cultivation step. This is outside the scope of this study, but will remain a focus area for analysis moving forward. In reality, a more practical out-year target case may lie somewhere in between the harvest points (and implicit productivity rates achievable in the future) attributed to the HCSD and HLSD cases. All modeling scenarios assume a delivered solids content of 20 wt% after upstream dewatering.

2.3 Modeling Basis Parameters

Similar to other biofuel pathways recently published in design reports under DOE-BETO efforts [6, 11], a number of key modeling assumptions are fixed constant to allow for comparable modeling outputs. These parameters include:

1. Design and construction time: 36 months
2. Facility start-up time: 0.5 year (6 months)
3. Cost-year dollar basis: 2011 dollars (all cost results presented here will be in 2011-year dollars)
4. Facility on-stream time: 90% (330 days/year or 7,920 hours/year). Note, this is intended to represent an n^{th}-plant facility, and may be a higher factor than what is achievable today, particularly when considering the algal cultivation step. This on-stream factor assumes year-round operation with approximately one month per year allocated to facility shut-down, whether due to maintenance, upstream pond upsets (such as culture crashes, freezes, etc.), or other downtime factors.

2.4 Design Report Conventions

2.4.1 Units

The Aspen Plus model we developed is based on the same set of units as used in NREL's 2013 biochemical hydrocarbon design report model [6], as it was constructed by modifying this model which also follows a similar process using dilute acid pretreatment and fermentation operations; these units include kg, kmol, atm, °C for materials, and MM kcal (Gcal) for energy. Values in this report that were retrieved 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, certain equipment specifications are 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. "Ton" also appears in Section 3.7 in the context of refrigeration, but this usage should be clear from the discussion.

2.4.2 Product Density 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 is done to maintain focus on *total fuel (energy)* yield from the relevant precursor components in the starting biomass, whether ultimately converted to diesel (RDB) or ethanol, and also because ethanol is intended to serve as a "representative" end-product of carbohydrate utilization, among other options, including biological or catalytic conversion of sugars to hydrocarbon fuels and/or products. Additionally, and more important from a TEA standpoint, the volumetric split between RDB and ethanol products is evenly divided, varying from 50% (HCSD) to 65% (HLSD) RDB versus ethanol. In such cases where two products are produced in similar volumes, it is most appropriate to consider the summative energy-equivalent (GGE) yield for both products, to avoid introducing dramatic sensitivity in the MFSP result if one of the products were instead assigned a coproduct value (this approach is consistent with a recently published woody biomass pyrolysis design report which also produces similar volumes of different fuel products [28]). Thus, the main two products, RDB and ethanol, are adjusted by heating values to a single GGE fuel yield, while a much smaller coproduct, naphtha, is appropriately assigned a coproduct value and generates a small amount of coproduct revenue.

To adjust to an energy-equivalent fuel yield, component heating values were set at 23.6 MJ/L (84,530 BTU/gal, HHV basis) for ethanol based on standard published values [29], and were calculated (HHV basis) from the Aspen model for the RDB product based on the component mixture. A higher heating value was selected in this case as it allowed for a more readily achievable comparison to published values for the RDB product, in order to validate a close resemblance between Aspen calculations and accepted energy content for similar products. Indeed, the calculated HHV value for the RDB product was 130,430 BTU/gal, which is roughly 0.3% lower than a published value for hydrotreated renewable diesel at 130,817 BTU/gal [29]. To subsequently translate to a GGE basis, a conventional gasoline heating value of 124,340 BTU/gal (46.54 MJ/kg, HHV basis) was applied [DOE 2012]; this corresponds to a lower heating value of 116,090 BTU/gal gasoline. Similarly, where densities are required to calculate product volumes, a standard published ethanol density of 0.789 kg/L at 20 °C (68 °F) was applied [3], while the Aspen model calculated the density for the RDB and naphtha products at 0.768 and 0.730 kg/L respectively (also at 20 °C).

3 Process Design and Cost Estimation Details

The process design described in this study is based upon experimental demonstration results observed first under the SABC consortium [14], then more recently in follow-up NREL work as described by Laurens et al. [13], projected out to plausible future improvements in key process areas. Some areas of the process draw from previous NREL experimental and analysis efforts, namely dilute acid pretreatment and fermentation operations, but aside from expertise gained in optimizing such operations and associate process integration related to their use, this process is unique given the presence of lipids and absence of lignocellulosic components such as lignin and pentose carbohydrates; this makes for a more ideal process in some respects but adds new challenges in others, as documented in the following sections. It is important to reiterate that the targets presented here are merely one set of conditions that would enable achieving the 2022 cost goal below $5/GGE and will help to inform near-term research directions, but do not necessarily represent the most optimal possible scenario. Additionally, given the scope and timeframe under which the present analysis effort was conducted, the report is more concise in some areas (such as detailed process flow diagrams and stream-level information) relative to earlier NREL design reports. This section describes the process as modeled and discusses the influence of specific R&D goals in the decision-making process.

With respect to costing estimates, given the more preliminary scope and nature of the design models established here, the engineering support provided by Harris Group focused primarily on the major unit operations (e.g. extraction and distillation columns, oil cleanup and upgrading "facility" costs, etc.), and did not obtain vendor pricing for minor supporting equipment such as pumps, heat exchangers, and agitators. For this minor equipment, Harris Group added a percentage factor to the provided cost estimate for the respective major operations, which varied according to process area as discussed below, or NREL estimated costs using Aspen Economic Evaluator [15] for a number of heat exchangers. Additionally, at the request of Harris Group who provided all relevant vendor estimates for cost quotations, vendor company names associated with a particular unit operation will not be provided in this report.

3.1 Area 100: Pretreatment and Conditioning

3.1.1 Overview

The process begins with algal biomass delivered to the conversion facility at 20 wt% solids after processing through upstream dewatering. As discussed above in Section 2.1, the raw feed rate to conversion varies seasonally over a range of roughly 5.4:1 between summer and winter, with fall and spring flowrates both roughly 40% lower than peak summer flows, reflecting the same degree of variance upstream in the cultivation step. Two options are presented above for handling this variance: (1) to feed the material straight to the facility with no pre-processing, thus requiring most capital expenses to be substantially over-designed for the remainder of the year (particularly in the winter), or (2) to divert, dry, and store the fraction of peak summertime capacity that is produced in excess of the spring season (the next highest production season), and to use it during the winter months to reduce the dramatic swings in throughputs and equipment over-design. Of these two options, the latter scenario to divert a portion of summer production for use in the winter is selected as the base case for the primary focus of this analysis. This is done to maintain consistency with the above-referenced algal HTL design report [11] as it allows for more realistic commercial operability without the need for substantial equipment turn-down ratio capabilities, and also because it ultimately achieves a lower fuel production cost, as will be presented later. It is recognized that this choice carries a sustainability penalty, namely the added requirement for natural gas drying of the diverted summertime material, but the overall greenhouse gas impact is diluted given that this step is only required for three months out of the year and not utilized for 75% of the year. Moving forward, NREL will work with partners at ANL to more fully characterize the complete LCA of the integrated system, and if the level of natural gas for drying in the summer proves to be unacceptable, the alternative option may instead be selected (we show in Section 5.2 that economics are still reasonable for the case without summer diversion). Additionally, alternative summer drying techniques may be possible such as solar drying to reduce the sustainability burden for natural gas drying, but this is not considered here. Storing dried biomass for 3-6 months also brings a risk for biomass spoilage via oxidation or microbial degradation, and may require additional mitigation measures to prevent which are not considered here.

The amount of delivered feedstock material diverted to be dried in the summer equates to 35% of total summertime feed rate (2,229 ton/day AFDW basis), thus 780 ton/day of dry biomass (AFDW basis) is diverted away to drying and 1,449 ton/day is sent on to the pretreatment fractionation step. Consistent assumptions for the drying operation were employed as documented in the algal HTL design report [11], namely a purchase cost of $905,000 per dryer (2013-dollars), each accommodating a flow up to 2,520 kg/hr of moisture-free biomass for drying the material from 80% to 10% moisture. The dryer is natural-gas-fired and operates at 77% thermal efficiency (i.e., heat required for drying versus heat content of the natural gas). These values are based on a quotation for a rotary drum dryer. Also consistent with the HTL report, the cost for storing the dried material until the winter is assumed to be accounted for in the indirect capital cost factors presented in Table 15. Subsequently, during the fall, winter, and spring seasons, all delivered material is sent straight to the fractionation step at a flow rate of 1,264, 416, and 1,449 ton/day (AFDW basis) respectively, with the 780 ton/day from storage added in the winter to constitute a total winter flow to fractionation of 1,196 ton/day.

The biomass pretreatment (fractionation) step is a crucial operation to enable successful recovery and conversion of both the carbohydrate and lipid components. In the context of lignocellulosic (terrestrial) biomass processing, dilute acid pretreatment breaks down primarily hemicellulose compounds to their monomeric sugar constituents (namely xylose, mannose, arabinose, and galactose) by hydrolysis reactions, while making the remaining biomass more amenable to further deconstruction downstream, such as enzymatic hydrolysis of cellulose to glucose [3, 6]. While such recalcitrant carbohydrates are largely not present here, a similar dilute acid operation has been shown in literature to enable sugar recovery from algal biomass [30, 31] and is required to disrupt the algal cell wall, achieve high soluble sugar recovery, and enable effective downstream solvent extraction. Polymeric glucan carbohydrates (primarily starch) have also been observed to hydrolyze to monomeric glucose through the dilute acid pretreatment step, and NREL experimental efforts have found that no amylase or cellulase enzymes are required (e.g., for targeting either starch or cellulose carbohydrates) following the pretreatment operation [32], as the majority of the carbohydrates are seen to be less recalcitrant than cellulose and hemicellulose components typically present in cellulosic feedstocks after processing through acid pretreatment. This eliminates the need for costly enzymatic saccharification operations that have been pursued in literature as an alternative means to algal sugar/ethanol production [33-35].

Pretreatment fractionation occurs at a temperature of 130°-180°C and corresponding bubble-point pressures, with a residence time ranging from 1 to 10 minutes. Following pretreatment, the hydrolysate slurry is flash-cooled, vaporizing a fraction of water which is condensed and routed to the water recycle pool. The hydrolysate is then sent to a conditioning reactor, where ammonium hydroxide is used to raise its pH to 5.

Figure 4 shows a simplified flow diagram of the pretreatment area.

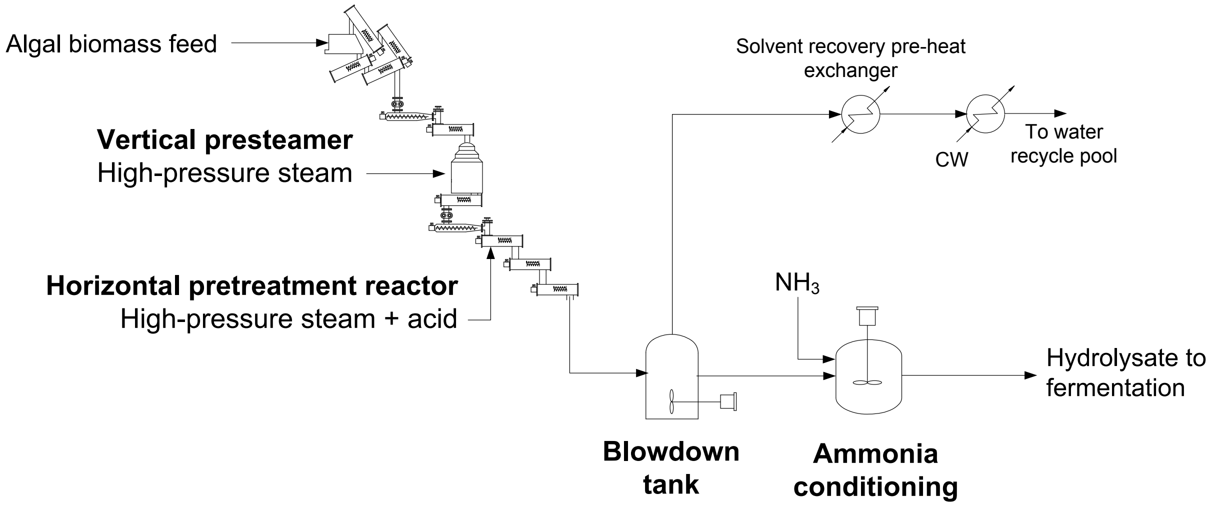


Figure 4. Simplified flow diagram of the pretreatment and conditioning process

3.1.2 Design Basis

In principle, the acid pretreatment reactor system is the same as that described in earlier NREL reports for biochemical conversion of corn stover [3, 6] and will not be repeated in such detail here. In summary, this 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 that 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, shown in Figure 5. 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. The current model assumes operation of the presteamer at 100°C such that no significant reactions occur in this section. It can be used if additional holdup 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.

While historically NREL has operated the pretreatment step at 30 wt% total solids for processing corn stover, in this case, the solids content both entering and exiting the reaction step (after flash) are 20 wt%. However, pilot-scale work has already shown effective operation of the unit near 20% solids when processing algal biomass using NREL's 120 L horizontal batch reactor, with trials conducted to date on up to 25 kg of dry weight biomass at 20% solids loading [36]. Thus, from an operational standpoint, the lower solids loading is not anticipated to be problematic. Acid loading in the reactor was set at a target 1 wt% relative to feed slurry liquor (water), based on assumed improvements relative to early experimental efforts using 2 wt%. This appears to be a reasonable target reduction, as it translates to an acid loading of 40 mg/g dry biomass, compared to corn stover processing at lower values closer to 9 mg/g [6]. The effective sulfuric acid concentration in the pretreatment reactor (estimated at roughly 0.8 wt% after dilution by condensing steam is accounted for) likely requires maintaining a high-cost Incoloy 825 cladding metallurgy, rather than switching to lower-cost stainless steel alloys, consistent with prior cost estimates for this operation. The potential impact of lower-cost metallurgy and other pretreatment reactor cost savings is addressed in Section 5.1. 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; however, given the seasonally variable feed rates discussed previously for this pathway, residence time may also fluctuate by a similar 20% (1 minute) between summer and winter seasons, although the screw speed may also be adjusted to maintain targeted residence time.

A comparison of current (experimental baseline) operational parameters for the pretreatment reactor compared to the target case is summarized in Table 2.

Table 2. Pretreatment Conditions Applied in Experimental Work to Date [13], Compared to this Design

Parameter	Experimental baseline	Present design target
Sulfuric acid loading (wt% of feed liquor rate)	2%	1%
Residence time	1-5 minutes	5 minutes
Temperature	145°C	150°C
Pressure	4.4 atm	4.6 atm
Total solids loading	20-25 wt%	20 wt%

Table 3 summarizes the reactions and conversions that take place in pretreatment. As noted previously, although starch carbohydrates have been measured to constitute 20%-25% of total fermentable carbohydrates in both HCSD and HLSD algal biomass materials (with the remainder being non-starch glucose and mannose carbohydrates), for modeling purposes, all fermentable carbohydrates were assumed to initially start out as starch in the feed biomass. This allows for calculating heat of reaction and subsequently setting high-pressure steam requirements and pretreatment heat balances based on a defined glucan carbohydrate (with the remaining carbohydrates not yet characterized as to their compositional nature or recalcitrance), as well as allows for easier tracking of overall sugar yields in the model. Thus, for this discussion, starch "glucan" represents the total fermentable carbohydrate portion for the HLSD biomass base case, constituting 38.2 wt% of the biomass as shown in Table 1. Likewise, with glucose and mannose constituting the large majority of monomeric sugars in the biomass (both hexose sugars), all monomeric sugars were modeled here as glucose for the sake of modeling simplicity. Being the same composition and molecular weight, glucose and mannose behave the same in the model and have both been experimentally demonstrated to be equally fermentable to ethanol when using *S. cerevisiae* [13, 37]. With these points in mind, the model targets a "glucan"-to-"glucose" yield of 90% (i.e., 90% recovery of all fermentable carbohydrates as soluble hexose sugar). Additionally, hydroxy-methyl furfural (HMF), a common fermentation inhibitor, is assumed to be produced at the same level as in prior design reports for corn stover [6]. Low levels (0.1-0.2 g/L) of HMF have been measured in hydrolysates from algal biomass pretreatment, but not at levels that became inhibitory to subsequent ethanol fermentation, and no furfural was detected [13]. Other inhibitors may also be produced from pretreatment, and additional characterization is required to further quantify these details moving forward.

Table 3. Pretreatment Hydrolysis Reactions and Assumed Conversions

Reaction	Reactant	% Converted to Product
$(\text{Glucan})_n + n\ H_2O \rightarrow n\ \text{Glucose}$ [a]	Glucan	90%
$(\text{Glucan})_n \rightarrow n\ \text{HMF} + 2n\ H_2O$	Glucan	0.3%

[a] "Glucan" is assumed in the model to represent all initial fermentable carbohydrates and "glucose" represents all hexose sugars, e.g., glucose and mannose

Relative to the 90% target for solubilized glucose yield (recovery from initial carbohydrates), current experimental work has demonstrated up to 77% recovery of sugar from algal carbohydrates to date when following the same reactor configuration modeled here (pretreatment prior to extraction) [13]. Thus, the target of 90% represents a reasonable improvement anticipated to be achieved through continued research and reactor optimization moving forward, e.g. optimizing acid loading, temperature, and time, or considering alternative mechanical or alkaline pretreatment steps. The pretreatment reactor is discharged to a flash tank. The pressure of the flash is controlled to keep the hydrolysate at 20 wt% total solids content. The flash is condensed by cross-exchange with the solvent recovery distillation preheater downstream, then cooled and routed to the water recycle pool, without the need for cleanup, as it contains only a minor fraction (~4 ppm) of HMF. After the flash, the hydrolysate whole slurry containing 20% total solids and 13% insoluble solids 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 from 117°C to 100°C. The composition of the stream at this point, and other major points throughout the process, is shown in the stream table information in Appendix D. The material from conditioning is conveyed to a hydrolysate storage tank.

3.1.3 Cost Estimation

The cost for the rotary drum dryer discussed above, was set based on the HTL design report at a purchase cost of $905,000 per dryer (2013-dollars), each accommodating a flow up to 2,520 kg/hr of moisture-free biomass [11]. As the facility is already at a slight heat deficit (requiring marginal natural gas to meet all facility utility heat demands), no excess heat is available to operate the dryer and summer drying requires additional natural gas during the three months that the unit operates, delivered to the dryer based on 77% thermal efficiency.

The acid pretreatment reactor design and cost basis was updated in NREL's 2013 biochemical hydrocarbon 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 design report [6]. 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:

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

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

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, 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, the recommended metallurgy from the vendor continues to be Incoloy-825, with a total installed capital cost of approximately $43.9MM (2011-dollars) for the system, including feedstock receiving bin, pre-steaming, pressurized heating, reaction, and flash cooling. This cost, and all subsequent downstream capital costs discussed in this report for the base case, are based on the "summer storage" scenario peak-season feed rate to pretreatment of 1,449 AFDW ton/day in the summer and spring seasons, after diverting excess summer capacity to the dryer (Sections 2.1 and 3.1.1). The reactor schematic is shown in Figure 5, and is similar to the configuration presented in earlier reports [3, 6].

In addition to the cost for the pretreatment reactor system itself, a "balance of plant" factor was applied to account for other minor equipment supporting the pretreatment fractionation system, including the neutralization tank, flash tank, and other minor tanks, pumps and agitators. This balance of plant factor was calculated as 4.6% of the pretreatment system cost based on costing details presented in the 2013 design report [6]. Finally, two heat exchangers which had not been employed in earlier analyses were added in this case; one, which cross-exchanges heat between the pretreatment reactor feed and effluent streams (after steam addition and pretreatment/neutralization reactions), and a second which condenses the pretreatment flash steam by exchanging with the downstream solvent distillation pre-heater. Both shell-and-tube exchangers were costed in Aspen Economic Evaluator [15], as typical for more minor operations such as these.

Area 100 contributes about $0.50/GGE to the MFSP, including dryers, acid pretreatment, and conditioning. About 56% of this contribution is attributed to capital cost, of which acid pretreatment equipment accounts for the large majority (70%) of total capital expenses.

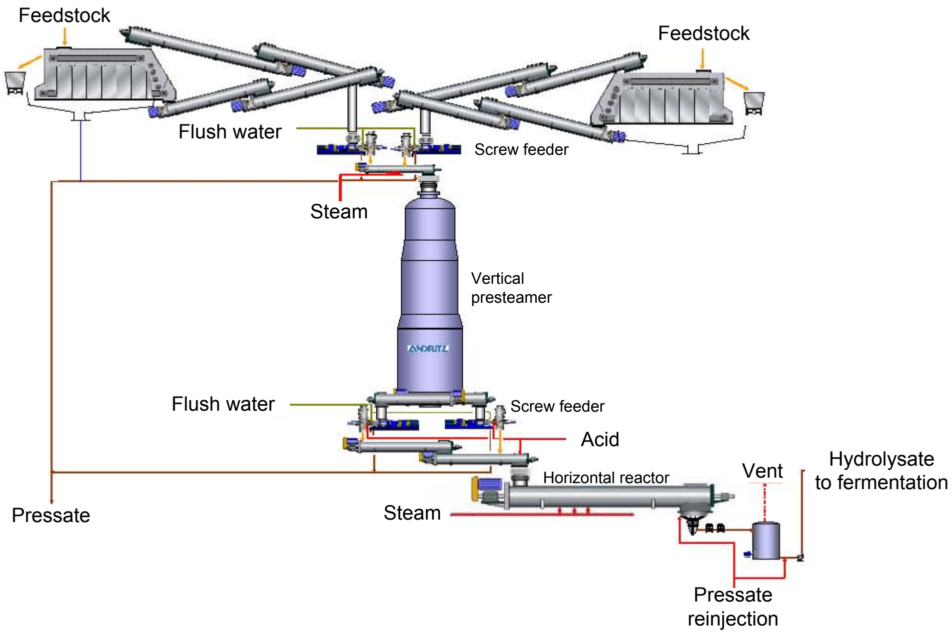


Figure 5. Horizontal acid pretreatment reactor design scheme

3.2 Area 200: Fermentation and Distillation

3.2.1 Overview

In this process area, released soluble sugars are fermented to ethanol using *Saccharomyces cerevisiae* D5A (yeast) which readily utilizes both glucose and mannose as the two primary sugars released from pretreatment. Fermentation proceeds in batch bioreactors at a total batch time of 1.5 days, after which point the fermentation broth is sent to ethanol purification consisting of beer and rectification distillation columns and vapor-phase molecular sieve adsorption, which ultimately concentrates the ethanol product up to 99.5%. The vapor overhead from the beer column and fermentation vents, containing CO_2, ethanol, and other minor components, is sent to a scrubbing column to recover the majority of the volatilized ethanol. The beer column stillage, containing water and all remaining algal components, is routed to lipid extraction. Figure 6 shows a simplified flow diagram of the fermentation and ethanol purification process.

Given the lack of lignin, xylan, and other "impurity" components present in the algal biomass which are typical for lignocellulosic feedstocks, the produced sugars entering the fermentation step are relatively clean and readily convertible to a wide range of energy or chemical products, and represent roughly 38 wt% of the starting biomass following 90% solubilization of fermentable glucan carbohydrates from upstream pretreatment. While DOE-BETO has begun to transition from a primary R&D emphasis on ethanol to hydrocarbon fuel and blend stock products in light of having sufficiently demonstrated R&D progress through 2012 on *cellulosic* ethanol technology pathways [38-40], ethanol is reasonably maintained in this particular pathway as a fuel coproduct alongside renewable diesel and is intended to serve as a *representative energy product* from monomeric sugars (with fuel yield translated to a GGE energy basis for both products). A number of other options exist to instead convert sugars to hydrocarbon fuels or chemicals, via either biological or catalytic routes [41, 42]. Such options could include commodity fuel precursor molecules such as fatty acids, fatty alcohols, isoprenoids, or paraffins with cost and yield implications as detailed in [6], or high-value coproducts including organic acids, diols, and other compounds for which even a small amount of production may garner significant revenue but must be balanced against market volume limitations (discussed further in Section 5.4). Moving forward, such options will also be considered, but in this analysis, the decision was made to focus on ethanol as the product from sugars given favorable in-house data already established for this step within the integrated algal fractionation process [13]. Nonetheless, this highlights another key advantage of this technology pathway in possessing process flexibility around utilization of carbohydrates, as well as lipids and protein fractions of algal biomass.

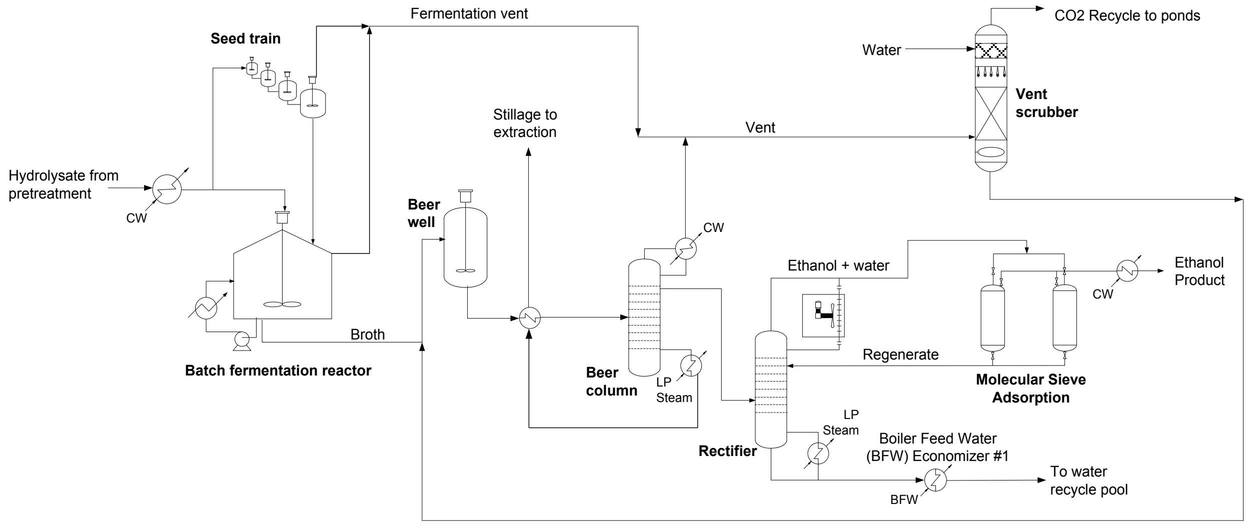


Figure 6. Simplified flow diagram of the ethanol fermentation and purification process

3.2.2 Design Basis

Ethanol fermentation is a well-characterized process that largely follows process and design details as documented in NREL’s 2011 cellulosic ethanol design report [3], with a number of minor differences. After first being cooled by cross-exchange with the feed stream entering pretreatment, the neutralized hydrolysate is further cooled by cooling water to the required fermentation temperature (37°C for *S. cerevisiae*). Ten percent of the hydrolysate stream is split off to inoculum seed growth to initiate fermentation, utilizing consistent design assumptions for the seed train system as described in [3] by analogy from previously modeled *Z. mobilis* propagation and ethanol fermentation. In summary, two parallel inoculum seed trains, each train consisting of five reactors in series, operate in batch mode with a 24-hour batch time and an additional 12-hour turnaround time. The seed reactors are cooled with chilled water from Area 700 to maintain the temperature at 37°C. The first vessel is inoculated with a seed culture from the lab. Its broth is used to inoculate a larger reactor, and so on. After five iterations, the cell mass from the last vessel is sufficient to inoculate the production vessel. However, at the design feed rate utilized here (1,449 ton/day in the summer case), a considerably lower amount of ethanol is produced than in the 2011 design case (3,263 gal/hr in the summer design case versus 7,257 gal/hr in the 2011 ethanol design case). Thus, the individual vessel sizes used for each step of the seed train were assumed to scale accordingly. At 45% of the basis ethanol production rate, the seed train utilized in the current design was assumed to initiate in a 17-gal vessel (rather than 40 gal) as the first step, eventually scaling up to 86,000 gal (rather than 200,000 gal) in the final step. Table 4 summarizes the seed train design specifications.

Table 4. Seed Train Specifications

Inoculum level	10 vol % of production vessel size
Batch time	24 h
Vessel turnaround time	12 h
Number of trains	2
Number of vessel stages	5
Maximum vessel volume	86,000 gal (326 m^3)
Diammonium phosphate (DAP) loading	0.67 g/L broth (whole slurry)

Table 5 gives the reactions and conversions used to describe the microorganism growth and sugar metabolism in the seed vessels. In contrast to *Zymomonas mobilis*, a bacterial organism frequently utilized for fermentative conversion of cellulosic sugars to fuels [43-45], *S. cerevisiae* typically consumes a larger fractional amount of sugar diverted toward biomass cell growth [46]. Thus, while the 2011 ethanol design report assumed 4% conversion of glucose to *Z. mobilis* in the seed train, translating to 0.4% diversion of glucose to cell mass in this step (given the split of 10% of total sugars sent to the seed train), the present design increases this fraction to 40%, i.e., 4% diversion of total sugar to *S. cerevisiae* biomass. This is likely a conservative assumption for such a fractionally higher amount of sugar diversion to cell growth when still fundamentally based on anaerobic fermentation. NREL has also investigated fermentation of algal sugars using *Z. mobilis*, and this may also be a viable candidate, although the use of *S. cerevisiae* is preferred as this organism also readily converts the mannose sugars hydrolyzed from the algal biomass while *Z. mobilis* does not. Lipids are assumed to be preserved across the fermentation stage, and available for subsequent extraction; in fact, additional lipids may be produced during yeast (*S. cerevisiae*) propagation and increase downstream extraction yields, but this has not been experimentally characterized and is not considered here. Unlike previous models based on lignocellulosic hydrolysate fermentation, the present analysis assumes sufficient nitrogen for organism propagation is present in the algal biomass hydrolysate (primarily protein) constituents, and does not require an external supply of nitrogen from corn steep liquor (CSL) or other sources; this is consistent with early experimental observations to date. Although the same also likely holds for phosphorus, the model conservatively assumes DAP is still required for organism growth, delivered to the seed train at a concentration of 0.67 g/L and to the fermentation stage at 0.33 g/L. This small DAP rate is ultimately reclaimed as additional DAP from the anaerobic digester and a fraction is available for recycle to the ponds.

Table 5. Seed Train Reactions and Assumed Conversions

Reaction	Reactant	% Converted to Product
Glucose → 2 Ethanol + 2 CO_2	Glucose	55%
Glucose + 0.3704 Protein + 0.018 DAP → 6 *S. cerevisiae* + 2.4 H_2O [a]	Glucose	40%

[a] For element balance closure purposes, *S. cerevisiae* was modeled as *Z. mobilis*. Sufficient nitrogen for cell propagation is assumed to be present in the algal biomass (protein) constituents, but the model assumes supplementary phosphorus is still required for biomass growth, delivered as DAP.

Sugar may be lost to side products by contaminating microorganisms in addition to being converted to cell mass and ethanol product. Sugar contamination losses were set consistent with the 2011 report basis, namely modeling a diversion of 3% sugar (glucose) losses to lactic acid upstream of the fermentors.

The main fermentation step assumes the same vessel design as the 2011 ethanol design report, namely parallel stirred-tank anaerobic batch reactors sized at 1-MM gal capacity each (950,000 gal working volume). The fermentation batch time is modeled as 1.5 days (36 hours). The fermentation temperature of 37°C is maintained using chilled water from the coolant loop (Area 700). Agitation is also set consistently with the previous ethanol design basis at a basis of 0.03 hp/1,000 gal (6 W/m^3), a fairly minimal power demand attributed to merely maintaining a reasonable suspension for anaerobic fermentation. Four 1-MM gal fermentation vessels are required in the peak (summer/spring) design case, and operate at fractionally lower working volume below the 95% design basis for lower seasonal throughputs. Table 6 provides a summary of fermentor design parameters utilized in this design.

Table 6. Guiding Bioreactor Design Basis Assumptions

Temperature	37°C (99°F)
Initial fermentation solids level	20% total solids (7% soluble, 13% insoluble)
Reactor size	1 MM gal (3.8 MM L) operating at 95% working volume under peak (summer/spring) seasonal throughputs
Residence time	1.5 days (36 h)
Inoculum level	10 vol %
Diammonium phosphate (DAP) level	0.33 g/L broth (whole slurry)

The model targets a fermentation ethanol yield at 95% of theoretical, i.e., a metabolic yield of 0.48 g/g glucose. This represents an improvement over 81%-83% yield experimentally demonstrated to date for algal hydrolysate sugars utilizing upstream dilute acid pretreatment [13], which may be achieved through continued process integration research (i.e. better tailoring upstream operations towards downstream conversion needs, tracking and mitigating fermentation inhibitors), optimizing fermentation conditions including further quantification of optimal nutrient C:N:P ratios for maximizing ethanol production, or evaluating alternative organisms. As noted previously, all monomeric fermentable sugars are modeled in this analysis as glucose, representing both glucose and mannose as measured experimentally. Given that these two sugars have the same molecular weight and behave similarly during fermentation using *S. cerevisiae*, with no discernable lag between consumption of one sugar and another (data not published), this simplifying assumption is appropriate. While a marginal amount of HMF is assumed to be produced as a byproduct from the upstream pretreatment step, no discernable inhibition from HMF during fermentation has been observed to date based on measured HMF levels of 0.1-0.2 g/L and no furfural (another fermentation inhibitor) has been measured in the hydrolysate material [13]. It is recognized that such inhibition, as well as overall fermentation performance (yield and rate) in general, is a function of feed and product concentrations entering and exiting fermentation, and the sugar concentration in the hydrolysate is fairly dilute in the base case model at 7.4% (69 g/L) sugars, translating to an ethanol concentration in the fermentor broth of 30 g/L. If the hydrolysate sugar concentration were to be increased, for example, by adjusting the pretreatment flash specification and/or adding a concentration step, levels of inhibitory components could also become more problematic. However, if such concentrations could be achieved with no negative impact on fermentation performance, this would likely lead to a more optimal system, primarily by reducing the energy-intensive distillation requirement downstream. Table 7 summarizes the reactions and conversions assumed in fermentation.

Table 7. Fermentation Reactions and Assumed Conversions

Reaction	Reactant	% Converted to Product
Glucose → 2 Ethanol + 2 CO_2	Glucose	95.0%
Glucose + 0.3704 Protein + 0.018 DAP → 6 *S. cerevisiae* + 2.4 H_2O [a]	Glucose	2.0%

[a] For element balance closure purposes, *S. cerevisiae* was modeled as *Z. mobilis*. Sufficient nitrogen for cell propagation is assumed to be present in the algal biomass (protein) constituents, but the model assumes supplementary phosphorus is still required for biomass growth, delivered as DAP.

The fermentation broth or "beer" has an ethanol concentration of 3.4 wt% (30 g/L) and is collected in the beer well. The beer well is designed for a residence time of 4 hours to provide surge capacity between fermentation and distillation. The material is then sent to the beer column, a distillation column with 32 stages operating at 48% efficiency that removes CO_2 and a small amount of ethanol into the vapor overhead, recovers roughly 98% of ethanol into the vapor side-draw (sent to the rectification column), and recovers roughly 90% of the water plus all residual algal components and other solids into the bottoms (stillage) stream. The feed to the beer column is preheated in an economizer heat exchanger by cross-exchange with the bottoms stream fed to the lipid extraction section. The ethanol side-draw from the beer column is routed to the rectification column, a second distillation column with 45 stages operating at 76% efficiency. The rectification column further purifies the ethanol to near-azeotropic concentrations (92.5 wt%) in the overhead, leaving 99.9% purity water in the bottoms stream. The bottoms stream is cooled to 30°C by cross-exchange with the high-pressure pretreatment steam economizer, then is combined into the water recycle pool. Further details on both distillation columns may be found in NREL's 2011 ethanol design report. Both columns are reboiled using low-pressure (2.7 atm) steam generated by recovering heat from the AD turbine flue gas. Also consistent with the original design specifications, the beer column is condensed to 60°C using cooling water, while the rectification column is condensed to 91°C using air cooling.

The CO_2 vent streams from both the fermentation and seed train operations contain a small amount of volatilized ethanol, and are routed along with the distillation beer column overhead to a water scrubber. The scrubber is specified to recover 99% of the volatilized ethanol in the vapor feed stream by varying the scrubbing water flow rate. The scrubber overhead product contains a high concentration (over 97 wt%) of CO_2, and makes an ideal candidate for recycling back to upstream cultivation ponds as it contains very low levels of inert components which otherwise represent a parasitic power draw on transporting the gas back to the ponds. Thus, this stream is destined for recycle to the cultivation ponds, and the CO_2 present in the vent stream is added to the CO_2 in the gas turbine flue gas for recycle coproduct credits (discussed further below). The liquid product (water plus recovered ethanol) is routed to the beer well along with the fermentor product broth.

The overhead vapor from the rectification column is fed to a molecular sieve adsorption unit, which is a system of columns packed with beds of adsorbent. Water is selectively adsorbed to the beds as the vapor flows through. The molecular sieve system is a packaged unit (further details discussed in the 2011 design report) and ultimately removes 95% of the water and a small amount of ethanol, leaving 99.5% pure ethanol vapor, which is condensed by cross-exchange with the regeneration condensate, then cooled by cooling water and pumped to storage. The regeneration condensate is recycled to the top of the rectification column. The feed entering the molecular sieve unit is slightly heated using additional low-pressure steam.

3.2.3 Cost Estimation

All cost estimates for the pertinent equipment included in the fermentation and distillation section were scaled from original vendor-supplied costs presented in the 2011 ethanol design case [3]. For fermentation, this includes (1) the seed train equipment, scaled to lower seed reactor sizes (noted above) using a 0.6 scaling factor per the methodology discussed in Section 4.2, (2) fermentation vessels, maintaining a consistent 1-MM gal fermentor size as per the original vendor-supplied design, and (3) accompanying agitation/cooling equipment. The majority of the equipment in the fermentation utilizes 304SS as the material of construction. Aside from this explicitly costed equipment, other minor equipment in the fermentation section was costed by applying a balance of plant factor of 8.6% relative to the primary fermentation equipment costs, which account for minor pumps, heat exchangers (primarily the cooling water-supplied hydrolysate feed cooler), and intermediate tankage, based on a similar factor from NREL's 2011 ethanol design report [3]. For distillation/purification, the earlier cost estimates include costs for the beer and rectification columns, as well as accompanying reboilers and condensers, the beer column feed/effluent exchanger, the vent scrubber, and the molecular sieve package. The distillation section utilizes primarily 316SS as the material of construction.

3.3 Area 300: Lipid Extraction and Solvent Recovery

3.3.1 Overview

This is a key process area targeting the extraction of the algal lipid fraction for subsequent cleanup and upgrading to RDB, as the primary contribution to overall GGE fuel yield. In this section, the stillage product from the beer column is routed to a liquid-liquid extraction system which uses hexane solvent to extract the lipids at high solvent-loading in a multi-stage counter-current extraction column. The extracted light oil phase contains solvent, lipids (both fatty acid lipids and polar lipid impurities), and a small amount of water, and is routed to a stripping column to recover the solvent, leaving a high purity (~99.7% total lipids) lipid stream that is sent to subsequent lipid cleanup. The aqueous product is sent to anaerobic digestion. Figure 7 presents a simplified flow diagram of the lipid extraction section.

An important distinction of this process pathway is a demonstrated improvement in extractability of the lipid material when placing this step downstream of dilute acid pretreatment, relative to the reverse order of extraction followed by pretreatment. Specifically, analysis by Laurens et al. presents a compelling contrast for FAME-lipid extractability depending on the sequence of process operations, whereby FAME extractability (FAME measured in extract relative to FAME in initial biomass) ranged between 6%-8% when the biomass was sent directly to solvent extraction first, but improved by *an order of magnitude* to 77%-93% when the biomass was first processed through upstream pretreatment [13]. While based on a limited amount of experimental data, this is a very encouraging result, and validates NREL's previous assumptions in earlier harmonization modeling analyses stipulating a high extraction efficiency [8], while also negating the need for the costly and energy-intensive cell disruption homogenization step assumed in that prior work to be required as part of the extraction process (based in turn on much earlier work from the Aquatic Species Program [47]). It should also be reiterated that this extraction process is done on a wet solids basis, without the need for further evaporative drying, thus dispelling the notion that lipid extraction must always be done on a dry-solids basis.

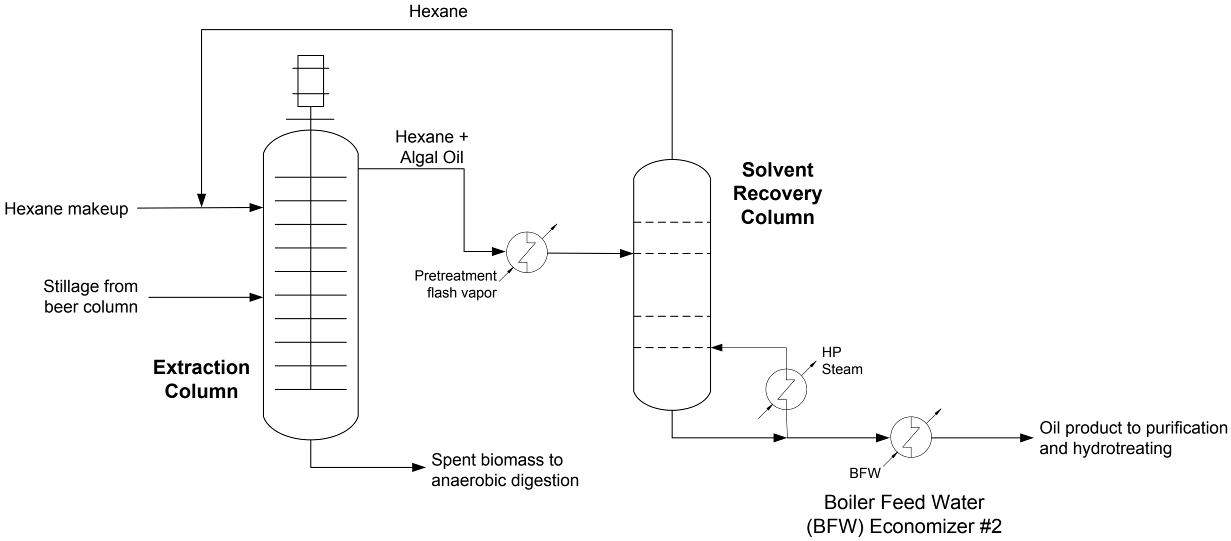


Figure 7. Simplified flow diagram of the lipid extraction and solvent recovery process

3.3.2 Design Basis

The beer column stillage contains roughly 84% water, 2% impurity/byproduct components introduced from upstream operations (primarily lactic acid from assumed contamination losses in fermentation, yeast biomass from ethanol production, plus HMF and ammonium sulfate formed as a byproduct and neutralization product from pretreatment respectively), and 14% residual algal components. This stream is cooled from 125°C to 50°C by cross-exchange with the beer column preheater, and then fed to the top of the liquid-liquid extraction column. Harris Group obtained a vendor design and accompanying budgetary estimate for this critical unit operation, specified as a reciprocating counter-current Karr column, which was quoted by the vendor to be capable of achieving greater than 95% oil extraction utilizing hexane solvent. Additionally, the column designed here is described to be well-suited to handle emulsions given the column's agitation action (the column is agitated rather than static), which is another important design feature as this process presents a risk of forming stable emulsions between the aqueous and lipid phases, depending on specific conditions and chemistry. However, the column is also limited to a fairly low capacity (the summer/spring design capacity in fact requires 16 parallel columns) and is also a substantial capital expense. Design and cost details for this piece of equipment are presented in the next section.

Hexane solvent is fed to the bottom of the extraction column at a 5:1 weight ratio (hexane versus total algal solids, excluding other solids components introduced from upstream processing). This represents a very large amount of solvent circulated through the column, but was deemed reasonable based on extrapolation from bench-scale extractions and to ensure sufficient solvent contact is achieved in the column. This did not appear to be an unreasonable assumption from the viewpoint of Harris Group or the extraction equipment vendor either, but will require further experimental demonstration, both to confirm efficacy at this solvent ratio as well as subsequently high solvent recovery and reuse. Hexane was selected for consistency with the experimental work that also used this solvent system with good results [13] and because it is an ideal choice both from a processing standpoint (hexane is easily recovered from lipids downstream with low miscibility) and a selectivity standpoint (hexane is a non-polar solvent and ideally targets extraction of neutral fatty acid lipids over polar lipids which must be removed later anyway). The model targets 95% extraction efficiency/recovery of the fatty acid lipid fraction and 33% extraction of polar lipid components, with a lower recovery of the polar impurity fraction expected given the use of a non-polar solvent such as hexane. This is still anticipated to err on the conservative side, as phospholipids and other polar components have been measured on part-per-million (ppm) levels in the hexane extract for mid- and high-lipid biomass samples (data not published), substantially less than a 33% yield into the extract phase. Translating to NREL's standard analytical method definitions [27], this equates to 95% extraction of "FAME-equivalent lipids" as defined above. Additional non-FAME (non-fatty acid) lipid components such as hydrocarbons and sterols have also been measured in the biomass, which would primarily partition into the hexane extract phase and ultimately get converted to additional fuel products through hydrotreating. As this fraction was not characterized in the dataset utilized for this analysis (Table 1), it was conservatively not considered in the model. However, this adds a small margin of "safety" to the 95% extraction target stipulated here, as the net *total* lipid yield in the extract would be equivalent to a marginally lower extraction of, for example, 90% FAME lipids combined with 90% of non-FAME (non-fatty acid but still extractable and convertible) lipids if this fraction were present at 2.5% of total biomass. Additionally, 2% water carryover into the solvent/oil phase is assumed.

A note is also warranted here regarding the choice for process configuration. While clear benefits around lipid extractability are enabled by placing extraction after pretreatment (discussed above), several options exist regarding stream routing downstream of the pretreatment fractionation operation. Earlier cursory TEA modeling for the system [13] followed the same configuration as utilized experimentally, namely including a solid-liquid centrifuge separation step immediately following pretreatment neutralization, which split the material into a soluble liquid (sugar) stream sent to fermentation and a high-solids (30%-35% insoluble solids) stream sent to extraction in parallel. While sensible for experimental validation of the technology potential, it is not an optimal configuration in the context of process integration as it leads to considerable product yield losses as well as operability issues at commercial scale. Without the use of a substantially more costly solid-liquid separation system, including a washing step, a centrifuge separation would lead to losses of 25% or more of the soluble sugars carried over into the solids phase given a centrifugation limit of 30%-35% insoluble solids in this phase. These sugars would then no longer be available for fermentation, but instead would ultimately go to anaerobic digestion. Furthermore, the specified extraction column design is limited to a maximum 20% insoluble solids content in the feed stream, which would require re-dilution if separated solids at >30% concentration were fed to extraction from an upstream separation step. In the present design, the whole slurry is first taken through fermentation to maximize sugar retention and ethanol yield, then the whole stillage is subsequently sent to lipid extraction at 14% insoluble solids and does not require any further dilution for operability purposes in the extraction column.

The process for whole-slurry fermentation also adds additional capital expenses by way of higher total fermentation tank volume relative to fermentation of only the liquor phase, but these costs are relatively small compared to other process operations and are outweighed by the benefit of higher ethanol yield using the whole-slurry approach. Additionally, a small amount of additional lipids may be present in the yeast (*S. cerevisiae*) biomass produced during fermentation which may then be available for additional lipid yields during extraction; however, this has not been characterized and is not included here.

Following the extraction operation, the oil/solvent product is sent to a stripping distillation column to recover the solvent. This column was also specified by Harris Group and designed as a packed column with an overall height of 20 ft. A total of three columns operate in parallel, primarily to accommodate high turndown ratios up to the original 5:1 basis if required (e.g., the “non-summer-storage” option). In the base case scenario, including diversion of excess summer capacity and associated reduction in turndown ratio to 1.2:1, a single distillation column would likely be required, which would likely bring economy-of-scale advantages. However, as the cost for the stripping columns alone does not represent a large fraction of overall system capital cost, the savings would likely be marginal. The feed to the distillation column is preheated from 52° to 66°C by cross-exchange with the pretreatment flash vapor condensate exchanger described previously. The reboiler is driven by high-pressure steam (665 psig, 46.3 atm) generated from AD turbine flue gas heat recovery.

The extraction column bottoms product or raffinate stream contains water, residual biomass solids, and a small amount of lipid and solvent losses equating to 67%, 5%, and 0.5% of polar lipids, fatty acid lipids, and solvent respectively, relative to the column feed rates. This stream is routed to anaerobic digestion. Fresh hexane solvent is mixed with the solvent recycle to make up for losses to the raffinate and recovered lipid streams. All key process assumptions for the extraction area, as well as comparisons to pertinent experimentally demonstrated values to date, are summarized in Table 8.

Table 8. Area 300 Guiding Design Basis Assumptions and Comparisons to Values Experimentally Demonstrated to Date [13]

Parameter	Baseline Experimental	Target
Solvent loading (solvent/biomass ratio, dry wt basis)	4.4-5.8	5.0
Extraction method	3-step extraction (extract, dry/recover solvent, re-extract)	Counter-current agitated LLE column
Insoluble solids to extraction	20%	14%
FAME lipid extraction yield [a]	77-93% [b]	95%
Polar lipid impurity extraction yield	ND	33%
Solvent carryover losses (% makeup required)	NA	0.54%

[a] FAME lipids as defined in [27]

[b] Values shown are for *Scenedesmus* strain; *Chlorella* strain extraction efficiencies were lower

3.3.3 Cost Estimation

Cost estimates for the primary unit operations in this section were provided by Harris Group based on input from vendors. Namely, four Karr extraction columns were originally specified to accommodate a total lipid feed rate of 5,894 kg/hr to the extraction system. Under the present scenario based on the maximum summer/spring capacity design, this would translate to 16 equivalent columns operating in parallel. Each column is designed with a 6-ft-diameter agitated zone, 9-ft-diameter expanded ends, and an overall height of 60 ft. Materials of construction are 304L for the column shell and 304SS for column internals. Each column requires 40 hp (30 KW) of power. The resulting purchase cost was quoted as $1.98 MM per column in 2013-year dollars, which translates to a total installed cost of $63.5MM for the extraction system, or roughly 30% of the total ISBL (inside battery limits) equipment installed costs, thus representing a significant expense for the overall facility; however, the vendor and Harris Group expressed confidence that this particular column design, while costly, is well-suited for this particular application with >95% expected performance for extraction efficiency (including the ability to handle emulsions if they were to form, as noted previously).

Harris Group also supplied cost estimates for the solvent stripping recovery columns and associated reboilers (designed with three each, as noted above). The costs were quoted as $714,000 and $150,000 for the distillation column and reboiler respectively, based on purchase costs per unit in 2009-dollars. The solvent recovery column is a packed tower with dimensions of 10-ft-diameter x 20-ft-height, with all equipment utilizing 304SS metallurgy. The extraction feed heat exchanger cost is already included with the previously-quoted cost under the distillation section, e.g., the beer column feed-effluent (stillage) heat exchanger, and the solvent recovery column preheater exchanger cost was already discussed above in Section 3.1.3 (cross-exchange with pretreatment flash condensate).

3.4 Area 400: Product Purification and Upgrading

3.4.1 Overview

Area 400 first purifies the recovered lipid product to remove impurities which would be problematic for the subsequent catalytic upgrading step. Lipid cleanup consists of a series of degumming, demetallization, and bleaching steps to remove gums, metals, and other impurities with the use of phosphoric acid, wash water, silica, and clay. These operations are similar to those utilized in the biodiesel industry, and although this system incurs non-trivial costs, it was deemed necessary to include, at a minimum until further experimental data becomes available on the hydroprocessing operation, to validate or refute its use in the context of integration with upstream operations (e.g., introduction of salts through pretreatment and neutralization and additional changes incurred through fermentation). The cost impact attributed to removing the lipid purification steps is considered below in the Sensitivity Analysis section. The resulting purified neutral lipid material is then sent to upgrading in a hydrotreater for purposes of deoxygenation and saturation, primarily yielding a diesel-range RDB product with a small yield of naphtha coproduct, as well as off-gas which is combusted in the biogas turbine. The process schematic for the Area 400 operations is depicted in Figure 8.

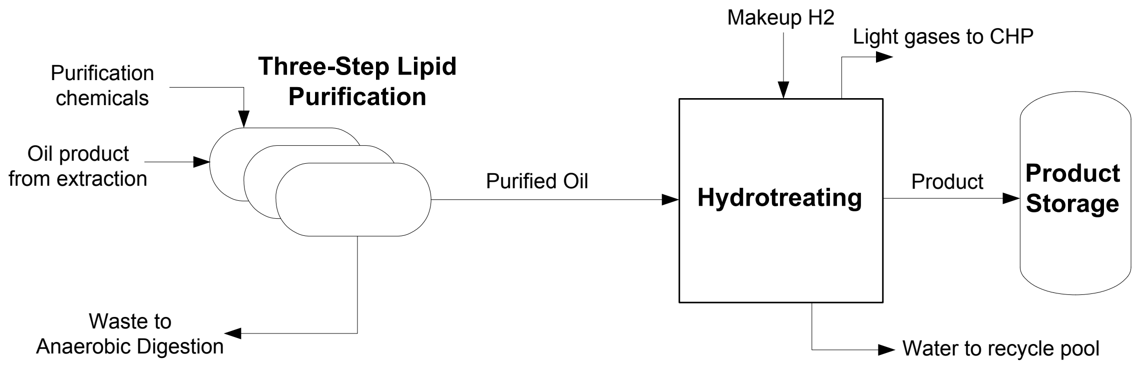


Figure 8. Simplified flow diagram of product purification and upgrading

3.4.2 Design Basis

As modeled, the bottoms product exiting the solvent recovery stripping column contains 98% neutral (fatty acid) lipid, 1.7% polar lipid impurity components, 0.3% hexane, and a trace amount of water. While phospholipids are of primary concern within the polar lipid fraction for their propensity to form gums and deactivate catalysts [48, 49], in reality the extracted polar lipid fraction has also been characterized to contain phytol (derived from chlorophyll) and nitrogen species (equivalent to ~300-400 ppm nitrogen), and the extracted lipids from *Scenedesmus* were also measured to contain >35 ppm metals, primarily sodium, iron, copper, phosphorus, and calcium, attributed to both upstream salt formation and trace nutrients in the algal biomass (data not published). In light of this information, Harris Group and the vendors they contacted thought it prudent to include a lipid cleanup step to remove these impurities for downstream catalyst protection in the hydrotreater using commercially available technology common to similar operations used in biodiesel processing.

A general block diagram for the purification system is shown in Figure 9. As shown, gums (including phospholipids) are removed in a degumming step with the addition of phosphoric acid followed by centrifugation with wash water. Subsequently, demetallization is performed by adding silica, forming a slurry, and then filtering out the spent silica. This step is followed by bleaching with clay to remove additional metals and impurities, which again involves forming a slurry with clay followed by filtration. Cakes formed from filtering the silica and clay from the oil, as well as the material removed from degumming, are routed to anaerobic digestion, with the small phosphoric acid content of this stream assumed to be available as additional phosphorus for subsequent recycle to the cultivation ponds. The vendors Harris Group spoke to noted that the phospholipid content in the product oil is expected to be less than 10 ppm, possibly as low as 2 to 3 ppm, based on the analytical characterization of the extracted oil furnished to the vendor. Additionally, they expect metals content to be less than 5 ppm for potassium (K), calcium (Ca), magnesium (Mg), and sodium (Na). Testing is required to specifically determine the expected performance of these systems. It should be noted that when treating vegetable oils, a neutralization step is performed during degumming wherein lye is added to remove free fatty acids. Here, free fatty acids are valuable feedstocks for hydrotreating and therefore this step is not included in the concept presented here. The process takes place at 110°C (230°F), with the incoming lipid material first cooled to this temperature after exiting the solvent recovery column by cross-exchange with a second high-pressure pretreatment steam economizer.

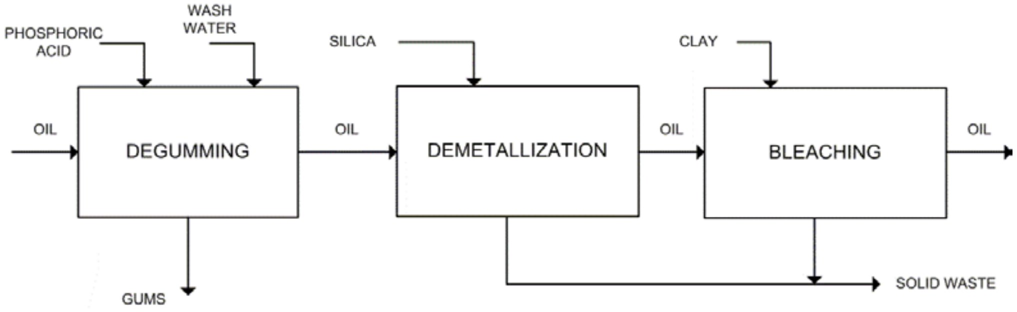


Figure 9. Block diagram schematic for lipid purification system

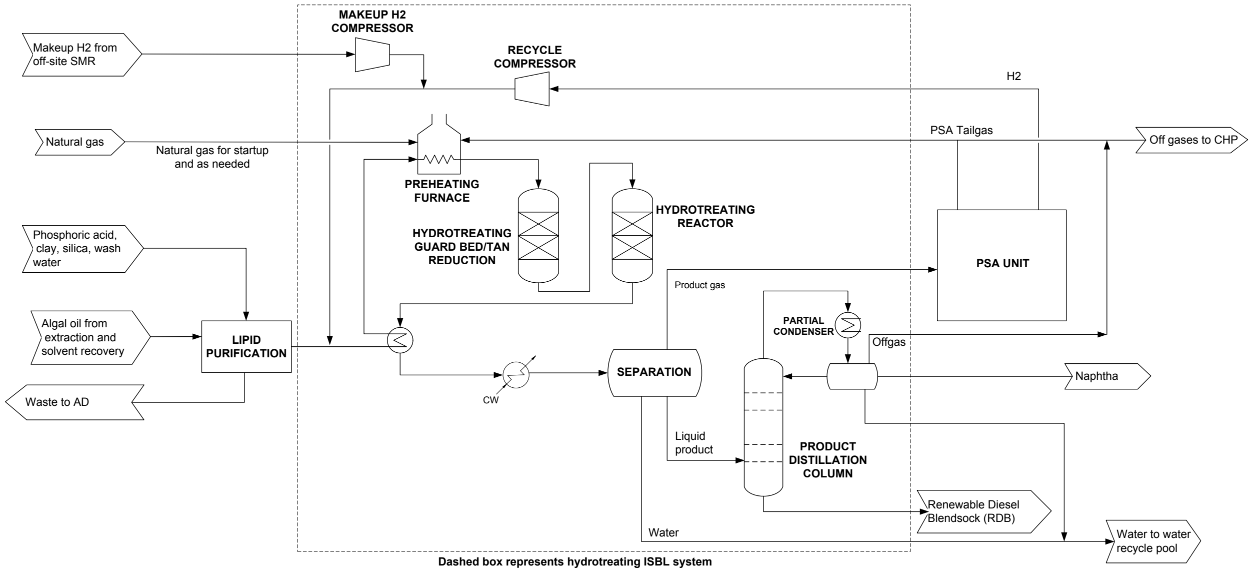
A summary of the operating conditions and chemical requirements for the lipid cleanup step is presented in Table 9. The resulting purified lipid product is modeled as essentially 100% neutral lipids (modeled here as the triglyceride triolein, although diglycerides, monoglycerides, and free fatty acids are also present), cooled down to 93°C after passing through the various cleanup operations including mixing and removing wash water.

Table 9. Operating Conditions and Material Requirements for Lipid Purification

Degumming Step	
Total lipid feed rate – design basis	22,576 kg/hr (summer/spring design case)
Feed temperature	110°C (230°F)
Phosphoric acid dosing (wt% of feed)	0.19%
Wash water demand (wt% of feed)	10%
Demetallization Step	
Silica dosing (wt% of feed)	0.1%
Bleaching Step	
Clay dosing (wt% of feed)	0.2%
Purified lipid product rate – design basis	22,124 kg/hr (summer/spring design case)
Total gums/waste product to AD	12.5% of feed (polar lipids + added materials)

Following lipid cleanup, the material is upgraded in a hydrotreater to refine the oxygenated intermediate material into saturated paraffin components suitable for blending as a diesel blendstock. The design and cost estimates applied for this section are largely consistent with the hydrotreating facility described in NREL's 2013 design report focused on hydrotreating free fatty acids [6], but will be repeated again here. In summary, Harris Group consulted with a number of vendors, including catalyst suppliers and process licensors, to obtain budgetary estimates for a hydrotreating facility that may be expected to process the feed as modeled here into deoxygenated diesel-range paraffins. A number of cost estimates were provided for such a facility, but the vendors declined to provide specifics without confidentiality restrictions regarding associated process and design conditions (e.g., reactor sizing dictated by space velocities, operating temperatures and pressures, hydrogen partial pressure, etc). Thus, to estimate reasonable process conditions for modeling purposes, a combination of literature and Harris Group estimates were applied. Namely, a report by Marker et al. describing experimental conditions and yields for hydroprocessing vegetable oils was leveraged which remains one of the most thorough reports on the subject to date, and was jointly authored by UOP, NREL, PNNL, and Michigan Technological University [50]. This report described relatively mild hydrotreating temperature and pressure conditions at 350°C (660°F) and 35 atm (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 480 psig at reactor inlet). As confirmed by Harris Group, high hydrogen feed is necessary in part to provide a quench to control the large exotherm associated with deoxygenation. A reasonable liquid hourly space velocity (LHSV) associated with this process would likely be in the range of 0.9-1.4 hr^{-1} [50], which was confirmed by an estimated value of 1.2 hr^{-1} from Harris Group.

A schematic diagram of the cleanup and hydrotreating process operations is shown in Figure 10. This diagram includes an assumption of a two-stage hydrotreater, with the first stage including a guard bed for metals, as well as a possible mechanism for handling the high acid (TAN) content as measured for extracted lipids from *Scenedesmus*. Additionally, a cost estimate was also included for a pressure swing adsorption (PSA) unit in the hydrotreater gas recycle loop, primarily to remove CO_2 formed by decarboxylation reactions in the hydrotreater, as well as propane produced from the glycerol backbone of the triglyceride lipids (discussed below). The design and cost estimates for the PSA unit were also provided by vendor input. In the Aspen model, the gas recycle feed stream to the PSA consists primarily of H_2, CO_2, CO, and propane (after cooling the reactor effluent and knocking out the majority of the produced water, which is routed to the water recycle pool). Given this gas composition, the PSA vendor provided a cost estimate for a patented PSA unit which utilizes a zeolite-based sorbent to purify the recycled hydrogen to roughly 97 wt%, with the product containing approximately 0.1% CO_2, 1% CO, and 2% propane, relative to the recycle feed stream to the PSA composed of 32% H_2, 47% CO_2, 2% CO, and 19% propane (mass basis), attributed to the hydrotreating product yields discussed below. Given this composition, particularly the presence of propane, the PSA system incurs a greater hydrogen loss than the NREL 2013 cellulosic design report (91% H_2 recovery versus >95% in the cellulosic case), requires a much greater power demand (2.3 MW in the peak summer/spring design case versus 35 KW in the cellulosic case), and is significantly more costly (discussed in the next section). This difference is because in the earlier cellulosic pathway report, the targeted product was exclusively free fatty acids which do not contain the three-carbon glycerol backbone present in lipid triglycerides, which is rejected as propane from the hydrotreater and makes for a more complicated hydrogen purification operation. The overall economic sensitivity to the assumed PSA hydrogen recovery is evaluated in Section 5.1. The PSA tailgas, containing approximately 4% H_2, 26% propane, 67% CO_2, and 3% CO (mass basis), is split to provide the necessary heat to the hydrotreater furnace (thus avoiding external natural gas to drive this operation), with the remainder sent to the biogas turbine (A500).



Dashed box represents hydrotreating ISBL system

Figure 10. Schematic diagram of hydrotreating operations included in vendor design estimate

Oxygen removal from the lipid feed may occur by way of hydrodeoxygenation (HDO) or decarboxylation (DCO) reactions in the hydrotreater. The relative degree of one reaction pathway over another may be tailored by catalyst and reaction conditions, with a tradeoff exhibited between higher hydrogen consumption to reject oxygen as water via HDO (this carries important implications particularly on the LCA/sustainability metrics of the system when sourcing hydrogen from conventional fossil sources) and decreased carbon yields to reject oxygen as CO_2 via DCO (with decreased product yield translating directly to cost impacts). For this work, hydrotreater product yields were based on the previously-cited report by Marker et al. [50], which presents an optimized experimental case favoring DCO over HDO by a factor of over 2:1 (72% of the reaction occurs via DCO and 28% via HDO), translating to high CO_2 yields relative to H_2O and lower fuel yields than would be possible with higher HDO ratios. This is a similar fractional split favoring the DCO pathway as presented in other sources [51-53]. The associated hydrogen consumption was set at 1.7 wt% of the oil feed rate, also consistent with other data favoring DCO [51, 52]; however, additional hydrogen is lost from the system via the PSA as noted above, with all hydrogen consumption/losses met with fresh makeup hydrogen. Sustainability implications for sourcing hydrogen from off-site natural gas steam methane reforming are presented in Section 5.3. The experimental literature data from which the modeled yields were derived indicated no heavier-than-diesel residual material (vacuum gas oil or VGO) was present in the hydrotreater product stream [50], thus avoiding the need for a second residual hydrocracking unit in this design. The resulting product yields are summarized in Table 10.

Table 10. Product Yields from Hydrotreating (derived from [50])

Product	Yield, wt% of oil feed
Total reactor product yield	101.7%
H_2 consumption	-1.7%
Reactor product composition:	
Fuel-range hydrocarbons	80.0%
Diesel-range	*78.5%*
Naphtha-range	*1.5%*
CO_2	12.1%
CO	0.6%
H_2O	1.7%
Off-gas (propane)	5.6%
Total	100.0%

The product yields summarized above were slightly adjusted from earlier NREL models for algal lipid upgrading [8, 54], to more precisely match the original literature data source [50] and to ensure 100% element balance closure across this step (based on the representative triolein component modeled for the neutral lipid fraction). Given the rigor of the original literature dataset from which this model is derived, as well as highly favoring DCO mechanisms for oxygen rejection, this is a reasonable, if not conservative, set of assumptions around hydrotreating process yields and compositions. Nonetheless, it will be important to validate or refine these assumptions based on experimental data specific to hydroprocessing the extracted algal oil from the HLSD strain/harvest basis assumed here. NREL has recently begun experimental efforts in this regard, with the primary objective of better quantifying hydroprocessing conditions, yields, and product characteristics associated with algal lipid upgrading, but data are not yet available at the time of this writing to leverage in the design report modeling effort. The assumptions employed here will be revisited as such data become available.

The diesel-range product from the hydrotreating operation is composed primarily of straight-chain paraffins with high cetane value, making it potentially valuable as a diesel-range blendstock. It is recognized that to produce a stand-alone final fuel product meeting ASTM diesel specifications, the straight-chain paraffinic product from hydrotreating would require further isomerization to address likely cloud point issues [14]; however, the high cetane and low sulfur content expected to be exhibited by the RDB product is expected to make such a product valuable for diesel blending purposes, and the removal of oxygen during hydrotreating suffices to meet DOE requirements consistent across other technology pathways for a finished product [9]. NREL experimental efforts will also work to better understand product quality and tradeoffs in further processing requirements brought about by adding a hydroisomerization step after deoxygenation.

The product stream exiting the hydrotreating reactor is split into gas, liquid, and aqueous phases, in which the gas phase is sent to the PSA for subsequent hydrogen recycle (as described above) and the aqueous phase (primarily water) is sent to the water recycle pool. The liquid product is fractionated in a distillation column, whose design and cost is included with the rest of the vendor-supplied hydrotreating facility quotation. The distillation column is steam-stripped using low-pressure steam produced from the turbine flue gas heat recovery steam generator. The RDB product fraction exiting the distillation column, modeled as paraffins primarily in the C13-C20 range, is diverted to an Aspen calculator block to calculate the higher heating value of the product, for purposes of translating to a GGE yield (see Section 2.4.2). The naphtha product fraction, modeled as paraffins primarily in the C6-C12 range, is produced as a small coproduct. The column off-gas is combined with the PSA tail gas and sent to the AD biogas turbine. Table 11 summarizes the key process assumptions discussed above for the hydrotreating step.

Table 11. Summary of Hydrotreating Process Assumptions

Oil feed rate (at standard conditions) – design capacity (summer/spring capacity)	154,040 gal/day (583 m^3/day)
Average reaction temperature	660°F (350°C)
Pressure	500 psig (35 atm)
Hydrogen partial pressure	480 psig (34 atm)/6,000 SCF/bbl
Reactor LHSV	~1.2 hr^{-1} (estimated) [a]
Reaction H_2 consumption	1.7 wt% of feed
Hydrogen loss across recycle PSA	9%
Makeup hydrogen rate – design capacity (summer/spring capacity)	5.4 MM SCFD (145,311 Nm^3/day)
RDB product rate – design capacity (summer/spring capacity)	145,758 gal/day (552 m^3/day)
Naphtha coproduct rate – design capacity (summer/spring capacity)	3,403 gal/day (13 m^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)

3.4.3 Cost Estimation

The bleaching, demetallization, and degumming cost estimates were furnished by vendors via Harris Group, and translate to an installed cost of ~$7MM based on the summer/spring design capacity. A marginal power demand of roughly 0.01 KWh/kg oil feed rate was applied, also per vendor feedback. Variable operating costs for the purification chemicals (and all chemicals for the facility) are discussed in Section 4.3. The original vendor quote also included the requirement for steam to heat the oil material to at least 85°-95°C (185°-200°F) from a lower temperature, but the present case sends the oil product directly to the cleanup stage from the solvent recovery bottoms; thus instead is first cooled down to a slightly higher temperature (110°C) by further preheating the feed water for the pretreatment high-pressure steam generator in a second economizer following the rectification column bottoms cooler. The cost for this heat exchanger was scaled from earlier NREL model costs for a plate-and-frame heat exchanger [3].

Consistent with NREL's 2013 biochemical hydrocarbon design report, the vendor-supplied hydrotreating facility cost is an "inside battery limits" (ISBL) cost estimate, and includes costs for reactors, makeup and recycle gas compressors, fired heater, separation vessels, and distillation. Additionally, while catalyst replacement cost is not explicitly specified, one vendor noted that catalyst would be an insignificant part of the total cost estimate for this particular system, at a typical catalyst lifetime of two years and the given lipid feed specification. Based on a number of vendor-supplied cost estimates, combined with Harris Group's expertise on the subject as well as typical industry costs for petroleum hydroprocessing, an original cost estimate of $23MM (installed cost) was utilized as consistent with NREL's 2013 design report focused on free fatty acid hydrotreating, which translates to $30MM installed in the present analysis based on the design summer/spring capacity. The uncertainty range associated with this cost estimate was quoted as -50%/+100%, thus is most appropriately viewed as an order-of-magnitude cost for processing this type of lipid/fatty acid feedstock. The overall cost impact associated with this given range of uncertainty in the hydrotreating cost estimate is evaluated in Section 5.1. 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 attempting 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. Given the relatively mild hydrotreating conditions listed in Table 11, which are in line with a relatively high-quality material as would be expected to be present from purified fatty acids/lipids (e.g. <0.05% nitrogen, <0.01% sulfur, ~10% oxygen based on internal analytical characterizations), both Harris Group and NREL thought it appropriate to utilize such a baseline cost estimate here.

The PSA unit was provided as a separate cost item, as this is not a typical unit operation in conventional hydrotreating systems, but is required here primarily to remove CO_2 and CO (via the DCO reaction which produces these products) as well as propane (via rejection of the 3-carbon glycerol backbone from triglycerides) formed during hydrotreating. The cost estimate for this operation was provided by a PSA vendor. As noted above, the design basis for this operation was based on a patented zeolite sorbent system, allowing for reasonable hydrogen recoveries and high rejection of the other components, albeit at considerable expense given the difficulty of removing propane. Indeed, the PSA system installed cost was estimated at nearly $12MM, or roughly 40% of the cost of the rest of the hydrotreating facility. This is a substantial cost increase for this unit relative to the 2013 design report which excluded propane formation. Harris Group also evaluated the cost for an amine system as an alternative CO_2 removal option, and found that this alternative may be more cost-effective relative to the PSA option, but this also is likely to exhibit a higher heat demand for amine regeneration, as well as require the addition of a purge stream to mitigate propane buildup in the recycle loop. As more definition becomes available for the intermediate lipid product characterization and hydrotreating specifics, the options for hydrogen purification and/or purge in the recycle gas line may be revisited. The power demand for the PSA unit in this case was estimated as 0.10 KWh/kg oil feed to the hydrotreating section.

3.5 Area 500: Anaerobic Digestion/CHP

3.5.1 Overview

The aqueous product (raffinate) exiting the lipid extraction column is combined with the waste stream from the lipid purification step, then cooled and routed to anaerobic digestion (AD). The AD step is used in part to provide a degree of wastewater treatment/ sludge stabilization, but of equal importance in the context of a fully integrated process with algae cultivation, AD is used as a means to recover nutrients (primarily nitrogen and phosphorus, but also other minor nutrients not considered in the model) for recycle to the cultivation step, as well as to recover residual carbon in the form of biogas for heat and power benefits. While AD alone is not envisioned to necessarily provide sufficient treatment of the facility wastewater for subsequent discharge or recycle to the *conversion* process, microalgae cultivation itself is also an effective water treatment mechanism [55]. Thus, similar to earlier NREL designs, including that utilized in the previously referenced harmonization report, AD is included as a means for spent biomass utilization and a precedent for enabling these important sustainability benefits (nutrient recycle and energy recovery), with the effluent intended for recycle to the algal growth ponds (maintaining cultivation outside the scope of this work). AD organisms break down the incoming organic carbon into biogas consisting primarily of methane and CO_2, as well as nitrogen and phosphorus fixed in the residual material into forms that may be assimilated by cells upon recycle to algal cultivation. Sulfate is also converted to H_2S, which partitions into the biogas.

While a high fraction of N and P nutrients are assumed to be recovered into the liquid effluent stream, the remaining fraction primarily allocates to a high-solids digestate cake.

The biogas stream is routed to a gas turbine, which combusts the material to generate electricity and heat. Sufficient power is generated to provide a net-export coproduct credit, when considering strictly the conversion facility boundaries here (i.e. negating upstream power demands for cultivation and dewatering). The turbine flue gas heat is also recovered to generate steam for meeting facility heating and live steam demands. This combined heat and power (CHP) benefit is another key sustainability driver that has been previously shown to provide important LCA advantages in an integrated algal processing pathway [8, 56, 57], justifying the continued use of AD/CHP in this case as well. The flue gas exits the turbine at high temperature and is passed through a series of three heat recovery steam generator exchangers, which generate high-pressure steam for live injection into the pretreatment system (Area 100), high-pressure steam for heating purposes in a closed-loop utility, and low-pressure steam for closed-loop heating, respectively. After considering all facility steam and heating demands, overall heat balances leave a small heat deficit which must be satisfied with supplemental natural gas co-fired in the turbine. This aspect must be taken into consideration in work that proposes higher value coproducts from the raffinate. Both the economic and LCA impacts from such a scheme demand careful attention to the tradeoffs implicit in each proposed coproduct. The flow diagram for the AD/CHP system is shown in Figure 11**.**

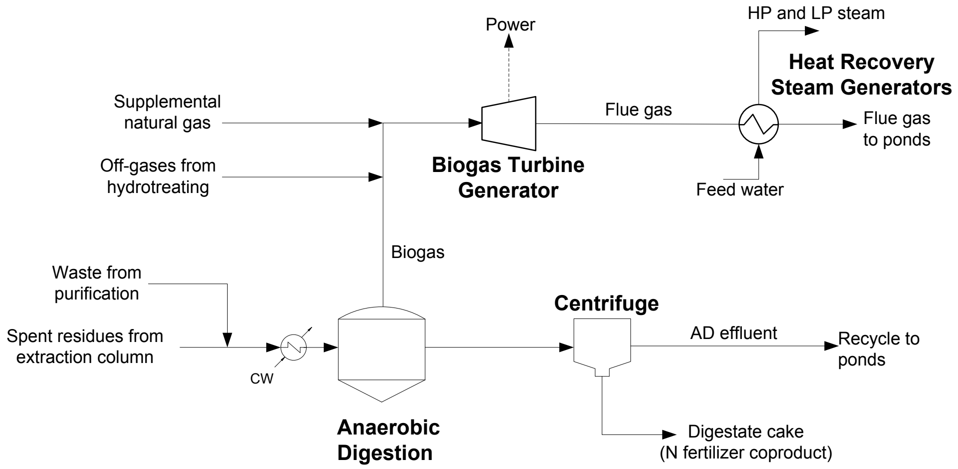


Figure 11. Simplified flow diagram of the AD/CHP system

3.5.2 Design Basis

The AD system fundamentally employs the same process assumptions as documented in detail in the 2012 algae harmonization report. This includes a typical 35°C operating temperature (achieved by cooling the AD feed stream down from 52°C in a cooling water exchanger, although thermophilic operation at 50-60°C is also possible), and 20-day hydraulic retention time (HRT) based on peak summer/spring capacity [58]. Applying the same guidance used in prior efforts as applied in the harmonization report, this translates to a volatile solids loading factor of roughly 3.3 g/L-day (4.4 g/L-day total solids basis) using standard industrial AD design criteria

[58], associated with a total feed solids content of approximately 9%. Based on peak summer/spring capacity flowrates, the resulting total AD volume required was calculated to be 32.1 MM gal (121,500 m^3), requiring two digester units each sized at 16.1 MM gal, assuming 27 MM gal maximum digester size as applied in the 2013 design report [6]. Relative to the harmonization baseline which otherwise forms the design basis for this step, the present model applies a higher solids loading rate (4.4 g/L-day versus 2.4 g/L-day originally) due to a proportionately higher total solids feed concentration (9% versus 5%). Volatile carbon destruction to biogas was maintained at 48%, consistent with the harmonization basis. While this translated to a biogas methane yield of 0.3 L CH_4/g TS in the harmonization model, it is reduced to 0.2 L/g TS in the present analysis (roughly 0.27 L/g volatile solids), due to the introduction of additional non-digestible solids in upstream processes (e.g., salts) as well as lower carbon content of the digestible solids attributed to removing the majority of both lipids *and* carbohydrates upstream, rather than only lipids. Consistent with the harmonization basis, the biogas is composed of 67% CH_4 and 33% CO_2. Finally, AD heat and power demands were set at 0.22 $KWh_{thermal}$/kg total solids and 0.085 KWh/kg total solids respectively as documented in the harmonization report. The AD operational and yield assumptions are summarized in Table 12.

Although the original assumptions around operating conditions (HRT and solids loading factor) and biogas yields as established in the harmonization baseline were largely derived from literature values on AD of spent algal residues with often incomplete or otherwise non-applicable datasets, experimental studies from NREL and Washington State University (WSU) utilizing lipid-extracted algae confirmed these assumptions with very close agreement, namely identical methane yield at the same HRT and roughly 15% higher total solids loading rate [59]. While this validates the original AD assumptions for lipid-extracted algae and the present assumptions are sensible in relation to the earlier basis (e.g., lower methane yield given less carbon available for digestion), the present assumptions carry a higher degree of uncertainty, particularly with respect to operability of the AD system, given a lower carbon-to-nitrogen (C/N) ratio in this case. As the AD feed here has been extracted of both lipids and carbohydrates, the residual material is enriched in protein and thus will have a lower C/N ratio than earlier model scenarios; currently estimated C/N ratio as modeled is roughly 8.9 (molar basis). It is known that AD systems exhibit operability issues as the C/N ratio decreases due to nitrogen inhibition [60], but a specific limit is not well-quantified for algal biomass residues. NREL/WSU's recent experimental work confirmed that no issues were observed for AD of lipid-extracted algal biomass, thus negating this concern up to a point, but further research is needed to confirm this remains the case in the present scenario.

The total slurry feed rate to the AD unit in the summer/spring design case is 243,500 kg/hr or 21,480 kg/hr total solids. Of the total solids, roughly 75% are digestible, with non-digestible solids identified as ammonium sulfate salt, ash, fermentation organism (*S. cerevisiae*), clay and silica materials used in lipid purification, and hexane. As noted above, the model is based on a freshwater algal strain and subsequent processing; if a saline scenario were considered, salt levels could translate to more challenging AD operation. Aside from methane and CO_2 in the biogas, H_2S is also present as a product of sulfur breakdown. This is subsequently converted to SO_2 after combustion in the gas turbine, present in the flue gas at approximately 280 ppm_v. As the flue gas is not vented to the atmosphere, but rather is destined for recycle to the ponds upstream (mixed with fresh flue gas from coal power plant operations which also contains SO_2), no desulfurization is included to remove sulfur after the turbine and SO_2 is assumed to be absorbed into the cultivation ponds. However, sulfur control may still be needed prior to the turbine for metallurgy corrosion mitigation purposes; this is not considered further here as the sulfur content will be highly dependent on source of cultivation CO_2, i.e. purified CO_2 versus bulk flue gas, with the measured sulfur content of the biomass presented in Table 1 based on carefully controlled growth conditions and media composition which may not reflect commercial systems.

Beyond the biogas fraction, consistent assumptions were employed for both the AD liquid effluent and solids digestate cake product streams as described in the harmonization report, namely 80% N and 50% P recovered in the effluent stream (relative to digestible N and P entering AD, excluding ammonium sulfate salt) with the remainder allocated to the digestate cake which exits the system at 30 wt% insoluble solids after passing through a centrifuge. A small amount of flocculant may be intermittently required to reach this solids content in the digestate, but the cost would be marginal and this is not considered here. Of the nitrogen, 5% is assumed subsequently lost to volatilization, translating to 76% net nitrogen recovery in the effluent. The solids digestate material has been identified as a key contributor to fugitive methane emissions when tracking this parameter [61], and results in a nearly negligible coproduct credit using the assumptions described below. As an alternative, the digestate cake fraction may be instead rerouted to the cultivation ponds along with the AD liquid effluent fraction, where early observations have indicated it substantially biodegrades given the highly aerobic conditions of the open ponds [62]. This option is not considered here to maintain consistency with prior work and as it is not yet known if it would negatively impact cultivation over a long term, but will be considered in future analysis.

The AD effluent stream is intended to be recycled back to the algal cultivation step upstream, thereby reducing the amount of makeup N/P nutrients required for biomass growth. As such, the stream is assigned a coproduct value after translating the elemental components to equivalent diammonium phosphate (DAP) and ammonia, with DAP and NH_3 coproduct values equal to each component's purchase cost as applied upstream. The digestate cake stream is also assigned a coproduct value, equating to \$500/tonne of bioavailable nitrogen (assuming 40% of the total nitrogen in the digestate cake is bioavailable), also consistent with the harmonization report and as originally described by [55]; this translates to roughly \$4/tonne total solids. It is justified to include coproduct credits here for AD digestate, nutrient recycle to cultivation, and CO_2 recycle to cultivation (discussed below), because these recycle credits were not included in the original \$430/ton biomass cost basis discussed above (documented in DOE's MYPP report, table B-4 [9]). Consequently, rather than reducing the biomass feedstock cost to account for these recycles, the equivalent cost savings are instead counted here as coproduct credits to (marginally) reduce the final fuel MFSP. The cost impact to final fuel MFSP attributed to removing any of these coproduct credits is considered below in the Sensitivity Analysis section (see Section 5.1).

Table 12. Anaerobic Digestion Assumptions

Metric	Basis	Source
Operating temperature	35 °C	[58]
Hydraulic retention time	20 days (design basis)	[58]
AD heat demand	0.22 KWh/kg total solids (thermal)	[8]
AD power demand	0.085 KWh/kg total solids	[8]
Volatile solids loading rate	3.3 g/L/day	Extrapolated from
(Total solids loading rate)	(4.4 g/L/day)	[58]
Total solids in feed	9%	Per Aspen model
CH_4 yield in biogas	0.2 L/g TS (48% carbon destruction)	Consistent with [8]
Biogas composition	67% CH_4 / 33% CO_2 (vol basis)	[8]
N recovery in effluent	76% net (80% - 5% volatilization loss)	[8]
P recovery in effluent	50%	[8]
Bioavailable N in digestate cake	40%	[55]
Digestate cake insoluble solids	30%	[6]

The biogas is next routed to a gas turbine generator, after mixing with hydrotreating off-gases including the PSA tail gas and distillation overhead off-gas. A gas turbine is employed here rather than alternatives such as an internal combustion engine, as it provides high electrical efficiencies (at large scale as is the case here) and avoids high fugitive methane emissions typically encountered with internal combustion engines [61]. The biogas is first compressed to 12.2 atm (165 psig) and mixed with injection air at a ratio to maintain combustion temperature (prior to turbine expansion) near 1155°C, as is typical for the specified unit [63]. In the previous harmonization baseline model, the turbine had been specified to let down to a slightly elevated pressure of 5 psig (relative to standard gas turbines exhausting near atmospheric pressure), in order to overcome hydraulic losses associated with flue gas cooling for steam generation, recycle to the cultivation ponds, and sparging through a membrane into a 1.5m sump. The present analysis also assumes the turbine flue gas is destined for recycle to the ponds in the same manner, but given additional steam generation demands in the present system (e.g. for live steam injection to the pretreatment reactor as well as low-pressure steam for ethanol distillation), the flue gas exhaust pressure was further elevated to 7 psig (1.5 atm). This difference of 2 psig incurs a loss of 17% in total amount of power generated, but also increases the exhaust temperature by 22°C (translating to more heat available for steam generation). After accounting for parasitic losses for biogas and air compression, the turbine electrical efficiency is roughly 20%, compared to 25% in the harmonization baseline at the lower exhaust pressure, or 33% as typical for standard gas turbines exhausting to atmospheric pressure. Thus, while incurring a substantial penalty in overall power generation potential, the assumptions employed here are reasonable in order to allow for all flue gas cooling and recycle requirements in the broader context of a fully integrated process with upstream cultivation/dewatering connections. Given that the turbine operating conditions were tailored to allow for flue gas recycle to the cultivation pond step (taking a significant electrical efficiency penalty), it is justified to then include a coproduct credit for flue gas recycle, as the recycled flue gas translates to a cost savings for the cultivation step by requiring less makeup flue gas delivered from offsite, which is more costly than on-site recycle [8].

The flue gas recycle credit was calculated as \$33.9/ton CO_2, based on back-calculating the explicit cost per ton of offsite CO_2 pipeline delivery in the original harmonization benchmark model. This is a reasonable value, relative to standard costs for flue gas carbon capture ranging from \$35-45/ton [64, 65]. Given the low CO_2 content of the flue gas (7 wt%) as typical for gas turbines operating with substantial excess air, the parasitic energy penalty for transporting the flue gas with high inert levels is not ideal, and further analysis is required to better understand if a more optimum recycle or venting scenario exists. Additionally, more detailed pond dynamics may dictate that recycle occur intermittently, for example based on variable pond pH, which would either require compressing and storing flue gas or venting during such intermittent downtimes (a more likely case). The high-concentration CO_2 from the ethanol fermentor vent is also added to this value to determine the total CO_2 recycled to the cultivation ponds. The cost impact to MFSP associated with removing this CO_2 recycle credit is evaluated in Section 5.1.

The flue gas exhaust exits the turbine at a temperature of 638°C. The exhaust is then used to generate steam for the various facility heating and live steam requirements in a series of heat recovery steam generators (HRSGs). The first exchanger cools the flue gas to 490°C to raise superheated steam at 268°C which is injected to the dilute acid pretreatment system. The second exchanger further cools the stream to 430°C to raise high-pressure saturated steam at 260°C for closed-loop utility heating serving the solvent recovery distillation reboiler and the hydrotreating distillation pre-heater. The final exchanger further cools the stream to 140°C to raise low-pressure saturated steam for closed-loop utility heating serving the ethanol distillation reboilers, the molecular sieve unit, and AD heating demands (this is the largest heat demand of the three exchangers). Specifics on the steam side of the exchangers and associated costing assumptions are discussed in Section 3.7. As mentioned previously, while the turbine generates sufficient electricity to make the conversion facility a net-power exporter to garner a small power coproduct revenue, the overall facility is slightly heat-limited given the heat raised by flue gas cooling relative to total heat/steam demands (driven primarily by ethanol distillation). Consequently, a small amount of supplemental natural gas is imported and co-fed to the gas turbine to satisfy the marginal heat deficit.

3.5.3 Cost Estimation

The design and cost estimates for the AD system are based on NREL's recent 2013 cellulosic design report model [6], which in turn was based on updated estimates for AD (and other operations) provided from an engineering subcontract with wastewater vendor Brown and Caldwell. Under that subcontract, Brown and Caldwell provided a rigorous review and detailed design of a number of wastewater treatment operations, drawing on support from technology suppliers. The AD system designed by Brown and Caldwell consisted of multiple AD vessels each sized at 27 MM gal, thus the same maximum size criteria was applied for this analysis. Four such AD units were specified in the original design at a total purchase cost of \$25.8MM (2012-dollars), translating to \$6.5MM per individual vessel; this cost also includes supporting equipment including recycle pumps, biogas blowers, mixers, centrifugation, and a biogas emergency flare. The original cost was specified to be scaled according to feed chemical oxygen demand (COD) to the AD system, which was an appropriate basis for scaling in the original cellulosic biochemical process model. However, in this case COD is a less certain value as estimated by the Aspen model given the underlying biomass component characterizations. Consequently, the cost was scaled in the present analysis according to required AD volume, calculated based on the volatile solids loading rate discussed above and constrained by a maximum size of 27 MM gal per single AD unit. The design and cost basis employed here is derived from a low-rate bulk volume flow (BVF) digester system designed for NREL's cellulosic biofuel model, which originally assumed long retention times on the order of >40 days; the high biogas yields and short retention times assumed in this design may require alternative AD designs, such as smaller (more costly) agitated dual-reactor systems. This will be a continued area of further investigation.

Added to the resulting overall AD system cost, a digestate centrifuge was also included as 5% of the cost of the AD system, consistent with the 2013 cellulosic design report. The gas turbine cost basis was left unchanged relative to the 2012 harmonization baseline, at an installed cost of $445/KW generated (2006-dollars) [63], which was originally presented in the *GTW 2006 Handbook* [66].

3.6 Area 600: Storage

3.6.1 Overview

This portion of the plant provides bulk storage for process chemicals and the liquid fuel products. The tanks explicitly costed include diesel, ethanol, and naphtha storage, as well as water for fire suppression. Additional tankage for chemical inputs is included as a balance of plant factor. All assumptions for storage costs and logistics are set consistently with other recent design report practices, and may not consider more detailed logistics challenges for loading/unloading, rail car transport, etc. (considered beyond the scope of this stage of analysis).

3.6.2 Design Basis

Table 13 shows the major storage requirements for the present design. Tanks for diesel, ethanol, and naphtha products were each sized to ensure at least seven days of storage, while the fire water storage tank was scaled from NREL's 2013 design report according to dry biomass feed rate (summer/spring capacity case) to the pretreatment reactor.

Table 13. Storage Requirements for Major Tanks

Material	Size
RDB product	Sufficient to contain >7 days of production: 1 carbon steel tank @ 750,000 gal
Ethanol product	Sufficient to contain >7 days of production: 1 carbon steel tank @ 750,000 gal
Naphtha product	Sufficient to contain >7 days of production: 1 carbon steel tank @ 25,000 gal
Fire water	4 hours of fire suppression @ 2,500 gpm: 1 glass-lined carbon steel tank @ 600,000 gal (scaled down to lower dry feed rate in present design)

Other supplemental tanks and pumps were not considered explicitly here, but were costed based on a 20% balance of plant factor relative to the total cost of the major tanks listed above, based on similar minor tankage costs considered in detail in NREL's 2013 design report.

3.6.3 Cost Estimation

The costs for the A700 storage section were based on NREL's prior cellulosic biochemical design reports [3, 6], and costs are scaled according to new material flowrates to estimate new tank prices if applicable.

3.7 Area 700: Utilities

3.7.1 Overview

Area 700 contains the utilities required by the conversion facility. Area 700 tracks cooling water, chilled water, process water, plant and instrument air, the clean-in-place (CIP) system, and steam generation. In the model, Area 700 also tracks the electricity usage throughout the plant.

The process water manifold in Area 700 provides fresh process water, which is assumed suitable for all plant users, at a constant pressure to the facility. Water is provided to the boiler for steam generation, the cooling tower makeup, the CO_2 vent water scrubber, and the lipid purification operations. 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 biogas turbine, have their own compressors specified. The CIP system provides hot cleaning and sterilization chemicals to pretreatment and fermentation operations.

3.7.2 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 14. The percentage of cooling duty contributed by each user is shown in Figure 12. The total cooling water demand is 20 MM kcal/hr, based on average cooling requirements over the four individual modeled seasons.

Table 14. Cooling Water Users

Chiller Condenser	The chilled-water loop requires cooling water to condense the refrigerant. The cooling water duty for the chiller condenser is set equal to the total load on the chilled-water loop.
Pretreatment Flash Cooler	Cools the pretreatment flash condensate to suitable temperature for mixing into the water recycle pool.
Fermentation Cooler	Cools the hydrolysate feed to fermentation (37°C).
Beer Column Condenser	Condenses the beer column reflux.
Molecular Sieve Cooler	Provides cooling and condensation during regeneration of the molecular sieve.
Anaerobic Digester Feed Cooler	Cools the AD feed stream to required operating temperature (35°C).
Hydrotreating Reactor Effluent Cooler	Cools the effluent stream of the hydrotreater following cross-exchange with the reactor inlet stream.

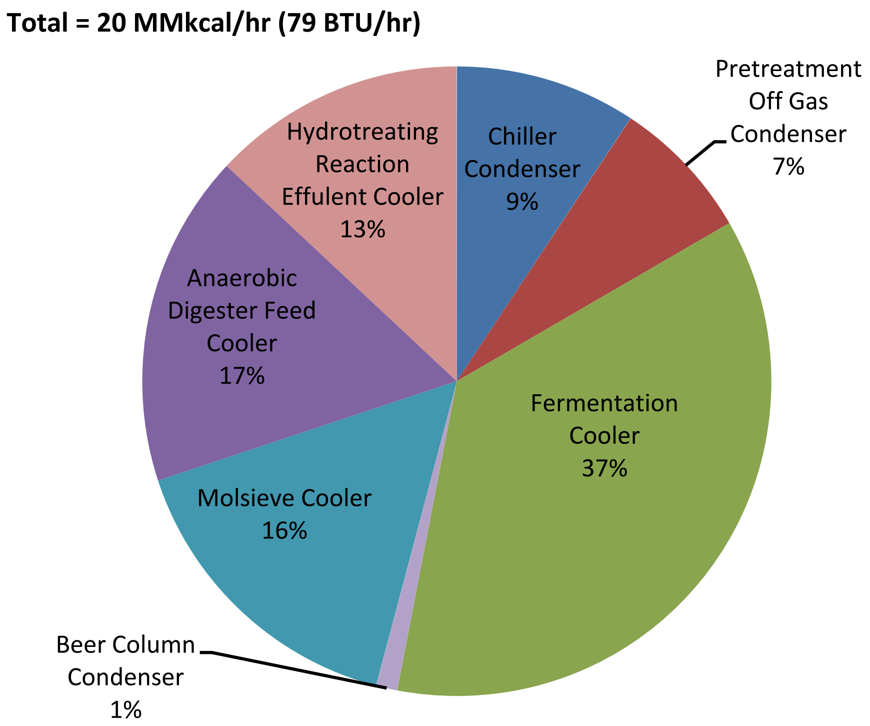


Figure 12. Average cooling water heat duty distribution between major users

The largest user of cooling water is the fermentation cooler, associated with a large heat duty to cool the pretreatment hydrolysate down from 68°C to 37°C after first recovering heat in the pretreatment feed-effluent exchanger. Other larger cooling demands include the AD feed cooler, molecular sieve cooler, and hydrotreating reactor effluent cooler, all operating between 35°-38°C on the process side. Aspen computes the cooling tower evaporation rate based on a temperature drop from 37°C to 28°C. Beyond evaporative water losses from the cooling tower, it was also 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 Trane centrifugal chillers, each originally specified at 2,350-ton of refrigerant cooling. 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 chilled-water loop provides cooling to the ethanol and seed train fermenters to maintain fermentation temperature at 37°C.

Fresh water is assumed to enter the facility at 33°C and is split in several ways to meet process needs. Clean water is used as boiler feed water to create steam for direct injection to the process (pretreatment reactor), as makeup for utility steam loop blowdown, as makeup water for the cooling tower, scrubbing water for the CO_2 vent water scrubber, and as wash water for use in the lipid purification unit. The process water tank is designed for an 8-hour residence time. The process water pump pumps water from the tank into the facility.

The plant and instrument air systems provide compressed air for pneumatic tools and cleanup, instrument operation, and air cooling systems. This does not include major air demands such as biogas combustion. The plant air compressor was originally sized for 400 SCFM at 125 psig, and scaled according to dry biomass feed rate to the facility. An instrument air dryer and surge tank were designed to provide clean dry air at a consistent pressure to the instrument air system.

Steam is used in the plant both as a utility heat source and for direct injection into the process in the pretreatment reactor system. Heat is recovered from the flue gas exhaust of the biogas turbine, which is supplemented with natural gas to meet the heating balance needs of the facility (discussed previously). High-pressure superheated steam (176 psig/ 13 atm, 268°C) is generated for direct injection into the pretreatment reactor system by cross-exchange with the first HRSG exchanger. Heating demands of other processes are satisfied by two separate steam loops that create saturated steam at high pressure and low pressure. The high-pressure saturated steam (665 psig, 46.3 atm) is used for the reboiler of the solvent recovery distillation column (Area 300), and to heat the inlet stream on the RDB product distillation column (Area 400). The low-pressure steam (39 psig, 2.7 atm) is used for heating the ethanol distillation beer and rectification column reboilers, the molecular sieve heat requirement, and the heating needs of the anaerobic digester. Additionally, a small amount of low-pressure superheated steam is also created for steam stripping of the A400 product distillation column.

About 62% of the electricity generated by the gas turbine is used throughout the plant to power items such as pumps, agitators and compressors. The surplus is sold to the grid for credit. The distribution of total plant power utilization among all areas is shown in Figure 13. The largest power users in the facility are the pretreatment and product purification/upgrading sections, each accounting for 19% of total power demands relative to total generation; in turn, these are driven primarily by power requirement for the pretreatment reactor feed system, the hydrotreating PSA unit, and the hydrotreating makeup hydrogen compressor. Similar to the cooling water values, the absolute value shown below for total power generated (18.5 MW) is the average of the four individual seasonal models, as is the allocation to process area. The average annual excess power produced and sold to the grid as a coproduct is roughly 6.9 MW, or more accurately, 54.9 MM KWh/year. While power is viewed as a net exported coproduct in the current work scope focused strictly on the conversion facility, when expanding the scope to a fully integrated process including upstream cultivation and dewatering, a net power *import* would still likely be required given high power demands upstream, e.g. for algal pond paddle wheels [8]. Thus the power export value indicated here would go towards decreasing overall power demand for the fully integrated process.

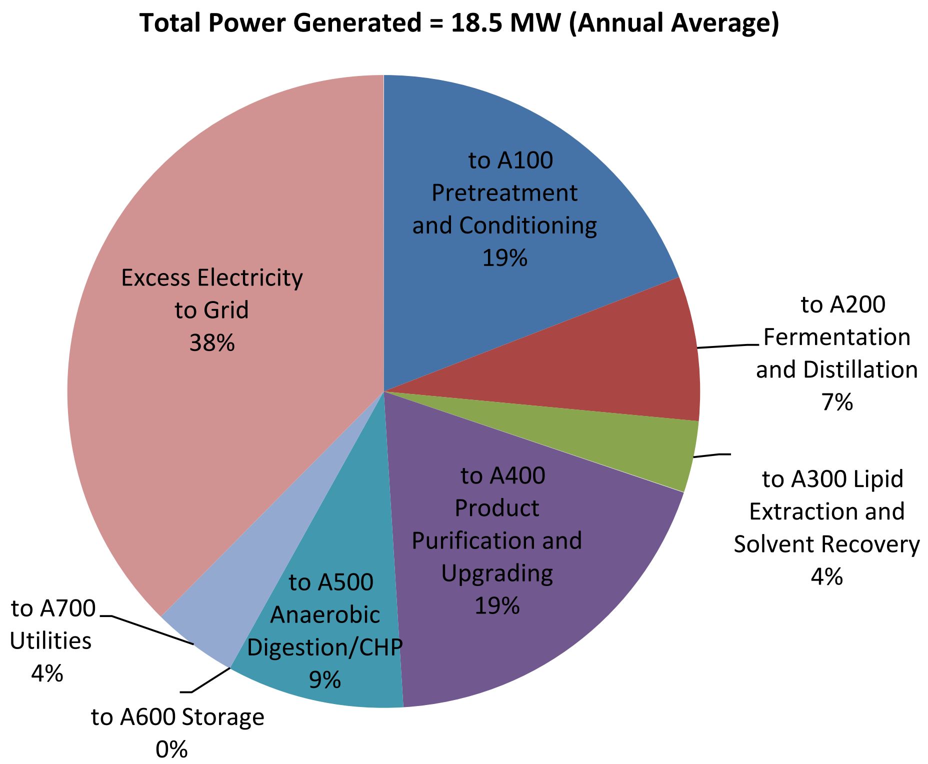


Figure 13. Average distribution of plant electricity utilization by process area

3.7.3 Cost Estimation

All cost estimates for the utility equipment in A700 were maintained consistent with the basis values used in NREL's 2011 and 2013 biochemical design reports. 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 reduced cooling water throughput estimated here. Harris Group had estimated costs for the cooling water circulation and makeup pumps using their historical cost database. The material of construction for the cooling water loop is carbon steel. The cost for the chiller came from a recent quote for a similarly-sized system. Aspen Capital Cost Estimator [15] was used to cost the steam HRSG equipment including the high- and low-pressure steam boiler heat exchangers. Harris Group had also used their historical cost 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 (MFSP) required to obtain a zero net present value (NPV) with a finite 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 NREL's 2011 and 2013 cellulosic design reports for assumptions which remain unchanged [3, 6].

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 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 stakeholders 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. 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-dollars through 2017 to permit comparison of future feedstocks, conversion technologies, and other alternative scenarios. However, the present equipment costs were obtained in 2012-dollars or 2013-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-dollars or 2010-dollars for unit operations which were previously provided by Harris Group and vendors in support of NREL's earlier design reports, and which 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 prior design reports. Thus, the details will not be repeated here, but will be summarized briefly. Capital costs provided in a year other than 2011-dollars were adjusted using the Plant Cost Index from *Chemical Engineering Magazine* [67] 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 2013-dollars, we assumed the same Plant Cost Index value from 2012 (all cost quotes which fall in this category were provided in the first half of 2013). Consistent with methods utilized for all other recent design reports, costs are not inflated over the 30-year facility lifetime (i.e. inflation is kept constant over the modeled facility lifetime). Similar to capital costs, for chemical costs we used the Industrial Inorganic Chemical Index from SRI Consulting [68]. Employee salaries were maintained from 2009-dollars and were scaled using the labor indices provided by the United States Department of Labor Bureau of Labor Statistics [69]. The general formula for converting nominal values to real values and for adjusting dollars to a common basis is:

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

4.2 Total Capital Investment (TCI)

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 analysis, 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 remains the same as described in the 2011 and 2013 cellulosic design reports, and again further detail can be found there which 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 listing of the equipment is provided in Appendix A, along with equipment purchased and installed costs. As described in prior design reports utilizing similar equipment, Harris Group obtained many package quotes for recent cost estimates, in which a given unit operation and all of its support equipment were quoted under one price. The installation factor for such packages can be relatively low because most of the engineering is already included in the price. Additionally, equipment furnished as a pre-fabricated skid generally has a lower installation cost relative to equipment cost.

The purchased cost for a given component reflects a baseline equipment size. As changes are made to the process, the required equipment size may be different than what was originally designed. 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 NREL’s 2011 ethanol report). 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). A good example of this is the new pretreatment reactor cost information discussed in Section 3, which has resulted in a modified cost estimate to utilize multiple discrete sections of the horizontal portion of the pretreatment reactor (e.g. the portion which is in contact with acid) to allow for variations in residence time through this section of the reactor.

Once the total equipment cost has been determined in the year of interest, we must add several other direct and indirect costs to determine the total capital investment (TCI). Site development and warehouse costs are based on the inside-battery-limits (ISBL) equipment costs (Areas 100-400) and are considered part of the total direct cost (TDC). Beyond ISBL Areas 100-400, the other process areas are considered "outside battery limits" (OSBL) including Areas 500-700 (AD/CHP, storage, and utilities). 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. Similar to the choice of 10% IRR, the cost factors applied in estimating FCI are based on assuming n^{th}-plant economics, particularly for less prescriptive elements such as project contingency factor (see Section 1.4). The sum of FCI and the working capital for the project is the TCI. Table 15 summarizes these categories and additional factors. The values assumed for each respective factor were maintained consistently with those discussed in NREL's 2011 and 2013 design reports.

Table 15. Additional Costs for Determining Total Capital Investment (TCI)

Item	Description	Amount
	Additional direct costs	
Warehouse	On-site storage of equipment and supplies.	4.0% of the installed cost of ISBL equipment (A100-400)
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 the 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, and unusual foundations; sales, use, and other taxes; freight, insurance in transit, and import duties on equipment, piping, steel and instrumentation; overtime pay during construction; field insurance; project team; transportation equipment, bulk shipping containers and plant vehicles.	10% of TDC

Table 16. Project Cost Worksheet Including Total Direct Costs and Total Capital Investment

Process Area		**Purchased Cost [a]**	**Installed Cost [a]**
Area 100: Pretreatment and Conditioning		$ 42,800,000	$ 65,600,000
Area 200: Fermentation and Distillation		$ 10,500,000	$ 19,600,000
Area 300: Lipid Extraction and Solvent Recovery		$ 34,900,000	$ 71,500,000
Area 400: Product Purification and Upgrading		$ 48,300,000	$ 48,300,000
Area 500: Anaerobic Digestion/CHP		$ 20,500,000	$ 21,600,000
Area 600: Storage		$ 2,700,000	$ 4,500,000
Area 700: Utilities		$ 4,200,000	$ 5,700,000
Totals (excl. Area 100)		**$ 163,900,000**	**$ 236,700,000**
Warehouse	4.0%	of ISBL	$ 8,200,000
Site development	9.0%	of ISBL	$ 18,400,000
Additional piping	4.5%	of ISBL	$ 9,200,000
Total Direct Costs (TDC)			**$ 272,600,000**
Prorateable expenses	10.0%	of TDC	$ 27,300,000
Field expenses	10.0%	of TDC	$ 27,300,000
Home office & construction fee	20.0%	of TDC	$ 54,500,000
Project contingency	10.0%	of TDC	$ 27,300,000
Other costs (start-up, permits, etc.)	10.0%	of TDC	$ 27,300,000
Total Indirect Costs			**$ 163,500,000**
Fixed Capital Investment (FCI)			**$ 436,100,000**
Land			$ 1,800,000
Working capital	5.0%	of FCI	$ 21,800,000
Total Capital Investment (TCI)			**$ 459,800,000**
Lang Factor (FCI/purchased equip cost)			2.7
TCI per annual GGE			$7.37/GGE
2011-dollars			
			HSLD + Store

[a] Capital costs are based on peak (summer/spring) design capacity following excess summer drying/storage scenario

4.3 Variable Operating Costs

Variable operating costs, which include feedstock, other raw materials, and co-product credits, are incurred only when the process is operating. In this pathway, variable operating costs by definition also vary as seasonal feed rates and associated equipment throughputs fluctuate, in contrast to fixed operating costs (and equipment capital costs) which do not. Quantities of raw materials used and coproducts produced were determined using the Aspen material balance, based on running individual Aspen models for each season and then taking the average of all four seasons in setting the net annual operating expenses and coproduct revenues. Table 17 documents the costs and sources of chemicals used in the process and Table 18 summarizes the variable input/output flowrates on a per-season basis, as well as the resulting variable costs and coproduct revenues on a per-year and per-GGE-of-total fuel basis.

The cost basis for all material costs in the present model, which were also used in NREL's most recent 2013 cellulosic design report, were left unchanged and similarly scaled to 2011-dollars. This includes sulfuric acid, ammonia and DAP (both as chemical inputs as well as coproduct credits translated from nitrogen and phosphorus balances allocated to the AD effluent stream), process water, and hydrogen. As discussed in the 2013 design report, hydrogen was assumed here to be purchased as a product from standard natural gas-derived steam methane reforming (SMR), priced according to a recent DOE Hydrogen Program report at \$1.57/kg (2012-dollars) [70]. Consistent with the 2013 report, given that hydrotreating is not yet well-defined with direct experimental data for lipid upgrading (but rather is based on literature data), the decision was made to maintain the assumption around purchased hydrogen rather than including on-site generation (from AD biogas and/or natural gas reforming), so as not to place too much burden on an alternative on-site hydrogen production system without first validating the hydrotreating operation, particularly hydrogen consumption. Additionally, as this would be considered an OSBL process, using sound literature values for hydrogen price reduces overall uncertainty relative to building a new hydrogen generation system into the model. Further details behind the rationale for the hydrogen price assumptions may be found in the 2013 cellulosic design report. The overall cost sensitivity to the assumed hydrogen price is presented in Section 5.1, using a reasonable maximum and minimum hydrogen cost range as presented in the referenced DOE report, namely \$1.10/kg-\$2.00/kg.

Aside from feedstock cost discussed earlier, other raw material inputs to the present model which were not utilized in prior analyses include hexane (costed based on an industrial spot price); phosphoric acid, clay, and silica for lipid purification (costed based on price estimates furnished by Harris Group); and natural gas. Cost for natural gas, used both for drying excess summer feedstock for use in the winter as well as for utility supplemental heating, was set at \$5.1/thousand ft^3 [11]. This is within the current price range although slightly on the conservative side relative to recently published prices of \$3.89 and \$4.66/thousand ft^3 in 2012 and 2013 respectively [71].

This design includes a number of coproducts, including AD digestate cake (valued at \$500/tonne bioavailable N as a fertilizer product, as discussed earlier), AD effluent nitrogen and phosphorus destined for recycle to upstream cultivation (translated to equivalent NH_3 and DAP rates, using coproduct credits equal to these material's costs, as noted above), flue gas and fermentor vent CO_2 also destined for recycle to upstream cultivation (valued at \$33.9/ton of CO_2, as discussed above), excess power sold to the grid, and naphtha. The products and recycle streams from AD being sent back to upstream cultivation are warranted to include for (small) coproduct revenues here, as they were not originally counted in the \$430/ton biomass feedstock cost basis, thus instead are accounted for here (see discussion in Section 3.5). The electricity export price, taken to be the average wholesale price as determined by the North American Electric Reliability Corporation, was left unchanged from prior NREL 2011 and 2013 design reports, set at 5.72 cents/KWh. This corresponds to roughly 84% of the EIA 2011 average industrial retail price for electricity of 6.8 cents/KWh [72], which is consistent with published values for wholesale electricity as a fraction of retail prices [73]. While diesel and ethanol products are summed together according to GGE heating values to form the overall GGE "fuel product" for reasons discussed above, naphtha was assigned as a separate coproduct given its much smaller production rate and likelihood that it may not be appropriate directly as a fuel or blendstock, but rather may require further processing. The naphtha coproduct value was set as \$3.25/gal as consistent with [11]. The sensitivity impact on overall fuel MFSP attributed to removing any of these coproduct credits is considered below in Section 5.1.

Table 17. Chemical Costs + Coproducts and Sources

Component	Cost (2011$)	Source
Inputs		
Biomass feedstock	$0.2150/lb	2013 MYPP, $430/dry ton AFDW @ 80% moisture [9]
Hexane	$0.5394/lb	SRI Chemical Economics Handbook [68]
Sulfuric acid, 93%	$0.0499/lb	Basic Chemical of Omaha via Harris Group [6]
Ammonia	$0.2496/lb	Terra Industries via Harris Group [6]
Diammonium phosphate	$0.5492/lb	Ronas Chemicals via Harris Group [6]
Natural Gas	$0.1051/lb	$5.1/thousand ft^3 [11]
Hydrogen	$0.6838/lb	DOE report, SMR derived H_2 [6, 70]
Process Water	$0.0001/lb	Peters & Timmerhaus [6]
Phosphoric acid (lipid cleanup)	$0.3600/lb	Harris Group cost database
Silica (lipid cleanup)	$1.0000/lb	Harris Group cost database
Clay (lipid cleanup)	$0.3000/lb	Harris Group cost database
Coproducts		
AD digestate cake	$0.2419/lb of N	$500/tonne of bioavailable N (40% of digestate N) [6, 55]
Ammonia recycle	$0.2496/lb	Set equal to input cost per above
Diammonium phosphate	$0.5492/lb	Set equal to input cost per above
Naphtha	$3.25/gal	Consistent with algal HTL design report [11]
Carbon dioxide recycle	$0.0169/lb	$33.89/ton CO_2; extrapolated from flue gas pipeline cost in [8]
Power coproduct to grid	$0.0572/KWh	Consistent with prior design reports [6]

Table 18. Variable Operating Costs. (Includes seasonal flowrates and resulting annual costs; HLSD summer drying/storage design basis.)

Process Area	Stream Description	Summer Usage (kg/h)	Fall Usage (kg/h)	Winter Usage (kg/h)	Spring Usage (kg/h)	Cost ($/ton) 2011$	MM$/yr (2011$)	Cent/GGE Fuel (2011$)
				Raw materials				
A100	Feedstock input [a]	84,224 [b]	47,767	15,705 [b]	54,749	430.00	190.03	304.74
	Sulfuric acid, 93%	2,433	2,099	1,985	2,406	99.85	1.95	3.12
	Ammonia	786	678	641	777	499.27	3.14	5.04
	Natural gas for summer drying	8,111	0	0	0	210.27	3.72	5.97
A200	DAP	112	97	92	111	1,098.38	0.99	1.58
A300	Hexane	993	856	810	982	1,078.85	8.58	13.75
	Phosphoric acid	43	37	35	42	720.00	0.25	0.40
	Silica	23	19	18	22	2,000.00	0.36	0.58
	Clay	45	39	37	45	600.00	0.22	0.35
A400	Hydrogen	546	471	446	540	1,367.51	5.98	9.59
A700	Process water	85,179	73,448	69,745	84,247	0.29	0.20	0.32
	Supplemental natural gas for heating	949	799	748	937	210.27	1.58	2.53
	Subtotal						**216.98**	**347.95**
				By-products and credits				
A400	Naphtha (gal/hr)	142	122	116	140	$3.25/gal	3.35	5.37
A500	AD Digestate N [c]	71	62	58	71	483.79	0.28	0.44
	AD Ammonia	740	638	604	732	499.27	2.96	4.74
	AD DAP	322	277	262	318	1,098.38	2.83	4.54
	CO_2 Recycle	34,005	29,279	27,673	33,618	33.89	9.22	14.78
	Grid electricity (KW)	7,637	6,465	6,116	7,528	5.72¢/KWh	3.14	5.04
	Subtotal						**21.77**	**34.91**
Total variable operating costs							**195.21**	**313.05**

[a] Feedstock flowrate is on an ash-free dry weight basis. Values represent raw feed rate of feedstock to the facility from dewatering, prior to diversion of summer excess to storage.

[b] Total amount of summer feedstock diverted to storage for winter usage is 29,475 kg/h.

[c] Flowrate of bioavailable N in total AD digestate cake stream

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 NREL's prior 2011 and 2013 cellulosic design cases [3, 6], which in turn were based in large part on NREL's 2002 ethanol design report [2] and/or Peters and Timmerhaus [74].

Table 19 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 NREL's 2011 ethanol report.

Table 19. Fixed Operating Costs

Position	2011 Salary	# Required	2011 Cost	MM$/yr (2011$)	Cent/GGE Fuel (2011$)
Labor and supervision					
Plant manager	155,400	1	155,400		
Plant engineer	74,000	2	148,000		
Maintenance supervisor	60,257	1	60,257		
Maintenance technician	42,286	12	507,429		
Lab manager	59,200	1	59,200		
Lab technician	42,286	4	169,143		
Shift supervisor	50,743	4	202,971		
Shift operators	42,286	20	845,714		
Yard employees	29,600	4	118,400		
Clerks and secretaries	38,057	3	114,171		
Total salaries			2,380,686	2.38	3.82
Labor burden (90%)			2,142,617	2.14	3.44
Other overhead					
Maintenance	3.0%	of ISBL	6,147,070	6.15	9.86
Property insurance	0.7%	of FCI	3,052,774	3.05	4.90
Total fixed operating costs				**13.72**	**22.01**

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 [68]. 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.

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

Discount Rate

For this analysis, the discount rate (which is also the internal rate of return [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. [75] 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."

Equity Financing

For this analysis, it was assumed that the plant would be 40% equity financed. The terms of the loan were taken to be 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 prior NREL design reports. Figure 14 illustrates the sensitivity of minimum fuel selling price to the percentage of equity financing and the after-tax discount rate (the IRR).

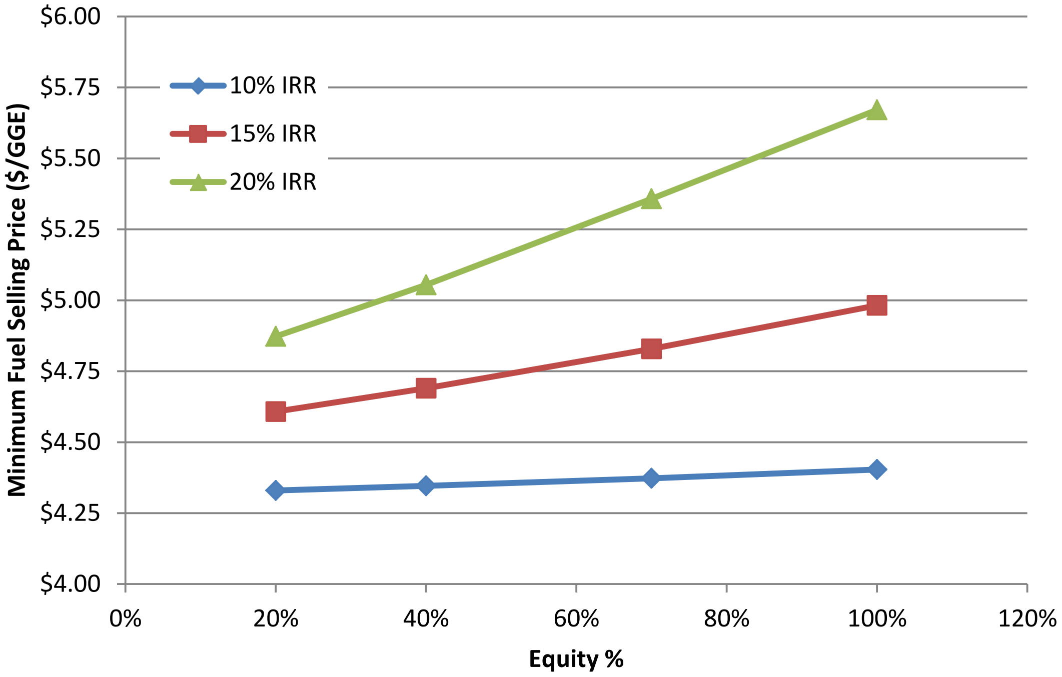


Figure 14. Sensitivity of MFSP to IRR and % equity (8% interest on a 10-year loan)

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 [76], a 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 power plant (i.e., gas turbine) equipment, which has a 20-year recovery period (Asset Class 49.13). Again, all assumptions employed here are maintained consistently with prior design reports.

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-gallon 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 nine 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%); this is consistent with previous design report practice standards and is applied consistently here as well.

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 cellulosic design reports. Perry and Green [77] indicates 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 a $1.5 billion investment) can be constructed in 24 months [78]. Certainly this algal biofuel process 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 20 summarizes the schedule for construction and the cash flow during that time. Twelve months are added before construction for planning and engineering.

Table 20. 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, and preliminary plant and equipment arrangements complete.	8%
12	24	All detailed engineering including foundations, structure, piping, electrical, site 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 based on typical 60 MM gal/yr grain-to-ethanol facility.

Start-Up Time
Perry and Green [77] indicates that for a moderately complex plant, start-up should be about 25% of the construction time, or 6 months in this case. While the 2011 ethanol model assumed a start-up time of 3 months under an n^{th} plant assumption, this is extended back to 6 months in the present 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.

Working Capital
Peters and Timmerhaus [74] defines 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 prior work, namely 5% of fixed capital investment.

Table 21 summarizes the parameters used in the discounted cash flow analysis. Using these parameters, plus the cost information in Table 16, Table 18, and Table 19, the resulting minimum fuel selling price of the primary fuel products translated to an energy-equivalent GGE yield is $4.35/GGE (2011-dollars). Based on individual product heating values, this equates to a selling price of $2.95/gal for ethanol and $4.57/gal for RDB. It is worthwhile to distinguish this MFSP result, based on 10% IRR and other economic inputs noted above, from the *cost of production* (i.e. breakeven cost to cover capital and operating expenses at 0% IRR), as these terms are often incorrectly used interchangeably; the associated cost of production for this case is estimated at $3.68/GGE. Table 22 summarizes the yields and conversion costs for the present design. According to the methodology of Cran [79], the expected accuracy of the TCI analysis is +/- 25%. If we apply this uncertainty to the TCI, the impact on the MFSP is +/-$0.26/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 15 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 $430/dry U.S. ton delivered to the conversion facility and has not been broken down).

Table 21. 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 (turbine) plant depreciation	150% DB
Steam (turbine) 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 start-up	50%
Variable costs incurred during start-up	75%
Fixed costs incurred during start-up	100%

Table 22. Summary of Yields, Rates, and Conversion Costs

Feedstock rate	1,339 dry U.S. ton/day – average annual rate (AFDW basis)
On-line time	7,920 h/yr (330 days/yr, i.e. 90% on-line factor)
Total fuel yield (RDB + ethanol)	141.1GGE/dry U.S. ton feedstock (AFDW)
Total fuel production rate (RDB + ethanol)	62.4 MM GGE/yr
RDB production rate	46.3MM GGE/yr (44.1 MM gal/yr)
Ethanol production rate	16.1 MM GGE/yr (23.7 MM gal/yr)
Total equipment cost	$237MM
Total capital investment (TCI)	$460MM
TCI per annual gallon	$7.37/GGE
Minimum Fuel Selling Price	**$4.35/GGE**
Feedstock contribution	$3.05/GGE
Conversion contribution	$1.30/GGE

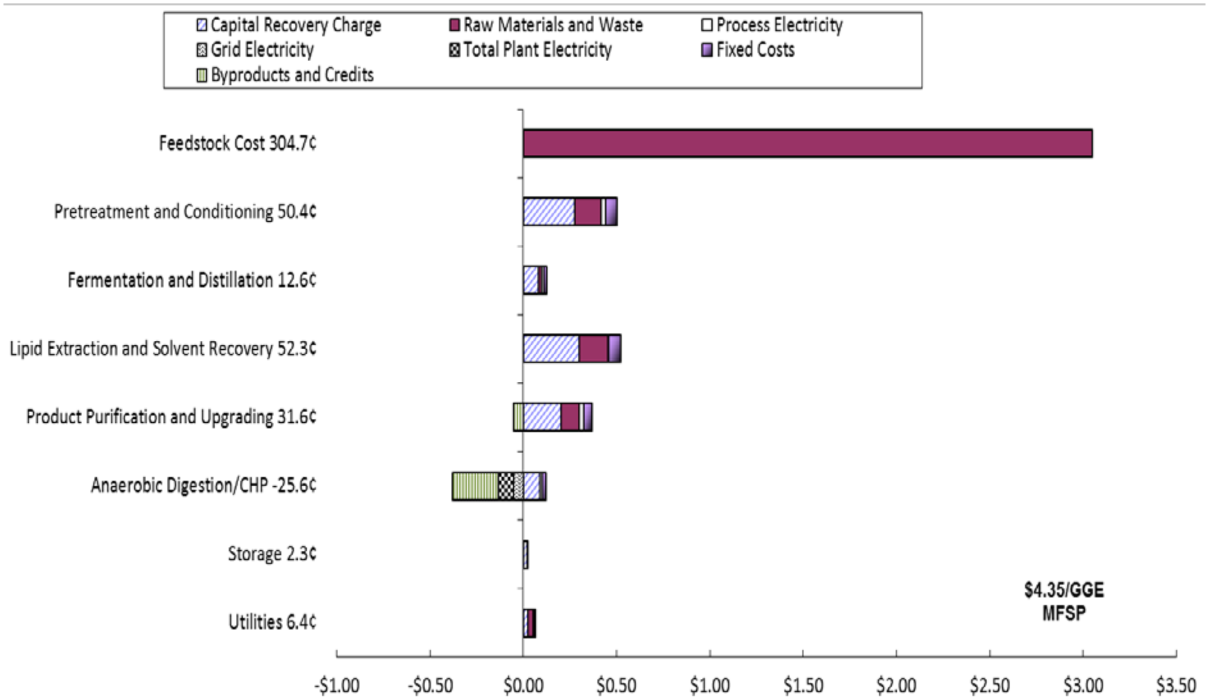


Figure 15. Cost contribution details from each process area (per GGE total fuel)

As shown in Figure 15, feedstock cost dominates overall system economics at the $430/ton price assumption employed here, accounting for 70% of total MFSP. This is to be expected, when comparing to terrestrial cellulosic biofuel pathways utilizing feedstock costs closer to $80/ton for materials such as corn stover or wood [6, 28], and highlights the key challenge facing algal biofuels being the high cost of biomass production (see [8] for further quantitative details on estimated cost allocations for upstream steps). Such a result is consistent with the recent algal HTL design report [11] as well as DOE's MYPP projections [9], where the latter case in fact estimated cultivation as contributing 80% to MFSP. In this light, cultivation and harvesting operations will continue to be a critical area of further analysis moving forward, and is anticipated to be a primary focus for NREL algal TEA modeling in 2015 to expand on earlier modeling efforts that considered front-end operations.

Beyond feedstock costs, the pretreatment fractionation and lipid extraction process areas are the largest process area contributors to MFSP cost, contributing 50¢/GGE and 52¢/GGE respectively. In both cases, capital costs are the largest driver at roughly 60% of these values, in turn dominated by the pretreatment reactor system and the extraction column as single line item costs in each respective area. The next cost driver is the lipid purification and upgrading area, with primary cost contributions from hydrotreating and PSA capital expenses, offset slightly by the small naphtha coproduct. Aside from naphtha, the remaining coproducts are all allocated to the AD/CHP system which overall provides for a *net negative* 26¢/GGE contribution to MFSP after accounting for all capital expenses, associated with coproduct credits from CO_2 and N/P nutrient recycles to upstream cultivation, AD digestate cake, and the exported power. Cost contributions attributed to fermentation and distillation are fairly small at 13¢/GGE.

It is noteworthy that even at a feedstock cost of $430/ton (five-fold higher than terrestrial cellulosic feedstocks), combined with relatively low credits from all coproducts under $0.35/GGE, the modeled pathway evaluated here still achieves a fuel MFSP of $4.35/GGE. This is a similar cost as the above-referenced algal HTL design report at $4.49/GGE, enabled in both cases by a highly favorable fuel yield at 138 GGE/ton AFDW in this design and 130 GGE/ton AFDW in the HTL case [11] (excluding contributions from naphtha in both cases). In comparison, cellulosic biofuel pathways have presented design case fuel yield targets of 45 GGE/ton via biochemical conversion of corn stover to hydrocarbons [6], 52 GGE/ton via biochemical conversion of corn stover to ethanol [3], 87 GGE/ton via pyrolysis of woody biomass to hydrocarbons [28], and 63 GGE/ton via thermochemical conversion of woody biomass to mixed alcohols [5]. The ALU fractionation and HTL design cases even surpass the original DOE 2022 target of 117 GGE/dry ton as presented in the May 2013 MYPP report [9], which was based only on lipids. Thus, the fuel yields presented here on the order of 1.5 to 3 times higher than the above-noted cellulosic pathways makes up for much of the higher feedstock cost. While the analysis presented here demonstrates a viable and realistic path to achieve cost targets below $5/GGE, additional improvements are required to ultimately reduce MFSP costs towards $3/GGE targets. A brief discussion of possible paths toward this goal is also included in Section 5.4.

5 Analysis and Discussion

5.1 Cost Sensitivity Analysis

A single-point sensitivity was performed using the variables and limits shown in Table 23. 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 results of the sensitivity analysis are displayed as a tornado plot in Figure 16.

Table 23. Assumptions Varied in the Sensitivity Analysis

Area	Assumption	Min	Baseline	Max
Biomass Delivery	Feedstock cost, \$/ton	300	430	550
	Average feed rate, ton/day	650	1339	2205
	Feed solids content, wt%	15%	20%	25%
	Dryer vs. No dryer [a]	No dryer	Dryer	-
Pretreatment	Pretreatment glucose yield	80%	90%	95%
	Pretreatment HMF yield	0%	0.3%	2%
	Pretreatment acid loading, wt% of liquor feed	0.5%	1%	2%
	Pretreatment residence time, min	2	5	10
	Pretreatment reactor metallurgy	Stainless steel	High alloy	-
Fermentation	Fermentation glucose to ethanol conversion	85%	95%	
	Fermentation contamination loss	0%	3%	6%
Extraction	Extraction lipid yield	85%	95%	98%
	Extraction CAPEX	-50%	0%	+50%
Refining/Upgrading	Lipid purification requirement	Not required	Required	-
	PSA H_2 recovery	85%	91%	99%
	H_2 price, \$/kg	1.2	1.6	2
	Hydrotreating CAPEX	-50%	0%	+100%
Anaerobic Digestion	AD volatile solids loading factor, g/L-day	1	3.3	7
Financial/Other	Naphtha coproduct credit, \$/gal	2.75	3.25	3.75
	AD digestate N credit, \$/tonne bioavailable N	30	500	700
	CO_2 recycle credit, \$/ton	0	34	
	AD effluent N/P recycle credit	Not included	Included	
	Power coproduct credit, ¢/KWh	5	5.7	7
	Total Capital Investment (TCI)	-25%	0%	+25%

[a] Dryer vs. no dryer scenario is still based on base case with excess summer storage; this case assumes storage without the dryer CAPEX and natural gas OPEX costs in order to demonstrate the cost impact associated with summer drying.

As shown in Figure 16, biomass feedstock cost translates to the strongest single MFSP influence, weighing on overall economics considerably more heavily than nearly any other parameter. A feedstock cost of \$550/ton increases MFSP by nearly \$1/GGE relative to the \$4.35/GGE baseline, while \$300/ton decreases MFSP by a similar amount, which would reduce cost to \$3.42/GGE. The baseline feedstock cost of \$430/ton is set consistent with the above-referenced algal HTL design report, in both cases based on DOE's published MYPP target values [9]. Additionally, average feed rate to the conversion facility also exhibits a noticeable impact on MFSP (although substantially less than feedstock cost), particularly at lower feed rates due to loss of economy of scale benefits. For the sensitivity analysis considered here, feed rate was varied from a low of 650 ton/day (roughly 50% of the base case) to a high of 2,205 ton/day (2,000 metric tonne/day) corresponding to standard cellulosic biomass modeling cases [3, 5, 6]. All feed rates here are based on annual average rates, with seasonal variability maintained consistent with the base case between individual seasons (see Section 2.1).

The targeted extraction lipid yield and capital cost also make significant impacts to MFSP. The extraction efficiency is significant primarily because it directly impacts RDB yield which is the primary fuel product from the integrated process, and thus greatly affects the overall fuel energy yield (GGE yield) more substantially than similar fractional differences in pretreatment sugar (glucose) yield. Extraction capital cost was also identified above as one of the most significant single line items of the overall facility capital cost estimates. Total capital investment (TCI) and hydrotreating capital costs also significantly impact MFSP, although these two variables are related, as hydrotreating capital cost is part of TCI. The most significant components of TCI are the extraction column, pretreatment reactor, and hydrotreating capital cost. The algae dryer also accounts for a significant portion of TCI, and is presented separately in the tornado plot when considering removal of the dryer capital cost and natural gas operating cost while maintaining the assumption of excess summer storage for use in the winter (i.e., primarily to demonstrate the overall cost impact associated with the summer drying requirements in isolation).

Out of the various coproduct credit inclusions, the primary coproduct with the largest impact on MFSP is the CO_2 recycle credit. If this credit were eliminated, MFSP would increase by roughly \$0.14/GGE. As discussed above it is justified to include this credit as it was not counted in the original \$430/ton biomass cost basis, but further analysis is required to better understand the most optimal case for recycling the low-concentration CO_2 in the turbine flue gas. Feed solids content exhibits a noticeable, though not drastic, impact on economics over a range from 15-25%, thus suggesting that there may be room for further upstream optimization to reduce dewatering costs from the 20% solids basis assumed here. The variation of combined carbohydrate-to-ethanol yield across pretreatment and fermentation also has a considerable impact on MFSP. The remaining sensitivity variables each account for less than ±\$0.10 change in MFSP, which are small in relation to the feedstock parameters given the dramatic contribution to overall MFSP that feedstock carries (roughly 70% of total MFSP as discussed above). However, taken in combination, these additional parameters are also non-trivial as they ultimately may impact overall process economics.

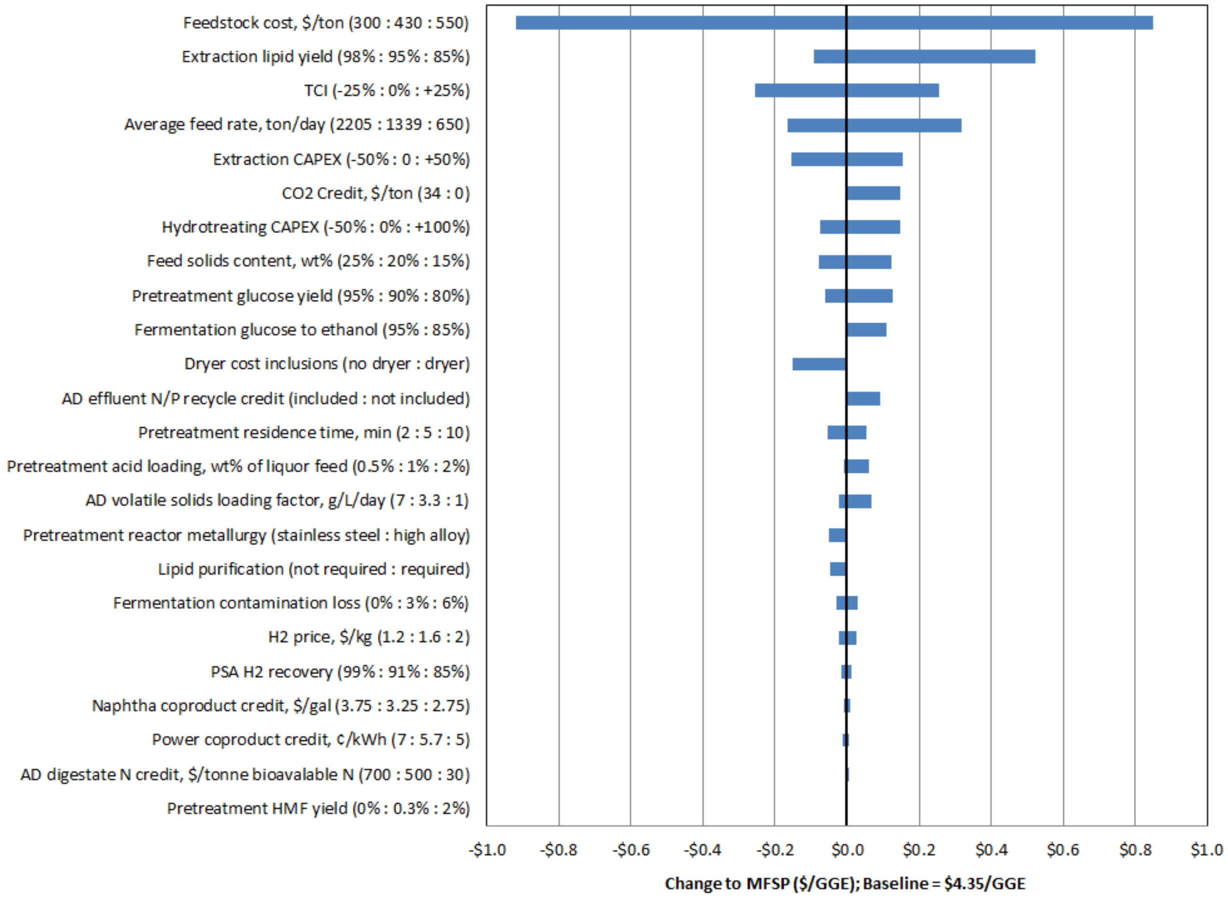


Figure 16. Tornado plot presenting results of the single-point sensitivity analysis on MFSP cost ($4.35/GGE reference case)

5.2 Alternative Composition and Processing Scenarios

While the high-lipid *Scenedesmus* (HLSD) strain with excess summer drying and storage was selected as the base case for the model, here we present the results when instead considering the alternative biomass composition basis (high-carbohydrate *Scenedesmus* – HCSD, Section 2.2) and/or the alternative processing scheme where raw seasonal feed rates from upstream cultivation are fed directly to the conversion facility *without* diverting, drying, and storing excess summer capacity for use in the winter (Section 2.1). Results for both MFSP and total fuel yield for all four scenarios are shown in Figure 17. Similar to the "HLSD + Store" base case, all other scenarios are based on individual Aspen models for all four seasons, setting maximum design capacity and associated capital and fixed operating costs according to maximum seasonal throughput, and taking the season-average results for variable operating costs and product/coproduct yields.

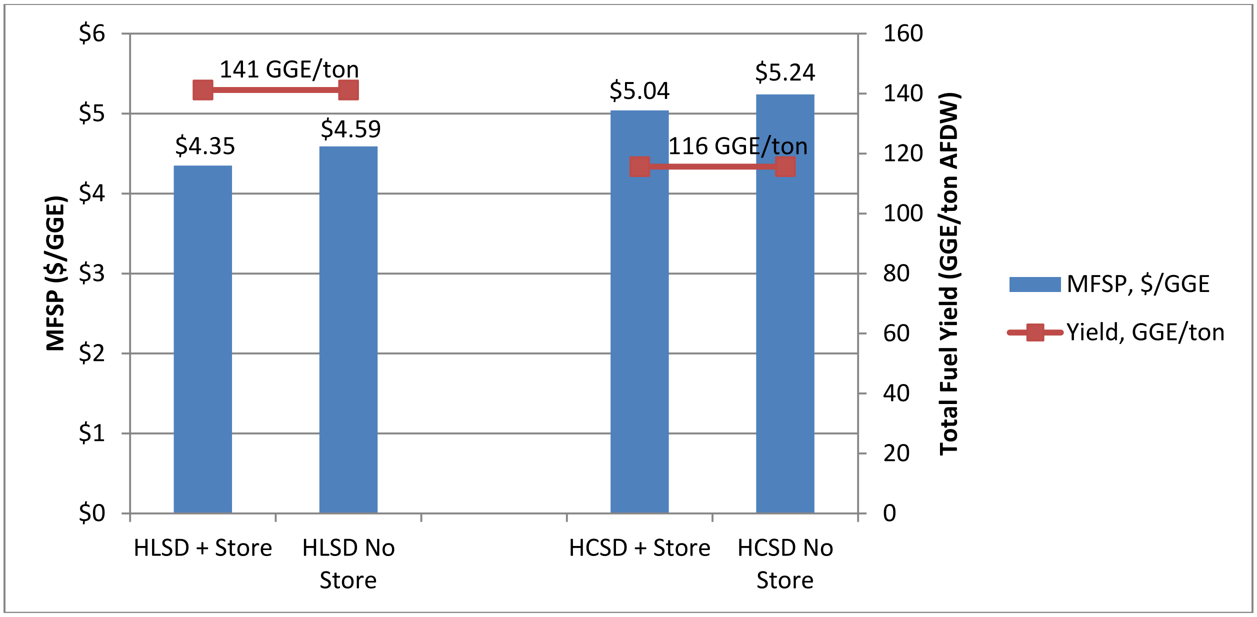


Figure 17. TEA modeling results for alternative composition and seasonal processing scenarios.
(See Sections 2.1-2.2 for definition of scenarios; base case = "*HLSD + Store*".)

As shown in Figure 17, eliminating the summer excess storage approach (and associated dryer capital cost and natural gas operating cost) in favor of sending the algal feed straight to the pretreatment reactor at higher summer versus winter variance results in an overall increase in MFSP on the order of 6% for the HLSD case and 4% for the HCSD case. This penalty is due to less efficient utilization of installed capital costs for three out of four seasons in the "No Store" case, with strong under-utilization of capital in the winter season at 80% throughput turndown versus 20% in the base case including summer storage (assuming that the equipment as installed would still operate effectively at such a high turndown). However, the "no store" alternative scenario also would improve overall system sustainability by elimination of the drying natural gas demand in the summer months (see Section 5.3 for the comparison in sustainability metrics). As noted previously, if sustainability as measured by greenhouse gas profile for the overall integrated process (including upstream cultivation and dewatering) proves to be too detrimental in the "summer storage" approach during subsequent LCA evaluation of the full system, switching to the "no storage" approach may still be a viable option, as it does not substantially increase cost. However, this would require further consideration for equipment operability to ensure the unit operations that are not operating with several units in parallel would be able to accommodate such high turndown ratios. It is also likely that the unfavorable winter effect would be less detrimental in select locations of the U.S., as well as equatorial regions outside the U.S.

Alternatively, modeling for the HCSD biomass composition basis resulted in a more substantial increase in MFSP relative to the HLSD basis, at a 14-16% cost penalty. This is primarily due to the associated overall fuel energy yield (ethanol + RDB, GGE/ton) decreasing by roughly 18%, given the lower energy content of ethanol relative to RDB, which increases to roughly 50% of total production by volume in the HCSD basis versus roughly 35% in the HLSD case. This will be a critical point to consider moving forward in understanding requirements to ultimately achieve economic viability, as the HCSD basis represents an earlier harvest point for the biomass than the HLSD basis, and thereby a higher biomass productivity rate (g/m^2/day) in cultivation.

An ultimately achievable combination of biomass cultivation productivity and lipid content may in fact lie between these two scenarios, resulting in both yields and MFSP costs within the range bounded by these points. Moving forward, when reconsidering a fully integrated model, tradeoffs between cultivation productivity and composition will be important to understand and quantify. This is a key issue that the recently-formed ATP^3 Consortium (of which NREL is a partner) is working to address, namely to provide publicly available data focused on productivity rates, compositional analysis, and cultivation characteristics for algal pond test-bed sites located across the U.S., based on long-term cultivation trials across all seasons using consistent methods [80]. The resulting data will be highly valuable in validating and improving models for algal biofuels, given a continued lack of such data in the public domain at sufficient scale and duration.

5.3 Sustainability Metrics for Base Model

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. Although sustainability metrics are presented here in keeping with other recent design report practices for cellulosic feedstock pathways [6], it is imperative to note that the results presented here are strictly confined to the conversion steps of the process, which is otherwise inseparably integrated from upstream cultivation and dewatering steps. Although it is reasonable to consider sustainability metrics for biorefinery pathways processing terrestrial cellulosic feedstocks, algal processes are more inherently integrated between feedstock production and logistics (cultivation and dewatering) operations and downstream conversion steps, including numerous recycles and heat/power integration between the process boundaries. As such, presentation of sustainability metrics attributed to only the conversion stage for algal processes do not carry as much meaning as for a fully integrated facility and carry a danger of being misinterpreted. A complete LCA of the fully integrated process, incorporating the details from this model as well as pertinent upstream operations, is required to better understand and quantify the complete sustainability profile of this technology pathway.

Direct biorefinery emissions (i.e., CO_2, NO_2 and SO_2), water consumption, and other process-related metrics were derived from the conversion process model described above. SimaPro v.8.0.2 software [81] was used to develop and link units quantifying life-cycle impacts as previously documented by Hsu et al. [82]. GHG basis values for natural gas, purchased hydrogen (derived via natural gas steam methane reforming) and the U.S. average electricity grid mix were applied consistently with the basis from Argonne National Laboratory's GREET model software [83]. Ecoinvent v.2.0 [84] and the U.S. Life Cycle Inventory (LCI) [85] processes were used to fill the data gaps. 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 predicted by the process model, as shown in Table 24. This LCI table and subsequent discussion of sustainability metrics considers both the base case (including summer excess capacity drying and storage) and the alternative case without summer storage, both for the HLSD biomass basis. Additionally, the LCI data presented in Table 24 is translated to an average hourly basis, i.e., dividing the annual rates by the operating factor of 7,920 hours/year; the hourly values shown here (such as GGE/hr and kg/hr) do not reflect actual hourly rates for any season of the year, but rather are the average of the hourly rates across all four seasons.

Table 24. Input and Output Data for Modeled Conversion Facility (HLSD Biomass Basis).
(Note: Hourly rates shown below are based on annual averages over all modeled seasons.)

Summer NG drying and storage scenario:	*Store*	*No Store*
Products		
RDB, GGE/hr	5,840	5,822
Ethanol, GGE/hr	2,034	2,027
Naphtha, gal/hr	130	130
Electricity coproduct, KW	6,937	6,860
Resource Consumption, kg/hr		
Feedstock (AFDW basis)	50,611	50,611
Hexane makeup	910	908
Acid (pretreatment)	2,231	2,224
Ammonia (pretreatment)	721	718
DAP (ethanol fermentation)	103	103
Supplemental natural gas (utility)	859	854
Natural gas for summer drying	2,028	NA
Hydrogen	501	499
Process water demands	78,155	77,901
Phosphoric acid (lipid purification)	39	39
Silica (lipid purification)	21	21
Clay (lipid purification)	41	41
Output Streams, kg/hr		
AD digestate cake (dry basis total flow)	8,325	8,300
AD digestate cake bioavailable N	65	65
AD effluent NH_3	678	676
AD effluent DAP	295	294
Recycle water excluding N/P nutrients	233,528	232,836
Air Emissions, kg/hr		
Water (H_2O)	13,297	13,252
VOC (primarily ethanol)	2	2
Ammonia (NH_3)	40	40
Oxygen (O_2)	47,900	48,116
Nitrogen (N_2)	234,877	235,323
Nitric Oxide (NO)	238	234
Nitrogen dioxide (NO_2)	7	7
Carbon monoxide (CO)	2.E-02	2.E-02
Carbon dioxide (CO_2), biogenic	25,999	28,331
Carbon dioxide (CO_2), fossil	10,707	2,712
Methane (CH_4)	Trace	Trace
Sulfur dioxide (SO_2)	199	198

In addition to the primary fuel products, the process also produces an excess amount of electricity as well as coproduct naphtha, AD effluent nitrogen and phosphorus (modeled as equivalent NH_3 and DAP), AD digestate cake, and flue gas/fermentor vent CO_2 for recycle to algal cultivation. It is assumed that the excess electricity would be sold to the grid translating to coproduct credits for avoided GHG emissions and fossil energy consumption. This exported electricity is treated as an avoided product using the product displacement method [86]. Coproduct displacement (also termed "system boundary expansion") is based on the concept of displacing the existing product with the new product. The excess electricity coproduct displaces an equivalent amount of grid electricity, thus avoiding a significant amount of GHG emissions as well as fossil energy consumption, assuming an average U.S. electricity grid mixture. The GHG and fossil energy consumption credits attributed to the displacement of an average U.S. electricity grid mixture is 0.65 kg $CO_{2\text{-equiv}}$/KWh and 7.46 MJ/KWh, respectively, as defined by GREET [83]. Similar to exported electricity, all other coproducts are also treated as avoided products using the product displacement method. The naphtha displacement coproduct credit is 0.46 kg $CO_{2\text{-equiv}}$/kg for GHG and 52.6 MJ/kg for fossil energy consumption. Additionally, pertinent GHG emission and fossil energy consumption credits are also applied to the AD effluent stream that contains nitrogen and phosphorus as NH_3 and DAP-equivalents (also accounting for nitrogen in DAP). AD digestate nitrogen credits are estimated based on pertinent information in the Ecoinvent database, in which 1 kg of bioavailable N equates to 6.45 kg calcium nitrate at 15.5% N content. The unit process inventory takes into account the use of energy resources needed for the production of calcium nitrate as a byproduct from the manufacture of NPK fertilizers. The corresponding GHG emission factor and fossil energy consumption for 1 kg of bioavailable N in the AD digestate cake is determined to be 4.21 kg $CO_{2\text{-equiv}}$ and 64.5 MJ, respectively.

Details of contributions to GHG emissions and fossil energy consumption at the conversion stage are presented in Table 25, which corresponds to the information in Table 24. Fossil GHG emissions associated with the conversion stage for the cases with and without summer natural gas (NG) drying and storage are determined to be 2.87 kg $CO_{2\text{-eq}}$/GGE and 0.93 kg $CO_{2\text{-eq}}$/GGE, respectively. The difference is largely attributed to the additional NG consumption and fossil CO_2 emissions associated with feedstock drying. Direct CO_2 emissions from the conversion process (shown in Table 24) consist of both fossil and biogenic CO_2 (i.e., CO_2 absorbed from the atmosphere and incorporated as algal biomass during the feedstock cultivation step). With its biomass origin, biogenic CO_2 does not contribute to the increase of GHG in the atmosphere [87] and is not considered in the IPCC global warming methodology [88]. Hence, the contributions to GHG at the conversion stage are solely from facility fossil CO_2 emissions, as well as the fossil GHG profile associated with the underlying LCI processes (e.g., material inputs/outputs to and from the facility to support process operations). Fossil CO_2 emissions shown in Tables 24 and 25 are attributed to combustion of natural gas and volatilized hexane in the turbine flue gas (via co-firing the AD biogas with supplemental natural gas for facility heat requirements) and combustion of natural gas in the summer drying scenario.

It is evident that natural gas, hydrogen, and ammonia consumption, as well as the facility direct fossil CO_2 emissions, contribute the most to the overall GHG emissions estimate. The total GHG coproduct credit attributed to electricity, naphtha, and AD product outputs is determined to be negative 0.93 kg $CO_{2\text{-eq}}$/GGE. Table 25 also shows the fossil energy consumption for the conversion process. Fossil energy consumption associated with the conversion stage for the cases with and without summer NG drying and storage are estimated to be 29.9 MJ/GGE and 15.0 MJ/GGE, respectively. Similar to the GHG emissions, the difference in total fossil energy consumption for both cases is largely attributed to the additional natural gas consumption associated with feedstock drying. The total fossil energy credits attributed to exported power, naphtha, and AD coproducts are comparable for both cases, at roughly negative 15.2 MJ/GGE.

Table 25. Conversion Process GHG Emissions and Fossil Energy Consumption per GGE Total Fuel Product (RDB + Ethanol)

	GHG Emission kg $CO_{2\text{-eq}}$/GGE		Fossil Energy Input MJ/GGE	
Summer NG drying and storage scenario:	***Store***	***No Store***	***Store***	***No Store***
Direct refinery emissions	1.36	0.35	0.00	0.00
Hexane	0.12	0.12	7.05	7.05
Sulfuric acid	0.02	0.02	0.33	0.33
Ammonia	0.20	0.20	3.81	3.81
Diammonium phosphate	0.04	0.04	0.75	0.75
Supplemental natural gas utility	0.39	0.39	6.30	6.29
Natural gas for summer drying	0.93	0.00	14.89	0.00
Hydrogen	0.71	0.71	11.64	11.64
Lipid purification chemicals	0.01	0.01	0.12	0.12
Biorefinery infrastructure	0.02	0.02	0.23	0.23
Electricity credits	-0.58	-0.57	-6.57	-6.52
Co-product naphtha credits	-0.02	-0.02	-2.39	-2.39
Co-product AD credits	-0.34	-0.34	-6.26	-6.26
Total	**2.87**	**0.93**	**29.90**	**15.05**

Table 26 summarizes the key sustainability metrics for the conversion process for the cases with and without summer natural gas drying and storage. On an energy basis, the GHG emissions at the conversion stage for the cases with and without summer drying are 32.2 kg $CO_{2\text{-eq}}$/GJ and 10.4 kg $CO_{2\text{-eq}}$/GJ, respectively. Similarly, the fossil energy consumption for the cases with and without summer drying is 0.33 MJ/MJ and 0.17 MJ/MJ, respectively.

In addition to GHG emissions and fossil energy consumption discussed above, water consumption (i.e., net water use during the biorefinery operation), total fuel efficiency (yield), and carbon-to-fuel efficiency are also reported in Table 26. 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), but it does not include water that is directly returned to surface water or groundwater resources. Aside from the water delivered to the conversion process with the high-moisture feedstock, facility process water consumption demands equate to 1,876 m^3/day as an average across all seasons, allocated to ethanol vent scrubber water (14%), wash water for lipid purification (3%), boiler water for live steam injection to pretreatment (31%), boiler makeup water to make up for blowdown losses (3%), and cooling tower makeup water (49%). Note that this value is prior to accounting for water evaporated during summer drying and consequently required to re-constitute the material for feed in the winter, which adds an additional 687 m^3/day when allocated evenly across a full year, for a total process water consumption of 2,563 m^3/day. Compared to this process water demand, total water rate in the "recycle pool," which is destined for recycle back to the upstream cultivation ponds, is approximately threefold higher, at 5,539 m^3/day averaged across all seasonal models, due to much of the feed water ultimately ending up in this recycle stream. Consequently, the net water balance for the conversion process alone is a net *negative* 2,976 m^3/day or 4.2 gal/GGE for the base case with summer drying; however, this is predicated on the large amount of water entering into the conversion process (at 80% moisture content) for "free" when viewed exclusively in the context of the conversion facility by itself. The true water footprint of the overall process must also take into account upstream cultivation and dewatering operations, which is where the vast majority of water losses occur attributed to evaporation from very large open pond surface areas (although the exact amount is highly dependent on facility location [8]). For a fully integrated facility including consideration of cultivation evaporative and blowdown water losses, the net water consumption is likely to be greater than zero, but would just be reduced by the amount as shown here for the conversion facility "negative" water balance. For the alternative case without summer drying, the conversion stage water balance reduces to a net negative 5.1 gal/GGE, attributed to a process water consumption of 1,876 m^3/day.

The total fuel yield is 141.1 GGE per dry ton and the corresponding carbon-to-fuel efficiency is determined to be 63% (or 64% when also including naphtha), calculated strictly as a ratio of carbon in the liquid fuel products relative to carbon in the feedstock. When also including non-algal carbon inputs attributed to natural gas (summer drying and utility heating), hexane makeup, and hydrogen (sourced from off-site natural gas steam methane reforming at a natural gas carbon-to-hydrogen molar ratio of 0.44:1 [89]), carbon efficiency decreases to 55% (56% including naphtha) for the summer storage base case, or 58% for the alternative case without summer drying and storage. A more detailed accounting of carbon balances is presented below in Figure 18, tracking seasonal-average carbon flows through the conversion process for the modeled *HLSD + Store* base case.

Table 26. Summary of Sustainability Metrics for the Process

Summer NG drying and storage scenario:	***Store***	***No Store***
GHG emissions (kg CO_{2e}/GJ) [a]	32.2	10.4
Conversion stage process water consumption (m^3/day)	2,563	1,876
Conversion stage process water consumption (gal/GGE fuel)	3.6	2.6
Conversion stage net water balance (gal/GGE fuel) [b]	-4.2	-5.1
Total fuel yield (GGE/dry ton)	141.1	141.1
Algal Carbon-to-fuel efficiency (C in fuel/C in biomass) [c]	63%	63%
Total Carbon-to-fuel efficiency (C in fuel/total input C) [d]	55%	58%
Net fossil energy consumption (MJ/MJ) [a]	0.33	0.17

[a] GHG and fossil energy values include recycle coproduct credits as documented above

[b] Table reports both direct process water consumption (inputs to satisfy facility water demands) as well as net water balance after also considering total water recycled to upstream growth ponds (counting biomass feed moisture content as "free" in the boundaries of this analysis)

[c] Algal carbon efficiency defined here as the ratio of carbon in primary fuel products (RDB + ethanol) relative to carbon in biomass feed; excludes naphtha (extra 1%)

[d] Total carbon efficiency defined here as the ratio of carbon in primary fuel products (RDB + ethanol) relative to total input carbon including natural gas, hydrogen (from off-site natural gas), and hexane; excludes naphtha (extra 1%)

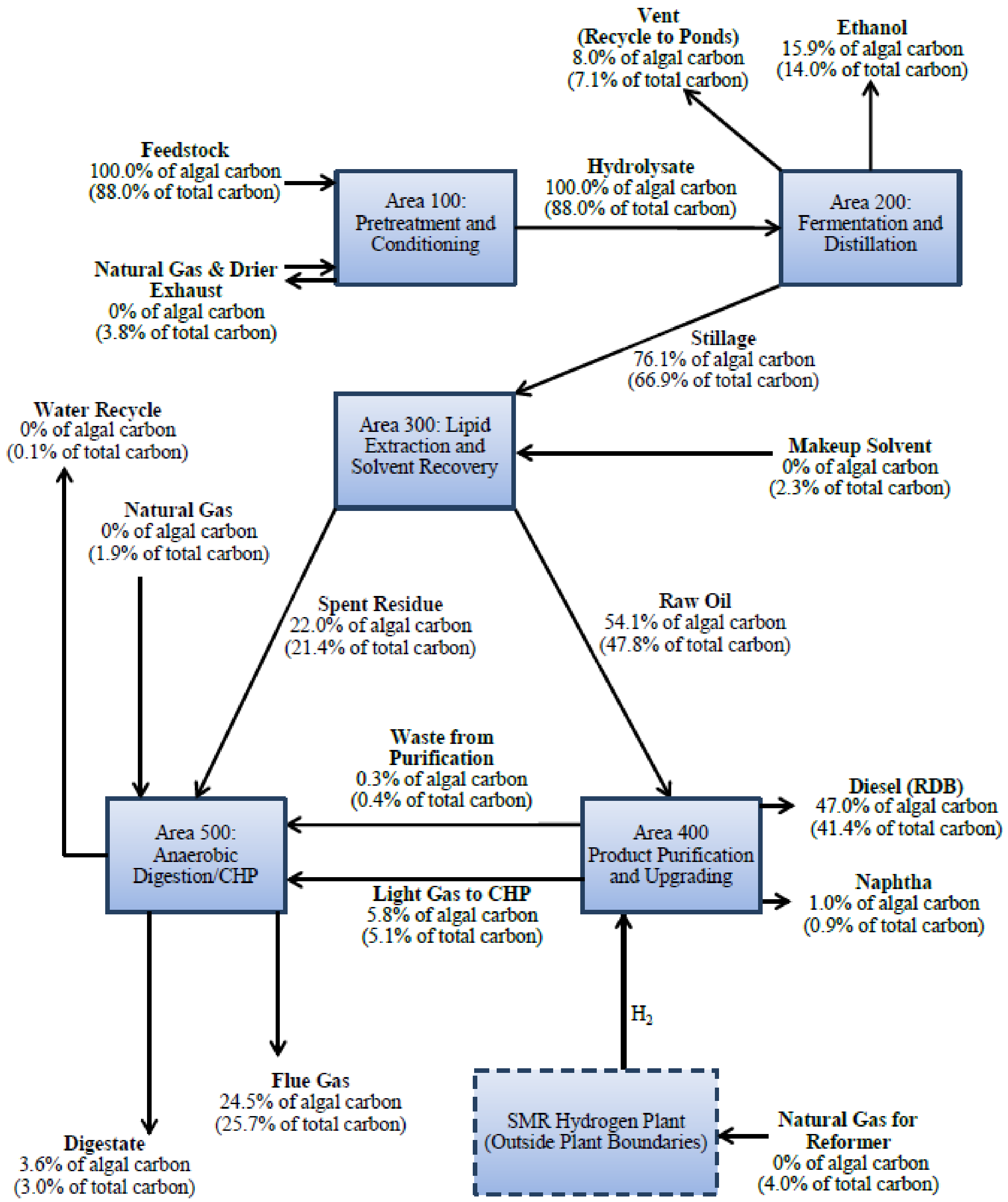


Figure 18. Carbon balance diagram tracking carbon flows across the modeled process (*HLSD* + *Store* base case, overall seasonal average flows).

5.4 Paths Forward to $3/GGE

The analysis presented here demonstrates a strategy for achieving an overall fuel selling price near $4.50/GGE, on par with published targets for algal hydrothermal liquefaction processing [11]. However, additional improvements will be required to further improve economics towards standard DOE-BETO targets closer to $3/GGE. First and foremost, feedstock cost reductions would translate to substantial cost savings, as presented above. For example, as shown in Section 5.1, reducing feedstock cost from $430/ton (AFDW) to $300/ton would reduce MFSP by nearly

$1/GGE. Such an improvement would require high cultivation productivity or a lower-cost cultivation system (per acre of cultivation area) relative to the open raceway pond design as detailed in [8, 55]. Similarly, employing a crop rotation strategy or other mechanism to reduce seasonal variability, particularly increasing winter season productivity, would lead to substantial improvements in economics; to a more limited extent in the conversion back-end (as seen by relatively marginal cost differences between the "summer store" and "no store" scenarios), but considerably more in the cultivation front-end, given the high costs for cultivation and necessity to maintain efficient utilization of installed cultivation capacity. This is also a critical point in improving sustainability, namely GHG emissions, of integrated algal biofuel processes [8, 10].

Beyond further improvements in biomass production, on the conversion side, as discussed previously a key advantage of the process described here is its flexibility in allowing for a "plug and play" processing approach for conversion of the carbohydrate, lipid, and protein fractions individually. Given that the process selectively targets solubilization/extraction and conversion of each of these biomass constituents rather than conversion of the whole biomass feedstock, a large number of options may be pursued to alternative products or coproducts. For example, substantial research efforts are underway to pursue biological or catalytic conversion routes of sugars to hydrocarbon fuel precursors, or to high-value chemicals [6, 90]. This could lead to either increased energy (GGE) fuel yield or high coproduct revenues, respectively. One example of precedent is NREL's 2013 biochemical cellulosic design report, which describes various coproduct pathway options for conversion of lignin components to value-added chemicals, including 1,3 butadiene, 1,4 butanediol, adipic acid, and cyclohexane, with a number of these options providing substantial coproduct revenue to assist in reducing MFSP towards $3/GGE targets [6]. Coproduct options from oil constituents are possible as well, with Cellana proving a noteworthy example whose business model in the near term focuses on garnering significant revenue from isolation of omega-3 fatty acid oils and selling for a substantial fraction of total product sales [91].

Similarly, a number of alternative options exist for utilization of the protein-rich residue. Rather than AD, the protein-enriched material may instead be potentially used for animal or aquaculture feed markets (if sufficiently high-value, e.g., >$350/tonne [8]), bioplastics [92], or a novel route described by Huo et al. [93] and subsequently investigated by SABC, Sandia National Laboratory, and others, in which the protein is hydrolyzed and fermented to butanol and higher alcohols, further increasing total fuel yield potential beyond that considered here. These options were briefly considered in earlier TEA feasibility studies and were not found to offer substantial benefits to justify their use in the present analysis, however they could become more beneficial for higher-protein feedstocks such as the early-harvest *Scenedesmus* biomass scenario presented in Table 1. Such options will continue to be investigated moving forward. In any case, the reliance on coproducts requires careful analysis and understanding of the material's market volume, to avoid saturating a given coproduct market with one or several biofuel facilities, given the very large commodity fuel market into which the biofuel industry will expand. Additionally, similar to the discussion in the algal HTL design report [11], off-site upgrading of the extracted algal oil at a central location or in concert with petroleum refinery blending may also present further opportunities for cost reduction.

6 Concluding Remarks

6.1 Summary

The present report establishes a plausible case for achieving a modeled cost goal under $5/GGE of upgraded renewable diesel blendstock and ethanol, coproduced via biochemical fractionation of algal biomass and selective conversion of carbohydrate and lipid constituents. The pathway is based on a single set of process R&D metrics and associated cost estimates for the base case, realizing that process alternatives exist, both in terms of sequence order of operations as well as different operations altogether to convert carbohydrates, lipids, and even proteins to alternative fuels or products. The pathway discussed here assumes biomass is fed at 20 wt% solids to a dilute acid pretreatment/fractionation step, the resulting whole hydrolysate slurry is sent through a fermentation step to convert solubilized sugars to ethanol, and the resulting whole fermentation stillage is processed in a wet extraction operation with remaining high-protein residue routed to anaerobic digestion for purposes of carbon and nutrient reclamation. The integrated process assumes high solubilization/recovery of both carbohydrates and lipids, and subsequent high conversion of each of these components to ethanol and RDB respectively. Some operations, primarily pretreatment and fermentation, build off of NREL's well-established operational experience and carry a reasonably high level of certainty in ultimately achieving their targeted performance metrics. Other units, such as lipid extraction and upgrading, are in an earlier stage of understanding based on in-house experimental data, but early results have been promising in proving out proof-of-concept and planning for follow-on experimental efforts. More open data flow between NREL and commercial practitioners of these technologies would greatly improve our understanding of the current state of technology.

With the combination of process conversion targets stipulated in this design (primarily 90% sugar solubilization in pretreatment, 95% conversion of sugar-to-ethanol after first assuming fractional sugar losses to fermentation organism propagation and contamination, 95% lipid extraction efficiency, and 80 wt% yield of RDB from purified lipids across hydrotreating), the resulting overall fuel yield was found to be 92.5% of theoretical based on the given algal biomass composition. Thus, while the process scenario allows for achieving MFSP costs below $5/GGE, relatively little room exists for further increasing yields from the carbohydrate or lipid fractions. Thus, additional improvements elsewhere will be required to ultimately reach a $3/GGE goal. These include (1) further optimization on the front-end, primarily achieving high algal productivity rates and/or identifying fundamentally lower-cost cultivation systems, (2) reducing seasonal variability by employing different strains in the winter or other engineering/design measures, (3) considering off-site centralized lipid upgrading options, or (4) pursuing higher-value coproduct options likely based either on the protein or carbohydrate fractions. The current technology pathway discussed here is readily amenable to such alternative options, and possesses a high degree of "plug and play" flexibility in pursuing other technology options for any given biomass component.

The end result of the techno-economic analysis was a predicted minimum fuel selling price (MFSP) of $4.35/GGE (2011-dollars) at a final upgraded product yield of 141.1 GGE/dry ton of biomass (ash-free dry weight basis) in the base case. This reflects a $3.05/GGE contribution from feedstock and a $1.30/GGE contribution from the conversion process. This modeled selling price is strictly representative of n^{th}-plant assumptions regarding biorefinery design, operation, and financing, and is not intended to reflect first-of-a-kind or early-entry commercial facilities.

The total fuel yields presented here are similar to fuel yield targets for HTL conversion of algal biomass, and in both cases are distinctly higher than terrestrial cellulosic pathways. This particular technology pathway achieves these yields via selective conversion of specific components to specific fuel products under relatively low-pressure/temperature operating conditions, however requires mid- to high-lipid content biomass to achieve these yields, which adds a burden to upstream cultivation performance and requires carefully balancing achievable biomass productivity rates against resulting biomass composition and fuel potential.

The base case considered here assumes processing of a late-harvest, high-lipid algal strain with diversion, drying, and storage of excess summer capacity for later use in the winter, to minimize seasonal fluctuations inherent to algal cultivation and thereby fluctuations in equipment processing capacity. However, the analysis also considered, in detail, a process alternative of sending the dewatered material straight to the conversion process without the drying step, substantially increasing the feed rate and downstream throughput variability from a range of 1.2:1 (summer versus winter) to 5.4:1. While maintaining essentially the same overall fuel yield, the alternative processing scenario increased MFSP costs by roughly 5%, due to a higher penalty in under-utilization of installed capital cost relative to a smaller savings in eliminating summer drying equipment and natural gas costs. Alternatively, switching to an earlier-stage harvest biomass composition with high carbohydrate and lower lipid content increased MFSP costs by 14%-16%, attributed to a concomitant GGE yield reduction of 18%. This implies that overall cost targets for an integrated process would be more difficult to achieve at the earlier-harvest biomass composition scenario; however, this must also be considered in reference to the ability to achieve a given algal cultivation productivity rate and thereby algal biomass cost, as well as our decision to valorize sugar as a biofuel rather than as a higher value bio-based chemical. In either case, this analysis sets the stage for continued consideration of these tradeoffs for a fully integrated process including the front-end cultivation and harvesting steps. It may be likely that a realistic optimum exists between the two scenarios considered here, i.e. a mid-to-late stage harvest with a resulting lipid content above 27% but below 41% while maintaining reasonable cultivation productivity rates by year 2022 and beyond.

In addition to providing an economic analysis, the present report also considers sustainability metrics for the baseline processing scenario with summer storage, as well as for the alternative without summer storage, by tracking and reporting on greenhouse gas emissions, fossil energy demand, and direct consumptive water use at the modeled biorefinery facility (conversion stage only). The sustainability analysis for the baseline scenario indicates a fossil GHG profile of 2.87 kg $CO_{2\text{-eq}}$/GGE, fossil energy consumption of 29.9 MJ/GGE, and water demand of *net negative* 4.2 gal/GGE (associated with high moisture content of the incoming feedstock). The alternative scenario without summer excess capacity drying results in GHG emissions, fossil energy consumption, and water demand of 0.93 $CO_{2\text{-eq}}$/GGE, 15.1 MJ/GGE, and net negative 5.1 gal/GGE respectively. Thus, the results indicate conflicting optima between economics (favoring summer storage) and sustainability (favoring no summer storage). However, as these values are only representative of a partial process focused on only the conversion operations in an otherwise inherently integrated process, the sustainability metrics values carry somewhat limited meaning and a full TEA and LCA is required for the overall well-to-wheel integrated process including upstream biomass production and harvesting operations, in order to present a more complete understanding of such tradeoffs. This will be a subject of continued modeling consideration moving forward.

It is worthwhile to reiterate that algal biofuels in general, and the proposed fractionation and conversion process discussed here for processing algal biomass in particular, are in an early stage of development and understanding, relative to other cellulosic technology pathways as well as HTL conversion technologies (whether for algal biofuels or otherwise). Thus, the absolute cost values established here, as well as the timeframe required to achieve these outcomes, inherently carries a somewhat higher degree of uncertainty given the nascent stage of research as presented in the public domain and conducted to date at NREL. This said, the economic potential as well as distinct technology advantages enabled by this approach make a strong case for continued consideration as a means to achieving a viable algal biofuels industry.

6.2 Future Work

Moving forward, to ultimately achieve cost goals and reduce uncertainty in key areas for the algal fractionation conversion pathway in the context of a fully integrated process with upstream cultivation, a number of important bottlenecks, uncertainties, and areas for further development are summarized below. Some of these points are repeated from NREL's recent technical memo for the algal lipid upgrading (ALU) process [94]:

- **Feedstock production and logistics:**
    - **Validate algae growth rates and biomass compositional analysis based on data from large scale demonstrations:** In order to ultimately be able to set realistic, achievable goals related to algal cultivation, as well as biomass quality (i.e., composition), a sound understanding of the initial baseline is first required. While NREL and other partners have done substantial analysis on this subject [8, 10], these two key parameters still largely remain limited to modeled or semi-empirical data. Among a number of entities working to address this issue, the recently-formed ATP3 and RAFT programs are focused on this goal as a critical priority, working to provide publicly available data for cultivation test-bed sites located in regions across the US. Data made available from these efforts will be leveraged in NREL's models to better quantify today's "state of technology" costs.
    - **Reduce cost and increase performance for cultivation and dewatering:** To ultimately realize economic viability for algal-derived biofuels regardless of the technology, substantial improvements in both performance (e.g. cultivation productivity, harvesting efficiency) and cost (e.g. alternative or lower-cost designs, removing pond liners) will be required. This will depend to a large extent on both locality and strain specifics, with high cost sensitivities to seasonal variability of the selected facility location. Additionally, continued attention must be afforded to protection of crop yield against culture crashes (losses to grazers and pathogens) through improving strain robustness and systems design [95].
- **Conversion operations:**
    - **Confirm efficacy of integrated process:** While promising experimental data have proven the efficacy of the process concepts when conducting the key conversion steps on separated solids and liquids from pretreatment [13], NREL has not yet considered an experimental approach consistent with the extrapolated basis used for modeling purposes, e.g., fermentation of the whole pretreatment

hydrolysate followed by lipid extraction of the whole fermentation stillage. This approach is anticipated to be equally viable, although may present additional challenges that must be overcome by continued research to demonstrate the high yield targets stipulated in the model for an integrated process. Further process intensification to target increased fermentation productivity and decreased capital costs may also contribute to economic improvements.

  - **Further characterize products/intermediates:** NREL has a reputable history of algal biomass characterization capabilities and the biomass component characterizations employed here are sound; however, less work has been done to date on characterizing the conversion products. This is particularly of interest for the extracted algal oil product, as it impacts subsequent requirements and costs both for lipid cleanup and upgrading.
  - **Demonstrate upgrading of algal oil:** The assumptions employed in the model described here are based on operating and yield assumptions from literature, for general fatty acid triglyceride materials. The model will incorporate experimental data as it becomes available from new NREL efforts focused on algal oil upgrading, which will better define specific requirements for lipid purification, as well as conversion both to paraffinic blendstocks produced from deoxygenation and to finished fuels produced from hydroisomerization.

- **Coproduct utilization:**
  - **Confirm AD performance:** The assumptions currently employed here for the anaerobic digestion (AD) operation, specifically metrics of biogas yield and nutrient fixation in the recycle effluent, are based on a compilation of literature studies, which have largely been validated by NREL experimental work when considering AD of lipid-extracted biomass. However, for biomass that is also largely devoid of carbohydrates, as in the process modeled here, additional operational challenges may be presented for the AD unit which are not yet well-understood. Further analysis is required to confirm the assumptions utilized here.
  - **Evaluate alternative coproduct opportunities:** Many coproduct opportunities have yet to be fully explored for economic and LCA trade-offs and should further be considered going forward. These include options for both alternative carbohydrate utilization scenarios to hydrocarbon fuels or chemicals (balanced by additional processing costs), as well as protein utilization options such as high-value animal or fish feed materials, bioplastics, or protein hydrolysis and fermentation to alcohols. Other high-value coproduct options may also be plausible; however, any coproduct strategies must be carefully considered relative to market volumes given the large-scale commodity fuel market that this process is intended to contribute to. Additionally, direct use of wastewater as a water and/or nutrient source may also provide additional economic benefits by way of wastewater treatment cost offsets.

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Appendix A. Individual Equipment Cost Summary

The following table shows abbreviated specifications, purchased cost and installed cost for each piece of equipment in this process design, with all capital costs set based on the maximum throughput design case, i.e. summer season flows. Although each piece of equipment has its own line, many were quoted as part of a package. 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						Scaled Installed Costs											
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
Area 100: Pretreatment and Conditioning																	
Dryer	Scaled from Jones et al. 2013 HTL report			12	$905,000	2013	$10,860,000	strm.algdry	30237	kg/hr	0.80	1.6	32872	1.09	$11,610,729	$11,632,576	$18,612,122
PT Flash/Solvent Dist Preheat HX	Costed in IPE, S-T heat exchanger			1	$48,019	2007	$48,019	heat.77	8.6	Gcal/h	0.6	3.0	10.6	1.23	$54,409	$60,653	$181,960
Pretreatment Feed-Effluent HX	Costed in IPE, S-T heat exchanger			1	$432,519	2007	$432,519	heat.80	12.3	Gcal/h	0.60	1.6	14.9	1.21	$484,484	$540,088	$864,140
Pretreatment Reactor	Per NREL 2013 design report			5	$16,400,000	2013	$31,200,000	DRYFD2	63166	kg/hr	0.60	1.5	56573	0.90	$29,203,298	$29,258,248	$43,887,372
Pretreament: Balance of Plant	Scaled from NREL 2013 design report for minor supporting equiptment				-		-	4.6% BOP factor per NREL 2013 BC design report							$1,343,352	$1,345,879	$2,018,819
														Area 100 Totals	$42,696,272	$42,837,445	**$65,564,413**

Mechanical Equipment List						Scaled Installed Costs											
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
Area 200: Fermentation and Distillation																	
Ethanol Fermentor	All CAPEX items per NREL 2011 ethanol design report				$10,128,000	2009	$10,128,000	SSFVES	12	vessels	1.00	1.5	4	0.33	$3,376,000	$3,788,701	$5,683,052
Ethanol Fermentor Agitator					$630,000	2009	$630,000	SSFVES	12	vessels	1.00	1.5	4	0.33	$210,000	$235,672	$353,507
Ethanol Fermentor Cooler					$86,928	2009	$86,928	SSFVES	12	vessels	1.00	1.5	4	0.33	$28,976	$32,518	$48,777
Ethanol Fermentor Recirc Pump					$47,200	2009	$47,200	SSFVES	12	vessels	1.00	1.5	4	0.33	$15,733	$17,657	$26,485
Ethanol Seed Train Agitator					$26,000	2009	$26,000	strm.515	21672	kg/hr	0.60	1.5	9744	0.45	$16,095	$18,062	$27,093
Ethanol Seed Train Agitator					$43,000	2009	$43,000	strm.515	21672	kg/hr	0.60	1.5	9744	0.45	$26,618	$29,872	$44,808
1st Ethanol Seed Fermentor					$75,400	2009	$75,400	strm.515	21672	kg/hr	0.60	1.8	9744	0.45	$46,674	$52,380	$94,284
2nd Ethanol Seed Fermentor					$116,600	2009	$116,600	strm.515	21672	kg/hr	0.60	1.8	9744	0.45	$72,178	$81,001	$145,802
3rd Ethanol Seed Fermentor					$157,600	2009	$157,600	strm.515	21672	kg/hr	0.60	1.8	9744	0.45	$97,558	$109,484	$197,071
4th Ethanol Seed Fermentor					$352,000	2009	$352,000	strm.515	21672	kg/hr	0.60	2.0	9744	0.45	$217,895	$244,532	$489,064
5th Ethanol Seed Fermentor					$1,180,000	2009	$1,180,000	strm.515	21672	kg/hr	0.60	2.0	9744	0.45	$730,444	$819,738	$1,639,476
Fermentation: Balance of Plant	Scaled from NREL 2011 ethanol design report for supporting equipment							8.6% BOP factor per NREL 2013 BC design report							$416,083	$466,947	$752,450
Vent Scrubber					$215,000	2009	$215,000	strm.18	22608	kg/hr	0.60	2.4	9814	0.43	$130,311	$146,240	$350,977
Ethanol Distillation Equipment	Includes beer and rectification columns				$3,407,000	2009	$3,407,000	strm.511	30379	kg/hr	0.60	2.4	13031	0.43	$2,050,333	$2,300,977	$5,522,345
Reboilers					INCLUDED												
Beer Condenser					INCLUDED												
Rectification Condenser					$487,000	2010	$487,000	heat.QAC-RECT	23	MM kcal/hr	0.60	2.8	13	0.56	$342,854	$364,578	$1,020,820
Beer Column HX Economizer					INCLUDED												
Ethanol Molecular Sieve					$2,601,000	2009	$2,601,000	strm.515	22687	kg/hr	0.60	1.8	9744	0.43	$1,566,458	$1,757,950	$3,164,311
														Area 200 Totals	$9,344,210	$10,466,310	**$19,560,322**

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
Area 300: Lipid Extraction and Solvent Recovery																	
Extraction Column	6' D AGITATED ZONE, 9' D EXPANDED ENDS, 60' OAH	160 HP	Shell: 304L Int: 304SS	16	$1,980,000	2013	$31,680,000	cmix.TAG.28	5894	kg/hr	1.00	2.0	23398	price per unit	$31,680,000	$31,739,610	$63,479,220
Solvent Recovery Column	10'D x 20'H PACKED TOWER		304SS	3	$714,000	2009	$714,000	cmix.TAG.42	5599	kg/hr	0.85	2.4	22228	3.97	$2,304,996	$2,586,771	$6,208,251
Solvent Recovery Reboiler	Harris Group Cost Estimate		304SS	3	$150,000	2009	$150,000	cmix.TAG.42	5599	kg/hr	0.85	3.3	22228	3.97	$484,243	$543,439	$1,793,350
														Area 300 Totals	$34,469,239	$34,869,821	$71,480,821

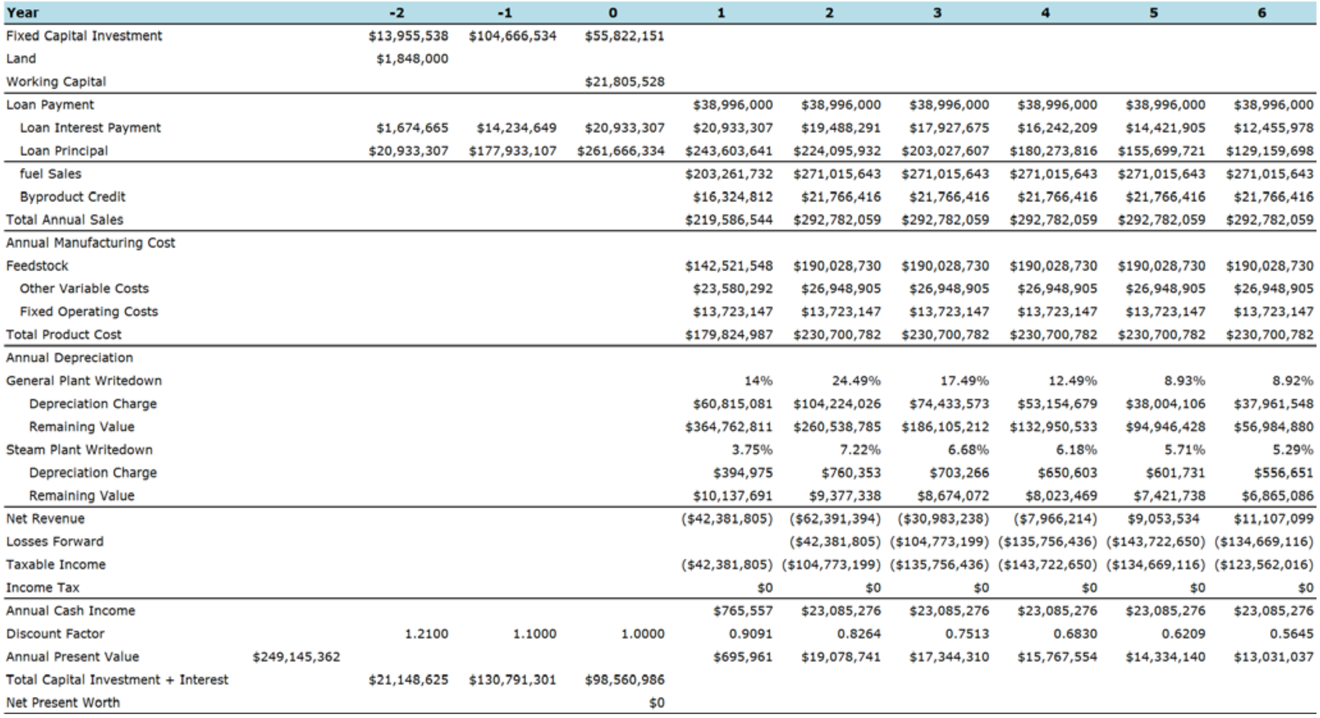
Mechanical Equipment List						Scaled Installed Costs											
EQUIPMENT TITLE	**DESCRIPTION**	**HP**	**MATERIAL**	**NUM REQD**	**$**	**Year of Quote**	**Purch Cost in Base Yr**	**Scaling Variable**	**Scaling Val**	**Units**	**Scaling Exp**	**Inst Factor**	**New Val**	**Size Ratio**	**Scaled Purch Cost**	**Purch Cost in Proj year**	**Inst Cost in Proj year**
Area 400: Product Purification and Upgrading																	
Hydrotreating Unit	ISBL package unit; 2013 Vendor quote via HGI				$30,000,000	2013	$30,000,000	strm.REFINE.OILFDSTD den.REFINE.OILFDSTD	109	gpm	0.50	1.0	107	0.98	$29,719,549	$29,775,470	$29,775,470
PSA Unit	ISBL package unit; 2013 Vendor quote via HGI				$12,000,000	2013	$12,000,000	strm.REFINE.OILFDSTD den.REFINE.OILFDSTD	109	gpm	0.60	1.0	107	0.98	$11,865,510	$11,887,836	$11,887,836
Bleaching/Degumming Unit	ISBL package unit; 2013 Vendor quote via HGI				$6,400,000	2013	$6,400,000	strm.tagstd den.tagstd	109	gpm	0.60	1.0	115	1.06	$6,621,025	$6,633,483	$6,633,483
														Area 400 Totals	$48,206,084	$48,296,790	**$48,296,790**

Mechanical Equipment List						Scaled Installed Costs											
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**
Area 500: Anaerobic Digestion/CHP																	
Anaerobic Digester	Scaled from NREL 2013 design report - includes AD supporting equipment e.g. pumps, blowers, mixers, centrifugation, and emergency flare			2	$6,450,000	2012	$6,450,000	Size Calculated Externally	27000000	gallons /unit	0.60	1.1	16050915	0.59	$9,914,267	$9,932,922	$11,025,544
Gen-Set	Per Dean et al. 2010, GTW 2006				$445	2006	$445	Work.WGEN	-1	kW	1.00	1.0	-20189	20189	$8,984,326	$10,532,666	$10,532,666
														Area 500 Totals	$18,898,593	$20,465,588	**$21,558,209**

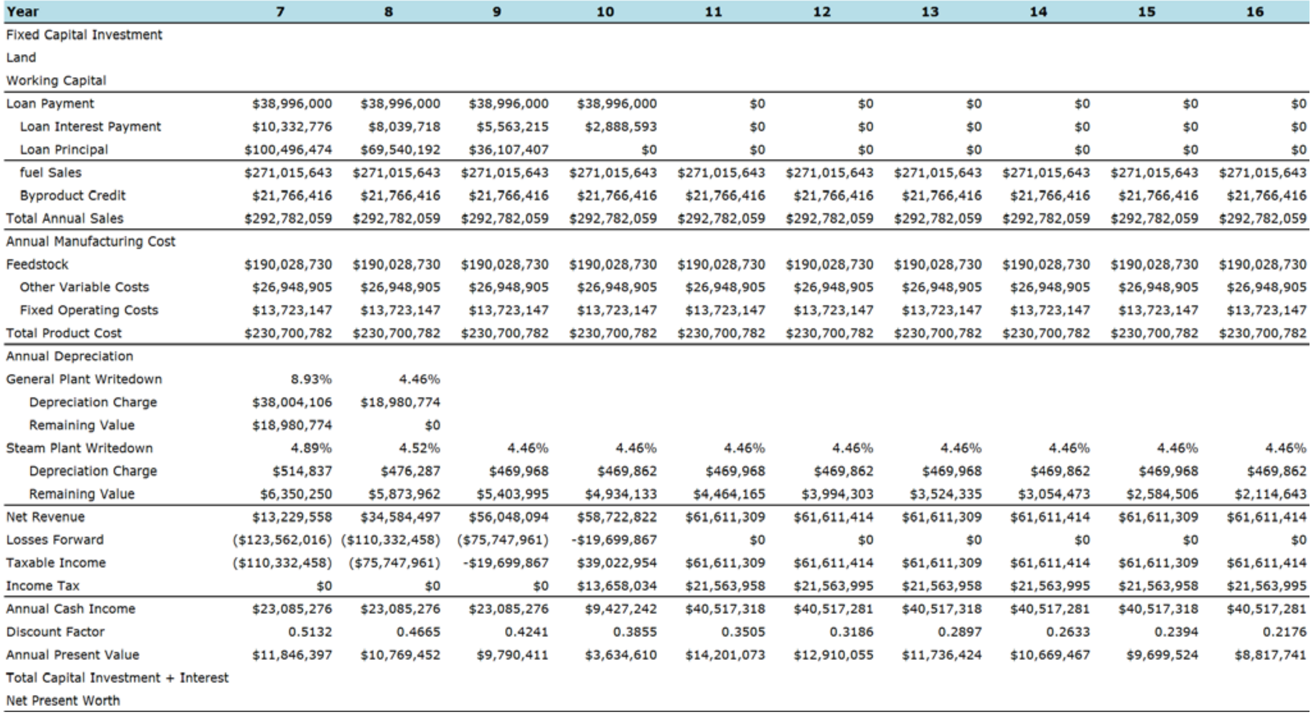
Mechanical Equipment List						Scaled Installed Costs											
EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
Area 600: Storage																	
Diesel Product Storage Tank	750,000 gal., 7+ day storage, Floating roof		A285C	1	$670,000	2009	$670,000	strm.diesel	11341	kg/hr	0.70	1.7	17661	1.56	$913,556	$1,025,234	$1,742,898
Ethanol Product Storage Tank	750,000 gal., 7+ day storage, Floating roof		A285C	1	$670,000	2009	$670,000	strm.515	11341	kg/hr	0.70	1.7	9744	0.86	$602,492	$676,143	$1,149,444
Naphtha Product Storage Tank	25,000 gal, 7+ day storage, Floating roof		A285C	1	$87,057	2009	$87,057	strm.naphtha	378	kg/hr	0.70	1.7	392	1.04	$89,271	$100,184	$170,313
Firewater Storage Tank	600,000 gal - 4 hrs @ 2500 gpm		Glass lined	1	$501,000	2009	$501,000	DRYFD2	83333	kg/hr	0.70	1.7	56573	0.68	$382,027	$428,728	$728,838
Tankage BOP	Scaled from NREL 2013 Design Report for minor chemical tanks							20% BOP factor per NREL 2013 BC design report							$397,469	$446,058	$758,299
Dried algae storage (3 months)	Aspen IPE; 3.7 MM gal each, silo-style tank			8	$13,500,000	2009	$0	strm.algdry	32750	kg/hr	0.60	1.8	32872	1.00	$0	$0	$0
														Area 600 Totals	$2,384,815	$2,676,348	**$4,549,792**

Mechanical Equipment List						Scaled Installed Costs											
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
Area 700: Utilities																	
Cooling Tower System		750 hp	FIBERGLASS	1	$1,375,000	2010	$1,375,000	strm.utility.4	10037820	kg/hr	0.60	1.5	2658568	0.26	$619,595	$658,854	$988,281
Plant Air Compressor		150 hp		1	$28,000	2010	$28,000	DRYFD2	83333	kg/hr	0.60	1.6	56573	0.68	$22,194	$23,600	$37,760
Chilled Water Package		3400 hp		1	$1,275,750	2010	$1,275,750	heat.utility.24	14	Gcal/hr	0.60	1.6	2	0.14	$398,090	$423,314	$677,303
CIP System			SS304/SS316	1	$694,000	2009	$694,000	DRYFD2	83333	kg/hr	0.60	1.8	56573	0.68	$550,094	$617,340	$1,111,212
Cooling Water Pump		500.0	CS	3	$283,671	2010	$283,671	strm.utility.4	10982556	kg/hr	0.80	3.1	2658568	0.24	$91,195	$96,973	$300,617
Makeup Water Pump		20.0	CS	1	$6,864	2010	$6,864	strm.utility.procwat	155564	kg/hr	0.80	3.1	85179	0.55	$4,240	$4,508	$13,975
Process Water Circulating Pump		75.0	CS	1	$15,292	2010	$15,292	strm.utility.procwat	518924	kg/hr	0.80	3.1	85179	0.16	$3,603	$3,831	$11,877
Instrument Air Dryer			CS	1	$15,000	2009	$15,000	DRYFD2	83333	kg/hr	0.60	1.8	56573	0.68	$11,890	$13,343	$24,018
Plant Air Receiver			CS	1	$16,000	2009	$16,000	DRYFD2	83333	kg/hr	0.60	3.1	56573	0.68	$12,682	$14,233	$44,121
Process Water Tank No. 1			CS	1	$250,000	2009	$250,000	strm.utility.procwat	451555	kg/hr	0.70	1.7	85179	0.19	$77,781	$87,290	$148,392
Steam Boiler HP Steam HRSG	Costed in IPE				$1,394,900	2009	$1,394,900	strm.steam.HP2	3975024	lb/day	0.60	1.0	752358	0.19	$513,800	$576,610	$576,610
Steam Boiler HP Steam HRSG	Costed in IPE				$1,394,900	2009	$1,394,900	strm.steam.3	3975024	lb/day	0.60	1.0	1367601.4	0.34	$735,385	$825,283	$825,283
Steam Boiler LP Steam HRSG	Costed in IPE				$976,430	2009	$976,430	strm.steam.20	3975024	lb/day	0.60	1.0	2732373	0.69	$779,761	$875,083	$875,083
HP Steam Econ1 HX	Scaled from NREL 2011; plate and frame HX			1	$23,900	2009	$23,900	heat.47	5.4	Gkal/h	0.60	2.5	1.9	0.36	$12,844	$14,414	$35,603
HP Steam Econ2 HX	Scaled from NREL 2011; plate and frame HX			1	$23,900	2009	$23,900	heat.4	5.4	Gkal/h	0.60	2.5	1.5	0.28	$11,077	$12,431	$30,705
														Area 700 Totals	$3,844,230	$4,247,107	**$5,700,840**

Appendix B. Discounted Cash Flow Rate of Return Worksheet



Year		-2	-1	0	1	2	3	4	5	6
Fixed Capital Investment		$13,955,538	$104,666,534	$55,822,151						
Land		$1,848,000								
Working Capital				$21,805,528						
Loan Payment					$38,996,000	$38,996,000	$38,996,000	$38,996,000	$38,996,000	$38,996,000
Loan Interest Payment		$1,674,665	$14,234,649	$20,933,307	$20,933,307	$19,488,291	$17,927,675	$16,242,209	$14,421,905	$12,455,978
Loan Principal		$20,933,307	$177,933,107	$261,666,334	$243,603,641	$224,095,932	$203,027,607	$180,273,816	$155,699,721	$129,159,698
fuel Sales					$203,261,732	$271,015,643	$271,015,643	$271,015,643	$271,015,643	$271,015,643
Byproduct Credit					$16,324,812	$21,766,416	$21,766,416	$21,766,416	$21,766,416	$21,766,416
Total Annual Sales					$219,586,544	$292,782,059	$292,782,059	$292,782,059	$292,782,059	$292,782,059
Annual Manufacturing Cost										
Feedstock					$142,521,548	$190,028,730	$190,028,730	$190,028,730	$190,028,730	$190,028,730
Other Variable Costs					$23,580,292	$26,948,905	$26,948,905	$26,948,905	$26,948,905	$26,948,905
Fixed Operating Costs					$13,723,147	$13,723,147	$13,723,147	$13,723,147	$13,723,147	$13,723,147
Total Product Cost					$179,824,987	$230,700,782	$230,700,782	$230,700,782	$230,700,782	$230,700,782
Annual Depreciation										
General Plant Writedown					14%	24.49%	17.49%	12.49%	8.93%	8.92%
Depreciation Charge					$60,815,081	$104,224,026	$74,433,573	$53,154,679	$38,004,106	$37,961,548
Remaining Value					$364,762,811	$260,538,785	$186,105,212	$132,950,533	$94,946,428	$56,984,880
Steam Plant Writedown					3.75%	7.22%	6.68%	6.18%	5.71%	5.29%
Depreciation Charge					$394,975	$760,353	$703,266	$650,603	$601,731	$556,651
Remaining Value					$10,137,691	$9,377,338	$8,674,072	$8,023,469	$7,421,738	$6,865,086
Net Revenue					($42,381,805)	($62,391,394)	($30,983,238)	($7,966,214)	$9,053,534	$11,107,099
Losses Forward						($42,381,805)	($104,773,199)	($135,756,436)	($143,722,650)	($134,669,116)
Taxable Income					($42,381,805)	($104,773,199)	($135,756,436)	($143,722,650)	($134,669,116)	($123,562,016)
Income Tax					$0	$0	$0	$0	$0	$0
Annual Cash Income					$765,557	$23,085,276	$23,085,276	$23,085,276	$23,085,276	$23,085,276
Discount Factor		1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830	0.6209	0.5645
Annual Present Value	$249,145,362				$695,961	$19,078,741	$17,344,310	$15,767,554	$14,334,140	$13,031,037
Total Capital Investment + Interest		$21,148,625	$130,791,301	$98,560,986						
Net Present Worth				$0						



Year	7	8	9	10	11	12	13	14	15	16
Fixed Capital Investment										
Land										
Working Capital										
Loan Payment	$38,996,000	$38,996,000	$38,996,000	$38,996,000	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$10,332,776	$8,039,718	$5,563,215	$2,888,593	$0	$0	$0	$0	$0	$0
Loan Principal	$100,496,474	$69,540,192	$36,107,407	$0	$0	$0	$0	$0	$0	$0
fuel Sales	$271,015,643	$271,015,643	$271,015,643	$271,015,643	$271,015,643	$271,015,643	$271,015,643	$271,015,643	$271,015,643	$271,015,643
Byproduct Credit	$21,766,416	$21,766,416	$21,766,416	$21,766,416	$21,766,416	$21,766,416	$21,766,416	$21,766,416	$21,766,416	$21,766,416
Total Annual Sales	$292,782,059	$292,782,059	$292,782,059	$292,782,059	$292,782,059	$292,782,059	$292,782,059	$292,782,059	$292,782,059	$292,782,059
Annual Manufacturing Cost										
Feedstock	$190,028,730	$190,028,730	$190,028,730	$190,028,730	$190,028,730	$190,028,730	$190,028,730	$190,028,730	$190,028,730	$190,028,730
Other Variable Costs	$26,948,905	$26,948,905	$26,948,905	$26,948,905	$26,948,905	$26,948,905	$26,948,905	$26,948,905	$26,948,905	$26,948,905
Fixed Operating Costs	$13,723,147	$13,723,147	$13,723,147	$13,723,147	$13,723,147	$13,723,147	$13,723,147	$13,723,147	$13,723,147	$13,723,147
Total Product Cost	$230,700,782	$230,700,782	$230,700,782	$230,700,782	$230,700,782	$230,700,782	$230,700,782	$230,700,782	$230,700,782	$230,700,782
Annual Depreciation										
General Plant Writedown	8.93%	4.46%								
Depreciation Charge	$38,004,106	$18,980,774								
Remaining Value	$18,980,774	$0								
Steam Plant Writedown	4.89%	4.52%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%
Depreciation Charge	$514,837	$476,287	$469,968	$469,862	$469,968	$469,862	$469,968	$469,862	$469,968	$469,862
Remaining Value	$6,350,250	$5,873,962	$5,403,995	$4,934,133	$4,464,165	$3,994,303	$3,524,335	$3,054,473	$2,584,506	$2,114,643
Net Revenue	$13,229,558	$34,584,497	$56,048,094	$58,722,822	$61,611,309	$61,611,414	$61,611,309	$61,611,414	$61,611,309	$61,611,414
Losses Forward	($123,562,016)	($110,332,458)	($75,747,961)	-$19,699,867	$0	$0	$0	$0	$0	$0
Taxable Income	($110,332,458)	($75,747,961)	-$19,699,867	$39,022,954	$61,611,309	$61,611,414	$61,611,309	$61,611,414	$61,611,309	$61,611,414
Income Tax	$0	$0	$0	$13,658,034	$21,563,958	$21,563,995	$21,563,958	$21,563,995	$21,563,958	$21,563,995
Annual Cash Income	$23,085,276	$23,085,276	$23,085,276	$9,427,242	$40,517,318	$40,517,281	$40,517,318	$40,517,281	$40,517,318	$40,517,281
Discount Factor	0.5132	0.4665	0.4241	0.3855	0.3505	0.3186	0.2897	0.2633	0.2394	0.2176
Annual Present Value	$11,846,397	$10,769,452	$9,790,411	$3,634,610	$14,201,073	$12,910,055	$11,736,424	$10,669,467	$9,699,524	$8,817,741
Total Capital Investment + Interest										
Net Present Worth										

Year	17	18	19	20	21	22	23	24
Fixed Capital Investment								
Land								
Working Capital								
Loan Payment	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$0	$0	$0	$0	$0	$0	$0	$0
fuel Sales	$271,015,643	$271,015,643	$271,015,643	$271,015,643	$271,015,643	$271,015,643	$271,015,643	$271,015,643
Byproduct Credit	$21,766,416	$21,766,416	$21,766,416	$21,766,416	$21,766,416	$21,766,416	$21,766,416	$21,766,416
Total Annual Sales	$292,782,059	$292,782,059	$292,782,059	$292,782,059	$292,782,059	$292,782,059	$292,782,059	$292,782,059
Annual Manufacturing Cost								
Feedstock	$190,028,730	$190,028,730	$190,028,730	$190,028,730	$190,028,730	$190,028,730	$190,028,730	$190,028,730
Other Variable Costs	$26,948,905	$26,948,905	$26,948,905	$26,948,905	$26,948,905	$26,948,905	$26,948,905	$26,948,905
Fixed Operating Costs	$13,723,147	$13,723,147	$13,723,147	$13,723,147	$13,723,147	$13,723,147	$13,723,147	$13,723,147
Total Product Cost	$230,700,782	$230,700,782	$230,700,782	$230,700,782	$230,700,782	$230,700,782	$230,700,782	$230,700,782
Annual Depreciation								
General Plant Writedown								
Depreciation Charge								
Remaining Value								
Steam Plant Writedown	4.46%	4.46%	4.46%	4.46%	2.23%			
Depreciation Charge	$469,968	$469,862	$469,968	$469,862	$234,984			
Remaining Value	$1,644,676	$1,174,814	$704,846	$234,984	$0			
Net Revenue	$61,611,309	$61,611,414	$61,611,309	$61,611,414	$61,846,293	$62,081,276	$62,081,276	$62,081,276
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$61,611,309	$61,611,414	$61,611,309	$61,611,414	$61,846,293	$62,081,276	$62,081,276	$62,081,276
Income Tax	$21,563,958	$21,563,995	$21,563,958	$21,563,995	$21,646,202	$21,728,447	$21,728,447	$21,728,447
Annual Cash Income	$40,517,318	$40,517,281	$40,517,318	$40,517,281	$40,435,074	$40,352,830	$40,352,830	$40,352,830
Discount Factor	0.1978	0.1799	0.1635	0.1486	0.1351	0.1228	0.1117	0.1015
Annual Present Value	$8,016,135	$7,287,389	$6,624,905	$6,022,636	$5,464,015	$4,957,183	$4,506,530	$4,096,845
Total Capital Investment + Interest								
Net Present Worth								

Year	25	26	27	28	29	30
Fixed Capital Investment						
Land						($1,848,000)
Working Capital						($21,805,528)
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	$271,015,643	$271,015,643	$271,015,643	$271,015,643	$271,015,643	$271,015,643
Byproduct Credit	$21,766,416	$21,766,416	$21,766,416	$21,766,416	$21,766,416	$21,766,416
Total Annual Sales	$292,782,059	$292,782,059	$292,782,059	$292,782,059	$292,782,059	$292,782,059
Annual Manufacturing Cost						
Feedstock	$190,028,730	$190,028,730	$190,028,730	$190,028,730	$190,028,730	$190,028,730
Other Variable Costs	$26,948,905	$26,948,905	$26,948,905	$26,948,905	$26,948,905	$26,948,905
Fixed Operating Costs	$13,723,147	$13,723,147	$13,723,147	$13,723,147	$13,723,147	$13,723,147
Total Product Cost	$230,700,782	$230,700,782	$230,700,782	$230,700,782	$230,700,782	$230,700,782
Annual Depreciation						
General Plant Writedown						
Depreciation Charge						
Remaining Value						
Steam Plant Writedown						
Depreciation Charge						
Remaining Value						
Net Revenue	$62,081,276	$62,081,276	$62,081,276	$62,081,276	$62,081,276	$62,081,276
Losses Forward	$0	$0	$0	$0	$0	$0
Taxable Income	$62,081,276	$62,081,276	$62,081,276	$62,081,276	$62,081,276	$62,081,276
Income Tax	$21,728,447	$21,728,447	$21,728,447	$21,728,447	$21,728,447	$21,728,447
Annual Cash Income	$40,352,830	$40,352,830	$40,352,830	$40,352,830	$40,352,830	$40,352,830
Discount Factor	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
Annual Present Value	$3,724,405	$3,385,822	$3,078,020	$2,798,200	$2,543,819	$2,312,562
Total Capital Investment + Interest						($1,355,549)
Net Present Worth						

Appendix C. Aspen Plus Properties

The table below is a list of the components used in the Aspen model. Previous versions of NREL models 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.

Component	Property	Quantity	Units	Reference
H2O	-	-	-	Native Aspen component
ETHANOL	-	-	-	Native Aspen component
GLUCOSE	-	-	-	Native Aspen component (dextrose)
SUCROSE	-	-	-	Native Aspen component; used to represent non-fermentable carbohydrates
HMF	MW	126.11		(5-hydroxymethyl furfural) Properties for HMF were estimated within Aspen using NIST TDE routines. Specify molecular structure, MW, TB and DHFORM.
	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		
LACID	-	-	-	Native Aspen component (lactic acid)
NH3	-	-	-	Native Aspen component
H2SO4	-	-	-	Native Aspen component
NH4SO4	-	-	-	Native Aspen component (ammonium sulfate)
DAP	-	-	-	Native Aspen component (diammonium phosphate)
NAOH	-	-	-	Native Aspen component
OIL	-	-	-	Native Aspen component (oleic acid); used to represent polar lipid impurity fractions
O2	-	-	-	Native Aspen component
N2	-	-	-	Native Aspen component
NO	-	-	-	Native Aspen component
NO2	-	-	-	Native Aspen component
CO	-	-	-	Native Aspen component

Component	Property	Quantity	Units	Reference
CO2	-	-	-	Native Aspen component
CH4	-	-	-	Native Aspen component
H2S	-	-	-	Native Aspen component
SO2	-	-	-	Native Aspen component
GLUCAN	-	-	-	Native Aspen component (starch); based on [6] CELLULOS component
PROTEIN	Formula	$CH_{1.57}O_{0.31}N_{0.29}S_{0.007}$		Wheat gliadin; all protein component properties left unchanged from [6]
	MW	22.8396		
	DHSFRM	-17618	cal/mol	From literature value of gliadin ΔH_c
	PLXANT/1	-1.00E+20	atm	Forces non-volatility
ASH	-	-	-	Native Aspen component CaO
ZYMO	Formula	$CH_{1.8}O_{0.5}N_{0.2}$		Used to represent *S. cerevisiae* cell mass based on [3, 6] ZYMO component
	MW	24.6264		
	DHSFRM	-31169.39	cal/mol	See [6]
BIOMASS	Formula	$CH_{1.64}O_{0.39}N_{0.23}S_{0.0035}$		Anaerobic digestion solids residue in digestate cake; based on [6]
	MW	23.238		
	DHSFRM	-23200.01	cal/mol	See [6]
CASO4	-	-	-	Gypsum, used to represent silica and clay for lipid purification; native Aspen component
TAG	-	-	-	FAME lipids. Native Aspen component Triolein
HEXANE	-	-	-	Native Aspen component (n-hexane)
H3PO4	Formula	H_3PO_4		Phosphoric acid for lipid purification. Native Aspen component
H2	Formula	H_2		Native Aspen component

Component	Property	Quantity	Units	Reference
C9	Formula	C_9H_{20}		Native Aspen component (n-nonane)
C10	Formula	$C_{10}H_{22}$		Native Aspen component (n-decane)
C11	Formula	$C_{11}H_{24}$		Native Aspen component (n-undecane)
C12	Formula	$C_{12}H_{26}$		Native Aspen component (n-dodecane)
C13	Formula	$C_{13}H_{28}$		Native Aspen component (n-tridecane)
C14	Formula	$C_{14}H_{30}$		Native Aspen component (n-tetradecane)
C15	Formula	$C_{15}H_{32}$		Native Aspen component (n-pentadecane)
C16	Formula	$C_{16}H_{34}$		Native Aspen component (n-hexadecane)
C17	Formula	$C_{17}H_{36}$		Native Aspen component (n-heptadecane)
C18	Formula	$C_{18}H_{38}$		Native Aspen component (n-octandecane)
C19	Formula	$C_{19}H_{40}$		Native Aspen component (n-nonadecane)
C20	Formula	$C_{20}H_{42}$		Native Aspen component (n-eicosane)
ALGCEL	-	-	-	Algal cell biomass. Duplicate of component ZYMO.

Appendix D. Process Flow Diagrams

High-level stream table information from Aspen Plus modeling output follows, for key streams associated with each process operation area. The stream table values are associated with the *HLSD + Store* base case scenario, for the summer season which is used to set the basis for the facility design capacity. 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%.

RDB	Paraffins targeted in the C13-C20 range
Naphtha	Paraffins targeted in the C6-C12 range
Soluble Fermentable Sugars	Glucose, mannose (modeled as 100% glucose)
Furfurals	Furfural, HMF
CO/SOx/NOx/H_2S	CO, SO_2, NO, NO_2, H_2S
Cell Mass	*S. cerevisiae*, algal cell mass (see Table 1), AD solids
Other Insoluble Solids	Silica, clay (lipid purification chemicals)

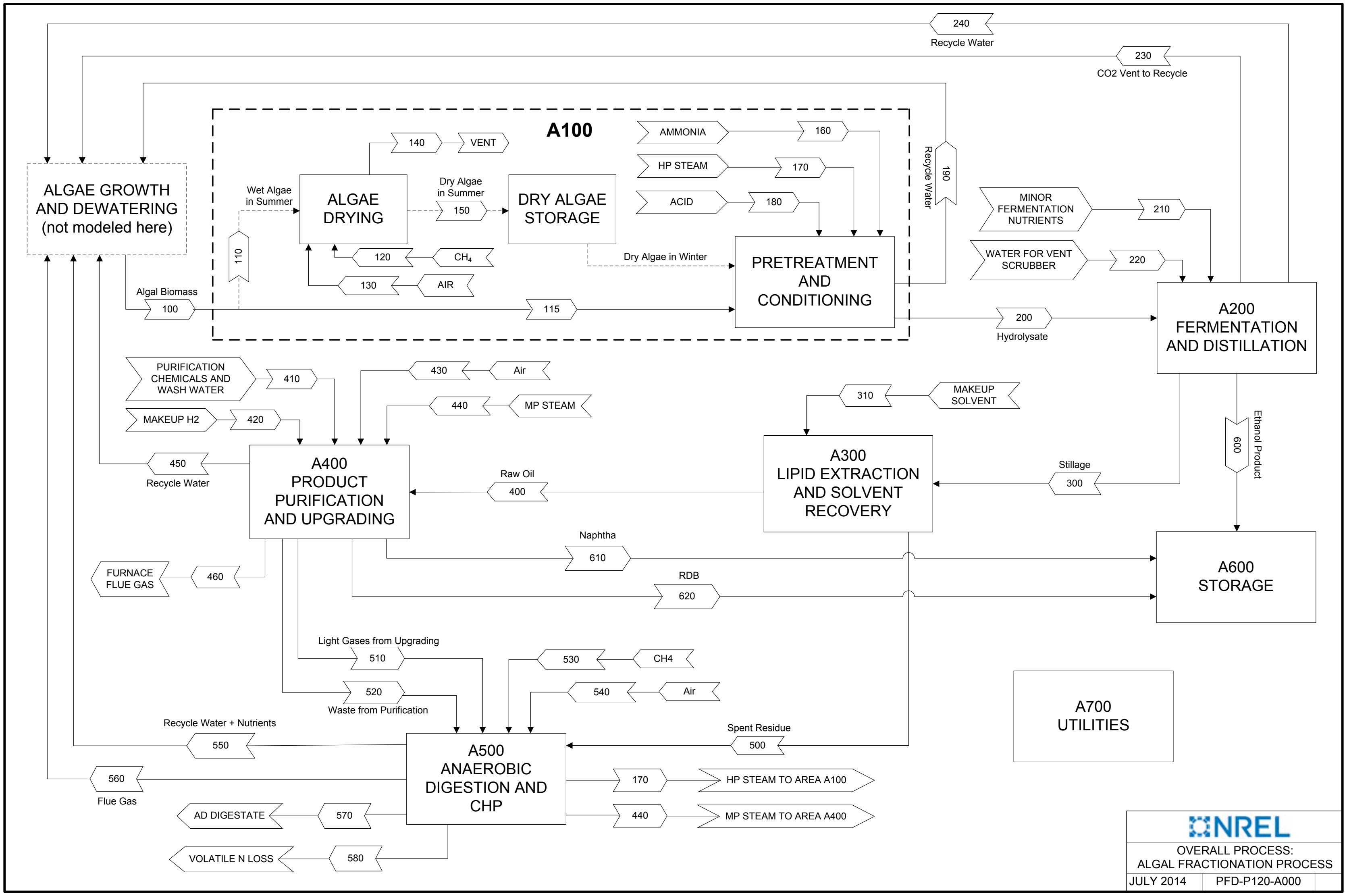


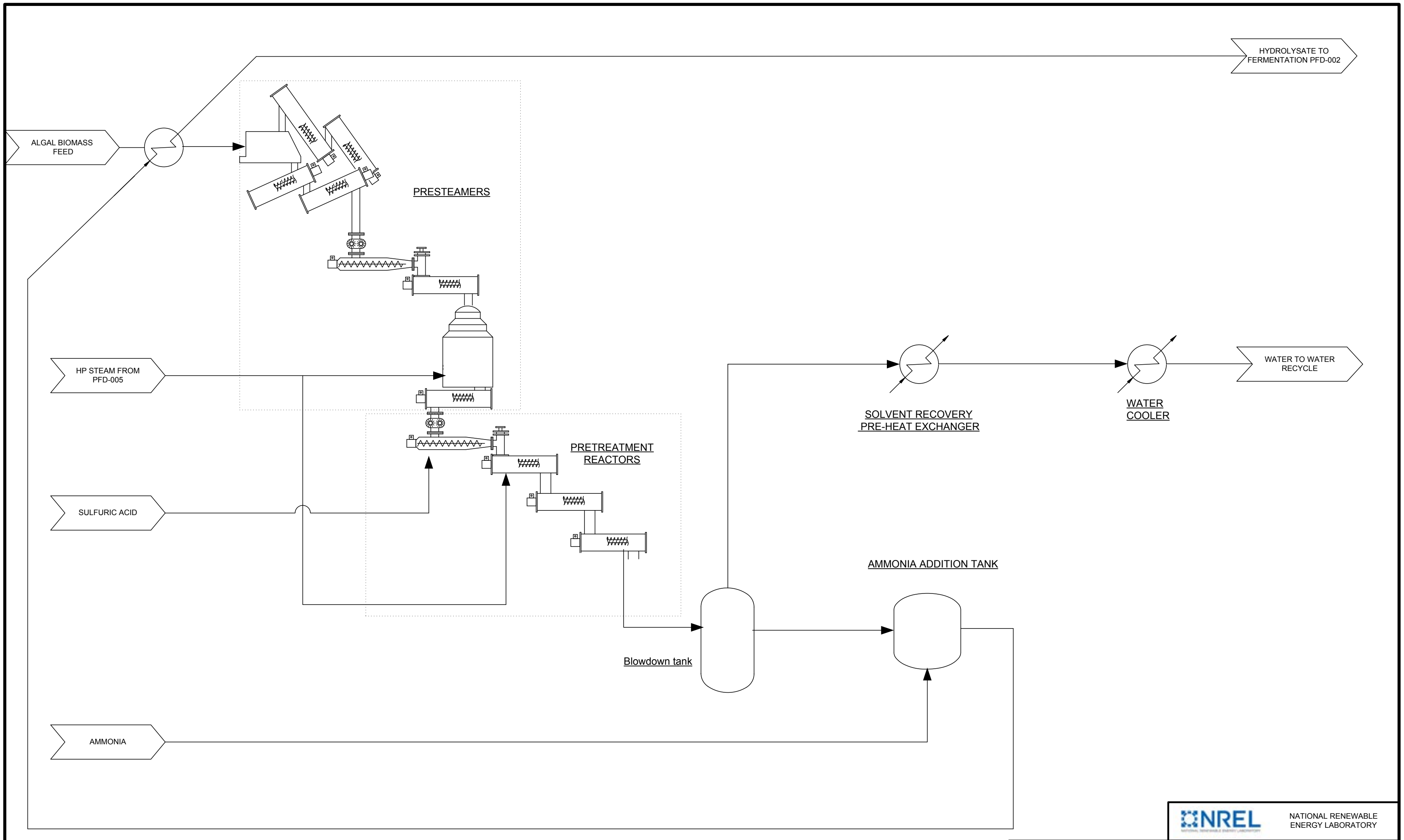
Figure 19. Overall process schematic for Aspen stream tables

Aspen Plus mass balance information for key stream tables:

COMPONENT	UNITS	100	110	115	120	130	140	150	160	170	180	190	200	210
Total Flow	kg/hr	430,242	147,375	282,867	8,111	154,354	276,968	32,872	786	25,847	2,433	19,960	291,973	112
Insoluble Solids	wt%	20.0%	20.0%	20.0%	0.0%	0.0%	0.0%	89.7%	0.0%	0.0%	0.0%	0.0%	11.9%	0.0%
Soluble Solids	wt%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	9.2%	100.0%
Temperature	°C	25	25	25	25	25	100	100	25	268	20	37	37	25
Pressure	atm	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	13.0	3.4	1.8	1.0	1.0
Vapor Fraction		0.0	0.0	0.0	1.0	1.0	1.0	0.0	1.0	1.0	0.0	0.0	0.0	0.0
RDB	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0
Naphtha	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethanol	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0
Water	kg/hr	344,194	117,900	226,294	0	0	132,720	3,397	0	25,847	170	19,956	230,210	0
Soluble Fermentable Sugars	kg/hr	0	0	0	0	0	0	0	0	0	0	0	21,594	0
Non-Fermentable Carbohydrates	kg/hr	3,382	1,158	2,223	0	0	0	0	0	0	0	0	2,223	0
FAME Lipids	kg/hr	35,443	12,141	23,303	0	0	0	0	0	0	0	5	23,298	0
Polar Lipid Impurities	kg/hr	1,773	607	1,165	0	0	0	0	0	0	0	0	1,165	0
Hexane	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0
Furfurals	kg/hr	0	0	0	0	0	0	0	0	0	0	0	50	0
Lactic Acid	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0
Ammonia	kg/hr	0	0	0	0	0	0	0	786	0	0	0	0	0
Sulfuric Acid	kg/hr	0	0	0	0	0	0	0	0	0	2,263	0	0	0
Ammonium Sulfate	kg/hr	0	0	0	0	0	0	0	0	0	0	0	3,049	0
Phosphoric Acid	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0
DAP	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	112
H2	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0
CH4	kg/hr	0	0	0	8,111	0	0	0	0	0	0	0	0	0
Propane	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	kg/hr	0	0	0	0	35,952	3,595	0	0	0	0	0	0	0
N2	kg/hr	0	0	0	0	118,402	118,402	0	0	0	0	0	0	0
CO2	kg/hr	0	0	0	0	0	22,251	0	0	0	0	0	0	0
CO/SOX/NOX/H2S	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbohydrate (IS)	kg/hr	32,845	11,251	21,594	0	0	0	0	0	0	0	0	2,095	0
Protein (IS)	kg/hr	7,710	2,641	5,069	0	0	0	0	0	0	0	0	5,069	0
Cell Mass (IS)	kg/hr	3,072	1,052	2,020	0	0	0	29,475	0	0	0	0	2,020	0
Ash	kg/hr	1,824	625	1,199	0	0	0	0	0	0	0	0	1,199	0
Other Insoluble Solids	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0

COMPONENT	UNITS	220	230	240	300	310	400	410	420	430	440	450	460	500
Total Flow	kg/hr	12,282	9,609	22,755	262,260	993	22,576	2,368	546	1,066	780	1,082	1,227	240,677
Insoluble Solids	wt%	0.0%	0.0%	0.0%	13.7%	0.0%	0.0%	2.9%	0.0%	0.0%	0.0%	0.0%	0.0%	5.5%
Soluble Solids	wt%	0.0%	0.0%	0.0%	2.5%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	0.0%	2.7%
Temperature	°C	33	67	30	125	25	110	25	25	32	232	78	380	52
Pressure	atm	5.0	1.2	1.6	2.3	1.0	1.0	1.0	1.0	1.0	2.4	1.7	1.2	1.0
Vapor Fraction		0.0	1.0	0.0	0.0	0.0	0.0	0.0	1.0	1.0	1.0	0.0	1.0	0.0
RDB	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0
Naphtha	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethanol	kg/hr	0	2	11	125	0	0	0	0	0	0	0	0	125
Water	kg/hr	12,282	245	22,738	219,759	0	0	2,258	0	0	780	1,082	128	219,759
Soluble Fermentable Sugars	kg/hr	0	0	0	569	0	0	0	0	0	0	0	0	569
Non-Fermentable Carbohydrates	kg/hr	0	0	0	2,223	0	0	0	0	0	0	0	0	2,223
FAME Lipids	kg/hr	0	0	5	23,293	0	22,124	0	0	0	0	0	0	1,170
Polar Lipid Impurities	kg/hr	0	0	0	1,165	0	385	0	0	0	0	0	0	781
Hexane	kg/hr	0	0	0	0	993	68	0	0	0	0	0	0	925
Furfurals	kg/hr	0	0	0	50	0	0	0	0	0	0	0	0	50
Lactic Acid	kg/hr	0	0	0	586	0	0	0	0	0	0	0	0	586
Ammonia	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0
Sulfuric Acid	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0
Ammonium Sulfate	kg/hr	0	0	0	3,049	0	0	0	0	0	0	0	0	3,049
Phosphoric Acid	kg/hr	0	0	0	0	0	0	43	0	0	0	0	0	0
DAP	kg/hr	0	0	0	96	0	0	0	0	0	0	0	0	96
H2	kg/hr	0	0	0	0	0	0	0	546	0	0	0	0	0
CH4	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0
Propane	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0
O2	kg/hr	0	0	0	0	0	0	0	0	248	0	0	41	0
N2	kg/hr	0	0	0	0	0	0	0	0	818	0	0	818	0
CO2	kg/hr	0	9,362	0	0	0	0	0	0	0	0	0	240	0
CO/SOX/NOX/H2S	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0	0
Carbohydrate (IS)	kg/hr	0	0	0	2,095	0	0	0	0	0	0	0	0	2,095
Protein (IS)	kg/hr	0	0	0	5,011	0	0	0	0	0	0	0	0	5,011
Cell Mass (IS)	kg/hr	0	0	0	3,039	0	0	0	0	0	0	0	0	3,039
Ash	kg/hr	0	0	0	1,199	0	0	0	0	0	0	0	0	1,199
Other Insoluble Solids	kg/hr	0	0	0	0	0	0	68	0	0	0	0	0	0

COMPONENT	UNITS	510	520	530	540	550	560	570	580	600	610	620
Total Flow	kg/hr	4,154	2,821	949	334,316	206,658	348,022	28,194	43	9,744	392	17,661
Insoluble Solids	wt%	0.0%	16.0%	0.0%	0.0%	0.1%	0.0%	29.7%	0.0%	0.0%	0.0%	0.0%
Soluble Solids	wt%	0.0%	0.0%	0.0%	0.0%	1.5%	0.0%	0.9%	0.0%	0.0%	0.0%	0.0%
Temperature	°C	30	93	25	25	35	140	35	35	20	20	20
Pressure	atm	1.7	1.0	1.0	1.0	1.0	1.5	1.0	1.0	1.0	1.0	1.0
Vapor Fraction		1.0	0.0	1.0	1.0	0.0	1.0	0.0	1.0	0.0	0.0	0.0
RDB	kg/hr	1	0	0	0	0	0	0	0	0	0	17,608
Naphtha	kg/hr	12	0	0	0	0	0	0	0	0	382	0
Ethanol	kg/hr	0	0	0	0	0	0	0	0	9,678	0	0
Water	kg/hr	25	2,258	0	0	202,636	14,287	19,114	0	49	2	53
Soluble Fermentable Sugars	kg/hr	0	0	0	0	0	0	0	0	0	0	0
Non-Fermentable Carbohydrates	kg/hr	0	0	0	0	0	0	0	0	0	0	0
FAME Lipids	kg/hr	0	0	0	0	0	0	0	0	0	0	0
Polar Lipid Impurities	kg/hr	0	385	0	0	0	0	0	0	0	0	0
Hexane	kg/hr	0	68	0	0	0	0	0	0	0	0	0
Furfurals	kg/hr	0	0	0	0	0	0	0	0	0	0	0
Lactic Acid	kg/hr	0	0	0	0	0	0	0	0	0	0	0
Ammonia	kg/hr	0	0	0	0	740	0	217	43	0	0	0
Sulfuric Acid	kg/hr	0	0	0	0	0	0	0	0	0	0	0
Ammonium Sulfate	kg/hr	0	0	0	0	2,786	0	263	0	0	0	0
Phosphoric Acid	kg/hr	0	43	0	0	0	0	239	0	0	0	0
DAP	kg/hr	0	0	0	0	322	0	0	0	0	0	0
H2	kg/hr	159	0	0	0	0	0	0	0	0	0	0
CH4	kg/hr	0	0	949	0	0	0	0	0	0	0	0
Propane	kg/hr	1,212	0	0	0	0	0	0	0	0	7	0
O2	kg/hr	0	0	0	77,868	0	52,247	0	0	0	0	0
N2	kg/hr	0	0	0	256,448	0	256,360	0	0	0	0	0
CO2	kg/hr	2,622	0	0	0	2	24,643	0	0	17	1	0
CO/SOX/NOX/H2S	kg/hr	123	0	0	0	2	485	0	0	0	0	0
Carbohydrate (IS)	kg/hr	0	0	0	0	0	0	0	0	0	0	0
Protein (IS)	kg/hr	0	0	0	0	0	0	0	0	0	0	0
Cell Mass (IS)	kg/hr	0	0	0	0	145	0	7,120	0	0	0	0
Ash	kg/hr	0	0	0	0	24	0	1,175	0	0	0	0
Other Insoluble Solids	kg/hr	0	68	0	0	1	0	66	0	0	0	0



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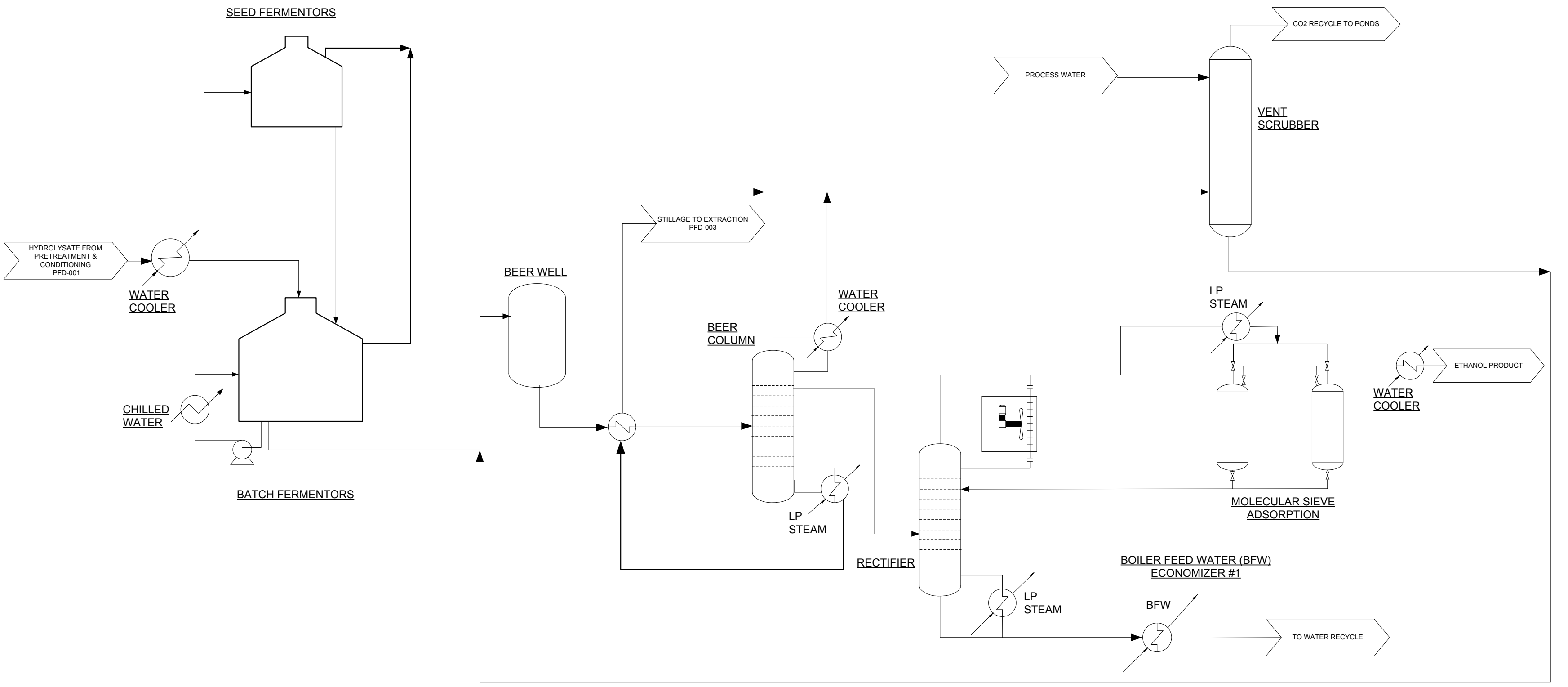
ADVANCED BIOFUELS ENGINEERING DESIGN
ALGAE FRACTIONATION TO FUELS

PROCESS FLOW DIAGRAM
A100 PRETREATMENT AND CONDITIONING

Subconsultant:	–		
Drawn: JAR	–	Check:	–
Engr: JAR	–	Check:	–
Appr:	–	PMgr:	–

Project No:	Drawing:	Rev:
	PFD-001	A



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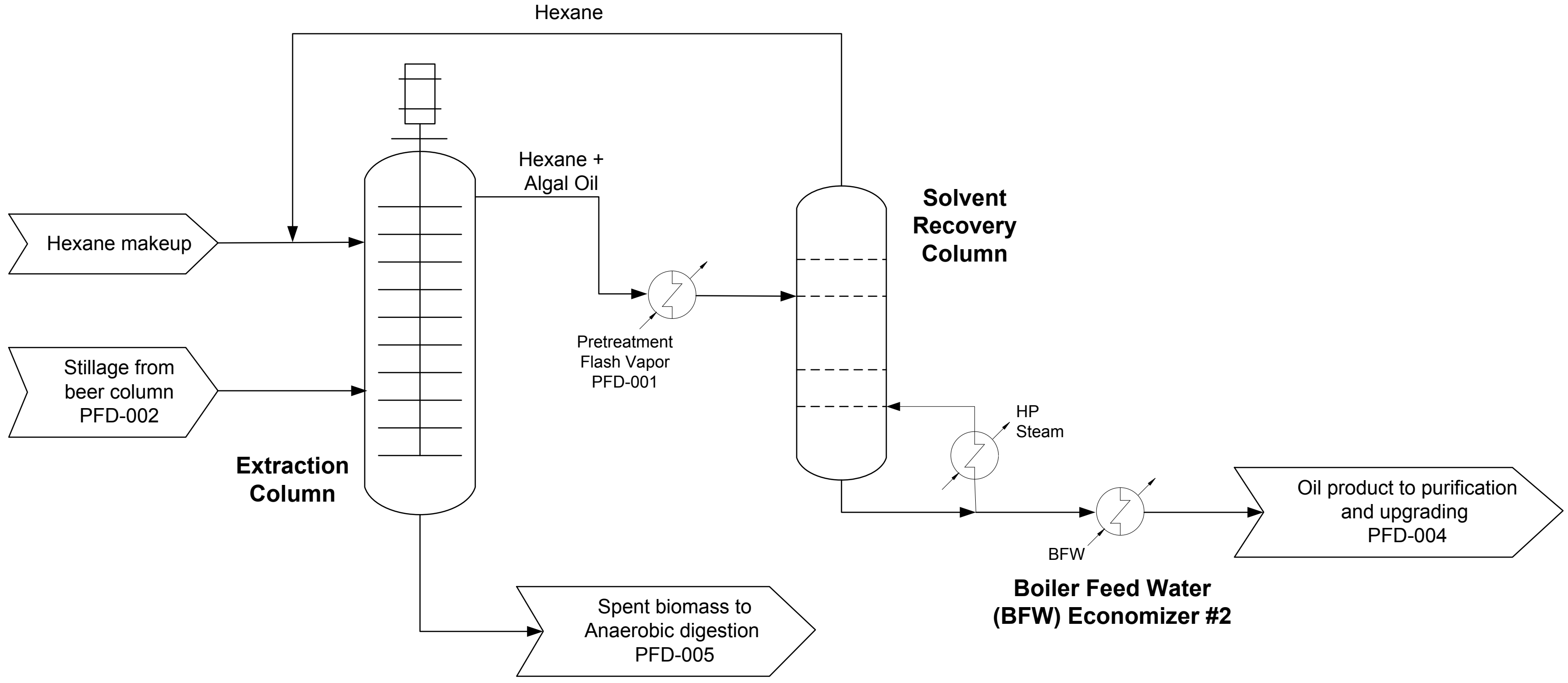
ADVANCED BIOFUELS ENGINEERING DESIGN
ALGAE FRACTIONATION TO FUELS

PROCESS FLOW DIAGRAM
A200 FERMENTATION AND SEPARATION

Project No:
Drawing: PFD-002
Rev: A



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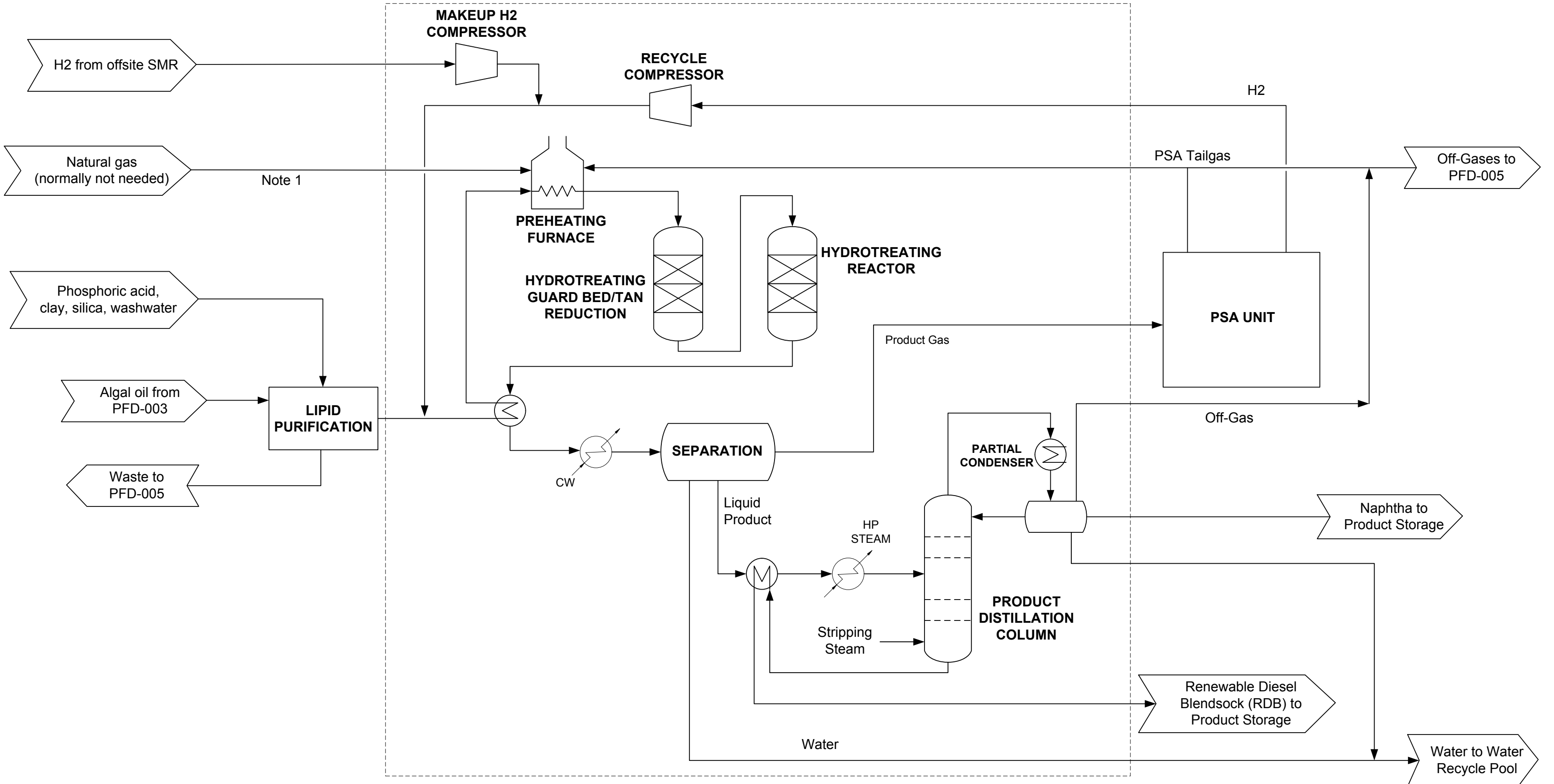
ALGAE FRACTIONATION TO FUELS

PROCESS FLOW DIAGRAM
A300: EXTRACTION AND SOLVENT RECOVERY

Subconsultant:	-		
Drawn: JAR	-	Check:	-
Engr: JAR	-	Check:	-
Appr:	-	PMgr:	-

Project No:	Drawing:	Rev:
	PFD-003	A



Dashed box represents hydrotreating ISBL system

Notes:
1. Natural gas for startup and as needed

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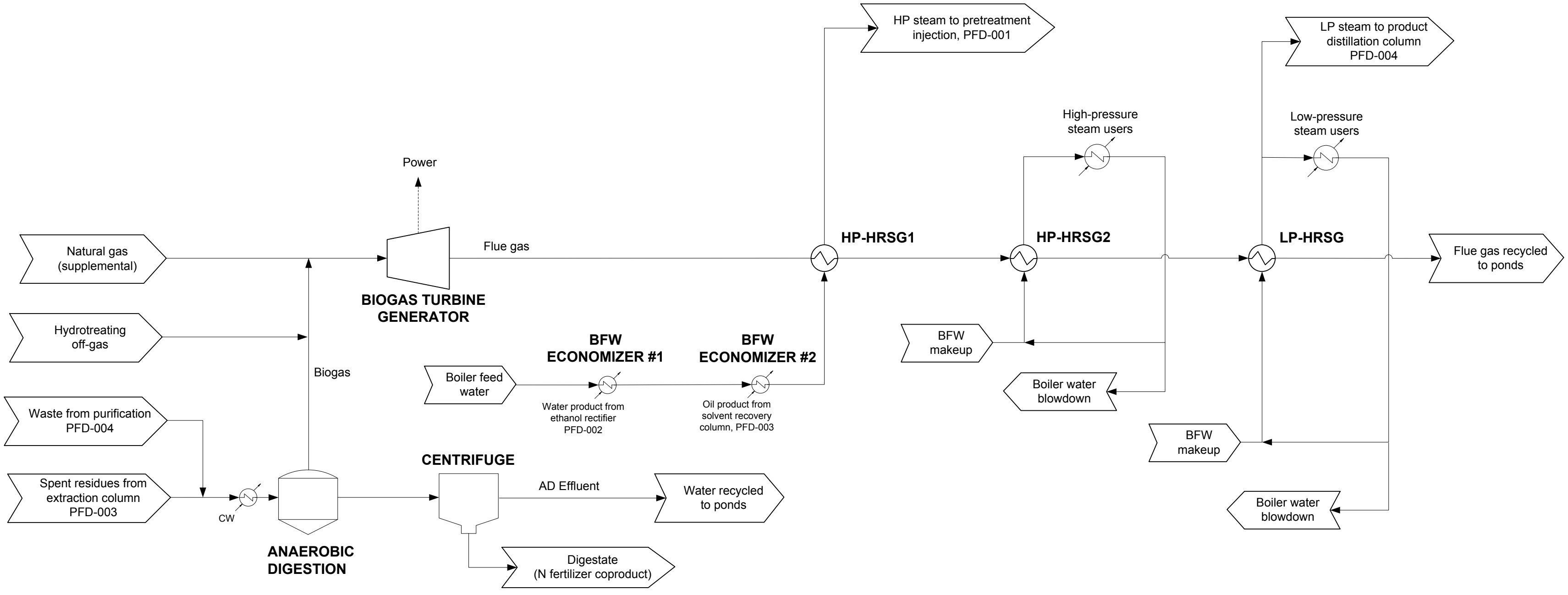
Engr: JAR	-	Check:	-
Appr:	-	PMgr:	-

ADVANCED BIOFUELS ENGINEERING DESIGN
ALGAE FRACTIONATION TO FUELS

PROCESS FLOW DIAGRAM
A400: PRODUCT PURIFICATION AND UPGRADING

Project No:	Drawing: PFD-004	Rev: A



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ADVANCED BIOFUELS ENGINEERING DESIGN
ALGAE FRACTIONATION TO FUELS

PROCESS FLOW DIAGRAM
A500: ANAEROBIC DIGESTION AND CHP

Subconsultant:	-		
Drawn: JAR	-	Check:	-
Engr: JAR	-	Check:	-
Appr:	-	PMgr:	-

Project No:
Drawing: PFD-005
Rev: A