



Conceptual Basis and Techno-Economic Modeling for Integrated Algal Biorefinery Conversion of Microalgae to Fuels and Products

2019 NREL TEA Update: Highlighting Paths to Future Cost Goals via a New Pathway for Combined Algal Processing

Ryan Davis,¹ Matthew Wiatrowski,¹ Christopher Kinchin,¹ and David Humbird²

1 National Renewable Energy Laboratory

2 DWH Process Consulting LLC

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List of Acronyms

ACCE	Aspen Capital Cost Estimator	IX	ion exchange
AD	anaerobic digestion	kcal	kilocalorie
AFDW	ash-free dry weight	kg	kilogram
atm	atmosphere	KOH	potassium hydroxide
BETO	Bioenergy Technologies Office	kWh	kilowatt hour
BOP	balance of plant	L	liter
Btu	British thermal unit	lb	pound
CAP	combined algal processing	LCA	life cycle analysis
CAPEX	capital expenditures	LLE	liquid-liquid extraction
cc	cubic centimeter	LP	low-pressure
CHNOS	carbon, hydrogen, nitrogen, oxygen, sulfur	m ³	cubic meter
CHP	combined heat and power	MBSP	minimum biomass selling price
CIP	clean-in-place	MFSP	minimum fuel selling price
CO ₂	carbon dioxide	mg	milligram
CSTR	continuous stirred-tank reactor	MJ	megajoule
DAP	diammonium phosphate	mm	millimeter
DCFROR	discounted cash flow rate of return	MM	million
DEOA	diethanolamine	MOT	mild oxidative treatment
DOE	U.S. Department of Energy	MW	megawatt
f ²	square feet	NaOH	sodium hydroxide
FAME	fatty acid methyl ester	Nb ₂ O ₅	niobium pentoxide
FCI	fixed capital investment	NG	natural gas
FFA	free fatty acid	NH ₃	ammonia
g	gram	N/P	nitrogen/phosphorous
gal	gallon	NREL	National Renewable Energy Laboratory
Gcal	gigacalorie	OPEX	operating expenses
GGE	gallon gasoline equivalent	OSBL	outside battery limits
GHG	greenhouse gas	ppm	parts per million
H ₂	hydrogen	psig	pounds per square inch gauge
H ₂ O ₂	hydrogen peroxide	Pt	platinum
H ₂ SO ₄	sulfuric acid	PU	polyurethane
H ₃ PO ₄	phosphoric acid	PUFA	polyunsaturated fatty acid
HCSD	high-carbohydrate (mid- harvest)	RDB	renewable diesel blend-stock
	<i>Scenedesmus</i>	SCFD	standard cubic feet per day
HDO	hydrodeoxygenation	SMR	steam methane reforming
HI	hydroisomerization	SO ₂	sulfur dioxide
HLSD	high-lipid (late harvest)	TAG	triglyceride
	<i>Scenedesmus</i>	TCI	total capital investment
HP	high-pressure	TDC	total direct cost
HPSD	high-protein (early harvest)	TDI	toluene diisocyanate
	<i>Scenedesmus</i>	TEA	techno-economic analysis
hr	hour	USDA	U.S. Department of Agriculture
INL	Idaho National Laboratory	WHSV	weight hourly space velocity
IRR	internal rate of return	wt%	weight percent
ISBL	inside battery limits	yr	year
		ZrO ₂	zirconium dioxide

Executive Summary

The U.S. Department of Energy promotes the production of a range of liquid fuels and chemicals derived 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 such processes. This includes pathways from lignocellulosic (terrestrial) biomass, as well as from algal (aquatic) biomass systems.

Over the past decade, NREL has maintained a program focusing on research and techno-economic analysis (TEA) for technologies to produce and convert algal biomass to fuels, and more recently bio-derived products. Algal conversion processes at NREL have largely fallen under a processing approach termed “combined algal processing” (CAP), generally defined as a fractionation approach borrowing from biochemical conversion concepts wherein individual constituents of algae (e.g., carbohydrates, lipids, and protein) are isolated and upgraded to fuels or products across a range of conversion technology options. Historically, NREL CAP conversion pathways have typically involved a dilute acid pretreatment step to hydrolyze carbohydrates into monomeric sugars and enable downstream lipid recovery, followed by fermentation of sugars and extraction of lipids, both of which could go toward producing either fuels or chemical products. Protein was traditionally relegated to anaerobic digestion given more limited options for utilization of protein for fuels or chemicals. While effective, this approach implicitly limited biomass feedstock options to low protein and moderate-to-high carbohydrate and lipid contents, which places more burdens on upstream cultivation in order to both reduce biomass costs (increase cultivation productivity rates) and simultaneously improve biomass compositional quality (move away from nutrient-replete, high-protein biomass harvesting).

The present report documents the conceptual basis for a new potential CAP design strategy which may allow more flexibility in accommodating different algal biomass feedstock compositions by enabling upgrading of both protein and carbohydrates in a single step, without a strict requirement for either component to be in soluble or monomeric form, while maintaining effective wet lipid extraction techniques to enable high lipid recoveries. In light of previously established constraints around algal biomass costs (which are significantly higher than lignocellulosic terrestrial biomass), the present CAP processing strategy reflects an integrated biorefinery concept producing both fuels and value-added chemical coproducts as a means to improve profitability and generate coproduct revenues to help drive down the minimum fuel selling price (MFSP) toward economically viable levels. Namely, this report highlights an integrated CAP biorefinery process and associated technical targets that would be required to achieve U.S. Department of Energy target MFSP goals of \$2.5/gallon gasoline equivalent (GGE) by 2030. This is accomplished by a process involving low-cost seasonal storage of algal biomass during high-growth seasons, rapid flash hydrolysis pretreatment of the biomass, solvent extraction of pretreated biomass, cleanup and fractionation of lipids into triglyceride and free fatty acid fractions, and a series of thermochemical conversion steps to upgrade carbohydrates and protein to hydrocarbon fuels. These steps include mild oxidative treatment (MOT), a process originally investigated at NREL for upgrading lignin, followed by catalytic ketonization, condensation, and hydrotreating of MOT products to fuels. Isolated triglycerides are sent to a coproduct train, with the base case focused on upgrading to polyurethane foams as a high-value, high-market-volume coproduct at a base case value of \$2.04/lb.

To evaluate economic feasibility of the above processing approach, NREL, in collaboration with subcontractors DWH Process Consulting and Nexant, conducted a TEA for this new CAP biorefinery concept. The modeled facility processes an average of 565 dry ash-free tons of algal biomass per day at a delivered cost of \$486/ton, reflective of an upstream algae farm co-located with the CAP facility achieving a targeted annual average cultivation productivity of 25 g/m²/day over 5,000 acres of production ponds, inclusive of biomass harvesting/dewatering to 18 wt% solids (ash-free dry weight); alternative scales are investigated as a sensitivity. The biorefinery achieves an overall fuel selling price of \$2.56/GGE in 2016 dollars as determined by modeled conversion performance targets and “*n*th-plant” project costs and financing, associated with a net fuel yield of 88 GGE/dry ton. These values are attributed to a continued focus at present around a high carbohydrate (53%), mid-level lipid (27%), low protein (13%)/low ash (2%) algal biomass feedstock composition, generally reflective of an early transitional stage of nutrient depletion between fully replete (high protein) and deplete (high lipid) harvesting that many strains undergo. However, other compositional scenarios were considered as part of a sensitivity analysis to begin understanding implications for accommodating alternate algae feedstocks.

The analysis reiterates that algal biofuel economics are driven strongly by feedstock costs, contributing over \$5/GGE to the fuel selling price at the feedstock costs and fuel yields indicated above. To offset such a substantial cost, polyurethanes are coproduced from the isolated lipid triglycerides, accounting for a net *reduction* of nearly \$6/GGE in MFSP (inclusive of both production costs and product revenues), leaving approximately \$3/GGE MFSP contributions from fuel-train CAP processing costs across pretreatment, lipid extraction and separation, MOT/catalytic upgrading, and supporting utilities. Given such high influence from the polyurethane coproduct train, any related factors impacting the output yields of coproduct versus fuel were generally found to exhibit strong sensitivities on overall MFSP, including delivered biomass composition (more strongly a reflection of lipid content and polyurethane potential than CAP fuel-train dependencies), the ratio of lipid triglycerides versus free fatty acids, polyurethane coproduct value, and lipid extraction yields. Additional strong cost drivers include algal feedstock costs (as noted above), facility scale (reflecting economy-of-scale dependencies), and MOT yields/selectivities to carboxylic acid products. The analysis also found that although this process approach allows for accommodating more variable biomass compositions, including higher-protein algae at a possible benefit to algae farm cultivation productivity and biomass cost, such a higher-protein composition delivered at a lower cost reflective of a hypothetically higher 30 g/m²/day productivity would ultimately translate to a considerable net MFSP penalty. Accordingly, such a scenario would require identifying additional or different coproducts to target in still enabling the potential to achieve the \$2.5/GGE target.

It should be emphasized that this assessment represents an early state of understanding and conceptual development for this CAP configuration, and accordingly this document is *not* a formal design report intended to set detailed out-year targets, and the outputs from the TEA modeling presented here should be interpreted with caution. In particular, uncertainties exist around the flash hydrolysis, MOT, and ketonization/condensation steps, as NREL experimental data is not yet well established for those operations. Namely, analytical data is needed to validate the asserted targets for MOT acid yields and speciation, as well as ketonization/condensation product structures and chain lengths as a function of MOT acid species and coupling reactions. As such details are better understood moving forward, this TEA work will be refined in future updates to guide research directions toward the most economically impactful priorities.

Combined Algal Processing to Fuels and Bioproducts Process Engineering Analysis

Cost Year Basis: 2016 \$
MFSP (Gasoline Equivalent Basis): \$2.56 /GGE

Total Fuel Production (RDB + Naphtha):	16.41 MMGGE/yr
RDB Production:	14.77 MMGGE/yr
Naphtha Production:	1.64 MMGGE/yr
Total Fuel Yield:	87.95 GGE / dry U.S. ton feedstock
<i>Polyurethane Coproduct Production:</i>	4871 kg/hr (full price product)
	203 kg/hr (scrap, 50% value)
	475 lb/dry ton feedstock (all PU production)
Feedstock Cost:	\$486 dry ton algal biomass (ash free)
Internal Rate of Return:	10%
Equity Percent of Total Investment:	40%

Capital Costs	
A100: Seasonal Storage and Pretreatment	\$7,300,000
A200: Lipid Extraction, Purification, and Separation	\$23,900,000
A300: MOT and Catalytic Upgrading	\$30,100,000
A400: Polyol/Polyurethane Production	\$78,300,000
A500: Boiler, Utilities, and Storage	\$12,400,000
Total Installed Equipment Cost	
	\$152,000,000
Added Direct + Indirect Costs	
	\$187,500,000
(% of TCI)	55.23%
Total Capital Investment (TCI)	
	\$339,500,000
Installed Equipment Cost/Annual GGE	
	\$9.26
Total Capital Investment/Annual GGE	
	\$20.69
Loan Rate	
	8%
Term (years)	
	10
Capital Charge Factor (Computed)	
	0.133
Carbon Retention Efficiencies	
Total Carbon Efficiency to Fuel Products (Fuel C/Biomass C)	44.8%
RDB (RDB C/Biomass C)	40.4%
Naphtha (Naphtha C/Biomass C)	4.3%
Total Carbon Efficiency to Polyurethane	
(Polyurethane C/Biomass C)	30.2%
(Polyurethane C from Biomass/Biomass C)	19.3%
Specific Operation Conditions	
Feedstock Composition	HCSD
TAG/FFA Ratio	50:50
Lipid Extraction Efficiency	96%
Operating Hours/Year	7920

Manufacturing Costs (cents/GGE)	
Feedstock	548.3
O2	28.5
Acid/Base for Ion Exchange Regeneration	15.0
Solvent Makeup	17.1
Oil Cleanup Chemicals	3.5
Hydrogen	17.3
Natural Gas	50.9
Process Water	2.6
TDI	264.5
Polyol/Polyurethane Production, excluding TDI	55.4
Waste Streams	4.4
Electricity	27.7
Coproduct Credits	-1149.2
Catalyst	21.4
Fixed Costs	74.0
Capital Depreciation	65.3
Average Income Tax	20.8
Average Return on Investment	188.4
Total	255.9
Manufacturing Costs (\$/yr)	
Feedstock	\$90,000,000
O2	\$4,700,000
Acid/Base for Ion Exchange Regeneration	\$2,500,000
Solvent Makeup	\$2,800,000
Oil Cleanup Chemicals	\$600,000
Hydrogen	\$2,800,000
Natural Gas	\$8,400,000
Process Water	\$400,000
TDI	\$43,400,000
Polyol/Polyurethane Production, excluding TDI	\$9,100,000
Waste Streams	\$700,000
Electricity	\$4,500,000
Coproduct Credits	-\$188,600,000
Catalyst	\$3,500,000
Fixed Costs	\$12,100,000
Capital Depreciation	\$10,700,000
Average Income Tax	\$3,400,000
Average Return on Investment	\$30,900,000
Total	\$41,900,000

Figure ES-1. Economic summary for modeled CAP biorefinery pathway. Calculation methodologies behind the details shown here may be found in NREL's public TEA models (<https://www.nrel.gov/extranet/biorefinery/aspen-models/>).

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1 Introduction

1.1 Background and Motivation

The U.S. Department of Energy (DOE) Bioenergy Technologies Office (BETO) promotes the production of liquid fuels and products 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. Within BETO's Advanced Algal Systems platform, these programs include laboratory research to improve biological characteristics (e.g., algal cultivation productivity, biomass composition, and 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 multiorganization consortia partnerships.

To support the DOE program, the National Renewable Energy Laboratory (NREL) investigates the process design and economics of numerous bioenergy production pathways through techno-economic analysis (TEA) modeling to estimate a product's plant gate price based on process and plant design assumptions consistent with applicable best practices in engineering, construction, and operation. For a biorefinery pathway converting biomass to fuels, 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 potential economic viability of a given biofuel technology compared to other technology options and in reference to projected wholesale fuel prices in the future. The TEA effort at NREL also helps to direct biomass conversion research by examining the sensitivity of the MFSP to alternative unit operations 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 BETO program cost targets. It also allows DOE to make more informed decisions about research proposals that claim to reduce costs.

For the past 8 years, NREL has maintained a program focused on TEA modeling for conceptual algal system designs, inclusive of both algal biomass production as well as conversion to fuels and coproducts. This effort has aided BETO, NREL, and other stakeholders to more quantitatively understand the economic implications for current technology performance benchmarks, future "best-case" technology improvements, and associated gaps and drivers in moving from the former to the latter. Within this timeframe, NREL published several notable TEA studies reflecting the focus of research and understanding of technology performance as has evolved over time. Initially, early TEA models had considered complete farm-to-fuel integrated systems reflective of both algal biomass production as well as conversion to fuels, albeit based on simplifying assumptions and assertions for technology performance that had not yet been well proven in realistic conditions or reasonable scales [1-3]. However, as research efforts caught up and more was learned about what such systems might be expected to look like at commercial scale, the models were refined with more detailed granularity and made more reflective of experimentally demonstrated approaches. On the biomass production side, NREL published an algae farm design report establishing algal cultivation and harvesting/dewatering performance targets which, if achieved in a hypothetical 5,000-acre algae farm under n^{th} -plant conditions,

would translate to a minimum biomass selling price (MBSP) below \$500/ton as an aspirational future target [4].

On the algal biomass conversion side, beginning in 2014 NREL established a research effort focused on selective isolation and conversion of key constituents in algal biomass (lipids, carbohydrates, protein, and other components) under a processing approach broadly referred to as “combined algal processing” (CAP) [5]. In contrast to whole-biomass thermochemical conversion techniques, the CAP process was more reflective of a biochemical conversion strategy under less severe operating conditions whereby discrete components of the biomass feedstock could be isolated and upgraded to fuels or products across a host of options for conversion pathways and product suites. Preliminary CAP configurations focused on well-understood operations including biomass pretreatment (required to enable effective fractionation/isolation of algal constituents, primarily utilizing dilute acid pretreatment), fermentation of algal sugars to ethanol, wet extraction of algal lipids subsequently hydrotreated to diesel-range hydrocarbon fuels, and anaerobic digestion of residual protein for nutrient recycling and combined heat and power (CHP) benefits [6]. Such a configuration was documented in a 2014 CAP design report, reflecting best-case TEA model details for a plausible set of algal biomass cost, composition, and conversion performance targets [7]. That work found that, although low risk from a convertibility standpoint, under even best-case projections such a configuration could only be expected to achieve an MFSP on the order of \$4.4–\$5/gallon gasoline equivalent (GGE) or more when focused on producing fuels alone (with relatively minimal coproduct revenue gains from anaerobic digestion of protein residuals).

Accordingly, over more recent years the TEA modeling and CAP research efforts have shifted in focus toward the inclusion of value-added coproducts which may be produced alongside fuels in a biorefinery configuration. Although this adds processing complexity relative to fuel-only configurations, TEA modeling has demonstrated that *when constrained to future target biomass costs of nearly \$500/ton, algal conversion processes focused on producing fuels alone cannot feasibly achieve MFSP levels below \$3.5–\$4.5/GGE, falling short of BETO algae platform goals of \$2.5/GGE in the future unless higher-value coproducts are also produced* [8, 9].

Within NREL’s CAP program, research and TEA modeling efforts have evaluated a number of algal coproduct opportunities to date, including succinic acid (from sugars), surfactants (from sterols), and initial investigations into bioplastics and mixotrophic algal biomass (from protein). More recently, focus has shifted to lipid-derived polyurethanes, given both relatively high product prices and market volumes, coupled with numerous end-use possibilities to enable tailoring to algal lipid feedstocks and potentially different chemistries/synthesis options. In light of these qualities, this study focuses on coproduct opportunities from algal lipid-derived polyurethanes, or alternatively polyols (intermediate precursors to polyurethanes), as the primary means to increase biorefinery revenues and decrease fuel selling prices.

Prior NREL CAP processing schematics historically have followed a general series of operations noted above based on dilute acid pretreatment (utilized both to hydrolyze carbohydrates into monomeric sugars and to enable effective lipid extraction), fermentation of sugars into fuels/fuel intermediates or products, extraction and upgrading of lipids to fuels, and anaerobic digestion of residual protein. NREL research is currently expanding to pursue new processing approaches. A key underlying factor in this evolution revolves around relaxing the dependency on low-protein/higher-carbohydrate and/or lipid algae as the above processing schematic is built upon,

as well as avoiding the need for high yields of monomeric sugars, which NREL research has shown is highly species- and media-dependent with respect to efficacy of carbohydrate hydrolysis through dilute-acid pretreatment [5]. Instead, the CAP approach considered here utilizes flash hydrolysis pretreatment (more severe conditions but shorter residence time and without utilizing acid), lipid extraction and upgrading (with optional diversion of a fraction of lipids to polyols/polyurethanes, similar to the prior CAP schematic), and oxidative thermochemical treatment to upgrade both solubilized and solid carbohydrates and protein to fuel precursors. In the context of the present study, this new CAP approach is still considered for its application to a lower-protein/higher-carbohydrate harvested algae composition in keeping with prior TEA models [8], but is also anticipated to provide a basis framework for future work moving into more nutrient-replete/higher-protein algal compositions from upstream harvesting.

The focus of this report is to document a plausible pathway model for conversion of algal biomass to fuel and chemical products, with high fractional carbon yield to hydrocarbon fuels while also coproducing value-added products (polyols/polyurethanes) at sufficient levels to demonstrate a means to achieve modeled MFSP costs at approximately \$2.5/GGE by 2030. Similar to prior TEA reports, the work continues NREL's practice of consulting with expert consultants to assist in design/cost estimates and model fidelity for key unit operations, in this case Nexant and DWH Process Consulting. In keeping with caveats noted in prior NREL reports, we stress that this conceptual design serves to describe a *single, feasible* conversion process and to transparently document the assumptions and details that went into it. But in contrast to previously published "design reports" which generally were built on reasonably to well-understood unit operations and conversion technologies (with accompanying level of detail in the reports) [7, 10, 11], this present report is reflective of a less well-established process, in some cases without significant operational or analytical data from in-house experimental work to yet fully define all details of each step with high confidence. Accordingly, this report is more concise and is not intended to convey a highly rigorous design case in exhaustive detail across all operations, but rather presents a conceptual design documenting feasibility-level TEA modeling and associated economics to demonstrate proof-of-concept for this new CAP conversion approach. Moving forward, as data and understanding continues to evolve for newer and more novel areas of the process, the models and economic tools developed for this report may be updated in a similar fashion as prior NREL TEA iterations have been refined.

1.2 Process Overview

The process described here uses rapid flash hydrolysis pretreatment of algal biomass delivered after upstream dewatering to 18 wt% solids (outside the scope of this analysis), followed by whole-slurry solvent extraction of lipids, and vacuum distillation to separate free fatty acids (upgraded to hydrocarbon fuels) from triglycerides (upgraded to polyols/polyurethanes) in the extracted lipids. The aqueous raffinate from extraction is routed to mild oxidative treatment, yielding mixed carboxylic acids which are subsequently upgraded across a series of catalysis steps to hydrocarbon fuels. Prior to flash hydrolysis, seasonally variable biomass delivery rates are normalized by diverting peak seasonal capacity to a wet anaerobic storage process for use in lower-capacity seasons. The process design also includes boiler heat generation, product storage, and required utilities. The process is divided into five areas (see Figure 1).

- *Area 100: Seasonal storage and pretreatment.* In this area, the algal biomass following upstream cultivation, harvesting, and dewatering is either split with a fraction diverted to

wet anaerobic storage (peak production seasons) or combined with biomass from storage (low production seasons) to eliminate seasonal fluctuations through the CAP process. The biomass is then routed to rapid flash hydrolysis at elevated temperature and pressure (but at neutral pH conditions without the addition of acid or caustic). The pretreated slurry is rapidly cooled and routed to extraction.

- *Area 200: Lipid extraction, purification, and separation.* After flashing the pretreatment hydrolysate to 20% total solids (TS, some of which are now present as solubilized proteins and carbohydrates), the slurry is cooled and routed to a wet extraction process. The process consists of a series of agitated reactors and phase separation centrifuges in the presence of a nonpolar solvent (hexane) and a polar co-solvent (ethanol). The nonpolar extract phase is routed to a stripping column to recover hexane and is then sent to bleaching and degumming operations to remove impurities. The lipids are then fractionated via vacuum distillation into free fatty acids (FFAs) and triglycerides (TAGs).
- *Area 300: Mild oxidative treatment and fuel catalytic upgrading.* The aqueous raffinate from extraction is routed to a separate distillation column to recover the ethanol co-solvent, and the stillage is then sent to thermocatalytic upgrading steps. First, the material is routed through mild oxidative treatment (MOT) in the presence of oxygen at elevated temperature/pressure, which converts carbohydrates and protein into mixed carboxylic acids generally in the C₁–C₄ range. The product from MOT is cooled, flashed, filtered of ash and residual solids, and routed through ion exchange (primarily capturing nitrogen and phosphorous liberated from MOT). It is then re-heated and sent to catalytic ketonization, converting carboxylic acids into ketones (primarily acetone and C₆-C₇ cyclic ketones) across a heterogeneous catalyst reactor. The resulting ketones phase separate from water and are then routed to condensation, forming larger-chain oxygenated cyclic intermediates in the jet-diesel range (C₉-C₁₅), followed by hydrodeoxygenation (HDO) hydrotreating alongside the FFAs isolated from lipids in Area 300, yielding a mixture of naphtha-to-diesel range hydrocarbons. A product distillation column fractionates final hydrocarbon fuel phases.
- *Area 400: Polyol/polyurethane coproduction.* The TAGs isolated from Area 300 are sent to a coproduct upgrading train, consisting first of a one-step epoxidation and ring-opening step in the presence of acetic acid and hydrogen peroxide, yielding polyols. The polyols may be sold as one coproduct option, or alternatively may be further upgraded to polyurethanes by reacting with isocyanate cross-linkers to produce polyurethane foams (the base case reflects polyurethane, with polyol coproducts considered as a sensitivity study).
- *Area 500: Boiler, utilities, and storage.* Biorefinery waste off-gas streams are combined as necessary with natural gas in a gas boiler, which generates low- and high-pressure steam for use in utility heating. Hot oil is also generated from natural gas as needed to satisfy high-temperature heating demands beyond the temperature limits achievable by high-pressure steam. Additionally, this area includes other supporting utility operations including a cooling water system, process water manifold, and power systems. This process area also provides bulk storage for fuels and products, as well as firewater and chemical inputs.

1.3 Techno-Economic Analysis Approach and Assumptions

The engineering approach taken in this work for modeling the conversion process is similar to that described in prior reports, and will not be repeated in the same detail here [7, 10, 11]. In summary, process flow diagrams are initially developed based on inputs and discussions with researchers around pertinent process configurations (i.e., block flow diagram in Figure 1 and more detailed diagram provided in Appendix C) from which process simulations are developed in Aspen Plus [12]. This software computes thermodynamically rigorous material and energy balances for each unit operation in this conceptual biorefinery. The material and energy balance data from the Aspen simulation are used to assist in determining the number and size of capital equipment items. As process conditions and flows change, baseline equipment costs are automatically adjusted in an Excel spreadsheet using a scaling factor. These baseline costs come from vendor quotes (a favored procedure when available for larger or nonstandard unit operations and packaged or skid-mounted subsystems), NREL and subcontractor historical cost data, or published engineering literature (when necessary). Final equipment costs for this report are tabulated in Appendix A.

Once equipment costs are determined, direct and indirect overhead cost factors (e.g., installation costs and project contingency) are applied to determine a feasibility-level estimate of total capital investment (TCI) in 2016 dollars. The TCI, along with the plant operating expenses (also developed using flow rates from the Aspen model), is used in a discounted cash flow rate of return (DCFROR) analysis to determine a plant gate price for total fuel at a given discount rate. The plant gate price, also called the minimum fuel selling price (MFSP), is defined as the sales price (in \$/GGE) required to obtain a net present value (NPV) of zero for a 10% internal rate of return (IRR) after taxes for a hypothetical n^{th} -plant model over a 30-year plant lifetime.

The product of the analysis is a techno-economic model that estimates a rational price for a pre-commercial process. The resultant MFSP is unique for the set of process conditions simulated, and it should be emphasized that some amount of uncertainty always exists around these chosen conditions, as well as around the assumptions made for capital and raw material costs. The underlying cost estimates and TEA modeling approach generally fall within an “engineering feasibility” level of analysis (generally AACE International Class 4 [13]), with an expected uncertainty of approximately $\pm 25\%$ in the TCI estimates [14]. (Sensitivity to this and other parameters are considered in Section 5.1.) Without a detailed understanding of the basis behind it, the computed MFSP carries a risk of being taken out of context. While the MFSP can be used to assess the marketplace competitiveness of a given process, it is best suited for comparing technological variations against one another or for performing sensitivity analyses that indicate where economic or process performance improvements are needed.

The analysis reported here assumes 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 to comment on their comparative economic impact, n^{th} -plant analysis avoids artificial inflation of project costs associated with risk financing, longer startups, equipment over-design, and other costs associated with first-of-a-kind or pioneer plants, lest these

overshadow the economic impact of the technical research advances in conversion or process integration. While these n^{th} -plant economics may not be reflective of pioneer plants and other early technology adopters, they should still help to provide justification and support the underlying technology.

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 an assertion of 90% process uptime (roughly 329 days operating per year). While some industries such as corn wet milling may operate as high as 96% uptime, 90% is viewed here as a reasonable n^{th} -plant target for a relatively complex integrated biorefinery as reflected in this effort. No concessions are included for equipment over-design beyond standard redundancy included for some equipment such as compressors. These assumptions were agreed upon by BETO and reflect standard metrics applied universally across multiple national laboratory TEA models to allow for a consistent basis for analysis. It should be emphasized, however, that these assumptions carry a degree of uncertainty and are subject to refinement.

While Aspen Plus is 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 operations, such as solid-liquid separations (across extraction) and nonthermal purification steps, were modeled as user-defined unit operation modules with a fixed performance based on empirical data or by standard engineering practices. Thermal/catalytic reactors were modeled using stoichiometric reactions of discrete feed and product components rather than rigorous kinetics or rate expressions, which satisfies mass balances and appropriate reactor sizing when coupled with targeted catalyst space velocities or other reactor sizing parameters. Additionally, some unit operations within the polyol/polyurethane coproduct train were handled with less granularity based on overall input/output flow rates and operating conditions furnished by Nexant or otherwise extrapolated from industry literature. The Aspen Plus simulation uses component physical properties internal to the software, including properties estimated by Aspen for custom-defined components not within the existing database.

2 Design Basis and Conventions

2.1 Plant Size and Feedstock Specifications

Similar to prior algae conversion designs [3, 15], the scope of this analysis begins with harvested algal biomass feedstock delivered to the conversion facility after dewatering to a target solids content, thus upstream cultivation/processing logistics are outside the scope of this assessment. (See details documented in NREL's 2016 algae farm design report [4] and more recent 2017 harmonization report [8] for a description of the process reflected in NREL's algal biomass production models and associated TEA results.) In summary, the algae farm is based on 5,000 acres of production pond cultivation area (not including inoculum production), utilizing large 10-acre ponds with minimal requirements for plastic liners aside from targeted areas for erosion control. Algal biomass is continuously harvested at a concentration of 0.5 g/L (0.05 wt% solids ash-free dry weight [AFDW]) and concentrated in the original case to 20 wt% (200 g/L AFDW) across a series of three dewatering steps including settling, membranes, and centrifugation. Cultivation productivity is targeted to achieve an annual average value of 25 g/m²/day, based on a maximum seasonal variability of 3:1 between summer versus winter productivity. Table 1 provides a summary of key cultivation/harvesting parameters implicit in the delivered algal biomass feedstock cost utilized here.

Given the BETO program's recent emphasis on saline strains over freshwater cultivation, the primary difference here relative to the above-cited 2016 algae farm design case framework is a shift in focus to saline cultivation. The primary impact from this adjustment is around the required amount of circulated water that must be removed from the algae farm as blowdown to mitigate salt buildup to intolerable levels. As this is a function of evaporation rates from the ponds, this dependency is highly location-specific, where more arid locations with higher evaporation will require more blowdown and incur higher salt disposal costs than less arid locations. To set seasonal evaporation rates, the above-cited 2017 harmonization report was referenced, running the algae model for the targets noted above through the coastal areas of site groups 5 (eastern Texas and Louisiana) and 7 (Florida) from that study's saline cultivation scenario, which represents the largest collections of identified site location possibilities along the U.S. Gulf Coast. After running both site location scenarios coupled with the algae farm design case framework targets summarized above and in Table 1, the resulting MBSP was taken as the average of both cases, calculated to be \$488/ton AFDW.

From there, the MBSP was slightly adjusted further, to reduce the targeted biomass concentration through final dewatering from 20% to 18%, given drivers described in Section 3.1.2 necessary to accommodate pretreatment heat balances through flash cooling. Relaxing the dewatering targets is only anticipated to save a minimal amount (~\$2/ton biomass costs), as this only impacts third-stage centrifugation costs in concentrating from 13 wt% exiting secondary dewatering (which in turn only represent approximately 1% or roughly \$5/ton of the targeted MBSP when concentrating up to 20 wt% [4]). This brings the MBSP down to **\$486/ton AFDW** as reflected in this study's algae transfer price for the CAP facility.

Table 1. Summary of Key Biomass Cultivation/Logistics Targets in Upstream Algae Farm TEA Model [7, 8]

Parameter	Target
Algae farm size (acres – production ponds excluding inoculum)	5,000
Cultivation annual average productivity (g/m ² /day AFDW)	25
Seasonal productivity (g/m ² /day AFDW – winter : spring : summer : fall)	11.7 : 28.5 : 35.0 : 24.9
Seasonal evaporation rates (cm/day – winter : spring : summer : fall)	0.05 : 0.19 : 0.23 : 0.11
Harvest density (g/L AFDW)	0.5
Dewatering steps @ outlet solids content (g/L AFDW)	Settling @ 10, Membranes @ 130, Centrifuge @ 180 ^a
Salinity levels (mg/L incoming makeup water vs. strain tolerance)	7.7 g/L makeup vs. 40 g/L tolerance
Minimum biomass selling price to CAP facility (\$/ton AFDW)	\$486/ton

^a Originally targeted at 200 g/L in algae farm design case; reduced here to 180 g/L for new CAP targets.

Upon delivery to the CAP facility, the dewatered biomass feed rates fluctuate following the seasonal cultivation productivity variations summarized in Table 1. At a maximum variability of 3:1 between summer versus winter, this would incur large equipment turndown ratios through conversion that in some cases could not be accommodated without utilizing multiple equipment trains at considerable economy-of-scale penalties. Even if such large equipment turndowns could be accommodated, this would require substantial equipment over-design and subsequently wasted capital capacity for off-peak seasons. Accordingly, peak biomass capacity in excess of the annual average delivery rate must be stored and used to supplement low-productivity periods to normalize these fluctuations. Prior algae TEA efforts had assumed the use of seasonal drying and storage (at the expense of natural gas and without supporting data on such an approach) [7], but more recent collaborations with partners at Idaho National Laboratory (INL) have demonstrated better economics and supporting data for wet biomass storage under anaerobic conditions [16]. While a fraction of biomass stored in this way over long periods of time incurs some degradation losses, at a 3:1 seasonal variability only 16% of total annual biomass production must be diverted to storage and subsequently subject to those losses, which translates to less than 4% overall biomass loss based on early experimental work [16] (this has been further reduced in more recent work). The resulting feed rates both “as delivered” and “as processed” through CAP after seasonal storage are shown in Figure 2, with biomass purchase costs incurred based on the former. The resulting rate processed through CAP is 548.2 tons/day AFDW across all seasons. More details for this operation are discussed in Section 3.1.

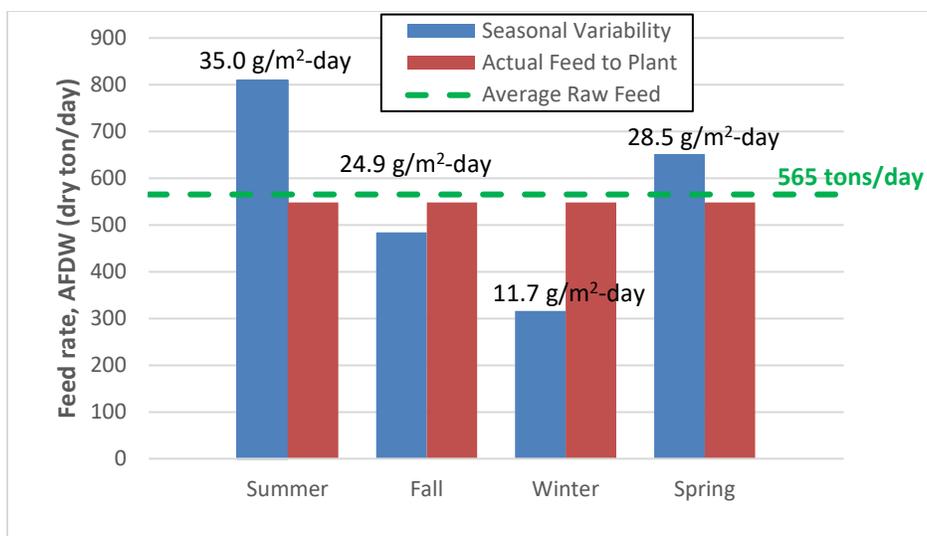


Figure 2. Seasonal and annual average feed rates to conversion facility. Full seasonal variability reflecting as-delivered biomass rates [blue bar] and as-processed through CAP [red bar] after excess peak capacity storage for use in low-production seasons. Average of delivered rates exceeds the rate processed through CAP due to degradation losses of seasonally stored biomass. Data labels indicate corresponding seasonal productivity from upstream cultivation.

The targeted algal biomass composition delivered from the upstream algae farm is maintained consistently with prior recent NREL TEA efforts [4, 8], reflecting a compositional profile exemplified by *Scenedesmus acutus* (LRB-AP 0401) based on biomass cultivated by Arizona State University. Specifically, the base case CAP model employed here reflects measured compositional data attributed to an early stage of nutrient deprivation for this species in moving between high-protein nutrient replete to low-protein/high-lipid nutrient deplete harvesting, historically termed “high-carbohydrate *Scenedesmus*” or HCSD [4, 17]. This composition was maintained here for two primary reasons:

1. While the new CAP process configuration reflected in this study is anticipated to accommodate a higher-protein biomass feedstock with minimal process modifications (as will be of increasing focus in future CAP work), this process has not yet been demonstrated experimentally for high-protein biomass. Thus, for this initial study, we opted to maintain consistency with prior TEA efforts focused on a lower-protein, mid-lipid, high-carbohydrate biomass profile as a reasonable target that still may be achievable by 2030.
2. For such a compositional profile, the HCSD basis represents the highest level of detailed granularity from NREL analytical data, and thus the lowest discrepancy from 100% mass closure, relative to other high-carbohydrate species characterized by NREL to date. This translates to lower uncertainties introduced by manual compositional adjustments to achieve 100% mass closures as required for modeling.

The compositional details both as-measured and after 100% closure as-modeled for this HCSD basis, as well as earlier and later stages of nutrient depletion for this species, are presented in Table 2 (see Davis et al. 2016 [4] for further details and context). While it is acknowledged that

the original compositional data for this *Scenedesmus* strain was based on freshwater cultivation and this study asserts use of non-freshwater algae, more recent cultivation trials under various NREL-ASU collaborations (including the recently-formed DISCOVER consortium [18]) have demonstrated the ability for numerous strains of *Scenedesmus* as well as closely-related *Desmodesmus* to be tolerant to at least brackish salinity conditions (at least 5 mg/L salinity tolerance or more), both of which are known to exhibit rapid shifts in carbohydrate accumulation upon early nutrient deprivation. Additionally, we emphasize that this does not imply a target scenario that must utilize *Scenedesmus*, but rather one that matches the *compositional profile* consistent with this example HCSD basis while also achieving the targeted annual productivity of 25 g/m²/day. Moving forward, future work is planned to focus on conceptual TEA modeling based alternatively on high-protein algal biomass, highlighting similarities and differences from the process described here for high-carbohydrate biomass conversion.

Table 2. Algal Biomass Elemental and Component Compositions for Early-Harvest (HPSD), Mid-Harvest (HCSD), and Late-Harvest (HLSD) *Scenedesmus*. (Left portion = raw analytical data, right portion = adjusted values to achieve 100% mass closure for modeling) [17, 19].)

<i>Scenedesmus</i> sp.	Measured:			Mass Balance Closure for Model:		
	Early (HPSD)	Mid (HCSD)	Late (HLSD)	Early (HPSD)	Mid (HCSD)	Late (HLSD)
Wt% composition (dry basis)						
C	49.3	52.9	56.3	52.2	54.0	57.4
H	7.1	8.0	8.5	7.5	8.2	8.7
N	8.3	1.8	1.6	8.8	1.8	1.6
O	28.7	34.8	31.2	30.4	35.5	31.8
S	0.0	0.2	0.2	0.0	0.2	0.2
P	0.97	0.22	0.20	1.03	0.22	0.20
Total ^a	94.4	97.9	98.0	100.0	100.0	100.0
Ash	6.7	2.3	2.1	6.7	2.4	2.1
Fermentable carbohydrates	24.3	46.3	37.9	26.2	47.8	38.2
Other carbohydrates ^b	3.4	1.6	1.3	12.8	5.0	3.9
Protein	34.5	12.8	8.9	43.2	13.2	9.0
Lipids (measured as FAME) ^c	6.6	26.5	40.9	8.3	27.4	41.2
Non-FAME lipid impurities	ND ^d	ND	ND	0.8	2.7	2.1
Cell mass	ND	ND	ND	1.9	1.6	3.5
Total	75.5	89.5	91.1	100.0	100.0	100.0

^a Measured element balance on dry weight basis; adjusted to AFDW basis for model (right portion of table).

^b Nonfermentable carbohydrates, but still convertible through MOT under new CAP pathway evaluated here.

^c Lipids measured as fatty acid methyl ester (FAME) per standard NREL analytical methods [20].

^d ND = not determined.

One key additional detail in the compositional profile that had not been as important in prior CAP models is the ratio of TAGs versus FFAs in the FAME-equivalent lipid fraction reflected in Table 2. As the new CAP configuration now focuses on splitting lipids to fuels and polyols/polyurethanes based on distilling FFA from TAG (see Section 3.2), the TAG:FFA ratio becomes an important factor in dictating yields of each product. Given that this ratio can be highly variable by strain and time of harvesting, in the present work it is evaluated over a range to evaluate overall impacts to the TEA outputs.

2.2 Updated Modeling Basis

In keeping with prior TEA report updates, several underlying model parameters are maintained consistently in the present work. These include a 0.5-year facility startup time and 90% process uptime (7,920 hours per year on-stream factor). It should be emphasized that these are reflective of the n^{th} -plant modeling basis employed here, as discussed above, and early pioneer plants may require significantly longer startup times or sustain shorter annual uptimes. Other conventions such as reported units and physical quantities, like total solids (TS) and insoluble solids (IS), are consistent with details described previously [21, 22]. All cost results are presented here in 2016 dollars.

The results from this analysis are reported primarily in terms of energy yields in gallons gasoline equivalent, e.g., \$/GGE, GGE/yr, and GGE/ton. This includes energy yield in both the diesel and naphtha product cuts from hydrotreating distillation, rather than yield of a single product with the other treated as a coproduct. This is done to maintain focus on total fuel (energy) yield from biomass carbon, to avoid introducing unnecessary subjectivity in decisions for distillation column product cuts, or over-interpreting the current projections for catalytic upgrading steps with regards to catalyst selectivity to diesel versus naphtha. Moreover, the primary product from upgrading of algal lipids is diesel (with some naphtha), while the current assertions for MOT and catalytic upgrading of carbohydrates and protein is in the gasoline range, although with an additional condensation step to further expand the chain length of ketones (see Section 3.1), this could be shifted into the diesel range. Lower heating values (LHVs) for both diesel- and naphtha-range products were calculated by the Aspen model, and correspond to similar values for standard petroleum-equivalent products [23]. To translate to a GGE basis, a conventional gasoline heating value of 116,090 BTU/gal (LHV basis) was applied [23]. All coproducts (polyols and polyurethanes) are reported on a mass basis, using mass-based product values to avoid ambiguity with other industry units such as board-foot.

3 Process Design and Cost Estimation Details

3.1 Area 100: Seasonal Storage and Pretreatment

3.1.1 Overview

The process begins with biomass delivered to the CAP conversion facility from upstream algae cultivation and dewatering (co-located at the same site). As discussed previously, the majority of the targeted cultivation and dewatering performance parameters implicit in the delivered MBSP remain consistent with NREL's 2016 algae farm design report, with the exception of added blowdown/salt disposal costs reflective of non-freshwater production, as well as associated evaporation rates tied to average Gulf Coast region figures [4, 8]. Additionally, final-stage dewatering (centrifugation) is slightly relaxed from a prior target of 20 wt% solids to 18 wt% more suitable for this CAP schematic. With a targeted seasonal productivity variation up to 3:1 between summer versus winter growth rates and associated biomass production outputs, seasonal swings in biomass throughput to the CAP facility must be mitigated to avoid untenable equipment turndown ratios or wasted capital capacity for processing maximum peak flows. As introduced in Section 2.1, rather than relying on diversion of excess peak capacity to a costly and energy-intensive drying step for dry biomass storage (without supporting data verifying stored biomass integrity) as was assumed in prior studies, this work makes use of a wet, anaerobic storage operation based on inputs from partners at INL [16]. As depicted in Figure 3, biomass from upstream dewatering and short-term surge storage is diverted as required to a low-cost in-ground pit with a liner and cover to reasonably maintain anaerobic conditions and prevent soil percolation. Optionally, acid may be added or an organic acid-producing organism could be inoculated as a means to prevent degradation similar to terrestrial biomass ensiling, but early published data did not observe discernable advantages for such an approach [16]. Stored biomass is then blended with fresh harvested biomass during lower-production seasons to normalize the flow through the biorefinery. Some degradation of stored biomass is incurred along with a compositional shift, but is a relatively low fraction of total annual biomass production.

The biomass slurry (now at a 1:1 throughput across all seasons) is sent to flash hydrolysis. In contrast to NREL's prior CAP models focused on dilute acid pretreatment (required if utilizing downstream fermentation of monomeric sugars), flash hydrolysis as employed here exhibits cost advantages including no acid or caustic neutralization chemicals (or subsequent salt formation), lower-cost metallurgy (stainless steel rather than Incoloy-clad metallurgy), and a simpler and much smaller reactor design (pressure vessel without any moving parts, with a residence time on the order of 10 seconds rather than 5–10 minutes). However, the reaction requires more severe operating conditions and also necessitates rapid heating and cooling before and after the reaction to avoid degradation products, raising important heat exchange design considerations. Although carbohydrate hydrolysis to sugars is low with this operation, this is acceptable as monomeric sugars are no longer required in this process; additionally, this still is anticipated to enable high lipid extraction yields (which remains a key step for this process) [24, 25]. In addition to cell disruption for downstream lipid extraction, the hydrolysis solubilizes a large fraction of protein solids and a smaller portion of carbohydrates. The pretreated hydrolysate is flash cooled by dropping the pressure and driving off steam, then combined with solvent for downstream extraction and flashed again, vaporizing solvent to further cool the hydrolysate. The vaporized solvent is condensed, combined with the flash liquid, and routed to extraction.

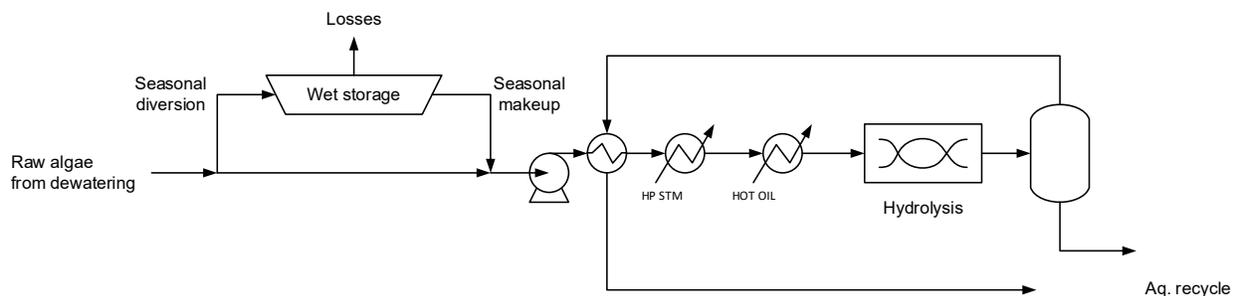


Figure 3. Simplified flow diagram of seasonal biomass storage and flash hydrolysis steps

3.1.2 Design Basis

The first step in the conversion process is to normalize the seasonal delivered algae flow from upstream cultivation by either diverting dewatered algae to storage, or drawing algae from storage, according to the flow rates presented in Figure 2. It is important to re-emphasize that dewatered algae is not equivalent to dried algae. The dewatered algae has passed through settling tanks, centrifuges, and membranes to remove enough water to achieve a solids content of 18%. This is lower than the typical 20% solids, which enables multiple economic benefits for the process. First, the higher water content allows for more heat recovery in the process-to-process heat exchanger that heats the reactor feed. Water is flashed off after the reaction to a 20% solids level and is then cross-exchanged with the feed, so having more water in the feed allows for more water to flash and more heat recovery. Of course, there is also a higher heating requirement for the feed, but there is a net benefit relative to a 20% solids feed basis. In addition to this, because the dewatering requirement is lower, the biomass would be available at a marginally lower price, as discussed in Section 2.1.

The dried algae distinction is generally reserved for algae processed through heated dryers such as rotary drum dryers. While dried algae storage had previously been assumed as the means for mitigating seasonal variability in prior NREL TEA reports [7], the present work reflects a more cost-effective wet storage process based on collaborations with INL. Excess algae from high-productivity months, such as late spring, all summer, and early fall, is diverted to covered wet storage pits. During storage, protein and lipid content of the algae are generally unaffected while carbohydrate content is reduced in evolving storage degradation products, primarily CO₂. The change in algae composition across the seasonal storage operation is presented in Table 3. A detailed explanation of the wet storage operation/conditions is beyond the scope of this report, however results of wet storage experimental testing and TEA implications can be found in Wendt et al. 2019 [16]. In summary, relative to a base case viewed as most realistic for seasonal drying and storage at an MFSP of \$3.72/GGE, an alternative for the most realistic wet storage option was found in that study to reduce the MFSP to \$3.40/GGE (over \$0.30/GGE reduction).

Table 3. Input Compositions to TEA Model Based on Delivered (Raw) and Wet Storage Composition [8]

	Raw Algae	Wet Storage Algae
Algae composition (wt%)		
Protein	13.2	14.2
Free fatty acids	26.0	27.5
Ash	2.4	3.1
Fermentable carbohydrates	47.8	46.2
Nonfermentable carbohydrates	3.2	1.7
Glycerol	3.0	3.0
Sterol	1.8	1.8
Nonfuel polar lipid impurities	1.0	1.0
Cell mass	1.6	1.6
Sum	100.0	100.0
Storage losses		22.8%
Acid produced per kg of whole algae (after storage)		
Succinic acid, kg		0.090
Lactic acid, kg		0.083

The normalized algae flow rate entering the conversion facility after storage adjustments is 548.2 tons/day AFDW. The first major unit operation in the conversion process is the flash hydrolysis reactor, which operates at 280°C and 1,200 psig (82.6 atm). To achieve these conditions, the algae feed stream is first pumped to pressure, then heated in three heat exchangers. The first heat exchanger is a feed-effluent heat exchanger that heats the algae to about 110°C using flash vapor (almost entirely steam) from a flash tank following the hydrolysis reactor. This is followed by a second heat exchanger, which uses high-pressure steam to increase the temperature up to the 260°C, and finally a hot oil heat exchanger to reach 280°C operating temperature of the flash hydrolysis reactor.

In the flash hydrolysis reactor, only 10% of carbohydrates are converted to sugars, but more importantly about two-thirds of the protein content is solubilized [26], and the algae cells are disrupted, making the biomass amenable to downstream lipid extraction. In order to minimize degradation products, the flash hydrolysis reactor residence time is very short, approximately 10 seconds [24, 25], therefore the hydrolysate is immediately and rapidly cooled in a flash tank, which reduces the pressure to 560 psig (39.2 atm). The flash vapor (steam) is used to heat the incoming algae in the feed-effluent heat exchanger. The hydrolysate is sent to lipid extraction, where a second flash occurs to bring the temperature down further after combining with hexane solvent (covered in Area 200). Under this approach, the critical rapid quench step is done entirely without heat exchangers but rather through evaporative cooling across flash let-down steps.

3.1.3 Cost Estimation

Capital costs for unit operations in the wet storage and flash hydrolysis areas were estimated by adapting capital costs reported in previous NREL publications. The construction costs for storage pits as well as the purchased cost of liners and covers for the pits were consistent with Wendt et al. 2019 [16]. The storage pits cost \$34,300 (2014 dollars) per 1,000 m³ to construct, and pond liners cost \$1.00 per ft² (2014 dollars). All storage costs were scaled to accommodate the required storage volume, and costs were adjusted to 2016 dollars. The original basis of the wet storage pit capital costs was leveraged from in-ground settling ponds in the 2015 algae farm model design report [4].

Capital costs for the preheat and trim heater exchangers were estimated by adjusting the size of shell and tube heat exchangers costed in previous TEA reports. The source of the original costs estimates was generally from Aspen Capital Cost Estimator (ACCE) [27]. The flash vessel following the hydrolysis reactor was costed in the same manner: scaling a flash vessel costed in a previous NREL techno-economic analysis with the original capital cost from ACCE.

As stated before, residence time in the flash hydrolysis reactor is 10 seconds, therefore the required volume of the flash hydrolysis reactor is small, only 0.38 m³ (380 L). As a result, only one reactor is required. ACCE generated a purchased cost of \$10,400. At this purchase price, standard installation factors would result in an installed cost of \$20,000 to \$30,000, which appears unreasonably low. Due to its small size the flash hydrolysis reactor incurs very low material requirements. As a result, installation costs are most likely a significant percentage of the installed costs. Therefore, the installation factor was assumed to be further increased for the flash hydrolysis reactor, resulting in a more conservative installed equipment cost of \$98,776.

Due to uncertainty in how algae will be transferred from the wet storage pits, the power requirement for the wet storage and pretreatment area was approximated by assuming the power required will be half that of the power required for the more complex pretreatment process presented in the 2011 NREL cellulosic ethanol design report [10]. The power required for the pretreatment area in the cellulosic ethanol design was 5.7 MW, which was halved to account for the reduced complexity of the process presented here without moving parts aside from pumps to move the stored material in and out of the storage pits. Additionally, the halved power requirement was further adjusted downward to account for the lower total solids flow rate in this process versus the cellulosic ethanol design. The final estimated power requirement for the storage and flash hydrolysis area is 0.7 MW.

3.2 Area 200: Lipid Extraction, Purification, and Separation

3.2.1 Overview

As noted above, the pretreatment hydrolysate is flash cooled first by driving off steam at reduced pressure, then by adding recirculated (condensed) hexane solvent and further dropping the pressure, which vaporizes hexane and further cools the hydrolysate, then by re-condensing the vaporized hexane and mixing with the liquid hydrolysate. This avoids the need for any physical heat exchangers to satisfy the rapid cooling demands of the flash hydrolysis unit, as well as any potential fouling or viscosity issues that would further complicate heat exchanger operation. The cooled slurry mixed with hexane is then combined with ethanol as a co-solvent and routed through a series of three agitation/phase separation steps to extract algal lipids. This operation

has been extensively carried out at NREL, achieving over 90%–96% lipid yields in high purities, with a target of 96% extraction yield set here. This approach employing dual solvents across a multistage agitation/separation process represents an improvement over prior NREL models that had assumed the use of a dedicated liquid-liquid extraction (LLE) column, which was both costly and riskier without supporting data at scale and with a higher risk for forming emulsions. The nonpolar extract phase exiting each separation step is routed to a hexane stripping column, where hexane is recovered and recycled back to flash cooling upstream (over 99.7% recovery). The polar raffinate phase is routed to a second column to recover and recycle ethanol.

The recovered lipids are then purified through a series of physical refining steps, namely degumming and bleaching, to remove impurities such as gums (including phospholipids) and metals following such steps commonly employed in the biodiesel industry. The purified lipids are then fractionated into lighter-boiling FFAs versus higher-boiling TAGs in a vacuum distillation column. The column employs steam-stripping, which helps to both reduce the column operating temperatures and also provides further purification for the TAG fraction, destined for polyol/polyurethane coproduction (vacuum steam stripping is another common operation in biodiesel refining for purifying TAGs). The FFA fraction is routed to hydrotreating for upgrading to fuels.

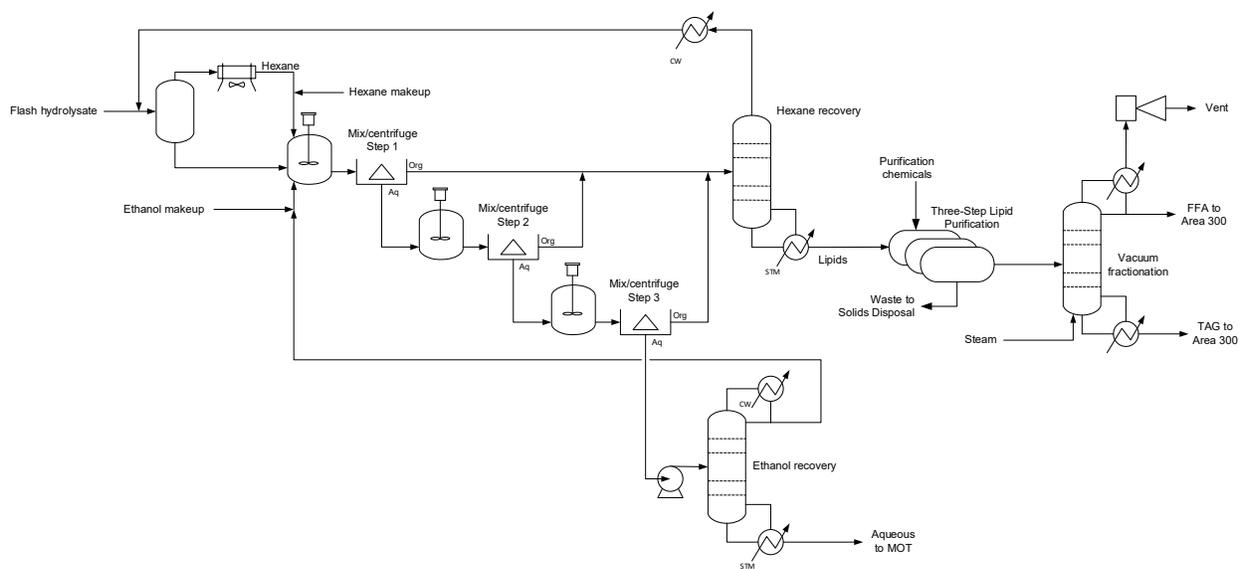


Figure 4. Simplified flow diagram of the lipid extraction, purification, and separation process

3.2.2 Design Basis

After flashing to 20 wt% solids, the pretreated hydrolysate enters Area 200 at roughly 250°C. The material is then mixed with the condensed recycle plus makeup hexane solvent required for the initial extraction step, and flashed to let the pressure down from 40 to 3 atm. This flash vaporizes hexane and water, reducing the temperature further to 120°C. The vaporized stream is re-condensed with cooling water, then combined with the flash liquid and the co-solvent stream (ethanol with residual water), further cooling the combined material to the requisite 55°C for extraction. As noted above, the prior 2014 CAP design report had originally assumed the use of an agitated LLE column for the lipid extraction step, which was costly and limited to fairly

small capacities, requiring 16 parallel columns [7]. Still, that unit was selected at the time based on guidance from an engineering subcontractor and vendor who were concerned about the potential for forming emulsions, with that particular reciprocating LLE column quoted as a suitable solution. The prior LLE column design was based on asserted lipid extraction performance targets, but without supporting experimental data for such a design.

Instead, the present work utilizes an improved lipid extraction approach which is both lower-cost and reflective of supporting NREL experimental data. In this approach, a nonpolar solvent (hexane or similar light hydrocarbon) is fed with a polar solvent (ethanol) through an agitated vessel (CSTR) followed by a phase-separation centrifuge, repeated three times in series. The extract phase from each step is routed to a solvent recovery stripping column, where the nonpolar solvent is stripped from the extracted lipids and recycled for subsequent use and the bottoms lipid product is routed to lipid cleanup and fractionation. The aqueous raffinate product from phase separation is then routed to the next CSTR extraction step along with makeup nonpolar solvent (the ethanol co-solvent remains with the aqueous phase across each step). After the third extraction/solvent recovery, the aqueous stream is routed to a second distillation column to recover and recycle ethanol, leaving water and lipid-extracted biomass to be routed to MOT. This approach for lipid extraction carries several advantages over the prior LLE basis, including (1) simpler and lower-cost unit operations which are more scalable to large commercial scales, (2) the use of ethanol co-solvent mitigates concerns about emulsion formation, as ethanol is known to break emulsions, and (3) the inclusion of a second solvent recovery distillation column recovers nearly 100% of both ethanol and any light hydrocarbon solvent carried over into the raffinate, thus reducing makeup costs for either solvent.

Each CSTR extraction step operates at 55°C for 15 minutes, utilizing a solvent ratio of 2.8 nonpolar to 1.1 ethanol to 1 biomass (dry weight, g/g/g). The operation is targeted to achieve an overall net lipid extraction yield of 96% (FAME lipids), with a FAME lipid purity of at least 94% (the remainder being sterols and other polar or non-FAME lipids, although in some strains a lower purity has been observed between 85-90%) [8, 9]. Table 4 provides a summary of key details for this operation. Recent NREL experimental work has demonstrated achievement of these targets across the use of three separate approaches for sourcing the “nonpolar” solvent, namely hexane, light naphtha, and heavy naphtha (unpublished data), where the use of naphtha could potentially offer further benefits by recycling a portion of the final light hydrocarbon product after lipid hydrotreating as its own solvent (eliminating the need for externally purchased makeup hexane). The resulting experimental data indicated nearly identical overall lipid extraction yields across all three nonpolar solvent options with inclusion of the ethanol co-solvent, achieving 95%–96% total FAME lipid recovery. The light naphtha (consisting of a roughly 50/50 split of C₅ and C₆ hydrocarbons) is intended to be recovered and recycled prior to re-entering hydrotreating, similar to the hexane configuration as shown in Figure 4. Alternatively, the heavy naphtha fraction (consisting of C₇–C₉ hydrocarbons) would instead allow for eliminating the nonpolar solvent recovery column and routing the full extract phase straight through hydrotreating. However, TEA modeling found that the cost savings in the latter case would be outweighed by significantly higher cost penalties for increasing the circulation rate through hydrotreating via heavy naphtha recycle, thus this option was not pursued further. This leaves hexane or light naphtha, which lead to comparable MFSP results in either case, given that hexane/nonpolar solvent losses are minimal under this new two-solvent system approach (i.e., any carryover losses of the solvent into the aqueous raffinate phase are ultimately recovered

and recycled with the addition of the ethanol recovery column). For sake of simplicity, hexane is assumed in this work.

Table 4. Lipid Extraction Targets (based on experimentally demonstrated data [8, 9])

Parameter	Value
Extraction configuration	3-stage CSTR + centrifugation with 2 solvents
Solvent loading (nonpolar: EtOH: dry biomass, wt)	2.7:1.1:1, g/g/g
CSTR extraction residence time	15 min
Temperature	55°C
FAME lipid extraction yield per stage ^a	74% - 65% - 56%
Total FAME lipid extraction yield ^a	96%
Nonsterol lipid impurity partition to extract	<11.5%

^a FAME lipids as defined in [20]. Extraction yields per stage reflect data for individual CSTR steps (3 in series).

The recovered lipids from the solvent stripper column bottoms are purified through lipid cleanup steps consisting of degumming and bleaching, followed by FFA distillation from TAG in a steam-stripped vacuum column (see Table 5). These operations largely draw from the biodiesel industry which commonly makes use of such steps through chemical/physical refining operations to purify biodiesel oil feedstocks. The use of these operations was suggested by Nexant, primarily for purposes of TAG cleanup prior to polyol/polyurethane upgrading. Given that such operations were also included in NREL's prior CAP design report [7] based on detailed inputs from Harris Group, design/cost details for the degumming and bleaching steps were largely maintained consistent with that prior work. In summary, gums (including phospholipids) are removed in a degumming step with the addition of phosphoric acid followed by centrifugation with wash water. This is followed by bleaching with clay to remove metals and impurities, which involves forming a slurry with clay followed by filtration. The vendors Harris Group had originally contacted in that prior CAP design case noted that the phospholipid content in the product oil was expected to be less than 10 ppm, possibly as low as 2 to 3 ppm, based on an analytical characterization of the extracted oil furnished to the vendor (based on a similar *Scenedesmus* strain as still assumed here). Additionally, they expected metals content to be less than 5 ppm for potassium, calcium, magnesium, and sodium. Although previous NREL research did not observe significant catalyst issues tied to lipid impurities when focused on exclusive HDO upgrading of lipids, more recent efforts have shown higher sensitivity when focused on both HDO and HI catalyst activity (unpublished data), with increased phosphoric acid dosing required at 1 wt% of feed to sufficiently remove chlorophyll and phospholipids (thus this parameter was increased relative to prior Harris Group inputs of 0.2%). The process takes place at 80°C (176°F), with the incoming lipid material first cooled to this temperature with cooling water after exiting the solvent recovery column.

As Nexant also noted, when treating vegetable oils an additional neutralization step is typically performed during TAG cleanup wherein caustic is added to remove FFAs. Given that in this process FFAs represent a significant fraction of the extracted lipids and are valuable feedstocks for conversion to fuels, FFA neutralization is not included in this work. Finally, the purified lipid material is sent to a steam-stripped distillation column to fractionate FFAs from TAGs. The column operates at 0.1 atm and a steam-to-bottoms ratio of 0.82 lb/lb. This operation achieves 99% FFA recovery in the distillate at close to 100% purity and 100% TAG recovery in the bottoms at 99% purity. As an alternative to steam stripping, the natural gas supplied to the boiler may also serve as the stripping agent, as is employed in the chemical industry. However, as the steam required for stripping is not substantial (1.2% of total facility steam demand), this basis

was employed here. A steam ejector draws a vacuum on the column, incurring further steam demands from the boiler. While the degumming/bleaching cleanup steps were conservatively placed before the FFA/TAG stripping column, as the lipid impurities would largely remain with the TAG fraction from this column (leaving relatively pure FFA in the overhead), those cleanup steps could alternatively be placed downstream on the TAG bottoms product line, which would save on costs by processing a smaller throughput based on TAG alone. Another potential impurity that could cause problems is organic chlorine components if present at elevated levels in the extracted lipids, which could cascade into downstream polyol/polyurethane processing challenges, though this is not presently known.

Table 5. Operating Conditions and Material Requirements for Lipid Purification (based on guidance from engineering subcontractor; 1% phosphoric acid dosing based on experimental data)

Degumming Step	
Total lipid feed rate – design basis	6,100 kg/hr (max case)
Feed temperature	80°C (176°F)
Phosphoric acid dosing (wt% of feed)	1.0%
Bleaching Step	
Clay dosing (wt% of feed)	0.2%
Purified lipid product rate – design basis	5,730 kg/hr (max case)
FFA Stripping Column	
Top stage pressure	0.1 atm
Steam:bottoms ratio (wt)	0.82

3.2.3 Cost Estimation

All major equipment items as utilized here for lipid extraction were based on prior vendor inputs documented in previous NREL TEA reports. Cost estimates for lipid extraction CSTR vessels and agitators are scaled from NREL’s 2011 ethanol design case for comparably small-sized CSTR vessels, originally quoted by a vendor as 2,000-gallon skid package vessels that were employed for use in an ethanol fermentation seed train [10]. The subsequent phase separation centrifuges were also based on previously furnished vendor quotations, as documented in NREL’s 2013 biochemical design report [22]. The hexane/nonpolar solvent recovery columns and reboilers were based on inputs supplied by Harris Group (utilizing three each), as established in NREL’s 2014 CAP design report [7]. The solvent recovery column is a packed tower 10 feet in diameter and 20 feet high, utilizing 304SS metallurgy. The nonpolar (ethanol) solvent recovery column is based on NREL’s 2011 ethanol design report, again based on a prior vendor quote provided via Harris Group (but reducing the resulting cost in this case as the original cost basis was provided as a package for both the beer and rectification columns, where only the former is needed here). Taking these capital costs into account in addition to the relevant operating costs, the total cost of extraction is estimated to be approximately \$59/ton of extracted lipids.

As noted above, the bleaching and degumming cost estimates were originally furnished by vendors via Harris Group in support of the 2014 CAP design case, and translate here to an installed cost of ~\$3.1 MM. A marginal power demand of roughly 0.01 KWh/kg oil feed rate was applied, also per vendor feedback. Variable operating costs for the extraction makeup solvents and purification chemicals (and all chemicals for the facility) are discussed in Section

4.2. Finally, costing for the vacuum FFA stripping column was done in ACCE, based on stainless steel metallurgy utilizing 10 trays for a total vessel height of 6.1 m with an internal diameter of 3.7 m (based on Aspen Plus column sizing at the modeled conditions). Likewise, the vacuum steam ejector was also costed in ACCE. Cost estimates for all heat exchangers in this process area were scaled from prior NREL models for a similar service unit.

3.3 Area 300: Mild Oxidative Treatment and Fuel Catalytic Upgrading

3.3.1 Overview

Area 300 contains several core unit operations involved in the fuel train, depicted in Figure 5. First, the bottoms stillage stream exiting the ethanol recovery column on the extraction raffinate product is routed to MOT. For this operation, the material is pumped back up to a milder pressure than flash hydrolysis and fed to a bubble column-type pressurized reactor. Oxygen is sparged into the reactor and is consumed along with carbohydrates and protein (both soluble and insoluble components from upstream flash hydrolysis), yielding mixed carboxylic acids. In place of pure oxygen, air or oxygen-enriched air could possibly be utilized, albeit requiring higher system pressures to maintain suitable oxygen partial pressures (this will be a key area for future investigation/optimization and cost tradeoff analysis). Acknowledging that this is one process step for which the TEA modeling is currently in front of the experimental/analytical work, several “representative” acid components are presently reflected as generally understood to constitute the key acid products; these include formic, acetic, and succinic acids. All three of these acids have been experimentally observed to constitute the major products of MOT to date (unpublished data). Other ketonizable acids have been observed in smaller amounts, represented by succinic acid for modeling purposes. Additionally, most of the nitrogen and phosphorous are liberated as ammonia and phosphate, respectively, and are expected to be available for capture and recycle back to the cultivation ponds (this has not yet been experimentally demonstrated, but anticipated to be achievable given appropriate ion exchange capture performance). CO₂ is also formed and is recycled from the vent stream. The product from MOT is cooled, filtered of ash and other residual solids, and sent through ion exchange to capture the liberated N/P species.

The material is then routed to catalytic ketonization, employing a fixed heterogeneous catalysis reactor to upgrade acids to ketones of larger chain length through coupling reactions. In principle, this operation is similar conceptually to NREL’s biochemical design case employing ketonization of butyric acid, but in this case it is an aqueous reaction in the presence of 89% water, and converts a larger range of acids to ketones. These ketone products are represented in the model by acetone (a primary product of acetic acid condensation) and 3-methyl-2-cyclopenten-1-one, chosen to represent a broader spectrum of C₆-C₇ cyclic ketones. These representative products were again selected based on preliminary experimental data; however, this is another step requiring further understanding in the context of upstream MOT reactions. The product stream from ketonization is then sent to a two-stage flash distillation to separate the CO₂ produced from the reactions. The resulting liquid is then separated further into acetone and a mixture of water and the C₆ cyclic ketone via distillation. The aqueous mixture is decanted to yield a relatively pure water stream, which is recycled to the algae ponds, and an organic stream. The organic stream, along with the acetone from the distillation, is sent to condensation.

In condensation, the ketones are reacted in a batch slurry reactor in the presence of a toluene solvent, producing cyclic enones (represented by aristolone and isophorone) and water. Given the above-noted uncertainties around representative compounds and their relative yields across MOT and ketonization, such uncertainties further cascade into condensation. The condensation products are thus also somewhat speculative, with the speciation based on early experimental observations with reasonable assumptions around balancing stoichiometry. The enone products are separated from the solvent and unreacted ketones via distillation and sent to hydrotreating, with the solvent and unreacted ketones recycled back to condensation. The condensation products, alongside recovered FFA lipids from Area 200, are reacted with hydrogen in hydrotreating, yielding cyclic hydrocarbons in the jet/diesel range from enones and straight/isomerized products from lipids largely in the diesel range. The HDO operation employs a single-step hydroprocessing reactor with a catalyst to deoxygenate and saturate the intermediates, based on prior NREL experimental work for this approach historically focused on algal lipids. The resulting hydrocarbon blend-stock products are fractionated in a distillation column into diesel and naphtha cuts, with off-gas also removed and routed to the boiler.

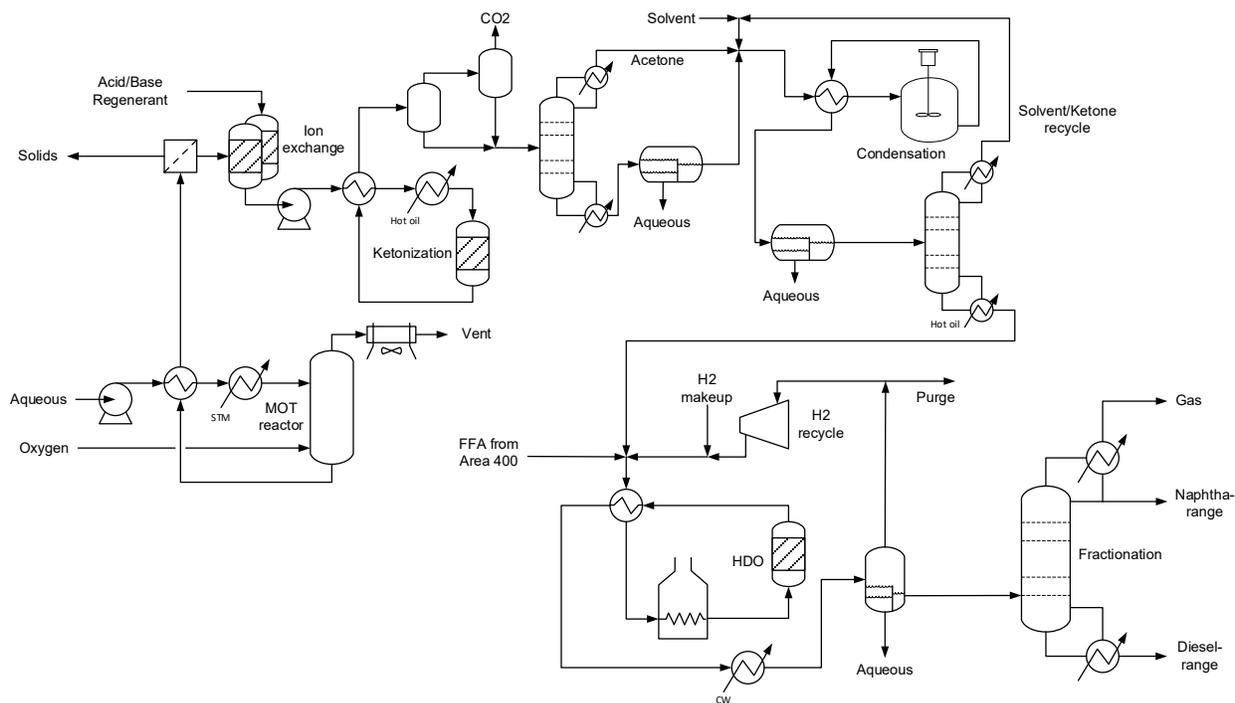


Figure 5. Simplified flow diagram of MOT and fuel catalytic upgrading

3.3.2 Design Basis

Mild Oxidative Treatment

The aqueous raffinate stream from extraction contains protein (existing as both solids and solubilized peptides), carbohydrates, sugars, generic algal cell matter, and ash, as well as a small fraction of unrecovered ethanol, hexane, and lipids. This raffinate stream is pumped up to 35 atm and heated to 225°C using a process-to-process heat exchanger with the MOT effluent stream. Pure oxygen is compressed and fed into a bubble column-type reactor. It is estimated that the MOT reactor will achieve close to 100% O₂ solubilization and subsequent utilization at these

elevated pressure conditions; therefore, O₂ is fed into the reactor at minimal stoichiometric excess (5%) [28]. This currently represents a significant area of uncertainty and need for further experimental understanding/optimization. Preliminary proof of concept experimental indications have shown promising MOT yields but so far achieved under low substrate concentrations (less than 2 wt% solids), with yields dropping off as solids content increases albeit partially offset by better yields as oxygen partial pressure increases (unpublished data). This may indicate a linkage to adequate oxygen dispersion and access to the substrate as a key factor dictating achievable MOT performance. Moving forward, further experimental and TEA work will be conducted to better understand these issues and opportunities to optimize between MOT substrate concentration, reactor pressure, and oxygen delivery ratio; or alternatively given high costs for sourcing pure oxygen from an air separation unit, also understanding opportunities for using air (no cost for oxygen but high cost for compression and more elevated reactor pressures) or oxygen-enriched air (falling between those bounds). Alternatively, other oxidants will also be explored that may offer benefits such as peroxide or ozone.

Further confounding oxygen dispersion and associated reactor design challenges, if processing a high-salinity algal substrate, a brine solution and a vapor phase could be present in the MOT reactor, which could be both corrosive and problematic for oxygen dissolution. Other corrosive components could also be evolved from the MOT process, such as sulfuric or hydrochloric acids. Further analytical characterization will be required to confirm the presence of such species, and accordingly further assessment for reactor design/metallurgy considerations.

Protein is modeled as a representative compound with the formula of C_{5.17}H_{9.86}N_{1.30}O_{2.57}S_{0.04}, as based on a weight-averaged amino acid distribution from Templeton and Laurens [29]. In MOT, the understood reaction mechanism for protein starts with hydrolyzing the various peptides into their constituent amino acids. Following this, it is understood that the peptides undergo decarboxylation on carboxyl groups adjacent to amine groups. Next, those amine groups are then deaminated, converting the adjacent carbon into a carboxyl group. In the interest of a manageable model, these reactions were simplified to satisfy a stoichiometric balance with product acid ratios consistent with preliminary data. The representative product compounds selected were formic, acetic, and succinic acid, with the former two compounds reflecting the majority acids observed to date experimentally, and succinic acid representing a broader range of similar acids (primarily succinic, but also glycolic, methylsuccinic, fumaric, maleic, levulinic, butyric, isobutyric, valeric, and isovaleric acids). Relative carbon yields of each acid are shown in Table 6 along with relevant reaction conditions. The three representative acids are produced in relative proportion consistent with early experimental data, albeit asserting higher overall carbon yields of 80% to be achieved in the future versus 63% experimental baseline (achieved under the same reaction conditions but more dilute MOT substrates to date).

Table 6. Carbon yields and reaction conditions for MOT (product C yields based on aspirational targets; acid distributions consistent with early experimental data conducted at similar conditions but more dilute substrate concentrations)

<i>Parameter</i>	<i>Model Target</i>
Carbon Yield	
Formic Acid	8.7%
Acetic Acid	34.0%
Other Ketonizable Acid ^a	37.3%
CO ₂	20.0%
Reaction Conditions	
Residence Time	60 min
Solids Loading	147 g/L
Temperature	225°C
Pressure	35 bar

^aRepresented by succinic acid in the model

The modeled reactions in MOT are stoichiometrically balanced so that any given feed component is reacted with oxygen to produce the specified carboxylic acids and CO₂ and is balanced with water and oxygen (CO₂ production is set from the aspirational target noted above of 80% carbon conversion to acids). Any other elements present are liberated as follows: nitrogen as ammonia (NH₃); sulfur as sulfur dioxide (SO₂); and phosphorous as phosphoric acid (H₃PO₄). In addition to glucose (produced from carbohydrate hydrolysis in both flash hydrolysis and MOT) and protein, MOT is also assumed to facilitate the degradation of nonfermentable carbohydrates, cell matter, and glycerol. The full list of modeled stoichiometric reactions is shown in Table 7. It should be noted that these are approximations for the purpose of keeping the element balance consistent within the model. Although these reactions are likely not entirely consistent with actual stoichiometries, they are deemed to be an appropriate approximation for this level of detail.

Table 7. MOT Reactions and Assumed Conversions (100% conversions based on aspirational targets; acid yields/distributions consistent with experimentally demonstrated data)

Reaction	Reactant	% Converted to Product
Protein ^a + 1.74 O ₂ + 0.67 H ₂ O → 0.45 Formic Acid + 0.88 Acetic Acid + 0.48 Succinic Acid + 1.03 CO ₂ + 1.3 NH ₃ + 0.04 SO ₂	Protein	100%
Glucose + 1.74 O ₂ → 0.52 Formic Acid + 1.02 Acetic Acid + 0.56 Succinic Acid + 1.20 CO ₂ + 1.76 H ₂ O	Glucose	100%
NF Carbs ^b + 3.48 O ₂ → 1.04 Formic Acid + 2.04 Acetic Acid + 1.12 Succinic Acid + 2.40 CO ₂ + 2.52 H ₂ O	Sucrose	100%
Glycerol + 1.37 O ₂ → 0.26 Formic Acid + 0.51 Acetic Acid + 0.28 Succinic Acid + 0.60 CO ₂ + 1.88 H ₂ O	Glycerol	100%
Algal Cell ^c + 1.19 O ₂ → 0.0016 H ₃ PO ₄ + 0.029 NH ₃ + 0.0015 SO ₂ + CO ₂ + 0.854 H ₂ O	Algal Cell	100%

^a Protein is represented by C_{5.17}H_{9.86}N_{1.30}O_{2.57}S_{0.04}

^b Non-fermentable carbohydrates are represented by Sucrose.

^c Algal cell mass is represented by CH_{1.8021}O_{0.4938}N_{0.0292}S_{0.0015}P_{0.0016}

The reactions occurring in MOT are highly exothermic. Instead of wasting this heat, it is utilized by operating the reactor adiabatically and controlling the temperature by the vent takeoff. The vent stream is then cross-exchanged with water to produce low-pressure (150°C) steam for the plant. After cross-exchange, the stream goes to a two-stage flash. The first flash tank recovers the

majority of the vaporized acids, while the second flash separates the gaseous CO₂ from the aqueous phase. Both streams from the second flash are used as recycles to the algae farm.

The liquid from MOT is cooled and then sent to a microfiltration unit, which removes ash and any residual solids (note that any consumable nonlipid components, listed in Table 7, are assumed to react to completion). Any lipids are also assumed to be precipitated across MOT and removed by this filter. This assumption comes from experimental observations that O₂ from MOT will form oxygen and peroxy linkages between fatty acids, making them insoluble in water and causing them to stick to the solid ash.

Following MOT, the NH₃ and H₃PO₄ is adsorbed by a dual bed ion exchange column (one acid resin, one base resin) to be recycled back to the algae ponds. These beds are regenerated by NaOH and H₂SO₄ to recover H₃PO₄ and NH₃, respectively, at a 1:1 molar ratio. This step has not yet been experimentally verified, nor has the efficacy of resulting N/P recycle to cultivation, and will be important areas for future experimental confirmation given potential challenges in recovering these components from complex organic mixtures. Another alternative that may be considered in the future is using a single ion exchange column which adsorbs the carboxylates rather than the nutrients, followed by elution with methanol; this may enable a lower energy use in subsequent heating operations. In the current case, though, we are left with an aqueous mixture of water and carboxylic acids in the C₁–C₄ range, which is pumped to 57 bar and heated to 350°C via cross-exchange with a hot oil system (given that the heating temperature demand exceeds temperatures achievable with high-pressure steam) before entering the ketonization reactor.

Ketonization

The ketonization reactor is a fixed bed reactor with a heterogenous catalyst of 0.1% Pt on Nb₂O₅ [30]. The reactor is operated at 350°C and 57 bar, with a weight hour space velocity (WHSV) of 1.7 hr⁻¹ and a catalyst replacement period of 2 years. The operation is similar on a high level to NREL's 2018 biochemical design case which also utilizes a ketonization step based on demonstrated experimental data, albeit that case focused on upgrading a single carboxylic acid (butyric acid) to a single ketone product (4-heptanone), processing pure butyric acid [11]. In the present application, the ketonization step is less straightforward, processing a more diverse feed of multiple acids in the presence of 86% water. In reality, there would likely be a broader spectrum of carboxylic acids present from upstream MOT than the representative components reflected for current modeling efforts, which would combine to form an even broader spectrum of ketones. However, because knowledge of this product slate is limited based on a dearth of analytical data available to date, it is assumed that the acetic acid reacts to form acetone (a well-understood reaction, assumed to reach 100% of the theoretical carbon yield) and the succinic acid reacts to form a cyclic ketone represented by 3-methyl-2-cyclopenten-1-one. The latter product was observed experimentally as one of the predominant C₆/C₇ cyclic ketones produced from a mock mixture of formic, acetic, and succinic acids.

Early experimental data has yielded a lower fraction of C₆/C₇ cyclic ketones; however, these are more desirable products (given their potential to constitute valuable jet fuel precursors) and future research will aim to shift the balance more towards these higher molecular weight molecules. In the modeled reaction for this product, two moles of succinic acid react to form the cyclic ketone, with two moles of carbon lost as CO₂ (achieving 100% of the theoretical carbon

yield, as in the acetone reaction). Oxygen is balanced with water. These assumptions are acknowledged to be a simplification of the full chemistries; however, the important parameters of the reaction in terms of TEA ramifications (namely, carbon yields and throughout for reactor sizing determination) are accurately captured.

Ketonization reactions are assumed to proceed to completion. The effluent stream from the ketonization reactor is cooled and sent to a two-stage flash distillation to separate the CO₂ produced by the reactions. This leaves a mixture of acetone, cyclic ketone, and water. This stream is first distilled to recover the acetone, followed by decanting of the column bottoms. The aqueous phase is recycled to algal cultivation and the organic phase (primarily the cyclic ketone intermediate) is combined with the acetone and routed to condensation.

Condensation

Given the small carbon chain length of the ketone products (C₃-C₆), a chain-elongation step is required to produce fuel-range products. This is also similar in concept to the above-cited NREL biochemical design case which also employs a condensation step (albeit again more simplistically converting a single C₇ ketone to a single C₁₄ enone in the biochemical design), and assumes a similar processing approach based on a batch slurry reactor, assuming the same catalyst as well as use of a solvent to promote good conversion constrained by slow kinetics for this step [11]. Here, this condensation step assumes acetone and the cyclic ketone are reacted to two main products: aristolone and isophorone (C₁₅ and C₉ oxygenated cyclic molecules). Again, these compounds have been experimentally observed under early work as some of the predominant products from condensation, but are also meant to represent a broader spectrum of similar products. Similar to the above steps, the experimental data for this operation is also limited to date in the context of these ketone molecules, so the model assumes full carbon preservation from ketones to products, with excess oxygen rejected as water (as is consistent with a general condensation reaction of ketones). Reaction yields are manipulated to produce a 60:40 ratio by mass of aristolone to isophorone, in line with early experimental data. The condensation reaction occurs in a slurry reactor with a heterogeneous Nb₂O₅ catalyst. Recent experiments have suggested that a Mn-Zr-O catalyst could alternatively be a promising catalyst for this reaction, consistent with results of Wu et al [31]; at present, the niobium catalyst basis is maintained consistent with the biochemical design case [11]. In future work, this will be revisited if additional experimental data support preferential use of the Mn-Zr-O option.

The condensation reaction occurs at 220°C and 11.2 bar with a residence time of 2.5 hours, and reactions are assumed to proceed to 60% conversion per-pass. After reaction, the mixture is separated in a decanter to remove the aqueous phase. The organic phase is sent to a distillation column to recover the solvent and unreacted ketones, and the bottom product stream proceeds to hydrotreating section. The overhead is recycled, translating to a net overall 99.1% conversion across condensation (ketone feed vs product to hydrotreating).

Hydrotreating

In this step, the condensation products and the free fatty acids from A200 are upgraded in a hydrotreater to refine the oxygenated intermediates into saturated hydrocarbon components suitable for blending into diesel, naphtha (gasoline), and possibly jet fuel pools. Prior NREL research demonstrated the potential to upgrade algal lipids through a one-step hydrotreating reaction using a 1% Pt/SAPO-11 catalyst with combined functionality to perform both HDO and

HI, with the latter enabling improved cold-flow properties such as cloud point relative to straight-chain C₁₄–C₁₈ paraffins otherwise formed from straight HDO of lipids. Compared to traditional (and prior) HDO catalysts that use metallic functional groups for hydrogenation, the HDO+HI catalyst combines a metallic (Pt) functional group for hydrogenation with an acidic (SAPO-11) functional support for isomerization of the hydrocarbon substrate. Previous TEA assessment coupled with biorefinery blending models found that performing HDO and HI in two separate reactor steps would incur more processing costs than the increased value as a blendstock was worth, but that combining both functions into a single reaction step would be justified to improve the blendability into the diesel pool and overall value to a refiner. Such an approach could be maintained here if it could be shown to impart isomerization to the lipid HDO products, though would not be necessary for the condensation HDO products (already highly isomerized cyclical structures). Either way, when constrained to a single-step reactor, the cost differences are not expected to be significant whether only HDO or combined HDO+HI activity is sought. More important to the economics would be the ability to perform hydrotreating for both the FFA and condensation products in a single combined step, rather than two parallel reactors. Given that the condensation product molecules would only require mild deoxygenation and saturation (similar to the FFA components), this combined operation is expected to be straightforward.

Table 8. Summary of Hydrotreating Process Assumptions (based on guidance from engineering subcontractor and experimental data)

Oil feed rate (at standard conditions)	43,992 gal/day (166.6 m ³ /day)
Average reaction temperature	707°F (375°C)
Pressure	435 psig (30.6 atm)
Hydrogen partial pressure	322 psig (22.6 atm)/5,864 SCF/bbl
Reactor LHSV	~1 hr ⁻¹ (estimated)
Reaction H ₂ consumption	3.1 wt% of intermediate feed
Hydrogen loss in purge	9.0%
Average makeup hydrogen rate	251,402 SCFD (6,730 N-m ³ /day)
Diesel Rate – design capacity	42,048 gallons/day (159.2 m ³ /day)
Naphtha Rate – design capacity	5,172 gallons/day (19.6 m ³ /day)

Operating details for the hydrotreating step are shown in Table 8. The free fatty acid portion of the feed is reacted to water, CO, CO₂, and a slate of saturated hydrocarbons ranging from C₃ to C₂₁ based on internal NREL experimental data. The condensation products (cyclic oxygenated intermediates) are assumed to react with hydrogen to produce saturated cyclic hydrocarbons of the same structure, with HDO oxygen rejection as water. The total final product yields from both substrates are shown in Table 9. An added advantage of splitting FFAs from TAGs with only the former sent to hydrotreating for fuel upgrading is that it saves on hydrogen consumption to remove the glycerol backbone (lost as propane), reflected as a low off-gas yield here. However, overall hydrogen consumption is relatively high due to hydrogen losses in the purge stream (as necessitated by the presence of CO and CO₂) in addition to HDO activity to process the oxygenated ketones based on the representative ketone structures noted above.

Table 9. Product Yields from Hydrotreating (lipid yields based on experimentally demonstrated data; MOT train intermediate HDO yields based on reasonable projections for 100% HDO conversion)

Product	Yield, wt% of intermediate feed
Total reactor product yield	103.1%
H ₂ consumption	-3.1%
Reactor product composition:	
Fuel-range hydrocarbons	87.3%
<i>Diesel-range</i>	78.2%
<i>Naphtha-range</i>	9.1%
CO ₂	1.0%
CO	1.0%
H ₂ O	9.8%
Off-gas (propane)	0.9%
Total	100.0%

The product stream exiting the hydrotreating reactor is split into gas, organic, and aqueous phases. The aqueous phase (primarily water produced from the HDO reaction per Table 9) is recycled to the pond and the gas phase, minus a 5% purge, is compressed and recycled back to the reactor to minimize losses of hydrogen. The organic phase is sent to a steam-stripped distillation column where it is split into diesel, gasoline, and off-gas fractions. The off-gas is sent to the steam boiler, while the diesel and gasoline are diverted to storage as the primary fuel products reflected in this design.

3.3.3 Cost Estimation

The MOT reactors were sized and costed using ACCE. Based on mixing and O₂ transfer requirements, it was determined that the optimal setup would be two 71-m³ parallel reactors, with a total purchase cost of \$1.3 MM for both reactors. Given the uncertainty associated with the specific engineering that would be associated with this reactor design, a contingency factor of 2X was also included, bringing the total purchase cost of both reactors to \$2.6 MM. Other equipment costs required for MOT such as heat exchangers, flash tanks, and O₂ compressors were taken from NREL's previous biochemical and CAP design reports [4, 11]. The microfiltration unit and the ion exchange column costs were obtained from subcontractor quotes (Harris Group), which were originally cited in NREL's 2015 catalytic biochemical conversion design report [21].

The ketonization reactor cost was scaled consistently from NREL's previous 2018 biochemical report. It should be noted that the previous reactor operated with pure acid in the vapor phase, while this ketonization reaction occurs in the liquid aqueous phase. However, the reactor design should not change drastically, and any change in throughput is accounted for with cost scaling on reactor feed. Similarly, the condensation reactor cost is also based on the 2018 biochemical report, with cost adjusted based on the difference in residence time/reactor volume. The distillation column costs for ketonization and condensation were estimated using ACCE. Other equipment required for ketonization and condensation, such as heat exchangers, flash tanks, the decanter, and the hot oil system, are also taken from the 2018 biochemical report.

Costs for all HDO/HI equipment were estimated using ACCE, based on consistent basis costs and scaling approaches as done for the HDO reactors in NREL's 2018 biochemical design

report. Costs for the catalyst (1% Pt/SAPO-11) were estimated at \$360/lb, with a 2-year catalyst lifetime and a 42% reclamation factor for subsequent catalyst replacement costs.

3.4 Area 400: Polyol/Polyurethane Coproduction

3.4.1 Overview

This area covers the coproduct processing train in parallel to fuel production, namely TAG upgrading to polyols or polyurethanes (both options are considered in this assessment, see Figure 6). After lipid purification and distillation to remove FFA, the remaining TAG fraction is routed to polyol upgrading. Prior NREL TEA models for this step [8] had originally assumed a different approach largely based on literature wherein lipids were assumed to be entirely constituted of FFAs (i.e., TAGs are first hydrolyzed to FFAs whether by biological functions of the strain or as a processing step), and polyunsaturated fatty acids (PUFAs) were selectively isolated via urea complexing [32-34]. Isolated PUFAs were then reacted with hydrogen peroxide to form epoxide rings from each double bond, and then epoxide rings were opened with methanol and water to form polyols [35, 36]. The polyols were subsequently esterified with glycerol to yield a polyester polyol of higher molecular weight [36]. Finally, in the case of polyurethane (PU) synthesis, the resulting polyester polyol was reacted with toluene diisocyanate (TDI) to produce polyurethane foams [37].

More recently, NREL consulted with an engineering subcontractor, Nexant, to solicit inputs on the assumptions and practicality of the above-described process. Through that work it was identified that such a process would likely be suboptimal for several reasons: (1) it is predicated on 100% FFA feedstocks (not typical for most bio-derived lipids, though it may be possible for algae depending on strain or additional processing to hydrolyze TAGs); (2) the urea complexing operation is largely based on patent literature and has not been demonstrated commercially (although other somewhat similar commercial processes exist for achieving PUFA isolation); (3) epoxidation and ring opening across two separate steps could be consolidated into a single-step operation as is practiced commercially; and (4) the resulting polyol structure from such a route would likely be more suitable for rigid than flexible foam applications upon conversion to PU, requiring a different PU process train design and cross-linker than TDI. Accordingly, Nexant provided guidance for a more optimal and straightforward configuration, based on isolating TAG (via simple vacuum distillation as discussed in Area 200) rather than PUFA and then upgrading TAG to polyol in a single-step epoxidation/ring-opening reaction, all following commercially demonstrated technologies [38]. The resulting polyol from this approach is more appropriately suited for the flexible foam market if electing to include polyol upgrading to PU, thus allowing to maintain a flexible foam reactor with TDI cross-linker as originally envisioned (but with costs modified per Nexant inputs).

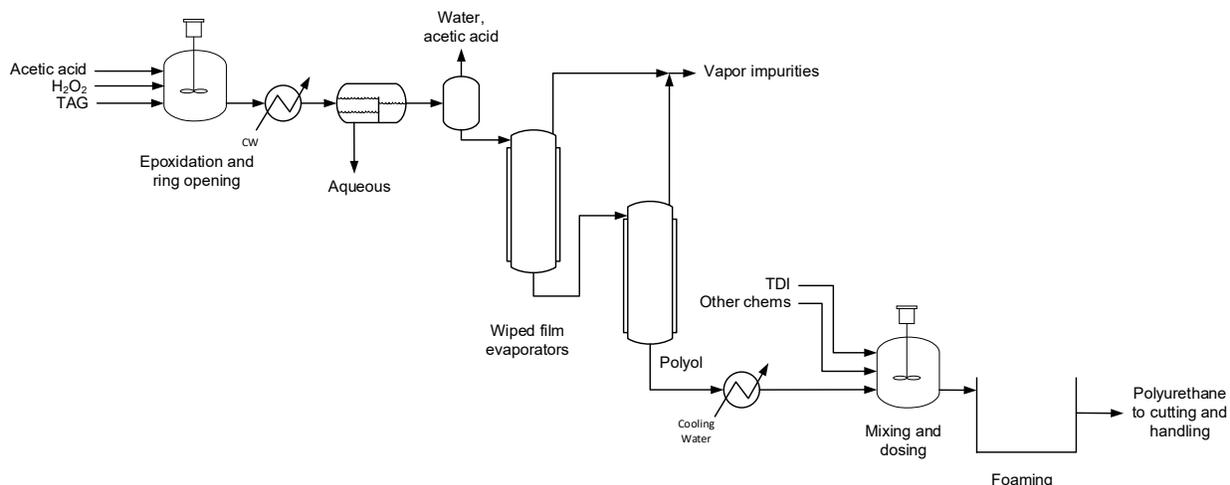


Figure 6. Simplified flow diagram of the polyol/polyurethane upgrading process

3.4.2 Design Basis

Polyol Production via Epoxidation and Ring Opening

The triglycerides separated from FFAs via vacuum distillation are sent to the polyol/polyurethane upgrading train. The fatty acid profile of the algal lipid fraction was determined analytically for the HCSD biomass sample reflected in Table 2 and is shown in Table 10. Given the wide array of fatty acids, compounded by the fact that any given triglyceride could have any combination of three fatty acid variations, the decision was made to not rigorously track TAG upgrading chemistries in the Aspen Plus models. However, thorough engineering analysis was still conducted based on consultations with Nexant to establish important factors such as yields, utilities, and raw materials needed across each step in this processing train, specific to this fatty acid profile as the underlying design basis. Any utility requirements were tied in with the overall plantwide utility system.

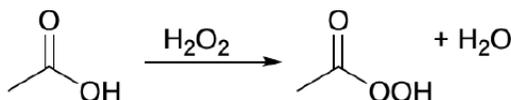
Table 10. Fatty Acid Profile of HCSD Algal Biomass (internal NREL data)

Fatty Acids Carbon Number	Degree of Unsaturation	Fatty Acid Profile (Percent of Incoming FFAs)
18	2	7.8%
18	3	6.2%
18	1	49.7%
18	0	4.2%
16	1	11.9%
16	0	19.2%
20	0	1.0%

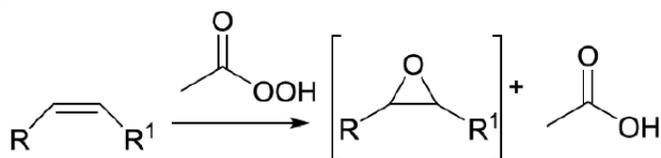
In epoxidation and ring opening, any unsaturated bonds are reacted with hydrogen peroxide (H_2O_2) and acetic acid to form and subsequently open epoxide rings in a single reactor step, leaving an alcohol group and an ester group (formed by the acetic acid) on either side of what

was previously an unsaturated bond. This process is described in a 2014 patent by Bio-based Technologies, since acquired by Cargill [38]. A mechanism is also proposed and is represented in Figure 7. Briefly, the acetic acid (or another organic acid having 1 to 3 carbons) reacts with the H₂O₂ to form peracetic acid. The peracetic acid reacts with the double bond on the lipid, forming an epoxide and regenerating the acetic acid. Next, the epoxide reacts with acetic acid to form the hydroxy ester, as shown in Figure 7. Alternatively, the epoxide can react with water (present via the peracetic acid formation or from the aqueous H₂O₂ solution) to form a diol. A third potential reaction would be the epoxide reacting with previously formed hydroxyl groups to form a hydroxyl ether. The rate of the epoxidation reaction can be controlled by altering the amount of H₂O₂ present in the reactor.

Peracetic Acid Formation:



Epoxidation of Vegetable Oil:



Ring Opening of Epoxidized Vegetable Oil:

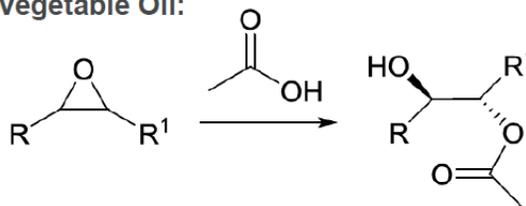


Figure 7. Mechanism for epoxidation and ring opening of vegetable oil (furnished under subcontract with Nexant)

The reaction occurs at ambient conditions and mild temperature. Once reaction is complete, a crude product mixture remains, consisting of polyether polyols, unreacted reactants, and other volatiles. This mixture is cooled to about 50°C–90°C and the aqueous phase is separated from the organic phase. The organic phase is sent to a flash evaporator under vacuum conditions (5–15 mm Hg) to remove the lighter volatiles, and is then sent to a degasser to remove the majority heavier components such as acetic acid and water.

Following this, the polyol mixture is sent to two wiped film evaporators in series, each at mild vacuum and temperatures up to 140°C and 260°C, respectively. The finished polyol product from the second evaporator is then cooled. In the case of producing polyols as the final coproduct, this stream is diverted to storage; if the final coproduct is polyurethane, the polyol mixture is sent to the polyurethane foam production line.

In either case, the average hydroxyl number of the polyol mixture is an important property dictating the optimal polyurethane product end use. The hydroxyl number is a metric used in industry to get a measure of the relative weight equivalence of hydroxyl groups, in units of mg KOH/g. The equation for the hydroxyl number is given below, where functionality represents the number of hydroxyl groups for a given polyol.

$$\text{hydroxyl number} = \frac{\text{functionality} * 56.1 * 1000}{\text{molecular weight}}$$

Polyols with higher hydroxyl numbers ranging from 300 to 500 mg KOH/g are used to make rigid polyurethane foams [39], while flexible foams are typically made from polyols with a hydroxyl number below 100 mg KOH/g [40]. Given the fatty acid profile in Table 10, the average hydroxyl number of a polyol from TAG would be approximately 148 mg KOH/g, assuming all epoxides react with acetic acid to form a hydroxy ester (as shown in Figure 7). This number could change in either direction depending on the relative reaction rates of the epoxide with water or previously formed hydroxy groups. The hydroxyl number can also be impacted by feeding less H₂O₂, decreasing the conversion of the reaction to epoxides. This keeps the molecular weight of the molecule close to the same, but reduces the final functionality. In this way, the final hydroxyl number can be controlled to some degree by changing the amount of H₂O₂ available during the reaction.

The predicted average hydroxyl number of 148 mg KOH/g in the present work is closer to the range for flexible foams and can be decreased by altering feed rate of H₂O₂. If the H₂O₂ feed rate is decreased such that fractional conversion of double bonds to epoxides is 0.80 or less (still assuming all epoxides are reacted with acetic acid to form a hydroxy ester), the hydroxyl number will be less than 100 mg KOH/g, placing the polyols in solid territory for flexible polyurethane production. Literature suggests that a typical fractional conversion of double bonds from epoxidation and ring opening is 75%–90% [41]. In a real production, testing would be done to determine the optimum conditions to produce the best quality polyol; however, the required raw materials would not change drastically. Table 11 shows the inputs required in the ring opening and epoxidation section of the process.

Table 11. Process Inputs for Ring Opening and Epoxidation (based on inputs from Nexant)

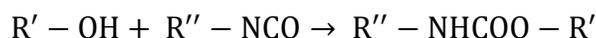
Raw Material	Feed Requirement, w/w Polyol Product
Triglyceride	0.90
Acetic acid	0.17
H ₂ O ₂	0.21
Other chemicals	\$46.70/ton Polyol
Utilities	Requirement per ton Polyol
Power	0.54 MWh
Cooling water	220 ton
Low-pressure steam	40 lb

Flexible Polyurethane Foam Production

By 2020, U.S. consumption of polyurethane foam is expected to reach 2.4 MM tonnes per year, split roughly 50/50 between flexible and rigid foams based on industry reports. Approximately 0.41 MM tonnes (17%) of that consumption is attributed to flexible bedding foams. In addition to bedding, flexible polyurethane is also used in transportation, furniture, flooring underlay, packaging, and textiles, with bedding being the largest market and transportation following close behind. The large market paired with a relatively high selling price makes flexible polyurethane boardstock a desirable coproduct candidate.

Based on this information and the hydroxyl number of the polyol, as previously discussed, the polyol is diverted to a dedicated flexible slabstock foam production facility, again based on guidance furnished from Nexant subcontractors. This facility consists of four sections: raw material tank farm, mixing and dosing section, foaming section, and the cutting and handling section. Additional foam cutting needs are performed as a later operation using simple hand tools to meet customer orders.

Polyol and the other required raw materials are sent from the raw material tank farm into the mixing and dosing section. The raw materials, including polyol, TDI, water, diethanolamine (DEOA, a surfactant), and additional surfactants, are sent to the mixing chamber. The TDI is a mix of two different isomers: 80% of the 2,4 isomer and 20% of the 2,6 isomer. In the mixing chamber, the raw materials are mixed by rotating impeller, creating bubble seeds that eventually become the cells in the finished foam. The homogenous liquid mixture, interspersed with bubble seeds, is sent to the foam growth section. Here, the chemical reactions occur. Linkages are created when the TDI reacts with the alcohol groups on the polyol. This polymerization reaction, shown below, is exothermic and acts to raise the temperature of the rising foam [37]. This rise in temperature helps expand the bubble seeds.



Separately, the isocyanate also reacts with water to produce CO₂ which acts as a blowing agent [37]. Gaseous CO₂ makes its way into the bubble seeds, expanding them. Eventually, the spherical cells come into contact and distort, creating a network of cells that fill 99% of the total volume of the foam. The foam expands in the early stages of the curing process; once the expansion is complete, the foam is allowed to cure for several hours more before further processing. At this time, the foam is sent to the cutting and handling section, where mechanical cutting equipment shapes the foam into the required size and shape.

Like the polyol plant, the polyurethane plant is not rigorously modeled in Aspen Plus based on specific components and reaction chemistries. However, key process input requirements, operating conditions, and yields were furnished by Nexant. Raw material and utility requirements are shown in Table 12.

Table 12. Process Inputs for Flexible Polyurethane Foam Production (based on inputs from Nexant)

Raw Material	Feed Requirement, w/w Polyurethane Product
Polyol	0.6111
TDI	0.3544
Water	0.0281
DEOA (98%)	0.0026
Surfactant	0.0049
Utilities	Requirement per tonne Polyol
Power	0.004 MWh

Polyurethane foam plants generally operate independently from polyol production plants, given that polyol plants are more of a liquid chemical production facility, while polyurethane is a more manpower-intensive solid manufacturing facility. However, there is no reason that both cannot be operated in close proximity and share the same utility systems. One of the main reasons that this is generally not done is the production schedule. A polyol plant, like most chemical production facilities, operates as a 24/7 facility (not including regular scheduled shutdowns). In contrast, a typical polyurethane plant operates on an 8-hours-per-day, 5-days-a-week schedule [42, 43]. The main reason for this is the maintenance requirement for the foam line, which can operate up to 8 hours at peak raw material throughput. After 8 hours, it must be taken offline for a 3-hour maintenance cycle, which involves disassembling the system, removing and cleaning multiple elements, and then reassembling. In addition to this, the foam is typically produced at a higher rate than the fabrication process and is therefore limited by storage space available [44].

The designed process overcomes these obstacles by operating at a higher throughput to make up for the 3-hour maintenance cycle per 8 hours of operation, and including the cost for a large 130,000-ft² warehouse capable of storing up to 5 days’ worth of polyurethane foam. This warehouse would provide a holdup spot in the supply chain to allow flexibility for the foam blocks to be shipped for final fabrication as needed, while also providing space for curing needs.

It is recognized that the current standard for polyurethane production, using some form of isocyanate (in this case, toluene diisocyanate), has negative health and environmental impacts [45, 46]. Accordingly, NREL is focusing current and future research on a non-isocyanate bio-based polyurethane linking agent. At this time, however, this research is in too early of a stage to faithfully provide a relevant techno-economic analysis on this process.

3.4.3 Cost Estimation

Cost estimations for the polyurethane train were largely based on guidance from Nexant subcontractors. Consolidated CAPEX values, estimated by Nexant, were used for each section and adjusted as needed based on process throughput. The epoxidation and ring opening section was based on a 30,000 tonne polyol/yr process with an installed inside battery limits (ISBL) cost of \$30.75 MM (2018 dollars), and the flexible foam polyurethane plant section was based on a 2,530 tonne polyurethane foam/yr process with an installed ISBL cost of \$4.75 MM. Both CAPEX costs were adjusted with a 0.65 scaling factor, as recommended by Nexant.

Separate from the normal warehouse cost calculations (calculated as 4% of ISBL), a dedicated rigorous warehouse cost was calculated based on keeping 5 days inventory of polyurethane. Approximately 1 day of inventory space is dedicated to curing needs for the foam, while the

remaining 4 days address supply chain concerns with downstream fabrication. The cost for this warehouse was estimated by calculating the required square footage of the warehouse and using a cost value of \$100/ft². This warehouse was calculated to cost an additional \$25.7 MM and was considered an outside battery limits (OSBL) cost.

Raw material costs for the polyurethane train were calculated using an average of values supplied by Nexant and other industry price sources. The selling price of polyurethane was set at \$2.04/lb, based on a 5-year average from industry sources, priced in \$/board/foot, and a density of 1.0 lb/ft³. Alternatively, drawing the boundary at polyols as the final output product was also considered over a range of values and yields, with results and context presented in Section 5.2.

3.5 Area 500: Boiler, Utilities, and Storage

3.5.1 Overview

Area 500 contains all OSBL operations to support the facility. This includes a gas boiler, cooling water, process water, plant and instrument air, the clean-in-place (CIP) system, and bulk storage for process chemicals and output products. The boiler generates high-pressure steam for utility heating, while low-pressure steam is generated with excess heat from MOT. Additionally, hot oil is generated to serve high-temperature heating demands in excess of high-pressure steam limitations. Any biorefinery off-gas streams with significant heating values are combusted in the boiler, with supplemental natural gas added as necessary to satisfy heat balances. Product storage tanks explicitly costed include diesel and naphtha. Polyurethane is produced as a solid foam slabstock, with storage costs accounted for separately (see Section 3.4.3).

3.5.2 Design Basis

Prior CAP models have employed anaerobic digestion (AD) coupled with biogas combustion in a gas turbine, with combined heat and power generation to raise steam from the turbine flue gas. In the current model, as there is no AD system in this design and accordingly no biogas, this approach has been replaced with a simple gas boiler to generate the necessary steam to satisfy facility demands. Namely, this includes high- and low-pressure steam systems generating saturated steam at 46.3 atm/260°C and 2.7 atm/130°C, respectively. Both the high-pressure and low-pressure systems are driven by combustion of biorefinery off-gas streams and supplemental natural gas; however, the low-pressure steam also takes advantage of the exothermic MOT reaction and is partially generated in the MOT vent heat exchanger. Figure 8 provides a breakdown of all facility heat demands, with roughly 33% allocated to low-pressure steam, 53% to high-pressure steam, and 14% to a hot oil utility system satisfying high-temperature heating demands in excess of high-pressure steam limits (specifically serving the ketonization reactor pre-heater at a process temperature of 350°C and the final flash hydrolysis pre-heater at a process temperature of 280°C). Heat integration is used wherever possible to minimize the heating duty required. Some of the notable users of high-pressure steam are A100 (driven by the large preheating requirement for the flash hydrolysis reactor) and A300 (driven by the preheating requirements prior to the MOT reactor). The low-pressure steam demand is dominated by A200 requirements, making up approximately 60% of low-pressure (LP) steam usage. Most of this demand comes from a few key items: the large reboiler duty of the ethanol recovery column, the preheating requirement for the hexane recovery column, and the steam ejector requirement for

the TAG-FFA vacuum column. The next largest user of LP steam comes from the distillation column following the ketonization reactor.

Total Heating Duty Required = 76.9 MMkcal/hr (304.9 MMBtu/hr)

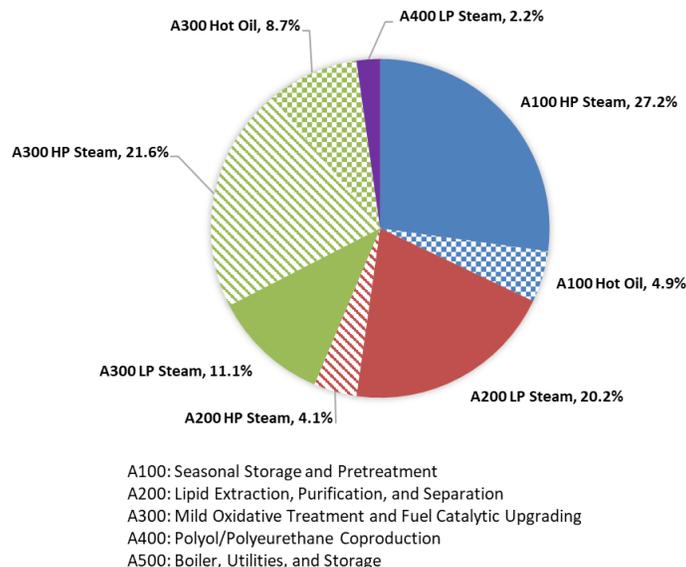


Figure 8. Average utility heating demand duty distribution between major users. (Note, some LP demands shown here are met by MOT heat evolution.)

Likewise, Figure 9 provides a breakdown of cooling water users throughout the facility. 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. 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.

The largest cooling water demand comes from Area 200, driven by the condensation of a large hexane recycle, as well as re-condensation of hexane when it is added to the flash hydrolysis effluent for rapid cooling. Other key users include the A300 section (mostly driven by the cooling requirement prior to ion exchange in addition to various smaller users) and polyol production section in A400 (requiring cooling water both after the reactor and after the wiped film evaporators).

Total Cooling Water Duty Required = 79.8 MMkcal/hr (316.3 MMBtu/hr)

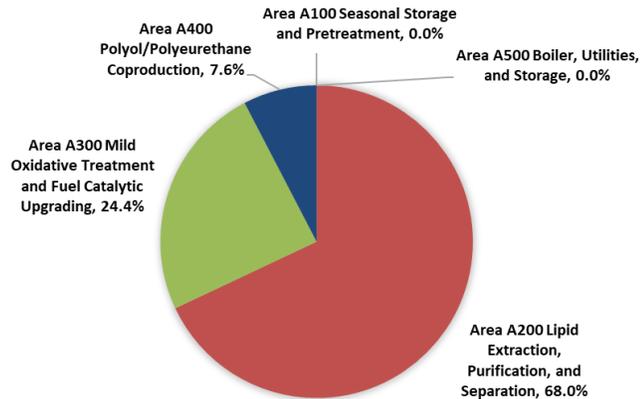


Figure 9. Average cooling water heat duty distribution between major users

Figure 10 provides a breakdown of average power demands by key process step. As there is no power generation from a CHP system in this design, all facility power demands are met by purchasing grid electricity. Most of the power demand is associated with pumps and compressors, with the largest demand coming from A500 (driven by the boiler air compressor, high- and low-pressure steam pumps, and cooling water pumps and tower fans). The power use in the other areas are all comparable to one another, with all being driven by a combination of pumps and compressors. Purchased power is costed at a rate of 6.8 ¢/KWh (see Section 4.2).

Total Power Required = 8.4 MW

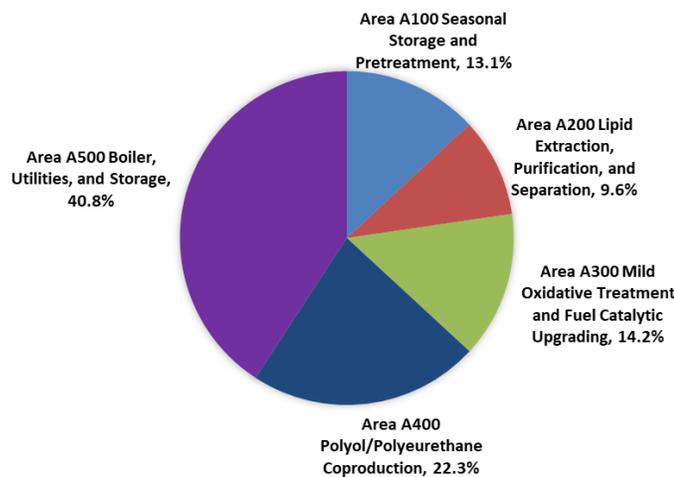


Figure 10. Average distribution of plant electricity utilization by process area

Table 13 shows the major storage requirements for the present design. Tanks for diesel and naphtha products were each sized to ensure at least 7 days of storage, while the fire water storage tank was scaled from NREL’s 2013 design report according to dry biomass feed rate to the pretreatment reactor following seasonal storage.

Table 13. Storage Requirements for Major Tanks

Material	Size
RDB product	Sufficient to contain >7 days of production: 1 carbon steel tank @ 300,000 gal
Naphtha product	Sufficient to contain >7 days of production: 1 carbon steel tank @ 40,000 gal
Fire water	4 hours of fire suppression @ 2,500 gal/minute: 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 prior NREL TEA reports.

3.5.3 Cost Estimation

The costs for most equipment within A500 were scaled from prior NREL design reports, namely the cooling tower, chilled water, and CIP systems, as well as product storage tanks [7, 22]. Costs for the boiler were estimated in ACCE, based on a package cost including draft fans, instrumentation and controls, burners, chemical injection system, steam drum, and other minor items.

4 Process Economics

This section summarizes the key economic inputs and results for the modeled biorefinery, with all cost results reflective of the coproduct boundary extending through polyurethanes in constituting a single base case (drawing the boundary earlier at polyols is reflected in the sensitivity analysis discussions). More details and supporting descriptions for the various economic parameters can be found in previously published NREL design reports [7, 10, 22]. The total capital investment (TCI) is first computed from the total equipment cost. Next, variable and fixed operating costs are determined. With these costs, a discounted cash flow analysis is used to determine the MFSP required to obtain a zero net present value (NPV) with a set internal rate of return (IRR). The analysis does not consider policy factors (e.g., subsidies and carbon credits) or early pioneer plant economics, but instead focuses strictly on the economic implications for the technology described here and associated future technical performance levels that must be achieved to realize those economics under a mature n^{th} -plant scenario.

The cost-year of 2016 was chosen for this analysis as consistent with recent NREL and other DOE-BETO TEA reports [11]. Accordingly, equipment and material costs based on information obtained in previous years are indexed to 2016. As presented in prior TEA studies and BETO Multi-Year Program Plan reports [10, 47], capital costs were adjusted using the Plant Cost Index from *Chemical Engineering Magazine* [48] to a common basis year of 2016. Similarly, for chemical/material costs we used the Producer Price Index for chemical manufacturing published by the U.S. Bureau of Labor Statistics [49]. Employee salaries were maintained from prior basis values and were scaled to 2016 using the labor indices provided by the U.S. Bureau of Labor Statistics [49]. The general formula for cost-year dollar back-casting is:

$$2016 \text{ Cost} = (\text{Base Cost}) \left(\frac{2016 \text{ Cost Index}}{\text{Base Year Index}} \right)$$

4.1 Total Capital Investment

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

In summary, each type of equipment utilizes a different installation factor to scale the given direct equipment purchased cost to a final installed cost. A complete list of the equipment is provided in Appendix A, along with equipment purchased and installed costs. The purchased cost for a given component reflects a baseline equipment size. As changes are made to the process, the equipment size required may be different than what was originally designed and costed. Consistent with previously published NREL TEA reports, a standard exponential cost

scaling equation was used to adjust for new process conditions or throughputs, using a scaling exponent n to account for economy-of-scale dependencies for a given operation:

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

Once the total equipment cost has been determined in the year of interest, several other direct and indirect costs were added to determine the TCI. Site development and warehouse costs, along with additional piping, are based on the ISBL equipment costs (Areas 100–400 in this design) and are considered part of the total direct cost (TDC). Beyond the ISBL operations, the supporting facilities under Area 500 are considered OSBL. Project contingency, field expenses, home office engineering and construction activities, and other costs related to construction are computed relative to the TDC and give the fixed capital investment (FCI) when summed. The sum of FCI and the working capital for the project is the TCI. These additional direct/indirect cost factor categories and values were maintained consistently with prior published NREL TEA reports [7, 11] and the reader is referred there for further details and descriptions of the various expenses included in each category. Beyond those factors, warehouse costs for curing and storage of the polyurethane product were calculated as a separate line item given that this is a significant expense beyond standard warehouse cost factors applied to all other CAP operations, following the details discussed in Section 3.4.3; this translates to a polyurethane warehouse “cost factor” of \$2,540 per kg/hr of polyurethane product (linear scaling assuming full warehouse capacity utilization). Table 14 summarizes the resulting installed capital, TDC, FCI, and TCI costs for the PU coproduct base case.

Table 14. Project Cost Worksheet Including Total Direct Costs and Total Capital Investment

Process Area		Purchased Cost ^a	Installed Cost ^a
Area 100: Seasonal Storage and Pretreatment		\$ 6,700,000	\$ 7,300,000
Area 200: Lipid Extraction/Purification/Separation		\$ 17,200,000	\$ 23,900,000
Area 300: MOT and Fuel Catalytic Upgrading		\$ 18,000,000	\$ 30,100,000
Area 400: Polyurethane Co-Production ^b		\$ 78,300,000	\$ 78,300,000
Area 500: Boiler, Utilities, and Storage		\$ 7,900,000	\$ 12,400,000
Totals		\$ 128,000,000	\$ 152,000,000
Warehouse	4.0%	of ISBL	\$ 4,400,000
Warehouse (polyurethane product) ^c	\$5,081	per kg/hr PU	\$ 25,800,000
Site development	9.0%	of ISBL	\$ 12,600,000
Additional piping	4.5%	of ISBL	\$ 6,300,000
Total Direct Costs (TDC)			\$ 201,000,000
Prorateable expenses	10.0%	of TDC	\$ 20,100,000
Field expenses	10.0%	of TDC	\$ 20,100,000
Home office & construction fees	20.0%	of TDC	\$ 40,200,000
Project contingency	10.0%	of TDC	\$ 20,100,000
Other costs (start-up, permits, etc.)	10.0%	of TDC	\$ 20,100,000
Total Indirect Costs			\$ 120,600,000
Fixed Capital Investment (FCI)			\$ 321,600,000
Land			\$ 1,800,000
Working capital	5.0%	of FCI	\$ 16,100,000
Total Capital Investment (TCI)			\$ 339,500,000
Lang Factor (FCI/purchased equipment cost)			2.51
TCI per annual GGE			\$20.69/GGE

^a Capital costs are based on peak throughputs.

^b Polyol/polyurethane process train costs provided by Nexant as total installed costs.

^c Only applicable for PU coproduction; not applicable when drawing the boundary at polyols.

4.2 Variable Operating Costs

Variable operating costs, which include raw materials, waste handling charges, and coproduct credits, are incurred only when the process is operating. Quantities of raw materials used and wastes produced were determined using the Aspen material balance. Table 15 documents the costs and sources of chemicals used in the process and Table 16 summarizes the variable costs on a per-year and per-GGE basis. Many input chemicals, as well as output disposal costs and imported power, have been utilized in prior recent NREL TEA reports, and thus the same cost basis was maintained in this work [4, 7, 8, 11]. Natural gas prices were updated to the latest 5-year average for industrial use per U.S. EIA data, at \$4.12/MM BTU. Nutrient and CO₂ recycle credits were set consistently with their purchase cost basis in the upstream algae farm model, to ensure overall consistency in the integrated pond-to-fuel supply chain between cultivation demands versus CAP outputs (nutrient/CO₂ recycle credits are not included in the algae farm model, but rather are credited to reduce MFSPs in downstream conversion). Other costs for new chemicals were based on inputs from Nexant as well as industry databases.

Polyol or polyurethane (PU) is coproduced in this design as a key enabling revenue stream to support the production of fuels at a viable MFSP. The base case assumes complete upgrading of isolated TAGs to PU beyond polyol intermediates, following guidance from Nexant for feasibility design/cost estimates on such a hypothetical integrated system. However, recognizing

that typically this is *not* an integrated process done all at once commercially (i.e., producing both the polyol and then the PU product in an integrated production process), and also given more uncertainties regarding final PU properties if produced exclusively from the isolated algal TAG feedstock, this study also considers an alternative scenario drawing the final product boundary at polyols (see Section 5.2). For the base case PU coproduction scenario, recognizing the variable nature for such commodity chemical prices, a 5-year average flexible foam PU value of \$2.04/lb (2016 dollars) was utilized in keeping with prior NREL TEA methodologies reflecting coproduction of a high-value chemical product.

Table 15. Chemical Costs + Coproducts and Sources

Component	Cost (2016\$)	Source
Inputs		
Biomass feedstock	\$0.241/lb	Algae farm cost targets [8], adjusted to 18% solids = \$486/ton
Oxygen	\$0.047/lb	Industry database
Sodium hydroxide (IX regeneration)	\$0.238/lb	Nexant (indexed from 2011\$ basis) [11]
Sulfuric acid (IX regeneration)	\$0.054/lb	Harris Group [10]
Natural gas	\$4.12/MM BTU	U.S. EIA 5-year average NG price for industrial use
Hexane (lipid extraction)	\$0.578/lb	Industry database
Ethanol (lipid extraction)	\$0.355/lb	Prior NREL analysis
Phosphoric acid (lipid cleanup)	\$0.381/lb	Harris Group cost database [7]
Silica (lipid cleanup)	\$1.060/lb	Harris Group cost database [7]
Clay (lipid cleanup)	\$0.318/lb	Harris Group cost database [7]
Toluene (condensation solvent)	\$0.330/lb	Industry database
Hydrogen	\$0.731/lb	DOE report, SMR H ₂ @ \$4/MM BTU NG [50]
Process water	\$0.0002/lb	Peters and Timmerhaus [51]
Acetic acid (polyol train)	\$0.274/lb	Average of Nexant and industry database values
Hydrogen peroxide (polyol train)	\$0.450/lb	Average of Nexant and industry database values
Catalysts/other chem (polyol train)	\$ scaling	Nexant inputs: \$0.05/kg polyol
Nitrogen (polyol train)	\$0.029/lb	Average of Nexant and industry database values
TDI (PU train)	\$1.382/lb	Average of Nexant and industry database values
DEOA (PU train)	\$1.028/lb	Nexant inputs
Surfactant (PU train)	\$1.249/lb	Nexant inputs
Ash/solids disposal	\$0.021/lb	Prior NREL TEA studies [10]
Electricity from grid	\$0.068/KWh	Prior NREL TEA studies [11]
Coproducts		
N recycle as NH ₃	\$0.409/lb	USDA fertilizer prices (5-year average) [4]
P recycle as DAP	\$0.334/lb	USDA fertilizer prices (5-year average) [4]
CO ₂ recycle	\$0.018/lb	NREL algae farm report @ \$45/tonne [4]
Polyurethane (flexible foam)	\$2.040/lb	Industry database (5-year average)
Polyurethane (foam scrap)	\$1.020/lb	Assuming 50% of main PU price for scrap after fabrication

Table 16. Variable Operating Costs. (Annual average across individual seasons.)

Process Area	Stream Description	Usage (kg/hr) ^a	MM\$/yr (2016\$)	¢/GGE (2016\$)
Raw Materials				
N/A	Feedstock	21,365	89.96	548.26
A200	Hexane	16	0.16	0.99
	Ethanol	271	1.68	10.26
	Phosphoric acid (cleanup)	59	0.39	2.40
	Silica (cleanup)	6	0.11	0.67
	Clay (cleanup)	12	0.07	0.40
A300	Oxygen	5,657	4.67	28.48
	NaOH (IX regen)	54	0.22	1.36
	Sulfuric acid (IX regen)	2,402	2.24	13.67
	Hot oil system (MM BTU/hr NG)	49	1.47	8.96
	Toluene (condensation solvent)	165	0.95	5.81
	Hydrogen	222	2.83	17.25
A400	Acetic acid	527	2.52	15.34
	Hydrogen peroxide	586	4.60	28.06
	Catalysts/other chem	\$ scaling	1.16	7.09
	Nitrogen	62	0.03	0.19
	TDI (PU)	1,798	43.40	264.50
	DEOA (PU)	13	0.24	1.44
	Surfactant (PU)	25	0.54	3.31
A500	Supplemental natural gas	4,352	6.89	41.96
	Boiler chemicals	0.04	0.00	0.01
	Cooling tower chemicals	2	0.01	0.03
	Process water	159,230	0.43	2.62
Power	Grid electricity (KW)	8,419	4.55	27.72
Subtotal			169.13	1,030.79
Waste Disposal				
A300	Disposal of ash	1,951	0.72	4.37
Subtotal			0.72	4.37
Coproducts and Credits				
A300	NH ₃ recycle	371	2.65	16.17
A300	DAP recycle	177	1.03	6.29
A300	CO ₂ recycle	23,948	7.70	46.94
A400	Polyurethane (foam)	4,871	173.55	1057.73
A400	Polyurethane (scrap)	203	3.62	22.04
Subtotal			188.55	1,149.17
Total Variable Operating Costs			-18.71	-114.02

4.3 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 labor costs were largely maintained consistently with other recent NREL CAP TEA models [8]. Reasonable additions were included here for laborers and operators to support the polyol/polyurethane train. Table 17 shows the recommended number of employees and associated salaries. Other fixed operating costs (e.g., labor overheads, maintenance, and property insurance) were also maintained consistent with other recent NREL TEA studies [10, 11].

Table 17. Fixed Operating Costs

Position	2016 Salary	# Required	2016 Cost	MM\$/yr (2016\$)	¢/GGE Fuel (2016\$)
Labor and supervision					
Plant manager	164,452	1	164,452		
Plant engineer	78,310	2	156,621		
Maintenance supervisor	63,767	1	63,767		
Maintenance technician	44,749	5	223,744		
Lab manager	62,648	1	62,648		
Lab technician	44,749	2	89,498		
Shift supervisor	53,699	2	107,397		
Shift operators	44,749	9	402,739		
Shift supervisor: polyols	53,699	2	107,397		
Shift operators: polyols	44,749	9	402,739		
Shift operators: PU	44,749	3	134,246		
Laborers: PU	31,324	30	939,724		
Yard employees	31,324	2	62,648		
Clerks and secretaries	40,274	2	80,548		
Total salaries		71	2,998,167	3.00	18.27
Labor burden (90%)			2,698,351	2.70	16.45
Other overhead					
Maintenance	3.0%	of ISBL	4,187,405	4.19	25.52
Property insurance	0.7%	of FCI	2,250,986	2.25	13.72
Total fixed operating costs				12.13	73.96

4.4 Discounted Cash Flow Analysis and the Minimum Fuel Selling Price

4.4.1 Discount Rate, Equity Financing, and Other Financial Metrics

Consistent with prior NREL TEA studies, the discount rate (also the IRR in this analysis) was set to 10% and the plant lifetime was set to 30 years. The 10% rate is consistent with all platforms across the BETO portfolio, and more context on its basis is discussed in prior reports [10, 11]. Also in keeping with other recent TEA reports, it was assumed that the plant would be 40% equity financed. The terms of the loan were maintained at 8% interest for 10 years. The principal is taken out in stages over the 3-year construction period. Interest on the loan is paid during this period, but principal is not paid back (this is another n^{th} -plant assumption, which says that this cash flow comes from the parent company until the plant starts up). Figure 11 illustrates the sensitivity of MFSP to the percentage of equity financing and the after-tax discount rate (the IRR). All other key financial metrics were also maintained consistently with prior NREL TEA reports, and more details and accompanying context may be found there [10, 11]. The primary exception is the federal tax rate, now reduced to 21% per the updated 2017 corporate tax code changes. Key parameters are summarized as follows:

- MACRS depreciation schedule, 20-year recovery period for steam boiler, 7-year period for all other facility equipment
- 21% federal corporate tax rate
- 3-year construction period: 12 months planning/engineering, followed by 24 months construction and commissioning (following n^{th} -plant economics)

- 6-month startup time
- Working capital = 5% of FCI.

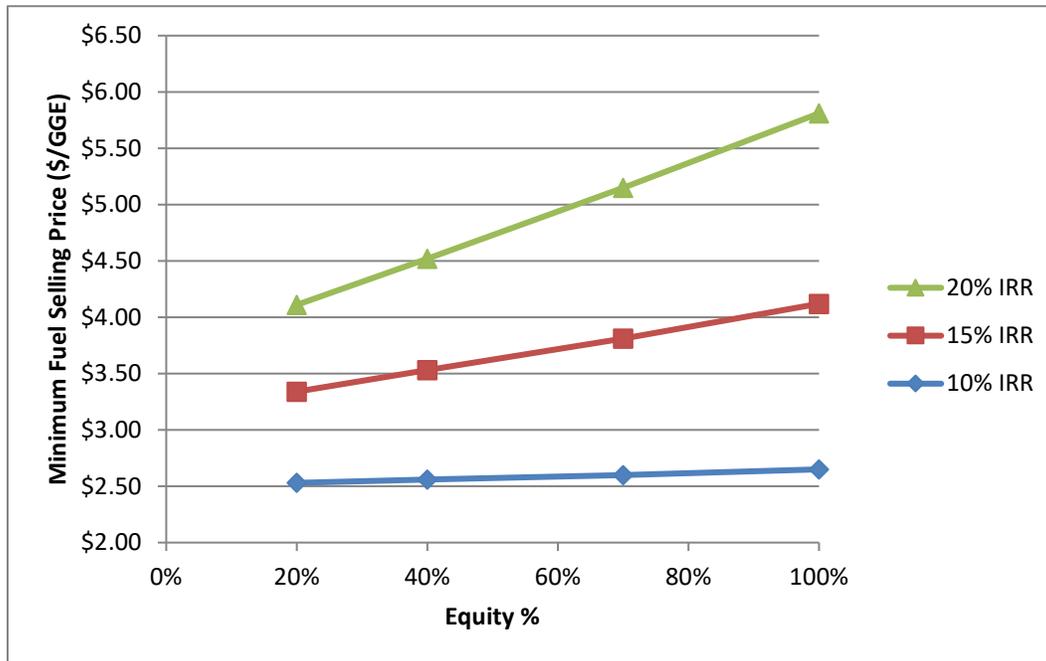


Figure 11. Sensitivity of MFSP to IRR and % equity. (8% interest on a 10-year loan.)

4.4.2 Base Case TEA Results

Based on the TEA parameters summarized above, the resulting **MFSP of total fuel products is estimated at \$2.56/GGE** for the polyurethane coproduct base case (2016 dollars), including all fuel-range product cuts from distillation adjusted by heating values (calculated in the Aspen model) to gasoline equivalents.

Table 18 summarizes the yields and conversion costs for the present effort. According to the methodology of Cran [14], the expected accuracy of the overall TCI analysis at least as applicable to more rigorous historical NREL design reports is $\pm 25\%$ (although with this report focused on more preliminary technology concepts and cost inputs, it is difficult to assert an expected TCI accuracy here). If such an uncertainty range were maintained here for the TCI, the impact on the cost of total fuel would be $\pm \$0.68/\text{GGE}$. The complete discounted cash flow summary worksheets are shown in Appendix B. The MFSP can be further broken down into the cost of each process area. Figure 12 illustrates the contribution to the overall cost by process area and capital, operations, and fixed costs (the bar for feedstock reflects the single feedstock cost of \$486/ton AFDW delivered to the CAP facility and is not broken down here).

Table 18. Summary of Yields, Rates, and Conversion Costs for PU Coproduct Base Case

Parameter	Value
Feedstock rate from algae farm	565 tons/day (AFDW)
Online time	7,920 h/yr (90% online factor)
Total fuel yield	88.0 GGE/dry ton feedstock
Total fuel production rate	16.4 MM GGE/yr
<i>Diesel-range production rate</i>	14.8 MM GGE/yr
<i>Naphtha-range production rate</i>	1.6 MM GGE/yr
Coproduct yield (PU)	475 lb/dry ton feedstock
Coproduct output rate (PU)	89 MM lb/yr
Total variable OPEX excluding coproduct ^a	\$170 MM/yr
Total coproduct revenue	\$189 MM/yr
Total fixed OPEX	\$12 MM/yr
Total equipment cost	\$152 MM
Total capital investment	\$339 MM
TCI per annual gallon	\$20.69/GGE
Minimum Fuel Selling Price	\$2.56/GGE
Feedstock contribution	\$5.48/GGE
Fuel conversion contribution	\$2.82/GGE
Coproduct conversion contribution	-\$5.74/GGE

^a Excludes coproduct revenue including PU and all nutrient/CO₂ recycles to cultivation (next row). Not including catalyst replacement schedules.

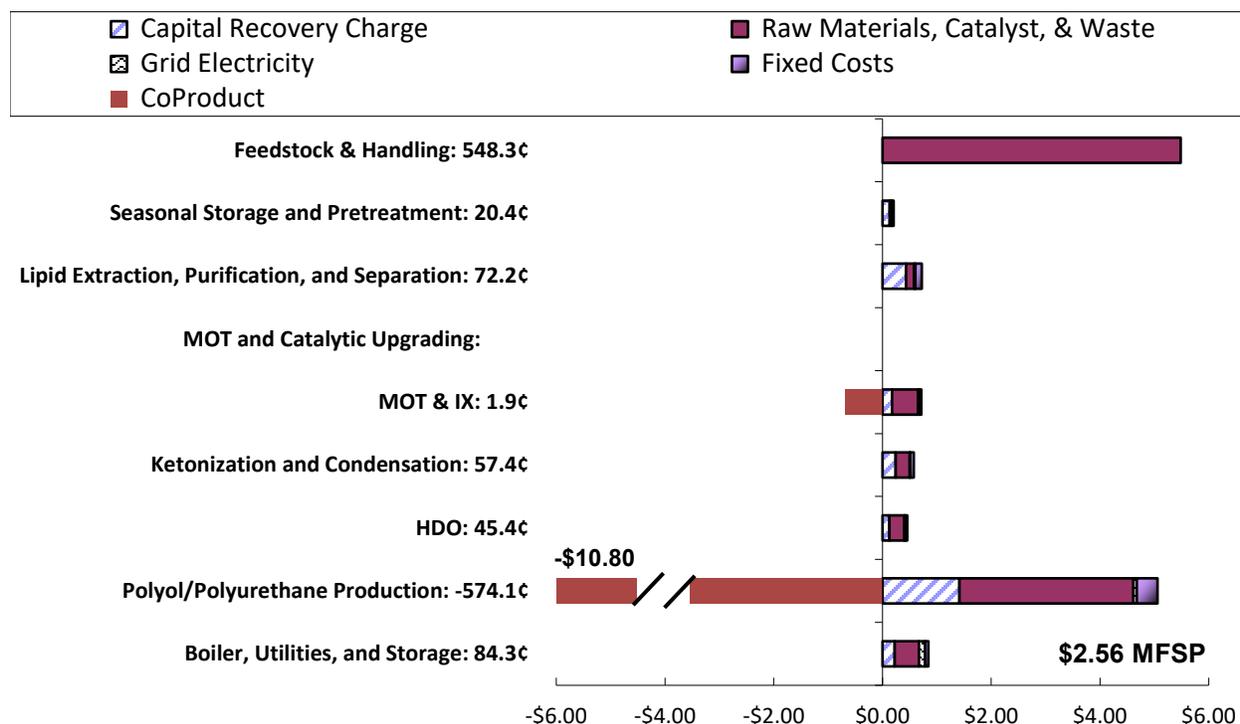


Figure 12. Base case cost contribution details from each process area (per GGE total fuel products)

As is typical for algal biofuel TEA models, feedstock cost represents the single largest contributor to MFSP, roughly \$5.5/GGE attributed to the delivered \$486/ton cost from upstream cultivation. This highlights the importance of including coproducts for reducing biorefinery fuel costs, as feedstock contribution to MFSP is strictly a reflection of feedstock cost and GGE/ton fuel yields, where even with a relatively high fuel yield of 88 GGE/ton in this process, the cost of algal biomass alone without any conversion processing costs already far exceeds the \$2.5/GGE MFSP goal. *The latter would require a 2.5-fold fuel yield improvement over that modeled here (over 200 GGE/ton, likely exceeding theoretical limits for what could be possible), just to reduce feedstock contributions alone below \$2.5/GGE*, again without any additional processing costs yet included. Barring a very high carbon/energy content algal biomass composition (generally implying an exceedingly high lipid content), such a fuel yield is not theoretically possible under any conversion processing approach. Thus, the only means to otherwise reduce MFSP toward the \$2.5/GGE threshold would be (1) lower-cost biomass (not likely possible for algae cultivated in traditional systems, aside from wastewater applications at more limited national scalability), or (2) additional revenues to offset these costs through higher-value coproducts produced alongside fuels. Indeed, the addition of the PU coproduct is shown in Figure 12 to more than offset the entire feedstock cost contribution to MFSP plus an additional \$0.26/GGE (thus leaving roughly \$2.80/GGE available for non-feedstock conversion cost contributions to MFSP to still enable achieving the \$2.5/GGE goal). As is typical for value-added coproducts [11, 52], the processing costs for the PU production train add considerable expenses to the biorefinery (over \$5/GGE), but are outweighed by coproduct revenues on a roughly 2:1 basis, translating to an overall net negative MFSP contribution of \$5.74/GGE.

Beyond the strong drivers for feedstock costs and coproducts, Area 300 (MOT and catalytic upgrading) represents the largest MFSP contribution within the biorefinery fuel production operations at \$1.05/GGE, in turn driven primarily by ketonization and condensation followed by HDO. MOT incurs the largest processing costs of all Area 300 operations, but is largely offset by coproduct credits for CO₂ as well as N/P nutrient recycles following ion exchange (IX), which are allocated to the MOT operation. The main cost contributors to all Area 300 steps are variable operating costs, namely oxygen (MOT), sulfuric acid (IX regeneration, allocated to MOT), and hydrogen (HDO). Area 500 (boiler/utilities) follows as the next strongest contributor, driven most substantially by natural gas boiler demands, and then Area 200 (lipid extraction and processing), primarily reflecting capital costs. Notably, Area 100 (storage and pretreatment) represents the smallest contributions to MFSP, at \$0.20/GGE, reflecting relatively low-cost wet seasonal storage and substantially less costly operations for flash hydrolysis than NREL's prior CAP approach utilizing dilute acid pretreatment reactors [7].

5 Analysis and Discussion

5.1 Cost Sensitivity Analysis

A single-point sensitivity was performed using the variables and limits shown in Table 19. The baseline for all variables 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 13.

Table 19. Assumptions Varied in the Sensitivity Analysis

Area	Assumption	Min MFSP	Baseline	Max MFSP
Biomass	Feedstock composition	HLSD	HCSO	HPSD
	Feedstock cost (\$/ton AFDW)	\$434	\$486	\$566
	Feedstock composition + cost (\$/ton AFDW)	HLSD @ \$566	HCSO @ \$486	HPSD @ \$434
	Scale (vs, base case 548 tons/day AFDW)	2X	1X	0.5X
Flash Hydrolysis (FH)	FH solids loading (wt% AFDW)	20%	18%	10%
	FH pressure (psig)	1,000	1,200	2,000
	FH residence time (seconds)	-	10	60
	FH protein solubilization (wt%)	75%	64%	50%
Lipid Extraction	Lipid extraction yield (wt% FAME lipids)	99%	96%	85%
	Extraction CSTR residence time (min)	10	15	20
MOT	Carbs to C ₂₊ acids yield (carbon %)	85%	71.3%	35%
	Protein to C ₂₊ acids yield (carbon %)	85%	71.3%	35%
	Overall MOT to C ₂₊ acids yield (carbon %)	85%	71.3%	35%
	MOT residence time (min)	30	40	50
	MOT temperature (°C)	200	225	-
	MOT oxygen consumption		Base	2X
	MOT CAPEX	-50%	Base	50%
	IX NaOH/H ₂ SO ₄ regeneration molar ratio	0.33	1	3
Fuel Upgrading	MOT/IX recycle of bioavailable N/P to ponds	-	90%	50%
	Ketonization temperature (°C)	330	350	370
	Ketonization catalyst	Mn/ZrO ₂	0.1% Pt/Nb ₂ O ₅	-
	Ketonization WHSV (hr ⁻¹)	2.5	1.7	1
	C ₂₊ acids to ketones yield (carbon %)	-	75%	50%
	Condensation single pass conversion		60%	80%
	HDO WHSV (hr ⁻¹)	2	1	0.5
	H ₂ price (\$/kg)	1	1.61	2
Polyurethane	Catalysis train overall capital cost	-50%	Base	50%
	Polyol synthesis CAPEX cost	-25%	Base	25%
	PU synthesis CAPEX cost	-25%	Base	25%
	Warehouse storage time	3 days	5 days	7 days
	TDI:Polyol feed ratio	-10%	Base	+10%
	TAG:FFA ratio	30:70	50:50	70:30
	Polyurethane value	\$2.26	\$2.04	\$1.80
Economics	Total capital investment	-25%	-	25%
	Income tax rate		21%	35%
	On-stream factor	96%	90%	80%

As shown in Figure 13, delivered biomass feedstock composition plays a critical role in overall MFSP for this process, with the high-lipid *Scenedesmus* (HLSD) composition as presented in Table 2 translating to a negative MFSP, or high-protein *Scenedesmus* (HPSD) adding nearly \$5/GGE to the base case MFSP reflective of high-carbohydrate *Scenedesmus* (HCSD). This is primarily driven by resulting PU coproduct yields as a function of lipid content, given the substantial role this coproduct plays in overall biorefinery economics as shown in Figure 12. **Such a strong sensitivity is *not* a reflection of this CAP processing approach itself, i.e., in comparison to prior CAP approaches focused on different process operations for utilization of carbohydrates, lipids, or protein, but rather is a reflection of the strong dependency that any configuration exhibits on the yield ratio of the high-value PU coproduct relative to fuels** (a function of algal lipid composition). This is evidenced by the allocated MFSP contributions *excluding* coproduct exhibiting a much smaller variance of $\pm\$0.74/\text{GGE}$ or 9% relative to the base case (\$8.30/GGE MFSP contributions before including the PU coproduct train, increasing to \$9.04/GGE or decreasing to \$7.83/GGE for HPSD and HLSD, respectively). Accordingly, the apparent strong MFSP dependency on algal composition would likewise manifest in other CAP configurations or any other processing approach incurring substantial swings in the yield of PU or another high-value coproduct.

The impact of composition is somewhat negated if considered simultaneously with feedstock cost variances, where feedstock costs were varied (both alone and in conjunction with composition) between \$434/ton up to \$566/ton, which represent upstream algae farm MBSPs attributed to cultivation productivity of 30 and 20 g/m²/day respectively, relative to the base case projection of 25 g/m²/day (\$486/ton MBSP at 18% harvested solids concentration) [4]. Thus, if cultivation productivity only achieved 20 g/m²/day but at the improved HLSD compositional quality (given known tradeoffs between cultivation productivity and composition), this would still translate to a substantial benefit on CAP biorefinery economics at an MFSP of under \$1/GGE. Alternatively, if cultivation productivity could instead reach 30 g/m²/day but at the higher-protein/lower-lipid HPSD compositional basis, biorefinery MFSP would still increase by over \$4/GGE. The same feedstock cost variances taken in isolation also still carry a significant cost impact, but less-so than the compositional variances as evaluated here.

Likewise, other parameters also impacting PU coproduct yields exhibit significant influence over the MFSP, including lipid extraction yield (namely on the lower side, with an \$0.84/GGE penalty if extraction yield reduces from 96% to 85%), TAG:FFA ratio (impacting MFSP by \$2-3/GGE moving in either direction to 30:70 or 70:30), and PU value (over \$1/GGE MFSP impact attributed to recent historical low and high price points for PU market values). Given the clear importance that this coproduct plays in the overall economic viability of this system, a separate sensitivity analysis expanding on coproduct yield and price drivers is presented below.

Beyond feedstock composition, cost, and PU coproduct-related parameters, the next factors that exhibit moderate to significant influence on MFSPs are related to fuel train yields, particularly around the MOT and ketonization steps, namely MOT yields to ketonizable acids and ketonization yields to ketones. Given fairly high uncertainties in the model for MOT yields due to the dearth of data for this operation (generally in an earlier stage of conceptual development), overall ketonizable acid carbon yields (defined here as C₂₊ acids, which contribute to fuel-range hydrocarbon yields downstream) from both carbohydrates and proteins were increased by 20% and decreased by 50% (carbon yields of 85% and 35%), translating to an MFSP savings of

\$0.21/GGE or penalty of \$1.02/GGE, respectively. Protein and carbohydrate yields were also manipulated separately, showing that carbohydrate yield had a significantly higher impact on MFSP (as expected, since the carbohydrate content in the algae feed is higher than the protein). These results demonstrate the importance achieving target yields moving forward for MOT conversion. Given the uncertainty around the oxygen requirement for MOT, a sensitivity for 2x oxygen consumption was also considered; this resulted in a \$0.29/GGE increase in MFSP. The ketonization carbon yield also had an impact on MFSP, though to a lesser degree than in MFSP. Since the base case value of 75% is equal to the theoretical maximum carbon yield, only the low yield case was analyzed; decreasing the carbon yield to 50% resulted in a \$0.45 increase in MFSP.

As typically observed in prior TEA studies, underlying economic/operational parameters also incur fairly significant impacts to MFSP. As noted above, a $\pm 25\%$ uncertainty in facility TCI translates to a $\pm \$0.63/\text{GGE}$ impact on MFSP. Increasing the incoming tax rate to 35% (as is consistent with prior models, preceding recent changes to the U.S. tax code) would result in a \$0.19/GGE increase. Finally, if the facility on-stream operating factor were increased from 90% to 96% (350 days/year), MFSP would reduce by \$0.20/GGE due to more efficient utilization of facility-installed capital costs. Alternatively, if this were reduced to 80% (292 days/year), for example due to factors such as upstream cultivation upsets, pond freezes, or contamination events beyond available biomass storage capacity at that time, MFSP would increase by \$0.44/GGE.

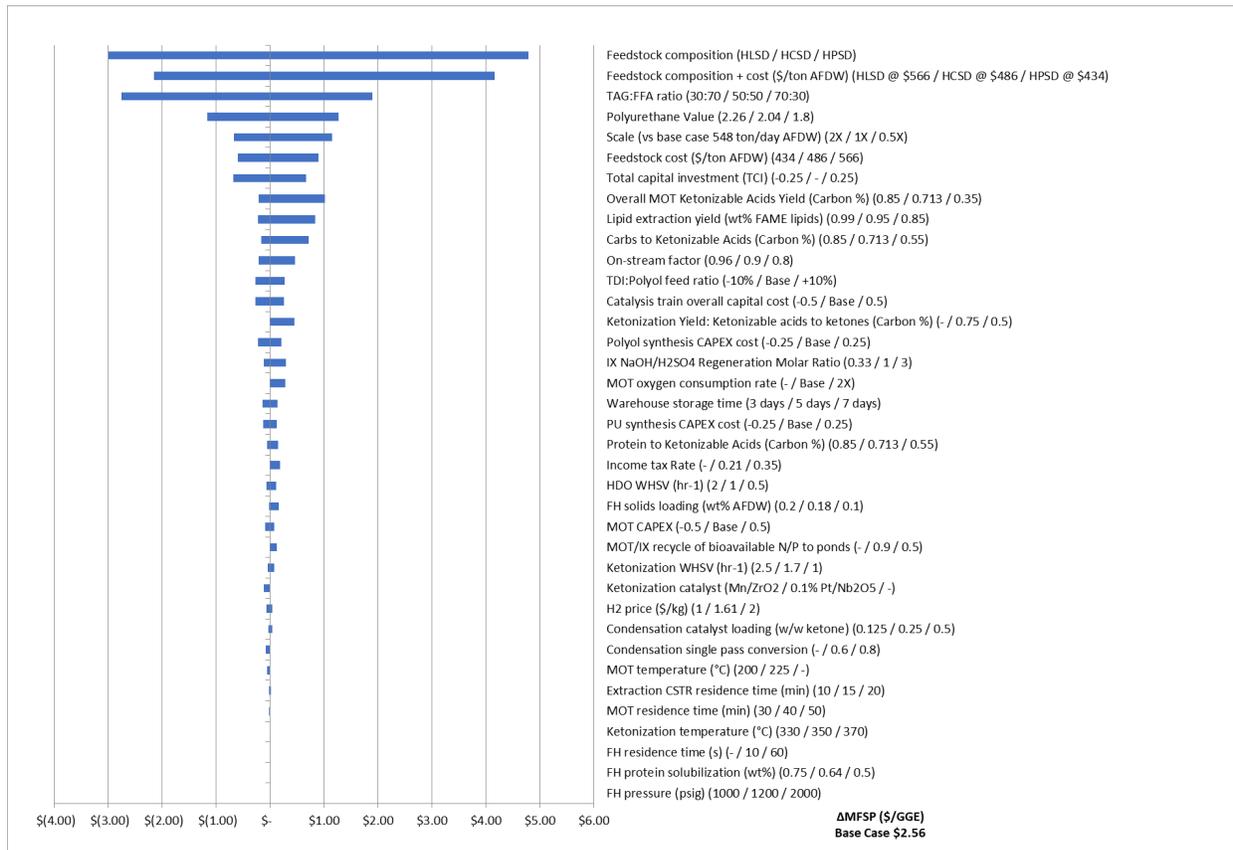


Figure 13. Tornado plot presenting results of the single-point sensitivity analysis on MFSP cost (\$2.56/GGE reference case)

5.2 Sensitivity Scans for Polyol vs. Polyurethane Values/Yields

As discussed previously, the base case scenario follows the TAG coproduct upgrading train all the way through to polyurethanes, in order to provide a complete picture of the economic potential for that process route and also given design/cost inputs from Nexant furnished across such a fully integrated process. However, as also cautioned by Nexant, given that it is not typical for an industrial PU process to be fully integrated between production and conversion of polyol intermediates, and also in light of uncertainties regarding final PU foam properties if sourced solely from algal triglycerides (i.e., commercial processes making use of other [nonalgae] renewable oils commonly include blending with petroleum polyols to achieve targeted PU properties [53, 54]), this assessment also considers drawing the product boundary earlier at polyols. Under that approach, all downstream costs for PU synthesis are removed, including reactor capital costs, TDI reactant co-feeds, PU curing/storage warehouse costs, and dedicated PU labor expenses. In this section, TEA implications are presented for both polyol and PU coproduct boundaries across a range of product values as well as yields, with the latter dictated by the underlying TAG:FFA ratio constituting the overall lipid fraction (Table 2).

Estimating prices for commodity chemicals is challenging as these prices can vary significantly over the course of a year, given factors such as fluctuations in feedstock costs, shifts in supply/demand, unplanned outages at production facilities, etc. Additionally, the market for PU products includes a large range of different product materials, properties, and end uses spanning elastomers, rigid and flexible foams, and other applications. In the context of the current process reflecting synthesis chemistries targeting flexible foams, prices have fluctuated less dramatically than other chemical products. Still, given the large impact coproduct prices exhibit on the resulting MFSP, this analysis considers the implications on biorefinery economics across a range of product prices. For the base case, the PU price was set at \$2.04/lb (2016 dollars) representing a 5-year average price for commodity flexible foam. This was varied here between \$1.80/lb (representing the minimum price over that 5-year timeframe) up to \$2.26/lb (a value on the upper range of prices that may be found in published literature [55, 56]). For polyol coproduction, the polyol product value was varied from a low of \$1.29/lb (based on published values for petroleum-derived polyols [57], corresponding closely with recent 5-year average prices, 2016 dollars) up to a high of \$1.81/lb (representing a high-quality polyol price inclusive of bio-derived “green premiums” as supported at \$4/kg = \$1.81/lb or more in certain foam product applications, per feedback from industry [58]; as similar feedback was not available for polyurethanes such a “green premium” was not considered in that case).

Figure 14 presents a TEA sensitivity scan for the above variables. First, this figure highlights the increased value for polyurethanes relative to polyols, with the upper limit for polyol price range as discussed above matching the lower limit for polyurethanes, albeit recognizing the increased production costs and mass addition from TDI in producing a unit mass of PU. This also demonstrates the high sensitivity for overall biorefinery economics on coproduct yields, with MFSPs reducing significantly as the TAG:FFA ratio increases (i.e., ratio of coproduct versus fuel produced from algal lipids) and also becoming more sensitive to the product value (steeper MFSP curve) at higher coproduct yields. At a ratio of 30:70 TAG:FFA, the \$2.5/GGE MFSP target cannot be achieved under the integrated CAP process schematic evaluated here, for either polyol or PU coproduction. However, at the 50:50 ratio, the PU curve intersects \$2.5/GGE at the midpoint of the PU price range (corresponding to the base case PU value of \$2.05/lb), but polyol

coproducts still cannot support achievement of the MFSP target. Finally, at a 70:30 TAG:FFA ratio, polyol coproduction could plausibly support the \$2.5/GGE target, albeit would require a value of roughly \$1.70/lb, which exceeds typical ranges for petroleum-derived polyols in approaching values that may be supported in some consumer markets where a bio-derived material garners a price premium as noted above. Notably, at this elevated TAG:FFA ratio, the PU curve would in fact translate to a *negative* MFSP at a PU value of \$2.02/lb (slightly below the base case PU value).

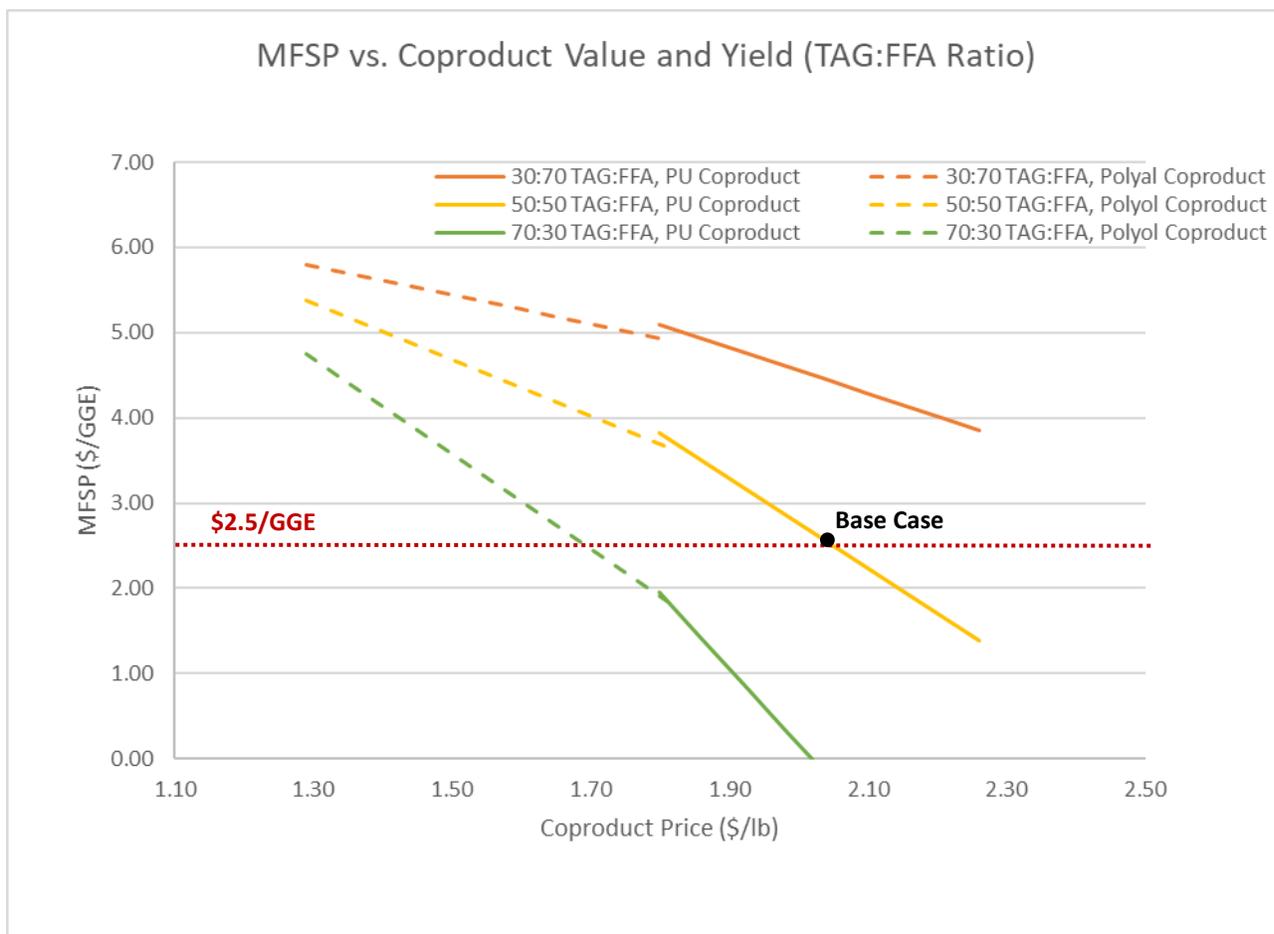


Figure 14. Sensitivity scan plot for MFSP vs. coproduct price at various TAG:FFA ratios

5.3 Sustainability Metric Indicators

This section presents primary sustainability metric indicators of the current conceptual process at the conversion stage. A seasonal average of inputs (raw materials and energy) and outputs (products, coproducts, emissions, and recycle streams to the algal ponds) are shown in Table 20. The information provided in Table 20 provides the necessary information required for performing life cycle analysis (LCA) modeling to quantify greenhouse gas (GHG) emissions and fossil energy consumption. The biorefinery GHGs and fossil energy consumption will be calculated separately under supply chain sustainability analysis efforts coordinated by partners at Argonne National Laboratory in the future and are not reported here. A complete well-to-wheel

or supply chain LCA evaluation is required to more fully understand the sustainability implications for the full supply chain based on this technology pathway, such as how the overall integrated biorefinery GHG emissions profiles compare to BETO goals relative to petroleum diesel. One key driver in the system LCA is the inclusion of a nonfuel coproduct (flexible polyurethane foam). When produced in significant quantities, as is the case here, it can strongly influence the overall GHG emissions (with the potential for substantial GHG credits relative to energy-intensive synthesis of a material such as fossil-derived polyol), depending on the coproduct handling method selected for the LCA [59].

Table 20. Input and Output Data for Modeled Conversion Facility.

(Note: Hourly rates shown below are based on annual averages over all modeled seasons.)

Products	Production Rate	Units
Diesel	5312	kg/hr (1865 GGE/hr)
Naphtha	577	kg/hr (207 GGE/hr)
Polyurethane	5074	kg/hr
NH ₃ Recycle	371	kg/hr
Diammonium Phosphate Recycle	177	kg/hr
CO ₂ Recycle (Fossil)	14490	kg/hr
CO ₂ Recycle (Biogenic)	9458	kg/hr
Resource Consumption	Flow Rate	Units
Biomass Feedstock (ash-free dry weight)	21365	kg/hr
Oxygen	5657	kg/hr
NaOH	54	kg/hr
Sulfuric Acid	2402	kg/hr
Hexane	16	kg/hr
Ethanol	91	kg/hr
Phosphoric Acid	59	kg/hr
Silica	6	kg/hr
Clay	12	kg/hr
Toluene	165	kg/hr
Hydrogen	222	kg/hr
Acetic Acid	527	kg/hr
H ₂ O ₂	586	kg/hr
Catalysts and Other Chemicals	Cost calculation	
Nitrogen	62	kg/hr
Toluene Diisocyanate	1798	kg/hr
DEOA	13	kg/hr
Surfactant	25	kg/hr
Process Water	159230	kg/hr
Energy Consumption	Rate	Units
Grid Electricity	8419	kW
Natural Gas	278	MM BTU/hr
Output Streams	Flow Rate	Units
Unreacted Solids	1951	kg/hr
Water Recycle	151140	kg/hr
Air Emissions	Flow Rate	Units
H ₂ O	139613	kg/hr

One takeaway from Table 20 is that there are virtually no direct emissions from the modeled process, other than the evaporative water losses from the cooling tower. The reason for this is that all the exit points for CO₂ can be collected and recycled to the algal ponds instead of being vented to the atmosphere. The CO₂ comes from just a few points in the process, with the majority coming from boiler flue gas (sourced from fossil fuels) and vents from the MOT, ketonization, and HDO sections (sourced from the algal biomass). Similarly, much of the water in the process can be recovered and recycled back to the ponds. Any potentially harmful components present in water recycles, such as ethanol, hexane, and various carboxylic acids, are diluted to the ppm level before making it back to the pond, at which point they will be diluted even further.

Table 21 summarizes the key sustainability metric indicators for the conversion process evaluated here. Biorefinery net water consumption in the modeled process comes from the makeup water for the cooling tower and high- and low-pressure steam loops; there are no other process water requirements. Despite this, there is still a relatively high water demand, mostly driven by a large cooling water requirement. This high demand may impose constraints on facility siting locations, requiring sufficient access to local water resources while also being located in areas suitable for meeting delivered feedstock compositional/cost targets discussed in Section 2.1.

Table 21. Summary of Sustainability Metric Indicators for the Modeled Conversion Process

Sustainability Metrics		
Fuel Yield by Weight of Biomass	88.0	GGE/dry ton biomass
Carbon Efficiency to Fuels	45%	% C in feedstock
Carbon Efficiency to Polyurethane ^a	19%	% C in feedstock
Electricity Import	4.1	KWh/GGE
Natural Gas Import	141.7	MJ/GGE
Water Consumption	20.3	gal/GGE
Water Consumption	3,822	m ³ /day

^a Represents the percentage of carbon in the feedstock that is incorporated into the polyurethane coproduct. Additional carbon in the coproduct is sourced from TDI and acetic acid and is not considered here.

It should be noted that the water consumption does not consider the 82% moisture in the biomass feed. From a sustainability perspective, the conversion facility should not be viewed as a standalone facility because it relies on the upstream algae farm for the feedstock. 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, and such losses during cultivation are made up with saline water, thus not counting against overall fresh water footprint [2]).

Figure 15 shows an overall carbon balance across the modeled process. From Figure 15, as well as Table 21, it can be seen that the fuel yield and carbon efficiency to fuels is lower than past CAP processes (e.g. 48% algal C yield to fuels under the prior CAP case [7]) due to a combination of factors. First, the carbon efficiency of the combined MOT, ketonization, condensation, and HDO is fairly low (56%) due to carbon being lost as formic acid (eventually

converting to CO₂ during ketonization) and as CO₂ in MOT, ketonization and HDO. This yield also includes the losses of lipids through the filter after MOT. In addition to this, half of the lipid fraction, which typically has a higher carbon efficiency to fuels, is diverted to the coproduct.

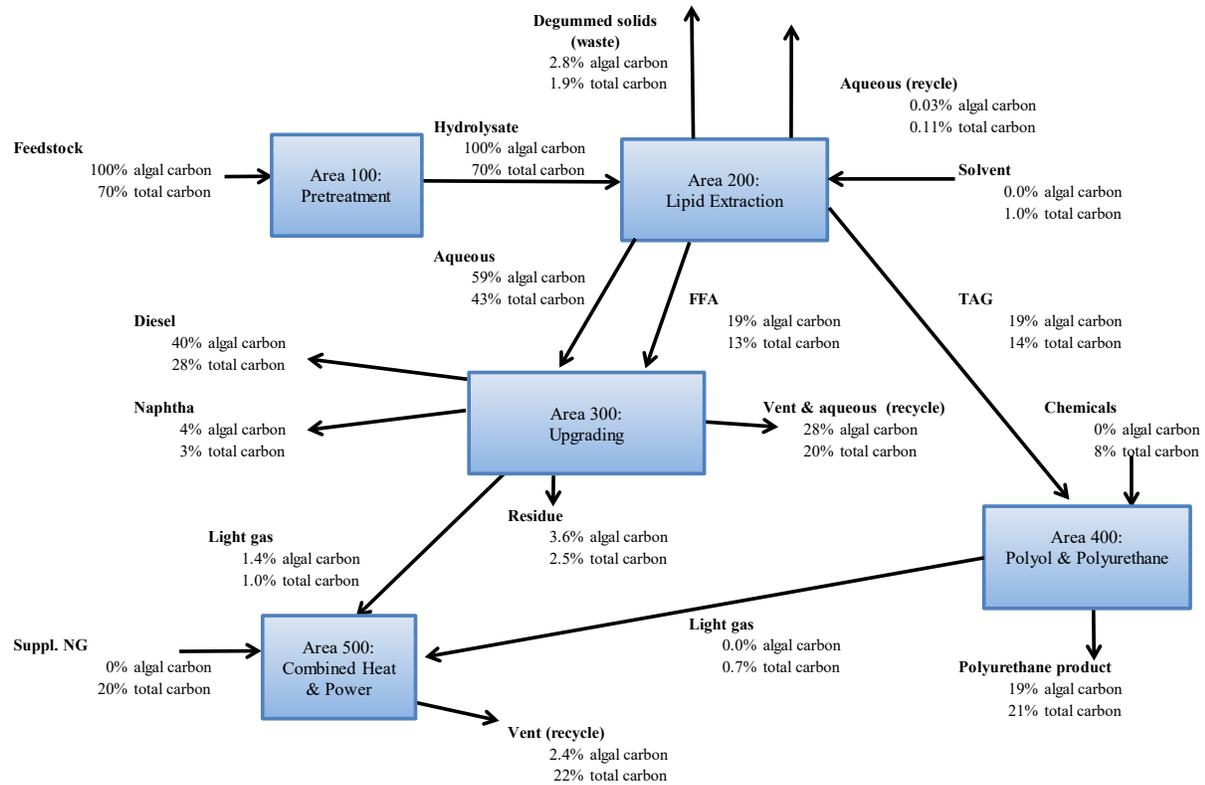


Figure 15. Carbon balance diagram tracking carbon flows across the modeled process (overall seasonal average flows)

6 Concluding Remarks

This report establishes a plausible conceptual design for achieving an ultimate \$2.5/GGE cost goal of upgraded renewable hydrocarbon fuels by 2030 via a new algae CAP conversion processing framework. Namely, the current pathway presents a way to diverge from NREL's prior CAP schematics predicated on producing and then fermenting monomeric sugars as a key pillar of the process, which inherently requires relatively high algal carbohydrate content to justify the expense of a dedicated sugar fermentation train, and accordingly relatively low protein content to leave enough room for carbohydrates as well as lipids (for upgrading to fuels and/or coproducts, still maintained here). The new CAP pathway replaces a higher-cost acid pretreatment reactor with a lower-cost flash hydrolysis vessel, and also consolidates two processing trains for separate carbohydrate (fuels/products) and protein (AD/CHP) utilization into a single process with flexibility to utilize both substrates in varying amounts and soluble/solid states, via MOT and catalytic upgrading to hydrocarbon fuels. The use of efficient wet lipid extraction based on demonstrated operations is maintained as another core pillar of this process, given the value of algal lipids in contributing either to high-GGE fuel yields with relatively minimal processing, or to convertible feedstocks for high-value polyol/polyurethane products (or in this case, both). While the present effort presents the resulting TEA implications for this process based again on a high-carbohydrate algal biomass feedstock (using a representative HCSD example case), the flexibility for accommodating alternative, potentially lower-cost, algal feedstocks with lower carbohydrates and higher protein is a beneficial attribute of this new CAP configuration which may be considered in more detail in a future update. However, we recognize that such an alternative composition would likely require investigating other or additional coproduct opportunities than strictly lipid-derived PU.

Although still considered a preliminary concept for such a new CAP design (in some cases not yet fully verified against experimental/analytical data which is still under development, namely for flash hydrolysis, MOT, and MOT acids catalytic upgrading), if the general operational parameters and yields projected for this process could ultimately be met in the future, a number of desirable traits for algae conversion to fuels could be met through this approach. These include (1) high fuel yields (88 GGE/ton); (2) high fractional conversion of algal carbon to fuels relative to coproducts (over a 2:1 ratio = 70%); (3) the ability to recover and recycle high fractions of nitrogen/phosphorous nutrients back to upstream algae cultivation; and (4) consolidating the number of individual steps and parallel processing trains relative to prior CAP iterations, to make for a less complex biorefinery facility, all while supporting achievement of \$2.5/GGE fuel selling prices based on algal biomass as the sole feedstock for conversion. This would require several key operational metrics to support high overall yields to both fuels and coproducts, including 71% overall carbon conversion of carbohydrates and protein to C₂₊ acids across MOT, 100% conversion of intermediate substrates across ketonization and catalytic HDO reactors plus 99% net conversion across condensation (inclusive of unconverted ketone recycling), and over 95% lipid extraction efficiency. To date, the lipid extraction yield metric has been experimentally demonstrated but the others have not.

The end result of the techno-economic analysis was a modeled MFSP of \$2.56/GGE for the base case process with coproduction of algal TAG-derived polyurethanes (2016 dollars), associated with a fuel yield of 88.0 GGE/ton and PU coproduct yield of 475 lb/ton (the latter which includes mass addition from external co-feeds, primarily TDI). Feedstock costs for production of algal

biomass from upstream farms constitute the largest contributor to this MFSP, at \$5.48/GGE. Such a large cost contributor is not new, but does reiterate the importance of incorporating value-added coproducts to garner additional revenue for the biorefinery, given that the base case algal biomass cost (\$486/ton) would dictate the need for impossibly high conversion fuel yields above 200 GGE/ton merely to bring costs for biomass feedstock alone below \$2.5/GGE without including any additional conversion processing costs. Accordingly, the inclusion of algal-derived PU coproducts provides a net reduction of \$5.74/GGE to the MFSP after including both PU train production costs and coproduct sale revenues. The MOT and catalytic upgrading process steps contribute the next most significant additions to MFSP at \$1.05/GGE overall, followed by the boiler/utilities (\$0.84/GGE), lipid extraction/fractionation (\$0.72/GGE), and finally, seasonal storage and pretreatment (\$0.20/GGE). The seasonal storage and pretreatment costs are minimal, in part because they are tied to the use of a less costly flash hydrolysis vessel versus prior work with a dilute acid pretreatment reactor.

A sensitivity analysis was conducted, highlighting that any parameters influencing yields of the high-value PU coproduct relative to fuels exhibit significant impact on overall biorefinery economics/MFSPs, including feedstock composition (weighted most heavily by algal lipid content and associated PU production potential), lipid TAG:FFA ratio (similarly influencing PU versus fuel production with TAGs routed to PU and FFAs to fuels), PU value, and lipid extraction yield. Processing scale also strongly influences MFSP, by way of economy-of-scale impacts around capital expenses as well as labor costs. For technical parameters related to the fuel train, the MOT operation carries relatively higher uncertainty around yields and component distributions, translating to high MFSP sensitivities attributed to overall MOT yields and more specifically carbohydrates to C₂₊ acid yields. Finally, feedstock cost also weighs strongly on MFSP sensitivities, given the large cost driver for this parameter noted above. Notably, for a lower feedstock cost reflective of 30 g/m²/day cultivation productivity upstream coupled with a higher-protein/lower-lipid composition (versus the base case HCSD composition at 25 g/m²/day), MFSP would still increase significantly, and vice-versa for a higher feedstock cost reflecting 20 g/m²/day coupled with low-protein/high-lipid biomass content, when maintaining the same lipid PU coproduction assumptions as in the base case. Again, this is more strongly a function of PU coproduct impacts than the CAP processing schematic, and implies that while this CAP approach allows flexibility to accommodate a higher-protein algal feedstock, it would require modifications to the coproduct(s) of focus in order to achieve the \$2.5/GGE MFSP target. Finally, rather than a fully integrated process integrating TAG upgrading to polyols and then to PU at a high product value, the MFSP goal could still potentially be maintained when moving the coproduct boundary to end at polyols (lower-value than PU), albeit would require a higher TAG:FFA ratio as well as a polyol selling price generally exceeding that for standard commodity polyols as may be supported in certain markets for a bio-renewable price premium.

Moving forward, a number of gaps and uncertainties exist, as do areas for improvement, when considering how to prioritize future efforts in R&D planning and TEA model refinements (as also may be appropriate to upgrade this conceptual report to a more formal target-setting “design report” in the future). These include:

1. **Improve understanding and fidelity of MOT upgrading:** This feasibility TEA exercise made simplifying assumptions based on early experimental work and published literature regarding yields and speciation of carboxylic acid products from both carbohydrates and

protein. Further data is needed to verify the reasonableness of these assumptions, or to improve the number or type of acids being reflected based on analytical mass balances. This includes the fate of carbohydrates and protein (both solid and soluble components), as well as residual lipids (both neutral and polar lipid components). It will also be important to verify the magnitude of reaction enthalpies predicted by Aspen, given a relatively large heat evolution reflected in the model, which suffices to cover all LP steam demands in the overall biorefinery. Additionally, a key area for future experimental and TEA investigation will focus on better understanding and optimizing tradeoffs between MOT substrate concentration, reactor design/pressure, and oxygen sourcing options (pure oxygen, air, oxygen-enriched air, or alternative oxidants such as peroxide or ozone). Early proof of concept data has begun to show promising MOT yield potential, albeit to date under low substrate concentrations that so far are not yet commercially practical, representing a priority for further development.

2. **Refine process details for post-MOT catalytic upgrading:** It is also not currently well-understood how the MOT acid products combine through ketonization, what types of ketone structures are produced, and how sensitive the reactions are to the ratios of feed acids (reflecting the ratios of incoming biomass constituents through MOT). Further experimental verification is required on the ketonization catalyst tolerance to water, given a difference in processing aqueous acid feed streams here versus pure acids in NREL's biochemical design case process. Likewise, additional experimental and analytical work will be required to quantify and optimize condensation product yields and performance, including catalyst and solvent options to maximize per-pass yields and minimize costs. It will also be important to verify (1) the capacity for the hydrotreating step to simultaneously accommodate both condensation products and lipids (FFAs) in a single reaction step, (2) the associated catalyst performance, and (3) the activity for performing both HDO and HI reactions so to avoid producing lower-value straight-chain paraffins.
3. **Further refine polyurethane TEA details, understand product properties, and consider alternative synthesis options:** While chemistries and manufacturing logistics are more straightforward for lipid upgrading to epoxides/polyols, such details become more complicated and feed/product-specific in upgrading polyols to polyurethanes, with a range of different PU types and associated manufacturing process differences. While NREL's subcontract with Nexant has improved our understanding for what such a process is expected to look like at commercial scale for "traditional" flexible foam PU synthesis on a high level, further room exists to improve the granularity and model fidelity in reflecting process, design, and cost details for key unit operations. Additionally, it is recognized that the final PU product value will depend on the properties of the produced foam. If it is insufficient based on the algal TAG polyol alone, blending with conventional petroleum polyols may be required to achieve desired product characteristics; product testing and/or feedback from industry would be required to address such questions. Finally, NREL experimental efforts are pursuing more novel routes to make the PU product fully renewable through non-isocyanate chemistries utilizing carbonated epoxides and amines (which may be produced from a fraction of algal proteins). This option was not reflected in the base case TEA given uncertainties for both processing costs as well as product properties/values, but may be considered in future TEA model updates as more work is conducted on that route.

4. **Evaluate TEA potential and process modification requirements for high-protein biomass:** Given continued challenges in improving both algal cultivation productivity (biomass cost) and compositional quality (biomass value) together, further TEA and experimental efforts are planned to understand optimal CAP configurations to accommodate higher-protein algal feedstocks than the “HCSD” baseline. A follow-on study to this one may be conducted in the future to investigate such a scenario and determine what process or coproduct changes would be required to highlight a path toward maintaining achievement of \$2.5/GGE MFSP goals.

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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 seasonal throughput case.

Mechanical Equipment List	Base Quoted Costs			Base Unit Sizing						Unit Costs		
	EQUIPMENT TITLE	Num Req'd	Year of Quote	Purch Cost (All units)	Scaling Value	Units	Scaling Exp	Inst Factor	New Value	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year
Seasonal Storage and Pretreatment												
Seasonal Storage												
Wet Seasonal Storage Pit		2014	\$34,300	264170	gal	1	1	43853629.8	166.0	\$5,693,983	\$5,353,984.76	\$5,353,985
Liner Cover		2014	\$1	1	ft^2	1	1	781691.0	781691.0	\$508,099	\$477,759.58	\$477,760
Flash Hydrolysis												
P2P HeatX - FH Preheat to FH Water vapor	1	2007	\$432,519	12.3	Gcal/hr	0.6	1.6	8.8	0.7	\$353,766	\$364,741.46	\$583,586
Flash Hydrolysis Trim Heater	1	2009	\$41,000	-2	Gcal/hr	0.7	2.2	-3.7	1.8	\$62,927	\$65,314.85	\$143,693
Flash Hydrolysis Reactor	1	2016	\$10,400	105000	kg/hr	0.6	8.7	123668.2	1.2	\$11,473	\$11,472.94	\$99,616
H2O Flash Tank	1	2013	\$90,000	12189	kg/hr	0.7	1.5	123668.2	10.1	\$455,665	\$435,102.53	\$652,654
Pretreatment: Balance of Plant	1	2013								\$528	\$528	\$4,582
<i>INSERT NEW ITEMS ABOVE THIS LINE</i>												
Seasonal Storage and Pretreatment Totals:										A100 Totals:	\$6,708,904	\$7,315,875

Mechanical Equipment List	Base Quoted Costs			Base Unit Sizing					Unit Costs			
	Num Req'd	Year of Quote	Purch Cost (All units)	Scaling Value	Units	Scaling Exp	Inst Factor	New Value	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
Lipid Extraction, Purification, and Separation												
Extraction												
Hexane Flash Tank	1	2013	\$90,000	12189	kg/hr	0.7	1.5	169077.3	13.9	\$567,185	\$541,589.91	\$812,385
Hexane Condenser	1	2010	\$487,000	23.2	Gcal/hr	0.6	2.8	19.5	0.8	\$438,167	\$430,927.46	\$1,206,597
CSTR 1 Vessel Agitator	2	2013	\$22,000	2000	gal	0.7	1.5	16613.0	8.3	\$96,831	\$92,461	\$138,692
					g/cc			0.8				
CSTR 2 Vessel Agitator	2	2013	\$22,000	2000	gal	0.7	1.5	15975.2	8.0	\$94,214	\$89,962	\$134,943
					g/cc			0.8				
CSTR 3 Vessel Agitator	2	2013	\$22,000	2000	gal	0.7	1.5	15825.7	7.9	\$93,596	\$89,372	\$134,058
					g/cc			0.8				
CSTR 1 Vessel	2	2009	\$157,600	2000	gal	0.7	1.8	16613.0	8.3	\$693,662	\$719,979	\$1,295,962
					g/cc			0.8				
CSTR 2 Vessel	2	2009	\$157,600	2000	gal	0.7	1.8	15975.2	8.0	\$674,913	\$700,519	\$1,260,933
					g/cc			0.8				
CSTR 3 Vessel	2	2009	\$157,600	2000	gal	0.7	1.8	15825.7	7.9	\$670,486	\$695,923	\$1,252,661
					g/cc			0.8				
Centrifuge 1	1	2009	\$500,000	250	gpm	1	1	1107.5	4.4	\$2,215,061	\$2,299,097	\$2,299,097
					g/cc			0.8				
Centrifuge 2	1	2009	\$500,000	250	gpm	1	1	1065.0	4.3	\$2,130,030	\$2,210,840	\$2,210,840
					g/cc			0.8				
Centrifuge 3	1	2009	\$500,000	250	gpm	1	1	1055.0	4.2	\$2,110,095	\$2,190,148	\$2,190,148
					g/cc			0.8				
Solvent Recovery Column	3	2009	\$714,000	5599	kg/hr	0.85	2.4	2856.4	0.5	\$402,952	\$418,240	\$1,003,775
Solvent Recovery Reboiler	3	2009	\$150,000	5599	kg/hr	0.85	3.3	2856.4	0.5	\$84,654	\$87,865	\$289,956
Ethanol Distillation Equipment	1	2009	\$1,135,667	30379	kg/hr	0.6	2.4	35562.5	1.2	\$1,248,252	\$1,295,608	\$3,109,460
Reboilers	1											
Beer Condenser	1											
Feed pump	2	2011	\$258,600	29274	kg/hr	0.8	1.42	140940.2	4.8	\$909,224	\$840,919	\$1,197,969
Heat exchanger	1	2009	\$41,000	-2	Gcal/hr	0.7	2	-0.2	0.1	\$8,941	\$9,280	\$18,561
Heat exchanger	1	2009	\$41,000	-2	Gcal/hr	0.7	2	-0.5	0.2	\$14,511	\$15,061	\$30,123
Heat exchanger	1	2009	\$41,000	-2	Gcal/hr	0.7	2	-0.5	0.2	\$14,512	\$15,063	\$30,126
Column Preheater-	1	2009	\$41,000	2	Gcal/hr	0.7	2	8.8	4.4	\$115,654	\$120,041	\$240,083
Heat exchanger	1	2009	\$41,000	2	Gcal/hr	0.7	2.2	18.4	9.2	\$194,033	\$201,395	\$443,068
Vacuum Distillation (TAG/FFA Separation)												
Heat exchanger - Degum Feed	1	2010	\$184,000	2	Gcal/hr	0.7	2.2	0.7	0.3	\$87,724	\$86,275	\$189,805
Bleaching/Degumming Unit	1	2013	\$6,400,000	109	gpm	0.6	1	35.4	0.3	\$3,258,345	\$3,111,309	\$3,111,309
					kg/hr			6103.5				
					g/cc			0.8				
Steam Ejector	1	2016	\$49,900	4636	kg/hr	0.6	1.6	22929.6	4.9	\$130,213	\$130,213	\$208,340
Vacuum Column	1	2016	908300	5500	kg/hr	0.7	1.75	5731.4	1.0	\$934,882	\$934,882	\$1,638,122
Reboiler	1	2010	\$184,000	-2	Gcal/hr	0.7	2.2	-2.3	1.1	\$201,152	\$197,828	\$435,222
Condenser	1	2009	\$82,000	2	Gcal/hr	0.7	2.2	2.1	1.1	\$86,115	\$89,382	\$196,640
P2P HX - TAG-FFA Vent/Extraction Column Preheat												
Bottoms pump	1	2009	\$9,200	22681	kg/hr	0.8	3.1	2826.3	0.1	\$1,739	\$1,805	\$5,595
Distillate Pump	1	2009	\$9,200	22681	kg/hr	0.8	3.1	2885.6	0.1	\$1,768	\$1,835	\$5,688
Heat Exchanger - Vent	1	2007	\$432,519	12.3	Gcal/hr	0.6	1.6	16.1	1.3	\$507,951	\$523,709	\$837,935
Lipid Extraction, Purification, and Separation Totals:										A200 Totals:	\$17,169,011	\$23,909,109

Mechanical Equipment List	Base Quoted Costs			Base Unit Sizing						Unit Costs		
	Num Req'd	Year of Quote	Purch Cost (All units)	Scaling Value	Units	Scaling Exp	Inst Factor	New Value	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
MOT and Catalytic Upgrading												
Mild Oxidative Treatment												
MOT Feed Pump	3	2010	\$283,671	10982556	kg/hr	0.8	3.1	105733.3	0.0	\$6,912	\$6,798	\$21,074
P2P Heat Exchanger - Reactor Feed/Flash Liq	1	2010	\$92,000	7.5	Gcal/hr	0.7	2.2	11.4	1.5	\$123,386	\$121,348	\$266,965
MOT Trim Heater	1	2007	\$432,519	-12.3	Gcal/hr	0.6	1.6	-10.1	0.8	\$383,599	\$395,500.05	\$632,800
MOT Reactor (includes vent)	2	2016	\$2,598,000	100000	kg/hr	0.8	1.6	111256.2	1.1	\$2,831,461	\$2,831,461	\$4,408,585
MOT Effluent Flash	1	2013	\$90,000	12189	kg/hr	0.7	1.5	63495.4	5.2	\$385,749	\$272,854	\$409,281
MOT Effluent Flash	1	2013	\$90,000	12189	kg/hr	0.7	1.5	5300.7	0.4	\$50,246	\$47,978	\$71,967
Polishing Filter		2014	\$440,000	53204	kg/hr	0.9	1.8	15043.9	0.3	\$141,165	\$132,736	\$238,925
liquid flow								104034.9				
water flow								88991.0				
O2 Compressor	1	2011	\$1,103,700	14665.49	kg/hr	0.6	1.1	5672.8	0.4	\$624,240	\$577,345	\$652,400
Ion Exchange Cooler	1	2009	\$41,000	2	Gcal/hr	0.7	2.2	7.0	3.5	\$98,588	\$102,328	\$225,122
Ion Exchange (Dual Resin)	1	2014	\$5,250,000	53204	kg/hr	0.9	1.8	14509.8	0.3	\$1,630,441	\$1,533,084	\$2,759,551
liquid flow								103500.8				
water flow								88991.0				
Ketonization and Condensation												
Ketonization Feed Pump	1	2011	\$29,242	290936	kg/hr	0.6	3.5	103500.8	0.4	\$15,729	\$14,547	\$50,915
Ketonization P2P Heat Exchanger - Preheat/Rxr Exit	1	2007	\$432,519	12.3	Gcal/hr	0.6	1.6	65.7	5.3	\$1,182,157	\$1,218,832.26	\$1,950,132
Ketonization Reactor	1	2014	\$2,044,000	53204	kg/hr	0.37	2.2	103146.6	1.9	\$2,611,319	\$2,455,391.92	\$5,401,862
Ketonization Effluent Flash	1	2013	\$90,000	12189	kg/hr	0.7	1.5	103146.6	8.5	\$401,313	\$383,203	\$574,805
Ketonization Effluent Flash 2	1	2013	\$90,000	12189	kg/hr	0.7	1.5	0.0	0.0	\$0	\$0	\$0
Ketonization Product Pump	1	2009	\$9,200	22681	kg/hr	0.8	3.1	96756.4	4.3	\$29,363	\$30,477	\$94,479
Ketonization Column	1	2016	\$221,824	88799	kg/hr	0.6	1.3	96769.8	1.1	\$233,565	\$233,565	\$303,138
Heat exchanger - Reboiler	1	2010	\$184,000	-2	Gcal/hr	0.7	2.2	-8.3	4.1	\$497,997	\$489,770	\$1,077,494
Heat exchanger - Condenser	1	2009	\$82,000	2	Gcal/hr	0.7	2.2	0.9	0.4	\$45,165	\$46,878	\$103,132
Decanter	1	2012	\$132,541	8096	kg/hr	0.6	1.3	93762.5	11.6	\$576,243	\$533,956	\$684,864
Condensation Feed Pump	1	2011	\$29,242	290936	kg/hr	0.6	3.5	0.0	0.0	\$0	\$0	\$0
Condensation Preheater	1	2010	\$184,000	-2	Gcal/hr	0.7	2.2	-5.8	2.9	\$386,328	\$379,945	\$835,880
Condensation Reactor (5 hour residence time)	1	2015	\$1,993,000	237365	kg/hr	1	1.8	34447.6	0.1	\$144,617.14	\$140,695	\$253,251
Residence Time Scaling				2.5	hr			0.5				
Decanter 2	1	2012	\$132,541	8096	kg/hr	0.6	1.3	34447.6	4.3	\$315,997	\$292,808	\$375,562
Condensation Column	1	2016	\$259,394	29122	kg/hr	0.6	1.3	33069.5	1.1	\$279,952	\$279,952	\$358,420
Heat exchanger - Reboiler	1	2010	\$184,000	-2	Gcal/hr	0.7	2.2	-4.5	2.2	\$324,285	\$318,927	\$701,640
Heat exchanger - Condenser	1	2009	\$82,000	2	Gcal/hr	0.7	2.2	5.3	2.7	\$162,296	\$168,453	\$370,597
Condensation Flash Tank	1	2013	\$90,000	12189	kg/hr	0.7	1.5	28893.9	2.4	\$164,675	\$157,244.24	\$235,866
Hydrodeoxygenation												
Feed pump	2	2011	\$258,600	29274	kg/hr	0.8	1.4	7002.0	0.2	\$82,341	\$76,156	\$108,491
Air Compressor	2	2011	\$34,600	3818.92118	kg/hr	0.6	1.8	331.0	0.1	\$7,977	\$7,377	\$13,433
Furnace	1	2011	\$241,400	2.41874182	Gcal/hr	0.7	1.5	0.2	0.1	\$42,629	\$39,427	\$60,103
HDO Trim heater	1	2009	\$41,000	-2	Gcal/hr	0.7	2.2	0.0	0.0	\$0	\$0	\$0
HDO Makeup Compressor	2	2011	\$3,242,400	389.6519	kg/hr	0.6	1.1	118.7	0.3	\$1,588,677	\$1,469,329	\$1,601,381
HDO Makeup Compressor spare	1	2011	\$1,621,200	389.6519	kg/hr	0.6	1.1	118.7	0.3	\$794,338	\$794,665	\$794,346
HDO Recycle Compressor	2	2011	\$2,207,400	14665.4901	kg/hr	0.6	1.1	1382.3	0.1	\$535,129	\$494,928	\$559,367
HDO Recycle Compressor Spare	1	2011	\$1,103,700	14665.4901	kg/hr	0.6	1.1	1382.3	0.1	\$267,565	\$247,464	\$273,159
HDO-Feed-Effluent HX	1	2011	\$353,600	-14.3341665	Gcal/hr	0.7	2.7	-1.7	0.1	\$80,936	\$74,856	\$198,803
HDO Reactor	1	2011	\$3,777,764	208.383118	kg/hr	0	0.7	45.8	0.2	\$1,308,441	\$1,210,146	\$2,420,292
Pressure Factor Base (P=1300psig)	1			29274	kg/hr							
Pressure Factor Current (P=435 psig)				0.889923842	g/cc							
Internals	1											
Other Factors	1											
HDO Effluent CW HX	1	2011	\$321,600	4.41497612	Gcal/hr	0.7	1.7	1.9	0.4	\$176,551	\$163,288	\$270,572
HDO CHPS	1	2011	\$328,500	39911.26633	L/hr	0.7	2.6	9189.4	0.2	\$117,509	\$108,681	\$281,182
				26114.4444	kg/hr			6425.0				
				0.696208579	g/cc			0.8				
				2395.26725	kg/hr			668.7				
				0.997299235	g/cc			1.0				
Separation												
Frac effluent HX	1	2011	\$92,400	1.67373049	Gcal/hr	0.7	2.5	0.3	0.2	\$25,989	\$24,037	\$59,285
Frac feed trim HX	1	2011	\$17,000	-2.3892083	Gcal/hr	0.7	5.0	-0.4	0.2	\$5,167	\$4,779	\$24,033
Frac Column	1	2011	\$311,600	25643.9602	kg/hr	0.6	1.7	6424.9	0.3	\$135,808	\$125,605	\$209,772
Frac reflux pump	2	2011	\$11,600	25643.9602	kg/hr	0.8	6.8	6424.9	0.3	\$3,833	\$3,545	\$24,144
Frac Air cool condenser	1	2011	\$15,600	2.50642349	Gcal/hr	0.7	4.6	0.6	0.3	\$5,948	\$5,501	\$25,460
Frac 3 phase separator	1	2011	\$33,300	25643.9602	kg/hr	0.7	5.7	6424.9	0.3	\$12,637	\$11,688	\$67,005
Naptha Air Cooler	1	2011	\$29,400	0.067765556	Gcal/hr	0.7	2.3	0.0	0.0	\$3,543	\$3,277	\$7,613
MOT and Catalytic Upgrading Totals:										A300 Totals:	\$17,970,827	\$30,057,149

Mechanical Equipment List	Base Quoted Costs			Base Unit Sizing						Unit Costs		
	Num Req'd	Year of Quote	Purch Cost (All units)	Scaling Value	Units	Scaling Exp	Inst Factor	New Value	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
Polyol/Polyurethane Production												
Epoxidation and Ring Opening												
CAPEX for all Epoxidation and RO	1	2018	\$30,750,000	1216.09589	kg/hr	0.65	1	2885.6	2.4	\$53,922,670	\$48,432,947	\$48,432,947
Polyurethane Reaction												
CAPEX for Flexible Foam	1	2013	\$4,750,000	289	kg/hr	0.65	1	5246.7	18.2	\$31,276,466	\$29,865,083	\$29,865,083
Polyol/Polyurethane Production Totals:										A400 Totals:	\$78,298,030	\$78,298,030

Mechanical Equipment List	Base Quoted Costs			Base Unit Sizing						Unit Costs		
	Num Req'd	Year of Quote	Purch Cost (All units)	Scaling Value	Units	Scaling Exp	Inst Factor	New Value	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
Boiler, Utilities, and Storage												
Boiler configuration												
Natural Gas Boiler -HP	1	2012	\$1,279,300	3975024	lb/day	0.6	1	6015189.3	1.5	\$1,640,281	\$1,519,911	\$1,519,911
Natural Gas Boiler -LP	1	2012	\$1,203,500	3975024	lb/day	0.6	1	4195698.1	1.1	\$1,243,154	\$1,151,927	\$1,151,927
LP Steam Exchanger with MOT	1	2013	\$542,500	5.761	Gcal/hr	0.7	2.2	29.4	5.1	\$1,696,780	\$1,620,211	\$3,613,070
Utilities												
Cooling Tower System	1	2010	\$1,375,000	10037820	kg/hr	0.6	1.5	9604970.8	1.0	\$1,339,111	\$1,316,987	\$1,975,481
Plant Air Compressor	1	2010	\$28,000	83333	kg/hr	0.6	1.6	21297.5	0.3	\$12,350	\$12,146	\$19,433
Chilled Water Package	1	2010	\$1,275,750	14.2	Gcal/hr	0.6	1.6	0.0	0.0	\$0	\$0	\$0
CIP System	1	2009	\$694,000	83333	kg/hr	0.6	1.8	21297.5	0.3	\$306,102	\$317,715	\$571,888
Cooling Water Pump	3	2010	\$283,671	10982556	kg/hr	0.8	3.1	9604970.8	0.9	\$254,829	\$250,619	\$776,919
Makeup Water Pump	1	2010	\$6,864	155563.5	kg/hr	0.8	3.1	167211.5	1.1	\$7,272	\$7,152	\$22,171
Process Water Circulating Pump	1	2010	\$15,292	518923.5	kg/hr	0.8	3.1	167211.5	0.3	\$6,180	\$6,078	\$18,842
Instrument Air Dryer	1	2009	\$15,000	83333	kg/hr	0.6	1.8	21297.5	0.3	\$6,616	\$6,867	\$12,361
Plant Air Receiver	1	2009	\$16,000	83333	kg/hr	0.6	3.1	21297.5	0.3	\$7,057	\$7,325	\$22,707
Process Water Tank No. 1	1	2009	\$250,000	451555	kg/hr	0.7	1.7	167211.5	0.4	\$124,718	\$129,449	\$220,064
Hot Oil System	1	2011	\$241,400	-2.4	MMkcal/hr	0.7	1.5	-10.1	4.2	\$656,599	\$607,273	\$923,055
Storage												
Diesel Product Storage Tank	1	2009	\$670,000	11340.5	kg/hr	0.7	1.7	5691.0	0.5	\$413,487	\$429,174.46	\$729,597
Naphtha Product Storage Tank	1	2009	\$87,057	378.0166667	kg/hr	0.7	1.7	597.6	1.6	\$119,962	\$124,513	\$211,672
Firewater Storage Tank	1	2009	\$501,000	83333	kg/hr	0.7	1.7	21297.5	0.3	\$192,795	\$200,110	\$340,186
Tankage BOP	1									\$145,249	\$150,759.34	\$256,291
Boiler, Utilities, and Storage Totals:										A500 Totals:	\$7,858,217	\$12,385,574

Appendix B. Discounted Cash Flow Rate of Return Worksheet

DCFROR Worksheet								
Year	-2	-1	0	1	2	3	4	5
Fixed Capital Investment	\$10,290,224	\$77,176,680	\$41,160,896					
Land	\$1,848,000							
Working Capital			\$16,078,475					
Loan Payment				\$28,754,003	\$28,754,003	\$28,754,003	\$28,754,003	\$28,754,003
Loan Interest Payment	\$1,234,827	\$10,496,028	\$15,435,336	\$15,435,336	\$14,369,843	\$13,219,110	\$11,976,318	\$10,634,104
Loan Principal	\$15,435,336	\$131,200,355	\$192,941,699	\$179,623,032	\$165,238,872	\$149,703,979	\$132,926,295	\$114,806,396
Fuel Sales				\$31,484,985.13	\$41,979,980	\$41,979,980	\$41,979,980	\$41,979,980
Coproduct Credit				\$141,414,754	\$188,553,005	\$188,553,005	\$188,553,005	\$188,553,005
Total Annual Sales				\$172,899,739	\$230,532,986	\$230,532,986	\$230,532,986	\$230,532,986
Annual Manufacturing Cost								
Feedstock				\$67,467,493	\$89,956,657	\$89,956,657	\$89,956,657	\$89,956,657
Catalyst Cost				\$19,566,729	\$0	\$6,142,096	\$0	\$6,142,096
Other Variable Costs				\$69,902,480	\$79,888,549	\$79,888,549	\$79,888,549	\$79,888,549
Fixed Operating Costs				\$12,134,910	\$12,134,910	\$12,134,910	\$12,134,910	\$12,134,910
Total Product Cost				\$169,071,612	\$181,980,115	\$188,122,211	\$181,980,115	\$188,122,211
Annual Depreciation								
General Plant Writedown				14.29%	24.49%	17.49%	12.49%	8.93%
Depreciation Charge				\$45,054,168	\$77,213,196	\$55,143,275	\$39,379,045	\$28,154,914
Remaining Value				\$270,230,423	\$193,017,227	\$137,873,952	\$98,494,906	\$70,339,992
Steam Plant Writedown				3.75%	7.22%	6.68%	6.18%	5.71%
Depreciation Charge				\$235,684	\$453,707	\$419,643	\$388,219	\$359,057
Remaining Value				\$6,049,224	\$5,595,516	\$5,175,873	\$4,787,654	\$4,428,597
Net Revenue				(\$56,897,061)	(\$43,483,876)	(\$26,371,254)	(\$3,190,712)	\$3,262,700
Losses Forward					(\$56,897,061)	(\$100,380,937)	(\$126,752,191)	(\$129,942,903)
Taxable Income				(\$56,897,061)	(\$100,380,937)	(\$126,752,191)	(\$129,942,903)	(\$126,680,203)
Income Tax				\$0	\$0	\$0	\$0	\$0
Annual Cash Income				(\$24,925,876)	\$19,798,867	\$13,656,772	\$19,798,867	\$13,656,772
Discount Factor	1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830	0.6209
Annual Present Value	\$184,268,737			(\$22,659,887)	\$16,362,700	\$10,260,535	\$13,522,893	\$8,479,781
Total Capital Investment + Interest	\$16,181,392	\$96,439,979	\$72,674,707					
Net Present Worth			\$0					

DCFROR Worksheet

Year	6	7	8	9	10	11	12	13
Fixed Capital Investment								
Land								
Working Capital								
Loan Payment	\$28,754,003	\$28,754,003	\$28,754,003	\$28,754,003	\$28,754,003	\$0	\$0	\$0
Loan Interest Payment	\$9,184,512	\$7,618,952	\$5,928,148	\$4,102,080	\$2,129,926	\$0	\$0	\$0
Loan Principal	\$95,236,904	\$74,101,854	\$51,275,999	\$26,624,077	\$0	\$0	\$0	\$0
Fuel Sales	\$41,979,980	\$41,979,980	\$41,979,980	\$41,979,980	\$41,979,980	\$41,979,980	\$41,979,980	\$41,979,980
Coproduct Credit	\$188,553,005	\$188,553,005	\$188,553,005	\$188,553,005	\$188,553,005	\$188,553,005	\$188,553,005	\$188,553,005
Total Annual Sales	\$230,532,986	\$230,532,986	\$230,532,986	\$230,532,986	\$230,532,986	\$230,532,986	\$230,532,986	\$230,532,986
Annual Manufacturing Cost								
Feedstock	\$89,956,657	\$89,956,657	\$89,956,657	\$89,956,657	\$89,956,657	\$89,956,657	\$89,956,657	\$89,956,657
Catalyst Cost	\$0	\$6,142,096	\$0	\$6,142,096	\$0	\$6,142,096	\$0	\$6,142,096
Other Variable Costs	\$79,888,549	\$79,888,549	\$79,888,549	\$79,888,549	\$79,888,549	\$79,888,549	\$79,888,549	\$79,888,549
Fixed Operating Costs	\$12,134,910	\$12,134,910	\$12,134,910	\$12,134,910	\$12,134,910	\$12,134,910	\$12,134,910	\$12,134,910
Total Product Cost	\$181,980,115	\$188,122,211	\$181,980,115	\$188,122,211	\$181,980,115	\$188,122,211	\$181,980,115	\$188,122,211
Annual Depreciation								
General Plant Writedown	8.92%	8.93%	4.46%					
Depreciation Charge	\$28,123,386	\$28,154,914	\$14,061,693					
Remaining Value	\$42,216,607	\$14,061,693	\$0					
Steam Plant Writedown	5.29%	4.89%	4.52%	4.46%	4.46%	4.46%	4.46%	4.46%
Depreciation Charge	\$332,157	\$307,206	\$284,204	\$280,433	\$280,370	\$280,433	\$280,370	\$280,433
Remaining Value	\$4,096,440	\$3,789,234	\$3,505,030	\$3,224,598	\$2,944,228	\$2,663,795	\$2,383,425	\$2,102,993
Net Revenue	\$10,912,816	\$6,329,702	\$28,278,826	\$38,028,262	\$46,142,574	\$42,130,342	\$48,272,500	\$42,130,342
Losses Forward	(\$126,680,203)	(\$115,767,388)	(\$109,437,686)	(\$81,158,860)	(\$43,130,598)	\$0	\$0	\$0
Taxable Income	(\$115,767,388)	(\$109,437,686)	(\$81,158,860)	(\$43,130,598)	\$3,011,976	\$42,130,342	\$48,272,500	\$42,130,342
Income Tax	\$0	\$0	\$0	\$0	\$632,515	\$8,847,372	\$10,137,225	\$8,847,372
Annual Cash Income	\$19,798,867	\$13,656,772	\$19,798,867	\$13,656,772	\$19,166,352	\$33,563,403	\$38,415,645	\$33,563,403
Discount Factor	0.5645	0.5132	0.4665	0.4241	0.3855	0.3505	0.3186	0.2897
Annual Present Value	\$11,175,944	\$7,008,083	\$9,236,318	\$5,791,804	\$7,389,459	\$11,763,768	\$12,240,408	\$9,722,122
Total Capital Investment + Interest								
Net Present Worth								

DCFROR Worksheet

Year	14	15	16	17	18	19	20	21
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	\$41,979,980	\$41,979,980	\$41,979,980	\$41,979,980	\$41,979,980	\$41,979,980	\$41,979,980	\$41,979,980
Coproduct Credit	\$188,553,005	\$188,553,005	\$188,553,005	\$188,553,005	\$188,553,005	\$188,553,005	\$188,553,005	\$188,553,005
Total Annual Sales	\$230,532,986	\$230,532,986	\$230,532,986	\$230,532,986	\$230,532,986	\$230,532,986	\$230,532,986	\$230,532,986
Annual Manufacturing Cost								
Feedstock	\$89,956,657	\$89,956,657	\$89,956,657	\$89,956,657	\$89,956,657	\$89,956,657	\$89,956,657	\$89,956,657
Catalyst Cost	\$0	\$6,142,096	\$0	\$6,142,096	\$0	\$6,142,096	\$0	\$6,142,096
Other Variable Costs	\$79,888,549	\$79,888,549	\$79,888,549	\$79,888,549	\$79,888,549	\$79,888,549	\$79,888,549	\$79,888,549
Fixed Operating Costs	\$12,134,910	\$12,134,910	\$12,134,910	\$12,134,910	\$12,134,910	\$12,134,910	\$12,134,910	\$12,134,910
Total Product Cost	\$181,980,115	\$188,122,211	\$181,980,115	\$188,122,211	\$181,980,115	\$188,122,211	\$181,980,115	\$188,122,211
Annual Depreciation								
General Plant Writedown								
Depreciation Charge								
Remaining Value								
Steam Plant Writedown	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	2.23%
Depreciation Charge	\$280,370	\$280,433	\$280,370	\$280,433	\$280,370	\$280,433	\$280,370	\$140,216
Remaining Value	\$1,822,623	\$1,542,191	\$1,261,821	\$981,388	\$701,019	\$420,586	\$140,216	\$0
Net Revenue	\$48,272,500	\$42,130,342	\$48,272,500	\$42,130,342	\$48,272,500	\$42,130,342	\$48,272,500	\$42,270,558
Losses Forward	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0
Taxable Income	\$48,272,500	\$42,130,342	\$48,272,500	\$42,130,342	\$48,272,500	\$42,130,342	\$48,272,500	\$42,270,558
Income Tax	\$10,137,225	\$8,847,372	\$10,137,225	\$8,847,372	\$10,137,225	\$8,847,372	\$10,137,225	\$8,876,817
Annual Cash Income	\$38,415,645	\$33,563,403	\$38,415,645	\$33,563,403	\$38,415,645	\$33,563,403	\$38,415,645	\$33,533,957
Discount Factor	0.2633	0.2394	0.2176	0.1978	0.1799	0.1635	0.1486	0.1351
Annual Present Value	\$10,116,040	\$8,034,812	\$8,360,364	\$6,640,340	\$6,909,391	\$5,487,885	\$5,710,241	\$4,531,463
Total Capital Investment + Interest								
Net Present Worth								

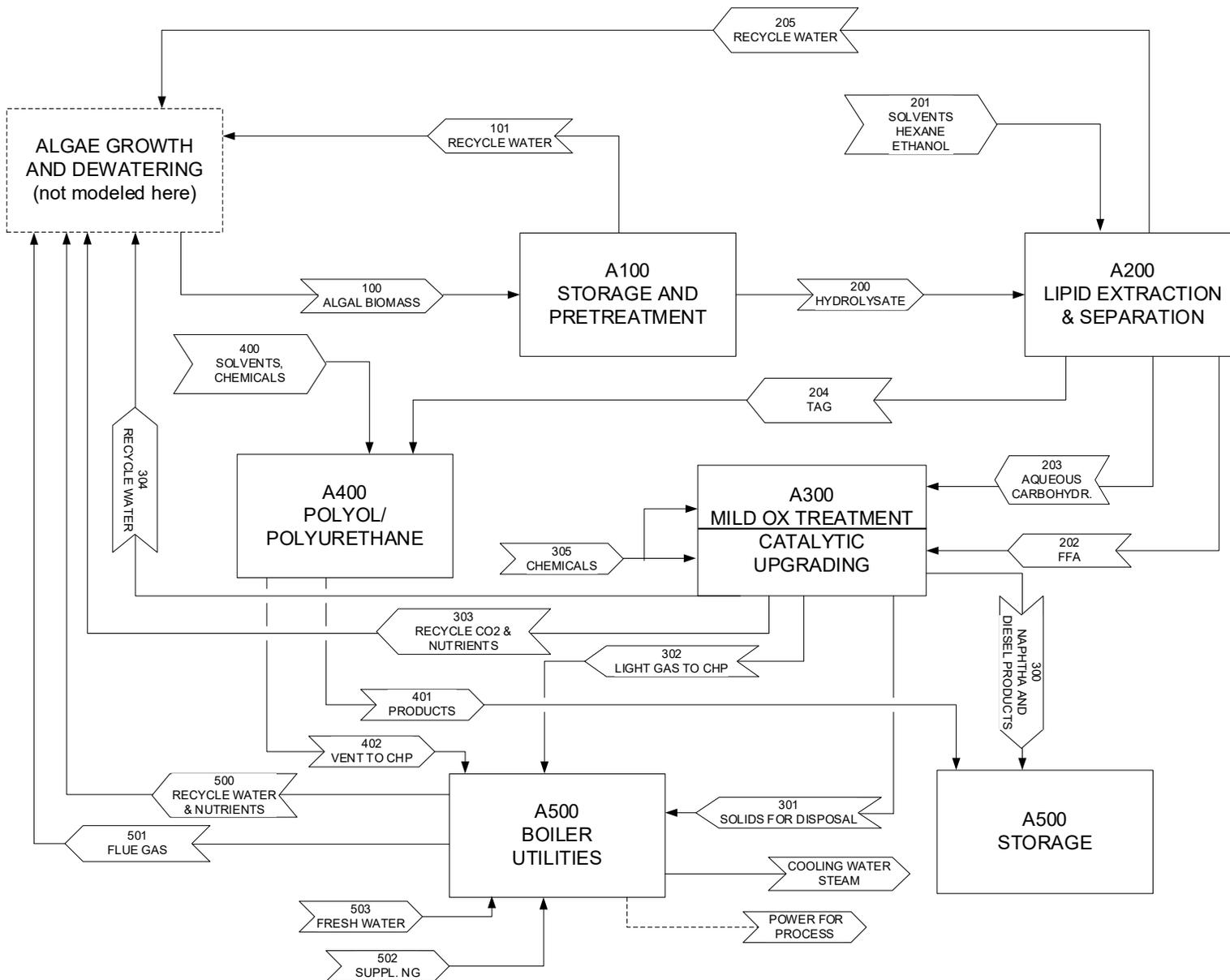
DCFROR Worksheet

Year	22	23	24	25	26	27	28	29	30
Fixed Capital Investment									
Land									-\$1,848,000
Working Capital									-\$16,078,475
Loan Payment	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0
Loan Interest Payment	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0
Loan Principal	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0
Fuel Sales	\$41,979,980	\$41,979,980	\$41,979,980	\$41,979,980	\$41,979,980	\$41,979,980	\$41,979,980	\$41,979,980	\$41,979,980
Coproduct Credit	\$188,553,005	\$188,553,005	\$188,553,005	\$188,553,005	\$188,553,005	\$188,553,005	\$188,553,005	\$188,553,005	\$188,553,005
Total Annual Sales	\$230,532,986	\$230,532,986	\$230,532,986	\$230,532,986	\$230,532,986	\$230,532,986	\$230,532,986	\$230,532,986	\$230,532,986
Annual Manufacturing Cost									
Feedstock	\$89,956,657	\$89,956,657	\$89,956,657	\$89,956,657	\$89,956,657	\$89,956,657	\$89,956,657	\$89,956,657	\$89,956,657
Catalyst Cost	\$0	\$6,142,096	\$0	\$6,142,096	\$0	\$6,142,096	\$0	\$6,142,096	\$0
Other Variable Costs	\$79,888,549	\$79,888,549	\$79,888,549	\$79,888,549	\$79,888,549	\$79,888,549	\$79,888,549	\$79,888,549	\$79,888,549
Fixed Operating Costs	\$12,134,910	\$12,134,910	\$12,134,910	\$12,134,910	\$12,134,910	\$12,134,910	\$12,134,910	\$12,134,910	\$12,134,910
Total Product Cost	\$181,980,115	\$188,122,211	\$181,980,115	\$188,122,211	\$181,980,115	\$188,122,211	\$181,980,115	\$188,122,211	\$181,980,115
Annual Depreciation									
General Plant Writedown									
Depreciation Charge									
Remaining Value									
Steam Plant Writedown									
Depreciation Charge									
Remaining Value									
Net Revenue	\$48,552,870	\$42,410,774	\$48,552,870	\$42,410,774	\$48,552,870	\$42,410,774	\$48,552,870	\$42,410,774	\$48,552,870
Losses Forward	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0	\$0
Taxable Income	\$48,552,870	\$42,410,774	\$48,552,870	\$42,410,774	\$48,552,870	\$42,410,774	\$48,552,870	\$42,410,774	\$48,552,870
Income Tax	\$10,196,103	\$8,906,263	\$10,196,103	\$8,906,263	\$10,196,103	\$8,906,263	\$10,196,103	\$8,906,263	\$10,196,103
Annual Cash Income	\$38,356,767	\$33,504,512	\$38,356,767	\$33,504,512	\$38,356,767	\$33,504,512	\$38,356,767	\$33,504,512	\$38,356,767
Discount Factor	0.1228	0.1117	0.1015	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
Annual Present Value	\$4,711,974	\$3,741,722	\$3,894,194	\$3,092,332	\$3,218,342	\$2,555,647	\$2,659,787	\$2,112,105	\$2,198,171
Total Capital Investment + Interest									(\$1,027,340)
Net Present Worth									

Appendix C. Key Aspen Stream Data Tables

High-level stream table information from Aspen Plus modeling output follows, for key streams associated with each process operation area under both pathway scenarios. 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%.

Solvent	Hexane, toluene, ethanol
Carboxylic acids	Formic acid, acetic acid, succinic acid, lactic acid
Ketones	Acetone, 3-methyl-2-cyclopenten-1-one
Light hydrocarbon gas	C ₂ -C ₅ alkanes
CO/SO _x /NO _x	CO, SO ₂ , NO ₂ , NO
Cell matter	<i>Scenedesmus</i> , algal cell mass; see Table 2



Component	Units	100	101	200	201	201	202	203	204	205	300	301	302
Total flow	kg/hr	118186	12798	105388	271	16	2734	99765	2791	32805	5889	1951	277
Insoluble Solids	%	17%	0%	17%	0%	0%	100%	12%	100%	0%	0%	50%	0%
Soluble Solids	%	0%	0%	3%	0%	0%	0%	3%	0%	0%	0%	0%	0%
Temperature	°C	25	40	252	25	25	102	65	350	25	20	104	48
Pressure	atm	1	1	40	2	3	1	2	1	1	1	35	1
Solvent	kg/hr	0	0	0	271	16	0	274	0	18	0	3	55
Carboxylic Acids	kg/hr	553	3	550	0	0	0	550	0	0	0	127	0
Ketones	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0
Renewable Hydrocarbons	kg/hr	0	0	0	0	0	0	0	0	0	5889	0	0
Light hydrocarbon gas	kg/hr	0	0	0	0	0	0	0	0	0	0	0	89
Water	kg/hr	97208	12795	84312	0	0	2	84312	0	32784	0	850	8
CO2	kg/hr	0	0	0	0	0	0	0	0	0	0	0	42
Methane	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0
H2	kg/hr	0	0	0	0	0	0	0	0	0	0	0	16
O2	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0
N2	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0
CO/SOx/NOx	kg/hr	0	0	0	0	0	0	0	0	0	0	0	67
H2O2	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0
TDI/DEOA/Surfactant	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0
Polyurethane	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0
NH3	kg/hr	0	0	0	0	0	0	0	0	0	0	5	0
Diammonium Phosphate	kg/hr	0	0	0	0	0	0	0	0	0	0	0	0
Glucose (SS)	kg/hr	0	0	1011	0	0	0	1011	0	0	0	0	0
Protein (SS)	kg/hr	0	0	1805	0	0	0	1805	0	0	0	0	0
Glycerol (IS)	kg/hr	313	0	313	0	0	0	313	0	0	0	0	0
Free fatty acids (IS)	kg/hr	2878	0	2877	0	0	2731	115	28	3	0	115	0
Triglycerides (IS)	kg/hr	2878	0	2878	0	0	0	115	2763	0	0	115	0
Polar Lipid Impurities (IS)	kg/hr	574	0	574	0	0	0	202	0	0	0	202	0
Fermentable carbs (IS)	kg/hr	9100	0	8190	0	0	0	8190	0	0	0	0	0
Non-fermentable carbs (IS)	kg/hr	967	0	967	0	0	0	967	0	0	0	0	0
Protein (IS)	kg/hr	2834	0	1029	0	0	0	1029	0	0	0	0	0
Ash (IS)	kg/hr	535	0	535	0	0	0	535	0	0	0	535	0
Cell matter (IS)	kg/hr	347	0	347	0	0	0	347	0	0	0	0	0

Component	Units	303	304	305	400	401	402	500	500	501	502	503
Total flow	kg/hr	12540	88318	6360	3012	5074	285	4025	13509	94165	4352	154907
Insoluble Solids	%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%
Soluble Solids	%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%	0%
Temperature	°C	30	30	25	25	25	118	100	28	130	25	25
Pressure	atm	1	1	1	1	1	1	1	5	1	1	1
Solvent	kg/hr	66	257	165	0	0	0	0	0	0	0	0
Carboxylic Acids	kg/hr	2	13	0	527	0	285	0	0	0	0	0
Ketones	kg/hr	156	0	0	0	0	0	0	0	0	0	0
Renewable Hydrocarbons	kg/hr	0	0	0	0	0	0	0	0	0	0	0
Light hydrocarbon gas	kg/hr	0	0	0	0	0	0	0	0	0	0	0
Water	kg/hr	164	88027	0	0	0	0	4025	13509	11479	0	154907
CO2	kg/hr	11006	12	0	0	0	0	0	0	12942	0	0
Methane	kg/hr	0	0	0	0	0	0	0	0	0	4352	0
H2	kg/hr	0	0	222	0	0	0	0	0	0	0	0
O2	kg/hr	295	0	5731	0	0	0	0	0	2038	0	0
N2	kg/hr	243	0	243	62	0	0	0	0	67705	0	0
CO/SOx/NOx	kg/hr	53	1	0	0	0	0	0	0	0	0	0
H2O2	kg/hr	0	0	0	586	0	0	0	0	0	0	0
TDI/DEOA/Surfactant	kg/hr	0	0	0	1836	0	0	0	0	0	0	0
Polyurethane	kg/hr	0	0	0	0	5074	0	0	0	0	0	0
NH3	kg/hr	379	6	0	0	0	0	0	0	0	0	0
Diammonium Phosphate	kg/hr	177	0	0	0	0	0	0	0	0	0	0
Glucose (SS)	kg/hr	0	0	0	0	0	0	0	0	0	0	0
Protein (SS)	kg/hr	0	0	0	0	0	0	0	0	0	0	0
Glycerol (IS)	kg/hr	0	0	0	0	0	0	0	0	0	0	0
Free fatty acids (IS)	kg/hr	0	0	0	0	0	0	0	0	0	0	0
Triglycerides (IS)	kg/hr	0	0	0	0	0	0	0	0	0	0	0
Polar Lipid Impurities (IS)	kg/hr	0	0	0	0	0	0	0	0	0	0	0
Fermentable carbs (IS)	kg/hr	0	0	0	0	0	0	0	0	0	0	0
Non-fermentable carbs (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
Ash (IS)	kg/hr	0	0	0	0	0	0	0	0	0	0	0
Cell matter (IS)	kg/hr	0	0	0	0	0	0	0	0	0	0	0