Techno-Economic Assessment for the Production of Hydrocarbon Fuels via Catalytic Upgrading of Furans

Bruno Klein, Ian McNamara, Ryan Davis, Ashutosh Mittal, and David Johnson

National Renewable Energy Laboratory

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Cover Photos by Dennis Schroeder: (clockwise, left to right) NREL 51934, NREL 45897, NREL 42160, NREL 45891, NREL 48097, NREL 46526.

Nomenclature

ACCE	Aspen Capital Cost Estimator
BCD	base-catalyzed deconstruction
BDO	2,3-butanediol
BTU	British thermal unit
CAPEX	capital expenditures
CO_2	carbon dioxide
CSL	corn steep liquor
CUBI	Catalytic Upgrading of Biochemical Intermediates
DMR	deacetylation and mechanical refining
FCI	fixed capital investment
gal	gallon
GGE	gallon gasoline equivalent
HDO	hydrodeoxygenation
HMF	hydroxymethylfurfural
IS	insoluble solids
ISBL	inside battery limits
lb	pound
LHV	lower heating value
MEK	methyl ethyl ketone (2-butanone)
MFSP	minimum fuel selling price
MM	million
MVR	mechanical vapor recompression
NREL	National Renewable Energy Laboratory
NRTL	nonrandom two-liquid
OPEX	operational expenses
PNNL	Pacific Northwest National Laboratory
PSA	pressure swing adsorption
SS	soluble solids
TCI	total capital investment
TDC	total direct cost
TEA	techno-economic analysis
VFP	vacuum filter press
WHSV	weight hourly space velocity
WWT	wastewater treatment

Executive Summary

The U.S. Department of Energy Bioenergy Technologies Office promotes the production of liquid fuels from lignocellulosic feedstocks by sponsoring programs in fundamental and applied research that aim to advance the state of biomass conversion technologies. Biomass that has been deconstructed into sugars and sugar-derived intermediates via hydrolysis and fermentation can be converted into transportation fuels and bioproduct chemical streams via a multitude of routes, including chemical catalysis. The multi-lab Catalytic Upgrading of Biochemical Intermediates (CUBI) project within the Chemical Catalysis for Bioenergy Consortium (ChemCatBio) addresses critical barriers linked to the catalytic conversion of sugars and their intermediates into hydrocarbon fuels and coproducts. When incorporated within the broader biochemical conversion deconstruction process in a biorefinery context, developments under CUBI show that the key catalytic upgrading strategies under investigation exhibit potential to achieve a targeted minimum fuel selling price (MFSP) under \$2.50 per gallon gasoline equivalent (GGE) if value is added to lignin in the form of coproducts.

To support such efforts, the National Renewable Energy Laboratory (NREL) investigates the process design and economics of modeled cellulosic biorefineries in order to develop a plant gate price for fuels and bio-derived products. This report documents the techno-economic analysis (TEA) implications of a biochemical/catalytic pathway for the production of long-chain hydrocarbon fuels and proposes two distinct conceptual biorefineries centered around the process encompassing sugar dehydration to furans, aldol condensation between furans and a ketone (methyl ethyl ketone [MEK]), and a final step of hydrotreating to obtain hydrocarbons in the C_{14}–C_{16} range. The **integrated** plant simultaneously produces both furans (furfural and hydroxymethylfurfural [HMF]) and ketone (MEK via 2,3-butanediol [BDO]) from sugars; on the other hand, a **dedicated** biorefinery focuses on exclusively producing furans from corn stover hydrolysate and procuring the ketone externally. In either plant, the main coproducts are adipic acid (derived from lignin) and sodium sulfate. This study is based on detailed process simulations and TEA to determine MFSP for the main hydrocarbon fuel product. Because this report considers long-term performance targets for the full pathway dedicated to sugar upgrading to fuels, the critical remaining research points needed to achieve future cost goals are also discussed.

TEA results estimate the MFSP of the **dedicated** and **integrated** biorefineries at **\$2.54/GGE** and **\$2.72/GGE** (2016-dollars), respectively, with corresponding fuel yields of 108.4 and 61.2 GGE/dry ton of biomass. The assessment also clearly reiterates the need for valorizing the lignin fraction to a high-value coproduct in order to achieve such MFSP levels: At a yield of 284 and 276 lb/dry ton of biomass, respectively, the coproduction of adipic acid contributes to overall MFSPs at credits corresponding to *negative* \$1.23/GGE and \$2.16/GGE for the respective cases. Although adipic acid has been chosen as the bio-derived lignin coproduct in this work in keeping with prior NREL TEA design case studies, a number of other molecules with industrial application could have the same positive effect on overall biorefinery economics. Notably, the MFSP results highlighted here are quite comparable to those reported in NREL's aforementioned design case focused on biological conversion of sugars to fermentation intermediates with subsequent catalytic upgrading of those intermediates to hydrocarbon fuels, at \$2.47–\$2.49/GGE, thus presenting another viable alternative pathway to achieve similar fuel cost targets through purely catalytic upgrading of sugars in this case.

A single-point sensitivity analysis around selected parameters was also conducted to identify the major cost drivers of the biorefineries. The main factor for both biorefining setups is adipic acid coproduct price, which could lead to large swings in MFSP depending on its market value. For sugar-to-fuel catalysis details in the context of the primary focus of this report, in the case of the **dedicated** plant, the price at which MEK is purchased could also lead to significantly different MFSP results. Whether produced internally via the **integrated** plant or procured externally for the **dedicated** plant, it will be crucial for both economic and life cycle considerations to ultimately source the ketone co-reactant through a renewable or otherwise low-carbon-intensity process. The furans pathway influences the economic performance of the biorefinery, especially through the yield achievable in sugar dehydration: A high yield is desirable to keep processing costs low on a per-GGE basis because a reduction in this parameter directly decreases the overall fuel yield of the plant.

Finally, the report provides a qualitative discussion on additional opportunities for cost reduction within the proposed biorefineries, namely through adding value to furans, to coproducts that may be obtained following BDO dehydration, and to lignin through alternative pathways.

Dedicated Biorefinery for Hydocarbon Fuel Production: Process Engineering Analysis

DMR Pretreatment, Enzymatic Hydrolysis & Catalytic Upgrading of Sugars, Lignin Conversion to Coproducts

All Values in 2016$

Minimum Fuel Selling Price (MFSP, Gasoline-Equivalent Basis):	**$2.54 /GGE**
Contributions: Feedstock	**$0.66 /GGE**
Fuel Conversion	**$3.11 /GGE**
Coproduct Conversion	**-$1.23 /GGE**
Fuel Production	78.5 MMGGE per year
Fuel Yield	108.4 GGE / dry U.S. ton feedstock
Adipic Acid Coproduct Yield	284 lb / dry U.S. ton feedstock
Feedstock + Handling Cost	$71.26 /dry U.S. ton feedstock
Internal Rate of Return (After-Tax)	10%
Equity Percent of Total Investment	40%

Capital Costs	
Area 200: Pretreatment	$48,700,000
Area 300: Sugar Hydrolysis and Conditioning	$59,800,000
Area 400: Enzyme Production	$11,500,000
Area 500: Furans Production and Upgrading	$43,000,000
Area 600: Wastewater	$31,300,000
Area 700: Lignin	$140,000,000
Area 800: Boiler	$60,100,000
Area 900: Utilities & Storage	$20,000,000
Total Installed Equipment Cost	$414,400,000
Added Direct + Indirect Costs	$372,500,000
(% of TCI)	47%
Total Capital Investment (TCI)	$786,900,000
Installed Equipment Cost/Annual GGE	$5.28
Total Capital Investment/Annual GGE	$10.02
Operating Hours Per Year (On-Stream Factor)	7884 (90%)
Loan Rate	8.0%
Term (years)	10
Capital Charge Factor (Computed)	0.131

Specific Operating Conditions	
Enzyme Loading (mg/g cellulose)	10.0
Net Electricity Import (kWh/GGE)	5.2
Plant Electricity Use (kWh/GGE)	6.9

Manufacturing Costs (cents/GGE fuel product)	
Feedstock + Handling	65.7
Sulfuric Acid	11.0
Caustic	54.2
Glucose (enzyme production)	10.8
Hydrogen	63.8
Electricity (import)	35.1
Natural Gas	12.0
Methyl Ethyl Ketone	61.8
Other Raw Materials	24.2
Catalysts	1.8
Waste Disposal	1.8
Na_2SO_4 Coproduct	-21.5
Adipic Acid coproduct ($0.86/lb)	-223.6
Fixed Costs	26.0
Capital Depreciation	31.7
Average Income Tax	9.9
Average Return on Investment	89.4
Total	253.9

Manufacturing Costs ($/yr)	
Feedstock + Handling	$51,600,000
Sulfuric Acid	$8,600,000
Caustic	$42,600,000
Glucose (enzyme production)	$8,400,000
Hydrogen	$50,100,000
Electricity (import)	$27,600,000
Natural Gas	$9,400,000
Methyl Ethyl Ketone	$48,500,000
Other Raw Materials	$19,000,000
Catalysts	$1,400,000
Waste Disposal	$1,400,000
Na_2SO_4 Coproduct	-$16,900,000
Adipic Acid coproduct ($0.86/lb)	-$175,700,000
Fixed Costs	$20,400,000
Capital Depreciation	$24,900,000
Average Income Tax	$7,800,000
Average Return on Investment	$70,200,000
Total	$199,300,000

Figure ES-1. Economic summary for catalytic upgrading of sugars in a dedicated biorefinery

Integrated Biorefinery for Hydocarbon Fuel Production: Process Engineering Analysis

DMR Pretreatment, Enzymatic Hydrolysis & Fermentation / Catalytic Upgrading of Sugars, Lignin Conversion to Coproducts

All Values in 2016$

Minimum Fuel Selling Price (MFSP, Gasoline-Equivalent Basis):	**$2.72 /GGE**
Contributions: Feedstock	**$1.17 /GGE**
Fuel Conversion	**$3.71 /GGE**
Coproduct Conversion	**-$2.16 /GGE**
Fuel Production	44.3 MMGGE per year
Fuel Yield	61.2 GGE / dry U.S. ton feedstock
Adipic Acid Coproduct Yield	276 lb / dry U.S. ton feedstock
Feedstock + Handling Cost	$71.26 /dry U.S. ton feedstock
Internal Rate of Return (After-Tax)	10%
Equity Percent of Total Investment	40%

Capital Costs	
Area 200: Pretreatment	$48,700,000
Area 300: Sugar Hydrolysis and Conditioning	$42,100,000
Area 400: Enzyme Production	$11,500,000
Area 500: Furans Production and Upgrading	$29,600,000
Area 550: MEK Production	$25,100,000
Area 600: Wastewater	$35,700,000
Area 700: Lignin	$134,700,000
Area 800: Boiler	$70,400,000
Area 900: Utilities & Storage	$18,000,000
Total Installed Equipment Cost	$415,800,000
Added Direct + Indirect Costs	$370,400,000
(% of TCI)	47%
Total Capital Investment (TCI)	$786,200,000
Installed Equipment Cost/Annual GGE	$9.39
Total Capital Investment/Annual GGE	$17.75
Operating Hours Per Year (On-Stream Factor)	7884 (90%)
Loan Rate	8.0%
Term (years)	10
Capital Charge Factor (Computed)	0.129

Specific Operating Conditions	
Enzyme Loading (mg/g cellulose)	10.0
Net Electricity Import (kWh/GGE)	6.4
Plant Electricity Use (kWh/GGE)	10.9

Manufacturing Costs (cents/GGE fuel product)	
Feedstock + Handling	116.5
Sulfuric Acid	19.0
Caustic	88.8
Glucose (enzyme production)	19.1
Hydrogen	63.5
Electricity (import)	42.3
Natural Gas	31.8
Other Raw Materials	32.9
Catalysts	4.3
Waste Disposal	3.2
Na_2SO_4 Coproduct	-38.7
Adipic Acid coproduct ($0.86/lb)	-385.8
Fixed Costs	45.3
Capital Depreciation	56.2
Average Income Tax	17.4
Average Return on Investment	155.8
Total	271.6

Manufacturing Costs ($/yr)	
Feedstock + Handling	$51,600,000
Sulfuric Acid	$8,400,000
Caustic	$39,300,000
Glucose (enzyme production)	$8,400,000
Hydrogen	$28,100,000
Electricity (import)	$18,700,000
Natural Gas	$14,100,000
Other Raw Materials	$14,600,000
Catalysts	$1,900,000
Waste Disposal	$1,400,000
Na_2SO_4 Coproduct	-$17,100,000
Adipic Acid coproduct ($0.86/lb)	-$170,900,000
Fixed Costs	$20,100,000
Capital Depreciation	$24,900,000
Average Income Tax	$7,700,000
Average Return on Investment	$69,000,000
Total	$120,200,000

Figure ES-2. Economic summary for catalytic upgrading of sugars in an integrated biorefinery

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1 Introduction

1.1 Background and Motivation

The U.S. Department of Energy Bioenergy Technologies Office promotes the production of liquid fuels from lignocellulosic feedstocks by sponsoring programs in fundamental and applied research that aim to advance the state of biomass conversion technology. These programs include research to develop improved bioconversion organisms, chemical catalysts, hydrolysis enzymes, and integrated unit operations through synthetic biology, catalyst development and testing, chemical and mechanical pretreatment work, detailed engineering studies of potential processes, and construction of pilot-scale demonstration and production facilities. To support this work, the National Renewable Energy Laboratory (NREL) investigates the process design and economics of modeled cellulosic biorefineries in order to develop a plant gate price for fuels and bio-derived products. Through process modeling and techno-economic analysis (TEA), a modeled biorefinery's minimum fuel selling price (MFSP) may be quantified to reflect the economic viability for a given conversion technology pathway attributed to a set of technical parameters for the process.

Generally speaking, biochemical conversion technology approaches that operate based on the conversion of targeted biomass constituents to targeted product molecules may be divided into biomass deconstruction operations (i.e., pretreatment/fractionation of biomass into convertible components) and intermediate upgrading operations (i.e., biological or catalytic upgrading of those convertible components into fuels or products). NREL maintains extensive focus on both deconstruction and conversion processes, with numerous TEA reports released over recent years reflecting various strategies for the biochemical deconstruction and conversion technology pathways that may be taken in an integrated biorefinery [1–4].

Biomass that has been deconstructed into sugars and sugar-derived intermediates via hydrolysis and fermentation can be converted into transportation fuels and bioproduct chemical streams via a multitude of routes, including chemical catalysis. The multi-lab Catalytic Upgrading of Biochemical Intermediates (CUBI) project within the Chemical Catalysis for Bioenergy Consortium (ChemCatBio) is dedicated to addressing critical barriers linked to the catalytic conversion of such biomass-derived sugars and their intermediates into hydrocarbon fuels and coproducts. Most biochemical conversion pathways involving fermentation of lignocellulosic sugars are challenged by (1) requiring complete hydrolysis of carbohydrates into monomeric sugars, and (2) carbon efficiency penalties incurred by evolution of carbon dioxide (CO_2) during the fermentation process. Chemical catalysis may offer the potential to alleviate some of these challenges, given appropriate insertion points to be integrated into biochemical conversion processes. When incorporated within the broader biochemical conversion deconstruction process and when coupled with opportunities for lignin coproduct valorization, developments under CUBI show that the key catalytic upgrading strategies under investigation exhibit potential to achieve a targeted MFSP under $2.50 per gallon gasoline equivalent (GGE).

Among the multiple catalytic upgrading strategies developed under CUBI, including the conversion of intermediates such as carboxylic acids, 2,3-butanediol (BDO) via butenes or methyl ethyl ketone (MEK), and furans, this report focuses on establishing TEA models for the latter approach via direct catalysis of sugars, with or without a parallel fermentation step (optionally included as a means of providing a ketone co-reactant for catalytic upgrading). Two main biorefinery options were designed to deconstruct and hydrolyze biomass to sugars and convert them into chemicals required for the synthesis of hydrocarbon fuels. The **integrated** plant produces both furans (furfural and hydroxymethylfurfural [HMF]) and a ketone (MEK) from sugars, whereas the **dedicated** biorefinery focuses on exclusively producing furans from corn stover hydrolysate and procuring the ketone externally. Either approach is an alternative route to synthesize hydrocarbon fuels, relative to pathways presented in a recent biochemical design report [1]. Figure 1 presents a simplified process flow diagram of the strategies assessed in this work.

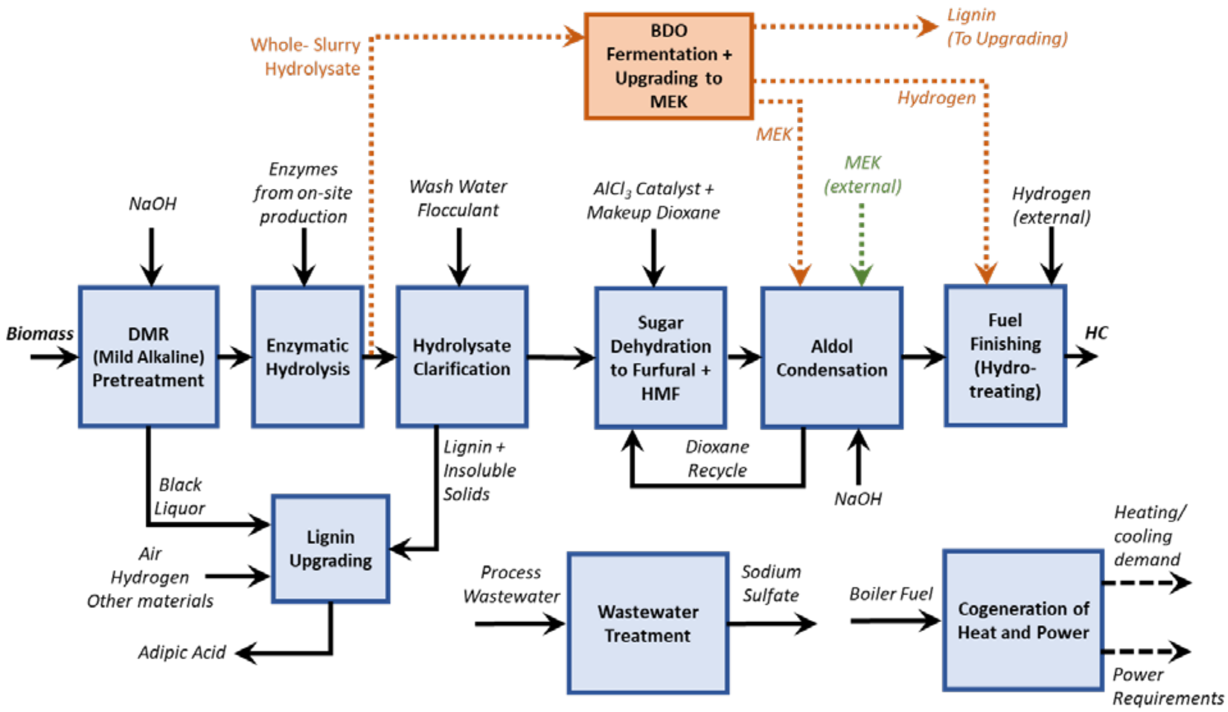

Figure 1. Simplified representation of the biorefining configurations assessed in this report. Blue boxes represent sections included in both biorefining configurations, whereas the green arrow is specific to the dedicated case and the orange objects are representative of the integrated plant.

As in previous efforts, the TEA models presented here solve for the MFSP of the hydrocarbon product as the main economic metric, after including coproduct revenues from the sale of adipic acid produced from lignin deconstruction and upgrading, set at a fixed market price. This is a useful metric to address the feasibility of each production route when compared to other technologies. Through process simulation and TEA, which is the basis for this assessment, it is possible to determine the critical remaining research needed to achieve future cost goals.

1.2 Process Overview

As previously indicated, this report addresses the production of a renewable hydrocarbon fuel blendstock through **two biorefining configurations**: a **dedicated** biorefinery, in which sugars are exclusively converted into furans and the MEK co-reactant is purchased externally, and an **integrated** biorefinery, which combines fermentative and catalytic pathways for conversion of sugars into fuel precursor intermediates (both furans and MEK at the necessary ratios). Either route draws much of the configuration from previous design reports, namely areas outside the conversion of sugars, and accordingly we defer to those prior reports for a detailed accounting of the associated process details and model inputs.

The process starts with a deacetylation/mild alkaline extraction and mechanical refining pretreatment of corn stover, followed by enzymatic hydrolysis (saccharification) of the cellulose and hemicellulose for solubilization of sugar.

In the **integrated** option, part of the hydrolysate is routed to BDO fermentation and further upgraded to MEK. The remainder of the slurry is filtered and concentrated, and the clarified hydrolysate is dehydrated to furfural and HMF. The resulting intermediates (furans and MEK) are then combined in an aldol condensation reaction to form higher-molecular-weight, oxygenated intermediates. The intermediates are subsequently catalytically upgraded to final hydrocarbon fuel products. In the **dedicated** biorefinery, in contrast, all of the enzymatic hydrolysis stream is sent to a vacuum filter press (VFP) and further concentrated prior to being routed to the production of furans; MEK is sourced externally for aldol condensation. Both biorefineries include lignin valorization in the setup. Lignin is initially deconstructed to soluble monomers and converted (along with other biomass residual components) to muconic acid, which is purified and further upgraded to adipic acid for sale as a coproduct. The facility also includes feedstock handling and storage, wastewater treatment (WWT), residual waste combustion, product storage, and utilities. The process is divided into 10 areas (Figure 2). Except for Area 500, which is the development focus of CUBI and the core section assessed in this report, and Area 550, which is also closely related to another pathway under investigation in CUBI, all other areas are only briefly summarized because they maintain the same details and modeling inputs as documented in the 2018 design report [1].

Area 100: Feedstock logistics and handling. The biorefinery is designed to process 2,000 dry metric tonnes of corn stover per day. All Area 100 processing aspects are outside the scope of this work and are combined into delivered feedstock costs at the pretreatment section (Area 200).

Area 200: Pretreatment. In this area, the biomass is processed with a deacetylation and mechanical refining (DMR) pretreatment approach, which comprises the preferential solubilization and removal of acetate and other biomass components with the use of an alkali extraction step, as well as sequential disc refining operations.

Area 300: Enzymatic hydrolysis and hydrolysate conditioning. Enzymatic hydrolysis is initiated in a high-solids continuous reactor using a cellulase enzyme produced on site and completed in parallel batch reactors. Part of these vessels are also employed for BDO fermentation (Area 550). In the **dedicated** plant, the hydrolysate is fully processed in a flocculant-assisted vacuum filter to remove solids (routed to the lignin train), and the clarified liquor is concentrated prior to use in Area 500. In the **integrated** biorefinery, part of the hydrolysate is diverted to Area 550 for whole-slurry BDO fermentation; the remainder of the hydrolysate is clarified and concentrated as previously indicated before being routed to Area 500.

Area 400: Enzyme production. An on-site enzyme production section was maintained in this design, consistent with details provided in prior design reports. The whole broth from Area 400, containing the enzyme secreted by *Trichoderma reesei*, is fed to Area 300 to carry out enzymatic hydrolysis.

Area 500: Furans production and upgrading. The **integrated** biorefinery reroutes around 55% of the clarified, concentrated hydrolysate from Area 300 to the sugar dehydration reaction to form furans. Furfural and HMF present in the reactional mixture are then sent to an aldol condensation reactor, in which heavier oxygenated compounds (condensates) are synthesized with MEK as an additional reactant. MEK for this reaction is sourced from Area 550. The **dedicated** plant follows a similar path, only differing in that all sugars in the concentrated hydrolysate are sent to sugar dehydration and MEK is purchased to carry out the aldol condensation reaction. In both biorefineries, condensates ultimately undergo catalytic upgrading to hydrocarbon fuels through hydrotreating.

Area 550: MEK production. Whole slurry reserved from Area 300 (approximately 45% of total sugars) is sent to a batch anaerobic fermentation to BDO. The resulting BDO broth is clarified in a lignin filter press and then catalytically converted into MEK and other coproducts. MEK is recovered and routed to the aldol condensation reaction. Area 550 is exclusive to the **integrated** approach.

Area 600: Wastewater treatment. Wastewater streams from Areas 500 and 700 are treated by aerobic digestion and reverse osmosis to remove organics and salts, respectively. Sodium sulfate salt recovered from the brine is sold as a secondary coproduct to offset a portion of the caustic and acid demands/costs incurred throughout the integrated design.

Area 700: Lignin upgrading. This section focuses on upgrading lignin monomers and other residual (soluble) components to muconic acid via fermentation and further upgrading to adipic acid (finished coproduct, which is sold). Carbon sources for bioconversion are the black liquor from DMR pretreatment and the residual solids separated downstream after enzymatic hydrolysis and (in the **integrated** case) BDO fermentation.

Area 800: Combustor, boiler, and turbogenerator. The residual solids, wastewater sludge, and off-gas streams are combusted to generate high-pressure steam for heat and power.

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

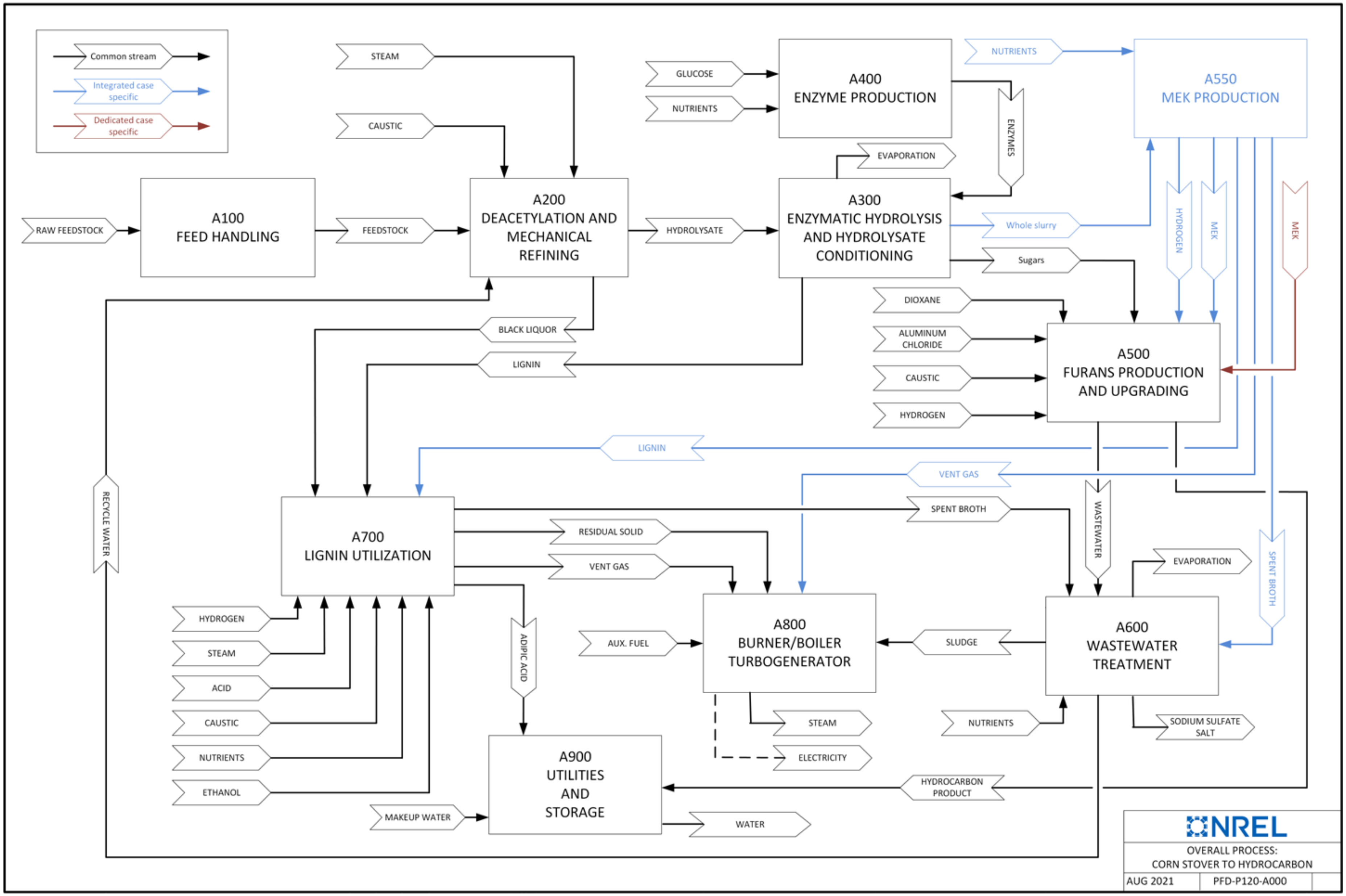

Figure 2. Simplified flow diagram of the overall processes

1.3 Techno-Economic Analysis Approach and Assumptions

The main goal of this analysis is to establish a techno-economic model to estimate an MFSP linked to the production of hydrocarbon fuels in a biorefinery context. The approach followed in this work is similar to previous efforts [3, 4].

Initially, process flow diagrams are established with the aid of researchers to ensure the validity of the proposed biorefining approach. The block flow diagram in Figure 2 is a direct result from this step. Then, process simulations are built and run in Aspen Plus to generate mass and energy balances, which are further employed to determine fixed and variable operational expenses (OPEX), as well as capital expenditures (CAPEX) with equipment purchase. The nonrandom two-liquid (NRTL) model is used as the base thermodynamic model throughout the simulations (with notable exceptions such as the Peng-Robinson or Soave-Redlich-Kwong equations of state and NRTL variations for specific unit operations). Baseline equipment costs may come from a multitude of sources, including vendor quotations, scientific literature, NREL and subcontractor historical cost data, or estimated with the built-in Aspen Capital Cost Estimator (ACCE) in Aspen Plus. Once equipment costs are determined, direct and indirect overhead cost factors (e.g., installation costs and project contingency) are applied to determine the total capital investment (TCI) in 2016$. The TCI and the plant OPEX are then used to establish a discounted cash flow rate of return (DCFROR) and to calculate the MFSP (in $/GGE) at a 10% internal rate of return over a 30-year plant lifetime. The underlying cost estimates and TEA modeling approach are compatible with an AACE Class 4 level of analysis [5], with an estimated uncertainty of approximately ±25% in the determined TCI [6].

2 Design Basis and Conventions

This section briefly presents considerations on biorefinery size, feedstock specifications, and modeling basis. Table 1 summarizes the main parameters linked to the commercial biorefinery modeled in this report. It is important to point out that such parameters are reflective of n^{th}-plant commercial-scale designs, which implies that several facilities of the same type have already been built and that the technology in question is mature, in agreement with what has been described in the 2018 design report. It should be clear that pioneer plants (first-of-a-kind) would present a different set of specifications that would influence its economic assessment.

Table 1. Design Basis for the Modeled Biorefineries

Parameter	Value
Plant nameplate capacity	2,205 dry U.S. tons/day (2,000 metric t/day)
Annual feedstock requirement	724,000 dry U.S. tons/year
Expected operation time	7,884 h/year
Equivalent uptime	90%
Facility startup time	0.5 years

Also consistent with prior recent design cases [1, 3, 4, 7], the delivered feedstock composition is left unchanged (shown in Table 2). More details on historical components used in Aspen Plus for process simulation and their modeled properties may be found in prior design reports [1].

Table 2. Delivered Feedstock Composition Assumed in the Present Design

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

[a] Represents acetyl groups present in the hemicellulose polymer, converted to acetic acid under low-pH conditions.

This report is consistent with previous efforts and communicates results in terms of energy yields in gallons gasoline equivalent (e.g., $/GGE, GGE/yr, GGE/ton) because the main product of the biorefineries is a fuel compatible with jet fuel and diesel cuts. All yields and MFSPs are normalized to a GGE basis according to their energy content to maintain the consistency of analysis among past and future NREL reports. Lower heating value (LHV) for the fuel product was calculated by an Aspen model and is similar to standard petroleum-equivalent (drop-in hydrocarbon) products [8]. To translate to a GGE basis, a conventional gasoline heating value of 116,090 British thermal units (BTU) per gallon (LHV basis) was considered [8].

3 Process Design and Cost Estimation Details

3.1 Area 100: Feedstock Logistics and Handling

Feedstock logistics and handling, as well as their impacts on the cost, volume, and quality of the biomass delivered to the biorefinery, are consistent with NREL's 2018 design report, based on inputs furnished by partners at Idaho National Laboratory. Feedstock cost is estimated at $71.26/dry U.S. ton (2016$) in this assessment, reflecting future 2030 goals for delivered biomass meeting the composition specifications listed in Table 2. Because this is a significant cost component when determining MFSP, further reductions in feedstock cost should have an important impact in achieving future MFSP goals below $2.50/GGE [1, 2].

3.2 Area 200: Pretreatment

Consistent with NREL's 2018 design report, pretreatment is accomplished through DMR, in which an alkaline extraction operation is utilized to remove acetyl groups, as well as a fraction of lignin, ash, biomass extractives, and other components. In comparison to previous studies centered around deacetylation and dilute acid pretreatment, DMR avoids feeding biomass into a high-pressure reactor, enables an effective ensuing enzymatic hydrolysis, and limits lignin degradation [1]. The latter is especially important in view of the additional purpose of the biorefinery on adding value to lignin in the form of adipic acid. In this design case, the pretreatment starts in a continuous counter-current extraction unit, in which biomass and a caustic solution are fed at opposite ends of a screw conveyor. The extracted solids are dewatered with a screw discharger at one end of the unit (leaving biomass at

30% total solids) and routed to mechanical refining. The caustic exits with the black liquor, which is sent to a base-catalyzed deconstruction step (along with residual downstream lignin solids) and then to aerobic fermentation to muconic acid in Area 700. Following alkaline extraction, DMR uses a two-stage mechanical processing operation and a disc refiner, followed by a secondary roller mill to open the biomass structure and render the fibers more available for enzymatic hydrolysis. Table 3 summarizes the main parameters used to model the DMR pretreatment step.

Table 3. DMR Pretreatment Conditions Applied in This Design [a]

Parameter	Current Design
Deacetylation (mild alkaline extraction)	
Temperature	92°C
Residence time	90 min
Total caustic (NaOH) loading	70 mg/g dry biomass
Net solubilized glucan to liquor (wt %)	2%
Net solubilized xylan to liquor (wt %)	10%
Net solubilized arabinan to liquor (wt %)	30%
Ash removal (wt %)	66%
Solubilized acetate (wt %)	100% (10 g/L)
Solubilized extractives (wt %)	100% (79 g/L)
Solubilized lignin (wt %)	47% (40 g/L)
Mechanical refining	
Solids loading (wt %)	30%
Power demand (kWh/dry tonne processed biomass)	200

[a] DMR parameters are generally based on NREL experimental data observed to date on DMR processing with batch deacetylation, but extrapolated to higher targeted black liquor concentrations and marginally lower carbohydrate losses targeted for counter-current deacetylation/alkaline extraction [1].

Costing of the units in Area 200 is kept consistent with the 2018 design report. The counter-current deacetylation unit was costed based on a Braunschweigische Maschinenbauanstalt AG sugar beet cossette extraction unit [9, 10]. Subsequently for mechanical refining, a total of 8 disc refiner units and 11 roller mill units are utilized with a purchase cost of around $2.5 million and $580,000 per unit, respectively, based on previous vendor quotations (2013$).

3.3 Area 300: Enzymatic Hydrolysis and Hydrolysate Conditioning

The enzymatic hydrolysis process is depicted in Figure 3, maintaining consistency for the hydrolysis operations with the details reflected in NREL's 2018 design report under the BDO pathway, namely initiating hydrolysis in a 24-hour vertical liquefaction vessel, after which point the material may be more easily mixed and accordingly is transferred to batch-stirred tank reactors. After mixing in the enzyme and associated broth from on-site enzyme production (Area 400), total solids loading into the continuous vessel is maintained at 25 wt %, consistent with prior targets [1]. The material is batched to one of eight 1-million (MM)-gallon agitated vessels (950,000 gal working volume), where enzymatic hydrolysis continues for another 96 hours (5 days total hydrolysis time). Temperature is controlled through all hydrolysis operations at 50°C. In the **integrated** biorefinery scenario, an additional amount of time (and thus reactor vessels) is devoted to sequential fermentation of sugars to BDO, with the associated additional tankage volume for fermentation allocated to Area 550. Total enzyme loading is targeted at 10 mg enzyme protein/g cellulose (including both cellulase and hemicellulase enzymes) with a target of 90% conversion of cellulose to glucose and xylan to xylose, as well as 85% arabinan to arabinose. The target design conditions and yields for enzymatic hydrolysis are summarized in Table 4.

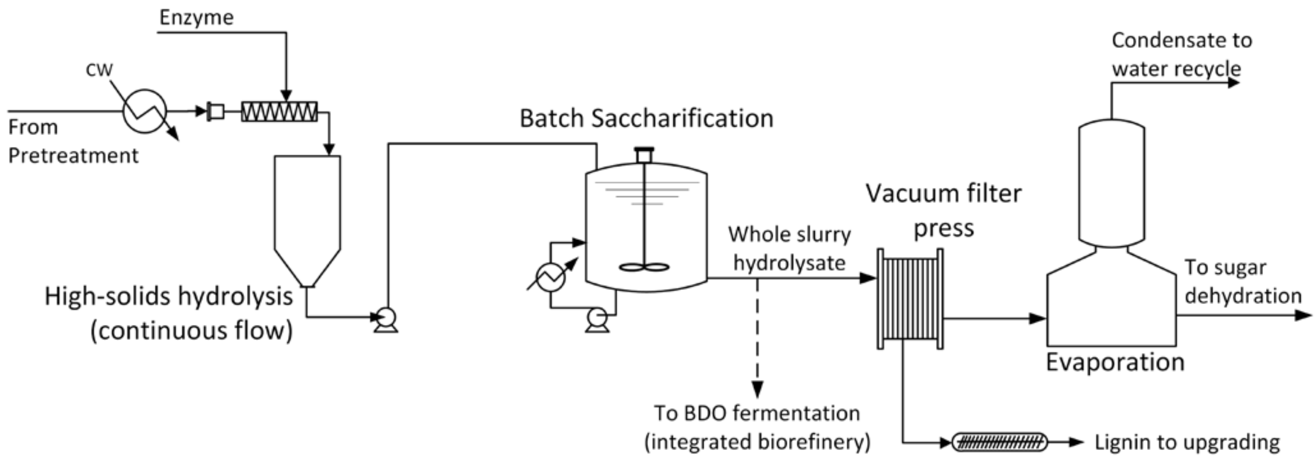

Figure 3. Process schematic diagram for enzymatic hydrolysis and hydrolysate conditioning operations

Table 4. Enzymatic Hydrolysis Conditions and Conversion Targets

Temperature	50°C (122°F)
Initial solids loading	25 wt % total solids
Residence time	5.0 days total (96 h)
Number and size of continuous vessels	6 @ 950 m^3 (250,000 gal) each
Number and size of batch vessels	8 @ 3,600 m^3 (950,000 gal) each
Total enzyme (cellulase + hemicellulase) loading	10 mg protein/g cellulose
Conversion: $(Glucan)_n + n\ H_2O \rightarrow n$ Glucose	90%
Conversion: $(Glucan)_n + n\ H_2O \rightarrow n$ Cellobiose	1.2%
Conversion: $(Xylan)_n + n\ H_2O \rightarrow n$ Xylose	90%
Conversion: $(Arabinan)_n + n\ H_2O \rightarrow n$ Arabinose	85%

Following batch saccharification, in the **dedicated** biorefinery scenario, the full hydrolysate stream is routed through a solids removal step, utilizing a vacuum filter press aided by a flocculant to support cake formation and wash water to support recovery of sugars in the liquor phase. Although costly, this operation is deemed necessary if placed sequentially following enzymatic hydrolysis, reflecting recent experimental learnings regarding the difficulty of filtering solids from DMR-pretreated hydrolysate. The **integrated** biorefinery case also routes the majority of the hydrolysate through the vacuum filter press operation for subsequent upgrading to furans, but a fraction (roughly 45%) carries on as raw whole slurry into BDO fermentation in order to generate the MEK co-reactant internally. The vacuum filter press operates at targeted parameters summarized in Table 5, achieving a 95% net recovery of sugars in the liquor phase (as may be found in NREL's sugar model scenarios available from https://www.nrel.gov/extranet/biorefinery/aspen-models/). The product liquor stream from the vacuum filter press is subsequently concentrated using a multistage mechanical vapor recompression (MVR) evaporator, achieving 49% sugar concentration (50% water), ultimately to enable appropriate targeted sugar concentrations in the feed stream to the furan catalytic upgrading reactor after combining with subsequent dioxane/water recycle. To avoid the possibility of sugar degradation at high temperatures [11], the evaporators are assumed to be operated under slight vacuum to keep the maximum temperature below 90°C. Key parameters for the MVR evaporator are also summarized in Table 6.

Table 5. Hydrolysate Clarification and Concentration Specifications

Clarification: flocculant loading	20 g/kg insoluble solids (IS)
Clarification: wash ratio	10 L/kg IS
Clarification: permeance	15 kg IS/m^2-h
Clarification: sugar recovery to liquors	95%
Concentration: product sugar concentration	49 wt % (50% water, 1% other solubles)
Concentration: maximum operating temperature	87°C
Concentration: evaporator technology	4-stage MVR
Concentration: electricity usage	5,657 kW (**dedicated** case) 3,191 kW (**integrated** case)

The cost assumptions for the hydrolysis reactors were maintained with prior NREL studies [1–3], namely empty towers for the continuous hydrolysis reactor based on a vendor quotation for flat-bottomed plug-flow reactors with a 10:1 height-to-diameter ratio, as well as 1-MM-gal batch hydrolysis reactors and agitators. As previously noted, in the **integrated** biorefinery case, the volume and tankage requirements for the batch hydrolysis portion of the process were allocated to Area 300 and additional volume/tank costs for subsequent BDO fermentation were allocated to Area 550. Capital and operating (flocculant) costs for the vacuum filter press and MVR evaporator were based on previous inputs furnished by vendors and engineering subcontractors.

3.4 Area 400: Enzyme Production

This process area produces cellulase and hemicellulase enzymes used in Area 300 to hydrolyze cellulose into glucose and xylan/arabinan into xylose/arabinose, respectively. Consistent with earlier design cases, the present design considers aerobic bioconversion of a *T. reesei*-like fungus on a purchased glucose substrate. Also as noted in the 2018 design report, one difference in the present work is that both cellulase and hemicellulase enzymes are required in the enzymatic hydrolysis step, but the same overall enzyme production process framework is maintained, assuming the costs of producing a quantity of enzyme protein are similar whether for cellulase or hemicellulase. The whole broth product is transferred to the hydrolysis tanks without an enzyme isolation step. Figure 5 provides a simplified flow diagram of the enzyme production section.

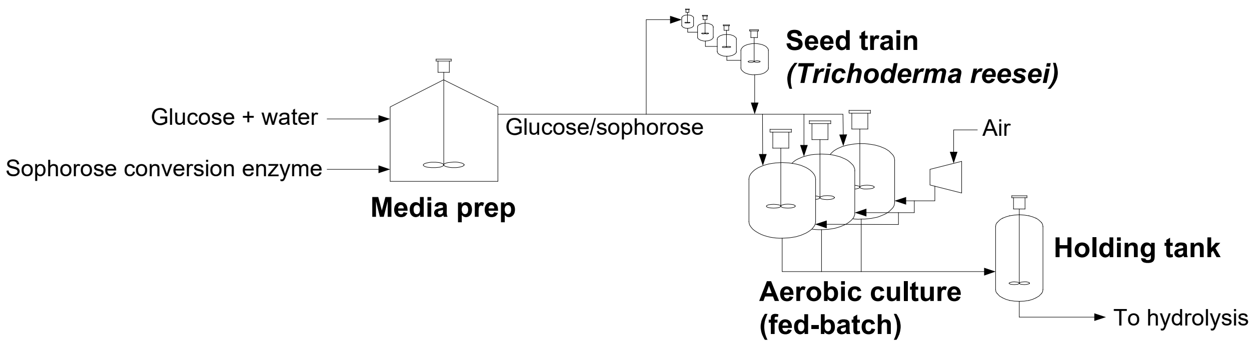

Figure 4. Simplified flow diagram of the enzyme production process

The key assumptions used in the current design are maintained consistently with prior models [1]. The targeted total cellulase and hemicellulase loading to enzymatic hydrolysis is maintained at 10 mg of enzyme protein per gram of cellulose, with an additional 10% produced to account for a slipstream provided to the media preparation tank. Each production vessel is sized at 300 m^3 (80,000 gal), operated in fed-batch between a 50% initial and 80% final volume, and with consistent cycle times documented in prior NREL reports. The required number of vessels are set for either biorefinery configuration, reflecting a productivity rate of 0.30 g protein/L-h. Required inputs to support enzyme production are supplied including glucose, corn steep liquor (CSL), ammonia, sulfur dioxide (SO_2), and corn oil antifoam. Temperature is maintained at 28°C using chilled water, and aeration inputs are set to satisfy required oxygen transfer rates based on governing oxygen solubilization considerations discussed previously [2]. Likewise, the inoculum seed train assumptions are also consistent with prior models, making use of four trains with three batch seed fermentors increasing in size from 0.3 to 3 to 30 m^3. The aeration demand is assumed to be 10% of the production aeration rate. The key design parameters, as well as targeted yields for enzyme production, are summarized in Table 6.

Table 6. Enzyme Production Bioreactor Design/Operating Parameters

Parameter	Assumption
Protein loading to enzymatic hydrolysis	10 mg protein/g cellulose
Reactor size	300,000 L @ 80% final working volume
Operating temperature	28°C
Enzyme titer at harvest	50 g/L
Mass yield of enzyme from glucose	0.24 kg enzyme/kg glucose
Enzyme production cycle time	120 h online, 48 h offline, 168 h total
Total electricity demand per kg protein (air compressors, agitators, chillers, pumps)	9 kWh/kg

The cost estimation for all equipment in Area 400 was left unchanged from prior NREL design reports making use of this process area. Quotations for the production bioreactors, internal cooling coils, production agitators and motors, skid-mounted seed fermentors, and air compressor were provided by vendors through a previous engineering subcontract. Not included in the enzyme production model are any costs for concentration, stabilization, or transportation of the enzyme to the plant, which would not be required in this case for on-site production. Applicable licensing fees are not included, with rationale discussed previously for implications such as this and other comparative details relative to externally purchased enzymes [2].

3.5 Area 500: Furans Production and Upgrading

3.5.1 Overview

Area 500 represents the core portion of the biorefinery of "new" focus in this work, in which sugars (either part of them in the integrated case or the totality of them in the dedicated case) are converted to furans and ultimately to hydrocarbon fuels in a series of catalytic steps.

In short, this is a thermocatalytic process converting sugars to hydrocarbons via furans (furfural and HMF) as intermediates. The furans pathway is centered around the catalytic production of furfural and HMF from pentose and hexoses, respectively, contained in clarified corn stover hydrolysate. The sugar dehydration reaction is carried out with a mixture of water and dioxane as a necessary solvent and $AlCl_3$ as the catalyst. The stream containing furfural and HMF is then mixed with NaOH and MEK to undergo aldol condensation. The reactional mixture is then sent to a distillation column for dioxane recovery, which yields an aqueous dioxane stream close to its azeotropic composition for recycle. The high-molecular-weight condensates from the bottom of the column are separated from the aqueous phase in a decanter. Finally, furans/MEK condensates are ultimately deoxygenated with H_2 in a high-pressure catalytic fixed-bed reactor to produce hydrocarbons in the C_{14}–C_{16} range. Figure 5 depicts the overall process diagram flow for this area and Figure 6 indicates the reactional steps to reach hydrocarbon fuels starting from hydrolysate sugars.

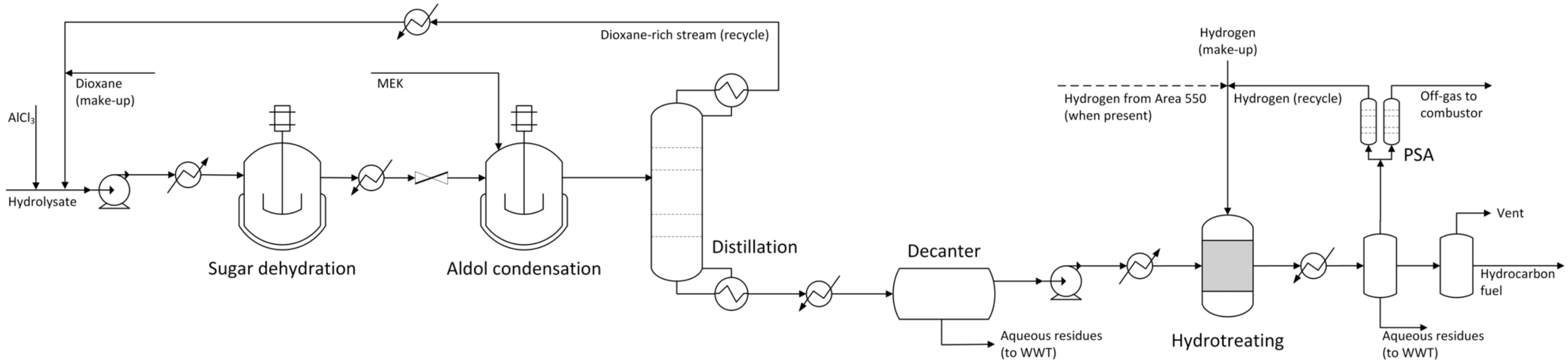

Figure 5. Simplified flow diagram of the furans production and upgrading section

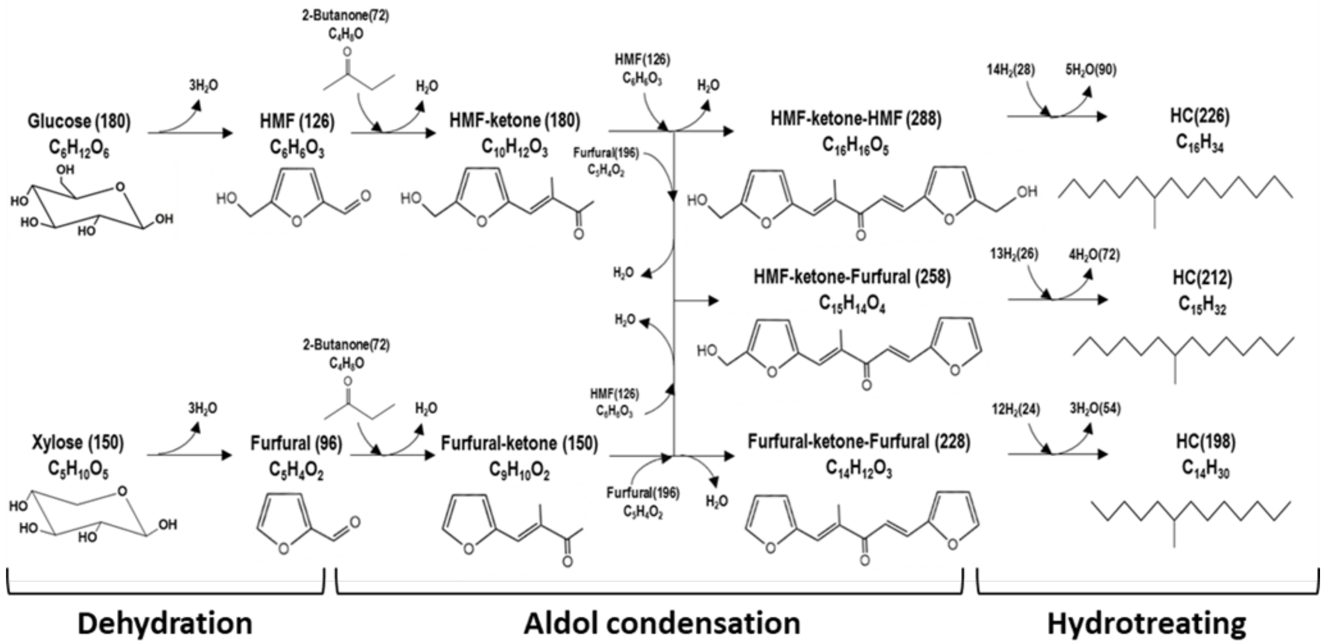

Figure 6. Main reactions taking place in furans production and upgrading section

3.5.2 Design Basis

Sugar Dehydration to Furans

The catalytic upgrading process starts with the hydrolysate that was conditioned in Area 300 through filtering to remove lignin and concentration via an MVR evaporator. The hydrolysate is concentrated up to a combined sugar fraction of 49 wt %. The analysis showed that there are marginal gains in terms of MFSP for increasing sugar concentration in the hydrolysate stream: passing from 30 wt % combined sugars up to 40 wt % reduces MFSP by \$0.03/GGE, whereas increasing it to 49 wt % means gaining another \$0.05/GGE. It should be highlighted that concentrating the hydrolysate up to 49 wt % combined sugars corresponds to an actual sugar concentration of 29 wt % in the dehydration reactor considering the water that is recycled with dioxane after solvent recovery. This is consistent with recent experimental developments that aim at a sugar feed concentration of 30 wt % for dehydration. The main reasons for this behavior are leaner equipment and lower energy requirements for heating, cooling, and pumping in Area 500. After mixing with the catalyst ($AlCl_3$) and dioxane, the stream is pressurized and heated before being fed into the sugar dehydration reactor. The operating conditions for the reactor are based on recent experimental work, including pressure of 24.8 atm (350 psi), temperature of 230°C, $AlCl_3$ utilized in catalytic quantities (1:36 molar ratio to sugars), as well as use of dioxane solvent at a volume ratio in relation to the full aqueous hydrolysate of 1.5:1 to promote high yields to furans. See Table 7 for a summary of the main considered parameters. In the reactor, the sugars are fully converted to the respective furans, both monomers (glucose, xylose, and arabinose) and oligomers (sucrose and cellobiose). At maximum conversion, furfural yield is 64 wt % and that of HMF corresponds 70 wt % relative to corresponding sugar feeds. Table 7 also presents the reactions occurring in the reactor and Figure 6 indicates the overall reactional path.

Table 7. Sugar Dehydration Conditions and Conversion Targets

Lewis acid ($AlCl_3$) loading	1:36 mol $AlCl_3$/mol sugars
Dioxane:hydrolysate ratio (v/v)	1.5:1
Temperature	230°C
Pressure	24.8 atm (350 psi)
Reaction time	5 min, continuous
Catalyst lifetime	N/A (homogeneous)
Conversion: Glucose → HMF + 3 H_2O	100% (glucose)
Conversion : Xylose → Furfural + 3 H_2O	100% (xylose)
Conversion: Arabinose → Furfural + 3 H_2O	100% (arabinose)

Aldol Condensation

The reactional mixture produced from the first step is conditioned through cooling and pressure reduction down to conditions required in the aldol condensation reaction (60°C and atmospheric pressure). NaOH is used as the catalyst and MEK is the ketone of choice to react with furfural and HMF (although other ketones are possible for this step as well). Both compounds are mixed before being routed to the reactor. The condition at which MEK is delivered to the aldol condensation reaction differs according to the setup of the biorefinery. In the **dedicated** case, MEK is considered to be purchase externally, entering the process at 99% purity (though such a high purity is not needed for this operation). On the other hand, in the **integrated** biorefinery, MEK is alternatively sourced from Area 550 and routed to the aldol condensation at 84.9 wt % purity, close to its azeotropic composition with water. This approach was chosen to avoid an azeotropic distillation setup in Area 550, thus minimizing CAPEX and energy consumption for the recovery of MEK. Besides, the distillation column for dioxane recovery after the aldol condensation reactions is able to handle the amount of water brought by the MEK-rich stream from Area 550. The full approach for separating MEK from water and minor coproducts from BDO dehydration is discussed in detail in Section 3.6.2.

Table 8 summarizes the main conditions employed in the aldol condensation reaction. It is important to highlight that the reaction reaches already high conversions in a relatively short time (around 20 min), but the total reaction time is set at 60 min to ensure the full conversion of furans into condensates. The furans/MEK condensates are formed by combining one MEK molecule and either two furfural molecules (F-MEK-F), two HMF molecules (H-MEK-H), or one of each furan (H-MEK-F). The reactional path is indicated in Figure 6, and Table 8 also presents the reactions considered in the simulations. MEK requirements are tailored to ensure full conversion of furans and eliminate residual MEK in the outlet stream of the reactor.

Table 8. Aldol Condensation Conditions and Conversion Targets

Conversion	100%
Catalyst	NaOH (homogeneous)
Residence time	60 min
Temperature	60°C
Pressure	1 atm
NaOH loading	3.6 g/L
Conversion: 2 HMF + MEK → H-MEK-H + 2 H_2O	28% (MEK)
Conversion: 2 Furfural + MEK → F-MEK-F + 2 H_2O	8% (MEK)
Conversion: HMF + Furfural + MEK → H-MEK-F + 2 H_2O	64% (MEK)

Dioxane Recovery

After aldol condensation, the reactional mixture contains mainly water, dioxane, furans/MEK condensates, and other minor solutes, namely $AlCl_3$ and NaOH. Because the process is dependent on large amounts of dioxane as a cosolvent, it is imperative to recycle it to the inlet of the sugar dehydration reactor while keeping dioxane makeup to a minimum. The proposed configuration for dioxane recovery nearly avoids dioxane losses and minimizes impurities in the dioxane recycle stream. The distillation column (depicted in Figure 5) is designed to recover 99.9% of the dioxane at approximately 78 wt % purity. These performance targets were deemed as optimal to avoid overburdening the distillation column with a high reflux ratio and therefore a high energy consumption. A preliminary setup including a second column with an extractive distillation using benzene as the entrainer was assessed to determine the feasibility of recovering and recycling dioxane at a higher purity (97.5 wt %). This configuration resulted in much larger heating and cooling demands and in the purchase of large quantities of external boiler fuel, which ultimately led to it being dropped from the final version of either biorefinery approach in favor of including some water in the recycle.

After dioxane removal from the reactional mixture at the top of the distillation column, the bottoms are routed to a decanter, in which the heavy furans/MEK condensates are easily separated from the aqueous phase. While the heavy organic molecules are sent to hydrotreating, the aqueous phase is sent to WWT (Area 600). NaOH added to the reaction and still present in this stream is partially recovered as sodium sulfate after treatment with H_2SO_4. $AlCl_3$ used as the catalyst in the sugar dehydration reaction is not recovered or removed prior to aldol condensation, being routed to Area 600 for disposal.

Catalytic Upgrading to Hydrocarbons

In the hydrotreating section, the condensates are pressurized and heated up to the conditions required by the final hydrotreating step. These compounds undergo deoxygenation over a Pd catalyst (supported by Al_2O_3/SiO_2) at a weight hourly space velocity (WHSV) of 2 h^{-1} with H_2 in stoichiometric excess to yield hydrocarbons. In the target case, all condensates are converted via hydrodeoxygenation, meaning that no CO_2 is lost through decarboxylation. The reactor outlet stream is routed to a flash vessel for degassing; gases are then sent to a pressure swing adsorption (PSA) unit to maximize H_2 recovery for recycling (95% recovery of surplus H_2). The tail gas from the PSA is burned in a dedicated furnace to partially supply the energy requirements of the hydrotreating feed heating section. Finally, the resulting hydrocarbons, mainly isoparaffins in the range of C_{14} to C_{16}, are dried and sent to storage. Table 9 summarizes the main conditions of the hydrotreating reactor and the reactions that take place.

Table 9: Hydrotreating Conditions and Conversion Targets

Conversion	100%
Catalyst	1% $Pd/Al_2O_3\text{-}SiO_2$
WHSV	2 h^{-1}
Temperature	300°C
Pressure	98.7 atm
Hydrodeoxygenation (HDO) vs. decarbonylation	100:0
Hydrocarbon fuel composition	C_{14}–C_{16} branched hydrocarbons
Conversion: H-MEK-H + 14 $H_2 \rightarrow C_{16}H_{34}$ + 5 H_2O	100% (H-MEK-H)
Conversion: F-MEK-F + 12 $H_2 \rightarrow C_{14}H_{30}$ + 3 H_2O	100% (F-MEK-F)
Conversion: H-MEK-F + 13 $H_2 \rightarrow C_{15}H_{32}$ + 4 H_2O	100% (H-MEK-F)

3.5.3 Cost Estimation

Processing in Area 500 occurs in some central pieces of equipment, namely the three reactors for sugar dehydration, aldol condensation, and hydrotreating. The sugar dehydration reactor is dimensioned for a 5-min residence time. The base equipment is a 30,000-gal tubular flow reactor able to withstand high pressures and temperatures estimated at $16.3 MM (2013$). The aldol condensation reaction is set at 60 min and the base reactor for scaling is a 200,000-gal, atmospheric pressure vessel estimated at $1.2 MM (2009$). Finally, the hydrotreating operation is based on an HDO reactor from previous efforts [1]. The base unit is costed at $6.5 MM (2011$) for a liquid flow of 33,000 L/h.

The dioxane recovery column, as well as the decanter, were costed using ACCE. Other auxiliary units (pumps, heaters, compressors, PSA unit) were retrieved from previous NREL cost databases and scaled accordingly. The full equipment list and the installed costs for the items in each biorefining approach (**dedicated** or **integrated**) after scaling can be found in Appendix A.

3.6 Area 550: MEK Production

3.6.1 Overview

Area 550 includes all processing steps for conversion of a fraction of sugars in the hydrolysate to BDO via biological conversion (anaerobic fermentation), as well as catalytic upgrading of this intermediate to MEK and its recovery. As noted, Area 550 is only included for the **integrated** biorefinery scenario to fully reflect the economic implications for a single 2,000-tonne/day biorefinery configured to coproduce both the furan and MEK intermediates together, while MEK is purchased in the **dedicated** case.

In the BDO pathway, after batch enzymatic hydrolysis is completed, the hydrolysate is cooled and batch fermentation is initiated in the same vessels utilizing the whole hydrolysate slurry with solids. BDO fermentation utilizes an engineered strain of *Zymomonas mobilis* to convert sugars to 2,3-BDO plus hydrogen as a byproduct for achieving cell redox balancing (the latter is purified from the fermentation off-gas and used in Area 500 to help meet catalytic upgrading hydrogen demands). The fermentor product broth is routed to a lignin press to remove solids, and then to a polishing filter to remove particle fines and ion exchange to mitigate ionic/salt species that may be problematic for the downstream catalyst. The aqueous BDO stream is then heated at elevated pressure and routed to catalytic BDO upgrading, producing MEK and minor byproducts (following guidance from Pacific Northwest National Laboratory [PNNL] collaborators focused on this BDO-to-MEK pathway). The MEK product is recovered in a series of distillation columns and sent to Area 500 to be employed as a reactant in the aldol condensation of furans. Figure 7 depicts the schematic for this process.

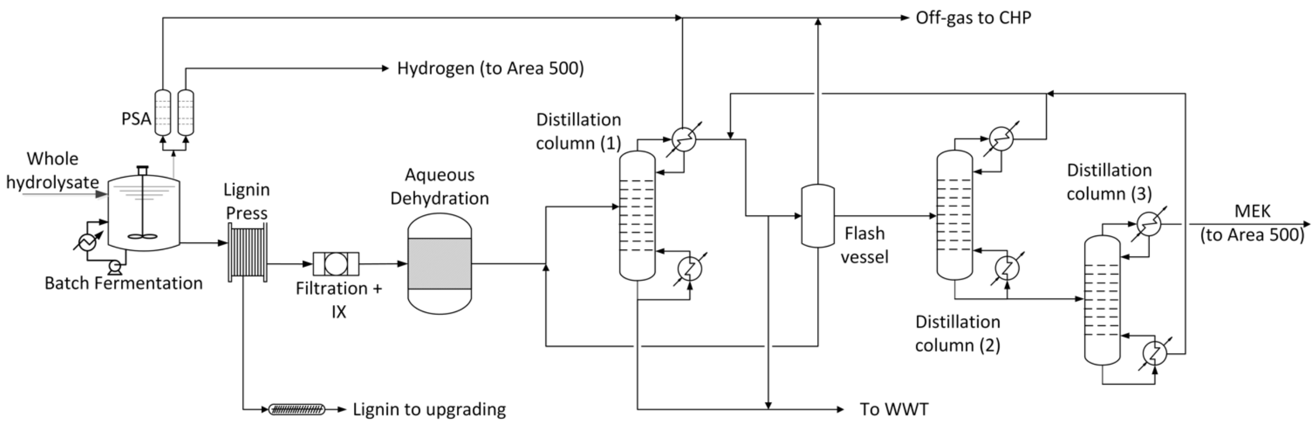

Figure 7. Simplified flow diagram of BDO fermentation, catalytic upgrading to MEK, and ketone recovery in the integrated biorefinery

3.6.2 Design Basis

Fermentation/Product Recovery: BDO

All details for the BDO fermentation and broth clarification/cleanup steps are identical to those documented in the 2018 design report, but at lower throughput scales based on a fraction of the hydrolysate sugars dedicated to this train in parallel to the furan synthesis train. In summary, an engineered strain of the fermenting bacterium *Z. mobilis* converts glucose, xylose, and arabinose sugars to 2,3-BDO, following similar details as NREL's previous models for ethanol fermentation using this organism [2]. However, in contrast to ethanol fermentation, in order for 2,3-BDO fermentation to be performed fully anaerobically, hydrogen is coproduced to satisfy cell redox balances, which is separated from the fermentor vent stream using a PSA unit and routed for use in Area 500. Additionally, given less toxicity tolerance sensitivity to BDO than ethanol, BDO fermentation is configured to achieve a target titer near 100 g/L based on maintaining elevated enzymatic hydrolysis solids concentration targets of 25 wt % upstream, with resultant hydrolysate processed through fermentation without solids removal.

BDO fermentation is performed in the same 1-MM-gallon agitated batch vessels as used for enzymatic hydrolysis, after cooling to 32°C (but as noted, the extra tankage volume and associated costs for fermentation are allocated to Area 550 beyond enzymatic hydrolysis tankage under Area 300). The process targets overall sugar utilizations of 95%, 90%, and 85% for glucose, xylose, and arabinose, respectively (as well as 95% utilization of the biomass sucrose), translating to a targeted process yield of 0.47 g BDO/g total sugars over a 1.5-day fermentation batch time. The assumptions for seed train design and conversions are also maintained consistently with the details documented in the 2018 design report, in turn reflective of seed train logistics for *Z. mobilis* cell mass inoculum production described in the 2011 ethanol report [2]. Namely, 10% of the hydrolysate to Area 550 is split to inoculum growth consisting of two trains of five reactors each, increasing in size up to 200,000 gal (757 m^3) operated in 24-hour batch cycles. The product from the seed train is recombined with the remainder of the hydrolysate for bioconversion in the main production fermentor vessels to BDO, after also accounting for a standard 3% loss of sugars to contamination represented as lactic acid. All key BDO fermentation conditions and parameters are summarized in Table 10.

Table 10. BDO Fermentation Conditions and Conversion Targets

Organism	Recombinant *Z. mobilis*
Temperature	32°C (96°F)
Initial fermentation solids level	25% total solids
Residence time	1.5 days (36 h) = 2.6 g/L-h productivity
Inoculum level	10 vol %
Inoculum production: number of seed trains	2
Inoculum production: number of batch stages	5
Inoculum production: maximum stage volume	200,000 gal (757 m^3)
Conversion: Glucose → BDO + H_2 + 2 CO_2	95%
Conversion: Glucose + 0.047 CSL + 0.018 DAP [a] → 6 *Z. mobilis* + 2.4 H_2O	2%
Conversion: 6 Xylose → 5 BDO +5 H_2 +10 CO_2	90%
Conversion: Xylose + 0.039 CSL + 0.015 DAP → 5 *Z. mobilis* + 2 H_2O	1.9%
Conversion: 6 Arabinose → 5 BDO +5 H_2 +10 CO_2	90%
Conversion: Arabinose + 0.039 CSL + 0.015 DAP → 5 *Z. mobilis* + 2 H_2O	1.9%
Overall BDO process yield, g/g sugars	0.47

[a] Diammonium phosphate

Following completion of the fermentation batch cycle, the fermentor broth is routed to a clarification step, employing a lignin press to remove lignin and other residual solids, utilizing consistent assumptions as the 2011 ethanol design report for this operation as was used to clarify ethanol beer stillage [2]. This step achieves 98% removal of insoluble solids and reduces water content in the solid material from 80 to 25 wt %; however, it also incurs a small 3% loss of BDO product. The majority of the *Zymomonas* biomass is also removed here, which is routed to the boiler and incinerated (as necessary for engineered organism destruction). Recent experimental efforts have suggested that the use of this low-cost lignin press may prove challenging for removal of solids (whether before or after fermentation) based on DMR pretreatment upstream and resulting small particle sizes that make filtration difficult. At present, this unit is maintained for consistency with the published TEA model details for the baseline BDO pathway in the 2018 design report, but this may be revisited moving forward, at which point for this pathway it may be more sensible to process the entire hydrolysate through the vacuum filter press (utilized for furan synthesis in Area 500), which also may allow for more flexibility in subsequent BDO fermentation approaches as may utilize clarified sugars through fed-batch processing to further increase BDO titers.

Catalytic Upgrading: BDO to MEK

The clarified BDO fermentation broth is further purified across a polishing filter (microfilter) to remove particle fines, followed by ion exchange to remove soluble cations and anions that may otherwise deactivate downstream catalysts. These operations were maintained here based on subcontractor guidance during NREL's 2015 catalytic upgrading design case focused on catalytic aqueous-phase reforming of sugars to fuels [4], where they were deemed necessary to ensure catalyst protection at least in the context of those reactor systems. The polishing filter consists of parallel crossflow microfiltration skids with a pore size of 0.1 microns. The filtrate is routed to ion exchange for further purification. A separate-bed ion exchange system is utilized to remove a range of ionic species. Two resin bed trains are included—one for anions and one for cations. There are two units installed in parallel to allow for regeneration of the resin, which is assumed to be required every 17 hours [1]. Resin is regenerated with acid and caustic. An additional 1% loss of BDO is assumed across the combination of these two purification steps. The purified stream is then routed to BDO upgrading.

The 2,3-BDO product may undergo catalytic upgrading either based on high-purity or bulk aqueous catalysis. The former would be preferred from a capital cost and energy standpoint for catalytic upgrading, but traditional BDO purification requires energy-intensive distillation to boil all the water off (roughly 90%) from the higher-boiling BDO component, which would incur unacceptably high heat demands. The Bioprocessing Separations Consortium (https://www.bioesep.org/) is conducting research in part focused on more cost- and energy-efficient ways to concentrate BDO, with future cost-trade-off studies planned to evaluate economic impacts between separation costs versus savings in catalytic upgrading. At present, aqueous catalytic upgrading is assumed, making use of the full clarified fermentation broth directly without BDO concentration, consistent with the BDO upgrading pathway documented in the 2018 design case.

For 2,3-BDO aqueous catalytic upgrading to MEK, bifunctional solid acid catalysts are used. This pathway is currently under investigation by PNNL under CUBI. The model for this process has been built based on inputs furnished by PNNL researchers for consistency purposes. In the present design, the aqueous BDO stream containing roughly 10 wt % BDO, 88 wt % water, and 2 wt % other components (primarily unconverted sugars) is considered to be converted adiabatically to dehydration intermediates over a heterogeneous catalyst [12]. The composition of products is dependent on the composition of the feed stream, the reaction temperature and pressure, space velocity, and catalyst type. The catalyst considered in this TEA effort is an acid-based one, such as one supported by SiO_2-ZrO_2 or zeolite. As shown in Table 11, 100% of the BDO is converted to MEK, isobutanol, 1,3-butadiene, isobutanol, and 3-buten-2-ol. Stoichiometric element balances are closed with water, CO_2, and O_2 formation, with model convergence based on the NRTL property package given the presence of oxygenated/polar components (this is the default property package in most of the biochemical model steps for both fuel train pathways).

Table 11. Product Distribution of the 2,3-BDO Upgrading Reaction (Future targets at high conversion and selectivity to desired alkenes); based on inputs from PNNL [13]

Compound	Yield (mol%)
MEK	80.3
Isobutanol	9.7
1,3-butadiene	3.1
Isobutanal	3.2
3-buten-2-ol	3.2
CO_2	2.0
H_2O	95.4
O_2 [a]	2.1

[a] O_2 added in small amounts for mass balance closure purposes

Key reaction parameters are summarized in Table 12. Pressure was set at 250°C in the reactor as a future target condition (the biorefinery is virtually unaffected by changing the reaction temperature to 280°C [current experimental condition], as discussed in Section 5.2.2). In a similar way, pressure is not a critical parameter to the reaction and was set at 10 atm. Detailed costs are presented in Section 4.2.

Table 12. 2,3-BDO Catalytic Upgrading Reactor Conditions

Parameter	Operating Condition
Operating temperature	250°C
Operating pressure	1 atm
WHSV (h^{-1}) [a]	0.24 h^{-1}
Catalyst type	Copper-based catalysts Cu/SiO_2-ZrO_2, Cu/zeolite
Catalyst lifetime	2 years

[a] WHSV of 0.24 h^{-1} refers to 2,3-BDO only (2.4 h^{-1} for the full 2,3-BDO-containing stream

The MEK product from the BDO catalytic upgrading step is sent to a series of distillation columns and a liquid-liquid separation to separate the MEK from water and the other byproducts of the reaction. This separation step was modeled using the NRTL model and the Redlich-Kwong equation (NRTL-RK) property method to properly reflect the azeotropes between water/MEK, water/isobutyraldehyde, and water/isobutanol. Figure 8 presents an overview of the three-step distillation system designed to achieve the required MEK recovery for further use in Area 500.

3.6.3 Cost Estimation

As noted, BDO fermentation takes place in the same physical vessels as hydrolysis (sequential hydrolysis and fermentation, similar to the 2011 ethanol case)— utilizing the 1-MM-gal batch-stirred tank bioreactors. However, for cost allocation purposes, the fraction of total batch time and associated volume spent in fermentation is allocated to Area 550 and vice versa for hydrolysis allocated to Area 300. All design/cost details for the main fermentation and seed equipment are described in the 2011 ethanol design report; briefly, the large 1-MM-gal fermentors, seed fermentors (including cooling coils), seed hold tank, and lignin press were quoted previously by vendors.

A filtered hydrolysate storage tank with a residence time of 20 minutes is included to provide intermediate storage for filtrate from the belt filter. The clarified BDO polishing filtration system includes two parallel skid microfiltration units, including clean-in-place system and backwash. The ion exchange system uses a separate-bed configuration, with strong acid cation and weak base anion resins based on scaling from cost quotations provided by an engineering subcontractor in NREL's 2015 design case (the provided cost did not break out equipment versus resin cost details separately).

The 2,3-BDO upgrading reactor cost is based on the aqueous phase reforming reactor in NREL's 2015 catalytic upgrading design case [4], sized based on the total aqueous flow rate into the reactor. The reactor operation consists of packed-bed pressure vessels, clad in 317L SS, and includes internals that support the catalyst and distribute the process fluid. Finally, equipment used in MEK recovery (distillation columns and flash vessel) were costed using ACCE, and auxiliary units (pumps and heaters) were fetched from previous NREL design reports.

3.7 Area 600: Wastewater Treatment

Wastewater is generated in the process from condensed pretreatment flash vapor in Area 200, the fermentation broths in Areas 550 and 700 (after separation of product and cells), water removed from separations within Area 500 (including water generated during catalytic upgrading reactions), and minor sources such as boiler and cooling tower blowdown. All such wastewater is sent to the WWT system in Area 600. After treatment, the effluent water is assumed clean and fully reusable by the process, reducing both the fresh makeup water requirement and discharge to the environment. All assumptions pertaining to WWT are maintained consistently with the 2018 design report. This also includes the prior justification to remove the anaerobic digestion unit, in light of much lower chemical oxygen demand loading in the combined WWT feed stream relative to prior designs, driven primarily by high utilization of sugars in the fuel train as well as lignin and unconverted organics in the lignin train, leaving lower residual organic matter to be processed through WWT. The updated simplified flow diagram is shown in Figure 8.

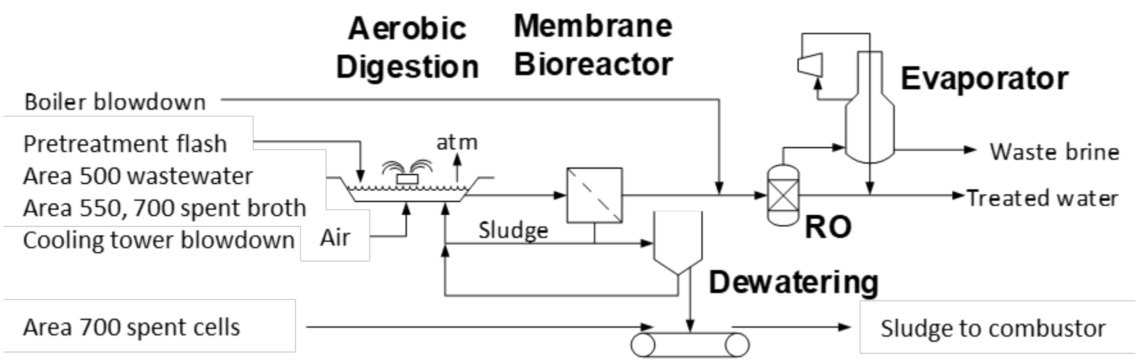

Figure 8. Simplified flow diagram of the WWT process

The aerobic system comprises several large basins or ponds that are aerated from the bottom via a grid sparger, removing 96% of soluble organic matter and converted to water, CO_2, and cell mass. The water in the basins is continuously cycled through membrane bioreactors, which are aeration tanks with ultrafiltration membranes that yield a clean water stream and a low-solids biomass sludge stream, which is mostly recycled to the basin. The wasted fraction of sludge that is not recycled passes first through a gravity belt thickener, then is subsequently centrifuged to >20% solids. The resulting digester sludge is routed to the boiler. The water effluent of the membrane system is sufficiently clean to feed directly to reverse osmosis to remove dissolved salts. The reverse osmosis permeate is recycled to the process and the reverse osmosis retentate is concentrated in an evaporator to produce a brine (sodium sulfate salt), sold in this design as a coproduct after further drying and crystallization. The evaporator condensate is combined with the reverse osmosis permeate for reuse in the process.

In previous NREL design reports, the individual unit costs were scaled to either the hydraulic flow rate or chemical oxygen demand load to anaerobic digestion. Similar to the 2018 design case, with the removal of the anaerobic digestion unit, the basis flows for costing were readjusted to reflect the appropriate feed location entering the aerobic digester. Additional equipment costs for sodium sulfate purification were scaled from a recent subcontract with Nexant for an evaporation flash drum, dryer, and centrifuge designed for this purpose.

3.8 Area 700: Lignin Upgrading

This process area includes the key unit operations for deconstruction and conversion of lignin (and other residual biomass components) to coproducts, as a critical element to enabling future MFSP targets. Similar to other operations outside of Areas 500/550, the lignin utilization process was maintained consistently with the 2018 design report, with more extensive documentation of all operations and assumptions documented there [1]. In summary, the black liquor product from DMR pretreatment is combined with the residual solids stream from the vacuum filter press and (in the **integrated** biorefinery case) solids from BDO fermentor broth clarification and routed to a more severe "base-catalyzed deconstruction" (BCD) unit employing a caustic soaking process at elevated temperature to further deconstruct high-molecular-weight lignin to metabolically accessible compounds. The caustic from DMR pretreatment that exits with the black liquor offsets a large portion of the caustic required to achieve the elevated pH in the BCD step. The soluble liquid BCD product is then routed to bioconversion, which converts soluble lignin and other residual organics to muconic acid through an aerobic pH-controlled bioconversion with *Pseudomonas putida*. The product is initially produced as a muconate salt at the given fermentation pH (i.e., the caustic from DMR/BCD neutralizes muconic acid to maintain near-neutral pH, forming sodium muconate), which is then isolated in high purity as muconic acid crystals following acid addition and low-pH crystallization. The muconic acid crystal is redissolved in a carrier solvent (ethanol) and hydrogenated under mild conditions to adipic acid, which is subsequently recovered through another crystallizer. A schematic diagram for the process is shown in Figure 9.

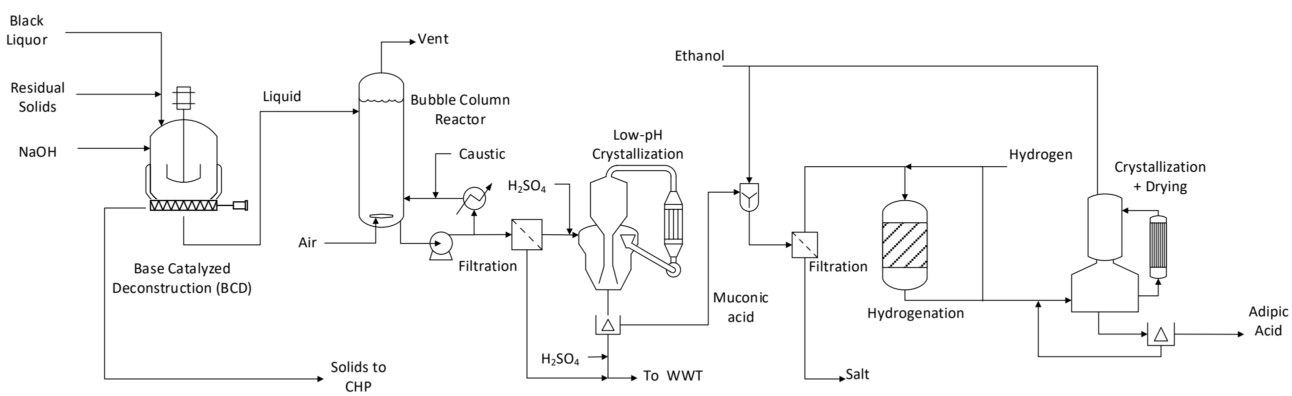

Figure 9. Schematic diagram of lignin deconstruction and conversion process to coproducts

In the BCD step, caustic (NaOH) is required at a minimum of 2 wt % loading based on the combined black liquor/residual solids feed mixture. Similar to the 2018 design case, after accounting for the amount of caustic provided in deacetylation upstream, the combined BCD feed stream has a caustic loading of 5.1%, in excess of the 2% threshold, and no additional caustic is needed. The treated mixture is pressurized to 6.3 atm, then reacted for 30 minutes at a temperature of 120°C (though higher temperatures may be reflected in the future if found to support better lignin deconstruction). In keeping with the 2018 design basis, 36 wt % of the solids entering the BCD reactor are solubilized to metabolically accessible monomers/oligomers. Of the solids present, this includes 48 wt % solubilization of carbohydrates and 53 wt % of the lignin. Residual solids are separated and sent to the boiler, while the solubilized product is cooled and then routed to fermentation. Table 13 summarizes the BCD reaction conditions and conversions used in the current design.

Table 13. Reaction Conditions and Key Parameters for Lignin BCD

Temperature	120°C
Pressure	6.3 atm
NaOH loading (minimum required), wt %	5.1% (2%)
Residence time	30 min
Total solubilization, wt % solids	36%
Deconstruction extents	
Carbohydrate deconstruction (cellulose, xylan, arabinan)	48%
Lignin deconstruction	53%

All bioconversion assumptions are also maintained consistent with those documented in the 2018 design report. This includes the use of a seed train reflecting three stages of sequentially larger reactor volumes from 0.3 to 3 to 100 m^3, consuming only sugars in a ratio of 46% conversion to cell mass and 54% conversion to muconic acid (applied equally to glucose, xylose, and arabinose sugars). This also includes routing the remaining majority of the feed to production bioreactors utilizing 1,000-m^3 bubble column vessels with a pump-around loop circulated through a cooler, and compressed air delivered to satisfy required oxygen transfer rates matched up with fermentation production rates. The net metabolic stoichiometry and conversion of the muconic acid fermentation is a combination of theoretical maximum metabolic yields and maintenance/respiration reactions dependent on the substrate consumed. For the production reaction, Table 14 lists the biological reactions and conversion to product for the carbon sources in the feed stream. The fermentation is run in a fed-batch mode, with an assumed average working volume of 70%. Temperature is controlled at 32°C and operates at a mild positive pressure of 1.34 atm, which assists in maintaining the axenic process. Overall, to completely convert the feed stream requires 18 and 19 bubble columns in the **integrated** and **dedicated** plants, respectively. The current design implements pH control to neutralize the muconic acid produced during the fermentation (as needed in excess of the caustic in solution) administered in the pump-around loop after cooling the broth. The overall fermentation is targeted to occur at a net productivity of 1 g muconic acid/L/h across all consumed substrates. Table 14 also summarizes the major reactor specifications for the production system.

Table 14. Lignin Fermentation Conditions and Conversion Targets

Seed train volume (stage 1, 2, 3)	0.3, 3, 100 m^3
Number of seed trains	3
Production reactor size	1,000 m^3
Production reactor temperature	32°C
Muconic acid productivity	1.0 g/L-h
Net muconic acid titer	68.5 g/L
Conversion: Glucose + 1.18 O2 + 0.28 NH_3 → 4.8 *P. putida* + 1.2 CO_2 + 1.98 H2O	46%
Conversion: Glucose + 1.94 O_2 → 0.74 Muconic + 1.57 CO_2 + 3.78 H2O	54%
Conversion: Xylose + 0.98 O2 + 0.23 NH_3 → 4 *P. putida* + 1 CO_2 + 1.64 H2O	46%
Conversion: Xylose + 1.57 O_2 → 0.62 Muconic + 1.26 CO_2 + 3.13 H2O	54%
Conversion: Arabinose + 0.98 O2 + 0.23 NH_3 → 4 *P. putida* + 1 CO_2 + 1.64 H2O	46%
Conversion: Arabinose + 1.57 O_2 → 0.62 Muconic + 1.26 CO_2 + 3.13 H2O	54%
Conversion: Sucrose + 2.35 O2 + 0.56 NH_3 → 9.6 *P. putida* + 2.4 CO_2 + 2.96 H2O	46%
Conversion: Sucrose + 3.8731 O_2 → 1.48 Muconic + 3.13 CO_2 + 6.57 H2O	54%
Conversion: Acetate +.39 O2 + 0.093 NH_3 → 1.6 *P. putida* + 0.4 CO_2 + 0.66 H2O	100%
Conversion: Extractives + 0.68 O2 + 0.28 NH_3 → 4.8 *P. putida* + 1.2 CO_2 + 1.98 H2O	46%
Conversion: Extractives + 1.44 O_2 → 0.74 Muconic + 1.57 CO_2 + 3.78 H2O	54%
Conversion: Lignin + 3 O_2 → 1 Muconic + 2 CO_2 + 1 H_2O	100%

After the fermentation, the collected broth is sent through an ultrafilter to remove debris and cell mass. The remaining solids are sent to wastewater treatment and eventually burned in the high-solids boiler. The recovered liquid is carbon filtered to remove coloring compounds and then proceeds to the muconic acid recovery system. The muconic acid is acidified and recovered via low-temperature crystallization in the acid form [14]. The crystallization occurs at a temperature of 15°C and a pH of 2, recovering 98.8% of the product. The entrained liquid in the crystal is removed via a fluidized-bed drier, and then redissolved into an ethanol solvent. The ethanol:muconic acid ratio is set at 4, constrained to remain above the solubility limit of muconic acid in ethanol. Table 15 lists the key crystallizer metrics. Finally, muconic acid is hydrogenated to adipic acid in a three-stage packed-bed reactor operating at 40 atm. Hydrogen is fed in excess at a molar ratio of 2.6 mols H_2:mol muconic acid to ensure complete hydrogenation. The reactor is operated over a 2% ruthenium on carbon catalyst at a mild temperature of 78°C to avoid cracking of the facile double bond backbone, as well as over hydrogenation of the acid end groups critical to the final polymer properties. After hydrogenation, the liquid proceeds to a flash evaporator, which concentrates the adipic acid to a ratio of 2.5 ethanol:adipic by mass at elevated temperatures. The concentrated adipic acid product stream is crystallized by lowering the temperature to 15°C [15, 16]; 73.4% of the adipic acid is removed per pass as crystals via centrifugation, and the mother liquor with the remaining uncrystallized adipic acid and ethanol is recycled back to the evaporation cycle until extinction. The recovered crystal product is sent to a drier to remove entrained ethanol, then stored on-site. Table 15 also summarizes the key design parameters for the hydrogenation reactor and adipic acid crystallizer.

Table 15. Muconic Acid Crystallizer Metrics

Muconic acid crystallizer target pH	2
Muconic acid crystallizer temperature	15°C
Muconic acid crystallization recovery	98.8%
Hydrogenation ethanol:muconic mass ratio	4:1
Hydrogenation temperature	78°C
Hydrogenation pressure	40 atm
Hydrogenation H_2:muconic molar ratio	2.6
Hydrogenation conversion	100%
Hydrogenation catalyst	2% Rh/C
Hydrogenation WHSV	5 h^{-1}
Adipic acid crystallizer ethanol:adipic ratio	2.5
Adipic acid crystallizer temperature	15°C
Adipic acid crystallization recovery (per-pass)	73.4%

As documented in the 2018 design report, The BCD operation is costed based on a 127-m^3 pulping reactor vertical pressure vessel with a design similar to the pre-steaming section of the dilute acid pretreatment reactor [3]. All fermentation units are constructed of 304 or 316 stainless steel. The initial two seed fermentors are costed from prior seed tanks based on vendor quotations. The third seed unit and production unit are bubble column reactor units with a length-to-diameter ratio of 6, estimated through a combination of industry quotations and ACCE cost modeling. Fermentation compressors were sized based on the required air flow rate for meeting the oxygen uptake rate demands, with a pressure increase determined by the reactor dimensions and hydraulic pressure at the bottom of the vessel. Ultrafiltration and carbon filtration units were estimated based on guidance from an engineering consultancy with Nexant and designed as a counter-current diafiltration package unit including feed pumps and controls. The unit is sized for a base membrane area of 53,820 ft^2 and includes an operating cost for membrane replacement. The crystallizer, centrifuge, and crystal drier unit capital costs were scaled from estimates provided by Nexant. The adipic acid hydrogenation reactor was quoted as a low-pressure fixed-bed hydrodeoxygenation unit based on a prior engineering subcontract, adjusted for lower operating pressure relative to the original quotation. The adipic acid concentrator was designed as an initial mixing/feed tank, followed by a heat exchanger and flash evaporation tank constructed of SS316 for additional corrosion resistance. Adipic acid crystallizers, centrifuges, and dryers assumed similar designs as the muconic acid unit but scaled to the proper product stream.

3.9 Area 800: Combustor, Boiler, and Turbogenerator

Again, all assumptions pertaining to Area 800 remain consistent with the 2018 design report. In brief, the purpose of the combined heat and power subsystem is to burn residual byproduct streams to produce steam and electricity. Combustible byproducts include unconverted lignin and carbohydrates from the feedstock, cell mass from fermentation and WWT, and off-gas streams from catalytic upgrading operations. Combustion of these byproduct streams generates steam to drive the upgrading and separation operations, and partially offsets the plant's electric power demand. The fuel streams are fed to a bubbling fluidized-bed combustor boiler capable of handling the wet solids. A fan moves air into the combustion chamber. Treated water enters the heat exchanger circuit in the combustor and is boiled and superheated to high-pressure steam at 900 psig. A multistage turbine and generator are used to generate electricity. Steam is extracted from the turbine at two different conditions for use in the process. In the final stage of the turbine, the remaining steam is taken down to a vacuum and condensed with cooling water for maximum energy conversion. The condensate is returned to the boiler feed water system along with condensate from the various process heat exchangers. The steam turbine turns a generator that produces power for all use in the plant. The balance of required power is purchased from the grid (see Area 900). NO_x emissions are mitigated with ammonia injection in a selective non-catalytic reduction system, and SO_x emissions are mitigated with flue-gas desulfurization.

Whereas the 2018 design case required some natural gas (to provide supplemental heating demands in the BDO pathway and to drive a hot oil system for high-temperature utility heating in the acids pathway), the present case requires substantially more natural gas co-fired in the boiler to satisfy increased heat demands incurred in Areas 500 and 550. These are driven by large temperature swings between the key unit operations in Area 500 while processing large volume throughputs inclusive of water and dioxane solvent (as well as subsequent solvent distillation recovery). Likewise, high heat demands are incurred in Area 550 for vaporizing the aqueous BDO stream containing roughly 90% water, for subsequent catalytic conversion to MEK in the **integrated** biorefinery scenario.

The cost basis for the Area 800 equipment remains the same as described in the prior NREL reports, reflecting cost estimates furnished from vendor quotations for both the boiler and turbine system, as well as for most other minor equipment. The boiler capital cost includes the boiler feed water preheater, flue gas desulfurization spray dryer, and baghouse for collection of ash and particulates from the flue gas. For the baghouse, bag replacement appears as a periodic charge in the cash flow worksheet.

3.10 Area 900: Utilities

Area 900 tracks all plant utilities except steam, which is provided by Area 800, including electric power, cooling water, chilled water, plant and instrument air, process water, and the clean-in-place system. The process water manifold in Area 900 mixes fresh water with treated wastewater and condensate from the sugar evaporation system (assumed suitable for all plant users) and provides this water at a constant pressure to the facility. The clean-in-place system provides hot cleaning and sterilization chemicals to hydrolysis, bioconversion, and the enzyme production section. Consistent with prior designs, the cooling water system is designed for a 28°C supply temperature with a 9°C temperature rise in coolers throughout the facility. This is an assumed average rise; the actual cooling water rises across each exchanger are not explicitly modeled in Aspen. The cooling water demands are summarized in Figure 10.

Similar to the trends in the 2018 design report, heat balances are considerably different for this pathway relative to prior NREL biochemical models, with the present cases resulting in a net heat deficit requiring supplemental natural gas. Accordingly, after extracting steam from intermediate turbine stages to satisfy process steam/heat demands, a minimal amount of steam remains passing through the final turbine stage and subsequent condenser (which historically had been the largest cooling demand and thus source of cooling tower losses/makeup water requirements in prior NREL models). In the present models, the core conversion operations in Areas 500 and 550 constitute the majority of the facility cooling requirements, representing over 50% of total cooling water demands in either biorefinery scenario. In the **dedicated** scenario, dioxane recovery constitutes the single largest cooling requirement (28%), followed by other Area 500 operations (26%), whereas in the **integrated** scenario, coolers used in the added MEK production train represent the largest cooling duty (24%), followed again by dioxane recovery (17%) and other Area 500 coolers (15%). The chiller condenser also represents a significant fraction of cooling demands, at 24% in both scenarios. The compressor electricity demand for the chiller was estimated at 0.56 kW/ton of refrigeration and the cooling water demand for the chiller system was assumed to be equal to the heat removed in the chilled-water loop.

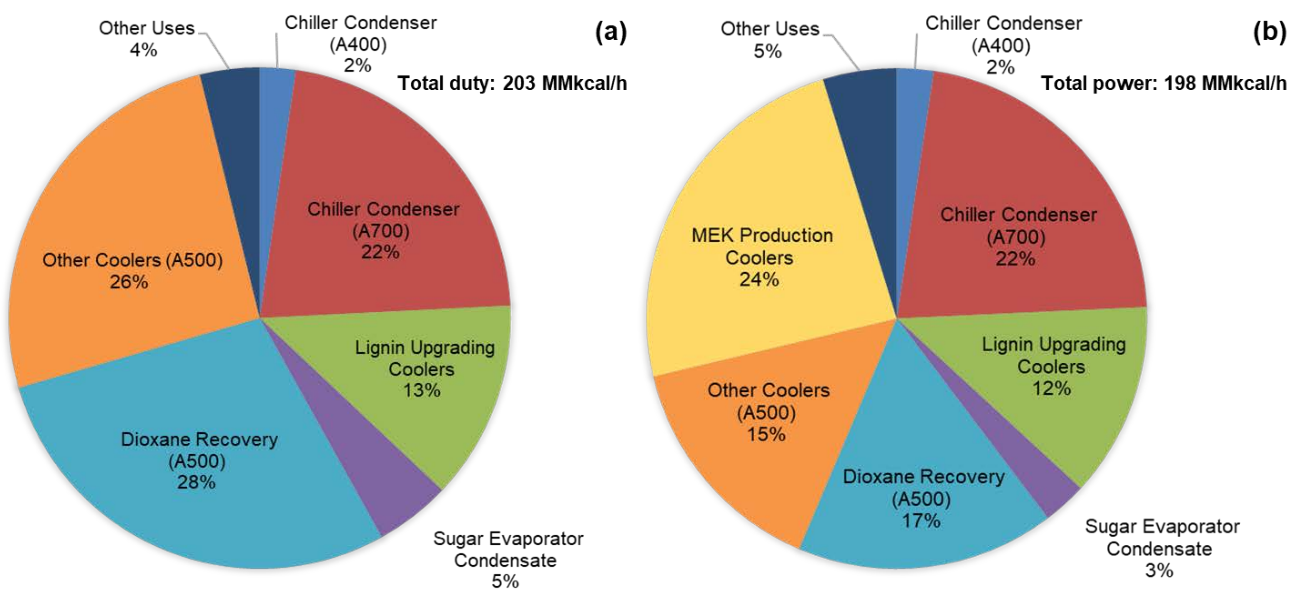

Figure 10. Cooling water heat duty distribution between major users in the (a) dedicated case and (b) integrated case

The electricity generated in Area 800 is used to partially offset the plant power demand throughout the facility to power pumps, agitators, compressors, etc. (68.7 MW total power required for the **dedicated** biorefinery scenario and 61.2 MW for the **integrated** scenario), but there is still a considerable power deficit on the order of roughly 35–52 MW that must be imported from the grid after considering the power generated on-site. The distribution of total plant power utilization among all areas is shown in Figure 11. Note that the cost of the power required by Area 100 is already assumed to be included in the feedstock cost and is subtracted from the plant's net electricity import. This is reflected in the economics by an operating cost credit equal to this amount of electricity.

Consistent with the 2018 design case pathways, for both biorefinery scenarios in the present analysis, Area 200 constitutes the single largest power demand, primarily due to the switch to DMR pretreatment in this design, with high power demands to drive the mechanical refining equipment. Area 700 also reflects a sizeable share of the facility power demands associated with compressors for aerobic fermentation and downstream coproduct recovery operations. The new Area 500 operations do not exhibit disproportionately high power requirements, as they are driven primarily by heat demands for elevated temperature catalysis operations.

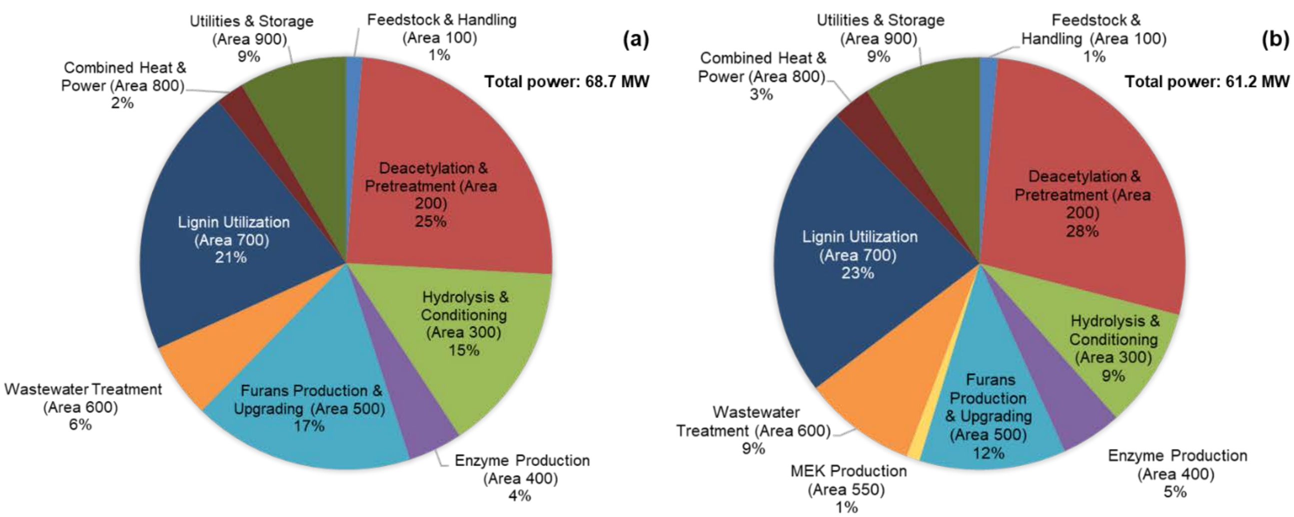

Figure 11. Distribution of plant electricity utilization by process area for the (a) dedicated case and (b) integrated case

All cost estimates for the utility equipment in Area 900 were maintained consistent with the basis values used in prior NREL reports. To summarize, the cooling tower was based on a cost estimate from a vendor for a fiberglass cooling tower capable of handling 44,000 gal/min; this cost is scaled to the respective cooling water throughputs estimated here. Most costs for the remaining operations in this section were furnished from an engineering subcontractor.

4 Process Economics

This section summarizes the key inputs and results for the modeled biorefinery scenarios, with more details available in previous reports [2]. The TCI is first computed from the total equipment cost. Next, variable and fixed operating costs are determined. With these costs, we use a discounted cash flow analysis to determine the minimum fuel selling price required to obtain a zero net present value with a fixed internal rate of return. The analysis does not consider any policy factors (e.g., subsidies and carbon credits) or early pioneer plant economics, but instead focuses strictly on the economic implications of the technical parameters modeled here as may factor into n^{th}-plant technology maturity levels.

The cost-year of 2016 was maintained for this analysis as consistent with the 2018 design report. As presented in prior design cases and other reports [2, 17], capital costs were adjusted using the Plant Cost Index from *Chemical Engineering Magazine* [18] 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 [19]. 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 [19]. The general formula for cost-year dollar back-casting is:

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

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. In keeping with standard NREL TEA methodology, a factored approach in which multipliers are applied to the purchased equipment cost is maintained for estimating installed costs. 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. Instead of re-costing in detail, a standard exponential scaling expression was used:

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

In this equation, the scaling exponent n varies depending on the type of equipment to reflect economy-of-scale dependencies. The basis for scaling is typically some characteristic of the equipment related to production capacity, such as flow or heat duty. Some equipment does not follow such a scaling-factor approach, namely when the capacity for a given operation is exceeded and requires multiple units in parallel, thus losing economy-of-scale benefits that are captured in this exponential expression.

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 (detailed in Table 16). Site development and warehouse costs, along with additional piping, are based on the inside-battery-limits (ISBL) equipment costs (Areas 200, 300, 400, 500, 550, and 700 in this design) and are considered part of the total direct cost (TDC). Beyond the ISBL operations, the other process areas are considered outside battery limits, including Areas 100 (rolled up into feedstock costs), 600, 800, and 900. Project contingency, field expenses, home office engineering and construction activities, and other costs related to construction are computed relative to the TDC and give the fixed capital investment (FCI) when summed. The categories and values for these additional cost escalation factors were maintained consistently with prior NREL design reports, and the reader is referred there for further details [2, 3].

Table 16. Project Cost Worksheet Including TDC and TCI (2016$)

Process Area	Dedicated Case Purchased Cost	Dedicated Case Installed Cost	Integrated Case Purchased Cost	Integrated Case Installed Cost
Area 100: Feedstock Storage and Handling [a]	Included in feedstock cost		Included in feedstock cost	
Area 200: Pretreatment	$35,800,000	$48,700,000	$35,800,000	$48,700,000
Area 300: Enzymatic Hydrolysis & Hydrolysate Conditioning	$34,000,000	$59,800,000	$24,000,000	$42,100,000
Area 400: Enzyme Production	$6,700,000	$11,500,000	$6,700,000	$11,500,000
Area 500: Furans Production and Upgrading	$26,700,000	$43,000,000	$18,600,000	$29,600,000
Area 550: MEK Production (Integrated case)	-	-	$12,700,000	$25,100,000
Area 600: Wastewater Treatment [b]	$16,500,000	$31,300,000	$18,800,000	$35,700,000
Area 700: Lignin Upgrading	$66,100,000	$140,000,000	$63,700,000	$134,700,000
Area 800: Combustor, Boiler, and Turbogenerator	$33,200,000	$60,100,000	$38,900,000	$70,400,000
Area 900: Utilities	$11,500,000	$20,000,000	$10,400,000	$18,000,000
Totals (Excl. Area 100)	**$230,600,000**	**$414,300,000**	**$229,600,000**	**$415,800,000**
Warehouse	4.0% of ISBL	$12,100,000		$11,700,000
Site development	9.0% of ISBL	$27,300,000		$26,200,000
Additional piping	4.5% of ISBL	$13,600,000		$13,100,000
Total Direct Costs		**$467,300,000**		**$466,900,000**
Proratable expenses	10.0% of TDC	$46,700,000		$46,700,000
Field expenses	10.0% of TDC	$46,700,000		$46,700,000
Home office and construction fee	20.0% of TDC	$93,500,000		$93,400,000
Project contingency	10.0% of TDC	$46,700,000		$46,700,000
Other costs (startup, permits, etc.)	10.0% of TDC	$46,700,000		$46,700,000
Total Indirect Costs		**$280,400,000**		**$280,100,000**
Fixed Capital Investment		**$747,600,000**		**$747,000,000**
Land		$1,800,000		$1,800,000
Working capital	5.0% of FCI	$37,400,000		$37,400,000
Total Capital Investment		**$786,900,000**		**$786,200,000**
Lang factor (TCI/purchased equip. cost)		3.6		3.7
TCI per annual gallon gasoline equivalent		$10.02/GGE		$17.75/GGE

[a] Feedstock handling not included in this calculation.
[b] Area 600 not included in Lang factor.

4.2 Variable Operating Costs

Variable operating costs, which include raw materials, waste handling charges, and byproduct credits, are incurred only when the process is operating. Quantities of raw materials used and wastes produced were determined using the Aspen material balance. Table 17 documents the costs and sources of chemicals used in the process and Table 18 summarizes the variable costs on a per-year and per-GGE basis. All costs for materials used in NREL's 2018 design report were maintained consistently here, including those that had been updated reflecting new cost information available at that time—i.e., ammonia, sulfuric acid, glucose (concentrated glucose syrup), diammonium phosphate, and sodium hydroxide, as well as sodium sulfate salt sold as an additional coproduct [1]. As discussed in that report, the sodium sulfate salt must be sold to offset elevated expenses incurred for substantial use of sodium hydroxide throughout the facility (representing an MFSP contribution of roughly $0.50–$0.90/GGE in the present biorefinery scenarios). Both pathways require a net power import after considering the amount of power generated through the combined heat and power system, costed consistently with prior design cases with grid imports at 6.8 ¢/kWh. For this assessment, natural gas costs were reduced from $5 to $3.50/MM BTU, reflecting an average of natural gas prices over more recent years [20]. Adipic acid is coproduced in the lignin train at a product purity over 99.7 wt %, with a sale price set at $1,710/short ton ($0.86/lb) in 2016$, reflective of a 15-year average price for this product. Further discussion on historical price fluctuations and rationale for the selection of this value is provided in the 2018 design report. Sensitivity on overall biorefinery MFSP to the adipic acid coproduct value is considered in Sections 5.2.1 and 5.2.2.

Costs for additional/new chemicals as required for this pathway are also reflected in Table 17. Key among them is the assumed cost for the MEK co-reactant applied in the **dedicated** biorefinery scenario when purchasing this component externally. Chemical-grade MEK derived from conventional chemistries may be purchased at a cost of roughly $0.77/lb (based on an average of recent market prices); however, such a cost would cause steep penalties in the resultant MFSP (reflected in the sensitivity analysis described later) and would also likely incur substantial penalties in the life cycle assessment for this pathway, given that conventional MEK is synthesized from petrochemical routes via butene. Accordingly, to achieve viability for this pathway, MEK or another ketone purchased externally should be bio-derived (to improve the life cycle assessment profile) and need not be available at chemical-grade purities (to reduce costs). In the present analysis, the **dedicated** biorefinery scenario assumes a “transfer price” of $0.30/lb MEK as an approximate calculation for the minimum selling price of bio-MEK produced from a separate stand-alone 2,000-tonne/day biorefinery configured for exclusive production of this component—i.e., for a dedicated facility mirroring Area 550 of this report inclusive of a similar lignin coproduct train, based on discussions with PNNL collaborators coordinating the research work for the BDO-to-MEK catalysis pathway [13]. Accordingly, the primary difference between the **dedicated** and **integrated** biorefinery scenarios is ultimately the economy of scale for a single 2,000-tonne/day biorefinery simultaneously producing both products or two separate biorefineries of that scale, each producing their respective intermediates (furans and MEK); although in the latter case, more MEK would be produced than needed for reaction with furans and could proceed on (e.g., to final fuel upgrading through subsequent catalysis steps).

Table 17. Chemical Costs and Sources

Component	Cost (2016$)	Source
Biomass delivered to reactor throat	$0.0285/lb	Idaho National Laboratory inputs, $71.26/dry ton @ 20% moisture
Sulfuric acid, 93%	$0.0430/lb	Industry database, 5-year average
Ammonia	$0.1900/lb	Industry database, 5-year average
Sodium hydroxide	$0.2384/lb	Nexant (indexed from 2011$ basis)
Ultrafilter replacement	0.0297 $/$ cost	Nexant (annual cost per $ membrane CAPEX)
Corn steep liquor	$0.0339/lb	Corn products via Harris Group
Diammonium phosphate	$0.1645/lb	Industry database, 5-year average
Corn oil (antifoam)	$0.6439/lb	Industry database
Glucose	$0.3670/lb	U.S. Department of Agriculture Economic Research Service, 5-year average [21]
SO_2	$0.1811/lb	Industry database
Enzyme nutrients	$0.4896/lb	Industry database (see 2011 design report for details)
Hydrogen	$0.7306/lb	U.S. Department of Energy report, steam methane reforming H_2 @ $4/MM BTU natural gas [22]
BDO upgrading catalyst [a]	$32.34/lb	NREL internal database
Dioxane	$0.764/lb	Industry database, 5-year average
Aluminum chloride	$0.1682/lb	Industry database, 5-year average
MEK	$0.30/lb	Estimated transfer price for biobased MEK
Hydrotreating catalyst	$105/lb	NREL internal database
Polymer for WWT	$2.6282/lb	Brown and Caldwell 2012 WWT design [23]
Ethanol	$0.3370/lb	Prior NREL analysis
Natural gas	$3.5/MM BTU	Average 2016 values retrieved from the U.S. Energy Information Administration [20]
Lime	$0.1189/lb	Harris Group
Boiler chemicals	$2.9772/lb	2002 design report [7]
Cooling tower chemicals	$1.7842/lb	2002 design report [7]
Fresh water	$0.0002/lb	Peters and Timmerhaus [24]
Sodium sulfate salt coproduct value	$0.0706/lb	Nexant (indexed from 2011$ basis)
Adipic acid coproduct value	$0.8554/lb	Average price over a 15-year cycle

[a] Price of catalyst assumed to be that of an H-ZSM-5 zeolite

Table 18. Variable Operating Costs

Process Area	Stream Description	Dedic. Usage (kg/h)[a]	Integr. Usage (kg/h)[a]	Dedic. MM$/yr (2016$)	Integr. MM$/yr (2016$)	Dedic. ¢/GGE (2016$)	Integr. ¢/GGE (2016$)
Raw Materials							
N/A	Feedstock	104,167	104,167	51.62	51.62	65.72	116.51
A200	Sulfuric acid, 93%	0	0	0	0	0	0
	Caustic (as pure)	5,833	5,833	24.17	24.17	30.78	54.57
	Ammonia	0	0	0	0	0	0
A300	Flocculant	345	195	6.35	3.58	8.09	8.09
A400	Glucose	1,324	1,324	8.45	8.45	10.75	19.06
	Corn steep liquor	90	90	0.05	0.05	0.07	0.12
	Corn oil	7	7	0.08	0.08	0.10	0.18
	Ammonia	63	63	0.21	0.21	0.26	0.47
	Host nutrients	37	37	0.31	0.31	0.40	0.71
	Sulfur dioxide	9	9	0.03	0.03	0.04	0.06
A500	Dioxane	209	120	2.77	1.59	3.53	3.58
	Aluminum chloride	949	535	2.77	1.56	3.53	3.53
	Caustic (as pure)	1,220	688	5.05	2.85	6.44	6.44
	Methyl ethyl ketone	9,301	-	48.51	-	61.76	-
	Hydrogen	3,522	1,799	44.73	22.84	56.95	51.56
	Hydrotreating catalyst [b]	1.2	0.7	0.36	0.20	0.46	0.46
A550	Corn steep liquor	-	265	-	0.16	-	0.35
	Diammonium phosphate	-	32	-	0.09	-	0.21
	BDO upgrading catalyst [b]	-	1.3	-	0.60	-	1.57
A600	Ammonia	0	55	0	0.18	0	0.41
	Polymer	0.4	0.9	0.02	0.04	0.02	0.10
A700	Caustic (as pure)	3,218	2,972	13.34	12.32	16.98	27.80
	Ammonia	95	91	0.31	0.30	0.40	0.68
	Diammonium phosphate	616	589	1.76	1.69	2.24	3.80
	Corn steep liquor	144	126	0.08	0.07	0.11	0.17
	Sulfuric acid, 93%	11,560	11,241	8.64	8.40	11.00	18.97
	Ultrafilter replacement	Cost	Cost	0.14	0.14	0.17	0.31
	Ethanol	39	38	0.23	0.22	0.29	0.51
	Hydrogen	426	417	5.41	5.30	6.88	11.95
	Hydrotreating catalyst [b]	0.7	0.7	0.84	0.82	1.07	1.85
A800	Boiler chemicals	0.2	0.2	0.01	0.01	0.02	0.03
	Flue gas desulfurization lime (SO_x control)	87	94	0.18	0.19	0.23	0.44
	Ammonia (NO_x control)	836	950	2.76	3.14	3.51	7.08
	Natural gas	6,500	9,700	9.45	14.10	12.03	31.82
A900	Cooling tower chemicals	5	5	0.14	0.14	0.18	0.32
	Makeup water	281,059	284,836	0.75	0.76	0.96	1.72
Power	Grid electricity (kW)	51,271	34,829	17.62	18.73	22.43	42.27
	Subtotal			**257.14**	**184.94**	**327.40**	**417.70**
Waste Disposal							
A800	Disposal of ash	4,252	4,265	1.40	1.41	1.78	3.17
	Subtotal			**1.40**	**1.41**	**1.78**	**3.17**
Coproducts and Credits							
A600	Sodium sulfate (98.5 wt %)	13,770	13,946	16.91	17.12	21.53	38.65
A700	Adipic acid (99.7 wt % pure)	11,812	11,494	175.65	170.93	223.65	385.80
	Subtotal			**192.56**	**188.05**	**245.18**	**424.45**
Total Variable Operating Costs				**65.98**	**−1.70**	**84.00**	**−3.58**

[a] For reference, to convert to kg/GGE basis, fuel outputs are 9,962 and 5,619 GGE/h for dedicated and integrated cases, respectively.

[b] Catalyst usage amortized to kg/h basis for consistency with rest of table.

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 fixed operating costs were maintained consistently with the 2018 design report, which in turn were based in large part on NREL's 2002 ethanol design report [7] and/or Peters and Timmerhaus [24]. Table 19 shows the recommended number of employees and associated salaries. The number of employees was estimated by considering the likely degree of automation for each area and adding a reasonable number of management and support employees. Because the model feedstock is predominately corn stover, salaries were estimated for rural regions of the U.S. Midwest. These estimates may vary depending on location. A 90% labor burden is applied to the salary total and covers items such as safety, general engineering, general plant maintenance, payroll overhead (including benefits), plant security, janitorial and similar services, phone, light, heat, and plant communications. The 90% estimate is the median of the general overhead range suggested in the 2008 Process Economics Program Yearbook produced by SRI Consulting (now IHS) [25]. Table 20 shows the full fixed operating costs associated with both biorefineries. Annual maintenance materials were estimated as 3% of the installed ISBL capital cost and property insurance, and local property tax was estimated as 0.7% of the fixed capital investment, based on the 1994 Chem Systems report described in NREL's 2011 ethanol report [2]. These factors are all consistent with those used in prior design reports.

Table 19. Positions and Salaries ($/yr) for Employees

Position	2016 Salary	# Required	2016 Cost
Plant manager	$164,452	1	$164,452
Plant engineer	$78,310	4	$313,241
Maintenance supervisor	$63,767	1	$63,767
Maintenance technician	$44,749	12	$536,985
Lab manager	$62,648	1	$62,648
Lab tech	$44,749	2	$89,498
Lab tech – enzyme	$44,749	2	$89,498
Shift supervisor	$53,699	4	$214,794
Shift operators	$44,749	24	$1,073,970
Shift operators – enzyme	$44,749	8	$357,990
Yard employees	$31,324	4	$125,297
Clerks and secretaries	$40,274	3	$120,822
Total salaries			$3,212,962
Labor burden (90%)			$2,891,655

Table 20. Fixed Operating Costs

Labor and Supervision	**Dedicated MM$/yr**	**Integrated MM$/yr**	**Dedicated ¢/GGE**	**Integrated ¢/GGE**
Total salaries	3.21	3.21	4.09	7.25
Labor burden (90%)	2.89	2.89	3.68	6.53
Other Overhead	**Dedicated MM$/yr**	**Integrated MM$/yr**	**Dedicated ¢/GGE**	**Integrated ¢/GGE**
Maintenance (3.0% of ISBL)	9.09	8.75	11.57	19.75
Property insurance (0.7% of FCI)	5.23	5.23	6.66	11.80
Total Fixed Operating Costs	14.32	20.08	26.1	45.33

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 standard NREL TEA practices, the discount rate (which is also the internal rate of return in this analysis) was maintained at 10% and the plant lifetime at 30 years. The 10% rate is consistent with all platforms across the Bioenergy Technologies Office portfolio, and more context on its basis is discussed in prior reports [1, 2]. Also consistent with other recent TEA reports, it was assumed that the plant would be 40% equity financed. The terms of the loan were established at 8% interest for 10 years. The principal is taken out in stages over the 3-year construction period. Interest on the loan is paid during this period, but principal is not paid back (this is another n^{th}-plant assumption, which says that this cash flow comes from the parent company until the plant starts up).

Again, the Internal Revenue Service Modified Accelerated Cost Recovery System (MACRS) basis for depreciation schedules is maintained in the present design, which uses a 7-year recovery period for the majority of the plant except for the steam plant equipment (Area 800), which uses a 20-year recovery period. The updated corporate tax rate of 21% is also maintained, without further consideration for state taxes. Likewise, the current analysis maintains the assumption of 12 months for planning and engineering, followed by 24 months for facility construction, with a startup time of 6 months and working capital of 5% relative to FCI (all reflecting n^{th}-plant assumptions).

4.4.2 Base Case TEA Results

Based on the TEA parameters summarized here, the resulting MFSP of total fuel products is **$2.54/GGE for the dedicated case** and **$2.72/GGE for the integrated case (2016$)**, representative of the hydrocarbon fuel product adjusted by heating values (calculated in the Aspen model) to gasoline equivalents. Such MFSP results are analogous to those reported in NREL's aforementioned design case [1] focused on biological conversion of sugars to fermentation intermediates with subsequent catalytic upgrading of those intermediates to hydrocarbon fuels, at $2.47–$2.49/GGE. This indicates that such a route is another viable alternative pathway to achieve similar fuel cost targets through purely catalytic upgrading of sugars in this case.

Table 21 summarizes the yields and conversion costs for the present designs. According to Cran's methodology [6], the expected accuracy of the overall TCI analysis is ±25% (although some specific pieces of equipment carry a higher degree of uncertainty in underlying cost estimates, as previously identified). If we apply this uncertainty to the TCI, the impact on the cost of total fuel is ±$0.33/GGE and ±$0.58/GGE for the **dedicated** and **integrated** cases, respectively. 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 and Figure 13 illustrate the contribution to the overall cost by process area and capital, operations, and fixed costs (the bar for feedstock plus handling reflects the single feedstock cost of $71.26/dry U.S. tons delivered to pretreatment and has not been broken down).

Table 21. Summary of Yields, Rates, and Conversion Costs for Both Biorefinery Configurations

	Dedicated Biorefinery	Integrated Biorefinery
Feedstock rate	2,205 dry U.S. tons/day	
Online time	7,884 h/yr (90% online factor)	
Total fuel yield	108.4 GGE/dry U.S. ton feedstock	61.2 GGE/dry U.S. ton feedstock
Total fuel production rate	78.5 MM GGE/yr	44.3 MM GGE/yr
Adipic acid coproduct yield	284 lb/dry U.S. ton feedstock	276 lb/dry U.S. ton feedstock
Adipic acid production rate	205 MM lb/yr	200 MM lb/yr
Total variable OPEX excluding coproducts	$269 MM/yr	$187 MM/yr
Coproduct revenue	$193 MM/yr	$188 MM/yr
Total fixed OPEX	$20 MM/yr	$20 MM/yr
Total equipment cost	$414 MM	$416 MM
Total capital investment	$787 MM	$786 MM
TCI per annual gallon	$10.02/GGE	$17.75/GGE
Minimum Fuel Selling Price	**$2.54/GGE**	**$2.72/GGE**
Feedstock contribution	$0.66/GGE	$1.17/GGE
Fuel conversion contribution	$3.11/GGE	$3.71/GGE
Coproduct conversion contribution	−$1.23/GGE	−$2.16/GGE

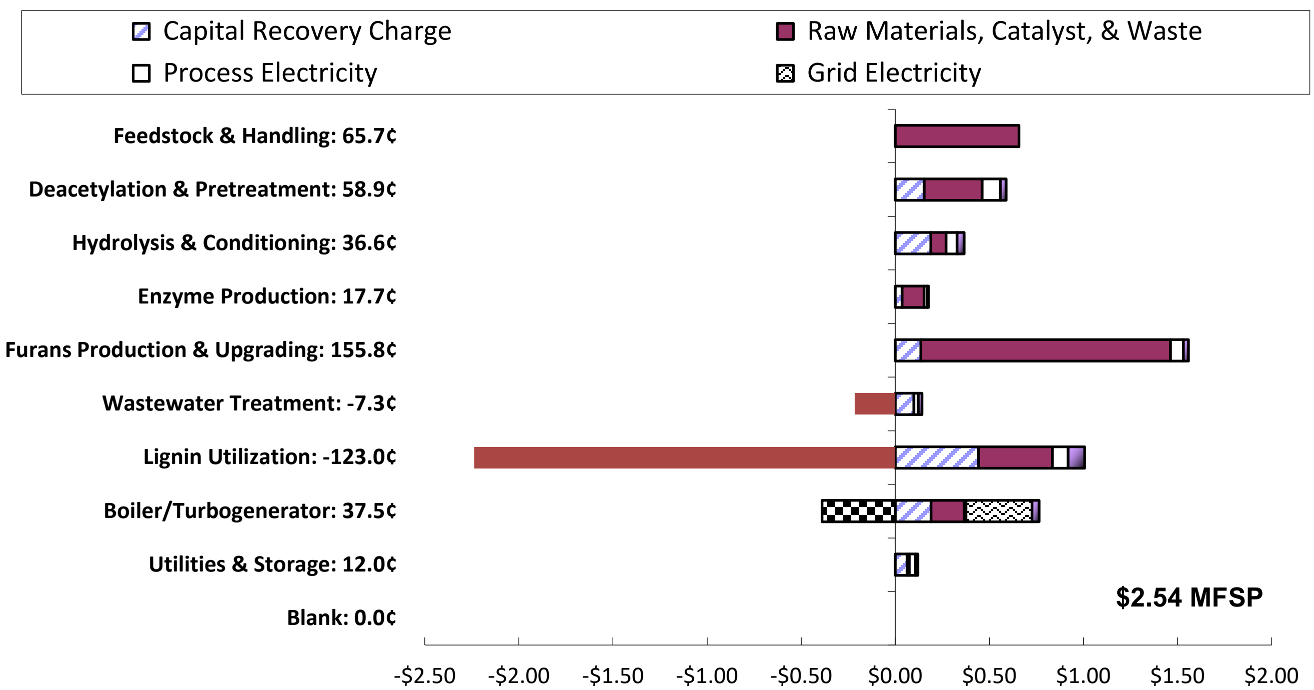

Figure 12. Dedicated case cost contribution details from each process area (per GGE total fuel)

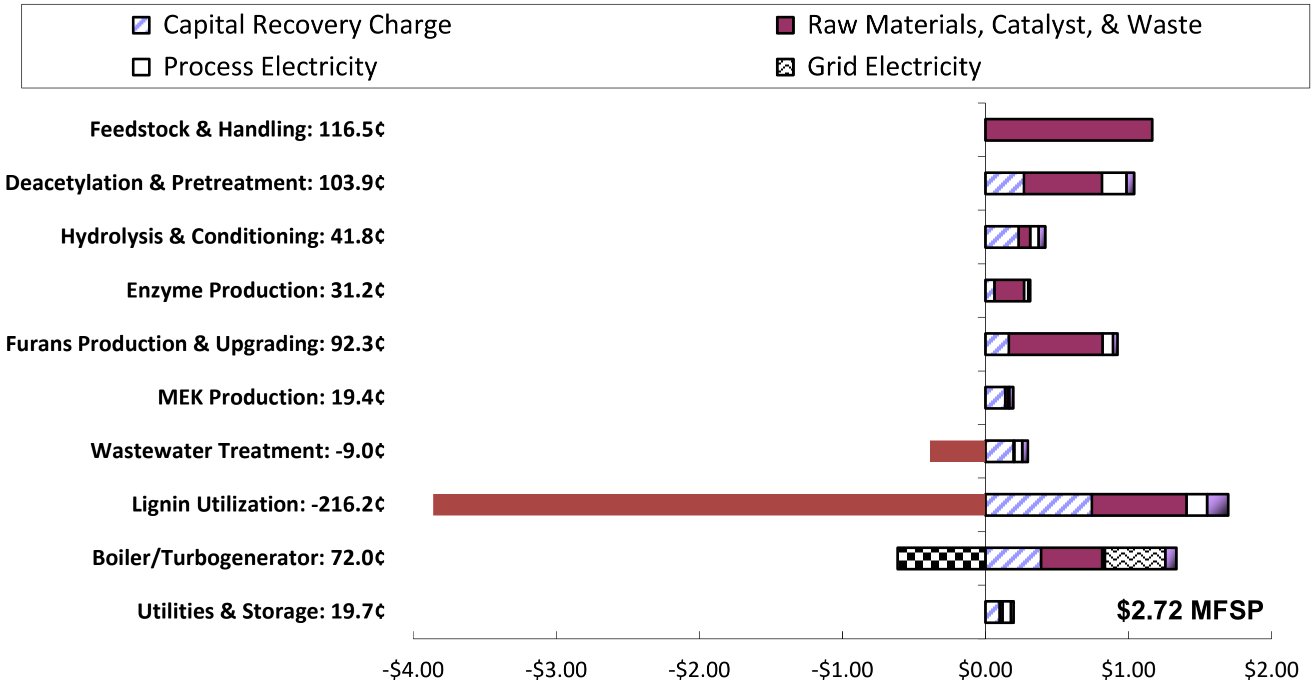

Figure 13. Integrated case cost contribution details from each process area (per GGE total fuel)

As shown in Table 21, Figure 12, and Figure 13, the MFSP estimates are seen to vary by approximately $0.18/GGE between the two biorefinery configurations. The **dedicated** biorefinery case assumes a purchase cost of $0.30/lb MEK, representing an estimated "transfer price" from a separate facility that is otherwise producing bio-MEK through the same processing steps as those modeled here but with the sugar conversion train focused exclusively on BDO-to-MEK production (mirroring Area 550 in the **integrated** case). Given this, this difference in MFSP is ultimately a reflection of economies of scale for a single 2,000-tonne/day biorefinery coproducing both the furan and MEK co-reactants simultaneously versus two separate biorefineries of this scale each producing the respective intermediates. As such, although the **dedicated** biorefinery views the MEK as an externally sourced chemical to support its own operations, unless that MEK or an alternate ketone could be sourced from a non-fossil feedstock that was also not a terrestrial biomass crop (e.g., a waste feedstock that may be upgraded to such a ketone), the **integrated** case ultimately represents the fuel yields that can be achieved per ton of starting biomass (corn stover) and other related considerations for the overall biorefinery supply chain.

Accordingly, the present **integrated** pathway demonstrates potential for substantially higher fuel yields than the pathways presented in the 2018 design report: roughly 61 GGE/ton versus 43–45 GGE/ton (36%–42% increase) [1]. Such a high fuel yield is not commonly encountered in a biochemical processing approach when focused only on the carbohydrate fraction of the biomass feedstock and is largely achieved owing to high carbon retention efficiencies across the furan catalytic upgrading steps by rejecting oxygen as water rather than CO_2. The **dedicated** biorefinery case could achieve even higher fuel yields of 108 GGE/ton, but as noted, this may be somewhat artificial as the ketone likely must be derived from a low-carbon-intensity source, and if this were through a biomass conversion process the *overall system* yield would still be near the 61-GGE/ton basis. This is a similar finding as a previously investigated design report focused on catalytic upgrading of sugars to hydrocarbon fuels via aqueous-phase reforming technology, which also maintained high carbon efficiencies through low CO_2 rejection and had found the potential for fuel yields as high as 78 GGE/ton when sourcing hydrogen externally, reducing to 45 GGE/ton with *in situ* hydrogen sourcing through a parallel carbohydrate upgrading train [4]. The main data for key process streams are presented in Appendix C.

Although these results demonstrate the potential for exceptionally high fuel yields through this pathway, they also carry higher processing costs than the pathways documented in the 2018 design report. Namely, capital expenditures based on TCI are between 4% and 13% higher than the 2018 design case pathways ($786–$787 MM for the two scenarios here versus $697–$758 MM for the 2018 cases), although this still supports lower MFSPs as it translates to a TCI per annual gallon of $10/GGE (**dedicated** case) and $18/GGE (**integrated** case) versus the 2018 cases at $22–$23/GGE. However, net operating expenses are also higher, at $96 MM/yr (**dedicated** case) and $19 MM/yr (**integrated** case) versus *negative* $15–$18 MM/yr inclusive of variable and fixed operating costs and coproduct credits. This is driven primarily by increased costs for hydrogen and natural gas required for the catalytic upgrading operations, together adding roughly $0.75/GGE and $0.95/GGE between both biorefinery scenarios, respectively.

As is typical for TEA models, feedstock constitutes the largest single MFSP contribution at roughly $0.66/GGE and $1.17/GGE for the dedicated and integrated scenarios respectively. In the dedicated case, externally purchased MEK also incurs large costs at $0.62/GGE. Similar to the 2018 design report, the ability to reduce MFSPs down to $3/GGE or lower is strongly contingent on valorizing the lignin fraction of the biomass for conversion to value-added coproducts, reducing net MFSPs by $1.24/GGE and $2.17/GGE in the **dedicated** and **integrated** cases, respectively. In either case, annual coproduct revenues are comparable to each other as well as (slightly more than) the 2018 design case pathways but appear lower on a per-GGE basis given the higher fuel yields in the present models, particularly for the **dedicated** biorefinery case. Ultimately, this means that the higher fuel yields translate favorably to a lower reliance on lignin-derived coproducts to achieve MFSP goals below $3/GGE, albeit trading off other challenges with higher facility operating expenses.

5 Analysis and Discussion

5.1 Carbon Balance

Table 22 shows the overall flow of carbon inputs and outputs, with a carbon balance closure very near unity (difference of approximately 0.2%). The biorefining cases assessed in this report differ significantly in terms of carbon input and utilization. It is also noteworthy to highlight that the **dedicated** and the **integrated** biorefineries process, respectively, 27% and 17% more carbon in comparison to the pathways presented in the 2018 design report. Apart from biomass, which makes up around 76% and 86% of the total processed carbon in the **dedicated** and **integrated** cases, respectively, both plants import a considerable amount of natural gas to be combusted in the boiler to supply process heat requirements. Additionally, the **dedicated** biorefinery imports around 13% of the total carbon in the form of methyl ethyl ketone for the aldol condensation reaction in Area 500. Other inputs (glucose and other chemicals) are minor contributors to the carbon balance. In terms of outputs, fermentor vents and other process off-gases are sent to the boiler in Area 800 and are accounted as flue gas. In the **dedicated** case, nearly 60% of the total carbon input leaves as the hydrocarbon fuel and the adipic acid coproduct. This high carbon conversion efficiency is due to the high yields of the reactions considered in Area 500 and to the effective merging of carbon in methyl ethyl ketone (external input) into the hydrocarbon fuel product. In the **integrated** case, the combined products account for around

42% of the total carbon input, which is a result from relying solely on the biomass feedstock to synthesize both furans and ketone to generate the hydrocarbon fuel. Other large carbon outlets in the biorefineries are the combustor stack and the aerobic digestion lagoons.

Table 22. Overall Carbon Balance for the Biorefineries

	Dedicated Case		Integrated Case	
Stream	Carbon Flow (kmol/h)	% of Carbon Flow	Carbon Flow (kmol/h)	% of Carbon Flow
Carbon inlets				
Biomass feedstock	3,087	76%	3,087	82%
Natural gas	406	10%	605	16%
Methyl ethyl ketone	516	13%	-	-
Glucose	44	1%	44	1%
Other chemical inputs	15	<1%	13	<1%
Total	4,068	100%	3,749	100%
Carbon outlets				
Area 500 fuel output	1,961	48%	1,106	29%
Area 700 coproduct	487	12%	474	13%
Combustor flue gas	1,571	39%	2,001	53%
Aerobic lagoons	55	1%	175	5%
Total	4,074	100%	3,756	100%

5.2 Cost Sensitivity Analysis

For each of the biorefining strategies discussed previously in the report, the techno-economic models were used to carry out sensitivity analyses on key model variables. Starting with the baseline for each variable as described in this report, minimum and maximum values were chosen to assess their impact on the MFSP of the whole process one parameter at a time, with all other variables held constant. The goal of the analyses presented herein is to focus on parameters linked to Areas 500/550 and on other variables that could have an impact on either plant when considering the full biorefining context. Some of the most significant drivers of MFSP in the 2018 design report were chosen to be included in this assessment (e.g., NaOH loading in DMR, enzyme loading in enzymatic hydrolysis, and factors related to lignin utilization in Area 700).

5.2.1 Single-Point Sensitivity Analysis: Dedicated Case

Table 22 presents the studied variables, their baseline values, and the associated minima/maxima for the dedicated biorefinery, whereas Figure 14 displays the sensitivities of MFSP in a tornado plot. The variables are ranked in order of the extent of their impact, from largest to smallest.

The largest impact on MFSP was due to the uncertainty in adipic acid price, which was varied in the same price range as that presented in the 2018 design report ($0.50–$1.25/lb). Coproduct credits generated in a biorefinery context are an important strategy to help cellulosic fuels achieve MFSP under $2.50/GGE. MEK price, a significant component of the biorefinery's OPEX, could also be a major cost driver if chemical-grade MEK was to be purchased in the market (this was debated in Section 4.2). Accordingly, if another low-cost source of MEK was to be developed, such as through the conversion of carboxylic acids from wastewater-based arrested anaerobic digestion, the MFSP could be further improved by $0.21/GGE if MEK was to be purchased at $0.20/lb. The furan yield from sugar dehydration also influences MFSP results due to a reduction in fuel yield (from 78.5 MM

GGE/year in the base case to 62.8 MM GGE/year in the 80% molar yield sensitivity case), which ends up increasing costs on a per-GGE basis. The uncertainty in capital cost associated with the approach used in this study (±25% TCI) shows the fourth-largest impact on MFSP. Next, parameters tied to the lignin valorization sections (such as muconic acid productivity during aerobic fermentation and the amount of biologically accessible lignin) also have a significant impact because they directly affect the amount of adipic acid that could be sold as a high-value coproduct. Other variables, including those related specifically to the fuel train, have less pronounced influence over MFSP. For information purposes, natural gas prices were varied between $3 and $5/MM BTU, which represents the general price range in the last 8 years [20]. Because hydrogen is considered to be produced via steam methane reforming, its purchase price was varied between $0.55 and $0.91/lb, corresponding to the minimum and maximum natural gas prices of $3 and $5/MM BTU, respectively.

Table 23. Assumptions Varied in the Sensitivity Analysis of the Dedicated Biorefinery

	Assumption	Min MFSP	Baseline	Max MFSP
Pretreatment	DMR NaOH loading (mg/g)	50	70	100
Enzymatic Hydrolysis and Conditioning	Vacuum filter press capital cost	−50%	-	+50%
	Flocculant loading (g/kg solid)	10	20	30
	Enzymatic hydrolysis enzyme loading (mg/g)	5	10	20
Furans Production and Upgrading	Sugar dehydration to furans (molar yield)	-	100%	80%
	Dehydration reaction residence time (min)	-	5	10
	Dioxane:hydrolysate loading (v/v)	1.0:1	1.5:1	-
	NaOH loading (g/L)	-	3.6	7
	HDO:decarbonylation split in hydrotreating	-	100:0	90:10
	Hydrotreating reactor capital cost	−50%	-	50%
	Hydrotreating WHSV (h^{-1})	10	2	1
	Furans production train capital cost	−50%	-	50%
	MEK price ($/lb)	0.20	0.30	0.77
Lignin Utilization	Muconic productivity (g/L/h)	2	1	0.5
	Metabolically accessible lignin (wt%)	0.675	0.533	0.405
	Adipic acid price ($/lb)	1.25	0.86	0.50
Economics	Total capital investment	−25.0%	-	25.0%
	Feedstock cost ($/dry ton)	60	71.26	80
	Hydrogen price ($/lb)	0.55	0.73	0.91
	Natural gas price ($/MM BTU)	3	3.50	5

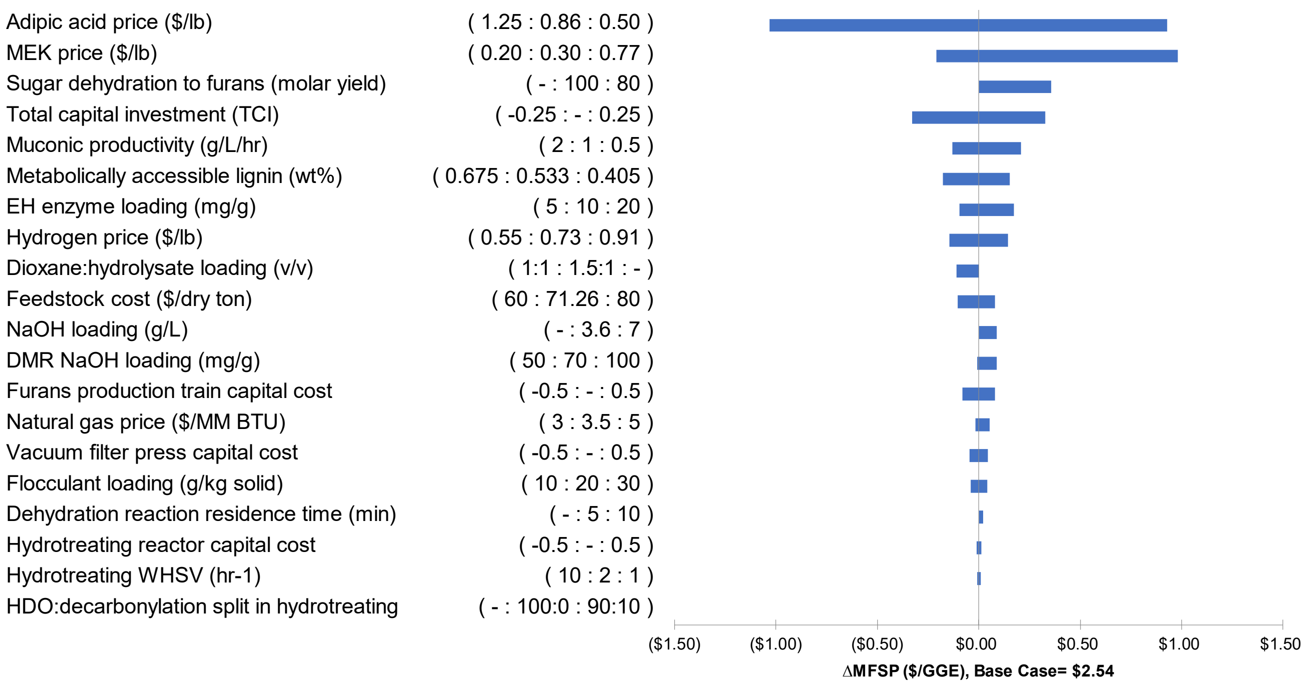

Figure 14. Dedicated biorefinery single-point sensitivity tornado chart for MFSP

5.2.2 Single-Point Sensitivity Analysis: Integrated Case

As for the **dedicated** biorefinery, Table 24 lists the studied variables, their baseline values, and the associated minima/maxima for the sensitivity analysis of the **integrated** plant. Figure 15 then presents the tornado plot for sensitivities of MFSP.

In a similar way to the **dedicated** case, the selling price associated with adipic acid is the main factor in the sensitivity analysis for the **integrated** biorefinery. This supports the narrative of needing to add value to all biomass fractions so fuel production from cellulosic biomass can reach low MFSP targets. Other factors (TCI, furan yield from sugar dehydration, and parameters linked to lignin fermentation) are also ranked high in their influence over MFSP for the same reasons as discussed in the previous section.

Table 24. Assumptions Varied in the Sensitivity Analysis of the Integrated Biorefinery

	Assumption	Min MFSP	Baseline	Max MFSP
Pretreatment	DMR NaOH loading (mg/g)	50	70	100
Enzymatic Hydrolysis and Conditioning	Vacuum filter press capital cost	−50%	-	+50%
	Flocculant loading (g/kg solid)	10	20	30
	Enzymatic hydrolysis enzyme loading (mg/g)	5	10	20
Furans Production and Upgrading	Sugar dehydration to furans (molar yield)	-	100%	80%
	Dehydration reaction residence time (min)	-	5	10
	Dioxane:hydrolysate loading (v/v)	1.0:1	1.5:1	-
	NaOH loading (g/L)	-	3.6	7
	HDO:decarbonylation split in hydrotreating	-	100:0	90:10
	Hydrotreating reactor capital cost	−50%	-	50%
	Hydrotreating WHSV (h^{-1})	10	2	1
	Furans production train capital cost	−50%	-	50%
BDO Fermentation and MEK Production	Fermentation glucose to product (wt%)	100%	95%	85%
	BDO to MEK (molar yield)	0.90	0.803	0.75
	BDO upgrading WHSV (h^{-1})	1	0.24	0.1
	BDO/MEK train capital cost	−50%	-	50%
Lignin Utilization	Muconic productivity (g/L/h)	2	1	0.5
	Metabolically accessible lignin (wt%)	0.675	0.533	0.405
	Adipic acid price ($/lb)	1.25	0.86	0.50
Economics	Total capital investment	−25.0%	-	25.0%
	Feedstock cost ($/dry ton)	60	71.26	80
	Hydrogen price ($/lb)	0.55	0.73	0.91
	Natural gas price ($/MM BTU)	3	3.50	5

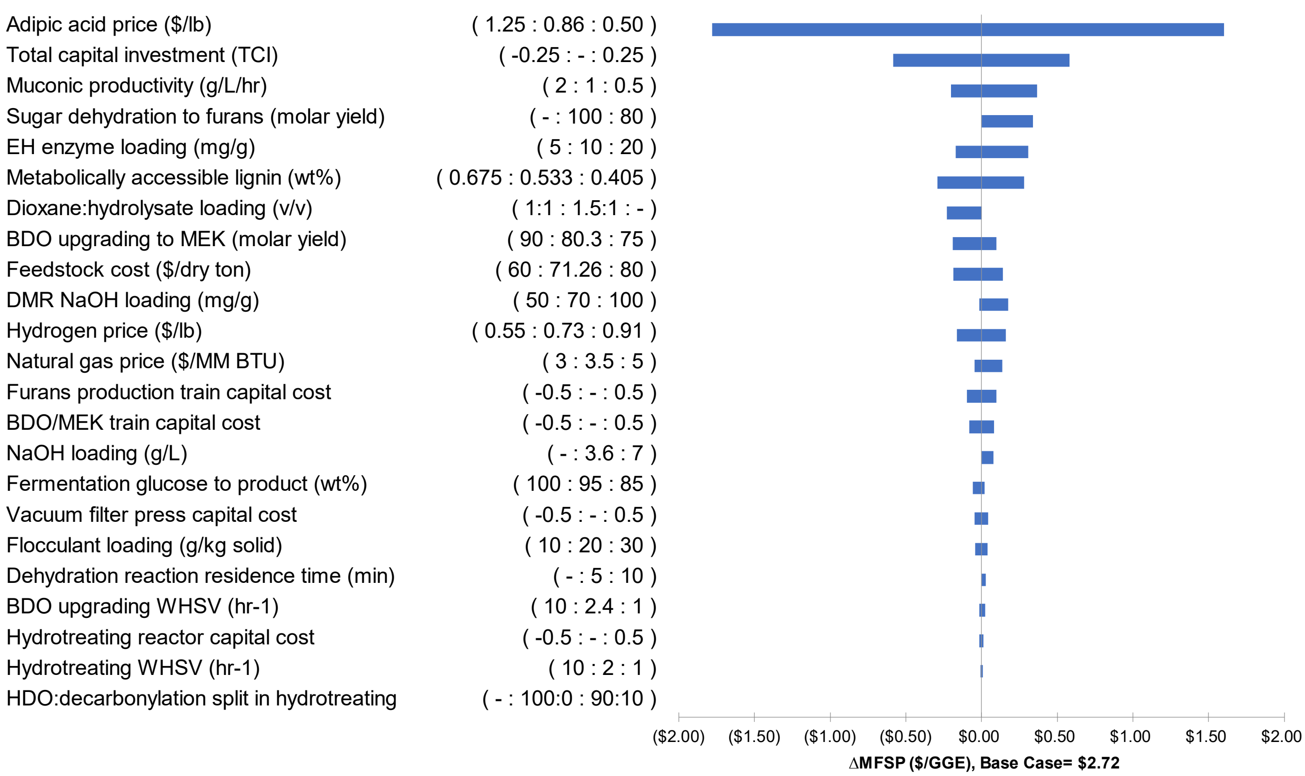

Figure 15. Integrated biorefinery single-point sensitivity tornado chart for MFSP

5.2.3 Case Study: Experimental Baseline

A case study reflective of the current level of experimental development linked to sugar dehydration, aldol condensation, and hydrotreating (all represented in Area 500 of the biorefineries) has also been conducted. By combining factors already explored in Sections 5.2.1 and 5.2.2 and summarized in Table 25, two new scenarios were generated, one for each biorefining setup, **dedicated** or **integrated**. They are deemed to be representative of the experimental baseline of the furans pathway prioritized under CUBI. For the comparison, all other parameters have been kept the same as in the base case (e.g., biomass deconstruction, lignin upgrading, and auxiliary sections).

Table 25. Parameters Varied in the Assessment of the Current Experimental Baseline

Parameter	Target case (2030)	Current experimental baseline
Dehydration		
HMF from C6 sugars (molar yield)	100%	72%
Furfural from C5 sugars (molar yield)	100%	90%
Reaction time (min)	5	15
Aldol condensation		
NaOH loading (g/L)	3.6	7.0
Hydrotreating		
Hydrotreating catalyst	1% Pd/Al_2O_3-SiO_2	5% Pd/Al_2O_3-SiO_2
HDO:decarbonylation split in hydrotreating	100:0	90:10

Table 26 shows the results for the **dedicated** biorefinery using current experimental parameters. The estimated MFSP is up to \$3.14/GGE, a delta of \$0.60/GGE. Following the discussion in Section 5.2.1, a reduction in the production of furans during sugar dehydration leads to a significant drop in fuel output (down by 12% to 61.4 MM GGE/year) and ends up being the main cost driver among the set of variables in Table 25.

Table 26. Summary of Yields, Rates, and Conversion Costs for the Dedicated Biorefinery Using Current Experimental Parameters

	Dedicated biorefinery	
	Target case (2030)	**Current experimental baseline**
Feedstock rate	2,205 dry U.S. tons/day	
Online time	7,884 h/yr (90% online factor)	
Total fuel yield	108.4 GGE/dry U.S. ton feedstock	84.7 GGE/dry U.S. ton feedstock
Total fuel production rate	78.5 MM GGE/yr	61.4 MM GGE/yr
Adipic acid coproduct yield	284 lb/dry U.S. ton feedstock	284 lb/dry U.S. ton feedstock
Adipic acid production rate	205 MM lb/yr	205 MM lb/yr
Total variable OPEX excluding coproducts	\$269 MM/yr	\$255 MM/yr
Coproduct revenue	\$193 MM/yr	\$190 MM/yr
Total fixed OPEX	\$20 MM/yr	\$21 MM/yr
Total equipment cost	\$414 MM	\$430 MM
Total capital investment	\$787 MM	\$815 MM
TCI per annual gallon	\$10.02/GGE	\$13.28/GGE
Minimum Fuel Selling Price	**\$2.54/GGE**	**\$3.14/GGE**
Feedstock contribution	\$0.66/GGE	\$0.84/GGE
Fuel conversion contribution	\$3.11/GGE	\$3.87/GGE
Coproduct conversion contribution	−\$1.23/GGE	−\$1.57/GGE

The results obtained for the case study related to the **integrated** biorefinery are summarized in Table 27. The assessment determined an MFSP of $3.31/GGE when considering current experimental data, corresponding to an increase of $0.59/GGE in comparison to the target case. A similar reasoning as for the **dedicated** biorefinery can be applied to explain the results, which are mainly dependent on lower fuel yields and higher processing costs.

Table 27. Summary of Yields, Rates, and Conversion Costs for the Integrated Biorefinery Using Current Experimental Parameters

	Integrated biorefinery	
	Target case (2030)	**Current experimental baseline**
Feedstock rate	2,205 dry U.S. tons/day	
Online time	7,884 h/yr (90% online factor)	
Total fuel yield	61.2 GGE/dry U.S. ton feedstock	52.6 GGE/dry U.S. ton feedstock
Total fuel production rate	44.3 MM GGE/yr	38.1 MM GGE/yr
Adipic acid coproduct yield	276 lb/dry U.S. ton feedstock	277 lb/dry U.S. ton feedstock
Adipic acid production rate	200 MM lb/yr	201 MM lb/yr
Total variable OPEX excluding coproducts	$187 MM/yr	$187 MM/yr
Coproduct revenue	$188 MM/yr	$187 MM/yr
Total fixed OPEX	$20 MM/yr	$21 MM/yr
Total equipment cost	$416 MM	$429 MM
Total capital investment	$786 MM	$811 MM
TCI per annual gallon	$17.75/GGE	$21.32/GGE
Minimum Fuel Selling Price	**$2.72/GGE**	**$3.31/GGE**
Feedstock contribution	$1.17/GGE	$1.36/GGE
Fuel conversion contribution	$3.71/GGE	$4.49/GGE
Coproduct conversion contribution	−$2.16/GGE	−$2.53/GGE

Results could be slightly improved if anaerobic digestion were employed as an additional step in WWT to harness part of the chemical oxygen demand present in the wastewater leaving Area 500 due to the partial conversion of sugars into furans. Carbon in that stream could be converted into biogas to help offset natural gas purchase to fuel the boiler section.

5.2.4 Case Study: BDO Fermentation with Clarified Hydrolysate

The final case study investigates the effect of changing the substrate for BDO fermentation in Area 550 from whole slurry (still with lignin mixed with sugars) to a clarified hydrolysate. Based on guidance from the research and development team, conducting fermentation on clarified sugars tends to be preferred over whole-slurry strategies in future experimental developments. The absence of lignin in the fermentors could improve substrate homogeneity and mass-transfer properties. Using a clarified hydrolysate could also allow for the recycle of cells in fermentation approaches that require this feature. The exact economic impact of these possibility is yet to be determined.

In this case study, clarified hydrolysate, both prior and after evaporation, is diverted to BDO fermentation instead of whole-slurry hydrolysate. Figure 16 presents the points at which sugar is retrieved in Area 300. In the simulation, 19.8% from the dilute hydrolysate and 30.2% from the concentrated sugar stream are sent to Area 550, resulting in a final sugar split of 56% to Area 500 and 44% to Area 550. For consistency purposes, BDO titer was maintained at 97 g/L as in the 2018 design report and in the base integrated biorefinery assessed in this report, both carried out with whole-slurry fermentation. Finally, the lignin press required for solids removal after whole-slurry fermentation in the base case is forfeited in this case study. Table 28 summarizes the main results from the TEA comparing both scenarios of the integrated biorefinery.

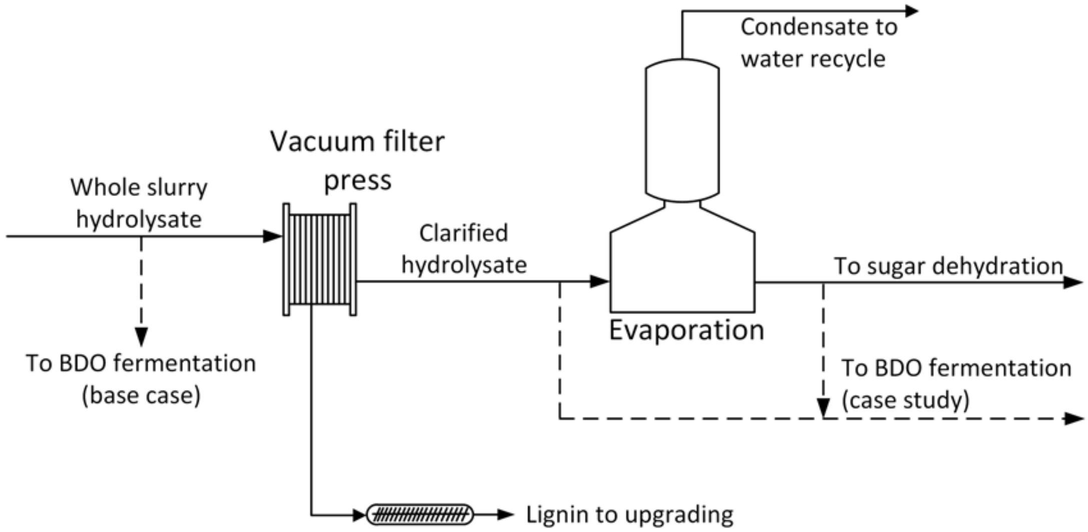

Figure 16. Diagram depicting the points in Area 300 at which sugars are diverted to BDO fermentation. The base case utilizes whole-slurry hydrolysate, whereas the case study employs a combination of diluted and concentrated clarified hydrolysate.

Table 28. Summary of Yields, Rates, and Conversion Costs for the Case Study on Clarified Hydrolysate Utilization

	BDO fermentation substrate	
	Whole slurry (base case)	**Clarified hydrolysate (case study)**
Feedstock rate	2,205 dry U.S. tons/day	
Online time	7,884 h/yr (90% online factor)	
Total fuel yield	61.2 GGE/dry U.S. ton feedstock	60.7 GGE/dry U.S. ton feedstock
Total fuel production rate	44.3 MM GGE/yr	44.0 MM GGE/yr
Adipic acid coproduct yield	276 lb/dry U.S. ton feedstock	284 lb/dry U.S. ton feedstock
Adipic acid production rate	200 MM lb/yr	205 MM lb/yr
Total variable OPEX excluding coproducts	$187 MM/yr	$191 MM/yr
Coproduct revenue	$188 MM/yr	$193 MM/yr
Total fixed OPEX	$20 MM/yr	$21 MM/yr
Total equipment cost	$416 MM	$432 MM
Total capital investment	$786 MM	$818 MM
TCI per annual gallon	$17.75/GGE	$18.59/GGE
Minimum Fuel Selling Price	**$2.72/GGE**	**$2.81/GGE**
Feedstock contribution	$1.17/GGE	$1.17/GGE
Fuel conversion contribution	$3.71/GGE	$3.86/GGE
Coproduct conversion contribution	−$2.16/GGE	−$2.22/GGE

The main result of the analysis is an MFSP that is only $0.09/GGE higher in comparison to the base integrated plant, amounting to $2.81/GGE. The clarified hydrolysate case has higher processing costs, both in terms of OPEX in the form of flocculant usage and electricity consumption, and in terms of capital investment due to the need for a large-sized VFP ($14-MM installed cost VFP in the base case, which rises to $24 MM in the updated case study). There are only marginal gains ($0.06/GGE) from additional adipic acid production. Also, it is worth noting that the increase of $0.09/GGE in MFSP due to the use of a VFP to clarify the slurry leaving the enzymatic hydrolysis is specific to this hybrid biorefining setup (considering both fermentation- and catalytic-based processes to convert sugars into hydrocarbon fuels). The result should be more pronounced in a purely fermentation-centered plant, such as that assessed in the 2018 design report for the BDO case. The main reason for this conclusion is essentially the lower fuel yields attained by such biorefineries, which would translate to a higher increase in processing cost on a per-GGE basis.

Provided that future experimental efforts with BDO fermentation (and that of other metabolites) will likely be increasingly based on clarified hydrolysates in detriment of the whole slurry, researchers should be aware of the potential TEA ramifications linked to choosing this strategy. Future TEA efforts could be devoted to examining in detail the impacts of switching to this strategy; for example, understanding the benefits of combining batch hydrolysis with VFP vs. whole-slurry BDO fermentation or continuous enzymatic hydrolysis and fermentation to carboxylic acids (cases described in the 2018 design report).

5.3 Sustainability Metric Indicators

This section presents primary sustainability metric indicators of the current conceptual process at the conversion stage derived from Aspen Plus simulations. Table 29 summarizes the key sustainability metric indicators for the two biorefining strategies assessed in this report. The processes differ significantly in terms of fuel yield, carbon efficiency to fuels, and natural gas imports. The **dedicated** biorefinery benefits from the purchase of externally sourced MEK to improve fuel yields to beyond 108 GGE/dry ton of biomass. In this plant, 26% of the carbon in the hydrocarbon fuel comes from the ketone, while the remainder is obtained from the biomass via the catalytic conversion of sugars. For the **integrated** plant, the fuel yield is 61 GGE/dry ton of biomass because it split efforts in obtaining both classes of compounds (furans and ketones) needed in the aldol condensation reaction. The overall combined carbon efficiency (to fuel and adipic acid) for the **dedicated** and **integrated** plants are 62.6% and 51.1%, respectively, considering only the carbon in the biomass feedstock. Both plants are energy-intensive and require natural gas supplementation in the boiler to supply the high energy demand. Finally, net water consumption is nearly 7,000 m^3 per day for either plant, which is high in absolute terms but comparable to the pathways in the 2018 design report on a per-GGE basis.

Table 29. Summary of Sustainability Metric Indicators for the Modeled Biorefining Configurations

		Biorefinery	
Sustainability Metrics		**Dedicated**	**Integrated**
Fuel yield by weight of biomass	GGE per dry ton biomass	108.4	61.2
Carbon efficiency to fuels	% C in biomass	46.8	35.8
Carbon efficiency to adipic acid	% C in biomass	15.8	15.3
Electricity import	kWh/GGE	5.2	6.4
Natural gas import	MJ/GGE	36.3	96.0
Water consumption	gal/GGE	7.4	13.4
Water consumption	m^3/day	6,745	6,836

5.4 Additional Opportunities for Cost Reduction

Beyond the process configurations considered in this report, several other opportunities could be followed to further reduce MFSP. Although outside the scope of the present TEA work, these potential strategies are briefly discussed and could be evaluated in more detail in the future.

Add Value to Furans

Both **dedicated** and **integrated** biorefineries assessed in this report aim to produce fuels via an approach centered around the production of furans through sugar dehydration as a means of obtaining reactive molecules for aldol condensation, with a ketone and further hydrotreating. The process could benefit from selling part of them directly to the market, because both furfural and HMF are products with industrial use. This is an especially interesting option in view of the high yields of furans expected from sugars in the hydrolysate stream in future scenarios; such biorefineries could afford diverting a portion of furans to this end. Alternatively, furans are also precursors to a multitude of other relevant products, as depicted in Figure 17. Such pathways, mostly catalytic ones, could give origin to compounds such as γ-valerolactone, caprolactam, and 2,5-furandicarboxylic acid (FDCA) from HMF [26] and 2-methyltetrahydrofuran (MTHF) and furfuryl alcohol from furfural [27]. In short, selling furans to the market or producing different compounds from furans would expand the biorefinery concept presented in this report to a multiproduct plant, besides the already produced hydrocarbon fuel, adipic acid, and sodium sulfate. The exact economic impact and feasibility of either approach should be assessed in future efforts.

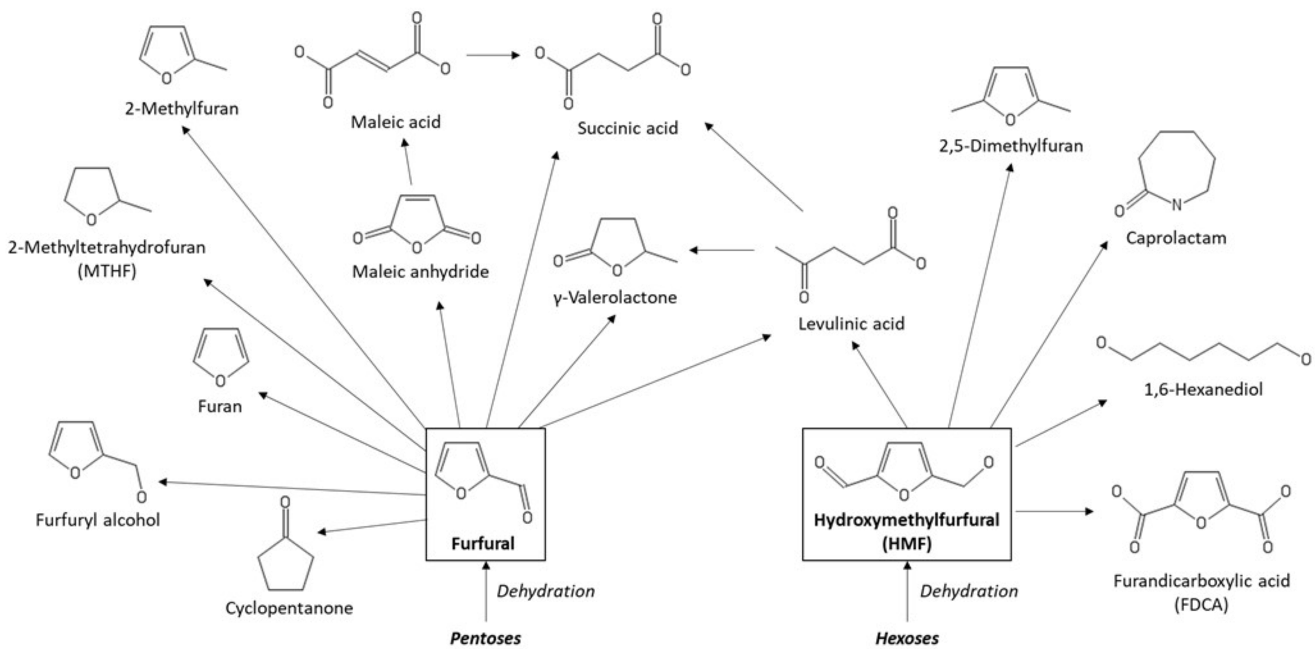

Figure 17. Pathways from furfural and HMF to products of industrial interest (based on [27] and [28])

Add Value to Coproducts from BDO Dehydration

The catalytic upgrading of BDO to MEK in Area 550 generates several byproducts, such as isobutanol (major compound) and minor fractions of 1,3-butadiene, isobutanal, and 3-buten-2-ol. The molar yield to each molecule is summarized in Table 11 (Section 3.6.2). The current sequence of distillation columns after BDO upgrading aims to recover MEK close to its azeotrope composition with water for further use in the aldol condensation reaction with furans in Area 500. Water and other organic compounds present in the stream leaving the BDO upgrading reactor are currently sent to WWT. Although recovering some of these compounds with the intention of selling them to market would entail additional CAPEX and OPEX, this possibility could ultimately have a positive impact on the carbon efficiency of the biorefinery, its sustainability metrics, and economic performance.

Add Value to Lignin with Alternative Pathways

As pointed out in this report, adipic acid is intended to represent one example of many other potential products from lignin. In a scenario with the deployment of multiple similar biorefineries dedicated to the production of hydrocarbon fuels, several of them would be expected to add value to lignin in different ways to be able to generate coproduct credits without oversupplying the market with a single compound. This could be achieved with a multitude of routes, either biological ones, catalytic processes, or a combination thereof [28].

6 Concluding Remarks

6.1 Summary

This report, an effort stemming from the CUBI project within the Chemical Catalysis for Bioenergy Consortium (ChemCatBio), proposes two conceptual biorefining configurations for the production of upgraded renewable hydrocarbon fuels coupled with bio-derived coproducts. The core of the biorefineries is a pathway that initiates with sugar dehydration to furans, an aldol condensation reaction with a ketone, and a final step of hydrotreating to obtain hydrocarbons in the C_{14}–C_{16} range. A **dedicated** biorefinery focuses at converting all sugars in the hydrolysate stream to furans (furfural and HMF) while purchasing external MEK to carry out the final stages of fuel production. On the other hand, an **integrated** biorefinery produces both furans and MEK from sugars to generate the hydrocarbon fuel; MEK is obtained after upgrading of BDO synthesized through whole-slurry fermentation. In addition to biomass deconstruction to sugars and lignin valorization through adipic acid production via muconic acid fermentation, the report sets future performance targets for the full pathway dedicated to sugar upgrading to fuels.

Final results of the TEA point to an estimated MFSP of **$2.54/GGE** for the **dedicated** biorefinery (2016$) at a final upgraded fuel product yield of 108.4 GGE/dry ton of biomass and an MFSP of **$2.72/GGE** for the **integrated** plant at a fuel product yield of 61.2 GGE/dry ton of biomass. Both results are comparable to those obtained in a previous NREL design case [1], indicating that the pathways presented herein are viable alternatives to achieve similar fuel cost targets through purely catalytic upgrading of sugars. The inclusion of adipic acid as a high-value, lignin-derived coproduct in the plants at a yield of 284 and 276 lb/dry ton of biomass, respectively, can significantly contribute toward reducing MFSP, corresponding to credits of *negative* $1.23/GGE and $2.16/GGE, respectively. A sensitivity analysis was also critical to pinpoint the major cost drivers of the biorefineries. The main factor for either biorefining setup is adipic acid price, which could lead to large swings in MFSP depending on its market price. The yield to furans in sugar dehydration is also a highly impactful parameter. For the **dedicated** plant specifically, the price at which MEK is purchased could also lead to significantly different MFSP results.

Finally, the report also presents key sustainability metric indicators for the modeled biorefining configurations, which are suggestive of the environmental performance of the systems. For the **dedicated** plant, these metrics were estimated at 48% carbon recovery in the form of hydrocarbon fuels, 5.2 kWh/GGE net power import, 36 MJ/GGE natural gas consumption, and 7.4 gal/GGE net water demands for the biorefinery. For the **integrated** biorefinery, the same metrics were estimated at 29% carbon recovery as fuels, 6.4 kWh/GGE power import, 96 MJ/GGE natural gas consumption, and 13.4 gal/GGE net water demand.

6.2 Future Work

Moving forward, certain possibilities for further development and assessment could ultimately achieve cost goals and reduce uncertainty in key areas for the modeled processes evaluated here:

- **Achieve future high-yield targets of furans from sugars:** Among the parameters directly tunable under CUBI, achieving (or maintaining) a high yield to furans via sugar dehydration stands among the top three priorities identified during the sensitivity analyses in Section 5.2.
- **Improve catalyst performance for BDO upgrading to MEK:** The unwanted conversion of BDO to minor coproducts during catalytic upgrading to MEK has several detrimental effects to the **integrated** biorefining configuration assessed in this report, namely the loss of carbon from BDO to other nonreactive compounds and the need for a sequence of distillation columns to effectively purify MEK for further use. A reduction in the number of coproducts would be highly beneficial to achieve lower MFSPs. This is also a parameter that could be modulated under CUBI.
- **Achieve future lignin deconstruction, bioconversion, and upgrading targets:** The assessment clearly shows the dependence of achieving low MFSPs on the commercialization of a high-value coproduct in the form of adipic acid from lignin. For the assumptions to be valid, the targets considered in the TEA for lignin deconstruction/conversion/upgrading should receive special attention from the associated experimental development team.
- **Assess (in detail) the impact of carrying out fermentations using clarified hydrolysate:** A preliminary assessment of this strategy was presented in Section 5.2.4. Further TEA work related to this fermentation approach could be conducted to examine in detail its benefits and disadvantage.
- **Assess options for product portfolio diversification:** In an integrated, multiproduct biorefinery, alternative options for diversifying product portfolio and improving the robustness of the plant could be possible. These may pass through adding value to furans, to coproducts obtained during BDO dehydration, and to lignin through different conversion pathways, as detailed in Section 5.4.

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Appendix A. Individual Equipment Costs Summary

The following tables show abbreviated specifications, purchased cost, and installed cost for each piece of equipment in this process design. Although each piece of equipment has its own line, many were quoted as part of a package, so their scaling calculations are not shown. NREL would like to acknowledge the subcontractors and equipment vendors who assisted us with cost estimates over recent years as were utilized for this report.

Dedicated Biorefinery

A200: Pretreatment	Mechanical Equipment List					Scaled Installed Costs											
EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
Deacetylation Tank Discharge Pump	1771 GPM, 150 FT TDH	100	316SS	1	$22,500	2009	$22,500	strm.a200.211a	402194	kg/hr	0.80	2.3	283668	0.71	$17,017	$17,663	$40,624
Deacetylation reactor conveyors	Feed and discharge drag conveyors	40 hp	SS316	3	$110,000	2013	$110,000	strm.a200.211a	277167	kg/hr	0.80	1.7	283668	1.02	$336,178	$321,008	$545,713
Pretreatment Sugar Beet Extruder				1	$5,424,000	2010	$5,424,000	SCIS.a200.DEAC-IN	38600	kg/hr	1.00	1.0	70483	1.83	$9,904,187	$9,740,555	$9,740,555
Flash Tank Agitator	Side-mounted, 3 x 75 hp. (170 kW)	170 kW	316LSS	3	$90,000	2009	$90,000	strm.a200.254	252891	kg/hr	0.50	1.5	192174	0.76	$78,455	$81,432	$122,148
Ammonia Addition Tank Agitator		10 hp	SS	1	$21,900	2009	$21,900	strm.a200.228	410369	kg/hr	0.50	1.5	230674	0.56	$16,419	$17,042	$25,563
Ammonia Static Mixer			SS	1	$5,000	2009	$5,000	strm.a200.275	157478	kg/hr	0.50	1.0	38501	0.24	$2,472	$2,566	$2,566
Pretreatment Water Heater	29.9 MMBtu		304SS	1	$92,000	2010	$92,000	Heat.A200.QH201	-8	Gcal/hr	0.70	2.2	-0.3	0.04	$9,867	$9,704	$21,348
Milling Equipment	200kw/dry ton			8	$2,466,700	2013	$2,466,700	SCIS.a200.211b	62942	kg/hr	0.60	1.5	57306	0.91	$19,733,600	$18,843,101	$28,264,651
Milling Equipment-Szego Mill				11	$578,000	2013	$578,000	SCIS.a200.211b	62942	kg/hr	0.60	1.4	57306	0.91	$6,358,000	$6,071,089	$8,499,524
Blowdown Tank Discharge Pump	1900 GPM, 150 FT TDH	125	316SS	1	$25,635	2010	$25,635	strm.a200.222	292407	kg/hr	0.80	2.3	193160	0.66	$18,398	$18,094	$41,617
Flash Tank Discharge Pump	900 GPM, 150 FT TDH	75	316SS	1	$30,000	2009	$30,000	strm.a200.254	204390	kg/hr	0.80	2.3	192174	0.94	$28,557	$29,640	$68,172
Hydrolyzate Pump	1771 GPM, 150 FT TDH	100	316SS	1	$22,500	2009	$22,500	strm.a200.228	402194	kg/hr	0.80	2.3	230674	0.57	$14,422	$14,969	$34,430
S/L Split Discharge Pump to WWT	900 GPM, 150 FT TDH	75	316SS	1	$30,000	2009	$30,000	strm.a200.4	204390	kg/hr	0.80	2.3	193160	0.95	$28,674	$29,762	$68,452
Flash Tank	23' x 48' - 110,000 gal.		SS316	1	$511,000	2009	$511,000	strm.a200.223	264116	kg/hr	0.70	2.0	193160	0.73	$410,493	$426,067	$852,133
Ammonia Addition Tank	118,000 gal, 1hr residence time		SS304	1	$236,000	2009	$236,000	strm.a200.228	410369	kg/hr	0.70	2.0	230674	0.56	$157,684	$163,667	$327,334
														Area 200 Totals	**$37,114,424**	**$35,786,358**	**$48,654,831**

A300: Hydrolysis and Fermentation	Mechanical Equipment List					Scaled Installed Costs											
EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
Batch Enzymatic Hydrolysis																	
Hydrolyzate Cooler	Plate & Frame 32.5 MMBtu/hr		SS 304	1	$85,000	2010	$85,000	heat.A300.EH.QC301	8	Gcal/hr	0.70	2.2	7	0.80	$72,529	$71,331	$156,928
Enzyme-Hydrolysate Mixer	inline mixer 1673 gpm	100 hp	SS316	1	$109,000	2009	$0	strm.a300.EH.t310fd	379938	kg/hr	0.50	1.7	2	0.00	$0	$0	$0
Saccharification Tank	250,000 gal each - 19' dia. x 120' tall		304SS	8	$3,840,000	2009	$3,840,000	strm.A300.EH.306	421776	kg/hr	0.70	2.0	238249	0.56	$2,574,520	$2,672,192	$5,344,385
Saccharification Transfer Pump	352 GPM, 150 FT TDH	20	316SS	5	$47,200	2009	$47,200	strm.a300.EH.306	421776	kg/hr	0.80	2.3	238249	0.56	$29,888	$31,022	$71,351
Enzymatic Hydrolysis Storage Tank	1,200,000 gallon		316SS	1	$1,317,325	2011	$1,317,325	strm.a300.EH.306A	328984	kg/hr	0.70	1.8	238249	0.72	$1,050,976	$972,023	$1,749,641
Fermentor Tank (saccharification contribution)			304SS	12	$10,128,000	2009	$10,128,000		12	ea	1	1.50	7.8	0.65	$6,623,653	$6,874,944	$10,312,415
Fermentor Agitator (saccharification contirbution)		30 hp	SS304	1	$52,500	2009	$52,500		1	ea	1	1.50	7.8	7.85	$412,016	$427,648	$641,471
Fermentation Cooler (saccharification contribution)	Plate & frame		304SS	12	$86,928	2009	$86,928		12	ea	1	2.20	7.8	0.65	$56,850	$59,007	$129,816
Fermentation Recirc/Transfer Pump (saccharification contribution)	340 GPM, 150 FT	20	316SS	5	$47,200	2009	$47,200		12	ea	0.8	2.30	7.8	0.65	$33,605	$34,880	$80,223
Vacuum filter press (VFP)	(4) 170 m2 Horizontal Belt Filters	660 hp ea	304SS	7	$2,152,500	2013					1	1.7		1.00	$15,067,500	$14,387,563	$24,458,858
Sugar Concentration		3600 kW	304SS	1	$6,370,000	2013	$6,370,000	STRM.A500.FURANS.SUG-CONC.301SUG	244084	kg/hr	0.70	2.0	361769	1.48	$8,390,032	$8,011,423	$16,022,847
Concentrated Sugar Storage Tank	5,500 gallons - 20 min residence time	20 hp	SS	1	$168,000	2011	$168,000	STRM.A500.FURANS.S16	76712	kg/hr	0.70	1.8	361769	4.72	$497,516	$460,140	$828,253
														Area 300 Totals	**$34,809,086**	**$34,002,173**	**$59,796,188**

A400: Enzyme Production	Mechanical Equipment List					Scaled Installed Costs											
EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
Cellulase Fermentor Agitators		800.0	SS316		$580,000	2009	$580,000	CLVESSEL	1	ea	1.00	1.5	5	5.00	$2,900,000	$3,010,021	$4,515,032
Cellulase Fermentor Agitators		0.75hp	SS316		$3,420	2009	$3,420	ICLSEED	1	ea	1.00	1.5	4	4.00	$13,680	$14,199	$21,298
Cellulase Fermentor Agitators		8 hp	SS316		$11,000	2009	$11,000	ICLSEED	1	ea	1.00	1.5	4	4.00	$44,000	$45,669	$68,504
Cellulase Fermentor Agitators		80 hp	SS316		$63,000	2009	$63,000	ICLSEED	1	ea	1.00	1.5	4	4.00	$252,000	$261,560	$392,341
Media-Prep Tank Agitator		7.5 hp	A285C	1	$8,500	2009	$8,500	strm.a400.402a	12255	kg/hr	0.50	1.5	0	0.00	$0	$0	$0
Cellulase Nutrient Mix Tank Agitator		3 hp	CS	1	$4,800	2009	$4,800	strm.a400.416	174	kg/hr	0.50	1.6	123	0.70	$4,030	$4,182	$6,692
Cellulase Hold Tank Agitator		10 hp	SS316	1	$26,900	2009	$26,900	strm.422	10930	kg/hr	0.50	1.5	7575	0.69	$22,395	$23,244	$34,866
Cellulase Fermentor	80,000 gal, 1 atm, 28 °C, Internal coil		SS316		$400,500	2009	$400,500	CLVESSEL	1	ea	1.00	2.0	5	5.00	$2,002,500	$2,078,471	$4,156,943
1st Cellulase Seed Fermentor	80 gallon skid complete - $46,000 ea		304SS		$46,000	2009	$46,000	ICLSEED	1	ea	1.00	1.8	4	4.00	$184,000	$190,981	$343,765
2nd Cellulase Seed Fermentor	800 gallon skid complete - $57,500 ea		304SS		$57,500	2009	$57,500	ICLSEED	1	ea	1.00	1.8	4	4.00	$230,000	$238,726	$429,706
3rd Cellulase Seed Fermentor	8,000 gallon skid complete - $95,400 ea		304SS		$95,400	2009	$95,400	ICLSEED	1	ea	1.00	1.8	4	4.00	$381,600	$396,077	$712,939
Cellulase Fermentation Cooler	Cooling coil included with Cellulase Fermenter		304SS		INCLUDED												
Media Prep Tank Cooler	Cooling coil included with Media Prep Tank		304SS	1	INCLUDED												
Fermenter Air Compressor Package	8000 SCFM @ 16 psig		CS	2	$350,000	2009	$350,000	strm.a400.450	33168	kg/hr	0.60	1.6	17839	0.54	$241,242	$250,395	$400,632
Cellulase Transfer Pump	59 gpm, 100 FT, TDH SIZE 2X1-10C	3	316SS	1	$7,357	2010	$7,357	strm.a400.420	13399	kg/hr	0.80	2.3	7575	0.57	$4,662	$4,585	$10,545
Cellulase Seed Pump	3 GPM, 100 FT TDH SIZE 2X1-10	2	316SS	4	$29,972	2010	$29,972	strm.a400.409	681	kg/hr	0.80	2.3	421	0.62	$20,408	$20,071	$46,163
Media Pump	63 GPM, 100 FT TDH SIZE 2X1-10C	3	316SS	1	$7,357	2010	$7,357	strm.a400.402a	14307	kg/hr	0.80	2.3	0	0.00	$0	$0	$0
Cellulase Nutrient Transfer Pump	Gear Pump 2 GPM, 100 FT	1	316SS	1	$1,500	2009	$1,500	strm.a400.416	454	kg/hr	0.80	2.3	123	0.27	$526	$546	$1,257
Cellulase Feed Pump	Gear Pump	1	316SS	1	$5,700	2009	$5,700	strm.a400.422	18168	kg/hr	0.80	2.3	7575	0.42	$2,831	$2,938	$6,758
Anti-foam Pump	Gear Pump 2 GPM, 100 FT	1	316SS	1	$1,500	2009	$1,500	strm.a400.444	11	kg/hr	0.80	2.3	7.3	0.69	$1,115	$1,157	$2,661
SO2 Storage Tank	1 ton cylinders, incl w/ delivery		SS304	1	$0												
Media-Prep Tank	20,000 gallon, incl. coil		304SS	1	$176,000	2009	$176,000	strm.a400.402a	12255	kg/hr	0.70	1.8	0	0.00	$0	$0	$0
Cellulase Nutrient Mix Tank	HDPE, 8,000 gal		HDPE	1	$9,000	2010	$9,000	strm.a400.416	224	kg/hr	0.70	3.0	123	0.55	$5,903	$5,806	$17,417
Cellulase Hold Tank	80,000 gal		304SS	1	$248,070	2009	$248,070	strm.a400.422	10930	kg/hr	0.70	1.8	7575	0.69	$191,920	$199,201	$358,562
														Area 400 Totals	**$6,502,812**	**$6,747,831**	**$11,526,082**

A500: Furans Production & Upgrading	Mechanical Equipment List					Scaled Installed Costs											
EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
Hydrolysate-Dioxane Pump				1	$95,660	2002	95660	STRM.A500.FURANS.5	349268	lb/hr	0.33	2.47	786957	2.25	$125,070	$171,259	$423,011
Pre-Dehydration Economizer	2-4 TEMA shell and tube HX		316SS	1	$353,600	2011	$353,600	HEAT.A500.FURANS.55	14.3	Mmkcal/hr	0.7	2.65582579	74	5.18	$1,117,655	$1,033,693	$2,745,308
Pre-Dehydration Heater				1	$76,800	2015	$76,800	HEAT.A500.FURANS.HP-3	5.1827	Gcal/hr	1	3.04296875	25	4.91	$377,265	$367,034	$1,116,872
Sugar Dehydration Reactor	Tubular flow reactor, 5 min RT			1	$6,926,760	2013	$6,926,760	STRM.A500.FURANS.5				1.7			$6,842,072	$6,533,317	$11,106,639
					5	min		DEN.A500.FURANS.5									
Post-Dehydration Cooler				1	$76,800	2015	$76,800	HEAT.A500.FURANS.60	5.1827	Gcal/hr	1	3.04296875	36	6.95	$533,734	$519,259	$1,580,090
Aldol condensation reactor	5th seed fermentor (Humbird, 2009)			1	$1,180,000	2009	$1,180,000	STRM.A500.FURANS.S8	200000	gal	0.7	2	93280	0.47	$691,850	$718,097	$1,436,194
								DEN.A500.FURANS.S8									
Dioxane Column Condenser	ACCE			1	$267,100	2016	$267,100	HEAT.A500.FURANS.SEPRTN.S7	58	Gcal/hr	0.44	2.47	53	0.92	$257,526	$257,526	$636,089
Dioxane Column Condenser (ACC)	ACCE			1	$39,500	2016	$39,500	HEAT.A500.FURANS.SEPRTN.S7	58	Gcal/hr	0.44	2.47	53	0.92	$38,084	$38,084	$94,068
Dioxane Column Reboiler	ACCE			1	$254,200	2016	$254,200	HEAT.A500.FURANS.SEPRTN.S6	-67	Gcal/hr	0.79	2.47	-62	0.93	$239,976	$239,976	$592,741
Dioxane Column Reflux Pump	ACCE			1	$23,300	2016	$23,300	STRM.A500.FURANS.SEPRTN.S8	274694	kg/hr	0.79	2.47	268824	0.98	$22,906	$22,906	$56,577
Dioxane Column Tower	ACCE			1	$628,600	2016	$628,600	STRM.A500.FURANS.SEPRTN.S1	373388	kg/hr	0.68	2.47	367573	0.98	$621,926	$621,926	$1,536,157
Hydrocarbon Decanter	ACCE			1	$32,100	2016	$32,100	STRM.A500.FURANS.SEPRTN.BOT-C	98695	kg/hr	0.60	2.47	101050	1.02	$32,557	$32,557	$80,417
Dioxane Recycle Cooler				1	$76,800	2015	$76,800	HEAT.A500.FURANS.61	5.1827	Gcal/hr	1	3.04296875	7	1.35	$103,462	$100,656	$306,293
Hydrocarbon Pump	190 GPM, 615 FT TDH	40	316SS	2	$24,300	2009	$24,300	STRM.A500.FURANS.S22	43149	kg/hr	0.8	2.3	34087	0.79	$40,247	$41,774	$96,080
Pre-Hydrotreating Economizer	2-4 TEMA shell and tube HX		316SS	1	$353,600	2011	$353,600	HEAT.A500.FURANS.47	-14.3	Mmkcal/hr	0.7	2.66	-6	0.41	$190,634	$176,313	$468,255
Pre-Dehydration Heater				1	$76,800	2015	$76,800	HEAT.A500.FURANS.48	5.1827	Gcal/hr	1	3.04296875	2	0.36	$27,648.98	$26,899.16	$81,853.30
Fresh H2 compressor	Reciprocating compressor (5 stages)			1	$1,621,200	2011	$1,621,200	STRM.A500.FURANS.H2-FRESH	389.6519	kg/hr	0.6	1.09	3522	9.04	$6,074,032	$5,617,728	$6,122,603
Fresh H2 compressor (spare)	Reciprocating compressor (5 stages)			1	$1,621,200	2011	$1,621,200	STRM.A500.FURANS.H2-FRESH	389.6519	kg/hr	0.6	1.08	3522	9.04	$6,074,032	$5,617,728	$6,074,090
Hydrotreating Reactor	Fixed bed reactor (Q3 FY17 milestone), base PF=2.5,208 BBL/hr			1	$6,513,387	2011	$6,513,387	STRM.A500.FURANS.S22	32894.95	L/hr	0.7	2.00	13102	0.40	$3,419,408	$3,162,529	$6,325,058
Pressure factor (via Guthrie)	*(>1000PSIG=2.5,900=2.3,800=1.9,700=1.8,600=1.6,500=1.45,400=1.35)*							DEN.A500.FURANS.S22									
Flash vessel	23' x 48' - 110,000 gal.		316SS	1	$511,000	2009	$511,000	STRM.A500.FURANS.43	264116	kg/hr	0.7	2.00	39288	0.15	$134,631	$139,738	$279,477
PSA unit	H2 recovery			1	$975,000	2013	$975,000	STRM.A500.FURANS.26	13528	kg/hr	0.6	1.90	1937	0.14	$303,745	$290,039	$551,073
Recycle H2 compressor	Centrifugal compressor			1	$1,103,700	2011	$1,103,700	STRM.A500.FURANS.RECYC	14665.49	kg/hr	0.6	1.13	1679	0.11	$300,670	$278,083	$314,289
Recycle H2 compressor (spare)	Centrifugal compressor			1	$1,103,700	2011	$1,103,700	STRM.A500.FURANS.RECYC	14665.49	kg/hr	0.6	1.10	1679	0.11	$300,670	$278,083	$306,957
Air compressor				2	$34,600	2011	$34,600	STRM.A500.FURANS.44	3818.9219	kg/hr	0.6	1.82	4002	1.05	$71,172	$65,825	$119,855
Off-gas boiler				1	$241,400	2011	$241,400	HEAT.A500.FURANS.46	2.4187418	MMkcal/hr	0.7	1.52	2	0.96	$234,317	$216,714	$330,368
Flash vessel	23' x 48' - 110,000 gal.		316SS	1	$511,000	2009	$511,000	STRM.A500.FURANS.S1	264116	kg/hr	0.7	2	28804	0.11	$108,338	$112,448	$224,897
														Area 500 Totals	**$28,283,628**	**$26,680,191**	**$43,005,310**

A600: WWT	Mechanical Equipment List					Scaled Installed Costs											
EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
Aeration Basin	Concrete and steel, not installed cost		Concrete	3	$4,804,854	2012	$4,804,854	Hydraulic flow	2.7	MGD	0.60	2.1	1.7	0.61	$3,578,308	$3,315,719	$6,863,538
Pump - Centrifugal, Aeration Basin Feed	852 gpm ea	45 hp	CS	4	$64,800	2012											
Aeration Grid	Full floor aeration grid		CS	1	$2,500,000	2012											
Caustic Feed System		1.5 hp	CS	4	$20,000	2012	$20,000	COD	5600	kg/hr	0.60	3.0	1956	0.35	$10,641	$9,860	$29,579
Blowers	15000 SCFM @ 10.3psig ea	1000 hp ea	CS	9	$2,070,000	2012	$2,070,000	COD	5600	kg/hr	0.60	2.0	1956	0.35	$1,101,310	$1,020,492	$2,040,984
Membrane Bioreactor	Includes membrane, CIP, Scour system	85 hp ea	CS	1	$4,898,500	2012	$4,898,500	Hydraulic flow	2.7	MGD	1.00	1.6	1.7	0.61	$2,997,268	$2,777,318	$4,554,801
Pump, Centrifugal , MBR, RAS		160 hp	CS	6	INCLUDED	2012											
Gravity Belt Thickeners	2m presses	48hp	CS	3	$750,000	2012	$750,000	COD	5600	kg/hr	0.60	1.6	1956	0.35	$399,025	$369,744	$587,892
Centrifuge		165 hp ea	CS	1	$686,800	2012	$686,800	COD	5600	kg/hr	0.60	2.7	1956	0.35	$365,401	$338,586	$910,798
Pump, Centrifugal, Centrifuge Feed	105 gpm	15hp	CS	2	INCLUDED	2012											
Pump, Submersible, Centrate	100 gpm	10 hp ea	CS	2	INCLUDED	2012											
Dewatering Polymer Addition	9.8 gph neat polymer	1 hp ea	CS	2	INCLUDED	2012											
Conveyor		10 hp ea	CS	1	$7,000	2012	$7,000	COD	5600	kg/hr	0.60	2.9	1956	0.35	$3,724	$3,451	$9,870
Reverse Osmosis			CS	7	$2,450,000	2012	$2,450,000	Hydraulic flow	2.7	MGD	1.00	1.8	1.7	0.61	$1,499,093	$1,389,084	$2,430,897
Evaporator	368 gpm	1480 hp ea	Titanium	1	$5,000,000	2012	$5,000,000	Hydraulic flow	2.7	MGD	0.60	1.6	1.7	0.61	$3,723,638	$3,450,385	$5,555,119
Ammonia Addition System	0.63 gpm	4.5 hp	CS	4	$195,200	2012	$195,200	COD	5600	kg/hr	0.60	1.5	1956	0.35	$103,853	$96,232	$148,197
Sodium Sulfate Puification																	
Evaporator feed tank	insulated, 6460 gal				$45,966	2011	$45,966	strm.A600.23	290932	kg/hr	0.60	2.50	2379966	8.18	$162,220	150033.3809	$375,083
Evaporator feed heater	shell and tube 1/2 pass				$274,818	2011	$274,818	heat.A600.31	-13	MMkcal/h	0.60	3.00	-8	0.57	$195,585	180891.736	$542,675
Evaporator flash drum	23' x 48' - 110,000 gal.		SS316	1	$511,000	2009	$511,000	strm.A600.23	264116	kg/hr	0.70	2.00	2379966	9.01	$2,381,026	2471358.137	$4,942,716
Centrifuge	Nexant quote sodium sulfate, 25410 lb/hr solids basis			1	$327,680	2011	$327,680	strm.A600.NA2SO4	11524	kg/hr	0.60	2.3	14458	1.25	$375,462	$347,256	$798,689
Dryer	Nexant quote Sodium sulfate, 25410 lb/hr solids basis			1	$555,008	2011	$555,008	strm.A600.PRD-SALT	11524	kg/hr	0.60	2.6	13770	1.19	$617,592	$571,197	$1,485,111
														Area 600 Totals	**$17,514,146**	**$16,491,606**	**$31,275,951**

A700: Lignin Utilization	Mechanical Equipment List					Scaled Installed Costs											
EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
A701: Lignin Conditioning																	
Neutralization Tank	2.6 atm, 130C operating30 min. hold = 30,000 gal		SS317	1	$236,000	2009	$236,000	strm.A700.A701.LIQUID1	410369	kg/hr	0.70	2.0	108129	0.26	$92,779	$96,299	$192,598
Pulping Reactor Tank	2.6 atm, 130C operating (up to 160C, 30 min)30 min. hold = 30,000 gal		SS316	1	$16,300,000	2013	$16,300,000	strm.A700.A701.PULP-OUT	323295	kg/hr	0.60	1.7	140578	0.43	$9,889,598	$9,443,319	$15,581,477
Flash/drain tank			SS317	1	$262,000	2013	$262,000	strm.A700.A701.PULP-OUT	323296	kg/hr	0.70	2.0	140578	0.43	$146,259	$139,659	$279,318
Black Liquor Storage Tank	1,200,000 gallon		316SS	1	$1,317,325	2011	$1,317,325	strm.A700.A701.5	328984	kg/hr	0.70	1.8	92878	0.28	$543,513	$502,682	$904,828
A702: Muconate Fermentation																	
1st Aerobic Seed	80 gallon skid complete - $46,000 ea		304SS		$46,000	2009	$46,000	NSD1000	1	ea	1.00	1.80	4	4.00	$184,000	$190,981	$343,765
1st Seed Vessel Agitator		0.75hp	SS316		$3,420	2009	$3,420	NSD1000	1	ea	1.00	1.50	4	4.00	$13,680	$14,199	$21,298
2nd Aerobic Seed	800 gallon skid complete - $57,500 ea		304SS		$57,500	2009	$57,500	NSD1000	1	ea	1.00	1.80	4	4.00	$230,000	$238,726	$429,706
2nd Seed Vessel Agitator		8 hp	SS316		$11,000	2009	$11,000	NSD1000	1	ea	1.00	1.50	4	4.00	$44,000	$45,669	$68,504
Bubble column seed fermenter	100 m3		316SS		$274,100	2014	$274,100	NSD1000	1	ea	1.00	2.30	4	4.00	$1,096,400	$1,030,932	$2,371,143
Seed circulation cooler	650 sqft		316SS		$8,400	2014	$8,400	NSD1000	1	ea	1.00	2.20	4	4.00	$33,600	$31,594	$69,506
Bubble column production fermenter	1000 m3		316SS		$1,691,400	2014	$1,691,400	NVES1000	1	ea	1.00	2.30	19	19.00	$32,136,600	$30,217,664	$69,500,627
Production circulation cooler	4500 sqft		316SS		$48,100	2014	$48,100	NVES1000	1	ea	1.00	2.20	19	19.00	$913,900	$859,329	$1,890,525
Production circulation pump	400 gpm		316SS		$11,500	2014	$11,500	NVES1000	1	ea	1.00	2.30	19	19.00	$218,500	$205,453	$472,542
Fermentation air compressor	25,000 ACFM @ 45psig; max size in ACCE		CS		$1,318,600	2014	$1,318,600	AIRV1000	13	m^3/s	1.00	1.60	12	0.88	$1,155,394	$1,086,403	$1,738,246
Fermentation air receiver	25,000 gal		CS		$104,600	2014	$104,600	AIRV1000	13	m^3/s	1.00	2.00	12	0.88	$91,653	$86,181	$172,361
Fermentation Surge tank	insulated cone bottom, 6460 gal				$45,966	2011	$45,966	strm.A700.A702.UF-FD	290932	kg/hr	0.60	2.50	187319	0.64	$35,295	$32,643	$81,608
Ultrafiltration membrane separator					$2,048,000	2011	$2,048,000	Volume Flow	1303	GPM	0.60	2.50	813	1.00	$2,048,000	$1,894,146	$4,735,366
membrane broth feed pump					INCLUDED			strm.A700.A702.UF-FD		kg/hr			187319				
membrane solvent feed pump					INCLUDED			den.A700.A702.UF-FD		gm/cc			1.0				
A703: Recovery and Upgrading																	
Carbon Filter	2 Vessels, for color removal				$345,234	2011	$345,234	Volume Flow	1347	GPM	0.60	2.50	748	1.00	$345,234	$319,299	$798,247
Initial carbon loading					see CATALYST	2011		strm.A700.A703.CFIL-FD		kg/hr			172423				
								den.A700.A703.CFIL-FD		gm/cc			1.0				
CCM Crystallizer	Oslo Type. 2 In series		316SS	2 series	$7,104,192	2011	$7,104,192	Volume Flow	190	GPM	0.60	2.50	45.396	0.24	$3,009,338	$2,783,265	$6,958,162
								strm.A700.A703.CRY1-PRD		kg/hr	0.60	2.50	12245				
								den.A700.A703.CRY1-PRD		g/cc	0.60	2.50	1.2				
CCM Centrifuge	Centrifuge Separator				$327,680	2011	$327,680	strm.A700.A703.CRY1-PRD	13403	kg/hr	0.60	2.30	12245	0.91	$310,384	$287,067	$660,254
CCM Drier	Fluidized bed drier parallel			2 parallel	$555,008	2011	$555,008	strm.A700.A703.DRY1-PRD	11526	kg/hr	0.60	2.60	11640	1.01	$558,302	$516,361	$1,342,537
Dissolution Tank	mixing tank to redissolve crystals in solvent (etOH)				$1,317,325	2011	$1,317,325	strm.A700.A703.FIL2-FD	328984	kg/hr	0.70	1.80	57591	0.18	$388,972	$359,751	$647,551
Dissolution Tank agitator	pump to retain crystal suspension	80	316SS		$63,000	2009	$63,000	work.A700.A703.W-ETOHMX	60	kW	1.00	1.50	60	1.00	$63,000	$65,390	$98,085
Filtration Centrifuge(salt removal)	removes precipitated solids after dissolution				$327,680	2011	$327,680	strm.A700.A703.FIL2-SLT	13403	kg/hr	0.60	2.30	159	0.01	$22,923	$21,201	$48,762
HDO feed tank	insulated, 6460 gal				$45,966	2011	$45,966	strm.A700.A703.FIL2-PRD	290932	kg/hr	0.60	2.50	57432	0.20	$17,364	$16,060	$40,149
HDO reactor pump					$802,861	2014	$802,861	strm.A700.A703.FIL2-PRD	208720	kg/hr	0.80	1.40	57432	0.28	$285,963	$268,888	$376,443
HDO Feed Effluent econimizer	2-4 TEMA shell and tube HX		316SS		$353,600	2011	$353,600	heat.A700.A703.QX-HDO	14	Mmkcal/hr	0.70	2.66	3	0.20	$113,073	$104,578	$277,742
HDO trim preheater			304SS		$41,000	2009	$41,000	heat.A700.A703.QH-TRIM	-2	MMkcal/hr	0.70	2.20	0.0	0.00	$0	$0	$0
HDO Fixed Bed Reactor	(Q3 FY17 milestone), base PF=2.5,208 BBL/hr				$4,168,568	2011	$4,168,568	Volume Flow (liquid)	32895	L/hr	0.70	2.00	71584	2.18	$7,184,024	$6,644,333	$13,288,666
Pressure Factor (via Guthrie)	(>1000PSIG=2.5,900=2.3,800=1.9,700=1.8,600=1.6,500=1.45,400=1.35)		1.6					strm.A700.A703.RXR-FD	29274	kg/hr			57432				
Internals								den.A700.A703.RXR-FD	1	GM/CC			0.8				
Hydrogenation Intercooler (bed1)					$2,353,181	2007	$2,353,181	heat.A700.A703.QC-BED1	32	MMkcal/hr	0.65	2.21	2	0.06	$392,088	$404,252	$893,398
Hydrogenation Intercooler (bed2)					$2,353,181	2007	$2,353,181	heat.A700.A703.QC-BED2	32	MMkcal/hr	0.65	2.21	3	0.10	$530,457	$546,913	$1,208,679
H2 Makeup Compressor	recipricating compressor(5 stages)				$1,621,200	2011	$1,621,200	strm.A700.A703.H2-MU	390	kg/hr	0.60	1.09	426	1.09	$1,709,652	$1,581,216	$1,723,323
H2 Makeup Compressor spare	recipricating compressor(5 stages)				$1,621,200	2011	$1,621,200	strm.A700.A703.H2-MU	390	kg/hr	0.60	1.08	426	1.09	$1,709,652	$1,581,216	$1,709,668
HHPS	Via Adipic model(via MB)				$436,000	2013	$436,000	strm.A700.A703.HHPS-FD	119841	kg/hr	1.00	1.50	57857	0.48	$210,494	$200,995	$301,493
HDO hot gas cooler					$321,600	2011	$321,600	heat.A700.A703.QAC-2	4	Mmkcal/h	0.70	1.66	0.0	0.01	$10,949	$10,127	$16,780
CHPS	3-Phase horizontal sep., demister, 3/16 SS316 cladding				$328,500	2011	$328,500	Volume Flow	39911	L/hr	0.70	2.59	96	0.00	$4,835	$4,472	$11,570
								strm.A700.A703.CHPS-LIQ		kg/hr			76				
								den.A700.A703.CHPS-LIQ		GM/CC			0.8				
AA evaporator feed tank	insulated, 6460 gal				$45,966	2011	$45,966	strm.A700.A703.EVAP-FD	290932	kg/hr	0.60	2.50	101631	0.35	$24,455	$22,618	$56,546
AA evaporator feed heater	shell and tube 1/2 pass				$274,818	2011	$274,818	heat.A700.A703.QH-EVAP	-13	MMkcal/h	0.60	3.00	-2	0.13	$81,004	$74,919	$224,756
AA evaporator flash drum	23' x 48' - 110,000 gal.		SS316	1	$511,000	2009	$511,000	strm.A700.A703.EVAP-FD	264116	kg/hr	0.70	2.00	101631	0.38	$261,867	$271,802	$543,605
AA condensor drum					$487,000	2010	$487,000	heat.A700.A703.QC-COND	23	MMkcal/h	0.60	2.80	9	0.41	$284,053	$279,360	$782,209
AA Crystallizer	Oslo Type. 2 In series		316SS	2 series	$7,104,192	2011	$7,104,192	Volume Flow	190	GPM	0.60	2.50	46.641	0.25	$3,058,619	$2,828,844	$7,072,110
								strm.A700.A703.CRY2-PRD		kg/hr	0.60	2.50	12434				
								den.A700.A703.CRY2-PRD		g/cc	0.60	2.50	1.2				
AA Centrifuge separator	Centrifuge Separator				$327,680	2011	$327,680	strm.A700.A703.CRY2-PRD	13403	kg/hr	0.60	2.30	12434	0.93	$313,251	$289,719	$666,353
AA Drier	Fluidized bed drier parallel			2 parallel	$555,008	2011	$555,008	strm.A700.A703.DRY2-PRD	11526	kg/hr	0.60	2.60	11843	1.03	$564,125	$521,746	$1,356,539
														Area 700 Totals	**$70,327,252**	**$66,120,273**	**$139,957,048**

A800: CHP	Mechanical Equipment List					Scaled Installed Costs											
EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
High Solids Burner and Turbine																	
Burner Combustion Air Preheater	INCLUDED			1	INCLUDED												
BFW Preheater	INCLUDED			1	INCLUDED												
Pretreatment/BFW heat recovery	9.4 MM Btu/hr		SS304	1	$41,000	2009	$41,000	heat.A800.A810.QH812	-2	Gcal/hr	0.70	2.2	-2	0.72	$32,711	$33,952	$74,694
Air Intake Fan	INCLUDED				INCLUDED												
Boiler	525,000 lb/hr @ 900 psig		CS	1	$28,550,000	2010	$28,550,000	strm.A800.A810.813c	238203	kg/hr	0.60	1.8	230378	0.97	$27,983,543	$27,521,215	$49,538,187
Combustion Gas Baghouse	Baghouse, Spray dryer scrubber, flues/ducting			1	$11,000,000	2013	$0	strm.A800.A810.812	238203	kg/hr	0.60	1.8	157987	0.66	$0	$0	$0
Turbine/Generator	23.6 kW, 2 extractions			1	$9,500,000	2010	$9,500,000	work.A900.wtotal	-42200	kW	0.60	1.8	-16528	0.39	$5,413,441	$5,324,003	$9,583,205
Hot Process Water Softener System				1	$78,000	2010	$78,000	strm.A800.A810.812	235803	kg/hr	0.60	1.8	157987	0.67	$61,339	$60,326	$108,586
Amine Addition Pkg.				1	$40,000	2010	$40,000	strm.A800.A810.812	235803	kg/hr	0.00	1.8	157987	0.67	$40,000	$39,339	$70,810
Ammonia Addition Pkg				1	INCLUDED												
Phosphate Addition Pkg.				1	INCLUDED												
Condensate Pump			SS316	2	INCLUDED												
Turbine Condensate Pump			SS304	2	INCLUDED												
Deaerator Feed Pump			SS304	2	INCLUDED												
BFW Pump			SS316	5	INCLUDED												
Blowdown Pump			CS	2	INCLUDED												
Amine Transfer Pump			CS	1	INCLUDED												
Condensate Collection Tank			A285C	1	INCLUDED												
Condensate Surge Drum			SS304	1	INCLUDED												
Deaerator	Tray type		CS;SS316	1	$305,000	2010	$305,000	strm.A800.A810.812	235803	kg/hr	0.60	3.0	157987	0.67	$239,852	$235,889	$707,668
Blowdown Flash Drum			CS	1	INCLUDED												
Amine Drum			SS316	1	INCLUDED												
														Area 800 Totals	**$33,770,886**	**$33,214,724**	**$60,083,151**

A900: Utilities & Storage	Mechanical Equipment List					Scaled Installed Costs											
EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
Utilities System																	
Cooling Tower System	44,200 gpm	750 hp	FIBERGLASS	1	$1,375,000	2010	$1,375,000	strm.a900.945	10037820	kg/hr	0.60	1.5	23291036	2.32	$2,278,412	$2,240,770	$3,361,155
Plant Air Compressor	400 SCFM@125 psig	150 hp		1	$28,000	2010	$28,000	DRY101	83333	kg/hr	0.60	1.6	83333	1.00	$28,000	$27,537	$44,060
Chilled Water Package	2 x 2350 tons (14.2 MM kcal/hr)	3400 hp		1	$1,275,750	2010	$1,275,750	heat.a900.qchwop	14	Gcal/hr	0.60	1.6	49	3.48	$2,693,705	$2,649,201	$4,238,722
CIP System	100,000 GAL		SS304/SS316	1	$421,000	2009	$421,000	strm.a900.914	63	kg/hr	0.60	1.8	145	2.30	$694,222	$720,560	$1,297,008
Cooling Water Pump	16,120 GPM, 100 FT TDH SIZE 20X20-28	500.0	CS	3	$283,671	2010	$283,671	strm.a900.945	10982556	kg/hr	0.80	3.1	23291036	2.12	$517,611	$509,059	$1,578,083
Make-up Water Pump	685 GPM, 75 FT TDH SIZE 6X4-13	20.0	CS	1	$6,864	2010	$6,864	strm.a900.904	155564	kg/hr	0.80	3.1	529217	3.40	$18,279	$17,977	$55,729
Process Water Circulating Pump	2285 GPM, 75 FT TDH SIZE 8X6-13	75.0	CS	1	$15,292	2010	$15,292	strm.a900.905	518924	kg/hr	0.80	3.1	803985	1.55	$21,706	$21,347	$66,177
Instrument Air Dryer	670 SCFM - CYCLING TYPE		CS	1	$15,000	2009	$15,000	DRY101	83333	kg/hr	0.60	1.8	83333	1.00	$15,000	$15,569	$28,024
Plant Air Receiver	3800 gal - 72" x 228" vertical		CS	1	$16,000	2009	$16,000	DRY101	83333	kg/hr	0.60	3.1	83333	1.00	$16,000	$16,607	$51,482
Process Water Tank No. 1	250,000 gal		CS	1	$250,000	2009	$250,000	strm.a900.905	451555	kg/hr	0.70	1.7	803985	1.78	$374,383	$388,586	$660,597
Storage																	
Ammonia Storage Tank	28,000 gal		SA- 516-70	2	$196,000	2010	$196,000	strm.A900.NH3-NET	1171	kg/hr	0.70	2.0	994	0.85	$174,699	$171,813	$343,626
Dioxane Storage Tank	28,000 gal		SA- 516-70	2	$196,000	2010	$196,000	strm.A500.FURANS.DIOXANE	1171	kg/hr	0.70	2.0	209	0.18	$58,617	$57,649	$115,297
CSL Storage Tank	70,000 gal		Glass lined	1	$70,000	2009	$70,000	strm.A900.CSL-NET	1393	kg/hr	0.70	2.6	234	0.17	$20,084	$20,846	$54,198
CSL Storage Tank Agitator		10 hp	SS304	1	$21,200	2009	$21,200	strm.A900.CSL-NET	1393	kg/hr	0.50	1.5	234	0.17	$8,690	$9,019	$13,529
CSL Pump	8 GPM, 80 FT TDH	0.5	CS	1	$3,000	2009	$3,000	strm.A900.CSL-NET	1393	kg/hr	0.80	3.1	234	0.17	$720	$747	$2,317
DAP Bulk Bag Unloader	Super sack unloader			1	$30,000	2009	$30,000	strm.A900.DAP-NET	163	kg/hr	0.60	1.7	616	3.78	$66,603	$69,130	$117,521
DAP Bulk Bag Holder	Super sack holder			1	INCLUDED												
DAP Make-up Tank	12,800 gal		SS304	1	$102,000	2009	$102,000	strm.A900.DAP-NET	1615	kg/hr	0.70	1.8	616	0.38	$51,941	$53,912	$97,042
DAP Make-up Tank Agitator		5.5 hp	SS304	1	$9,800	2009	$9,800	strm.A900.DAP-NET	163	kg/hr	0.50	1.5	616	3.78	$19,049	$19,772	$29,657
DAP Pump	2 GPM, 100 FT TDH	0.5	CS	1	$3,000	2009	$3,000	strm.A900.DAP-NET	163	kg/hr	0.80	3.1	616	3.78	$8,689	$9,018	$27,957
Sulfuric Acid Pump	5 GPM, 150 FT TDH SIZE 2X1-10	0.5	SS316	1	$7,493	2010	$7,493	strm.A900.ACID-NET	1981	kg/hr	0.80	2.3	11560	5.84	$30,727	$30,220	$69,505
Sulfuric Acid Storage Tank	12,600 gal, 12' dia x15' H		SS	1	$96,000	2010	$96,000	strm.A900.ACID-NET	1981	kg/hr	0.70	1.5	11560	5.84	$330,012	$324,560	$486,840
Caustic Storage Tank	12,600 gal, 12' dia x15' H		SS	1	$96,000	2011	$96,000	strm.A900.BASE-NET	1981	kg/hr	0.70	1.5	10271	5.18	$303,801	$280,978	$421,467
AlCl3 Storage Tank	12,600 gal, 12' dia x15' H		SS	1	$96,000	2011	$96,000	STRM.A500.FURANS.ALCL3	1981	kg/hr	0.70	1.5	949	0.48	$57,330	$53,023	$79,534
Firewater Storage Tank	600,000 gal - 4 hrs @ 2500 gpm		Glass lined	1	$803,000	2009	$803,000	strm.A900.H2O-FIRE	8343	kg/hr	0.70	1.7	8343	1.00	$803,000	$833,464	$1,416,890
Firewater Pump	2500 GPM, 150 FT TDH	125.0	CS	1	$15,000	2009	$15,000	strm.A900.H2O-FIRE	8343	kg/hr	0.80	3.1	8343	1.00	$15,000	$15,569	$48,264
Diesel storage tank	750,000 gal., 7 day storage, Floating roof		A285C	1	$670,000	2009	$670,000	strm.PRD-500	11341	kg/hr	0.70	1.7	28139	2.48	$1,265,750	$1,313,770	$2,233,409
MEK storage tank	750,000 gal., 7 day storage, Floating roof		A285C	1	$670,000	2009	$670,000	CMIX.MEK.A500.FURANS.MEK	11341	kg/hr	0.70	1.7	9301	0.82	$583,183	$605,308	$1,029,024
Co-Product Storage Tank (Adipic)				1	$690,900	2007	$690,900	strm.PRD-700	23322.902	kg/hr	0.65	1.850	11843	0.507792502	$444,745	$458,543	$848,305
Co-Product Storage Tank (Sodium Sulfate)				1	$690,900	2007	$690,900	strm.PRD-600	23322.902	kg/hr	0.65	1.850	13770	0.590406709	$490,528	$505,746	$935,630
Glucose Storage Tank	70,000 gal		Glass lined	1	$70,000	2009	$70,000	strm.a400.401	1393	kg/hr	0.70	2.6	1557	1.12	$75,686	$78,557	$204,249
														Area 900 Totals	**$11,466,172**	**$11,518,859**	**$19,955,298**

Integrated Biorefinery

A200: Pretreatment	Mechanical Equipment List					Scaled Installed Costs											
EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
Deacetylation Tank Discharge Pump	1771 GPM, 150 FT TDH	100	316SS	1	$22,500	2009	$22,500	strm.a200.211a	402194	kg/hr	0.80	2.3	283668	0.71	$17,017	$17,663	$40,624
Deacetylation reactor conveyors	Feed and discharge drag conveyors	40 hp	SS316	3	$110,000	2013	$110,000	strm.a200.211a	277167	kg/hr	0.80	1.7	283668	1.02	$336,178	$321,008	$545,713
Pretreatment Sugar Beet Extruder				1	$5,424,000	2010	$5,424,000	SCIS.a200.DEAC-IN	38600	kg/hr	1.00	1.0	70483	1.83	$9,904,187	$9,740,555	$9,740,555
Flash Tank Agitator	Side-mounted, 3 x 75 hp. (170 kW)	170 kW	316LSS	3	$90,000	2009	$90,000	strm.a200.254	252891	kg/hr	0.50	1.5	192174	0.76	$78,455	$81,432	$122,148
Ammonia Addition Tank Agitator		10 hp	SS	1	$21,900	2009	$21,900	strm.a200.228	410369	kg/hr	0.50	1.5	230674	0.56	$16,419	$17,042	$25,563
Ammonia Static Mixer			SS	1	$5,000	2009	$5,000	strm.a200.275	157478	kg/hr	0.50	1.0	38501	0.24	$2,472	$2,566	$2,566
Pretreatment Water Heater	29.9 MMBtu		304SS	1	$92,000	2010	$92,000	Heat.A200.QH201	-8	Gcal/hr	0.70	2.2	-0.3	0.04	$9,867	$9,704	$21,348
Milling Equipment	200kw/dry ton			8	$2,466,700	2013	$2,466,700	SCIS.a200.211b	62942	kg/hr	0.60	1.5	57306	0.91	$19,733,600	$18,843,101	$28,264,651
Milling Equipment-Szego Mill				11	$578,000	2013	$578,000	SCIS.a200.211b	62942	kg/hr	0.60	1.4	57306	0.91	$6,358,000	$6,071,089	$8,499,524
Blowdown Tank Discharge Pump	1900 GPM, 150 FT TDH	125	316SS	1	$25,635	2010	$25,635	strm.a200.222	292407	kg/hr	0.80	2.3	193160	0.66	$18,398	$18,094	$41,617
Flash Tank Discharge Pump	900 GPM, 150 FT TDH	75	316SS	1	$30,000	2009	$30,000	strm.a200.254	204390	kg/hr	0.80	2.3	192174	0.94	$28,557	$29,640	$68,172
Hydrolyzate Pump	1771 GPM, 150 FT TDH	100	316SS	1	$22,500	2009	$22,500	strm.a200.228	402194	kg/hr	0.80	2.3	230674	0.57	$14,422	$14,969	$34,430
S/L Split Discharge Pump to WWT	900 GPM, 150 FT TDH	75	316SS	1	$30,000	2009	$30,000	strm.a200.4	204390	kg/hr	0.80	2.3	193160	0.95	$28,674	$29,762	$68,452
Flash Tank	23' x 48' - 110,000 gal.		SS316	1	$511,000	2009	$511,000	strm.a200.223	264116	kg/hr	0.70	2.0	193160	0.73	$410,493	$426,067	$852,133
Ammonia Addition Tank	118,000 gal, 1hr residence time		SS304	1	$236,000	2009	$236,000	strm.a200.228	410369	kg/hr	0.70	2.0	230674	0.56	$157,684	$163,667	$327,334
														Area 200 Totals	$37,114,424	$35,786,358	$48,654,831

A300: Hydrolysis and Fermentation	Mechanical Equipment List					Scaled Installed Costs											
EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
Batch Enzymatic Hydrolysis																	
Hydrolyzate Cooler	Plate & Frame 32.5 MMBtu/hr		SS 304	1	$85,000	2010	$85,000	heat.A300.EH.QC301	8	Gcal/hr	0.70	2.2	7	0.80	$72,519	$71,321	$156,906
Enzyme-Hydrolysate Mixer	inline mixer 1673 gpm	100 hp	SS316	1	$109,000	2009	$0	strm.a300.EH.t310fd	379938	kg/hr	0.50	1.7	2	0.00	$0	$0	$0
Saccharification Tank	250,000 gal each - 19' dia. x 120' tall		304SS	8	$3,840,000	2009	$3,840,000	strm.A300.EH.306	421776	kg/hr	0.70	2.0	238222	0.56	2574312.146	$2,671,977	$5,343,954
Saccharification Transfer Pump	352 GPM, 150 FT TDH	20	316SS	5	$47,200	2009	$47,200	strm.a300.EH.306	421776	kg/hr	0.80	2.3	238222	0.56	$29,886	$31,019	$71,345
Enzymatic Hydrolysis Storage Tank	1,200,000 gallon		316SS	1	$1,317,325	2011	$1,317,325	strm.a300.EH.306A	328984	kg/hr	0.70	1.8	238222	0.72	$1,050,891	$971,944	$1,749,500
Fermentor Tank (saccharification contribution)			304SS	12	$10,128,000	2009	$10,128,000		12	ea	1	1.50	6.9	0.58	$5,843,797	$6,065,501	$9,098,252
Fermentor Agitator (saccharification contirbution)		30 hp	SS304	1	$52,500	2009	$52,500		1	ea	1	1.50	6.9	6.92	$363,506	$377,297	$565,946
Fermentation Cooler (saccharification contribution)	Plate & frame		304SS	12	$86,928	2009	$86,928		12	ea	1	2.20	6.9	0.58	$50,157	$52,060	$114,532
Fermentation Recirc/Transfer Pump (saccharification contribution)	340 GPM, 150 FT	20	316SS	5	$47,200	2009	$47,200		12	ea	0.8	2.30	6.9	0.58	$30,400	$31,554	$72,574
Vacuum filter press (VFP)	(4) 170 m2 Horizontal Belt Filters	660 hp ea	304SS	4	$2,152,500	2013					1	1.7		1.00	$8,610,000	$8,221,465	$13,976,490
Sugar Concentration		3600 kW	304SS	1	$6,370,000	2013	$6,370,000	STRM.A500.FURANS.SUG-CONC.301SUG	244084	kg/hr	0.70	2.0	204062	0.84	$5,619,465	$5,365,881	$10,731,762
Concentrated Sugar Storage Tank	5,500 gallons - 20 min residence time	20 hp	SS	1	$168,000	2011	$168,000	STRM.A500.FURANS.S4	76712	kg/hr	0.70	1.8	49053	0.64	$122,849	$113,620	$204,516
														Area 300 Totals	$24,367,783	$23,973,640	$42,085,776

A400: Enzyme Production	Mechanical Equipment List					Scaled Installed Costs											
EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
Cellulase Fermentor Agitators		800.0	SS316		$580,000	2009	$580,000	CLVESSEL	1	ea	1.00	1.5	5	5.00	$2,900,000	$3,010,021	$4,515,032
Cellulase Fermentor Agitators		0.75hp	SS316		$3,420	2009	$3,420	ICLSEED	1	ea	1.00	1.5	4	4.00	$13,680	$14,199	$21,298
Cellulase Fermentor Agitators		8 hp	SS316		$11,000	2009	$11,000	ICLSEED	1	ea	1.00	1.5	4	4.00	$44,000	$45,669	$68,504
Cellulase Fermentor Agitators		80 hp	SS316		$63,000	2009	$63,000	ICLSEED	1	ea	1.00	1.5	4	4.00	$252,000	$261,560	$392,341
Media-Prep Tank Agitator		7.5 hp	A285C	1	$8,500	2009	$8,500	strm.a400.402a	12255	kg/hr	0.50	1.5	0	0.00	$0	$0	$0
Cellulase Nutrient Mix Tank Agitator		3 hp	CS	1	$4,800	2009	$4,800	strm.a400.416	174	kg/hr	0.50	1.6	123	0.70	$4,030	$4,182	$6,692
Cellulase Hold Tank Agitator		10 hp	SS316	1	$26,900	2009	$26,900	strm.422	10930	kg/hr	0.50	1.5	7575	0.69	$22,395	$23,244	$34,866
Cellulase Fermentor	80,000 gal, 1 atm, 28 °C, Internal coil		SS316		$400,500	2009	$400,500	CLVESSEL	1	ea	1.00	2.0	5	5.00	$2,002,500	$2,078,471	$4,156,943
1st Cellulase Seed Fermentor	80 gallon skid complete - $46,000 ea		304SS		$46,000	2009	$46,000	ICLSEED	1	ea	1.00	1.8	4	4.00	$184,000	$190,981	$343,765
2nd Cellulase Seed Fermentor	800 gallon skid complete - $57,500 ea		304SS		$57,500	2009	$57,500	ICLSEED	1	ea	1.00	1.8	4	4.00	$230,000	$238,726	$429,706
3rd Cellulase Seed Fermentor	8,000 gallon skid complete - $95,400 ea		304SS		$95,400	2009	$95,400	ICLSEED	1	ea	1.00	1.8	4	4.00	$381,600	$396,077	$712,939
Cellulase Fermentation Cooler	Cooling coil included with Cellulase Fermenter		304SS		INCLUDED												
Media Prep Tank Cooler	Cooling coil included with Media Prep Tank		304SS	1	INCLUDED												
Fermenter Air Compressor Package	8000 SCFM @ 16 psig		CS	2	$350,000	2009	$350,000	strm.a400.450	33168	kg/hr	0.60	1.6	17839	0.54	$241,242	$250,395	$400,632
Cellulase Transfer Pump	59 gpm, 100 FT, TDH SIZE 2X1-10C	3	316SS	1	$7,357	2010	$7,357	strm.a400.420	13399	kg/hr	0.80	2.3	7575	0.57	$4,662	$4,585	$10,545
Cellulase Seed Pump	3 GPM, 100 FT TDH SIZE 2X1-10	2	316SS	4	$29,972	2010	$29,972	strm.a400.409	681	kg/hr	0.80	2.3	421	0.62	$20,408	$20,071	$46,163
Media Pump	63 GPM, 100 FT TDH SIZE 2X1-10C	3	316SS	1	$7,357	2010	$7,357	strm.a400.402a	14307	kg/hr	0.80	2.3	0	0.00	$0	$0	$0
Cellulase Nutrient Transfer Pump	Gear Pump 2 GPM, 100 FT	1	316SS	1	$1,500	2009	$1,500	strm.a400.416	454	kg/hr	0.80	2.3	123	0.27	$526	$546	$1,257
Cellulase Feed Pump	Gear Pump	1	316SS	1	$5,700	2009	$5,700	strm.a400.422	18168	kg/hr	0.80	2.3	7575	0.42	$2,831	$2,938	$6,758
Anti-foam Pump	Gear Pump 2 GPM, 100 FT	1	316SS	1	$1,500	2009	$1,500	strm.a400.444	11	kg/hr	0.80	2.3	7.3	0.69	$1,115	$1,157	$2,661
SO2 Storage Tank	1 ton cylinders, incl w/ delivery		SS304	1	$0												
Media-Prep Tank	20,000 gallon, incl. coil		304SS	1	$176,000	2009	$176,000	strm.a400.402a	12255	kg/hr	0.70	1.8	0	0.00	$0	$0	$0
Cellulase Nutrient Mix Tank	HDPE, 8,000 gal		HDPE	1	$9,000	2010	$9,000	strm.a400.416	224	kg/hr	0.70	3.0	123	0.55	$5,903	$5,806	$17,417
Cellulase Hold Tank	80,000 gal		304SS	1	$248,070	2009	$248,070	strm.a400.422	10930	kg/hr	0.70	1.8	7575	0.69	$191,920	$199,201	$358,562
														Area 400 Totals	$6,502,812	$6,747,831	$11,526,082

A550: MEK Production	Mechanical Equipment List					Scaled Installed Costs											
EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
BDO - Fermentation																	
Fermentor Feed Cooler	Plate & frame		SS304	1	$86,928	2009	$86,928	heat.A500.BDO2MEK.FERM.QC310	5.404	Gcal/hr	0.7	1.80	1.6	0.30	$37,554	$38,979	$70,162
Seed Hold Tank Agitator		15 hp	SS304	1	$31,800	2009	$31,800	strm.A500.BDO2MEK.FERM.304	40414	kg/hr	0.5	1.50	10434.4	0.26	$16,158	$16,771	$25,157
4th Seed Vessel Agitator		7.5 hp	SS	2	$26,000	2009	$26,000		2	ea	0.5	1.50	2.0	1.00	$26,000	$26,986	$40,480
5th Seed Vessel Agitator		10 hp	SS	2	$43,000	2009	$43,000		2	ea	0.5	1.50	2.0	1.00	$43,000	$44,631	$66,947
Beer Surge Tank Agitator		20 hp	SS304	2	$68,300	2009	$68,300	strm.A500.BDO2MEK.FERM.PRD	425878	kg/hr	0.5	1.50	84186.9	0.20	$30,367	$31,519	$47,278
1st Seed Fermentor	20 gallon skid complete - $37,700 ea		304SS	2	$75,400	2009	$75,400		2	ea	0.7	1.80	2.0	1.00	$75,400	$78,261	$140,869
2nd Seed Fermentor	200 gallon skid complete - $58,300 ea		304SS	2	$116,600	2009	$116,600		2	ea	0.7	1.80	2.0	1.00	$116,600	$121,024	$217,842
3rd Seed Fermentor	2000 gallon skid complete - $78,800 ea		304SS	2	$157,600	2009	$157,600		2	ea	0.7	1.80	2.0	1.00	$157,600	$163,579	$294,442
4th Seed Fermentor	20,000 gallon, incl. coil - $176,000 ea		304SS	2	$352,000	2009	$352,000		2	ea	0.7	2.00	2.0	1.00	$352,000	$365,354	$730,709
4th Seed Fermentor Coil	incl. w/ tank		304SS	1	INCLUDED												
5th Seed Fermentor	200,000 gallon, incl. coil - $590,000 ea		304SS	2	$1,180,000	2009	$1,180,000		2	ea	0.7	2.00	2.0	1.00	$1,180,000	$1,224,767	$2,449,534
5th Seed Fermentor Coil	incl. w/ tank		304SS	1	INCLUDED												
Seed Hold Transfer Pump	190 GPM, 150 FT TDH	10	316SS	1	$8,200	2009	$8,200	strm.A500.BDO2MEK.FERM.304	43149	kg/hr	0.8	2.30	10434.4	0.24	$2,634	$2,734	$6,288
Seed Hold Tank	300,000 gallon		316SS	1	$439,000	2009	$439,000	strm.A500.BDO2MEK.FERM.304	40414	kg/hr	0.7	1.80	10434.4	0.26	$170,145	$176,600	$317,881
Seed Transfer Pump	190 GPM, 615 FT TDH	40	316SS	2	$24,300	2009	$24,300	strm.A500.BDO2MEK.FERM.304	43149	kg/hr	0.8	2.30	10434.4	0.24	$7,806	$8,102	$18,634
Bioreactor Seed Blowers	750 SCFM @ 11psig	61 hp	CS	1	$19,000	2013	$19,000		1	ea	0.6	1.60	2.0	2.00	$28,799	$27,499	$43,998
Fermentor Tank (Anaerobic)			304SS	12	$10,128,000	2009	$10,128,000		12	ea	1	1.50	0.9	0.08	$779,856	$809,442	$1,214,164
Fermentor Agitator		30 hp	SS304	1	$52,500	2009	$52,500		1	ea	1	1.50	0.9	0.92	$48,510	$50,350	$75,526
Fermentation Cooler	Plate & frame		304SS	12	$86,928	2009	$86,928		12	ea	1	2.20	0.9	0.08	$6,693	$6,947	$15,284
Fermentation Recirc/Transfer Pump	340 GPM, 150 FT	20	316SS	5	$47,200	2009	$47,200		12	ea	0.8	2.30	0.9	0.08	$6,069	$6,300	$14,489
Bioreactor Transfer Pump	2152 GPM, 171 FT TDH	125	316SS	1	$26,800	2009	$26,800	strm.A500.BDO2MEK.FERM.PRD	488719	kg/hr	0.8	2.30	84186.9	0.17	$6,563	$6,812	$15,667
Bioreactor Storage Tank	1,200,000 gallon		316SS	1	$1,317,325	2011	$1,317,325	strm.A500.BDO2MEK.FERM.PRD	156789	kg/hr	0.7	1.80	84186.9	0.54	$852,400	$788,364	$1,419,056
PSA				1	$975,000	2013	$975,000	strm.A500.BDO2MEK.FERM.FERM-VNT	13528	kg/hr	0.6	1.90	9074.0	0.67	$767,264	$732,640	$1,392,017
BDO - SLS (whole-slurry lignin press)																	
Filtrate Tank Agitator		7.5 hp	SS	1	$26,000	2009	$26,000	strm.A500.BDO2MEK.FERM.SLS.572	337439	kg/hr	0.5	1.50	84186.9	0.25	$12,987	$13,479	$20,219
Lignin Wet Cake Conveyor	Belt 100 ft. long x 24" wide, enclosed	10	SS304	1	$70,000	2009	$70,000	strm.A500.BDO2MEK.FERM.SLS.571	28630	kg/hr	0.8	1.70	10865.8	0.38	$32,247	$33,471	$56,900
Lignin Wet Cake Screw	Screw conveyor - 25 ft lg x 14" dia	15	SS304	1	$20,000	2009	$20,000	strm.A500.BDO2MEK.FERM.SLS.571	28630	kg/hr	0.8	1.70	10865.8	0.38	$9,213	$9,563	$16,257
Pressure Filter Pressing Compr	460 SCFM, 300 psig	150 hp		1	$75,200	2009	$75,200	strm.A500.BDO2MEK.FERM.SLS.SQAIRIN	808	kg/hr	0.6	1.60	196.4	0.24	$32,186	$33,407	$53,452
Pressure Filter Drying Compr	4000 SCFM, 130 psig (ea)	700 hp ea.		2	$405,000	2009	$405,000	strm.A500.BDO2MEK.FERM.SLS.557	12233	kg/hr	0.6	1.60	2939.6	0.24	$172,150	$178,681	$285,890
Filtrate Tank Discharge Pump	590 GPM, 100 FT TDH SIZE 4X3-13		SS	1	$13,040	2010	$13,040	strm.A500.BDO2MEK.FERM.SLS.571	31815	kg/hr	0.8	2.30	10865.8	0.34	$5,521	$5,430	$12,489
Feed Pump	1014 GPM 230 FT TDH SIZE 8X6-15	100 hp	SS	1	$18,173	2010	$18,173	strm.A500.BDO2MEK.FERM.SLS.571	31815	kg/hr	0.8	2.30	10865.8	0.34	$7,694	$7,567	$17,405
Manifold Flush Pump		100 hp	SS	1	$17,057	2010	$17,057	strm.A500.BDO2MEK.FERM.SLS.571	31815	kg/hr	0.8	2.30	10865.8	0.34	$7,222	$7,103	$16,336
Cloth Wash Pump		150 hp	SS	1	$29,154	2010	$29,154	strm.A500.BDO2MEK.FERM.SLS.571	31815	kg/hr	0.8	2.30	10865.8	0.34	$12,344	$12,140	$27,921
Filtrate Discharge Pump	590 GPM, 100 FT TDH SIZE 4X3-13	75 hp	SS	1	$13,040	2010	$13,040	strm.A500.BDO2MEK.FERM.SLS.571	31815	kg/hr	0.8	2.30	10865.8	0.34	$5,521	$5,430	$12,489
Pressure Filter	384 sq. m fitration area ea incl packing		SS316	2	$3,294,700	2010	$3,294,700	strm.A500.BDO2MEK.FERM.SLS.571	31815	kg/hr	0.8	1.70	10865.8	0.34	$1,394,955	$1,371,909	$2,332,244
Filtrate Tank	13,750 gal 14' dia x 12' H		SS	1	$103,000	2010	$103,000	strm.A500.BDO2MEK.FERM.SLS.571	31815	kg/hr	0.7	2.00	10865.8	0.34	$48,556	$47,753	$95,507
Feed Tank	20,300 gal 14' dia x 18' H		SS	1	$174,800	2010	$174,800	strm.A500.BDO2MEK.FERM.SLS.571	31815	kg/hr	0.7	2.00	10865.8	0.34	$82,403	$81,042	$162,083
Recycled Water Tank	4000 gal.		HDPE	1	$1,520	2010	$1,520	strm.A500.BDO2MEK.FERM.SLS.571	31815	kg/hr	0.7	3.00	10865.8	0.34	$717	$705	$2,114
Pressing Air Compressor Receiver	1350 gal., 300 psig design		CS	1	$8,000	2010	$8,000	strm.A500.BDO2MEK.FERM.SLS.571	31815	kg/hr	0.7	3.10	10865.8	0.34	$3,771	$3,709	$11,498
Drying Air Compressor Receiver	9,000 gal., 150 psig design		CS	2	$17,000	2010	$17,000	strm.A500.BDO2MEK.FERM.SLS.571	31815	kg/hr	0.7	3.10	10865.8	0.34	$8,014	$7,882	$24,433
BDO -Dehydration and Oligomerization																	
Polishing Filter	Ceramic microfiltration		316L	4	$440,000	2014	$440,000	non water flow	53204	kg/hr	0.90	1.800	10539	0.20	$409,904	$385,428	$693,771
liquid flow								strm.A500.BDO2MEK.AQUEOUS.23BDODUP					84187				
water flow								cmix.H2O.A500.BDO2MEK.AQUEOUS.23BDODUP					73648				
Polished Hydrolysate Storage Tank	8500 gal - 20 min residence time		SS	1	$168,000	2011	$168,000	non water flow	76712	kg/hr	0.70	1.800	10539	0.14	$41,866	$38,721	$69,698
liquid flow								strm.A500.BDO2MEK.AQUEOUS.23BDODUP					84187				
water flow								cmix.H2O.A500.BDO2MEK.AQUEOUS.23BDODUP					73648				
Ion Exchange	Strong acid cation/weak base anion			1	$5,250,000	2014	$5,250,000	non water flow	53204	kg/hr	0.90	1.800	10539	0.20	$1,222,726	$1,149,715	$2,069,486
liquid flow								strm.A500.BDO2MEK.AQUEOUS.23BDODUP					84187				
water flow								cmix.H2O.A500.BDO2MEK.AQUEOUS.23BDODUP					73648				
Deionized Sugar Storage Tank	8500 gal - 20 min residence time		SS	1	$168,000	2011	$168,000	non water flow	76712	kg/hr	0.70	1.800	10539	0.14	$41,866	$38,721	$69,698
liquid flow								strm.A500.BDO2MEK.AQUEOUS.23BDODUP					84187				
water flow								cmix.H2O.A500.BDO2MEK.AQUEOUS.23BDODUP					73648				
Aqueous Upgrading Reactor	BC 2014 design report		317L Clad	1	$2,044,000	2014	$2,044,000	strm.A500.BDO2MEK.AQUEOUS.28	53204	kg/hr	0.37	2.200	83098	1.56	$2,410,631	$2,266,687	$4,986,712
HX BDO Feed-Btm Economizer	29.9 MMBtu		304SS	1	$92,000	2010	$92,000	Heat.A500.BDO2MEK.AQUEOUS.S26	8	Gcal/hr	0.70	2.200	49	6.17	$328,978	$323,543	$711,795
HX BDO Feed Trim Heater	9.4 MM Btu/hr		304SS	1	$92,000	2010	$92,000	Heat.A500.BDO2MEK.AQUEOUS.Q1-OG2	-2.4	Gcal/hr	0.70	2.200	-3	1.25	$107,303	$105,530	$232,166
MEK Recovery																	
1st Column Condenser	ACCE			1	$26,400	2016	$26,400	Heat.A500.BDO2MEK.AQUEOUS.S23	7	Gcal/hr	0.44	2.47	6	0.88	$24,927	$24,927	$61,571
1st Column Condenser (ACC)	ACCE			1	$12,300	2016	$12,300	Heat.A500.BDO2MEK.AQUEOUS.S23	7	Gcal/hr	0.44	2.47	6	0.88	$11,614	$11,614	$28,686
1st Column Reboiler	ACCE			1	$188,400	2016	$188,400	Heat.A500.BDO2MEK.AQUEOUS.S2	-12	Gcal/hr	0.79	2.47	-11	0.92	$176,962	$176,962	$437,096
1st Column Reflux Pump	ACCE			1	$5,500	2016	$5,500	STRM.A500.BDO2MEK.AQUEOUS.C501-D	8790	kg/hr	0.79	2.47	8950	1.02	$5,579	$5,579	$13,779
1st Column Tower	ACCE			1	$120,000	2016	$120,000	STRM.A500.BDO2MEK.AQUEOUS.C501FD	84620	kg/hr	0.68	2.47	83562	0.99	$118,978	$118,978	$293,876
Heat Exchanger				1	$76,800	2015	$76,800	Heat.A500.BDO2MEK.AQUEOUS.S22	5.1827	Gcal/hr	1	3.04296875	3	0.65	$49,593	$48,247.94	$146,816.99
Flash Vessel	23' x 48' - 110,000 gal.		SS316	1	$511,000	2009	$511,000	STRM.A500.BDO2MEK.AQUEOUS.D501-FD	264116	kg/hr	0.7	2	56550.5599	0.21	$173,728	$180,319	$360,638
2nd Column Condenser	ACCE			1	$116,700	2016	$116,700	Heat.A500.BDO2MEK.AQUEOUS.S21	10	Gcal/hr	0.44	2.47	10	1.01	$116,991	$116,991	$288,969
2nd Column Condenser (ACC)	ACCE			1	$22,800	2016	$22,800	Heat.A500.BDO2MEK.AQUEOUS.S21	10	Gcal/hr	0.44	2.47	10	1.01	$22,857	$22,857	$56,457
2nd Column Reboiler	ACCE			1	$71,200	2016	$71,200	Heat.A500.BDO2MEK.AQUEOUS.S17	-13	Gcal/hr	0.79	2.47	-13	1.00	$71,368	$71,368	$176,279
2nd Column Reflux Pump	ACCE			1	$8,900	2016	$8,900	STRM.A500.BDO2MEK.AQUEOUS.S9	16925	kg/hr	0.79	2.47	17027	1.01	$8,942	$8,942	$22,088
2nd Column Tower	ACCE			1	$317,300	2016	$317,300	STRM.A500.BDO2MEK.AQUEOUS.D501-1	55296	kg/hr	0.68	2.47	54935	0.99	$315,892	$315,892	$780,254
3rd Column Condenser	ACCE			1	$158,600	2016	$158,600	Heat.A500.BDO2MEK.AQUEOUS.S20	29	Gcal/hr	0.44	2.47	28	0.98	$157,257	$157,257	$388,426
3rd Column Condenser (ACC)	ACCE			1	$35,300	2016	$35,300	Heat.A500.BDO2MEK.AQUEOUS.S20	29	Gcal/hr	0.44	2.47	28	0.98	$35,001	$35,001	$86,453
3rd Column Reboiler	ACCE			1	$189,700	2016	$189,700	Heat.A500.BDO2MEK.AQUEOUS.S18	-29	Gcal/hr	0.79	2.47	-28	0.98	$186,833	$186,833	$461,478
3rd Column Reflux Pump	ACCE			1	$19,200	2016	$19,200	STRM.A500.BDO2MEK.AQUEOUS.OLIGPROD	6139	kg/hr	0.79	2.47	6180	1.01	$19,301	$19,301	$47,674
3rd Column Tower	ACCE			1	$364,400	2016	$364,400	STRM.A500.BDO2MEK.AQUEOUS.S10	38371	kg/hr	0.68	2.47	37908	0.99	$361,406	$361,406	$892,672
														Area 550 Totals	**$12,967,425**	**$12,717,755**	**$25,140,199**

A500: Furans Production & Upgrading	Mechanical Equipment List						Scaled Installed Costs										
EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
Furans production and upgrading																	
Hydrolysate-Dioxane Pump				1	$95,660	2002	95660	STRM.A500.FURANS.5	349268	lb/hr	0.33	2.47	443963	1.27	$103,541	$141,780	$350,197
Pre-Dehydration Economizer	2-4 TEMA shell and tube HX		316SS	1	$353,600	2011	$353,600	HEAT.A500.FURANS.55	14.3	Mmkcal/hr	0.7	2.65582579	42	2.92	$748,605	$692,367	$1,838,805
Pre-Dehydration Heater				1	$76,800	2015	$76,800	HEAT.A500.FURANS.HP-3	5.1827	Gcal/hr	1	3.04296875	14	2.77	$212,733	$206,964	$629,785
Sugar Dehydration Reactor	Tubular flow reactor, 5 min RT			1	$4,913,234	2013	$4,913,234	STRM.A500.FURANS.5				1.7			$4,853,164	$4,634,160	$7,878,072
					5	min		DEN.A500.FURANS.5									
Post-Dehydration Cooler				1	$76,800	2015	$76,800	HEAT.A500.FURANS.60	5.1827	Gcal/hr	1	3.04296875	20	3.92	$301,097	$292,931	$891,380
Aldol condensation reactor	5th seed fermentor (Humbird, 2009)			1	$1,180,000	2009	$1,180,000	STRM.A500.FURANS.S8	200000	gal	0.7	2	52863.3839	0.26	$464,911	$482,549	$965,097
								DEN.A500.FURANS.S8									
Dioxane Column Condenser	ACCE			1	$267,100	2016	$267,100	HEAT.A500.FURANS.SEPRTN.S7	58	Gcal/hr	0.44	2.47	30	0.52	$200,674	$200,674	$495,664
Dioxane Column Condenser (ACC)	ACCE			1	$39,500	2016	$39,500	HEAT.A500.FURANS.SEPRTN.S7	58	Gcal/hr	0.44	2.47	30	0.52	$29,677	$29,677	$73,301
Dioxane Column Reboiler	ACCE			1	$254,200	2016	$254,200	HEAT.A500.FURANS.SEPRTN.S6	-67	Gcal/hr	0.79	2.47	-35	0.53	$153,409	$153,409	$378,920
Dioxane Column Reflux Pump	ACCE			1	$23,300	2016	$23,300	STRM.A500.FURANS.SEPRTN.S8	274694	kg/hr	0.79	2.47	151672	0.55	$14,574	$14,574	$35,998
Dioxane Column Tower	ACCE			1	$628,600	2016	$628,600	STRM.A500.FURANS.SEPRTN.S1	373388	kg/hr	0.68	2.47	208246	0.56	$422,608	$422,608	$1,043,842
Hydrocarbon Decanter	ACCE			1	$32,100	2016	$32,100	STRM.A500.FURANS.SEPRTN.BOT-C	98695	kg/hr	0.60	2.47	57872	0.59	$23,303	$23,303	$57,558
Dioxane Recycle Cooler				1	$76,800	2015	$76,800	HEAT.A500.FURANS.61	5.1827	Gcal/hr	1	3.04296875	4	0.76	$58,346	$56,764	$172,731
Hydrocarbon Pump	190 GPM, 615 FT TDH	40	316SS	2	$24,300	2009	$24,300	STRM.A500.FURANS.S22	43149	kg/hr	0.8	2.3	19228	0.45	$25,458	$26,423	$60,774
Pre-Hydrotreating Economizer	2-4 TEMA shell and tube HX		316SS	1	$353,600	2011	$353,600	HEAT.A500.FURANS.47	-14.3	Mmkcal/hr	0.7	2.66	-3.3	0.23	$127,081	$117,534	$312,150
Pre-Dehydration Heater				1	$76,800	2015	$76,800	HEAT.A500.FURANS.48	5.1827	Gcal/hr	1	3.04296875	1.0525161	0.20	$15,596.74	$15,173.77	$46,173.31
Fresh H2 compressor	Reciprocating compressor (5 stages)			1	$1,621,200	2011	$1,621,200	STRM.A500.FURANS.H2-FRESH	389.6519	kg/hr	0.6	1.09	1986.54962	5.10	$4,308,144	$3,984,500	$4,342,593
Fresh H2 compressor (spare)	Reciprocating compressor (5 stages)			1	$1,621,200	2011	$1,621,200	STRM.A500.FURANS.H2-FRESH	389.6519	kg/hr	0.6	1.08	1986.54962	5.10	$4,308,144	$3,984,500	$4,308,185
Hydrotreating Reactor	Fixed bed reactor (Q3 FY17 milestone), base PF=2.5,208 BBL/hr			1	$6,513,387	2011	$6,513,387	STRM.A500.FURANS.S22	32894.95	L/hr	0.7	2.00	7390.8	0.22	$2,290,349	$2,118,289	$4,236,579
Pressure factor (via Guthrie)	*(>1000PSIG=2.5,900=2.3,800=1.9,700=1.8,600=1.6,500=1.45,400=1.35)*							DEN.A500.FURANS.S22									
Flash vessel	23' x 48' - 110,000 gal.		SS316	1	$511,000	2009	$511,000	STRM.A500.FURANS.43	264116	kg/hr	0.7	2.00	22162.117	0.08	$90,176	$93,597	$187,195
PSA unit	H2 recovery			1	$975,000	2013	$975,000	STRM.A500.FURANS.26	13528	kg/hr	0.6	1.90	1092.52755	0.08	$215,439	$205,717	$390,862
Recycle H2 compressor	Centrifugal compressor			1	$1,103,700	2011	$1,103,700	STRM.A500.FURANS.RECYC	14665.49	kg/hr	0.6	1.13	947.079615	0.06	$213,258	$197,237	$222,917
Recycle H2 compressor (spare)	Centrifugal compressor			1	$1,103,700	2011	$1,103,700	STRM.A500.FURANS.RECYC	14665.49	kg/hr	0.6	1.10	947.079615	0.06	$213,258	$197,237	$217,717
Air compressor				2	$34,600	2011	$34,600	STRM.A500.FURANS.44	3818.9219	kg/hr	0.6	1.82	2257.53985	0.59	$50,480	$46,688	$85,010
Off-gas boiler				1	$241,400	2011	$241,400	HEAT.A500.FURANS.46	2.4187418	MMkcal/hr	0.7	1.52	1.30758596	0.54	$156,947	$145,157	$221,283
Flash vessel	23' x 48' - 110,000 gal.		SS316	1	$511,000	2009	$511,000	STRM.A500.FURANS.S1	264116	kg/hr	0.7	2	16248.2684	0.06	$72,565	$75,318	$150,637
														Area 500 Totals	**$19,673,536**	**$18,559,130**	**$29,593,424**

A600: WWT	Mechanical Equipment List						Scaled Installed Costs										
EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
Aeration Basin	Concrete and steel, not installed cost		Concrete	3	$4,804,854	2012	$4,804,854	Hydraulic flow	2.7	MGD	0.60	2.1	1.9	0.70	$3,872,991	$3,588,778	$7,428,769
Pump - Centrifugal, Aeration Basin Feed	852 gpm ea	45 hp	CS	4	$64,800	2012											
Aeration Grid	Full floor aeration grid		CS	1	$2,500,000	2012											
Caustic Feed System		1.5 hp	CS	4	$20,000	2012	$20,000	COD	5600	kg/hr	0.60	3.0	6758	1.21	$22,388	$20,745	$62,236
Blowers	15000 SCFM @ 10.3psig ea	1000 hp ea	CS	9	$2,070,000	2012	$2,070,000	COD	5600	kg/hr	0.60	2.0	6758	1.21	$2,317,180	$2,147,137	$4,294,275
Membrane Bioreactor	Includes membrane, CIP, Scour system	85 hp ea	CS	1	$4,898,500	2012	$4,898,500	Hydraulic flow	2.7	MGD	1.00	1.6	1.9	0.70	$3,419,849	$3,168,889	$5,196,978
Pump, Centrifugal , MBR, RAS		160 hp	CS	6	INCLUDED	2012											
Gravity Belt Thickeners	2m presses	48hp	CS	3	$750,000	2012	$750,000	COD	5600	kg/hr	0.60	1.6	6758	1.21	$839,558	$777,948	$1,236,938
Centrifuge		165 hp ea	CS	1	$686,800	2012	$686,800	COD	5600	kg/hr	0.60	2.7	6758	1.21	$768,811	$712,393	$1,916,338
Pump, Centrifugal, Centrifuge Feed	105 gpm	15hp	CS	2	INCLUDED	2012											
Pump, Submersible, Centrate	100 gpm	10 hp ea	CS	2	INCLUDED	2012											
Dewatering Polymer Addition	9.8 gph neat polymer	1 hp ea	CS	2	INCLUDED	2012											
Conveyor		10 hp ea	CS	1	$7,000	2012	$7,000	COD	5600	kg/hr	0.60	2.9	6758	1.21	$7,836	$7,261	$20,766
Reverse Osmosis			CS	7	$2,450,000	2012	$2,450,000	Hydraulic flow	2.7	MGD	1.00	1.8	1.9	0.70	$1,710,448	$1,584,930	$2,773,627
Evaporator	368 gpm	1480 hp ea	Titanium	1	$5,000,000	2012	$5,000,000	Hydraulic flow	2.7	MGD	0.60	1.6	1.9	0.70	$4,030,290	$3,734,533	$6,012,599
Ammonia Addition System	0.63 gpm	4.5 hp	CS	4	$195,200	2012	$195,200	COD	5600	kg/hr	0.60	1.5	6758	1.21	$218,509	$202,474	$311,810
Sodium Sulfate Puification																	
Evaporator feed tank	insulated, 6460 gal				$45,966	2011	$45,966	strm.A600.23	290932	kg/hr	0.60	2.50	1351858	4.65	$115,537	$106,857	$267,143
Evaporator feed heater	shell and tube 1/2 pass				$274,818	2011	$274,818	heat.A600.31	-13	MMkcal/h	0.60	3.00	-7	0.56	$193,611	$179,067	$537,200
Evaporator flash drum	23' x 48' - 110,000 gal.		SS316	1	$511,000	2009	$511,000	strm.A600.23	264116	kg/hr	0.70	2.00	1351858	5.12	$1,602,568	$1,663,366	$3,326,733
Centrifuge	Nexant quote sodium sulfate, 25410 lb/hr solids basis			1	$327,680	2011	$327,680	strm.A600.NA2SO4	11524	kg/hr	0.60	2.3	14643	1.27	$378,328	$349,907	$804,786
Dryer	Nexant quote Sodium sulfate, 25410 lb/hr solids basis			1	$555,008	2011	$555,008	strm.A600.PRD-SALT	11524	kg/hr	0.60	2.6	13946	1.21	$622,307	$575,557	$1,496,448
														Area 600 Totals	**$20,120,213**	**$18,819,842**	**$35,686,644**

A700: Lignin Utilization	Mechanical Equipment List					Scaled Installed Costs											
A701: Lignin Conditioning																	
Neutralization Tank	2.6 atm, 130C operating30 min. hold = 30,000 gal		SS317	1	$236,000	2009	$236,000	strm.A700.A701.LIQUID1	410369	kg/hr	0.70	2.0	97536	0.24	$86,319	$89,594	$179,188
Pulping Reactor Tank	2.6 atm, 130C operating (up to 160C, 30 min)30 min. hold = 30,000 gal		SS316	1	$16,300,000	2013	$16,300,000	strm.A700.A701.PULP-OUT	323295	kg/hr	0.60	1.7	130653	0.40	$9,464,528	$9,037,431	$14,911,762
Flash/drain tank			SS317	1	$262,000	2013	$262,000	strm.A700.A701.PULP-OUT	323296	kg/hr	0.70	2.0	130653	0.40	$138,952	$132,681	$265,363
Black Liquor Storage Tank	1,200,000 gallon		316SS	1	$1,317,325	2011	$1,317,325	strm.A700.A701.5	328984	kg/hr	0.70	1.8	92878	0.28	$543,513	$502,682	$904,828
A702: Muconate Fermentation																	
1st Aerobic Seed	80 gallon skid complete - $46,000 ea		304SS		$46,000	2009	$46,000	NSD1000	1	ea	1.00	1.80	4	4.00	$184,000	$190,981	$343,765
1st Seed Vessel Agitator		0.75hp	SS316		$3,420	2009	$3,420	NSD1000	1	ea	1.00	1.50	4	4.00	$13,680	$14,199	$21,298
2nd Aerobic Seed	800 gallon skid complete - $57,500 ea		304SS		$57,500	2009	$57,500	NSD1000	1	ea	1.00	1.80	4	4.00	$230,000	$238,726	$429,706
2nd Seed Vessel Agitator		8 hp	SS316		$11,000	2009	$11,000	NSD1000	1	ea	1.00	1.50	4	4.00	$44,000	$45,669	$68,504
Bubble column seed fermenter	100 m3		316SS		$274,100	2014	$274,100	NSD1000	1	ea	1.00	2.30	4	4.00	$1,096,400	$1,030,932	$2,371,143
Seed circulation cooler	650 sqft		316SS		$8,400	2014	$8,400	NSD1000	1	ea	1.00	2.20	4	4.00	$33,600	$31,594	$69,506
Bubble column production fermenter	1000 m3		316SS		$1,691,400	2014	$1,691,400	NVES1000	1	ea	1.00	2.30	18	18.00	$30,445,200	$28,627,261	$65,842,699
Production circulation cooler	4500 sqft		316SS		$48,100	2014	$48,100	NVES1000	1	ea	1.00	2.20	18	18.00	$865,800	$814,101	$1,791,023
Production circulation pump	400 gpm		316SS		$11,500	2014	$11,500	NVES1000	1	ea	1.00	2.30	18	18.00	$207,000	$194,640	$447,671
Fermentation air compressor	25,000 ACFM @ 45psig; max size in ACCE		CS		$1,318,600	2014	$1,318,600	AIRV1000	13	m^3/s	1.00	1.60	11	0.85	$1,115,549	$1,048,938	$1,678,300
Fermentation air receiver	25,000 gal		CS		$104,600	2014	$104,600	AIRV1000	13	m^3/s	1.00	2.00	11	0.85	$88,493	$83,209	$166,417
Fermentation Surge tank	insulated cone bottom, 6460 gal				$45,966	2011	$45,966	strm.A700.A702.UF-FD	290932	kg/hr	0.60	2.50	176573	0.61	$34,066	$31,507	$78,766
Ultrafiltration membrane separator					$2,048,000	2011	$2,048,000	Volume Flow	1303	GPM	0.60	2.50	766	1.00	$2,048,000	$1,894,146	$4,735,366
membrane broth feed pump					INCLUDED			strm.A700.A702.UF-FD		kg/hr			176573				
membrane solvent feed pump					INCLUDED			den.A700.A702.UF-FD		gm/cc			1.0				
A703: Recovery and Upgrading																	
Carbon Filter	2 Vessels, for color removal				$345,234	2011	$345,234	Volume Flow	1347	GPM	0.60	2.50	704	1.00	$345,234	319298.7157	$798,247
Initial carbon loading					see CATALYST	2011		strm.A700.A703.CFIL-FD		kg/hr			162344				
								den.A700.A703.CFIL-FD		gm/cc			1.0				
CCM Crystallizer	Oslo Type. 2 In series		316SS	2 series	$7,104,192	2011	$7,104,192	Volume Flow	190	GPM	0.60	2.50	44.174	0.23	$2,960,473	2738070.993	$6,845,177
								strm.A700.A703.CRY1-PRD		kg/hr	0.60	2.50	11915				
								den.A700.A703.CRY1-PRD		g/cc	0.60	2.50	1.2				
CCM Centrifuge	Centrifuge Separator				$327,680	2011	$327,680	strm.A700.A703.CRY1-PRD	13403	kg/hr	0.60	2.30	11915	0.89	$305,344	$282,406	$649,533
CCM Drier	Fluidized bed drier parallel			2 parallel	$555,008	2011	$555,008	strm.A700.A703.DRY1-PRD	11526	kg/hr	0.60	2.60	11327	0.98	$549,237	$507,976	$1,320,738
Dissolution Tank	mixing tank to redissolve crystals in solvent (etOH)				$1,317,325	2011	$1,317,325	strm.A700.A703.FIL2-FD	328984	kg/hr	0.70	1.80	56041	0.17	$381,614	$352,946	$635,303
Dissolution Tank agitator	pump to retain crystal suspension	80	316SS		$63,000	2009	$63,000	work.A700.A703.W-ETOHMX	60	kW	1.00	1.50	60	1.00	$63,000	$65,390	$98,085
Filtration Centrifuge(salt removal)	removes precipitated solids after dissolution				$327,680	2011	$327,680	strm.A700.A703.FIL2-SLT	13403	kg/hr	0.60	2.30	155	0.01	$22,550	$20,856	$47,970
HDO feed tank	insulated, 6460 gal				$45,966	2011	$45,966	strm.A700.A703.FIL2-PRD	290932	kg/hr	0.60	2.50	55886	0.19	$17,082	$15,799	$39,497
HDO reactor pump					$802,861	2014	$802,861	strm.A700.A703.FIL2-PRD	208720	kg/hr	0.80	1.40	55886	0.27	$279,790	$263,083	$368,316
HDO Feed Effluent econimizer	2-4 TEMA shell and tube HX		316SS		$353,600	2011	$353,600	heat.A700.A703.QX-HDO	14	Mmkcal/hr	0.70	2.66	3	0.19	$110,926	$102,593	$272,470
HDO trim preheater			304SS		$41,000	2009	$41,000	heat.A700.A703.QH-TRIM	-2	MMkcal/hr	0.70	2.20	0.0	0.00	$0	$0	$0
HDO Fixed Bed Reactor	(Q3 FY17 milestone), base PF=2.5,208 BBL/hr				$4,168,568	2011	$4,168,568	Volume Flow (liquid)	32895	L/hr	0.70	2.00	69659	2.12	$7,048,215	$6,518,726	$13,037,453
Pressure Factor (via Guthrie)	(>1000PSIG=2.5,900=2.3,800=1.9,700=1.8,600=1.6,500=1.45,400=1.35)		1.6					strm.A700.A703.RXR-FD	29274	kg/hr			55886				
Internals								den.A700.A703.RXR-FD	1	GM/CC			0.8				
Hydrogenation Intercooler (bed1)					$2,353,181	2007	$2,353,181	heat.A700.A703.QC-BED1	32	MMkcal/hr	0.65	2.21	2	0.06	$385,095	$397,042	$877,463
Hydrogenation Intercooler (bed2)					$2,353,181	2007	$2,353,181	heat.A700.A703.QC-BED2	32	MMkcal/hr	0.65	2.21	3	0.10	$521,131	$537,298	$1,187,429
H2 Makeup Compressor	recipricating compressor(5 stages)				$1,621,200	2011	$1,621,200	strm.A700.A703.H2-MU	390	kg/hr	0.60	1.09	417	1.07	$1,688,438	$1,561,596	$1,701,939
H2 Makeup Compressor spare	recipricating compressor(5 stages)				$1,621,200	2011	$1,621,200	strm.A700.A703.H2-MU	390	kg/hr	0.60	1.08	417	1.07	$1,688,438	$1,561,596	$1,688,454
HHPS	Via Adipic model(via MB)				$436,000	2013	$436,000	strm.A700.A703.HHPS-FD	119841	kg/hr	1.00	1.50	56303	0.47	$204,839	$195,595	$293,393
HDO hot gas cooler					$321,600	2011	$321,600	heat.A700.A703.QAC-2	4	Mmkcal/h	0.70	1.66	0.0	0.01	$10,961	$10,138	$16,798
CHPS	3-Phase horizontal sep., demister, 3/16 SS316 cladding				$328,500	2011	$328,500	Volume Flow	39911	L/hr	0.70	2.59	96	0.00	$4,841	$4,477	$11,583
								strm.A700.A703.CHPS-LIQ		kg/hr			76				
								den.A700.A703.CHPS-LIQ		GM/CC			0.8				
AA evaporator feed tank	insulated, 6460 gal				$45,966	2011	$45,966	strm.A700.A703.EVAP-FD	290932	kg/hr	0.60	2.50	98895	0.34	$24,058	$22,251	$55,628
AA evaporator feed heater	shell and tube 1/2 pass				$274,818	2011	$274,818	heat.A700.A703.QH-EVAP	-13	MMkcal/h	0.60	3.00	-2	0.13	$79,691	$73,704	$221,113
AA evaporator flash drum	23' x 48' - 110,000 gal.		SS316	1	$511,000	2009	$511,000	strm.A700.A703.EVAP-FD	264116	kg/hr	0.70	2.00	98895	0.37	$256,914	$266,660	$533,321
AA condensor drum					$487,000	2010	$487,000	heat.A700.A703.QC-COND	23	MMkcal/h	0.60	2.80	9	0.40	$279,441	$274,824	$769,508
AA Crystallizer	Oslo Type. 2 In series		316SS	2 series	$7,104,192	2011	$7,104,192	Volume Flow	190	GPM	0.60	2.50	45.386	0.24	$3,008,955	$2,782,910	$6,957,276
								strm.A700.A703.CRY2-PRD		kg/hr	0.60	2.50	12099				
								den.A700.A703.CRY2-PRD		g/cc	0.60	2.50	1.2				
AA Centrifuge separator	Centrifuge Separator				$327,680	2011	$327,680	strm.A700.A703.CRY2-PRD	13403	kg/hr	0.60	2.30	12099	0.90	$308,165	$285,014	$655,533
AA Drier	Fluidized bed drier parallel			2 parallel	$555,008	2011	$555,008	strm.A700.A703.DRY2-PRD	11526	kg/hr	0.60	2.60	11525	1.00	$554,965	$513,274	$1,334,512
														Totals:	**$67,738,496**	**$63,681,817**	**$134,722,047**

A800: CHP	Mechanical Equipment List					Scaled Installed Costs											
High Solids Burner and Turbine																	
Burner Combustion Air Preheater	INCLUDED			1	INCLUDED												
BFW Preheater	INCLUDED			1	INCLUDED												
Pretreatment/BFW heat recovery	9.4 MM Btu/hr		SS304	1	$41,000	2009	$41,000	heat.A800.A810.QH812	-2	Gcal/hr	0.70	2.2	-2	0.77	$34,030	$35,321	$77,706
Air Intake Fan	INCLUDED				INCLUDED												
Boiler	525,000 lb/hr @ 900 psig		CS	1	$28,550,000	2010	$28,550,000	strm.A800.A810.813c	238203	kg/hr	0.60	1.8	288713	1.21	$32,041,811	$31,512,435	$56,722,383
Combustion Gas Baghouse	Baghouse, Spray dryer scrubber, flues/ducting			1	$11,000,000	2013	$0	strm.A800.A810.812	238203	kg/hr	0.60	1.8	241921	1.02	$0	$0	$0
Turbine/Generator	23.6 kW, 2 extractions			1	$9,500,000	2010	$9,500,000	work.A900.wtotal	-42200	kW	0.60	1.8	-25526	0.60	$7,026,246	$6,910,162	$12,438,292
Hot Process Water Softener System				1	$78,000	2010	$78,000	strm.A800.A810.812	235803	kg/hr	0.60	1.8	241921	1.03	$79,208	$77,899	$140,219
Amine Addition Pkg.				1	$40,000	2010	$40,000	strm.A800.A810.812	235803	kg/hr	0.00	1.8	241921	1.03	$40,000	$39,339	$70,810
Ammonia Addition Pkg				1	INCLUDED												
Phosphate Addition Pkg.				1	INCLUDED												
Condensate Pump			SS316	2	INCLUDED												
Turbine Condensate Pump			SS304	2	INCLUDED												
Deaerator Feed Pump			SS304	2	INCLUDED												
BFW Pump			SS316	5	INCLUDED												
Blowdown Pump			CS	2	INCLUDED												
Amine Transfer Pump			CS	1	INCLUDED												
Condensate Collection Tank			A285C	1	INCLUDED												
Condensate Surge Drum			SS304	1	INCLUDED												
Deaerator	Tray type		CS;SS316	1	$305,000	2010	$305,000	strm.A800.A810.812	235803	kg/hr	0.60	3.0	241921	1.03	$309,723	$304,606	$913,819
Blowdown Flash Drum			CS	1	INCLUDED												
Amine Drum			SS316	1	INCLUDED												
														Area 800 Totals	**$39,531,018**	**$38,879,762**	**$70,363,228**

A900: Utilities & Storage	Mechanical Equipment List					Scaled Installed Costs											
Utilities System																	
Cooling Tower System	44,200 gpm	750 hp	FIBERGLASS	1	$1,375,000	2010	$1,375,000	strm.a900.945	10037820	kg/hr	0.60	1.5	22653975	2.26	$2,240,813	$2,203,792	$3,305,688
Plant Air Compressor	400 SCFM@125 psig	150 hp		1	$28,000	2010	$28,000	DRY101	83333	kg/hr	0.60	1.6	83333	1.00	$28,000	$27,537	$44,060
Chilled Water Package	2 x 2350 tons (14.2 MM kcal/hr)	3400 hp		1	$1,275,750	2010	$1,275,750	heat.a900.qchwop	14	Gcal/hr	0.60	1.6	50	3.50	$2,707,030	$2,662,306	$4,259,690
CIP System	100,000 GAL		SS304/SS316	1	$421,000	2009	$421,000	strm.a900.914	63	kg/hr	0.60	1.8	145	2.30	$694,222	$720,560	$1,297,008
Cooling Water Pump	16,120 GPM, 100 FT TDH SIZE 20X20-28	500.0	CS	3	$283,671	2010	$283,671	strm.a900.945	10982556	kg/hr	0.80	3.1	22653975	2.06	$506,253	$497,889	$1,543,456
Make-up Water Pump	685 GPM, 75 FT TDH SIZE 6X4-13	20.0	CS	1	$6,864	2010	$6,864	strm.a900.904	155564	kg/hr	0.80	3.1	567275	3.65	$19,324	$19,004	$58,913
Process Water Circulating Pump	2285 GPM, 75 FT TDH SIZE 8X6-13	75.0	CS	1	$15,292	2010	$15,292	strm.a900.905	518924	kg/hr	0.80	3.1	722284	1.39	$19,923	$19,594	$60,740
Instrument Air Dryer	670 SCFM - CYCLING TYPE		CS	1	$15,000	2009	$15,000	DRY101	83333	kg/hr	0.60	1.8	83333	1.00	$15,000	$15,569	$28,024
Plant Air Receiver	3800 gal - 72" x 228" vertical		CS	1	$16,000	2009	$16,000	DRY101	83333	kg/hr	0.60	3.1	83333	1.00	$16,000	$16,607	$51,482
Process Water Tank No. 1	250,000 gal		CS	1	$250,000	2009	$250,000	strm.a900.905	451555	kg/hr	0.70	1.7	722284	1.60	$347,327	$360,504	$612,856
Storage																	
Ammonia Storage Tank	28,000 gal		SA- 516-70	2	$196,000	2010	$196,000	strm.A900.NH3-NET	1171	kg/hr	0.70	2.0	1104	0.94	$188,053	$184,946	$369,892
Dioxane Storage Tank	28,000 gal		SA- 516-70	2	$196,000	2010	$196,000	strm.A500.FURANS.DIOXANE	1171	kg/hr	0.70	2.0	120	0.10	$39,668	$39,013	$78,025
CSL Storage Tank	70,000 gal		Glass lined	1	$70,000	2009	$70,000	strm.A900.CSL-NET	1393	kg/hr	0.70	2.6	415	0.30	$29,967	$31,104	$80,871
CSL Storage Tank Agitator		10 hp	SS304	1	$21,200	2009	$21,200	strm.A900.CSL-NET	1393	kg/hr	0.50	1.5	415	0.30	$11,565	$12,004	$18,006
CSL Pump	8 GPM, 80 FT TDH	0.5	CS	1	$3,000	2009	$3,000	strm.A900.CSL-NET	1393	kg/hr	0.80	3.1	415	0.30	$1,138	$1,181	$3,661
DAP Bulk Bag Unloader	Super sack unloader			1	$30,000	2009	$30,000	strm.A900.DAP-NET	163	kg/hr	0.60	1.7	621	3.81	$66,955	$69,495	$118,142
DAP Bulk Bag Holder	Super sack holder			1	INCLUDED												
DAP Make-up Tank	12,800 gal		SS304	1	$102,000	2009	$102,000	strm.A900.DAP-NET	1615	kg/hr	0.70	1.8	621	0.38	$52,262	$54,245	$97,640
DAP Make-up Tank Agitator		5.5 hp	SS304	1	$9,800	2009	$9,800	strm.A900.DAP-NET	163	kg/hr	0.50	1.5	621	3.81	$19,133	$19,859	$29,788
DAP Pump	2 GPM, 100 FT TDH	0.5	CS	1	$3,000	2009	$3,000	strm.A900.DAP-NET	163	kg/hr	0.80	3.1	621	3.81	$8,750	$9,082	$28,154
Sulfuric Acid Pump	5 GPM, 150 FT TDH SIZE 2X1-10	0.5	SS316	1	$7,493	2010	$7,493	strm.A900.ACID-NET	1981	kg/hr	0.80	2.3	11241	5.67	$30,046	$29,550	$67,965
Sulfuric Acid Storage Tank	12,600 gal, 12' dia x15' H		SS	1	$96,000	2010	$96,000	strm.A900.ACID-NET	1981	kg/hr	0.70	1.5	11241	5.67	$323,605	$318,259	$477,389
Caustic Storage Tank	12,600 gal, 12' dia x15' H		SS	1	$96,000	2011	$96,000	strm.A900.BASE-NET	1981	kg/hr	0.70	1.5	9494	4.79	$287,515	$265,916	$398,874
AlCl3 Storage Tank	12,600 gal, 12' dia x15' H		SS	1	$96,000	2011	$96,000	STRM.A500.FURANS.ALCL3	1981	kg/hr	0.70	1.5	535	0.27	$38,400	$35,516	$53,273
Firewater Storage Tank	600,000 gal - 4 hrs @ 2500 gpm		Glass lined	1	$803,000	2009	$803,000	strm.A900.H2O-FIRE	8343	kg/hr	0.70	1.7	8343	1.00	$803,000	$833,464	$1,416,890
Firewater Pump	2500 GPM, 150 FT TDH	125.0	CS	1	$15,000	2009	$15,000	strm.A900.H2O-FIRE	8343	kg/hr	0.80	3.1	8343	1.00	$15,000	$15,569	$48,264
Diesel storage tank	750,000 gal., 7 day storage, Floating roof		A285C	1	$670,000	2009	$670,000	strm.PRD-500	11341	kg/hr	0.70	1.7	15873	1.40	$847,806	$879,970	$1,495,949
Co-Product Storage Tank(Adipic)				1	$690,900	2007	$690,900	strm.PRD-700	23322.902	kg/hr	0.65	1.850	11525	0.49	$436,930	$450,485	$833,397
Co-Product Storage Tank (Sodium Sulfate)				1	$690,900	2007	$690,900	strm.PRD-600	23322.902	kg/hr	0.65	1.850	13946	0.60	$494,585	$509,929	$943,370
Glucose Storage Tank	70,000 gal		Glass lined	1	$70,000	2009	$70,000	strm.a400.401	1393	kg/hr	0.70	2.6	1557	1.12	$75,686	$78,557	$204,249
														Area 900 Totals	**$10,363,957**	**$10,381,506**	**$18,025,715**

Appendix B. Discounted Cash Flow Rate of Return Worksheet

Dedicated Biorefinery

DCFROR Worksheet

Year		-2	-1	0	1	2	3	4	5
Fixed Capital Investment		$23,923,976	$179,429,817	$95,695,902					
Land		$1,848,000							
Working Capital				$37,381,212					
Loan Payment					$66,850,835	$66,850,835	$66,850,835	$66,850,835	$66,850,835
Loan Interest Payment		$2,870,877	$24,402,455	$35,885,963	$35,885,963	$33,408,774	$30,733,409	$27,844,015	$24,723,469
Loan Principal		$35,885,963	$305,030,688	$448,574,541	$417,609,670	$384,167,609	$348,050,183	$309,043,363	$266,915,998
Fuel Sales					$149,571,091	$199,428,121	$199,428,121	$199,428,121	$199,428,121
By-Product Credit					$144,419,846	$192,559,795	$192,559,795	$192,559,795	$192,559,795
Total Annual Sales					$293,990,937	$391,987,916	$391,987,916	$391,987,916	$391,987,916
Annual Manufacturing Cost									
Feedstock					$38,712,485	$51,616,647	$51,616,647	$51,616,647	$51,616,647
Periodic Costs (Catalyst, etc)					$16,174,947	$0	$1,506,720	$238,632	$1,506,720
Other Variable Costs					$188,725,096	$215,685,824	$215,685,824	$215,685,824	$215,685,824
Fixed Operating Costs					$20,426,101	$20,426,101	$20,426,101	$20,426,101	$20,426,101
Total Product Cost					$264,038,630	$287,728,572	$289,235,292	$287,967,204	$289,235,292
Annual Depreciation									
General Plant Writedown					14%	24.49%	17.49%	12.49%	8.93%
Depreciation Charge					$98,249,621	$168,378,811	$120,250,936	$85,873,881	$61,397,419
Remaining Value					$589,291,463	$420,912,652	$300,661,716	$214,787,835	$153,390,416
Steam Plant Writedown					3.75%	7.22%	6.68%	6.18%	5.71%
Depreciation Charge					$2,253,118	$4,337,403	$4,011,752	$3,711,336	$3,432,550
Remaining Value					$57,830,033	$53,492,630	$49,480,878	$45,769,542	$42,336,992
Net Revenue					($106,436,395)	($101,865,644)	($52,243,472)	($13,408,520)	$13,199,186
Losses Forward						($106,436,395)	($208,302,039)	($260,545,511)	($273,954,032)
Taxable Income					($106,436,395)	($208,302,039)	($260,545,511)	($273,954,032)	($260,754,846)
Income Tax					$0	$0	$0	$0	$0
Annual Cash Income					-$36,898,527	$37,408,509	$35,901,790	$37,169,877	$35,901,790
Discount Factor		1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830	0.6209
Annual Present Value	$425,588,258				-$33,544,116	$30,916,123	$26,973,546	$25,387,526	$22,292,187
Total Capital Investment + Interest		$34,657,852	$224,215,499	$168,963,077					
Net Present Worth				$0					

DCFROR Worksheet

Year	6	7	8	9	10	11	12	13
Fixed Capital Investment								
Land								
Working Capital								
Loan Payment	$66,850,835	$66,850,835	$66,850,835	$66,850,835	$66,850,835	$0	$0	$0
Loan Interest Payment	$21,353,280	$17,713,475	$13,782,487	$9,537,019	$4,951,914	$0	$0	$0
Loan Principal	$221,418,443	$172,281,084	$119,212,736	$61,898,921	$0	$0	$0	$0
Fuel Sales	$199,428,121	$199,428,121	$199,428,121	$199,428,121	$199,428,121	$199,428,121	$199,428,121	$199,428,121
By-Product Credit	$192,559,795	$192,559,795	$192,559,795	$192,559,795	$192,559,795	$192,559,795	$192,559,795	$192,559,795
Total Annual Sales	$391,987,916	$391,987,916	$391,987,916	$391,987,916	$391,987,916	$391,987,916	$391,987,916	$391,987,916
Annual Manufacturing Cost								
Feedstock	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647
Periodic Costs (Catalyst, etc)	$511,169	$1,745,352	$0	$1,506,720	$238,632	$2,017,888	$0	$1,745,352
Other Variable Costs	$215,685,824	$215,685,824	$215,685,824	$215,685,824	$215,685,824	$215,685,824	$215,685,824	$215,685,824
Fixed Operating Costs	$20,426,101	$20,426,101	$20,426,101	$20,426,101	$20,426,101	$20,426,101	$20,426,101	$20,426,101
Total Product Cost	$288,239,741	$289,473,924	$287,728,572	$289,235,292	$287,967,204	$289,746,460	$287,728,572	$289,473,924
Annual Depreciation								
General Plant Writedown	8.92%	8.93%	4.46%					
Depreciation Charge	$61,328,665	$61,397,419	$30,664,332					
Remaining Value	$92,061,751	$30,664,332	$0					
Steam Plant Writedown	5.29%	4.89%	4.52%	4.46%	4.46%	4.46%	4.46%	4.46%
Depreciation Charge	$3,175,395	$2,936,864	$2,716,960	$2,680,910	$2,680,309	$2,680,910	$2,680,309	$2,680,910
Remaining Value	$39,161,597	$36,224,733	$33,507,773	$30,826,862	$28,146,553	$25,465,643	$22,785,333	$20,104,423
Net Revenue	$17,890,836	$20,466,233	$57,095,565	$90,534,695	$96,388,489	$99,560,545	$101,579,035	$99,833,082
Losses Forward	($260,754,846)	($242,864,010)	($222,397,776)	($165,302,212)	-$74,767,517	$0	$0	$0
Taxable Income	($242,864,010)	($222,397,776)	($165,302,212)	-$74,767,517	$21,620,972	$99,560,545	$101,579,035	$99,833,082
Income Tax	$0	$0	$0	$0	$4,540,404	$20,907,715	$21,331,597	$20,964,947
Annual Cash Income	$36,897,341	$35,663,158	$37,408,509	$35,901,790	$32,629,473	$81,333,741	$82,927,747	$81,549,045
Discount Factor	0.5645	0.5132	0.4665	0.4241	0.3855	0.3505	0.3186	0.2897
Annual Present Value	$20,827,587	$18,300,839	$17,451,346	$15,225,863	$12,580,074	$28,506,980	$26,423,336	$23,621,853
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	$199,428,121	$199,428,121	$199,428,121	$199,428,121	$199,428,121	$199,428,121	$199,428,121	$199,428,121
By-Product Credit	$192,559,795	$192,559,795	$192,559,795	$192,559,795	$192,559,795	$192,559,795	$192,559,795	$192,559,795
Total Annual Sales	$391,987,916	$391,987,916	$391,987,916	$391,987,916	$391,987,916	$391,987,916	$391,987,916	$391,987,916
Annual Manufacturing Cost								
Feedstock	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647
Periodic Costs (Catalyst, etc)	$0	$1,506,720	$749,801	$1,506,720	$0	$1,745,352	$0	$2,017,888
Other Variable Costs	$215,685,824	$215,685,824	$215,685,824	$215,685,824	$215,685,824	$215,685,824	$215,685,824	$215,685,824
Fixed Operating Costs	$20,426,101	$20,426,101	$20,426,101	$20,426,101	$20,426,101	$20,426,101	$20,426,101	$20,426,101
Total Product Cost	$287,728,572	$289,235,292	$288,478,373	$289,235,292	$287,728,572	$289,473,924	$287,728,572	$289,746,460
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	$2,680,309	$2,680,910	$2,680,309	$2,680,910	$2,680,309	$2,680,910	$2,680,309	$1,340,455
Remaining Value	$17,424,114	$14,743,204	$12,062,894	$9,381,984	$6,701,675	$4,020,764	$1,340,455	$0
Net Revenue	$101,579,035	$100,071,714	$100,829,234	$100,071,714	$101,579,035	$99,833,082	$101,579,035	$100,901,000
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$101,579,035	$100,071,714	$100,829,234	$100,071,714	$101,579,035	$99,833,082	$101,579,035	$100,901,000
Income Tax	$21,331,597	$21,015,060	$21,174,139	$21,015,060	$21,331,597	$20,964,947	$21,331,597	$21,189,210
Annual Cash Income	$82,927,747	$81,737,564	$82,335,404	$81,737,564	$82,927,747	$81,549,045	$82,927,747	$81,052,245
Discount Factor	0.2633	0.2394	0.2176	0.1978	0.1799	0.1635	0.1486	0.1351
Annual Present Value	$21,837,468	$19,567,323	$17,918,583	$16,171,341	$14,915,284	$13,333,920	$12,326,681	$10,952,636
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									($37,381,212)
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	$199,428,121	$199,428,121	$199,428,121	$199,428,121	$199,428,121	$199,428,121	$199,428,121	$199,428,121	$199,428,121
By-Product Credit	$192,559,795	$192,559,795	$192,559,795	$192,559,795	$192,559,795	$192,559,795	$192,559,795	$192,559,795	$192,559,795
Total Annual Sales	$391,987,916	$391,987,916	$391,987,916	$391,987,916	$391,987,916	$391,987,916	$391,987,916	$391,987,916	$391,987,916
Annual Manufacturing Cost									
Feedstock	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647
Periodic Costs (Catalyst, etc)	$238,632	$1,506,720	$0	$1,745,352	$511,169	$1,506,720	$238,632	$1,506,720	$0
Other Variable Costs	$215,685,824	$215,685,824	$215,685,824	$215,685,824	$215,685,824	$215,685,824	$215,685,824	$215,685,824	$215,685,824
Fixed Operating Costs	$20,426,101	$20,426,101	$20,426,101	$20,426,101	$20,426,101	$20,426,101	$20,426,101	$20,426,101	$20,426,101
Total Product Cost	$287,967,204	$289,235,292	$287,728,572	$289,473,924	$288,239,741	$289,235,292	$287,967,204	$289,235,292	$287,728,572
Annual Depreciation									
General Plant Writedown									
Depreciation Charge									
Remaining Value									
Steam Plant Writedown									
Depreciation Charge									
Remaining Value									
Net Revenue	$104,020,712	$102,752,624	$104,259,344	$102,513,992	$103,748,175	$102,752,624	$104,020,712	$102,752,624	$104,259,344
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$104,020,712	$102,752,624	$104,259,344	$102,513,992	$103,748,175	$102,752,624	$104,020,712	$102,752,624	$104,259,344
Income Tax	$21,844,350	$21,578,051	$21,894,462	$21,527,938	$21,787,117	$21,578,051	$21,844,350	$21,578,051	$21,894,462
Annual Cash Income	$82,176,362	$81,174,573	$82,364,882	$80,986,054	$81,961,059	$81,174,573	$82,176,362	$81,174,573	$82,364,882
Discount Factor	0.1228	0.1117	0.1015	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
Annual Present Value	$10,095,035	$9,065,427	$8,362,144	$7,474,689	$6,876,980	$6,191,808	$5,698,384	$5,117,197	$4,720,212
Total Capital Investment + Interest									($2,248,169)
Net Present Worth									

Integrated Biorefinery

DCFROR Worksheet

Year		-2	-1	0	1	2	3	4	5
Fixed Capital Investment		$23,902,581	$179,269,356	$95,610,323					
Land		$1,848,000							
Working Capital				$37,347,783					
Loan Payment					$66,791,051	$66,791,051	$66,791,051	$66,791,051	$66,791,051
Loan Interest Payment		$2,868,310	$24,380,632	$35,853,871	$35,853,871	$33,378,897	$30,705,924	$27,819,114	$24,701,359
Loan Principal		$35,853,871	$304,757,905	$448,173,390	$417,236,210	$383,824,056	$347,738,929	$308,766,992	$266,677,300
Fuel Sales					$90,240,368	$120,320,491	$120,320,491	$120,320,491	$120,320,491
By-Product Credit					$141,036,844	$188,049,126	$188,049,126	$188,049,126	$188,049,126
Total Annual Sales					$231,277,213	$308,369,617	$308,369,617	$308,369,617	$308,369,617
Annual Manufacturing Cost									
Feedstock					$38,712,485	$51,616,647	$51,616,647	$51,616,647	$51,616,647
Periodic Costs (Catalyst, etc)					$16,422,580	$0	$1,280,169	$2,249,147	$1,280,169
Other Variable Costs					$116,478,855	$133,118,692	$133,118,692	$133,118,692	$133,118,692
Fixed Operating Costs					$20,084,934	$20,084,934	$20,084,934	$20,084,934	$20,084,934
Total Product Cost					$191,698,855	$204,820,273	$206,100,441	$207,069,420	$206,100,441
Annual Depreciation									
General Plant Writedown					14%	24.49%	17.49%	12.49%	8.93%
Depreciation Charge					$96,685,057	$165,697,484	$118,336,015	$84,506,394	$60,419,703
Remaining Value					$579,907,365	$414,209,881	$295,873,866	$211,367,473	$150,947,769
Steam Plant Writedown					3.75%	7.22%	6.68%	6.18%	5.71%
Depreciation Charge					$2,638,621	$5,079,521	$4,698,153	$4,346,337	$4,019,851
Remaining Value					$67,724,607	$62,645,086	$57,946,933	$53,600,596	$49,580,745
Net Revenue					($95,599,192)	($100,606,558)	($51,470,916)	($15,371,648)	$13,128,262
Losses Forward						($95,599,192)	($196,205,750)	($247,676,666)	($263,048,314)
Taxable Income					($95,599,192)	($196,205,750)	($247,676,666)	($263,048,314)	($249,920,052)
Income Tax					$0	$0	$0	$0	$0
Annual Cash Income					-$27,212,693	$36,758,293	$35,478,124	$34,509,146	$35,478,124
Discount Factor		1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830	0.6209
Annual Present Value	$425,209,568				-$24,738,812	$30,378,755	$26,655,240	$23,570,211	$22,029,124
Total Capital Investment + Interest		$34,628,858	$224,014,987	$168,811,977					
Net Present Worth				$0					

DCFROR Worksheet

Year	6	7	8	9	10	11	12	13
Fixed Capital Investment								
Land								
Working Capital								
Loan Payment	$66,791,051	$66,791,051	$66,791,051	$66,791,051	$66,791,051	$0	$0	$0
Loan Interest Payment	$21,334,184	$17,697,635	$13,770,161	$9,528,490	$4,947,485	$0	$0	$0
Loan Principal	$221,220,433	$172,127,017	$119,106,127	$61,843,566	$0	$0	$0	$0
Fuel Sales	$120,320,491	$120,320,491	$120,320,491	$120,320,491	$120,320,491	$120,320,491	$120,320,491	$120,320,491
By-Product Credit	$188,049,126	$188,049,126	$188,049,126	$188,049,126	$188,049,126	$188,049,126	$188,049,126	$188,049,126
Total Annual Sales	$308,369,617	$308,369,617	$308,369,617	$308,369,617	$308,369,617	$308,369,617	$308,369,617	$308,369,617
Annual Manufacturing Cost								
Feedstock	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647
Periodic Costs (Catalyst, etc)	$511,169	$3,529,316	$0	$1,280,169	$2,249,147	$1,791,337	$0	$3,529,316
Other Variable Costs	$133,118,692	$133,118,692	$133,118,692	$133,118,692	$133,118,692	$133,118,692	$133,118,692	$133,118,692
Fixed Operating Costs	$20,084,934	$20,084,934	$20,084,934	$20,084,934	$20,084,934	$20,084,934	$20,084,934	$20,084,934
Total Product Cost	$205,331,441	$208,349,589	$204,820,273	$206,100,441	$207,069,420	$206,611,610	$204,820,273	$208,349,589
Annual Depreciation								
General Plant Writedown	8.92%	8.93%	4.46%					
Depreciation Charge	$60,352,044	$60,419,703	$30,176,022					
Remaining Value	$90,595,725	$30,176,022	$0					
Steam Plant Writedown	5.29%	4.89%	4.52%	4.46%	4.46%	4.46%	4.46%	4.46%
Depreciation Charge	$3,718,697	$3,439,355	$3,181,825	$3,139,607	$3,138,904	$3,139,607	$3,138,904	$3,139,607
Remaining Value	$45,862,048	$42,422,694	$39,240,869	$36,101,261	$32,962,358	$29,822,751	$26,683,847	$23,544,240
Net Revenue	$17,633,251	$18,463,336	$56,421,336	$89,601,078	$93,213,808	$98,618,400	$100,410,441	$96,880,421
Losses Forward	($249,920,052)	($232,286,801)	($213,823,466)	($157,402,130)	-$67,801,052	$0	$0	$0
Taxable Income	($232,286,801)	($213,823,466)	($157,402,130)	-$67,801,052	$25,412,756	$98,618,400	$100,410,441	$96,880,421
Income Tax	$0	$0	$0	$0	$5,336,679	$20,709,864	$21,086,193	$20,344,888
Annual Cash Income	$36,247,124	$33,228,977	$36,758,293	$35,478,124	$29,172,467	$81,048,143	$82,463,152	$79,675,140
Discount Factor	0.5645	0.5132	0.4665	0.4241	0.3855	0.3505	0.3186	0.2897
Annual Present Value	$20,460,557	$17,051,719	$17,148,015	$15,046,188	$11,247,249	$28,406,880	$26,275,301	$23,079,050
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	$120,320,491	$120,320,491	$120,320,491	$120,320,491	$120,320,491	$120,320,491	$120,320,491	$120,320,491
By-Product Credit	$188,049,126	$188,049,126	$188,049,126	$188,049,126	$188,049,126	$188,049,126	$188,049,126	$188,049,126
Total Annual Sales	$308,369,617	$308,369,617	$308,369,617	$308,369,617	$308,369,617	$308,369,617	$308,369,617	$308,369,617
Annual Manufacturing Cost								
Feedstock	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647
Periodic Costs (Catalyst, etc)	$0	$1,280,169	$2,760,316	$1,280,169	$0	$3,529,316	$0	$1,791,337
Other Variable Costs	$133,118,692	$133,118,692	$133,118,692	$133,118,692	$133,118,692	$133,118,692	$133,118,692	$133,118,692
Fixed Operating Costs	$20,084,934	$20,084,934	$20,084,934	$20,084,934	$20,084,934	$20,084,934	$20,084,934	$20,084,934
Total Product Cost	$204,820,273	$206,100,441	$207,580,589	$206,100,441	$204,820,273	$208,349,589	$204,820,273	$206,611,610
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	$3,138,904	$3,139,607	$3,138,904	$3,139,607	$3,138,904	$3,139,607	$3,138,904	$1,569,804
Remaining Value	$20,405,336	$17,265,729	$14,126,825	$10,987,218	$7,848,314	$4,708,707	$1,569,804	$0
Net Revenue	$100,410,441	$99,129,568	$97,650,125	$99,129,568	$100,410,441	$96,880,421	$100,410,441	$100,188,203
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$100,410,441	$99,129,568	$97,650,125	$99,129,568	$100,410,441	$96,880,421	$100,410,441	$100,188,203
Income Tax	$21,086,193	$20,817,209	$20,506,526	$20,817,209	$21,086,193	$20,344,888	$21,086,193	$21,039,523
Annual Cash Income	$82,463,152	$81,451,966	$80,282,502	$81,451,966	$82,463,152	$79,675,140	$82,463,152	$80,718,484
Discount Factor	0.2633	0.2394	0.2176	0.1978	0.1799	0.1635	0.1486	0.1351
Annual Present Value	$21,715,125	$19,498,953	$17,471,812	$16,114,837	$14,831,723	$13,027,522	$12,257,622	$10,907,535
Total Capital Investment + Interest								
Net Present Worth								

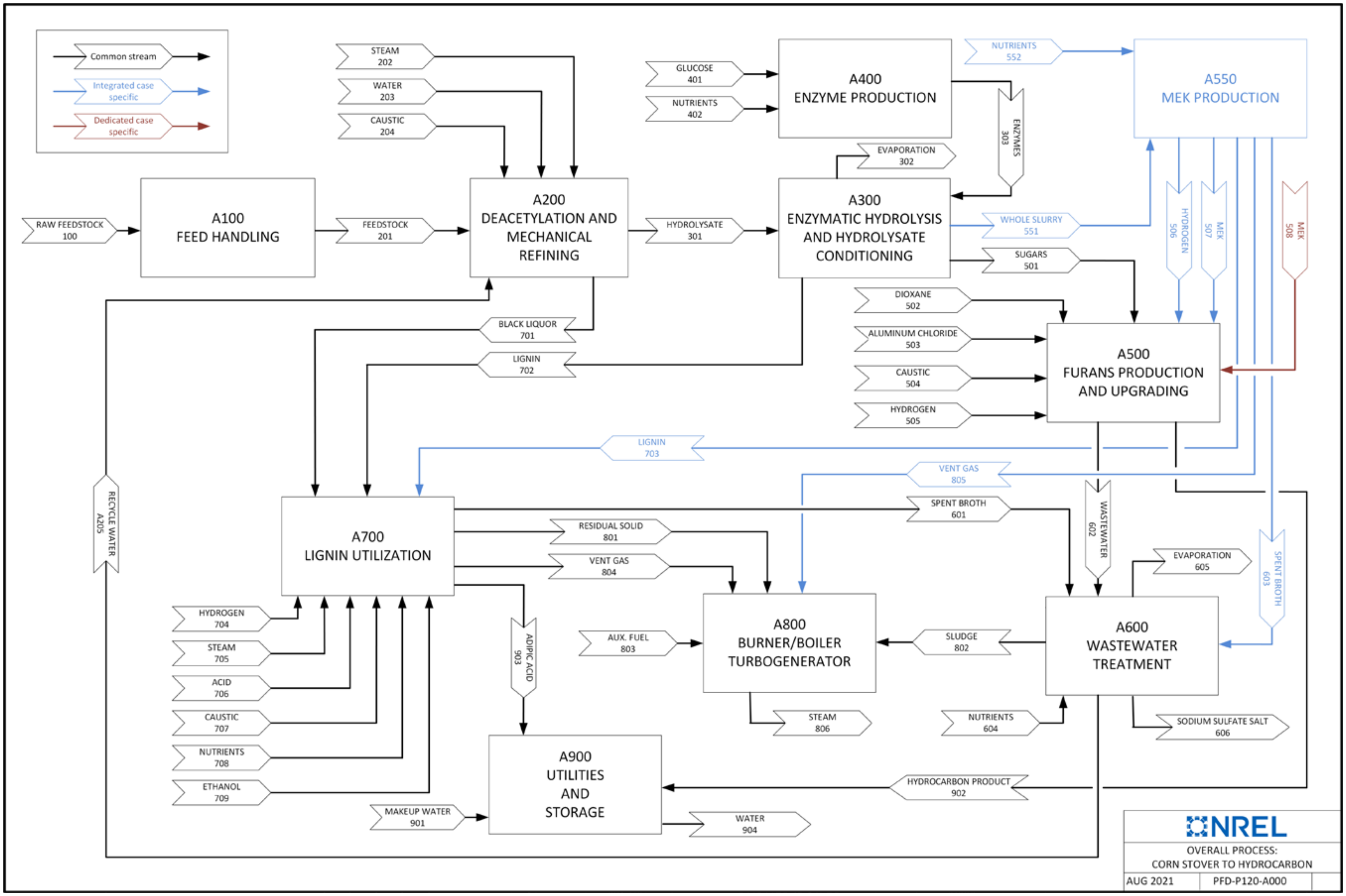
DCFROR Worksheet

Year	22	23	24	25	26	27	28	29	30
Fixed Capital Investment									
Land									($1,848,000)
Working Capital									($37,347,783)
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	$120,320,491	$120,320,491	$120,320,491	$120,320,491	$120,320,491	$120,320,491	$120,320,491	$120,320,491	$120,320,491
By-Product Credit	$188,049,126	$188,049,126	$188,049,126	$188,049,126	$188,049,126	$188,049,126	$188,049,126	$188,049,126	$188,049,126
Total Annual Sales	$308,369,617	$308,369,617	$308,369,617	$308,369,617	$308,369,617	$308,369,617	$308,369,617	$308,369,617	$308,369,617
Annual Manufacturing Cost									
Feedstock	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647	$51,616,647
Periodic Costs (Catalyst, etc)	$2,249,147	$1,280,169	$0	$3,529,316	$511,169	$1,280,169	$2,249,147	$1,280,169	$0
Other Variable Costs	$133,118,692	$133,118,692	$133,118,692	$133,118,692	$133,118,692	$133,118,692	$133,118,692	$133,118,692	$133,118,692
Fixed Operating Costs	$20,084,934	$20,084,934	$20,084,934	$20,084,934	$20,084,934	$20,084,934	$20,084,934	$20,084,934	$20,084,934
Total Product Cost	$207,069,420	$206,100,441	$204,820,273	$208,349,589	$205,331,441	$206,100,441	$207,069,420	$206,100,441	$204,820,273
Annual Depreciation									
General Plant Writedown									
Depreciation Charge									
Remaining Value									
Steam Plant Writedown									
Depreciation Charge									
Remaining Value									
Net Revenue	$101,300,197	$102,269,175	$103,549,344	$100,020,028	$103,038,176	$102,269,175	$101,300,197	$102,269,175	$103,549,344
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$101,300,197	$102,269,175	$103,549,344	$100,020,028	$103,038,176	$102,269,175	$101,300,197	$102,269,175	$103,549,344
Income Tax	$21,273,041	$21,476,527	$21,745,362	$21,004,206	$21,638,017	$21,476,527	$21,273,041	$21,476,527	$21,745,362
Annual Cash Income	$80,027,156	$80,792,649	$81,803,982	$79,015,822	$81,400,159	$80,792,649	$80,027,156	$80,792,649	$81,803,982
Discount Factor	0.1228	0.1117	0.1015	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
Annual Present Value	$9,831,014	$9,022,774	$8,305,198	$7,292,844	$6,829,917	$6,162,676	$5,549,351	$5,093,121	$4,688,068
Total Capital Investment + Interest									($2,246,254)
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%.

Other sugars (SS)	Arabinose, cellobiose, mannose, galactose, sucrose
Other organic acids (SS)	Acetic, succinic, muconic, and lactic acids, extractives
Fermentation nutrients (SS)	Ammonia, CSL, diammonium phosphate, other minor nutrients
Other chemicals (SS)	Aluminum chloride, other minor compounds
Solvent	Dioxane, ethanol
$CO/SO_X/NO_X/H_2S$	NO, NO_2, SO_2, CO, H_2S
Other structural carbohydrates (IS)	Xylan, arabinan, galactan, mannan
Protein (IS)	Corn protein, enzymes, denatured enzymes
Cell mass (IS)	Cell biomass for fermentation, enzyme, and wastewater organisms
Other insoluble solids (IS)	Ash, lime



Common stream
Integrated case specific
Dedicated case specific
STEAM 202
WATER 203
CAUSTIC 204
GLUCOSE 401
NUTRIENTS 402
A400 ENZYME PRODUCTION
NUTRIENTS 552
A550 MEK PRODUCTION
ENZYMES 303
EVAPORATION 302
RAW FEEDSTOCK 100
A100 FEED HANDLING
FEEDSTOCK 201
A200 DEACETYLATION AND MECHANICAL REFINING
HYDROLYSATE 301
A300 ENZYMATIC HYDROLYSIS AND HYDROLYSATE CONDITIONING
WHOLE SLURRY 551
SUGARS 501
HYDROGEN 506
MEK 507
MEK 508
DIOXANE 502
ALUMINUM CHLORIDE 503
CAUSTIC 504
HYDROGEN 505
A500 FURANS PRODUCTION AND UPGRADING
BLACK LIQUOR 701
LIGNIN 702
LIGNIN 703
VENT GAS 805
SPENT BROTH 601
RECYCLE WATER A205
A700 LIGNIN UTILIZATION
RESIDUAL SOLID 801
VENT GAS 804
WASTEWATER 602
EVAPORATION 605
SPENT BROTH 603
HYDROGEN 704
STEAM 705
ACID 706
CAUSTIC 707
NUTRIENTS 708
ETHANOL 709
ADIPIC ACID 903
AUX. FUEL 803
A800 BURNER/BOILER TURBOGENERATOR
SLUDGE 802
A600 WASTEWATER TREATMENT
STEAM 806
NUTRIENTS 604
SODIUM SULFATE SALT 606
A900 UTILITIES AND STORAGE
MAKEUP WATER 901
HYDROCARBON PRODUCT 902
WATER 904
NREL
OVERALL PROCESS:
CORN STOVER TO HYDROCARBON
AUG 2021
PFD-P120-A000

Dedicated Biorefinery

DEDICATED BIOREFINERY		100	201	202	203	204	205	301	302	303	401	402	501	502
Total flow	kg/h	104,167	104,167	17,982	39,884	5,833	155,649	230,647	274,768	7,575	1,557	25,828	86,974	209
Insoluble solids (IS)	kg/h	70,483	70,483					57,306		398		86	86	
Soluble solids (SS)	kg/h	12,850	12,850			5,833		321		22	1,324	1,386	43,363	
MEK	kg/h													
Renewable hydrocarbon	kg/h													
Adipic acid	kg/h													
Water	kg/h	20,833	20,833	17,982	39,884		155,649	173,020	274,768	7,147	234	7,682	43,517	
Glucose (SS)	kg/h										1,324	1,264	27,190	
Xylose (SS)	kg/h												14,230	
Other sugars (SS)	kg/h	642	642					321					1,923	
Other organic acids (SS)	kg/h	12,208	12,208							22		25	21	
Fermentation nutrients (SS)	kg/h											97		
Sulfuric acid (SS)	kg/h													
Sodium hydroxide (SS)	kg/h					5,833								
Sodium sulfate (SS)	kg/h													
Other chemicals (SS)	kg/h													
Soluble lignin	kg/h													
Solvent	kg/h													209
Other organics	kg/h									7		8	7	
Carbon dioxide	kg/h													
Methane	kg/h													
H_2	kg/h													
O_2	kg/h											3,880		
N_2	kg/h											12,777		
$CO/SO_x/NO_x/H_2S$	kg/h											9		
Cellulose (IS)	kg/h	29,205	29,205					28,621					13	
Other struct. carbohydr. (IS)	kg/h	19,948	19,948					17,725					17	
Acetate (IS)	kg/h	1,508	1,508											
Lignin (IS)	kg/h	13,132	13,132					6,960					35	
Protein (IS)	kg/h	2,583	2,583					2,583		23		25	13	
Cell mass (IS)	kg/h									59		25		
Other insoluble solids (IS)	kg/h	4,108	4,108					1,417		317		36	9	

DEDICATED BIOREFINERY		503	504	505	508	601	602	604	605	606	701	702	704	705
Total flow	kg/h	949	1,220	3,341	9,395	174,104	73,209		248,158	13,770	92,868	47,698	426	2,965
Insoluble solids (IS)	kg/h						86				11,669	17,155		
Soluble solids (SS)	kg/h	949	1,220			17,044	2,189			13,770	19,871	2,282		
MEK	kg/h				9,301									
Renewable hydrocarbon	kg/h													
Adipic acid	kg/h													
Water	kg/h				94	156,882	70,716		248,157		61,329	28,260		2,965
Glucose (SS)	kg/h					59						1,431		
Xylose (SS)	kg/h					59						749		
Other sugars (SS)	kg/h					899					321	101		
Other organic acids (SS)	kg/h					452	21				13,716	1		
Fermentation nutrients (SS)	kg/h					5								
Sulfuric acid (SS)	kg/h													
Sodium hydroxide (SS)	kg/h		1,220				1,220				5,833			
Sodium sulfate (SS)	kg/h					15,570				13,770				
Other chemicals (SS)	kg/h	949					949							
Soluble lignin	kg/h					134								
Solvent	kg/h						210							
Other organics	kg/h						7							
Carbon dioxide	kg/h					37								
Methane	kg/h													
H_2	kg/h			3,341		5							426	
O_2	kg/h													
N_2	kg/h					2								
$CO/SO_x/NO_x/H_2S$	kg/h													
Cellulose (IS)	kg/h						13				584	2,506		
Other struct. carbohydr. (IS)	kg/h						17				2,222	3,347		
Acetate (IS)	kg/h													
Lignin (IS)	kg/h						35				6,172	6,925		
Protein (IS)	kg/h						13					2,593		
Cell mass (IS)	kg/h											59		
Other insoluble solids (IS)	kg/h						9				2,691	1,725		

DEDICATED BIOREFINERY		706	707	708	709	801	802	803	804	806	901	902	903	904
Total flow	kg/h	11,560	3,218	855	39	14,896	2,236	6,500	51,290	230,378	248,158	28,139	11,843	522,926
Insoluble solids (IS)	kg/h			36		7,448	447							
Soluble solids (SS)	kg/h	10,751	3,218	746		843	129		12					
MEK	kg/h													
Renewable hydrocarbon	kg/h											27,746		
Adipic acid	kg/h												11,812	
Water	kg/h	809		72		6,598	1,658		1,221	230,378	248,158	393		522,926
Glucose (SS)	kg/h					3								
Xylose (SS)	kg/h					3								
Other sugars (SS)	kg/h					39								
Other organic acids (SS)	kg/h			36		516			12					
Fermentation nutrients (SS)	kg/h			711										
Sulfuric acid (SS)	kg/h	10,751												
Sodium hydroxide (SS)	kg/h		3,218			283	9							
Sodium sulfate (SS)	kg/h						113							
Other chemicals (SS)	kg/h						7							
Soluble lignin	kg/h					6								
Solvent	kg/h				39		1		8				31	
Other organics	kg/h													
Carbon dioxide	kg/h					2			11,345					
Methane	kg/h							6,500						
H_2	kg/h								95					
O_2	kg/h								1,820					
N_2	kg/h								36,789					
$CO/SO_x/NO_x/H_2S$	kg/h													
Cellulose (IS)	kg/h						13							
Other struct. carbohydr. (IS)	kg/h						17							
Acetate (IS)	kg/h													
Lignin (IS)	kg/h						35							
Protein (IS)	kg/h			36		69								
Cell mass (IS)	kg/h					7,378	374							
Other insoluble solids (IS)	kg/h						9							

Integrated Biorefinery

INTEGRATED BIOREFINERY		100	201	202	203	204	205	301	302	303	401	402	501	502	503	504
Total flow	kg/h	104,167	104,167	17,982	39,884	5,833	155,649	230,647	155,009	7,575	1,557	25,828	49,053	120	535	688
Insoluble solids (IS)	kg/h	70,483	70,483					57,306		398		86	49			
Soluble solids (SS)	kg/h	12,850	12,850			5,833		321		22	1,324	1,386	24,462		535	688
MEK	kg/h															
Renewable hydrocarbon	kg/h															
Adipic acid	kg/h															
Water	kg/h	20,833	20,833	17,982	39,884		155,649	173,020	155,008	7,147	234	7,682	24,539			
Glucose (SS)	kg/h										1,324	1,264	15,338			
Xylose (SS)	kg/h												8,027			
Other sugars (SS)	kg/h	642	642					321					1,085			
Other organic acids (SS)	kg/h	12,208	12,208							22		25	12			
Fermentation nutrients (SS)	kg/h											97				
Sulfuric acid (SS)	kg/h															
Sodium hydroxide (SS)	kg/h					5,833										688
Sodium sulfate (SS)	kg/h															
Other chemicals (SS)	kg/h														535	
Soluble lignin	kg/h															
Solvent	kg/h													120		
Other organics	kg/h									7		8	4			
Carbon dioxide	kg/h															
Methane	kg/h															
H_2	kg/h															
O_2	kg/h											3,880				
N_2	kg/h											12,777				
$CO/SO_x/NO_x/H_2S$	kg/h											9				
Cellulose (IS)	kg/h	29,205	29,205					28,621					7			
Other struct. carbohydr. (IS)	kg/h	19,948	19,948					17,725					9			
Acetate (IS)	kg/h	1,508	1,508													
Lignin (IS)	kg/h	13,132	13,132					6,960					20			
Protein (IS)	kg/h	2,583	2,583					2,583		23		25	7			
Cell mass (IS)	kg/h									59		25				
Other insoluble solids (IS)	kg/h	4,108	4,108					1,417		317		36	5			

INTEGRATED BIOREFINERY		505	506	507	551	552	601	602	603	604	605	606	701	702	703	704
Total flow	kg/h	1,798	188	6,180	103,838		163,967	42,167	77,848	55	282,439	13,946	92,868	26,907	10,865	417
Insoluble solids (IS)	kg/h				7,515			49	159				11,669	9,677	7,769	
Soluble solids (SS)	kg/h				19,896		16,598	1,235	1,591	55		13,946	19,871	1,287	59	
MEK	kg/h			5,247					172							
Renewable hydrocarbon	kg/h															
Adipic acid	kg/h															
Water	kg/h			933	76,423		146,883	40,761	74,389		282,438		61,329	15,942	2,714	
Glucose (SS)	kg/h				12,475		48		232					807	9	
Xylose (SS)	kg/h				6,529		53		466					422	17	
Other sugars (SS)	kg/h				882		890		221				321	57	8	
Other organic acids (SS)	kg/h				10		461	12	647				13,716		24	
Fermentation nutrients (SS)	kg/h						5		25	55						
Sulfuric acid (SS)	kg/h															
Sodium hydroxide (SS)	kg/h							688					5,833			
Sodium sulfate (SS)	kg/h						15,140					13,946				
Other chemicals (SS)	kg/h							535								
Soluble lignin	kg/h						133									
Solvent	kg/h							118								
Other organics	kg/h				3		310	4	1,405						324	
Carbon dioxide	kg/h						35		68							
Methane	kg/h															
H_2	kg/h	1,798	188				5									417
O_2	kg/h								63							
N_2	kg/h						2									
$CO/SO_x/NO_x/H_2S$	kg/h															
Cellulose (IS)	kg/h				1,098			7	22				584	1,414	1,076	
Other struct. carbohydr. (IS)	kg/h				1,466			9	29				2,222	1,888	1,437	
Acetate (IS)	kg/h															
Lignin (IS)	kg/h				3,034			20	61				6,172	3,907	2,973	
Protein (IS)	kg/h				1,136			7	24					1,462	1,157	
Cell mass (IS)	kg/h				26				8					33	384	
Other insoluble solids (IS)	kg/h				756			5	15				2,691	973	741	

INTEGRATED BIOREFINERY		**705**	**706**	**707**	**708**	**709**	**801**	**802**	**803**	**804**	**805**	**806**	**901**	**902**	**903**	**904**
Total flow	kg/h	2,791	11,241	2,972	806	38	14,230	5,389	9,700	49,543	11,994	288,713	282,439	15,873	11,525	437,448
Insoluble solids (IS)	kg/h				31		7,115	1,078								
Soluble solids (SS)	kg/h		10,454	2,972	712		833	251		12						
MEK	kg/h										16					
Renewable hydrocarbon	kg/h													15,652		
Adipic acid	kg/h														11,494	
Water	kg/h	2,791	787		63		6,261	4,047		1,177	277	288,713	282,439	222		437,448
Glucose (SS)	kg/h						2									
Xylose (SS)	kg/h						2									
Other sugars (SS)	kg/h						39									
Other organic acids (SS)	kg/h				31		510			12						
Fermentation nutrients (SS)	kg/h				680											
Sulfuric acid (SS)	kg/h		10,454													
Sodium hydroxide (SS)	kg/h			2,972			279	11								
Sodium sulfate (SS)	kg/h							232								
Other chemicals (SS)	kg/h							8								
Soluble lignin	kg/h						6									
Solvent	kg/h					38		2		8					30	
Other organics	kg/h						14	11			54					
Carbon dioxide	kg/h						2			11,023	8,666					
Methane	kg/h								9,700							
H_2	kg/h									95	10					
O_2	kg/h									1,708	716					
N_2	kg/h									35,520	2,254					
$CO/SO_x/NO_x/H_2S$	kg/h															
Cellulose (IS)	kg/h							29								
Other struct. carbohydr. (IS)	kg/h							39								
Acetate (IS)	kg/h															
Lignin (IS)	kg/h							80								
Protein (IS)	kg/h				31		67									
Cell mass (IS)	kg/h						7,048	910								
Other insoluble solids (IS)	kg/h							20								