Technology Case Study: Economic, Sustainability, and Deployment Considerations for Sustainable Aviation Fuels Produced via Lignocellulosic Sugar Catalysis

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1 National Renewable Energy Laboratory
2 Argonne National Laboratory

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Office of Energy Efficiency & Renewable Energy
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NREL/TP-5100-92412
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List of Acronyms

AD	anaerobic digestion
ANL	Argonne National Laboratory
APR	aqueous-phase reforming
BETO	Bioenergy Technologies Office
CCS	carbon capture and sequestration
CEH	continuous enzymatic hydrolysis
CH_4	methane
CI	carbon intensity
CO_2	carbon dioxide
CO_2e	carbon dioxide equivalent
COD	chemical oxygen demand
DAC	direct air capture
DDA	deacetylation and dilute acid
DHOG	dehydration-oligomerization
DMR	deacetylation and mechanical refining
GGE	gallon gasoline equivalent
GHG	greenhouse gas
HDO	hydrodeoxygenation
IECM	Integrated Environmental Control Model
INL	Idaho National Laboratory
IRA	Inflation Reduction Act
IRR	internal rate of return
IS	insoluble solids
ISBL	inside battery limits
LCA	life cycle analysis
LCFS	Low Carbon Fuel Standard
MAC	marginal abatement cost
MFSP	minimum fuel selling price
Na_2CO_3	sodium carbonate
NaOH	sodium hydroxide
NREL	National Renewable Energy Laboratory
R&D GREET	R&D Greenhouse gases, Regulated Emissions, and Energy use in Technologies
RFS	Renewable Fuel Standard
RIN	renewable identification number
SAF	sustainable aviation fuel
SMR	steam methane reforming
SOT	state of technology
TCI	total capital investment
TEA	techno-economic analysis
WWT	wastewater treatment

Executive Summary

For over two decades, the U.S. Department of Energy (DOE) has sponsored critical research on biofuel technology pathways, rooted in and driven by engineering analysis to highlight research directions. The nature and focus areas for analysis have continually evolved over much of that time, beginning with lignocellulosic (second-generation) ethanol, then shifting to hydrocarbons, primarily emphasizing diesel and gasoline upon successful pilot-scale demonstration of ethanol process/cost targets. Again more recently, in light of the changing landscape for light-duty transportation including the adoption of electric vehicles, the DOE Bioenergy Technologies Office (BETO) program has further evolved to focus more specifically on hard-to-electrify transportation modes such as jet and marine fuels, in concert with broader governmental priorities such as the Sustainable Aviation Fuel (SAF) Grand Challenge. Simultaneously, the focus has broadened from a prior context that had historically focused on economic analysis and resultant fuel cost targets as a primary objective, shifting further toward deep decarbonization, with a goal of at least 50%—and ideally 70%—greenhouse gas (GHG) reduction compared to jet and other petroleum fuel benchmarks.

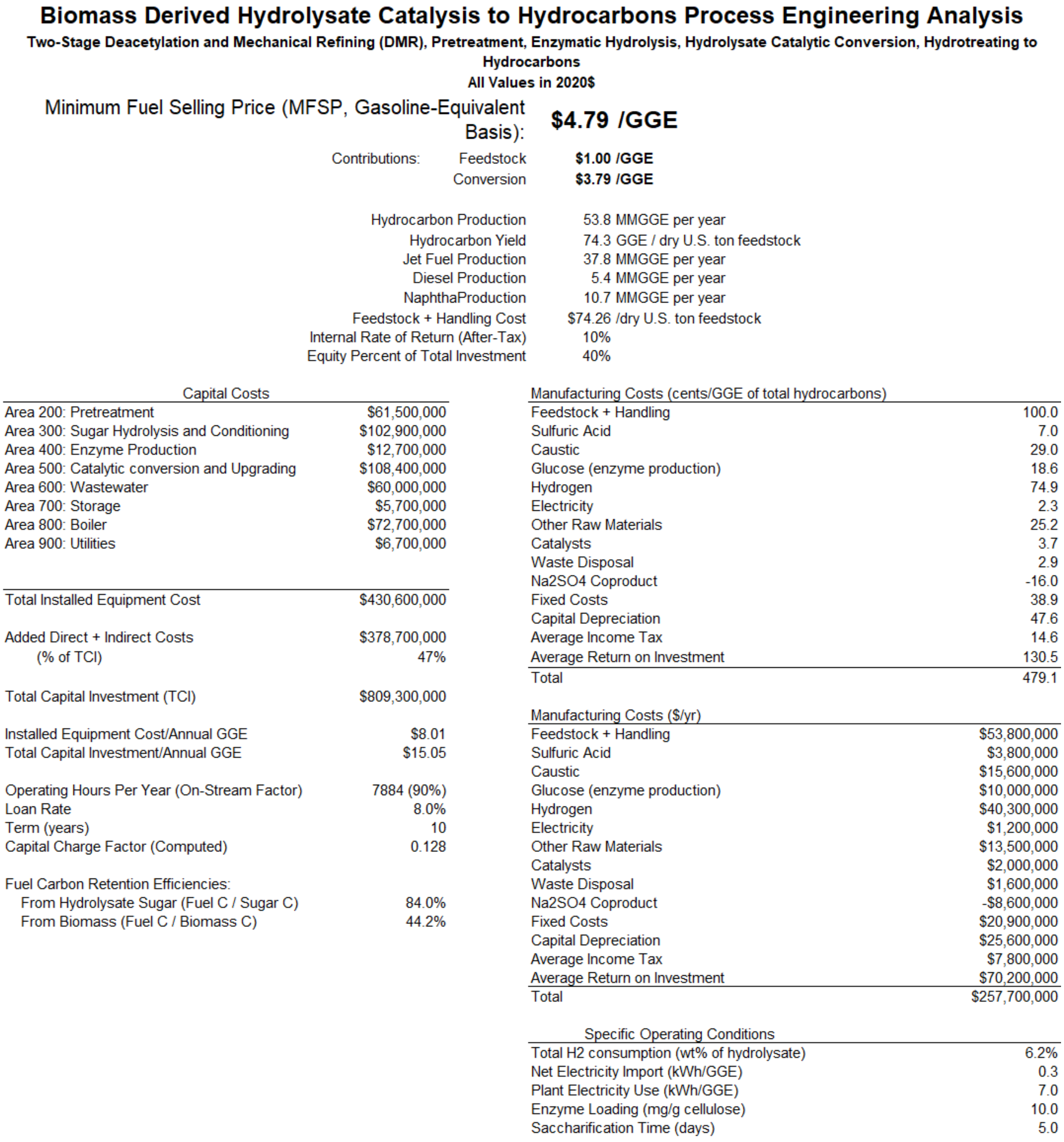
This has prompted a shift away from in-depth "design reports" focusing exclusively on deep-dive engineering design and economic analyses for a single conversion technology coupled with annual state-of-technology benchmarking updates for that technology (recognizing specific design choices reflected in that technology package may or may not align exactly with industry approaches). In its place, DOE national laboratories, including the National Renewable Energy Laboratory (NREL), have been tasked with publishing more concise "technology case studies," which may span multiple technology options and/or design decisions, with less in-depth technical design details but more comprehensive analysis metrics considered simultaneously. This report presents the first of such a series of studies, reflecting one exemplary representative pathway for the conversion of lignocellulosic sugars to SAF via aqueous-phase reforming (APR) catalysis, considered within a broader integrated biorefinery framework based on biochemical processing operations. While far from the only option for converting sugars to SAF, this pathway was selected as a case study here based on its relatively high technology maturity and reasonably simplistic processing approach (avoiding complex separations or other equipment scalability challenges), coupled with the potential for high fuel yields and favorable costs/carbon intensities with opportunities for further near-term optimization.

The biorefinery processing schematic reflected in the present study is based on herbaceous biomass (maintaining corn stover as a representative feedstock) processed through a 2,000-dry-tonne-per-day conversion facility. The biomass undergoes pretreatment, considering two leading options—deacetylation and mechanical refining (DMR), and deacetylation and dilute acid (DDA) pretreatment—followed by batch enzymatic hydrolysis maintaining historical NREL assumptions with on-site enzyme production costs. The hydrolysate undergoes solids removal, concentration, and purification prior to APR catalytic upgrading. The APR operations include sugar hydrogenation/APR, followed by dehydration and oligomerization reactions, and finally hydrotreating. The process yields a hydrocarbon fuel product targeting 70 wt % selectivity to SAF, with 20% to naphtha (gasoline) and 10% to diesel, based on modifying previously published APR studies based on feedback from NREL and industry stakeholders. The biorefinery also includes on-site wastewater treatment, utilities, and combustion of lignin and other residual materials for heat and power generation. We emphasize that this decision for

lignin combustion was made in line with the present study's emphasis on near-term scale-up and deployment, while recognizing that lignin valorization to coproducts or fuels remains an active (and important) area of research over the longer term.

After modeling the biorefinery system in Aspen Plus, coupled with techno-economic analysis (TEA) and life cycle analysis (LCA) following NREL and Argonne National Laboratory (ANL) methodologies, base case minimum fuel selling price (MFSP) metrics were estimated at \$4.79 per gallon gasoline equivalent (GGE) for both DMR and DDA pretreatment, associated with GHG emissions of 63.0 and 57.7 grams of carbon dioxide equivalent (CO_2e) per megajoule, for DMR and DDA respectively. As these metrics do not heavily rely on coproducts to offset costs or GHG emissions, they represent favorable initial base case results, owing to very high fuel yields on the order of 70–74 GGE/dry ton biomass (a more than 50% increase over leading biological conversion pathways given high carbon retention efficiencies). Moreover, incorporating a number of strategies that are implementable in the relatively near term could offer substantial opportunities to reduce GHG intensity, often at nominal cost increases. For example, switching to renewable power as well as on-site production of sodium hydroxide could reduce GHG emissions by 11% (DMR) to 13% (DDA) while remaining within 2% of base case MFSPs. GHG intensity could be further substantially reduced by another 28%–49% (DMR) or 31%–54% (DDA) by subsequently switching the hydrogen source from natural-gas-based steam methane reforming (SMR) to either SMR with CO_2 capture ("blue" hydrogen) or to electrolysis with renewable power ("green" hydrogen), albeit at higher MFSP penalties ranging from \$0.14–\$1.26/GGE depending on the hydrogen cost. Finally, biorefinery carbon capture and sequestration (CCS) could be added at a 16% or lower increase to incremental MFSPs, which offers substantial GHG reduction potential to enable carbon-negative GHG intensity, with a best-case potential of *negative* 60 gCO_2e/MJ (DMR) or *negative* 66 gCO_2e/MJ (DDA) if combined with other strategies noted above (more than 160% GHG reduction from conventional fuel benchmarks).

While the initial base cases do not initially qualify for significant decarbonization-based policy incentives at 29%–35% GHG reduction versus petroleum fuel benchmarks (DMR versus DDA), upon implementing a combination of alternative options described above, GHG improvements can easily exceed 50–60% reduction thresholds triggering more substantial policy credits. As an example, combining renewable power, on-site sodium hydroxide, and green hydrogen, several policy incentives applicable to renewable fuels production could be stacked together to reduce MFSPs by more than \$4/GGE in the most optimal cases. This could potentially increase further to more than \$5.5/GGE net MFSP reduction in a best-case scenario after also including biorefinery CCS, leading to MFSPs well within the range of economic viability compared to conventional fuel benchmarks of \$2.5/GGE while also achieving net-negative SAF production. Even without policy incentives, the marginal GHG abatement costs (a measure of the incremental cost to reduce GHG emissions by combining MFSP and GHG metrics) could be reduced from \$597 to \$718/tonne CO_2e (DDA vs. DMR pretreatment) to as low as \$175–\$179/tonne CO_2e for the same optimal case with green hydrogen and CCS. While there is not a formal carbon credit market in the United States to compare this against, this metric compares favorably to alternative decarbonization technologies such as direct air capture for direct GHG abatement, at near-term cost benchmarks of \$600/tonne CO_2 removed or more.



Biomass Derived Hydrolysate Catalysis to Hydrocarbons Process Engineering Analysis

Two-Stage Deacetylation and Mechanical Refining (DMR), Pretreatment, Enzymatic Hydrolysis, Hydrolysate Catalytic Conversion, Hydrotreating to Hydrocarbons

All Values in 2020$

Minimum Fuel Selling Price (MFSP, Gasoline-Equivalent Basis): **$4.79 /GGE**

Contributions:	Feedstock	**$1.00 /GGE**
	Conversion	**$3.79 /GGE**

Hydrocarbon Production	53.8 MMGGE per year
Hydrocarbon Yield	74.3 GGE / dry U.S. ton feedstock
Jet Fuel Production	37.8 MMGGE per year
Diesel Production	5.4 MMGGE per year
NaphthaProduction	10.7 MMGGE per year
Feedstock + Handling Cost	$74.26 /dry U.S. ton feedstock
Internal Rate of Return (After-Tax)	10%
Equity Percent of Total Investment	40%

Capital Costs	
Area 200: Pretreatment	$61,500,000
Area 300: Sugar Hydrolysis and Conditioning	$102,900,000
Area 400: Enzyme Production	$12,700,000
Area 500: Catalytic conversion and Upgrading	$108,400,000
Area 600: Wastewater	$60,000,000
Area 700: Storage	$5,700,000
Area 800: Boiler	$72,700,000
Area 900: Utilities	$6,700,000
Total Installed Equipment Cost	$430,600,000
Added Direct + Indirect Costs	$378,700,000
(% of TCI)	47%
Total Capital Investment (TCI)	$809,300,000
Installed Equipment Cost/Annual GGE	$8.01
Total Capital Investment/Annual GGE	$15.05
Operating Hours Per Year (On-Stream Factor)	7884 (90%)
Loan Rate	8.0%
Term (years)	10
Capital Charge Factor (Computed)	0.128
Fuel Carbon Retention Efficiencies:	
From Hydrolysate Sugar (Fuel C / Sugar C)	84.0%
From Biomass (Fuel C / Biomass C)	44.2%

Manufacturing Costs (cents/GGE of total hydrocarbons)	
Feedstock + Handling	100.0
Sulfuric Acid	7.0
Caustic	29.0
Glucose (enzyme production)	18.6
Hydrogen	74.9
Electricity	2.3
Other Raw Materials	25.2
Catalysts	3.7
Waste Disposal	2.9
Na2SO4 Coproduct	-16.0
Fixed Costs	38.9
Capital Depreciation	47.6
Average Income Tax	14.6
Average Return on Investment	130.5
Total	479.1

Manufacturing Costs ($/yr)	
Feedstock + Handling	$53,800,000
Sulfuric Acid	$3,800,000
Caustic	$15,600,000
Glucose (enzyme production)	$10,000,000
Hydrogen	$40,300,000
Electricity	$1,200,000
Other Raw Materials	$13,500,000
Catalysts	$2,000,000
Waste Disposal	$1,600,000
Na2SO4 Coproduct	-$8,600,000
Fixed Costs	$20,900,000
Capital Depreciation	$25,600,000
Average Income Tax	$7,800,000
Average Return on Investment	$70,200,000
Total	$257,700,000

Specific Operating Conditions	
Total H2 consumption (wt% of hydrolysate)	6.2%
Net Electricity Import (kWh/GGE)	0.3
Plant Electricity Use (kWh/GGE)	7.0
Enzyme Loading (mg/g cellulose)	10.0
Saccharification Time (days)	5.0

Figure ES-1. Economic summary for APR pathway via DMR pretreatment

Biomass Derived Hydrolysate Catalysis to Hydrocarbons Process Engineering Analysis

Deacetylation and Dilute Acid Pretreatment (DDA), Enzymatic Hydrolysis, Hydrolysate Catalytic Conversion, Hydrotreating to Hydrocarbons
All Values in 2020$

Minimum Fuel Selling Price (MFSP, Gasoline-Equivalent Basis): **$4.79 /GGE**

Contributions:	Feedstock	**$1.05 /GGE**
	Conversion	**$3.74 /GGE**

Hydrocarbon Production	51.0 MMGGE per year
Hydrocarbon Yield	70.4 GGE / dry U.S. ton feedstock
Jet Fuel Production	35.8 MMGGE per year
Diesel Production	5.1 MMGGE per year
NaphthaProduction	10.1 MMGGE per year
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	$58,000,000
Area 300: Sugar Hydrolysis and Conditioning	$87,200,000
Area 400: Enzyme Production	$12,700,000
Area 500: Catalytic conversion and Upgrading	$104,500,000
Area 600: Wastewater	$70,500,000
Area 700: Storage	$5,300,000
Area 800: Boiler	$73,700,000
Area 900: Utilities	$7,100,000
Total Installed Equipment Cost	$419,000,000
Added Direct + Indirect Costs	$363,900,000
(% of TCI)	46%
Total Capital Investment (TCI)	$782,900,000
Installed Equipment Cost/Annual GGE	$8.22
Total Capital Investment/Annual GGE	$15.35
Operating Hours Per Year (On-Stream Factor)	7884 (90%)
Loan Rate	8.0%
Term (years)	10
Capital Charge Factor (Computed)	0.128
Fuel Carbon Retention Efficiencies:	
From Hydrolysate Sugar (Fuel C / Sugar C)	82.9%
From Biomass (Fuel C / Biomass C)	41.9%

Manufacturing Costs (cents/GGE of total hydrocarbons)	
Feedstock + Handling	105.5
Sulfuric Acid	3.3
Caustic	14.6
Glucose (enzyme production)	20.0
Hydrogen	74.8
Electricity	-1.7
Other Raw Materials	23.7
Catalysts	3.6
Waste Disposal	3.4
Na2SO4 Coproduct	-3.9
Fixed Costs	39.4
Capital Depreciation	48.6
Average Income Tax	14.9
Average Return on Investment	132.7
Total	479.1

Manufacturing Costs ($/yr)	
Feedstock + Handling	$53,800,000
Sulfuric Acid	$1,700,000
Caustic	$7,500,000
Glucose (enzyme production)	$10,200,000
Hydrogen	$38,200,000
Electricity	-$800,000
Other Raw Materials	$12,100,000
Catalysts	$1,900,000
Waste Disposal	$1,700,000
Na2SO4 Coproduct	-$2,000,000
Fixed Costs	$20,100,000
Capital Depreciation	$24,800,000
Average Income Tax	$7,600,000
Average Return on Investment	$67,700,000
Total	$244,500,000

Specific Operating Conditions	
Total H2 consumption (wt% of hydrolysate)	6.1%
Net Electricity Export (kWh/GGE)	0.3
Plant Electricity Use (kWh/GGE)	7.6
Enzyme Loading (mg/g cellulose)	10.0
Saccharification Time (days)	5.0

Figure ES-2. Economic summary for APR pathway via DDA pretreatment

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1 Introduction

1.1 Background and Motivation

Over the past 10-20 years, the National Renewable Energy Laboratory (NREL) has published a series of "design reports" to establish reference design cases for the conversion of lignocellulosic biomass to fuels and products. These design cases were structured around detailed engineering design and economic considerations for hypothetical commercial-scale biorefineries—with the approach typically first establishing a base case for a given technology pathway [1], and then updating the design as data and technology learnings evolve [2–5]. In all cases, the design reports were intended to be forward-looking to reflect process performance targets that may plausibly be achieved as research progresses in future years beyond currently demonstrated benchmarks, coupled with a model framework assuming *nth*-plant economics (asserting a sufficient number of biorefineries have already been built and operated to mitigate operational and cost/financial risks that inflate pioneer plant economics).

Within the biochemical conversion technology area broadly based on biomass deconstruction/fractionation to sugars and subsequent upgrading to fuel, NREL first published a series of design reports focused on lignocellulosic ethanol production, reflective of NREL's research focus under the U.S. Department of Energy's Bioenergy Technologies Office (BETO) program in the 2002–2012 time frame [2,3]. Following successful demonstration of ethanol cost targets in 2012 [6], the BETO program shifted focus to fungible hydrocarbon fuels, primarily focused at the time on gasoline and diesel blendstocks. Initial design cases focused on aerobic biological conversion of sugars via lipid/fatty acid intermediates (subsequently upgraded to renewable diesel blendstocks), which were subsequently modified via anaerobic biological conversion to carboxylic acid or butanediol intermediates (again subsequently upgraded through more intensive catalysis steps to hydrocarbon fuels) [5,7]. Additionally, a design case was also established focused on catalytic conversion of sugars to diesel/gasoline fuels via aqueous-phase reforming (APR) [4]. Across all pathways, the primary focus of the work was on techno-economic analysis (TEA) to prioritize designs toward achieving minimum fuel selling price (MFSP) goals set by the BETO program.

While the above-described design report series proved highly useful as public reference cases for integrated biorefinery TEA modeling and to guide research priorities around *cost* goals [8], they placed less emphasis on other metrics such as greenhouse gas (GHG) emissions or sustainable process design for near-term deployment (e.g., process simplification or sourcing choices for key input materials). More recently, the BETO program has further pivoted to prioritize deep decarbonization of at least 70% GHG emissions reduction compared to petroleum fuel benchmarks, particularly emphasizing commercial deployability for pathways maximizing sustainable aviation fuel (SAF) outputs [9,10]. MFSP and other TEA metrics continue to also be important in demonstrating reasonable economics, but are no longer tied to specific cost targets, recognizing the multitude of dynamic factors that can influence a technology's market competitiveness (including but not limited to petroleum prices, economic conditions, costs of input raw materials, values of products/coproducts, and availability/application of policy incentives).

In light of these renewed priorities, this report documents an *example case study* for one promising technology pathway which exhibits good potential to meet high GHG reduction levels under reasonable costs, enabled by high carbon retention/fuel yields and relatively simplistic processing steps as may be readily scaled to commercial deployment. The selected pathway of focus here is an updated version of the APR design case noted above, now modified to maximize selectivity to SAF fuel blend components and incorporating additional modifications/refinements to other key biorefinery steps relative to the above-cited 2015 APR design report. Also relative to that and other previous design reports, this study focuses less on a discussion of technical/design details and more on a broader reporting of analysis metrics and their interplays across economic, sustainability, and deployment considerations, ultimately with a goal to highlight an example path to achieving ≥70% GHG emission reductions for SAF and other fuels relative to petroleum benchmarks. While some considerations are still made to evaluate specific technology options and their tradeoffs around base case cost/GHG metrics (for example investigating two pretreatment methods within the overall integrated biorefinery schematic), the analysis shows that from a high-level business perspective, such nuanced technical details are largely outweighed by the need to incorporate larger wholesale decarbonization approaches to the overall biorefinery in order to ultimately reach viability. Future reports will document similar strategies for other case study examples, including technologies representing longer-term but high-novelty research opportunities.

1.2 Process Overview

On a high level, the modeled process employs unit operations largely consistent with those described in the 2015 APR report [4], though with some modifications to operating conditions/yields for a number of key steps as documented in subsequent sections of this report. We refer the reader to the 2015 report and other design cases cited herein for a more complete discussion of the process/design details, which otherwise remain the same in this study. The process is divided into nine areas (Figure 1).

Area 100: Feed handling. Corn stover biomass feedstock is delivered to the biorefinery following upstream biomass preprocessing steps (grinding/milling), with all preprocessing costs included in the delivered feedstock price as required to meet the feedstock compositional specifications outlined in Section 2.1.

Area 200: Pretreatment and conditioning. Two options are evaluated for biomass pretreatment in this study. The first employs deacetylation and dilute acid (DDA) similar to the 2015 study, wherein a mild alkaline soaking step is utilized primarily to remove acetate along with fractions of lignin and other components of the biomass, followed by treatment with acid and steam to partially deconstruct hemicellulose carbohydrates to sugars. The second option employs deacetylation and mechanical refining (DMR), which utilizes a slightly higher-severity alkaline extraction step to remove additional lignin followed by mechanical refining to physically pretreat the biomass with minimal carbohydrate hydrolysis.

Area 300: Enzymatic hydrolysis, hydrolysate conditioning, and purification. Enzymatic hydrolysis is performed consistently with the 2015 APR report, based on a continuous liquefaction step followed by batch saccharification using enzymes produced on-site. The hydrolysate is clarified to remove lignin and other solids using a vacuum filter press with added flocculant and wash water, then concentrated in a mechanical vapor recompression evaporator. The clarified/concentrated sugars are then purified through a polishing filter to remove residual solids, followed by ion exchange to remove cations/anions for downstream catalyst protection.

Area 400: Cellulase enzyme production. On-site enzyme production is again maintained in this design consistent with prior NREL design reports, utilizing purchased glucose to synthesize cellulase/hemicellulase enzymes via aerobic fed-batch bioconversion.

Area 500: Catalytic conversion and upgrading. The catalysis section employs a similar series of reaction steps as documented in the 2015 APR report, albeit under modified conditions to focus on maximizing selectivity to SAF (albeit still producing diesel and gasoline fuel cuts as well). In summary, there are four reaction stages condensed into three separate reactor vessels, namely hydrogenation and APR of sugars (Reactor 1), condensation and oligomerization (Reactor 2), and hydrotreating (Reactor 3). Each stage consists of packed-bed reactor vessels with hydrogen addition, allowing for oxygen rejection as water rather than carbon dioxide (CO_2).

Area 600: Wastewater treatment (WWT). Wastewater streams are treated by anaerobic and aerobic digestion, generally following the same schematic as employed in the 2015 APR report, though now including a dryer and centrifuge to crystallize the recovered salt (primarily sodium sulfate) for sale as a coproduct, similar to NREL's 2018 design report [5].

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

Area 800: Combustor, boiler, and turbogenerator. The solids from the filter press and WWT are combusted along with the biogas from anaerobic digestion and process off-gases to generate high-pressure steam for electricity production and process heat. The boiler produces more steam than is required for heating, with the surplus converted to electricity through a turbogenerator to partially offset net power import demands. All details for Area 800 are also maintained consistently with the latest 2018 design case [5].

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

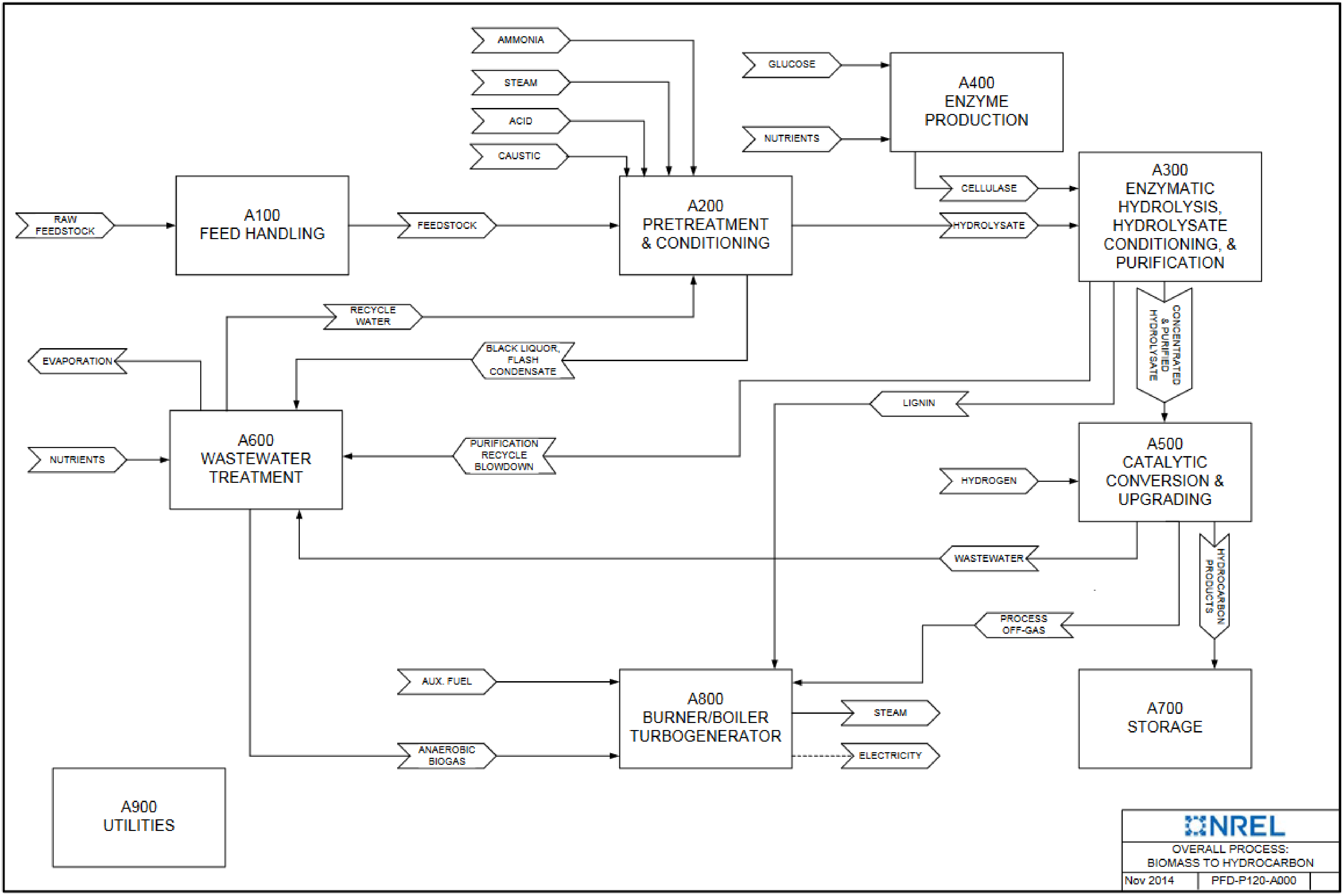


Figure 1. Simplified flow diagram of the overall process

1.3 Analysis Approach and Basis

While encompassing a broader set of metrics than prior design cases, the overall approach to the analysis remains consistent with those described previously and will not be repeated in such detail here [3,5]—namely, the foundation of the analysis is rooted in a thermodynamically rigorous process simulation model conducted in Aspen Plus [11]. This step remains crucial in informing mass and energy flow rates at the unit and biorefinery level as needed to establish capital/operating costs, life cycle inventories, yields, and CO_2 emissions. In turn, that information is used to conduct the following analyses (following information flows as depicted in Figure 2):

- **Techno-economic analysis:** TEA is conducted following standard NREL methodologies as described in prior design cases [3,5]. In summary, unit-level flow rates are used to determine the number and size of processing equipment, whose capital costs are then estimated by scaling from prior cost estimates sourced from vendors/subcontractors, costing software (e.g., Aspen Capital Cost Estimator), or engineering literature, while operating costs are based on unit prices and quantities for raw material inputs and outputs. Indirect cost factors are applied on top of the direct capital cost figures to yield total capital investment (TCI), which then may be combined alongside operating and labor/maintenance costs to be run through cash flow calculations to ultimately solve for the MFSP required to achieve a net present value (NPV) of zero for a target internal rate of return (IRR) (maintained here at a fixed 10% IRR target).

- **Life cycle analysis (LCA):** LCA is now explicitly included in this study's analysis scope, rather than deferred for subsequent consideration, as a key priority to quantify the pathway's carbon intensity (CI). LCA modeling is conducted by Argonne National Laboratory (ANL) collaborators using the R&D Greenhouse gases, Regulated Emissions, and Energy use in Technologies (R&D GREET) model [12]. The R&D GREET model incorporates life cycle inventories for all biorefinery inputs and outputs from the Aspen process models (i.e., materials, energy, waste outputs, products, and coproducts) to calculate direct (biorefinery) and indirect (embodied) GHG emissions, fossil energy consumption, and water footprint based on a field-to-gate system boundary. Further discussion on LCA modeling methodology is provided in Section 3.2.

- **Marginal abatement cost (MAC):** The marginal cost of GHG abatement ties together the TEA and LCA results into a single metric, measuring the incremental cost of reducing GHG emissions relative to a reference case (petroleum fuels). This is a useful metric that may in turn be compared to other GHG abatement strategies when GHG reduction is a primary goal (or alternatively may be viewed as a CO_2 mitigation credit required to reach cost parity with the incumbent technology, under an alternative carbon credit market scenario). MAC is calculated as defined in Eq. 1:

$$Marginal\ GHG\ abatement\ cost = \frac{cost\ of\ biofuel\ \left(\$/_{MJ}\right) - cost\ of\ reference\ fuel\ \left(\$/_{MJ}\right)}{GHG\ emissions\ of\ reference\ fuel\ \left(^{g\ CO_2e}/_{MJ}\right) - GHG\ emissions\ of\ biofuel\ \left(^{g\ CO_2e}/_{MJ}\right)} \qquad \text{(Eq. 1)}$$

- **Policy incentives:** Finally, this study includes consideration for the incorporation of applicable policy incentives as may reduce the MFSP attributed to decarbonization of the fuel supply chain. Past NREL TEA efforts have generally excluded policy incentives, as they can (and often do) fluctuate both in terms of the specific applicable credits and the value of such credits. However, in light of further momentum building on this topic over recent years in terms of new or expanded policy incentives for producing SAF and other renewable fuels, coupled with practical considerations that commercial biorefineries would capitalize on such incentives in reality, this study reports MFSPs both with and without inclusion of applicable policy credits given the study's emphasis on commercial deployment drivers. Many such incentives as currently structured are based on the degree of GHG reduction relative to petroleum benchmarks, discussed further in Section 3.4.

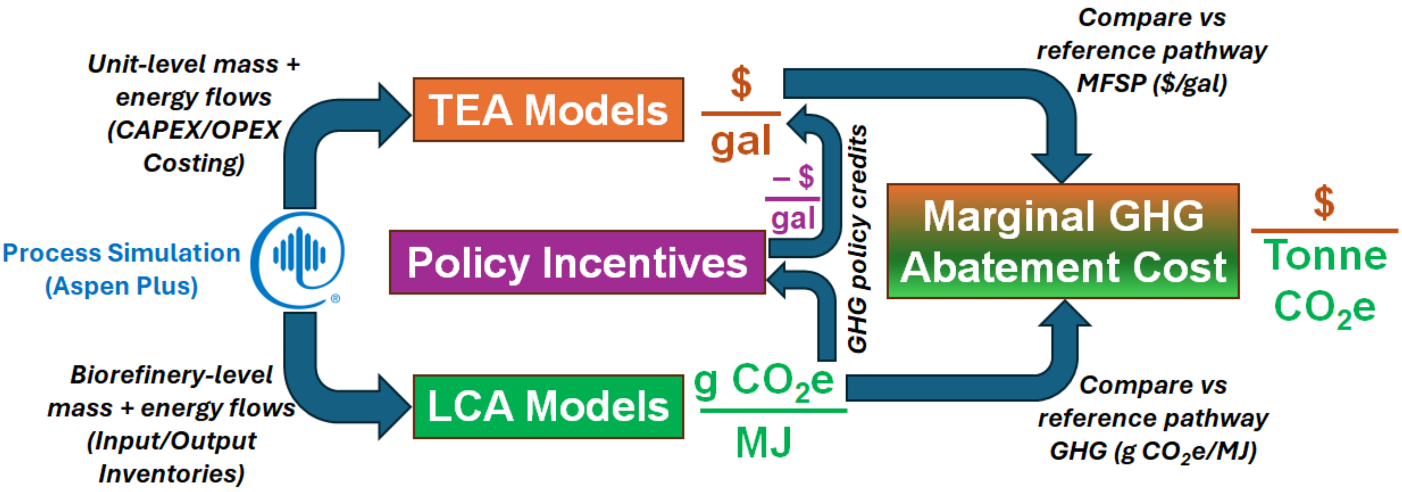


Figure 2. Information flows connecting analysis stages evaluated in this study

As in prior design reports, this study continues to assume an *n*th-plant basis for TEA modeling (and associated tie-ins to MAC and policy incentive calculations). As noted above and described in further detail in past reports [3], *n*th-plant economics reflect commercially mature technology that is generally well understood and does not pose out-of-the-ordinary risks for construction and operation for the scale at which it is designed to run. This is in contrast to first-of-a-kind or pioneer plants, which may incur any number of economic challenges tied to (for example) higher downtime caused by unplanned interruptions or shutdowns, non-steady-state operation leading to equipment upsets or running below capacity, the need to overdesign key equipment and/or include equipment redundancies to ensure continuous operation, higher financing costs and/or higher required rate of return, etc. It is beyond the scope of the current study to explore *n*th versus pioneer plant trade-offs as this is difficult to objectively quantify without commercial data or industry guidance, but future work may at least begin investigating specific considerations such as overdesign for key equipment.

Also consistent with prior recent design cases, given the production of multiple fuel cuts from the modeled pathway beyond SAF alone, all fuel outputs (naphtha, jet, and diesel range) are combined together on a lower heating value basis to be reported as gallons gasoline equivalent

(GGE) at a standard value of 116,090 Btu/gal gasoline. While SAF is the primary focus of this work and not gasoline, GGE remains a common basis for reporting fuel yields across the BETO program, also recognizing the naphtha and diesel fuel products remain important in decarbonizing road and marine transportation as well. However, for additional granularity, the yields of all individual fuel products are reported, as well as resultant \$/gal selling prices corresponding to overall \$/GGE MFSPs. Finally, the biorefinery scale is again fixed at 2,000 dry metric tonnes per day of corn stover processed through the biorefinery at a 90% operating uptime, in keeping with prior historical analyses.

2 Process Design and Cost Estimation Details

A simplified block diagram schematic of the modeled biorefinery process is depicted in Figure 3, followed by a brief summary of key process steps/design considerations. Again, these discussions are intentionally kept more concise, particularly for process sections that remain largely consistent with prior design reports, and we refer to those reports for a more thorough technical discussion [3–5].

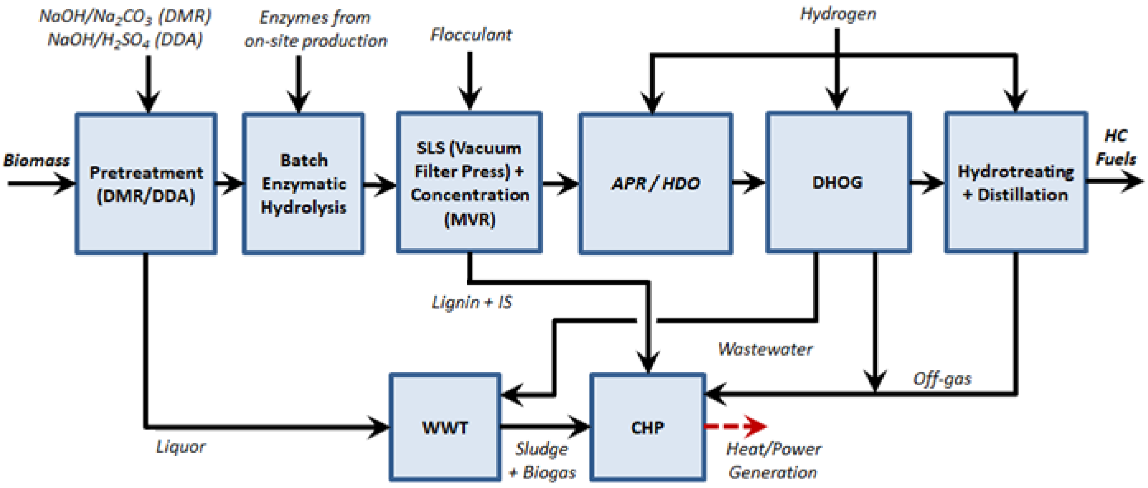


Figure 3. Block flow diagram for the APR-to-SAF pathway

2.1 Area 100: Feedstock Logistics and Handling

All details regarding biomass feedstock scale, composition, and cost as delivered to the biorefinery remain unchanged relative to prior design reports [4,5]. This acknowledges, however, that work continues to progress in investigating and optimizing biomass preprocessing logistics via collaborators at Idaho National Laboratory (INL), to identify optimal preprocessing steps to meet or exceed the feedstock material attribute specifications and/or to further reduce future feedstock cost targets. For example, recent work at INL has investigated the addition of an air classification step to fractionate whole corn stover into its constituent anatomical fractions (stalks, cobs, husks, and leaves), finding beneficial reductions in grinding energy demand which can offset the incremental cost of the fractionation step, translating to minimal or no net increase in delivered feedstock cost while meeting or surpassing material attribute specifications [8,13]. Such specifications primarily include $\geq$59 wt % carbohydrates and $\leq$4.9 wt % ash, as well as 20 wt % moisture content and a particle size of ¼-inch following hammer mill grinding [13]; additional compositional details are summarized in Table 1. Under the above example, fractionated tissues of corn stover could either be sequentially processed under biorefinery campaigns, ideally tailoring operating conditions to suit each respective fraction [14], or could be recombined at optimal ratios for conversion. Such options are being evaluated under the Feedstock-Conversion Interface Consortium, and findings from that work may subsequently be leveraged in future iterations of technology design case studies such as this.

All such costs for preprocessing logistics are rolled into the delivered cost of biomass feedstock at the throat of the biorefinery pretreatment reactor (deacetylation in this case). The feedstock cost was maintained at $71.26/dry U.S. ton as originally established in 2016 dollars, representing 2030 feedstock cost targets consistent with preprocessing projections described in NREL's most recent 2018 design report [5]. However, this cost was updated to 2020 dollars for consistency with this report's cost-year basis, using the U.S. Bureau of Labor Statistics' Producer Price Index for corn products [15]. The delivered feedstock cost reflects NREL's standard n^{th}-plant biorefinery scale of 2,205 dry U.S. tons/day (2,000 dry metric tonnes/day) as evaluated in prior design cases, equating to 724,000 dry U.S. tons/year at a 90% facility uptime. Also as in prior studies, the feedstock is again assumed to be primarily corn stover (either as whole biomass or fractionated tissues as discussed above), though with the possibility of blending with other herbaceous feedstocks (e.g., switchgrass) or waste materials (e.g., municipal solid waste) to reduce costs and/or carbon intensity. However, any such savings in using blended feedstocks would need to be weighed against impacts on convertibility or operating severity incurred through pretreatment and enzymatic hydrolysis.

Table 1. Delivered Feedstock Composition Assumed in the Present Design [3–5]

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.

2.2 Area 200: Pretreatment

2.2.1 Overview

Consistent with prior NREL design cases, the modeled biorefinery begins with delivered biomass feedstock to the pretreatment section (Area 200), again with all preprocessing/handling logistics (Area 100) included in the feedstock price as noted above. In earlier design cases [3,4,7], pretreatment was based on dilute acid pre-hydrolysis, which later evolved into deacetylation and dilute acid (DDA) utilizing a mild deacetylation step to remove acetate along with minor amounts of lignin and other biomass constituents [6]. Dilute acid/DDA is a relatively well-studied pretreatment process, and has generally been demonstrated to achieve favorable yields with low operating costs and moderate capital costs [6]. However, its elevated operating temperature/pressure (160°C and 5.5 atm) requires the use of screw feeders to feed biomass across a pressure envelope. This can become a key bottleneck in the process, facing a host of operational reliability challenges in both feeding heterogeneous biomass like corn stover, as well as plugging and rapid wear of the screw feeder, which can lead to shutdowns or high maintenance costs at commercial scale [16,17].

In light of the above challenges, more recently NREL began focusing research on deacetylation and mechanical refining (DMR) pretreatment, which operates at ambient pressure and can thus avoid pressurized feeding challenges encountered for dilute acid treatment [5]. Compared to DDA, DMR typically employs somewhat higher-severity "deacetylation" (alkaline extraction) conditions, primarily in its caustic loading, as it aims to remove a higher fraction of lignin to enable good downstream enzymatic hydrolysis performance. Mechanical refining is a two-stage operation utilizing disc refiners followed by roller milling, though research has investigated optional removal of the secondary milling stage. DMR also has been shown to achieve very high yields with similar capital costs as DDA but higher operating expenses, tied primarily to elevated usage of caustic (typically sodium hydroxide [NaOH]) [18]. DMR also faces higher CI challenges tied to high power demands for mechanical refining, as well as GHG-intensive NaOH demands, although recent research has identified a way to partially mitigate this challenge through the use of two-stage deacetylation with sodium carbonate (Na_2CO_3) as a mild caustic, allowing lower requirements for NaOH [19]. In light of the above trade-offs for these two leading pretreatment technologies, this study considers both DDA and DMR options within the broader integrated biorefinery framework.

2.2.2 Deacetylation and Dilute Acid

While DDA pretreatment has not been the subject of substantial research focus at NREL in recent years, it was revisited in this study to more directly consider TEA/LCA trade-offs alongside DMR, while being mindful of its operational reliability challenges at scale noted above. Table 2 summarizes key inputs for DDA pretreatment assumed here as reasonable near-term parameters, based on extrapolating from prior NREL data (e.g., the 2012 ethanol pilot plant demonstration [6]). This includes deacetylation operated at 20 wt % total solids using 30 kg of NaOH per dry tonne biomass, followed by dilute acid treatment at 160°C and 30 wt % solids using 9 kg/dry tonne loading of sulfuric acid. Expected solubilization yields into the liquor phase are also summarized below. While differences in operational reliability between DDA and DMR (noted above) are not explicitly reflected given a lack of operational data for either operation at commercial scale, such implications may be deduced from the sensitivity analysis to the biorefinery on-stream factor presented later in this report (Section 4.2.1).

Table 2. Key Parameters for DDA Pretreatment

Parameter	Current Design
Deacetylation caustic loading	30 kg NaOH/dry tonne biomass
Total solids content	20 wt % deacetylation, 30 wt % dilute acid pretreatment
Dilute acid pretreatment temperature	160°C
Dilute acid pretreatment acid loading	9 kg sulfuric acid/dry tonne biomass
Deacetylation solubilizations to black liquor	
Extractives	100%
Ash	75%
Acetate	*88%*
Lignin	20%
Xylan	2%
Glucan	0%
Arabinan	0%

2.2.3 Deacetylation and Mechanical Refining

For DMR pretreatment, as noted above, recent NREL research has moved to a two-stage deacetylation process to reduce the requirements for expensive and GHG-intensive NaOH, via a pre-extraction with Na_2CO_3 as a less costly and lower-GHG alkali. Under this configuration, Na_2CO_3 acts as a "sacrificial alkali" to neutralize acetate and other acidic constituents in the biomass feedstock, allowing NaOH to be more efficiently utilized for delignification. This generates two separate liquor streams from the two alkali extraction steps; while these could be processed differently, in the current design both liquor streams are combined and routed to WWT. Following a number of NREL experimental trials, the optimal conditions identified to date utilize a loading of 80 kg Na_2CO_3 plus 24 kg NaOH per dry tonne biomass [20]. Both deacetylation and mechanical refining are operated at 20 wt % total solids in this design as envisioned to be performed at commercial scale and supported by pilot-scale data (though bench-scale NREL data is often performed under more dilute conditions) [21], maintaining mechanical refining at a total power usage of 200 kWh/dry tonne of biomass (based on previous external guidance for optimal mechanical refiner electricity consumption at commercial scale). Table 3 summarizes the key parameters for the DMR process applied here, including targeted liquor solubilizations.

Table 3. Key Parameters for DMR Pretreatment

Parameter	Current Design
Deacetylation caustic loading	80 kg Na_2CO_3 + 24 kg NaOH/dry tonne
Total solids content	20 wt % deacetylation and mechanical refining
Mechanical refining power demand	200 kWh/dry tonne biomass
Deacetylation solubilizations to black liquor	
Extractives	100%
Ash	11%
Acetate	100%
Lignin	20%
Xylan	10%
Glucan	2%
Arabinan	30%

2.3 Area 300: Enzymatic Hydrolysis and Hydrolysate Conditioning

The pretreated solids slurry is cooled to 50°C and routed to enzymatic hydrolysis, maintaining consistent details for this step as documented in prior reports [3,4]. In summary, saccharification is initiated using a 24-hour liquefaction tank (continuous vertical vessel) followed by batch hydrolysis in stirred-tank reactors. Hydrolysis is conducted at 20 wt % total solids for a total of 5 days (1 day continuous plus 4 days batch), although it is possible to conduct hydrolysis at 25 wt % solids as a longer-term goal as projected in the latest 2018 design report [5]. Also as described in the 2018 report, an alternative approach continues to offer a novel opportunity to reduce saccharification costs in the future via continuous enzymatic hydrolysis (CEH), employing continuous circulation of hydrolysate through microfiltration/ultrafiltration membranes. The CEH approach may enable key operational/cost reductions relative to batch enzymatic hydrolysis by (1) continuously removing sugars as they are produced, improving kinetics by reducing feedback inhibition, (2) retaining and recycling enzymes to be used more than once as in the batch enzymatic hydrolysis approach, and (3) enabling *in situ* solids removal from sugars as part of the CEH operation, saving on costly downstream solids clarification equipment and added flocculant demands when solids must otherwise be removed following batch enzymatic hydrolysis. Although promising, given the focus in the present study prioritizing near-term deployment at commercial scale, CEH was not selected for use in the current analysis, as it has not yet been demonstrated at a scale beyond a 30-L pilot skid.

Batch hydrolysis conditions for the DMR case are based on recent NREL state-of-technology (SOT) results [8,20], applying a total enzyme loading of 10 mg/g cellulose (including both cellulase and hemicellulase enzymes at an 80/20 ratio) to achieve 88%, 95%, and 80% conversion of glucan, xylan, and arabinan to their respective sugar monomers. This was extrapolated from the same two-stage deacetylation approach for DMR—i.e., 80 g Na_2CO_3 and 24 g NaOH per kilogram dry biomass—with reasonable adjustments to the data in the context of future targets (xylose yields are slightly higher than 93% SOT basis at these DMR loadings, as original experimental data indicated up to 96% yields, and arabinose yields were adjusted lower versus 91% per the 2021 SOT, as this parameter was found to subsequently be reduced to 70% in more recent SOT cases, with the caveat that arabinose measurements are not routinely scrutinized to the same level of detail). For the DDA case, batch hydrolysis parameters were set at 80%, 80%, and 75% conversion of glucan, xylan, and arabinan to sugars, respectively, at a 10-mg/g enzyme loading, based on extrapolating to reasonable targets relative to older 2012 SOT data (the last publicly available DDA data from NREL trials) that used older generations of enzyme packages at that time [6].

The resulting hydrolysate is next routed to a solid/liquid separation step to clarify the sugars. This is performed using vacuum belt filtration with wash water and added flocculant as necessary to remove solids while minimizing sugar losses to the filter cake. Based on recent laboratory experience, this operation requires more costly operating conditions to achieve a targeted 95% sugar retention when processing hydrolysate pretreated by DMR versus DDA, given finer solids particles generated from DMR. Accordingly, the filter press employs a wash ratio of 17.5 L/kg insoluble solids (IS), flocculant loading of 20 g/kg IS, and filter capacity of 12 kg IS/$m^2 \cdot h$ for the DMR case, relaxed to 10 L/kg, 10 g/kg IS, and 20 kg IS/$m^2 \cdot h$, respectively, for the DDA case. Next, the clarified sugar stream is concentrated using a mechanical vapor recompression evaporator, maintaining all design/cost parameters for that operation consistent with prior studies [5]. The evaporator reduces the water content in the clarified hydrolysate stream to 50 wt %, which requires additional evaporation for DMR starting at 93 wt % water (compared to DDA at 90%) given higher wash ratios required in the filtration step. Finally, the concentrated sugar stream is sent to microfiltration to remove residual fine particulates followed by ion exchange to remove cations/anions for downstream catalyst protection, consistent with all details presented in the 2015 APR design report [4].

All key parameters for enzymatic hydrolysis and hydrolysate conditioning are summarized in Table 4.

Table 4. Key Parameters for Enzymatic Hydrolysis, Clarification, and Concentration

Parameter	Current Design
Enzymatic Hydrolysis	
Temperature	50°C
Residence time	5 days (1 continuous + 4 batch)
Total solids loading (wt %)	20%
Enzyme loading (mg/g cellulose)	10/10 (DMR/DDA)
Glucan to glucose conversion	88%/80% (DMR/DDA)
Xylan to xylose conversion	95%/80% (DMR/DDA)
Arabinan to arabinose conversion	80%/75% (DMR/DDA)
Hydrolysate Clarification (Vacuum Belt Filtration)	
Sugar recovery in liquor	95%
Wash ratio (L/kg IS)	17.5/10 (DMR/DDA)
Flocculant loading (g/kg IS)	20/10 (DMR/DDA)
Filter capacity (kg IS/m^2·h)	12 / 20 (DMR/DDA)
Hydrolysate Concentration (Mechanical Vapor Recompression Evaporation)	
Water content in concentrated hydrolysate (wt %)	50%
Polishing Filtration (Microfilter)	
Filtrate maximum particle size (microns)	0.1
Filter recovery rate of sugars/soluble solids	90%
Ion Exchange	
Bed type	Separate bed (strong acid cation/weak base anion)
Caustic (NaOH)/acid (HCl) regeneration demand (g/kg soluble solids feed)	12.8/22.4

2.4 Area 400: Enzyme Production

Consistent with prior design cases, the current study again maintains the assumption for on-site production of enzymes to support the saccharification step. Also as noted in those prior reports [3,4], this is not a suggestion that biorefineries would or should follow such a practice, as it is more likely that commercial enzyme needs will be met by commercial enzyme suppliers (particularly in the near term). Rather, it is handled this way in an effort to capture the true cost of enzyme production in a transparent way based on underlying capital and operating costs. Thus, other costs for off-site commercial sourcing are also not represented here, including licensing/royalty fees or stabilization/transportation costs.

The enzyme production module is again left unchanged from prior reports, consisting of 300,000-L (80,000-gal) aerobic bioreactor tanks operating in fed-batch mode, producing enzyme protein via a *T. reesei* cell host. Purchased glucose (corn syrup) is utilized as the carbon source, partially converted to sophorose to induce enzyme production. Aeration rate is set by oxygen transfer rate demands, in turn a function of bioreactor vessel design and productivity, as described in detail previously [3,4]. Four trains of three seed bioreactors provide inoculum to the main enzyme production bioreactors, with the latter producing enzyme protein at a productivity rate of 0.3 g/L·h, thus requiring five total production bioreactors with a total enzyme loading of 10 mg/g for both cases. Additional nutrients are supplied at consistent ratios with those assumed in prior studies, including ammonia, corn steep liquor, and sulfur dioxide, along with corn oil, which acts as an anti-foaming agent. Temperature is maintained at 28°C using chilled water. Key parameters for on-site enzyme production are summarized in Table 5.

Table 5. Key Parameters for On-Site Enzyme Production System

Parameter	Assumption
Protein loading to enzymatic hydrolysis	10 mg protein/g cellulose
Reactor size	300,000 L at 80% final working volume (fed-batch)
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 (0.3-g/L·h productivity based on total cycle time)

2.5 Area 500: Hydrolysate Catalytic Conversion and Upgrading

The modeling effort for hydrolysate catalytic upgrading in the present analysis focused on modifying the 2015 design case by Davis et al. [4], which presented a similar APR concept. As noted previously, the pathway presented in the 2015 design report targeted diesel and high-octane gasoline products, based largely on patent literature data published by industry. For this technical assessment, the pathway models include modifications to the reaction steps to target maximizing a SAF bio-blendstock. The SAF-selective yield updates are explored across both DDA and DMR pretreatment strategies.

The process modeling basis for the catalytic conversion and upgrading steps were derived from the 2015 design case and updated for selectivity to SAF product using model compounds and future target fractionated yield estimates informed and reviewed by NREL and industry experts. The pathway updates for the catalytic upgrading of carbohydrates to SAF focus on three primary reactor stages:

- APR/hydrodeoxygenation (HDO) to deoxygenate highly oxygenated compounds (sugars and other organics in the clarified hydrolysate stream) with high reactivity to lower-oxygen-content molecules with moderate reactivity.
- Dehydration-oligomerization (DHOG) condensation to condense products from the APR step and increase carbon chain lengths of feed molecules (<C6) to those in the jet fuel range (approximately C9–C19).
- Hydrotreating to remove remaining oxygen and produce hydrocarbon fuels that meet respective gasoline, jet, and diesel specifications.

Figure 4 presents the block flow diagram for the overall modeled pathway from purified hydrolysate to finished hydrocarbon products. The goal of the successive catalytic steps is to remove oxygen or "de-functionalize" carbohydrates and oligomerize them to hydrocarbon molecules. The APR/HDO reactor system is modeled in Aspen Plus as a single reactor with an aqueous recycle loop for exotherm control. However, system designs would likely be constructed with two-stage reactor systems as shown in the figure. Capital and operating costs in this analysis are also based on the two-stage configuration.

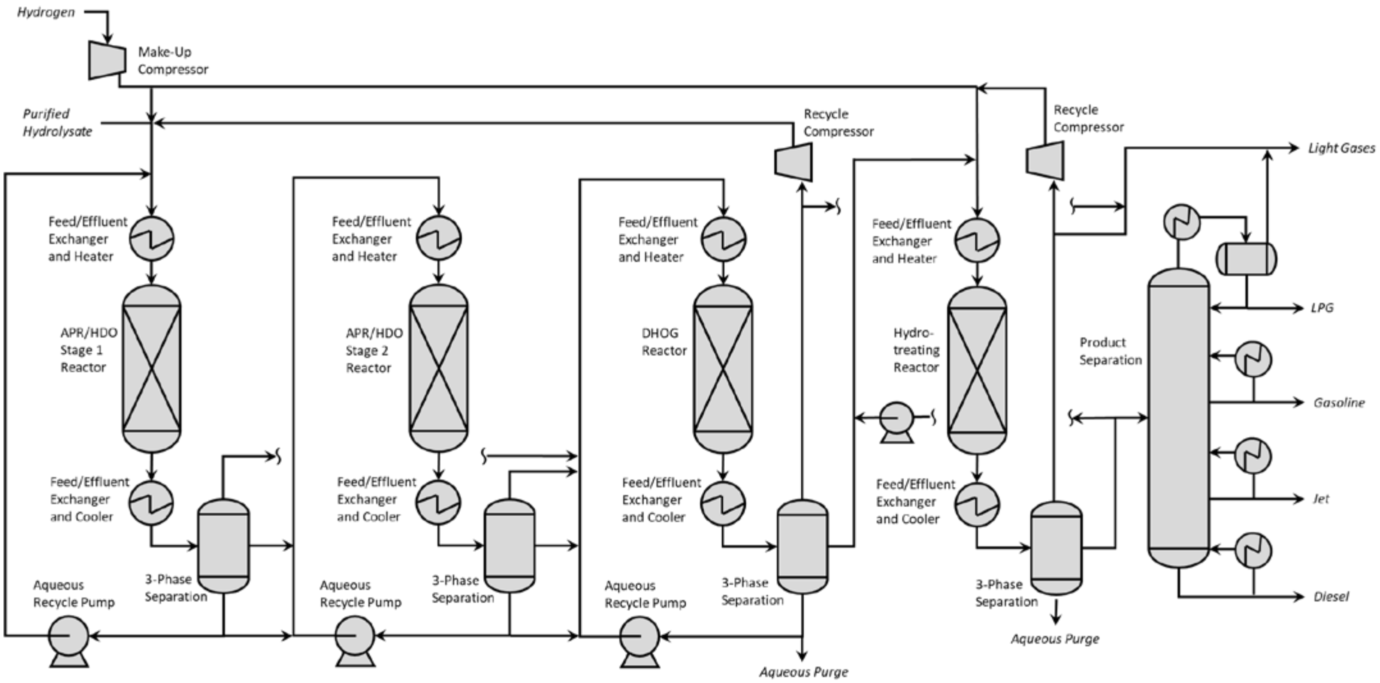


Figure 4. Block flow diagram for the APR-to-SAF pathway from hydrolysate to products

Steps for modeling SAF production yields involved matching component compositional targets by modifying the reactor stoichiometric basis for the APR outlet composition and SAF product composition. First, modified stoichiometry for the APR reactor was defined based on the target APR products profile reflected in the histogram presented in Figure 5. As shown in the figure, there are unknown compounds in the intermediate stream based on the data provided. Figure 6 shows the modeled APR outlet composition with the unknown compounds assumed to be represented by alcohols. Operating pressure for the APR reactor is 70 atm. Outlet temperature for the exothermic reactor is controlled at 240°C using aqueous-phase recycle to the reactor inlet.

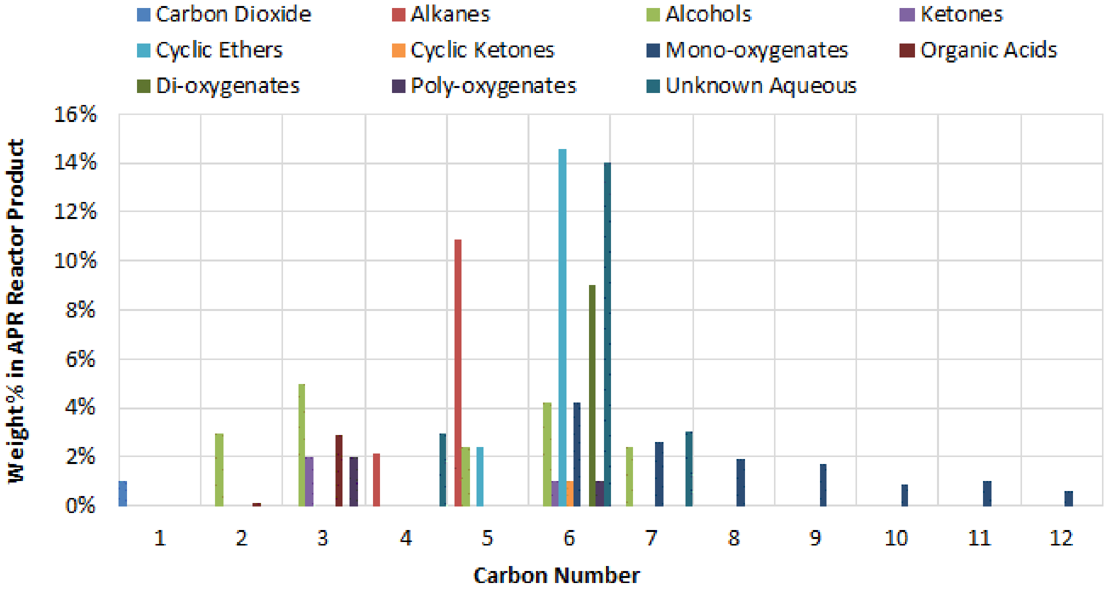


Figure 5. Target APR/HDO outlet composition

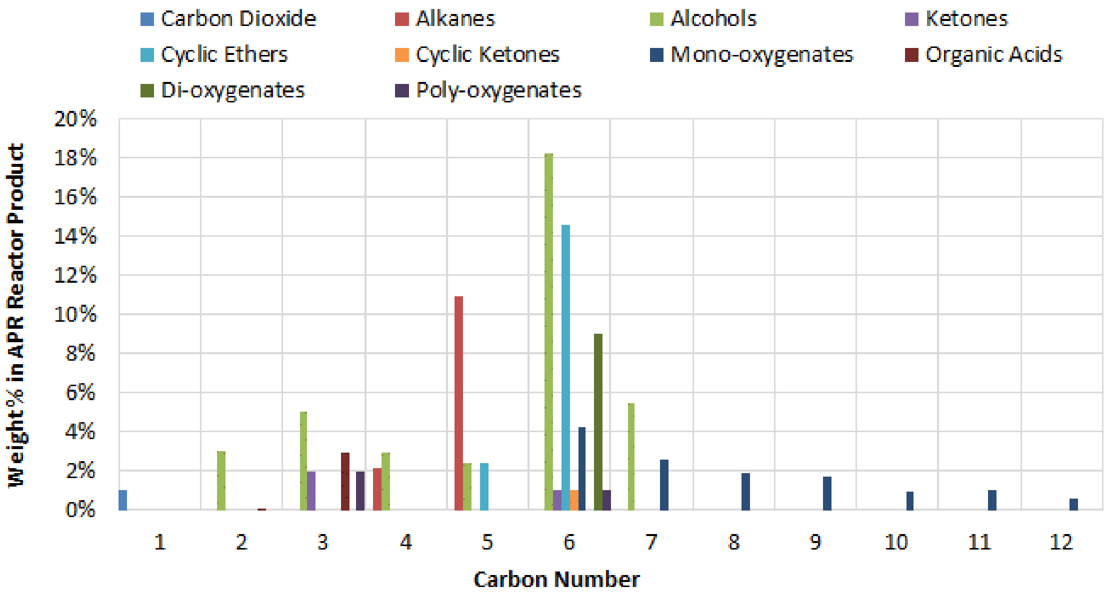


Figure 6. Modeled APR/HDO outlet composition

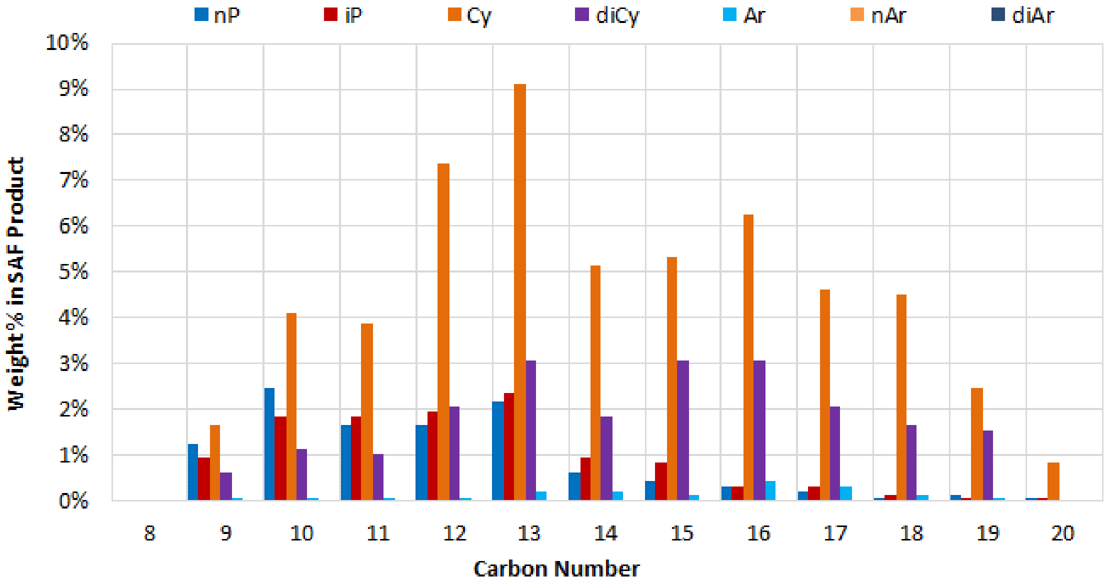


Figure 7. Target and modeled SAF product following DHOG and hydrotreating steps. Component families represented in the figure are *n*-paraffins (nP), iso-paraffins (iP), cyclo-paraffins (Cy), dicyclo-paraffins (diCy), aromatics (Ar), naphthenic aromatics (nAr), and di-aromatics (diAr).

Next, stoichiometry for the DHOG and hydrotreating reactors were updated based on the target SAF composition shown in Figure 7. The models were updated to exactly match the SAF compositional basis, as there were no unknowns in the dataset. DHOG compositional targets were not specifically defined. Therefore, the conversion extents in the DHOG and hydrotreating reactors were defined by splitting the hydrogen demands between the two reactor systems evenly. Operating pressure for the DHOG reactor is modeled at 65 atm, and outlet temperature is controlled at 260°C using aqueous-phase recycle to the reactor inlet. The hydrotreating reactor pressure is 58 atm, with reactor outlet temperature controlled at 370°C with cooled liquid effluent recycle. Heat generated from the hydroprocessing exothermic reactions is controlled through hydrogen addition and liquid product recycle to the reactor inlet. In practice, recycling light components may also help to optimize desired end-product selectivities, but this was not implemented in the models based on the degree of yield granularity specified in the models.

Overall yield targets for liquid products were aligned to a base reference DDA pretreatment scenario (previously established as the default case in the 2015 APR design report) of 70 gallons per ton (gasoline, jet, and diesel) using model compounds and yield structures informed and reviewed by NREL and industry experts. The overall target fuel yield was matched by changing aqueous and vapor carbon losses through the reaction stoichiometry while ensuring APR and SAF *compositional* targets were maintained. The initial yield basis of 70 gallons per ton corresponds with the hydrolysate substrate reflected in the 2015 NREL design case. In turn, yields for the updated DDA and DMR pretreatment scenarios utilize the same conversion losses from APR hydrolysate feed to hydrocarbon products, though now reflecting different upstream sugar yields from enzymatic hydrolysis and subsequent losses (e.g., through hydrolysate solids removal). The primary parameter influencing yields across these pathway scenarios is total carbon in the concentrated carbohydrates stream to catalytic upgrading, alongside catalyst performance to achieve the targeted conversions asserted here. Product selectivity targets for this modified catalytic upgrading pathway to gasoline, jet, and diesel are 20 wt %, 70 wt %, and 10 wt %, respectively. The selectivity targets are applied to all pretreatment approaches.

We note that these catalyst performance metrics have not been investigated or demonstrated internally at NREL, but rather are based on aspirational targets projected from state of technology data benchmarks furnished by industry. Catalyst performance on DDA or DMR-pretreated hydrolysate represents a key opportunity for experimental investigation in the future, to better understand challenges and bottlenecks that may be overcome through further research. For example, DDA pretreatment would impart elevated levels of sulfur in the hydrolysate feed (due to upstream use of sulfuric acid), which could pose challenges for catalyst activity or lifetimes. Likewise DMR could result in higher levels of sodium. Such considerations and potential impacts on catalyst yields, space velocities, and lifetimes are not well-understood based on public domain data, but are investigated through sensitivity analysis in Section 4.2.1.

Hydrogen consumption allocations to APR, DHOG, and hydrotreating were calculated by modeled yields and overall elemental balances. Within the process model, APR/HDO is represented as a single reactor with aqueous recycle loop for exotherm control. However, the capital and annual catalyst costs are based on the two-reactor system. The modeled DHOG reactor system did not indicate the need for liquid recycle in exotherm control, as the water present already serves as a sufficient heat sink as modeled. There are no liquid pumps other than liquid recycles in the catalytic conversion train and reboiler and reflux for distillation columns. The operating conditions and design bases for the reactors are summarized for the DDA and DMR scenarios in Table 6 and Table 7, respectively.

Table 6. Summary of Reactor Conditions and Design Parameters for DDA Scenario

Parameter	*Units*	*HDO/APR 1*	*HDO/APR 2*	*DHOG*	*Hydrotreater*
Reactor feed rate (total organics) [a]	kg/h	65,200	65,200	43,700	45,500
Reactor inlet temperature	°C	140		250	300
Reactor outlet temperature	°C		240	260	370
Operating pressure	atm	70	70	65	58
Weight hourly space velocity	h^{-1}	1.2	1.0	1.2	2.5
Hydrogen consumption	kg/h	2,400 (total HDO/APR + DHOG)			350
Catalyst loading weight	kg	54,400	65,200	36,400	8,700
Catalyst lifetime	years	1	2	2	2
Catalyst material [4] [b]		Precious metal + carbon support	Base metal + mixed oxide support	Precious metal + mixed oxide support	Precious metal + mixed oxide support
Catalyst cost (2020 $)	$/kg	139	18	49	49
Catalyst replacement cost (2020 $)	$MM	0.8	1.2	0.2	0.04

[a] Space velocities based on mass throughput rate of organics (dry wt excluding water).

[b] Catalyst materials and costs are maintained consistently with the details described in the 2015 APR report [4].

Table 7. Summary of Reactor Conditions and Design Parameters for DMR Scenario

Parameter	*Units*	*HDO/APR 1*	*HDO/APR 2*	*DHOG*	*Hydrotreater*
Reactor feed rate (total organics) [a]	kg/h	69,700	69,700	45,500	51,600
Reactor inlet temperature	°C	140		250	300
Reactor outlet temperature	°C		240	260	370
Operating pressure	atm	70	70	65	58
Weight hourly space velocity	h^{-1}	1.2	1.0	1.2	2.5
Hydrogen consumption	kg/h	2,500 (total HDO/APR + DHOG)			390
Catalyst loading weight	kg	58,100	69,700	37,900	9,200
Catalyst lifetime	years	1	2	2	2
Catalyst material [4] [b]		Precious metal + carbon support	Base metal + mixed oxide support	Precious metal + mixed oxide support	Precious metal + mixed oxide support
Catalyst cost (2020 $)	$/kg	139	18	49	49
Catalyst replacement cost (2020 $)	$MM	0.8	1.3	0.2	0.05

[a] Space velocities based on mass throughput rate of organics (dry wt excluding water).

[b] Catalyst materials and costs are maintained consistently with the details described in the 2015 APR report [4].

Similarly to the process model updates, the capital costs were informed by the 2015 design report and conservatively updated for SAF production with input from NREL and industry experts. Costs for the updated DDA and DMR scenarios were extrapolated from initial basis values using typical scaling calculations and the Aspen Plus heat and material balances, and updated to 2020 dollars. The resulting capital cost estimates are summarized in Table 8.

Table 8. Capital Cost Estimates for Updated APR-to-SAF Production Pathway (Total Purchased Equipment Cost | Total Installed Cost)

Reactor System	*DDA MM$ (2020)*	*DMR MM$ (2020)*
APR/HDO Stage 1 reactor section	14.8 | 29.7	15.5 | 31.1
APR/HDO Stage 2 reactor section	7.4 | 14.8	7.8 | 15.5
DHOG reactor section	26.8 | 53.5	27.5 | 55.1
Hydrotreating and distillation	3.2 | 6.5	3.4 | 6.8
Catalytic conversion and upgrading	**52.2 | 104.5**	**54.2 | 108.4**

Operating cost estimates were updated per the new Aspen Plus heat and material balances utilizing new raw material cost values as summarized in Section 3.1. Catalyst unit costs and general assumed material formulations were maintained consistently with the 2015 design case after updating to 2020 dollars, then applied to this study's catalyst targets for weight hourly space velocity listed above on a water-free basis. However, one update was made in the present study to incorporate a precious metal reclamation factor, set to recover 90% of the catalysts' precious metals during catalyst change-outs (not applicable to the APR-2 step) and thus reducing subsequent catalyst load costs by the same factor, consistent with more recent approaches as also assumed in NREL's 2018 design case [5]. Hydrotreating catalyst costs, which were not explicitly included in the 2015 report, were assumed at the same price basis as the DHOG step (in both cases reflecting a precious metal on mixed oxide support).

2.6 Area 600: Wastewater Treatment

Similar to other supporting units, all process/design assumptions for WWT remain consistent with prior design cases that include anaerobic digestion (AD) [4,7]. Wastewater streams generated from pretreatment flash condensation and deacetylation black liquor are combined with blowdown from the boiler and cooling tower, as well as hydrolysate processing steps (ion exchange backwash and APR/hydrotreating-produced water streams) and routed to WWT. WWT maintains the design and costing updates provided by Brown and Caldwell in 2013 [22], primarily consisting of anaerobic and aerobic digestion, membrane filtration, and salt removal/concentration via reverse osmosis, evaporation, and crystallization. Caustic species, primarily NaOH and Na_2CO_3 (in the case of two-stage DMR), are first neutralized with sulfuric acid before entering the WWT plant. The total chemical oxygen demand (COD) entering WWT is approximately 112 g/L (19,580 kg/h) and 52 g/L (17,590 kg/h) for the DMR and DDA cases, respectively, similar to prior NREL designs and higher than the latest 2018 design case (which included conversion of soluble organics [lignin and otherwise] to coproducts prior to reaching WWT), consequently requiring reintroduction of the AD step, which was assumed to be removed in the 2018 design. However, the model has been better automated to perform AD conversions based on overall yields to biogas and cell mass, replacing individual stoichiometric reactions that required manual specifications in prior design cases—maintaining overall COD destruction (75%–84% conversion to biogas—methane [CH_4] and CO_2—plus 5% to cell mass) and yielding a ratio of 300–292 gCH_4 and 883–906 gCO_2 per kilogram COD removed in the DMR and DDA cases, respectively.

The liquid from AD is pumped to the aerobic activated sludge basin, removing 96% of the remaining soluble organic matter (74% converted to water and CO_2 and 22% to cell mass). The output is then routed to a membrane bioreactor for clarification, followed by a gravity belt thickener and centrifuge to dewater the solids sludge to 20% total solids, split to internal recycle and disposal to the combustor/boiler. The clarified water is then routed to reverse osmosis to remove salts, with the salt stream further concentrated to 50% solids via a mechanical vapor recompression evaporator. The brine is further processed through a dryer and centrifuge to crystallize the salt for sale as a coproduct, in keeping with the rationale described in the 2018 design case [5]. Capital cost estimates were likewise maintained consistently with the 2013 updates, with the cost for AD and aerobic digestion scaled as a function of COD rate, while the costs for the membrane bioreactor, dewatering, and salt processing steps were scaled with hydraulic flow. The DMR case represents a deviation from prior designs in that the nitrogen content in the WWT feed is essentially zero, in contrast to the DDA case, which utilizes ammonium hydroxide for acid neutralization upstream (or prior design cases with DMR plus biological conversion of sugars/lignin, both of which utilize nitrogen nutrient sources and produce fermentation cell mass). Accordingly, a larger supplemental ammonia feed rate is required to supply the AD organisms, as shown in Table 11. Key parameters pertinent to the WWT design are summarized in Table 9.

Table 9. Key Parameters for WWT System

Parameter	Value (DMR/DDA)
Hydraulic load to WWT	1.1/2.1 MMgal/day
Total COD in WWT feed	112/52 g/L
Total Kjeldahl nitrogen	2.4×10^{-6}/302 mg/L
COD destruction through AD	80%/89% (75%/84% to biogas + 5% to cell mass)
AD methane yield	300/292 g/kg COD removed
Residual COD destruction through aerobic basin	96% (74% to water and CO_2 + 22% to cell mass)

2.7 Area 700: Product and Feed Chemical Storage

Area 700 continues to represent general costs for input and product storage tanks. This includes 7 days of storage for all fuel streams (jet, diesel, and naphtha cuts), holding tanks for feed chemicals ranging from 13,000 to 70,000 gallons, and a 600,000-gallon firewater tank.

2.8 Area 800: Combustor, Boiler, and Turbogenerator

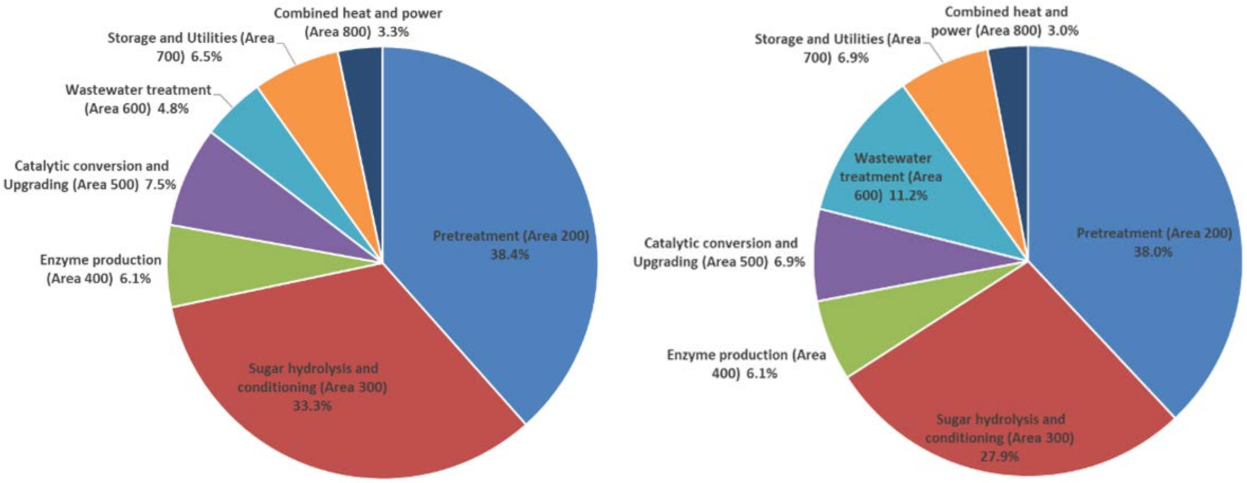
Given the focus in this study to prioritize process simplification and near-term deployment, the biorefinery design forgoes lignin valorization to high-value chemical coproducts (e.g., base-catalyzed deconstruction and bioconversion to adipic acid as outlined in the 2018 design case [5] or other lignin conversion pathways being more recently investigated). Instead, this design reverts back to the use of a combustor and boiler for combined heat and power integration, maintaining all cost and design details consistently with prior NREL reports [3,4] including explicit costing of ammonia for control of nitrogen oxides (NO_x) via selective non-catalytic reduction [5]. The combustor burns biorefinery waste byproducts including insoluble lignin and unconverted carbohydrates removed following enzymatic hydrolysis, biogas and sludge streams from WWT, and off-gases from the conversion operations. The combustor is designed to accommodate high-moisture solids with a boiler efficiency near 80%, with the flue gas used to raise superheated steam at 875 psig following flue gas desulfurization.

The steam is then routed through a multistage turbogenerator with offtake points to divert superheated steam away for process purposes (primarily pretreatment, requiring more steam in the DDA case than DMR), as well as medium- and high-pressure saturated steam for utility heating. The final turbine stage operates at vacuum pressure (0.1 atm) and is condensed with cooling water to maximize power generation before being recycled back to the boiler. The turbine generator efficiency is maintained at 85% as in prior designs, generating 46.0 and 51.0 MW of power in the DMR and DDA cases, respectively. After considering biorefinery power demands discussed in more detail below, the resulting net power balance translates to a small power import for DMR and export for DDA. In the DDA case, the exported power is treated as a coproduct sold to the grid. If energy self-sufficiency were a priority for the biorefinery, in the DMR case a small amount of additional corn stover feedstock could be brought in and bypassed directly to the boiler to fully meet biorefinery power demands, but this would incur additional cost and may not be warranted as the DMR net power import is already small and thus does not incur substantial GHG impacts. Ash and residual lime and calcium sulfate from flue gas desulfurization are collected in a baghouse (including periodic cost for bag replacement) and disposed to a landfill (costed as an operating expense).

2.9 Area 900: Utilities

Also consistent with prior design cases, Area 900 accounts for all plant utilities aside from steam (handled in Area 800), including electric power, cooling water, chilled water, plant and instrument air, process water, and the clean in place (CIP) system [5]. Freshwater makeup is combined with treated wastewater and condensate from sugar evaporation (again assumed to be suitable for all facility uses) and provided to the cellulase production unit, boiler and cooling tower makeup, the CIP system, and the wash for the lignin separation vacuum filter press. The cooling water system is maintained the same as prior design cases, designed for a 28°C supply and a 9°C temperature rise as an average across coolers throughout the facility. A full accounting of all cooling water demands across the facility is not repeated in this report, but consistent with prior studies, the large majority of cooling water duty requirements is associated with the steam turbine condenser (73% and 78% of total biorefinery cooling demands for the DMR and DDA cases, respectively).

For biorefinery power balances, the DMR case incurs a 1.9 MW net power import requirement given facility power demands (47.9 MW) outweighing the power generated from the steam turbine (46.0 MW), but this reverses to a 1.9 MW net export to the grid in the DDA case given the avoidance of power-intensive mechanical refining (49.1 MW biorefinery power demands vs. 51.0 MW power generation). A breakdown of facility power allocations is presented in Figure 8 for both the DMR and DDA base cases.



(a) Total power 47.9 MW

(b) Total power 49.1 MW

Figure 8. Distribution of plant electricity utilization by process area: (a) DMR case and (b) DDA case

3 Modeling Methods and Assumptions

3.1 Process Economics/Techno-Economic Analysis

All methodologies for TEA calculations remain consistent with prior design reports and will not be repeated in such detail here [3,5]; however, the present analysis also considers inclusion of policy incentives in addition to standard MFSP metrics as outlined below. Briefly, TCI is calculated from installed plus other direct and indirect equipment costs, and combined with variable and fixed operating costs. These values are then run through a discounted cash flow analysis to determine the required MFSP in order to obtain a zero net present value (NPV) at a fixed 10% internal rate of return (IRR); any of these three metrics can be solved as a variable upon fixing the other two (e.g., IRR could be solved for when selling fuels at a given price). The biorefinery plant lifetime is maintained at 30 years, along with the assumption of 40% equity/60% debt financing with the debt portion serviced by a 10-year, 8% interest loan. The federal income tax rate is maintained at 21% (prior to considering policy incentives), as is the n^{th}-plant basis for 3-year construction time and 6-month facility startup period (variances to these timeframes are considered in the sensitivity analysis in Section 4.2.1). Installed equipment capital costs were largely based on prior analyses [4,5,8] with the exception of Area 500 catalysis reactors, which were updated based on industry feedback as described in Section 2.5. Costs were scaled from their original basis values using the standard exponential scaling expression as a function of size (typically dictated by throughput) and scaling exponent n:

$$New\ Cost = (Base\ Cost) \times (New\ Size/Base\ Size)^n.$$

Costs are adjusted to 2020 dollars in the present analysis, using similar inflation indices as in past reports but now in more granularity for operating costs using indices more specific to key inputs. Namely, key cost inputs were updated by taking multiyear average costs (typically the most recent 5 years when available) for key inputs including glucose, hydrogen sourced via steam methane reforming (SMR), NaOH, ammonia, electricity, and natural gas (sourced externally in a sensitivity case for on-site NaOH production), and updating to 2020 dollars using commodity price indices specific to each input [23–30]. All other operating cost inputs were adjusted to 2020 dollars using the standard Producer Price Index (PPI) for chemical manufacturing published by the U.S. Bureau of Labor Statistics [31]. Capital costs were adjusted based on our standard approach using the Plant Cost Index from *Chemical Engineering Magazine* [32], while employee salaries were scaled to 2020 dollars using labor indices provided by the U.S. Bureau of Labor Statistics [33]. Additional direct and indirect costs were added to the calculated installed equipment capital costs, maintaining standard cost factors as defined and described in prior reports [5]. Details for base case capital costs, as well as variable and fixed operating costs, are summarized in Table 10, Table 11, and Table 12, respectively.

Table 10. Capital Cost Worksheet for Biorefinery Base Cases

Process Area		DMR Pretreatment		DDA Pretreatment	
		Purchased Cost	**Installed Cost**	**Purchased Cost**	**Installed Cost**
Area 100: Feedstock Storage and Handling [a]		Included in feedstock cost		Included in feedstock cost	
Area 200: Pretreatment		$40,100,000	$61,500,000	$37,800,000	$58,000,000
Area 300: Enzymatic Hydrolysis and Hydrolysate Conditioning		$58,000,000	$102,900,000	$49,100,000	$87,200,000
Area 400: Enzyme Production		$7,400,000	$12,700,000	$7,400,000	$12,700,000
Area 500: Catalytic Conversion and Upgrading		$54,200,000	$108,400,000	$52,200,000	$104,500,000
Area 600: Wastewater Treatment [b]		$40,100,000	$60,000,000	$45,600,000	$70,500,000
Area 700: Storage		$3,300,000	$5,700,000	$3,000,000	$5,300,000
Area 800: Combustor, Boiler, and Turbogenerator		$40,200,000	$72,700,000	$40,700,000	$73,700,000
Area 900: Utilities		$3,800,000	$6,700,000	$4,000,000	$7,100,000
Totals (Excl. Area 100)		**$247,200,000**	**$430,600,000**	**$240,000,000**	**$419,000,000**
Warehouse	4.0% of ISBL [c]		$11,400,000		$10,500,000
Site development	9.0% of ISBL		$25,700,000		$23,600,000
Additional piping	4.5% of ISBL		$12,800,000		$11,800,000
Total Direct Costs (TDC)			**$480,600,000**		**$464,900,000**
Proratable expenses	10.0% of TDC		$48,100,000		$46,500,000
Field expenses	10.0% of TDC		$48,100,000		$46,500,000
Home office and construction fee	20.0% of TDC		$96,100,000		$93,000,000
Project contingency	10.0% of TDC		$48,100,000		$46,500,000
Other costs (e.g., startup, permits)	10.0% of TDC		$48,100,000		$46,500,000
Total Indirect Costs			**$288,400,000**		**$279,000,000**
Fixed Capital Investment (FCI)			**$769,000,000**		**$743,900,000**
Land			$1,800,000		$1,800,000
Working capital	5.0% of FCI		$38,400,000		$37,200,000
Total Capital Investment (TCI)			**$809,300,000**		**$782,900,000**
Lang factor (TCI/purchased equipment cost)			3.8		4.0
TCI per annual GGE			$15.05/GGE		$15.35/GGE

[a] Feedstock handling not included in this calculation.

[b] Area 600 not included in Lang factor.

[c] ISBL = inside battery limits (Areas 200, 300, 400, and 500.

Table 11. Variable Operating Costs for Biorefinery Base Cases

Process Area	Stream Description	Cost (2020 $/lb) [a]	Basis	DMR MM$/yr (2020 $)	DDA MM$/yr (2020 $)
N/A	Feedstock	$74.26/ dry ton	$71.26/ton (2016 $) updated to 2020 $ via corn Producer Price Index [15]	53.79	53.79
A200	Sulfuric acid, 93%	$0.023	5-year avg. updated to 2020 $ [28]	0	0.33
	NaOH (as pure)	$0.120	5-year avg. updated to 2020 $ [28]	4.17	5.22
	Na_2CO_3 (as pure)	$0.089	5-year avg. updated to 2020 $ [28]	10.26	0
	Ammonia	$0.278	5-year avg. updated to 2020 $ [27]	0	1.26
A300	Flocculent	$0.600	NREL internal database	4.24	2.25
	HCl for IX regeneration, 35%	$0.059	Updated to 2020 $ from [28]	1.00	0.97
	Caustic for IX regeneration	$0.120	Updated to 2020 $ from [28]	1.16	1.12
A400	Glucose	$0.435	5-year avg. updated to 2020 $ [26]	10.01	10.22
	Corn steep liquor	$0.037	Updated to 2020 $ from [4]	0.06	0.06
	Corn oil	$0.700	Updated to 2020 $ from [4]	0.09	0.09
	Ammonia	$0.278	See above	0.30	0.31
	Host nutrients	$0.530	Updated to 2020 $ from [4]	0.34	0.35
	Sulfur dioxide	$0.200	Updated to 2020 $ from [4]	0.03	0.03
A500 [b]	Hydrogen	$0.769	5-year avg. updated to 2020 $ [25]	40.27	38.15
A600	Sulfuric acid	$0.023	See above	3.79	1.37
	Ammonia	$0.278	See above	1.76	0.29
	NaOH (as pure)	$0.120	See above	0	1.13
	Polymer	$2.84	Updated to 2020 $ from [4]	0.74	0.77
A800	Boiler chemicals	$3.22	Updated to 2020 $ from [4]	0.01	0.01
	Flue gas desulfurization lime (SO_x control)	$0.129	Updated to 2020 $ from [4]	0.33	1.04
	Ammonia (NO_x control)	$0.278	See above	3.53	3.58
A900	Cooling tower chemicals	$1.93	Updated to 2020 $ from [4]	0.08	0.08
	Makeup water	$0.0002	Updated to 2020 $ from [4]	1.03	1.01
Power	Electricity import	8.2 ¢/kWh	5-year avg. industrial retail power price, updated to 2020 $ [29]	1.23	0
	Subtotal			**138.22**	**123.42**
A800	Disposal of ash	$0.021	Updated to 2020 $ from [4]	1.55	1.74
	Subtotal			**1.55**	**1.74**
A600	*Sodium sulfate coproduct*	*$0.041*	*5-year avg. updated to 2020 $* [28]	*−8.60*	*−1.98*
Power	*Electricity coproduct*	*5.8 ¢/kWh*	*5-year avg. power export price to grid, updated to 2020 $* [30]	*0*	*−0.85*
	Subtotal			***−8.60***	***−2.83***
Total Variable Operating Costs				**131.17**	**122.33**
Net Contribution to MFSP ($/GGE)				**$2.44**	**$2.40**

[a] Costs in $/lb unless otherwise noted.

[b] Does not include catalyst costs (included in cash flow calculations separately, reflecting costs in Table 6 and Table 7).

Table 12. Fixed Operating Costs for Biorefinery Base Cases

Position	2020 Salary	# Required	2020 Cost	MM$/yr (2020 $)	DMR ¢/GGE (2020 $)	DDA ¢/GGE (2020 $)
Labor and Supervision						
Plant manager	$188,566	1	$188,566			
Plant engineer	$89,793	4	$359,172			
Maintenance supervisor	$73,117	1	$73,117			
Maintenance technician	$51,310	12	$615,724			
Lab manager	$71,834	1	$71,834			
Lab tech	$51,310	2	$102,621			
Lab tech – enzyme	$51,310	2	$102,621			
Shift supervisor	$61,572	4	$246,290			
Shift operators	$51,310	24	$1,231,448			
Shift operators – enzyme	$51,310	8	$410,483			
Yard employees	$35,917	4	$143,669			
Clerks and secretaries	$46,179	3	$138,538			
Total salaries			$3,684,083	3.68	6.85	7.22
Labor burden (90%)			$3,315,674	3.32	6.16	6.50
Other Overhead			**DMR MM$/yr**	**DDA MM$/yr**	**DMR ¢/GGE**	**DDA ¢/GGE**
Maintenance		3.0% of ISBL	8.56	7.87	15.92	15.44
Property insurance		0.7% of FCI	5.38	5.21	10.01	10.21
Total Fixed Operating Costs			**20.95**	**20.08**	**38.94**	**39.37**

3.2 Life Cycle Analysis

ANL analyzed the GHG emissions associated with fuel production, measured in grams of carbon dioxide equivalent (CO_2e) per megajoule, and biorefinery-level GHG emissions, measured in kilograms of CO_2e per tonne of dry biomass feedstock processed through the conversion biorefinery. ANL's R&D GREET model was used to conduct the LCA and estimate the life cycle GHG emissions including the feedstock-to-fuel supply chain CO_2, CH_4, and nitrous oxide (N_2O) emissions based on their respective 100-year global warming potentials (1, 29.8, and 273 [34]). Biogenic CO_2 uptake during biomass growth and emissions from downstream conversion/combustion steps are counted as zero and not explicitly tracked in the calculations.

While coproduct outputs were small in this study (sodium sulfate salt recovered from WWT and exported power in the DDA case), the coproduct effects were addressed using two different methods: the displacement method and the biorefinery-level method [35]. The displacement method attributes all the supply chain emissions to the fuel product while crediting the biofuel for the emissions avoided by producing the coproduct using conventional technologies. This credit is referred to as the "*coproduct displacement credit.*" Alternatively, a biorefinery-level approach estimates the total emissions from producing both the biofuel and coproducts, and the corresponding overall emissions reductions from displacing the same quantity of conventional fuels and products with these bio-based alternatives. As the primary coproduct, sodium sulfate, is more appropriately viewed as an unintended byproduct than an intentionally co-produced output, and does not otherwise exert substantial influence on overall GHGs when being handled via the displacement method, the application of other alternative coproduct methods such as economic or energy allocation were not deemed necessary to consider in this context.

For LCA modeling purposes, the feedstock logistics supply assumes a corn stover feedstock blend that comprises 70% two-pass corn stover and 30% three-pass corn stover [36]. As corn stover is considered a waste material, it does not share any LCA burdens with corn cultivation, reflecting a supply chain boundary beginning at corn stover collection from the field. The feedstock production and logistics involve the harvesting and collecting of corn stover, followed by storage, preprocessing, handling, and blending before it is transported to the biochemical conversion plant. Figure 9 illustrates the LCA system boundary of the corn stover-to-hydrocarbon fuels supply chain. There are minimal direct fossil CO_2 emissions from the biorefinery, thus the majority of non-biogenic CO_2 emissions are attributed to indirect emissions from the sourcing and disposal of inputs and outputs.

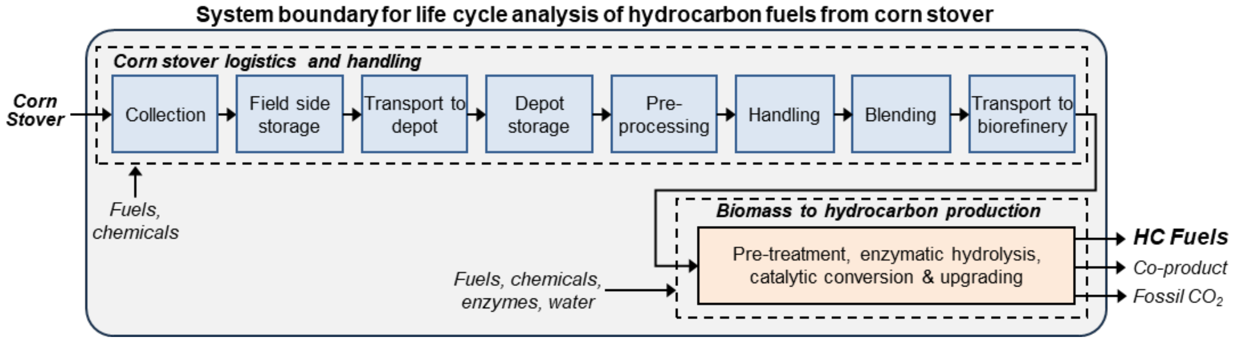


Figure 9: LCA system boundary of the supply chain for hydrocarbon fuels and co-products from corn stover. There are minimal direct fossil CO_2 emissions from the biorefinery, thus the majority of non-biogenic CO_2 emissions reflect indirect emissions associated with inputs and outputs.

3.3 Marginal Cost of Greenhouse Gas Abatement

The marginal GHG abatement cost is defined previously in Eq. 1. As described in Section 1.3, MAC provides a means of quantitatively combining the MFSP with corresponding GHG emissions reductions when comparing both metrics against a reference cost and GHG baseline, in this case for petroleum fuels. The incremental *increase* in the cost of biofuel (translated from \$/GGE to \$/MJ based on a standard lower heating value of 122.5 MJ/GGE) is divided by the incremental *reduction* in the GHG emissions enabled by the biofuel (as reported in gCO_2e/MJ), to quantify the incremental cost of reducing GHG emissions as supported by adopting the modeled technology over the incumbent baseline (\$/tonne CO_2e). Particularly when prioritizing maximum decarbonization potential as is considered over a range of scenarios here, MAC provides useful information to understand the cost of achieving such decarbonization as may be compared to other technologies. Direct air capture (DAC) of atmospheric CO_2 is one basis often used as a comparator in this context, with a benchmark MAC reported in literature of roughly \$600/tonne CO_2e as a reference case focused strictly on ambient CO_2 drawdown and sequestration (with one study noting current reported costs between \$500 and \$600/tonne and another between \$600 and \$1,000/tonne), plausibly to be reduced to a range of \$100–\$600/tonne in the future [37–39]. A SAF/biofuel conversion pathway such as this that could meet or improve on the MAC relative to DAC would reflect a preferential strategy for GHG abatement, as it would simultaneously provide a useful service to support aviation and other transportation fuel needs.

3.4 Policy Incentives

Over recent years, the United States and European Union have proposed significant policy legislation supporting the SAF industry. While the aim of the policies developed in these two parts of the world are the same, their approach and configuration are different. In the United States, the federal government is implementing incentives to drive down the SAF cost for airlines and boost supply, whereas Europe is setting industrial targets and blending mandates. Currently available tax credits in the United States are briefly summarized below, and considered as an alternative case for inclusion in the TEA models to evaluate their implications on MFSPs.

The United States has the potential to sustainably produce a significant amount of biomass from various sources, enabling the production of around 55–80 billion gallons of renewable liquid biofuels annually [40]. Major research initiatives and consortia are working together to address challenges in biofuel feedstocks, conversion processes, and engine optimization. These efforts connect to the SAF Grand Challenge in aiming to accelerate SAF production and deployment.

The U.S. federal government has implemented policies like the Renewable Fuel Standard (RFS) and Inflation Reduction Act (IRA) to support biofuel production at the federal level. These policies offer incentives such as tax credits, grants, and loans. Additionally, states like California have their own policies, such as the Low Carbon Fuel Standards (LCFS), that complement federal efforts. Some states are proposing tax credits specifically for SAF. To promote innovation and overcome barriers in the biofuels sector, collaboration between government, academia, and industry is crucial. The R&D GREET model is being updated by Argonne National Laboratory to provide a consistent methodology for calculating the life cycle GHG emissions of biofuels like SAF across different policies and incentive programs.

In the United States, the IRA, passed in August 2022, defines a SAF fuel credit for the sale or use of a qualified blend instead of a SAF mandate, and provides grants to support the production, storage, and distribution of SAF. Two relevant tax credit schemes are introduced with the IRA (40B and 45Z) [41].

IRA Section 40B (2023–2024) provides a tax credit for the sale or use of SAF from 2023 through 2024 of $1.25/gal, based on achieving a life cycle GHG emissions reduction of at least 50% compared with petroleum-based jet fuel [41]. SAF that achieves a greater reduction in lifecycle GHG emissions is eligible for an additional $0.01/gal credit for each percentage point of emissions reductions, up to $1.75/gal of SAF at 100% GHG reduction.

IRA Section 45Z (2025–2027) provides a tax credit for fuel (both non-SAF and SAF) produced after 2024 and used or sold before 2027. The value of an incentive for transportation fuels is determined by multiplying the base or alternative credit by an emissions factor. The emissions factor is calculated by subtracting the emissions rate of the fuel from 50 $kgCO_2e/MMBtu$ and dividing by 50 $kgCO_2e/MMBtu$ [42]. The emissions factor is used to determine the lifecycle GHG emissions reduction of the fuel, and the policy gives an option to carry out numerical roundings, which is not done in this work. The greater the reduction in emissions, the higher the value of the incentive. Fuels with zero emissions (100% reduction) can receive a credit of \$1.00/gal for non-aviation fuel or \$1.75/gal for SAF, reducing proportionately until a minimum emissions threshold of 50 kgCO$_2$e/MMBtu. This credit can be further extended beyond \$1.75/gal for SAF by the same emissions factor calculation in the case of net-negative GHG emissions rates. To be eligible for credits, a transportation fuel must have a life cycle GHG emissions of at most 50 kgCO$_2$e/MMBtu [42,43]. Additionally, both 40B and 45Z necessitate SAF to meet the requirements under ASTM D7566 1 (which is possible through a time-intensive qualification process, ASTM D4054). Recognizing this technology pathway is not currently ASTM-certified, these policies are not included under base-case calculations for SAF policy credit considerations; however, for purposes to understand maximum *potential* policy implications, these policies are alternatively considered later in a sensitivity case assuming that the fuel produced here were to meet ASTM certification and thus qualifies for all currently available credits.

The **RFS** was the first federal policy action to set volumetric requirements for biofuels in efforts to increase biofuel usage nationwide, as first established by the U.S. Environmental Protection Agency in 2005 and later amended in 2007 [44]. SAF is eligible for RFS credits depending on the feedstocks and pathways used to generate the fuel, as well as the ability to meet the required GHG emissions reduction, for the D3/D7, D4, D5, and D6 categories of renewable identification numbers (RINs) [45]. The Environmental Protection Agency sets target volumes for different categories of renewable fuels, including cellulosic biofuel, biomass-based diesel, advanced biofuel, and renewable fuel. The variability of these target volumes results in variability in the credit prices from year to year. The SAF produced here is eligible to generate D3 RIN credits as a cellulosic fuel and must meet a minimum 60% reduction in life cycle GHG emissions compared to petroleum benchmarks [44]. However, a specific volumetric target is not presently set for SAF [44]. Credits through the RFS that are earned by producing SAF can be stacked with other federal tax credits for SAF, such as those provided by the 2022 IRA, as well as state credits relevant to SAF [46].

State and local policies play a crucial role in supporting the production and consumption of SAF by offering additional credits and incentives on top of federal initiatives. These policies can benefit state economies while contributing to the overall sustainability of the aviation industry. Moreover, the variations in state policies can create competitive advantages for states in terms of SAF production capacity and use by airlines. States with more favorable policies may attract investment in SAF production facilities, leading to job creation and economic growth. Additionally, airlines operating in states with supportive policies are more likely to incorporate SAF into their fuel mix, enhancing their environmental credentials and meeting sustainability targets.

The **California LCFS** was established in 2009 to reduce GHG emissions in the transportation sector and develop a range of low-carbon and renewable alternatives to reduce petroleum dependency [47]. SAF is eligible under the LCFS program as an “opt-in” fuel. SAF producers can create and sell LCFS credits to “obligated parties,” or fossil jet fuel producers, for revenue [47].

Oregon adopted a Clean Fuels Program in 2016 with a target for a 20% reduction in CI of transportation fuels from 2015 levels by 2030 and 37% by 2035. The program includes SAF as eligible for credits for production and importation [48].

The **Washington** state Clean Fuel Standard is designed to incentivize the reduction in the CI of transportation fuels to 20% below 2017 levels by 2038. The tax credit for SAF is \$1.00/gal with a minimum of 50% GHG reduction relative to fossil Jet A, and an additional \$0.02/gal for each additional percentage point reduction past 50%, capped at \$2.00/gal [49]. However, these tax credits can only be applied when one or more facilities in the state achieve a production capacity of 20 million gallons per year of alternative jet fuel. Total current SAF production in Washington is less than 20 million gallons per year.

Illinois adopted a SAF purchase credit in 2023, which is valid between June 2023 and June 2033. The credit is \$1.50/gal for SAF sold to or used by an air carrier in Illinois [50].

Minnesota introduced a SAF tax credit limited to \$7.4 million in 2025 and \$2.1 million in 2026 and 2027. The credit is \$1.50/gal for SAF produced or blended in Minnesota and used in an aircraft in Minnesota, with a minimum 50% life cycle GHG emissions versus petroleum jet [51].

New Mexico adopted a Clean Fuel Standard in March 2024, but details of this state rulemaking are still being finalized [52].

An evaluation of the combined value of federal and state tax credits demonstrates the contributions from those different tiers of policies. Using California as a representative example, the combination of federal with state credits bolsters the total potential credit for biofuels, as illustrated in Figure 10. The RFS and LCFS policies are seen to be “mutually reinforcing” as they may be stacked to provide greater benefit for the same effort and cost to comply [53]. The combined federal and state credit values have more moderate overall fluctuations over recent periods, mainly due to the fluctuations from RFS and California LCFS policies. As one possible factor for this relatively decreased value fluctuation, the LCFS setting of compliance thresholds for a longer term may reduce market uncertainty and improve biofuel producer confidence, compared to the RFS approach of setting obligated volumes every 2–3 years. When considering both federal and federal-plus-state combined credit scenarios, SAF currently has an overall advantage over renewable diesel at more significant GHG reduction values. However, the RFS and state LCFS credit policies tend to favor renewable diesel over SAF. Despite this, Figure 10 highlights that the stacking of available federal and state credits over the past 5 years, including IRA 40B and 45Z, plays a significant role in providing substantial incentive credits to improve biofuel production economics while shifting the advantage toward SAF in these scenarios. It should also be noted that this figure strictly reflects the *values* of policy incentives as may be combined to garner overall biorefinery credits, while emphasizing there would be additional *costs* for implementing decarbonization strategies in order to access larger CI reduction credits (i.e. moving toward the “zero CI” curves).

These policy incentives are not only available to SAF but also to other fuels. Table 13 provides a list of policy credits applicable to all the fuels produced here, including SAF, diesel and naphtha, based on their estimated carbon footprints for four scenarios defined around the source of electricity and hydrogen, and whether biorefinery carbon capture and sequestration (CCS) is applied and onsite caustic production is carried out (these scenarios are further discussed in Section 4.2.2). The base-case analysis assumes that 2023 is the first year of operations for the 30-year facility lifetime. Naphtha is regarded as a gasoline blendstock that is eligible to receive credit under the policies considered in Table 13, except for 40B, which is specifically for SAF. It is further supposed that this production plant is part of a larger entity that has a tax liability greater than the revenue that 40B and 45Z can generate, as these policy options essentially provide *non-refundable* tax credits. Otherwise, it would not be possible for such new facilities to have tax liabilities before they could start making profit and be eligible for these two policy incentives. Additional considerations for the application of these policy incentives are provided in the footnotes of Table 13.

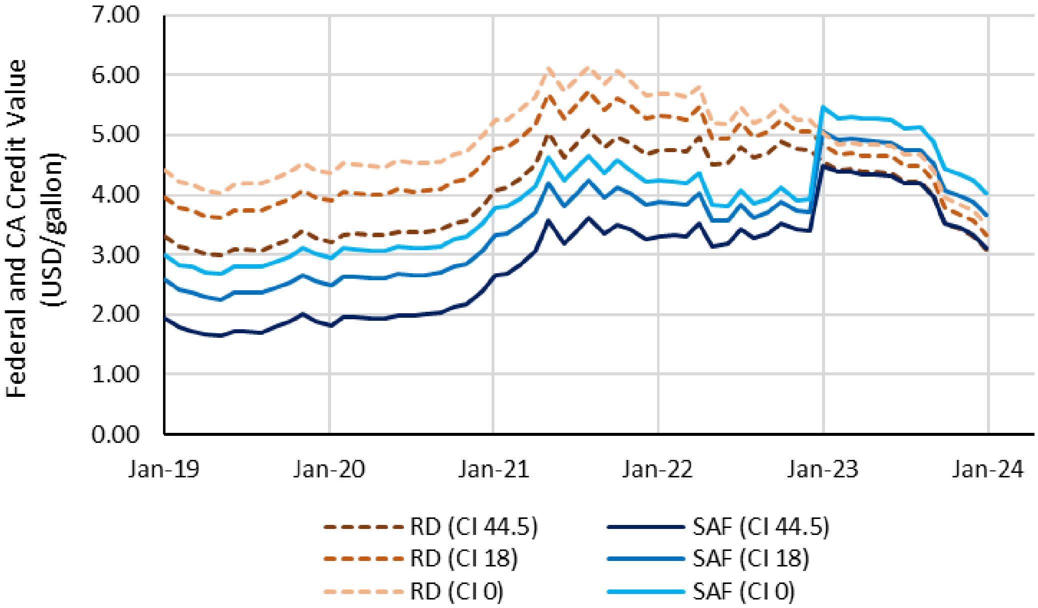


Figure 10. Example case demonstrating combined federal and California state biofuel policy incentive values ($/gal) for the time period from January 2019 to January 2024.

Incentives include: RFS RIN credits (applicable to renewable diesel [RD] and SAF), biodiesel mixture credit (RD), SAF credit (SAF, 2023–2024), California LCFS credit (RD, SAF), and LCFS avoided deficits for petroleum diesel obligated through LCFS (RD). Not included are avoided deficits for petroleum diesel through the California cap-and-trade program, which could be applied to the combined value for RD.

Table 13. Scenarios and Basis of Policy Credits as Applied to These Scenarios. Applicable credits are stacked in combination to calculate total biorefinery incentives. (Note that policy options that are in italics are included only in a subsequent sensitivity case. See footnote [b].)

		Scenario I	**Scenario II**	**Scenario III**	**Scenario IV**
Electricity		Renewable	Renewable	Renewable	Renewable
Hydrogen		Green	Blue	Green	Blue
CCS		Yes	Yes	No	No
Caustic		On-site	On-site	On-site	On-site
Policy Type		**Credits ($/gallon of fuel)**			
40B credit [a]		Years 1 and 2 (2023 and 2024)			
- SAF [b]	*DMR*	*2.42*	*2.29*	*1.42*	*1.29*
	DDA	*2.50*	*2.37*	*1.49*	*1.36*
45Z credit [c]		Years 3-5 (2025-2027)			
- Naphtha/Diesel	DMR	2.27	2.02	0.39	0.14
	DDA	2.40	2.16	0.51	0.26
- SAF [b]	*DMR*	*3.97*	*3.54*	*0.68*	*0.25*
	DDA	*4.20*	*3.77*	*0.89*	*0.46*
California LCFS credit [d]		Years 1-30 (2023-2052)			
- Naphtha	DMR	1.06-2.28	0.97-2.10	0.36-0.89	0.18-0.70
	DDA	1.10-2.37	1.02-2.19	0.45-0.97	0.27-0.79
- SAF	DMR	1.21-2.62	1.12-2.41	0.42-1.03	0.21-0.82
	DDA	1.26-2.73	1.17-2.52	0.52-1.13	0.32-0.92
- Diesel	DMR	1.28-2.75	1.17-2.53	0.44-1.08	0.22-0.86
	DDA	1.33-2.87	1.23-2.65	0.55-1.19	0.33-0.97
D3 RIN credit [e,f]		Year 1 (2023)	Year 2 (2024)	Years 3+ (2025-2052)	
- Naphtha		2.58	3.11	2.46	
- SAF		4.13	4.98	3.94	
- Diesel		4.39	5.29	4.19	

[a] 40B SAF credit is currently set to expire in 2024. Value of credit is dependent on the reduction in life cycle GHG emissions with respect to a conventional fuel benchmark: an initial credit of $1.25/gal of SAF with a minimum 50% reduction in life cycle GHG emissions plus $0.01/gal for each additional percentage point reduction. The Carbon Offsetting and Reduction Scheme for International Aviation (CORSIA) baseline of 89 gCO$_2$e/MJ [54] is used in these calculations.

[b] It is required that the fuel be included in the ASTM D7566 Standard to be eligible for these SAF credits, which is not the case as of this writing. However, a similar fuel (HDO-SAK, Hydrodeoxygenation Synthetic Aromatic Kerosene) is currently undergoing the ASTM D4054 qualification process before it can be included in D7566, providing a justifiable basis to assume these credits could be extended in the near-term to a pathway such as this. For this reason, these policy options are only considered in a sensitivity case.

[c] 45Z credit will be available from 2025 and is currently set to expire in 2027. A credit of $1.75/gal for SAF and $1.00/gal for other fuels is applied to relative life cycle GHG savings from 47.4 gCO$_2$e/MJ (50 kgCO$_2$e/MMBtu), *e.g. SAF credit = $1.75/gal × (47.4 – CI)/47.4*.

[d] Based on average weekly carbon prices (in $/tonne) published by the California Air Resources Board (CARB) [55]. Price averages from 2023 and 2024 (until November 11, 2024) are used to determine the 2023 and 2024 credits, respectively. A single five-year price average (from November 18, 2019 to November 11, 2024) is used to estimate all the future credits for 2025 and onwards. It is assumed that California LCFS will be continued until 2052 and that this carbon price will be applicable during that period, even though it will change over time. The credit itself is calculated by multiplying the carbon price by the absolute life cycle GHG savings from a conventional fuel footprint, which is determined by CARB for all three fuel cuts and currently set to decrease linearly every year until 2030. It is further assumed that this linear trend will continue until 2052. Hence, a different credit applies each year. This table lists the calculated minimum and maximum credits until 2052.

[e] Based on individual D3 RIN prices (in $/gal) from all transactions (verified and unverified) reported by the U.S. Environmental Protection Agency (EPA) [56]. Averages of weekly averaged prices from 2023 and 2024 (until November 21, 2024) are used to determine the 2023 and 2024 RIN credits, respectively. A single five-year price average (from the week of November 18, 2019 to November 11, 2024) is used as the RIN credit for 2025 and onwards. It is assumed that the EPA's Renewable Fuels Standard (RFS2) will be continued until 2052 and that this

RIN price will be applicable during that period, even though it will change over time. An equivalence factor of 1, 1.6 and 1.7 is applied to these RIN prices to determine the credit for naphtha (as gasoline blendstock), SAF and diesel, respectively.

[f] These RIN credits are tied to the year and not to the scenario like the other policy options in this table. However, in order for these fuels to be eligible for these D3 RIN credits, their relative life cycle GHG savings from EPA's conventional fuel baselines must be at least 60%, which happens to be the case for all the scenarios except for the DMR case in Scenario IV (which is not eligible for these credits). EPA's conventional fuel GHG baselines are 93.1 gCO2e/MJ (98.2 kg CO2e/MMBtu) for gasoline/naphtha and 91.9 gCO2e/MJ (97 kg CO2e/MMBtu) for diesel and jet fuel [57].

Finally, while not directly related to the applicability of decarbonization policy incentives, a note is warranted regarding broader regulatory policies and their impacts on new biorefinery projects. A critical element for new biorefinery project siting and construction depends heavily on the regulatory permitting process, particularly air permitting, the complexity of which can vary widely depending on specific location. This point is reiterated in NREL's recently published "SAF State of Industry" report, which highlights the air permitting and approval process as a major constraint challenging the deployment of SAF biorefineries based on comprehensive industry feedback [46]. Accordingly, such considerations could play a prominent role in the timeline to construct new biorefinery facilities and in turn the applicable decarbonization policy incentives which may be available at that time, given several shorter-term policies set to expire in coming years as enumerated above.

4 Results and Discussion

4.1 Base Case Results

The above-described base cases translate to a net MFSP of **$4.79/GGE across either pretreatment scenario**. The resulting MFSP parity between the two pretreatment cases is a coincidental outcome of yield versus cost tradeoffs between the two approaches, with a slightly lower (5%) fuel yield in the DDA case (reflecting lower projected sugar yields vs. DMR) offset by 4% lower capital and fixed operating costs and 10% lower variable operating costs (a function of lower pretreatment chemical costs and subsequent WWT neutralization costs, as well as lower flocculant loadings for hydrolysate solids removal). Figure 11 provides a breakdown of contributions to MFSP by both process area and cost category, with further TEA details presented in Table 14. For both pretreatment scenarios, the catalytic upgrading step incurs the greatest process area costs, responsible for roughly 30% of net MFSPs, followed by feedstock costs (roughly 20%) and then pretreatment and enzymatic hydrolysis/hydrolysate processing (both similarly contributing roughly 16%–17% in the DMR case, reducing to 13%–15% in the DDA case due to slightly lower capital costs for DDA pretreatment and hydrolysate clarification). These results compare favorably to other leading biochemical conversion pathways such as biological and catalytic upgrading of corn stover sugars via ethanol-to-jet at a published MFSP of $5.37/GGE (though based on differing process details and cost-year basis published in 2017) [58].

When viewed by cost category, variable operating costs represent the strongest driver, contributing more than 50% to MFSP in both pretreatment cases, in turn driven primarily by feedstock (21–22%) and hydrogen costs (16%). The strong feedstock contribution is typical for most biofuel conversion processes [3,5], but the outsized influence from hydrogen costs is unique for biochemical pathways (at the $1.69/kg hydrogen cost basis), as the APR pathway consumes substantially more hydrogen than other historical configurations. However, this is a trade-off in order to achieve substantially higher fuel yields owing to much higher carbon retention efficiencies than other pathways, particularly involving biological conversion of sugars. Base case fuel yields in this study reach **70 and 74 GGE/dry ton** biomass corresponding to **83%–84% carbon efficiency** from sugars to fuels (DDA vs. DMR cases, respectively), compared to 43–45 GGE/ton in the 2018 biological pathway design cases at 50%–51% sugar-to-fuel carbon efficiency [5], or 50 GGE/ton for the above-referenced ethanol-to-jet pathway [58]. This also represents an opportunity for substantial decarbonization improvements depending on choices for hydrogen sourcing, as discussed further below. Following variable operating costs, capital-related expenses add roughly 40% to net MFSPs for either pretreatment case (driven most strongly by catalysis reactor costs), while fixed operating costs (labor/overhead, maintenance, and property insurance/taxes) contribute 8%. Policy incentives are not included in these MFSP considerations, given that neither base case scenario yet reaches a 50% GHG reduction threshold required to trigger inclusion of policy credits. However, given potential to substantially further reduce GHG emissions based on alternative scenarios presented below, policy credits can then be introduced as highlighted in Section 4.2.2.

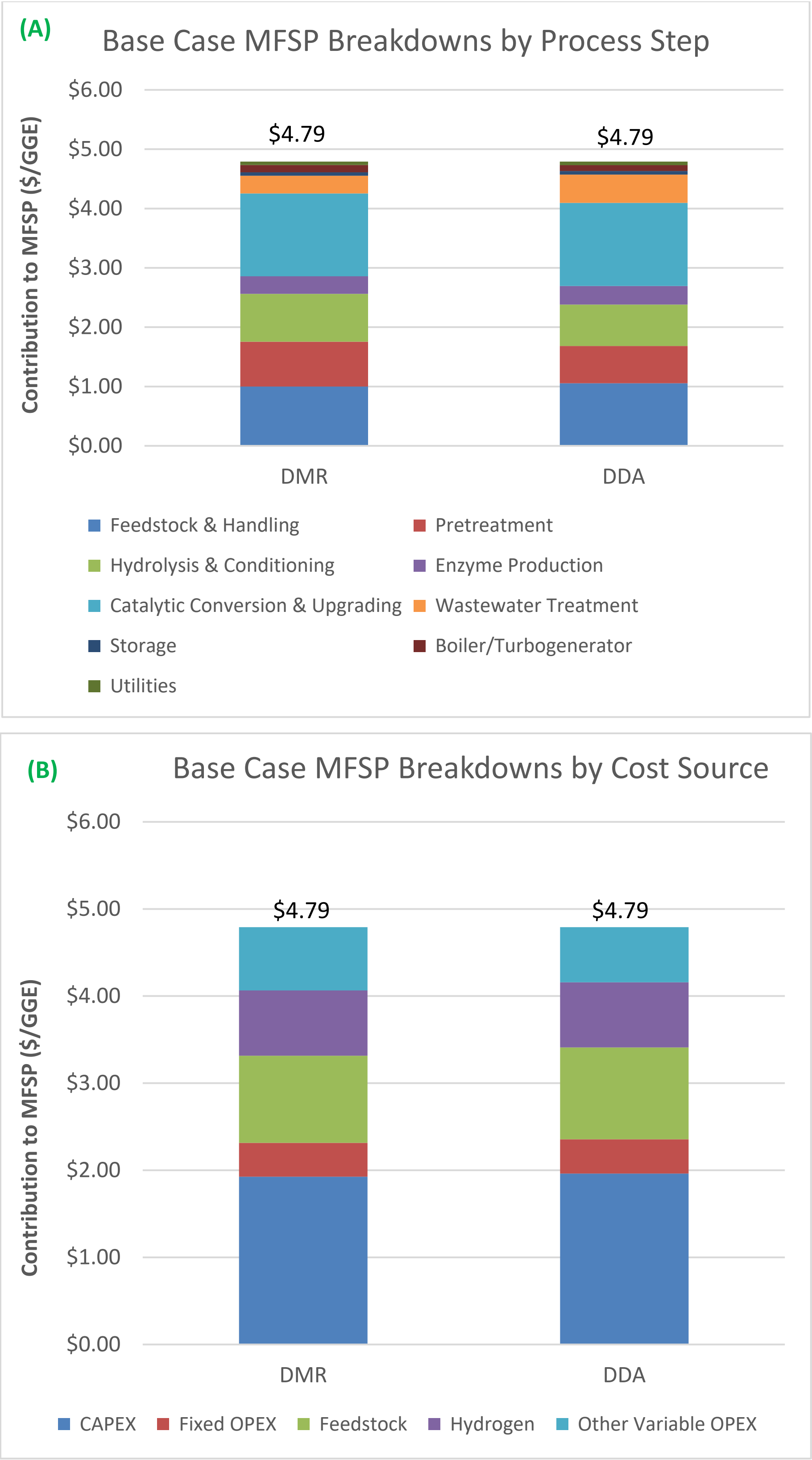


Figure 11. Contributions to base case MFSPs by (A) process step and (B) cost category.

CAPEX: capital expenses; OPEX: operating expenses

Table 14. Key TEA Results for Base Case Scenarios

Parameter	DMR	DDA
Feedstock rate (dry U.S. tons/day)	2,205	2,205
Online time (hours/yr)	7,884 (90%)	7,884 (90%)
Total fuel yield (GGE/dry U.S. ton feedstock)	**74.3**	**70.4**
Total fuel production rate (MM GGE/yr)	53.8	51.0
SAF (MM GGE/yr)	*37.8*	*35.8*
Diesel (MM GGE/yr)	*5.4*	*5.1*
Naphtha (MM GGE/yr)	*10.7*	*10.1*
Overall carbon efficiency, biomass to fuels	44.2%	41.9%
APR carbon efficiency, sugar to fuels	84.0%	82.9%
Minimum fuel selling price ($/GGE)	**$4.79**	**$4.79**
Feedstock contribution ($/GGE)	*$1.00*	*$1.05*
Conversion contribution ($/GGE)	*$3.79*	*$3.74*
Corresponding MFSP for SAF ($/gal)	*$5.28*	*$5.28*
Corresponding MFSP for diesel ($/gal)	*$5.55*	*$5.55*
Corresponding MFSP for naphtha ($/gal)	*$4.63*	*$4.62*
Total capital investment ($MM)	$809.3	$782.9
Variable operating costs (excluding coproducts) ($MM/yr) [a]	$141.8	$127.0
Coproduct revenues ($MM/yr) [b]	$8.6	$2.8
Fixed operating costs ($MM/yr)	$21.0	$20.1

[a] Includes amortized catalyst and baghouse replacement costs

[b] Includes salt recovery from WWT in both cases plus net power export to grid in DDA case.

Figure 12 shows the supply chain GHG emissions of the base cases using the coproduct displacement method, along with their key contributing processes, compared to a life cycle carbon intensity benchmark of 89 gCO_2e/MJ for petroleum jet fuels as a commonly-accepted industry baseline [54]. The GHG emissions range from 63.0 gCO_2e/MJ for the DMR pretreatment scenario down to 57.7 gCO_2e/MJ for the DDA pretreatment scenario, representing a 29%–35% reduction in supply chain GHG emissions relative to those of petroleum jet fuels. The DDA pretreatment scenario achieves 8% lower GHG emissions than the DMR scenario, primarily owing to its lower caustic requirements (including the avoidance of fossil CO_2 emissions incurred with the use of sodium carbonate upon neutralization) and lower biorefinery power demands translating to a net power export to the grid. Under the displacement method, all chemical usage and associated emissions are attributed to fuel, while receiving a small displacement credit of 2.7 gCO_2e/MJ from the sodium sulfate salt coproduct for DMR and 1.7 gCO_2e/MJ from both the coproduced sodium sulfate salt and surplus electricity for DDA.

Breaking down the GHG emissions of the biorefinery (the red bar in the figure below), the use of hydrogen (assumed in the base case to be sourced from SMR using natural gas without sequestration of evolved CO_2) contributes the most to the overall carbon intensity, accounting for 59% of the biorefinery's GHG emissions with DMR and 67% with DDA pretreatment. This is followed by chemical usage in the pretreatment step, contributing 27% for DMR and 20% for DDA. Meanwhile, enzyme production and hydrolysis each account for approximately 10% of total GHG emissions in both cases. While pretreatment chemicals (particularly in the DMR case) impart a non-trivial contribution to GHG emissions, to date alternative pretreatment methods with less intensive chemical use such as steam explosion have not been observed to achieve improvements due to inferior sugar yields, but this continues to be an area of research. Additionally, there is no carbon sequestration in the base case designs as fuels (destined for combustion) are the only carbon-containing outputs from the biorefinery (CO_2 capture and sequestration of biorefinery flue gas is considered as an alternative scenario further below).

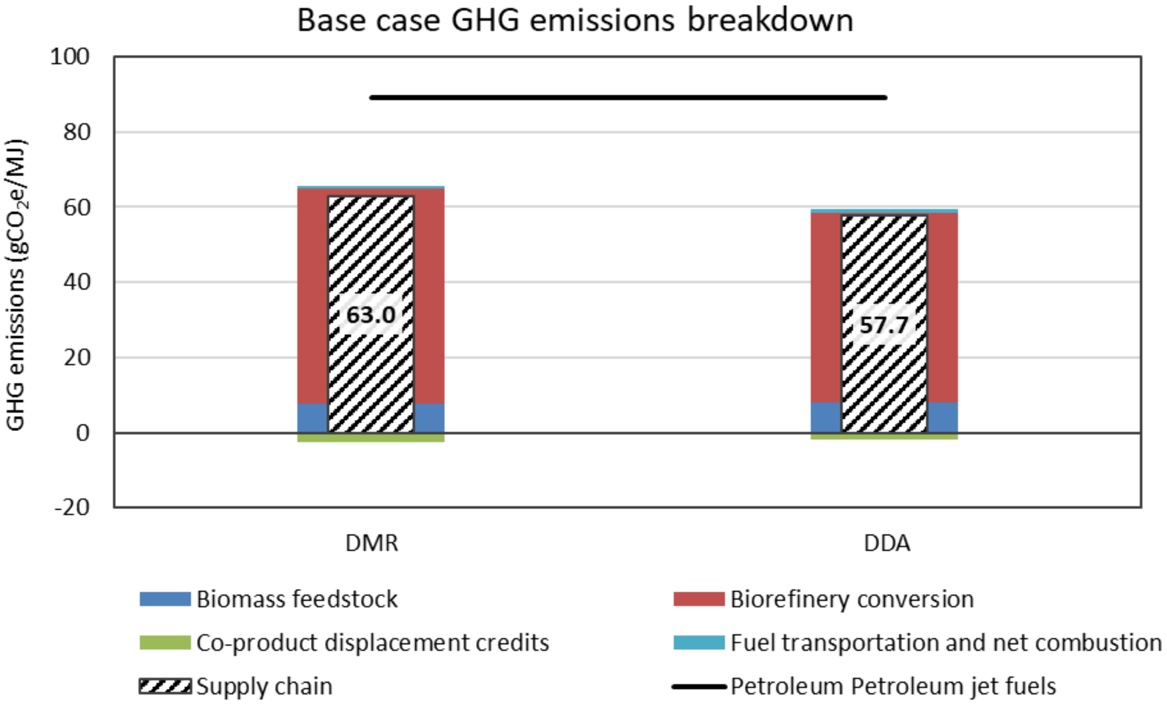


Figure 12. Supply chain GHG emissions for base case pathways, using displacement method to address effects of coproducts. Only non-biogenic CO_2 inputs and emissions are reflected.

The biorefinery-level emissions under the DMR and DDA pretreatment scenarios differ due to variations in fuel and coproduct yields, as well as their carbon intensities. Figure 13 compares the biorefinery-level GHG emissions with those from producing an equivalent amount of conventional fuels and products, measured in kg CO_2e per dry tonne of feedstock processed through the biorefinery. The DMR scenario results in a reduction of about 329 kgCO_2e per dry tonne, or 36% relative to conventional counterparts. Meanwhile, the DDA pretreatment scenario achieves a reduction of roughly 350 kgCO_2e of GHG emissions per dry tonne of feedstock processed, or 41% reduction relative to the incumbent. This highlights that both scenarios offer favorable initial GHG emission reductions, again with a slight advantage going to the DDA scenario.

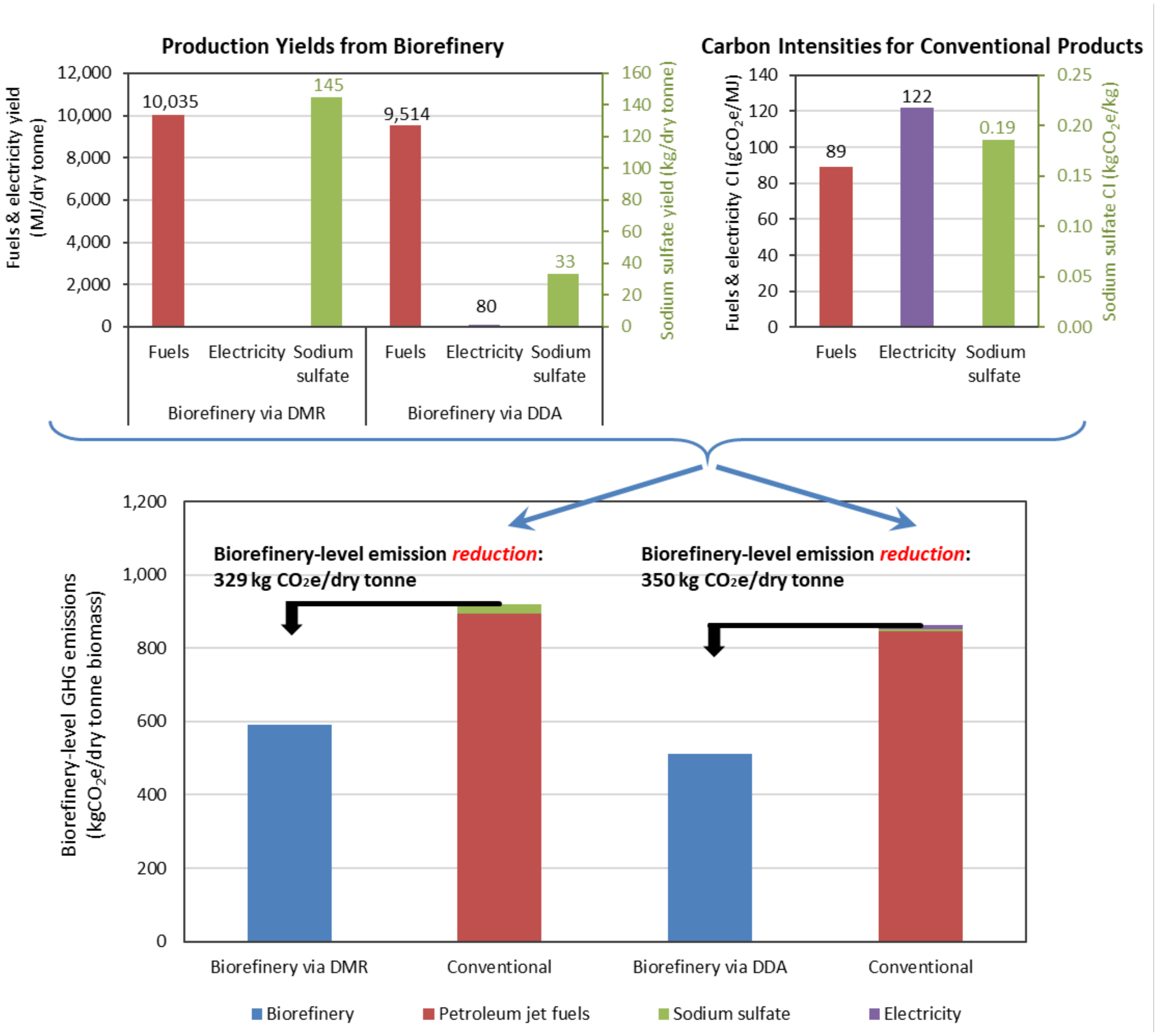


Figure 13. Base case biorefinery-level GHG emissions and reductions per dry tonne of biomass processed

Given the base case MFSP and GHG results presented above, the resulting marginal GHG abatement costs are calculated as presented in Table 15. Given moderate MFSP premiums over conventional fuel prices set here to be \$2.50/GGE based on weighted-average wholesale fuel projections through 2050 [5,59] coupled with less than 50% GHG reductions compared to conventional benchmarks (89 gCO$_2$e/MJ for jet fuel [54]), MAC values are somewhat elevated for the initial base cases, calculated at \$597–\$718/tonne CO$_2$e. However, given opportunities for substantial further improvement, particularly for GHG intensity based on alternative scenarios considered below, MAC metrics in turn also can be significantly reduced as presented further below.

Table 15. Marginal Cost of GHG Abatement Results for Base Case Scenarios

	DMR	DDA	Units
MFSP without policy incentives	$4.79	$4.79	$/GGE
Average fuel market selling price	$2.50	$2.50	$/GGE
Incremental cost of fuel	$2.29	$2.29	$/GGE
Translation at 122.5 MJ/GGE	$0.0187	$0.0187	$/MJ
CO_2 reduction of fuel	26.0	31.3	gCO_2e/MJ
Marginal Cost of GHG Abatement	**$718**	**$597**	**$/tonne CO_2e**

4.2 Sensitivity Analysis

Beyond the base case scenarios presented above for either pretreatment configuration, in this section a number of sensitivity cases are investigated. First, MFSP sensitivities to key economic and process parameters are considered based on varying individual parameters over a range of reasonable conditions as could be encountered within the context of the biorefinery base case designs, highlighting resultant drivers on MFSPs via tornado plots. Subsequently, sensitivities are evaluated on MFSP/GHG trade-offs attributed to alternative scenarios allowing for modifying base case assumptions around sourcing of key inputs and other biorefinery design options.

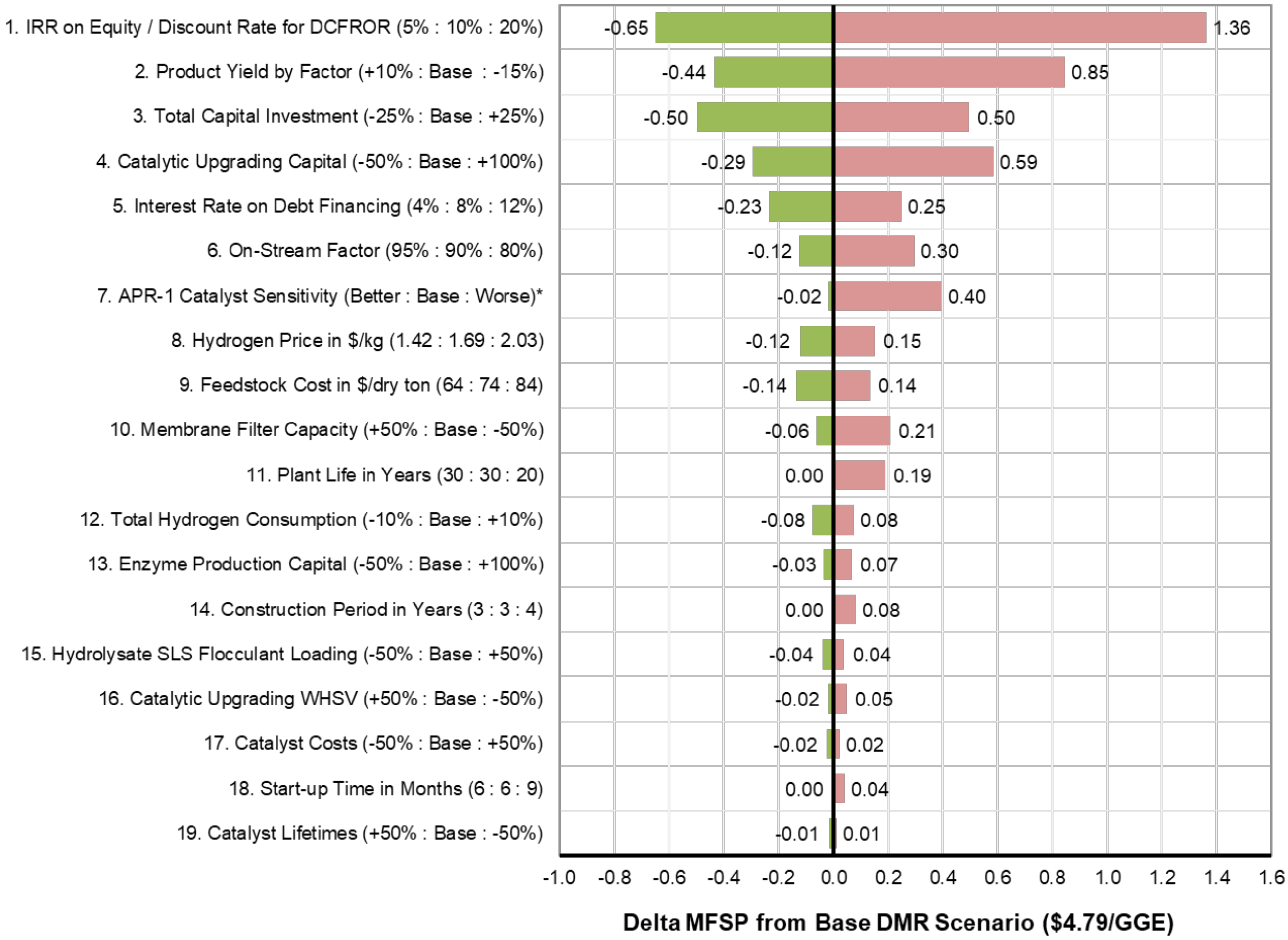
4.2.1 Single-Point Sensitivity Analysis

Figure 14 and Figure 15 present tornado plots highlighting MFSP sensitivities to a selected range of parameters for both pretreatment scenarios. Most parameters are self-explanatory and will not be discussed individually, beyond highlighting key points. Of the top five MFSP drivers, four are directly related to capital expenditures based on the variable ranges selected. The strongest single parameter evaluated was shown to be the discount rate (IRR), adding more than $1.3/GGE in either pretreatment case upon doubling the IRR from 10% to 20%, or proportionately reducing MFSPs by roughly half that amount if cutting the IRR in half from 10% to 5%. The overall uncertainty range in TCI of approximately ±25% would translate to a roughly ±$0.50/GGE MFSP impact, while potentially larger uncertainties in the capital costs attributed to the catalytic upgrading steps (conservatively varied from −50% to +100%) would add a similar amount to MFSP on the high side, or subtract roughly $0.30/GGE on the low side. Additionally, adding or subtracting four percentage points from the base case 8% debt financing interest rate would incur a roughly $0.25/GGE impact to MFSP.

Beyond capital-related expenses, the largest MFSP driver evaluated was shown to be overall product yields achievable through the catalytic upgrading steps, reflecting an MFSP impact of −$0.44/GGE up to +$0.85/GGE in both cases when varying yields by +10% or -15%, highlighting the importance of optimizing catalyst conditions for maximum fuel yields (this parameter did not differentiate between SAF and other fuels, as they are all combined to a total GGE basis; however, yield was varied to a greater extent on the downside recognizing these are already future aspirational projections and may be challenged in the near-term by current catalyst performance and/or sensitivities to hydrolysate impurities). In general, DMR sensitivities were shown to be marginally lower than DDA in many cases when varying the same parameter, due to slightly higher fuel yields in the DMR case and thus a larger $/GGE denominator serving as a buffer against individual cost impacts. However, membrane filter capacity for hydrolysate solids removal had a somewhat more outsized impact on the DMR case, as this is a more costly step in general for separating DMR-pretreated solids particles compared to DDA, as discussed above. Both feedstock cost and hydrogen purchase price were shown to incur a moderate influence on MFSPs, owing primarily to a tighter range of variance considered for either parameter (recognizing they both add nontrivial contributions to MFSPs as noted in Section 4.1). Namely, feedstock costs were varied by ±$10/dry ton relative to the target of $74/ton, translating to a ±$0.14/GGE MFSP impact, while hydrogen costs were varied between the low and high point prices for standard hydrogen production over the 5-year period that the average cost was calculated from, resulting in a −$0.12/GGE savings or +$0.15/GGE increase to MFSP accordingly.

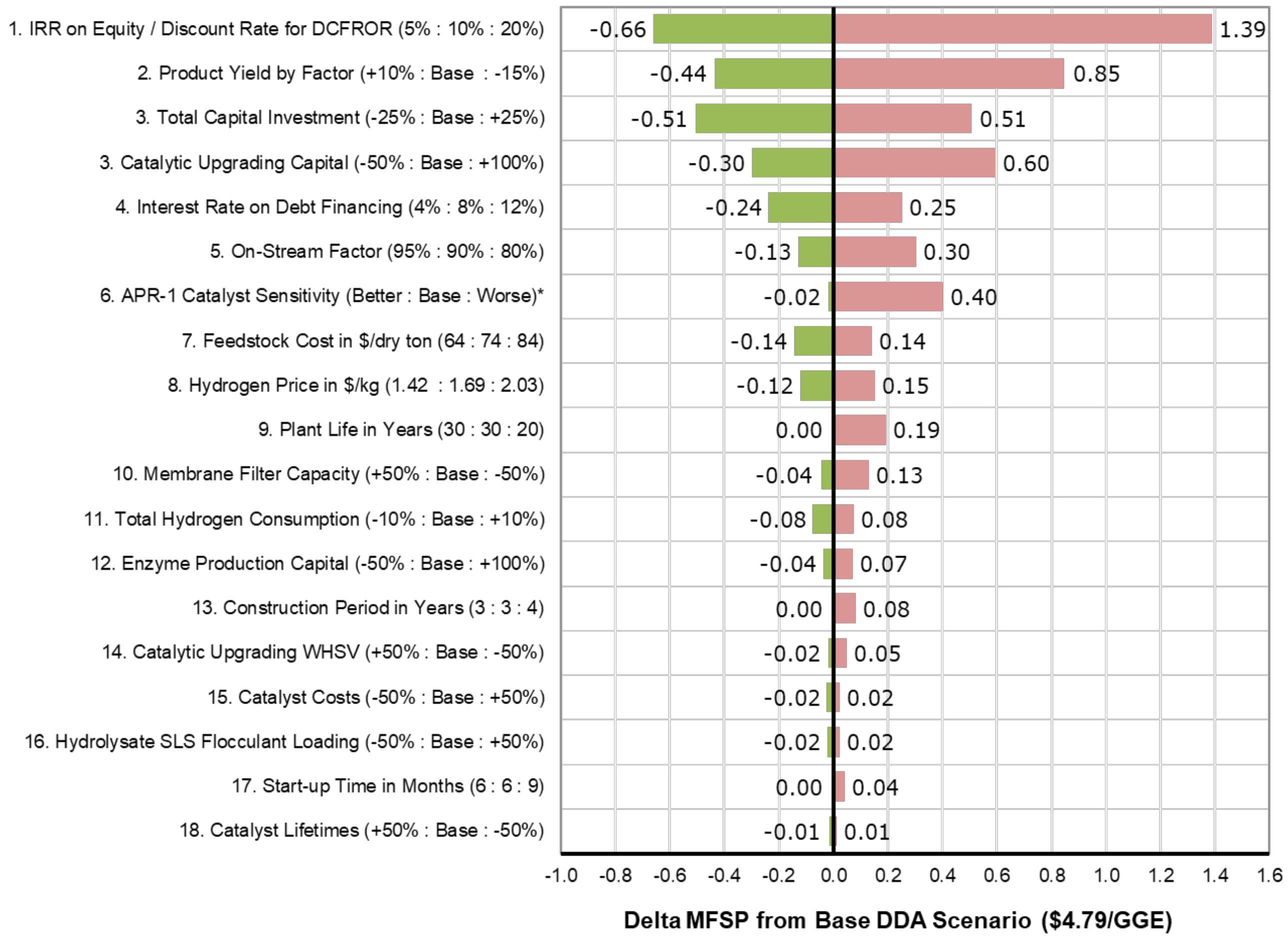
As noted above, NREL has not performed direct experimental work to establish the current state of performance for the catalytic steps e.g. regarding catalyst sensitivities on DMR or DDA-pretreated hydrolysates. However, applying a ± 50% variation in catalyst lifetimes or space velocities (as two key metrics tied to such considerations) did not lead to significant impacts in overall MFSPs when varied in isolation. As this consideration could particularly impact the first reaction stage processing the initial hydrolysate, an additional sensitivity case was carried out for the APR-1 catalyst lifetime parameter to be reduced by up to 75% while simultaneously increasing facility downtime from 10% to 20% (decreasing the on-stream factor from 90% to 80%). This combination is considered recognizing the APR-1 catalyst is already assumed to be replaced annually, thus any more frequent replacement needs than once per year could ultimately translate in turn to necessitating overall facility shutdowns more than once per year and challenging the ability for the biorefinery to operate at 90% on-stream capacity. In this case, MFSP impact would be more pronounced, adding $0.40/GGE in both cases, though the majority of this impact would be attributed to the lower biorefinery on-stream factor exhibiting a similar magnitude of MFSP sensitivity on its own (thus still incurring less substantial MFSP impacts due to catalyst costs).

Finally, a number of parameters were also shown *not* to incur substantial MFSP sensitivities under the range of variances considered. These include flocculant loading for the hydrolysate solids removal step, enzyme production capital expenses, hydrogen consumption, and catalyst costs. Sensitivities to longer facility construction or startup times were also shown to be minimal, i.e. if even n^{th}-plant assertions were too aggressive in reflecting a 3-year construction period followed by a 6-month startup time. However, it should be noted that hydrogen consumption is not an insignificant parameter (contributing $0.76/GGE to the MFSP in either pretreatment case, only marginally less than feedstock costs), but strictly that it does not significantly influence MFSP if the range in variability for this parameter is only 10% while maintaining the same fuel yields.



*Better: WHSV +50%, catalyst lifetime +50%, on-stream factor 90%. Worse: WHSV -75%, catalyst lifetime -75%, on-stream factor 80%.

Figure 14. Local sensitivity analysis for the base DMR case



*Better: WHSV +50%, catalyst lifetime +50%, on-stream factor 90%. Worse: WHSV -75%, catalyst lifetime -75%, on-stream factor 80%.

Figure 15. Local sensitivity analysis for the base DDA case

4.2.2 Alternative Scenarios: Input Sourcing, Carbon Capture, and Policy Incentives

While the base case reflecting default assumptions regarding biorefinery configurations and sourcing of inputs highlights favorable yields, a straightforward and simple processing schematic, and moderate GHG reductions compared to petroleum fuel benchmarks, it does not yet achieve the primary goal of 50%–70% GHG savings and thus also would not initially qualify for policy incentives except for California's LCFS to further improve economic viability. However, several opportunities exist to significantly further reduce the carbon intensity beyond base case levels, as may reach and ultimately exceed thresholds for consideration of policy incentives—these options are summarized in Table 16. As a key driver identified above, hydrogen is evaluated across several scenarios expanding beyond "gray" hydrogen derived via steam methane reforming (SMR) from natural gas—namely "blue" hydrogen (SMR with capture of released CO_2) and "green" hydrogen (derived via water electrolysis using zero-GHG power), reflecting cost and GHG intensity values from literature (considering a range of potential costs for green hydrogen spanning current versus future cost potential at $4.50 and $2.00/kg) [12,60–64]. These prices generally reflect production costs alone without explicit inclusion of transportation or storage, which could add incremental additional costs, but this was not considered here as such cost increases would fall well within the large range of sourcing costs already evaluated. Direct biorefinery power consumption is also evaluated to be obtained from zero-GHG sources, nominally assuming no change in electricity costs relative to standard grid power as expected to continue approaching cost parity in the near future [65].

Beyond simpler changes such as cost and CI for alternative power or hydrogen sourcing options, several additional considerations were also implemented requiring additional process/modeling granularity. First, recognizing strong CI drivers attributed to NaOH usage in deacetylation (for both DMR and DDA), an alternative scenario is considered for on-site production of NaOH via the chlor-alkali process—an electrolysis process converting water and sodium chloride to NaOH, as well as hydrogen and chlorine gas byproducts. Unlike purchased NaOH, which is assumed in LCA to be produced by a market average represented by 66% via diaphragm cell technology and 34% via membrane cell process technology [66], the on-site NaOH case adopts the membrane cell process technology, which is more energy efficient than the diaphragm cell process. Capital and operating costs as well as corresponding NaOH, hydrogen, and chlorine byproduct yields, were estimated based on industry database information, translating to a net NaOH-equivalent cost of $0.49/kg relative to externally purchased NaOH at $0.26/kg. Because the primary driver behind the GHG intensity of chlor-alkali NaOH is power for electrolysis, this scenario is only considered assuming zero-GHG electricity as a prerequisite for this approach to be practical. We also note that under a future *n*th-plant scenario, it may be more convenient for the biorefinery to purchase “green NaOH” from a producer offering this via their own chlor-alkali plant operated on zero-GHG power. While the electricity source is the strongest driver on the CI of NaOH inputs, the savings in incremental costs and complexity would need to be weighed against other on-site NaOH integration benefits such as co-production of hydrogen for use within the biorefinery. Other market factors may also play into such decisions, such as keeping NaOH and chlorine markets in balance, as chlorine is a second major chlor-alkali product.

Table 16. Cost and GHG Intensity of Base Case vs. Alternative Integration Options

Input Sourcing	Base Case Cost	Base Case GHG Intensity	Alternative Case Cost	Alternative Case GHG Intensity
Hydrogen (base: gray, alternative: blue, green)	$1.69/kg	9.44 kgCO$_2$e/kg	Blue: $3.00/kg Green: $2.00–$4.50/kg (future target vs. near term)	Blue: 5.00 kgCO$_2$e/kg Green: 1.70 kgCO$_2$e/kg
Electricity (base: U.S. grid average, alternative: zero-CI power)	Purchasing: 8.17 ¢/kWh Selling: 5.80 ¢/kWh	440 gCO$_2$e/kWh	Purchasing: 8.17 ¢/kWh	Negligible
NaOH (base: off-site NaOH purchase, alternative: on-site NaOH production with zero-CI power)	$0.26/kg	2.05 kgCO$_2$e/kg	$0.49/kg	0.97 kgCO$_2$e/kg
Biorefinery CCS (not used in base case)	N/A	N/A	See Table 17. for correlation details	

Finally, to consider maximum decarbonization potential, we also consider a scenario implementing carbon capture and sequestration (CCS) of biorefinery CO_2 emissions. The Integrated Environmental Control Model (IECM) [67], a software program created by Carnegie Mellon University (currently being developed by the University of Wyoming), was used as the primary data source to construct this CCS scenario. IECM enables users to build and analyze their own power plant models by choosing from technology options that are available in three fossil fuel power plant configurations: pulverized coal, integrated gasification combined cycle, and natural gas combined cycle. One of these plant components is "CO_2 capture and storage" with more than 10 different capture technologies, such as amine capture or membrane separation, which become available to choose from based on the plant configurations put together by the user. IECM has already been used to build capture-technology-agnostic correlations for the cost of CO_2 [68,69]. Here, we follow a similar approach but reflecting additional granularity to gather detailed mass and energy flows along with capital costs for each unit around a single capture technology, as well as the cost for transportation and storage. Cost correlations can be directly integrated into MFSP calculations; however, mass and energy flows are required to conduct an LCA.

Flue gas CO_2 compositions from these three power plants vary—approximately 4.2% for natural gas combined cycle (v/v), 12% for pulverized coal, and 37% for integrated gasification combined cycle [68]. As a comparison, the primary point source in our models (combustor flue gas) emits more than 96% of the biorefinery CO_2 that can be captured, at a flue gas CO_2 concentration of 10.9%–11.8% (v/v) in the DMR and DDA cases, respectively. Hence, IECM data for typical natural gas combined cycle and pulverized coal power plants were leveraged, which offer two capture technologies in common: amine- or ammonia-based absorption systems. Correlations shown in Table 17. were fitted for an amine-based system by varying the percent carbon capture between 60% and 90%, and the amount of flue gas through different plant sizes. The plant size is altered by varying the number of turbines in the natural gas combined cycle case from one to five, and in the pulverized coal case by directly varying the gross electrical output between 100 and 650 MW. In both cases, the capacity factor is set to 90% (7,884 hours/year) and the cost year to 2020, in line with our modeling efforts, while keeping all other parameters at their default values/definitions. A capture rate of 90% is assumed in this work and applied to each correlation in Table 17. . These correlations were applied as an external layer to the models, e.g. additional steam and energy (and all other) inputs for CCS were assigned costs and GHG emissions assuming external sourcing without explicitly tying into Aspen Plus model energy integrations.

Table 17. Fitting Coefficients from IECM for Installed Equipment Costs, Mass and Energy Flows, and Other Cost Items for Carbon Capture via Amine Absorption and Geological Storage.

α: percent capture rate, β: weight ratio of CO_2 in flue gas, γ: flue gas flow rate (tons/h)

			$10^{a+b\log\alpha+c\log\beta+d\log\gamma}$				
Installed Equipment Cost	**Correlation Yields**	**Unit Price**	**a**	**b**	**c**	**d**	**R^2**
Direct contact cooler	2020 $		3.84	0.45	−0.70	0.83	0.99
Flue gas blower			3.25	0.44	−0.58	0.83	0.99
CO_2 absorber vessel			4.40	0.44	−0.58	0.83	0.99
Heat exchangers			2.40	1.10	−0.01	0.83	0.99
Circulation pumps			2.71	1.10	−0.01	0.83	0.99
Sorbent regenerator			3.28	1.10	−0.01	0.83	0.99
Reboiler			2.17	1.10	−0.10	1.08	1.00
Steam extractor			3.14	0.98	0.06	0.56	0.87
Sorbent reclaimer			3.57	0.60	0.85	0.83	0.99
Sorbent processing			5.01	0.21	0.84	0.60	1.00
Drying and Compression Unit			3.90	0.68	-0.01	0.90	0.99
Mass and Energy Flows							
Monoethanolamine (MEA)	ton/h	$1.22/lb [a]	−3.01	0.28	1.39	1.00	1.00
Activated carbon	ton/h	$1.13/lb [a]	−6.13	1.00	0.00	1.00	1.00
Caustic	ton/h	12.0 ¢/lb	−2.75	0.26	1.49	1.00	1.00
Reclaimer waste	ton/h	12.0 ¢/lb [a]	−1.88	0.16	1.91	1.00	1.00
Electricity	MWh/h	8.17 ¢/kWh	−2.89	0.05	−1.12	1.00	1.00
Water	gal/h	0.14 ¢/gal	−0.52	0.00	−2.30	1.00	1.00
Steam for amine regeneration [b]	MWh/h	8.17 ¢/kWh	−2.83	1.03	−0.15	1.00	1.00
Compression electricity	MWh/h	8.17 ¢/kWh	−3.07	1.00	0.00	1.00	1.00
Other cost items							
Pipeline transport cost	$/ton CO_2 captured		4.27	−0.87	−0.01	−0.86	1.00
Geological storage cost			4.12	−0.81	0.30	−0.73	0.98

[a] Unit prices from IECM (for others, see Table 11).

[b] Electricity equivalent.

Upon implementing the alternative sourcing/configuration options described above, the resulting trade-offs between MFSPs and GHG emissions are depicted in Figure 16 and Figure 17 for DMR and DDA pretreatment, respectively. Given the particular prominence of hydrogen sourcing assumptions and recognizing this parameter need not be binary (i.e., 100% of one source or another), these figures are constructed to consider a range of inputs between 100% gray hydrogen (right-most point on curves) and 100% blue or green hydrogen (again with the latter considering two cost scenarios; left-most point on curves). Based on these figures, the base case could reach 50% GHG emissions reduction through sourcing roughly 66% of the hydrogen demands via green hydrogen in the DMR pretreatment case, and sourcing 47% via green or 83% via blue hydrogen in the DDA pretreatment case. Renewable power for biorefinery operations marginally reduces GHG emissions, but subsequently including on-site NaOH production (also with renewable power for chlor-alkali electrolysis) can achieve a more meaningful reduction of roughly 7 gCO_2e/MJ while incurring a fairly minimal 2% increase in MFSP. These modifications are based simplistically on swapping grid power for renewable power and external NaOH for on-site NaOH (also with renewable power), though not accounting for any requirements such as battery storage or additional NaOH storage to manage a variable supply of renewable power. This subsequently allows for an additional reduction of 27 gCO_2e/MJ when coupled with green hydrogen sourcing in both DDA and DMR cases, though still not quite enough to reach 70% GHG reduction thresholds except for DDA with green hydrogen. However, the carbon intensity improves much more dramatically to reach *net-negative* GHG emissions when considering incorporation of biorefinery CCS, reaching a range approximately between –12 and –33 (DMR) or –18 and –39 (DDA) gCO_2e/MJ between the base case versus inclusion of renewable power and on-site NaOH production for 100% gray hydrogen—corresponding to an MFSP increase of \$0.85–\$0.91/GGE from the base case for DMR versus DDA pretreatment. This could be further reduced to values of –40 to –60 (DMR) or –45 to –66 (DDA) gCO_2e/MJ for either configuration basis in a best-case scenario with 100% green hydrogen.

Figure 18 and Figure 19 present an alternative means of evaluating such TEA/LCA trade-offs through waterfall plots, reflecting sequential implementations of these strategies generally ordered from the most readily implementable (renewable electricity) to the least (CCS, which is dependent on local geology, access to pipeline infrastructure, and regulatory policies, and has recently faced a number of challenges in commercial deployment [70]. Supply constraints for green or blue hydrogen (coupled with similar logistical constraints with CCS for blue hydrogen) may also make such options more challenging to implement in the immediate future [71]. Again, first through inclusion of renewable power and on-site NaOH production, GHG intensities for either pretreatment option can be reduced by 11%–13% with minimal (2%) impact on MFSP increases. Subsequently, GHG emissions could be reduced more substantially by an additional 28%–49% (DMR) or 31%–54% (DDA) upon switching to 100% blue or green hydrogen sourcing, although at more significant MFSP penalties ranging between 12% and 26% based on near-term hydrogen costs—albeit at a much more moderate 3% MFSP penalty based on potential future costs for green hydrogen. Alternatively, MFSPs with green hydrogen at a cost as low as \$1/kg reflecting longer-term DOE goals by 2050 [72] would be \$4.56/GGE and \$4.59/GGE for DMR and DDA respectively, but could increase to as high as \$6.80/GGE and \$6.83/GGE for near-term costs for green hydrogen as high as \$6/kg (after still including renewable power and on-site NaOH production following the same waterfall plot sequence). Including lower-GHG blue or green hydrogen can exceed 50% and in some cases 70% GHG reduction thresholds (for DDA pretreatment with green hydrogen), after first including renewable power and on-site NaOH production, thus triggering the potential to include decarbonization policy incentives (considered further below, although LCFS credits would still apply at lower GHG reduction levels). And finally, after then including biorefinery CCS on top of all other prior modifications, GHG emissions could greatly reduce to less than zero gCO_2e/MJ for all cases, while controlling MFSP increases to 16% or less. Alternatively, Figure 18 and Figure 19 also highlight that merely moving to CCS without any other modifications could potentially achieve the most favorable single cost-benefit trade-off given substantial GHG reductions under relatively moderate MFSP penalties (though recognizing other factors influencing the viability of CCS implementation including pipeline transportation costs and logistical, regulatory, and community acceptance considerations).

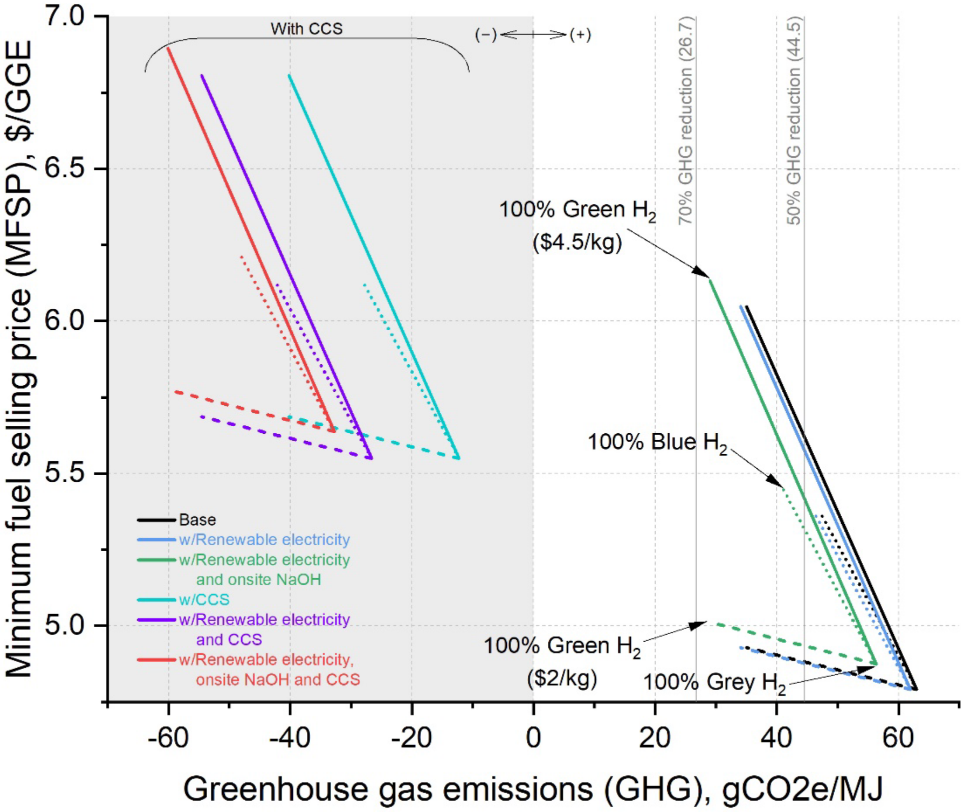


Figure 16. Impacts on MFSP vs. GHG emissions trade-offs reflecting alternative sourcing/configuration options: DMR pretreatment

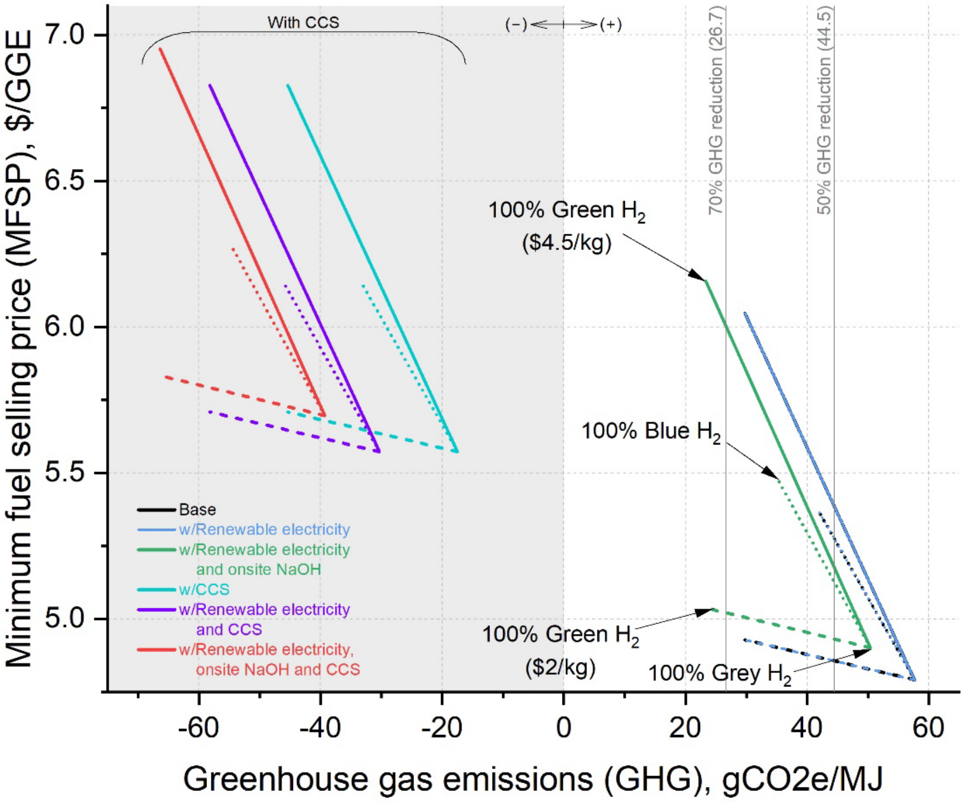


Figure 17. Impacts on MFSP vs. GHG emissions trade-offs reflecting alternative sourcing/configuration options: DDA pretreatment. Note: base case overlaps with "base case plus renewable electricity" curve, thus a separate base case curve does not appear visible in this plot.

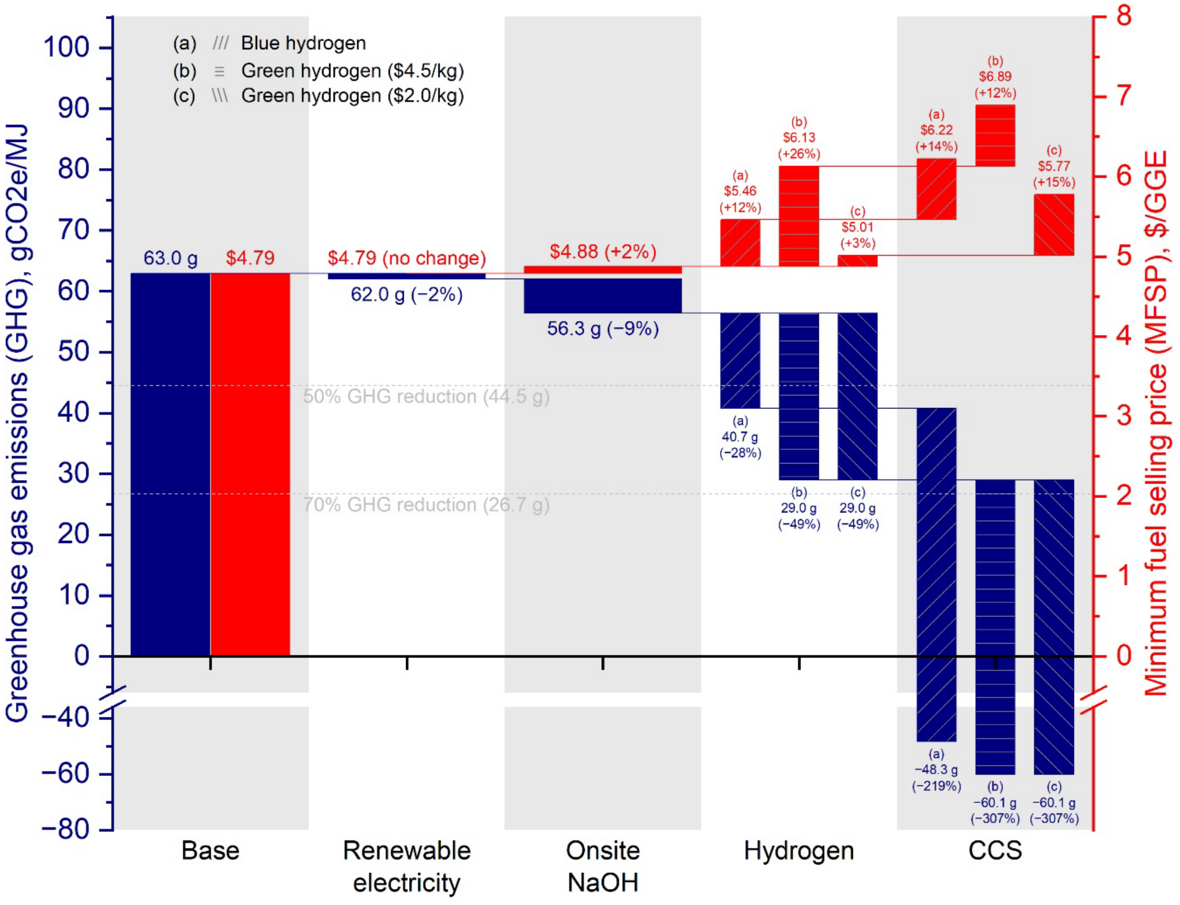


Figure 18. Waterfall plot depicting TEA/LCA trade-offs for sequential implementation of alternative biorefinery sourcing/configuration options: DMR pretreatment

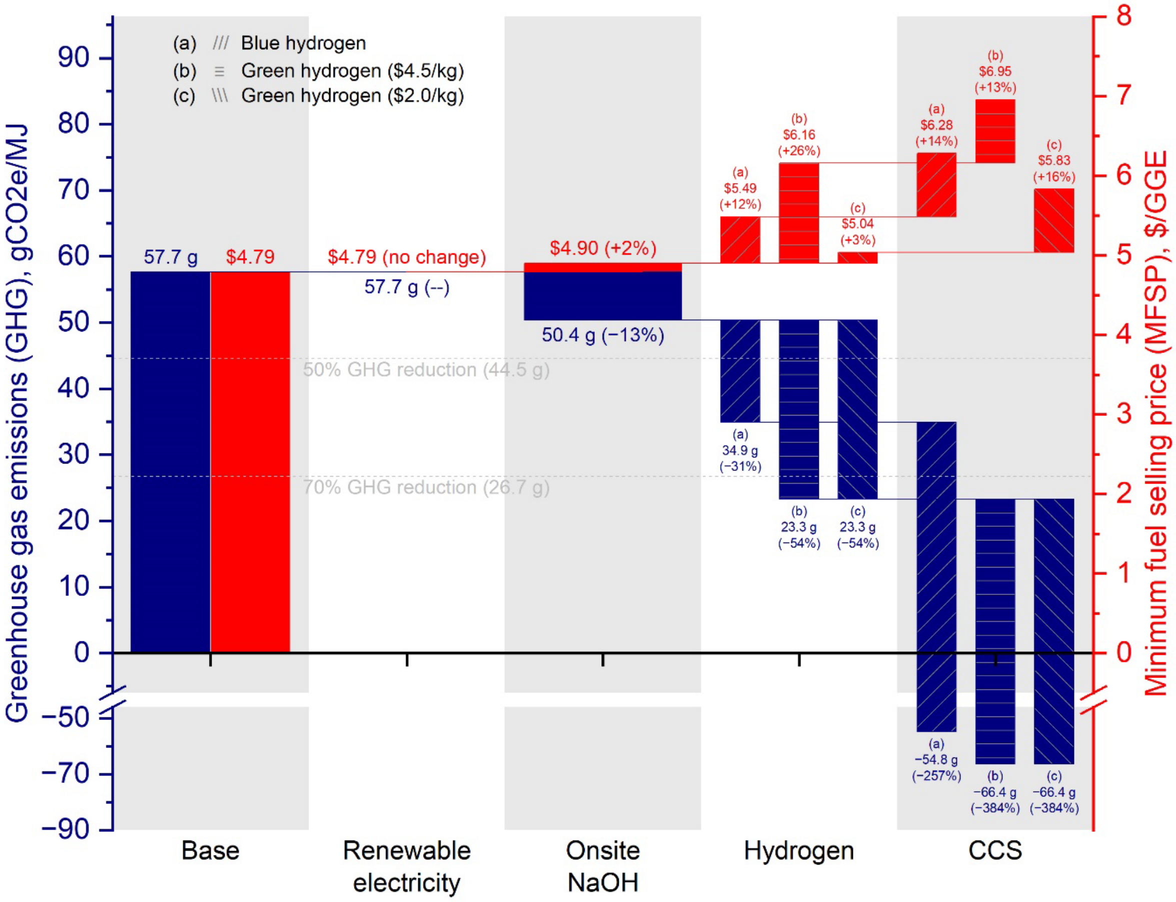


Figure 19. Waterfall plot depicting TEA/LCA trade-offs for sequential implementation of alternative biorefinery sourcing/configuration options: DDA pretreatment

Marginal GHG abatement costs for the above sequential scenarios are summarized in Table 18. Relative to base case MACs in the range of $597–$718/tonne CO_2e, the TEA/LCA trade-offs presented above would translate to GHG benefits outweighing MFSP penalties, even without inclusion of policy incentives, resulting in sequentially lower MAC projections upon incorporation of the same sequential GHG reduction strategies. After including renewable electricity, on-site NaOH production, and blue or green hydrogen sourcing, MAC costs could be reduced to a range of $316–$501/tonne CO_2e between the two pretreatment scenarios. From there, upon subsequently including biorefinery CCS, MAC values could be further improved to $175–$241/tonne CO_2e. Although there is not a carbon credit market in the United States for comparison of these values, this indicates quite favorable potential for decarbonization within reasonable cost constraints, again when viewed in comparison to DAC on the order of $600/tonne or greater for current costs or $100–$600/tonne for plausible future DAC costs (with the lower range of this likely to represent highly optimistic costs) [37–39].

Table 18. Marginal Cost of GHG Abatement Results for Alternative Scenarios

Results for DMR / DDA	Incremental Cost of Fuel ($/GGE)	Incremental CO_2 Reduction of Fuel (gCO_2e/MJ)	Marginal Cost of GHG Abatement ($/tonne CO_2e)
Base case	$2.29 / $2.29	26.0 g / 31.3 g	$718 / $597
1: Renewable electricity	$2.29 / $2.29	27.0 g / 31.3 g	$692 / $597
2: (1) + on-site NaOH	$2.38 / $2.40	32.7 g / 38.6 g	$594 / $508
3a: (2) + blue hydrogen	$2.96 / $2.99	48.3 g / 54.1 g	$501 / $451
3b: (2) + green hydrogen @ $4.50/kg	$3.63 / $3.66	60.0 g / 65.7 g	$494 / $455
3c: (2) + green hydrogen @ $2.00/kg	$2.51 / $2.54	60.0 g / 65.7 g	$342 / $316
4a: (3a) + CCS	$3.72 / $3.78	137.3 g / 143.8 g	$221 / $215
4b: (3b) + CCS	$4.39 / $4.45	149.1 g / 155.4 g	$241 / $234
4c: (3c) + CCS	$3.27 / $3.33	149.1 g / 155.4 g	$179 / $175

As evidenced above, opportunities exist to achieve deep decarbonization of this biorefinery pathway even into net-negative CI ranges through a combination of one or more of the above strategies, albeit at varying degrees of cost penalties before considering policy incentives. However, the latter is also important to consider in terms of cost-benefit trade-offs to the biorefinery. Based on incorporation of the various policy credits discussed in Section 3.4 (except for 40B and 45Z-SAF which are not initially included in the base case calculations), impacts to MFSPs are summarized below for the key scenarios upon initially reaching pertinent GHG reduction thresholds to trigger applicable incentives. Figure 20 and Figure 21 repeat the same waterfall plots as Figure 18 and Figure 19 (based on sequential inclusion of each modification in order), but now also including consideration of decarbonization policy credits based on the assumptions outlined previously for their implementation in the present analysis.

Upon achieving 50% or greater GHG reduction levels enabled by switching to blue or green hydrogen (in addition to renewable electricity and on-site NaOH production), the combined stacking of all relevant policy incentives applied to SAF production translates to a potential MFSP reduction of roughly $0.56–$4.14/GGE for blue hydrogen, increasing up to roughly $4.20–4.30/GGE for green hydrogen reflective of larger GHG reductions in the latter case. This credit would offset or in most cases outweigh the added expenses of either hydrogen sourcing option, even in the near-term scenario for high-cost green hydrogen, resulting in MFSPs lower than the original base cases (except in the DMR case with blue hydrogen, which would almost exactly offset the cost due to the fact that at 40.7 gCO_2e/MJ, it is not eligible to receive D3 RIN credits as that requires a minimum of 60% emissions savings, thus achieves substantially less MFSP reduction benefits than the other cases exceeding the 60% threshold). Moreover, this trend would continue upon implementation of biorefinery CCS, enabling credits of approximately $5.50/GGE or more, with a "best case" scenario reaching MFSPs below $1/GGE tied to future/low-cost green hydrogen sourcing plus CCS.

Beyond these base case policy considerations, a number of additional policy scenarios were also investigated as summarized in Table 19. Even though SAF from this conversion technology is not assumed in the base case to be eligible to receive 40B or 45Z-SAF credits, as previously discussed, in an alternative case assuming eligibility via ASTM certification, these additional credits could potentially further bring down the MFSPs to negative values, e.g. to −\$0.68/GGE (DMR) and −\$0.76/GGE (DMR), when green hydrogen priced at \$2/kg is employed (see Table 19). Without the inclusion of these additional SAF policy credits, starting up the facility for production after the 45Z has expired in 2028 (instead of 2023), on the other hand, would add 15 cents per GGE to the lowest MFSPs from the both pretreatment cases indicated in Figure 20 and Figure 21, under the current assumptions, as also shown in Table 19. While we caution that such favorable results as presented here may represent a hypothetical best-case scenario with respect to the value and duration of policy incentives as discussed in Section 3.4 as the future of such incentives remains unclear, this still highlights the potential for favorable deployment in the near term under such policy applications as a key factor that would undoubtedly be taken into consideration if this technology were to be commercialized under the current policy structures.

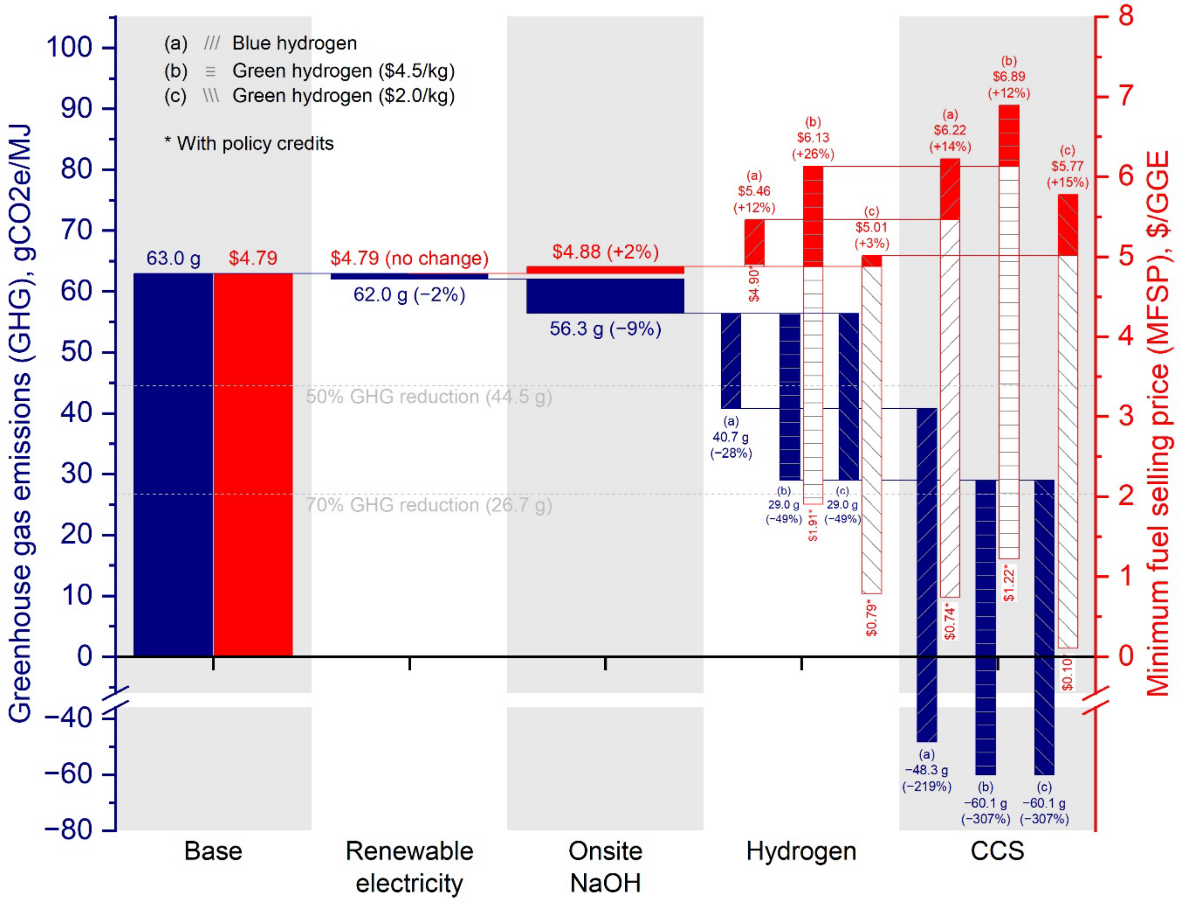


Figure 20. Waterfall plot depicting TEA/LCA trade-offs for sequential implementation of alternative biorefinery sourcing/configuration options, with inclusion of policy credits (white hashed bars): DMR pretreatment

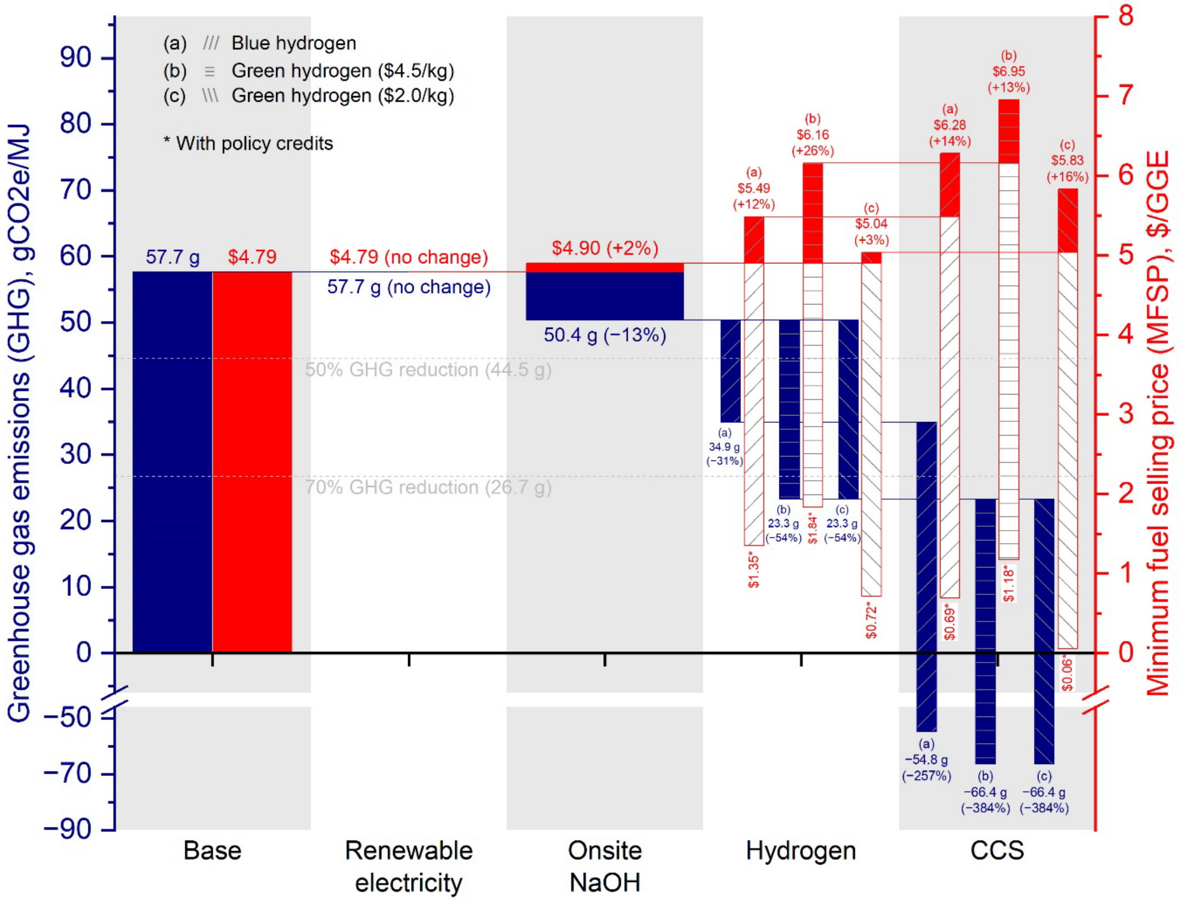


Figure 21. Waterfall plot depicting TEA/LCA trade-offs for sequential implementation of alternative biorefinery sourcing/configuration options, with inclusion of policy credits (white hashed bars): DDA pretreatment

Table 19. Additional Scenarios Evaluating the Impact of Construction and Start-up Periods, Start of Operations, and the Applicability of 40B and 45Z-SAF Policy Credits on Net MFSP ($/GGE)

Net MFSP ($/GGE)	**Without 40B or 45Z-SAF**		**With 40B and 45Z-SAF**	
	DMR	DDA	DMR	DDA
Reference case*	0.10	0.06	-0.68	-0.76
4-year construction	0.20	0.16	-0.57	-0.65
9-month start-up	0.15	0.10	-0.63	-0.71
Production starting in 2028	0.25	0.21	0.25	0.21

*Reference case: Renewable electricity, onsite NaOH, green H_2 ($2/kg), CCS; three-year construction period, six-month start-up period, production starting in 2023.

5 Concluding Remarks

5.1 Summary

In keeping with recent pivots to present more comprehensive analysis packages spanning economic, sustainability, and near-term deployment considerations emphasizing SAF decarbonization, this report establishes the first in a planned series of technology case studies meeting these goals. Focusing on an example biorefinery pathway coupling relatively mature deconstruction technologies for hydrolyzing lignocellulosic biomass carbohydrates to sugars, coupled with APR catalytic upgrading of sugars to SAF, we present TEA/LCA trade-offs between MFSP and GHG reduction metrics for initial base case configurations utilizing either DMR or DDA pretreatment, alongside alternative scenario cases offering the potential for further GHG reduction. Base case results representing the most readily deployable scenario were found to achieve an MFSP of $4.79/GGE for either pretreatment option (balancing tradeoffs between yields versus costs) and GHG emissions reductions of 29% and 35% relative to fossil jet fuel benchmarks for DMR and DDA pretreatment, respectively. As these cases otherwise target maximum process simplicity with respect to lignin disposition (combusting lignin for heat and power rather than lignin upgrading to fuels or products, though recognizing the latter will offer substantial breakthroughs in the future), this represents a favorable initial result for a SAF-focused conversion pathway from sugars. This is primarily due to high carbon retention efficiencies translating to very high fuel yields on the order of 70–74 GGE/dry ton biomass, a notable increase of roughly 60%–70% beyond target fuel yields achievable through biological sugar conversion pathways investigated in prior NREL analyses.

Moving beyond the base cases, the application of optional sourcing and design decisions was shown to offer substantial opportunities to further reduce GHG intensities. First, moving to renewable power coupled with on-site production of NaOH (i.e., lower-CI NaOH produced via electrolysis using renewable power) could reduce GHG intensity between 11% and 13% while only increasing MFSPs by 2%. Subsequently moving to blue or green hydrogen sourcing could more substantially reduce GHG intensity by an additional 28%–54% depending on the case, albeit at more substantial MFSP penalties between 12% and 26% given large hydrogen demands for this pathway as a trade-off for avoiding CO_2 rejection losses in the fuel upgrading process. However, longer-term cost projections for green hydrogen could mitigate MFSP penalties to no more than 3%. Finally, incorporating biorefinery CCS could substantially reduce GHG intensities by more than 200% while limiting MFSP increases to 16% or less, allowing the potential for strongly net-negative GHG emissions (substantially surpassing 70% GHG reduction targets). When taken all together, these strategies offer the potential to reduce marginal GHG abatement costs from $718 and $597/tonne CO_2e in the base cases for DMR and DDA pretreatment, respectively, to as low as $179 and $175/tonne CO_2e if based on green hydrogen under future cost projections at $2/kg plus biorefinery CCS. Both the base cases and particularly the optimized cases compare favorably against direct air capture at near-term benchmark costs of $600 or more per tonne CO_2 captured. Moreover, a preliminary assessment of applicable policy incentives indicated the potential to reduce MFSPs to well below $2.5/GGE thresholds for economic viability in either pretreatment case for the same best-case scenario with CCS and low-cost green hydrogen sourcing.

5.2 Future Work

Moving forward, a number of opportunities exist to further optimize the process beyond base case and alternative scenarios considered here, as may further improve TEA/LCA metrics and strengthen the case for deployment relative to other existing incumbent technologies (e.g., ethanol-to-jet as an example pertinent to similar biochemical processing pathways). These represent opportunities for future consideration, including:

- **Evaluate opportunities to optimize with feedstock logistics:** While the present study maintains the same underlying assumptions for feedstock type (whole corn stover blends) and scale (2,000 dry tonne/day integrated biorefinery) consistent with prior historical design cases, future work may benefit from revisiting these assumptions in light of ongoing feedstock logistics research. As discussed in this report, recent analyses by INL and NREL have shown promise for alternative approaches such as air classification to separate corn stover into its anatomical fractions, blend stover with other herbaceous feedstocks, or move to a more localized feedstock preprocessing depot approach.
- **Reduce pretreatment GHG intensity:** Pretreatment remains a key driver behind biorefinery GHG emissions attributed to high energy usage (power in the case of DMR, steam heating in the case of DDA) and more significantly to caustic demands (applicable in both pretreatment strategies that utilize deacetylation, but particularly so for DMR given elevated caustic requirements). As such, additional opportunities remain to further decarbonize pretreatment. This could include chemical-free strategies (e.g., hot water or steam explosion) or chemical recycling strategies (e.g., enabling caustic recycle either through separation strategies or other means). These remain active areas of NREL research.
- **Reduce costs for enzymatic hydrolysis:** While capital/operating costs for the enzymatic hydrolysis step itself are fairly minimal, costs for enzymes as well as downstream solid separations remain expensive. As such, opportunities remain to optimize this step in connection with related unit operations through process consolidation. Another topic of NREL research discussed in this report focuses on a novel continuous enzymatic hydrolysis process, circulating hydrolysate through membrane filters to continuously remove sugars as they are produced [5]. This concept has the potential to recover/recycle enzymes for more than one use, increase sugar yields relative to batch hydrolysis, and enable process consolidation via simultaneous solid/liquid separation without the need for costly downstream clarification equipment and flocculants.
- **Improve catalyst stability and selectivity:** As with any catalytic system, opportunities always remain to reduce catalyst costs, such as through improved stability (particularly as may relate to impurities in real hydrolysate substrates and tailoring or ideally reducing the need for hydrolysate cleanup steps), catalyst lifetime, and weight hourly space velocity. As NREL has not directly investigated catalyst performance on corn stover hydrolysates for this technology pathway, this presents a key opportunity for future research to better understand bottlenecks and challenges that may be overcome through further R&D tied to such considerations. Additionally, while fuel yields are already targeted to maximize SAF at 70 wt % selectivity versus diesel- and naphtha-range fuels, conditions or catalyst details may potentially have room for further optimization to increase SAF selectivity in light of priorities shifting toward this component.

- **Continue research on lignin utilization pathways:** While lignin valorization via deconstruction/upgrading to coproducts or fuels was not included in the present study, this is not a broader reflection on the viability for such strategies, but rather a design decision in keeping with this study's emphasis on near-term scale-up and deployment (which favors simply burning the lignin for heat and power). Rather, lignin upgrading remains an active area of NREL and partner research efforts, with the potential to be a significant breakthrough in fundamentally boosting biorefinery economics and carbon conversion efficiencies beyond what is possible from carbohydrate components alone. While base-catalyzed deconstruction remains one possible option for lignin deconstruction, oxidative and reductive strategies are also being investigated, including numerous configurations for reductive catalytic fractionation, showing promising results to date [73]. Alternatively, models could also be constructed to prioritize energy self-sufficiency e.g. by feeding additional biomass to the boiler to fully satisfy all biorefinery energy demands.
- **Evaluate alternative sugar conversion options:** As discussed throughout this report, the present analysis opted to focus on APR as one example representative pathway for upgrading sugars to SAF with good potential to meet 50%–70% or higher GHG reduction goals under reasonable economic conditions. While this establishes a near-term benchmark as may offer competitive viability against other established SAF pathways (ethanol-to-jet, among others), other opportunities may offer "game-changer" potential, albeit based on longer-term research at a lower technology readiness level. This could include enzymatic or biological conversion routes (maintaining key advantages for such routes to yield specific single products) at yields exceeding current metabolic limits with the potential to reduce or eliminate CO_2 evolution (e.g., via cell-free or metabolic engineering strategies). Such options may serve as the basis for a future study, to compare against other benchmarks such as ethanol-to-jet or the present APR pathway.
- **Expand TEA modeling scope for near-term deployment/risk analysis:** Further opportunities exist to expand the scope of TEA modeling to consider tradeoffs between *n*th-plant versus pioneer plant economics, as well as additional means (such as Monte-Carlo modeling) to incorporate financial risk into the analysis, for this or any alternative biorefinery pathway.

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Appendix A. Individual Equipment Costs Summary

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

DMR Case:

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
A200: Pretreatment	Mechanical Equipment List					Scaled Installed Costs											
Deacetylation																	
Deacetylation Tank Discharge Pump	1771 GPM, 150 FT TDH	100	316SS	2	$22,500	2009	$22,500	strm.a200.211a	402194	kg/hr	0.80	2.3	425771	1.06	$47,098	$53,803	$123,748
Deacetylation reactor	14' x 30' vessel, quad discharge screw live btm	15hp ea	SS316	6	$780,000	2013	$780,000	strm.a200.211a	277167	kg/hr	0.70	1.7	425771	1.54	$6,320,480	$6,642,465	$11,292,190
Deacetylation reactor conveyors	Feed and discharge drag conveyors	40 hp	SS316	6	$110,000	2013	$110,000	strm.a200.211a	277167	kg/hr	0.80	1.7	425771	1.54	$930,447	$977,847	$1,662,340
Pretreatment																	
In-line Sulfuric Acid Mixer	Kynar Lined - 600 gpm H2O - 5 gpm acid		SS304	1	$6,000	2009	$6,000	strm.a200.214	136260	kg/hr	0.50	1.0	0	0.00	$0	$0	$0
Blowdown Tank Agitator	Side-mounted, 3 x 75 hp. (170 kW)	170 kW	316LSS	1	INCLUDED												
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	315735	1.25	$100,563	$114,879	$172,319
Ammonia Addition Tank Agitator		10 hp	SS	1	$21,900	2009	$21,900	strm.a200.228	410369	kg/hr	0.50	1.5	315835	0.77	$19,213	$21,948	$32,922
Ammonia Static Mixer			SS	1	$5,000	2009	$5,000	strm.a200.275	157478	kg/hr	0.50	1.0	100	0.0006	$126	$144	$144
Pretreatment Water Heater	29.9 MMBtu		304SS	1	$92,000	2010	$92,000	Heat.A200.QH201	-8	Gcal/hr	0.70	2.2	-0.6	0.07	$14,998	$16,234	$35,715
Acid Pretreatment Reactor	3,000 mm x 10,300 mm each - scaled to RT	135 hp ea	Incoloy 825-		$31,200,000	2013	$0	SCIS.a200.211b	63166	kg/hr	0.60	1.5	63091	1.00	$0	$0	$0
Residence Time (min)					SIZING			Residence Time	5	min							
Milling Equipment	200kw/dry ton			9	$2,466,700	2013	$2,466,700	SCIS.a200.211b	62942	kg/hr	0.60	1.5	63091	1.00	$22,200,300	$23,331,251	$34,996,877
Milling Equipment-Szego Mill				13	$578,000	2013	$578,000	SCIS.a200.211b	62942	kg/hr	0.60	1.4	63091	1.00	$7,514,000	$7,896,786	$11,055,501
Sulfuric Acid Pump	9 GPM, 245 FT TDH		316SS	1	$8,000	2009	$8,000	strm.A200.232	3720	kg/hr	0.80	2.3	0	0.00	$0	$0	$0
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	317019	1.08	$27,347	$29,601	$68,083
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	315735	1.54	$42,483	$48,531	$111,620
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	315835	0.79	$18,544	$21,184	$48,723
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	317019	1.55	$42,621	$48,689	$111,984
Blowdown Tank	23' x 48' - 25 min. - 110,000 gal.		SS316	1	INCLUDED												
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	317019	1.20	$580,663	$663,329	$1,326,658
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	315835	0.77	$196,477	$224,448	$448,897
														Area 200 Totals	$38,055,360	$40,091,139	$61,487,720

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
A300: Sugar Hydrolysis and Conditioning	Mechanical Equipment List							Scaled Installed Costs									
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	13	1.58	$117,026	$126,672	$278,678
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	323410	0.77	3188608.277	$3,642,553	$7,285,105
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	323410	0.77	$38,166	$43,600	$100,279
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	323410	0.98	$1,301,661	$1,324,996	$2,384,993
Hydrolysis Tank			304SS	12	$10,128,000	2009	$10,128,000	SSFVES	12	ea	1	1.50	9	0.750	$7,596,000	$8,677,400	$13,016,100
Hydrolysis Agitator		30 hp	SS304	1	$52,500	2009	$52,500	SSFVES	1	ea	1	1.50	9	9.000	$472,500	$539,767	$809,651
Hydrolysis Cooler	Plate & frame		304SS	12	$86,928	2009	$86,928	SSFVES	12	ea	1	2.20	9	0.750	$65,196	$74,478	$163,851
Hydrolysis Recirc/Transfer Pump	340 GPM, 150 FT	20	316SS	5	$47,200	2009	$47,200	SSFVES	12	ea	0.8	2.30	9	0.750	$37,497	$42,835	$98,520
Hydrolysate Conditioning																	
SLS Feed Cooler	Plate & frame		SS304	1	$86,928	2009	$86,928	heat.A500.BDO-1.FERM.QC310	5.404	Gcal/hr	0.7	1.80	5.4	0.993	$86,507	$98,822	$177,880
Lignin Filter (after enzymatic hydrolysis)	(4) 170 m2 Horizontal Belt Filters	660 hp ea	304SS	1	$2,152,500	2013	$2,152,500	Number	10	ea	1	1.70	10	10.00	$21,525,000	$22,621,549	$38,456,634
sizing: solids flow					sizing			SCIS.A500.BDO-1.FERM.306B	20376.24								
sizing: area					sizing			area per unit	170	m^2							
sizing: permeance					sizing			permeance	12	kg/hr/m^2							
Sugar Concentration (MVR evaporation)		3600 kW	304SS	1	$6,370,000	2013	$6,370,000	STRM.A500.BDO-1.FERM.SUG-CONC.301SUG	244084	kg/hr	0.7	2.00	626051.9	2.56	$12,316,552	$12,943,995	$25,887,989
Concentrated Sugar Storage Tank	5,500 gallons - 20 min residence time	20 hp	SS	1	$168,000	2011	$168,000	STRM.A500.BDO-1.FERM.S4	76712	kg/hr	0.7	1.80	97557.9	1.27	$198,787	$202,351	$364,232
Polishing Filter	Ceramic microfiltration		316L	4	$440,000	2014	$440,000	non water flow	53204	kg/hr	0.90	1.800	53415	1.00	$1,766,288	$1,827,913	$3,290,243
liquid flow								STRM.REFINE.01					97555				
water flow								CMIX.H2O.REFINE.01A					44140				
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	97555	1.27	$198,783	$202,347	$364,224
liquid flow								STRM.REFINE.01					97555				
water flow								CMIX.H2O.REFINE.01A					44140				
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	53415	1.00	$5,268,756	$5,452,581	$9,814,647
liquid flow								STRM.REFINE.01					97555				
water flow								CMIX.H2O.REFINE.01A					44140				
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	97555	1.27	$198,783	$202,347	$364,224
liquid flow								STRM.REFINE.01					97555				
water flow								CMIX.H2O.REFINE.01A					44140				
														Area 300 Totals	$54,376,109	$58,024,205	$102,857,250

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
A400:Enzyme Production	Mechanical Equipment List					Scaled Installed Costs											
Cellulase Fermentor Agitators		800.0	SS316		$580,000	2009	$580,000	CLVESSEL	1	ea	1.00	1.5	5	5.00	$2,900,000	$3,312,857	$4,969,285
Cellulase Fermentor Agitators		0.75hp	SS316		$3,420	2009	$3,420	ICLSEED	1	ea	1.00	1.5	4	4.00	$13,680	$15,628	$23,441
Cellulase Fermentor Agitators		8 hp	SS316		$11,000	2009	$11,000	ICLSEED	1	ea	1.00	1.5	4	4.00	$44,000	$50,264	$75,396
Cellulase Fermentor Agitators		80 hp	SS316		$63,000	2009	$63,000	ICLSEED	1	ea	1.00	1.5	4	4.00	$252,000	$287,876	$431,814
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	$4,603	$7,365
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	$22,394	$25,583	$38,374
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	$2,287,585	$4,575,170
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	$210,195	$378,351
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	$262,744	$472,939
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	$435,926	$784,667
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	$275,587	$440,939
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	$5,046	$11,606
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	$22,090	$50,808
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	$601	$1,383
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	$3,234	$7,438
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	$1,273	$2,929
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	$6,390	$19,169
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	$191,919	$219,241	$394,634
														Area 400 Totals	$6,502,810	$7,426,723	$12,685,708

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
A500: Catalytic conversion and Upgrading	Mechanical Equipment List					Scaled Installed Costs											
APR-1 Reactor System	Costed as ISBL package (reactors and supporting equipment)		317L Clad	1	$11,702,128	2007	$11,702,128	REFINE.03	55709	kg/hr	0.70	2.0	69674	1.25	$13,685,708	$15,529,918	$31,059,837
APR-2 Reactor System	Costed as ISBL package (reactors and supporting equipment)		317L Clad	1	$5,851,064	2007	$5,851,064	REFINE.03	55709	kg/hr	0.70	2.0	69674	1.25	$6,842,854	$7,764,959	$15,529,918
Condensation Reactor System	Costed as ISBL package (reactors and supporting equipment)		317L Clad	1	$20,212,766	2007	$20,212,766	REFINE.14	35078	kg/hr	0.70	2.0	45520	1.30	$24,257,222	$27,525,992	$55,051,983
Hydrotreating Facility	Costed as ISBL package (reactors and supporting equipment)			1	$2,925,532	2007	$2,925,532	strm.REFINE.27	22730	kg/hr	0.70	2.0	23404	1.03	$2,985,973	$3,388,347	$6,776,693
														Area 500 Totals	$47,771,758	$54,209,216	$108,418,431

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
A600: WWT	**Mechanical Equipment List**					**Scaled Installed Costs**											
Anaerobic System	(4) 27 MG reactors, 8 recycle pumps, 5 blowers, 8 mixers, 1 centrifuge	795 hp	Concrete	4	$25,800,000	2012	$25,800,000	COD	27211.0	kg/hr	0.60	1.1	19580.7	0.72	$21,177,255	$21,597,468	$23,973,189
Biogas Emergency Flare				4	INCLUDED	2012											
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	1.1	0.41	$2,817,408	$2,873,313	$5,947,757
Pump - Submersible, Anaerobic Feed	2500 gpm ea	15 hp ea	CS	2	INCLUDED	2012											
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	19581	3.50	$42,385	$43,226	$129,679
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	19581	3.50	$4,386,864	$4,473,911	$8,947,822
Surface Aerators		100 hp ea	CS	3	$150,000	2012	$150,000	Hydraulic flow	3	MGD	0.60	2.7	1.1	0.41	$87,955	$89,700	$239,500
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	1.1	0.41	$2,012,237	$2,052,165	$3,365,550
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	19581	3.50	$1,589,444	$1,620,982	$2,577,362
Centrifuge		165 hp ea	CS	1	$686,800	2012	$686,800	COD	5600	kg/hr	0.60	2.7	19581	3.50	$1,455,506	$1,484,388	$3,993,002
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	19581	3.50	$14,835	$15,129	$43,269
Reverse Osmosis			CS	7	$2,450,000	2012	$2,450,000	Hydraulic flow	2.7	MGD	1.00	1.8	1.1	0.41	$1,006,426	$1,026,396	$1,796,194
Evaporator	368 gpm	1480 hp ea	Titanium	1	$5,000,000	2012	$5,000,000	Hydraulic flow	2.7	MGD	0.60	1.6	1.1	0.41	$2,931,835	$2,990,010	$4,813,917
Ammonia Addition System	0.63 gpm	4.5 hp	CS	4	$195,200	2012	$195,200	COD	5600	kg/hr	0.60	1.5	19581	3.50	$413,679	$421,888	$649,707
Sodium Sulfate Puification																	
Evaporator feed tank	insulated, 6460 gal				$45,966	2011	$45,966	strm.A600.23	290932	kg/hr	0.60	2.50	132735	0.46	$28,705	29219.28894	$73,048
Evaporator feed heater	shell and tube 1/2 pass				$274,818	2011	$274,818	heat.A600.31	-13	MMkcal	0.60	3.00	-4	0.27	$124,290	126518.4355	$379,555
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	132735	0.50	$315,685	360627.5224	$721,255
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	12669	1.10	$346,841	$353,058	$812,034
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	12065	1.05	$570,513	$580,741	$1,509,926
														Area 600 Totals	$39,321,864	$40,138,741	$59,972,768

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
A700: Storage	**Mechanical Equipment List**					**Scaled Installed Costs**											
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	793	0.68	$149,167	$161,462	$322,923
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	90	0.06	$10,292	$11,757	$30,568
CSL Storage Tank Agitator		10 hp	SS304	1	$21,200	2009	$21,200	strm.A900.CSL-NET	1393	kg/hr	0.50	1.5	90	0.06	$5,390	$6,158	$9,237
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	90	0.06	$335	$383	$1,188
DAP Bulk Bag Unloader	Super sack unloader			1	$30,000	2009	$30,000	strm.A900.DAP-NET	163	kg/hr	0.60	1.7	0	0.00	$0	$0	$1
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	0	0.00	$0	$0	$0
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	0	0.00	$1	$1	$1
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	0	0.00	$0	$0	$0
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	0	0.00	$0	$0	$0
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	0	0.00	$0	$0	$0
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	2000	1.01	$96,644	$98,376	$147,564
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	$917,319	$1,559,442
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	$17,135	$53,120
Naphtha storage tank	750,000 gal., 7 day storage, Floating roof		A285C	1	$670,000	2009	$670,000	strm.NAPH-F-C	11341	kg/hr	0.70	1.7	3837	0.34	$313,781	$358,452	$609,369
Jet fuel storage tank	750,000 gal., 7 day storage, Floating roof		A285C	1	$670,000	2009	$670,000	strm.JET-F-C	11341	kg/hr	0.70	1.7	13473	1.19	$755,885	$863,496	$1,467,943
Diesel storage tank	750,000 gal., 7 day storage, Floating roof		A285C	1	$670,000	2009	$670,000	strm.DIES-F-C	11341	kg/hr	0.70	1.7	1918	0.17	$193,120	$220,613	$375,042
Co-Product Storage Tank (Sodium Sulfate)				1	$690,900	2007	$690,900	strm.PRD-600	23322.9	kg/hr	0.65	1.850	12065	0.517317692	$450,150	$510,810	$944,998
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	$86,461	$224,798
														Totals:	$2,868,450	$3,252,423	$5,746,194

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
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	0.94	$39,205	$44,787	$98,531
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	212918	0.89	$26,690,991	$28,891,011	$52,003,820
Turbine/Generator	23.6 kW, 2 extractions			1	$9,500,000	2010	$9,500,000	work.A900.wtotal	-42200	kW	0.60	1.8	-45965	1.09	$9,999,840	$10,824,083	$19,483,349
Hot Process Water Softener System				1	$78,000	2010	$78,000	strm.A800.A810.812	235803	kg/hr	0.60	1.8	212903	0.90	$73,363	$79,410	$142,937
Amine Addition Pkg.				1	$40,000	2010	$40,000	strm.A800.A810.812	235803	kg/hr	0.00	1.8	212903	0.90	$40,000	$43,297	$77,935
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	212903	0.90	$286,867	$310,512	$931,535
Blowdown Flash Drum			CS	1	INCLUDED												
Amine Drum			SS316	1	INCLUDED												
														Area 800 Totals	$37,130,266	$40,193,099	$72,738,107

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
A900: Utilities	**Mechanical Equipment List**					**Scaled Installed Costs**											
Utilities System																	
Cooling Tower System	44,200 gpm	750 hp	FIBERGLAS	1	$1,375,000	2010	$1,375,000	strm.a900.945	10037820	kg/hr	0.60	1.5	11554515	1.15	$1,496,133	$1,619,452	$2,429,178
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	$30,308	$48,493
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	5	0.33	$655,752	$709,803	$1,135,684
CIP System	100,000 GAL		SS304/SS3	1	$421,000	2009	$421,000	strm.a900.914	63	kg/hr	0.60	1.8	145	2.30	$694,222	$793,055	$1,427,499
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	11554515	1.05	$295,429	$319,780	$991,319
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	486451	3.13	$17,088	$18,496	$57,338
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	486451	0.94	$14,522	$15,718	$48,727
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	$17,136	$30,844
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	$18,278	$56,661
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	486451	1.08	$263,372	$300,867	$511,474
														Area 900 Totals	$3,495,518	$3,842,893	$6,737,217

DDA Case:

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
A200: Pretreatment	Mechanical Equipment List					Scaled Installed Costs											
Deacetylation																	
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	420318	1.05	$23,308	$26,626	$61,239
Deacetylation reactor	14' x 30' vessel, quad discharge screw live btm	15hp ea	SS316	3	$780,000	2013	$780,000	strm.a200.211a	277167	kg/hr	0.70	1.7	420318	1.52	$3,131,851	$3,291,397	$5,595,375
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	420318	1.52	$460,450	$483,907	$822,642
Pretreatment																	
Blowdown Tank Agitator	Side-mounted, 3 x 75 hp. (170 kW)	170 kW	316LSS	1	INCLUDED												
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	214825	0.85	$82,950	$94,760	$142,139
Ammonia Addition Tank Agitator		10 hp	SS	1	$21,900	2009	$21,900	strm.a200.228	410369	kg/hr	0.50	1.5	326137	0.79	$19,523	$22,303	$33,454
Ammonia Static Mixer			SS	1	$5,000	2009	$5,000	strm.a200.275	157478	kg/hr	0.50	1.0	111312	0.7068	$4,204	$4,802	$4,802
Transfer Conveyor	800 mm x 7600 mm (2'-8" x 25')	60 hp ea	316LSS	2	INCLUDED												
Distribution Conveyor	800 mm x 7600 mm (2'-8" x 25')	60 hp ea	316LSS	2	INCLUDED												
Overfeed Conveyor	800 mm x 7600 mm (2'-8" x 25')	60 hp ea	316LSS	4	INCLUDED												
Pressurized Heating Screw	2500 mm x 9500 mm (8' x 31')	100 hp ea	Dup. 2205	1	INCLUDED												
Pressurized Pre-heater Discharge	850 mm x 3500 mm (2'-10" x 12')	50 hp ea.	Dup. 2205	2	INCLUDED												
Pressurized Transport - No. 1	900 mm x 3500 mm (3' x 12')	25 hp ea.	Incoloy 825-	1	INCLUDED												
Pressurized Transport - No. 2	1200 mm x 3500 mm (4' x 12')	50 hp ea.	Incoloy 825-	1	INCLUDED												
Pretreatment Water Heater	29.9 MMBtu		304SS	1	$92,000	2010	$92,000	Heat.A200.QH201	-8	Gcal/hr	0.70	2.2	-14.9	1.98	$148,489	$160,728	$353,602
Doffing Roll Storage Bins	60 cu. m. (2100 cu.ft.) with conveyors/scrapers	54 hp ea	304/316SS	2	INCLUDED												
Pin Drum Feeder		7.5 hp ea	316LSS	2	INCLUDED												
Plug Screw Feeder		1510 hp ea	316LSS	2	INCLUDED												
Prehydrolysis / Vertical Preheater	16' x 62' - 10 min. residence time		Dup. 2205	1	INCLUDED												
Pin Drum Feeder		7.5 hp ea	Incoloy 825-	2	INCLUDED												
Plug Screw Feeder		1510 hp ea	Incoloy 825-	2	INCLUDED												
Acid Pretreatment Reactor	3,000 mm x 10,300 mm each - scaled to RT	135 hp ea	Incoloy 825-	1	$31,200,000	2013	$31,200,000	SCIS.a200.211b	63166	kg/hr	0.60	1.5	62942	1.00	$31,200,000	$32,789,424	$49,184,135
Residence Time (min)					SIZING			Residence Time	5	min							
Milling Equipment	200kw/dry ton			8	$2,466,700	2013	$0	SCIS.a200.211b	62942	kg/hr	0.60	1.5	62942	1.00	$0	$0	$0
Milling Equipment-Szego Mill				12	$578,000	2013	$0	SCIS.a200.211b	62942	kg/hr	0.60	1.4	62942	1.00	$0	$0	$0
Sulfuric Acid Pump	9 GPM, 245 FT TDH		316SS	1	$8,000	2009	$8,000	strm.A200.232	3720	kg/hr	0.80	2.3	806	0.22	$2,355	$2,690	$6,187
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	243384	0.83	$22,135	$23,959	$55,107
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	214825	1.05	$31,219	$35,664	$82,026
Oligomer Hold Tank Discharge	900 GPM, 150 FT TDH	75	316SS	1	$17,408	2010	$0	strm.a200.223	292407	kg/hr	0.80	2.3	243384	0.83	$0	$0	$0
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	326137	0.81	$19,026	$21,735	$49,990
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	242578	1.19	$34,406	$39,304	$90,400
Blowdown Tank	23' x 48' - 25 min. - 110,000 gal.		SS316	1	INCLUDED												
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	243384	0.92	$482,580	$551,283	$1,102,565
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	326137	0.79	$200,942	$229,549	$459,097
														Area 200 Totals	$35,863,438	$37,778,130	$58,042,762

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
A300: Sugar Hydrolysis and Conditioning	**Mechanical Equipment List**					**Scaled Installed Costs**											
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	0.85	$76,101	$82,373	$181,221
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	333867	0.79	3260435.641	$3,724,606	$7,449,211
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	333867	0.79	$39,150	$44,724	$102,865
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	333867	1.01	$1,330,982	$1,354,843	$2,438,718
Hydrolysis Tank			304SS	12	$10,128,000	2009	$10,128,000	SSFVES	12	ea	1	1.50	9	0.750	$7,596,000	$8,677,400	$13,016,100
Hydrolysis Agitator		30 hp	SS304	1	$52,500	2009	$52,500	SSFVES	1	ea	1	1.50	9	9.000	$472,500	$539,767	$809,651
Hydrolysis Cooler	Plate & frame		304SS	12	$86,928	2009	$86,928	SSFVES	12	ea	1	2.20	9	0.750	$65,196	$74,478	$163,851
Hydrolysis Recirc/Transfer Pump	340 GPM, 150 FT	20	316SS	5	$47,200	2009	$47,200	SSFVES	12	ea	0.8	2.30	9	0.750	$37,497	$42,835	$98,520
Hydrolysate Conditioning																	
SLS Feed Cooler	Plate & frame		SS304	1	$86,928	2009	$86,928	heat.A500.BDO-1.FERM.QC310	5.404	Gcal/hr	0.7	1.80	5.5	1.024	$88,362	$100,942	$181,695
Lignin Filter (after enzymatic hydrolysis)	(4) 170 m2 Horizontal Belt Filters	660 hp ea	304SS	1	$2,152,500	2013	$2,152,500	Number	6	ea	1	1.70	7	7.00	$15,067,500	$15,835,085	$26,919,644
sizing: solids flow					sizing			SCIS.A500.BDO-1.FERM.306B	21633.41								
sizing: area					sizing			area per unit	170	m^2							
sizing: permeance					sizing			permeance	20	kg/hr/m^2							
Sugar Concentration (MVR evaporation)		3600 kW	304SS	1	$6,370,000	2013	$6,370,000	STRM.A500.BDO-1.FERM.SUG-CONC.301SU	244084	kg/hr	0.7	2.00	493697.4	2.02	$10,430,005	$10,961,341	$21,922,683
Concentrated Sugar Storage Tank	5,500 gallons - 20 min residence time	20 hp	SS	1	$168,000	2011	$168,000	STRM.A500.BDO-1.FERM.S4	76712	kg/hr	0.7	1.80	95876.5	1.25	$196,383	$199,903	$359,826
Polishing Filter	Ceramic microfiltration		316L	4	$440,000	2014	$440,000	non water flow	53204	kg/hr	0.90	1.800	52033	0.98	$1,725,101	$1,785,289	$3,213,520
liquid flow								STRM.REFINE.01					95873				
water flow								CMIX.H2O.REFINE.01A					43840				
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	95873	1.25	$196,378	$199,898	$359,817
liquid flow								STRM.REFINE.01					95873				
water flow								CMIX.H2O.REFINE.01A					43840				
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	52033	0.98	$5,145,897	$5,325,436	$9,585,785
liquid flow								STRM.REFINE.01					95873				
water flow								CMIX.H2O.REFINE.01A					43840				
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	95873	1.25	$196,378	$199,898	$359,817
liquid flow								STRM.REFINE.01					95873				
water flow								CMIX.H2O.REFINE.01A					43840				
														Area 300 Totals	$45,923,865	$49,148,819	$87,162,925

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
A400:Enzyme Production	**Mechanical Equipment List**							**Scaled Installed Costs**									
Cellulase Fermentor Agitators		800.0	SS316		$580,000	2009	$580,000	CLVESSEL	1	ea	1.00	1.5	5	5.00	$2,900,000	$3,312,857	$4,969,285
Cellulase Fermentor Agitators		0.75hp	SS316		$3,420	2009	$3,420	ICLSEED	1	ea	1.00	1.5	4	4.00	$13,680	$15,628	$23,441
Cellulase Fermentor Agitators		8 hp	SS316		$11,000	2009	$11,000	ICLSEED	1	ea	1.00	1.5	4	4.00	$44,000	$50,264	$75,396
Cellulase Fermentor Agitators		80 hp	SS316		$63,000	2009	$63,000	ICLSEED	1	ea	1.00	1.5	4	4.00	$252,000	$287,876	$431,814
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	125	0.72	$4,071	$4,650	$7,440
Cellulase Hold Tank Agitator		10 hp	SS316	1	$26,900	2009	$26,900	strm.422	10930	kg/hr	0.50	1.5	7730	0.71	$22,622	$25,842	$38,764
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	$2,287,585	$4,575,170
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	$210,195	$378,351
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	$262,744	$472,939
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	$435,926	$784,667
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	18203	0.55	$244,184	$278,948	$446,316
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	7730	0.58	$4,738	$5,128	$11,795
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	430	0.63	$20,741	$22,450	$51,635
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	125	0.28	$535	$611	$1,406
Cellulase Feed Pump	Gear Pump	1	316SS	1	$5,700	2009	$5,700	strm.a400.422	18168	kg/hr	0.80	2.3	7730	0.43	$2,877	$3,287	$7,560
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	7.5	0.70	$1,133	$1,294	$2,976
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	125	0.56	$5,987	$6,481	$19,442
Cellulase Hold Tank	80,000 gal		304SS	1	$248,070	2009	$248,070	strm.a400.422	10930	kg/hr	0.70	1.8	7730	0.71	$194,652	$222,364	$400,254
														Area 400 Totals	$6,509,319	$7,434,129	$12,698,652

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
A500: Catalytic conversion and Upgrading	**Mechanical Equipment List**							**Scaled Installed Costs**									
APR-1 Reactor System	Costed as ISBL package (reactors and supporting equipment)		317L Clad	1	$11,702,128	2007	$11,702,128	REFINE.03	55709	kg/hr	0.70	2.0	65240	1.17	$13,070,152	$14,831,414	$29,662,827
APR-2 Reactor System	Costed as ISBL package (reactors and supporting equipment)		317L Clad	1	$5,851,064	2007	$5,851,064	REFINE.03	55709	kg/hr	0.70	2.0	65240	1.17	$6,535,076	$7,415,707	$14,831,414
Condensation Reactor System	Costed as ISBL package (reactors and supporting equipment)		317L Clad	1	$20,212,766	2007	$20,212,766	REFINE.14	35078	kg/hr	0.70	2.0	43707	1.25	$23,576,734	$26,753,805	$53,507,609
Hydrotreating Facility	Costed as ISBL package (reactors and supporting equipment)			1	$2,925,532	2007	$2,925,532	strm.REFINE.27	22730	kg/hr	0.70	2.0	22034	0.97	$2,862,551	$3,248,292	$6,496,584
														Area 500 Totals	$46,044,513	$52,249,217	$104,498,434

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
A600: WWT	Mechanical Equipment List					Scaled Installed Costs											
Anaerobic System	(4) 27 MG reactors, 8 recycle pumps, 5 blowers, 8 mixers, 1 centrifuge	795 hp	Concrete	4	$25,800,000	2012	$25,800,000	COD	27211.0	kg/hr	0.60	1.1	17591.0	0.65	$19,858,536	$20,252,582	$22,480,365
Biogas Emergency Flare				4	INCLUDED	2012											
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	2.1	0.79	$4,173,751	$4,256,570	$8,811,099
Pump - Submersible, Anaerobic Feed	2500 gpm ea	15 hp ea	CS	2	INCLUDED	2012											
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	17591	3.14	$39,746	$40,534	$121,603
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	17591	3.14	$4,113,692	$4,195,318	$8,390,637
Surface Aerators		100 hp ea	CS	3	$150,000	2012	$150,000	Hydraulic flow	3	MGD	0.60	2.7	2.1	0.79	$130,298	$132,883	$354,799
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	2.1	0.79	$3,873,829	$3,950,696	$6,479,141
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	17591	3.14	$1,490,468	$1,520,043	$2,416,868
Centrifuge		165 hp ea	CS	1	$686,800	2012	$686,800	COD	5600	kg/hr	0.60	2.7	17591	3.14	$1,364,871	$1,391,954	$3,744,356
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	17591	3.14	$13,911	$14,187	$40,575
Reverse Osmosis			CS	7	$2,450,000	2012	$2,450,000	Hydraulic flow	2.7	MGD	1.00	1.8	2.1	0.79	$1,937,508	$1,975,953	$3,457,917
Evaporator	368 gpm	1480 hp ea	Titanium	1	$5,000,000	2012	$5,000,000	Hydraulic flow	2.7	MGD	0.60	1.6	2.1	0.79	$4,343,266	$4,429,447	$7,131,410
Ammonia Addition System	0.63 gpm	4.5 hp	CS	4	$195,200	2012	$195,200	COD	5600	kg/hr	0.60	1.5	17591	3.14	$387,919	$395,616	$609,249
Sodium Sulfate Puification																	
Evaporator feed tank	insulated, 6460 gal				$45,966	2011	$45,966	strm.A600.23	290932	kg/hr	0.60	2.50	2024680	6.96	$147,223	149862.3002	$374,656
Evaporator feed heater	shell and tube 1/2 pass				$274,818	2011	$274,818	heat.A600.31	-13	MMkcal	0.60	3.00	-2	0.17	$94,554	96248.99118	$288,747
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	2024680	7.67	$2,126,249	2428950.949	$4,857,902
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	2914	0.25	$143,607	$146,182	$336,218
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	2775	0.24	$236,217	$240,452	$625,175
														Area 600 Totals	$44,475,645	$45,617,480	$70,520,719

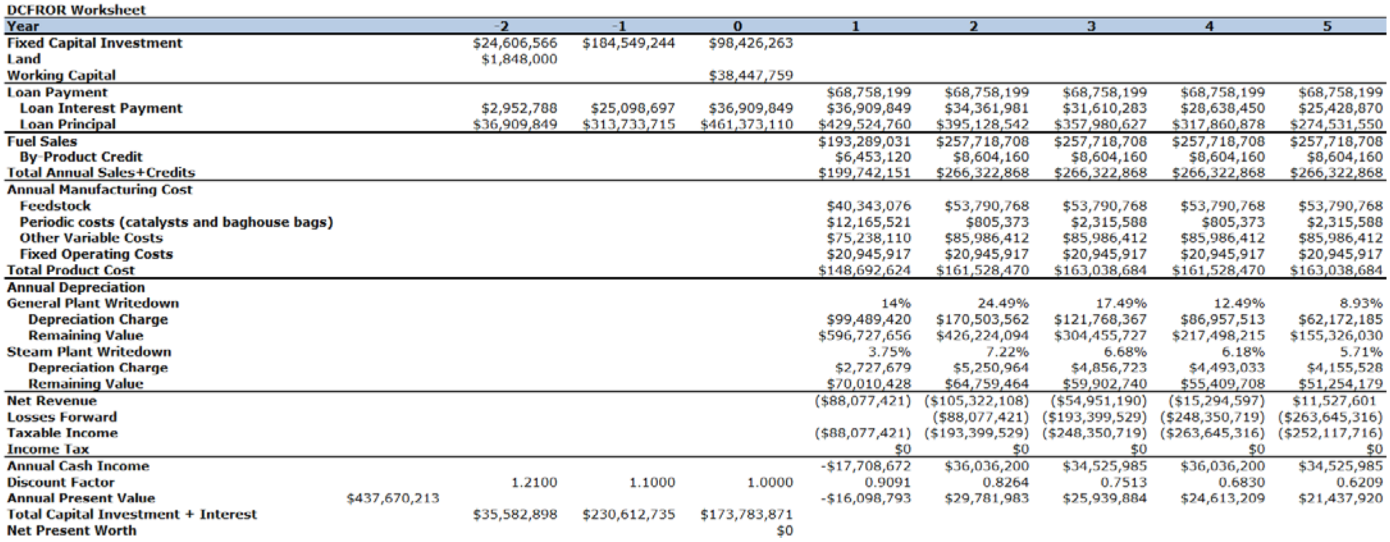
EQUIPMENT TITLE	DESCRIPTION	HP	MATERIAL	NUM REQD	$	Year of Quote	Purch Cost in Base Yr	Scaling Variable	Scaling Val	Units	Scaling Exp	Inst Factor	New Val	Size Ratio	Scaled Purch Cost	Purch Cost in Proj year	Inst Cost in Proj year
A700: Storage	Mechanical Equipment List					Scaled Installed Costs											
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	1065	0.91	$183,419	$198,537	$397,075
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	92	0.07	$10,438	$11,925	$31,004
CSL Storage Tank Agitator		10 hp	SS304	1	$21,200	2009	$21,200	strm.A900.CSL-NET	1393	kg/hr	0.50	1.5	92	0.07	$5,445	$6,220	$9,330
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	92	0.07	$341	$389	$1,207
DAP Bulk Bag Unloader	Super sack unloader			1	$30,000	2009	$30,000	strm.A900.DAP-NET	163	kg/hr	0.60	1.7	0	0.00	$0	$0	$1
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	0	0.00	$0	$0	$0
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	0	0.00	$1	$1	$1
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	0	0.00	$0	$0	$0
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	807	0.41	$3,651	$3,952	$9,090
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	807	0.41	$51,178	$55,397	$83,095
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	3043	1.54	$129,645	$131,969	$197,953
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	$917,319	$1,559,442
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	$17,135	$53,120
Naphtha storage tank	750,000 gal., 7 day storage, Floating roof		A285C	1	$670,000	2009	$670,000	strm.NAPH-F-C	11341	kg/hr	0.70	1.7	3636	0.32	$302,182	$345,202	$586,844
Jet fuel storage tank	750,000 gal., 7 day storage, Floating roof		A285C	1	$670,000	2009	$670,000	strm.JET-F-C	11341	kg/hr	0.70	1.7	12781	1.13	$728,497	$832,209	$1,414,755
Diesel storage tank	750,000 gal., 7 day storage, Floating roof		A285C	1	$670,000	2009	$670,000	strm.DIES-F-C	11341	kg/hr	0.70	1.7	1818	0.16	$186,015	$212,497	$361,245
Co-Product Storage Tank (Sodium Sulfate)				1	$690,900	2007	$690,900	strm.PRD-600	23322.9	kg/hr	0.65	1.850	2775	0.118987426	$173,177	$196,514	$363,550
Glucose Storage Tank	70,000 gal		Glass lined	1	$70,000	2009	$70,000	strm.a400.401	1393	kg/hr	0.70	2.6	1589	1.14	$76,764	$87,692	$228,000
														Totals:	$2,668,754	$3,016,959	$5,295,712

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	-1	0.26	$15,767	$18,011	$39,625
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	211469	0.89	$26,581,907	$28,772,935	$51,791,283
Turbine/Generator	23.6 kW, 2 extractions			1	$9,500,000	2010	$9,500,000	work.A900.wtotal	-42200	kW	0.60	1.8	-51006	1.21	$10,644,109	$11,521,456	$20,738,621
Hot Process Water Softener System				1	$78,000	2010	$78,000	strm.A800.A810.812	235803	kg/hr	0.60	1.8	211448	0.90	$73,061	$79,083	$142,350
Amine Addition Pkg.				1	$40,000	2010	$40,000	strm.A800.A810.812	235803	kg/hr	0.00	1.8	211448	0.90	$40,000	$43,297	$77,935
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	211448	0.90	$285,688	$309,236	$927,708
Blowdown Flash Drum			CS	1	INCLUDED												
Amine Drum			SS316	1	INCLUDED												
														Area 800 Totals	$37,640,532	$40,744,019	$73,717,521

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
A900: Utilities	Mechanical Equipment List					Scaled Installed Costs											
Utilities System																	
Cooling Tower System	44,200 gpm	750 hp	FIBERGLAS	1	$1,375,000	2010	$1,375,000	strm.a900.945	10037820	kg/hr	0.60	1.5	12491185	1.24	$1,567,766	$1,696,990	$2,545,485
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	$30,308	$48,493
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	5	0.34	$663,745	$718,455	$1,149,527
CIP System	100,000 GAL		SS304/SS3	1	$421,000	2009	$421,000	strm.a900.914	63	kg/hr	0.60	1.8	145	2.30	$694,222	$793,055	$1,427,499
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	12491185	1.14	$314,438	$340,356	$1,055,103
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	644232	4.14	$21,394	$23,157	$71,787
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	644232	1.24	$18,181	$19,679	$61,006
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	$17,136	$30,844
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	$18,278	$56,661
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	644232	1.43	$320,606	$366,249	$622,624
														Area 900 Totals	$3,659,352	$4,023,662	$7,069,029

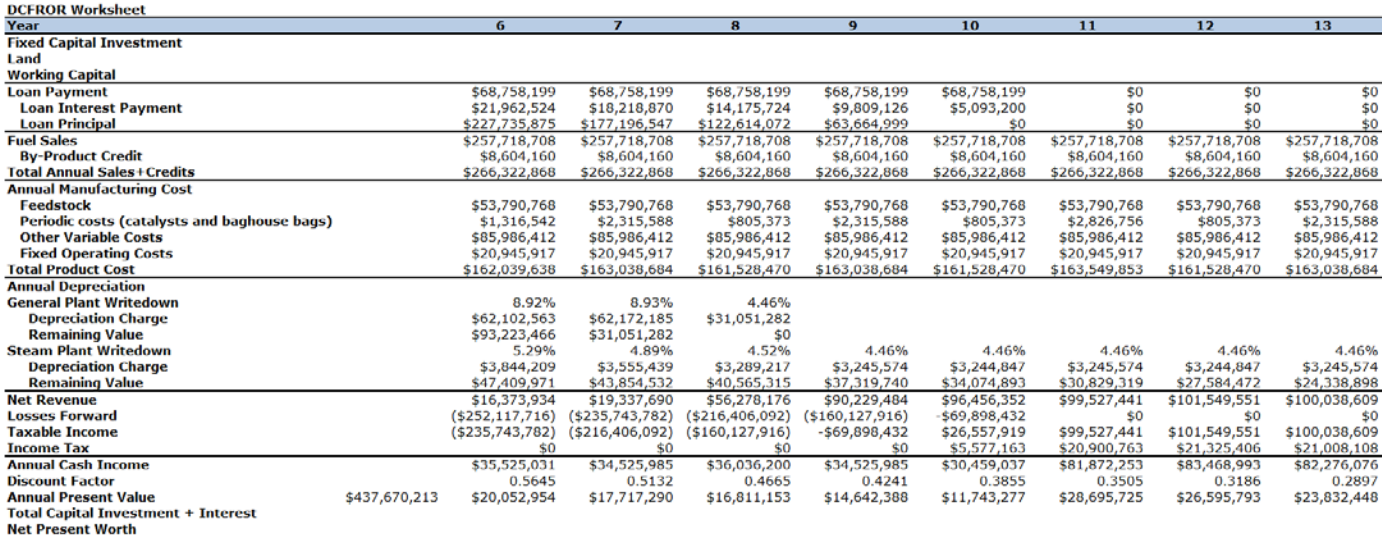
Appendix B. Discounted Cash Flow Rate of Return Worksheet

DMR Case:



DCFROR Worksheet

Year		**-2**	**-1**	**0**	**1**	**2**	**3**	**4**	**5**
Fixed Capital Investment		$24,606,566	$184,549,244	$98,426,263					
Land		$1,848,000							
Working Capital				$38,447,759					
Loan Payment					$68,758,199	$68,758,199	$68,758,199	$68,758,199	$68,758,199
Loan Interest Payment		$2,952,788	$25,098,697	$36,909,849	$36,909,849	$34,361,981	$31,610,283	$28,638,450	$25,428,870
Loan Principal		$36,909,849	$313,733,715	$461,373,110	$429,524,760	$395,128,542	$357,980,627	$317,860,878	$274,531,550
Fuel Sales					$193,289,031	$257,718,708	$257,718,708	$257,718,708	$257,718,708
By-Product Credit					$6,453,120	$8,604,160	$8,604,160	$8,604,160	$8,604,160
Total Annual Sales+Credits					$199,742,151	$266,322,868	$266,322,868	$266,322,868	$266,322,868
Annual Manufacturing Cost									
Feedstock					$40,343,076	$53,790,768	$53,790,768	$53,790,768	$53,790,768
Periodic costs (catalysts and baghouse bags)					$12,165,521	$805,373	$2,315,588	$805,373	$2,315,588
Other Variable Costs					$75,238,110	$85,986,412	$85,986,412	$85,986,412	$85,986,412
Fixed Operating Costs					$20,945,917	$20,945,917	$20,945,917	$20,945,917	$20,945,917
Total Product Cost					$148,692,624	$161,528,470	$163,038,684	$161,528,470	$163,038,684
Annual Depreciation									
General Plant Writedown					14%	24.49%	17.49%	12.49%	8.93%
Depreciation Charge					$99,489,420	$170,503,562	$121,768,367	$86,957,513	$62,172,185
Remaining Value					$596,727,656	$426,224,094	$304,455,727	$217,498,215	$155,326,030
Steam Plant Writedown					3.75%	7.22%	6.68%	6.18%	5.71%
Depreciation Charge					$2,727,679	$5,250,964	$4,856,723	$4,493,033	$4,155,528
Remaining Value					$70,010,428	$64,759,464	$59,902,740	$55,409,708	$51,254,179
Net Revenue					($88,077,421)	($105,322,108)	($54,951,190)	($15,294,597)	$11,527,601
Losses Forward						($88,077,421)	($193,399,529)	($248,350,719)	($263,645,316)
Taxable Income					($88,077,421)	($193,399,529)	($248,350,719)	($263,645,316)	($252,117,716)
Income Tax					$0	$0	$0	$0	$0
Annual Cash Income					-$17,708,672	$36,036,200	$34,525,985	$36,036,200	$34,525,985
Discount Factor		1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830	0.6209
Annual Present Value	$437,670,213				-$16,098,793	$29,781,983	$25,939,884	$24,613,209	$21,437,920
Total Capital Investment + Interest		$35,582,898	$230,612,735	$173,783,871					
Net Present Worth				$0					



DCFROR Worksheet

Year		**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**
Fixed Capital Investment									
Land									
Working Capital									
Loan Payment		$68,758,199	$68,758,199	$68,758,199	$68,758,199	$68,758,199	$0	$0	$0
Loan Interest Payment		$21,962,524	$18,218,870	$14,175,724	$9,809,126	$5,093,200	$0	$0	$0
Loan Principal		$227,735,875	$177,196,547	$122,614,072	$63,664,999	$0	$0	$0	$0
Fuel Sales		$257,718,708	$257,718,708	$257,718,708	$257,718,708	$257,718,708	$257,718,708	$257,718,708	$257,718,708
By-Product Credit		$8,604,160	$8,604,160	$8,604,160	$8,604,160	$8,604,160	$8,604,160	$8,604,160	$8,604,160
Total Annual Sales+Credits		$266,322,868	$266,322,868	$266,322,868	$266,322,868	$266,322,868	$266,322,868	$266,322,868	$266,322,868
Annual Manufacturing Cost									
Feedstock		$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768
Periodic costs (catalysts and baghouse bags)		$1,316,542	$2,315,588	$805,373	$2,315,588	$805,373	$2,826,756	$805,373	$2,315,588
Other Variable Costs		$85,986,412	$85,986,412	$85,986,412	$85,986,412	$85,986,412	$85,986,412	$85,986,412	$85,986,412
Fixed Operating Costs		$20,945,917	$20,945,917	$20,945,917	$20,945,917	$20,945,917	$20,945,917	$20,945,917	$20,945,917
Total Product Cost		$162,039,638	$163,038,684	$161,528,470	$163,038,684	$161,528,470	$163,549,853	$161,528,470	$163,038,684
Annual Depreciation									
General Plant Writedown		8.92%	8.93%	4.46%					
Depreciation Charge		$62,102,563	$62,172,185	$31,051,282					
Remaining Value		$93,223,466	$31,051,282	$0					
Steam Plant Writedown		5.29%	4.89%	4.52%	4.46%	4.46%	4.46%	4.46%	4.46%
Depreciation Charge		$3,844,209	$3,555,439	$3,289,217	$3,245,574	$3,244,847	$3,245,574	$3,244,847	$3,245,574
Remaining Value		$47,409,971	$43,854,532	$40,565,315	$37,319,740	$34,074,893	$30,829,319	$27,584,472	$24,338,898
Net Revenue		$16,373,934	$19,337,690	$56,278,176	$90,229,484	$96,456,352	$99,527,441	$101,549,551	$100,038,609
Losses Forward		($252,117,716)	($235,743,782)	($216,406,092)	($160,127,916)	-$69,898,432	$0	$0	$0
Taxable Income		($235,743,782)	($216,406,092)	($160,127,916)	-$69,898,432	$26,557,919	$99,527,441	$101,549,551	$100,038,609
Income Tax		$0	$0	$0	$0	$5,577,163	$20,900,763	$21,325,406	$21,008,108
Annual Cash Income		$35,525,031	$34,525,985	$36,036,200	$34,525,985	$30,459,037	$81,872,253	$83,468,993	$82,276,076
Discount Factor		0.5645	0.5132	0.4665	0.4241	0.3855	0.3505	0.3186	0.2897
Annual Present Value	$437,670,213	$20,052,954	$17,717,290	$16,811,153	$14,642,388	$11,743,277	$28,695,725	$26,595,793	$23,832,448
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		$257,718,708	$257,718,708	$257,718,708	$257,718,708	$257,718,708	$257,718,708	$257,718,708	$257,718,708
By-Product Credit		$8,604,160	$8,604,160	$8,604,160	$8,604,160	$8,604,160	$8,604,160	$8,604,160	$8,604,160
Total Annual Sales+Credits		$266,322,868	$266,322,868	$266,322,868	$266,322,868	$266,322,868	$266,322,868	$266,322,868	$266,322,868
Annual Manufacturing Cost									
Feedstock		$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768
Periodic costs (catalysts and baghouse bags)		$805,373	$2,315,588	$1,316,542	$2,315,588	$805,373	$2,315,588	$805,373	$2,826,756
Other Variable Costs		$85,986,412	$85,986,412	$85,986,412	$85,986,412	$85,986,412	$85,986,412	$85,986,412	$85,986,412
Fixed Operating Costs		$20,945,917	$20,945,917	$20,945,917	$20,945,917	$20,945,917	$20,945,917	$20,945,917	$20,945,917
Total Product Cost		$161,528,470	$163,038,684	$162,039,638	$163,038,684	$161,528,470	$163,038,684	$161,528,470	$163,549,853
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		$3,244,847	$3,245,574	$3,244,847	$3,245,574	$3,244,847	$3,245,574	$3,244,847	$1,622,787
Remaining Value		$21,094,051	$17,848,477	$14,603,630	$11,358,055	$8,113,208	$4,867,634	$1,622,787	$0
Net Revenue		$101,549,551	$100,038,609	$101,038,383	$100,038,609	$101,549,551	$100,038,609	$101,549,551	$101,150,228
Losses Forward		$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income		$101,549,551	$100,038,609	$101,038,383	$100,038,609	$101,549,551	$100,038,609	$101,549,551	$101,150,228
Income Tax		$21,325,406	$21,008,108	$21,218,060	$21,008,108	$21,325,406	$21,008,108	$21,325,406	$21,241,548
Annual Cash Income		$83,468,993	$82,276,076	$83,065,169	$82,276,076	$83,468,993	$82,276,076	$83,468,993	$81,531,467
Discount Factor		0.2633	0.2394	0.2176	0.1978	0.1799	0.1635	0.1486	0.1351
Annual Present Value	$437,670,213	$21,979,995	$19,696,238	$18,077,401	$16,277,883	$15,012,632	$13,452,796	$12,407,134	$11,017,394
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										($38,447,759)
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		$257,718,708	$257,718,708	$257,718,708	$257,718,708	$257,718,708	$257,718,708	$257,718,708	$257,718,708	$257,718,708
By-Product Credit		$8,604,160	$8,604,160	$8,604,160	$8,604,160	$8,604,160	$8,604,160	$8,604,160	$8,604,160	$8,604,160
Total Annual Sales+Credits		$266,322,868	$266,322,868	$266,322,868	$266,322,868	$266,322,868	$266,322,868	$266,322,868	$266,322,868	$266,322,868
Annual Manufacturing Cost										
Feedstock		$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768
Periodic costs (catalysts and baghouse bags)		$805,373	$2,315,588	$805,373	$2,315,588	$1,316,542	$2,315,588	$805,373	$2,315,588	$805,373
Other Variable Costs		$85,986,412	$85,986,412	$85,986,412	$85,986,412	$85,986,412	$85,986,412	$85,986,412	$85,986,412	$85,986,412
Fixed Operating Costs		$20,945,917	$20,945,917	$20,945,917	$20,945,917	$20,945,917	$20,945,917	$20,945,917	$20,945,917	$20,945,917
Total Product Cost		$161,528,470	$163,038,684	$161,528,470	$163,038,684	$162,039,638	$163,038,684	$161,528,470	$163,038,684	$161,528,470
Annual Depreciation										
General Plant Writedown										
Depreciation Charge										
Remaining Value										
Steam Plant Writedown										
Depreciation Charge										
Remaining Value										
Net Revenue		$104,794,398	$103,284,184	$104,794,398	$103,284,184	$104,283,230	$103,284,184	$104,794,398	$103,284,184	$104,794,398
Losses Forward		$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income		$104,794,398	$103,284,184	$104,794,398	$103,284,184	$104,283,230	$103,284,184	$104,794,398	$103,284,184	$104,794,398
Income Tax		$22,006,824	$21,689,679	$22,006,824	$21,689,679	$21,899,478	$21,689,679	$22,006,824	$21,689,679	$22,006,824
Annual Cash Income		$82,787,575	$81,594,505	$82,787,575	$81,594,505	$82,383,752	$81,594,505	$82,787,575	$81,594,505	$82,787,575
Discount Factor		0.1228	0.1117	0.1015	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
Annual Present Value	$437,670,213	$10,170,120	$9,112,324	$8,405,058	$7,530,846	$6,912,446	$6,223,840	$5,740,768	$5,143,669	$4,744,436
Total Capital Investment + Interest										($2,309,292)
Net Present Worth										

DDA Case:

DCFROR Worksheet

Year		**-2**	**-1**	**0**	**1**	**2**	**3**	**4**	**5**
Fixed Capital Investment		$23,804,223	$178,531,676	$95,216,894					
Land		$1,848,000							
Working Capital				$37,194,099					
Loan Payment					$66,516,211	$66,516,211	$66,516,211	$66,516,211	$66,516,211
Loan Interest Payment		$2,856,507	$24,280,308	$35,706,335	$35,706,335	$33,241,545	$30,579,572	$27,704,641	$24,599,715
Loan Principal		$35,706,335	$303,503,849	$446,329,189	$415,519,314	$382,244,648	$346,308,009	$307,496,439	$265,579,943
Fuel Sales					$183,261,337	$244,348,449	$244,348,449	$244,348,449	$244,348,449
By-Product Credit					$2,123,535	$2,831,380	$2,831,380	$2,831,380	$2,831,380
Total Annual Sales+Credits					$185,384,872	$247,179,829	$247,179,829	$247,179,829	$247,179,829
Annual Manufacturing Cost									
Feedstock					$40,343,076	$53,790,768	$53,790,768	$53,790,768	$53,790,768
Periodic costs (catalysts and baghouse bags)					$11,470,966	$754,127	$2,172,948	$754,127	$2,172,948
Other Variable Costs					$62,450,563	$71,372,072	$71,372,072	$71,372,072	$71,372,072
Fixed Operating Costs					$20,079,014	$20,079,014	$20,079,014	$20,079,014	$20,079,014
Total Product Cost					$134,343,620	$145,995,982	$147,414,803	$145,995,982	$147,414,803
Annual Depreciation									
General Plant Writedown					14%	24.49%	17.49%	12.49%	8.93%
Depreciation Charge					$95,766,501	$164,123,276	$117,211,764	$83,703,541	$59,845,686
Remaining Value					$574,397,959	$410,274,683	$293,062,919	$209,359,378	$149,513,691
Steam Plant Writedown					3.75%	7.22%	6.68%	6.18%	5.71%
Depreciation Charge					$2,764,407	$5,321,668	$4,922,119	$4,553,531	$4,211,482
Remaining Value					$70,953,114	$65,631,446	$60,709,327	$56,155,796	$51,944,314
Net Revenue					($83,195,991)	($101,502,642)	($52,948,428)	($14,777,866)	$11,108,143
Losses Forward						($83,195,991)	($184,698,633)	($237,647,062)	($252,424,927)
Taxable Income					($83,195,991)	($184,698,633)	($237,647,062)	($252,424,927)	($241,316,784)
Income Tax					$0	$0	$0	$0	$0
Annual Cash Income					-$15,474,958	$34,667,637	$33,248,816	$34,667,637	$33,248,816
Discount Factor		1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830	0.6209
Annual Present Value	$423,468,627				-$14,068,144	$28,650,939	$24,980,327	$23,678,462	$20,644,899
Total Capital Investment + Interest		$34,495,564	$223,093,182	$168,117,328					
Net Present Worth				$0					

DCFROR Worksheet

Year		**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**
Fixed Capital Investment									
Land									
Working Capital									
Loan Payment		$66,516,211	$66,516,211	$66,516,211	$66,516,211	$66,516,211	$0	$0	$0
Loan Interest Payment		$21,246,395	$17,624,810	$13,713,498	$9,489,281	$4,927,127	$0	$0	$0
Loan Principal		$220,310,127	$171,418,727	$118,616,014	$61,589,084	$0	$0	$0	$0
Fuel Sales		$244,348,449	$244,348,449	$244,348,449	$244,348,449	$244,348,449	$244,348,449	$244,348,449	$244,348,449
By-Product Credit		$2,831,380	$2,831,380	$2,831,380	$2,831,380	$2,831,380	$2,831,380	$2,831,380	$2,831,380
Total Annual Sales+Credits		$247,179,829	$247,179,829	$247,179,829	$247,179,829	$247,179,829	$247,179,829	$247,179,829	$247,179,829
Annual Manufacturing Cost									
Feedstock		$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768
Periodic costs (catalysts and baghouse bags)		$1,265,296	$2,172,948	$754,127	$2,172,948	$754,127	$2,684,117	$754,127	$2,172,948
Other Variable Costs		$71,372,072	$71,372,072	$71,372,072	$71,372,072	$71,372,072	$71,372,072	$71,372,072	$71,372,072
Fixed Operating Costs		$20,079,014	$20,079,014	$20,079,014	$20,079,014	$20,079,014	$20,079,014	$20,079,014	$20,079,014
Total Product Cost		$146,507,150	$147,414,803	$145,995,982	$147,414,803	$145,995,982	$147,925,971	$145,995,982	$147,414,803
Annual Depreciation									
General Plant Writedown		8.92%	8.93%	4.46%					
Depreciation Charge		$59,778,670	$59,845,686	$29,889,335					
Remaining Value		$89,735,021	$29,889,335	$0					
Steam Plant Writedown		5.29%	4.89%	4.52%	4.46%	4.46%	4.46%	4.46%	4.46%
Depreciation Charge		$3,895,971	$3,603,312	$3,333,506	$3,289,276	$3,288,539	$3,289,276	$3,288,539	$3,289,276
Remaining Value		$48,048,343	$44,445,031	$41,111,524	$37,822,249	$34,533,710	$31,244,434	$27,955,896	$24,666,620
Net Revenue		$15,751,643	$18,691,217	$54,247,508	$86,986,470	$92,968,182	$95,964,582	$97,895,309	$96,475,751
Losses Forward		($241,316,784)	($225,565,142)	($206,873,924)	($152,626,416)	-$65,639,946	$0	$0	$0
Taxable Income		($225,565,142)	($206,873,924)	($152,626,416)	-$65,639,946	$27,328,236	$95,964,582	$97,895,309	$96,475,751
Income Tax		$0	$0	$0	$0	$5,738,929	$20,152,562	$20,558,015	$20,259,908
Annual Cash Income		$34,156,468	$33,248,816	$34,667,637	$33,248,816	$28,928,707	$79,101,296	$80,625,833	$79,505,119
Discount Factor		0.5645	0.5132	0.4665	0.4241	0.3855	0.3505	0.3186	0.2897
Annual Present Value	$423,468,627	$19,280,436	$17,061,900	$16,172,708	$14,100,743	$11,153,269	$27,724,522	$25,689,875	$23,029,801
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		$244,348,449	$244,348,449	$244,348,449	$244,348,449	$244,348,449	$244,348,449	$244,348,449	$244,348,449
By-Product Credit		$2,831,380	$2,831,380	$2,831,380	$2,831,380	$2,831,380	$2,831,380	$2,831,380	$2,831,380
Total Annual Sales+Credits		$247,179,829	$247,179,829	$247,179,829	$247,179,829	$247,179,829	$247,179,829	$247,179,829	$247,179,829
Annual Manufacturing Cost									
Feedstock		$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768
Periodic costs (catalysts and baghouse bags)		$754,127	$2,172,948	$1,265,296	$2,172,948	$754,127	$2,172,948	$754,127	$2,684,117
Other Variable Costs		$71,372,072	$71,372,072	$71,372,072	$71,372,072	$71,372,072	$71,372,072	$71,372,072	$71,372,072
Fixed Operating Costs		$20,079,014	$20,079,014	$20,079,014	$20,079,014	$20,079,014	$20,079,014	$20,079,014	$20,079,014
Total Product Cost		$145,995,982	$147,414,803	$146,507,150	$147,414,803	$145,995,982	$147,414,803	$145,995,982	$147,925,971
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		$3,288,539	$3,289,276	$3,288,539	$3,289,276	$3,288,539	$3,289,276	$3,288,539	$1,644,638
Remaining Value		$21,378,081	$18,088,805	$14,800,267	$11,510,991	$8,222,452	$4,933,177	$1,644,638	$0
Net Revenue		$97,895,309	$96,475,751	$97,384,140	$96,475,751	$97,895,309	$96,475,751	$97,895,309	$97,609,220
Losses Forward		$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income		$97,895,309	$96,475,751	$97,384,140	$96,475,751	$97,895,309	$96,475,751	$97,895,309	$97,609,220
Income Tax		$20,558,015	$20,259,908	$20,450,669	$20,259,908	$20,558,015	$20,259,908	$20,558,015	$20,497,936
Annual Cash Income		$80,625,833	$79,505,119	$80,222,009	$79,505,119	$80,625,833	$79,505,119	$80,625,833	$78,755,922
Discount Factor		0.2633	0.2394	0.2176	0.1978	0.1799	0.1635	0.1486	0.1351
Annual Present Value	$423,468,627	$21,231,302	$19,032,893	$17,458,647	$15,729,664	$14,501,265	$12,999,722	$11,984,516	$10,642,333
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										($37,194,099)
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		$244,348,449	$244,348,449	$244,348,449	$244,348,449	$244,348,449	$244,348,449	$244,348,449	$244,348,449	$244,348,449
By-Product Credit		$2,831,380	$2,831,380	$2,831,380	$2,831,380	$2,831,380	$2,831,380	$2,831,380	$2,831,380	$2,831,380
Total Annual Sales+Credits		$247,179,829	$247,179,829	$247,179,829	$247,179,829	$247,179,829	$247,179,829	$247,179,829	$247,179,829	$247,179,829
Annual Manufacturing Cost										
Feedstock		$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768	$53,790,768
Periodic costs (catalysts and baghouse bags)		$754,127	$2,172,948	$754,127	$2,172,948	$1,265,296	$2,172,948	$754,127	$2,172,948	$754,127
Other Variable Costs		$71,372,072	$71,372,072	$71,372,072	$71,372,072	$71,372,072	$71,372,072	$71,372,072	$71,372,072	$71,372,072
Fixed Operating Costs		$20,079,014	$20,079,014	$20,079,014	$20,079,014	$20,079,014	$20,079,014	$20,079,014	$20,079,014	$20,079,014
Total Product Cost		$145,995,982	$147,414,803	$145,995,982	$147,414,803	$146,507,150	$147,414,803	$145,995,982	$147,414,803	$145,995,982
Annual Depreciation										
General Plant Writedown										
Depreciation Charge										
Remaining Value										
Steam Plant Writedown										
Depreciation Charge										
Remaining Value										
Net Revenue		$101,183,848	$99,765,026	$101,183,848	$99,765,026	$100,672,679	$99,765,026	$101,183,848	$99,765,026	$101,183,848
Losses Forward		$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income		$101,183,848	$99,765,026	$101,183,848	$99,765,026	$100,672,679	$99,765,026	$101,183,848	$99,765,026	$101,183,848
Income Tax		$21,248,608	$20,950,656	$21,248,608	$20,950,656	$21,141,263	$20,950,656	$21,248,608	$20,950,656	$21,248,608
Annual Cash Income		$79,935,240	$78,814,371	$79,935,240	$78,814,371	$79,531,416	$78,814,371	$79,935,240	$78,814,371	$79,935,240
Discount Factor		0.1228	0.1117	0.1015	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
Annual Present Value	$423,468,627	$9,819,722	$8,801,844	$8,115,473	$7,274,251	$6,673,120	$6,011,778	$5,542,977	$4,968,411	$4,580,973
Total Capital Investment + Interest										($2,237,446)
Net Present Worth										

Appendix C. Life Cycle Inventories and Applicable Emission Factors from the R&D GREET Model

Life cycle inventories (LCI) and applicable emission factors are presented below for the scenarios as defined in the following table:

I	With purchased caustic and without carbon capture (CC)
II	With onsite caustic via chlor-alkali process and without CC
III	With purchased caustic and CC (90% capture)
IV	With onsite caustic via chlor-alkali process and CC (90% capture)

Table 20. Life cycle inventories (LCI) and applicable emission factors from the R&D GREET Model [12]

	DMR Pretreatment				DDA Pretreatment					Carbon Intensities		
Scenarios	**I**	**II**	**III**	**IV**	**I**	**II**	**III**	**IV**	**Units**	**Values**	**Units**	**Notes**
Products												
Hydrocarbon fuel (*LHV)	19,228 200 836,225	19,228 200 836,225	19,228 200 836,225	19,228 200 836,225	18,235 189 792,826	18,235 189 792,826	18,235 189 792,826	18,235 189 792,826	kg/hr MMkcal/hr* MJ/hr*			
Co-products												
Recovered sodium sulfate salt from WWT	12,065	12,065	12,065	12,065	2,775	2,775	2,775	2,775	kg/hr	-0.1850	kg CO_2e/kg	
Chlorine (Cl_2)	-	2,326	-	2,358	-	3,260	-	3,293	kg/hr	-1.7622	kg CO_2e/kg	
Sulfuric acid (83.6%)	-	34	-	34	-	48	-	48	kg/hr	-0.0451	kg CO_2e/kg	
Sequestered fossil CO_2	-	-	2,491	2,491	-	-	<1	<1	kg/hr			
Sequestered fossil CO_2 (chlor-alkali)	-	-	-	621	-	-	-	868	kg/hr			
Sequestered biogenic carbon	-	-	69,218	69,218	-	-	70,682	70,682	kg/hr			
Export Electricity	-	-	-	-	1,856	-	-	-	kWh/hr	-0.4395	kg CO_2e/kWh	Grid electricity
Resource Consumption												
Biomass Feedstock (20% moisture)	104,167	104,167	104,167	104,167	104,167	104,167	104,167	104,167	kg/hr	0.0759	kg CO_2e/kg	Dry biomass
Sulfuric Acid, 93%	9,376	9,376	9,376	9,376	4,208	4,208	4,208	4,208	kg/hr	0.0451	kg CO_2e/kg	As pure
Caustic (as pure)	2,556	-	2,590	-	3,581	-	3,617	-	kg/hr	2.0521	kg CO_2e/kg	
Sodium Carbonate	6,667	6,667	6,667	6,667	-	-	-	-	kg/hr	0.7040	kg CO_2e/kg	

	DMR Pretreatment				DDA Pretreatment					Carbon Intensities		
Scenarios	**I**	**II**	**III**	**IV**	**I**	**II**	**III**	**IV**	**Units**	**Values**	**Units**	**Notes**
Ammonia	1,157	1,157	1,157	1,157	1,124	1,124	1,124	1,124	kg/hr	2.3847	kg CO_2e/kg	
Hydrochloric Acid (35.2%)	975	975	975	975	944	944	944	944	kg/hr	1.9873	kg CO_2e/kg	As pure
Glucose syrup	1,324	1,324	1,324	1,324	1,351	1,351	1,351	1,351	kg/hr	0.7539	kg CO_2e/kg	
Corn Steep Liquor	90	90	90	90	92	92	92	92	kg/hr	1.6119	kg CO_2e/kg	
Corn oil	7	7	7	7	7	7	7	7	kg/hr	0.0644	kg CO_2e/kg	
Host nutrients	37	37	37	37	38	38	38	38	kg/hr	2.3847	kg CO_2e/kg	
Sulfur Dioxide	9	9	9	9	9	9	9	9	kg/hr	6.9319	kg CO_2e/kg	
Flocculant	408	408	408	408	216	216	216	216	kg/hr	1.2740	kg CO_2e/kg	
Hydrogen	3,014	2,952	3,014	2,951	2,856	2,769	2,856	2,768	kg/hr	9.4447 5.0038 1.7000	kg CO_2e/kg	Grey H_2 Blue H_2 Green H_2
APR/HDO catalyst: Precious metal on carbon support	<1	<1	<1	<1	<1	<1	<1	<1	kg/hr	9.2390	kg CO_2e/kg	Mo/Co/Al_2O_3 catalyst as surrogate
APR/HDO catalyst: Base metal on mixed oxide support	4	4	4	4	4	4	4	4	kg/hr	0.6294	kg CO_2e/kg	
DHOG catalyst: Precious metal on mixed oxide support	<1	<1	<1	<1	<1	<1	<1	<1	kg/hr	9.2390	kg CO_2e/kg	
HDT catalyst: Precious metal on mixed oxide support	<1	<1	<1	<1	<1	<1	<1	<1	kg/hr	9.2390	kg CO_2e/kg	
Boiler Chemicals	<1	<1	<1	<1	<1	<1	<1	<1	kg/hr	6.9319	kg CO_2e/kg	ZSM-5 as surrogate
FGD Lime	148	148	148	148	467	467	467	467	kg/hr	1.2824	kg CO_2e/kg	
WWT Polymer	15	15	15	15	16	16	16	16	kg/hr	1.2740	kg CO_2e/kg	
Cooling Tower Chemicals	2	2	2	2	2	2	2	2	kg/hr	6.9319	kg CO_2e/kg	ZSM-5 as surrogate
Makeup Water	355,556	523,273	360,949	530,938	346,650	581,654	351,591	588,951	kg/hr	0.0000	kg CO_2e/kg	
Natural Gas (chlor-alkali)	-	226	-	229	-	316	-	319	kg/hr	0.5980	kg CO_2e/kg	
Calcium chloride (chlor-alkali)	-	30	-	30	-	42	-	42	kg/hr	6.9319	kg CO_2e/kg	ZSM-5 as surrogate
Hydrochloric Acid (pure) (chlor-alkali)	-	51	-	52	-	71	-	72	kg/hr	1.9873	kg CO_2e/kg	
Rock Salt (chlor-alkali)	-	3,872	-	3,925	-	5,425	-	5,481	kg/hr	0.2416	kg CO_2e/kg	
Sodium Carbonate (chlor-alkali)	-	146	-	148	-	204	-	206	kg/hr	0.7040	kg CO_2e/kg	
Sulfuric Acid, 98% (chlor-alkali)	-	29	-	29	-	41	-	41	kg/hr	0.0451	kg CO_2e/kg	As pure
Carbon Tetrachloride (chlor-alkali)	-	<1	-	<1	-	<1	-	<1	kg/hr	6.9319	kg CO_2e/kg	ZSM-5 as surrogate
Inert Gas (chlor-alkali)	-	<1	-	<1	-	<1	-	<1	kg/hr	8.5891	kg CO_2e/kg	
Sorbent (monoethanolamine, MEA) (CC)	-	-	25	25	-	-	25	26	kg/hr	2.6430	kg CO_2e/kg	
Activated Carbon (CC)	-	-	5	5	-	-	5	5	kg/hr	-2.2148	kg CO_2e/kg	
Import Grid Electricity	1,916	8,130	27,347	33,839	-	6,850	23,091	32,156	kWh/hr	0.4395 0.0000	kg CO_2e/kWh	Grid electricity Wind/solar electricity

Waste Streams

	DMR Pretreatment				DDA Pretreatment					Carbon Intensities		
Scenarios	**I**	**II**	**III**	**IV**	**I**	**II**	**III**	**IV**	**Units**	**Values**	**Units**	**Notes**
Disposal of Ash	4,355	4,355	4,355	4,355	4,887	4,887	4,887	4,887	kg/hr			
Offsite wastewater treatment (chlor-alkali)	-	5,197	-	5,269	-	7,282	-	7,356	kg/hr			
Reclaimer Waste Disposal (CC)	-	-	80	82	-	-	84	86	kg/hr			
Air Emissions												
H_2O	134,318	134,318	134,318	134,318	108,324	108,324	108,324	108,324	kg/hr			
N_2	304,483	304,483	304,483	304,483	380,014	380,014	380,014	380,014	kg/hr			
CO_2 (fossil)	2,767	2,767	277	277	<1	<1	<1	<1	kg/hr			
CO_2 (fossil) (chlor-alkali)	-	681	-	69	-	954	-	96	kg/hr			
CO_2 (biogenic)	78,469	78,469	9,251	9,251	80,243	80,243	9,561	9,561	kg/hr			
O_2	30,165	30,165	30,165	30,165	51,298	51,298	51,298	51,298	kg/hr			
NO_2	14	14	14	14	15	15	15	15	kg/hr			
SO_2	9	9	9	9	28	28	28	28	kg/hr			
CO (as CO_2) (biogenic)	109	109	109	109	111	111	111	111	kg/hr			
CH_4 (biogenic)	1	1	1	1	2	2	2	2	kg/hr			

Appendix D. Key Aspen Stream Data Tables

Figure 22 provides a high-level process flow diagram (PFD) of the integrated process showing key streams, which is followed by Table 21 and Table 22 providing mass and component flowrates of these streams from the Aspen Plus models for DMR and DDA pretreatments, respectively.

Note that some streams in Figure 22 are representative of more than one stream of the same nature (for instance, Stream 610 represents two separate wastewater streams leaving Area 300). Hence, the following tables do not show temperature, pressure and/or vapor fraction data when they are not identical between the individual streams as reflected in Aspen Plus. Also note that only the final products going to Storage (A700) are shown in the PFD.

Component classes for the stream tables are defined as follows:

IS	Insoluble solids
SS	Soluble solids
Cell Mass	Waste water treatment sludge, *Trichoderma reesei* from cellulase production
Furfurals	Furfural, hydroxymethylfurfural (HMF)
Hydrocarbons	Hydrocarbons that make up the final fuel cuts
Inorganic Soluble Solids	Ammonium sulfate, sodium nitrate
Organic Soluble Solids	Organic extractives, solubilized lignin, lactic acid, cellulase nutrient mix
Other Insoluble Solids	Ash, tar, lime, gypsum
Other Structural Carbohydrates	Galactan, mannan, arabinan
Other Sugars	Galactose, mannose, arabinose, sucrose
Oxygenates	Oxygenates in the fuel intermediates
Protein	Corn protein, cellulase, denatured cellulase
Sugar Oligomers	Oligomers of glucose, galactose, mannose, xylose and arabinose

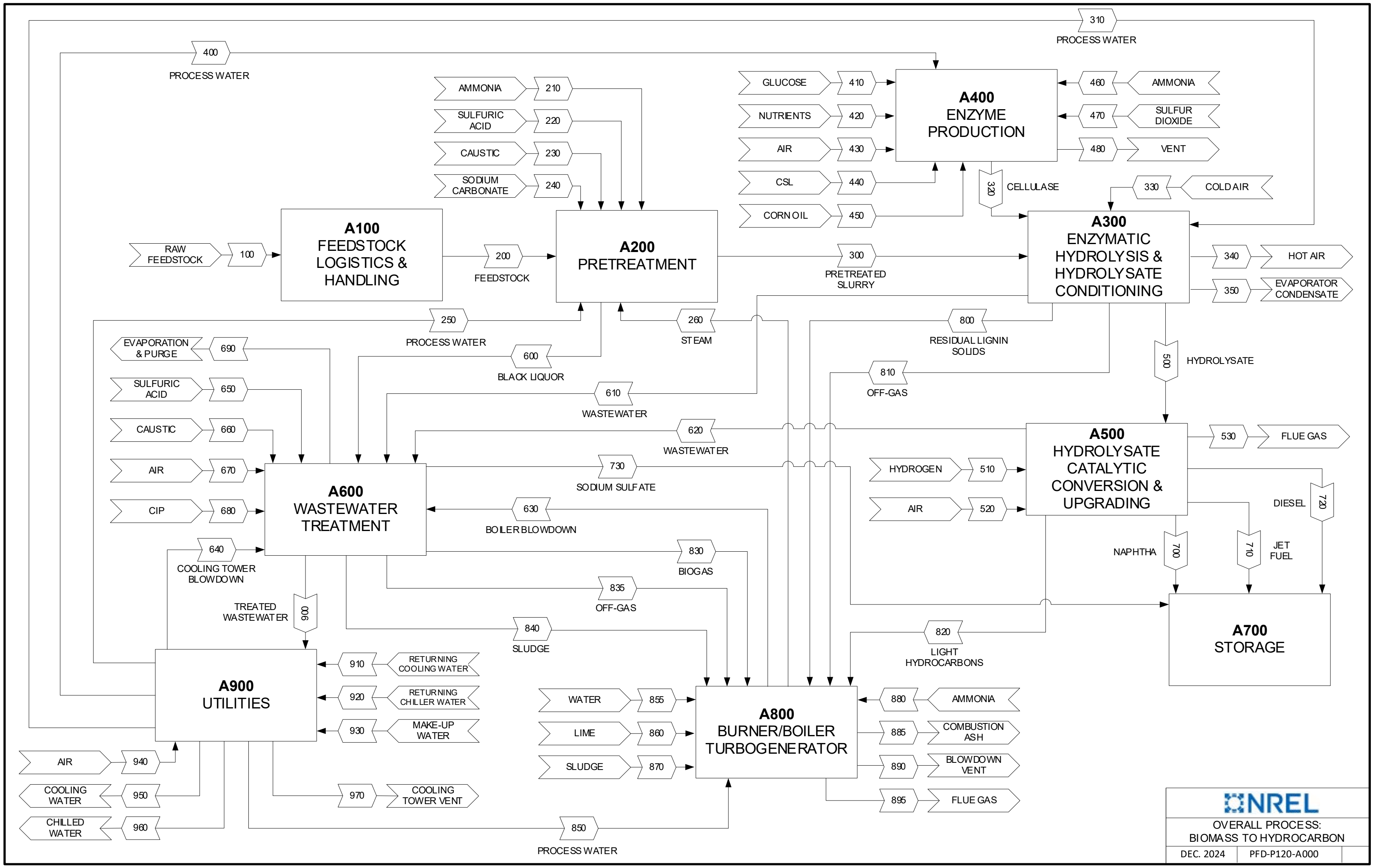


Figure 22. Overall process flow diagram for stream tables

Table 21. Details of Major Streams from the Aspen Plus Model: DMR Pretreatment

Stream #	100	200	210	220	230	240	250	260	300	310	320	330	340	350	400	410
Total Flow, kg/hr	104,167	104,167	<1	<1	2,000	6,667	283,263	29,776	315,835	354,137	7,575	14,085,665	14,085,502	528,494	6,252	1,557
Insoluble Solids (IS), kg/hr	70,483	70,483	-	-	-	-	-	-	63,091	-	398	-	-	-	-	-
Soluble Solids (SS), kg/hr	12,850	12,850	-	-	-	-	-	-	321	-	22	-	-	<1	-	1,324
Temperature, °C	25	25	20	20	20	20	33	266	97	33	29	25	55	70	33	28
Pressure, atm	1.0	1.0	9.0	3.4	1.0	1.0	5.0	13.0	8.0	5.0	1.7	1.0	1.0	2.0	5.0	1.0
Mass Vapor Fraction	0	0	0	0	0	0	0	1	0	0	0	1	1	0	0	0
Components, kg/hr																
Acetate (IS)	1,508	1,508	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Acetic Acid	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ammonia	-	-	<1	-	-	-	-	-	-	-	<1	-	-	<1	-	-
Carbon Dioxide	-	-	-	-	-	-	-	-	-	-	<1	-	-	<1	-	-
Cell Mass (IS)	-	-	-	-	-	-	-	-	-	-	59	-	-	-	-	-
Cellulose (IS)	29,205	29,205	-	-	-	-	-	-	28,621	-	-	-	-	-	-	-
$CO/SO_X/NO_X/H_2S$	-	-	-	-	-	-	-	-	-	-	<1	-	-	<1	-	-
Corn Oil	-	-	-	-	-	-	-	-	-	-	7	-	-	<1	-	-
Furfurals	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Glucose (SS)	-	-	-	-	-	-	-	-	-	-	-	-	-	<1	-	1,324
Hydrocarbons	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Hydrogen	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Inorganic Soluble Solids (SS)	-	-	-	-	-	-	-	-	<1	-	-	-	-	-	-	-
Lignin (IS)	13,132	13,132	-	-	-	-	-	-	10,505	-	-	-	-	-	-	-
Methane	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Nitrogen	-	-	-	-	-	-	-	-	-	-	<1	10,804,874	10,804,749	<1	-	-
Organic Soluble Solids (SS)	12,208	12,208	-	-	-	-	-	-	-	-	22	-	-	<1	-	-
Other Insoluble Solids (IS)	4,108	4,108	-	-	-	-	-	-	3,656	-	-	-	-	-	-	-
Other Struct. Carbohydr. (IS)	3,675	3,675	-	-	-	-	-	-	3,080	-	-	-	-	-	-	-
Other Sugars (SS)	642	642	-	-	-	-	-	-	321	-	-	-	-	<1	-	-
Oxygen	-	-	-	-	-	-	-	-	-	-	<1	3,280,791	3,280,753	<1	-	-
Oxygenates	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Protein (IS)	2,583	2,583	-	-	-	-	-	-	2,583	-	339	-	-	-	-	-
Sodium Carbonate	-	-	-	-	-	6,667	-	-	-	-	-	-	-	-	-	-
Sodium Hydroxide	-	-	-	-	2,000	-	-	-	-	-	-	-	-	-	-	-
Sodium Sulfate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sugar Oligomers (SS)	-	-	-	-	-	-	-	-	-	-	-	-	-	<1	-	-
Sulfuric Acid	-	-	-	<1	-	-	-	-	<1	-	-	-	-	-	-	-
Water	20,833	20,833	-	<1	-	-	283,263	29,776	252,423	354,137	7,147	-	-	528,493	6,252	234
Xylan (IS)	16,273	16,273	-	-	-	-	-	-	14,646	-	-	-	-	-	-	-
Xylose (SS)	-	-	-	-	-	-	-	-	-	-	-	-	-	<1	-	-

Table 21. Details of Major Streams from the Aspen Plus Model: DMR *(Continued)*

Stream #	**420**	**430**	**440**	**450**	**460**	**470**	**480**	**500**	**510**	**520**	**530**	**600**	**610**	**620**	**630**	**640**
Total Flow, kg/hr	37	17,839	90	7	63	9	18,279	87,676	3,014	2,040,827	2,041,695	110,037	4,963	69,612	4,245	17,118
Insoluble Solids (IS), kg/hr	-	-	23	-	-	-	-	-	-	-	-	3,258	51	-	-	-
Soluble Solids (SS), kg/hr	37	-	22	-	-	-	<1	43,536	-	-	-	15,156	2,440	-	-	-
Temperature, °C	20	25	20	20	25	28	N/A	63	N/A	N/A	N/A	90	N/A	N/A	98	28
Pressure, atm	1.0	1.0	1.0	1.0	1.0	3.0	1.0	72.4	N/A	1.0	N/A	1.0	72.4	N/A	1.0	5.1
Mass Vapor Fraction	0	1	0	0	1	1	1	0	1	1	1	0	0	0	0	0
Components, kg/hr																
Acetate (IS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Acetic Acid	-	-	-	-	-	-	-	-	-	-	-	1,508	-	-	-	-
Ammonia	-	-	-	-	63	-	<1	-	-	-	-	-	<1	-	-	-
Carbon Dioxide	-	-	-	-	-	-	1,304	-	-	-	1,650	-	<1	-	-	-
Cell Mass (IS)	-	-	-	-	-	-	-	-	-	-	-	-	<1	-	-	-
Cellulose (IS)	-	-	-	-	-	-	-	-	-	-	-	584	5	-	-	-
CO/SOX/NOX/H2S	-	-	-	-	-	9	<1	-	-	-	-	-	<1	-	-	-
Corn Oil	-	-	-	7	-	-	<1	-	-	-	-	-	7	-	-	-
Furfurals	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Glucose (SS)	-	-	-	-	-	-	<1	25,120	-	-	-	-	1,396	-	-	-
Hydrocarbons	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Hydrogen	-	-	-	-	-	-	-	-	3,014	-	-	-	-	-	-	-
Inorganic Soluble Solids (SS)	-	-	-	-	-	-	-	-	-	-	-	-	<1	-	-	-
Lignin (IS)	-	-	-	-	-	-	-	-	-	-	-	-	26	-	-	-
Methane	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Nitrogen	-	13,416	-	-	-	-	13,416	-	-	1,565,484	1,565,484	-	-	-	-	<1
Organic Soluble Solids (SS)	37	-	22	-	-	-	<1	-	-	-	-	14,835	21	-	-	-
Other Insoluble Solids (IS)	-	-	-	-	-	-	-	-	-	-	-	452	9	-	-	-
Other Struct. Carbohydr. (IS)	-	-	-	-	-	-	-	-	-	-	-	595	1	-	-	-
Other Sugars (SS)	-	-	-	-	-	-	-	2,939	-	-	-	321	163	-	-	-
Oxygen	-	4,074	-	-	-	-	3,135	-	-	475,343	473,401	-	-	-	-	<1
Oxygenates	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Protein (IS)	-	-	23	-	-	-	-	-	-	-	-	-	7	-	-	-
Sodium Carbonate	-	-	-	-	-	-	-	-	-	-	-	6,667	-	-	-	-
Sodium Hydroxide	-	-	-	-	-	-	-	-	-	-	-	2,000	-	-	-	-
Sodium Sulfate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sugar Oligomers (SS)	-	-	-	-	-	-	-	1,285	-	-	-	-	71	-	-	-
Sulfuric Acid	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Water	-	349	45	-	-	-	424	44,140	-	-	1,160	81,449	2,466	69,612	4,245	17,117
Xylan (IS)	-	-	-	-	-	-	-	-	-	-	-	1,627	2	-	-	-
Xylose (SS)	-	-	-	-	-	-	-	14,192	-	-	-	-	788	-	-	-

Table 21. Details of Major Streams from the Aspen Plus Model: DMR *(Continued)*

Stream #	**650**	**660**	**670**	**680**	**690**	**700**	**710**	**720**	**730**	**800**	**810**	**820**	**830**	**835**	**840**	**850**
Total Flow, kg/hr	9,376	<1	70,028	145	82,944	3,837	13,473	1,918	12,065	56,411	165	982	17,960	35,154	6,499	36,954
Insoluble Solids (IS), kg/hr	-	-	-	-	-	-	-	-	-	20,274	-	-	-	-	4,703	-
Soluble Solids (SS), kg/hr	-	-	-	-	51	<1	<1	<1	-	2,547	<1	<1	<1	<1	11	-
Temperature, °C	25	20	25	33	N/A	16	16	16	107	21	21	N/A	37	100	40	33
Pressure, atm	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	N/A	1.0	1.0	2.0	5.0
Mass Vapor Fraction	0	0	1	0	N/A	0	0	0	0	0	1	N/A	1	1	0	0
Components, kg/hr																
Acetate (IS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Acetic Acid	-	-	-	-	2	-	-	-	-	-	-	-	<1	185	<1	-
Ammonia	-	-	-	-	<1	-	-	-	-	<1	<1	-	<1	<1	<1	-
Carbon Dioxide	-	-	-	-	1,560	-	-	-	-	<1	<1	220	12,957	2,766	<1	-
Cell Mass (IS)	-	-	-	-	-	-	-	-	-	59	-	-	-	-	1,395	-
Cellulose (IS)	-	-	-	-	-	-	-	-	-	1,998	-	-	-	-	589	-
CO/SOX/NOX/H2S	-	-	-	-	10	-	-	-	-	<1	<1	-	20	<1	<1	-
Corn Oil	-	-	-	-	<1	-	-	-	-	<1	<1	-	<1	<1	<1	-
Furfurals	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Glucose (SS)	-	-	-	-	2	-	-	-	-	1,469	<1	-	<1	<1	1	-
Hydrocarbons	-	-	-	-	-	3,769	13,473	1,918	-	-	-	717	-	-	-	-
Hydrogen	-	-	-	-	-	-	-	-	-	-	-	31	-	-	-	-
Inorganic Soluble Solids (SS)	-	-	-	-	<1	-	-	-	-	<1	-	-	-	-	<1	-
Lignin (IS)	-	-	-	-	-	-	-	-	-	10,479	-	-	-	-	26	-
Methane	-	-	-	-	1	-	-	-	-	-	-	-	4,424	-	<1	-
Nitrogen	-	-	52,665	-	52,663	-	-	-	-	<1	124	-	2	-	<1	-
Organic Soluble Solids (SS)	-	-	-	-	18	-	-	-	-	1	<1	-	<1	<1	9	-
Other Insoluble Solids (IS)	-	-	-	-	-	-	-	-	-	3,647	-	-	-	-	461	-
Other Struct. Carbohydr. (IS)	-	-	-	-	-	-	-	-	-	446	-	-	-	-	596	-
Other Sugars (SS)	-	-	-	-	30	<1	<1	<1	-	172	<1	<1	<1	<1	<1	-
Oxygen	-	-	15,991	-	14,977	-	-	-	-	<1	38	-	<1	-	<1	-
Oxygenates	-	-	-	-	-	<1	<1	<1	-	-	-	<1	-	-	-	-
Protein (IS)	-	-	-	-	-	-	-	-	-	2,915	-	-	-	-	6	-
Sodium Carbonate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sodium Hydroxide	-	<1	-	-	<1	-	-	-	-	-	-	-	-	-	<1	-
Sodium Sulfate	-	-	-	-	86	-	-	-	12,065	-	-	-	-	<1	157	-
Sugar Oligomers (SS)	-	-	-	-	<1	-	-	-	-	75	<1	-	<1	<1	<1	-
Sulfuric Acid	8,719	-	-	-	9	-	-	-	-	-	-	-	<1	<1	<1	-
Water	656	<1	1,372	145	13,585	68	<1	<1	-	33,588	3	13	557	32,202	1,626	36,954
Xylan (IS)	-	-	-	-	-	-	-	-	-	730	-	-	-	-	1,629	-
Xylose (SS)	-	-	-	-	1	-	-	-	-	830	<1	-	<1	<1	<1	-

Table 21. Details of Major Streams from the Aspen Plus Model: DMR *(Continued)*

Stream #	**855**	**860**	**870**	**880**	**885**	**890**	**895**	**900**	**910**	**920**	**930**	**940**	**950**	**960**	**970**
Total Flow, kg/hr	592	148	304,700	730	4,355	2,340	418,986	130,895	11,554,515	438,014	709,693	6,768,640	11,554,515	438,014	6,911,525
Insoluble Solids (IS), kg/hr	-	148	-	-	4,355	-	-	-	-	-	-	-	-	-	-
Soluble Solids (SS), kg/hr	-	-	-	-	-	-	<1	-	-	-	-	-	-	-	-
Temperature, °C	33	25	25	80	193	100	198	26	37	15	13	25	28	4	37
Pressure, atm	5.0	1.0	1.0	1.1	1.0	1.0	1.0	1.0	5.0	4.1	4.0	1.0	5.0	4.1	1.0
Mass Vapor Fraction	0	0	1	1	0	1	1	0	0	0	0	1	0	0	1
Components, kg/hr															
Acetate (IS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Acetic Acid	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ammonia	-	-	-	730	-	-	<1	-	-	-	-	-	-	-	-
Carbon Dioxide	-	-	-	-	-	-	76,722	-	-	-	-	-	-	-	-
Cell Mass (IS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Cellulose (IS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CO/SOX/NOX/H2S	-	-	-	-	-	-	396	-	-	-	-	-	-	-	-
Corn Oil	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Furfurals	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Glucose (SS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Hydrocarbons	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Hydrogen	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Inorganic Soluble Solids (SS)	-	-	-	-	-	-	<1	-	-	-	-	-	-	-	-
Lignin (IS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Methane	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Nitrogen	-	-	229,152	-	-	-	230,441	<1	133	-	-	5,090,418	133	-	5,090,418
Organic Soluble Solids (SS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Other Insoluble Solids (IS)	-	148	-	-	4,355	-	-	-	-	-	-	-	-	-	-
Other Struct. Carbohydr. (IS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Other Sugars (SS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Oxygen	-	-	69,580	-	-	-	11,577	<1	77	-	-	1,545,654	77	-	1,545,654
Oxygenates	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Protein (IS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sodium Carbonate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sodium Hydroxide	-	-	-	-	-	-	<1	-	-	-	-	-	-	-	-
Sodium Sulfate	-	-	-	-	-	-	157	-	-	-	-	-	-	-	-
Sugar Oligomers (SS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sulfuric Acid	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Water	592	-	5,968	-	-	2,340	99,693	130,895	11,554,304	438,014	709,693	132,568	11,554,304	438,014	275,453
Xylan (IS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Xylose (SS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-

Table 22. Details of Major Streams from the Aspen Plus Model: DDA Pretreatment

Stream #	**100**	**200**	**210**	**220**	**230**	**240**	**250**	**260**	**300**	**310**	**320**	**330**	**340**	**350**	**400**	**410**
Total Flow, kg/hr	104,167	104,167	260	807	2,500	-	453,030	3,356	326,137	214,857	7,730	10,633,632	10,633,469	397,821	6,379	1,589
Insoluble Solids (IS), kg/hr	70,483	70,483	-	-	-	-	-	-	45,886	-	406	-	-	-	-	-
Soluble Solids (SS), kg/hr	12,850	12,850	-	-	-	-	-	-	19,743	-	23	-	-	<1	-	1,351
Temperature, °C	25	25	20	20	20	20	33	266	75	33	29	25	55	70	33	28
Pressure, atm	1.0	1.0	9.0	3.4	1.0	1.0	5.0	13.0	7.8	5.0	1.7	1.0	1.0	2.0	5.0	1.0
Mass Vapor Fraction	0	0	0	0	0	0	0	1	0	0	0	1	1	0	0	0
Components, kg/hr																
Acetate (IS)	1,508	1,508	-	-	-	-	-	-	181	-	-	-	-	-	-	-
Acetic Acid	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ammonia	-	-	260	-	-	-	-	-	-	-	<1	-	-	<1	-	-
Carbon Dioxide	-	-	-	-	-	-	-	-	-	-	<1	-	-	<1	-	-
Cell Mass (IS)	-	-	-	-	-	-	-	-	-	-	60	-	-	-	-	-
Cellulose (IS)	29,205	29,205	-	-	-	-	-	-	29,205	-	-	-	-	-	-	-
CO/SOX/NOX/H2S	-	-	-	-	-	-	-	-	-	-	<1	-	-	<1	-	-
Corn Oil	-	-	-	-	-	-	-	-	-	-	7	-	-	<1	-	-
Furfurals	-	-	-	-	-	-	-	-	343	-	-	-	-	294	-	-
Glucose (SS)	-	-	-	-	-	-	-	-	-	-	-	-	-	<1	-	1,351
Hydrocarbons	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Hydrogen	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Inorganic Soluble Solids (SS)	-	-	-	-	-	-	-	-	1,011	-	-	-	-	-	-	-
Lignin (IS)	13,132	13,132	-	-	-	-	-	-	10,505	-	-	-	-	-	-	-
Methane	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Nitrogen	-	-	-	-	-	-	-	-	-	-	<1	8,156,878	8,156,753	<1	-	-
Organic Soluble Solids (SS)	12,208	12,208	-	-	-	-	-	-	-	-	23	-	-	<1	-	-
Other Insoluble Solids (IS)	4,108	4,108	-	-	-	-	-	-	1,061	-	-	-	-	-	-	-
Other Struct. Carbohydr. (IS)	3,675	3,675	-	-	-	-	-	-	341	-	-	-	-	-	-	-
Other Sugars (SS)	642	642	-	-	-	-	-	-	3,669	-	-	-	-	<1	-	-
Oxygen	-	-	-	-	-	-	-	-	-	-	<1	2,476,754	2,476,716	<1	-	-
Oxygenates	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Protein (IS)	2,583	2,583	-	-	-	-	-	-	2,583	-	346	-	-	-	-	-
Sodium Carbonate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sodium Hydroxide	-	-	-	-	2,500	-	-	-	-	-	-	-	-	-	-	-
Sodium Sulfate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sugar Oligomers (SS)	-	-	-	-	-	-	-	-	566	-	-	-	-	<1	-	-
Sulfuric Acid	-	-	-	750	-	-	-	-	<1	-	-	-	-	-	-	-
Water	20,833	20,833	-	56	-	-	453,030	3,356	260,165	214,857	7,292	-	-	397,526	6,379	238
Xylan (IS)	16,273	16,273	-	-	-	-	-	-	2,009	-	-	-	-	-	-	-
Xylose (SS)	-	-	-	-	-	-	-	-	14,498	-	-	-	-	<1	-	-

Table 22. Details of Major Streams from the Aspen Plus Model: DDA *(Continued)*

Stream #	**420**	**430**	**440**	**450**	**460**	**470**	**480**	**500**	**510**	**520**	**530**	**600**	**610**	**620**	**630**	**640**
Total Flow, kg/hr	38	18,203	92	7	64	9	18,652	85,036	2,856	1,951,987	1,952,943	237,983	5,922	67,906	4,216	18,505
Insoluble Solids (IS), kg/hr	-	-	23	-	-	-	-	-	-	-	-	3,406	196	-	-	-
Soluble Solids (SS), kg/hr	38	-	23	-	-	-	<1	41,166	-	-	-	15,156	3,268	-	-	-
Temperature, °C	20	25	20	20	25	28	N/A	64	N/A	N/A	N/A	83	N/A	N/A	98	28
Pressure, atm	1.0	1.0	1.0	1.0	1.0	3.0	1.0	72.4	N/A	1.0	N/A	1.0	72.4	N/A	1.0	5.1
Mass Vapor Fraction	0	1	0	0	1	1	1	0	1	1	1	0	0	0	0	0
Components, kg/hr																
Acetate (IS)	-	-	-	-	-	-	-	-	-	-	-	-	142	-	-	-
Acetic Acid	-	-	-	-	-	-	-	-	-	-	-	1,327	-	-	-	-
Ammonia	-	-	-	-	64	-	<1	-	-	-	-	-	<1	-	-	-
Carbon Dioxide	-	-	-	-	-	-	1,331	-	-	-	1,538	-	<1	-	-	-
Cell Mass (IS)	-	-	-	-	-	-	-	-	-	-	-	-	<1	-	-	-
Cellulose (IS)	-	-	-	-	-	-	-	-	-	-	-	-	12	-	-	-
CO/SOX/NOX/H2S	-	-	-	-	-	9	<1	-	-	-	-	-	<1	-	-	-
Corn Oil	-	-	-	7	-	-	<1	-	-	-	-	-	7	-	-	-
Furfurals	-	-	-	-	-	-	-	30	-	-	-	341	2	-	-	-
Glucose (SS)	-	-	-	-	-	-	<1	23,303	-	-	-	-	1,295	-	-	-
Hydrocarbons	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Hydrogen	-	-	-	-	-	-	-	-	2,856	-	-	-	-	-	-	-
Inorganic Soluble Solids (SS)	-	-	-	-	-	-	-	-	-	-	-	-	960	-	-	-
Lignin (IS)	-	-	-	-	-	-	-	-	-	-	-	-	26	-	-	-
Methane	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Nitrogen	-	13,690	-	-	-	-	13,689	-	-	1,497,336	1,497,336	-	-	-	-	<1
Organic Soluble Solids (SS)	38	-	23	-	-	-	<1	-	-	-	-	14,835	21	-	-	-
Other Insoluble Solids (IS)	-	-	-	-	-	-	-	-	-	-	-	3,081	3	-	-	-
Other Struct. Carbohydr. (IS)	-	-	-	-	-	-	-	-	-	-	-	-	<1	-	-	-
Other Sugars (SS)	-	-	-	-	-	-	-	3,293	-	-	-	321	183	-	-	-
Oxygen	-	4,157	-	-	-	-	3,199	-	-	454,651	452,851	-	-	-	-	<1
Oxygenates	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Protein (IS)	-	-	23	-	-	-	-	-	-	-	-	-	7	-	-	-
Sodium Carbonate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sodium Hydroxide	-	-	-	-	-	-	-	-	-	-	-	2,500	-	-	-	-
Sodium Sulfate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sugar Oligomers (SS)	-	-	-	-	-	-	-	1,557	-	-	-	<1	87	-	-	-
Sulfuric Acid	-	-	-	-	-	-	-	-	-	-	-	<1	-	-	-	-
Water	-	357	46	-	-	-	432	43,840	-	-	1,219	215,253	2,449	67,906	4,216	18,505
Xylan (IS)	-	-	-	-	-	-	-	-	-	-	-	325	5	-	-	-
Xylose (SS)	-	-	-	-	-	-	-	13,014	-	-	-	<1	723	-	-	-

Table 22. Details of Major Streams from the Aspen Plus Model: DDA *(Continued)*

Stream #	**650**	**660**	**670**	**680**	**690**	**700**	**710**	**720**	**730**	**800**	**810**	**820**	**830**	**835**	**840**	**850**
Total Flow, kg/hr	3,401	1,086	165,906	145	174,323	3,636	12,781	1,818	2,775	59,941	165	792	18,380	-	11,923	11,763
Insoluble Solids (IS), kg/hr	-	-	-	-	<1	-	-	-	-	21,564	-	-	-	-	4,771	-
Soluble Solids (SS), kg/hr	-	-	-	-	1,397	<1	<1	<1	-	2,459	<1	<1	<1	-	36	-
Temperature, °C	25	20	25	33	N/A	16	16	16	109	22	22	N/A	38	N/A	31	33
Pressure, atm	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	N/A	1.0	1.0	2.0	5.0
Mass Vapor Fraction	0	0	1	0	N/A	0	0	0	0	0	1	N/A	1	0	0	0
Components, kg/hr																
Acetate (IS)	-	-	-	-	<1	-	-	-	-	39	-	-	-	-	<1	-
Acetic Acid	-	-	-	-	<1	-	-	-	-	-	-	-	<1	-	<1	-
Ammonia	-	-	-	-	<1	-	-	-	-	<1	<1	-	10	-	2	-
Carbon Dioxide	-	-	-	-	1,707	-	-	-	-	<1	<1	33	13,298	-	1	-
Cell Mass (IS)	-	-	-	-	-	-	-	-	-	60	-	-	-	-	1,311	-
Cellulose (IS)	-	-	-	-	-	-	-	-	-	4,661	-	-	-	-	12	-
CO/SOX/NOX/H2S	-	-	-	-	140	-	-	-	-	<1	<1	-	133	-	1	-
Corn Oil	-	-	-	-	<1	-	-	-	-	<1	<1	-	<1	-	<1	-
Furfurals	-	-	-	-	<1	-	-	-	-	17	<1	-	<1	-	<1	-
Glucose (SS)	-	-	-	-	<1	-	-	-	-	1,363	<1	-	<1	-	<1	-
Hydrocarbons	-	-	-	-	-	3,566	12,781	1,818	-	-	-	737	-	-	-	-
Hydrogen	-	-	-	-	-	-	-	-	-	-	-	5	-	-	-	-
Inorganic Soluble Solids (SS)	-	-	-	-	1,128	-	-	-	-	51	-	-	<1	-	19	-
Lignin (IS)	-	-	-	-	-	-	-	-	-	10,479	-	-	-	-	26	-
Methane	-	-	-	-	2	-	-	-	-	-	-	-	4,334	-	<1	-
Nitrogen	-	-	124,771	-	124,767	-	-	-	-	<1	124	-	4	-	<1	-
Organic Soluble Solids (SS)	-	-	-	-	238	-	-	-	-	1	<1	-	<1	-	15	-
Other Insoluble Solids (IS)	-	-	-	-	-	-	-	-	-	1,058	-	-	-	-	3,084	-
Other Struct. Carbohydr. (IS)	-	-	-	-	-	-	-	-	-	341	-	-	-	-	<1	-
Other Sugars (SS)	-	-	-	-	30	<1	<1	<1	-	193	<1	<1	<1	-	<1	-
Oxygen	-	-	37,886	-	35,952	-	-	-	-	<1	38	-	2	-	<1	-
Oxygenates	-	-	-	-	-	<1	<1	<1	-	-	-	<1	-	-	-	-
Protein (IS)	-	-	-	-	-	-	-	-	-	2,922	-	-	-	-	6	-
Sodium Carbonate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sodium Hydroxide	-	543	-	-	<1	-	-	-	-	-	-	-	-	-	<1	-
Sodium Sulfate	-	-	-	-	1,438	-	-	-	2,775	-	-	-	-	-	105	-
Sugar Oligomers (SS)	-	-	-	-	<1	-	-	-	-	91	<1	-	<1	-	<1	-
Sulfuric Acid	3,163	-	-	-	9	-	-	-	-	-	-	-	<1	-	<1	-
Water	238	543	3,249	145	8,909	70	<1	<1	-	35,898	3	17	598	-	7,004	11,763
Xylan (IS)	-	-	-	-	-	-	-	-	-	2,004	-	-	-	-	330	-
Xylose (SS)	-	-	-	-	<1	-	-	-	-	761	<1	-	<1	-	<1	-

Table 22. Details of Major Streams from the Aspen Plus Model: DDA *(Continued)*

Stream #	**855**	**860**	**870**	**880**	**885**	**890**	**895**	**900**	**910**	**920**	**930**	**940**	**950**	**960**	**970**
Total Flow, kg/hr	1,866	467	309,671	740	4,887	2,324	399,058	297,582	12,491,185	446,949	561,507	7,317,269	12,491,185	446,949	7,471,736
Insoluble Solids (IS), kg/hr	-	467	-	-	4,887	-	-	-	-	-	-	-	-	-	-
Soluble Solids (SS), kg/hr	-	-	-	-	-	-	19	-	-	-	-	-	-	-	-
Temperature, °C	33	25	25	80	188	100	194	27	37	15	13	25	28	4	37
Pressure, atm	5.0	1.0	1.0	1.1	1.0	1.0	1.0	1.0	5.0	4.1	4.0	1.0	5.0	4.1	1.0
Mass Vapor Fraction	0	0	1	1	0	1	1	0	0	0	0	1	0	0	1
Components, kg/hr															
Acetate (IS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Acetic Acid	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Ammonia	-	-	-	740	-	-	-	-	-	-	-	-	-	-	-
Carbon Dioxide	-	-	-	-	-	-	75,667	-	-	-	-	-	-	-	-
Cell Mass (IS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Cellulose (IS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
CO/SOX/NOX/H2S	-	-	-	-	-	-	421	-	-	-	-	-	-	-	-
Corn Oil	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Furfurals	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Glucose (SS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Hydrocarbons	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Hydrogen	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Inorganic Soluble Solids (SS)	-	-	-	-	-	-	19	-	-	-	-	-	-	-	-
Lignin (IS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Methane	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Nitrogen	-	-	232,891	-	-	-	234,199	<1	144	-	-	5,503,020	144	-	5,503,020
Organic Soluble Solids (SS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Other Insoluble Solids (IS)	-	467	-	-	4,887	-	-	-	-	-	-	-	-	-	-
Other Struct. Carbohydr. (IS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Other Sugars (SS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Oxygen	-	-	70,715	-	-	-	11,712	<1	83	-	-	1,670,936	83	-	1,670,936
Oxygenates	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Protein (IS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sodium Carbonate	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sodium Hydroxide	-	-	-	-	-	-	<1	-	-	-	-	-	-	-	-
Sodium Sulfate	-	-	-	-	-	-	105	-	-	-	-	-	-	-	-
Sugar Oligomers (SS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Sulfuric Acid	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Water	1,866	-	6,065	-	-	2,324	76,936	297,581	12,490,957	446,949	561,507	143,313	12,490,957	446,949	297,780
Xylan (IS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
Xylose (SS)	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-