Ex Situ Catalytic Fast Pyrolysis of Lignocellulosic Biomass to Hydrocarbon Fuels: 2018 State of Technology and Future Research

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1 National Renewable Energy Laboratory
2 DWH Process Consulting
3 Pacific Northwest National Laboratory
4 Idaho National Laboratory
5 Argonne National Laboratory

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NOTICE

Acknowledgments

Nomenclature

AAEM	alkali and alkaline earth metal
ANL	Argonne National Laboratory
Btu	British thermal unit
CFP	catalytic fast pyrolysis
DCFROR	discounted cash flow rate of return
FCI	fixed capital investment
GGE	gallon gasoline equivalent
GHG	greenhouse gas
HGF	hot gas filter
INL	Idaho National Laboratory
IRR	internal rate of return
ISBL	inside battery limits
LHV	lower heating value
MFSP	minimum fuel selling price
MM	million
NG	natural gas
PSA	pressure swing adsorption
RTO	regenerative thermal oxidizer
SCSA	supply chain sustainability analysis
SOT	state of technology
TCI	total capital investment
TDC	total direct cost
TEA	techno-economic analysis
TIC	total installed cost
TPEC	total purchased equipment cost
WHSV	weight hourly space velocity

Executive Summary

This report documents the progress in research funded by the U.S. Department of Energy's Office of Energy Efficiency and Renewable Energy Bioenergy Technologies Office, for the conversion of biomass to infrastructure-compatible liquid hydrocarbon fuels via catalytic fast pyrolysis (CFP). This research is focused on an *ex situ* CFP pathway where biomass undergoes a rapid deconstruction in a fast pyrolysis reactor at approximately 500°C (932°F), followed by the separation of vapors from solids (char and mineral matter); the vapors are then sent to an *ex situ* catalytic reactor for upgrading. Upgrading involves deoxygenation, hydrogenation, and carbon-carbon coupling, and this renders the vapors significantly less reactive and more amenable to further processing upon condensation (condensation produces CFP oil). Solids removal prior to the *ex situ* upgrading step provides an advantage with respect to catalyst stability and choices; catalyst choices can be further broadened to include noble metals in fixed bed systems. The effectiveness of this *ex situ* vapor upgrading step for CFP oil quality improvement has been verified, with experiments proving that single-step hydrotreating can deoxygenate the liquid product to <1wt% oxygen. Catalyst stability during hydrotreating of raw fast pyrolysis bio-oil is a major challenge; it requires multiple hydrotreating steps unless vapor upgrading (CFP) is included.

Significant advancements have been made with this research since it started in 2014. Catalyst development and testing has resulted in a >60% relative increase in carbon efficiency. This has allowed a quicker reduction in the modeled minimum fuel selling price (MFSP) compared to initial out-year projections documented in previous Bioenergy Technologies Office Multi-Year Program Plans. Modeled reduction in the MFSP since 2014, based on bench-scale experimental results, are shown in Figure ES-1; further details for the 2018 state of technology (SOT) are in Table ES-1 and Table ES-2. Updated 2022 goals show future technical and cost projections based on an extrapolation of the current research and its estimated trajectory (Figure ES-1, Table ES-1, and Table ES-3).
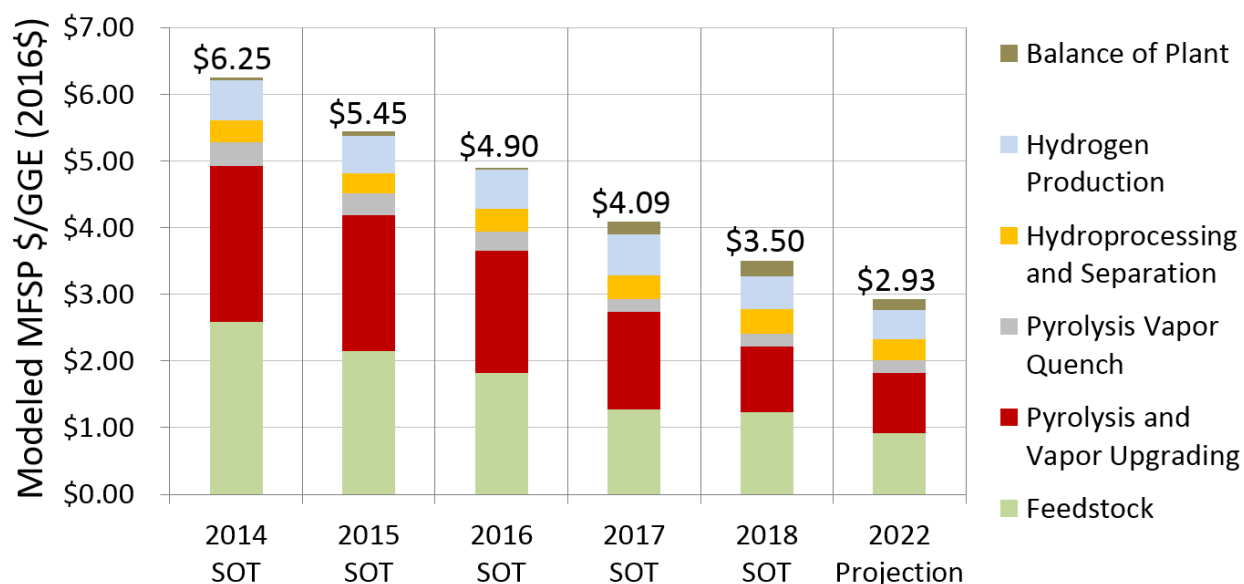
Figure ES-1. Modeled MFSP, based on experimental results for 2014–2018, and 2022 projection

With the recent gains in process efficiency, research focus through 2022 will include considerations of future industrial relevance, in addition to further yield improvements primarily by reducing carbon loss to light gases (CO and CO_2). Research will include: establishing longer CFP catalyst lifetimes and longer onstream times before requiring regeneration; developing more rapid regeneration; enabling the use of lower cost (and less pristine) feedstocks to allow further cost reduction and added diversity in the feedstock supply chain; and targeting improvements in fuel quality making the product fuel blendstocks more desirable for end use in transportation. Research through 2022 and beyond will include the improvement and tailoring of CFP oil composition through further catalyst research; this can enable reduced hydroprocessing costs including the option of coprocessing at petroleum refineries, improvements in downstream fuel quality, and the production of valuable and separable coproducts.

Table ES-1. Key Process Metrics for 2018 SOT and 2022 Projections (Updated and Previous)

	2018 SOT[a]	Updated 2022 Projection	Previous 2022 Projection[b]
Fast Pyrolysis Intermediate[c]			
Gas Species – CO, CO_2, C_1-C_4 (wt% of dry biomass)	13	13	13
Organics (wt% of dry biomass)	64	64	64
Water (wt% of dry biomass)	11	11	11
Char (wt% of dry biomass)	12	12	12
Vapor Upgrading	**Fixed Bed**	**Fixed Bed**	**Fluidized Bed**
Gas (wt% of dry biomass)	31	31	23
Aqueous Phase (wt% of dry biomass)	23	23	30
Carbon Loss (% of C in biomass)	*5*	*3*	*1.3*
Organic Phase (wt% of dry biomass)	31	31	27
H/C Molar Ratio	*1.2*	*1.2*	*1.6*
Oxygen (wt% in organic phase)	*18.5*	*16.4*	*6.4*
Carbon Efficiency (%)	*45*	*47*	*44*
Solid Losses, Char + Coke (wt% of dry biomass)	12 + 3	12 + 3	12 + 8
Final Fuel Blendstock			
Yield (%, w/w dry biomass)	23	25	25
Hydroprocessing Carbon Efficiency (%)	89	91	94
Overall Carbon Efficiency (% of C in biomass)	39.7	42.4	41.5
Overall Carbon Efficiency (% of C in biomass + NG)	39.7	42.3	41.5
Total Product (GGE/dry U.S. ton)	72	77	78
Gasoline-Range Product (gallons/dry U.S. ton)	36	38	36
Diesel-Range Product (gallons/dry U.S. ton)	34	37	39
Gasoline/Diesel-Range Product (% GGE basis)	48/52	48/52	45/55
Oxygen Content in Cumulative Product (wt%)	0.5	0.6	0.4
Minimum Fuel Selling Price ($/GGE)	3.50	2.93	3.23
Natural Gas[†] and Electricity			
Natural Gas Energy Input (% of biomass, LHV basis)	0.2	0.2	0.2
Natural Gas Cost Contribution (¢/GGE)	0.3	0.2	0.2
Surplus Electricity Credit (¢/GGE)	2	3	3
Fuel Blendstock Production Efficiencies			
Biomass Feedstock (%, LHV basis)	52	56	57
Biomass + Natural Gas[d] (%, LHV basis)	52	56	57
Biomass + Natural Gas[d] + Electricity (%, LHV basis, all electrical energy converted to heat)	53	57	57

[a] Some model adjustments were made to 2018 SOT experimental results to allow overall mass and atomic balance closures. [b] 2015 design report (https://www.nrel.gov/docs/fy15osti/62455.pdf) updated to 2016$ and 21% tax rate (with clean pine feedstock). [c] Fast pyrolysis intermediate yields were maintained from the 2015 design report. [d] Negligible natural gas. NG = natural gas; GGE = gallon gasoline equivalent; LHV = lower heating value.

Table ES-2. Economic Summary (Modeled) for the 2018 SOT

**Process Engineering Analysis for Hydrocarbon Fuel Production
via Ex Situ Upgrading of Fast Pyrolysis Vapors**

Potential Research-Driven Pathway for Cost-Competitiveness by 2022
2,000 Dry Metric Tonnes Biomass per Day
All Values in 2016$

Minimum Fuel Selling Price (MFSP)	**$3.36 /Gallon Gasoline Blendstock**
	$3.82 /Gallon Diesel Blendstock
	$3.50 /Gallon Gasoline Equivalent (GGE)

Gasoline Blendstock Production	26.0 MM Gal per Year	35.9 Gal per Dry US Ton Feedstock
Diesel Blendstock Production	24.8 MM Gal per Year	34.3 Gal per Dry US Ton Feedstock
Total Gasoline Equivalent Production	52.0 MM GGE per Year	71.7 GGE per Dry US Ton Feedstock

Delivered Feedstock Cost	$87.82 per Dry U.S. Ton (Includes Capital Up to Throat of Pyrolyzer)
Internal Rate of Return (After-Tax)	10.0%
Equity Percent of Total Investment	40.0%
On-Stream Factor	90.0%

Capital Costs		Operating Costs (¢ / GGE Product)	
100: Feedstock (Additional Dryer & Blower Only)	$520,000	Feedstock	121.5
200: Fast Pyrolysis & Vapor Upgrading	$116,250,000	Natural Gas	0.3
300: Pyrolysis Vapor Quench	$23,570,000	Catalysts	15.9
400: Hydroprocessing & Separation	$41,380,000	Sand	0.5
500: Hydrogen Plant	$68,620,000	Other Raw Materials	1.3
600: Steam System & Power Generation	$49,770,000	Waste Disposal	2.0
700: Cooling Water & Other Utilities	$8,960,000	Purchased Electricity	-
800: Water Management	$25,930,000	Fixed Costs	54.6
Total Installed Equipment Cost (TIC)	$334,990,000	Electricity Coproduct Credit	(1.7)
		Capital Depreciation	56.8
Land (115 Acres at $14000 per Acre)	$1,600,000	Average Income Tax	12.0
Site Development	$18,170,000	Average Return on Investment	87.2
(% of ISBL)	10.0%		
Indirect Costs & Project Contingency	$221,170,000	Operating Costs ($ / Year)	
(% of TIC)	66.0%	Feedstock	$63,100,000
		Natural Gas	$100,000
Fixed Capital Investment (FCI)	$589,780,000	Catalysts	$8,290,000
Working Capital	$29,490,000	Sand	$300,000
Total Capital Investment (TCI)	$619,260,000	Other Raw Materials	$680,000
		Waste Disposal	$1,060,000
Loan Interest Rate	8.0%	Purchased Electricity	$0
Loan Term (Years)	10	Fixed Costs	$28,350,000
		Electricity Coproduct Credit	-$910,000
Total Installed Equipment Cost per Annual GGE	$6.45	Capital Depreciation	$29,490,000
Fixed Capital Investment per Annual GGE	$11.35	Average Income Tax	$6,230,000
		Average Return on Investment	$45,310,000
Plant Operating Hours per Year	7884		
On-Stream Percentage	90.0%	Total Plant Electricity Usage (kW)	44,429
		Electricity Produced on Site (kW)	46,399
Maximum Yield Based on HHV of Feedstock + Natural Gas		Electricity Purchased from Grid (kW)	0
Theoretical GGE Production (MM GGE / Year)	105.1	Electricity Sold to Grid (kW)	1,970
Theoretical Yield (GGE / Dry Ton)	145.1		
Current Yield (Actual / Theoretical)	49.5%	Plant Electricity Use (kWh /GGE)	6.74

Overall Plant Efficiency - HHV %	51.6%	Specific Operating Conditions		
Overall Plant Efficiency - LHV %	52.1%	Feed Rate	Dry Tonnes / Day	2,000
			Dry Tons / Day	2,205
		Feedstock Cost	$/Dry Ton	$87.82
Version:	PyVPU-v218h ES FixedBed-v49-r046-AP10-FY18SOT (2016$)-V09d		$/Moisture+Ash Free Ton	$88.64

Table ES-3. Economic Summary (Modeled) for the Updated 2022 Projection

**Process Engineering Analysis for Hydrocarbon Fuel Production
via Ex Situ Upgrading of Fast Pyrolysis Vapors**

Potential Research-Driven Pathway for Cost-Competitiveness by 2022
2,000 Dry Metric Tonnes Biomass per Day
All Values in 2016$

Minimum Fuel Selling Price (MFSP)	**$2.83 /Gallon Gasoline Blendstock**
	$3.20 /Gallon Diesel Blendstock
	$2.93 /Gallon Gasoline Equivalent (GGE)

Gasoline Blendstock Production	27.6 MM Gal per Year	38.0 Gal per Dry US Ton Feedstock
Diesel Blendstock Production	26.9 MM Gal per Year	37.1 Gal per Dry US Ton Feedstock
Total Gasoline Equivalent Production	56.0 MM GGE per Year	77.3 GGE per Dry US Ton Feedstock

Delivered Feedstock Cost	$70.31 per Dry U.S. Ton (Includes Capital Up to Throat of Pyrolyzer)
Internal Rate of Return (After-Tax)	10.0%
Equity Percent of Total Investment	40.0%
On-Stream Factor	90.0%

Capital Costs		Operating Costs (¢ / GGE Product)	
100: Feedstock (Additional Dryer & Blower Only)	$460,000	Feedstock	90.3
200: Fast Pyrolysis & Vapor Upgrading	$117,270,000	Natural Gas	0.2
300: Pyrolysis Vapor Quench	$24,150,000	Catalysts	12.5
400: Hydroprocessing & Separation	$39,810,000	Sand	0.5
500: Hydrogen Plant	$65,240,000	Other Raw Materials	1.2
600: Steam System & Power Generation	$48,410,000	Waste Disposal	1.9
700: Cooling Water & Other Utilities	$8,850,000	Purchased Electricity	-
800: Water Management	$19,810,000	Fixed Costs	49.5
Total Installed Equipment Cost (TIC)	$323,990,000	Electricity Coproduct Credit	(2.5)
		Capital Depreciation	51.1
Land (115 Acres at $14000 per Acre)	$1,600,000	Average Income Tax	10.7
Site Development	$18,170,000	Average Return on Investment	77.9
(% of ISBL)	10.0%		
Indirect Costs & Project Contingency	$214,560,000	Operating Costs ($ / Year)	
(% of TIC)	66.2%	Feedstock	$50,520,000
		Natural Gas	$100,000
Fixed Capital Investment (FCI)	$572,150,000	Catalysts	$6,990,000
Working Capital	$28,610,000	Sand	$300,000
Total Capital Investment (TCI)	$600,760,000	Other Raw Materials	$700,000
		Waste Disposal	$1,070,000
Loan Interest Rate	8.0%	Purchased Electricity	$0
Loan Term (Years)	10	Fixed Costs	$27,700,000
		Electricity Coproduct Credit	-$1,410,000
Total Installed Equipment Cost per Annual GGE	$5.79	Capital Depreciation	$28,610,000
Fixed Capital Investment per Annual GGE	$10.22	Average Income Tax	$6,010,000
		Average Return on Investment	$43,590,000
Plant Operating Hours per Year	7884		
On-Stream Percentage	90.0%	Total Plant Electricity Usage (kW)	41,944
		Electricity Produced on Site (kW)	45,007
Maximum Yield Based on HHV of Feedstock + Natural Gas		Electricity Purchased from Grid (kW)	0
Theoretical GGE Production (MM GGE / Year)	104.0	Electricity Sold to Grid (kW)	3,063
Theoretical Yield (GGE / Dry Ton)	143.6		
Current Yield (Actual / Theoretical)	53.8%	Plant Electricity Use (kWh /GGE)	5.91
Overall Plant Efficiency - HHV %	55.7%	Specific Operating Conditions	
Overall Plant Efficiency - LHV %	56.1%	Feed Rate Dry Tonnes / Day	2,000
		Dry Tons / Day	2,205
		Feedstock Cost $/Dry Ton	$70.31
Version: PyVPU-v218h ES FixedBed-v49-r046-AP10-2022 Target (2016$)-V06d		$/Moisture+Ash Free Ton	$70.96

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1 Introduction

The 2015 catalytic fast pyrolysis (CFP) design report [1] detailed (1) *in situ* and (2) *ex situ* catalytic fast pyrolysis as two potential research options for the conversion of biomass to liquid transportation fuels. Catalysts are included within the fast pyrolysis reactor in an *in situ* process. On the other hand, biomass-derived solid material (char and inorganic matter) are separated after fast pyrolysis in an *ex situ* process; removal of all solids from pyrolysis vapors is desirable before catalytic upgrading in an *ex situ* reactor. The 2015 design report helped outline the basis for technical improvements necessary for future economic viability; associated modeled costs were presented to help understand the potential value of the research improvements [2]. Experimental results were used to understand the 2014 state of technology (SOT) for both the *in situ* and *ex situ* pathways. Subsequent research under the U.S. Department of Energy's Bioenergy Technologies Office focused on the *ex situ* pathway because this pathway could expedite the scientific understanding of catalytic chemistry and accelerate related development by eliminating the overwhelming impact of biomass-derived solid material on catalysts in an *in situ* environment.

Ex situ research under this project was initiated on fluidized systems with zeolite-based (primarily ZSM-5 and metal impregnated ZSM-5) catalysts. This was based on historic precedence of experimental work documenting some of the best yields using ZSM-5 catalysts for catalytic fast pyrolysis [3]. Circulating fluidized bed systems with a combustor for coke burn-off are ideal for ZSM-5 catalysts. To broaden the research and explore other bifunctional catalyst options [4,5], such as those with noble metals in their formulations as one such option, an analogous fixed bed approach was proposed, and its feasibility was analyzed by Dutta et al. [6]. Consequent catalyst research and associated experimental performance showed significant yield improvements using a Pt/TiO_2 catalyst [7]. Hence, the current process configuration for the 2018 SOT and projections to 2022 use the Pt/TiO_2 catalyst in a fixed bed as the base configuration.

1.1 Techno-Economic Analysis Approach

The techno-economic analysis (TEA) approach for this work is similar to those detailed previously [1,6]. Overviews of process and economic assumptions and methods are provided below. Further details are available in the previous publications.

1.1.1 Financial Assumptions

The modeled projections in this report are based on the technology being implemented in a mature or n^{th} plant; additional costs associated with pioneer plants are thus not included because the purpose of this TEA is to understand the potential impact and relevance of the research in the context of future industrial implementation. A consistent set of assumptions are used for all SOT and projections. Key assumptions are listed in Table 1.

Table 1. Summary of Financial Assumptions for Techno-Economic Analysis

Description of Assumption	Assumed Value
Cost year	2016
Internal rate of return on equity	10%
Plant financing by equity/debt	40%/60% of total capital investment
Plant life	30 years
Income tax rate	21%
Interest rate for debt financing	8.0% annually
Term for debt financing	10 years
Working capital cost	5.0% of fixed capital investment (excluding land purchase cost)
Depreciation schedule	7-year MACRS schedule [8]
Steam plant depreciation	20-year MACRS schedule [8]
Construction period (spending schedule)	3 years (8% Y1, 60% Y2, 32% Y3)
Plant salvage value	No value
Startup time	6 months
Revenue and costs during startup	Revenue = 50% of normal Variable costs = 75% of normal Fixed costs = 100% of normal
Onstream percentage after startup	90% (7,884 operating hours per year)

MACRS = modified accelerated cost recovery system

There were two significant changes to the financial assumptions compared to the previous publications [1,6]: (1) 21% tax rate (versus a previous 35% tax rate) and (2) 2016-dollars cost basis was used.

1.1.2 Estimation of Capital and Operating Costs

Detailed capital costs of individual equipment and their sources were listed in the 2015 design report [1] and additional fixed bed equipment costs were presented in the subsequent analysis for fixed bed systems [6]. Note that the fixed bed system cost for this analysis was based on the #1 upstream reactor in Dutta et al. [6], at approximately $2.5 MM base cost per 50% capacity reactor in 2013 dollars, a scaling exponent of 0.7, and an installation factor of 1.62.

Equipment costs were scaled based on process flows in the Aspen Plus process model using a scaling exponent:

$$Scaled\ Equipment\ Cost = Base\ Equipment\ Cost \left(\frac{Scaled\ Capacity}{Base\ Capacity}\right)^n$$

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The scaling exponent, n, is typically in the range of 0.6 to 0.7 for process equipment. However, it varies with equipment type, base size, and with other factors that affect scalability. Scaling factors are documented in Appendix B of the 2015 design report [1].

Total installed cost (TIC) of the equipment, which includes associated piping, instrumentation and controls, electrical systems, buildings, yard improvements, and direct labor, were derived from the equipment cost by applying an installation factor ($f_{installation}$).

$$Total\ Installed\ Cost\ (TIC) = f_{installation} * Total\ Purchased\ Equipment\ Cost\ (TPEC)$$

Installation factors are also documented in Appendix B of the 2015 design report.

Costs were converted to 2016 dollars using:

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

Operating costs were adjusted using the Producer Price Index for Chemical Manufacturing [9] and capital costs were adjusted using the *Chemical Engineering's* Plant Cost Index [10].

The total capital investment (TCI) was derived from the TIC in 2016 dollars after applying additional factors for overhead and contingency.

1.1.3 *Minimum Fuel Selling Price*

The TCI along with plant operating costs were used for a discounted cash flow analysis. Those costs along with the gallons gasoline equivalent (GGE) of total fuel blendstock product were used to derive the minimum fuel selling price (MFSP) in $/GGE.

1.1.4 *The Process Model*

The process was modeled in Aspen Plus with detailed a detailed accounting of all mass and energy flows. Details about the Aspen Plus [11] process model for *ex situ* CFP were documented in the 2015 design report [1] and the subsequent fixed bed publication [6]. The base models from the previous work were maintained for this analysis. Process assumption updates and other key aspects are described in the following sections.

2 Plant Design Basis

2.1 Feedstock Specifications and Plant Size

Feedstock information for this process was provided by Idaho National Laboratory (INL). Two different feedstocks were used for the 2018 SOT and the 2022 projection. The plant size was maintained at 2,000 dry metric tons per day.

The 2018 SOT feedstock was based on clean pine at $87.82/dry U.S. ton in 2016 dollars. It is projected that the process will be capable of using lower cost feedstock by 2022. The 2022 feedstock cost in this analysis is based on a blend of 75% air-classified forest residues and 25% clean pine, with an ash content of 0.51%; the modeled cost for this feedstock is $70.31/dry U.S. ton in 2016 dollars. Given the low ash and the specified moisture content in both the 2018 and 2022 feedstocks, the specification assumption in the process model was unaltered from the 2015 design report [1], with an elemental analysis of: C:50.94%, H:6.04%, N:0.17%, S:0.03%, O:41.90%, Ash:0.92% on a dry basis, and 10% moisture at the plant gate.

The final 2022 feedstock will need to be determined via iterative research between the National Renewable Energy Laboratory and INL, with an objective to find an optimal balance between feedstock quality (affecting CFP yields) and feedstock cost.

2.2 Process Overview
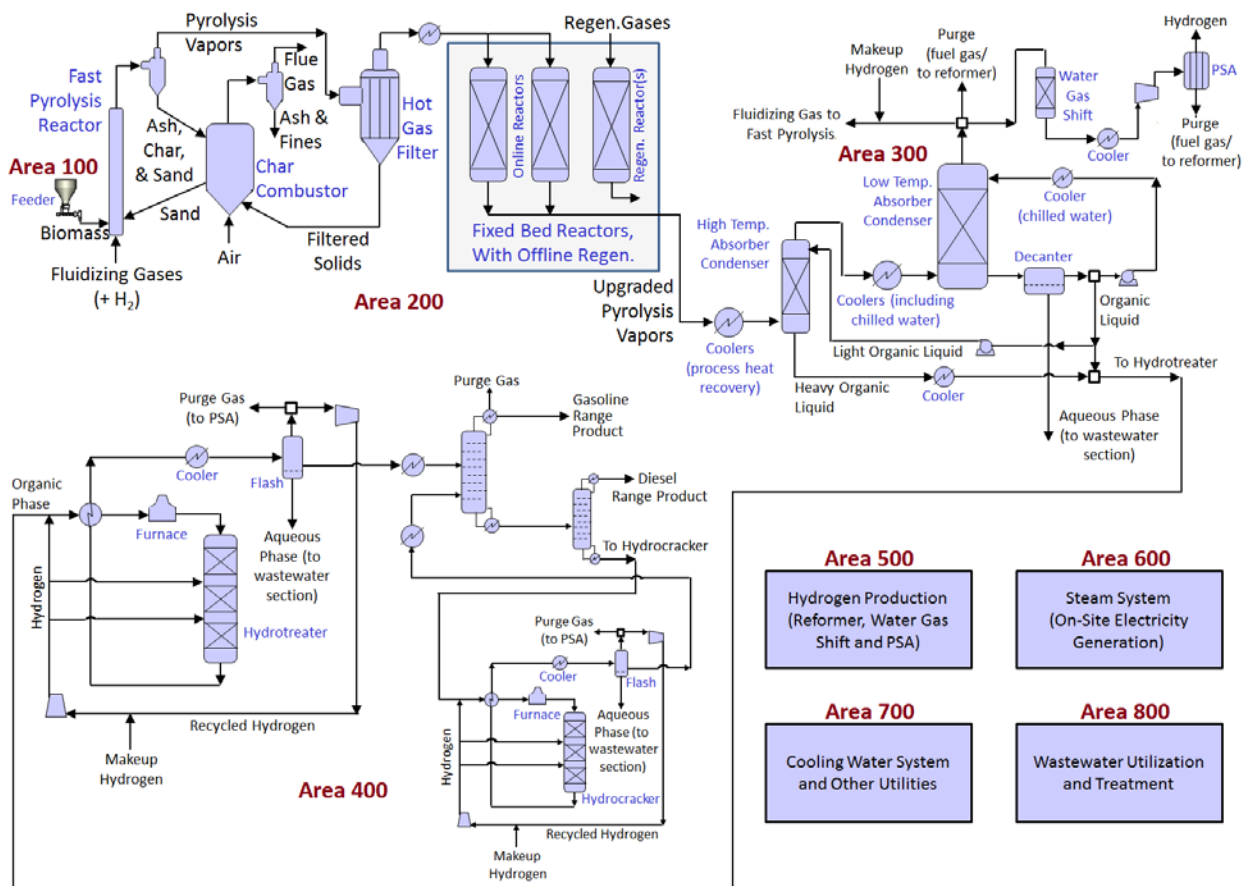
Figure 1. Simplified process flow diagram for fixed bed e*x situ* catalytic fast pyrolysis

A block flow diagram for the fixed bed *ex situ* catalytic fast pyrolysis process is shown in Figure 1. The design includes eight process areas, with four core operations:

- **A100:** Feedstock Handling (most of the processing occurs off-site and the TEA accounts for the feedstock delivered to the throat of the reactor through a cumulative cost)
- **A200:** Fast Pyrolysis, Hot Gas Filtration, and *Ex Situ* Catalytic Vapor Upgrading
- **A300:** CFP Product Condensation (with the separation of the organic liquid CFP oil from an aqueous wastewater stream, and use of separated permanent gases in the process)
- **A400:** CFP Oil Hydrotreating, Hydrocracking, and Product Distillation.

Supporting operations include:

- **A500:** Hydrogen Production (from process off-gases)
- **A600:** Steam System and Electricity Generation (from available excess heat)
- **A700:** Cooling Water and Other Utilities
- **A800:** Wastewater Utilization and Treatment (regenerative thermal oxidizer used to combust the organic content in the wastewater).

Further descriptions are included in the following section.

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3 Process Design

3.1 Area 100: Feed Handling

As mentioned in Section 2.1, two different feedstock materials were used in the 2018 and 2022 TEA models; both feedstocks had low ash (<1%) and 10% moisture, and the modeled feedstock composition was thus unaltered in both 2018 and 2022 process models presented in this report. All feedstock growth, handling, and processing costs are included in the cumulative feedstock costs presented below. A nominal feedstock size of 2 mm is specified for this process and necessary grinding costs are included in INL's feedstock cost [12]. The only minor feedstock handling related cost added to the plant equipment is a cross-flow dryer for warming the feedstock prior to feeding to the fast pyrolysis reactor.

Currently, it is estimated that there are 21,218,792 dry U.S. tons of pine feedstocks available nationally; 11,804,620 dry U.S. tons are planted pine and 9,414,172 dry U.S. tons are pine forest residues. The amount of material that can be aggregated at a cost of approximately $87/dry U.S. ton to the reactor throat can support 8 biorefineries of 2,000 dry metric tons per day, given no competition for the resource [12].

The cost summary for the 2018 clean pine feedstock is presented in Table 2.

Table 2. INL Modeled Feedstock Cost for Clean Pine in 2016$ (used for 2018 SOT)

Cost Summary ($/Dry U.S. Ton) (2016$)	
	2018 SOT
Grower Payment	$9.48
Harvest and Collection	$9.87
Field-Side Preprocessing	$2.82
Transportation	$31.56
Preprocessing	$27.14
Storage	$0.86
Handling	$2.62
Preprocessing Construction	$3.47
Quality Dockage	$0.00
Grand Total	**$87.82**

An alternate feedstock using a mix of clean pine and forest residues was modeled. This scenario estimates a total of 35.9 million U.S. tons available nationwide. If we consider only the volume that is aggregable within a 725,000 dry U.S. ton supply shed (necessary for a 2,000-dry-metric-tons-per-day plant operating at a 90% onstream factor) and ignore "stranded" resources, there is enough forest residue to supply 17 biorefineries at a size of 2,000 dry metric tons per day. However, it is currently believed that the quality of forest residue-based feedstock is insufficient for use in fast pyrolysis applications, with both overall ash and alkali and alkaline earth metal (AAEM) concentrations being elevated compared to clean pine. The forest residues can be made

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to compositionally resemble clean pine by undergoing a process of air classification and leaching. The air classification and leaching process is detailed in Hu et al. [13]. This process results in forest residues that have an overall ash content of <0.9 wt% and AAEM less than 1,300 ppm, which is similar to average values for clean pine. The projected feedstock for 2022 is a blend of clean pine and air-classified and leached forest residues that meet the quality specifications for catalytic fast pyrolysis. Blending 25% clean pine with 75% forest residue resulted in the lowest cost material with an ash content of 0.51 wt% and AAEM <1,300 ppm. The cost summary for the 2022 feedstock with 75% air-classified forest residues and 25% clean pine is presented in Table 3. Additional feedstock options will be studied for the *ex situ* CFP process in the coming years, and higher ash material will be considered if they can be justified through TEA based on experimental results.

Table 3. INL Modeled Feedstock Cost for 75% Forest Residues Plus 25% Clean Pine in 2016$ (used for the 2022 projection model)

Cost Summary ($/Dry U.S. Ton) (2016$)	
	2022 Projection
Grower Payment	$7.64
Harvest and Collection	$2.47
Field-Side Preprocessing	$9.81
Transportation	$13.32
Preprocessing	$31.12
Storage	$0.58
Handling	$2.09
Preprocessing Construction	$3.28
Quality Dockage	$0.00
Grand Total	**$70.31**

3.2 Area 200: Fast Pyrolysis and Catalytic Vapor Upgrading

The process model for Area 200 includes a circulating fluidized bed fast pyrolysis reactor. The dual bed reactor system includes a riser reactor for fast pyrolysis of biomass at approximately 500°C (932°F), with short biomass residence times of approximately 2 seconds in the riser, and a char combustor for providing heat to the endothermic fast pyrolysis reactions; circulating sand is heated in the char combustor and sent to the riser reactor where it heats the biomass to pyrolysis temperatures. The solids (char and mineral matter) from fast pyrolysis are removed from the hot vapors by cyclones. An additional hot gas filter (HGF) is also included to remove any residual solids. This HGF is necessary because of the downstream fixed bed *ex situ* catalytic vapor upgrading reactor that can easily plug from any residual solids. The catalytic fixed bed reactor system includes a Pt/TiO_2 catalyst with 0.5 wt% Pt loading. A 2-year catalyst lifetime is assumed in the model, along with a 70% cost recovery at the end of 2 years. A catalyst cost model, called CatCost, [14] developed under the Chemical Catalysis for Bioenergy Consortium [15] was used to estimate the cost of the Pt/TiO_2 catalyst. Note that the 2015 design report [1] included a circulating fluidized bed *ex situ* reactor design with zeolite catalyst.

The CFP bench-scale experimental setup and analytical methods used to generate experimental results for the 2018 SOT are described by Griffin et al. [7]. As a brief overview, a 2-inch fluidized pyrolyzer was followed by an HGF and a fixed bed Pt/TiO_2 vapor upgrading reactor. For the 2018 experiments, the reactor was operated with a continuous biomass feed until the catalyst deactivated. Approximately 100 g of catalyst was loaded in the fixed bed reactor, and 150 g/h of biomass was fed to the pyrolyzer. The system was operated at near atmospheric pressure with an 85% H_2/15% N_2 (by volume) gas flow at the rate of 17.6 standard liters per minute. Upon deactivation, the catalyst was regenerated using an air and nitrogen mixture and controlling the flows and inlet temperature so that the outlet temperature remained <480°C (896°F). Catalyst activity remained fairly stable over 90 regeneration cycles. The major developments in 2018 related to the fixed bed *ex situ* reactor system included the reduction in the catalyst Pt loading from 2% in 2017 to 0.5% in 2018. In addition, more efficient regeneration allowed a model assumption of two online and three regenerating reactors in 2018 versus two online and five regenerating reactors in 2017 (each reactor has a capacity to handle 50% of the vapor stream). Conservative assumptions were made when closing the carbon balance for the process model (the model needs a 100% closure): the carbon balance gap was not prorated among the measured quantities in each phase (solid char and coke, light gases, and organic and aqueous liquid phases) and the organic liquid carbon yield was kept closer to the experimental values rather than the higher value that would otherwise be obtained from a prorated distribution. The modeled yield values are shown in Table ES-1. Additional analytical equipment has been installed to further improve the experimental carbon balance closures in future years.

There are differences between the experimental system and the process model assumptions, primarily because of safety concerns with operating the available experimental equipment at elevated pressures. The modeled hydrogen partial pressure was ˜5 bar, compared to the experimental partial pressure of ˜0.85 bar; the experimental results at a lower pressure suggest that the modeled hydrogen partial pressure can be reduced, which will allow further reduction in the modeled cost. Note that the reactor in the process model is at 8 bar total pressure to reduce reactor system (including hot gas filter) size and cost. The impact of maintaining a lower hydrogen partial pressure (˜0.85 bar) at a higher system pressure (which will also result in a higher proportion of diluting gases) needs to be experimentally quantified before modifying future model assumptions.

The process model uses a weight hourly space velocity (WHSV) of 5 h^{-1} based on total vapor and gas mass flow into the reactor, with an added 70% overdesign [6]. This WHSV corresponds to a process model WHSV of approximately 3 h^{-1} on a dry biomass flow rate basis. Increasing online times for feasible industrial operations will be part of future research efforts.

The projected improvements by 2022 using the fixed bed *ex situ* vapor upgrading reactor system are also shown in Table ES-1, alongside the earlier design report projections [1] based on a fluidized *ex situ* system. The overall yields are comparable with the initial 2015 design. Additional significant cost reduction is projected based on improving the CFP system to handle lower cost (but poorer quality) feedstocks.

3.3 Area 300: CFP Product Condensation

The process design for this section was maintained from the 2015 design report [1]. The system consists of two direct quench absorber/condensers. The upgraded vapors from the *ex situ* reactors

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are initially cooled via indirect heat exchange up to the modeled dew point of the vapor stream. A heavy organic liquid is then condensed in the first absorber/condenser; the light organic liquid product from the second condenser is used as the quench liquid. The uncondensed light vapors from the first condenser, as well as the vaporized quench liquid are then sent through heat exchangers. The partially condensed vapors enter the second absorber/condenser column for a final quench using a stream of recycled (and cooled) light organic liquid. The bottom product of the second condenser is separated into an aqueous waste stream (sent to Area 800) and an organic product. As previously mentioned, part of this light organic liquid product is also recycled for use as a quench liquid for both the absorber/condensers. In this design the heavy organic liquid from the first condenser and light organic liquid from the second condenser are mixed and sent to Area 400 for hydroprocessing.

3.4 Area 400: CFP Oil Hydroprocessing

The 2015 design report was written based on a premise that a single reactor system can handle the hydrotreating of the CFP organic liquid. Experiments in 2017 and 2018 have proven that this is a valid assumption and it is possible to get to less than 1% oxygen content in the CFP oil after a single hydrotreating step. This is made possible by *ex situ* vapor upgrading that eliminates almost all highly oxygenated and reactive species such as sugars. Although aldehydes are also reactive, they are less prevalent in the CFP oil compared to ketones; ketones are less problematic. Other oxygenates present, such as phenolics, are significantly less prone to reacting and fouling hydrotreating catalysts. In fact, hydrotreating experiments at Pacific Northwest National Laboratory in 2018 showed that catalyst performance was maintained during the entire duration of 120 hours until all the CFP oil feed was exhausted; catalyst lifetime assumptions require further verification with longer experiments. The normalized carbon efficiency during hydrotreating for the 2018 SOT was 91%. An 89% carbon efficiency was assumed in the model, allowing for additional losses, including for any subsequent hydrocracking of the remaining heavy fraction. Hydrocracking experiments have not been conducted thus far. Some hydrocracking experiments are expected to be included in the coming years and the model will be recalibrated once experimental results are available. Note that the carbon efficiencies for hydrotreating were previously estimated based on a linear interpolation [1]; those estimates have been replaced by estimates based on experimental results. A hydrotreating WHSV of 0.47 h^{-1} [1] was used to match the experimental liquid hourly space velocity of 0.2 h^{-1}, after factoring densities of the catalyst and the CFP oil. A commercial Ni-Mo sulfide catalyst was used for hydrotreating. Additional information related to the experimental setup is available in a recent publication [7].

3.5 Area 500: Hydrogen Production

Hydrogen demands in the process were met (in the process models) without importing additional natural gas. Off-gases, primarily from catalytic fast pyrolysis, and other parts of the process, were processed in a steam reformer to produce hydrogen, and purified hydrogen was produced using pressure swing adsorption (PSA) units. Process design details for Area 500 are consistent with the 2015 design report [1].

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3.6 Area 600: Steam System and Electricity Generation

Heat available in the modeled process was used to generate electricity. Excess electricity, after meeting process demands, was sold to the grid. Process design details are consistent with the 2015 design report [1].

3.7 Area 700: Cooling Water and Other Utilities

Air cooling was the major cooling method in the process design when in-process heat recovery was not feasible (reflected by the pinch analysis diagrams in Section 3.9). Process heat exchange and air-cooling costs are included with the costs of the respective process areas. Water cooling was used primarily for cooling process streams below 140°F; process stream temperatures of 110°F were achieved after water cooling. Chilled water was used for cooling below 110°F. Process design details for Area 700 are consistent with the 2015 design report [1].

3.8 Area 800: Wastewater Utilization and Treatment

Organic species in the wastewater stream were oxidized in a regenerative thermal oxidizer (RTO) to allow discharge of the stream contents in an environmentally acceptable manner. Other methods of aqueous carbon utilization, both biological [16] and catalytic [17], are being explored but not included in the base case design at this time. Overall water balance for the system is shown in Table 4.

Table 4. Water Balance for the 2018 SOT and 2022 Projection Models

Area	Stream	Source/Destination	*2018*	*2022*	
A100	Scrubber Water	Input	30,000	30,000	lb/h
	To Flue Gas	Loss	10,784	5,354	lb/h
	Scrubber Blowdown	To Point Treatment	19,216	24,646	lb/h
	COD		*0*	*0*	*lb/h*
A300	Aqueous Phase	To Boiler A800	55,602	58,147	lb/h
	COD		*12,056*	*7,598*	*lb/h*
	WGS Steam Demand	From Boiler A800	40,989	40,206	lb/h
	PSA Water Knockout	To Point Treatment	24,634	22,361	lb/h
	COD		*1,003*	*583*	*lb/h*
A400	Hydrotreating Aq Phase	To Full Treatment	10,010	9,363	lb/h
	COD		*1*	*3*	*lb/h*
	Hydrocracking Aq Phase	To Full Treatment	171	287	lb/h
	COD		*0*	*0*	*lb/h*
A500	Reformer Steam Demand	From Boiler A800	31,228	29,972	lb/h
	PSA Water Knockout	To Boiler A800	10,236	9,091	lb/h
	COD		*0*	*0*	*lb/h*
A600	Makeup	Input	6,788	6,584	lb/h
	Makeup Chemicals (In Model)	Input	2	2	lb/h
	Steam Drum Knockout	To Cooling Tower Basin	6,790	6,586	lb/h
A700	Steam Drum Knockout	From A600	6,790	6,586	lb/h
	Makeup	Input	52,301	60,163	lb/h
	Blowdown	To Point Treatment	10,786	12,180	lb/h
	Drift	Loss	5,163	5,851	lb/h
	Evaporation	Loss	43,143	48,720	lb/h
A800	*Boiler*				
	BFW Makeup		29,814	19,984	lb/h
	Total Boiler Feed	To Boiler A800	85,416	78,131	lb/h
	COD		*12,056*	*7,598*	*lb/h*
	Process Steam Generated	To SMR+WGS	72,217	70,179	lb/h
	Remaining to RTO	Loss	13,199	7,952	lb/h
	COD		*0*	*0*	*lb/h*
	Point Treatment				
	Total Water for Treatment	75% Recovered	64,871	68,279	lb/h
	Water Recycled	Offsets Inputs	48,654	51,209	lb/h
	Remainder	To Full Treatment	16,218	17,070	lb/h
	Full Treatment				
	Total	Loss	26,398	26,720	lb/h
	COD		*1*	*3*	*lb/h*
	Total Water Makeup		118,905	116,734	lb/h
	Net Water Makeup		70,252	65,525	lb/h
	Net Water Makeup (Annual)		66.6	62.1	MMgal/y
	Water Usage		1.28	1.11	gal/GGE
	Liq Water Emission		26,398	26,720	lb/h
	Liq Water Emission (Annual)		25.0	25.3	MMgal/y

COD = chemical oxygen demand; WGS = water gas shift; BFW = boiler feedwater; SMR = steam methane reforming

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3.9 Pinch Analysis and Process Heat Exchange Cost

A detailed heat exchange network was developed for the *ex situ* process and documented in the 2015 design report [1]. A pinch analysis was done for process heat exchange in subsequent analyses, including the 2018 SOT and 2022 projections presented in this report. The results are shown in Figure 2. The 2018 SOT case had upper and lower pinch temperatures of 308.5°F and 285.0°F, with a ΔT_{min} of 23.5°F. The 2022 projection had upper and lower pinch temperatures of 326.0°F and 308.8°F, with a ΔT_{min} of 17.1°F. This analysis confirms that a feasible network is possible based on the process conditions. In addition, since the thermal profile of this system is similar to that shown for the earlier *ex situ* process [1], a cost estimate for the heat exchange network for the new models can be derived by scaling the costs from the previously detailed analysis using the total process heat exchange duty (as the scaling basis).
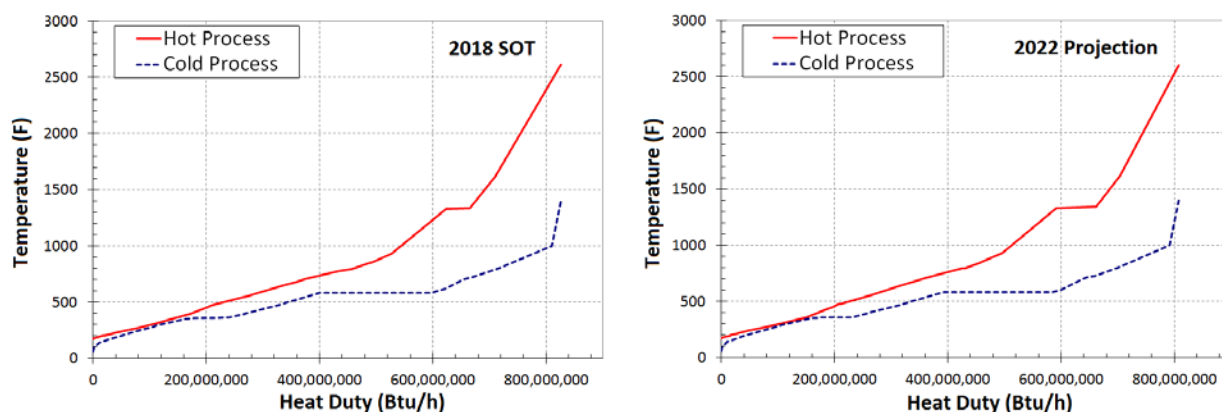
Figure 2. Pinch analysis composite curves for the 2018 SOT and 2022 projection models

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3.10 Energy Balance
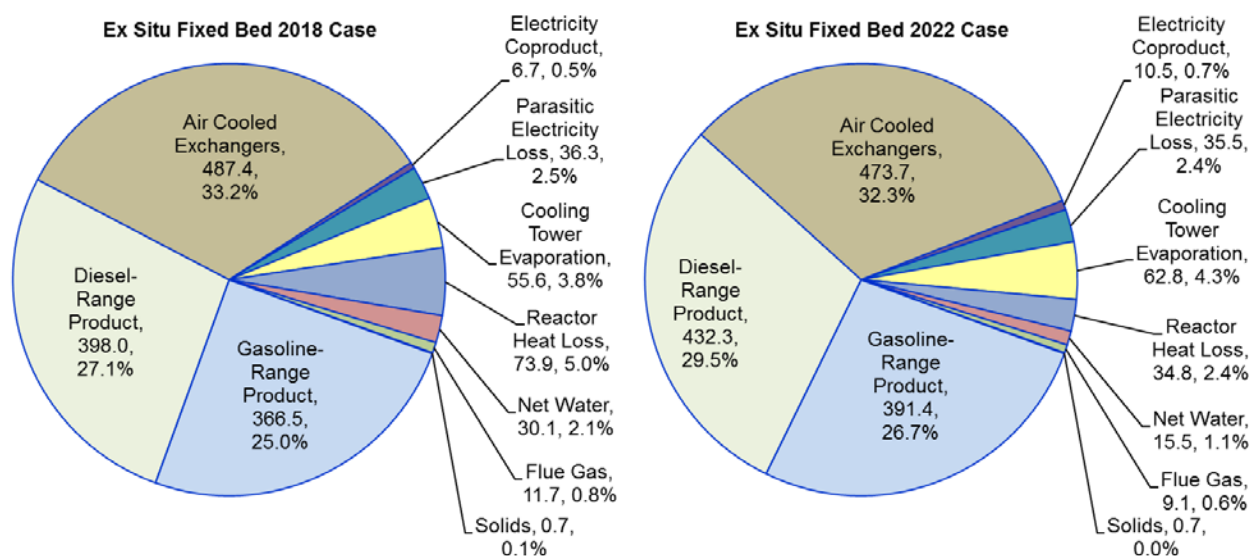
Figure 3. Overall energy balance (based on total energy input: LHV of wood + natural gas, at 1,467 MMBtu/h). Energy flows values are in MMBtu/h along with % distribution (values rounded to nearest tenth). Natural gas LHV was negligible (0.2% of wood) in both the 2018 and 2022 cases.

Energy balance diagrams based on the 2018 SOT and 2022 projection process models are shown in Figure 3. Thermal dissipation accounts for most of the losses. While this may be optimized to some extent, such losses are to be expected in any high-temperature conversion process with intermediate heating and cooling, such as in this process.

4 Process Economics

Capital and operating costs are listed in this section. Note that most of the information presented here is based on previously documented details [1,6]. Costs were updated to a 2016-dollars basis and scaled based on specific stream flows in the process models (as stated in Section 1.1.2).

4.1 Total Capital Investment

Installed capital costs are shown in Table 5, with per area total purchased equipment cost (TPEC), installation factors, and TIC.

Table 5. Total Installed Equipment Costs for the 2018 SOT and 2022 Projection Models

Area	Process Description	2018 SOT			2022 Projection		
		TPEC (MM$)	$f_{install}$	TIC (MM$)	TPEC (MM$)	$f_{install}$	TIC (MM$)
100	Feed handling and drying[a]	0.3	1.96	0.5	0.2	1.96	0.5
200	Fast pyrolysis and vapor upgrading	47.1	2.47	116.2	45.5	2.58	117.3
300	Pyrolysis vapor quench and product recovery	13.2	1.79	23.6	13.4	1.80	24.1
400	Hydroprocessing and product separation	23.9	1.73	41.4	22.6	1.76	39.8
500	Hydrogen plant	35.1	1.96	68.6	33.4	1.95	65.2
600	Steam system and power generation	26.9	1.85	49.8	26.3	1.84	48.4
700	Cooling water and other utilities	4.4	2.02	9.0	4.4	2.01	8.8
800	Wastewater management and recycle	11.1	2.34	25.9	8.4	2.35	19.8
ISBL	**(Areas 100–400)**	**84.4**	**2.15**	**181.7**	**81.8**	**2.22**	**181.7**
OSBL	**(Areas 500–800)**	**77.5**	**1.98**	**153.3**	**72.5**	**1.96**	**142.3**
Total		**161.9**	**2.07**	**335.0**	**154.3**	**2.10**	**324.0**

[a] Most investment costs for feed handling and drying are included in the per-unit woody feedstock price. This cost is for a secondary biomass dryer that serves to recover heat. ISBL = inside battery limits; OSBL = outside battery limits.

The sum of equipment purchases and installation/construction costs is defined as the total direct cost (TDC). Indirect costs, such as project management and engineering, procurement, and construction services, are estimated with factors on the TDC as shown in Table 6.

Table 6. Cost Factors for Indirect Costs

Indirect Costs	% of TDC*
Prorated expenses	10.0
Home office and construction fees	20.0
Field expenses	10.0
Project contingency	10.0
Other costs (startup and permits)	10.0
Total Indirect Costs	**60.0**

* Excluding land purchase cost.

The sum of direct and indirect costs is defined as the fixed capital investment (FCI). The working capital is estimated to be 5% of the FCI. The sum of FCI and working capital is the TCI. Table 7 presents a summary of these capital quantities for the 2018 SOT and 2022 projection.

Table 7. Total Capital Investment Calculations

		2018 SOT	2022 Projection
Total purchased equipment cost (TPEC)		**$161,870,000**	**$154,290,000**
Installation factor		2.070	2.100
Total installed cost (TIC)		**$334,990,000**	**$323,990,000**
Other direct costs			
Land (not depreciated)		$1,610,000	$1,610,000
Warehouse	4.0% of ISBL	$7,270,000	$7,270,000
Site development	10.0% of ISBL	$18,170,000	$18,170,000
Additional piping	4.5% of ISBL	$8,180,000	$8,180,000
Total direct costs (TDC)		**$368,610,000**	**$357,600,000**
Indirect costs	% of TDC (ex land)		
Prorated expenses	10.0%	$36,860,000	$35,760,000
Home office and construction fees	20.0%	$73,720,000	$71,520,000
Field expenses	10.0%	$36,860,000	$35,760,000
Project contingency	10.0%	$36,860,000	$35,760,000
Other costs (startup and permits)	10.0%	$36,860,000	$35,760,000
Total indirect costs	**60.0%**	**$221,170,000**	**$214,560,000**
Fixed capital investment (FCI)		**$589,780,000**	**$572,150,000**
Working capital	5.0% of FCI (ex land)	$29,490,000	$28,610,000
Total capital investment (TCI)		**$619,260,000**	**$600,760,000**
TCI/TPEC		3.826	3.894
FCI Lang Factor = FCI/ISBL TPEC		6.987	6.995
TCI Lang Factor = TCI/ISBL TPEC		7.336	7.345

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4.2 Operating Costs

Variable operating cost assumptions are shown in Table 8. The major addition to the table in the 2015 design report [1] is the Pt/TiO₂ catalyst used in the fixed bed *ex situ* reactor.

Table 8. Variable Operating Cost Assumptions

Variable	Information and Operating Cost (cost year in parentheses)
Fluidized bed media	For fast pyrolysis, the bed medium is sand. Sand price: $45.74/U.S. ton (2011$) [18]; initial fill, then makeup for attrition.
Fixed bed *ex situ* vapor upgrading catalyst	Pt/TiO₂ with 0.5% Pt loading. Catalyst unit cost $92.35/lb (2014$) [14]. A 2-year lifetime is assumed, with 70% cost recovery at the end of 2 years. Quantity of catalyst was determined using a WHSV of 5 h^{-1} and an additional 70% overdesign.
Hydrotreating and hydrocracking catalysts	To determine the amount of catalyst inventory, the hydroprocessors were sized for a WHSV of 0.5 h^{-1} based on the expected hydroprocessing severity. Initial fill is then replaced every 2 years. Price: $20/lb (2011$) based on National Renewable Energy Laboratory calculations using metals pricing and costs for manufacturing processes, and some buffer for modifications.
Steam methane reformer catalysts	Based on a literature value of price per unit hydrogen produced. Price: $7.80/U.S. ton hydrogen (2011$) [19].
Natural gas	Purchased from pipeline for feed to steam methane reformer for hydrogen production. Natural gas has an insignificant cost contribution because of minimal process use. Price: $239/U.S. ton ($5/MMBtu) (2011$).
Solids disposal	Price: $33/tonne (1998$) [20].
Diesel fuel	Usage: 10 gal/h plant-wide use. 2012 price projection: $21.29/MMBtu (2009$) [21] = $2.86/gal at 0.85 specific gravity.
Water makeup	Price: $0.22/tonne (2001$) [22] = $0.20/U.S. ton.
Chemicals	Boiler feedwater chemicals – Price: $75/MM lb blowdown (2014 vendor estimate). Cooling tower chemicals – Price: $33.84/yr per U.S. ton of cooling capacity (2014 vendor estimate). Caustic – Price: $150/dry U.S. ton (2010$) [23].
Wastewater	Most wastewater is cleaned using a reverse osmosis system and recycled. Additional treatment is assumed for the balance. Price: $0.022/gallon (2011$). Based on Humbird et al. [23].

Note: Costs shown were updated to 2016 dollars using the Producer Price Index for chemical manufacturing [9].

Fixed operating costs related to salaries are shown in Table 9. Number of personnel and their salaries were maintained from the 2015 design report [1], and salaries were adjusted using a labor cost index [24].

Table 9. Fixed Operating Costs

Cost Item	Factor	2018 SOT	2022 Projection
Salaries (plus 10% escalation)		$3,444,000	$3,444,000
Benefits and overhead	90.0% of total salaries (after adding 10%)	$3,100,000	$3,100,000
Maintenance	3.0% of fixed capital investment (FCI*)	$17,693,000	$17,165,000
Insurance and taxes	0.7% of fixed capital investment (FCI*)	$4,128,000	$4,005,000
Total fixed operating costs (2016$/year)		**$28,365,000**	**$27,714,000**

* Percentages of FCI exclude land purchase cost.

4.3 Discounted Cash Flow Analysis and the Minimum Fuel Selling Price

Once the capital and operating costs are determined, the GGE of fuel production is used to calculate an MFSP (in $/GGE) using a discounted cash flow rate of return (DCFROR) analysis. Further details are available in the 2015 design report [1].

4.4 Value of Hydrocarbon Fuel Products

The results of the cash flow analyses for the 2018 SOT and the 2022 projection are summarized in Table 10. Gasoline and diesel are normalized by lower heating value to represent a single gasoline-equivalent product and MFSP. To calculate individual selling prices for gasoline and diesel, the MFSP per GGE is ratioed back to these products by lower heating value. The densities of the blendstocks were assumed to be those of U.S. conventional gasoline (2,819 grams/gallon or 6.215 lb/gallon) and diesel fuels (3,167 grams/gallon or 6.982 lb/gallon) [25].

Table 10. Projected Selling Prices of Hydrocarbon Blendstocks

	2018 SOT	2022 Projection
Lower heating value for gasoline-range products (simulation result)	111,269 Btu/gal	112,061 Btu/gal
Lower heating value for diesel-range products (simulation result)	126,432 Btu/gal	126,795 Btu/gal
Calculated gasoline-equivalent MFSP	**$3.50/GGE**	**$2.93/GGE**
Calculated actual MFSP for gasoline-range products	**$3.36/gal**	**$2.83/gal**
Calculated actual MFSP for diesel-range products	**$3.82/gal**	**$3.20/gal**
Gasoline lower heating value for GGE normalization		*116,090 Btu/gal*

5 Process Economics Summary and Sensitivity Analysis

The contribution of individual areas toward the MFSP on a $/GGE basis are shown in Figure 4 (2018 SOT) and Figure 5 (2022 projection).
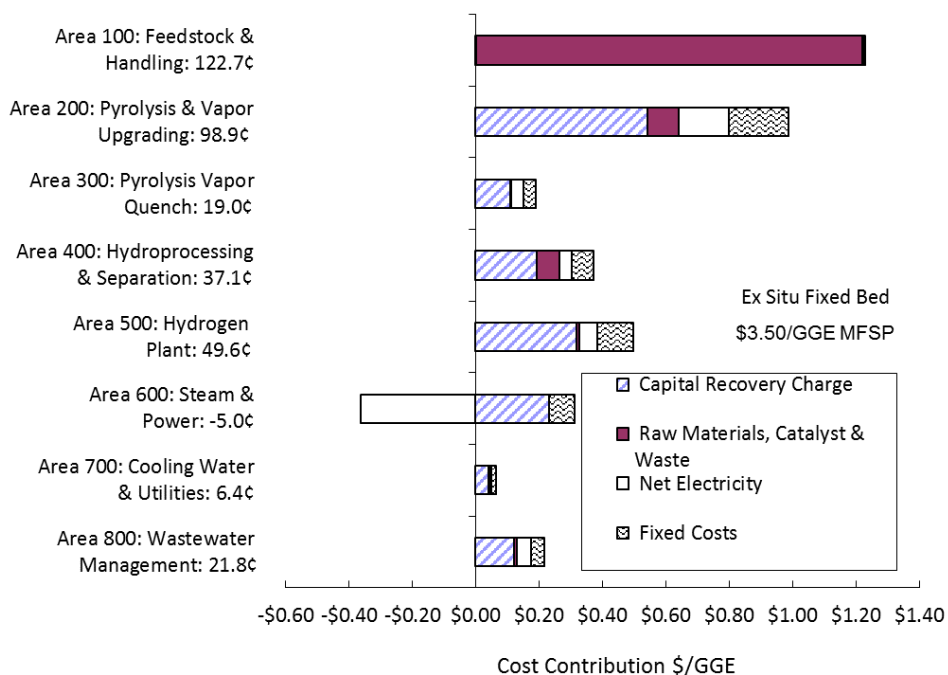
Figure 4. Cost contribution details from each process area for the 2018 SOT
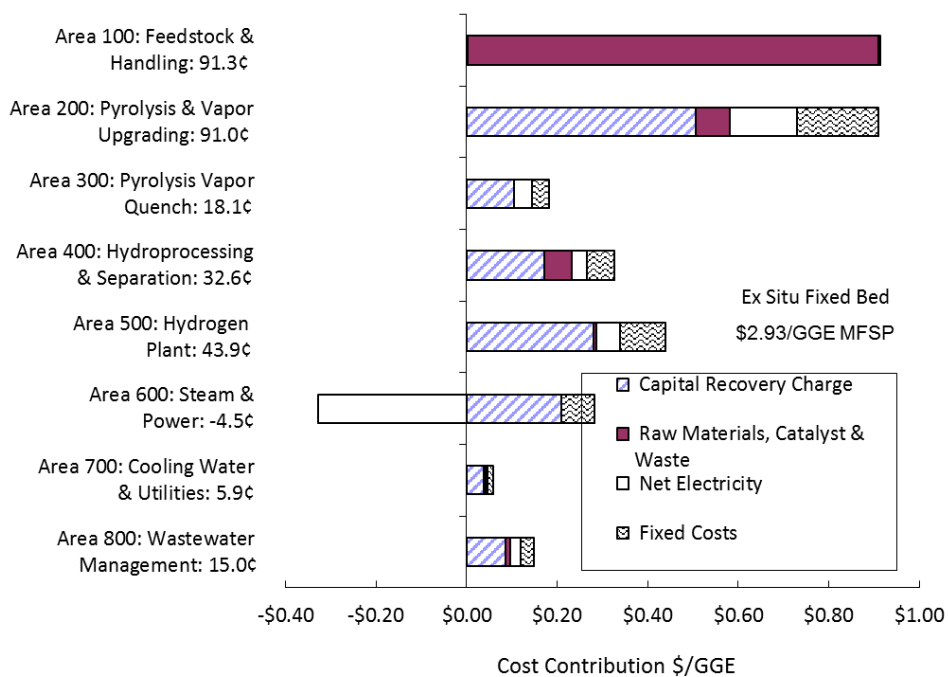
Figure 5. Cost contribution details from each process area for the 2022 projection

The single largest area of cost contribution is from the biomass feedstock (Area 100). It is projected that there will be a significant reduction in this area if research can enable the process to handle lower cost, poorer quality feedstocks. The 2022 projection reflects a cost reduction over the 2018 SOT, with a significant contribution from the use of a blended feedstock with 75% air-classified forest residues (the 2018 SOT used clean pine). The CFP operation (Area 200) is the next highest cost area. Research has already enabled a >50% modeled cost reduction in Area 200 compared to the 2014 SOT (Appendix A). Further cost reduction and additional robustness, along with yield improvements are the key goals of further research in this area. A discussion of the impacts of key parameters on the cost is included in the following sections.

5.1 Sensitivity Analysis

As previously mentioned, feedstock costs are the single-largest contributor to the MFSP. INL provided the related cost sensitivity information presented in Section 5.1.1. Conversion process sensitivities for the 2022 projection and impacts on the MFSP are presented in Section 5.1.2.

5.1.1 Feedstock Cost Sensitivity Analysis

Sensitivity analysis related to the clean pine feedstock used for the 2018 SOT is shown in Figure 6. The modeled cost for this feedstock was $87.82/dry U.S. ton in 2016 dollars.
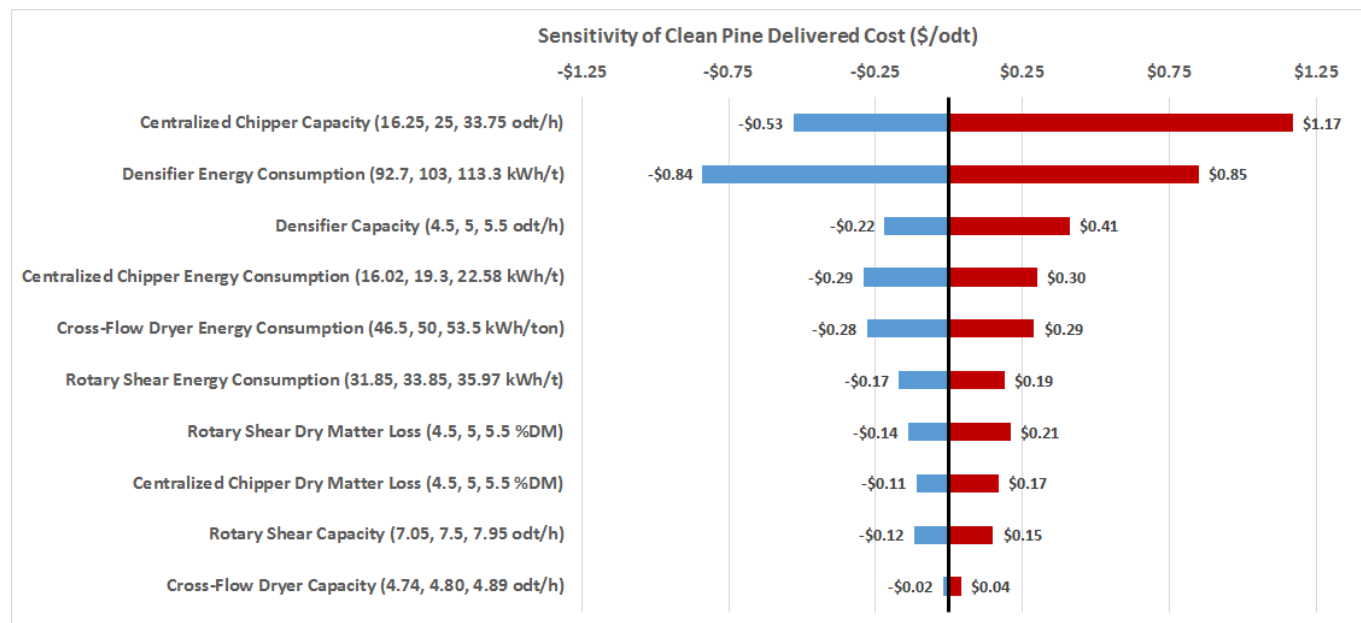
odt = oven dry ton

Figure 6. Sensitivity analysis for the 2018 SOT clean pine feedstock, with a modeled cost of $87.82/dry U.S. ton in 2016$ at the throat of the fast pyrolysis reactor

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Sensitivity analysis related to the blended feedstock modeled for the 2022 projection is shown in Figure 7. The modeled cost for this feedstock was $70.31/dry U.S. ton for this blended material consisting of 25% clean pine and 75% air-classified forest residues.
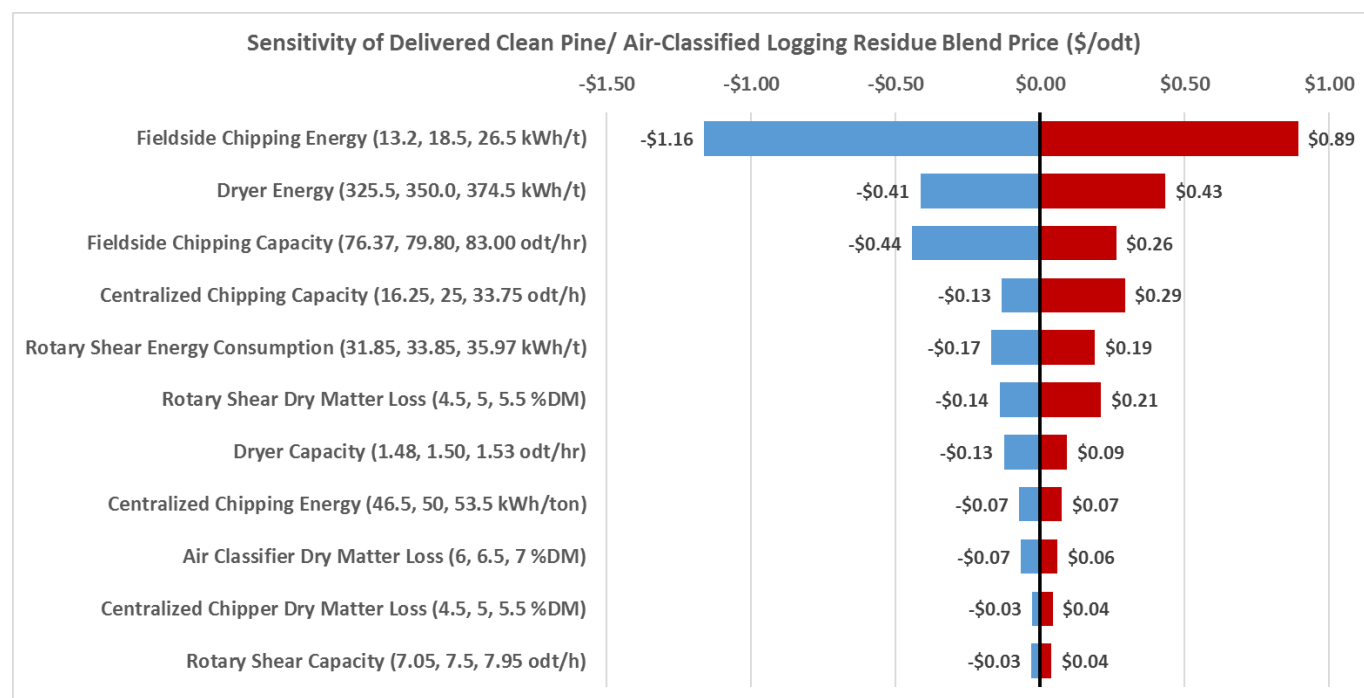
odt = oven dry ton

Figure 7. Sensitivity analysis for the 2022 projection blended feedstock, with a modeled cost of $70.31/dry U.S. ton in 2016$ at the throat of the fast pyrolysis reactor

5.1.2 Feedstock Cost Impact on the MFSP

The impact of the feedstock cost on the MFSP is shown in the next section (Item 4 in Figure 8). Increasing the feedstock cost from $70.31/dry U.S. ton to $100/dry U.S. ton increases by MFSP by 13.2%. Using clean pine at $87.82/dry U.S. ton for the 2022 case (with the same yield) increases the MFSP from $2.93/GGE to $3.16/GGE.

1. Plant Size (10,000 : 2,000 : 1,000 dry metric tonnes/day)	-20.9%	15.6%
2. Total Capital Investment (-15% : base : +30%)	-8.9%	17.7%
3. IRR on Equity / Discount Rate for DCFROR (5 : 10 : 15 %)	-16.3%	16.8%
4. Feedstock Cost, $/dry U.S. ton (50 : 70.31 : 100)	-9.0%	13.2%
5. CFP Organic Liq. Yield; C Efficiency % (34.1;50.6 : 31.4;46.6 : 27.3;40.6)	-4.8%	10.5%
6. Fast Py. & Ex Situ Reactor Capital (-20% : base : +40%)	-4.6%	9.2%
7. Time on Stream (94% : 90% : 80%)	-2.9%	8.5%
8. Interest Rate on the 60% Debt Financing (4% : 8% : 12%)	-5.7%	6.0%
9. Hydroprocessing C Efficiency % (91 : 91 : 88)	0.0%	5.1%
10. Plant Life (30 : 30 : 20 years)	0.0%	4.6%
11. No. of 50% Capacity Ex Situ Reactors Online:Regen (2:1 : 2:2 : 2:5)	-1.4%	4.3%
12. Organic Liq. Oxygen %, Same Overall Yield (16.4 : 16.4 : 18.3)	0.0%	3.8%
13. Hydrotreating Catalyst Unit Cost in 2011$, $/lb (10 : 20 : 60)	-0.9%	3.6%
14. Hydrogen Plant Capital (-20% : base : +30%)	-2.1%	3.2%
15. Pt/TiO2 Catalyst Unit Cost in 2014$, $/lb (50 : 92.35 : 200)	-1.2%	3.2%
16. Hot Gas Filter Capital Cost Only (Base : Base : x2)	0.0%	3.2%
17. Hydroprocessing & Distillation Capital (-20% : base : +40%)	-1.6%	3.1%
18. Steam & Power Plant Capital (-20% : base : +30%)	-1.6%	2.4%
19. Dry Biomass Basis WHSV for Fixed Bed Ex Situ Reactor (5 : 3 : 2)	-1.8%	2.2%
20. Lifetime of Pt/TiO2 Catalyst (4 years : 2 years : 1 year)	-0.8%	1.7%
21. Wastewater Management Capital (-20% : base : +50%)	-0.7%	1.6%
22. Pt/TiO2 Catalyst Reclamation Value after Lifetime (90% : 70% : 40%)	-1.0%	1.5%
23. Wt% Loss as Coke (vs. Gas), Same Organic Liq. Yield (2.1% : 3.2% : 4.5%)	-1.2%	1.5%
24. Hydrocracking Catalyst Unit Cost in 2011$, $/lb (10 : 20 : 60)	-0.2%	0.8%

Finance etc.
Cat. Fast Pyrolysis
Hydroprocessing
Balance of Plant

% Change to MFSP from the Ex Situ 2022 Projection ($2.93/GGE)

Figure 8. Sensitivity analysis for the 2022 projection, with a base case MFSP of $2.93/GGE

5.1.3 Financial and Overall Assumptions

Item 1 (Figure 8) shows that a larger plant can significantly reduce the production cost, provided feedstock is available at the same price. However, a constant feedstock cost assumption for larger scales does not usually hold for biomass, and feedstock cost increases can offset the gains from economies of scale. However, other options, such as deferring hydroprocessing and hydrogen production to a central facility, or a petroleum refinery, can help get some of the benefits of larger scale operations.

Favorable financing and a lower internal rate of return (IRR) expectation can help lower the MFSP (Items 3 and 8). Increased capital costs (Item 2) can increase the MFSP significantly. An onstream factor of 90% was assumed based on a mature plant scenario; lowering the onstream factor assumption from 90% to 80% (Item 7) can increase the MFSP by 8.5% because of low capital utilization. A reduced plant life (Item 10) will similarly affect the MFSP; DCFROR analysis using a longer plant life with paid off capital reduces the calculated MFSP.

5.1.4 Catalytic Fast Pyrolysis

CFP carbon efficiency to organic liquid (Item 5) has one of the biggest impacts on the MFSP. Research improvements and catalyst development since 2014 are the key reasons for the

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modeled cost reduction from $6.25/GGE in 2014 to $3.50/GGE in 2018 (Appendix A). Catalyst improvements have helped increase the carbon efficiency from 27% in 2014 to approximately 45% in 2018 (Appendix A). Note that carbon losses to coking and gases formed via cracking in an HGF are factored into this experimental efficiency reported at the bench scale; carbon losses in a scaled-up HGF system may be different because of a different vapor flow schematic and that will be studied during further scale-up. Additional deoxygenation during CFP (Item 12), while maintaining overall fuel yields, can help reduce the MFSP though reduced hydroprocessing, hydrogen production, and wastewater treatment costs.

Coke reduction in the fixed bed vapor upgrading system leads to efficiency gains toward liquid organic products compared to a ZSM-5 based fluidized system with higher coking. In fact, reduction in coking is a necessity for operational feasibility of the fixed bed system. Reduced coking can allow acceptable run durations before requiring regeneration. While liquid yield increase is the primary benefit from reduced coking, Item 23 shows that losses to the gas phase, rather than losses to the solid phase (coke or char), are economically beneficial; the primary benefit of gases versus solids in this design come from the natural compatibility of gases for hydrogen production, some of it via water gas shift, as well as savings from avoiding solids handling and combustion (in the case of additional combustion in the char combustor). Besides, the greatest benefit of coke prevention will be longer reactor online times enabling operational feasibility.

Item 6 (Area 200 capital) and Item 16 (hot gas filter capital) capture the impacts of changing the estimated capital costs. The required number of offline reactors (Item 11) and the weight hourly space velocity or WHSV on a dry biomass basis in units of h^{-1} (Item 19) affect both the capital equipment and catalyst requirement. The impacts of the unit cost of the catalyst (Item 15), catalyst lifetime (Item 20), and reclamation value of the Pt/TiO_2 catalyst (Item 22) are also shown.

5.1.5 Hydroprocessing

Similar to the CFP step, carbon efficiency during hydroprocessing has a significant impact on the MFSP (Item 9). This efficiency can be changed not only during hydroprocessing, but also in the upstream CFP step by changes to the composition of the CFP oil.

Impact of the hydrotreating, hydrocracking, and distillation (Area 400) capital is shown in Item 17. Changing hydrotreating (Item 13) and hydrocracking (Item 24) catalyst cost assumptions also affect the MFSP.

5.1.6 Other Plant Operations

Changes in capital cost estimates for the hydrogen plant (Item 14), steam and power production (Item 18) and wastewater management (Item 21) are also shown, and the order of their sensitivity impacts in Figure 8 is dictated by their base capital estimates. Co-location with larger facilities, such as petroleum refineries or chemical plants, can bring down these costs through economies of scale and resource integration.

6 Sustainability and Life Cycle Analysis

This section presents sustainability metric indicators for the 2018 SOT and 2022 projection conversion process models. Direct air emissions from the biorefinery (CO_2, NO_2, and SO_2), water consumption, and other process-related metrics were taken from the Aspen Plus models. The material and energy flow information from the conversion models also help capture the impacts of input raw materials, and outputs such as fuel yields, waste, and coproducts. Pertinent flows are shown in Table 11.

The input/output inventories in Table 11 also provide the necessary information required for performing life cycle and supply chain sustainability modeling to quantify greenhouse gas (GHG) emissions and fossil energy consumption. The biorefinery GHGs and fossil energy consumption are quantified separately under supply chain sustainability analysis (SCSA) efforts by Argonne National Laboratory (ANL). A complete well-to-wheel or supply chain life cycle analysis evaluation is required to understand the sustainability implications for the full supply chain based on this technology pathway and quantify associated reduction in GHG emissions from the production of the biomass-derived liquid fuel blendstock (compared to petroleum-derived liquid fuels).

Table 11. Material and Energy Flows in the Conversion Process

	2018 SOT	2022 Projection
Products	lb/h	lb/h
Gasoline Fuel	20,481	21,718
Diesel Fuel	21,997	23,815
Byproducts	kW	kW
Excess Electricity	1970	3063
Resource Consumption	**Flow rate**	**Flow rate**
	lb/h	lb/h
Blended Woody Biomass (wet)	204,131	204,131
Blended Woody Biomass (dry)	183,718	183,718
Sand Makeup	158	158
Natural Gas	147	110
Zeolite Catalyst	0.0E+00	0.0E+00
Fixed-Bed VPU Catalyst (Pt/TiO_2)	17	13
Hydrotreating Catalyst (sulfided CoMo or NiMo)	23	20
Hydrocracking Cat. (crystalline Si-Al with rare earth metals)	3	4
ZnO (reforming cleanup)	3.7E-02	2.8E-02
HDS (reforming cleanup)	1.6E-02	1.2E-02
Steam Reforming Catalyst	8.3E-02	6.2E-02
Shift Catalyst	1.1E-01	8.5E-02
PSA Adsorbent	2.8E+00	2.1E+00
50 wt% Caustic	294	296
Net Water Makeup	70252	65525

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Boiler Feedwater Chemicals	2	2
Cooling Tower Chemicals	1	1
No. 2 Diesel Fuel	71	71
Waste Streams	lb/h	lb/h
Solids Purge from Fluidized Bed Reactors	2063	2062
Wastewater	26398	26720
Air Emissions	lb/h	lb/h
CO_2 (fossil)	404	303
CO_2 (biogenic)	205833	197182
CH4	0.0E+00	0.0E+00
CO	0.0E+00	0.0E+00
NO_2	9	12
SO_2	107	108
H_2O	138466	124245
H_2S	0.0E+00	0.0E+00
Combustor Feed Stream Heating Values		
Char Combustor % Biogenic Carbon	100%	100%
Char Combustor LHV (MMBtu/h)	300.05	335.7
Reformer Fuel Combustor % Biogenic Carbon	99.65%	99.74%
Reformer Fuel Combustor LHV (MMBtu/h)	313.5	285.1

Table 12 shows efficiency indicators and water usage metrics in the conversion process. Carbon and energy efficiencies also reflect the sustainability of the process; conversion of biomass feedstock to desirable products benefits both the economics and sustainability. The process does not require natural gas, and the model predicts that there will be a net electricity export after in-process consumption (the electricity is produced from excess process heat).

Table 12. Sustainability and Process Efficiency Metrics for the Conversion Process

Model	2018 SOT		2022 Projection	
Electricity Credit	Yes	No	Yes	No
GHG Emissions (g CO₂e/MJ) [a]	-0.3	0.9	-0.8	0.8
Net Fossil Energy Consumption (MJ/MJ) [a]	-0.004	0.11	-0.01	0.009
Fuel Yield (% w/w of dry biomass)	23.1		24.8	
Carbon Efficiency to Fuels (% of C in biomass)	39.7		42.3	
Overall Energy Efficiency to Fuels (% LHV basis)	52.1		56.1	
Water Consumption (gal/GGE fuel)	1.3		1.1	
Total Fuel Yield (GGE/dry U.S. ton)	71.7		77.3	
Electricity Production (kWh/GGE)	7.0		6.3	
Electricity Consumption (kWh/GGE)	6.7		5.9	
Wastewater Generation (gal/GGE)	0.48		0.45	

[a] Calculated by ANL using Greenhouse gases, Regulated Emissions, and Energy use in Transportation v. 2018 [26] at the conversion step (i.e., at the biorefinery or "gate-to-gate"), excluding upstream and downstream processes in the supply chain. The full SCSA results are reported separately [27].

An SCSA for the *ex situ* CFP pathway was conducted using ANL's 2018 version of the Greenhouse gases, Regulated Emissions, and Energy use in Transportation model [26]. The SCSA incorporated the respective feedstocks modeled by INL for the 2018 SOT (100% clean pine) and the 2022 projection (75% air-classified and leached logging residues and 25% clean pine) [12]. For the conversion step, when the displacement credit of coproduced electricity is excluded, fossil energy consumption is approximately 0.11 and 0.009 MJ/MJ for the 2018 SOT and 2022 projection, respectively; GHG emission intensities are approximately 0.9 and 0.8 g CO₂e/MJ, respectively, for the 2018 SOT and 2022 projection. When the displacement credit of coproduced electricity is included, net fossil energy consumption is approximately -0.004 and -0.01 MJ/MJ for the 2018 SOT and 2022 projection, respectively; net GHG emission intensities are approximately -0.3 and -0.8 g CO₂e/MJ for the 2018 SOT and 2022 design cases, respectively. Energy self-sufficient conversion processes contribute to the low fossil energy consumption and low GHG emission intensities of the *ex situ* CFP conversion technology.

ANL's SCSA [27] showed that the overall modeled GHG reduction relative to petroleum-derived gasoline is greater than 60% for both the 2018 SOT and 2022 projection; feedstock choices and related preprocessing can have significant impacts on the SCSA results.

7 Conclusions and Future Work

Catalyst development and related experimental efforts have led to significant gains toward future feasibility of the *ex situ* catalytic fast pyrolysis process for biomass conversion presented in this report (and related previous publications). A more than 60% relative improvement in carbon efficiency has been experimentally demonstrated at the bench scale in 2018 (compared to the initial 2014 benchmark). The use of a Pt-based catalyst in a catalytic fixed bed *ex situ* vapor upgrading reactor, with a low Pt loading of 0.5% and high-carbon efficiencies, allowed reductions in the modeled MFSP.

With the recent gains in process efficiency, research focus through 2022 will include considerations of future industrial relevance, in addition to further yield improvements primarily by reducing carbon loss to light gases (CO and CO_2). Research will include: establishing longer CFP catalyst lifetimes and longer onstream times before requiring regeneration; developing more rapid regeneration; enabling the use of lower cost (and less pristine) feedstocks to allow further cost reduction and added diversity in the feedstock supply chain; and targeting improvements in fuel quality making the product fuel blendstocks more desirable for end use in transportation. Research through 2022 and beyond will include the improvement and tailoring of CFP oil composition through further catalyst research; this can enable reduced hydroprocessing costs including the option of coprocessing at petroleum refineries, improvements in downstream fuel quality, and the production of valuable and separable coproducts.

References

1. Dutta, A., A. Sahir A, E. Tan, D. Humbird, L. Snowden-Swan, P. Meyer, J. Ross, D. Sexton, R. Yap, and J. Lukas J. 2015. *Process Design and Economics for the Conversion of Lignocellulosic Biomass to Hydrocarbon Fuels—Thermochemical Research Pathways with in situ and ex situ Upgrading of Fast Pyrolysis Vapors*. NREL/TP-5100-62455, PNNL-23823. Golden, CO: National Renewable Energy Laboratory.

2. U.S. Department of Energy. 2016. *Bioenergy Technologies Office Multi-Year Program Plan*. Washington, DC: U.S. Department of Energy.

3. Lappas, A.A., M.C. Samolada, D.K. Iatridis, S.S. Voutetakis, and I.A. Vasalos. 2002. "Biomass pyrolysis in a circulating fluid bed reactor for the production of fuels and chemicals." *Fuel* 81, no. 16: 2087–2095. https://doi.org/10.1016/S0016-2361(02)00195-3.

4. Wan, S., T. Pham, S. Zhang, L. Lobban, D. Resasco, and R. Mallinson. 2013. "Direct catalytic upgrading of biomass pyrolysis vapors by a dual function Ru/TiO_2 catalyst." *AIChE Journal* 59, no. 7: 2275–2285. https://doi.org/10.1002/aic.14038.

5. Wang, K., D.C. Dayton, J.E. Peters, and O.D. Mante. 2017. "Reactive catalytic fast pyrolysis of biomass to produce high-quality bio-crude." *Green Chem* 19: 243–3251. https://doi.org/10.1039/C7GC01088E.

6. Dutta, A., J. Schaidle, D. Humbird, F. Baddour, and A. Sahir. 2016. "Conceptual Process Design and Techno-Economic Assessment of *Ex Situ* Catalytic Fast Pyrolysis of Biomass: A Fixed Bed Reactor Implementation Scenario for Future Feasibility." Topics in Catalysis 59, no. 1: 2–18. https://doi.org/10.1007/s11244-015-0500-z.

7. Griffin, M.B., K. Iisa, H. Wang, A. Dutta, K.A. Orton, R.J. French, and D.M. Santosa et al. 2018. "Driving towards cost-competitive biofuels through catalytic fast pyrolysis by rethinking catalyst selection and reactor configuration." *Energy Environ Sci* 11: 2904–2918. http://dx.doi.org/10.1039/C8EE01872C.

8. Internal Revenue Service. 2009. "How to Depreciate Property." Internal Revenue Service Publication 946. Washington, DC: United States Department of the Treasury.

9. U.S. Bureau of Labor Statistics. 2018. "PPI industry sub-sector data for Chemical Manufacturing." Accessed October 26, 2018: https://data.bls.gov/timeseries/PCU325---325---?amp%253bdata_tool=XGtable&output_view=data&include_graphs=true.

10. Chemical Engineering. 2018. "Chemical Engineering Plant Cost Index." Accessed September 2018.

11. Aspen Plus. n.d. Bedford, MA: AspenTech.

12. Hartley, Damon S., David N. Thompson, Hongqiang Hu, and Hao Cai. 2018. *Woody Feedstock 2018 State of Technology Report.* INL/EXT-18-51655. Idaho Falls, ID: Idaho National Laboratory.

13. Hu, Hongqiang, Tyler L. Westover, Robert Cherry, John E. Aston, Jeffrey A. Lacey, and David N. Thompson. 2017. "Process simulation and cost analysis for removing inorganics from wood chips using combined mechanical and chemical preprocessing." *BioEnergy Research* 10, no. 1: 237–247. https://doi.org/10.1007/s12155-016-9794-3.

14. Baddour, F., and L. Snowden-Swan. 2017. "Catalyst Cost Model Development." Presented at Bioenergy Technologies Office 2017 Project Peer Review, Denver, CO, March 7, 2017. https://www.energy.gov/sites/prod/files/2017/05/f34/thermochem_baddour_2.5.4.301-302.pdf.

15. Chemical Catalysis for Bioenergy Consortium. n.d. https://www.chemcatbio.org.

16. Beckham, G.T. 2017. "Biological Conversion of Thermochemical Aqueous Streams." Presented at Bioenergy Technologies Office 2017 Project Peer Review, Denver, CO, March 8, 2017.

17. Mukarakate, C., R.J. Evans, S. Deutch, T. Evans, A.K. Starace, J. ten Dam, M.J. Watson, and K. Magrini. 2017. "Reforming Biomass Derived Pyrolysis Bio-oil Aqueous Phase to Fuels." *Energy Fuels* 31, no. 2: 1600–1607. https://doi.org/10.1021/acs.energyfuels.6b02463.

18. "Sand and Gravel (Industrial)." 2014. U.S. Geological Survey. Accessed August 12, 2014: http://minerals.usgs.gov/minerals/pubs/commodity/silica/mcs-2014-sandi.pdf.

19. Suresh, B., R. Gubler, Y. Yamaguchi, and X. He. 2013. "Hydrogen." In *Chemical Economics Handbook.* Englewood, CO: IHS Chemical.

20. Peters, M.S., K.D. Timmerhaus, and R. West. 2004. *Plant Design and Economics for Chemical Engineers, International Edition.* 5th edition. New York: McGraw-Hill, Inc.

21. U.S. Energy Information Administration. 2011. "Annual Energy Outlook 2011. Data Table A3 – Energy Prices by Sector and Source." Washington, DC: U.S. Energy Information Administration.

22. Perry, R.H., D.W. Green, and J.O. Maloney. 1997. *Perry's Chemical Engineers' Handbook.* 7th edition. New York: McGraw-Hill.

23. Humbird, D., R. Davis, L. Tao, C. Kinchin, D. Hsu, A. Aden, P. Schoen, et al. 2011. *Process Design and Economics for Biochemical Conversion of Lignocellulosic Biomass to Ethanol: Dilute-Acid Pretreatment and Enzymatic Hydrolysis of Corn Stover.* NREL/TP-5100-47764. Golden, CO: National Renewable Energy Laboratory.

24. U.S. Bureau of Labor Statistics. 2018. "Databases, Tables & Calculators by Subject." Accessed 2018: http://data.bls.gov/cgi-bin/srgate (enter CEU3232500008 series id).

25. Argonne National Laboratory. 2014. "GREET 2016 Model." Accessed 2018: https://greet.es.anl.gov/publication-greet-model.

26. Wang, Michael, Amgad Elgowainy, Pahola Thathiana Benavides, Andrew Burnham, Hao Cai, Qiang Dai, Troy Robert Hawkins et al. 2018. *Summary of Expansions and Updates in GREET 2018*. ANL-18-38. Lemont, IL: Argonne National Laboratory.

27. H. Cai, T. Benavides, U. Lee, M. Wang, E. Tan, R. Davis, A. Dutta et al. 2018. *Supply Chain Sustainability Analysis of Renewable Hydrocarbon Fuels via Indirect Liquefaction, Ex Situ Catalytic Fast Pyrolysis, Hydrothermal Liquefaction, Combined Algal Processing, and Biochemical Conversion: Update of the 2018 State-of-Technology Cases and Design Case*s. ANL/ESD-18/13. Lemont, IL: Argonne National Laboratory.

Appendix A: 2014–2018 SOT and 2022 Projection

Processing Area Cost Contributions & Key Technical Parameters	Units	2014 SOT	2015 SOT	2016 SOT	2017 SOT	2018 SOT	2022 Projection
Process Concept: Hydrocarbon Fuel Production via Ex Situ Upgrading of Fast Pyrolysis Vapors		Clean Pine	Clean Pine	Clean Pine	Clean Pine	Clean Pine	Residues + Pine
Year $ Basis		2016	2016	2016	2016	2016	2016
Projected Minimum Fuel Selling Price▲	$/GGE*	$6.25	$5.45	$4.90	$4.09	$3.50	$2.93
Conversion Contribution	$/GGE*	$3.66	$3.30	$3.08	$2.82	$2.28	$2.02
Total Project Investment per Annual GGE	$/GGE-yr	$18.50	$16.46	$14.94	$12.17	$11.35	$10.22
Plant Capacity (Dry Feedstock Basis)	metric tons/day	2,000	2,000	2,000	2,000	2,000	2,000
Total Gasoline Equivalent Yield	GGE/dry US ton	42	46	51	69	72	77
Diesel-Range Product Proportion (GGE* basis)	% of fuel product	15%	15%	15%	52%	52%	52%
Feedstock							
Total Cost Contribution	$ / GGE	$2.59	$2.15	$1.82	$1.27	$1.23	$0.91
Capital Cost Contribution	$ / GGE	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00
Operating Cost Contribution	$ / GGE	$2.58	$2.14	$1.81	$1.27	$1.22	$0.91
Feedstock Cost	$ / Dry US Ton	$108.43	$98.56	$92.69	$87.82	$87.82	$70.31
Feedstock Moisture at Plant Gate	Wt % H_2O	10%	10%	10%	10%	10%	10%
Feed Moisture Content to Pyrolyzer	wt % H_2O	10%	10%	10%	10%	10%	10%
Energy Content (LHV, Dry Basis)	BTU / lb	8,000	8,000	8,000	8,000	8,000	8,000
Pyrolysis and Vapor Upgrading							
Total Cost Contribution	$ / GGE	$2.34	$2.03	$1.84	$1.46	$0.99	$0.91
Capital Cost Contribution	$ / GGE	$0.95	$0.82	$0.74	$0.65	$0.54	$0.51
Operating Cost Contribution	$ / GGE	$1.39	$1.21	$1.09	$0.80	$0.45	$0.40
Ex Situ Reactor Configuration	reactor type	Fluidized Bed	Fluidized Bed	Fluidized Bed	Fixed Bed	Fixed Bed	Fixed Bed
Ratio of Online:Regenerating Fixed Bed Reactors	ratio	N/A	N/A	N/A	2:5	2:3	2:2
Gas Phase	wt % of dry biomass	35%	36%	34%	31%	31%	31%
Aqueous Phase	wt % of dry biomass	25%	25%	24%	27%	23%	23%
Carbon Loss	% of C in biomass	2.9%	2.9%	3.4%	2.9%	5.0%	3.0%
Organic Phase	wt % of dry biomass	17.5%	18.6%	21.8%	28.3%	30.8%	31.4%
H/C Molar Ratio	ratio	1.1	1.1	1.1	1.2	1.2	1.2
Oxygen	wt % of organic phase	15.0%	13.3%	16.8%	16.5%	18.5%	16.4%
Carbon Efficiency	% of C in biomass	27%	29%	33%	42%	45%	47%
Solid Losses (Char + Coke)	wt % of dry biomass	23%	21%	20%	14%	15%	15%
Char	wt % of dry biomass	12.0%	11.0%	12.0%	10.4%	11.7%	11.7%
Coke	wt % of dry biomass	11.0%	9.5%	8.3%	3.3%	3.3%	3.2%
Pyrolysis Vapor Quench							
Total Cost Contribution	$ / GGE	$0.35	$0.33	$0.28	$0.20	$0.19	$0.18
Capital Cost Contribution	$ / GGE	$0.20	$0.19	$0.16	$0.12	$0.11	$0.10
Operating Cost Contribution	$ / GGE	$0.15	$0.14	$0.12	$0.08	$0.08	$0.08

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Hydroprocessing and Separation							
Total Cost Contribution	$ / GGE	$0.33	$0.31	$0.34	$0.35	$0.37	$0.33
Capital Cost Contribution	$ / GGE	$0.17	$0.16	$0.18	$0.19	$0.19	$0.17
Operating Cost Contribution	$ / GGE	$0.15	$0.14	$0.16	$0.16	$0.18	$0.15
Carbon Efficiency of Organic Liquid Feed to Fuels	%	88%	90%	87%	91%	89%	91%
Hydrotreating Pressure	psia	2,000	2,000	2,000	1,900	1,900	1,900
Oxygen Content in Cumulative Fuel Product	wt %	0.8%	0.8%	0.8%	0.6%	0.5%	0.6%
Hydrogen Production							
Total Cost Contribution	$ / GGE	$0.61	$0.56	$0.60	$0.62	$0.50	$0.44
Capital Cost Contribution	$ / GGE	$0.39	$0.36	$0.38	$0.41	$0.32	$0.28
Operating Cost Contribution	$ / GGE	$0.22	$0.20	$0.22	$0.21	$0.18	$0.16
Additional Natural Gas**	% of biomass LHV	0.3%	0.1%	0.2%	0.1%	0.2%	0.2%
Balance of Plant							
Total Cost Contribution	$ / GGE	$0.04	$0.07	$0.03	$0.20	$0.23	$0.16
Capital Cost Contribution	$ / GGE	$0.80	$0.71	$0.56	$0.43	$0.39	$0.33
Operating Cost Contribution	$ / GGE	($0.76)	($0.64)	($0.54)	($0.23)	($0.16)	($0.17)
Electricity Production from Steam Turbine (credit included in op. cost above)	$/GGE*	($1.12)	($0.96)	($0.78)	($0.42)	($0.36)	($0.33)
Sustainability and Process Efficiency Metrics							
Fuel and Coproducts Yield by Weight of Biomass	% w/w of dry biomass	13.7%	15.0%	16.5%	22.2%	23.1%	24.8%
Carbon Efficiency to Fuels and Coproducts	% C in Feedstock	23.5%	25.9%	28.3%	38.1%	39.7%	42.3%
Overall Carbon Efficiency to Fuels and Coproducts	% C in Feedstock + NG	23.5%	25.9%	28.3%	38.1%	39.7%	42.3%
Overall Energy Efficiency to Fuels and Coproducts	% LHV of Feedstock + NG	30.4%	33.4%	37.0%	50.2%	52.1%	56.1%
Electricity Production	kWh/GGE	21.0	18.0	14.7	8.0	7.0	6.3
Electricity Consumption (Entire Process)	kWh/GGE	12.7	11.0	9.6	6.4	6.7	5.9
Water Consumption in Conversion Process	gal H$_2$O/GGE	1.4	1.4	1.3	1.5	1.3	1.1
TEA Reference File		PyVPU-v218g ES - Fluidized Bed - FY14 SOT (2016$)-V02d.xlsm	PyVPU-v218g ES - Fluidized Bed - FY15 SOT (2016$)-V02d.xlsm	PyVPU-v218g ES - Fluidized Bed - FY16 SOT (2016$)-V02d.xlsm	PyVPU-v218h ES FixedBed-v49-r046-AP10-FY17SOT (2016$)-V02d.xlsm	PyVPU-v218h ES FixedBed-v49-r046-AP10-FY18SOT (2016$)-V09d.xlsm	PyVPU-v218h ES FixedBed-v49-r046-AP10-2022 Target (2016$)-V06d.xlsm

▲Conceptual design result. *Gallon gasoline equivalent on a lower heating value basis. **A negligible stream was maintained in the model to allow natural gas use if necessary.