

Ex Situ Catalytic Fast Pyrolysis of Lignocellulosic Biomass to Hydrocarbon Fuels: 2019 State of Technology and Future Research

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National Renewable Energy Laboratory
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List of Acronyms

AAEM	alkali and alkaline earth metal
ANL	Argonne National Laboratory
Btu	British thermal unit
CFP	catalytic fast pyrolysis
Co-HP	co-hydroprocessing
DCFROR	discounted cash flow rate of return
FCI	fixed capital investment
FY	fiscal year
GGE	gallon gasoline equivalent
GHG	greenhouse gas
HGF	hot gas filter
INL	Idaho National Laboratory
ISBL	inside battery limits
LCA	life-cycle analysis
LHV	lower heating value
MEK	methyl-ethyl-ketone
MFSP	minimum fuel selling price
NREL	National Renewable Energy Laboratory
PNNL	Pacific Northwest National Laboratory
MM	million
PSA	pressure swing adsorption
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

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Executive Summary

Structure of This Report

The goal of this executive summary is to present recent research outcomes for the biomass *ex situ* catalytic fast pyrolysis (CFP) conversion pathway in the context of techno-economic analysis (TEA), and provide a clear understanding of our rationale behind research decisions, without the reader having to search for details within the report. Key summary tables are also included at the end of this executive summary. Quantitative details are presented in the main body of this report with focus on the 2019 State of Technology (SOT) and a 2020 projection case with co-hydroprocessing (co-HP); background information is provided for anyone familiar with biomass CFP to comprehend the text without having to refer to additional material. Interested readers, and those unfamiliar with biomass CFP, are provided with more detailed references for additional background. Detailed cost breakdown information is presented in the Appendix.

Background Information

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 CFP; the focus is on research learnings since a 2018 SOT publication.ⁱ

In the ex situ CFP pathway, biomass undergoes rapid deconstruction in a fast pyrolysis reactor at approximately 500°C (932°F), followed by the separation of produced solids (char and mineral matter) from vapors (including permanent gases); 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 after condensation (the condensation product is known as CFP oil). Solids removal prior to the ex situ upgrading step provides an advantage with respect to catalyst stability and choices (relative to in situ upgrading where the catalyst mixes with biomass-derived solids within the fast pyrolysis reactor); catalyst choices can be further broadened to include supported 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 less than 1wt% oxygen. Catalyst stability during hydrotreating of raw fast pyrolysis bio-oil is a major challenge; multiple hydrotreating steps are required unless the catalytic vapor upgrading step is included. Initial research efforts associated with this project (beginning in 2014) were focused on zeolite catalysts in a fluidized reactor system; this was detailed in a 2015 design report.ⁱⁱ The 2018 SOT report documented significant liquid-range product yield improvements in 2017 using a Pt/TiO₂ catalyst that led to the choice of a fixed bed ex situ configuration for further research improvements beyond 2017. The significant modeled minimum fuel selling price (MFSP) reduction in 2018 was because of a reduced Pt loading of 0.5% on the Pt/TiO₂ catalyst, compared to 2% in 2017, and increased ratio of online to regeneration time (reduction from 2:5 to 2:3).

ⁱ NREL/TP-5100-71954 available at: <u>https://www.nrel.gov/docs/fy19osti/71954.pdf.</u>

ⁱⁱ NREL/TP-5100-62455, PNNL-23823 available at: <u>https://www.nrel.gov/docs/fy15osti/62455.pdf.</u>

Research and Development Since the 2018 SOT Report

Carbon Balance Closure

Through 2018, there were uncertainties regarding the product distribution for the CFP step because experimentally measured carbon balance closures were 88% and 87% in the 2017 and 2018 SOT experiments, respectively. A decision was made to tackle this uncertainty by deploying additional analytical equipment (a Polyarc-GC-MS system) in 2019, which helped achieve a nearly 100% carbon balance closure. Our previous approach was to prorate the missing carbon among the different phases (solid char, condensable liquids, and permanent gases), with some conservatism toward the carbon allocated to the liquid phase fuel-precursor products because fuel yield improvements significantly benefit the economics of the process. The 2018 SOT MFSP was initially evaluated at \$3.50/gallon gasoline equivalent (GGE), with sensitivity analysis capturing the effect of lower yields of liquid fuel. The new measurements in 2019 showed that the majority of the missing carbon was in light oxygenated species. This quantification had a significant impact on the economics of the conceptual process design because hydrotreating those light oxygenates would result in gaseous hydrocarbons that do not count towards our desired liquid-range hydrocarbon fuel products. The 2018 SOT was revised after discounting the organic phase products that do not produce liquid hydrocarbons; the carbon efficiency towards the organic intermediate used for producing liquid hydrocarbon fuels was revised downward from 44.7% to 40.4%, resulting in an upward revision of the 2018 SOT MFSP from \$3.50/GGE to \$3.80/GGE. This revision did not include any efforts towards the recovery and utilization of the newly quantified light oxygenates that account for >10% of the biomass carbon.

Recovery of Light Oxygenates

The additional quantification of product species required research towards the effective recovery and utilization of the light oxygenates to benefit the economics of the process. The results showed that there is significant selectivity toward three specific compounds (among all the light oxygenates and some light hydrocarbons): (a) acetaldehyde, (b) acetone, and (c) 2-butanone (or methyl-ethyl-ketone [MEK]) together accounted for ~90% of the mass of C₄- species, and >75% of the mass after including C₅₊ species of light compounds. Yields to these three compounds accounted for >7.5% of the total initial biomass carbon. The high selectivity offered an option to recover, separate, and sell these oxygenated products into existing markets. The removal of these compounds would also reduce hydrogen demand for the subsequent hydroprocessing step to produce hydrocarbon fuel blendstocks. The TEA model was modified to reflect this approach. The separation strategy was based on an adsorption/desorption cycle to trap the light oxygenates, followed by distillation of the relatively smaller stream of desorbed condensables to recover purified products. Model compound mixtures were used in experiments to measure adsorption/desorption performance; different adsorbents were tested to prove that we indeed have an adsorbent that can capture the desired compounds. Adsorption/desorption experiments on biomass CFP vapors will be performed in the near future to validate the TEA assumptions (TEA assumptions are currently based on model compounds). In addition, there are uncertainties related to effective distillation and purification of acetone and MEK from the desorbed stream containing other compounds. Contingencies were added to the TEA to cover additional costs; as noted, the desorbed stream will have a relatively small flow rate and additional capital and operating costs will likely be covered under our assumed contingency. We will however need to

fully quantify the desorbed stream to prove our approach of final product purification via distillation. The detailed stream composition will be used in our Aspen Plus model to quantify the potential product purity and recovery; follow-up experiments will be conducted if necessary. While acetone and MEK have existing markets that can absorb/consume the production from multiple biorefineries at the 2,000 dry metric tons per day scale (the basis of the conceptual design models used for the TEA), acetaldehyde does not have a significant market and cannot be sold directly after separation. Thus, acetaldehyde was not considered for near term valorization, although significant research opportunities exist for the conversion of a nearly pure acetaldehyde stream into other marketable products. The sale of these coproducts (acetone and MEK at this point) helped reduce the MFSP for fuel blendstocks by 52 cents/GGE in the 2019 SOT model.

CFP Catalyst Onstream Time

Another major development was the use of a modified support structure for the Pt/TiO₂ catalyst for CFP experiments in 2019. This catalyst was able to maintain a predetermined performance threshold for 8 hours before requiring regeneration; this is a significant improvement over the 2018 catalyst. The previous 2-hour onstream time in 2018 was a major concern for commercial implementation of this fixed bed *ex situ* CFP system. Further improvements are being made to extend the onstream time. The current performance allows a 1:1 ratio of online:regenerating reactors. The enhanced onstream time was achieved despite a decrease in the CFP oil oxygen content compared to the FY18 SOT, from 19 wt% to 15 wt% (dry basis). However, this new catalyst resulted in a reduction in the CFP oil carbon yield to 35% in 2019 compared to the revised 2018 SOT yield of 40%. It is likely that there is scope for optimization of the 2019 catalyst performance via adjustments in operating conditions and space velocities; this is expected to be explored during a 500-hour bench-scale experimental campaign in 2020.

Use of a Lower-Cost Feedstock

Another 2019 SOT highlight was the successful use of a lower-quality (and lower-cost) feedstock that incorporated 50% forest residues with 50% clean pine (compared to 100% clean pine used in previous years). This decreased the feedstock cost to \$70/dry US ton vs the 2018 value of \$88/dry US ton. The modeled 2019 SOT MFSP was \$3.33/GGE using the \$70/dry US ton feedstock cost; using the \$88/dry US ton feed cost would have given a modeled MFSP of \$3.63/GGE. Long-term impacts of the additional mineral matter from forest-residues on the *ex situ* CFP catalyst will be studied as part of the 500-hour run planned in 2020. Additional operations to further clean up mineral matter from forest-residues are being explored at the Idaho National Laboratory (INL). The current 2021-2022 projections *tentatively* use a 75% forest residues/25% clean pine blend with mineral matter content reduced below 1%. The impacts of any additional cleanup on CFP performance will be studied. The final feedstock choice for a 2022 verification will be dictated by the performance of the current 50% forest residues/50% clean pine mix during the 500-hour run.

CFP Oil Co-Hydroprocessing Strategy

Past SOTs have quantified standalone single-stage hydrotreating performance of CFP oils. Cohydroprocessing the CFP oil in petroleum refineries is also an option being considered. This approach can facilitate significant cost-savings (this is in addition to recovering coproducts and using lower cost feedstocks discussed above). CFP-oil can be co-hydroprocessed with streams at petroleum refineries by utilizing existing hydroprocessing capital and hydrogen infrastructure and avoiding the investment in these facilities at a smaller scale (with higher unit processing costs) at the biorefinery. However, streams such as CFP oil with high oxygen content of 15-20% have never been introduced into mainstream petroleum refining processes. The key challenges for a co-hydroprocessing approach include: (1) proving the feasibility of the coprocessing with appropriately compatible petroleum refinery units/streams to achieve desired product quality (including reduction in oxygen content and associated reactivity), (2) identifying potential disruptive impacts on petroleum refinery operations (e.g., operational upsets such as plugging, corrosion, catalyst life, fuel quality) and developing/proving ways to mitigate them, and (3) engaging petroleum refining industry experts to allow buy-in of this approach through diligent joint investigation. Before investigating this approach from a TEA perspective, preliminary cohydrotreating experimental data generated at PNNL under the Strategies for Co-Processing in Refineries (SCR) projectⁱⁱⁱ and NREL were used to understand whether this approach would be feasible. Experiments with a straight-run diesel fraction and 5-20 wt% CFP oil (from two different catalysts) showed effective deoxygenation of the product, maintenance of carbon efficiencies attributed to the CFP oil portion (compared to standalone hydrotreating of CFP oil), and similarities in fuel properties obtained from hydrotreating a straight run diesel stream (distillation curves with small parallel shifts were observed for one experimental data set and acceptable cetane numbers of greater than 40 were measured for a different experiment). The TEA approach for co-hydroprocessing used hydrogen consumption, the anticipated primary cost driver for this step at a petroleum refinery, as the key variable. Refinery costs were estimated using hydrogen consumption as the basis variable, with information derived from Gary et al.^{iv} The TEA results showed a reduction in the modeled MFSP from \$3.33/GGE for the 2019 SOT to \$3.09/GGE in 2020 (option case presented in this report), with the cost savings attributed entirely to the switch from standalone hydroprocessing to co-hydroprocessing. Life cycle assessment for this scenario (presented as a 2020 option in this report) showed a greater than 50% greenhouse gas (GHG) reduction (compared to petroleum derived gasoline) for the entire production chain from biomass to hydrocarbon fuels (after also accounting for the light-oxygenated coproducts mentioned previously). The hydrogen estimates will be updated in the future; lower hydrogen consumption can allow further GHG reduction. The other area for GHG reduction is during feedstock preprocessing, which will also be considered in an integrated manner as part of life cycle impacts for the entire supply chain.

Research Plans

Current research plans include a potential verification of key parts of this technology in 2022. This verification is designed to facilitate future scale-up by interested industrial entities. A 500-hour bench scale experimental campaign for CFP is expected to be conducted in 2020 to allow the selection of various components to be used during the verification. Corresponding hydroprocessing of the CFP oil, with 500-hour or longer run-times, is also planned. The TEA projections for 2020 presented below include two options for achieving an MFSP of approximately \$3.09/GGE. One of the options includes co-hydroprocessing with petroleum refinery streams, while the other involves an increase in fuel yields via the optimization of

ⁱⁱⁱ Baldwin et al. Bioenergy Technologies Office. Peer Review presentation March 2019. Available at: <u>https://www.energy.gov/sites/prod/files/2019/04/f61/Strategies%20for%20Co-</u> <u>Processing%20in%20Refineries%20%28SCR%29_NL0032422.pdf.</u>

^{iv} Gary, J.H; Handwerk, G.E; Kaiser, M.J. Petroleum Refining: Technology and Economics, 5th Edition. CRC Press.

process conditions. Further cost reduction towards achieving \$3/GGE will be explored based on the 2020 experimental results.

Summary of Key Results

The following figures and tables summarize key results from experiments and related analysis described above. The tables below include information about the updated 2018 SOT and the 2019 SOT, and two options for a 2020 projection of \$3.09/GGE. Additional information, including information for 2014–2017 SOTs and projections for 2021–2022 are presented in the Appendix. Note that the 2021–2022 projections presented are preliminary and will be updated based on the outcome and learnings from the 500-hour run and a decision about our choice of one of the two options presented for 2020. An option with projections based on co-hydroprocessing with additional light oxygenated coproducts recovery improvements was used as the placeholder for 2021–2022; additional benefits may result from using a lower ash feedstock at a comparable cost and a modest increase in fuel blendstock yield. The coproduct recovery improvements can be shown outside the scale-up efforts for the 2022 verification, thus minimizing additional changes to that setup. It should be noted that other research improvements are also expected to continue at the bench scale during the 2021–2022 time frame and some of those aligned improvements may be leveraged to scientifically justify additional cost reduction possible at the verification scale.

2019 SOT

The primary uncertainty for the 2019 SOT assessment lies in our assumption regarding coproduct recovery and purification, since mass and carbon balances are well quantified at this point. The base case MFSP reported for the 2019 SOT is \$3.33/GGE. The following variations were considered because of our current use of model compound information in the TEA:

- The acetone and MEK product prices assumed for the TEA were derived from 5-year averages. A 90% of a 5-year average price was used for both acetone^v and MEK,^{vi} with respective coproduct values of 39.8 ¢/lb and 69.4 ¢/lb respectively; the 90% factor was used in this particular instance to cover uncertainties related to product purity and associated value. Decreasing the coproduct values further to 70% (which may happen in the event that more significant impurities are not mitigated by additional purification steps at the biorefinery) increased the MFSP to \$3.44/GGE;
- Reducing the equipment cost contingency by \$3.5 million dropped the MFSP to \$3.28/GGE, while adding \$3.0 million to the equipment cost increased the MFSP to \$3.37/GGE; and
- Dropping acetone and MEK recovery by 20% increased the MFSP to \$3.43/GGE.

Based on these uncertainties, we report a modeled MFSP range of \$3.28/GGE to \$3.44/GGE associated with the base case value \$3.33/GGE.

^v Pampell, M.; Spyra, T. Acetone. Chemical Economics Handbook, IHS Markit. August 15, 2018.

^{vi} Zhang, E.; Bland, A.; Greiner, E.; Kumamoto, T. Methyl Ethyl Ketone (MEK). Chemical Economics Handbook, IHS Markit. August 31, 2018.

Sensitivity Analysis

Sensitivity analysis for key uncertainties for the 2020 co-hydroprocessing case are shown in Figure ES- 2 below. Some of the sensitivity analyses included in the 2018 SOT reportⁱ are still relevant and all those cases are not repeated in this report. The 2020 petroleum refinery coprocessing option is used as a basis for Figure ES- 2 because the current analysis covers some additional uncertainties related to co-hydroprocessing. Note that a CFP oil cost of \$121/barrel gasoline equivalent product is estimated for this 2020 case. The (barrel gasoline equivalent product) unit was used for volume to avoid confusion about the different density of CFP oil intermediate compared to conventional petroleum refinery streams; the product basis in this case is the final hydrocarbon fuel output from the CFP oil.

Summary of Figures and Tables Presented Below

- Figure ES- 1: Waterfall chart showing previous cost reduction and future projections
- Table ES-1: Key metrics for the updated 2018 SOT, 2019 SOT, and 2020 projections with two options
- Table ES-2: Modeled economic summary for the updated 2018 SOT case
- Table ES-3: Modeled economic summary for the 2019 SOT case
- Table ES- 4: Modeled economic summary for the 2020 projection with a co-hydroprocessing option
- Figure ES- 2: Sensitivity analysis for the 2020 model with co-hydroprocessing
- Table ES- 5: Modeled economic summary for the 2020 projection with a yield increase option.



Figure ES-1. Modeled MFSP for 2014–2019 SOTs and 2020–2022 projections. Two options are presented for cost reduction in 2020: (1) coproducts with refinery co-hydroprocessing and (2) coproducts with a higher fuel yield at the biorefinery. 2030 projection is very preliminary based on additional coproducts and CFP yield improvements.

	Updated 2018 SOT ^a	FY 2019 SOT	FY 2020 Co-HP	FY 2020 Yield ↑
Fast Pyrolysis Intermediate ^b				
Gas Species—CO, CO ₂ , C ₁ -C ₄ (wt% of dry biomass)	13	13	13	13
Organics (wt% of dry biomass)	64	64	64	64
Water (wt% of dry biomass)	11	11	11	11
Char (wt% of dry biomass)	12	12	12	12
Fixed Bed—Online: Regen Reactors ^e	2:3	2:2	2:2	2:2
Gas, Includes Condensables (wt% of dry biomass)	35	38	38	34
Aqueous Phase (wt% of dry biomass)	22	24	24	26
Carbon Loss (% of C in biomass)	5	4.4	4.4	4.3
Organic Phase (wt% of dry biomass)	28	23	23	26
H/C Molar Ratio	1.2	1.2	1.2	1.2
Oxygen (wt% in organic phase)	18.6	15	15	15
Carbon Efficiency (%)	40.4	35 ^f	35 ^f	39 ^f
Solid Losses, Char + Coke (wt% of dry biomass)	12 + 4 ^c	12 + 2	12 + 2	12 + 2
Final Fuel Blendstock				
Yield (%, w/w dry biomass)	21	19	19	21
Hydroprocessing Carbon Efficiency (%) ^e	89	93.5	93.5 ^g	93.5
Overall Carbon Efficiency to Fuel (% of C in biomass)	36	33	33	36
Overall C-Efficiency to Fuel (% of C in biomass + NG)	36	33	33	36
Total Product (GGE/dry U.S. ton)	65	59	59	65
Gasoline-Range Product (gallons/dry U.S. ton)	33	32	32	35
Diesel-Range Product (gallons/dry U.S. ton)	31	26	26	29
Gasoline/Diesel-Range Product (% GGE basis)	48/52	52/48	52/48	52/48
Oxygen Content in Cumulative Product (wt%)	0.5	0.5	0.5	0.5
Acetone + MEK Coproduct Credit (¢/GGE)	-	52	52	47
Minimum Fuel Selling Price (\$/GGE)	3.80	3.33	3.09	3.09
Natural Gas [†] and Electricity				
Natural Gas Energy Input (% of biomass, LHV basis)	0.3	0.1	0.5	1.3
Surplus Electricity Credit (¢/GGE)	7	2	21	-3
Fuel Blendstock Production Efficiencies				
Biomass Feedstock (%, LHV basis)	47	43.6	43.6	48
Biomass + Natural Gas ^d (%, LHV basis)	47	43.6	43.4	47.4
Biomass + Natural Gas ^d + Electricity (%, LHV basis, all electrical energy converted to heat)	49	44.1	48.1	47.2

Table ES-1. Key Metrics for Updated Fiscal Year (FY) 2018 SOT, 2019 SOT, and 2020 Projections

^a Updated from <u>https://www.nrel.gov/docs/fy19osti/71954.pdf</u>. ^b Fast pyrolysis intermediate yields maintained from 2015 design report (<u>https://www.nrel.gov/docs/fy15osti/62455.pdf</u>). ^c Coke value is rounded up, and model assumption is higher than experimental result. ^d Small amount of NG. ^e Model separation & distribution may differ from experiments because of differences in configurations and other assumptions. Parameters with greatest economic impacts are matched closely. ^f Additional >10% carbon in condensables not shown. ^g Co-hydroprocessing. Abbreviations: NG = natural gas; GGE = gallon gasoline equivalent; LHV = lower heating value.

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

\$3.64 /Gallon Gasoline Blendstock \$4.14 /Gallon Diesel Blendstock \$3.80 /Gallon Gasoline Equivalent (GGE)

Gasoline Blendstock Production Diesel Blendstock Production Total Gasoline Equivalent Production

Delivered Feedstock Cost Internal Rate of Return (After-Tax) Equity Percent of Total Investment **On-Stream Factor**

•	
23.8 MM Gal per Year	32.8 Gal per Dry US Ton Feedstock
22.3 MM Gal per Year	30.7 Gal per Dry US Ton Feedstock
47.0 MM GGE per Year	64.9 GGE per Dry US Ton Feedstoc

5 Ton Feedstock \$87.82 per Dry U.S. Ton (Includes Capital Up to Throat of Pyrolyzer)

201.0Z	he
10.0%	
40.0%	
90.0%	

Capital Costs		Ор	erating Costs (¢ / GGE Product)	
100: Feedstock (Additional Dryer & Blower Only)	\$590,000	Feedstock		134.3
200: Fast Pyrolysis & Vapor Upgrading	\$116,470,000	Natural Gas		0.4
300: Pyrolysis Vapor Quench	\$24,520,000	Catalysts		16.8
400: Hydroprocessing & Separation	\$38,100,000	Sand		0.6
500: Hydrogen Plant	\$64,200,000	Other Raw Materia	lls	1.4
600: Steam System & Power Generation	\$54,900,000	Waste Disposal		2.2
700: Cooling Water & Other Utilities	\$9,000,000	Purchased Electrici	ty	-
800: Water Management	\$25,170,000	Fixed Costs		60.0
Total Installed Equipment Cost (TIC)	\$332,930,000	Electricity Coprodu	ict Credit	(7.5)
		Capital Depreciation	on	62.3
Land (115 Acres at \$14000 per Acre)	\$1,600,000	Average Income Ta	x	13.2
Site Development	\$17,970,000	Average Return on	Investment	95.8
(% of ISBL)	10.0%			
Indirect Costs & Project Contingency	\$219,700,000		Operating Costs (\$ / Year)	
(% of TIC)	66.0%	Feedstock		\$63,100,000
		Natural Gas		\$200,000
Fixed Capital Investment (FCI)	\$585,870,000	Catalysts		\$7,900,000
Working Capital	\$29,290,000	Sand		\$300,000
Total Capital Investment (TCI)	\$615,160,000	Other Raw Materials		\$680,000
		Waste Disposal		\$1,040,000
Loan Interest Rate	8.0%	Purchased Electrici	ty	\$0
Loan Term (Years)	10	Fixed Costs		\$28,200,000
		Electricity Coprodu	ict Credit	-\$3,500,000
Total Installed Equipment Cost per Annual GGE	\$7.08	Capital Depreciation	on	\$29,290,000
Fixed Capital Investment per Annual GGE	\$12.47	Average Income Tax		\$6,210,000
		Average Return on	Investment	\$45,010,000
Plant Operating Hours per Year	7884			
On-Stream Percentage	90.0%	Total Plant Electricity Usage (kW)		44,446
		Electricity Produ	iced on Site (kW)	52,038
Maximum Yield Based on HHV of Feedstock + Natural	Gas	Electricity Purch	ased from Grid (kW)	0
Theoretical GGE Production (MM GGE / Year)	105.3	Electricity Sold t	o Grid (kW)	7,593
Theoretical Yield (GGE / Dry Ton)	145.3			
Current Yield (Actual / Theoretical)	44.7%	Plant Electricity Us	e (kWh/GGE)	7.46
Overall Plant Efficiency - HHV %	46.6%		Specific Operating Conditions	
Overall Plant Efficiency - LHV %	47.1%	Feed Rate	Dry Tonnes / Day	2,000
			Dry Tons / Day	2,205
		Feedstock Cost	\$/Dry Ton	\$87.82
Version:			S/Moisture+Ash Free Ton	\$88.64

PyVPU-v218h ES FixedBed-v49-r046-AP10-FY18SOT (2016\$)-V09d-Revised 40.4pct C-Eff

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Table ES-3. Economic Summary (Modeled) for the 2019 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)

\$3.21 /Gallon Gasoline Blendstock \$3.63 /Gallon Diesel Blendstock \$3.33 /Gallon Gasoline Equivalent (GGE)

Gasoline Blendstock Production Diesel Blendstock Production Total Gasoline Equivalent Production Light Oxygenated Coproducts

Delivered Feedstock Cost Internal Rate of Return (After-Tax) On-Stream Factor

Capital Costs			
100: Feedstock (Additional Dryer & Blower Only)	\$520,000		
200: Fast Pyrolysis & Vapor Upgrading	\$112,690,000		
300: Vapor Quench & CFP Co-Products	\$30,430,000		
400: Hydroprocessing & Separation	\$29,250,000		
500: Hydrogen Plant	\$69,300,000		
600: Steam System & Power Generation	\$47,280,000		
700: Cooling Water & Other Utilities	\$8,770,000		
800: Water Management	\$24,250,000		
Total Installed Equipment Cost (TIC)	\$322,470,000		
Land (115 Acres at \$14000 per Acre)	\$1,600,000		
Site Development	\$18,120,000		
(% of ISBL)	10.0%		
Indirect Costs & Project Contingency	\$218,570,000		
(% of TIC)	66.1%		
Fixed Capital Investment (FCI)	\$582,860,000		
Working Capital	\$29,140,000		
Total Capital Investment (TCI)	\$612,000,000		
Loan Interest Rate	8.0%		
Loan Term (Years)	10		
Total Installed Equipment Cost per Annual GGE	\$7.48		
Fixed Capital Investment per Annual GGE	\$13.53		
Plant Operating Hours per Year	7884		
On-Stream Percentage	90.0%		
Maximum Yield Based on HHV of Feedstock + Natural	Gas		
Theoretical GGE Production (MM GGE / Year)	103.3		
Theoretical Yield (GGE / Dry Ton)	142.6		
Current Yield (Actual / Theoretical)	41.7%		
Overall Plant Efficiency - HHV %	43.3%		
Overall Plant Efficiency - LHV %	43.6%		

Version: PyVPU-v218h ES FixedBed-v49-r046-AP10-FY19SOT-V18-Acetone-MEK-\$3.33

23.1 MM Gal per Year31.9 Gal per Dry US Ton Feedstock19.1 MM Gal per Year26.4 Gal per Dry US Ton Feedstock43.1 MM GGE per Year59.5 GGE per Dry US Ton Feedstock49.1 MM Ib per year3.4 % w/w of Dry Biomass2.7 % Acetone/Dry Biomass0.6 % MEK/Dry Biomass\$70.15 per Dry U.S. Ton (Includes Capital Up to Throat of Pyrolyzer)10.0% Equity, % of Total Investment40.0%

Operating Costs (¢ / GGE Product)		
Feedstock	117.0	
Natural Gas	0.1	
Catalysts	14.2	
Sand	0.6	
Other Raw Materials	1.6	
Waste Disposal	3.3	
Purchased Electricity	-	
Fixed Costs	65.2	
Chemical Coproduct Credit	(51.7)	
Electricity Coproduct Credit	(2.1)	
Capital Depreciation	67.6	
Average Income Tax	14.2	
Average Return on Investment	102.9	

Operating Costs (\$ / Year)			
Feedstock		\$50,410,000	
Natural Gas		\$0	
Catalysts		\$6,130,000	
Sand		\$300,000	
Other Raw Material	S	\$700,000	
Waste Disposal		\$1,420,000	
Purchased Electricit	у	\$0	
Fixed Costs		\$28,090,000	
Chemical Coproduct	t Credit	-\$22,290,000	
Electricity Coproduc	ct Credit	-\$920,000	
Capital Depreciation	n	\$29,140,000	
Average Income Tax	(\$6,110,000	
Average Return on Investment		\$44,320,000	
Total Plant Electricit	ty Usage (kW)	40,542	
Electricity Produced on Site (kW)		42,540	
Electricity Purchased from Grid (kW)		0	
Electricity Sold to Grid (kW)		1,997	
Plant Electricity Use	kWh /GGE)	7.42	
Specific Operating Conditions			
Feed Rate	Dry Tonnes / Day	2,000	
	Dry Tons / Day	2,205	
Feedstock Cost	\$/Dry Ton	\$70.15	

\$/Moisture+Ash Free Ton

\$71.39

Table ES-4. Economic Summary (Modeled) for the 2020 Projection with Co-Hydroprocessing Option

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)

\$2.98 /Gallon Gasoline Blendstock \$3.37 /Gallon Diesel Blendstock

23.1 MM Gal per Year

19.1 MM Gal per Year

49.1 MM lb per year

Fixed Costs

Refinery Coprocessing Cost Chemical Coproduct Credit

Electricity Coproduct Credit **Capital Depreciation**

Average Return on Investment Total Plant Electricity Usage (kW)

Electricity Produced on Site (kW)

Electricity Sold to Grid (kW)

Electricity Purchased from Grid (kW)

Average Income Tax

90.0%

43.1 MM GGE per Year

0.95 MM bbl per Year

\$3.09 /Gallon Gasoline Equivalent (GGE)

10.0% Equity, % of Total Investment 40.0%

2.7 % Acetone/Dry Biomas: 0.6 % MEK/Dry Biomass

CFP Oil (\$/bbl GE Product) \$120.78

\$70.15 per Dry U.S. Ton (Includes Capital Up to Throat of Pyrolyzer)

Gasoline Blendstock Production **Diesel Blendstock Production** Total Gasoline Equivalent Production Catalytic Fast Pyrolysis Oil Light Oxygenated Coproducts

Delivered Feedstock Cost Internal Rate of Return (After-Tax) **On-Stream Factor**

Canital Costs

cupital costs	
100: Feedstock (Additional Dryer & Blower Only)	\$590,000
200: Fast Pyrolysis & Vapor Upgrading	\$112,380,000
300: Vapor Quench & CFP Co-Products	\$30,210,000
400: Hydroprocessing & Separation	\$0
500: Hydrogen Plant	\$50,140,000
600: Steam System & Power Generation	\$60,850,000
700: Cooling Water & Other Utilities	\$8,940,000
800: Water Management	\$23,580,000
Total Installed Equipment Cost (TIC)	\$286,690,000
Land (115 Acres at \$14000 per Acre)	\$1,600,000
Site Development	\$15,880,000
(% of ISBL)	10.0%
Indirect Costs & Project Contingency	\$199,030,000
(% of TIC)	65.8%
Fixed Capital Investment (FCI)	\$530,740,000
Working Capital	\$26,540,000
Total Capital Investment (TCI)	\$557,280,000
Loan Interest Rate	8.0%
Loan Term (Years)	10
Total Installed Equipment Cost per Annual GGE	\$6.65
Fixed Capital Investment per Annual GGE	\$12.32
Plant Operating Hours per Year	7884
On-Stream Percentage	90.0%
Maximum Yield Based on HHV of Feedstock + Natura	ıl Gas
Theoretical GGE Production (MM GGE / Year)	103.8
Theoretical Yield (GGE / Dry Ton)	143.3
Current Yield (Actual / Theoretical)	41.5%
Overall Plant Efficiency - HHV %	43.1%
Overall Plant Efficiency - LHV %	43.4%

Operating Costs (¢ / GGE Product) Feedstock 117.0 Natural Gas 0.8 Catalysts 8.2 Sand 0.6 Other Raw Materials 1.5 Waste Disposal 3.1 Purchased Electricity **Fixed Costs** 60.7 Refinery Coprocessing Cost 21.3 Chemical Coproduct Credit (51.7)

31.9 Gal per Dry US Ton Feedstock

26.4 Gal per Dry US Ton Feedstock

59.5 GGE per Dry US Ton Feedstock

1.3 bbl per Dry US Ton Feedstock

3.4 % w/w of Dry Biomass

Electricity Coproduct Credit	(21.4)
Capital Depreciation	61.6
Average Income Tax	13.1
Average Return on Investment	94.0
Operating Costs (\$ / Year)	
Feedstock	\$50,410,000
Natural Gas	\$400,000
Catalysts	\$3,550,000
Sand	\$300,000
Other Raw Materials	\$660,000
Waste Disposal	\$1,350,000
Purchased Electricity	\$0

\$26,160,000

\$9,160,000

-\$22,290,000 -\$9,240,000

\$26,540,000

\$5,650,000 \$40,490,000

38.981

59,012

20,031

0

	Plant Electricity Use (kWh /GGE)		
Version:			
PyVPU-v218h ES FixedBed-v49-r046-AP10-FY20Target-V18-Acetone-MEK-HTcoproc-03		Specific Operating Conditions	
	Feed Rate	Dry Tonnes / Day	2,000
		Dry Tons / Day	2,205
	Feedstock Cost	\$/Dry Ton	\$70.15
		\$/Moisture+Ash Free Ton	\$71.39



% Change to MFSP from the ex situ base case (\$3.09/GGE)

Figure ES-2. Sensitivity analysis for the 2020 model with coproducts and co-hydroprocessing

Table ES-5. Economic Summary (Modeled) for the 2020 Projection with Yield Increase Option

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\$

90.0%

Minimum Fuel Selling Price (MFSP)

\$2.98 /Gallon Gasoline Blendstock \$3.38 /Gallon Diesel Blendstock \$3.09 /Gallon Gasoline Equivalent (GGE)

10.0% Equity, % of Total Investment 40.0%

Gasoline Blendstock Production 25.6 MM Gal per Year **Diesel Blendstock Production** 20.8 MM Gal per Year **Total Gasoline Equivalent Production** 47.4 MM GGE per Year Light Oxygenated Coproducts 48.7 MM lb per year 2.7 % Acetone/Dry Biomass

Delivered Feedstock Cost Internal Rate of Return (After-Tax) **On-Stream Factor**

Capital Costs		Op	perating Costs (¢ / GGE Product)
100: Feedstock (Additional Dryer & Blower Only)	\$510,000	Feedstock	
200: Fast Pyrolysis & Vapor Upgrading	\$112,710,000	Natural Gas	
300: Vapor Quench & CFP Co-Products	\$29,340,000	Catalysts	
400: Hydroprocessing & Separation	\$31,640,000	Sand	
500: Hydrogen Plant	\$72,420,000	Other Raw Materia	ls
600: Steam System & Power Generation	\$42,610,000	Waste Disposal	
700: Cooling Water & Other Utilities	\$8,710,000	Purchased Electricit	ty
800: Water Management	\$24,170,000	Fixed Costs	
Total Installed Equipment Cost (TIC)	\$322,110,000	Chemical Coproduc	t Credit
		Electricity Coprodu	ct Credit
Land (115 Acres at \$14000 per Acre)	\$1,600,000	Capital Depreciatio	n
Site Development	\$18,220,000	Average Income Ta	x
(% of ISBL)	10.0%	Average Return on	Investment
Indirect Costs & Project Contingency	\$218,260,000	-	
(% of TIC)	66.1%		
			Operating Costs (\$ / Year)
Fixed Capital Investment (FCI)	\$582,030,000	Feedstock	
Working Capital	\$29,100,000	Natural Gas	
Total Capital Investment (TCI)	\$611,130,000	Catalysts	
		Sand	
Loan Interest Rate	8.0%	Other Raw Materials	
Loan Term (Years)	10	Waste Disposal	
		Purchased Electricit	ty
Total Installed Equipment Cost per Annual GGE	\$6.79	Fixed Costs	
Fixed Capital Investment per Annual GGE	\$12.27	Chemical Coproduc	t Credit
		Electricity Coprodu	ct Credit
Plant Operating Hours per Year	7884	Capital Depreciatio	n
On-Stream Percentage	90.0%	Average Income Ta	x
		Average Return on	Investment
Maximum Yield Based on HHV of Feedstock + Natur	al Gas		
Theoretical GGE Production (MM GGE / Year)	104.6	Total Plant Electrici	ty Usage (kW)
Theoretical Yield (GGE / Dry Ton)	144.5	Electricity Produ	ced on Site (kW)
Current Yield (Actual / Theoretical)	45.3%	Electricity Purch	ased from Grid (kW)
		Electricity Sold t	o Grid (kW)
Overall Plant Efficiency - HHV %	47.0%		
Overall Plant Efficiency - LHV %	47.4%	Plant Electricity Use	e (kWh/GGE)
			Specific Operating Conditions
Version:		Feed Rate	Dry Tonnes / Day
PyVPU-v218h ES FixedBed-v49-r046-AP10-FY20Targe	t-V18-Acetone-MEK-3pct-higher	-C-eff-03-\$3.09	Dry Tons / Day
		Feedstock Cost	\$/Dry Ton

59.2 (46.6) 2.5 61.4 12.8 93.4 / Year) \$50,410,000 \$900,000 \$6,380,000 \$300,000

\$/Moisture+Ash Free Ton

\$70.15 per Dry U.S. Ton (Includes Capital Up to Throat of Pyrolyzer)

35.4 Gal per Dry US Ton Feedstock

28.8 Gal per Dry US Ton Feedstock 65.5 GGE per Dry US Ton Feedstock

> 106.3 1.9 13.5 0.6 1.5 3.0

\$700,000 \$1,440,000 \$0 \$28,060,000 -\$22,100,000 \$1,190,000 \$29,100,000 \$6,080,000 \$44,280,000

> 40,193 37,978 2,215 0 6.68

> > 2,000

2,205

\$70.15

\$71.39

3.4 % w/w of Dry Biomass

0.6 % MEK/Dry Biomass

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1 Introduction

The 2015 catalytic fast pyrolysis (CFP) design report [1] detailed *in situ* and *ex situ* CFP 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]. Since 2015, the focus of the research has been on *ex situ* systems because it affords more opportunity to understand and affect the chemistry through catalyst development in the absence of biomass-derived solid material present in an *in situ* system. Bench-scale experimental results have been used to update the state of technology (SOT) for the *ex situ* pathway since 2014; significant yield improvements have been achieved via catalyst development.

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 burnoff 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 SOT experiments since 2017 and future projections were based on the Pt/TiO_2 catalyst in a fixed bed as the base configuration.

The Executive Summary provides information about our research findings since the 2018 SOT update [8]; those details are not repeated here.

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 these previous process design reports [1,6].

1.1.1 Financial Assumptions

The modeled projections in this report are based on the technology being implemented in a mature or nth 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 is used for all SOT and projections. Key assumptions are listed in Table 1.

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 (FCI)
	(excluding land purchase cost)
Depreciation schedule	7-year MACRS schedule [9]
Steam plant depreciation	20-year MACRS schedule [9]
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 syst	em

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

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. These assumptions in Table 1 are consistent with the 2018 SOT report [8].

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 [6,8]. 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 million 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
$$\left(\frac{Scaled Capacity}{Base Capacity}\right)^n$$

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 [10] and capital costs were adjusted using the *Chemical Engineering's* Plant Cost Index [11].

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, was used for a discounted cash flow analysis. Those costs along with the gallons gasoline equivalent (GGE) of the 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 a detailed accounting of all mass and energy flows. Details about the Aspen Plus [12] 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). Feedstock blends costs used for the 2019-2020 cases and proposed for the 2021-2022 projections (included in the Appendix) were both modeled at approximately \$70/dry US ton. The plant size was maintained at 2,000 dry metric tons per day.

The 2019 SOT feedstock was based on a 50/50 blend of forest residues and clean pine at \$70.15/dry U.S. ton in 2016 dollars with a relatively high 1.75% modeled ash content (also reflected in the conversion process model by a prorated reduction of the other elements in the elemental analysis). Another feedstock blend option is being considered by INL and the 2021-2022 feedstock cost projection is based on a blend of 75% air-classified forest residues and 25% clean pine, with additional cleanup steps for ash reduction to below 1%; the modeled cost for this feedstock is \$70.31/dry U.S. ton in 2016 dollars. Given the low ash in this 2021-2022 feedstock option, 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.

Although the 2021-2022 models include a different blend in the current projections, the final determination of the feedstock to be used in 2021-2022 will be made based on the performance of the current 50/50 blend feedstock during the 500-hour run, as discussed in the Executive Summary.

2.2 Process Overview



Figure 1. Simplified process flow diagram for fixed bed ex situ catalytic fast pyrolysis, coproduct recovery, and hydroprocessing

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). Beginning 2019, oxygenated coproducts recovery and purification was added to this area
- 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 Section 3.

3 Process Design

3.1 Area 100: Feed Handling

As mentioned in Section 2.1, two different blends of forest residues and clean pine were considered in the 2019-2020 and 2021-2022 TEA models; blending forest residues with clean pine enables lower costs. The moisture content was 10% in both cases, but with a higher modeled ash content of 1.75% in the 2019-2020 cases. 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 [13]. The only minor feedstock handling-related cost added to the plant equipment is a cross-flow dryer for warming the feedstock before feeding to the fast pyrolysis reactor.

Currently, it is estimated that there are 21.2 million dry U.S. tons of pine feedstocks available nationally; 11.8 million dry U.S. tons are planted pine and 9.4 million dry U.S. tons are pine forest residues. This amount of material can be aggregated to support approximately 25 biorefineries of 2,000 dry metric tons per day, given no competition for the resource [13]; however, not all material would be available within the required cost envelope to enable the current cost targets.

The modeled cost summary for a lower cost 2019-2020 feedstock with a mix of 50% forest residues and 50% clean pine with an aggregated ash content of 1.75% (modeled) is presented in Table 2.

Cost Summary (\$/Dry U.S. Ton) (2016\$)		
	2019 SOT	
Grower Payment	\$9.74	
Harvest and Collection	\$4.94	
Field-Side Preprocessing	\$8.41	
Transportation	\$12.22	
Preprocessing	\$28.55	
Storage	\$0.68	
Handling	\$2.65	
Preprocessing Construction	\$2.96	
Quality Dockage	\$0.00	
Grand Total	\$70.15	

Table 2. INL Modeled Feedstock Cost for 50% Forest Residues Plus 50% Clean Pine in 2016\$
(used for 2019-2020 models)

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, the quality of higher blend levels of forest residues may be problematic for the conversion process because of a higher overall ash and alkali and alkaline earth metal (AAEM) concentrations compared to clean pine. If necessary, additional cleaning steps can help reduce this ash content. Cleaning steps are detailed in Hu et al. [14]. These steps can reduce the ash content below 1%. Experimental and model results will be used to co-optimize feedstock cost and quality, performance during the conversion process, and the impact of feedstock preprocessing steps on the LCA. The modeled cost summary for a 2022 feedstock option with 75% air-classified forest residues and 25% clean pine is presented in Table 3.

Cost Summary (\$/Dry U.S. Ton) (2016\$)		
	2022 Projection	
Grower Payment	\$6.84	
Harvest and Collection	\$2.47	
Field-Side Preprocessing	\$9.81	
Transportation	\$13.88	
Preprocessing	\$31.63	
Storage	\$0.58	
Handling	\$1.90	
Preprocessing Construction	\$3.21	
Quality Dockage	\$0.00	
Grand Total	\$70.31	

 Table 3. INL Modeled Feedstock Cost for 75% Forest Residues Plus 25% Clean Pine in 2016\$

 (alternate mix used for the 2021–2022 projection models)

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/TiO₂ 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, [15] developed under the Chemical Catalysis for Bioenergy Consortium [16] was used to estimate the cost of the Pt/TiO₂ 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] and in the 2018 SOT report [8]. As a brief overview, a 2-inch fluidized pyrolyzer was followed by an HGF and a fixed-bed Pt/TiO₂ catalyst vapor upgrading reactor.

The 50% forest residues/50% clean pine blend was converted at a fast pyrolysis temperature of 500°C, the fixed bed *ex situ* reactor setpoint temperature was maintained at 400°C, and the biomass:catalyst (B:C) ratio was 12. The catalyst had 0.5 wt% Pt on a TiO₂ support. CFP oil with an oxygen content of 15 wt% on dry basis was produced with a carbon yield of 35%. Compared to the FY18 SOT, the CFP oil had a lower oxygen content, as intended (15 wt% vs. 19 wt%) but was also produced at a lower carbon yield (35% vs. 40%). In spite of a four times higher B:C ratio during the FY19 experiments compared to FY18, the coke formation was similar or slightly lower (1.9 wt% vs. 2.1 wt%).

In addition to the CFP oil, 10.8% of the *biomass carbon* was present in light condensables. There were high carbon yields to three specific compounds: acetone 3.8%, acetaldehyde 2.8%, and MEK 1.2%. The combined carbon yield for the CFP oil, acetone, acetaldehyde, and MEK was 43%. The CFP oil contained only insignificant amounts of these three compounds. Additional metrics are shown in Table ES- 1.

3.3 Area 300: CFP Product Condensation and Coproducts

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 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.

A major addition to this section in the 2019 model was the recovery of light oxygenates from the gaseous stream of the second condenser. In the process design, the stream is sent to an adsorption system to remove the light oxygenates and other species, while letting the lighter gases pass through. The adsorbed species are then desorbed in a swing system. The desorbed stream is sent to a series of distillation columns to recover acetone and MEK as coproducts. The adsorption/desorption system cost was estimated from an ethanol dehydration mol-sieve system documented by Humbird et al. [22]. The scaling variable was moles of adsorbed species adjusted for partial pressure. The uptake of the desired species by the adsorbent is a key variable for determining the size of this system and model compound experiments showed that the uptake of acetone and MEK was in a similar range compared to the ethanol dehydration system. Experiments will be conducted to determine the uptake of acetone and MEK from biomass CFP vapors (after the condensation of heavier species).

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 2017–2019 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. Additional discussion was included in the 2018 SOT report [8]. For the 2019 SOT, a hydrotreating carbon efficiency of 95% was reported experimentally. The TEA model assumption was slightly lower to allow for some additional losses during hydrocracking of the heavier-than-diesel products; an overall hydrotreating + hydrocracking carbon efficiency of 93.5% was used in the model. The 2020 co-hydroprocessing TEA model maintains the same carbon efficiency assumptions for the CFP oil portion.

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 CFP 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]. The co-hydroprocessing case assumed the use of natural gas for hydrogen production at the petroleum refinery.

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. 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 to allow discharge of the stream contents in an environmentally acceptable manner.

3.9 Process Heat Exchange Cost

A detailed heat exchange network was developed for the *ex situ* process and documented in the 2015 design report [1]. Cost estimates for the heat exchange networks in subsequent models were derived by scaling the costs from the design report using the total process heat exchange duty as the scaling basis.

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). Details for the 2019 SOT model and 2020 projection with the co-HP option are presented below. Information for the other models can be obtained from the tables in the Executive Summary and the Appendix.

4.1 Total Capital Investment

Installed capital costs are shown in Table 4, with per area total purchased equipment cost (TPEC), installation factors, and TIC.

	2019 SOT		2020 C	o-HP Pro	jection		
Area	Process Description	TPEC	f _{install}	TIC	TPEC	f _{install}	TIC
		(MM\$)		(MM\$)	(MM\$)		(MM\$)
100	Feed handling and drying ^a	0.3	1.96	0.5	0.3	1.96	0.6
200	Fast pyrolysis and vapor upgrading	44.8	2.52	112.7	44.7	2.52	112.4
300	Pyrolysis vapor quench, condensation and coproduct recovery	21.0	1.84	38.7	24.9	1.84	45.8
400	Hydroprocessing and product separation	16.5	1.77	29.2	0 ^b	-	0 ^b
500	Hydrogen plant	35.4	1.95	69.3	24.8	2.03	50.1
600	Steam system and power generation	25.7	1.84	47.3	33.0	1.84	60.8
700	Cooling water and other utilities	4.3	2.02	8.8	4.4	2.02	8.9
800	Wastewater management and recycle	10.3	2.35	24.2	10.1	2.34	23.6
ISBL (Areas 100–400)		82.5	2.19	181.2	69.9	2.27	158.8
OSBL (Areas 500–800)		75.8	1.97	149.6	72.3	1.98	143.5
Total		158.4	2.09	330.8	142.2	2.13	302.3

Table 4. Total Installed Equipment Costs for the 2019 SOT and 2020 Projection with Co-HP

^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. ^b Capital at the biorefinery is zero because co-hydroprocessing occurs at the petroleum refinery in this model.

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 5.

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

Table 5. Cost Factors for Indirect Costs

* 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 6 presents a summary of these capital quantities for the 2019 SOT and 2020 co-HP projection.

		2019 SOT	2020 Co-HP Projection
Total purchased equipment cost (TPEC)		\$158,380,000	\$142,210,000
Installation factor		2.088	2.126
Total installed cost (TIC)		\$330,770,000	\$302,330,000
Other direct costs			
Land (not depreciated)		\$1,610,000	\$1,610,000
Warehouse	4.0% of ISBL	\$7,250,000	\$6,350,000
Site development	10.0% of ISBL	\$18,120,000	\$15,880,000
Additional piping	4.5% of ISBL	\$8,150,000	\$7,150,000
Total direct costs (TDC)		\$364,290,000	\$331,710,000
Indirect costs	% of TDC (ex land)		
Prorated expenses	10.0%	\$36,430,000	\$33,170,000
Home office and construction fees	20.0%	\$72,860,000	\$66,340,000
Field expenses	10.0%	\$36,430,000	\$33,170,000
Project contingency	10.0%	\$36,430,000	\$33,170,000
Other costs (startup and permits)	10.0%	\$36,430,000	\$33,170,000
Total indirect costs	60.0%	\$218,570,000	\$199,030,000
Fixed capital investment (FCI)		\$582,860,000	\$530,740,000
Working capital	5.0% of FCI (ex land)	\$29,140,000	\$26,540,000
Total capital investment (TCI)		\$612,000,000	\$557,280,000
TCI/TPEC		3.864	3.919
FCI Lang Factor = FCI/ISBL TPEC		7.060	7.592
TCI Lang Factor = TCI/ISBL TPEC		7.413	7.972

 Table 6. Total Capital Investment Calculations

4.2 Operating Costs

Variable operating cost assumptions are shown in Table 7. The major addition to the table in the 2015 design report [1] is the Pt/TiO_2 catalyst used in the fixed-bed *ex situ* reactor.

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\$) [17]; initial fill, then make up for attrition.			
Fixed-bed <i>ex situ</i> vapor upgrading catalyst	Pt/TiO ₂ with 0.5% Pt loading. Catalyst unit cost \$92.35/lb (2014\$) [15]. A 2-year lifetime is assumed, with 70% cost recovery at the end of 2 years. Quantity of catalyst was determined using a weight hourly space velocity (WHSV) of 5 h^{-1} based on vapor flow 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 NREL 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\$) [18].			
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 at biorefinery: \$239/U.S. ton (\$5/MMBtu) (2011\$).			
	For the co-hydroprocessing case, natural gas was \$3.5/MMBtu, based on a 10- year average for Texas using EIA information (<u>https://www.eia.gov/naturalgas/</u>).			
Solids disposal	Price: \$33/tonne (1998\$) [19].			
Diesel fuel	Usage: 10 gal/h plant-wide use.			
	2012 price projection: \$21.29/MMBtu (2009\$) [20] = \$2.86/gal at 0.85 specific gravity.			
Water makeup	Price: \$0.22/tonne (2001\$) [21] = \$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\$) [22].			
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. [22].			

Table 7. Variable Operating Cost Assumptions

Note: Costs shown were updated to 2016 dollars using the Producer Price Index for chemical manufacturing [10].

Fixed operating costs related to salaries are shown in Table 8. Number of personnel and their salaries were maintained from the 2015 design report [1], and salaries were adjusted using a labor cost index [23].

Cost Item	Factor	2019 SOT	2020 Co-HP Projection
Salaries		\$3,440,000	\$3,440,000
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*)	\$17,486,000	\$15,922,000
Insurance and taxes	0.7% of fixed capital investment (FCI*)	\$4,080,000	\$3,715,000
Total fixed operating co	osts (2016\$/year)	\$28,092,000	\$26,164,000

Table 8. Fixed Operating Costs

* 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 2019 SOT and the 2020 co-HP projection are summarized in Table 9. 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) [24].

	2019 SOT	2020 Co-HP Projection
Lower heating value for gasoline-range products (simulation result)	111,862 Btu/gal	111,882 Btu/gal
Lower heating value for diesel-range products (simulation result)	126,739 Btu/gal	126,739 Btu/gal
Calculated gasoline-equivalent MFSP	\$3.33/GGE	\$3.09/GGE
Calculated actual MFSP for gasoline-range products	\$3.21/gal	\$2.98/gal
Calculated actual MFSP for diesel-range products	\$3.63/gal	\$3.37/gal
Gasoline lower heating value for GGE normalization		116,090 Btu/gal

Table 9. Projected Selling Prices of Hydrocarbon Blendstocks

5 Process Economics Summary

The contribution of individual areas toward the MFSP on a \$/GGE basis are shown in Figure 2 (2019 SOT) and Figure 3 (2020 projection for co-HP option).



Cost Contribution \$/GGE

Figure 2. Cost contribution details from each process area for the 2019 SOT



-\$0.80-\$0.60-\$0.40-\$0.20 \$0.00 \$0.20 \$0.40 \$0.60 \$0.80 \$1.00 \$1.20 \$1.40

Cost Contribution \$/GGE

Figure 3. Cost contribution details from each process area for the 2020 co-HP projection

Note the biggest changes in the 2020 co-HP projection case presented above compared to the 2019 SOT case: (1) savings from lower hydroprocessing and hydrogen costs; (2) additional electricity generation at the biorefinery because of the availability of off-gases otherwise used for additional hydrogen production; and (3) additional contingency at the biorefinery (lumped in Area 300) related to any changes in configuration to allow co-hydroprocessing.

6 Sustainability and Life Cycle Analysis

This section presents sustainability metric indicators for the 2019 SOT and 2020 co-HP projection conversion process models. Direct air emissions from the biorefinery (CO₂, NO₂, and SO₂), 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 10.

The input/output inventories in Table 10 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 supply chain life cycle analysis (LCA) is required to understand the sustainability implications of 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).

	2019 SOT	2020 Co-HP Projection
Products	lb/h	lb/h
Gasoline Fuel	18,189	18,189
Diesel Fuel	16,913	16,913
Byproducts		
Excess Electricity (kW)	1997	20031
MEK (lb/h)	1178	1178
Acetone (lb/h)	5046	5046
Resource Consumption at Petroleum Refinery	Flow rate	Flow rate
Fuel Natural Gas (MMBtu/h)		239
Electricity (kW)	-	2719
Steam (lb/h)	-	-4507
Cooling Water Makeup (gal/h)	-	6014
Hydrotreating Catalyst (lb/h)	-	14
Hydrocracking Catalyst (lb/h)	-	2
Resource Consumption at Biorefinery	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	48	357
Zeolite Catalyst	0.0E+00	0.0E+00
Fixed-Bed VPU Catalyst (Pt/TiO ₂)	7	7
Hydrotreating Catalyst (sulfided CoMo or NiMo)	14	-
Hydrocracking Cat. (crystalline Si-Al with rare earth metals)	2	-
ZnO (reforming cleanup)	1.2E-02	9.1E-02
HDS (reforming cleanup)	5.2E-03	3.9E-02
Steam Reforming Catalyst	2.7E-02	2.0E-01
Shift Catalyst	3.7E-02	2.7E-01
PSA Adsorbent	9.2E-01	6.9E+00
50 wt% Caustic	291	291
Net Water Makeup	75,835	53,545
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	3,741	3,741

Table 10. Material and Energy Flows in the Conversion Process

Wastewater	25,189	22,147
Air Emissions	lb/h	lb/h
CO ₂ (fossil)	132	980
CO ₂ (biogenic)	196,717	197,545
CH4	0.0E+00	0.0E+00
CO	0.0E+00	0.0E+00
NO ₂	9	11
SO ₂	106	106
H ₂ O	135,165	150,488
H ₂ S	0.0E+00	0.0E+00
Combustor Feed Stream Heating Values		
Char Combustor % Biogenic Carbon	100%	100%
Char Combustor LHV (MMBtu/h)	295.94	295.94
Reformer Fuel Combustor % Biogenic Carbon	99.89%	99.17%
Reformer Fuel Combustor LHV (MMBtu/h)	305.43	458.66

Table 11 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 at the biorefinery, and the model predicts that there will be a net electricity export after in-process consumption (the electricity is produced from excess process heat); however, natural gas and electricity are consumed at the petroleum refinery for the co-HP case, as shown in Table 10.

Model		I9 SOT	2020 Co-HP Projection			
Electricity & Coproduct Credit	Yes	No	Yes	No		
GHG Emissions (g CO2e/MJ) ª	-19.8	1.0	-5.1 ^b	29.4 ^b		
Net Fossil Energy Consumption (MJ/MJ) ^a	-0.26	0.01	0.02	0.45		
Fuel Yield (% w/w of dry biomass)		19.1	19.1			
Carbon Efficiency to Fuels (% of C in biomass)	;	33.0	32.9			
Overall Energy Efficiency to Fuels (% LHV basis)		43.6	43.4			
Water Consumption (gal/GGE fuel)		1.7	1.2 °			
Total Fuel Yield (GGE/dry U.S. ton)	:	59.5	59.5			
Electricity Production (kWh/GGE)		7.8	10.8			
Electricity Consumption (kWh/GGE)		7.4	7.1			
Wastewater Generation (gal/GGE)	(0.55	0.49			

Table 11. Sustainability and Process Efficiency Metrics for the Conversion Process

^a Calculated by ANL using Greenhouse gases, Regulated Emissions, and Energy use in Transportation v. 2018 [25] 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 will be reported separately by ANL. ^bIncludes emissions at petroleum refinery. ^c At biorefinery.

An SCSA for the ex situ CFP pathway was conducted using ANL's Greenhouse gases, Regulated Emissions, and Energy use in Transportation or GREET model [25]. The SCSA incorporated the feedstock modeled by INL for the 2019 SOT (50% forest residues and 50% clean pine) [13]. For the conversion step, when the displacement credit of coproduced electricity is excluded, fossil energy consumption is approximately 0.01 and 0.45 MJ/MJ for the 2019 SOT and 2020 co-HP projection, respectively; GHG emission intensities are approximately 1.0 and 29.4 g CO₂e/MJ, respectively, for the 2019 SOT and 2020 co-HP projection. When the displacement credit of coproduced electricity and coproducts is included, net fossil energy consumption is approximately -0.26 and 0.02 MJ/MJ for the 2019 SOT and 2020 co-HP projection, respectively; net GHG emission intensities are approximately -19.8 and -5.1 g CO₂e/MJ for the 2019 SOT and 2020 co-HP projection 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 at the biorefinery. The acetone and MEK coproducts provide further GHG benefits based on the LCA product displacement method [26]; however, there is natural gas and grid electricity consumption at the petroleum refinery that increase GHG emissions for the production of final fuel for the 2020 co-HP case. It should be noted that the hydrogen consumption estimates at the petroleum refinery will be revisited; this report provides initial estimates.

ANL's SCSA shows that the overall modeled GHG reduction relative to a petroleum-derived gasoline baseline (95.3 gCO2e/MJ) is greater than 60% for the 2019 SOT (forthcoming publication [27]). ANL's calculations also quantified a 52% reduction for the 2020 co-HP projection. Feedstock choices and related preprocessing can have significant impacts on the SCSA results and can thus be an area of improvement in this respect. Co-hydroprocessing at a petroleum refinery using an assumption of hydrogen production from natural gas also impacts the GHG reduction for the 2020 co-HP case.

7 Conclusions and Future Work

There were significant updates to this conversion pathway since the 2018 SOT report. These include the accounting of nearly 100% of the biomass carbon in the CFP products; as a result of this quantification, it was found that there are significantly more light oxygenates in the products than previously quantified. Acetaldehyde, acetone, and MEK constitute a significant proportion of those light oxygenates. This necessitated the investigation of ways to recover, purify, and derive value from the light oxygenates by selling acetone and MEK as coproducts. Model compound experiments allowed the selection of a suitable adsorbent to capture these key coproducts and experiments will be conducted using biomass CFP vapors in the near future; this will help validate adsorption performance and related TEA assumptions. Other significant achievements include the successful use of low-cost feedstocks (that included forest residues) during bench scale experiments and a significant extension of onstream times of the fixed-bed ex situ CFP catalyst. Stand-alone hydroprocessing experiments for CFP oil have been successful so far with respect to oxygen content reduction in a single hydrotreating step. Co-hydroprocessing of CFP oil with petroleum refinery streams is being considered as an option to help reduce costs and enhance commercial relevance of this conversion pathway. Pilot-scale verification of key parts of this technology is expected in 2022; a 500-hour bench scale experimental campaign will be conducted in 2020 to prove the feasibility of scaling up this technology.

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Appendix A: 2014–2019 SOT and 2020–2022 Projection

Processing Area Cost Contributions & Key Technical Parameters	Units	2014 SOT	2015 SOT	2016 Target	2017 SOT ¹¹	2018 SOT	2019 SOT	2020 Projection Option 1 Co-HT	2020 Projection Option 2 Yield ↑	2021 Projection	2022 Projection	2030 ^{##} Projection
Process Concept: Hydrocarbon Fuel Production via Ex Situ Upgrading of Fast Pyrolysis Vapors		Clean Pine	Clean Pine	Clean Pine	Clean Pine	Clean Pine	50%Residues/ 50%Pine ^{††}	50%Residues/ 50%Pine ^{††}	50%Residues/ 50%Pine ^{††}	75%Residues/ 25%Pine	75%Residues/ 25%Pine	75%Residues/ 25%Pine
Year \$ Basis		2016	2016	2016	2016	2016	2016	2016	2016	2016	2016	2016
Projected Minimum Fuel Selling Price [▲]	\$/GGE*	\$6.27	\$5.44	\$4.90	\$4.09	\$3.80	\$3.33	\$3.09	\$3.09	\$3.05	\$3.00	\$2.48
Conversion Contribution	\$/GGE*	\$3.66	\$3.30	\$3.08	\$2.82	\$2.44	\$2.14	\$1.90	\$2.02	\$1.87	\$1.83	\$1.34
Total Project Investment per Annual GGE	\$/GGE-yr	\$18.50	\$16.46	\$14.94	\$12.17	\$12.47	\$13.53	\$12.32	\$12.27	\$12.19	\$12.07	\$11.13
Plant Capacity (Dry Feedstock Basis)	metric tons/day	2,000	2,000	2,000	2,000	2,000	2,000	2,000	2,000	2,000	2,000	2,000
Total Gasoline Equivalent Yield	GGE/dry US ton	42	46	51	69	65	59	59	65	60	61	62
Diesel-Range Product Proportion (GGE* basis)	% of fuel product	15%	15%	15%	52%	52%	48%	48%	48%	48%	48%	52%
Feedstock												
Total Cost Contribution ¹¹	\$ / GGE	\$2.60	\$2.14	\$1.82	\$1.27	\$1.36	\$1.18	\$1.19	\$1.08	\$1.18	\$1.17	\$1.14
Capital Cost Contribution ¹¹	\$ / GGE	\$0.00	\$0.00	\$0.00	\$0.00	\$0.00	\$0.00	\$0.00	\$0.00	\$0.00	\$0.00	\$0.00
Operating Cost Contribution ¹¹	\$ / GGE	\$2.60	\$2.14	\$1.81	\$1.27	\$1.35	\$1.18	\$1.18	\$1.07	\$1.18	\$1.16	\$1.13
Feedstock Cost [#]	\$ / Dry US Ton	\$109.01	\$98.31	\$92.70	\$87.82	\$87.82	\$70.15	\$70.15	\$70.15	\$70.31	\$70.31	\$70.31
Feedstock Moisture at Plant Gate	Wt % H ₂ O	10%	10%	10%	10%	10%	10%	10%	10%	10%	10%	10%
Feed Moisture Content to Pyrolyzer	wt % H ₂ O	10%	10%	10%	10%	10%	10%	10%	10%	10%	10%	10%
Energy Content (LHV, Dry Basis)	BTU / Ib	8,000	8,000	8,000	8,000	8,000	7,900	7,900	7,900	8,000	8,000	8,000
Pyrolysis and Vapor Upgrading		·										
Total Cost Contribution	\$ / GGE	\$2.34	\$2.03	\$1.84	\$1.46	\$1.10	\$1.14	\$1.14	\$1.06	\$1.13	\$1.12	\$1.14
Capital Cost Contribution	\$ / GGE	\$0.95	\$0.82	\$0.74	\$0.65	\$0.60	\$0.63	\$0.63	\$0.57	\$0.62	\$0.62	\$0.63
Operating Cost Contribution	\$ / GGE	\$1.39	\$1.21	\$1.09	\$0.80	\$0.50	\$0.51	\$0.51	\$0.49	\$0.50	\$0.50	\$0.51
Ex Situ Reactor Configuration	reactor type	Fluidized Bed	Fluidized Bed	Fluidized Bed	Fixed Bed	Fixed Bed	Fixed Bed	Fixed Bed	Fixed 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	2:2	2:2	2:2	2:2	2:2
Gas Phase	wt % of dry biomass	35%	36%	34%	31%	35%	38%	38%	34%	38%	38%	31%
Aqueous Phase	wt % of dry biomass	25%	25%	24%	27%	22%	24%	24%	26%	24%	24%	23%
Carbon Loss	% of C in biomass	2.9%	2.9%	3.4%	2.9%	5.0%	4.4%	4.4%	4.3%	4.4%	4.4%	3.0%
Organic Phase	wt % of dry biomass	17.5%	18.6%	21.8%	28.3%	27.9%	23.2%	23.2%	25.6%	23.4%	23.4%	31.4%
H/C Molar Ratio	ratio	1.1	1.1	1.1	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2
Oxygen	wt % of organic phase	15.0%	13.3%	16.8%	16.5%	18.6%	15.1%	15.1%	15.2%	15.1%	15.1%	16.4%
Carbon Efficiency	% of C in biomass	27%	29%	33%	42%	40%	35%	35%	39%	35%	35%	47%
Solid Losses (Char + Coke)	wt % of dry biomass	23%	21%	20%	14%	15%	14%	14%	14%	14%	14%	15%
Char	wt % of dry biomass	12.0%	11.0%	12.0%	10.4%	11.7%	11.6%	11.6%	11.6%	11.7%	11.7%	11.7%
Coke	wt % of dry biomass	11.0%	9.5%	8.3%	3.3%	3.7%	2.3%	2.3%	2.2%	2.3%	2.3%	3.2%
Vapor Quench, Co-Product Recovery + Contingency		_										
Total Cost Contribution	\$ / GGE	\$0.35	\$0.33	\$0.28	\$0.20	\$0.22	\$0.34	\$0.42	\$0.30	\$0.40	\$0.40	\$0.23
Capital Cost Contribution	\$ / GGE	\$0.20	\$0.19	\$0.16	\$0.12	\$0.13	\$0.22	\$0.26	\$0.19	\$0.24	\$0.24	\$0.13
Operating Cost Contribution	\$ / GGE	\$0.15	\$0.14	\$0.12	\$0.08	\$0.09	\$0.12	\$0.16	\$0.11	\$0.16	\$0.16	\$0.10

Table A-1. Processing Area Cost Contributions and Key Technical Parameters

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Hydroprocessing & Separation / Refinery Co-Processing												
Total Cost Contribution	\$ / GGE	\$0.33	\$0.31	\$0.34	\$0.35	\$0.38	\$0.30	\$0.21	\$0.30	\$0.21	\$0.21	\$0.04
Capital Cost Contribution	\$ / GGE	\$0.17	\$0.16	\$0.18	\$0.19	\$0.20	\$0.16	\$0.00	\$0.16	\$0.00	\$0.00	\$0.00
Operating Cost Contribution	\$ / GGE	\$0.15	\$0.14	\$0.16	\$0.16	\$0.18	\$0.14	\$0.21	\$0.14	\$0.21	\$0.21	\$0.04
Carbon Efficiency of Organic Liquid Feed to Fuels	%	88.4%	89.5%	87.2%	91.0%	89.0%	93.5%	93.5%	93.5%	93.5%	93.5%	91.0%
Hydrotreating Pressure	psia	2,000	2,000	2,000	1,900	1,900	1,900	1,900	1,900	1,900	1,900	1,900
Oxygen Content in Cumulative Fuel Product	wt %	0.8%	0.8%	0.8%	0.6%	0.5%	0.5%	0.5%	0.5%	0.5%	0.5%	0.6%
Hydrogen Production												
Total Cost Contribution	\$ / GGE	\$0.61	\$0.56	\$0.60	\$0.62	\$0.51	\$0.61	\$0.44	\$0.60	\$0.44	\$0.44	\$0.46
Capital Cost Contribution	\$ / GGE	\$0.39	\$0.36	\$0.38	\$0.41	\$0.33	\$0.39	\$0.28	\$0.37	\$0.28	\$0.28	\$0.28
Operating Cost Contribution	\$ / GGE	\$0.22	\$0.20	\$0.22	\$0.21	\$0.18	\$0.22	\$0.16	\$0.24	\$0.16	\$0.16	\$0.17
Additional Natural Gas (NG) at the Biorefinerv**	% of biomass LHV	0.3%	0.1%	0.2%	0.1%	0.3%	0.1%	0.5%	1.3%	0.6%	0.6%	0.2%
CoProducts	•	•										
Total Cost Contribution	\$ / GGE						(\$0.52)	(\$0.52)	(\$0.47)	(\$0.52)	(\$0.53)	(\$0.74)
Capital Cost Contribution *	\$ / GGE											\$0.06
Operating Cost Contribution •	\$ / GGE											(\$0.81)
CoProduct Credit	\$/GGE*						(\$0.52)	(\$0.52)	(\$0.47)	(\$0.52)	(\$0.53)	(\$0.83)
Balance of Plant												
Total Cost Contribution	\$ / GGE	\$0.04	\$0.07	\$0.03	\$0.20	\$0.23	\$0.27	\$0.20	\$0.21	\$0.20	\$0.19	\$0.22
Capital Cost Contribution	\$ / GGE	\$0.80	\$0.71	\$0.56	\$0.43	\$0.46	\$0.45	\$0.52	\$0.38	\$0.52	\$0.51	\$0.41
Operating Cost Contribution	\$ / GGE	(\$0.76)	(\$0.64)	(\$0.54)	(\$0.23)	(\$0.23)	(\$0.18)	(\$0.32)	(\$0.17)	(\$0.32)	(\$0.32)	(\$0.20)
Electricity Production from Steam Turbine (credit included in op.	\$/GGE*	(\$1.12)	(\$0,96)	(\$0,78)	(\$0.42)	(\$0.45)	(\$0,40)	(\$0.57)	(\$0.37)	(\$0.57)	(\$0.57)	(\$0.41)
cost above)				(11 - 17	(1) (1) (1) (1) (1) (1) (1) (1) (1) (1)	0.0.07	100 - 17	(1117)	() · · · /		(11 × 7	() · · · /
Sustainability and Process Efficiency Metrics	0//	40.7%	45.0%	40.5%	00.0%	00.0%	00.5%	00.5%	04.49/	00.7%	00.0%	04.0%
Fuel and Coproducts Field by Weight of Biomass	% W/W of dry biomass	13.7%	15.0%	16.5%	22.2%	20.9%	22.5%	22.5%	24.4%	22.7%	22.8%	24.8%
Carbon Emclency of Biomass to Fuels and Coproducts	% C in Feedstock	23.5%	25.9%	28.3%	38.1%	35.9%	37.2%	37.2%	40.5%	37.2%	37.3%	42.4%
Overall Carbon Efficiency to Liquid Hydrocarbon Fuels	% C In Feedstock	23.5%	25.9%	28.3%	38.1%	35.9%	33.0%	33.0%	36.3%	33.0%	33.0%	33.9%
Overall Energy Efficiency to Liquid Hydrocarbon Fuels	% LHV OF Feedstock	30.5%	33.4%	37.1%	50.3%	47.2%	43.6%	43.6%	48.0%	43.6%	43.6%	44.9%
Electricity Production	kWh/GGE	21.0	18.0	14.7	8.0	8.7	7.8	10.8	6.3	10.8	10.7	7.9
Electricity Consumption (Entire Process)	KWN/GGE	12.7	11.0	9.6	6.4	7.5	7.4	7.1	6.7	7.1	7.0	7.4
Water Consumption in Conversion Process	gal H ₂ O/GGE	1.4	1.4	1.3	1.5	1.4	1.7	1.2	1.5	1.2	1.2	1.4
TEA Reference File		PyVPU-v218g ES - Fluidized Bed - FY14 SOT (2016\$)- V02d-feed- 109.01.xlsm	PyVPU-v218g ES - Fluidized Bed - FY15 SOT (2016\$)- V02d-feed- 98.31.xlsm	PyVPU-v218g ES - Fluidized Bed - FY16 SOT (2016\$)- V02d-feed- 92.70.xlsm	PyVPU-v218h ES FixedBed- v49-r046-AP10 FY17SOT (2016\$)-V02d- b.xlsm	PyVPU-v218h ES FixedBed- v49-r046-AP10- FY18SOT (2016\$)-V09d- Revised 40.4pct C-Eff- b.xlsm	PyVPU-v218h ES FixedBed- v49-r046-AP10- FY19SOT-V18 Acetone-MEK- \$3.33.xlsm	PyVPU-v218h ES FixedBed- v49-046-AP10- FY20Target- V18-Acetone- MEK- HTcoproc-03b- \$3.09.xlsm	PyVPU-v218h ES FixedBed- v49-r046-AP10- FY20Target- V18-Acetone- MEK-3pct- higher-C-eff-03- \$3.09.xlsm	PyVPU-v218h ES FixedBed- v49-r046-AP10- FY21Target- V18-Acetone- MEK- HTcoproc-03b- Low Ash- \$3.05.xlsm	PyVPU-v218h ES FixedBed- v49-r046-AP10- FY22Target- V18-Acetone- MEK- HTcoproc-03b- Low Ash- \$3.00.xlsm	PyVPU-v218h ES FixedBed- v49-r046-AP10- Prelim 2030 Target from 2022-20% mixed phenolics-no HT capital (2016\$)-V06d- b.xlsm

^AConceptual design result. [†]SOT: State of Technology. ^{*}Gallon Gasoline Equivalent (GGE) on a Lower Heating Value (LHV) basis. ^{**} Natural gas stream was negligible in most of the biorefinery models. This was included to maintain model flexibility to allow natural gas use as an option. ^{#‡}2030 projections are based on high-level estimates and will be modeled in detail in future years. It is proposed that co-hydroprocessing of CFP oil will occur at a petroleum refinery. Capital for hydrogen production is included, while natural gas feed for hydrogen production is not included because credit is not taken for an equivalent amount of fuel gas from the CFP biorefinery. Coproduct credit is based on a preliminary estimate of diverting 20% CFP Oil to produce coproducts, including from the organic liquid phase. [•]Capital and operating costs for coproduct recovery in the 2019-2022 models are included in the "Vapor Quench, Co-Product Recovery + Contingency" section. ^{††} Modeled ash is 1.75% for 2019 and 2020, and less than 1% for all other years. ^{II} An additional biomass heater is included as a small additional in-plant cost, as shown in <u>https://www.nrel.gov/docs/fy15osti/62455.pdf</u>. ^{II} For the 2017 SOT unquantified portion of CFP yields were prorated to solids, liquids, and gases using measured yields. ^{#†} Small adjustments made to previously published feedstock cost estimates for 2014-2016.