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Catalytic Fast Pyrolysis Oil Stand-Alone and Co-Hydroprocessing Case Studies: Design, Cost, and Sustainability Based on Process Model Predictions

Abhijit Dutta,[1] Michael Talmadge,[1] Kristiina Iisa,[1] Cody Wrasman,[1] Xiaolin Chen,[1] Michael Griffin,[1] Calvin Mukarakate,[1] Suphat Watanasiri,[2] Longwen Ou,[3] and Hao Cai[3]

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Preface

Refined liquid fuels provide convenience for road, rail, air, and marine transportation due to their high energy density, ease of transport and storage, and existing production and distribution infrastructure. Biomass conversion technologies can help diversify and increase the supply of liquid fuels toward a more robust energy future. The backbone of such development must include reliable, high-volume biomass supply chains and efficient utilization of those resources. These activities, if pursued, will lead to rural infrastructure development and new jobs.

In this study we focus on woody biomass conversion via catalytic fast pyrolysis (CFP) to produce CFP oil, a partly stabilized oxygenated liquid. The CFP oil fuel intermediate can then be hydroprocessed to produce liquid hydrocarbon fuels compatible with existing petroleum fuel infrastructure. CFP involves fast pyrolysis with the rapid heating of biomass to ~500°C to produce vapors that are catalytically upgraded to increase the stability of the CFP oil produced after vapor condensation. The projections in this report are based on bench-scale experimental results used in conceptual process models, assuming the successful scale-up of bench-scale performance. A well-developed biomass supply chain and mature CFP technology are assumed for future projections; results are influenced by the underlying assumptions presented. Uncertainties associated with pioneer plants are not addressed in this report. Thus, the results represent the future potential of this technology pathway, and not its current status. The production and transport of biomass, along with preprocessing for compatibility with the CFP process and conversion to CFP oil, are the most expensive steps in this pathway; hence, related R&D can open the biggest opportunities for cost reduction.

The focus of this report is on understanding process implications and associated costs of introducing CFP oils into hydroprocessing units. Hydroprocessing is a family of well-established petroleum refinery operations involving reactions of petroleum feedstocks with hydrogen. These operations help refiners achieve fuel quality and quantity goals based on market demand and associated fuel regulations. Leveraging the existing extensive petroleum infrastructure to process CFP oils can be economically advantageous. However, the differences in properties of incumbent petroleum feedstocks with CFP oils, including a significant proportion of constituent oxygen in CFP oils, require careful R&D to understand the changes necessary when CFP oils are introduced in hydroprocessing operations designed to handle petroleum feedstocks. CFP oil hydroprocessing, both with 100% CFP oil and coprocessing with petroleum feedstocks, is a topic of ongoing research. Additional public domain experimental data are necessary to develop a more comprehensive understanding of reaction behavior in various setups; this can help develop predictive reactor models that can accelerate process development, scale-up, and refinery adoption of CFP oils. This report does not address the required experimental efforts, but explores other process requirements, such as changes necessary to control the reactor temperature because of the different exothermic behavior of CFP oils compared to petroleum feedstocks.

In summary, commercial advancement of this biomass conversion pathway requires (1) the development of an efficient, high-volume supply chain that addresses biomass feedstock quality requirements for CFP; (2) large-volume, reliable, and efficient production of CFP oils from biomass; and (3) further experimentation to understand CFP oil reaction behavior in hydroprocessing units to facilitate adoption by refiners after making necessary changes to address risks.

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National Renewable Energy Laboratory	Zia Abdullah Frederick Baddour Robert Baldwin
Pacific Northwest National Laboratory	Huamin Wang
Alder Renewables	Daniel Szeezil Derek Vardon
BP	Mary Biddy Casey Hetrick John Shabaker
BTG Biomass Technology Group BV	Robbie Venderbosch
Johnson Matthey	Xavier Baucherel Luke Tuxworth
Par Pacific	William Mok
POET	Dave Carlson

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We thank KBR for providing capital cost estimates for equipment in the hydroprocessing section, under a subcontract. Estimates provided by KBR are for informing a broad set of analysis and do not reflect any specific industrial installation. Estimates for every industrial implementation will need to be considered based on their unique assumptions to come up with accurate and specific estimates.

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List of Acronyms

bbl	barrel
BDE	barrels diesel equivalent
BGE	barrels gasoline equivalent
BJE	barrels jet equivalent
BPD	barrels per day
Btu	British thermal unit
CAPEX	capital expense
CFP	catalytic fast pyrolysis
CFP-A	CFP Case A presented in this report
CFP-B	CFP Case B presented in this report
CH_4	methane
CI	carbon intensity
CO	carbon monoxide
CO_2	carbon dioxide
CO_2e	carbon dioxide equivalent
COD	chemical oxygen demand
CoMo	cobalt molybdenum (catalyst)
DCFROR	discounted cash flow rate of return
DCN	derived cetane number
ΔT	delta temperature, or temperature difference
FCC	fluid catalytic cracking
FCI	fixed capital investment
$f_{install}$	installation factor
f_{Lang}	Lang factor
FP	fast pyrolysis
gal	gallon
GDE	gallon diesel equivalent
GGE	gallon gasoline equivalent
GHG	greenhouse gas
GJE	gallon jet equivalent
GREET	Greenhouse gases, Regulated Emissions, and Energy use in Technologies
h	hour
H_2	hydrogen
H_2O	water
H_2S	hydrogen sulfide
HC	hydrocrack
H:C	hydrogen-to-carbon molar ratio
HDN	hydrodenitrogenation
HDS	hydrodesulfurization
HEFA	hydroprocessed esters and fatty acids
HP	high pressure
HT	hydrotreat
HT+HC	hydrotreating followed by hydrocracking
H-TC	inter-bed hydrogen quench temperature control (for hydroprocessing)
ICN	indicated cetane number

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INL	Idaho National Laboratory
k	kilo (multiplied by 1,000)
kg	kilogram
kW	kilowatt
kWh	kilowatt-hour
lb	pound (mass)
LCO	light cycle oil
LHV	lower heating value
LP	low pressure
L-TC	inter-bed liquid product recycle temperature control (for hydroprocessing)
MEK	methyl ethyl ketone or 2-butanone
MFISP	minimum fuel intermediate selling price
MFSP	minimum fuel selling price
MJ	megajoule
MM	million (e.g., MMBtu or $__MM)
mol %	mole percent
MP	medium pressure
MT	million dry short tons (of biomass feedstock)
MW	megawatt
N_2O	nitrous oxide
NH_3	ammonia
NiMo	nickel molybdenum (catalyst)
NiW	nickel tungsten (catalyst)
NREL	National Renewable Energy Laboratory
OPEX	operating expense
PSA	pressure swing adsorption
R&D	research and development
RON	research octane number
scf	standard cubic feet
SRD	straight-run diesel
TEA	techno-economic analysis
TPEC	total purchased equipment costs
VGO	vacuum gas oil
VLP	very low pressure
vol %	volume percent
WGS	water-gas shift
WHSV	weight hourly space velocity
wt %	weight percent
y	year

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Executive Summary

Refined liquid fuels provide convenience for road, rail, air, and marine transportation due to their high energy density, ease of transport and storage, and existing production and distribution infrastructure. Distillate-range hydrocarbon fuels are expected to continue being the primary fuel for most transportation modes in the United States. In the case of aviation, liquid jet fuels will remain indispensable for the foreseeable future. Diversification of refinery feedstocks and co-utilization of domestic biomass feedstocks for liquid fuels can be important aspects of a robust energy future.

In this report we explore catalytic fast pyrolysis (CFP) oil as a potential refinery feedstock that can be utilized to supplement petroleum sources toward the supply of infrastructure-compatible liquid fuels. CFP oils require deoxygenation to achieve hydrocarbon fuel standards and specifications, as well as for compatibility with any additional downstream petroleum refinery processing. The substantial capital and infrastructure in petroleum refineries and the associated fuel distribution network can play important roles in the practical adoption of this new feedstock source. We present cases with conceptual model results based on bench-scale experimental performance observed to date in the production and hydroprocessing of CFP oils. The overall scope of this study is shown in Figure ES-1, starting from the biomass source and ending with the production of liquid transportation fuels.

Discussion on the production of CFP oils is presented in Section 2; previous work on this topic is leveraged to present two cases that produce CFP oils with different oxygen contents (CFP-A with 21.6 wt % oxygen and CFP-B with 16.0 wt % oxygen). Both cases are based on bench-scale experiments at the National Renewable Energy Laboratory (NREL). CFP-B provides better yields and economics compared to the more near-term CFP-A technology. CFP-B is thus considered a future case and CFP-A a near-term case. The conceptual design results presented reflect the potential for this approach in a future scenario with a mature and efficient biomass supply chain in the United States, along with the successful and reliable scale-up of CFP technologies.

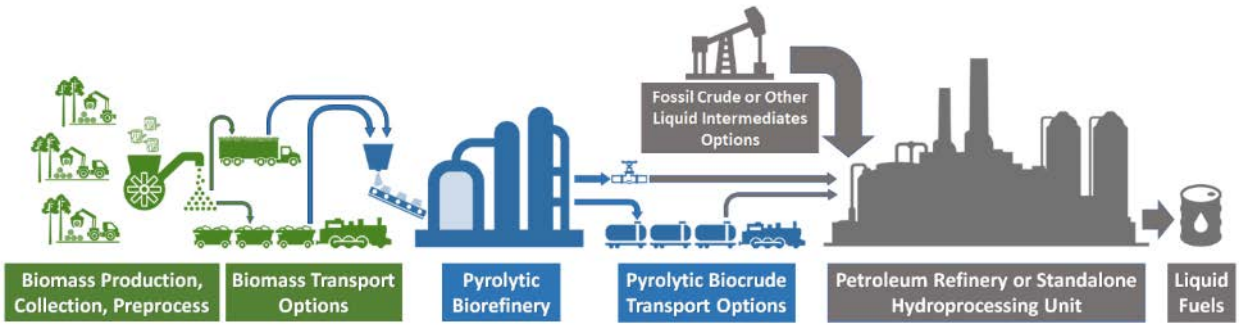

Figure ES-1. Scope of this field-to-fuel study: biomass production and transport, CFP oil production from biomass and transportation to a hydroprocessing facility, and the production of infrastructure-compatible liquid fuels.

Section 2 is followed by a study of CFP oil hydroprocessing cases in Section 3, which is the primary focus of this report. Process models serve as the basis for all analysis of CFP oil hydroprocessing. The models are exercised to highlight anticipated design and operational needs for (1) new units constructed for stand-alone hydroprocessing and (2) understanding the impacts

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on existing refinery hydroprocessing units if they are adapted to include CFP oils as a new feedstock. Potential design and operational impacts are explored for different feed combinations using a few simplified hydroprocessing configurations. Modeled greenhouse gas (GHG) emissions are also quantified as part of the analysis. This report does not focus on experimental research on catalyst development and other R&D work necessary for long-term operability of the hydroprocessing units with CFP oil feed. The process model assumptions are based on extrapolating short-duration hydroprocessing experiments mostly conducted using commercial petroleum refining hydroprocessing catalysts. These experiments have shown promise of feasibility with respect to achieving fuel quality compatible with petroleum-derived products.

Some of the biggest biomass feedstock challenges are the aggregation and processing of large quantities of solid biomass situated at dispersed locations. This report assumes separate facilities for the production of CFP oil, located closer to biomass resources, and potentially large, centralized hydroprocessing facilities capable of handling CFP oil aggregated from multiple CFP oil production facilities. Note that this analysis does not consider any policy incentives because those incentives can vary with time and location. It also does not consider the limitations of a less developed biomass supply chain or the technical challenges that need to be overcome in pioneer plants; these aspects will negatively impact the economics of initial implementations and need to be addressed via both R&D and industrial scale-up. Most financial assumptions are consistent with previous reports sponsored by the Bioenergy Technologies Office; details are provided in the main report along with any changes made to previous assumptions. Minimum selling prices are reported for the CFP oil "fuel intermediate" (minimum fuel intermediate selling price [MFISP]) and the final "fuel" (minimum fuel selling price [MFSP]) after CFP oil hydroprocessing. A discounted cash flow rate of return (DCFROR) method is the basis for price estimates after accounting for inputs, outputs, capital costs, etc. based on process model flows. The minimum selling prices and other costs are reported in 2020 U.S. dollars. Yields are reported on energy and volume bases in this report; alternate energy units are included in the summary tables. The common use of gallons gasoline equivalent (GGE) for energy quantification does not imply that gasoline is the primary fuel product; it is presented because it is a relatable metric in the United States.

Production of CFP Oils and Associated Modeling

In fast pyrolysis (FP), biomass is rapidly heated to ~500°C to break down its constituent polymeric structures into smaller molecules. These molecules include permanent gases, condensable vapors, and solid char. The vapors yield a liquid product upon cooling, and this liquid is of primary interest for conversion to liquid fuels. However, the liquid is unstable because it retains significant proportions of oxygen from its parent biomass feedstock (~40 wt % oxygen in woody biomass) in the form of reactive functional groups. CFP includes a catalyst within the FP reactor (*in situ* CFP) or in a separate reactor (*ex situ* CFP) following the FP reactor; CFP helps upgrade the vapors by reacting and reducing the oxygen content in a catalytic environment. The *ex situ* configuration provides advantages allowing tailored reaction conditions independent of the FP reactor and processes only the vapor/gas phase after almost all solids in the FP reactor are removed via cyclones and other equipment; removal of solids is advantageous for catalyst performance and maintenance. In this report we consider two different *ex situ* CFP conceptual designs for potential implementation: a near-term catalyst capability case (CFP-A) with ZSM-5 catalyst in a fluid bed and a potential future case (CFP-B) with improved catalyst capabilities for higher CFP oil yields, along with the production of valuable chemical coproducts

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with Pt/TiO_2 catalyst in a fixed bed. Assumptions in the models are based on bench-scale experimental data. A simplified block flow diagram for *ex situ* CFP is shown in Figure ES-2.

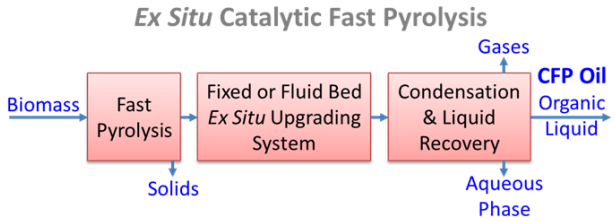

Figure ES-2. Simplified block flow diagram for *ex situ* CFP

CFP oils are considered fuel intermediates that can be hydroprocessed to produce hydrocarbon fuels. Some of the assumptions, metrics, and results from the analysis for the production of CFP oils are included in Table ES-1. The respective modeled MFISPs for CFP-A and CFP-B are $216/barrel and $148/barrel, with 800 miles of rail transportation included for the CFP oil.

Table ES-1. Metrics for CFP Oil Production for Near-Term (CFP-A) and Future (CFP-B) Catalysts With the Assumption of a Scaled-Up Efficient Biomass Supply Chain and Mature CFP Technology

Metric	CFP-A	CFP-B
Woody biomass plant size, dry metric tonnes/day	2,000	2,000
CFP oil production, barrels/day	2,225	2,870
Biomass to CFP oil energy efficiency, % lower heating value (LHV) basis	32.9%	42.2%
CFP oil production, gal/dry U.S. ton (barrels/dry U.S. ton)	42.4 (1.01)	54.7 (1.30)
Coproducts	Electricity	Electricity, chemicals
CFP oil oxygen content, wt %	21.6%	16.0%
CFP oil hydrogen-to-carbon molar ratio (H:C)	1.18	1.30
CFP oil LHV, Btu/lb	12,408	13,908
CFP oil specific gravity	1.151	1.053
CFP oil energy density, GGE/gal	1.026	1.053
MFISP with 800 miles of rail transport, $/GGE	$5.00	$3.35
MFISP with 800 miles of rail transport, $/barrel	$215.62	$148.30
MFISP without CFP oil transportation, $/barrel	$203.92	$137.60
GHG emissions for biomass production and preprocessing, CFP oil production with 800 miles rail transport, g CO_2e/MJ	6.87	8.88

Note: GGE has an energy value of 116,090 Btu.

For success of this technology, (1) the development of a commoditized and efficient biomass supply chain will be important to ensure a reliable supply of biomass feedstock with requisite quality, and (2) CFP oil yield improvements through R&D, development of valuable coproducts, and scale-up efforts will be necessary to increase yields, lower costs, reduce risks, and enable maturation and scale-up. Colocation of CFP oil production facilities with other industrial units can help increase options for sharing utilities and allow more optimal use of off-gases, thermal

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energy, and other coproducts from different CFP processes.[1] This can help improve both economics and sustainability while also leveraging existing industrial logistics to reduce costs and infrastructure uncertainties.

Hydroprocessing CFP Oils and Associated Modeling

In this report, the term hydroprocessing represents the broad family of refining processes involving hydrotreating and hydrocracking. Hydroprocessing has traditionally been applied in refining operations with petroleum feedstocks to improve product quality of refinery blendstocks by (1) removing impurities such as sulfur, nitrogen, oxygen, and metals; (2) saturating olefinic and aromatic compounds; (3) isomerizing long carbon chains; and (4) cracking to optimize product splits.[2] Hydroprocessing oxygenated liquid intermediates from biomass and waste organics has been a topic of growing interest. Section 3 of this report presents modeled process metrics for CFP oil hydroprocessing, both in a stand-alone configuration and with petroleum feedstocks like straight-run diesel (SRD), light cycle oil (LCO), and vacuum gas oil (VGO). The petroleum refinery-centric approach applied here is intended to complement prior biorefinery-centric analyses[3] and highlight process metrics and potential benefits of co-hydroprocessing CFP oil. Refinery capital investment requirements for both stand-alone and coprocessing units were estimated with input from the engineering company KBR (through a subcontract). Some relevant hydroprocessing experimental data are included for reference (Appendix A). However, not enough information is available at this point to allow the development of reactor models that can predict yields for different choices of process conditions. In the absence of predictive reactor models, we made a set of base assumptions for reactor performance. Those base assumptions, guided by experimental data, include complete deoxygenation and 90% carbon efficiency during the conversion of CFP oils to hydrocarbon fuels. Potential partial deoxygenation toward the production of marine[4] and other fuels is not discussed in this report. A hydrogen-to-carbon molar ratio (H:C) of ~1.8 for the total fuel product after hydroprocessing (for all liquid feedstock combinations, both petroleum and CFP oils) was assumed to allow consistent interpretation of all the results. The assumptions of nominal reaction temperatures and pressures in the models should not be viewed in the context of reactor performance and yields. We explore sensitivity cases to present modeled process impacts of variations from the various base assumptions. Two simplified reactor configurations are considered in this report. For SRD feed we consider a hydrotreater, while we include a pre-hydrotreater followed by a hydrocracker when modeling the heavier feedstocks, LCO and VGO.

[1] A. Dutta, et al. 2023. "Assessment of location and energy utility options for the implementation of pyrolytic biocrude production." *Sustainable Energy Fuels* 7: 4955–4966. doi.org/10.1039/D3SE00745F.
[2] J. H. Gary, G. E. Handwerk, and M. J. Kaiser. 2007. *Petroleum Refining: Technology and Economics*. Fifth Edition, CRC Press.
[3] A. Dutta, et al. 2023. "Model quantification of the effect of coproducts and refinery co-hydrotreating on the economics and greenhouse gas emissions of a conceptual biomass catalytic fast pyrolysis process." *Chemical Engineering Journal* 451(1): 138485. doi.org/10.1016/j.cej.2022.138485.
[4] X. Chen, et al. 2025. "Sustainable marine fuel production through mild hydrotreating of catalytic fast pyrolysis bio-oil." *Fuel*, 398: 135455. doi.org/10.1016/j.fuel.2025.135455.

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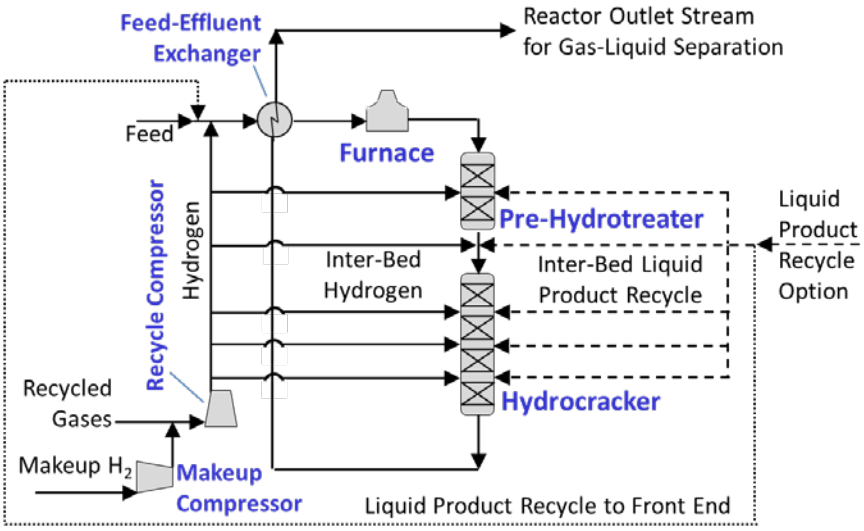

Figure ES-3. Configuration with a pre-hydrotreater followed by a hydrocracker. Reactor temperature control options using inter-bed hydrogen and inter-bed and front-end liquid product recycle are discussed in the report; not all options may be exercised within the same setup.

Reactor heat management is a key concern when CFP oils are introduced in conventional hydroprocessing designs. This concern is due to the significantly higher exotherm when processing CFP oils that involve two major classes of exothermic reactions: (1) hydrogenation and (2) hydrodeoxygenation. Based on existing design options in petroleum refineries, we explored cooling via inter-bed quench using (1) hydrogen and/or (2) liquid product recycle to the reactor (Figure ES-3). These configurations are referred to as hydrogen temperature control (H-TC) and liquid product recycle temperature control (L-TC). An option to recycle product liquid to only the front end of the reactor is also explored in the main report, as depicted among the options in Figure ES-3.

Space velocity adjustments are necessary when handling CFP oils vs. petroleum feedstocks. We adjusted weight hourly space velocities (WHSVs) based on chemical hydrogen demand (net stoichiometric H_2 consumed by the chemical reactions; this can also be calculated by subtracting reactor outlet H_2 from inlet H_2) because chemical hydrogen use can be representative of reaction demands on the catalyst. Chemical hydrogen consumption in our 100% SRD hydrotreater model, 615 scf/barrel of fresh SRD feed, and a corresponding WHSV of 2 h^{-1} was used as the basis. WHSV for other cases were prorated based on their modeled chemical hydrogen demands; thus, lower space velocities were used for the higher chemical hydrogen demand cases to account for more reaction time needed. This generic approach was adopted to allow consistency of WHSV assumptions across the broad range of cases explored; the availability of additional experimental data may help guide more specific estimates in the future.

Modeled product distillation includes three cuts: (1) gasoline, (2) jet, and (3) diesel/heavies. The distillation model was biased toward maximizing the jet cut because of the favorable jet fuel properties imparted by constituent cycloalkanes in woody biomass CFP-derived fuels and the projected continued growth in jet fuel demand; modeled distributions among the cuts are included in Appendix K. We emphasize the energy value of the total liquid product and not the distribution among the three cuts because the overlap of boiling ranges allows changing the cuts based on refinery objectives. The modeled cuts are based purely on boiling ranges; matching fuel

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properties of the product streams in the models with the desired product cuts was not a focus in the current work but can be revisited with the availability of more experimental data and predictive reactor models. For simplicity, only a fuel gas coproduct is assumed from the gas-phase outputs from hydroprocessing.

Life cycle GHG emissions are calculated by combining results for CFP oil production with additional contributions from hydroprocessing and final fuel combustion. All these calculations were guided by Argonne National Laboratory's Greenhouse gases, Regulated Emissions, and Energy use in Technologies (GREET) tool.[5] As expected, hydrogen consumption is the dominant contributor toward GHG emissions during hydroprocessing. We assumed hydrogen production via natural gas steam reforming (without any carbon capture) for use during hydroprocessing. Other options for large industrial-scale hydrogen production may become available in the future and help reduce GHG emissions.

Process parameters and some simple configuration variations for hydroprocessing were explored to help identify impacts of alternate assumptions in the models. The discussion in Section 3.2 and results presented in Section 3.4 may help provide approximate estimates for different potential implementation strategies:

- Stand-alone hydroprocessing of CFP oils.
- Repurposing refinery units for processing oxygenated feedstocks.
- Coprocessing CFP oils with petroleum feedstocks in existing refinery units, with minimal necessary modifications within existing constraints.
- Augmenting refinery units to alleviate constraints for different proportions of oxygenated liquid intermediates like CFP oils.

Catalyst development may improve hydroprocessing yields for specific CFP oil feedstocks in the future, but that aspect is not explored in this report. Dedicated experimental research is necessary for catalyst improvements; that aspect cannot be addressed via the process modeling-centric approach in this work. We assume the use of commercially relevant catalysts currently used for petroleum feedstocks; these catalysts have been used in many of the bench-scale CFP oil hydroprocessing experiments so far in both stand-alone and co-hydroprocessing configurations, with positive results reported in many short-term tests (Appendix A). Some process adjustments that may be necessary are mentioned in this report, but the list of adaptations needs to expand with improved understanding of this process; feed system modifications are also called out. Estimated costs for equipment modifications, because of changes in process variables like flow rates, are also factored into cost estimates for each case (included in Appendix K).

Hydroprocessing Results

Figure ES-4 summarizes estimated costs from CFP oil, petroleum feedstocks, hydrogen consumption, and remaining contributions when various volumetric proportions of CFP-A and CFP-B oils are processed with SRD in a hydrotreating configuration, and with a 50/50 volumetric mixture of LCO/VGO in a pre-hydrotreating followed by hydrocracking configuration. The modeled results below are with H-TC.

[5] M. Wang et al. 2023. *Summary of Expansions and Updates in R&D GREET 2023*. greet.anl.gov/publication-greet-2023-summary.

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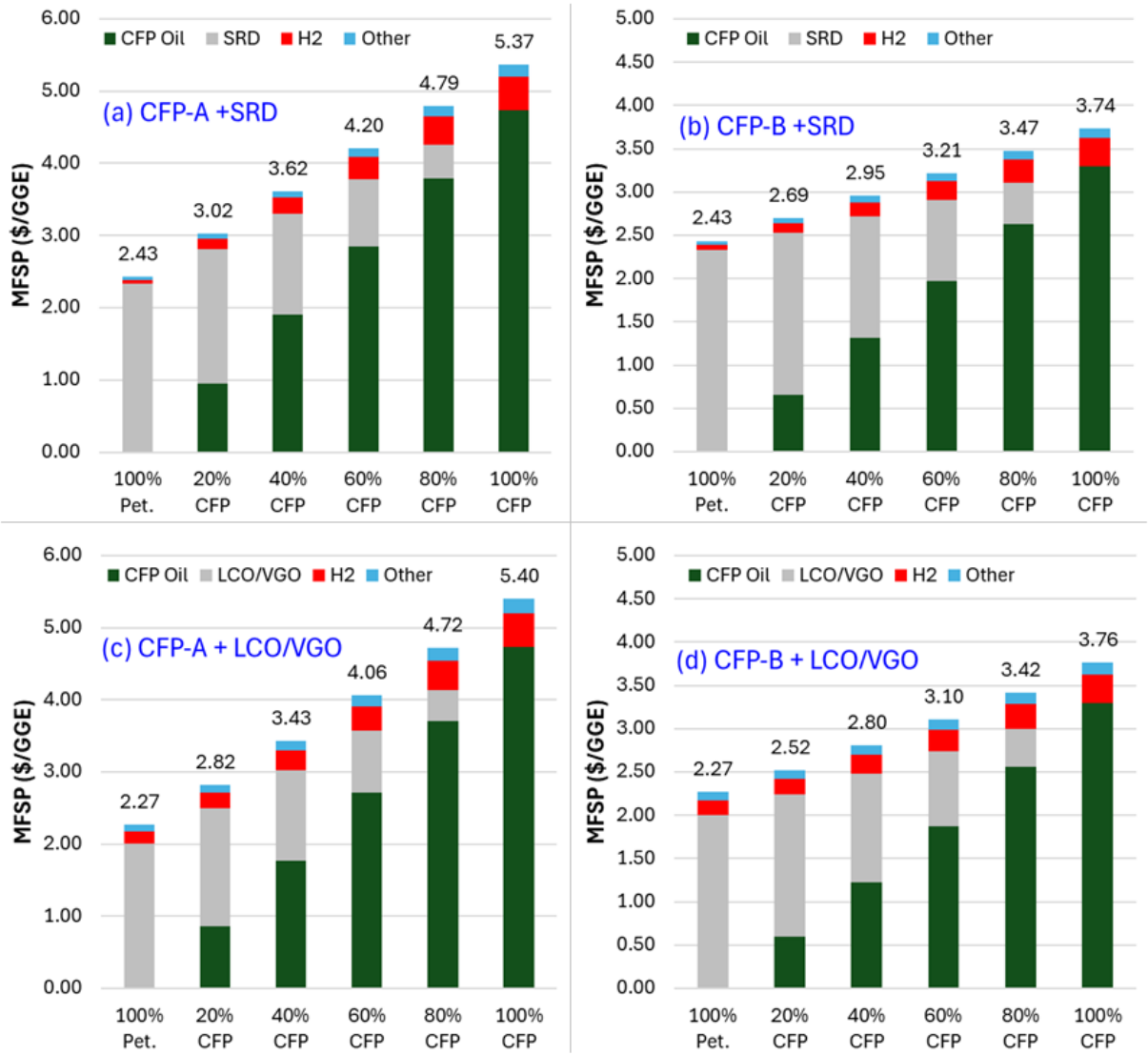

Figure ES-4. Estimated MFSPs with cost contributions from CFP oil feed, petroleum (Pet.) feed, H_2 input, and other costs for (a) CFP-A oil with SRD, (b) CFP-B oil with SRD, (c) CFP-A oil with LCO/VGO, and (d) CFP-B oil with LCO/VGO in 75k barrels per day (BPD) systems. Further details in Section 3.4.3.

Figure ES-4 shows the dominance of CFP oil and petroleum feedstock costs along with hydrogen costs in all cases at a large scale (75k BPD). Other scales and assumptions of new vs. repurposed units are explored in the main report, and the results show that there is a $0.01–$0.02/GGE benefit to the MFSP from larger capacities within the 30k–75k BPD range. Further capacity reductions to the 3k–20k BPD range result in larger increases in the MFSP. Practical implementations for CFP oil coprocessing will likely include liquid product recycle to the reactor to maintain operating parameters such as gas and liquid flows within the design constraints of existing equipment or to minimize deviations from conventional design parameters. Note that any liquid recycle will likely require a corresponding curtailment of fresh liquid feed input. Sample results for two cases with 20% CFP oil and 80% SRD feed in an inter-bed liquid recycle configuration are shown in Figure ES-5. The H-TC cases include 20 vol % liquid recycle with fresh feed curtailed from 75k BPD to 60k BPD. The amount of liquid recycle for the L-TC cases

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was determined by a temperature control design specification in the model; in this case the fresh feed was curtailed to 37k BPD for 20% CFP-A/80% SRD feed and 43k BPD for 20% CFP-B/80% SRD feed. More liquid recycle and fresh feed curtailment were necessary for CFP-A because of the higher oxygen content of CFP-A oil (21.6%) vs. CFP-B oil (16.0%). Removal of more oxygen increases the heat generated during exothermic hydrodeoxygenation.

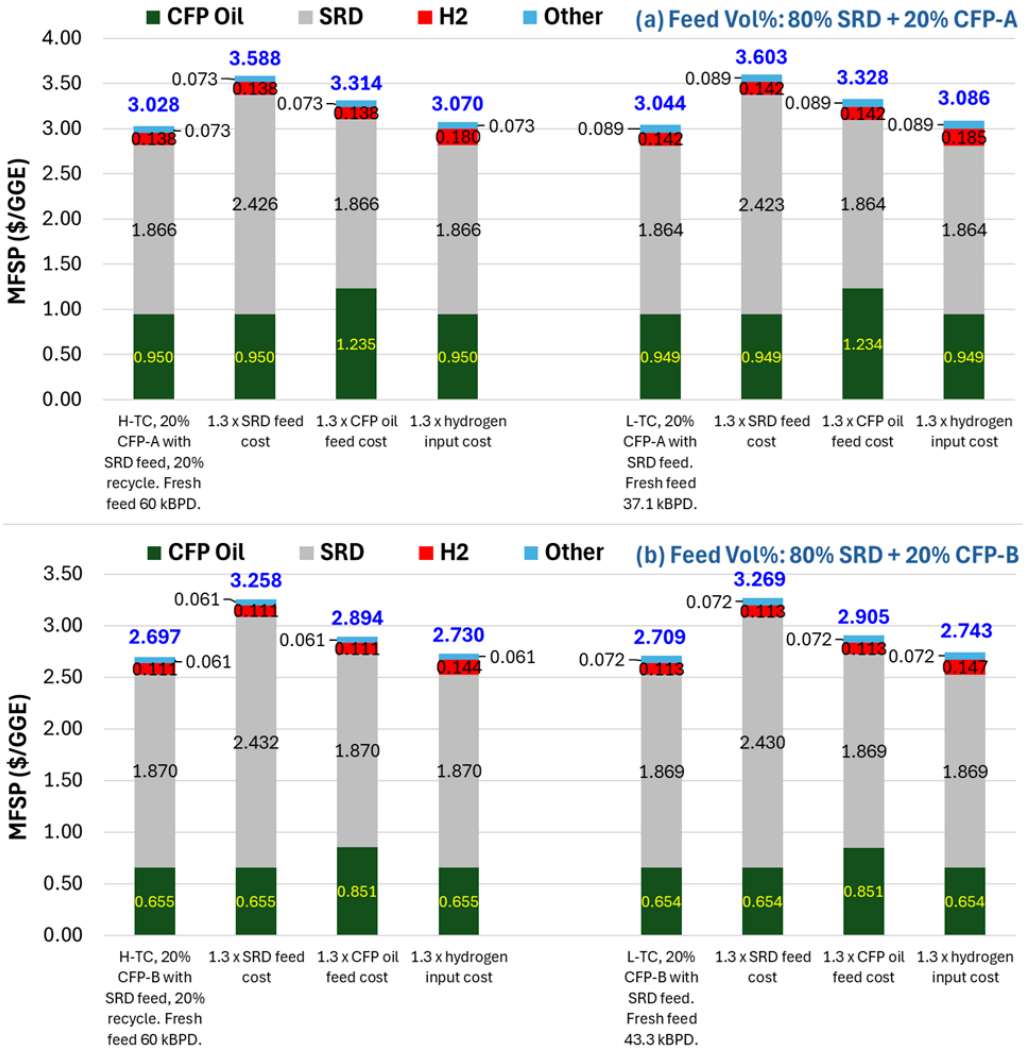

Figure ES-5. MFSPs upon changing unit costs of the three major contributors: (1) petroleum feed, (2) CFP oil, and (3) hydrogen by 30% (1.3×) each in both H-TC (sets of four on the left) and L-TC (sets of four on the right) configurations. Unit cost variation impacts for these three major cost categories can be estimated by simple proration. Results shown for (a) 20 vol % CFP-A oil + 80% SRD feed and (b) 20 vol % CFP-B oil with 80% SRD feed. Total liquid feed, including liquid recycle, is 75k BPD. Further details are available in Section 3.4.4.15.

The dominance of the three major contributors is further illustrated in Figure ES-5; as shown, impacts of changes in unit cost can be estimated by simple prorations within each of those three cost categories. The choices of refinery feedstocks and products, based on current market prices, are the biggest levers for increasing refinery margins. The required hydrogen input will largely be determined by chemical hydrogen consumption but may be influenced to a small extent by system design/operation choices. After those choices are made, additional improvements to the

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margins are typically dictated by the small "Other" category, which includes capital charges; other variable operating costs such as utilities, raw materials, and waste treatment; and fixed operating costs that include maintenance costs, salaries, benefits, and insurance. In addition to margins, the refinery on-stream factor and capacity utilization can have major economic impacts; whenever margins are positive, lower throughputs due to reasons such as equipment downtime or feed curtailment result in lost opportunity for increasing net revenue.

The small margins in play under the "Other" category implies that the total cost contribution of the other parameters explored in this report is relatively small per gallon for large-scale operations. Refinery unit designs and their operations can influence the "Other" category to impact the bottom line; although changes in this category may be small per gallon, the overall impact for the refinery unit can be substantial when large volumes are processed.

More than 100 hydroprocessing cases with process/parameter variations were explored in this report; these variations primarily impact the "Other" category and, in some cases, the hydrogen consumption. These sensitivity cases helped quantify the impacts of (1) control strategies; (2) CFP oil oxygen content; (3) petroleum feedstock types; (4) different proportions of CFP oil and petroleum feed during coprocessing; (5) hydrogen purity; (6) carbon efficiencies during CFP oil conversion; (7) space velocities and catalyst lifetime; (8) H:C in the total liquid product; (9) reactor temperature, pressure, and allowable temperature increase in the reactor (ΔT); (10) scale; (11) capital and key operating costs; (12) new stand-alone units; (13) repurposed units; and (14) on-stream factor. Because the reactor model is not predictive, each of these sensitivity cases reflects the effect of single parameter variations. These results are reported in Section 3.4.4.

Conclusion

This study provides high-level cost estimates for the production of infrastructure-compatible liquid hydrocarbon fuels from woody biomass via CFP followed by hydroprocessing of the CFP oil. The impacts of many design/operating parameters are assessed through process modeling. The focus of this report is on the hydroprocessing of CFP oils. While this report provides process insights into a broad range of hydroprocessing options, there remain many gaps that need to be addressed via experimental research, scale-up, and situation-specific risk assessments in order to advance CFP oil stand-alone and co-hydroprocessing technologies. R&D is necessary for enabling higher yields and valuable coproducts from CFP, along with increased process reliability. Reliable production of reasonably priced CFP oil is the most critical step for commercial success because CFP oil hydroprocessing has a smaller cost contribution compared to the production of CFP oil. Finally, the development of a robust biomass supply chain is a bottleneck for CFP technology implementation and for increasing CFP oil supply volumes for related hydroprocessing scale-up. Results presented in this report are for conceptual mature plants with well-developed biomass supply chains, and do not incorporate the effects of pioneer plant uncertainties that will increase costs in initial implementations. The benefits of successful development of this pathway can be multifaceted, reflecting the diversity of technologies and logistics involved. Furthering biomass resource and supply chain development can lead to rural infrastructure development and the creation of associated jobs. In the long run, the addition of CFP oils to the portfolio of liquid feedstocks for refineries can help augment energy options with domestic U.S. resource utilization.

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Section 1
Introduction

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1 Introduction

1.1 Motivation and Background

Pyrolysis processes for biomass conversion hold promise to produce significant volumes of oxygenated liquid intermediate feedstocks that can be further hydroprocessed to infrastructure-compatible liquid fuels, including gasoline, jet, diesel, and marine fuel. Fats, oils, and greases constitute the bulk of alternate liquid feedstocks that are currently hydroprocessed to liquid hydrocarbon fuels. It is projected in the near term that fats, oils, and greases and oilseed crops can help meet the U.S. goal to produce 3 billion gallons of alternate jet fuel by 2030 [1]. However, the projected supply growth of these resources is limited. It is necessary to diversify and include additional liquid feedstocks/fuel intermediates for refining to augment the contribution of alternate sources to the future U.S. fuel pool. Biomass and organic waste-derived pyrolysis liquid intermediates have the potential to boost that supply pool. Although pyrolysis pathways are not currently ASTM approved for jet fuel production, the potential abundance of cycloalkanes in the product after hydroprocessing catalytic fast pyrolysis (CFP) oils holds promise for future inclusion in jet fuel, in addition to use as fuel for terrestrial and marine vehicles.

From a biomass supply and availability perspective, conversion technologies will compete for the utilization of existing biomass resources. Technologies that prove higher efficiencies with lower production costs will be the preferred consumers of available, but limited, supply of biomass feedstocks. The 2023 Billion-Ton Study [2] provides biomass resources available in the near term, along with future growth potential. Because we assume the use of woody feedstocks for near-term pyrolysis conversion, based on the 2023 Billion-Ton Study we estimate a total near-term U.S. woody feedstock potential of approximately 175 million dry short tons (MT) per year of woody resources from forestry/wood (144 MT), forest processing waste (1 MT), logging residues (19 MT), other forest waste (8 MT), and small-diameter trees (3 MT). Thus, successful pyrolytic conversion and hydroprocessing technology development can be impactful toward meeting the U.S. 2050 liquid fuel demand of 70 billion gallons per year for not easily electrifiable sectors, including 35 billion gallons per year of aviation fuel [3]. Hypothetically, for assessing an optimistic potential impact of this technology, if pyrolysis-based processes to liquid hydrocarbon fuels have an efficiency of 60 gallons per dry ton of woody biomass feedstock [4], one can estimate nearly 10 billion gallons of liquid fuel per year from the available 175 MT of woody biomass in the United States. There are existing uses of woody resources from forestry, but additional woody resource availability is projected in the future, such as from woody energy crops (34–103 MT) and small-diameter trees (projected to grow from 3 MT to 35 MT) per Table ES-2 in the 2023 Billion-Ton Study.

In this report, we explore both stand-alone and co-hydroprocessing of stable pyrolysis liquid intermediates to inform potential transitions in this technology as more liquid feedstock options become available for future transportation fuels. It is noted that raw fast pyrolysis liquids, without any stabilization, are unstable and difficult to process in conventional hydroprocessing setups [5]. The trajectory of potential implementation and success of this technology for stable pyrolysis liquids depend on many factors, technical and beyond. There are advantages and disadvantages of both stand-alone and coprocessing strategies for new refinery feedstocks like CFP oils. Stand-alone processing offers the advantage of focused R&D and optimization for

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specific liquid intermediates, as well as clean accounting for feedstock-specific economic credits at locations where they are offered. On the other hand, in the near term, coprocessing with the introduction of small proportions of CFP oil in an existing petroleum refining unit can help understand impacts and modification needs without significant new capital expenditure or guaranteed supplies of significant volumes of CFP oil. Stand-alone processing may be possible without large capital expenditure if existing petroleum units are repurposed and dedicated to CFP oil processing, but this may not be possible until the supply of CFP oil is guaranteed. Similarly, building new and dedicated CFP oil hydroprocessing units will be contingent on CFP oil availability, and this approach also needs economic justification for the capital expenditure. Besides helping reduce capital expenditure, coprocessing can enable interoperability of CFP oil with petroleum sources and offer greater flexibility to maintain long-term stability with changes in supply dynamics. Figure 1 provides an overview of the scope of this study.

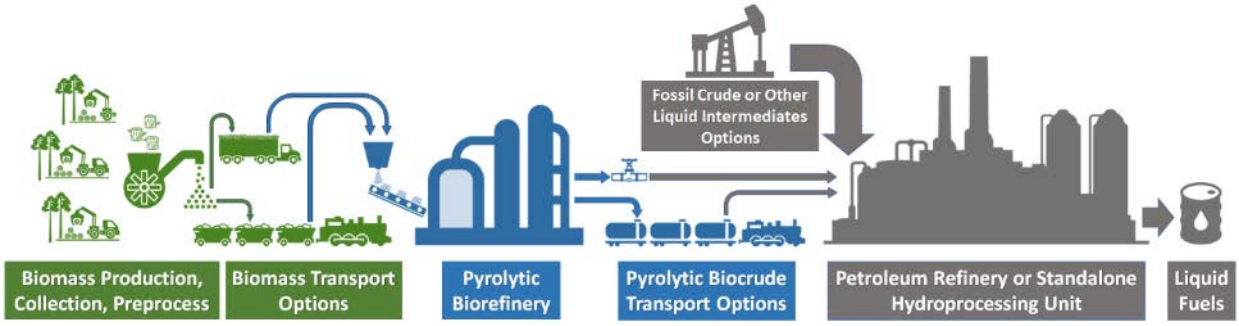

Figure 1. Scope of the field-to-fuel analysis in this study

1.1.1 Pyrolysis Liquid Intermediates

Within the family of pyrolysis liquids, we focus on CFP oils as the representative stable pyrolysis liquid intermediate because some hydroprocessing data are available from small-scale experiments, allowing us to extrapolate from that information to further explore uncertainties through sensitivity studies. Past and current scale-up of CFP oil production includes efforts by Kior [6], Anellotech [7], and Valmet [8]. Other non-catalytic pyrolysis liquids not considered in this report include stabilized pyrolysis liquids [9] or improved pyrolysis liquids after fractionation [10]. Hydropyrolysis at higher pressures [11] is not included in this report.

Consistent with a potential future scenario with scaled-up, commoditized, biomass-derived liquid intermediates production, we present costs and sustainability assessments for the entire field-to-fuel supply chain. We assume separate facilities for (1) the production of liquid intermediates and (2) subsequent hydroprocessing to final fuels, although those facilities may be colocated. One of the advantages of colocation is the flexibility in sourcing and generation of utilities, which can lead to cost and sustainability benefits [12]. We present two previously explored *ex situ* CFP design bases for CFP oil production: (1) a scenario with ZSM-5 catalyst [13], and (2) a potential future higher-carbon-efficiency scenario for biomass conversion via CFP and hydroprocessing to liquid fuels, along with the recovery of valuable coproducts, represented in this report by a process developed at the National Renewable Energy Laboratory (NREL) using a Pt/TiO_2 CFP catalyst [4,12]. Note that this Pt/TiO_2 CFP conceptual process, based on bench-scale experimental results, included the introduction of hydrogen during CFP, thus increasing the carbon efficiency through light hydrotreatment of CFP vapors; this CFP scenario is used to

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project potential improvements in CFP in future years. Options for tighter integration of utility sources and sinks and resulting potential benefits for this pathway were discussed in a separate article [12].

1.1.2 Hydroprocessing Toward Fuels Production

While hydroprocessing is a major processing option for the production of hydrocarbon fuels from oxygenated biomass and its derivatives, it is not the only potentially compatible refinery operation for processing oxygenated liquid intermediates. Fluid catalytic cracking (FCC) processing technology has been explored for non-catalytic pyrolysis liquids [14] and CFP oils [15]. Hydroprocessing is explored here because it usually provides higher carbon efficiencies, because the use of reactive hydrogen in a catalytic environment helps avoid significant carbon loss as CO, CO_2, and coke during deoxygenation. Gasoline is typically the dominant product from FCC processing, while less cracking during hydroprocessing of the same liquid feed results in heavier fuel products such as jet and diesel. Hydroprocessing also increases the hydrogen-to-carbon molar ratio (H:C) in the product, which is a necessity for meeting fuel property specifications of incumbent fuels. Based on up to 5 vol % CFP oil coprocessing with VGO in a Davison Circulating Riser representing an FCC setup [16], roughly 65% or lower proportion of the carbon in the CFP oil is retained in various liquid cuts. The remaining ~35% carbon is distributed between the gas phase and produced coke; the ratio of carbon in gas phase to coke was approximately 3.5 for the overall product from CFP oil + VGO product. The gas yields included C3 and C4 species. The gas and coke yields may be lower in commercial FCCs with petroleum feedstocks, ~26%, including coke, fuel gas, and liquefied petroleum gas, as reported by Gary et al. [17]. It is noted that other refinery operations such as FCCs and cokers [17] can be attractive options for specific liquid feedstocks and product objectives and are separate areas of research.

1.2 Techno-Economic and Sustainability Analysis Approaches

The techno-economic analysis (TEA) approach for this work is consistent with that detailed in prior NREL biofuels technical reports [18,19]; those reports evaluated the economic potential of efficient stand-alone biorefineries of the future that included CFP oil production followed by hydroprocessing within the same conversion facility that may be remote and located close to biomass feedstock production sites. In the work presented here, the team leveraged previous work for reporting biomass feedstock production and preprocessing followed by conversion to a CFP oil intermediate in stand-alone, greenfield biorefineries. Unlike our previous studies for refinery integration projections [20] that used prorated assumptions from stand-alone hydrotreating evaluations, the work presented in this report uses a petroleum-refinery-centric view to evaluate units that import CFP oil for stand-alone or co-hydroprocessing with base petroleum feedstocks. Most of these options may involve the transportation of significant volumes of CFP oil to centralized hydroprocessing locations to be better aligned with potential future volumetric growth for this conversion technology and its biomass feedstocks supply backbone. We include an assumption of 800 miles of rail transportation of CFP oil to represent potential future growth where economies of larger-scale operations are leveraged by collecting CFP oil within a large radius.

Life cycle analysis was conducted to estimate the greenhouse gas (GHG) emissions associated with hydrocarbon fuels produced from woody biomass via CFP and hydroprocessing, using

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Argonne National Laboratory's R&D Greenhouse gases, Regulated Emissions, and Energy use in Technologies (R&D GREET) model [21]. This analysis focuses on the 100-year global warming potential, measuring estimated emissions in carbon dioxide equivalent (CO_2e) through the biomass feedstock-to-hydrocarbon fuel supply chain. Emissions of CO_2, CH_4, and N_2O were estimated with 100-year global warming potential factors of 1, 29.8, and 273, respectively [22]. Sustainability impacts in this study include the accounting of biogenic carbon incorporated in the liquid fuels, as well as the generation of gases for heating utilities (e.g., fuel gas and steam). The results presented later in this report provide life cycle inventories including accounting for biogenic gases and impact of their use for utilities within an industrial infrastructure. For example, one cannot export steam or fuel gas at a remote location where there are no other major industrial consumers, which may be the case for CFP oil production facilities. In that case, electricity export may be the only option for utilizing excess energy sources like fuel gas. However, biogenic off-gases produced during hydroprocessing at a petroleum refinery may be used as fuel gas and for steam production, thus providing economic and sustainability benefits. In this analysis, coproduct contributions are accounted via displacement credits, because they are small (significantly less than 50%) in comparison to the main fuel products, as detailed in the life cycle inventory tables provided later in this report.

1.2.1 Financial Assumptions

The modeled projections in this report are based on the mature or n^{th}-plant technology. 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 mature industrial implementation. A consistent set of assumptions is used for all projections. Key assumptions are listed in Table 1. In the current analysis, the reference cost year for time-value-of-money and all cost indices is 2020 U.S. dollars.

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

Description of Assumption	Assumed Value
Cost year	2020
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 IRS Modified Accelerated Cost Recovery System schedule [23]
Steam plant depreciation	20-year IRS Modified Accelerated Cost Recovery System schedule [23]
Construction period (spending schedule)	3 years (8% Y1, 60% Y2, 32% Y3)
Plant salvage value	No value

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Description of Assumption	Assumed Value
Startup time	6 months
Revenue and costs during startup	Revenue = 50% of normal Variable costs = 75% of normal Fixed costs = 100% of normal
On-stream factor after startup	90% (7,884 operating hours per year) for CFP 95% (8,322 operating hours per year) for hydroprocessing

1.2.2 Estimation of Capital and Operating Costs

Information for the CFP oil production processes was obtained from our previous studies [18,19]. Details for the two cases included in this report are presented in the results sections, with further details included in Appendix E and Appendix F.

Capital costs of hydroprocessing equipment were provided by KBR; details are included in Appendix B. Specific evaluation is necessary for each location because bottlenecks, existing spare capacity, and other factors, along with locational costs, will be major considerations for choosing actual designs, changes to existing facilities, and the optimization of costs. For this report, we utilize a consistent methodology for hydroprocessing capital costs, as detailed in Appendix B. The cost of repurposing and retrofitting refinery hydroprocessing units for handling CFP oils is derived simplistically from the cost of new equipment, after subtracting the sunk cost of existing capital for the same equipment at a refinery unit. Potential specific differences, including more expenses for upgrading very old and often fully depreciated refinery equipment compared to new equipment, are not accounted in the analysis presented.

As detailed in Appendix B, KBR provided NREL with bare equipment costs (without installation) for hydroprocessing equipment (Table B-1) based on some specific sample process models from NREL that provided flow rates, duties, power consumption, etc. for various equipment. Because NREL-modeled cases explored in this report span a wide range of conditions, it was necessary to develop a methodology to quickly estimate costs using information from process simulations for different cases. NREL used the cost data points provided by KBR to estimate scaling exponents to allow cost calculations for the different processing capacities (Table B-2). Pressure corrections were applied to these estimated bare equipment costs based on guidance from KBR, as detailed in Table B-3 and its footnotes. KBR further provided Lang factor (f_{Lang}, below) relations to estimate the total FCI from the bare equipment costs for various families of process equipment. These FCI estimates were used by NREL for capital investment estimates. Details for hydroprocessing operating costs are provided in Section 3 and related appendices.

Economic calculations were performed in Excel. New equipment costs are scaled based on process flows in the Aspen Plus models using a scaling exponent:

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

The scaling exponents for each piece of equipment, n, are documented in Appendix B, along with the base value of the variable used for scaling.

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For the CFP models, we followed our previous approach [18] of calculating a total installed cost by applying an installation factor ($f_{install}$):

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

Further factors were applied to arrive at an FCI for the CFP plant, as quantified in Section 2.

For the hydroprocessing models, discussed in Section 3, the total FCI was derived from the purchased equipment cost by applying Lang factors (f_{Lang}):

$$Fixed\ Capital\ Investment\ (FCI) = f_{Lang} * Total\ Purchased\ Equipment\ Cost\ (TPEC)$$

Lang factor estimates include:

- Total direct cost items: Piping, civil, steel, instrumentation, electrical, insulation, and paint.
- Indirect field costs: Construction management, startup and commissioning, scaffolding, equipment rental, vendor representatives, field services and controls, business expenses, construction utilities, and firewatch.
- Other costs: Freight, taxes and permits, engineering, 2 years of escalation, and contingency.

Bare equipment costs and basis used for scaling various equipment are included in Table B-2. Lang factor estimation correlations and pressure correction information are listed in Table B-3.

Costs were converted to 2020 dollars using:

$$Cost\ in\ 2020\$ = Base\ Cost\ \left(\frac{2020\ Cost\ Index\ Value}{Base\ Year\ Cost\ Index\ Value}\right)$$

Equipment costs, operating costs, and labor costs were converted to 2020$ using the Chemical Engineering Plant Cost Index [24], Producer Price Index for Chemical Manufacturing [25], and U.S. labor cost for chemical manufacturing [26], respectively.

1.2.3 Minimum Selling Price for CFP Oil and Fuel Products

The total capital investment, along with plant operating costs, are integrated with a discounted cash flow analysis. Those costs, along with the gallons gasoline equivalent (GGE) of the total fuel intermediate or the fuel blendstock products, were used to derive the minimum fuel selling price (MFSP) in dollars per GGE. GGE is a unit of energy and easily converted to other energy bases like gallon jet equivalent (GJE) and gallon diesel equivalent (GDE). These energy units use lower heating values (LHVs) and are useful for providing a market-relatable cost (per gallon) when discussing volumes and prices of liquid fuel products. In this report we use 116,090 Btu/gal for GGE, 124,307 Btu/gal for GJE, and 128,450 Btu/gal for GDE, based on GREET [27].

1.2.4 Process Models

Aspen Plus models for stand-alone CFP biorefineries and the hydroprocessing units provide the process basis for assessing capital costs for all scenarios. Section 2 (CFP) and Section 3

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(hydroprocessing) describe the models and their assumptions. The Peng-Robinson with Boston-Mathias modifications equation of state was used throughout most of the process simulations. The ASME 1967 steam table correlations (STEAM-TA) were used for the steam cycle calculations. Compound slates used in the models are listed in Appendix C (for the CFP models) and Appendix D (for the hydroprocessing models).

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Section 2

Biomass to CFP Oil: Techno-Economics and Sustainability for CFP Oil Production

Modeling and Results

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2 Biomass Feedstock to CFP Oil

This section presents our assumptions for the conversion of woody biomass feedstocks to CFP oil. This research area was funded by the Bioenergy Technologies Office since 2014; previous experimental work, techno-economic and sustainability analyses are leveraged in this report. For completeness of this report, key information for woody feedstocks and the conversion process is provided or repeated here; references are provided for the interested reader to find more details.

2.1 Woody Feedstock and CFP Plant Size

We maintained woody feedstock assumptions developed from previous CFP work, as documented by Hartley et al. 2021 [28]. Modeled costs and sustainability metrics were developed for woody feedstocks with a 50:50 mixture of clean pine and forest residues for a 2,000 dry metric tonnes CFP plant located near the woody feedstock source. Feed composition and cost assumptions are shown in Table 2. All costs for collection, transport, and preprocessing including drying and size reduction are included in the modeled estimates shown below.

Table 2. Delivered Woody Feedstock Composition and Costs for CFP [28]

Component	Composition (dry wt %)
Carbon	50.51
Hydrogen	5.99
Nitrogen	0.17
Sulfur	0.03
Oxygen	41.55
Ash	≤1.75
Other Metrics	
Moisture (bulk wt %)	10
Particle size (inches)	≤0.08
Cost	**$/dry U.S. ton**
Clean pine (2016$)	$71.62
Logging residue (2016$)	$62.44
Total for 50:50 mix (2016$)	$67.03
Total for 50:50 mix (2020$)[a]	$72.53

[a] Derived using cost index [25].

2.2 CFP

CFP is a process that converts solid biomass into partially deoxygenated oils suitable for downstream upgrading. In CFP, biomass is thermally broken down in a non-oxidizing environment. Those vapors are further decomposed and deoxygenated by acidic, metallic, or basic catalyst active sites through a variety of mechanisms [29,30]. While CFP has been studied in a variety of reactor configurations and geometries, the most distinctive feature of a CFP process is whether the catalyst is included in the pyrolysis reactor with the biomass, called *in situ* CFP, or if the catalyst is contained in a separate reactor, called *ex situ* CFP [18,31]. *In situ* CFP

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offers the advantage of operational simplicity but can be challenging to operate due to rapid catalyst deactivation by solid contaminants (char and mineral matter) originating from the biomass feedstock. Instead, many processes study *ex situ* CFP which offers the opportunity to remove the solid contaminants after the fast pyrolysis reactor before primarily only the vapor/gas products stream enter a separate *ex situ* vapor upgrading reactor; solids removal options include cyclones for primary removal followed by guard beds or hot gas filters to further remove the solids that remain in the vapor stream. The *ex situ* configuration has added benefits of allowing independent control of pyrolysis and catalytic vapor upgrading reaction conditions, including temperature. Other options such as hydropyrolysis at higher pressures with the inclusion of hydrogen are being explored toward commercialization through the IH^2 process [11], but not discussed in this report; that process produces a significantly more deoxygenated oil and related conversion and hydroprocessing metrics will be very different compared to the 16.0% and 21.6% CFP oil oxygen contents presented in this report.

2.2.1 Ex Situ CFP in Fluid and Fixed-Bed Configurations

This work presents data collected in NREL's 2-inch fluidized bed reactor under two unique *ex situ* reactor configurations. These form the basis for two conceptual CFP plants:

> **CFP-A:** A current case using ZSM-5 CFP catalyst with 21.6 wt % oxygen content in the CFP oil, with integrated electricity production to use excess gases and heat from the CFP process [13].
>
> **CFP-B:** A potential future case using a Pt/TiO_2 catalyst in the presence of hydrogen, with improved CFP yields and 16.0 wt % oxygen content in the CFP oil and multiple utility integration options (including hydrogen) for excess gases and heat from CFP [12].

We extend bench-scale CFP performance for the two configurations to a scaled-up conceptual 2,000 dry metric tonnes per day woody biomass plant and quantify associated cost and sustainability metrics for a field to CFP oil assessment in this section (Section 2); metrics for the entire field to fuel supply chain, including hydroprocessing, is presented in the next section (Section 3).

In both CFP cases, pine feedstocks were pyrolyzed in a 2-inch reactor with bubbling bed of sand, and char was removed from the produced vapors in a cyclone. The vapors were then sent to two different catalytic upgrading reactor configurations described in Figure 2. Entrained-flow reactors are shown in Figure 2, reflecting potential commercial implementations; the bench-scale lab equipment included simpler bubbling bed systems instead of entrained systems. In the first configuration shown in Figure 2a, an acidic zeolite was used for pyrolysis vapor upgrading in a continually refreshed bubbling bed under an inert atmosphere. In this operating mode, deactivated catalyst was continuously withdrawn, and fresh catalyst was continuously added to the bubbling bed, allowing for longer operational times. Formed catalyst spheres from Johnson Matthey composed of 80 wt % ZSM-5 with a SiO_2/Al_2O_3 ratio of 30, and 20 wt % Al_2O_3 binder, with a mean diameter of 0.5 mm were used in the bubbling bed. Use of zeolite catalysts in a fluid catalyst bed is similar to Anellotech's Bio-TCat *in situ* CFP [7], and KiOR's BCC *in situ* CFP technologies [6], and similar catalysts have been used throughout the development of the biomass CFP technology [32]. The other CFP configuration, shown in Figure 2b, uses a fixed bed of Pt/TiO_2 in a H_2 rich atmosphere to achieve pyrolysis vapor deoxygenation. Pyrolysis

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vapors were first passed through a 2 µm stainless steel hot gas filter before entering a fixed bed of 0.5 wt % Pt/TiO_2. Note that large-scale long-term operations may require ceramic filter elements instead of stainless steel to avoid iron carryover via corrosion [33]. Guard beds may also be used instead of hot gas filters to prevent particle carryover into the fixed bed reactor. In both cases, products were collected in a condensation system consisting of combinations of either an ambient air or a propylene glycol cooled condenser at −2°C, an electrostatic precipitator, three dry ice cooled condensers, and an ice cooled coalescing filter. The CFP process produces char, catalyst coke, organic and aqueous liquids, condensable gases and permanent gases which can be quantified to give carbon balances exceeding 95% in most cases.

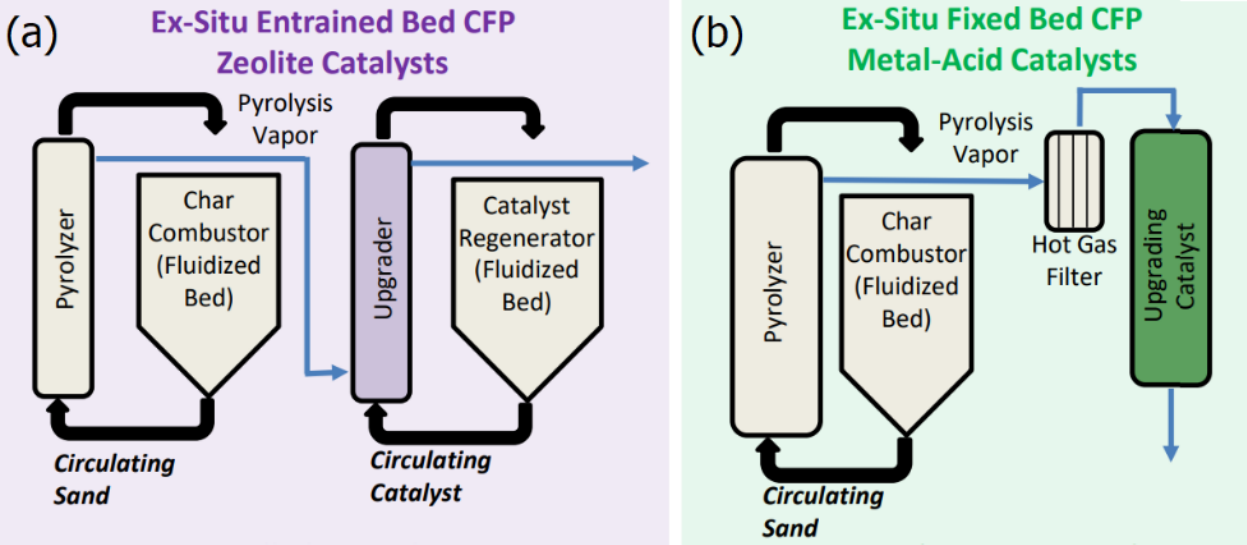

Figure 2. Configuration options for *ex situ* CFP. In bench-scale experiments, bubbling fluidized bed reactors were used without the recirculation loops for sand and catalyst [34].

2.2.2 Experimental Data and Model Assumptions for Two CFP Cases

Significant CFP data has been collected at NREL in the 2014–2024 time frame, and we chose two specific experimental results for our process modeling to represent two sufficiently different CFP oils where the effect of oxygen contents and yields on techno-economics and sustainability are apparent.

Table 3 collects operating conditions, biomass information, and oil properties produced under the two CFP configurations shown in Figure 2. A sample result from literature for fast pyrolysis alone, where the raw pyrolysis vapors are fed directly into the condensation system, are also included for comparison. Of particular interest is the oxygen content of the oil, which can be used to estimate the degree of deoxygenation required in downstream upgrading to produce hydrocarbon products. Without catalytic upgrading, fast pyrolysis oils have significantly higher oxygen contents and are chemically less stable, leading to more difficult downstream upgrading. Furthermore, the use of H_2 over Pt/TiO_2 reduces coke formation and favors pyrolysis vapor deoxygenation relative to ZSM-5 catalysts in an inert atmosphere; thus, the biomass-to-catalyst-mass ratio is higher with Pt/TiO_2. However, these yields can be influenced by changing the biomass-to-catalyst-mass ratio for both catalytic processes. Generally, a lower ratio, using more catalyst per unit biomass, results in a more deoxygenated oil with a lower carbon yield to oil [35].

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The presence of mineral matter in the biomass feedstock, indicated by the ash content, can lower the yield of liquid products during fast pyrolysis and consequently *ex situ* CFP. Historically, most of the NREL experiments were conducted with woody biomass containing <1 wt % ash, which is less that the ≤1.75% limit indicated in Table 2.

Table 3. Comparison of Operating Conditions and Resulting Oil Properties for Fast Pyrolysis and Two Configurations of *Ex Situ* CFP

	Fast Pyrolysis [a]	ZSM-5 in Inert CFP [b]	Pt/TiO_2 in H_2 CFP [c]
Biomass-to-catalyst-mass ratio	N/A	2.8	12
Operating gas	N_2	N_2	85 vol. % H_2/15 vol. % N_2
Pyrolyzer temperature, °C	500	500	500
Catalyst temperature, °C	N/A	500	400
Catalytic reactor type	N/A	Bubbling bed	Fixed bed
Biomass feedstock	Oak	50:50 wt % clean pine and forest residues	50:50 wt % clean pine and forest residues
Biomass elemental composition (dry basis)			
Ash, wt %	0.39	1.0	0.8
C, wt %	49.6	50.9	50.7
H, wt %	6.03	6.3	6.7
O, wt %	43.9	41.7	43.9
N, wt %	0.08	0.2	0.1
S, wt %	0.01	Not measured	0.03
Oil elemental composition			
C, wt % dry	54.9	71.5	77
H, wt % dry	6.28	6.9	7.0
O, wt % dry	38.7	21.6	15.0
N, wt % wet	0.07	0.1	0.2
S, ppm wet	<50	27	≤20
Oil water content, wt %	20.3	4.4	2.2
Oil H:C molar ratio, dry	1.36	1.15	1.21
Oil density, g/mL	1.24	1.15	1.01
Oil viscosity at 40°C, cP	57	285	29 [e]
Oil carboxylic acid number, mg KOH/g	91[d]	30	34 [e]

Data from: [a] Elliott et al. [36], [b] Griffin et al. [37] with additional previously unpublished information, [c] Yung et al. [38], [d] Black et al. [39]. [e] From similar experiments. The number of decimal places shown are as reported in literature.

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Oil oxygen content is important because it indicates the likelihood of an intermediate oil polymerizing due to reactions between oxygenates [40]. These polymerization reactions can lead to rapid viscosity increases with consequent equipment and transfer line plugging. The increased oxygen content of non-catalytic fast pyrolysis liquids also corresponds to increased acidity, which increases the metallurgical requirements for transporting and storing fast pyrolysis liquids, as well as downstream processing equipment relative to CFP oils [41]; previous work has shown that steels rich in Cr are most capable of resisting corrosion by organic acids [42]. Beyond oxygen content, fast pyrolysis liquids also have higher water contents than CFP oils because they generally do not separate into organic and aqueous phases, in part due to the highly oxygenated nature of their organic molecules (hence we refer to that liquid intermediate as fast pyrolysis liquid instead of fast pyrolysis oil). It has been shown that non-catalytic pyrolysis liquids can contain trace inorganic elements such as Na, K, and Si from the biomass feedstock [36]. These inorganics are often below the limit of detection in CFP oils [40], potentially as a result of deposition onto the catalyst surface [43]. However, inorganics retained in CFP oils will differ for different technologies and producers.

In addition to bulk properties of the produced oil, the distribution of biomass feedstock mass and carbon between different CFP products can have a major impact on the economic performance of the process [4]. Table 4 and Table 5 collect the mass and carbon yield of each product for the three operating modes described above. All operating modes produce similar carbon yields to char as the char-producing fast pyrolysis reaction step is conserved among all three configurations. In addition to not producing an aqueous phase, fast pyrolysis produces fewer permanent gases than CFP as these gases are typically produced by catalytic deoxygenation reactions. In the case of ZSM-5 upgrading, permanent gases are primarily CO and CO_2 produced by decarbonylation and decarboxylation reactions that aid in vapor deoxygenation. In the case of Pt/TiO_2 in the presence of H_2, a larger fraction of the biomass carbon is retained in the oil phase. The ability of Pt to activate hydrogen inhibits the deposition of carbon onto the surface of the Pt/TiO_2 catalyst, leading to a significantly lower coke carbon yield relative to the ZSM-5 catalyzed process. As with the oil properties, the selection of operating conditions influences the distribution of carbon throughout the product fractions.

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Table 4. Product Yield Distribution for the Fast Pyrolysis and CFP Processes.

Yields are on dry biomass feedstock basis.

Product	Fast Pyrolysis [a]	ZSM-5 in Inert CFP [b]	Pt/TiO_2 in H_2 CFP [c]
Char, wt %	9.9	12.7	11.1
Coke on catalyst, wt %	N/A	7.1	1.7
Oil, wt %	63.5	20.2	24.8
Aqueous, wt %	N/A	26.6	22.2
Condensable gases,[d] wt %	N/A	6.0	10.8
Permanent gases,[d] wt %	19.1	24.1	29.7
Balance, wt %	92.4	96.7	102.4

Data from: [a] Elliott et al. [36], [b] Griffin et al. [37] with additional previously unpublished information, [c] Yung et al. [38]. [d] Condensable gases are vapors that could condense at close to room temperature and pressure, but the condensation is limited by the presence of other gases that lower their partial pressure. Permanent gases typically have critical temperatures significantly lower than room temperature and cannot be condensed at ambient conditions even if not diluted by other gases. Portions of the condensable gases can be recovered in the liquid product by optimizing the condensation train. Condensable gases not reported separately for Fast Pyrolysis.

Table 5. Carbon Product Distribution for the CFP Processes Described Above

Product	ZSM-5 in Inert CFP [a]	Pt/TiO_2 in H_2 CFP [b]
Char, %	19.3	17.2
Coke on catalyst, %	12.1	3.3
Oil, %	28.1	35.7
Aqueous, %	6.3	1.7
Condensable gases, %	6.9	12.9
Permanent gases, %	21.8	28.1
Balance, %	94.5	98.9

Data from: [a] Griffin et al. [37] with additional previously unpublished information, [b] Yung et al. [38].

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2.3 Conceptual Models and Assumptions

Woody biomass conversion to CFP oil is discussed here. Significant previous documentation exists for similar conceptual designs and the reader can reference those for further details [18,19]. Area numbers are consistent with documentation in our previous reports.

Conceptual processes for the two CFP options are shown in Figure 3 (CFP-A with fluidized ZSM-5 catalyst) and Figure 4 (CFP-B with Pt/TiO_2 catalyst in a fixed bed).

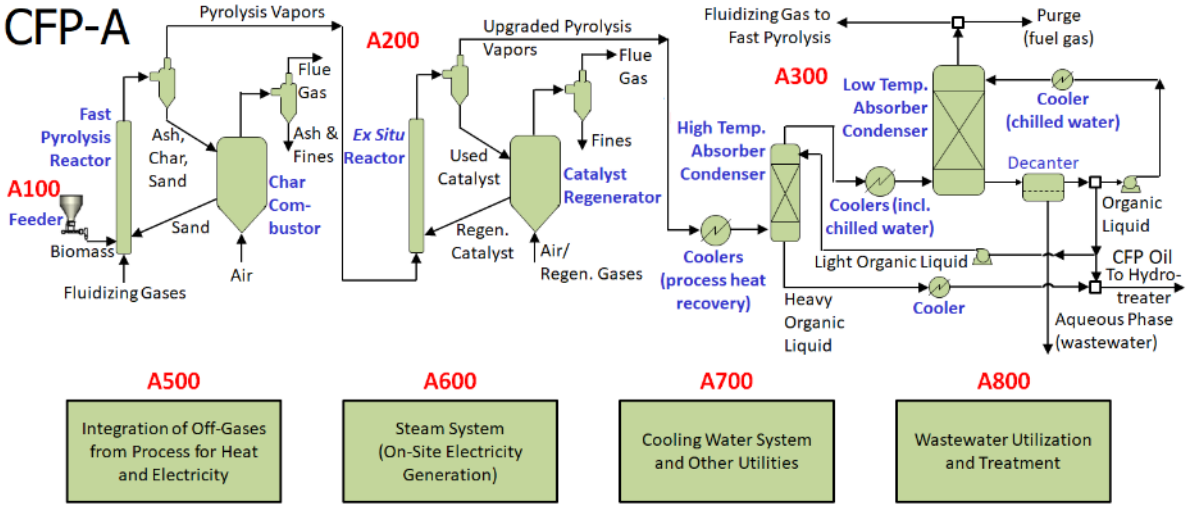

Figure 3. CFP-A conceptual design for *ex situ* CFP system with fluidized ZSM-5 catalyst

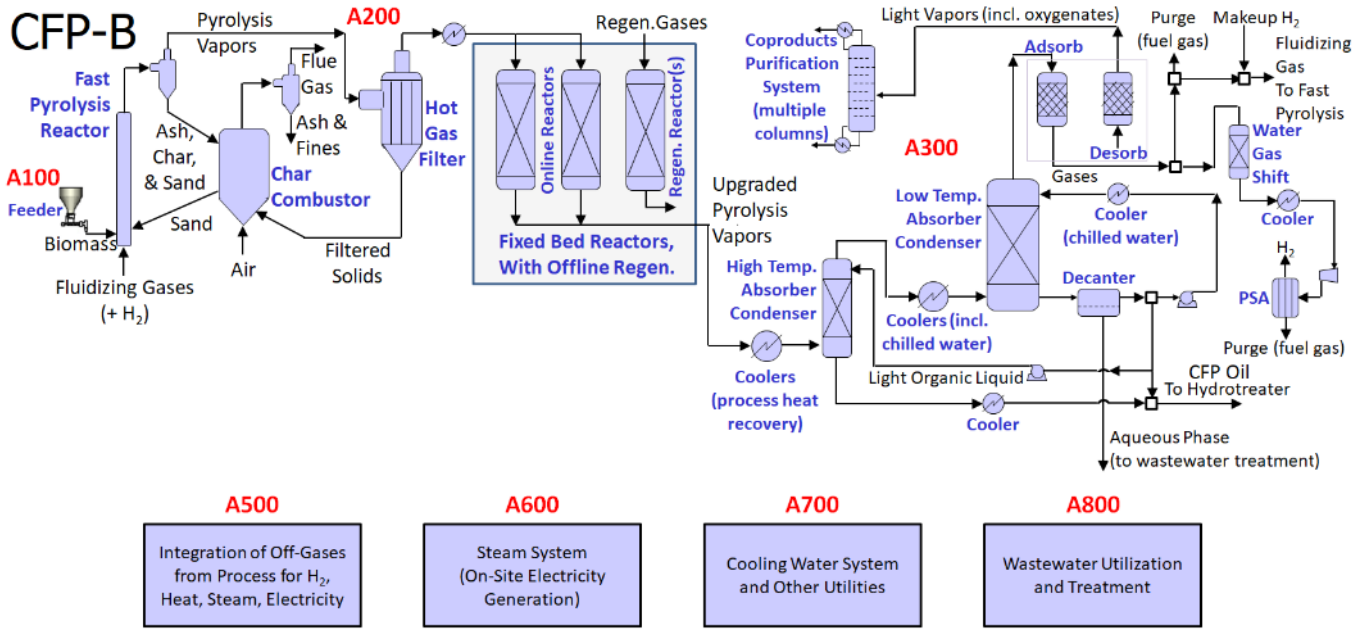

Figure 4. CFP-B conceptual design for *ex situ* CFP system with Pt/TiO_2 catalyst in fixed bed

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Area (denoted by prefix A) numbers for different operations are included below:

- **A100:** Biomass feedstock handling
- **A200:** Fast pyrolysis and *ex situ* catalytic vapor upgrading
- **A300:** CFP products condensation
- **A400:** Hydroprocessing CFP oil is discussed in Section 3

Supporting operations include:

- **A500:** Utilization of off-gases
- **A600:** Steam system and electricity generation
- **A700:** Cooling water and other utilities
- **A800:** Wastewater utilization and treatment

The following sections summarize the basis and assumptions for each area. Some of the descriptions are repeated from previous publications [4,18,19] to provide an overview to the reader without referring to other reports.

2.3.1 Area 100: Biomass Feedstock Handling

All woody biomass feedstock-related costs were modeled by Idaho National Laboratory (INL) and emissions by Argonne National Laboratory [28]. The specifications and cost for a 50% clean pine + 50% logging residues blend supplied to a conceptual 2,000 dry metric tonnes per day CFP plant are shown in Table 2. A woody feedstock with 10 wt % moisture content and a nominal particle size of 2 mm were specified. Ash content was specified as <1.75 wt % for modeling; actual ash contents were closer to 1 wt %, as shown in Table 3. Higher ash can be detrimental for liquid yields for both fast pyrolysis (FP) and CFP [44,45]. Woody feedstock preprocessing for careful control of both the proportion of ash and its constituent mineral species are necessary for FP and CFP. The direct physical impacts of ash on CFP catalysts and their deactivation may be mitigated in an *ex situ* configuration by ash removal via cyclones (and filters) after the FP step (Figure 3 and Figure 4).

Table 6. INL-Modeled Woody Feedstock Cost for 50% Forest Residues and 50% Clean Pine

	$/dry U.S. ton, 2016$
Grower payment	$9.74
Harvest and collection	$4.94
Field-side preprocessing	$8.41
Transportation	$12.22
Preprocessing	$25.43
Storage	$0.68
Handling	$2.65
Preprocessing construction	$2.96
Total cost (2016$)	**$67.03**
Total cost (2020$)[a]	**$72.53**

[a] Derived using cost index [25].

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Modeled cost breakdown [28] for the woody feedstock is reproduced in Table 6; this is included as an operating cost in the conversion TEA. The only costs added for Area 100 in the conversion process model are for a flue gas blower and a cross-flow dryer for warming the biomass feedstock using process flue gases.

2.3.2 Area 200: Fast Pyrolysis and Ex Situ Catalytic Vapor Upgrading

2.3.2.1 Area 200 Overview

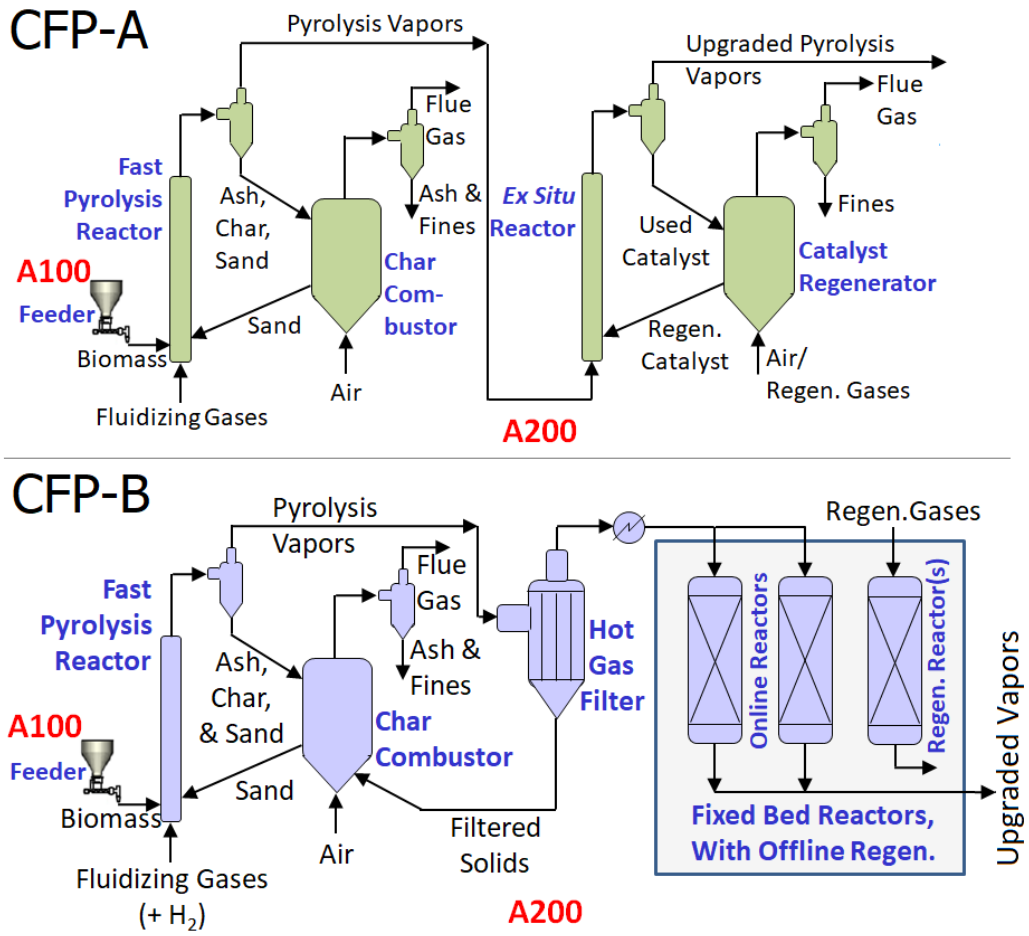

Figure 5. Fast pyrolysis followed by fluid (CFP-A) and fixed (CFP-B) bed *ex situ* catalytic vapor upgrading reactor systems

The modeled fast pyrolysis circulating dual fluidized bed system with sand as a heat carrier includes an entrained-flow pyrolyzer and a fluidized bed char combustor, modeled to operate at approximately 500°C and 720°C, respectively [18]. A simplified sample model is available for interested readers [46]. A portion of the non-condensable off-gases available after the condensation system (Area 300) is used for fluidization of biomass and sand in the entrained-flow pyrolyzer. Entrained solids are removed from the pyrolysis vapors via two cyclones before the vapors are sent to the *ex situ* catalytic upgrading reactors. The bed material (sand) and produced char, removed by the cyclones, are sent to a fluidized bed char combustor to generate heat for the endothermic pyrolysis reaction; the sand temperature increases upon char

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combustion, and the hot sand is circulated back to the pyrolysis reactor. Note that char sequestration and other potential uses of biochar are not explored with this design choice.

For the circulating dual fluidized bed CFP system with ZSM-5 catalyst (CFP-A), a lower pressure of approximately 43 psia (3 bar) is specified for the FP reactor in the model; the 43 psia (3 bar) pressure is conservative, and could be lower, if downstream equipment pressure drops alone dictate the choice of pressure. Note the configuration of the fluidized catalytic reactor system downstream of the FP reactor is nearly identical to the FP reactor system, with circulating catalyst, instead of sand used in the pyrolysis reactor. Vapor upgrading reactions may be exothermic, depending on the chemistry made possible by the catalyst. Heat is also generated via the combustion of coke deposited on the catalyst. Heat removal via catalyst coolers is necessary in our model, since excess heat is produced in this system.

The future case fixed-bed system with Pt/TiO_2 catalyst (CFP-B) is modeled to operate at a higher pressure of 121 psia (8.3 bar) at the FP reactor, allowing potential capital cost savings from smaller volumetric flow rates and equipment sizes. A hot gas filter is included prior to the fixed-bed reactors in CFP-B to capture particulates and alkali and alkaline earth metals from the FP reactor that are not removed by the cyclones, thus preventing plugging of the downstream catalytic upgrading fixed-bed reactor and/or deactivating the CFP catalyst; alternate equipment, such as swing guard beds, may be considered instead of a hot gas filter. A hot gas filter or guard bed may also be included in the CFP-A configuration (fluidized system) to reduce the carryover of solid contaminants from the fast pyrolysis reactor and help produce a better-quality CFP oil with fewer particulates; however, it is not required in CFP-A because a portion of the catalyst is purged on a regular basis, thus also allowing an outlet for some of the particulates. For CFP-B, the fluidizing gases are significantly enriched using hydrogen, since the Pt/TiO_2 catalyst is designed to provide higher yields by activating hydrogen, including for partial deoxygenation and hydrogenation, and coke prevention. This has been demonstrated in bench-scale experiments at near-atmospheric pressures [38], and the assumption of higher pressure in the model is conservative with respect to yields of CFP oil, since higher pressures can further activate hydrogen and increase yields. Performance at higher pressure was not demonstrated experimentally due to constraints of existing equipment and related safety in our lab-scale setup. A 0.5% Pt loading on TiO_2 [38] was used in the experiments and that was also the basis for this model. In the current model the reactor requires heat removal because of exothermic upgrading reactions. Heat management and related design and operations are also important during catalyst regeneration; previous work showed thermal excursions during regeneration leading to fragmentation of the catalyst at the top of the bed [47].

The CFP-A configuration is similar to the system described in detail in our 2015 design report [18], except for higher pressure and hydrogen enrichment included in the 2015 design report (not included in CFP-A). Equipment costs for the current system were updated by scaling based on the volumetric flow rates of the gas/vapor phase (included in Appendix E).

For CFP-B the design basis was maintained from Dutta et al. 2016 [19], with design parameters for reactor #1 specified in Table 2 of the reference [19] as the basis for reactor cost assumptions. Cost details are provided in Appendix F.

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2.3.2.2 Area 200 Design Basis

Table 7 and Table 8 list the model design parameters for the pyrolysis and vapor upgrading operations in the CFP-A and CFP-B configurations. Since both the FP and CFP-A vapor upgrading reactors are circulating dual fluidized beds with similar parameters, they are both shown in Table 7. Parameters for the fixed-bed catalytic upgrading reactor in CFP-B are shown in Table 8.

Table 7. Model Parameters for Circulating Dual Fluidized Bed Reactors for Fast Pyrolysis (CFP-A and CFP-B) and Catalytic Vapor Upgrading in CFP-A Configuration

	FP Reactor for CFP-A and CFP-B	CFP-A Vapor Upgrading Reactor
Reactor		
Parallel trains for reactor and combustor	2	1
Total biomass feed rate per train, dry metric tonnes/day	1,000	N/A
Biomass moisture (wt %)	10	N/A
Exit temperature, °C (°F)	500 (932)	500 (932)
Pressure, psia (bar)	43 (2.96) for CFP-A 121 (8.34) for CFP-B	37 (2.55)
Bed solids material	Sand	Catalyst (ZSM-5)
Bed solids inventory (ton)	99 for CFP-A 120 for CFP-B	77
Catalyst replacement (% of reactor system inventory/day)	N/A	2
Additional catalyst attrition (% of reactor system inventory/day)	N/A	1.6
Ratio of solids circulation to biomass feed with 10% moisture (w/w)	5.8 for CFP-A 7.1 for CFP-B	4.5
Number of cyclones per reactor	2	2
Stripping steam (lb/1,000 lb catalyst)	-	3
Fluidizing gases to dry biomass (w/w)	1.6 for CFP-A 0.7 for CFP-B	-
H_2 partial pressure in fluidizing gas feed, psia (bar)	N/A for CFP-A 94 (6.5) for CFP-B	N/A
Combustor/Regenerator		
Temperature, °C (°F)	720 (1328)	650 (1202)
Pressure, psia (bar)	26 (1.8) for CFP-A 117 (8.1) for CFP-B	26 (1.8)
Excess air (%)	20	20
Solids temperature before transfer to reactor, °C (°F)	720 (1328)	562 (1044)
No. of cyclones per combustor	2	2

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For both CFP-A and CFP-B the yields are informed by bench-scale experimental information presented in Section 2.2.2. CFP-B model metrics are slightly different from the metrics shown in Section 2.2.2, because the model is based on an average of three samples; details are available in the "FY 2020 SOT" column of Table ES-1 of a previous report [4]. Since NREL experiments were conducted in semi-batch systems without online catalyst regeneration, assumptions needed to be adapted for the continuous system in the process model. Catalyst circulation in CFP-A was dictated by reactor/regenerator temperature control in the conceptual model. Although the catalyst flow is higher in the model compared to the bench-scale experimental data, the actual activity of equilibrium catalyst in a scaled-up system may be lower and is an important parameter that will ultimately affect performance.

Similarly for the fixed-bed system in CFP-B, a WHSV of 5 h^{-1} was assumed based on total vapors + fluidizing gases flowing through the reactor (including extra catalyst for contingency [19]), catalyst loading was 110,214 lb in the two online reactors and 110,214 lb in the two offline reactors undergoing regeneration. Thus, the modeled catalyst loading was 110,214 lb for processing 183,718 lb/h biomass (for a 2,000 dry metric tonnes/day biorefinery). The experimental biomass to catalyst ratio was 12 (Table 3); thus, the estimated reactor cycle time is approximately 7.2 hours. Steam generation was included for heat removal from the fixed-bed reactor system.

Table 8. Model Parameters for Fixed-Bed Catalytic Vapor Upgrading in CFP-B Configuration

Parameter for Fixed-Bed Reactor	CFP-B Vapor Upgrading Reactor
Reactors on stream (parallel)	2
Spares for regeneration	2
WHSV (h^{-1})	5
H_2 concentration at reactor inlet	61 mol %
Outlet pressure, psia (bar)	107 (7.4)
Inlet and outlet temperature, °C (°F)	375 (707) and 425 (797)

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2.3.3 Area 300: Pyrolysis Vapor Quench and Products Recovery

2.3.3.1 Area 300 Overview

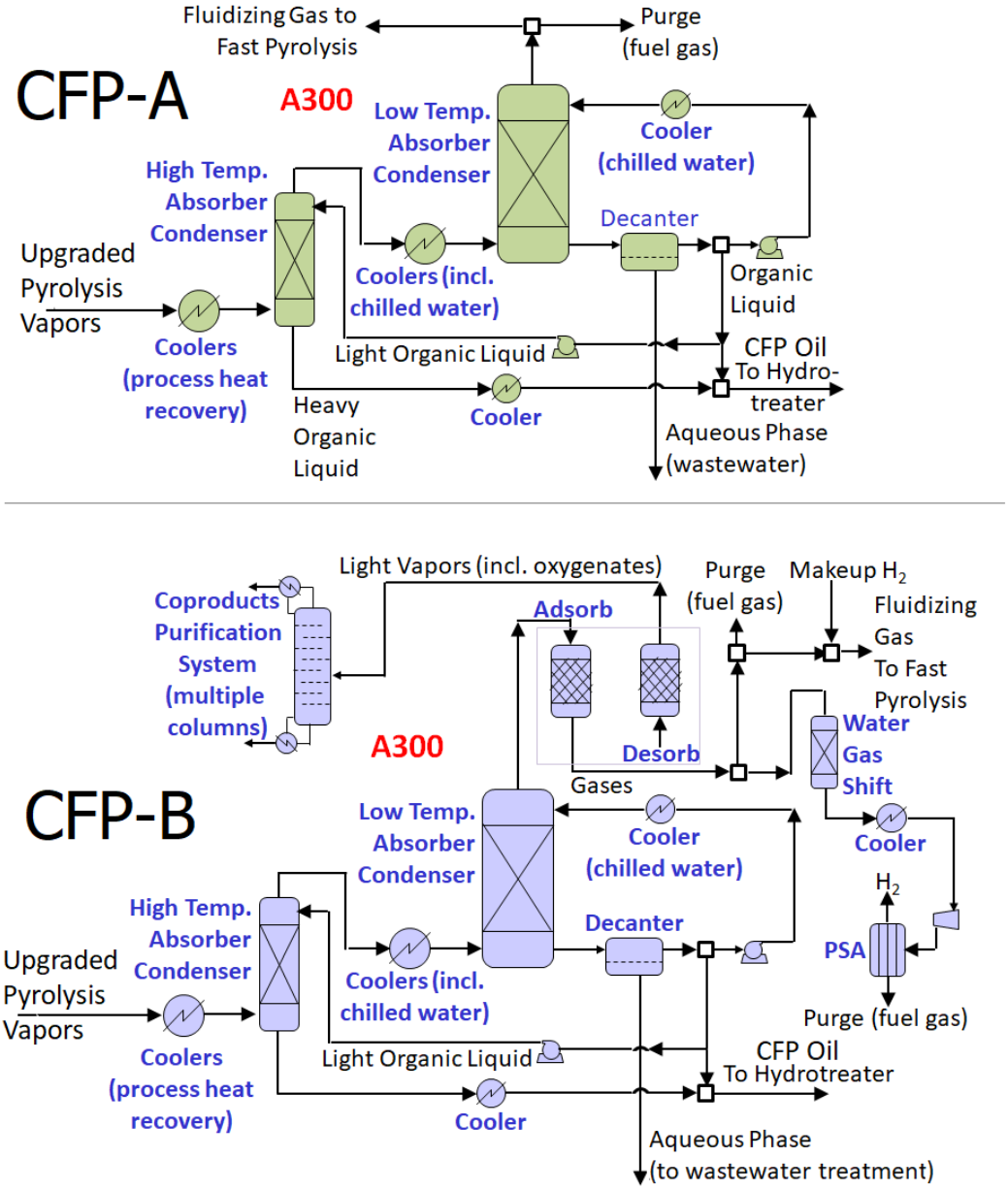

Figure 6. Condensation and products recovery in the CFP-A and CFP-B configurations

Figure 6 shows the condensation trains for CFP-A and CFP-B for the separation of vapors from non-condensable gases. CFP-B has additional operations compared to CFP-A because of the recovery of hydrogen and two light oxygenated coproducts, acetone and 2-butanone (methyl ethyl ketone [MEK]); recovery of the coproducts improve the economics and sustainability metrics for the process.

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In CFP-B, the gases are enriched with supplemental hydrogen in a deliberate attempt to create a hydrogen-rich atmosphere to favorably impact the chemistry in the presence of the Pt/TiO_2 catalyst. Water constitutes a large fraction of the condensable vapors. The choice of direct quench systems was guided by concerns about fouling indirect heat exchange equipment during the condensation of the heavy fraction of the organic vapors. In the model, heat is recovered via indirect heat exchange up to the estimated dew point of the vapors, 249°C (480°F) for CFP-A and 246°C (475°F) for CFP-B, and then followed by direct quench. The first high temperature absorber/condenser recovers the heavy organic fraction using a quench stream of recycled liquid consisting of the light organic fraction from the bottom of the second absorber/condenser supplied at 33°C (91°F) and 49°C (120°F) for CFP-A and CFP-B, respectively. Approximately 45 wt % (CFP-A) and 54 wt % (CFP-B) of the organic product is condensed out and recovered at the bottom of the absorber/condenser. The overhead products from the first absorber/condenser exit at 174°C (346°F) and 203°C (397°F) for the CFP-A and CFP-B cases, respectively, and enter an indirect heat exchanger train to cool down the remaining vapors. Process heat recovery, air cooling to 60°C (140°F), water cooling to 43.3°C (110°F) and chilled water cooling to 15.6°C (60°F) for CFP-A and 32°C (90°F) for CFP-B are used in sequence. This stream then enters a second absorber/condenser where a light organic phase and an aqueous phase are recovered as the bottoms product and separated by decantation. The recycled light organic product is chilled to 10°C (50°F) for CFP-A and 32°C (90°F) for CFP-B, and then used as the quench liquid for this second condenser. Chilled water cooling for CFP-B is deliberately at a higher temperature to allow a larger proportion of light oxygenates to remain uncondensed for subsequent recovery from the gas phase. The light and heavy organic fractions are mixed as the product CFP oil for hydroprocessing (discussed in Section 3). The aqueous fraction is sent to wastewater treatment. Filters are included for all three product streams (heavy organic, light organic, aqueous). Proper filtration to remove particulates from the organic phases is important for reducing the reactivity of CFP oils during storage and for the maintenance of downstream hydroprocessing catalysts.

Off-Gas Recycle in CFP-A

CFP-A does not include the recovery of light oxygenates as coproducts or the addition of hydrogen in the CFP system. Off-gases from FP/CFP reactions, along with fluidizing gases, recovered after condensation are split; 84% of these gases recycled to the fast pyrolysis reactor for fluidization, while the remainder is combusted in Area 500 for heat, steam, and power generation.

Chemical Coproducts Recovery in CFP-B

NREL experimental results for fixed-bed CFP using a Pt/TiO_2 catalyst [4] showed that condensable products retained in the gas phase after the bench-scale experimental condensation train were dominated by three light oxygenates: acetaldehyde, acetone, and MEK. These three compounds constituted >75 wt % of the condensables. Only acetone and MEK were targeted for recovery and purification as saleable products because of their compatibility with current market sizes. In 2020–2021, preliminary NREL experiments showed that an adsorption/desorption strategy can be feasible for the capture of light oxygenates. Although no problems were encountered in the two to three cycles of operation, we understand that additional testing is necessary to prove long-term robustness before scale-up of the system. Near-100% recovery of acetone and MEK was achieved during the adsorption/desorption operations using silica gel as the adsorbent, thus proving the ability to selectively adsorb light oxygenates from the complex

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gas/vapor stream. Following the recovery of a mixed stream of acetone, MEK, and other compounds via desorption, further steps were necessary to achieve chemical-grade purities of the saleable products (acetone and MEK). An extensive phase-equilibrium-based separations model (including several distillation columns and liquid-liquid separations) was developed in Aspen Plus for the purification of acetone and MEK using the detailed experimental speciation of the desorbed stream. The UNIFAC-Dortmund [48] property method was used as the basis for predicting phase equilibrium behavior. The complexities of obtaining high-purity products in the presence of azeotropes, associated losses and energy consumption, and the use of solvents to favorably affect phase equilibrium were captured in the model. In general, comparison of sampled binary systems phase behavior between UNIFAC predictions and experimentally fitted models showed reasonable parity, thus lending credence that our modeling adequately captured the losses, achievable purities, and costs. The modeled purities of acetone and MEK were 97.7 and 99.6 wt %, respectively, with approximately 90% and 65% recovery rates of those purified coproducts, respectively. Additional experiments, if pursued in the future, can prove long-term robustness and further validate property predictions; increasing the recovery of saleable coproducts can significantly benefit the economics and lower the CFP oil price. TEA information for this section, based on the process modeling efforts described above, are included in Table 10 (cost information is included in Appendix F).

Hydrogen Recovery and Recycle in CFP-B

Following the recovery of condensable light oxygenated coproducts, the non-condensable gases are split into (1) a purge stream (1%) to the reformer, (2) recycle to the fast pyrolysis reactor for fluidization (65%), and (3) the balance sent to a sour water-gas shift (WGS) reactor. The inlet temperature of the WGS reactor is 240°C (464°F) [49]. Steam for the WGS reactor is generated from the collected waste aqueous streams using a reboiler (discussed later under Area 800). The product from the WGS reactor is cooled, and condensed water is sent to the wastewater section. The gases are then compressed to approximately 190 psia and cooled to 43.3°C (110°F) before being sent to a PSA unit with assumptions of 84% hydrogen recovery at 99% purity. The PSA off-gases are sent to the reformer section for use either as fuel or reformer feed; this split is determined by the amount of supplemental hydrogen that needs to be generated for maintaining the desired hydrogen partial pressure in the fluidizing gas for CFP-B and for satisfying the fuel heat requirement of the reformer system if that becomes the limiting constraint (this is not the limiting constraint in the CFP-B model). The hydrogen recovered in the PSA is recycled with the fluidizing gas feed to the FP reactor. Supplemental hydrogen generated in the off-gas reformer is added to maintain a H_2 partial pressure of approximately 94 psia (6.5 bar) in the fluidizing gas feed into the FP reactor (Table 7).

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2.3.3.2 Area 300 Design Basis

Table 9 summarizes the design basis and other key information for this area.

Table 9. Key Model Parameters and Flows for Condensation System

	CFP-A	CFP-B
Heavy fraction condenser-absorber		
Temperature, °F	346	397
Pressure, psia	30	100
Feed rate, lb/h	498,980	321,012
Recycle liquid flow, lb/h	78,615	52,335
Vapor product, lb/h	560,760	349,570
Bottoms liquid product, lb/h	16,835	23,778
Light fraction condenser-absorber		
Feed temperature, °F	60	90
Overhead pressure, psia	21	91
Recycle liquid flow, lb/h	726,606	218,064
Vapor product, lb/h	390,898	219,016
Bottoms liquid flow, lb/h	896,467	348,619
Mole % H_2 overhead in gas	-	64%
Total aqueous phase		
Flow rate, lb/h	70,657	57,841
Mass percent C	8.1	2.9
Total organic phase		
Flow rate, lb/h	37,387	44,130
Mass percent C, H, O	71%, 7%, 22%	76%, 8%, 16%

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Table 10. Key Model Parameters for Light Oxygenates Recovery and Hydrogen Purification in CFP-B

	CFP-B
Light Oxygenates Recovery	
Acetone product % recovery	90%
Acetone product (lb/h)	5,336
MEK product % recovery	65%
MEK product (lb/h)	926
Utilities (included in overall CFP model)	
Electricity (kW)	260
Air cooler duty (MMBtu/h)	45.6
Cooling water duty (MMBtu/h)	2.6
Chilled water duty (MMBtu/h)	9.4
Cold streams for heat integration (MMBtu/h)	140
Hot streams for heat integration (MMBtu/h)	80
Water-Gas Shift	
Gas flow rate, lb/h	66,671
Inlet temperature, °F	464
Outlet temperature, °F	651
Pressure, psia	81
Steam flow rate, lb/h	42,659
Temperature approach to equilibrium, °F	35
Wastewater (condensate), lb/h	27,698
Pressure Swing Adsorption (PSA)	
Inlet flow rate, lb/h	81,633
Inlet hydrogen mole %	69%
Hydrogen recovery %	84%
Hydrogen purity %	99%
Pure H_2 in product stream, lb/h	7,583

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2.3.4 Area 400: CFP Oil Hydroprocessing

Our previous publications [18,19] presented CFP oil hydroprocessing under Area 400. This topic is the primary focus of this report. Section 3 of this report is dedicated to the discussion of the hydroprocessing of CFP oils and petroleum feedstocks in stand-alone and co-hydroprocessing configurations.

2.3.5 Area 500: Off-Gas Utilization

2.3.5.1 Area 500 Overview

In CFP-A, electricity is the only valuable coproduct. Thus, the primary use of off-gases from CFP is for combustion to produce heat, steam, and electricity for use in the process. Excess energy is converted to electricity and sold to the grid.

In the CFP-B configuration with Pt/TiO_2 catalyst, hydrogen is also produced to meet CFP needs. Since we assume a stand-alone CFP facility for this base case, no excess hydrogen is produced because of the assumed absence of local consumers or a pipeline for H_2 transportation. The presence of local industrial consumers utilities in a colocated facility may open up additional options for the use of off-gases as fuel gas, as well as the consumption of excess heat, steam, and hydrogen. These options and potential benefits of colocation are discussed in detail in Dutta et al. 2023 [12].

Hydrogen Production in CFP-B

A fixed-bed steam methane reformer package was used for hydrogen production. There is sufficient off-gas fuel available in the process to provide all the heat duty necessary to run the reformer. The reformer was modeled as a Gibbs equilibrium reactor operating at 115 psia and 880°C (1,616°F) with a 30°C (54°F) approach to equilibrium. Steam is provided from a wastewater recycle reboiler in Area 800. The product gases are quenched and sent to high temperature and low temperature shift reactors in series with intermediate heat exchange for temperature control. Following the water gas shift reactors, the gases are cooled and condensate removed. The gases are then compressed to 278 psia before feeding to a PSA. The PSA design in this steam reformer section assumes 80% hydrogen recovery with 99% purity (note this PSA is different from the PSA in the CFP gas recycle loop for CFP-B). Off-gases from the PSA are used as combustion fuel. PSA recovery can vary depending on the feed gas composition and other parameters; in an energy and hydrogen self-contained design such as this, where excess off-gas is combusted for steam and power generation, the PSA recovery parameter should not have a significant impact on the process economics and sustainability.

The process impacts of using biogenic off-gases for the production of hydrogen in a steam methane reformer configuration were assessed by Dutta 2023 [50]. As in previous reports [18,19], some of the additional capital costs beyond a steam methane reformer configuration are accounted here through the cost of heat integration via pinch analysis, added on top of a standard steam methane reformer cost estimate.

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2.3.5.2 Area 500 Design Basis

Table 11 summarizes Area 500 hydrogen production parameters for CFP-B. Further details, including equipment costs, are available in Appendix F.

Table 11. Hydrogen Production Parameters for CFP-B

Hydrogen Plant Design Parameters	
Hydrogen production rate	14.8 MMscfd
Hydrogen outlet purity	99 mol %
Hydrogen outlet pressure	266 psia

2.3.6 Area 600: Steam System and Power Generation

2.3.6.1 Area 600 Overview

As in previous designs [18,19], a conventional steam cycle uses process heat to produce electricity. It would be ideal if most of the energy in the biomass feedstock could be converted to CFP oil. However, depending on the efficiency and process design, different amounts of excess energy are available after the CFP process. In this report, CFP-B is more efficient than CFP-A for the production of CFP oil. However, there is a significant amount of heat available from both the CFP-A and CFP-B processes that needs to be managed efficiently to positively impact the economics. The big sources of heat are the char combustion during fast pyrolysis and coke burnoff from the CFP catalysts, cooling of upgraded pyrolysis vapors (including fluidization gases), off-gas combustion, and reformed gas quench (in CFP-B). In both the CFP-A and CFP-B designs there is excess electricity available after all in-plant demands are met. This electricity is sold to the grid and reduces the MFSP by $0.60/GGE and $0.20/GGE for the CFP-A and CFP-B cases, respectively. CFP-B has a lower electricity credit compared to CFP-A because more energy is retained in the CFP-B oil product due to higher yield and lower CFP-B oil oxygen content, and a portion of the off-gases in CFP-B are used for coproducts recovery and hydrogen production instead of the production of electricity.

Excess heat available for power generation was determined by conducting an overall plant heat balance. The heat balance assumptions are checked for thermodynamic viability by drawing heating and cooling curves for pinch analysis and ensuring there is no crossover between the curves. Electricity is generated using two steam turbines, with intermediate reheat. Pre-heaters, steam generators, and super-heaters are integrated within the process design to generate the steam from boiler feed water. Process condensate is recycled to the steam cycle, de-gassed, treated, and combined with makeup water.

The steam turbine efficiencies are assumed to be 75% and the generator mechanical efficiencies are assumed to be 97%. The first stage turbine drops the steam pressure from 1,321 psia to 350 psia. The second stage turbine drops the pressure from 345 psia to 3.5 psia. The steam inlets to both turbine stages are preheated to 1,000°F.

Low pressure (3.5 psia) steam exhaust from the turbine is cooled by exchanging heat with other process streams and is condensed in an air-cooled condenser. A condensate collection tank gathers the condensate returns from indirect heat exchange along with the steam turbine

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condensate. The condensate from the collection tank is treated in the hot condensate polishing unit, mixed with treated makeup water, and sent to the condensate surge tank. From the surge tank, the water is pumped, heated to the saturation temperature, and sent to the deaerator to remove any dissolved gases from the water. The water from the deaerator is pumped to a pressure of 1,351 psia and then preheated to its saturation (bubble point) temperature using a series of exchangers. Steam is generated by heat exchange with hot streams in the process. This saturated steam is collected in the steam drum. To prevent buildup of solids, water is periodically discharged from the steam drum (also known as blowdown). Saturated steam from the steam drum is superheated with another series of exchangers to 538°C (1,000°F) before it goes to the steam turbine.

The Aspen Plus model increases or decreases the boiler feed water flow rate through the steam cycle in order to close the heat balance. The control system in the plant will function in a similar fashion by controlling the flow rate of boiler feed water makeup to the steam system in order to maintain levels in the deaerator and steam drum.

All compressors, pumps, fans, etc., utilize electric motor drives and are modeled and quoted accordingly. An allowance of ~3 MW of excess power is specified to account for miscellaneous usage which may not be accounted for in the simulation and general electric needs (such as lights and computers).

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2.3.6.2 Area 600 Design Basis

Table 12 and Table 13 present summaries of the power requirements for each area for the CFP-A and CFP-B cases, respectively. Excess electricity is sold to the grid. Process efficiency metrics are presented later in Section 2.3.13.

Table 12. Plant Power Requirements for CFP-A

Plant Area	Generated (kW)	Consumed (kW)
Feed handling and drying [a]		650
Pyrolysis		15,437
Condensation and product separation		1,764
Hydrogen production and management		2,476
Steam system and power generation	73,472	2,374
Cooling water and other utilities		457
Wastewater management		6,317
Miscellaneous		2,948
Balance to grid		*41,049*
Plant totals	**73,472**	**73,472**

[a] Power costs for all feed handling, and for drying outside the plant, are included in the woody feedstock cost provided by INL.

Table 13. Plant Power Requirements for CFP-B

Plant Area	Generated (kW)	Consumed (kW)
Feed handling and drying [a]		383
Pyrolysis		15,354
Condensation and product separation		5,009
Hydrogen production and management		5,487
Steam system and power generation	50,905	1,670
Cooling water and other utilities		300
Wastewater management		1,538
Miscellaneous		2,974
Balance to grid		*18,190*
Plant totals	**50,905**	**50,905**

[a] Power costs for all feed handling, and for drying outside the plant, are included in the woody feedstock cost provided by INL.

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2.3.7 Area 700: Cooling Water and Other Utilities

2.3.7.1 Area 700 Overview

Equipment list for this area is included in Appendix E (CFP-A) and Appendix F (CFP-B). A mechanical draft cooling tower (M-701) provides cooling water to several heat exchangers in the plant. The tower utilizes fans to force air through circulated water.

An instrument air system is included to provide compressed air for both service and instruments. The system consists of an air compressor (K-701), dryer (S-701), and receiver (T-701). The instrument air is delivered at a pressure of 115 psia and a moisture dew point of −40°C (−40°F), and it is oil free.

Other miscellaneous items that are taken into account in the design include the following:

- Firewater storage tank (T-702) and firewater pump (P-702) for fire emergencies
- Diesel tank (T-703) and diesel pump (P-703) to fuel the front loaders
- Ammonia storage tank (T-704), ammonia pump (P-704), boiler feed water chemical storage tank (T-708), and boiler feed water chemical pump (P-707) for boiler feed water treatment.
- Caustic (NaOH) storage tank (T-705) and pump (P-705).

2.3.7.2 Area 700 Design Basis

Table 14 summarizes the design basis parameters for the plant's cooling water system.

Table 14. Design Basis for Cooling Water System

Cooling Water System Design Parameters	
Cooling water supply temperature (T_S)	90°F
Cooling water return temperature (T_R)	110°F
Losses from evaporation (F_{EVAP})	$0.00085 \times F_S \times (T_R - T_S)$
Losses from drift	$0.002 \times F_S$
Cycles of concentration (C)	5
Required blowdown	$F_{EVAP}/(C - 1)$
Cooling water makeup pressure	14.7 psia
Cooling water makeup temperature	60°F

Note: Fs = Cooling water supply rate in gallons per minute.

Water losses from the cooling water system include evaporation, drift (water entrained in the cooling tower exhaust air), and tower basin blowdown. Drift losses were estimated to be 0.2% of the cooling water supply rate. Evaporation losses and blowdown were calculated based on information and equations in Perry et al. [51] (as shown in Table 14).

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2.3.8 Area 800: Wastewater Management and Recycle

2.3.8.1 Area 800 Overview

There are differences in the Area 800 designs for CFP-A and CFP-B because CFP-B includes hydrogen use and production; hydrogen production requires the use of steam in the reformer and the WGS reactor. The general design for both systems is based on the use of a regenerative thermal oxidizer for the combustion of organics, as detailed in our previous design report [18].

For CFP-B Area 800 has three process sub-areas for treating and recycling water discharges from the plant: a wastewater boiler, a regenerative thermal oxidizer, and a reverse osmosis unit for point treatment. The majority of the soluble compounds containing carbon is found in the aqueous phase of the pyrolysis quench loop in Area 300. This stream, with a significantly smaller makeup of clean recycled water is sent to a waste heat boiler system to first generate steam for process use in the sour WGS reactor in Area 300. Any H_2S present in the wastewater is expected to be stripped out of the water during this step. A subsequent boiler generates steam for the reformer; small proportions of carbon species are also part of the steam and are expected to be utilized in the pre-reformer/reformer system for hydrogen production. Generating steam from the CFP quench water serves to recycle some of the volatile carbon species back to the process, and also to concentrate the remaining soluble species like phenol. To avoid fouling in the boiler, not all the wastewater is evaporated. This stream, more concentrated with carbon species, is sent to a regenerative thermal oxidizer to combust the organics. For CFP-A the steam generation step is skipped, and the entire CFP aqueous phase is sent to the regenerative thermal oxidizer.

An on-site reverse osmosis unit is specified for point treatment and recycle of streams with low contamination. These include the cooling tower blowdown, water knockout from the PSA compressor in Area 300, and the flue gas scrubber effluent from Area 100. We assume that 75% of the water can be recovered and recycled to the flue gas scrubber, Area 600 boiler, and cooling tower to offset the total water demand for the process. The remaining 25% is collected for treatment, along with the hydroprocessing water knockout streams, which were considered unsuitable for the boiler and regenerative thermal oxidizer due to nitrogen (ammonia) content. While we did not perform a detailed design for an on-site wastewater treatment plant, it is expected that such treatment will entail demineralization of the reverse osmosis retentate streams and probably activated sludge denitrification of the hydroprocessing water, followed by aerobic digestion and polishing. Anaerobic digestion is not likely to be economical due to the low remaining chemical oxygen demand (COD). We estimated an operating cost charge for wastewater using the aggregate CAPEX+OPEX cost contribution of the detailed wastewater facility designed for the 2011 biochemical ethanol design report [52]. At $0.022/gal wastewater, this conservatively estimated cost of treating the relatively less contaminated streams from the pyrolysis plant is a factor of 10 higher than the value cited in Peters and Timmerhaus [53] for treatment of boiler blowdown streams.

Table 15 summarizes the water balance for the plant. The consumptive water demand for the process is 2.80 gal/GGE for CFP-A and 1.19 gal/GGE for CFP-B. Along with uncertainties surrounding the choice of equipment in Area 800, we note that there could also be significant future opportunities to recover valuable products from the wastewater using processes such as aqueous phase reforming or biological upgrading if the CFP process is commercially established.

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Table 15. CFP Plant Water Balance

Area	Stream	Source/Destination	CFP-A	CFP-B	Unit
A100	Scrubber water	Input	30,000	30,000	lb/h
	To flue gas	Loss	15,017	8,311	lb/h
	Scrubber blowdown	To point treatment	14,983	21,689	lb/h
	COD		*0*	*0*	*lb/h*
A200	Catalyst steam stripper	From boiler A800	2,756		lb/h
A300	Aqueous phase	To boiler A800	62,003	55,311	lb/h
	COD		*14,539*	*4,613*	*lb/h*
	WGS steam demand	From boiler A800		42,325	lb/h
	PSA water knockout	To point Treatment		27,419	lb/h
	COD			*0*	*lb/h*
A500	Reformer steam demand	From boiler A800		25,958	lb/h
	PSA water knockout	To boiler A800		8,683	lb/h
	COD			*0*	*lb/h*
A600	Makeup	Input	10,748	7,447	lb/h
	Makeup	Input	4	3	lb/h
	Steam drum knockout	To cooling tower basin	10,752	7,449	lb/h
A700	Steam drum knockout	From A600	10,752	7,449	lb/h
	Makeup	Input	72,182	47,085	lb/h
	Blowdown	To point treatment	15,140	9,951	lb/h
	Drift	Loss	7,236	4,778	lb/h
	Evaporation	Loss	60,558	39,806	lb/h
A800	*Boiler*				
	Boiler feed water makeup			18,501	lb/h
	Total boiler feed	To boiler A800		73,812	lb/h
	COD			*4,613*	*lb/h*
	Process steam generated	To steam methane reformer + WGS		68,283	lb/h
	Remaining to regenerative thermal oxidizer	Loss	62,003	5,529	lb/h
	COD		*14,539*	*2,687*	*lb/h*
	Point treatment				
	Total water for treatment	75% recovered	30,122	67,743	lb/h
	Water recycled	Offsets inputs	22,592	50,807	lb/h
	Remainder	To full treatment	7,531	16,936	lb/h
	COD		*0*	*527*	*lb/h*
	Full treatment				
	Total	Loss	7,531	16,936	lb/h
	COD		*0*	*527*	*lb/h*
	Total water makeup		115,689	103,035	lb/h
	Net water makeup		93,097	52,228	lb/h
	Net water makeup (annual)		88.2	49.5	MMgal/y
	Water usage		2.80	1.19	gal/GGE
	Liquid water emission		7,531	16,936	lb/h
	Liq. water emission (annual)		7.1	16.0	MMgal/y

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2.3.9 Additional Process Design Information

Table 16 contains additional information used in the Aspen Plus model and design basis.

Table 16. Utility and Miscellaneous Design Information

Item	Design Information
Ambient air conditions [a,b,c]	Pressure: 14.7 psia $T_{dry\ bulb}$: 90°F $T_{wet\ bulb}$: 80°F Air composition (mol %): N_2: 75.7% O_2: 20.3% Ar: 0.9% CO_2: 0.03% H_2O: 3.1%

[a] In the GPSA *Engineering Data Book* [54], see Table 11.4 for typical design values for dry bulb and wet bulb temperature by geography. Selected values would cover summertime conditions for most lower 48 states.

[b] In Weast [55], see F-172 for composition of dry air. Nitrogen value is adjusted slightly to force mole fraction closure using only nitrogen gas (N_2), oxygen (O_2), argon (Ar), and carbon dioxide (CO_2) as air components.

[c] In Perry et al. [51], see psychrometric chart, Figure 12-2, for moisture content of air.

2.3.10 Pinch Analysis

While an energy balance was performed within the Aspen Plus process simulation, it was also necessary to ensure that there was no temperature crossover when the assumption was made that the cold process streams can be heated using the available heat in hot process streams. A pinch analysis was performed to show the temperatures at which hot and cold duties are available. This step is included in our thermochemical analysis to ensure thermodynamic integrity of our heat exchange assumptions [18]. For CFP-A and CFP-B the ΔT_{min} were 39.2°F and 16.8°F, respectively, with total duties of in-process exchange 969 MMBtu and 999 MMBtu, respectively. The heat exchange network cost was estimated by using the in-process exchange duties as scaling variables and a scaling exponent of 0.65, with base cost and duty information derived from the *ex situ* network in the 2015 design report [18].

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2.3.11 Modeled CFP Process Performance Metrics Summary

A summary of process performance is presented in Table 17. The information reflects the assumptions based on bench-scale experimental data discussed previously in Section 2.2.2. Additional metrics from the analysis are included in the following sections.

Table 17. Model Metrics for CFP-A and CFP-B

Metric	CFP-A	CFP-B [a]
Gas with some condensables (wt % of dry biomass)	34.0%	41.8%
Aqueous phase (wt % of dry biomass)	25.8%	20.5%
Carbon loss (% of C in biomass)	*6.31%*	*1.79%*
Organic phase (wt % of dry biomass)	20.2%	24.0%
H/C molar ratio	*1.18*	*1.30*
Oxygen (wt % in organic phase, dry basis)	*21.6%*	*16.0%*
Carbon efficiency, with some condensables (%)	*28.8%*	*35.7%*
Solid losses (char + coke)	12.7% + 7.1%	11.1% + 1.7%
CFP oil LHV (Btu/lb, based on model estimate)	12,408	13,908
Specific gravity (based on experiments)	1.151	1.053

[a] The data represent the average of three experimental repeats; the information presented in Section 2.2.2 was from one of the three repeats.

2.3.12 Process Economics

This section describes how the process model, detailed design, and cost estimation discussed so far is used to determine the economics of CFP oil production. The analysis is based entirely on estimated equipment costs and regular construction and project management costs associated with building a plant of this kind along with associated operating costs. The analysis does not consider any policy factors such as carbon credits, subsidies, or mandates.

The FCI is first computed using a comprehensive list of process and mechanical equipment in the plant and its expected construction and project costs. Variable and fixed operating costs associated with plant operation and maintenance are determined next. With these estimated costs, a discounted cash flow analysis is used to determine a minimum fuel intermediate selling price (MFISP) that resolves to a net present value of zero for the entire project at some assumed nonzero discount rate, also referred to in this report as the internal rate of return. In other words, given the production rates determined in the process model, the CFP oil fuel intermediate from the plant must be sold at its MFISP in order for the project to return the specified internal rate of return and no more. This section describes the calculations and assumptions made in completing the discounted cash flow analysis.

2.3.12.1 Total Capital Investment

This section summarizes the purchased and installed equipment costs for CFP-A and CFP-B. For the CFP processes in this report, the total installed costs for the plant serve as the basis for determining the total capital investment. Most of the equipment costs are from previous work [18,19] and were derived from various sources including equipment suppliers and licensors, cost estimates from Harris Group, published literature, and Aspen Capital Cost Estimator software. The reader should refer to the previous references [18,19] for further details. A complete listing

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of the equipment, along with its purchased and installed costs are included in Appendix E (CFP-A) and Appendix F (CFP-B). Table 18 presents a summary of the total purchased equipment cost (TPEC), aggregate installation factors ($f_{install}$), and total installed cost for each area of the plant.

Table 18. Cost Factors for Total Installed Equipment Costs

		CFP-A			CFP-B		
Area	**Process Description**	**TPEC (MM$)**	**$f_{install}$**	**Total Installed Cost (MM$)**	**TPEC (MM$)**	**$f_{install}$**	**Total Installed Cost (MM$)**
100	Feed handling and drying [a]	0.37	1.95	0.73	0.26	1.96	0.51
200	Pyrolysis	68.02	2.79	189.53	47.96	2.40	114.90
300	Pyrolysis vapor quench	6.99	1.85	12.90	36.59	1.55	56.80
500	Off-gas utilization	4.27	2.63	11.25	27.21	2.02	55.00
600	Steam plant and power	42.61	1.82	77.69	31.70	1.85	58.81
700	Cooling water and other utilities	5.42	2.02	10.94	4.54	2.01	9.13
800	Wastewater management	13.33	2.32	30.90	6.84	2.35	16.09
Inside Battery Limits	**Areas 100–300**	75.38	2.70	203.16	84.81	2.03	172.21
Outside Battery Limits	**Areas 500–800**	65.64	1.99	130.78	70.31	1.98	139.02
Total		141.03	2.37	333.94	155.12	2.01	311.23

[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 waste heat.

The sum of equipment purchases and installation/construction costs is defined as the total direct cost. Indirect capital costs are non-construction costs typically associated with project management and engineering, procurement, and construction services. Indirect costs are estimated with factors on the total direct cost as shown in Table 19. The sum of direct and indirect costs is defined as the FCI. Finally, working capital is determined to be 5% of the FCI. The sum of FCI and working capital is the total capital investment. Table 20 presents a summary of these capital quantities for CFP-A and CFP-B.

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Table 19. Cost Factors and Indirect Costs

Indirect Costs	% of Total Direct Cost [a]
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**

[a] Excluding land purchase cost.

Table 20. Project Cost Worksheet

		CFP-A	CFP-B
Total purchased equipment cost (TPEC)		**$141,030,000**	**$155,120,000**
Installation factor		2.368	2.006
Total Installed Cost		**$333,940,000**	**$311,230,000**
Other direct costs			
Land (not depreciated)		$1,610,000	$1,850,000
Warehouse	4.0% of inside battery limits (ISBL)	$8,130,000	$6,890,000
Site development	10.0% of ISBL	$20,320,000	$17,220,000
Additional piping	4.5% of ISBL	$9,140,000	$7,750,000
Total direct costs (TDC)		**$371,520,000**	$343,090,000
Indirect costs	% of TDC (ex land)		
Prorated expenses	10.0%	$37,150,000	$34,310,000
Home office and construction fees	20.0%	$74,300,000	$68,620,000
Field expenses	10.0%	$37,150,000	$34,310,000
Project contingency	10.0%	$37,150,000	$34,310,000
Other costs (startup and permits)	10.0%	$37,150,000	$34,310,000
Total indirect costs	**60.0%**	**$222,910,000**	**$205,850,000**
FCI		**$594,430,000**	**$548,940,000**
Working capital	5.0% of FCI (ex land)	$29,720,000	$27,450,000
Total capital investment (TCI)		**$624,150,000**	**$578,240,000**
TCI/TPEC		4.426	3.728
FCI Lang factor = FCI/ISBL TPEC		7.886	6.473
TCI Lang factor = TCI/ISBL TPEC		8.280	6.818

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2.3.12.2 Variable Operating Costs

Variable operating costs, which include raw materials, waste handling charges, and byproduct credits, are incurred only when the process is operating. Quantities of raw materials used and wastes produced were determined using the Aspen Plus modeled material balance.

Table 21 summarizes the basis for variable operating costs considered in this analysis. A detailed summary of variable plant operating costs is presented in Appendix E (CFP-A) and Appendix F (CFP-B) on a per-year and per-GGE basis.

Table 21. Summary of Variable Operating Costs

Variable	Information and Operating Cost (cost year in parentheses)
Fluidized bed media	For fast pyrolysis, the bed medium is sand. For CFP-A the medium is a zeolite catalyst. Initial fill then a replacement rate of 2 wt % of circulation per day. Sand price: $45.74/ton (2011) [18]. Zeolite catalyst price: $2.05/lb in 2020$ (estimated using CatCost [56] tool) For the fixed-bed system in CFP-B: Pt/TiO_2 catalyst (0.5% Pt loading). Catalyst price: $99.14/lb in 2020$ using CatCost [56], with 70% salvage value during replacement. Catalyst lifetime is assumed 2 years before replacement.
Hydrogen	Not purchased from external sources in CFP model. Placeholder price of $1.5/kg in 2020$ used if necessary for assessing purchase or sale of hydrogen.
Steam methane reformer catalysts	Based on a literature value of price per unit hydrogen produced. Price: $7.80/ton hydrogen (2011) [18].
Natural gas	Purchased from pipeline for feed to steam methane reformer for hydrogen production. No or minimal use as placeholder in CFP model. Price: $3.5/MMBtu (2016).
Solids disposal	Price: $33/tonne (1998) [53].
Diesel fuel	Usage: 10 gal/h plantwide use. 2012 price projection: $21.29/MMBtu (2009) [18] = $2.86/gal at 0.85 specific gravity.
Water makeup	Price: $0.22/tonne (2001) [18] = $0.20/ton.
Chemicals	Boiler feed water chemicals – Price: $75/MMlb blowdown (2014 vendor estimate) [18]. Cooling tower chemicals – Price: $33.84/year per ton of cooling capacity (2014 vendor estimate) [18]. Caustic – Price: $150/dry ton (2010) [18].
Wastewater	Most wastewater is cleaned using a reverse osmosis system and recycled. Additional treatment is assumed for the balance. Price: $0.022/gal (2011). Based on Humbird et al. [52].
Chemical coproducts	Acetone: $0.442/lb; MEK: $0.771/lb, approximate from past 5-year averages [4]. Same prices assumed in 2020$ for this report.
CFP oil rail transport	5.1 cents/ton-mile in 2007$ [57], assuming an 800-mile distance for the base cases. Update to 2020$ using producer price index for chemical manufacturing [25], because the ratio of indices for 2007$ and 2020$ was similar to the producer price index for rail transportation (data.bls.gov/cgi-bin/srgate series ID: PCU4821--4821--).

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2.3.12.3 Fixed Operating Costs

Fixed operating costs, which include employee salaries and benefits, overhead, plant maintenance costs, insurance, and taxes (other than income taxes), are generally incurred in full whether or not the plant is producing at full capacity. Table 22 lists the expected number of employees for the plant and their associated salaries. The salaries listed do not include benefits, so a general overhead factor is applied, as shown in Table 23, along with fixed costs for insurance, taxes, and general plant maintenance, such as security, janitorial services, and communications.

Table 22. Salary Costs for Plant Employees

Position Title	Salary (2007)	Number of Positions	Total Cost (2020)
Plant manager	$147,000	1	$196,000
Plant engineer	$70,000	1	$93,000
Maintenance supervisor	$57,000	1	$76,000
Laboratory manager	$56,000	1	$75,000
Shift supervisor	$48,000	5	$320,000
Lab technician	$40,000	2	$107,000
Maintenance technician	$40,000	16	$852,000
Shift operators	$48,000	20	$1,278,000
Yard employees	$28,000	12	$447,000
Clerks and secretaries	$36,000	3	$144,000
Total salaries (2020$/year)			**$3,587,000**

Note: Labor costs are indexed to values from the U.S. Bureau of Labor Statistics (data.bls.gov/cgi-bin/srgate CEU3232500008).

Table 23. Fixed Operating Costs

Cost Item	Factor	CFP-A	CFP-B
Salaries		$3,587,000	$3,587,000
Benefits and overhead	90.0% of total salaries	$3,228,000	$3,228,000
Maintenance	3.0% of FCI [a]	$17,833,000	$16,468,000
Insurance and taxes	0.7% of FCI [a]	$4,161,000	$3,843,000
Total fixed operating costs (2020$/year)		**$28,809,000**	**$27,126,000**

[a] Percentages of FCI exclude land purchase cost.

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2.3.12.4 Discounted Cash Flow Analysis and the MFISP

Once the capital and operating costs are determined, an MFISP is determined using a discounted cash flow rate of return (DCFROR) analysis. The general methodology used is the same as that applied in previous NREL reports [4,18], and example spreadsheets are available publicly [46]. We normalize the CFP oil intermediate on an LHV energy basis. In this case we use a single gasoline-equivalent product for calculations, although any other basis could be used, and costs can easily be converted to other energy units of interest. The discounted cash flow analysis determines the minimum selling price of a gallon of this gasoline-equivalent product, which yields a net present value of zero for the project. The financial parameters used to compute the net present value are as follows.

Discount Rate/Internal Rate of Return: For this analysis, the cash flow discount rate was set to an internal rate of return of 10%.

Equity Financing: For this analysis, it was assumed that the plant would be 40% equity financed. The terms of the loan on the remaining 60% debt were taken to be 8% interest for 10 years. The principal is taken out in stages over the 3-year construction period. Interest on the loan is paid during this 3-year period, but principal is not paid back. (This implies that cash flow comes from a parent company until the plant starts up.)

Depreciation: Capital depreciation is computed according to the IRS Modified Accelerated Cost Recovery System. The bulk of the plant capital is depreciated over a 7-year recovery period. Because the plant described here is a net exporter of electricity, the steam plant and power generation equipment is depreciated over a 20-year recovery period, according to the IRS.

Taxes: The federal corporate tax rate used in our analysis is 21%. Income tax is averaged over the plant life and that average is calculated on a per-GGE basis. The amount of income tax to be paid varies annually with changes in the volume of product produced and the allowable depreciation deduction. In fact, no income tax is paid in the first 8 years of operation because depreciation and loan interest deductions are greater than the net income. State taxes are not considered, primarily because the location of the plant has not been determined and tax rates vary from state to state (from 0% to 12%).

Construction Time: The construction time was taken to be 24 months. Twelve months are added before construction for planning and engineering.

Startup Time: Startup was taken to be 25% of the construction time, or 6 months in this case. It is assumed that the plant achieves 50% production during the startup period while incurring 75% of variable expenses and 100% of fixed expenses.

Lifetime: The plant and all equipment are assumed to have a 30-year life. No salvage value is assumed at the end of 30 years.

Working Capital: Working capital was taken to be 5% of the FCI.

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2.3.12.5 CFP Process TEA Summaries

TEA summaries for CFP-A and CFP-B are presented in Table 24 (CFP-A) and Table 25 (CFP-B). As expected, CFP-B has a lower MFISP than CFP-A because of higher CFP oil yield and the benefits from the co-production of saleable chemical coproducts.

Table 24. TEA Summary for CFP-A

Case CFP-A: Process Engineering Analysis for CFP Oil Production via Fluid Bed Ex Situ Upgrading of Fast Pyrolysis Vapors

Potential CFP Oil Production as a Hydroprocessing Feedstock for Liquid Hydrocarbon Fuels Production
2,000 Dry Metric Tonnes Biomass per Day
All Values in 2020$

Minimum Fuel-Intermediate Selling Price (MFISP)	**$5.13**	**/Gallon CFP Oil with 800 miles Transportation**	
	$216	**/Barrel CFP Oil with Transportation**	
	$204	**/Barrel CFP Oil without Transportation**	
	$5.00	**/Gallon Gasoline Equivalent (GGE) with Transport**	
CFP Oil Production	30.7	MM Gal per Year	42.4 Gal per Dry US Ton Feedstock
Total Gasoline Equivalent Production	31.5	MM GGE per Year	43.5 GGE per Dry US Ton Feedstock
CFP Oil Oxygen Content (wt% dry basis)	21.6%	Energy Content	1.026 GGE/gal (for 116090 Btu/GGE)
Delivered Feedstock Cost (2020$)	$72.53	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	
100: Feedstock (Additional Dryer & Blower Only)	$730,000
200: Fast Pyrolysis & Vapor Upgrading	$189,530,000
300: Pyrolysis Vapor Quench	$12,900,000
400: Hydroprocessing & Separation	$0
500: Off-Gas Utiization	$11,250,000
600: Steam System & Power Generation	$77,690,000
700: Cooling Water & Other Utilities	$10,940,000
800: Water Management	$30,900,000
Total Installed Equipment Cost (TIC)	$333,940,000
Land (115 Acres at $14000 per Acre)	$1,600,000
Site Development	$20,320,000
(% of ISBL)	10.0%
Indirect Costs & Project Contingency	$222,910,000
(% of TIC)	66.8%
Fixed Capital Investment (FCI)	$594,430,000
Working Capital	$29,720,000
Total Capital Investment (TCI)	$624,150,000
Loan Interest Rate	8.0%
Loan Term (Years)	10
Total Installed Equipment Cost per Annual GGE	$10.60
Fixed Capital Investment per Annual GGE	$18.87
Plant Operating Hours per Year	7884
On-Stream Percentage	90.0%

Specific Operating Conditions		
Feed Rate	Dry Tonnes / Day	2,000
	Dry Tons / Day	2,205
Feed Moisture and Dry Basis Ash, wt%		10% moist, 0.97% ash
Feedstock Cost	$/Dry Ton	$72.53
	$/Moisture+Ash Free Ton	$73.23

Version: FluidBed-ExSitu-22pct_O-DR002-XL004

Operating Costs (¢ / GGE Product)	
Feedstock	166.7
Natural Gas	0.0
Catalysts	11.7
Sand	0.8
Other Raw Materials	5.6
Waste Disposal	2.5
CFP Oil Transportation	27.1
Fixed Costs	91.4
Electricity Coproduct Credit	(60.1)
Capital Depreciation	94.3
Average Income Tax	19.9
Average Return on Investment	140.1

Operating Costs ($ / Year)	
Feedstock	$52,520,000
Natural Gas	$0
Catalysts	$3,700,000
Sand	$200,000
Other Raw Materials	$1,780,000
Waste Disposal	$780,000
CFP Oil Transportation	$8,550,000
Fixed Costs	$28,810,000
Electricity Coproduct Credit	-$18,940,000
Capital Depreciation	$29,720,000
Average Income Tax	$6,280,000
Average Return on Investment	$44,150,000

Total Plant Electricity Usage (kW)	32,424
Electricity Produced on Site (kW)	73,472
Electricity Purchased from Grid (kW)	0
Electricity Sold to Grid (kW)	41,049
Plant Electricity Use (kWh /GGE)	8.11
Overall Plant Efficiency to CFP Oil - HHV %	32.1%
Overall Plant Efficiency to CFP Oil - LHV %	32.9%

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Table 25. TEA Summary for CFP-B

Case CFP-B: Process Engineering Analysis for CFP Oil Production via Fixed Bed Ex Situ Upgrading of Fast Pyrolysis Vapors

Potential CFP Oil Production as a Hydroprocessing Feedstock for Liquid Hydrocarbon Fuels Production
2,000 Dry Metric Tonnes Biomass per Day
All Values in 2020$

Minimum Fuel-Intermediate Selling Price (MFISP)	**$3.53**	**/Gallon CFP Oil with 800 miles Transportation**	
	$148	**/Barrel CFP Oil with Transportation**	
	$138	**/Barrel CFP Oil without Transportation**	
	$3.35	**/Gallon Gasoline Equivalent (GGE) with Transport**	
CFP Oil Production	39.6	MM Gal per Year	54.7 Gal per Dry US Ton Feedstock
Total Gasoline Equivalent Production	41.7	MM GGE per Year	57.6 GGE per Dry US Ton Feedstock
CFP Oil Oxygen Content (wt% dry basis)	16.0%	Energy Content	1.053 GGE/gal (for 116090 Btu/GGE)
Delivered Feedstock Cost (2020$)	$72.53	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	
100: Feedstock (Additional Dryer & Blower Only)	$510,000
200: Fast Pyrolysis & Vapor Upgrading	$114,900,000
300: Pyrolysis Vapor Quench	$56,800,000
400: Hydroprocessing & Separation	$0
500: Off-Gas Utilization	$55,000,000
600: Steam System & Power Generation	$58,810,000
700: Cooling Water & Other Utilities	$9,130,000
800: Water Management	$16,090,000
Total Installed Equipment Cost (TIC)	$311,230,000
Land (115 Acres at $14000 per Acre)	$1,800,000
Site Development	$17,220,000
(% of ISBL)	10.0%
Indirect Costs & Project Contingency	$205,850,000
(% of TIC)	66.1%
Fixed Capital Investment (FCI)	$548,940,000
Working Capital	$27,450,000
Total Capital Investment (TCI)	$578,240,000
Loan Interest Rate	8.0%
Loan Term (Years)	10
Total Installed Equipment Cost per Annual GGE	$7.47
Fixed Capital Investment per Annual GGE	$13.17
Plant Operating Hours per Year	7884
On-Stream Percentage	90.0%

Specific Operating Conditions		
Feed Rate	Dry Tonnes / Day	2,000
	Dry Tons / Day	2,205
Feed Moisture and Dry Basis Ash, wt%		10% moist, 1.74% ash
Feedstock Cost	$/Dry Ton	$72.53
	$/Moisture+Ash Free Ton	$73.81

Version: FixedBed-CFP-1-R009-XL027-(2020$_Feed$72.53_OilTransport800)-SA

Operating Costs (¢ / GGE Product)	
Feedstock	126.0
Natural Gas	0.0
Catalysts	9.5
Sand	0.7
Other Raw Materials	4.3
Waste Disposal	3.2
Chemical Coproducts Credit	(58.2)
Electricity (& Energy Utility) Coproduct Credits	(20.2)
CFP Oil Transportation	24.2
Fixed Costs	65.1
Capital Depreciation	65.9
Average Income Tax	13.8
Average Return on Investment	101.0

Operating Costs ($ / Year)	
Feedstock	$52,520,000
Natural Gas	$0
Catalysts	$3,960,000
Sand	$300,000
Other Raw Materials	$1,810,000
Waste Disposal	$1,350,000
Chemical Coproducts Credit	-$24,240,000
Electricity (& Energy Utility) Coproduct Credits	-$8,410,000
CFP Oil Transportation	$10,090,000
Fixed Costs	$27,130,000
Capital Depreciation	$27,450,000
Average Income Tax	$5,740,000
Average Return on Investment	$42,100,000
Total Plant Electricity Usage (kW)	32,715
Electricity Produced on Site (kW)	50,905
Electricity Purchased from Grid (kW)	0
Electricity Sold to Grid (kW)	18,190
Plant Electricity Use (kWh /GGE)	6.19
Overall Plant Efficiency to CFP Oil - HHV %	41.5%
Overall Plant Efficiency to CFP Oil - LHV %	42.2%

Sensitivity analysis for CFP had been presented in previous reports [18,19] and the general trends still apply to these cases. The interested reader can review the previous publications for CFP TEA sensitivities. The impact of changing capital investment assumptions on the MFISP are shown in Appendix M (Comment 32).

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2.3.13 Sustainability and Efficiency Metrics

We present the life cycle inventory along with R&D GREET 2023 [27] based emission factors for GHG calculations. The values presented are for a 2,000 dry metric tonnes per day plant for CFP-A (Table 26) and CFP-B (Table 27). CFP-A produces more electricity at the expense of a lower CFP oil yield. From a high-level U.S. resource standpoint, the desirable outcome is to produce more liquid fuels via CFP-B from scarce biomass resources instead of producing electricity, even though current emission factors indicate that CFP-A is more sustainable from a GHG emissions standpoint. It is expected that the emission factor for displacing the electricity will continue to go down in the future as solar and wind deployments increase, and the GHG emissions accounting difference between CFP-A and CFP-B may progressively narrow down as overall carbon intensity for U.S. grid electricity declines.

Table 26. Life Cycle Inventory, GHG Emissions, and Efficiency Metrics for CFP-A

	Flow Rate/ Value	GHG Emission Factor	Emission Factor Unit	GHG Emission kg CO_2e/h
Biorefinery Resource Consumption				
Blended woody biomass				
Mass flow (wet), lb/h	204,131	0.00000	kg CO_2e/lb	-
Mass flow (dry), lb/h	183,718	0.10267	kg CO_2e/lb	18,863
Rail transport of biomass for colocated biorefinery, miles	0	1.433E-05	kg CO_2e/lb/mile	-
Sand makeup, lb/h	126	0.01030	kg CO_2e/lb	1
Natural gas as process input (not combust), MMBtu/h	0.0	11.64288	kg CO_2e/MMBtu	0
Zeolite catalyst, lb/h	229	5.47425	kg CO_2e/lb	1,252
50 wt % caustic, lb/h	1,565	0.46544	kg CO_2e/lb	728
Net water makeup, lb/h	93,097	0.00000	kg CO_2e/lb	-
Boiler feed water chemicals, lb/h	2	3.14443	kg CO_2e/lb	6
Cooling tower chemicals, lb/h	1	3.14443	kg CO_2e/lb	3
Purchased hydrogen (pure H_2 basis), lb/h	0	4.28425	kg CO_2e/lb	-
No. 2 diesel fuel (includes combustion), lb/h	71	1.73309	kg CO_2e/lb	123
Air Emissions From Process Model				
CO_2 (fossil), lb/h	0	0.45359	kg CO_2e/lb	0
CO_2 (biogenic), lb/h	236,382	0.00000	kg CO_2e/lb	-
Byproducts for Displacement Credit				
Exported hydrogen (pure H_2 basis), lb/h	0	−4.28425	kg CO_2e/lb	-
Exported biogenic fuel gas, MMBtu/h	0	−72.79877	kg CO_2e/MMBtu	-
Exported steam				
Energy (cond. to liquid at same pres.), MMBtu/h	0	−91.29024	kg CO_2e/MMBtu	-
Mass flow, lb/h	0			
Excess electricity (positive is export), kW	41,049	−0.43952	kg CO_2e/kwh	(18,042)

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	Flow Rate/ Value	GHG Emission Factor	Emission Factor Unit	GHG Emission kg CO_2e/h
MEK, lb/h	0	−0.47696	kg CO_2e/lb	-
Acetone, lb/h	0	−1.15579	kg CO_2e/lb	-
Products				
CFP oil				
LHV, MMBtu/h	464			
LHV, Btu/lb	12,408			
LHV, MJ/h	489,806			
LHV, GGE/h	3,996			
Mass flow, lb/h	37,387			
Density, lb/gal	9.602			
Volumetric flow, gal/h	3,894			
Volumetric flow, barrels/day	2,225			
Energy density, GGE/gal	1.026			
Bio-oil transportation, miles	800	1.433E-05	kg CO_2e/lb/mile	429
GHG Emissions (in different units)				
Total emissions without transport of CFP oil, kg CO_2e/h				2,934
Total emissions with 800 miles rail transport, kg CO_2e/h				3,363
GHG emissions without transport of CFP oil, g CO_2e/MJ				5.9910
GHG emissions with 800 miles rail transport, g CO_2e/MJ				6.8663
GHG emissions without transport of CFP oil, kg CO_2e/GGE				0.7344
GHG emissions with 800 miles rail transport, kg CO_2e/GGE				0.8416
GHG emissions without transport of CFP oil, kg CO_2e/lb				0.0785
GHG emissions with 800 miles rail transport, kg CO_2e/lb				0.0900
GHG emissions without transport of CFP oil, kg CO_2e/gal				0.7537
GHG emissions with 800 miles rail transport, kg CO_2e/gal				0.8638
Efficiency				
CFP oil from biomass (LHV basis)	32.9%			
Electricity from biomass (LHV of biomass basis)	9.9%			
CFP oil from biomass (carbon basis)	28.8%			
CFP oil from dry biomass (mass basis)	20.4%			

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Table 27. Life Cycle Inventory, GHG Emissions, Efficiency Metrics for CFP-B

	Flow Rate/ Value	GHG Emission Factor	Emission Factor Unit	GHG Emission kg CO_2e/h
Biorefinery Resource Consumption				
Blended woody biomass				
Mass flow (wet), lb/h	204,131	0.0000	kg CO_2e/lb	-
Mass flow (dry), lb/h	183,718	0.1027	kg CO_2e/lb	18,863
Rail transport of biomass for colocated biorefinery, miles	0	1.433E-05	kg CO_2e/lb/mile	-
Sand makeup, lb/h	159	0.0103	kg CO_2e/lb	2
Natural gas as process input (not combust), MMBtu/h	0	11.6429	kg CO_2e/MMBtu	0
Zeolite catalyst, lb/h	0	5.4743	kg CO_2e/lb	-
Fixed-bed vapor-phase upgrading catalyst (0.5% Pt/TiO_2), lb/h	14	57.3522	kg CO_2e/lb	802
Zinc oxide (ZnO) (reforming cleanup), lb/h	2	2.8244	kg CO_2e/lb	6
HDS (reforming cleanup), lb/h	1	3.1444	kg CO_2e/lb	3
Steam reforming catalyst, lb/h	5	5.9337	kg CO_2e/lb	29
Shift catalyst, lb/h	7	3.1444	kg CO_2e/lb	21
PSA adsorbent, lb/h	167	−1.0047	kg CO_2e/lb	(167)
Solvent 1 for coproduct recovery, lb/h	111	0.5508	kg CO_2e/lb	61
Solvent 2 for coproduct recovery, lb/h	3	2.3875	kg CO_2e/lb	7
50 wt % caustic, lb/h	231	0.4654	kg CO_2e/lb	107
Net water makeup, lb/h	52,226	0.0000	kg CO_2e/lb	-
Boiler feed water chemicals, lb/h	2	3.1444	kg CO_2e/lb	6
Cooling tower chemicals, lb/h	1	3.1444	kg CO_2e/lb	3
Purchased hydrogen (pure H_2 basis), lb/h	0	4.2842	kg CO_2e/lb	-
No. 2 diesel fuel (includes combustion), lb/h	71	1.7331	kg CO_2e/lb	123
Air Emissions From Process Model				
CO_2 (fossil), lb/h	0	0.4536	kg CO_2e/lb	0
CO_2 (biogenic), lb/h	204,023	0.0000	kg CO_2e/lb	-
Byproducts for Displacement Credit				
Exported hydrogen (pure H_2 basis), lb/h	3	−4.2842	kg CO_2e/lb	(11)
Exported biogenic fuel gas, MMBtu/h	0	−72.799	kg CO_2e/MMBtu	-
Exported steam				
Energy (cond. to liquid at same pres.), MMBtu/h	0	−91.290	kg CO_2e/MMBtu	-
Mass flow, lb/h	0			
Excess electricity (positive is export), kW	18,190	−0.4395	kg CO_2e/kwh	(7,995)
MEK, lb/h	926	−0.4770	kg CO_2e/lb	(442)
Acetone, lb/h	5,336	−1.1558	kg CO_2e/lb	(6,167)

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	Flow Rate/ Value	GHG Emission Factor	Emission Factor Unit	GHG Emission kg CO_2e/h
Products				
CFP oil				
LHV, MMBtu/h	614			
LHV, Btu/lb	13,908			
LHV, MJ/h	648,060			
LHV, GGE/h	5,287			
Mass flow, lb/h	44,130			
Density, lb/gal	8.788			
Volumetric flow, gal/h	5,022			
Volumetric flow, barrels/day	2,870			
Energy density, GGE/gal	1.053			
Bio-oil transportation, miles	800	1.433E-05	kg CO_2e/lb/mile	506
GHG Emissions (in different units)				
Total emissions without transport of CFP oil, kg CO_2e/h				5,251
Total emissions with 800 miles rail transport, kg CO_2e/h				5,757
GHG emissions without transport of CFP oil, g CO_2e/MJ				8.1020
GHG emissions with 800 miles rail transport, g CO_2e/MJ				8.8828
GHG emissions without transport of CFP oil, kg CO_2e/GGE				0.9931
GHG emissions with 800 miles rail transport, kg CO_2e/GGE				1.0888
GHG emissions without transport of CFP oil, kg CO_2e/lb				0.1190
GHG emissions with 800 miles rail transport, kg CO_2e/lb				0.1304
GHG emissions without transport of CFP oil, kg CO_2e/gal				1.0456
GHG emissions with 800 miles rail transport, kg CO_2e/gal				1.1463
Efficiency				
CFP oil from biomass (LHV basis)	42.2%			
Electricity from biomass (LHV of biomass basis)	4.3%			
CFP oil from biomass (carbon basis)	35.7%			
Chemical coproducts from biomass (carbon basis)	4.2%			
CFP oil from dry biomass (mass basis)	24.0%			
Chemical coproducts from dry biomass (mass basis)	3.4%			

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Section 3

CFP Oil Hydroprocessing: Techno-Economics and Supply Chain Sustainability Metrics for Hydrocarbon Fuel Product

Modeling and Results

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3 Hydroprocessing

We use the term hydroprocessing to represent the broad family of refining processes involving hydrotreating and hydrocracking. Hydroprocessing has traditionally been applied in refining operations with petroleum feedstocks to improve product quality of refinery blendstocks by (1) removing impurities such as sulfur, nitrogen, oxygen, and metals, (2) saturating olefinic and aromatic compounds, (3) isomerizing long carbon chains, and (4) cracking to optimize product splits [17]. Hydroprocessing oxygenated liquid intermediates from biomass and waste organics has been a topic of recent interest for fuels and chemicals production. Publications on this topic continue to grow with the availability of experimental results for specific feedstocks. A 2013 review article on this topic discussed a broad set of feedstocks [58]; additional references relevant to this report are included in the following discussions. The focus of this report is on process metrics when CFP oils are hydroprocessed either in a stand-alone configuration or coprocessed with other feedstocks like straight-run diesel (SRD), light cycle oil (LCO), and vacuum gas oil (VGO). Catalyst development to increase effectiveness is not a focus area of this report; we assume the use of commercially relevant catalysts for petroleum feedstocks, since limited experimental data so far has shown that those catalysts also work for upgrading CFP oils [13,38,59,60]. Research on catalyst activity and maintenance is necessary for long-term success, but we are not at a point where predictive reaction behavior can be incorporated into process models. Design and operational impacts are assessed in this study by including a consistent set of assumptions for reactor behavior; impacts of changing those assumptions are explored via sensitivity analysis. This report includes a premise of co-utilization or reutilization of petroleum technology and infrastructure for processing CFP oils; this is based on short-duration bench-scale experiments with CFP oils [13,59,61]; further R&D is necessary to assess, validate or invalidate, and provide solutions to enable the use of current infrastructure. Along with the differences in reaction behavior of CFP oil constituents, hydrogen consumption and related reactor heat management stand out as the most significant process impacts because of high hydrogen requirement for hydroprocessing CFP oils to hydrocarbon fuels. In this context, it should be noted that the range of oxygen contents in the CFP oils in this report (16 to 22 wt %) are significantly higher than typical proportions of N and S heteroatoms present in petroleum hydroprocessing feedstocks; in addition, oxygen removal from difficult-to-hydrodeoxygenate phenolic bonds in CFP oils, usually needs to be facilitated by the saturation of aromatic rings before the removal of that oxygen [62]. As a concurrent benefit, the saturation of aromatic rings result in cycloalkanes that are desirable for jet fuel [63] and also improve the fuel's cetane number [64]. Temperature control using hydrogen and/or liquid recycle injection into the reactor are explored. Additional anticipated process impacts are also discussed in this report. These hydroprocessing challenges exist for other oxygenated feedstocks and the learnings from this report may be extended to other liquid feedstocks.

Our goal is to present cases and results that can be adapted to understand key process considerations associated with various potential implementations, which may include:

- Stand-alone hydroprocessing of CFP oils
- Repurposing refinery units for processing oxygenated feedstocks
- Coprocessing CFP oils with petroleum feedstocks in existing refinery units with minimal necessary modifications within existing constraints

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- Augmenting refinery units to alleviate constraints for processing large volumes of oxygenated liquid intermediates like CFP oils

Given the breadth, depth, and diversity in implementation, we attempt to illustrate these aspects through variations within a set of a few base configurations discussed in Sections 3.2.2 and 3.2.4. Consistent assumptions are maintained to understand potential impacts. Those assumptions may differ for specific implementations, but the directional trends reported will likely be maintained for a broad set of conditions and configurations. In some cases, these assumptions lead to modeled operational parameters that are beyond the feasibility of current implementations; those results are useful for informing the adoption of some generic alternate strategies and may help inform the most effective choices unique to each installation. When trying to understand the potential use of any existing hydroprocessing facility for processing or coprocessing CFP oils, the parameter space for those studies will be more specific and evident and may include some of the aspects discussed later in this report.

3.1 Experimental Data for CFP Oil Hydroprocessing

CFP oils have been hydrotreated in continuous trickle-bed reactors for both stand-alone hydrotreating and co-hydroprocessing scenarios. Data for stand-alone hydrotreating is compiled in Appendix A, Table A-1. A wide variety of CFP oils, produced over HZSM-5 [13,65], Pt/TiO_2 [60,65], alumina [66,67], and red mud [68,69,70] catalysts and with oxygen contents of 16–29 wt % on a dry basis (db) have been processed. The longest published times on stream exceed 300 h, and these comprise processing CFP oils with high oxygen contents of 25-28 wt % oxygen (db) [67,68]. NREL has since completed CFP oil hydrotreating for 588 h with CFP oil (19–23 wt % O db) without plugging or fouling issues reported. A single-stage hydrotreating possible with CFP oils contrasts with stand-alone hydroprocessing of non-catalytic fast pyrolysis liquids where a separate reactor with catalysts such as Ru/C are used to first stabilize the liquid before further hydroprocessing [9].

Sulfided NiMo has been the catalyst of choice for most of the studies though sulfided CoMo [68] gave stable operation for over 300 h for a product with 1 wt % oxygen. Sulfur contents of CFP oils from woody biomass are typically lower than necessary to keep the catalyst sulfided, and H_2S or sulfiding agents need to be added during stand-alone processing; coprocessing with high sulfur petroleum feed may also provide the necessary sulfur. A co-hydrotreating study showed better hydrogenation with NiMo than with CoMo, leading to a higher H:C ratio in the product and a higher cetane number in the diesel-range cut [60]. Hydrogenation of aromatic rings is also an advantage for jet fuel since cycloalkanes have more favorable properties than aromatics [65].

Stand-alone hydrotreating temperatures approaching 400°C are required for good deoxygenation; operations at 300°C or below lead to the product oxygen content increasing significantly during the run (e.g., from 1 to 11 wt %) [66,67], whereas the product oxygen contents remained stable at 385°C–400°C in most of the experiments. Phenols are prevalent compounds in CFP oils and their deoxygenation is well known to require high temperatures [71]. Hydrogenation of the aromatic rings in phenols at lower temperatures produces cyclohexanols, and those alcohols can be hydrodeoxygenated at lower temperatures than phenols. However, at lower temperatures, the hydrogenation activity of the catalysts decreases with time on stream and therefore, the fraction of phenols in the product increases, leading to the increase in the product oxygen content measured over time at low temperatures.

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Experiments at 385°C–400°C have yielded product oxygen contents of <0.01 to approximately 1 wt %, with a tendency of CFP oils with oxygen contents above 20 wt % yielding products with ~1 wt % oxygen and those with <20 wt % O (db) giving lower values at similar conditions (Figure 7a). The reported carbon efficiencies to liquid hydrocarbons ranged from 83 to 93%, and in general decreased as the CFP oil oxygen content increased (Figure 7a). The experimental pressures have been at least 1,450 psi (100 bar), with a typical range of 1,800–2,000 psi (124–138 bar), and the hourly space velocities (liquid or weight hourly space velocities documented in Appendix A) have ranged from 0.1 to 0.35 h^{-1}. Some experiments at NREL have included a non-isothermal section prior to the final hydrotreating section, which resulted in longer problem-free operations (tested during a ~600 h experiment without any disruptions or significant decline in catalyst activity), very low oxygen contents (down to <0.01 wt %), enhanced hydrogenation, and yielding cycloalkanes as the main product group [13,60,65].

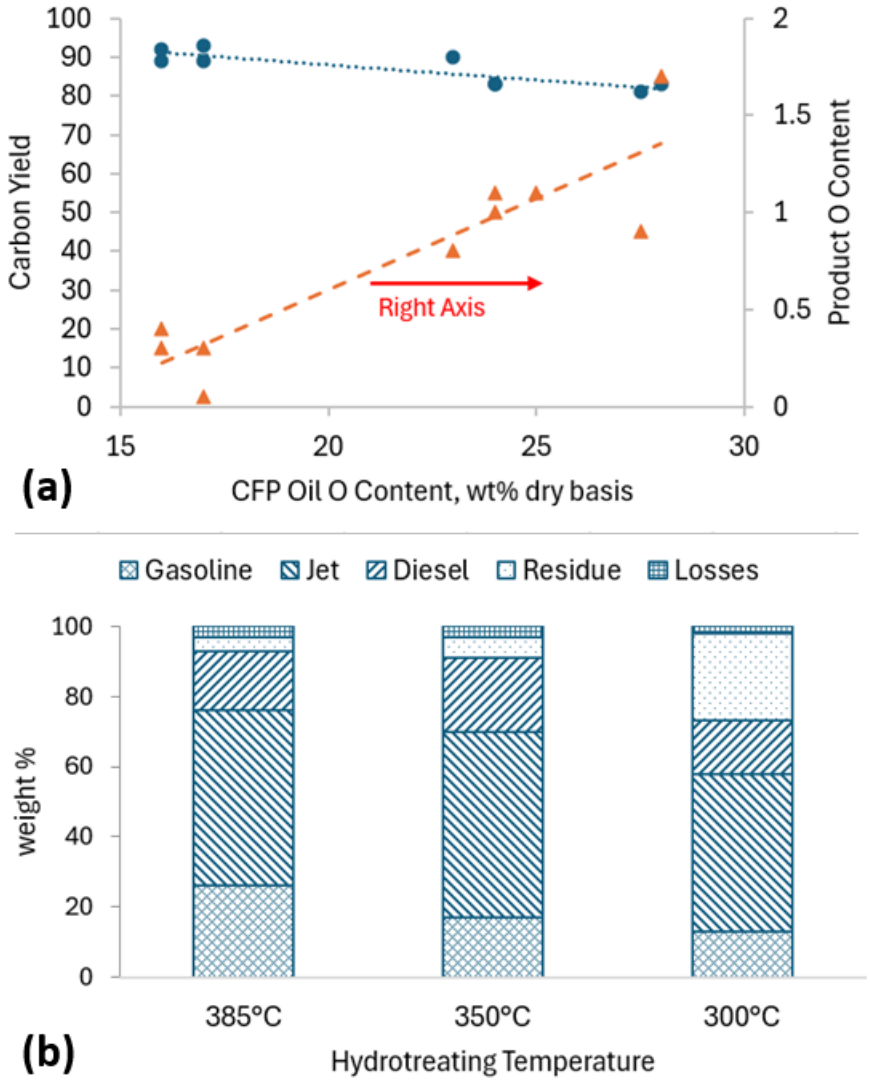

Figure 7. Stand-alone hydrotreating of CFP oils: (a) Carbon efficiencies and product oxygen contents for stand-alone hydrotreating of CFP oils. Data from [13,60,65,68,69,70,72]. (b) Impact of hydrotreating temperature on fuel fraction distribution. Data from [13].

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The jet-range fraction of the hydrotreated product has ranged from 39 to 53 wt % by distillation, with gasoline-range product constituting most of the rest (26–36 wt %) [13,65]. Increasing the hydrotreating temperature enhanced cracking and produced more gasoline and jet fractions, and less diesel and residue fractions (Figure 7b) [13]. When only gasoline and diesel were the intended targets, the gasoline fraction typically encompassed 40–50 wt % and diesel 39–45 wt % of the hydrotreated product [60,69,70,72].

The gasoline-range product research octane numbers (RON) have been estimated at 59–83 based on detailed hydrocarbon analysis by gas chromatography–mass spectrometry [68,70,72]. The highest values have been reported for HZSM-5-based CFP oils, which produce aromatic compounds. Diesel from isothermal hydrotreating had low cetane numbers (derived cetane number or DCN) of 13–26 [68,70,72], whereas including the initial non-isothermal section for hydrogenation was shown to increase the cetane number (indicated cetane number or ICN) to 45 [60]. While hydrogenation of aromatic compounds enhances diesel properties, it decreases the gasoline fraction octane number. The gasoline-range product may also be utilized for producing aromatics (e.g., benzene, toluene, xylene) or as solvents [65].

The distilled jet fractions from the experiments with a non-isothermal initial zone and final isothermal zone at 385°C–400°C have been measured for selected aviation fuel properties (ASTM D7566 finished fuel blend properties and D4054 Tier 1 properties) [13,65]. The jet fraction met ASTM guidelines for simulated distillation, flash point, freezing point, density, and viscosity. The heats of combustion were slightly lower than the guidelines for some samples (41.8–42.5 vs. ≥42.8 MJ/kg), as was the ICN. However, this does not preclude products with these properties from being approved as blendstocks. These properties can also be improved by further hydrogenation by adjusting reactor operations. Cycloalkanes were the main compound group (up to 91%) and further increasing their fraction enhances the heat of combustion and cetane number.

Data from continuous co-hydroprocessing experiments are summarized in Appendix A, Table A-2. These include co-hydrotreating with SRD [38,60, previously unpublished data from NREL] and co-hydrotreating/co-hydrocracking with VGO [59] or a mixture of VGO and LCO [previously unpublished data from NREL]. For co-hydrotreating with SRD, blend levels up to 30% have been tested under conditions resembling those in commercial operations: isothermal at 325°C–340°C (617°F–644°F), 800–1,200 psi (55–83 bar), 1 h^{-1}. Notable is the effective deoxygenation with oxygen contents typically below detection limits that has been achieved under these conditions that are significantly milder (e.g., temperature) than those required for stand-alone hydrotreating. This could be due to synergetic effects with the more hydrogen-rich petroleum stream, or more dispersed reactive oxygenated molecules and lower oxygen flow per catalyst active site.

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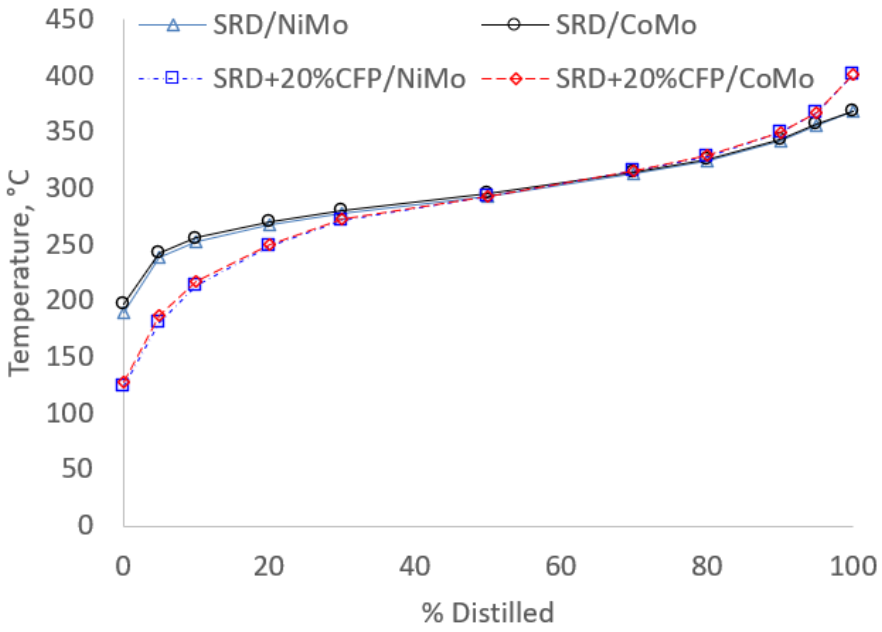

Figure 8. Impact of co-hydrotreating of CFP oil with SRD on simulated distillation. Data from [60].

Carbon-14 measurements suggest approximately 95% incorporation of carbon from the CFP oil into the final product [38,60]. The lower hydrotreating temperature may contribute to the high carbon efficiency, which exceeds most values reported for stand-alone hydrotreating. Close to 50% of the biogenic carbon has been reported in the gasoline range and 44% in the diesel range [38]. This is similar to the distribution of products for stand-alone hydrotreating and suggests that co-hydrotreating does not impact the cracking of CFP oil. Another study confirmed increases in the gasoline range and also in the heavy fractions (Figure 8) [60]. The ICN values of the diesel product have been in the range of 42–50; these represented decreases from the values of SRD-only hydrotreating [60].

Unpublished data from NREL includes hydrotreating CFP oil with SRD at blend levels of 10%–30% for the production of jet-range fuel; 15%–17% of the products boiled in the jet range, and the majority of the product was in the heavy diesel range boiling above the jet fraction. Compared to SRD hydrotreating, there was no statistically significant increase in the jet fraction. The jet fraction density increased, viscosity decreased, and heat of combustion decreased as the blend level increased, as would be expected based on stand-alone hydrotreating results.

Santosa et al. [59] co-hydroprocessed CFP oils with low oxygen contents of 13%–14.5% (db) over sulfided NiMo catalyst at 400°C at blend levels of 6%–20%. The products had oxygen contents below the detection limit (<0.5 wt %), and the blending did not reduce the desulfurization or cracking of the VGO. The jet-range product constituted 23%–27%, and residue was the largest fraction at 65%–73%. The fraction of gasoline and jet increased as the blend level increased. It should be noted that these experiments were performed over a typical hydrotreating catalyst, and this resulted in limited VGO cracking. The biogenic carbon incorporation was not explicitly given in the paper, but significant fractions of biogenic carbon were measured in the product.

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Previously unpublished data from NREL with 20% CFP oil and the remaining 80% of petroleum (50/50 mixture of VGO/LCO) feed includes co-hydrotreating over sulfided NiMo, combined with co-hydrocracking over NiW/zeolite. The jet fraction was significantly higher (41%) and residue lower (25%), compared to the study by Santosa et al. [59], which utilized only a hydrotreating catalyst. While the measured jet fuel properties were good, the experiment was of short duration (118 h total, of which 69 h with VGO and 50 h for coprocessing).

Co-hydrotreating and co-hydrocracking of a different biointermediate (hydrothermal liquefaction oil with 1.4 wt % O and 3.2 wt % S) at a blend level of 7.5% showed that higher temperatures (by 8°C) were required during hydrocracking to obtain the same conversion as with VGO only [73], and the same may be expected for CFP oil as well. However, only minor changes in the quality, hydrocarbon type, or hydrogen consumption were reported.

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3.1.1 Properties of Jet Cuts From Hydroprocessed CFP Oil Compared to ASTM Specifications

Below are the measured properties of jet cuts from some NREL experiments, with additional data available in the literature including in Griffin et al. [13], supporting the premise that hydroprocessed CFP oils can meet stringent jet fuel specifications. Distillation strategies and choice of cut points will determine the properties of the remaining cuts, and those may be blended into the gasoline and diesel pools to the extent dictated by specifics of distillation choices.

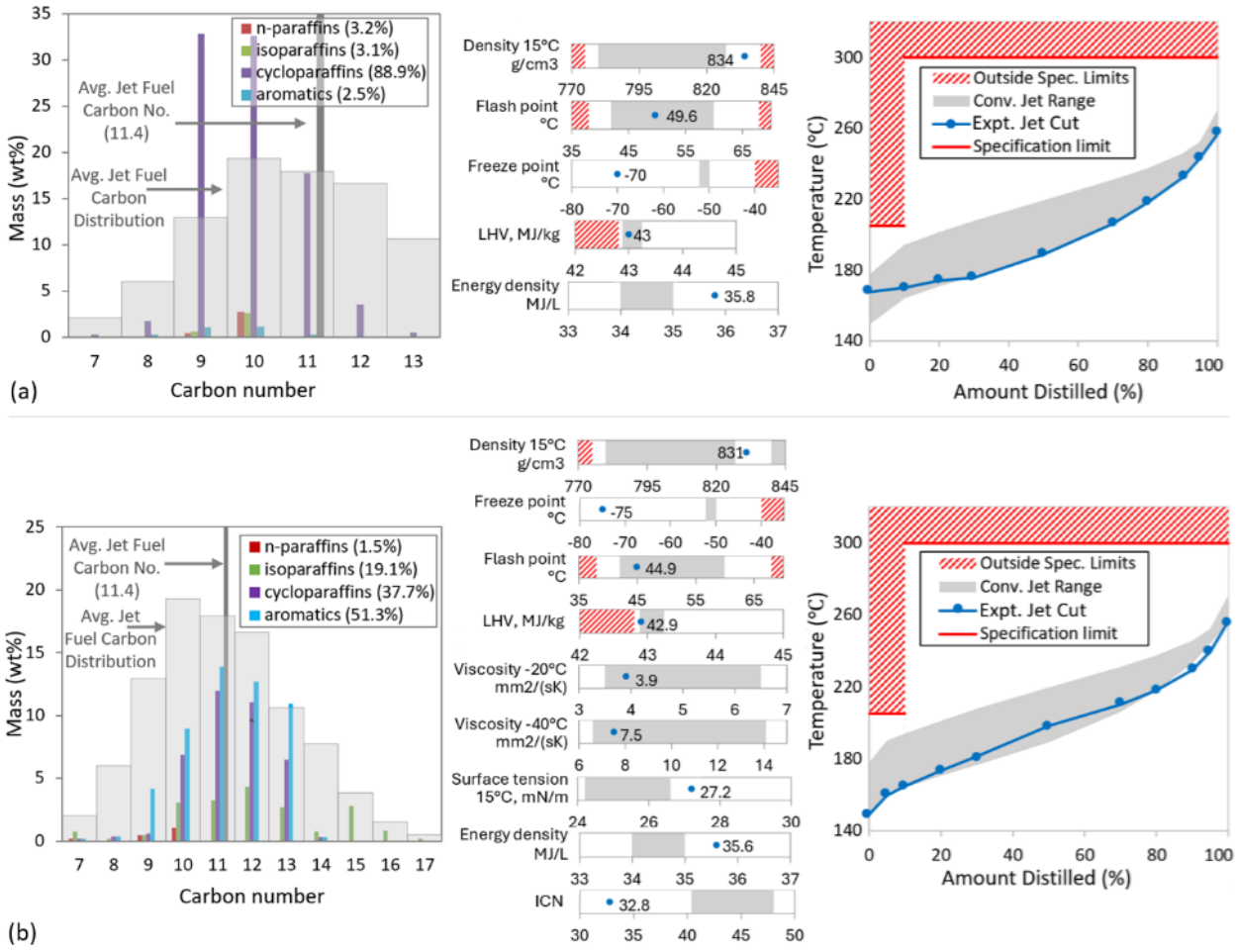

Figure 9. (a) Data for jet cut from hydrotreated HZSM-5 CFP oil from Chen et al. [65] and Watanasiri et al. [74]. Carbon numbers are based on two-dimensional gas chromatography. Distillation data from Simdist D86 correlation. (b) Data for jet cut from hydrotreated and hydrocracked fuel product with 20% HZSM-5 CFP oil co-hydroprocessed with 80% petroleum feed (50% LCO + 50% VGO). Carbon numbers are based on gas chromatography–vacuum ultraviolet spectroscopy. Distillation data from Simdist D86 correlation. The format of the plots was adapted from literature; the grey regions represent the range of conventional jet fuels, i.e., POSFs 10264, 10325, 10289; the red regions represent outside specification limits [75,76]. POSFs are numbers assigned by the U.S. Air Force Research Laboratory to identify specific batches of fuel.

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3.2 Hydroprocessing Conceptual Models

Hydroprocessing designs can be diverse and implemented differently at each refinery or for stand-alone operations. Configuration choices and process parameters are dictated primarily by the types of feedstock and products intended for the unit, as well as the design choices of the technology provider. A summary of the ranges of key operating variables in petroleum refinery hydroprocessing operations, hydrotreating (HT) and hydrocracking (HC) are presented in Table 28. Our choices of modeled operating parameters were dictated by bench-scale experiments, and we later discuss modeled results in the context of the ranges shown in Table 28. The base hydroprocessing models in Aspen Plus were designed for 75,000 barrels per day (BPD); our cost spreadsheet is designed to adapt for evaluating other throughputs and we present results for other scales.

Table 28. Ranges for Key Operating Parameters in Petroleum Refinery Hydroprocessing Units [77]

Process Parameter	Kerosene /Jet HT	Virgin Distillate HT	Cracked Distillate HT	Distillate HC	Gas Oil or FCC Feed HT	Cracked Gas Oil HC	Resid. HC
Feed gravity (API)	40–50	30–45	15–30	15–30	15–25	0–20	(10)–15
Temp. (°C) Temp. (°F)	290–370 554–698	310–400 590–752	340–400 644–752	340–450 644–842	340–430 644–806	340–450 644–842	400–450 752–842
Pressure (psia) Pressure (bara)	700–1,000 48.3–68.9	700–1,500 48.3–103.4	1,000–2,500 68.9–172.4	1,000–2,500 68.9–172.4	800–1,200 55.2–82.7	1,500–3,000 103.4–206.8	2,400–3,000 165.5–206.8
H_2 partial pres. (psia) H_2 partial pres. (bara)	500–900 34.5–62.1	500–1,200 34.5–82.7	800–1,800 55.2–124.1	800–2,300 55.2–158.6	600–1,000 41.4–68.9	1,200–2,700 82.7–186.2	1,900–2,700 131.0–186.2
Liquid hourly space velocity (h^{-1})	3–6	1–4	1–2	0.5–2	0.5–1	0.5–2	0.5–1
H_2 consumption (scf/bbl)	150–500	300–700	700–1,700	500–2,000	500–1,000	1,200–3,500	1,200–3,500
H_2 to oil ratio (scf/bbl)	600–2,500	1,200–3,000	2,000–7,000	2,000–8,000	1,500–4,000	4,000–10,000	4,000–10,000

3.2.1 Petroleum Refining Hydroprocessing Configurations

Hydroprocessing configurations vary, and several options are presented in the literature [17,78]. Most hydroprocessing units for gasoline, jet fuel, and diesel fuel production utilize fixed trickle-bed reactors in which feedstocks and hydrogen are fed to the top of downflow reactors and through packed catalyst beds [78,79]. The two primary process configurations for fixed-bed hydroprocessing units are single-stage and two-stage; two-stage configurations typically allow intermediate adjustments before the second stage, thus enabling targeted processing of specific portions of product from the first stage for value enhancement. Figure 10 presents process flow diagrams for the single and two-stage hydroprocessing unit configurations.

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Single-stage units (Figure 10a) apply one or multiple reactors in series within a single hydrogen circulation loop. These units generally have lower capital costs, simpler operations, and smaller footprint requirements compared to the more complex two-stage designs since there are fewer reactors and associated process control loops. Thus, single-stage units may generally have less operational flexibility for deep desulfurization/denitrification in the conversion of challenging and diverse feedstocks. However, the effectiveness of any configuration depends on the situational ability to meet key objectives for converting specific feedstocks and the ability to allow operational adaptations with changes in feedstock properties.

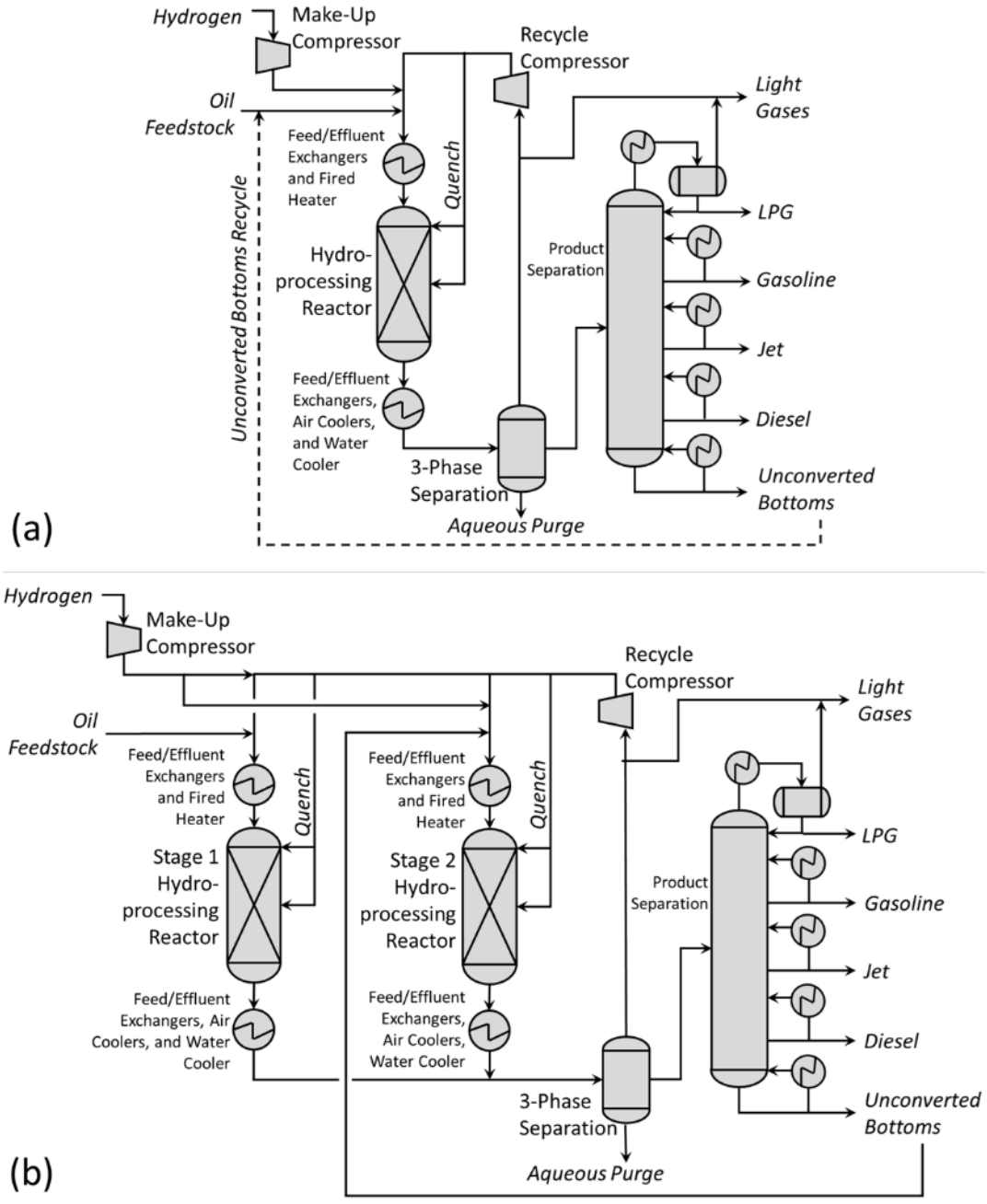

Figure 10. Simplified configurations for (a) single-stage and (b) two-stage hydroprocessing units

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Two-stage units (Figure 10b) are designed with separate reactor systems with hydrogen makeup to each stage and interstage separation to remove impurities before the second stage, which often includes aromatics saturation, isomerization and/or hydrocracking catalysts for property and quality improvement. The advantages of two-stage units include higher levels of operational flexibility and feedstock variability since there are two reactor systems available to reach quality targets. However, these units are more capital intensive, and complex for operation and process control.

A refiner's decision between a single-stage and a two-stage hydroprocessing unit depends on feedstock quality and variability, product specifications, investment budget, and refinery objectives. Single-stage units are cost-effective for moderate processing needs, while two-stage units are ideal for stringent fuel regulations and complex feedstocks. Flexibility may be beneficial when processing CFP oils, but at this point we do not have enough data to determine the most advantageous design decisions when processing or coprocessing CFP oils; meeting product quality and maximizing desired products will guide those decisions and justify additional investments for a two-stage system.

Our process models are based on single-stage configurations since most of the data available in the literature so far is from once-through experiments. We added the capability to recycle the total condensed liquid product (prior to fractionation) as a potential means for cooling the reactor. This is achieved through intermediate injection of the cold liquid or by recycling the liquid product into the front-end reactor feed to suppress reactivity of the total liquid fed into the reactor. This analysis can be extended to other configurations when additional data becomes available.

3.2.2 Generic Conceptual Configuration for Hydrotreating

We used a generic single-stage hydroprocessing configuration for the process model in this study. While there are variations in single-stage configurations, a general setup will usually include the units shown in Figure 11; this figure also reflects the setup of the Aspen Plus model. The same process configuration is carried through for all feeds included in this study. For hydrotreating (referred as HT in this report) we include SRD as the petroleum feedstock. The surrogate model compound slates for the two CFP oils from CFP-A and CFP-B processes (discussed in Section 2) are included in Appendix C. The SRD surrogate model compound slate is included in Appendix D. This hydroprocessing configuration was modified from the configuration presented in the 2015 design report [18]; petroleum feedstocks and coprocessing were not included in that report. The following description includes parameters common for all cases, with additional specifics for select cases included in separate tables in the following subsections. Note that many of these parameters are subsequently varied as part of sensitivity analysis to assess their impacts. Stream summaries from the process models are included in Appendix L for bookend cases with 100% CFP oil and 100% petroleum feed.

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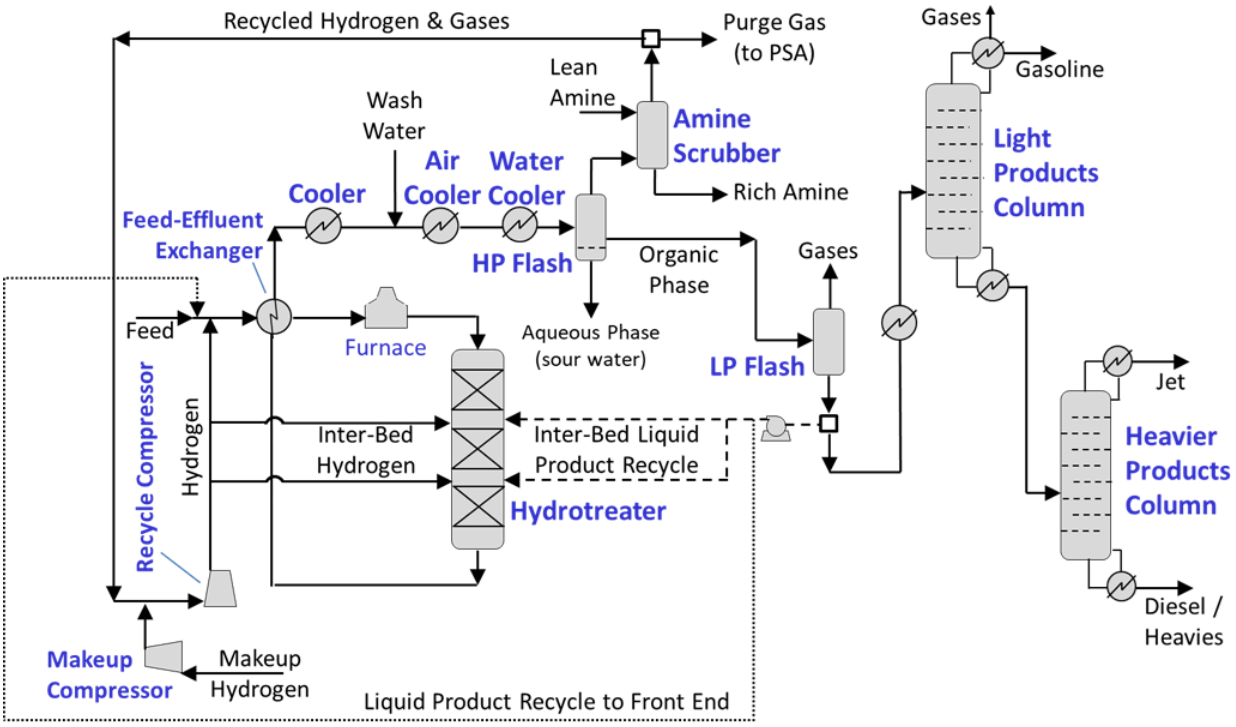

Figure 11. Conceptual design setup for hydrotreating (HT) configuration. Not all reactor cooling options shown may be exercised within the same setup.

Figure 11 shows the cold liquid feed stream going through a feed-effluent exchanger to preheat the feed while cooling the hot reactor effluent. The preheated feed, mixed with hydrogen and liquid recycle (if that option is chosen) is then further heated to the desired reactor inlet temperature in a furnace. The mix of feed and hydrogen then enters the hydrotreating reactor. The hydrotreating reactor includes inter-bed cooling via hydrogen and/or product liquid recycle quench in the various cases presented later; in the model, both these inter-bed streams are cooled to 43°C (110°F) before entering the reactor. The reactor specifications include an inlet pressure of 1,045 psia (72 bara), and an inlet temperature of 340°C (644°F). Three beds are included, with a 22°C (40°F) temperature rise allowed in each bed, by design. The reactor system outlet pressure is 1,015 psia (70 bara). Stoichiometric reactions with associated conversions are included in the model, with adjustments made for every case to achieve a consistent H:C ratio of ~1.8 in the total liquid (that includes all distillation cuts) fuel product and a 90% CFP oil hydroprocessing carbon efficiency. In the process model the reactor includes three beds with options for inter-bed injection of hydrogen and/or liquid recycle for cooling. The reactor products after initial cooling in the feed-effluent exchanger go through another cooler which is designed to generate steam in the process model. This is the source of steam coproducts at different pressures included in the economic and sustainability analysis. Four steam pressure options are included in the model: 15, 150, 350, and 900 psig (1, 10, 24, and 62 barg); the amounts of steam at each pressure are dictated by the temperature-enthalpy distribution during the cooling process. Wash water is then introduced, followed by an air cooler that cools the stream to 60°C (140°F) and water cooler that further cools the stream to 43.3°C (110°F). The cooled streams are separated in a high-pressure (HP) flash drum at 995 psia (68.6 bara), with organic liquid, aqueous sour water, and a gaseous stream recovered. The gases are sent to an amine scrubber for the removal of acid gases before being recycled to the reactor; a portion of

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the gas stream is then purged with the amount of purge determined by the target proportion of hydrogen at the reactor inlet (80 mol %). The main gas stream (non-purged portion) is sent to a recycle compressor with makeup hydrogen added to cover chemical hydrogen consumption and purge losses. In some cases, the hydrogen purity remains above the 80 mol % design specification, and a minimal amount of purge is included in those cases. The purge stream is sent to a pressure swing adsorption (PSA) unit to recover hydrogen for reuse in the process. The PSA specification is set to recover 85% of the hydrogen and the recovered hydrogen is accounted within the model boundary to determine the net hydrogen consumption. The purge gas from the PSA, along with gases recovered from the subsequent low-pressure (LP) flash unit and the gas stream from final product distillation, are claimed as fuel gas coproducts. For simplicity, no liquefied petroleum gas product/coproduct is included in the model. The organic liquid stream from the HP flash unit is sent to a LP flash unit at 55 psia where additional light species are recovered in the gas phase as a fuel gas product, as mentioned above. The liquid product is sent for distillation, with the option of recycling a portion of this organic liquid product for reactor heat management after pumping it to the higher reactor pressure.

3.2.3 Product Distillation

For simplicity of modeling, two distillation columns are included with three liquid fuel product cut ranges: gasoline, jet, and diesel/heavies. The jet portion is maximized in the model, based on the earlier discussion about potential future demand growth. Emphasis is not placed on the cuts because they can be altered based on fuel demand, provided they meet specifications; a total liquid fuel product with an H:C ratio of ~1.8 is produced and the goal of this analysis is to understand process implications of meeting this criterion using various petroleum and CFP oil feed options. Jet fuel properties are emphasized in Section 3.1 because of stringent requirements and projected future demand. The properties of other cuts, gasoline and diesel/heavies will be altered depending on the distillation strategy. Section 3.1 also provided references for gasoline and diesel cut properties measured after some CFP oil hydrotreating experiments where a jet cut was not included in the product slate. We have ongoing efforts to match product qualities and fuel properties with experimental results [74,80] as more data become available, but that was not a focus of the modeling work for this report. Thus, one should loosely interpret the proportions of modeled cuts presented in Appendix K and product stream flows in Appendix L because they do not reflect the experimental distillation properties reported in Section 3.1.

The liquid product after the LP flash unit is preheated via heat recovery from the distillation products, followed by a preheat furnace. The feed temperature for column 1 is set at 450°F, and it has 20 theoretical stages with feed entering above stage 6. The column overhead pressure is set at 35 psia. The feed temperature for column 2 is 561°F, with no additional heating when the bottoms product of column 1 is sent to column 2. Column 2 has 20 theoretical stages, with feed entering above stage 16. The gases recovered from the overhead of column 1 are claimed as a fuel gas coproduct. The gasoline cut is the top product of column 1, while jet and diesel/heavies respectively are top and bottom products of column 2. As mentioned earlier, the H:C of the total liquid product was ~1.8 for all cases in order to ensure similar product qualities across all cases. The product cuts in the model were based on boiling points and not tied to quality requirements for each cut. Design specifications in the RADFRAC blocks in Aspen Plus were used to get rough product cut distributions. Column 1 specifications included: (i) 5 mole % recovery of model compounds DECALIN + D-DECALN + L-NC10 + P-C10H20 + P-IC4CY6 in the bottoms product, (ii) 2 mole % recovery of model compounds DIMDODEC + D-DIMDOD, P-

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3C3CY6 + P-NC15 + P-C14H20 in the top product. Column 2 specifications included: (i) 0.05 mole % recovery of model compounds DECALIN + D-DECALN + L-NC10 + P-C10H20 + P-IC4CY6 in the bottoms product, (ii) 0.5 mole % recovery of model compounds MEFLOREN + D-NC18 + P-C17H26 + L-NC18 in the top product. Specified recoveries are based on feeds to each of the columns. Model compound names corresponding to the component IDs listed above are included in Appendix D, Table D-6.

3.2.4 Generic Configuration for Hydrotreating Followed by Hydrocracking

A similar single-stage configuration (Figure 12) was adopted for a setup with pre-hydrotreating followed by hydrocracking (referred as HT+HC in this report) for stand-alone and coprocessing of CFP oil (with petroleum feed comprised of 50 vol % LCO + 50 vol % VGO). This configuration was adopted from one of many options presented in the literature [78]. We used this setup for the process model after collecting data from preliminary experiments to show that this can be a potentially viable approach to produce fuels, including in the jet range [61]; these experiments are mentioned in the discussion in Section 3.1 and some of the results reported in Appendix A. The general description for this configuration is mostly identical to the hydrotreating configuration described above, except for higher pressures in the reactor section of the process and the inclusion of two reactors in series with different catalysts for pre-hydrotreating and subsequent hydrocracking. Commercial catalysts, sulfided $NiMoS_x/Al_2O_3$ for hydrotreating and NiW/zeolite for hydrocracking, were used in NREL experiments [61].

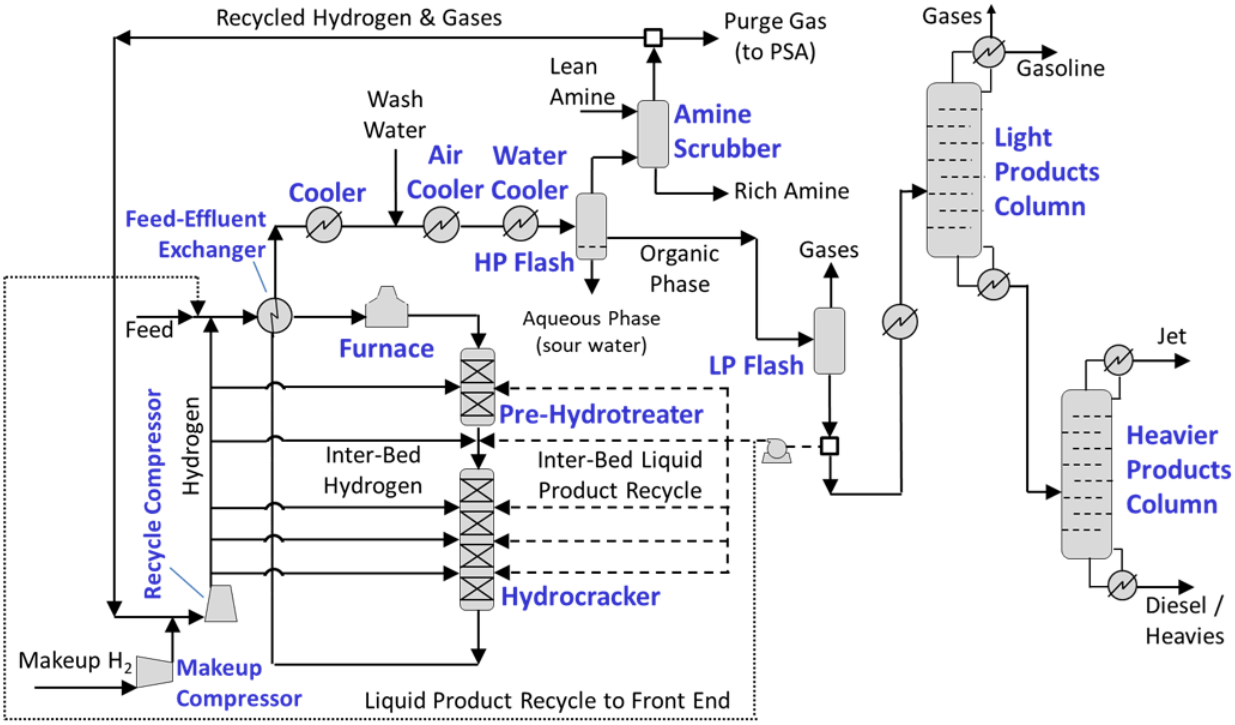

Figure 12. Hydroprocessing in a pretreat followed by hydrocracking (HT+HC) conceptual configuration (adapted from Hsu and Robinson [78]). Not all reactor cooling options shown may be exercised within the same setup.

The reactor inlet pressure was 2,045 psia (141 bara), with the HP flash unit at 1,995 psia (137.5 bara). The liquid product train pressure downstream of the HP flash unit is consistent with the

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hydrotreating configuration. The stoichiometric reactor model configuration and the product specification (H:C ratio of ~1.8 in the total liquid fuel products) were maintained, consistent with the hydrotreating model, with additional reactions added to capture hydrogenation and cracking of LCO/VGO. Surrogate model compounds for representing LCO/VGO are listed in Appendix D.

For the reactor system we duplicated the cost for the hydrotreating and hydrocracking reactors, i.e., in this configuration we used two separate, but identical, reactor costs as shown in Appendix B, Appendix H, and the HT+HC cases in subsequent Appendices I and K. Note that the reactor cost estimates are adjusted for higher pressures.

3.2.5 Quench Systems

Heat and catalyst temperature management are some of the biggest constraints in hydroprocessing systems and usually impact end-of-life and major maintenance decisions. With the introduction of a considerably higher proportion of heteroatoms (oxygen in CFP oils) than are present in most petroleum-derived feedstocks, careful design and operations decisions need to be made if CFP oil is introduced in existing hydroprocessing systems that were originally designed for petroleum feedstocks. Greater adjustments will be necessary for systems designed to handle relatively lighter and hydrogen-saturated (e.g., paraffinic) feedstocks like SRD that require less hydroprocessing, because there is lower hydrogen uptake for SRD compared to systems with heavier and less hydrogen-saturated feedstocks like LCO/VGO that consume significantly more hydrogen (Table 28). In other words, refinery systems that already have high hydrogen consumption may be more adaptable to the high-hydrogen-consumption scenarios associated with the hydroprocessing of CFP oils.

Quench system designs that include the injection of cold hydrogen between reactor beds are widely used and have been studied for some time [81]. This design is included in both Figure 11 and Figure 12. Another option is to recycle and inject a portion of the post-hydroprocessed liquid in between the reactor beds, depicted by dashed lines in Figure 11 and Figure 12 [82]. The post-hydroprocessed liquid can also be included with the front-end feed stream, thus reducing processing needs and associated heat of reaction for the overall feed when it includes post-processed recycle. We considered both options in this study, with the model designed for either (a) hydrogen or (b) liquid recycle as the manipulated variable to control reactor temperatures. When inter-bed injection of recycled liquid is already included within a reactor system design, there may be significant additional flexibility to adapt the control system to handle diverse feeds with different exothermic behavior. Inter-bed liquid recycle injection can provide additional heat removal compared to introducing with the front-end liquid feed because the cold recycle liquid can further absorb heat from the reactor products through sensible temperature gain and vaporization within the reactor. *However, any liquid recycle option will typically entail capacity curtailment for fresh feed.* In many petroleum refining configurations processing heavier feedstocks such as LCO/VGO, the liquid recycle stream is a heavier cut after distillation that requires further cracking; that is not the case in our simplified conceptual design, and we only included a split stream from the total liquid product prior to fractionation.

Hydrogen addition alone can also be an option for cooling when inter-bed liquid recycle cannot be considered. When processing oxygenated feedstocks like CFP oil, that configuration will typically also require either capacity curtailment via liquid product recycle to the front end or

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equipment debottlenecking to handle the requirement of higher gas flow rates. In general, the introduction of CFP oils into petroleum units will require higher hydrogen flows for both heat management via inter-bed cold H_2 injection and for satisfying the reaction stoichiometry, while maintaining appropriate hydrogen partial pressures.

Reactors may be designed to allow both hydrogen and liquid insertion between beds [83]. The process model includes that dual option but allows only one of the two flows for temperature control in any given configuration. When hydrogen is used for temperature control the liquid recycle fraction can be specified in the model. When liquid recycle is used for temperature control, the hydrogen concentration at the reactor outlet can be specified for modeling purposes; during actual operations, hydrogen supply will align based on process operations and reaction requirements, and the need for modifications will be dictated by the adequacy of reactor and process performances toward achieving production and catalyst maintenance goals.

3.2.6 Hydrogen Purity

This study includes two makeup hydrogen purities, 99 mol % and 85 mol %, to assess their impact on process parameters; the chosen purity level is reported with the results for the various cases (and included in the code for each case, discussed later). The lower purity makeup is common in refining operations, especially for relatively low hydrogen demand systems, such as SRD hydrotreating. Higher purity makeup gas can help reduce purge and may affect other parameters like recycled gas volume (depending on factors such as the temperature control strategy); this can become an important tool during process modifications as the hydrogen demand increases with changes in feedstock properties. A PSA unit can be added to increase the makeup hydrogen purity and can be a stand-alone retrofit outside the main hydroprocessing unit.

Note that we maintained hydrogen purity of 80 mol % in the gas feed into the reactor for most of the cases reported, and this variable is controlled in the model by manipulating the quantity of treat gas purge. In some cases (examples available in Appendix K), the hydrogen purity into the reactor was >80% when 99 mol % hydrogen makeup is introduced in the relatively lower hydrogen demand processes.

3.2.7 Space Velocity Adjustments and Catalyst Requirement

Experimental data in Appendix A shows the range of space velocities from available experimental CFP oil hydroprocessing results. Carefully planned parametric studies are necessary for evaluating catalyst performance for both stand-alone and coprocessing configurations. Long-term decisions should not be based on short-duration bench-scale assessments, and all operating variables including reactor temperature and pressure, hydrogen partial pressure, and space velocities need to be varied to quantify their impacts. However, long-term pilot tests may be expensive, and proper experimental design at different scales is important to maximize learnings to enable successful scale-up. As discussed in Section 3.1, short-duration CFP oil stand-alone- and co-hydroprocessing experiments at NREL and Pacific Northwest National Laboratory using catalysts that are commercially relevant for petroleum feedstocks have identified many potential challenges. Bench-scale experiments have also helped understand some potential mitigation strategies for longer duration operations, such as the inclusion of an initial non-isothermal reaction zone when processing CFP oils and other reactive oxygenated feedstocks.

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While space velocities necessary to achieve deoxygenation and hydrogenation to meet product specifications may be directly inferred in the case of stand-alone CFP oil hydroprocessing at a chosen set of conditions, the parameter space and combinations for understanding the limits during co-hydroprocessing are more extensive and include various possible feed options and compositions. Space velocity at lower levels of CFP oil cofeed may need no adjustments if the space velocity for the unit has spare capacity for meeting stringent specifications, e.g., for ultra-low-sulfur diesel. While the CFP oil may be converted to on-spec fuel product, there may be competing or inhibitory effects of CFP oil on petroleum feedstock hydroprocessing that may cause changes in hydrodesulfurization (HDS) and hydrodenitrogenation (HDN). Overall, there may be inhibitory impacts and competition among HDS, HDN, and hydrodeoxygenation, and those effects need to be understood through experimental studies. Although this phenomenon was pronounced for biocrude from hydrothermal liquefaction from waste (sewage sludge and food waste in this reference) [72] and not for CFP oils, this should remain a consideration as the quality of CFP oils may vary in the future and low-sulfur and low-nitrogen CFP oils may not always be produced from different biomass sources. Similarly, it was noticed that the product H:C ratio for VGO/LCO hydroprocessing was lowered in the presence of 20 vol % CFP oil [61]. Given these impacts it is necessary to assess coprocessing situations individually, although the template of such parametric studies may become immediately evident for any specific configuration and feedstock option. Additional adjustment factors may need to be added, e.g., if catalyst activity-inhibiting CO gas is present and accumulates in the recycle system (CO is discussed separately); there may also be inhibitory effects from specific components or impurities in the CFP oil, depending on the source.

Given this uncertainty and the absence of extensive parametric experimental studies, we chose a general approach to cover the space velocity uncertainty. Reaction demands on the catalyst can be estimated from the chemical hydrogen demand in a hydroprocessing system. Our SRD model quantified a chemical hydrogen consumption of approximately 615 scf/bbl. We assumed a weight hourly space velocity (WHSV) of 2 h^{-1} for this system following general ranges and guidelines from literature (Table 28); catalyst loading for the remaining cases were based on prorating the chemical hydrogen consumption with this SRD system. Thus, as an example, if the chemical hydrogen consumption is double of 615 scf/bbl, at 1,230 scf/bbl, the WHSV required for the reactor system is assumed to be half of 2 h^{-1}, at 1 h^{-1}. This information is included in the result summaries provided for key cases later in this report, and for all the cases in Appendix K. Based on the experimental discussion in Section 3.1, we assume the use of commercial catalysts; catalyst unit cost assumptions are stated later under operating costs. For the HT+HC configuration (Figure 12), we assumed for reactor and catalyst costing purposes that the space velocity calculated based on the total chemical hydrogen consumption was doubled for each reactor, since the reactions are spread across the two reactors. Note that the WHSVs reported in Appendix K include metrics based on both fresh feed and total feed (with recycle); the metric based on the total feed can be indicative of deviations from the status quo of petroleum feedstock processing when units are considered for repurposing with the introduction of CFP oils or other oxygenated feedstocks, along with any specific proportion of product recycle.

3.2.8 Operating Temperature and Pressure

Typical reported hydroprocessing temperatures range from 300°C to 400°C (572°F to 752°F), with more severe cracking operations extending to 450°C (842°F), as shown in Table 28. The units are also designed to limit temperature increases (ΔT) within specifications to optimize

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performance and catalyst maintenance from the start to end of run with a catalyst load. The design operating pressure of the unit is determined based on the required severity for the feedstock being handled, with higher design pressures (and higher associated H_2 partial pressures) applied to more difficult heavier feeds. These reactor temperature profile and ΔT are changed during the lifetime of a catalyst load to achieve reactor performance goals.

As previously mentioned, this study is focused on understanding process impacts with specified reactor performance but does not include predictive reactor behavior. Thus, the specifications of reactor temperature, ΔT, and pressure specifications do not impact the model yields and product quality. We picked values within the ranges for hydroprocessing units (Table 28) and then varied them in sensitivity cases to understand process and cost impacts of choosing different design and operating temperatures and pressures. The impacts of varying model yields are considered in separate sensitivity cases.

The base pressures chosen for HT (Section 3.2.2) and HT+HC (Section 3.2.4) reactor inlets are 1,045 psia (72 bara) and 2,045 psia (141 bara), respectively. The base reactor inlet temperature was 340°C (644°F), with a ΔT of 22.2°C (40°F). Note that this assumption was maintained in both configurations. In practice the two reactors in the HT+HC configuration will likely be operated at different temperatures to facilitate different target chemistries in the two reactors, as was during the NREL experiments mentioned in Section 3.1 and in Appendix A (Table A-2), with hydrotreating at 385°C (725°F) and hydrocracking at 400°C (752°F) [61].

3.2.9 CO and CO_2 Formation From Oxygenated Species

As mentioned earlier, the primary advantage of hydroprocessing over FCC processing of CFP oils is higher carbon efficiencies because hydrodeoxygenation (e.g., removal of oxygen as water) is the primary mode of oxygen removal, versus the formation of CO and CO_2 in an FCC where oxygen removal is accompanied by carbon loss. However, the formation of CO and CO_2 to some extent during hydroprocessing is a possibility because of the mix of compounds present in the process. Hydroprocessing results for CFP oils from NREL and Pacific Northwest National Laboratory have so far reported no measured CO formation, and small amounts of CO_2 were reported in some cases [59,84]. Since an acid gas removal system will already be included in the design for H_2S removal, CO_2 accumulation may not be a concern. However, differences in amine reactivity with H_2S and CO_2 may need to be assessed based on specific situations. One must also be mindful of potential concerns such as formation of carboxylic acid and ammonium carbonate salts; associated corrosion may require reevaluation of metallurgy in the reactor effluent cooling sections and wash water requirements to avoid salt precipitation [85]. On the other hand, if CO is formed, it can potentially accumulate in the gas recycle loop; in that case, increasing the proportion of purge gas may be one of the primary options to reduce CO concentrations. The literature indicates that CO formation can inhibit catalyst activity, with one source showing reduced hydrotreatment activity with straight-run gas oil. They reported 5% decrease in saturation, 4%–15% decrease in HDS, 5%–10% decrease in HDN activities upon spiking the hydrogen feed with approximately 1.4 vol % CO [86]. Other studies where CO_2 and/or CO were artificially introduced during gas oil hydrotreating also captured inhibitory effects on HDS, HDN, and hydrogenation [87,88]. However, as noted, we do not expect significant levels of CO to accumulate during CFP oil hydroprocessing, based on bench-scale results reported; acid gas removal included in the gas recycle loop can reduce CO_2 concentrations.

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3.2.10 Water Wash in Reactor Effluent Cooling and Acid Gas Removal

Salts from sulfur, nitrogen, and chlorine (from reaction with hydrogen) have the potential to form highly concentrated and corrosive solutions as hydroprocessing reactor effluents are cooled and condensed to liquids. Therefore, water wash is often required to (1) maintain adequate free water in the effluent train and (2) control the concentration of salts in the condensed aqueous phase to non-corrosive levels. Based on industry guidance, the modeling targets include minimum 25% free water at salt deposition temperature and maximum 2 wt % NH_4HS. This target corresponds to approximately 4.7 wt % ammonium bisulfate $(NH_4)HSO_4$; the goal is to maintain this concentration well below the corrosion limit of 8 wt %. Water produced in hydrodeoxygenation reactions contributes to the effluent wash water, thus reducing the additional water wash makeup injection required to meet the modeled targets. It is noted that CFP oil from woody biomass is low in sulfur, nitrogen, and chlorine, and coprocessing CFP oils of this type is not expected to aggravate concerns of salt formation, unless the biomass feedstock for CFP oil production is switched from a woody source. Operational adjustments may be necessary even if the changes are small and the existing design is capable of accommodating those changes. As mentioned in the previous section, concerns about carboxylic acid and ammonium carbonate salts may need to be addressed if a significant amount of CO_2 is formed during hydroprocessing. Acid gas and ammonia separations were modeled using simple separator (SEP) blocks in Aspen Plus, equimolar amounts of NH_3 and H_2S are reported in the stream summary tables (Appendix L) to represent NH_4HS, with additional H_2S removal reported in a separate stream. For simplicity, complete removal of NH_3, H_2S and CO_2 is included in the SEP blocks. Operating cost estimation basis and amine makeup for acid gas removal are included in Section 3.3.1. Capital costs for amine scrubbers are included in the equipment cost estimates provided later.

3.2.11 Sour Wastewater Impacts

Sour water is a product from hydroprocessing due to water in the feed streams, produced water from deoxygenation, and supplemental wash water. There is a cost of wastewater treatment included in the TEA basis. However, requirements or processing strategies for treatment of new wastewater species introduced with CFP oil is not addressed in the scope of this work. It is assumed that the existing refinery water treatment facility could meet water discharge permit specifications without incurring major capital or operating costs. Cost assumption per gallon of wastewater treatment is included under operating costs (Table 30).

Phenolic compounds may be a major concern for treatment of wastewater from bio-oil upgrading processes where near-complete deoxygenation is not achieved. Phenols are organic acids formed when an aromatic phenyl group (C6H5-) bonds to a hydroxyl group. The most common example is carbolic acid, more commonly known simply as phenol. When phenol dissolves in water it dissociates into a hydrogen ion and phenolate ion or phenoxide, resulting in an acidic and potentially corrosive and toxic solution. With the near-complete deoxygenation requirement for the production of jet fuel, it may be expected that most of the phenolic compounds will be saturated and/or deoxygenated at the exit of the hydroprocessing reactors, thus alleviating this concern regarding significant discharge of phenolic compounds. This may however require additional wastewater treatment considerations if more phenols partition into the aqueous phase when deliberate partial deoxygenation is one of the goals, e.g., for potential production of marine fuels that can tolerate some fuel oxygen [61,89], but more experimental data is necessary before

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drawing conclusions. Partial deoxygenation during hydroprocessing is not a strategy discussed in this report.

3.2.12 Additional Assumptions for Refinery Integration Scenarios

There are several additional assumptions for refinery integration as outlined below:

Amine Systems: The refinery provides lean amine and regeneration of rich amine from existing facility capabilities to the hydroprocessing unit.

Acid Gas: Any CO_2 yielded from CFP oil processing in the refinery will be captured by the amine scrubbing system on the hydrogen recycle loop of the hydroprocessing unit. Incremental CO_2 may impact the apparent capacity of the sulfur recovery unit. However, the impact is expected to be minimal and offset by reduced sulfur load to the refinery with CFP oil processing. Depending on the biogenic feedstock's chemical composition, larger proportions of CO_2 may be produced during coprocessing, and the CO_2 may not be completely removed without augmenting amine system capacity, as may be the case during hydroprocessed esters and fatty acids (HEFA) coprocessing [90].

Catalysts: Any specially designed catalyst technologies for optimization of CFP oil upgrading, if necessary, are assumed to negligibly impact catalyst unit cost; we assumed no changes to commercial catalysts for this report. Increased catalyst consumption for CFP oil upgrading is addressed through space velocity adjustments discussed previously. However, frequent catalyst changeouts can negatively impact the on-stream factor of the unit, as discussed in Section 3.4.4.14.

Product Quality and Blending: The fuel properties of products from CFP oil are assumed to meet quality specifications of fossil derived equivalents after adequate hydroprocessing. This aspect is discussed in Section 3.1 and Section 3.2.3.

3.2.13 Feed System Modification

NREL bench-scale experiments and input from industry partners have led to the conceptual CFP oil feeding strategy proposed in Figure 13. The design aims to minimize undesired thermal polymerization reactions in CFP oil during feeding into a heated system, which leads to coking, plugging, and other operational problems. The CFP oil feed system includes dedicated tankage, pumps, static mixers with hydrogen addition, and a small stabilization reactor upstream of the main hydroprocessing reactor as insurance for handling higher oxygen content CFP oils. This proposed design may also address common mixing and miscibility challenges due to non-homogeneity of hydrocarbon and oxygenate mixtures. The costs of these systems are included in our estimates. More extensive designs of stabilization units may be warranted based on experimental data for specific feedstocks. The current system is based on minimally ensuring that the CFP oil is in the presence of hydrogen, in a catalytic environment within enhanced piping, as it gets heated while being fed into the reactor system; this can prevent polymerization and plugging as has been demonstrated in bench-scale experiments. Catalyst costs are not estimated separately for this section, with an assumption that the catalyst needs for this section can be absorbed with the catalyst cost for the main hydroprocessing reactors since our assumption is that significant hydroprocessing will not occur in this section beyond avoidance of polymerization. The equipment shown in Figure 13 are listed in Appendix B.

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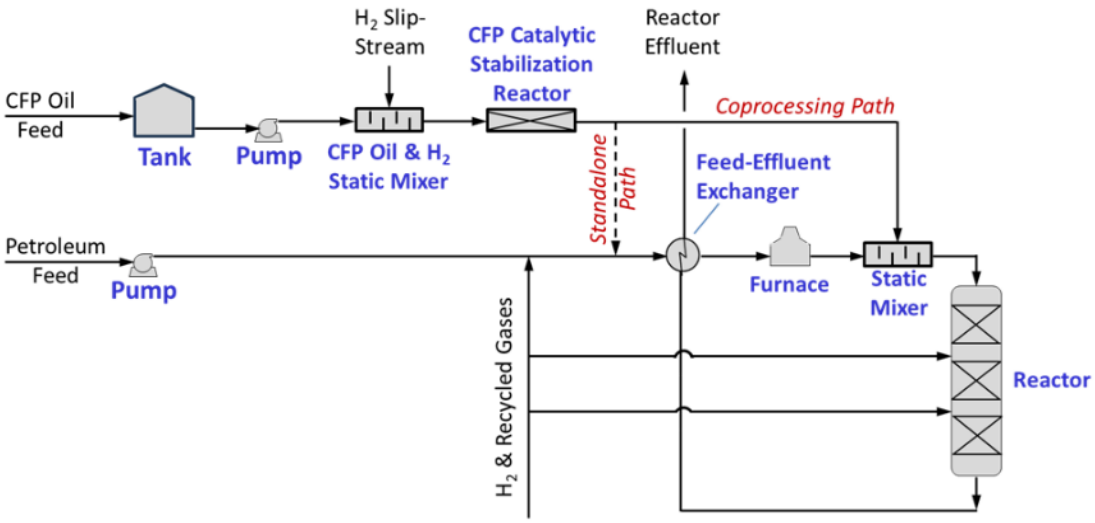

Figure 13. CFP oil feed design for upgrading by hydroprocessing (feed filters not shown)

3.2.14 Density and H:C Ratio Assumptions for Key Streams

Petroleum refineries typically quantify capacities based on volumetric flow rates (BPD). Table 29 lists density assumptions in the model and the H:C ratios of the feed streams based on surrogate model compounds. Note that the final total liquid product after distillation, including all cuts, was targeted to have a H:C ratio of ~1.8, unless otherwise noted.

Table 29. Density Assumptions and Dry Basis H:C Ratio in Hydroprocessing Model

Position Title	Specific Gravity	H:C Ratio [a]	Basis
CFP-B oil with 16.0 wt % oxygen (dry basis)	1.053	1.29	NREL experiments
CFP-A oil with 21.6 wt % oxygen (dry basis)	1.151	1.12	NREL experiments
SRD petroleum feedstock	0.840	1.67	NREL Experiments
LCO petroleum feedstock	0.960	1.28	See Appendix D
VGO petroleum feedstock	0.930	1.51	See Appendix D
Hydroprocessed product recycle for heat management	0.830 [b]	~1.80 [c]	See footnote

[a] Actual H:C ratios shown, without any corrections to reflect effective H:C ratios (i.e., no accounting for oxygen).
[b] Assumed similar, but slightly lower than SRD due to hydrogenation.
[c] Consistent model assumption for total liquid product in all cases (except sensitivity analysis specific to this metric).

3.3 Hydroprocessing Process Economics and Sustainability Analysis

The general framework for the estimation of process economics and sustainability was described in Section 1.2. Additional specifics for hydroprocessing are stated in this section. As noted in Section 1.2.2, capital costs for hydroprocessing were based on KBR estimates that were further correlated for scaling by NREL; details are provided in Appendix B.

3.3.1 Operating Costs

Variable operating cost assumptions are shown in Table 30. These inputs are integrated with the Aspen Plus heat and material balances in the TEA framework to calculate annual variable operating costs for hydroprocessing CFP oil and petroleum feeds.

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Table 30. Variable Operating Cost Inputs

Variable	Information and Operating Cost (cost year in parentheses)
Petroleum feedstocks	Costs for SRD, LCO, and VGO are estimated from a West Texas Intermediate (WTI) benchmark price input and historical data for spot pricing from OPIS [91]: WTI at $85/bbl results in estimates for SRD at $105.88/bbl, LCO at $102.95/bbl, and VGO at $100.78/bbl (2020$).
CFP oil	Prices, including 800 miles of rail transport, are estimated in Section 2: CFP-A Oil: $215.62/bbl (2020$). See Table 24. CFP-B Oil: $148.30/bbl (2020$). See Table 25.
Hydrogen	$1.5/kg (2020$), average price assumption based on steam methane reformers [92]. This price is on a pure H_2 mass basis and does not add costs for other impurities in the supplied gas. In the model, on a volume basis: (i) for 99% pure H_2, the remaining gases are 0.5% CO and 0.5% CO_2, (ii) for 85% pure H_2, the remaining gases are 7.5% CH_4, 5% C_2H_6, 2% C_3H_8, 0.5% C_4H_{10}, approximated based on hydrogen compositions in some refinery operations.
Electricity	$0.082/kWh represents the 5-year average price from EIA Annual Energy Outlook 2023 (2020$).
Hydrotreating and hydrocracking catalysts	SRD hydrotreating WHSV of 2 h^{-1} was used as a basis with corresponding chemical hydrogen consumption of 615 scf/bbl. WHSV and catalyst inventory for other cases were prorated based on the chemical hydrogen consumption for each case. 20% excess catalyst was included. For HT+HC (shown in Section 3.2.4), the WHSV was split equally among the two reactors, i.e., the calculated overall WHSV was doubled for each reactor. Initial fill is replaced every 2 years for hydrotreating and every 3 years for hydrocracking catalysts. Price: $16.6/lb (2014$) [93], (also estimated using CatCost [56] for verification of approximate cost of a NiMo hydrotreating catalyst). Converted to $17.82/lb (2020$).
Natural gas and refinery fuel gas	$3.23/MMBtu for natural gas, 10-year average from EIA (Henry Hub Spot Price) [94].
Steam	Price ranges for different steam pressures (HP, MP, LP, VLP) approximated using natural gas price (above) and applying a 85.7% efficiency toward steam [95]. HP Steam (900 psig): $0.0045/lb; MP Steam (350 psig): $0.0045/lb; LP Steam (150 psig): $0.0045/lb; VLP Steam (15 psig): $0.0044/lb. Other methods and specific situations/uses will change these valuations.
Solids disposal	Calculated from input for reclamation discount in catalyst calculations. Base value for discount is 0%.
Diesel fuel	Usage estimated at 10 gal/h plantwide. $2.57/gal at $85/bbl WTI price (2020$).
Water makeup and cooling water circulation	Price: $0.22/tonne = $0.20/U.S. ton (2004$) [18]. Cooling water circulation estimated cost $0.00768/ton (2020$) circulation based on an assumption of 2.3% of circulation loss, and water makeup price above.
Wastewater	Price: $0.022/gal (2011$). Based on Humbird et al. [52].
Acid gas removal	Amine makeup cost of $0.001226/lb mol inlet gas (2020$) estimated based on makeup rate of 2.5 lb/million scf inlet gas in reference [17] and methyl diethanolamine price of $8.38/gal (2009$) in reference [96].

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Variable	Information and Operating Cost (cost year in parentheses)
	Other cumulative acid gas removal operating costs of $0.477/lb mol acid gas removed were estimated based on reference [17] and unit costs for natural gas, electricity, and cooling water circulation as stated above. Amine makeup estimate for life cycle inventory: 2.5 lb/million scf inlet gas [17].

Note: Costs were updated to 2020 dollars using Producer Price Index for chemical manufacturing [25].

Fixed operating cost basis for salaries, benefits, overhead, plant maintenance, insurance, and taxes are summarized in Table 31. Salaries were adjusted to 2020$ using the U.S. Bureau of Labor Statistics labor cost index [26].

Table 31. Fixed Operating Cost Inputs

Description	Cost Basis
Salaries	$2.55 million/year for 43.2 total positions [a]
Benefits and overhead	90.0% of total salaries
Maintenance	3.0% of FCI
Insurance and taxes	0.7% of FCI

[a] Supervisory and management positions may be split between multiple refinery units and/or complexes.

3.3.2 Carbon Intensity Quantification

Carbon intensity (CI) calculations are coupled with the TEA framework for this analysis. The NREL analysis team, in consultation with Argonne National Laboratory, incorporated greenhouse gas (GHG) emissions factors based on GREET [27] for modeled process inputs and coproducts, including petroleum and CFP oil feedstocks, to enable direct CI calculations for the yielded products. Coproducts, steam and fuel gas, are accounted via displacement credits. An example is provided in Section 3.4.2.2 to illustrate the methodology and show our assumptions for emission factors.

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3.4 Hydroprocessing Results

A significant number of cases were explored for this analysis. Results are categorized to help the reader understand the modeled impacts of various parameters.

3.4.1 Interpretation of Case IDs

Case IDs are provided with the results. A guide is provided in Table 32, allowing the interested reader to independently interpret the Case IDs.

Table 32. Interpretation of Case IDs

Position →	1	2	3	4 and 5	6	7	8
	Petroleum Feed Type	**CFP Oil Type**	**H_2 Makeup Purity**	**Vol % CFP Oil in Feed**	**Liquid Recycle**	**Reactor Inlet H_2 Mol %**	**Process Model Total Feed Rate**
Options	1 = SRD 2= LCO 3 = VGO 4 = LCO+ VGO	1 = CFP-B (16.0% O) 2 = CFP-A (21.6% O)	1,3 = 99% 2,4 = 85% H_2 Temp Control, H_2 feed:quench 1,2 = 33:67 3,4 = 60:40	01 = 100% Petroleum 99 = 100% CFP Oil xx = % Feed CFP Oil	0=0%; 1=10%; 2=20%; 3=30%; 4=40%; 5=50%; 8=Liquid Quench Temp. Control	1=80%; 2=75%; 3=85%; 4=90%; 5=70%; 6=65%; 7=60%	1=75k BPD 2=30k BPD 3=10k BPD 4=3k BPD
Example: 12120211	SRD	CFP-A	99%, 33:67	20% CFP	H_2 temp. control 20% recycle	80%	75k BPD

The example Case ID 12120211 reflects 80 vol % SRD + 20 vol % CFP-A oil hydrotreating with 99% purity hydrogen makeup supply. Reactor cooling is achieved using inter-bed hydrogen quench for temperature control (H-TC) and aided by 20% inter-bed recycle of the total condensed liquid product. The total capacity of the unit is 75k BPD, with 20 vol % or 15k BPD liquid product recycle and the balance of 60k BPD fresh feed.

3.4.2 Example Case To Illustrate Format of Detailed Results in Appendices

An example case is presented here to help familiarize the reader with the format of detailed results presented in the Appendix K. The TEA information below is split into separate tables for description; the appendices include all the information in a single table.

3.4.2.1 TEA Metrics Tables

Table 33. Feed Information for Example Case 12120211

	Value	**Unit**
Case ID	**12120211**	
Total Feed to Reactor	75,000	**BPD**
SRD feed at $105.88/barrel	48,000	BPD
LCO feed at $102.95/barrel	-	BPD
VGO feed at $100.78/barrel	-	BPD
CFP oil (16% O) feed at $215.62/barrel	12,000	BPD

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	Value	Unit
CFP oil feed volumetric proportion of fresh feed	20%	
Liquid product recycle	15,000	BPD
Liquid product recycle volumetric proportion of total feed	20%	
CFP oil oxygen content (dry basis)	21.6%	wt %

Table 33 shows the feed information to the unit. It includes 20 vol % CFP-A oil and 80 vol % SRD in fresh feed. This case includes temperature control using inter-bed hydrogen injection and also has 20 vol % inter-bed liquid recycle to help reactor heat management. In the process model, the total feed to the reactor is 75k BPD, with 20 vol % recycled product liquid (15k BPD) and 80 vol % (60k BPD) fresh feed. Of the 60k BPD, 12k BPD (20 vol %) is CFP-A oil and 48k BPD (80 vol %) is SRD.

Table 34 shows modeled product information for the case. As mentioned elsewhere, although distillation is included in the model to provide a nominal idea of approximate distributions among the three potential products: gasoline, jet, and diesel/heavies, the cuts may be changed based on fuel demand and quality specifications. The primary criteria for all the hydroprocessing cases are (1) the removal of all oxygen and (2) having an H:C molar ratio of ~1.8 in the total fuel product (including all three cuts). The units of barrels gasoline, jet, and diesel equivalents (BGE, BJE, and BDE) are provided to allow the reader to gauge the total energy content of the products in different units. These units are derived from GGE, GJE, and GDE mentioned earlier by using a conversion of 42 gal/bbl. Information about weight precents in each cut in the model are included but not emphasized as a strict metric for this report.

Table 34. Product Information for Example Case 12120211.

Model outputs, based on underlying assumptions, are marked “calc”; unmarked values are model specifications or targets.

	Value	Unit
Total Liquid Fuel Product	719,338 (calc)	lb/h
Barrels gasoline equivalent (total fuel product)	64,833 (calc)	BGE/day
Barrels jet equivalent (total fuel product)	60,548 (calc)	BJE/day
Barrels diesel equivalent (total fuel product)	58,595 (calc)	BDE/day
Total liquid product hydrogen-to-carbon ratio	1.79	
Gasoline product (modeled cut)	24% (calc)	wt %
Jet product (modeled cut)	47% (calc)	wt %
Diesel/heavies product (modeled cut)	29% (calc)	wt %

Table 35 shows information related to hydroprocessing and catalyst assumptions. It includes the SRD hydrotreating chemical hydrogen consumption of 615 scf/bbl and WHSV of 2 h^{-1} used as the basis for prorating the WHSV for the cases presented. For the HT+HC cases (Figure 12) the calculated WHSV is doubled because the reactions occur over two reactors. In this case the hydrogen consumption increases to 1,316 scf/bbl of total feed (including recycle) and the

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capacity of fresh feed is reduced from 75k BPD to 60k BPD (with 20% liquid recycle). As a result, the WHSV based on fresh feed is reduced from 2 h^{-1} for the 100% SRD case to 0.75 h^{-1}.

Recycling liquid product into intermediate points of the reactor helps cool the reactor, as is the case here with 20% liquid product recycle. As discussed later, the cooling advantage is lost if the full recycle is fed to the front end of the reactor, as may be the case when reusing a reactor that is not already equipped for inter-bed recycle of liquid. In both inter-bed and front-end liquid recycle configurations the recycled product liquid reduces reactivity of the total feed, thus reducing the hydroprocessing exotherm; the recycled liquid also helps maintain the liquid flow profile in the reactor at the designed flow rate, even though fresh feed rate is curtailed.

Table 35. Hydroprocessing and Catalyst Information for Case 12120211.

Model outputs, based on underlying assumptions, are marked "calc"; unmarked values are model specifications or targets.

	Value	Unit
Catalyst type	Commercial	
SRD hydrotreating chemical H_2 consumption basis	615	scf/bbl
SRD hydrotreating WHSV basis	2.00	h^{-1}
Chemical H_2 consumption based on total feed	1,316 (calc)	scf/bbl
Hydroprocessing inlet pressure	1,045	psia
Hydroprocessing inlet temperature	644	°F
Hydroprocessing outlet temperature	684	°F
Hydrotreating		
WHSV based on fresh feed (prorated from SRD basis above, using chemical H_2), doubled if HT+HC configuration	0.75 (calc)	h^{-1}
Catalyst load (with 20% excess)	1,268,916 (calc)	lb
Catalyst replacement period	2.00	years
Catalyst unit cost (2020$)	17.82	$/lb
Hydrocracking		
WHSV (prorated from SRD basis above, using chemical H_2), doubled if HT+HC configuration	N/A	h^{-1}
Catalyst load (with 20% excess)	N/A	lb
Catalyst replacement period	N/A	years
Catalyst unit cost (2020$)	17.82	$/lb
Total catalyst cost per feed barrel	0.52 (calc)	$/bbl
H_2 temperature control with 20% liquid recycle		
Hydroprocessing carbon efficiency for CFP oil	90.0%	
Hydroprocessing carbon efficiency for petroleum feed	98.6%	

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Table 35 also shows the modeled carbon efficiencies for fossil petroleum and CFP oil feeds. Note that ~90% carbon efficiency from CFP oil has been reported in many NREL experiments and was thus chosen as the base case efficiency; this assumption is varied later in sensitivity cases.

Table 36 includes metrics for hydrogen use in the process. In our economic spreadsheet, chemical hydrogen consumption in standard cubic feet (scf) of hydrogen per barrel (bbl) of fresh feed to the unit is used to prorate the WHSV and calculate the catalyst load, based on the stream flow rate of the fresh feed for scaling (the calculation may also be performed using a total feed basis since this involves simple and appropriate algebraic adjustments). The purity of hydrogen used in each case is shown. The hydrogen purity at the reactor inlet (modeled value) is also shown. In some cases, the usual specification of 80 mol % cannot be met with a purge, especially when the makeup hydrogen purity is 99 mol % and hydrogen consumption is relatively low, such as during 100% SRD hydroprocessing. For the petroleum base case with 100% SRD hydrotreating (case 12201011) we use 85 mol % purity hydrogen makeup.

Table 36. Hydrogen Metrics for Case 12120211.

Model outputs, based on underlying assumptions, are marked “calc”; unmarked values are model specifications or targets.

	Value	Unit
Makeup H_2 purity	99%	mol %
H_2 purity at reactor inlet gas feed	80%	mol %
Chemical H_2 consumption, based on total feed	1,316 (calc)	scf/bbl
Chemical H_2 consumption, based on fresh feed	1,645 (calc)	scf/bbl
Reactor H_2 feed, based on total feed	6,448 (calc)	scf/bbl
Reactor H_2 feed, based on fresh feed	8,060 (calc)	scf/bbl
Makeup H_2, based on total feed	1,572 (calc)	scf/bbl
Makeup H_2, based on fresh feed	1,965 (calc)	scf/bbl
Net H_2 after PSA recovery, based on total feed	1,389 (calc)	scf/bbl

Table 37 includes a summary of TEA results and sustainability metrics. The MFSP is calculated on an energy basis of the total fuel products and is shown in GGE. GJE and GDE values are also shown to allow a quick estimate of prices of different fuels from this process, with the simple assumption that costs will be prorated on an energy content basis. In practice volumetric units are used by refiners; the GGE, GJE, and GDE units are included here to provide information using approximate average equivalent volumes of different fuels while avoiding potential errors in estimating densities. It is noted that other factors, including market forces, determine fuel retail prices. Refinery operations are adjusted based on market prices and delta yield analyses, which are not included in this analysis. The values shown are for 75k BPD plants, which is the total capacity of the base process model. Our economic spreadsheets are enabled to assess TEA impacts with scaled capital costs estimated at other lower capacities; this is discussed later under sensitivity cases.

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Liquid feed is the dominant cost, but as expected and shown later, hydrogen costs increase as the proportion of CFP oil increases. Capital costs will also increase relative to existing petroleum processing units if CFP oil is introduced without appropriately curtailing feed capacity to allow reuse of existing capacities of unit operations. Further discussion about constraints is included later in this report.

An MFSP estimate with the reuse of existing paid-up capital for a 100% petroleum feedstock unit with expenses for subsequent modifications is also projected for each case. In this context, the 100% petroleum SRD hydrotreating base case is used for calculating capital savings when existing paid-up equipment is repurposed for CFP oil hydrotreating or co-hydrotreating (configuration shown in Section 3.2.2). The base case number used for estimating repurposed capital in each case is included in Appendix K. The base GHG emissions and corresponding reductions are also based on the 100% SRD hydrotreating case. There is a corresponding petroleum hydroprocessing baseline for LCO/VGO feed in the HT+HC configuration for assessing capital savings and GHG reduction in the other configuration (configuration shown in Section 3.2.4).

Table 37. TEA and Sustainability Metrics for Case 12120211.

Model outputs, based on underlying assumptions, are marked “calc”; unmarked values are model specifications or targets. Note that the fresh feed was 80 vol % SRD and 20 vol % CFP-A.

	Value	Unit
Capital		
Fixed capital investment	$275.7 (calc)	million $
Working capital	$13.8 (calc)	million $
Operating costs	$2.99 (calc)	/GGE product
Petroleum feed cost	$1.87 (calc)	/GGE product
CFP oil feed cost	$0.95 (calc)	/GGE product
Net hydrogen makeup cost	$0.14 (calc)	/GGE product
Fixed operating costs	$0.02 (calc)	/GGE product
Coproduct/utility credits	($0.03) (calc)	/GGE product
Other operating costs	$0.06 (calc)	/GGE product
Capital charges, income tax, etc.	$0.04 (calc)	/GGE product
MFSP, repurposing 100% petroleum unit with paid-up capital costs for the unit ($/GGE)	$3.00 (calc)	/GGE product
MFSP, with full capital costs ($/GGE)	$3.03 (calc)	/GGE product
MFSP, with full capital costs ($/GJE)	$3.24 (calc)	/GJE product
MFSP, with full capital costs ($/GDE)	$3.35 (calc)	/GDE product
Fuel life cycle CI	76.0 (calc)	g CO_2e/MJ prod
GHG emissions for petroleum diesel case	90.8 (calc)	g CO_2e/MJ prod
GHG reduction over petroleum diesel case	16.3% (calc)	

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Additional economic calculation worksheets with capital costs using methods outlined in Appendix B, operating costs, and DCFROR calculations for MFSP are included for select cases in Appendices G–J. These detailed worksheets are limited to some bookend cases to limit this report to a reasonable size; the interested reader can get interpolated estimates for other cases using some of these bookend cases presented.

3.4.2.2 Life Cycle Inventory and GHG Emissions Calculations

Table 38 includes the life cycle inventory and corresponding life cycle GHG emission calculations for the example case above. The emission factors are based on R&D GREET 2023 [27] and are used for calculating GHG emissions for all the cases reported. The emission factors are not repeated in following tables and only final calculated GHG emissions are presented for subsequent results. For cases where exact emission factors are not available for a specific situation, e.g., for SRD, LCO, and VGO after these intermediate refinery streams are produced, but before they are hydroprocessed to fuels, a small adjustment was made based on available information for hydroprocessed SRD to remove the estimated GHG emissions from hydroprocessing, as calculated by our model, thus matching the overall diesel baseline life cycle GHG emissions modeled with R&D GREET 2023. Specifically, the R&D GREET 2023 value of 15.629 kg CO_2e/MMBtu for diesel was adjusted to 13.9066 kg CO_2e/MMBtu in Table 38 after removing SRD hydroprocessing emissions estimated by our model. Note that hydroprocessing-generated emissions are accounted via the life cycle inventory of our process models, and this adjustment avoids double-counting. This is important for the other cases evaluated, because the introduction of CFP oil changes the calculated GHG emissions for hydroprocessing, and those emissions are evaluated using the life cycle inventory from the process model for each case. The emissions associated with the production of untreated LCO and VGO streams were assumed identical to SRD (on an energy basis) and also use the derived value for untreated SRD (13.9066 kg CO_2e/MMBtu).

Table 38. Life Cycle Inventory and GHG Emission Calculation Worksheet for Case 12120211.

All flow rates/values are from the process model or otherwise calculated.

	Flow Rate or Value	Emission Factor	Emission Factor Unit	Emission kg CO_2e/h
Resource Consumption				
SRD				
Mass flow, lb/h	588,851			
LHV, MMBtu/h	10,611	13.90660	kg CO_2e/MMBtu	147,561
LCO				
Mass flow, lb/h	0			
LHV, MMBtu/h	0	13.90660	kg CO_2e/MMBtu	0
VGO				
Mass flow, lb/h	0			
LHV, MMBtu/h	0	13.90660	kg CO_2e/MMBtu	0
CFP oil				
Mass flow, lb/h	201,717	0.08996	kg CO_2e/lb	18,145

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	Flow Rate or Value	Emission Factor	Emission Factor Unit	Emission kg CO_2e/h
LHV, MMBtu/h	2,466			
H_2 (pure H_2 basis) from natural gas, lb/h	23,087	4.28425	kg CO_2e/lb	98,912
Hydrotreating catalyst, lb/h	76	4.19094	kg CO_2e/lb	320
Hydrocracking catalyst, lb/h	0	3.11017	kg CO_2e/lb	0
Amine makeup, lb/h	74	3.14443	kg CO_2e/lb	232
Water makeup, lb/h	229,014	0.00000	kg CO_2e/lb	0
Boiler chemicals, lb/h	8	3.14443	kg CO_2e/lb	24
Cooling tower chemicals, lb/h	4	3.14443	kg CO_2e/lb	12
Diesel fuel, lb/h	71	1.73309	kg CO_2e/lb	123
Grid electricity, kW	17,251	0.43952	kg CO_2e/kWh	7,582
Fuel and Natural Gas Accounting				
Fuel gas produced from hydroprocessing				
Biogenic source, MMBtu/h	415			
Fossil source (incl. impurities), MMBtu/h	365			
Impurities from H_2 supply (fossil), MMBtu/h	45			
Net fuel gas produced (fossil), MMBtu/h	320			
Consumption, MMBtu/h	314			
Consumption (fossil), MMBtu/h	314			
Excess fuel gas (fossil), MMBtu/h	6			
Additional biogenic used, MMBtu/h	0			
Net biogenic fuel gas produced, MMBtu/h	415			
Coproducts for Displacement Credit				
Excess fuel gas (fossil), MMBtu/h	6	−11.85826	kg CO_2e/MMBtu	−73
Net biogenic fuel gas produced, MMBtu/h	415	−72.79877 [a]	kg CO_2e/MMBtu	−30,213
Liquefied petroleum gas (biogenic)	0	−72.36334	kg CO_2e/MMBtu	0
Liquefied petroleum gas (fossil)	0	−4.29854	kg CO_2e/MMBtu	0
HP steam (900 psig), MMBtu/h	0	−91.29027	kg CO_2e/MMBtu	0
MP steam (350 psig), MMBtu/h	0	−91.29027	kg CO_2e/MMBtu	0
LP steam (150 psig), MMBtu/h	141	−91.29027	kg CO_2e/MMBtu	−12,847
VLP steam (15 psig), MMBtu/h	220	−91.29027	kg CO_2e/MMBtu	−20,051
Products LHV energy basis accounting				
Gasoline (biogenic), MMBtu/h	720			
Distribution		0.31477	kg CO_2e/MMBtu	227
Combustion		0.04405	kg CO_2e/MMBtu	32

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	Flow Rate or Value	Emission Factor	Emission Factor Unit	Emission kg CO_2e/h
Jet (biogenic), MMBtu/h	1,261			
Distribution		0.31477	kg CO_2e/MMBtu	397
Combustion		0.04405	kg CO_2e/MMBtu	56
Diesel (biogenic), MMBtu/h	644			
Distribution		0.31477	kg CO_2e/MMBtu	203
Combustion		0.04405	kg CO_2e/MMBtu	28
Heavy fuel (biogenic), MMBtu/h	0			
Distribution		0.31477	kg CO_2e/MMBtu	0
Combustion		0.04405	kg CO_2e/MMBtu	0
Gasoline (fossil), MMBtu/h	2,457			
Distribution		0.31477	kg CO_2e/MMBtu	773
Combustion		79.82602	kg CO_2e/MMBtu	196,105
Jet (fossil), MMBtu/h	4,907			
Distribution		0.31477	kg CO_2e/MMBtu	1,545
Combustion		79.82602	kg CO_2e/MMBtu	391,693
Diesel (fossil), MMBtu/h	3,183			
Distribution		0.31477	kg CO_2e/MMBtu	1,002
Combustion		79.82602	kg CO_2e/MMBtu	254,081
Heavy fuel (fossil), MMBtu/h	0			
Distribution		0.31477	kg CO_2e/MMBtu	0
Combustion		79.82602	kg CO_2e/MMBtu	0
Total GHG Emissions (kg CO_2e/h)				**1,055,868**
Total liquid fuel products (biogenic), MMBtu/h	2,625			
Total liquid fuel products (fossil), MMBtu/h	10,546			
Total liquid fuel products, MMBtu/h	13,171			
Percent biogenic liquid fuel (energy basis)	19.93			
Emissions, kg CO_2e/MMBtu	80.2			
Emissions, g CO_2e/MJ	76.0			

[a] Note that in addition to production, this is taking credit for subsequent *avoidance of fossil carbon emissions when the fuel gas is used by the consumer for displacing fossil derived fuel gas*. It should not be confused with carbon sequestration.

3.4.3 100% Petroleum to 100% CFP Oil Feed With H_2 Temperature Control

There is a vast parameter space that can be covered even within the limits of the single-stage configurations presented in this report (Figure 11 and Figure 12). It should be noted again that reactor performance assumptions are maintained for all the cases reported here, with 90% carbon

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efficiency for CFP oils to hydrocarbon fuels, and a final H:C ratio of ~1.8 in the overall fuel product. Improving reactor performance and with the adjustment of operating parameters is an area of experimental research; here we focus on other process implications of varying operating parameters, understanding that underlying reactor performance effects are not captured automatically in the model. At this point we do not have enough experimental and/or predictive information to make concurrent reactor performance adjustments in the process model.

This section presents a general sweep of the parameter space as we transition from 100% petroleum (SRD and LCO/VGO) to 100% CFP oils (CFP-A and CFP-B) to get a general idea of the boundaries and limitations for processing CFP oils. We use a base model setup with 75k BPD hydroprocessing capacity and hydrogen temperature control, irrespective of the feed combinations. No liquid product recycle is included in the cases presented in this section. The cost and sustainability estimation methods discussed earlier allow us to present corresponding MFSP and GHG metrics. Process parameters stretch significantly from the 100% petroleum to 100% CFP oil cases and are often beyond conventional design boundaries, but this theoretical exercise helps inform subsequent discussions regarding process adjustments and key sensitivity parameters. Figure 14 and Figure 15 show economic and GHG estimates that include capital costs, without any assumptions of paid-up capital at the refinery. As is expected, at a large scale (75k BPD), capital and fixed costs have minor impacts. A sensitivity case is included later showing projections at smaller scales. These modeled cases with temperature control using hydrogen quench, have a third of the hydrogen fed to the front end with additional H_2 fed to intermediate points. The front-end hydrogen feed proportion can vary for different reactor designs and corresponding catalysts; the effect of changing this split ratio is discussed in a subsequent sensitivity case. Hydrogen starvation needs to be avoided throughout the reactor profile during operations because it can lead to coking and other undesirable reaction behavior. Although the process model includes only two hydrogen injection points after the feed point (for simplicity), refinery reactor designs can include more hydrogen injection points, thus achieving the desired cooling and avoiding hydrogen starvation through the profile of the reactor. It is noted that the modeled cases processing CFP oils generally have higher hydrogen to feed ratios (Appendix K) compared to typical petroleum refining setups (Table 28); this is especially higher when hydrogen quench is the only means for cooling a reactor processing CFP oils because hydrogen supply for meeting the cooling requirements exceed the reactor's hydrogen feed requirement (as estimated from the ratios of H_2 feed to chemical H_2 consumption in Table 28). Total hydrogen fed to the reactor was at least 3.5x the chemical hydrogen demand in all cases, with significantly higher values for most of the cases (range of values 3.58 to 8.37 for all the cases in Appendix K). These ratios are in alignment or often higher than the ranges for petroleum feed processing (Table 28), thus likely avoiding hydrogen starvation for the overall reactor in the modeled cases. Optimization to use lower hydrogen to feed ratios while meeting reactor performance and maintenance requirements as well as ensuring proper distribution of hydrogen through the profile of the reactor may be topics of future studies for CFP oil hydroprocessing.

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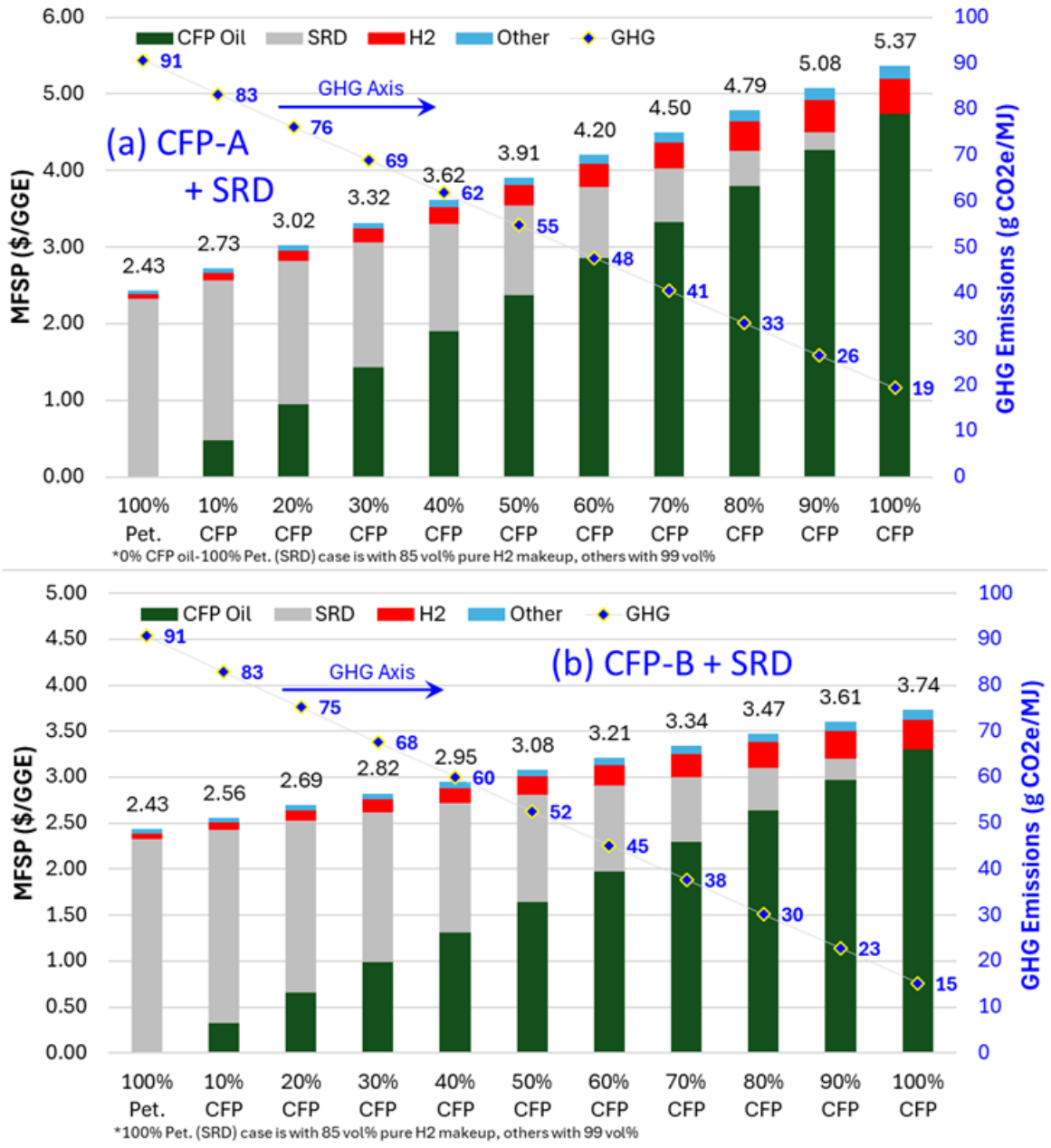

Figure 14. Projected costs and GHG emissions if CFP oils (CFP-A and CFP-B) and SRD are hydrotreated in various proportions with H_2 temperature control. Note that many of the cases are hypothetical and not feasible without further operational adjustments and/or equipment modifications in repurposed units with conventional design/operational guidelines.

Figure 14 and Figure 15 show a linear trend for key cost contributors (MFSP, CFP oil feed, petroleum feed, and other costs) and GHG emissions. This shows that metrics for any intermediate case within this configuration can be estimated via interpolation from the bookend cases; this is convenient from a reporting standpoint. Details are included for the four bookend cases with 100% each of CFP-A oil (Appendix E), CFP-B oil (Appendix F), SRD (Appendix G), and LCO/VGO (Appendix H). The figures show that besides CFP oil costing more per barrel than petroleum feed, hydrogen requirements for processing increases for CFP oil since

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significant hydrodeoxygenation is necessary. The linear trends also point to directional validity of the TEA models that were exercised over a wide range of parameters.

With consistent assumptions of achieving an H:C ratio of ~1.8 in the total liquid product, it is apparent that the lower oxygen content CFP-B oil (16.0 wt % O) costs less to process and requires less hydrogen than CFP-A oil (21.6 wt % O); the unit cost of CFP-A is also higher based on the estimates presented in Section 2.3.12.5.

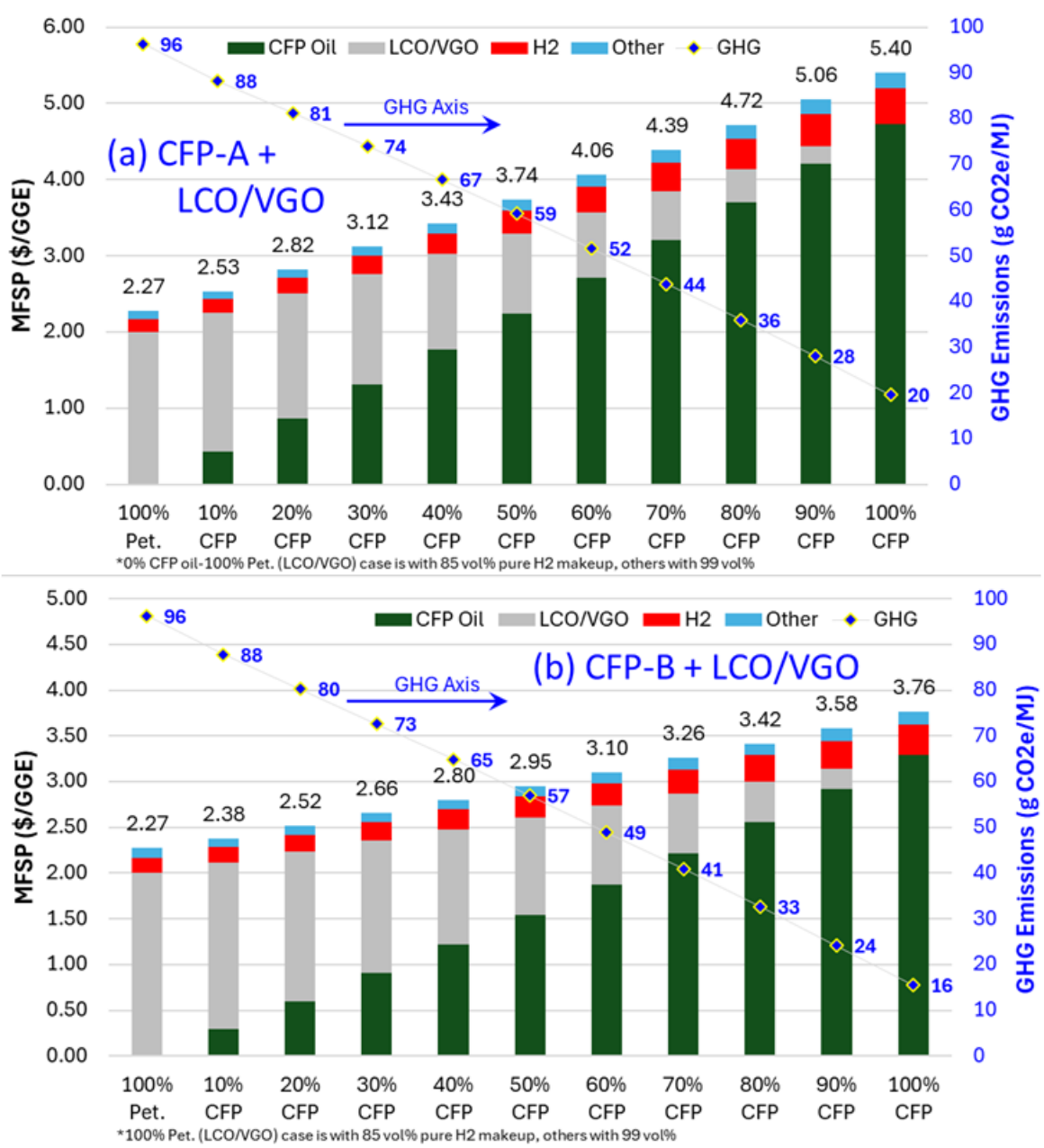

Figure 15. Projected costs and GHG emissions if CFP oils (CFP-A and CFP-B) and 50/50 vol % LCO/VGO are pre-hydrotreated and hydrocracked in various proportions with H_2 temperature control. Note that many of the cases are hypothetical and not feasible without further operational adjustments and/or equipment modifications in repurposed units with conventional design/operational guidelines.

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SRD costs less to process from a hydrogen use standpoint and correspondingly has lower GHG emissions. However, the heavier LCO/VGO feedstock produces more fuel (with the H:C 1.8 product assumption) on an energy basis after hydroprocessing and hence has a lower MFSP than the SRD derived fuel (besides the slightly lower cost/barrel of LCO and VGO compared to SRD). Note that the LCO/VGO process includes capital costs for an additional hydrocracker and the reactor system costs more because of higher pressure; however, the total fuel product energy and volume are significantly higher, resulting in a lower overall MFSP for the LCO/VGO derived fuel compared to SRD at the 75k BPD scale when using feed cost assumptions shown in Table 30. GHG emissions are also higher for LCO/VGO compared to SRD because of the higher H_2 demand to meet the same H:C ratio of ~1.8 in the products.

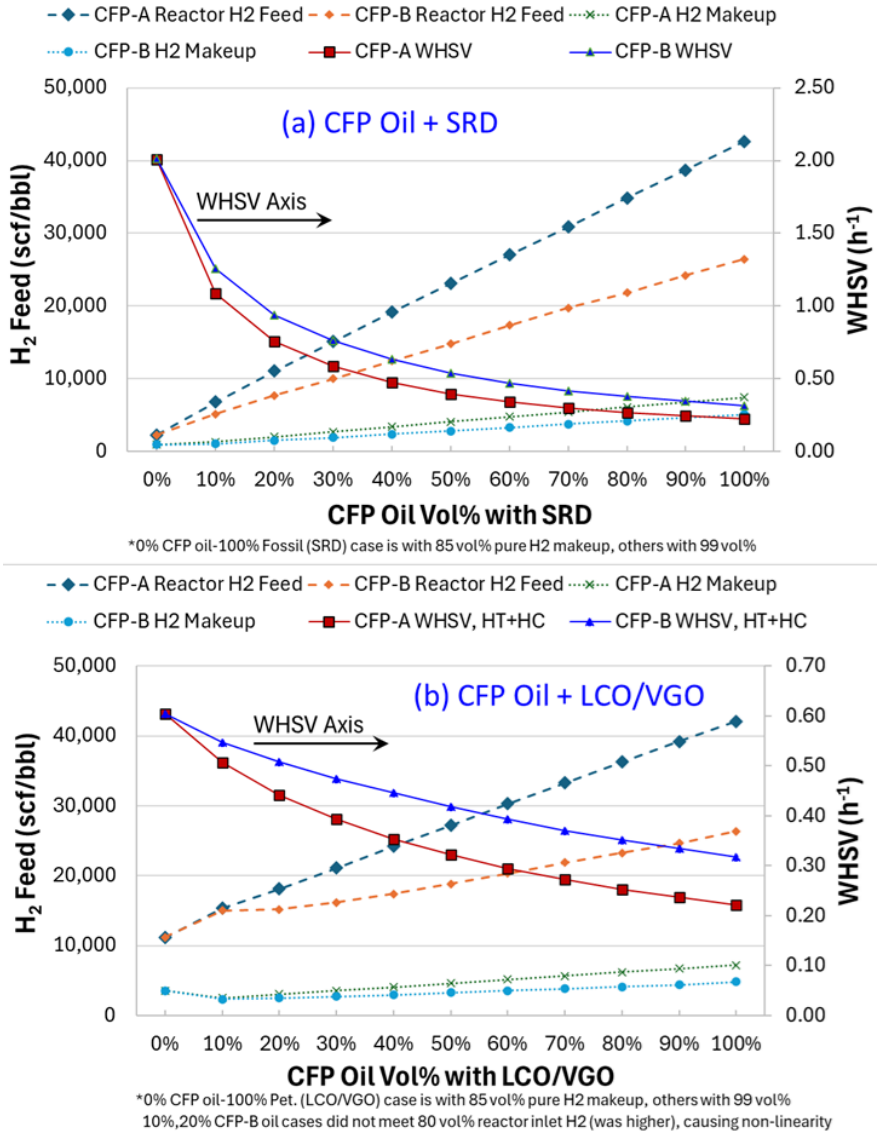

Figure 16. Key hydrogen parameters for CFP oil in different volume proportions with (a) SRD and (b) LCO/VGO (50 vol % LCO + 50 vol % VGO) feed. Reason for non-linearities in H_2 feed (scf/bbl) are mentioned in footnotes, with further exploration in subsequent sensitivity cases. Non-linearities in WHSV (mass basis) caused by changing density of the mixture (volume basis on x-axis); there is greater difference in densities of SRD vs. CFP oil compared to LCO/VGO vs. CFP oil.

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Additional parameters related to H_2 consumption and WHSV are shown in Figure 16. Non-linearities and discontinuities are explained in the caption and footnotes below the figures. For Figure 16b, the significant change in trends for 0% and 10% CFP oil cases are due to higher than the target 80 mol % H_2 concentration in the reactor gas feed and the use of 85 mol % purity H_2 makeup in the 0% CFP oil cases (vs 99% in the other cases); sensitivities to these two parameters are discussed in further detail in Sections 3.4.4.4 and 3.4.4.5.

3.4.4 Sensitivity Cases

The previous section included a sweep of different liquid feed compositions through the process model using a fixed set of assumptions. Many of those cases presented need to operate beyond current hydroprocessing design boundaries. Technology providers may prefer to maintain similarities with current designs that have evolved over decades to minimize risks even as hydroprocessing technologies adapt to support new feedstocks. For repurposing existing units, it is imperative to adjust process operations to help maintain conditions that mimic incumbent use. The reactor is at the heart of the process and its design and adaptation is key for adjusting remaining operations around the unit. From a reactor process parameters standpoint, gas and liquid flows and the capability of the loaded catalyst to process the feed material under similar temperatures and pressures are the most important considerations. The following subsections explore sensitivities to various parameters to help assess process impacts from various choices. Summaries of these cases are provided in Appendix K using the template explained in Section 3.4.2.1.

One of the key handles for repurposing existing units for processing or coprocessing CFP oils is the use of liquid product recycle to alleviate temperature control challenges arising from the significantly higher exotherm during the hydroprocessing of highly oxygenated feedstocks such as CFP oil. As discussed later, both hydrogen and liquid quench streams are used in hydroprocessing setups for temperature control. The acronyms H-TC for hydrogen quench temperature control and L-TC for recycled product liquid quench temperature control were mentioned earlier. All the cases in the previous section used H-TC without liquid product recycle. For temperature control via inter-bed injection of hydrogen and/or recycled product liquid, those injected streams are modeled at 43.3°C (110°F), assuming cooling water as the means for cooling those streams via indirect heat exchange.

The model is capable of exploring a large parameter space beyond what is included in this report. The cases presented in this section were selected to help understand general trends and impacts of key parameters/assumptions. They are mostly based on single parameter variations from a chosen base case and since the reactor model is not predictive, associated changes are not convoluted with changes in other parameters. Thus, the reader may combine the results of impacts from multiple parameter changes to get a reasonable estimate of their combined impact.

3.4.4.1 Sensitivity Results Plot Template Using 100% SRD Hydrotreating Example

Results from the 100% SRD case are shown in Figure 17 to help familiarize the reader with the plots in subsequent sections where results are presented using the same template.

The first bar shows operating costs (OPEX), without feed. Since feed cost is the dominant contributor to OPEX (see Figure 14 and Figure 15), removing the feed cost helps understand sensitivity of other operating cost contributions. The feed cost impact can be prorated based on the fresh feed input and proportions of petroleum and CFP oil feed (discussed in Section

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3.4.4.15). Unit costs of different feed material are shown in Table 30 and further details regarding feed material and their flows are included for each case in Appendix K. The second bar shows the cost of net hydrogen (after PSA recovery of hydrogen from purge gas). Net hydrogen consumption is the dominant contributor toward the previous metric (OPEX without feed); other significant contributors to the OPEX without feed include electricity, fuel gas, royalties, fixed operating costs, and catalyst costs; note that steam and fuel gas credits are also lumped into this metric. Cost calculation details for this 100% SRD case are shown in Appendix G. The third bar shows the life cycle CI of fuel produced from the process; depending on the specific case, net hydrogen consumption is usually one of the primary drivers of CI along with life cycle emissions from liquid feedstocks production and fuel product end-use combustion (Table 38). The fourth bar shows the H_2 feed per total barrel of liquid feed into the reactor (including recycle) and is a metric that can help understand gas flow changes with respect to associated equipment constraints; this can guide operating parameter adjustments to move the operations within the constraints. The fifth bar shows the FCI; this is shown separately because capital charges per gallon are otherwise small and not very noticeable in the MFSP breakdown when compared to operating costs for feed and hydrogen inputs in large-scale systems, such as the 75k BPD cases presented.

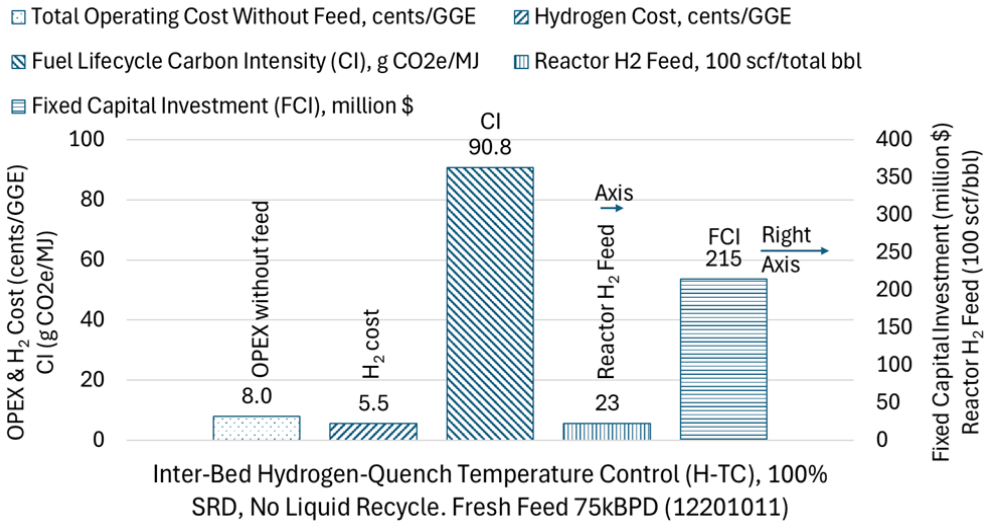

Figure 17. Model results for the 100% SRD hydrotreating case

3.4.4.2 Liquid Product Recycle in H-TC and L-TC Configurations

The case configurations in Section 3.4.3 included inter-bed H_2 injection for reactor temperature control (H-TC), with no liquid product recycle to the reactor. For high H_2 demand processes with high exothermicity, cooling via H_2 injection alone will result in very high gas flows into the reactor system, far exceeding the requirements to achieve the reactor target chemistry. This can result in unconventional design parameters. The H-TC approach needs to be supplemented with liquid product recycle for additional cooling in these cases. Liquid product recycle may reduce the reactivity of the total feed liquid, depending on the stream chosen for recycle. In this study we recycle a portion of the total condensed liquid product after the LP flash unit; hence, it reduces conversion requirements and H_2 demand for processing this recycled portion. Other configurations, not included here, often include the recycle of the distilled heavy fraction after

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hydrocracking; sending the heavy fractions through another hydrocracking reaction cycle helps maximize desired fuel cuts, but also involves additional reactions and associated exothermicity.

The assumption for most of the cases presented in this study include inter-bed recycle of the LP flash liquid product. The illustration in Section 3.4.2 includes 20% inter-bed liquid recycle. Front-end recycle, instead of inter-bed recycle is also an option. This option is included in the third case presented in Figure 18 and Table 39. This may be the case when repurposing for CFP oil processing, where retrofitting an existing reactor for inter-bed recycle is not feasible. The front-end recycle can still help reduce reactivity of the total reactor feed, helping maintain both liquid and gas flow rates into the reactor within existing design limits. However, liquid inter-bed injection is more effective in many respects, as illustrated in Figure 18 (third case with front end vs. second case with inter-bed recycle of 20% liquid product). The model assumes (in all the cases presented in this report) the curtailment of the unit's fresh feed capacity to compensate for the additional liquid introduced via recycle. A case without liquid recycle is also shown (first case in Figure 18) for comparison. The fourth case in Figure 18 and Table 39 involves L-TC and is discussed in the following paragraph.

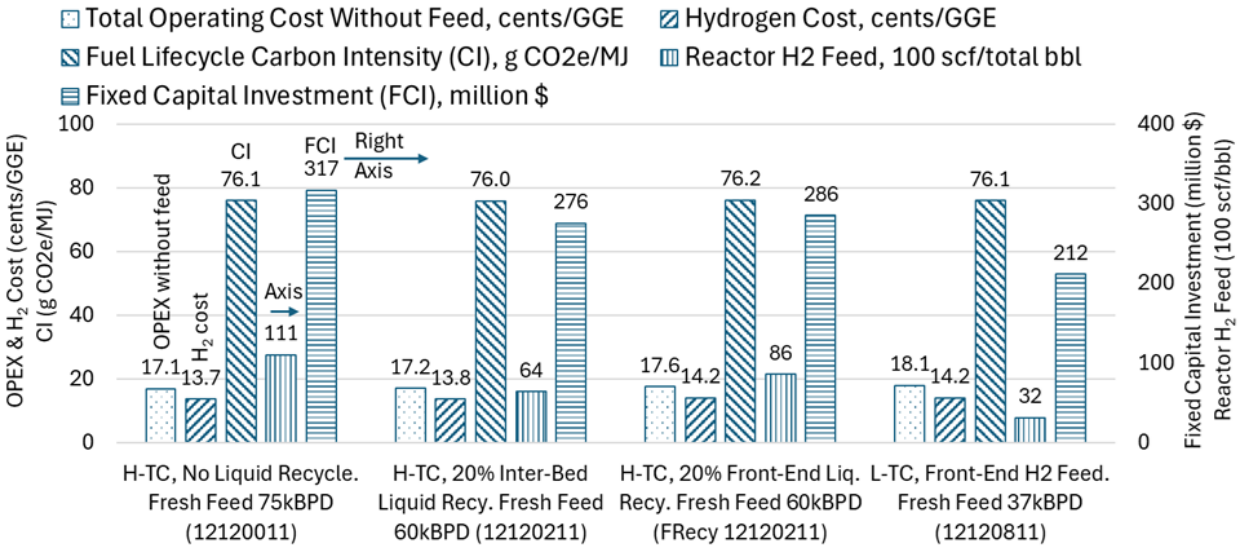

Figure 18. Impacts of hydrogen (H-TC) and liquid recycle (L-TC) quench temperature control on some process, cost, and sustainability metrics. All cases include 20 vol % CFP-A oil with SRD feed. Total feed rate is 75k BPD (includes any liquid recycle to reactor).

Liquid Recycle Temperature Control (L-TC)

Case 12120811 (fourth case) in Figure 18 and Table 39 introduces temperature control by varying the inter-bed liquid recycle (L-TC). In this case all hydrogen is fed to the front end of the reactor (i.e., no inter-bed H_2 injection). Hydrogen flow can be set independently based on reactor design, including any variations for the type of feed. For the modeled cases with liquid recycle control we determined the flow rate by maintaining the hydrogen flow into the reactor at approximately four times of the chemical hydrogen demand, generally aligning with ranges in Table 28. The same conversion criteria stated earlier was maintained, including near-complete deoxygenation of CFP oil and ~1.8 H:C ratio in the total product fuel (noting again that the reactor models are not predictive). The temperature control setup determines the liquid recycle rate to achieve the required cooling with the other chosen design/operating parameters. In this

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case 50% of the liquid product is recycled (Table 39). Hydrogen flow can be adjusted independently based on experimentally determined requirements/sensitivity of the catalyst for different liquid feeds.

Table 39. Results for Liquid Product Recycle Cases (Additional Metrics in Appendix K)

Case ID	12120011	12120211	FRecy 12120211	12120811
Temperature control	H-TC	H-TC	H-TC	L-TC
Total feed to reactor, BPD	75,000	75,000	75,000	75,000
SRD feed at $105.88/barrel, BPD	60,000	48,000	48,000	29,718
CFP oil feed at $215.62/barrel, BPD	15,000	12,000	12,000	7,430
CFP oil feed vol. proportion of fresh feed	20%	20%	20%	20%
Liquid product recycle, BPD	0	15,000	15,000	37,852
Liquid prod. recycle vol. proportion of total feed	0%	20%	20%	50%
Liquid recycle injection point	Inter-bed	Inter-bed	*Front feed*	Inter-bed
CFP oil oxygen content (dry basis), wt %	21.6%	21.6%	21.6%	21.6%
Total liquid fuel product after distillation, lb/h	899,086	719,338	715,835	445,758
Total liquid product H:C ratio	1.79	1.79	1.79	1.80
Base SRD case H_2 consumption, scf/bbl	615	615	615	615
Base SRD case WHSV assumption, h^{-1}	2	2	2	2
Chemical H_2 consumption, based on total feed, scf/bbl	1,630	1,316	1,337	834
Chemical H_2 consumption, based on fresh feed, scf/bbl	1,630	1,645	1,671	1,684
WHSV (fresh feed basis), h^{-1}	0.75	0.75	0.74	0.73
Net H_2 after PSA recovery, based on fresh feed, scf/bbl	1,721	1,736	1,770	1,784
Net H_2 based on total feed, scf/bbl	1,721	1,389	1,416	884
Hydroprocessing inlet pressure, psia	1,045	1,045	1,045	1,045
Hydroprocessing inlet temperature, °F	644	644	644	644
Hydroprocessing outlet temperature, °F	684	684	684	684
Hydroprocessing carbon efficiency for CFP oil	90.00%	90.00%	90.00%	90.01%
Hydroprocessing carbon efficiency for pet. feed	98.65%	98.60%	98.26%	98.47%
Makeup H_2 purity, mol %	99%	99%	99%	99%
H_2 purity at reactor inlet, mol %	80%	80%	80%	80%
Reactor H_2 feed, based on total feed, scf/bbl	11,066	6,448	8,626	3,170
Reactor H_2 feed, based on fresh feed, scf/bbl	11,066	8,060	10,782	6,399
Makeup H_2, based on total feed, scf/bbl	2,004	1,572	1,628	940
Makeup H_2, based on fresh feed, scf/bbl	2,004	1,965	2,035	1,897
Net H_2 after PSA recovery, based on total feed, scf/bbl	1,721	1,389	1,416	884
FCI, million $	317.5	275.7	285.6	211.8
MFSP, with full capital costs, per GGE	$3.02	$3.03	$3.05	$3.04
Fuel life cycle CI, g CO_2e/MJ	76.1	76.0	76.2	76.1

For repurposing existing units to handle different proportions of new feedstocks, an L-TC configuration may be advantageous for the automatic adjustment of the system to a certain degree while maintaining flows or deviating gradually from the base liquid and gas flows within

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the boundaries of the original design. However, some expenditure toward retrofitting and capacity augmentation of equipment will be unavoidable, and the extent of changes will be dictated by the repurposing goals. If inter-bed H_2 quench is also introduced with an L-TC approach (not presented here), the additional cooling from H_2 can help reduce liquid recycle and hence also reduce the fresh feed capacity curtailment.

3.4.4.3 Distribution of Front-Feed and Inter-Bed Quench H_2 for Temperature Control

A previous assumption carried forward from the 2015 design report [18] was that one-third of the hydrogen supply to the reactor is fed to the front end of the reactor and remaining is supplied at intermediate points to cool the reactor; this assumption was applied to the cases shown in Section 3.4.3 and other cases explored in this report. The ratio was chosen in the previous conceptual model [18] because cold hydrogen was used to control the significant exothermicity of 100% CFP oil hydroprocessing by providing required cooling; as a result, the supply of hydrogen usually exceeded the requirement for reactions in this configuration (H-TC without liquid recycle). Thus, having a higher proportion of hydrogen for inter-bed quench can reduce the total hydrogen feed when reactor cooling is the constraint. Note that considerations such as catalyst performance, reactivity maintenance over time, and coking will dictate the required reaction hydrogen and hydrogen feed to the front end of the reactor.

Exothermicity during hydroprocessing differs for different petroleum feedstocks, hence this ratio (feed/quench H_2) likely varies for different designs, but this information may be proprietary and vary for different technology providers. As a sensitivity case we show the impact of a changing the 33:67 front-feed to quench hydrogen ratio to 60:40; the 60:40 ratio may be more in line with some petroleum applications [97]. This new ratio was chosen for illustration, noting that other ratios may be more relevant for specific installations.

Figure 19 and Table 40 shows the advantage of a higher proportion of inter-bed H_2 quench for accomplishing temperature control. There is a significantly higher H_2 flow rate requirement to the reactor in the 60:40 case compared to the 33:67 case for the same reaction heat removal. However, before using the split ratio for more optimal temperature control, an overriding primary factor for determining this split is to ensure that the front end of the reactor is not starved of H_2 due to a low H_2 flow with the feed, which will impact long-term catalyst maintenance. This design parameter may be considered for adjustment when new units are designed for feedstocks with different hydroprocessing exothermicity compared to petroleum feedstocks.

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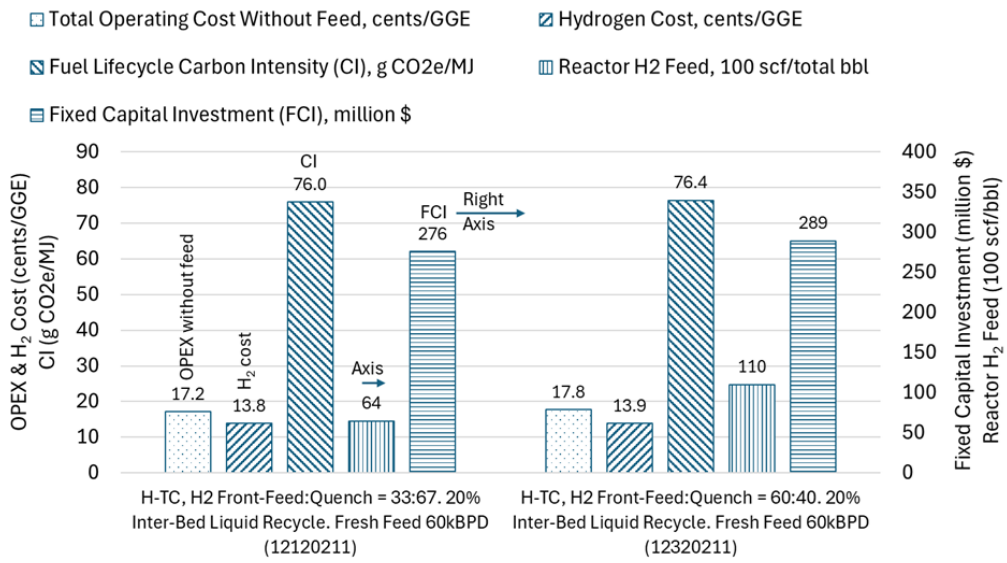

Figure 19. Impact of the distribution of front-feed H_2 and inter-bed quench H_2 for temperature control. All cases with 20 vol % CFP-A oil with SRD, 99% pure makeup H_2, and 80% pure H_2 feed.

Table 40. H_2 Front-End Feed-to-Quench Ratio Sensitivity With 33:67 (Case 12120211) and 60:40 (Case 12320211). Further Details Provided in Appendix K.

Case ID	12120211	12320211
Total feed to reactor, BPD	75,000	75,000
SRD feed at $105.88/barrel, BPD	48,000	48,000
CFP oil feed at $215.62/barrel, BPD	12,000	12,000
Liquid product recycle (inter-bed liquid feed), BPD	15,000	15,000
CFP oil oxygen content (dry basis), wt %	21.6%	21.6%
Hydroprocessing inlet temperature, °F	644	644
Hydroprocessing outlet temperature, °F	684.1	684.3
Hydrogen With Front-End Liquid Feed-to-Quench Ratio	***33:67***	***60:40***
Makeup H_2 purity, mol %	99%	99%
H_2 purity at reactor inlet, mol %	80%	80%
Reactor H_2 feed, based on total feed, scf/bbl	6,448	11,014
Reactor H_2 feed, based on fresh feed, scf/bbl	8,060	13,767
Makeup H_2, based on total feed, scf/bbl	1,572	1,619
Makeup H_2, based on fresh feed, scf/bbl	1,965	2,024
Net H_2 after PSA recovery, based on total feed, scf/bbl	1,389	1,396
Net H_2 based on fresh feed, scf/bbl	1,736	1,746
FCI, million $	275.7	289.2
MFSP, with full capital costs, per GGE	$3.03	$3.04
Fuel life cycle CI, g CO_2e/MJ	76.0	76.4

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3.4.4.4 Purity of H_2 in Reactor Feed Gas

One of the design specifications in the process model controls the purity of hydrogen in the gas feed into the hydroprocessing reactors. The base assumption is to maintain an 80 mol % H_2 purity in the reactor gas feed, constituting recycled treat gas after some purge and makeup H_2. The purge helps prevent the accumulation of gases produced during hydroprocessing. Lowering the purity of H_2 can be advantageous from a process standpoint, as shown below. However, one needs to first ensure that reactor performance is still acceptable after this change; lowering the hydrogen purity also lowers hydrogen partial pressure, thus affecting reactor performance and catalyst maintenance, and the threshold for hydrogen purity should thus be determined by those criteria. Maintaining purity close to the lowest allowable threshold may be advantageous by helping lower treat gas purge and thus hydrogen consumption. Net H_2 after PSA recovery of H_2 from the purge gases is also reported in Table 41; 85% H_2 recovery is assumed in the PSA. Note that the H_2 purity of supplied makeup hydrogen is specified separately in the model. As a reminder, the modeled reactor performance is not predictive with respect to H_2 concentration, and the same conversion/yield performance is maintained across all the cases shown in Table 41 and Figure 20. H_2 starvation is likely avoided in the modeled cases, based on the H_2 to feed ratios used in the models, as discussed in Section 3.4.3; the H_2 partial pressure (concentrations) for satisfying reaction needs should be determined and verified experimentally for each situation.

Table 41. Impact of H_2 Purity in Reactor Feed Gas. Further Details Provided in Appendix K.

Case ID	12210811	12210821	12210851	12220011	12220021
Temperature control	L-TC	L-TC	L-TC	H-TC	H-TC
H_2 purity at reactor inlet, mol %	80%	75%	70%	80%	75%
Total feed to reactor, BPD	75,000	75,000	75,000	75,000	75,000
SRD feed at $105.88/barrel, BPD	40,803	40,802	40,604	60,000	60,000
CFP oil feed at $215.62/barrel, BPD	4,534	4,534	4,512	15,000	15,000
CFP oil feed volumetric proportion of fresh feed	10%	10%	10%	20%	20%
Liquid product recycle, BPD	29,664	29,665	29,884	0	0
Liquid product recycle volumetric proportion of total feed	40%	40%	40%	0%	0%
CFP oil oxygen content (dry basis), wt %	21.6%	21.6%	21.6%	21.6%	21.6%
Total liquid fuel product, lb/h	543,297	543,323	540,672	897,075	896,945
Total liquid product H:C ratio	1.80	1.80	1.80	1.79	1.79
Base SRD case H_2 consumption, scf/bbl	615	615	615	615	615
Base SRD case WHSV assumption, h^{-1}	2	2	2	2	2
Chem. H_2 based on total feed, scf/bbl	703	703	699	1,622	1,626
Hydroprocessing inlet pressure, psia	1,045	1,045	1,045	1,045	1,045
Hydroprocessing inlet temperature, °F	644	644	644	644	644
Hydroprocessing outlet temperature, °F	684	684	684	684	685
WHSV (fresh feed basis), h^{-1}	1.06	1.06	1.06	0.76	0.76
Hydroprocessing carbon efficiency for CFP oil	89.99%	90.00%	90.00%	90.00%	90.00%
Hydroprocessing carbon efficiency for petroleum feed	98.26%	98.32%	98.32%	98.12%	98.27%

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Case ID	12210811	12210821	12210851	12220011	12220021
Makeup H_2 purity, mol %	85%	85%	85%	85%	85%
Chemical H_2 consumption, based on total feed, scf/bbl	703	703	699	1,622	1,626
Chemical H_2 consumption, based on fresh feed, scf/bbl	1,163	1,164	1,161	1,622	1,626
Reactor H_2 feed, based on total feed, scf/bbl	2,681	2,682	2,663	10,266	9,340
Reactor H_2 feed, based on fresh feed, scf/bbl	4,435	4,437	4,427	10,266	9,340
Makeup H_2, based on total feed, scf/bbl	1,272	985	858	4,005	2,840
Makeup H_2, based on fresh feed, scf/bbl	2,104	1,630	1,427	4,005	2,840
Net H_2 after PSA recovery, based on total feed, scf/bbl	826	781	756	2,020	1,847
FCI, million $	225.9	218.0	214.4	372.6	349.6
MFSP, with full capital costs, per GGE	$2.76	$2.75	$2.75	$3.07	$3.05
Fuel life cycle CI, g CO_2e/MJ	84.1	83.8	83.6	77.7	76.8

Note that the L-TC cases include 10 vol % and H-TC cases include 20 vol % CFP-A oil feed with SRD.

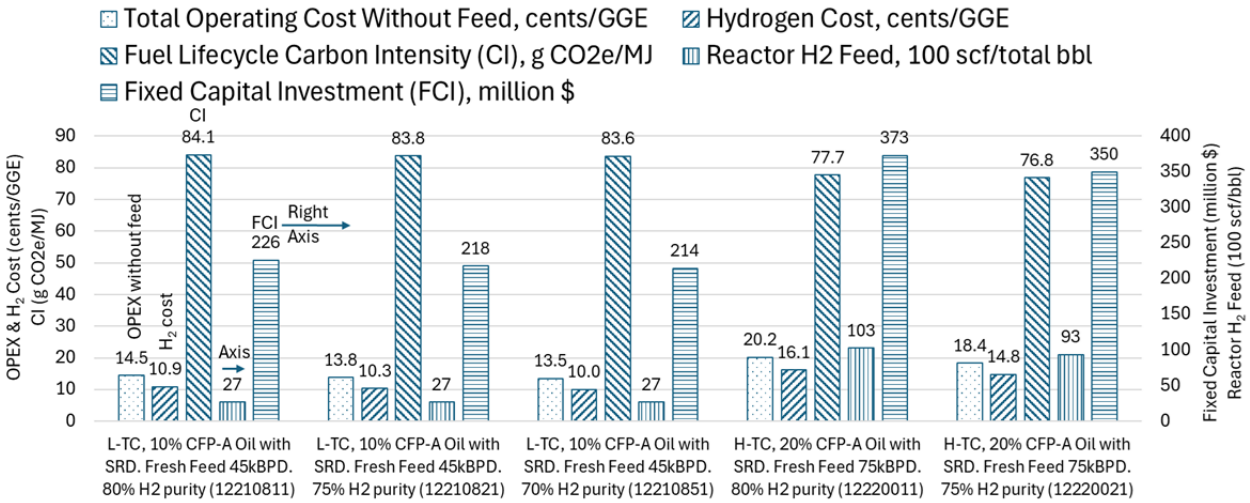

Figure 20. Process impacts of H_2 purity in reactor feed gas for L-TC and H-TC cases. L-TC cases with 10 vol % and H-TC 20 vol % CFP-A oil feed with SRD. Makeup H_2 purity 85 mol % in all cases.

3.4.4.5 Makeup Hydrogen Purity

Directionally, lower treat gas purge is necessary with higher purity H_2 makeup because less impurities are added from H_2 makeup supply impurities. The cases shown in Table 42 are with H-TC. Note that more cooling is accomplished in the model with lower purity H_2 supply because the composite impurities (see Table 30) in the 85% pure H_2 supply have a higher average molecular weight compared to the circulating off-gases from hydroprocessing (hence higher specific heat on a molar basis). Capital costs are higher for the 85% pure H_2 case because of the more gas flow and higher cost of gas processing/supply equipment such as larger compressors and PSA; our assumption that the reactor cost is scaled using the mass flow into the reactor (impacted by the molecular weight of gas mentioned above) also adds to the cost. Note that the

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impurities in the 85% pure H_2 are assumed to flow through the process and are not included either for coproduct value or for CI credit. Results are shown in Figure 21 and Table 42. The benefit from higher purity H_2 makeup will be more pronounced in cases with higher H_2 demand.

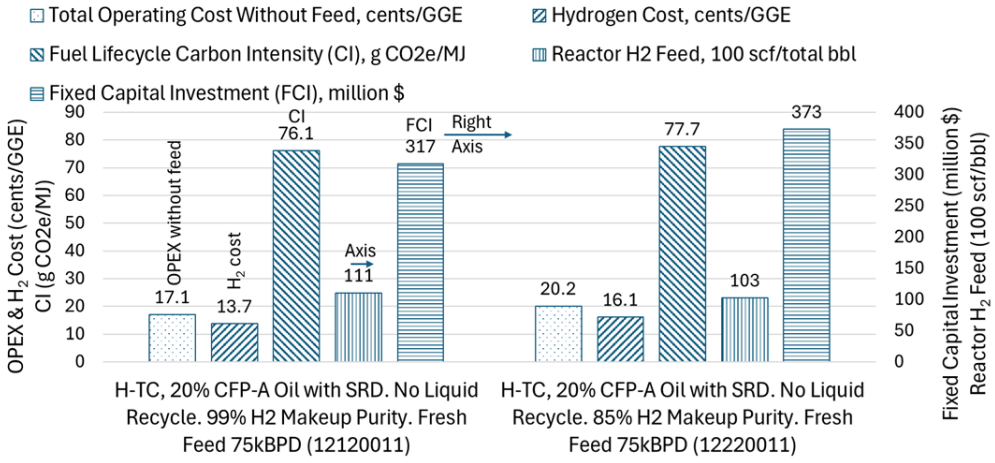

Figure 21. Process impacts of makeup hydrogen purity. Makeup purity of 99 mol % and 85 mol % in the first (12120011) and second (12220011) cases.

Table 42. Impact of Makeup H_2 Purity. Further Details Provided in Appendix K.

Case ID	12120011	12220011
Total feed to reactor, BPD	75,000	75,000
SRD feed at $105.88/barrel, BPD	60,000	60,000
CFP oil (21.6% O) feed at $215.62/barrel, BPD	15,000	15,000
Total liquid product H:C ratio	1.79	1.79
WHSV (fresh feed basis), h^{-1}	0.75	0.76
Makeup H_2 purity, mol %	99%	85%
H_2 purity at reactor inlet, mol %	80%	80%
Chemical H_2 consumption, based on total feed, scf/bbl	1,630	1,622
Reactor H_2 feed, based on total feed, scf/bbl	11,066	10,266
Makeup H_2, based on total feed, scf/bbl	2,004	4,005
Net H_2 after PSA recovery, based on total feed, scf/bbl	1,721	2,020
FCI, million $	317.5	372.6
MFSP, with full capital costs, per GGE	$3.02	$3.07
Fuel life cycle CI, g CO_2e/MJ	76.1	77.7

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3.4.4.6 CFP Oil Oxygen Content

Table 43 and Figure 22 show results for co-hydrotreating SRD with the two CFP oils from Section 2, CFP-A (21.6% O) and CFP-B (16.0%) in H-TC and L-TC configurations.

Table 43. Impact of CFP Oil Oxygen Content. Further Details Provided in Appendix K.

Case ID	12120211	11120211	12120811	11120811
CFP oil oxygen content (dry basis), wt %	21.6%	16.0%	21.6%	16.0%
Temperature control	H-TC	H-TC	L-TC	L-TC
Total feed to reactor, BPD	75,000	75,000	75,000	75,000
SRD feed at $105.88/barrel, BPD	48,000	48,000	29,718	34,643
CFP oil, $215.62 (21.6%O), $148.30 (16.0%O), BPD	12,000	12,000	7,430	8,661
CFP oil feed volumetric proportion of fresh feed	20%	20%	20%	20%
Liquid product recycle, BPD	15,000	15,000	37,852	31,696
Liquid prod. recycle vol. proportion of total feed	20%	20%	50%	42%
Total liquid fuel product, lb/h	719,338	717,264	445,758	518,038
Total liquid product H:C ratio	1.79	1.80	1.80	1.80
Base SRD case H_2 consumption, scf/bbl	615	615	615	615
Base SRD case WHSV assumption, h^{-1}	2	2	2	2
Hydroprocessing inlet pressure, psia	1,045	1,045	1,045	1,045
Hydroprocessing inlet temperature, °F	644	644	644	644
Hydroprocessing outlet temperature, °F	684	684	684	684
WHSV (fresh feed basis), h^{-1}	0.75	0.93	0.73	0.92
Hydroprocessing carbon efficiency for CFP oil	90.00%	90.00%	90.01%	89.99%
Hydroprocessing carbon efficiency for petroleum feed	98.60%	98.61%	98.47%	98.52%
Makeup H_2 purity, mol %	99%	99%	99%	99%
H_2 purity at reactor inlet, mol %	80%	80%	80%	80%
Chem. H_2 consumption, based on total feed, scf/bbl	1,316	1,056	834	776
Chem. H_2 consumption, based on fresh feed, scf/bbl	1,645	1,320	1,684	1,344
Reactor H_2 feed, based on total feed, scf/bbl	6,448	3,780	3,170	2,926
Reactor H_2 feed, based on fresh feed, scf/bbl	8,060	4,725	6,399	5,068
Makeup H_2, based on total feed, scf/bbl	1,572	1,153	940	824
Makeup H_2, based on fresh feed, scf/bbl	1,965	1,441	1,897	1,427
Net H_2 based on fresh feed, scf/bbl	1,736	1,383	1,784	1,418
Net H_2 after PSA recovery, based on total feed, scf/bbl	1,389	1,106	884	819
FCI, million $	275.7	246.6	211.8	209.3
Net hydrogen makeup cost, per GGE	$0.14	$0.11	$0.14	$0.11
MFSP, with full capital costs, per GGE	$3.03	$2.70	$3.04	$2.71
Fuel life cycle CI, g CO_2e/MJ	76.0	75.1	76.1	75.3

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All reactor assumptions include 90% carbon efficiency for CFP oil conversion to hydrocarbon fuel; the impact of reduced carbon efficiency is considered later in a separate sensitivity case. Note that the 90% efficiency assumption may not be valid for poorer quality CFP oils with higher oxygen contents and the presence of associated reactive oxygenated functional groups like sugars, acids, and aldehydes; at some point the poorer quality of CFP oils with higher oxygen contents will render it un-processible via direct conventional hydroprocessing and will require separate stabilization steps. The ~90% hydroprocessing carbon efficiency was achieved experimentally for the CFP-A and CFP-B oils.

As expected, processing the lower oxygen content CFP oil (CFP-B) requires less hydrogen, has higher WHSV, lower capital and operating costs, a slightly lower CI because of less hydrogen use. Note that an additional major contributor to the MFSP is the cost of CFP oil, which is also lower for CFP-B (Table 43).

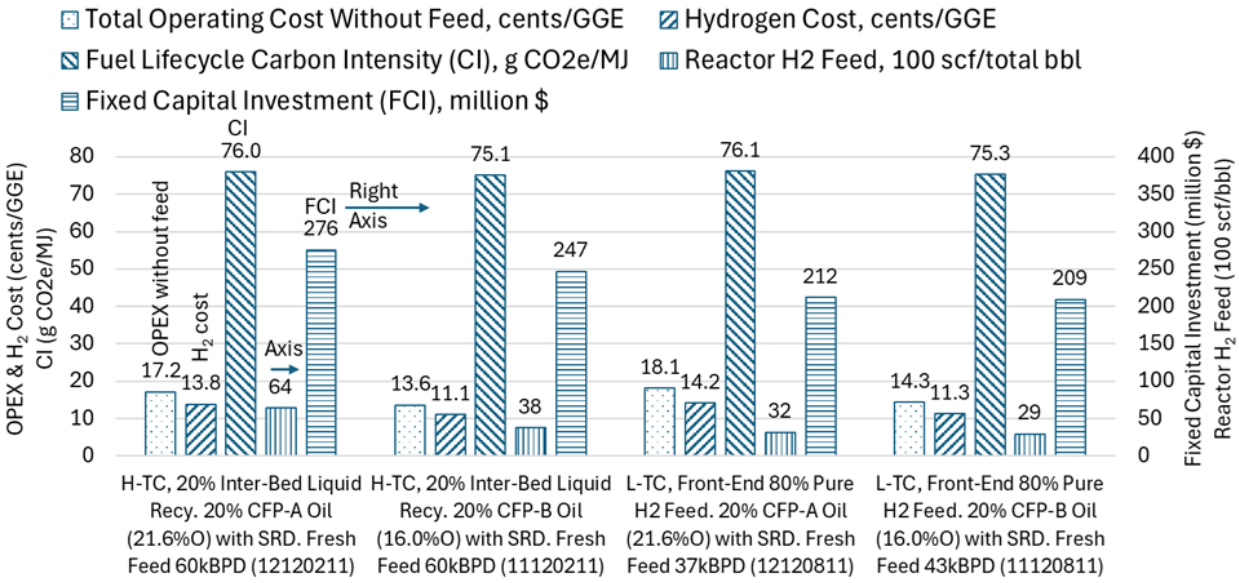

Figure 22. Impacts of CFP oil oxygen content in H-TC and L-TC setups. All the cases include 20 vol % CFP oil with 80 vol % SRD feed.

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3.4.4.7 Petroleum Feed Type

Figure 23 and Table 44 show results for different petroleum feed types, SRD and 50/50 LCO/VGO in combination with 20 vol % CFP-A and CFP-B oils. LCO/VGO have lower H:C ratios (Table 29), hence require more hydrogen and higher severity during hydroprocessing.

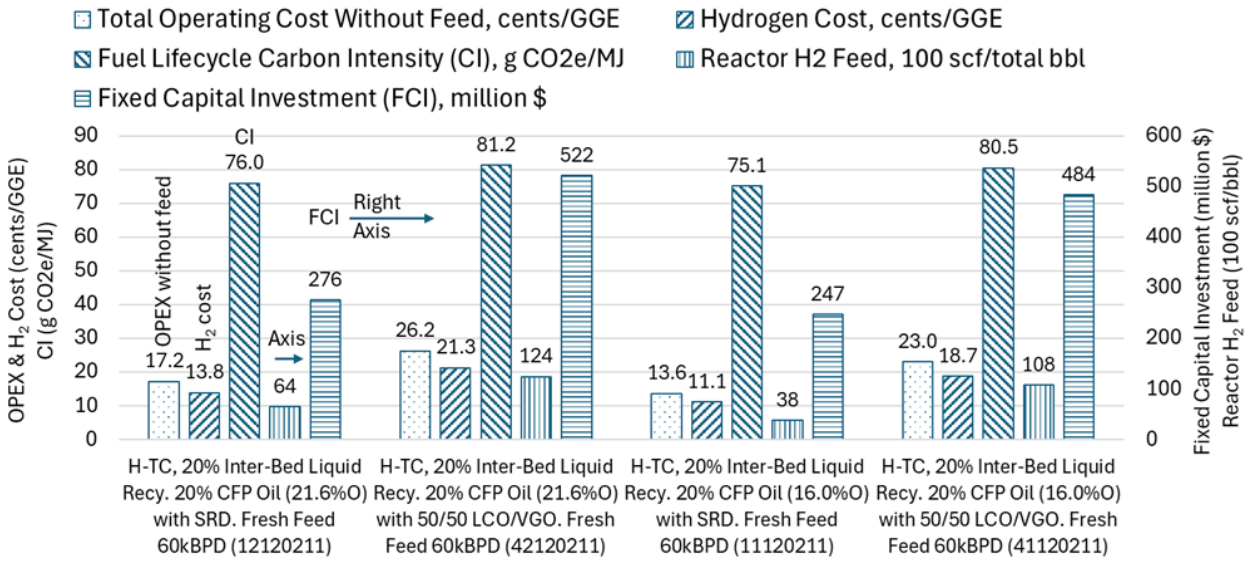

Figure 23. Impacts of processing petroleum feed types, SRD and LCO/VGO, with 20 vol % CFP-A (21.6% O) and CFP-B (16.0% O) oils in H-TC setup with 20 vol % liquid product recycle.

The SRD cases are based on the hydrotreating configuration (Figure 11), while the LCO/VGO cases are based on the HT+HC configuration (Figure 12). Note again, for ease of comparison we assumed that the product H:C molar ratio is ~1.8 in all cases. Table 44 provides relevant metrics, and as noted earlier, it is assumed that the catalyst in the HT+HC configuration is spread across two reactors.

Table 44. Impact of Petroleum Feed Types. Further Details Provided in Appendix K.

Case ID	12120211	42120211	11120211	41120211
Total feed to reactor, BPD	75,000	75,000	75,000	75,000
SRD feed at $105.88/barrel	48,000	0	48,000	0
LCO feed at $102.95/barrel	0	24,000	0	24,000
VGO feed at $100.78/barrel	0	24,000	0	24,000
CFP oil feed (CFP-A $148.3, CFP-B $215.62/bbl), BPD	12,000	12,000	12,000	12,000
CFP oil feed volumetric proportion of fresh feed	0.2	0.2	0.2	0.2
Liquid product recycle, BPD	15,000	15,000	15,000	15,000
Liquid product recycle vol. proportion of total feed	0.2	0.2	0.2	0.2
CFP oil oxygen content (dry weight basis)	21.6%	21.6%	16.0%	16.0%
Total liquid fuel product, lb/h	719,338	788,895	717,264	787,456
Total liquid product H:C ratio	1.79	1.80	1.80	1.80
Base SRD case H_2 consumption, scf/bbl	615	615	615	615
Base SRD case WHSV assumption, h^{-1}	2	2	2	2
Hydroprocessing inlet pressure, psia	1,045	2,045	1,045	2,045

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Case ID	12120211	42120211	11120211	41120211
Hydroprocessing inlet temperature, °F	644	644	644	644
Hydroprocessing outlet temperature, °F	684	685	684	684
Hydrotreating				
WHSV (fresh feed basis), h^{-1}	0.75	0.88	0.93	1.00
Catalyst load (with 20% excess), lb	1,268,916	1,180,871	995,793	1,014,088
Catalyst replacement period, years	2	2	2	2
Catalyst unit cost, $/lb	16.6	16.6	16.6	16.6
Hydrocracking				
WHSV (fresh feed basis), h^{-1}	N/A	0.88	N/A	1.00
Catalyst load (with 20% excess), lb	N/A	1,180,871	N/A	1,014,088
Catalyst replacement period, years	N/A	3	N/A	3
Catalyst unit cost, $/lb	N/A	16.6	N/A	16.6
Catalyst cost per feed barrel, $/bbl	0.52	0.80	0.41	0.69
Net H_2 after PSA recovery, based on fresh feed, scf/bbl	1,736	2,927	1,383	2,575
Net H_2 based on total feed, scf/bbl	1,389	2,342	1,106	2,060
FCI, million $	275.7	521.6	246.6	483.6
MFSP, with full capital costs ($/GGE)	3.03	2.83	2.70	2.53
Fuel life cycle CI, g CO_2e/MJ	76.0	81.2	75.1	80.5

3.4.4.8 CFP Oil to Hydrocarbon Fuels Carbon Efficiency and Product H:C Ratio

This section presents cases with different assumptions for CFP oil to fuel carbon efficiency and the hydrogen-to-carbon ratio in the product fuel.

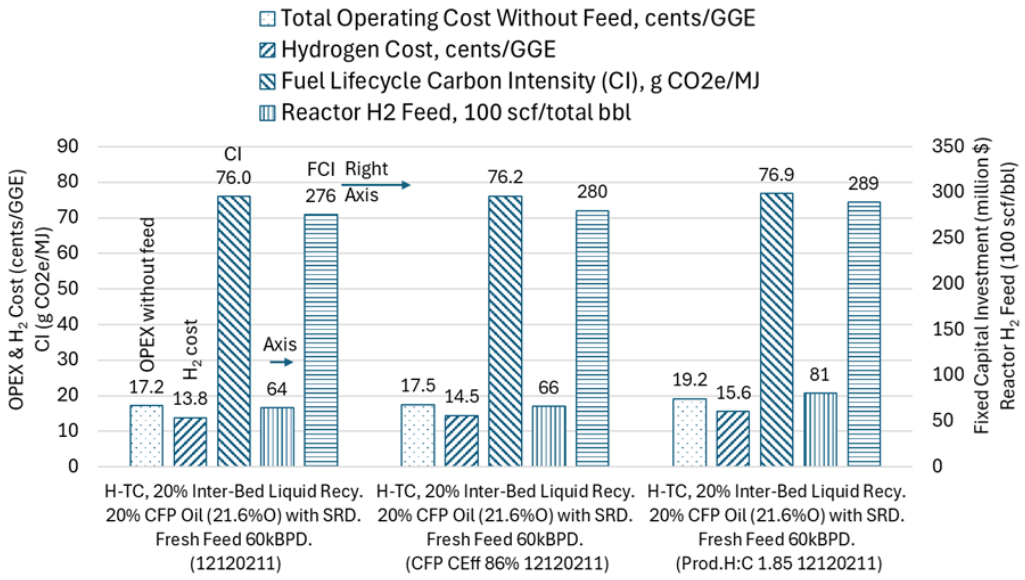

Figure 24. Impacts of CFP oil conversion carbon efficiency and H:C ratio of the fuel products.

Results for these sensitivity cases are shown in Figure 24 and Table 45.

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Table 45. Impacts of CFP Oil Carbon Efficiency to Fuel and H:C Ratio of Fuel Products. Further Details Provided in Appendix K.

Case ID	12120211	CFP Carbon Efficiency 86% 12120211	Prod. H:C 1.85 12120211
CFP oil oxygen content (dry basis), wt %	21.60%	21.60%	21.60%
Temperature control	H-TC	H-TC	L-TC
Total feed to reactor, BPD	75,000	75,000	75,000
SRD feed at $105.88/barrel, BPD	48,000	48,000	48,000
CFP oil, $215.62 (21.6%O), $148.30 (16.0%O), BPD	12,000	12,000	12,000
CFP oil feed volumetric proportion of fresh feed	20%	20%	20%
Liquid product recycle, BPD	15,000	15,000	15,000
Liquid prod. recycle vol. proportion of total feed	20%	20%	20%
Total liquid fuel product, lb/h	719,338	712,309	722,908
Total liquid product H:C ratio	1.79	1.79	1.85
Base SRD case H_2 consumption, scf/bbl	615	615	615
Base SRD case WHSV assumption, h^{-1}	2	2	2
Hydroprocessing inlet pressure, psia	1,045	1,045	1,045
Hydroprocessing inlet temperature, °F	644	644	644
Hydroprocessing outlet temperature, °F	684	684	684
WHSV (fresh feed basis), h^{-1}	0.75	0.73	0.66
Hydroprocessing carbon efficiency for CFP oil	90.00%	86.01%	89.99%
Hydroprocessing carbon efficiency for petroleum feed	98.60%	98.55%	98.59%
Makeup H_2 purity, mol %	99%	99%	99%
H_2 purity at reactor inlet, mol %	80%	80%	80%
Chem. H_2 consumption, based on total feed, scf/bbl	1,316	1,356	1,501
Chem. H_2 consumption, based on fresh feed, scf/bbl	1,645	1,695	1,877
Reactor H_2 feed, based on total feed, scf/bbl	6,448	6,609	8,051
Reactor H_2 feed, based on fresh feed, scf/bbl	8,060	8,261	10,064
Makeup H_2, based on total feed, scf/bbl	1,572	1,680	1,768
Makeup H_2, based on fresh feed, scf/bbl	1,965	2,099	2,210
Net H_2 after PSA recovery, based on total feed, scf/bbl	1,389	1,439	1,577
FCI, million $	275.7	280.1	289.3
Net hydrogen makeup cost, per GGE	$0.138	$0.145	$0.156
MFSP, with full capital costs, per GGE	$3.03	$3.06	$3.02
Fuel life cycle CI, g CO_2e/MJ	76.0	76.2	76.9

As expected, lower carbon efficiency negatively impacts all metrics because of lower liquid fuel production. Figure 24 and Table 45 illustrate this with a change in the carbon efficiency from 90% in the first case to 86% in the second case. This is a point sensitivity using a specific set of

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assumptions (H-TC with 20 vol % CFP-A oil feed with 20 vol % liquid product recycle); the negative impacts of CFP oil hydroprocessing carbon efficiency on the metrics will be more pronounced with an increase in the proportion of CFP oil in the feed.

The third case illustrates the effect of using the same reactor setup to increase the liquid product molar H:C ratio from ~1.8 (first case) to 1.85, with the hydrogen price assumption of $1.5/kg (Table 30). The energy value of the fuel produced increases and that improves the economics, even though processing costs and hydrogen consumption increase. Correspondingly, the CI, driven primarily by hydrogen consumption, also increases when H:C ratio is increased. Note that the modeled WHSV is also decreased (per our methodology discussed in Section 3.2.7) for the higher hydrogen consumption in the 1.85 H:C ratio case.

3.4.4.9 Reactor Temperature and ΔT

Figure 25 shows that increasing both the reactor temperature and the allowable reactor temperature increase (ΔT) help reduce the gas feed into the reactor, primarily because of lower gas flow requirements for cooling the reactor in a H-TC setup. Note that the higher reactor temperature case benefits from our assumption that the inter-bed H_2 quench streams are at a constant temperature of 43.3°C (110°F); thus, additional sensible heat can be absorbed into the gas feed when the reactor is operating at a higher temperature. Although not shown, the L-TC case will also benefit for the same reason, thus requiring a smaller liquid product recycle to achieve the same amount of cooling; this will help increase the fresh feed throughput. In Figure 25, the slightly lower FCI is a result of lower gas circulation and gas feed into the reactor in the H-TC setup. Note that the overall H_2 consumption is roughly the same in all the cases, after accounting for H_2 recovered from the purge gas via PSA.

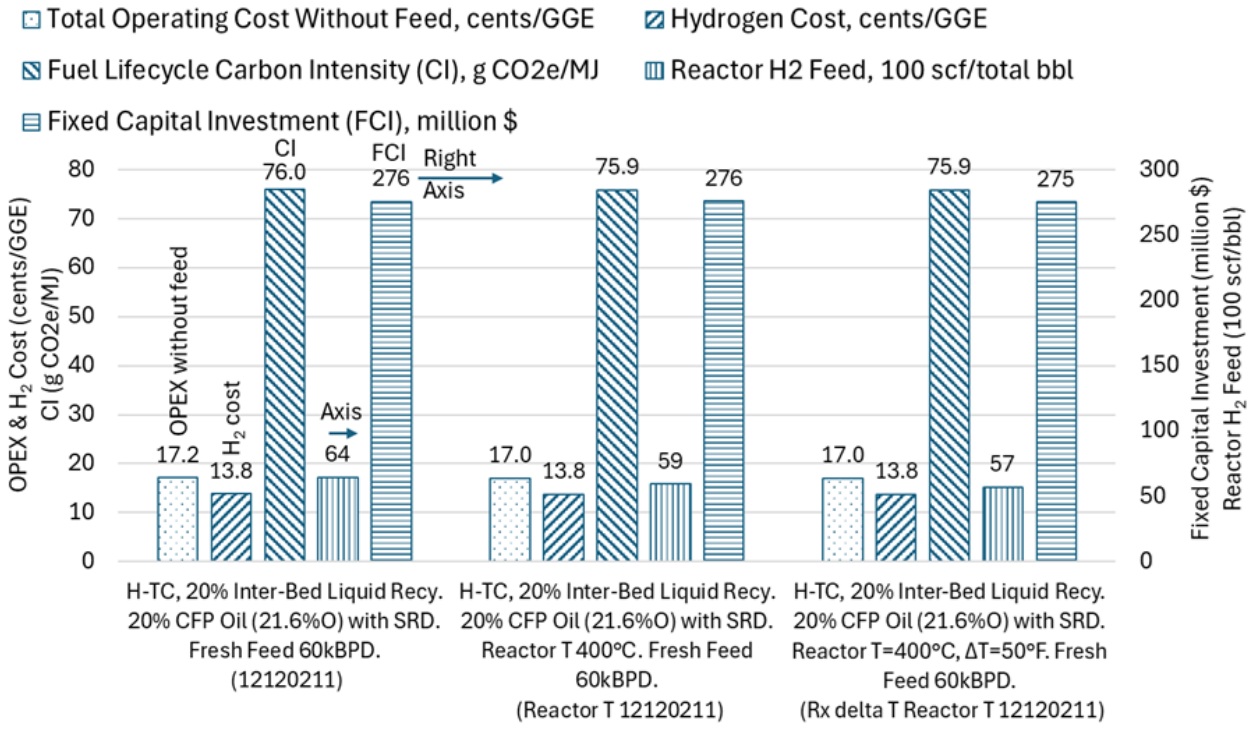

Figure 25. Impacts of reactor temperature and ΔT

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3.4.4.10 Reactor Pressure

The choice of reactor pressure in a hydroprocessing unit is determined by the feedstock and the design requirements for processing target feedstocks to desired product cuts and quality. Figure 26 shows that decreasing reactor pressure can reduce capital costs along with other costs such as gas compression costs.

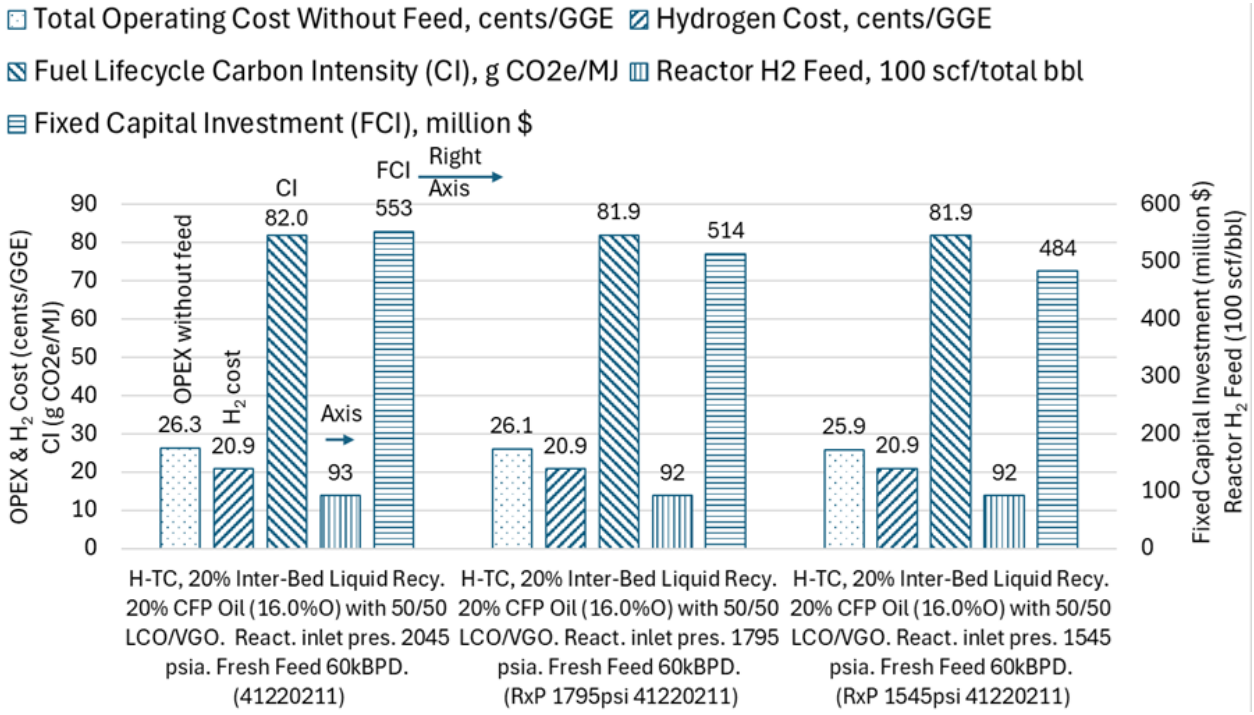

Figure 26. Sensitivity to reactor pressure in a HT+HC configuration with H-TC, 20 vol % CFP-B oil (16.0 wt % O) with 50/50 LCO/VGO and 20 vol % liquid product recycle.

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3.4.4.11 On-Stream Factor

Figure 27 shows the impacts of changing the 95% on-stream factor base assumption for hydroprocessing. On-stream factors affect capital and fixed cost utilizations. Hence the impact is larger at smaller scales where those costs have more dominant contributions to the MFSP.

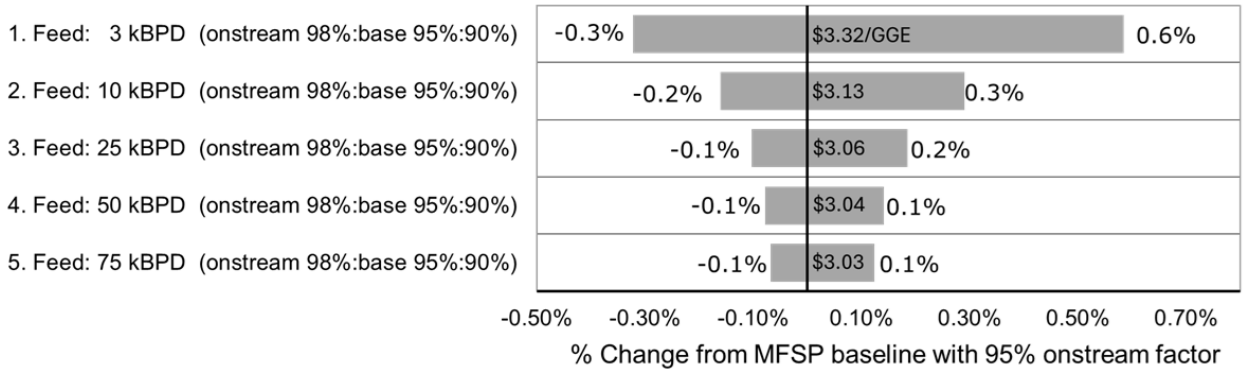

Figure 27. Sensitivity to on-stream factors at different scales for Case 12120211

Although the impact on the MFSP is not as significant in larger-scale systems based on our methods and assumptions, the on-stream factor is a high-priority consideration in those larger systems because lower on-stream factors result in lost revenue for the business, thus affecting its overall profitability. Typically, refining volumes are large, but margins are small, and lower volumes processed with the small margins in play is economically detrimental; the MFSP changes with the on-stream factor appear to be small and do not capture this impact. As mentioned previously, the refining unit has no control over the cost of the feed material once it has been chosen and has minimal control over hydrogen consumption. The onstream factor is one of the key handles the refinery unit can control and it is important to keep the unit running while meeting product cut objectives and specifications when new liquid feedstocks such as CFP oils are introduced.

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3.4.4.12 Scale, Capital Cost, and Repurposing Existing Units

Figure 28 shows the sensitivity to different scales, (a) without and (b) with 30% higher capital costs, and potential savings from repurposing existing units for coprocessing CFP oils.

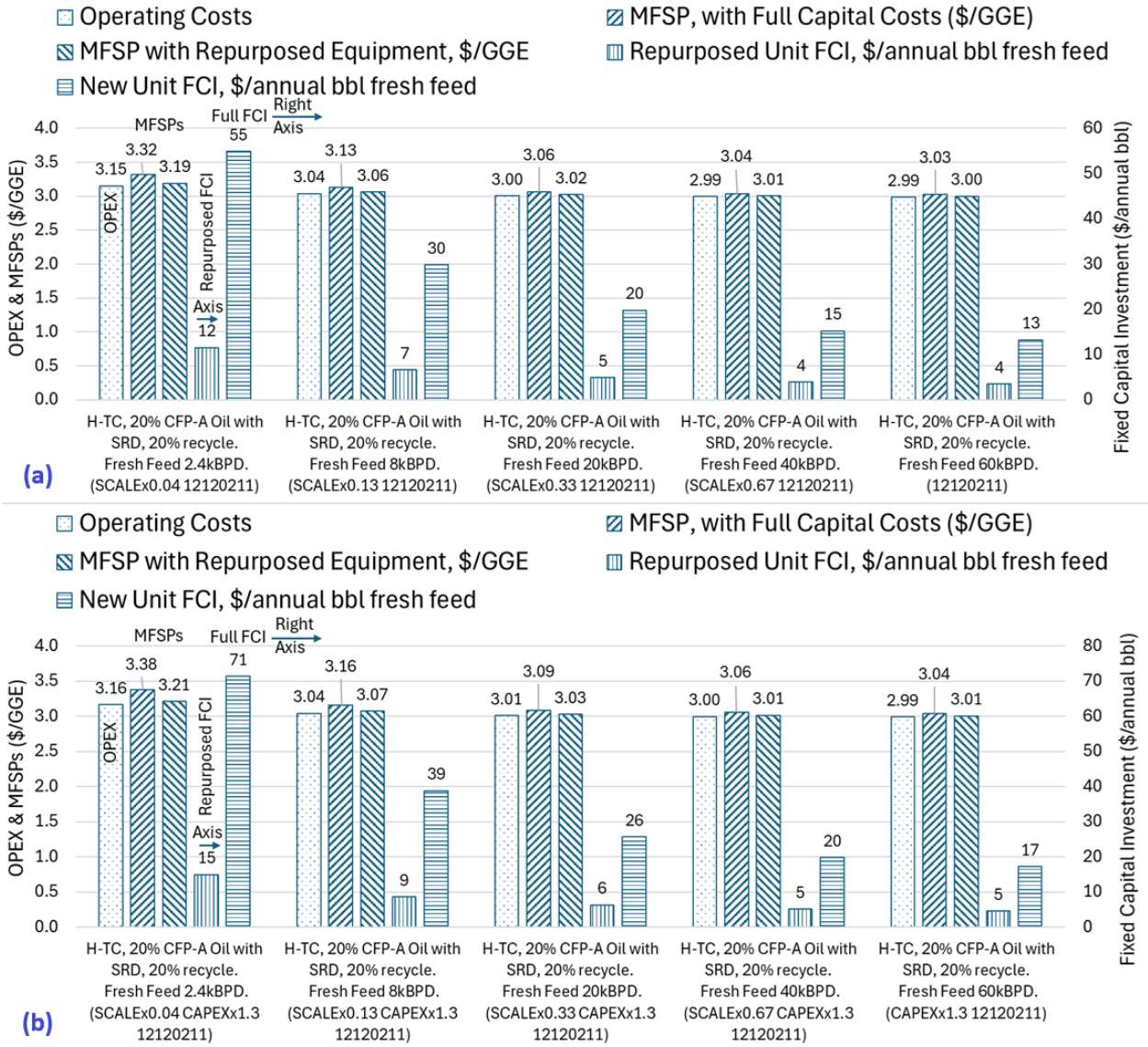

Figure 28. Impacts of (a) scale, (b) scale and 30% higher capital costs, along with estimated potential savings upon repurposing existing units. All cases are based on 12120211, which includes H-TC with 20 vol % CFP-A oil with SRD feed, and 20 vol % liquid product recycle. Base unit sizes processing 3k, 10k, 25k, 50k, 75k BPD of 100% SRD feed (without liquid recycle) were used for the five cases, respectively, to estimate savings from repurposing those units.

These sensitivity cases are around a base case with 75k BPD total reactor capacity, where 20 vol % liquid product is recycled and 60k BPD of fresh feed processed by the unit. The feed includes 20 vol % CFP-A oil (21.6% oxygen content) with SRD. Note that the relevant metrics presented in Figure 28 are different from the set of metrics reported in previous figures.

As expected, capital investment dollars per annual barrel increases significantly at smaller scales. Fixed operating costs also have a greater impact on the economics and overall operating costs increase as a result. Comparing corresponding cases between Figure 28 (a) without and (b) with

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a 30% CAPEX increase, shows that capital cost escalations have larger impacts at smaller scales. Note that based on our assumptions, the product CI is minimally impacted by scale.

Repurposing existing units results in savings in capital costs and has a more significant impact on smaller scale units. Note that capital cost savings estimates were estimated via a simple subtraction of the FCI of each equipment in the base unit that is repurposed, from the FCI of the corresponding equipment in the cases presented. Differences that are negative were ignored when deriving the repurposed FCI of each unit. The estimated FCI for repurposed units are listed in Appendix K in the row "Delta Capital for Repurposed Unit" and the base unit that was used as the basis for repurposing is included the following row, "Repurposed Case ID". Note that additional expenditure that may be necessary for making equipment workable for large turndown ratios below ranges of equipment design capacity are not included in these estimates; those can be estimated on a case-by-case basis when specific configurations of interest are identified. Feed system modifications to account for the reactivity of CFP oil will need to be included; the system discussed in Section 3.2.13 is factored into this analysis. When repurposing, capacity (volume of feed) reduction will be necessary in most situations in order to maximize direct reuse with minimal changes, since the hydrogen demand and reaction/catalyst requirements for processing the CFP oils presented in this study exceed corresponding requirements for processing incumbent petroleum feedstocks. Additional modifications for changing materials of construction in specific sections may be necessary, beyond the reactor feed system modifications included in our estimates.

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3.4.4.13 Stand-Alone CFP Oil Hydroprocessing

The first three cases in Figure 29 show results for hydrotreating 7k BPD CFP-A oil fresh feed with 3k BPD liquid product recycle (10k BPD total), with different base nameplate capacity SRD units in each case (75k, 30k, and 10k BPD, respectively) for repurposed cost estimates. The last two cases show results for the L-TC configuration with capacity curtailment to 2.2k BPD and 16.2k BPD starting from 10k BPD and 75k BPD SRD hydrotreating units. The same methods mentioned in the previous Section 3.4.4.12 are used for estimating savings for repurposed units, with roughly the capital cost for building the existing unit subtracted to determine FCI savings.

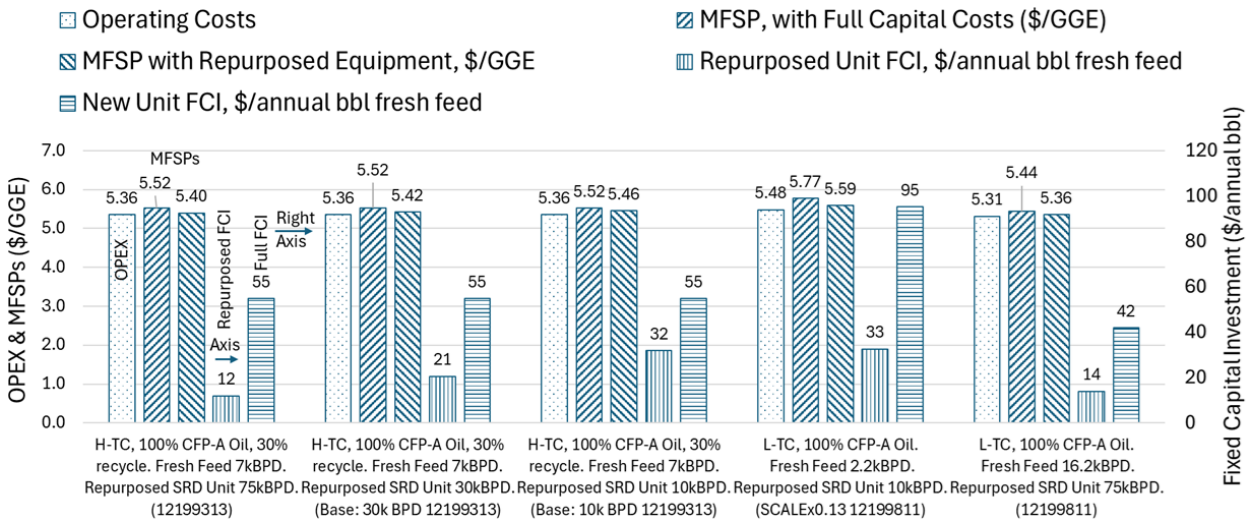

Figure 29. Stand-alone CFP-A oil (21.6 wt % O) hydroprocessing in the H-TC configuration (first three cases) with 10k BPD total reactor feed, 7k BPD fresh feed and 30 vol % product liquid recycle. Sizes of repurposed units are included in each case caption. The last two cases show the CFP-A oil feed processed in the L-TC configuration, with capacity curtailment dictated by the inter-bed liquid product recycle for temperature control; 10k BPD and 75k BPD SRD systems are shown as the base units for estimating benefits of repurposed equipment.

Matching gas and liquid flows with the requirements for new feed material will be important when making repurposing design decisions; in this case we did not emphasize that aspect, although they can be modeled when specific objectives are presented. Compatibility of sizes is important because there may be wasted capacity and capital if the base unit is too large compared to the equipment capacity requirements for processing the volume of CFP oil available; coprocessing may help better capacity utilization in such cases.

Note that 3k BPD is approximately the volume of CFP oil produced by a single 2,000 dry metric tonnes per day biomass plant (2,225 BPD of CFP-A oil listed in Table 26 and 2,870 BPD of CFP-B oil listed in Table 27). Aggregation of CFP oil may be an economical option to take advantage of larger-scale hydroprocessors using CFP oil from multiple biomass to CFP oil conversion facilities. In our models 800 miles of rail transport is included in the costs of CFP oils and the transportation impact is accounted in the estimated GHG emissions.

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3.4.4.14 WHSV and Catalyst Lifetime

As mentioned earlier (Section 3.2.7), the WHSV for each case was prorated from a base assumption for SRD hydrotreating with a WHSV of 2 h^{-1} for a modeled 615 scf/bbl chemical hydrogen consumption. Chemical hydrogen consumption in each case is used for prorating the WHSV. Since the reactor model is not predictive, changes in WHSV do not affect reactor performance in the model. The goal for this sensitivity case is to understand cost and GHG impacts with changes in the amount of catalyst necessary to achieve hydroprocessing performance in industrial units. The effect on the MFSP from changing this assumption by different factors are shown in Figure 30 using a specific case (12120211). The results shown here represent catalyst cost changes only; sensitivity to capital cost changes is discussed in Section 3.4.4.12. The results show that the contribution from catalyst costs can increase significantly for this case if a factor of less than 0.5 is applied to the WHSV estimated by the method stated above. The actual WHSV necessary will need to be determined for each feedstock based on experimental performance. Based on assumptions presented in this report, WHSV changes did not have any significant impact on the CI for this case, with the calculated CIs of 76.19 and 75.97 g CO2e/MJ for the two end points, 0.1xWHSV and 3xWHSV (in Figure 30), respectively.

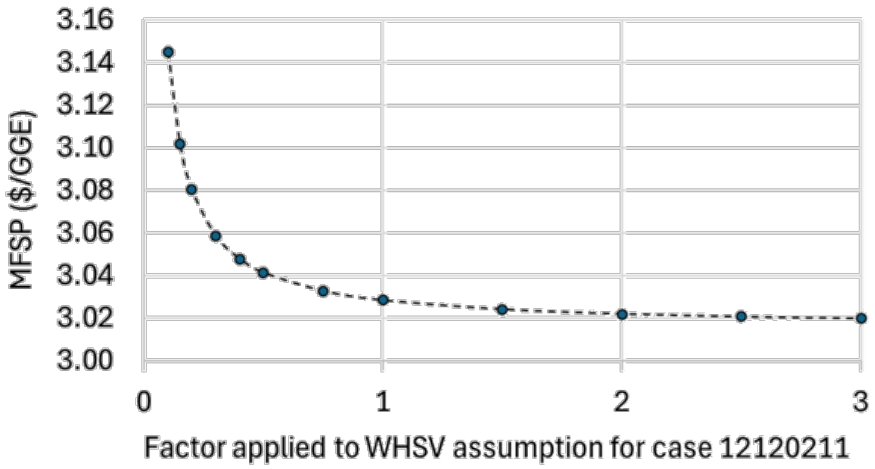

Figure 30. Impact of changing estimated WHSV in the model by different factors for a H-TC setup with 20 vol % CFP-A oil co-hydrotreated with 80 vol % SRD, with 20 vol % liquid product recycle.

A catalyst lifetime of two years was assumed, but this has not been proven through extended testing with CFP oils or mixtures of CFP oils and petroleum feedstocks. As mentioned previously, hydroprocessing CFP oils require significant deoxygenation and this increases the reaction demands on the catalyst. The impacts on the catalyst need to be assessed experimentally for specific catalysts used and CFP oils being processed; associated hydrogen partial pressure may be an important parameter for maintaining catalyst activity during the targeted catalyst lifetime. Long catalyst lifetimes that match existing refinery operations will be desirable to avoid more frequent replacements and turnarounds than the refinery's maintenance schedule. The model estimates an MFSP increase from \$3.0285/GGE to \$3.0401/GGE, and a minor CI increase from 75.98 to 76.01 g CO_2/MJ, when the hydrotreating catalyst lifetime is changed from the baseline of two years to one year for the same case 12120211 shown in Figure 30.

The magnitude of impacts of lower WHSV and catalyst lifetime assumptions on the refinery's economics are not adequately captured by the MFSP increase projections alone. Lower WHSVs

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lead to refinery throughput curtailment and lessened catalyst lifetimes require frequent catalyst changeouts, both of these reduce the annual capacity of the refinery. Refineries typically operate with high throughputs and small margins, and reduced annual capacities will have similar impacts as lower on-stream factors discussed in Section 3.4.4.11. Any curtailments that adversely impact operations and capacity utilization will negatively impact revenues and profits.

3.4.4.15 Petroleum, CFP Oil, and Hydrogen Prices

As shown in Section 3.4.3, (a) petroleum and (b) CFP oil feedstocks along with (c) hydrogen are expected to be the key contributors to the MFSP for these processes. For large-scale operations where capital and fixed operating costs have smaller impacts, the MFSP will be close to the sum of these three cost contributors. Given the dominance of these inputs, it is useful to see the effects of changing unit prices of these inputs from their base values (Table 30). Petroleum feed stream prices are tied to crude prices, the hydrogen price usually tied to natural gas price, and CFP oil prices from an established mature process will depend primarily on biomass feedstock prices. Hydrogen sourced from refinery processes may have a different pricing structure and price dependency. Figure 31 shows that the impact of different unit prices of these three input categories can easily be estimated by prorating their contributions as illustrated with 30% increases in the prices of each of these inputs using (a) H-TC (12120211) and L-TC (12120811) cases for CFP-A and (b) H-TC (11120211) and L-TC (11120811) cases for CFP-B. Note how switching from H-TC to L-TC (with lower throughput of fresh feed) has a minor impact on the absolute value of the MFSP; the small difference is primarily because of lower capital utilization in the L-TC case where fresh feed is curtailed to 37.1k BPD compared to 60k BPD in the H-TC case. As discussed in the previous section, feed curtailment will impact the refinery's profitability beyond the impact on MFSP changes.

Although contributors other than these three inputs have relatively small impacts, those are important because changes of fractions of a cent can have a big impact on the refining margins because of their large-scale operations. The prices of the three major inputs are largely determined by external factors. The refiner may determine the feed mix based on overall profitability and may have marginal ability to influence hydrogen consumption through design and operational choices. Some of the parameters explored in this study may be influenced by the refiner during both design and/or operation stages. Thus, it is important to optimize the other parameters described in this report for more optimal performance and to help improve refining margins.

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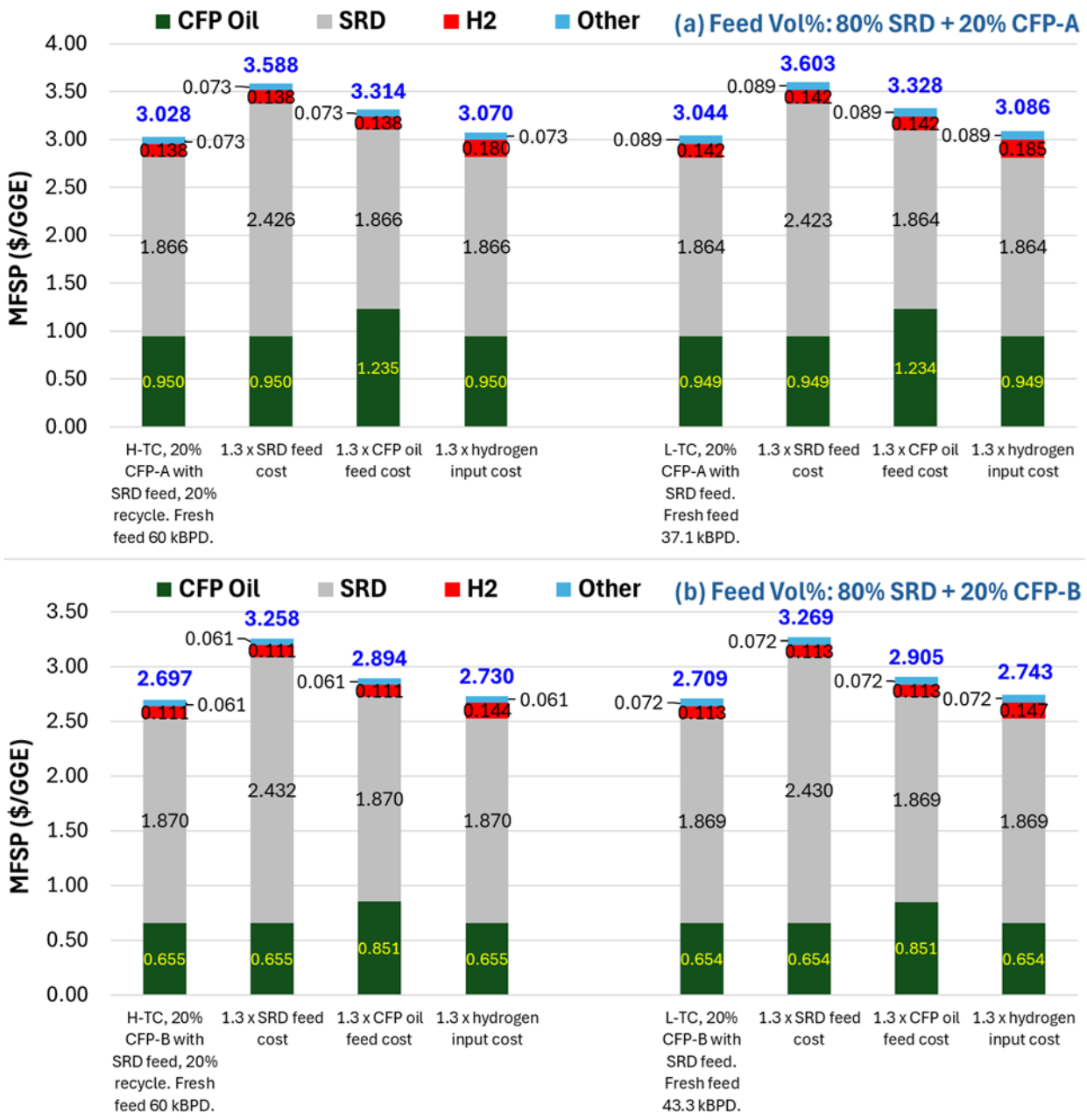

Figure 31. MFSP changes upon changing unit costs of the three major contributors (petroleum feed, CFP oil, hydrogen) by 30% (1.3x) each; these unit cost variation impacts can be estimated by simple proration. The two sets show (a) 20 vol % CFP-A oil + 80 vol % SRD feed with H-TC base case 12120211 and L-TC base case 12120811, and (b) 20 vol % CFP-B oil + 80 vol % SRD feed with H-TC base case 11120211 and L-TC base case 11120811. The very small differences among the H-TC and L-TC cases for these 3 cost contributions within each set (in the third decimal place) are due to slight differences in modeled product H:C ratios and carbon efficiencies. The total GGE output from CFP-A is slightly higher for CFP-A due to higher liquid density of the CFP-A compared to CFP-B oil.

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Section 4
Conclusion

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4 Conclusion

The results presented include modeled costs and sustainability metrics for the entire supply chain for the production of liquid hydrocarbon fuels from woody biomass via CFP conversion. Biomass production assumptions are based on detailed reports prepared by INL. Two sample CFP oil production designs presented (CFP-A and CFP-B) are based on conceptual scale-up assumptions from NREL bench-scale experimental results. Hydroprocessing assumptions are based on publicly available experimental results, including those from NREL and Pacific Northwest National Laboratory.

The analysis shows that increasing the efficiency of the supply chain for CFP oil production from biomass offers the biggest opportunities for cost reduction, while subsequent successful hydroprocessing of the CFP oil with high carbon efficiencies, leveraging existing capital to help minimize costs, can help maximize the end value for fuel producers and consumers.

The focus of this report is on the hydroprocessing step. If CFP oils can be produced at reasonable costs and significant volumes, further development and adaptation of the design and operations of hydroprocessing units can help optimize CFP oil processing. Heat management in hydroprocessing reactors is a key consideration when processing CFP oils, and careful engineering decisions need to be made to develop optimal design/operations strategies for specific implementations. Beyond feed costs for petroleum and CFP oil, hydrogen is the most substantial operating cost in hydroprocessing; this is a necessary expense because experiments have shown that the CFP oil intermediate requires adequate hydroprocessing in order to meet current fuel specifications. If CFP oil production at commercial scales results in quality variations beyond acceptable limits, then pre-stabilization operations need to be added within the scope before bulk CFP oil supplies are introduced into regular downstream hydroprocessing units. Although FCC coprocessing of CFP oils is not the subject of this report, it can serve as a low-risk option for consideration in the early stages, at the expense of lower carbon efficiencies compared to hydroprocessing, to help understand and develop refinery logistics requirements for the introduction of a new feed category of CFP oils into petroleum refineries.

Results from the hydroprocessing models are intended to provide insights into the impacts of design/operational parameters for CFP oil integration. Simplified designs were chosen for modeling with the recognition that there is significant diversity in hydroprocessing unit designs, feedstocks, and operating objectives. The parameter space can be more expansive for complicated and high-severity units. Modeling efforts are theoretical and informed by experimental data from short-duration laboratory or pilot scale runs; final yields and product quality (reflected by the H:C ratio) are fixed in a non-predictive reactor model, with sensitivity cases exploring the impact of parameter variations. Further experimental research is necessary for understanding and improving long-term reactor performance and catalyst maintenance as it is critical for successful commercial deployment. If continued technology and supply chain development contributes significantly to the availability of CFP oils, then subsequent collection of more experimental data to enable the development of predictive reactor models can greatly benefit the design and operation of future systems, and prolonged operations at incrementally larger scales can increase confidence toward successful adoption and scale-up. Opportunities exist for catalyst improvements if CFP oils, or other oxygenated liquid feedstocks, are commoditized and processed in significant volumes in the future. Other risks should be assessed carefully and mitigated, such as changing materials of construction in specific sections of

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incumbent processes to handle new liquid feedstocks. Results for CFP oil processing in stand-alone and co-hydroprocessing configurations are explored, and results are presented for different proportions of CFP oil + petroleum feed mixes, covering 100% petroleum feed to 100% CFP oil. Many of the cases estimate design/operating parameters that are beyond incumbent design ranges. Optimal use of existing equipment will be an important consideration when introducing CFP oil feedstocks in existing petroleum hydroprocessing equipment; thus, gas and liquid flow rates may need to be managed by adjusting operating parameters, such as liquid recycle rates, in order to successfully constrain operations within the equipment design parameters. An approach based on specific constraint-mitigation for an existing unit was not the focus in this project. However, the modeling framework is designed for such analysis that can be explored in future efforts. The analysis presented here may be adapted for other oxygenated liquid feedstocks that may be considered by refineries.

As with any analysis, all results should be viewed within the context of underlying assumptions and limitations. We attempted to provide sufficient details of base parameter assumptions and their variations in the sensitivity cases to allow interested readers to get approximate results by prorating for other specific sets of assumptions that may not be presented in this report. These parameters and configurations include (1) control strategies; (2) CFP oil oxygen content; (3) petroleum feedstock types; (4) different proportions of CFP oil and petroleum feed during coprocessing; (5) hydrogen purity; (6) carbon efficiencies during CFP oil conversion; (7) space velocities and catalyst lifetime; (8) H:C ratio in the total liquid product; (9) reactor temperature, pressure, and allowable temperature increase in the reactor (ΔT); (10) scale; (11) capital and key operating costs; (12) new stand-alone units; (13) repurposed units; and (14) on-stream factor.

Further development and scale-up of the front-end biomass supply chain, including production, collection, preprocessing, transportation, and storage, will be critical for improving reliability, increasing supply volumes, and reducing the cost of biomass feedstocks. Based on our assumptions in the two CFP cases presented, biomass feedstock cost constitutes approximately one-third of the total cost of CFP oil. Major challenges for biomass feedstocks include distributed production, heterogeneity, contaminants, and low energy density compared to petroleum. Improving efficiency of collection, logistics, and preprocessing will thus be key toward lowering costs. The development of a robust supply chain at scale can help standardize and bring reliability to biomass feedstock supplies and also provide economic benefits to rural areas.

Improving conversion yields of CFP oil from biomass, along with the production of valuable coproducts, will be necessary for lowering the cost of the liquid CFP oil intermediate. The production of reasonably priced CFP oil is critical for commercial success. Maintaining CFP oil quality within acceptable ranges for downstream hydroprocessing unit designs will be key for large-scale industrial acceptance of this technology. Locating CFP conversion facilities near other existing industrial infrastructure can help reduce risks, lower costs, and improve sustainability, such as through the optimization of utilities [12].

Overall, the development and extensive demonstration of diverse technologies, with associated construction and environmental permitting, is necessary for the success of the entire supply chain required to support this conversion pathway. Multiple sectors must devise efficient custody transfer while ensuring quality and reliability. Key handoff points will include commoditized

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biomass supplied to the CFP conversion plant and standardized CFP oils supplied to hydroprocessing units. Many of the technologies throughout the supply chain are evolving, and it will be important to consolidate learnings from previous setbacks and develop best practices for scale-up in order to reduce implementation risks and facilitate success.

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Appendices

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Appendix A. CFP Oil Stand-Alone Hydrotreating and Co-Hydroprocessing Experimental Data

Table A-1. Data for Stand-Alone CFP Oil Hydrotreating in Continuous Reactors

CFP Catalyst	CFP Oil O, wt % db	HT Catalyst	HT Temp °C	HT Pres. bar	HT Liquid Hourly Space Velocity L/(Lh)	Duration, h	C Yield %	Product O, wt %	Distillation	Notes	Ref
Pt/TiO_2	16	$NiMoS_x$	400	130	0.2	82	89	0.4	Gasoline 45% Diesel 39% Residue 16%	Total 140 h, rest at 0.3 L/(Lh). Gasoline RON 67 Diesel DCN 24	1
Red mud	25	$CoMoS_x$/ ZrO	400	124	0.13–0.15	307	87–96	1.0–1.2	Simdist: Gasoline 50% Diesel 40% Residue 10% Jet 29%		2
Al_2O_3-based	19.5	$NiMoS_x$	350	100	0.5	17	77	0–2		Plugged	3
Al_2O_3-based	19.5	$NiMoS_x$	300	138	0.25	103	83	2–9		O increased during run	3
Al_2O_3-based	19.5	$NiMoS_x$	290	138	0.25	132	89	1–11		O increased during run	3
Red mud	24	$NiMoS_x$	400	124	0.2	90	83	1.0	Simdist: Gasoline 42% Diesel 41% Residue 17% Jet 28%		4
Red mud	23	$NiMoS_x$	400	124	0.2	90	90	0.8	Simdist: Gasoline 48% Diesel 42% Residue 11% Jet 28%		4
Red mud	24	$NiMoS_x$	400	124	0.2		83	1.1	Gasoline 37% Diesel 43% Residue 19%	Total 104 h, rest at 0.3 L/(L h) Gasoline RON 59 Diesel DCN 26	4,5

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CFP Catalyst	CFP Oil O, wt % db	HT Catalyst	HT Temp °C	HT Pres. bar	HT Liquid Hourly Space Velocity L/(Lh)	Duration, h	C Yield %	Product O, wt %	Distillation	Notes	Ref
Red mud	27.5	$NiMoS_x$	400	124	0.2		81	0.9	-		4
Red mud	28	$NiMoS_x$	400	124	0.2		83	1.7	-		4
γ-Al_2O_3	28	$NiMoS_x$	300	138	0.25	103	83	1.5–9	-	O increased during run	6
γ-Al_2O_3	25.5	$NiMoS_x$	290	138	0.25	365	89	2–8	-	O increased during run	6
γ-Al_2O_3	17	$NiMoS_x$	300	138	0.25	118	88	0.3–1.5	-	O increased during run	6
γ-Al_2O_3	28.8	$NiMoS_x$	290	138	0.35	62	-	1–9	-	Plugged, O increased during run	6
HZSM-5	17	$NiMoS_x$	400	128	0.2	80	93	0.05	Gasoline 50% Diesel 39% Residue 11%	Total 102 h, rest at 0.3 L/(L h) Gasoline RON 83 Diesel DCN 13	5
Pt/TiO_2	17	$NiMoS_x$/Al_2O_3	150–385	125	0.16 WHSV		89	≤0.3	Gasoline 51% Diesel 45% Residue 4%	Half of catalyst at final temperature Diesel ICN 45	7
Pt/TiO_2	16	$NiMoS_x$/Al_2O_3	150–385	125	0.11 WHSV	70	92	0.3	Gasoline 34% Jet 40% Diesel 11% Residue 12%	Half of catalyst at final temperature	8
HZSM-5	18	$NiMoS_x$/Al_2O_3	90–385	125	0.12 WHSV	50	91.5	0.05	Gasoline 36% Jet 39% Diesel 14% Residue 8%	Half of catalyst at final temperature	8
HZSM-5	20	$NiMoS_x$/Al_2O_3	150–385	125	0.1 WHSV	72	92	0.04	Gasoline 26% Jet 46% Diesel 21% Residue 3%	Half of catalyst at final temperature	9
HZSM-5	17	$NiMoS_x$/Al_2O_3	150–385	125	0.1 WHSV	72	91	0.02	Gasoline 26% Jet 50% Diesel 17% Residue 4%	Half of catalyst at final temperature Total 216 h	9

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CFP Catalyst	CFP Oil O, wt % db	HT Catalyst	HT Temp °C	HT Pres. bar	HT Liquid Hourly Space Velocity L/(Lh)	Duration, h	C Yield %	Product O, wt %	Distillation	Notes	Ref
HZSM-5	17	$NiMoS_x/Al_2O_3$	150–350	125	0.1 WHSV	72			Gasoline 17% Jet 53% Diesel 21% Residue 6%	Half of catalyst at final temperature Total 216 h	9
HZSM-5	17	$NiMoS_x/Al_2O_3$	150–300	125	0.1 WHSV	72			Gasoline 13% Jet 45% Diesel 15% Residue 25%	Half of catalyst at final temperature Total 216 h	9
HZSM-5	19.5	$NiMoS_x/Al_2O_3$	150–400	125	0.1 WHSV	423	93	<0.01	Gasoline 34% Jet 53% Diesel 4% Residue 7%	Half of catalyst at final temperature Total 588 h	NREL, previously unpublished

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Table A-2. Data for Co-Hydroprocessing of CFP Oils With Petroleum Streams

CFP Catalyst	CFP Oil O, wt % db	Petroleum Stream	CFP %	HT Conditions	Overall C Yield, %	CFP Oil C Incorporation by C-14	Product O, wt %	Distillation	Notes	Ref
Pt/TiO_2	17	SRD	10%	$NiMoS_x/Al_2O_3$ 340°C, 83 bar 1 g/(g_{cat} h) 600 sL H_2/L	100	95%	<0.5	Gasoline 10% Diesel 82% Residue 9%	Biogenic C distribution: Gasoline 49% Diesel 44% Residue 7% Diesel ICN 50	10
Pt/TiO_2	17	SRD	20%	$NiMoS_x/Al_2O_3$ 325°C, 55 bar 1 g/(g_{cat} h) 600 sL H_2/L	100	95%	≤0.3	-	ICN 45	7
Pt/TiO_2	17	SRD	20%	$CoMoS_x/Al_2O_3$ 325°C, 55 bar 1 g/(g_{cat} h) 600 sL H_2/L	95	95%	≤0.3	-	Normalized C yield 99% ICN 42	7
HZSM-5	22	SRD	10%	$NiMoS_x/Al_2O_3$ 340°C, 83 bar 1 g/(g_{cat} h) 600 sL H_2/L	99 normalized	30% (error?)	<0.01	Gasoline 2% Jet 16% Diesel 61% Residue 19%	Jet: 4.7 mm^2/(sK) at −20°C 9.5 mm^2/(sK) at −40°C 824 g/cm^3 at 15°C 27 mN/m at 20°C 43.0 MJ/kg Flash 44°C Freezing −65°C	NREL, previously unpublished

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CFP Catalyst	CFP Oil O, wt % db	Petroleum Stream	CFP %	HT Conditions	Overall C Yield, %	CFP Oil C Incorporation by C-14	Product O, wt %	Distillation	Notes	Ref
HZSM-5	22	SRD	20%	$NiMoS_x/Al_2O_3$ 340°C, 83 bar 1 g/(g_{cat} h) 600 sL H_2/L	98 normalized	94%	0.3, <0.01 in jet fraction	Gasoline 6% Jet 15% Diesel 56% Residue 21%	Jet fraction: 4.3 mm^2/(sK) at −20°C 8.5 mm^2/(sK) at −40°C 835 g/cm^3 at 15°C 27 mN/m at 20°C 42.7 MJ/kg Flash 46°C Freezing −69°C	NREL, previously unpublished
HZSM-5	22	SRD	30%	$NiMoS_x/Al_2O_3$ 340°C, 83 bar 1 g/(g_{cat} h) 600 sL H_2/L	97 normalized	99%	<0.01	Gasoline 8% Jet 17% Diesel 46% Residue 26%	Jet: 4.0 mm^2/(sK) at −20°C 7.8 mm^2/(sK) at −40°C 839 g/cm^3 at 15°C 27 mN/m at 20°C 42.6 MJ/kg Flash 45°C Freezing −69°C	NREL, previously unpublished
Zeolite	13	VGO	6%	$NiMoS_x/Al_2O_3$ (200°C) 400°C 83 bar 2 L VGO/(L_{cat} h) 2,000 sL H_2/L	88	4.6% biogenic C in product	<0.5	Simdist: Gasoline 4% Jet 24% Residue 72%	Transition zone starting at 200°C with Minimal impact on VGO processing (HDS, cracking)	11
Zeolite	13	VGO	11%	$NiMoS_x/Al_2O_3$ (200°C) 400°C 83 bar 2 L VGO/(L_{cat} h) 2,000 sL H_2/L	96	8.3% biogenic C in product	<0.5	Simdist: Gasoline 5.5% Jet 25% Residue 69%	Minimal impact on VGO processing (HDS, cracking)	11

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CFP Catalyst	CFP Oil O, wt % db	Petroleum Stream	CFP %	HT Conditions	Overall C Yield, %	CFP Oil C Incorporation by C-14	Product O, wt %	Distillation	Notes	Ref
Zeolite	13	VGO	20%	$NiMoS_x/Al_2O_3$ (200°C) 400°C 83 bar 2 L VGO/(L_{cat} h) 2,000 sL H_2/L	95	16.4% biogenic C in product	<0.5	Simdist: Gasoline 7% Jet 27% Residue 66%	Minimal impact on VGO processing (HDS, cracking)	11
Zeolite	14.5	VGO	6%	$NiMoS_x/Al_2O_3$ (200°C) 400°C 83 bar 2 L VGO/(L_{cat} h) 2,000 sL H_2/L	96		<0.5	Simdist: Gasoline 4% Jet 23% Residue 73%	Minimal impact on VGO processing (HDS, cracking)	11
Zeolite	14.5	VGO	11%	$NiMoS_x/Al_2O_3$ (200°C) 400°C 83 bar 2 L VGO/(L_{cat}h) 2,000 sL H_2/L	94		<0.5	Simdist: Gasoline 5% Jet 24% Residue 72%	Minimal impact on VGO processing (HDS, cracking)	11
P/ZSM-5	22%	50% VGO 50% LCO	20%	HT+HC (1) $NiMoS_x/Al_2O_3$ 385°C 0.93 WHSV (2) NiW/zeolite 400°C 0.93 WHSV 131 bar 3,000 sL H_2/L	96		0.05	Gasoline 16% Jet 41% Diesel 14% Residue 25%	Jet: 3.2 mm^2/(sK) at −20°C 7.5 mm^2/(sK) at −40°C 831 g/cm^3 at 15°C 27 mN/m at 20°C 42.9 MJ/kg Flash 45°C Freezing <−75°C ICN 33	NREL, previously unpublished

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Appendix B. Capital Cost Estimates for Hydroprocessing Units

The overall process flow diagram for the analysis is shown in Figure B-1. The corresponding equipment list is provided in Table B-1.

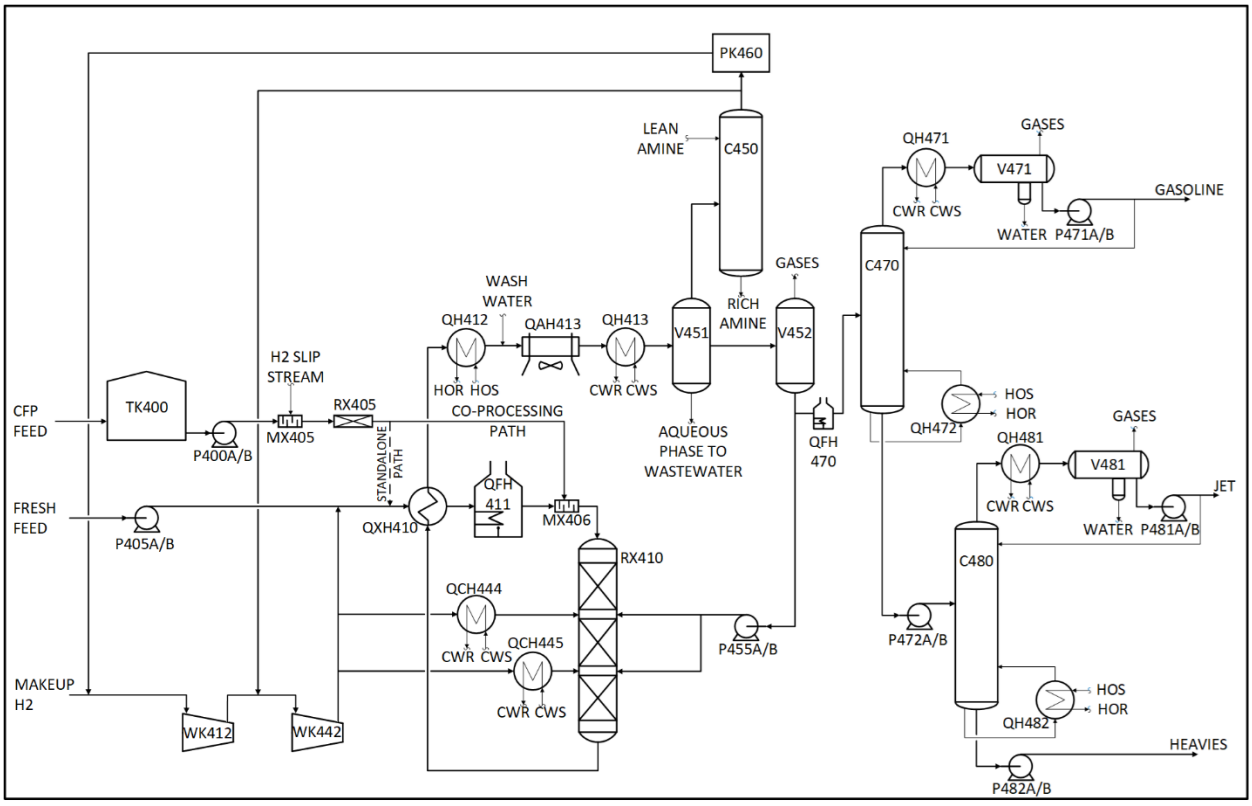

Figure B-1. Process flow diagram for petroleum feedstock and CFP oil hydroprocessing. Equipment types associated with the equipment numbers are shown in the following Table B-1.

The capital cost basis for the refinery hydroprocessing scenarios represent the estimated capital requirements to either enable a refinery unit for coprocessing or repurpose an existing refinery unit for dedicated CFP oil hydroprocessing. KBR[6] worked with NREL to develop the capital cost bases for the hydroprocessing facilities. The KBR estimates were based on U.S. Gulf Coast labor rates and first Quarter 2024 Equipment cost, and NREL used the overall 2023 Chemical Engineering Plant Cost Index published by *Chemical Engineering* for deriving cost information in 2020 U.S. dollars.

[6] Estimates provided by KBR are for informing a broad set of analysis and do not reflect any specific industrial installation. Estimates for every industrial implementation will need to be considered based on their unique assumptions to come up with accurate and specific estimates.

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Table B-1. Hydroprocessing Unit Equipment List

C450, Amine Scrubber	PK460, Pressure Swing Adsorption
C470, Product Recovery Column 1	QAH413, Rx Effluent Air Cooler
C480, Product Recovery Column 2	QCH413, Rx Effluent Water Cooler
MX405, CFP and H2 Static Mixer	QCH444, Water Cooler H2 Quench 1
MX406, Rx Inlet Static Mixer	QCH445, Water Cooler H2 Quench 2
P400A, Catalytic Pyrolysis Oil Pump	QFH411, Reactor Feed Furnace
P400B, Catalytic Pyrolysis Oil Pump	QH412, Process Rx Eff Cooler
P405A, Liquid Feed Pump	QFH470, Column 1 Feed Preheat Furnace
P405B, Liquid Feed Pump	QH471, Product Recovery Condenser 1
P455A, Liquid Recycle Pump	QH472, Product Recovery Reboiler 1
P455B, Liquid Recycle Pump	QH481, Product Recovery Condenser 2
P471A, Product Recovery Reflux Pump 1	QH482, Product Recovery Reboiler 2
P471B, Product Recovery Reflux Pump 1	QXH410, Feed Preheater to Furnace
P472A, Product Recovery Bottoms Pump 1	RX405, CFP Cat Stabilization Rx
P472B, Product Recovery Bottoms Pump 1	RX410, Reactor System
P481A, Product Recovery Reflux Pump 2	TK400, Cat Pyrolysis Oil Storage
P481B, Product Recovery Reflux Pump 2	V451, High Pressure Flash Drum
P482A, Product Recovery Bottoms Pump 2	V452, Low Pressure Flash Drum
P482B, Product Recovery Bottoms Pump 2	V471, Prod Recovery Overhead Drum 1
PIPE412, H2 MU Comp Jumpover Line	V481, Prod Recovery Overhead Drum 2
PIPE442, H2 Rec Comp Jumpover Line	WK412, H2 Make-up Compressor
PIPE500, CFP Pipe from Storage	WK442, H2 Recycle Compressor

KBR provided NREL with bare equipment costs (without installation) for the list in Table B-1 equipment based on some sample process conditions from NREL's models. Model-informed flow rates, duties, power consumption etc. were used for cost estimates. Since NREL-modeled cases span a wide range of conditions, it was necessary to develop a methodology to quickly estimate costs for different process conditions. NREL used the cost data points provided by KBR to estimate scaling exponents for each equipment to allow direct cost calculations for any condition within the range of the original estimates (Table B-2). Pressure corrections were applied to these estimated bare equipment costs based on guidance from KBR, as detailed in Table B-3 and its footnotes.

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Table B-2. Scaling Information Derived by NREL Based on KBR Cost Estimates

Equip ID	Equipment Description	Base Bare Equip for 1 unit, 1Q24 (2023$ used for year conversion)	Base Equip Pres. (psig)	Base Scaling Value	Scaling Unit	Scaling Expo-nent	Pres. Correct Active?
C450	Amine Scrubber	$111,500		241	H2S+CO_2 (lbmol/h)	0.8050	0
C470	Product Recovery Column 1	$5,165,700		946725	Feed (lb/h)	0.6096	0
C480	Product Recovery Column 2	$3,695,600		678905	Feed (lb/h)	0.5792	0
MX405	CFP and H_2 Static Mixer	$6,200		10000	CFPO Feed (BPD)	0.4449	0
MX406	Rx Inlet Static Mixer	$16,100		7500	CFPO Feed (BPD)	0.5683	0
P400A/B	Cat Pyrolysis Oil 2 Pumps	$625,000	1193.4	10000	CFPO Feed (BPD)	0.1840	1
P405A/B	Liquid Feed 2 Pumps	$1,539,000	1193.4	75000	Fresh Feed (BPD)	0.5836	1
P455A/B	Liquid Recycle 2 Pumps	$992,400	1193.4	24908	Recycle (BPD)	0.5836	1
P471A/B	Prod Recov Reflux 2 Pumps - 1	$61,200	23.5	946725	C470 Feed (lb/h)	0.2973	1
P472A/B	Prod Recov Btms 2 Pumps - 1	$65,200	59.7	946725	C470 Feed (lb/h)	0.3189	1
P481A/B	Prod Recov Reflux 2 Pumps - 2	$55,600	1.5	678905	C480 Feed (lb/h)	0.2965	1
P482A/B	Prod Recov Btms 2 Pumps - 2	$42,800	37.7	678905	C480 Feed (lb/h)	0.2202	1
PIPE412	H2 MU Comp Jumpover Line	$6,399		12177	Base H2 Mkup (lb/h)	0.7000	0
PIPE442	H2 Rec Comp Jumpover Line	$11,668		75784	Base H2 Recycle (lb/h)	0.7000	0
PIPE500	CFP Pipe from Storage	$48,398		10000	CFPO Feed (BPD)	0.7000	0
PSA460	Pressure Swing Adsorption	$6,056,874		83533220	H2 Mkup (scf/d)	0.4950	0
QAH413	Rx Effluent Air Cooler	$2,157,000		126687455	Air Cool (Btu/h)	0.8885	0
QCH413	Rx Effluent Water Cooler	$1,086,020	500.0	39278409	Trim Cool (Btu/h)	0.8507	1
QCH444	Water Cooler H2 Quench 1	$189,540	500.0	4732610	Int. Fed Gas Cool1 (Btu/h)	0.8371	1
QCH445	Water Cooler H2 Quench 2	$189,540	500.0	4732610	Int. Fed Gas Cool2 (Btu/h)	0.8371	1
QFH411-L	Reactor Preheat Furnace - Liquid Feed	$1,384,900		42461918	Furnace Duty (Btu/h)	0.5120	0
QFH411-G	Reactor Preheat Furnace - Gas Feed	$693,300		11864372	Furnace Duty (Btu/h)	0.4860	0
QH412	Process Rx Eff Cooler	$3,754,270	500.0	137084553	Process Cool (Btu/h)	0.9856	1
QH471	Prod Recov Condenser 1	$4,334,600	150.0	946725	C470 Feed (lb/h)	0.8230	0
QH472	Prod Recov Reboiler 1	$3,942,300	150.0	946725	C470 Feed (lb/h)	0.8180	0
QH481	Prod Recov Condenser 2	$2,184,900	150.0	678905	C480 Feed (lb/h)	0.7739	0
QH482	Prod Recov Reboiler 2	$1,664,400	150.0	678905	C480 Feed (lb/h)	0.7761	0
QXH410	Feed Preheater to Furnace	$1,258,530	500.0	366906999	Rx/Eff Duty (Btu/h)	0.8846	1
RX405	CFP Cat Stabilization Rx	$18,500	1193.8	10000	CFPO Feed (BPD)	0.2100	1
RX410-HT	Reactor System - Hydrotreater	$11,276,100	1150.0	1274842	Gas+Liq (lb/h)	0.9091	1
RX410-HC	Reactor System - Hydrocracker	$11,276,100	1150.0	1274842	Gas+Liq (lb/h)	0.9091	1
TK400	Cat Pyrolysis Oil Storage	$2,504,600		10000	CFPO Feed (BPD)	0.8689	0
V451	High Pressure Flash Drum	$811,000	500.0	1274842	Feed Gas+Liq (lb/h)	0.9548	1
V452	Low Pressure Flash Drum	$220,600		990018	Feed Gas+Liq - Gas Recycle (lb/h)	0.7344	0
V471	Prod Recov Ovhd Drum 1	$55,100		946725	C470 Feed (lb/h)	0.3352	0
V481	Prod Recov Ovhd Drum 2	$67,900		678905	C480 Feed (lb/h)	0.3389	0
WK412	H2 Make-up Compressor	$11,426,400		13345	H2 Compr. (hp)	0.7962	0
WK442	H2 Recycle Compressor	$3,054,900		3977	Recy. Compr. (hp)	0.3038	0
QH470	Dist. Col 1 Liq. Feed Preheat Furnace (reused QFH411-L)	$1,384,900	150.0	42461918	Furnace Duty (Btu/h)	0.5120	0

KBR further provided Lang factor relations to estimate the total fixed capital investment (FCI) from the bare equipment costs for various families of process equipment. NREL used the average of the low and high FCI estimates, unless there is specific guidance otherwise. These estimates, along with pressure adjustment correlations are included in Table B-3 and its

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footnotes. In some cases, NREL's sensitivity cases included reactor flows beyond the valid range for the KBR pressure correction correlations. NREL removed the pressure correction for the hydrotreater (RX-410-HT) in those cases since the pressure in the hydrotreating-only configuration is close to the original pressure used for the reactor cost estimates.

Table B-3. Pressure Correction Correlations and Lang Factors for FCI Calculations

Equipment Type	Pressure Factor	Lang Factor Low	Lang Factor High	Footnotes
Column	-	$815x^{-0.37}$	$3360x^{-0.45}$	1, 4
Static Mixer	-	90	100	2, 5
Pump	Correlation	$30x^{0.845}$	$80x^{0.800}$	3, 6
Pipe	-	6.5	7	2, 7
PSA	-	3.6	4.4	2, 4
Air Cooler	-	4.75	5.75	2, 5
Shell and Tube Exchanger	Correlation	$50x^{0.843}$	$200x^{0.763}$	3, 8
Liquid-Dominated-Feed Furnace	-	$630x^{-0.36}$	$3050x^{-0.44}$	1, 9
Gas-Dominated-Feed Furnace	-	13.2	15.8	2, 10
Stabilization Reactor	Correlation	35.5	45.5	2, 8
Hydrotreater Reactor	Correlation	3.5	5.3	2, 8
Tank	-	6.0	7.2	2, 4
LP Flash Vessel	-	$1000x^{-0.36}$	$5000x^{-0.44}$	1, 9
HP Flash Vessel	Correlation	$1000x^{-0.36}$	$5000x^{-0.44}$	1, 11
Compressor	-	3.0	3.8	2, 12

Footnotes:

(1) Lang factor based power curve where X = bare equipment cost.

(2) Lang factor is a constant multiplier.

(3) Equation to calculate FCI, not Lang Factor. X = bare equipment cost. Formula results in FCI directly.

(4) No changes in equipment design pressure.

(5) Changes in design pressure have negligible impact on bare equipment cost.

(6) Three formulas are needed to scale bare equipment cost based on pressure. The variables are defined as:

B = Bare Equipment Cost at Known Pressure

P = Known Pressure (psig)

x = New Desired Pressure (psig)

y = Bare Equipment Cost at New Desired Pressure

D and C are factors calculated from the above parameters and used in third equation to calculate y. *For pumps:*

$$D = 0.001262 - 7.44e^{-10} * \left(\frac{B}{e^{(P * 0.00109)}}\right)$$

$$C = \left(\frac{B}{e^{(P * D)}}\right)$$

$$y = Ce^{Dx}$$

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(7) Impacts to piping costs from pressure variation were not evaluated.

(8) Equation for scaling bare equipment cost on pressure where:

B = Bare Equipment Cost at Known Pressure

P = Known Pressure (psig)

x = New Desired Pressure (psig)

y = Bare Equipment Cost at New Desired Pressure

For shell and tube exchangers:

$$y = x * \left(\frac{0.64 * B}{P} + 260\right) + \left(\frac{550 * B}{P} - 325000\right)$$

For stabilization reactor:

$$y = B + 4.1004 * (x - P)$$

For hydroprocessing reactor:

$$y = B + 5228 * (x - P)$$

(9) FCI at varying pressures fell within range given by low and high factors on base case (1,085 psig hydrotreater operating pressure). Low Lang factor should be used for pressures below 1,085 psig and high Lang factor should be used for pressures above 1,085 psig.

(10) Impacts to gas furnace from pressure variation were not evaluated.

(11) Equation for scaling bare equipment cost on pressure where:

B = Bare Equipment Cost at Known Pressure

P = Known Pressure (psig)

x = New Desired Pressure (psig)

y = Bare Equipment Cost at New Desired Pressure

A = Scaling factor that varies depending on pressure.

If initial and new pressures are both <1,750 psig, A = 1,225.9

If initial and new pressures are both >1,750 psig, A = 2,467.1

If initial and new pressures fall into separate pressure ranges (0–1,750 psig and 1,750+ psig), the following steps are needed to scale bare equipment cost:

1. The common data point for the two pressure curves is 207.08 psig. Use the appropriate A value based on initial pressure to calculate y based on x = 207.08 psig.
2. Use the bare equipment cost calculated in step 1 and initial pressure of 207.08 psig (these are now the B and P values, respectively, for the new calculation) to scale bare equipment cost to the desired pressure. Select the A value based on the desired pressure.

For high-pressure flash vessel:

$$y = B + A * (x - P)$$

(12) A correlation for impacts from pressure variations on compressor bare equipment cost was not determined. Pressure variations resulted in greater variability in bare equipment cost so the low and high Lang factors were adjusted to accommodate these variations.

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Appendix C. CFP Oil Surrogate Compounds in Aspen Plus Models

Surrogate mode compounds used to represent the CFP-A and CFP-B oils in the Aspen Plus models are shown in Table C-1.

Table C-1. Mass Percentages of Surrogate Compounds Used To Represent CFP-A and CFP-B Oils

Name	Formula	CFP-A	CFP-B
Hydrogen	H2		0.002348%
Water	H2O	0.831567%	0.825528%
Carbon Monoxide	CO	0.015095%	0.010715%
Carbon Dioxide	CO2	0.029084%	0.088160%
Methane	CH4	0.001911%	0.007635%
Formaldehyde	CH2O	0.001857%	
Formic-Acid	CH2O2	5.387388%	
Glyoxal	C2H2O2	0.249593%	0.000000%
Ethylene	C2H4	0.012963%	0.010174%
Acetaldehyde	C2H4O	0.493885%	0.634521%
Glycol-Aldehyde	C2H4O2	0.156957%	0.000000%
Acetic-Acid	C2H4O2	0.153211%	0.000000%
Ethane	C2H6	0.001878%	0.009246%
Propylene	C3H6	0.057342%	0.041143%
Acetone	C3H6O	1.290069%	1.096863%
Acetol	C3H6O2	0.043269%	0.000000%
Propionic-Acid	C3H6O2		0.000000%
Propane	C3H8	0.009597%	0.015394%
Furan	C4H4O	0.352389%	0.180944%
Trans-Crotonaldehyde	C4H6O	0.031935%	0.270075%
2-Buten-3-One	C4H6O	0.920391%	
Diacetyl-Sulfide	C4H6O2S	1.533775%	
1-Butene	C4H8	0.051555%	0.033899%
Tetrahydrofuran	C4H8O		0.156162%
Methyl-Ethyl-Ketone	C4H8O		0.608461%
N-Butyraldehyde	C4H8O		0.199415%
Furfural	C5H4O2	0.037105%	
2-Methylfuran	C5H6O	0.634893%	0.621674%
3-Methylfuran	C5H6O		0.098493%
Cyclopentenone	C5H6O	0.468633%	0.163926%
Furfuryl-Alcohol	C5H6O2		0.000000%
1,3-Cyclopentanedione	C5H6O2	0.149723%	
Cyclopentanone	C5H8O	0.017397%	0.378491%
2-Pentenal	C5H8O		0.518068%
Trans-3-Penten-2-One	C5H8O		0.156733%
Methyl-N-Propyl-Ketone	C5H10O		0.218840%
Diethyl-Ketone	C5H10O		0.132348%
N-Pentane	C5H12		0.011161%
Benzene	C6H6	0.362210%	0.299024%
Phenol	C6H6O	0.002591%	5.086820%
1,2-Benzenediol	C6H6O2	0.460630%	0.765510%

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Name	Formula	CFP-A	CFP-B
5-Methylfurfural	C6H6O2	0.026620%	
Hydroxymethylfurfural	C6H6O3		0.000000%
1,3-Cyclohexadiene	C6H8		0.009666%
2,5-Dimethyloxole	C6H8O		0.253147%
2-Methyl-2-Cyclopenten-1-One	C6H8O	0.162409%	0.118644%
2-Ethylfuran	C6H8O		0.228404%
2-Hydroxy-3-Methyl-2-Cyclopenten-1-one	C6H8O2		0.000000%
Cyclohexene	C6H10	0.183453%	
Cyclohexanone	C6H10O		0.000000%
2-Methylcyclopentanone	C6H10O		0.156492%
Levoglucosan	C6H10O5	3.147458%	0.000000%
1-Hexanal	C6H12O		0.151875%
2-Hexanone	C6H12O		0.073219%
3-Hexanone	C6H12O		0.067565%
Cyclohexane	C6H12		0.765082%
4-Methylpiperidine	C6H13N	1.010897%	
Toluene	C7H8	1.351127%	0.223328%
O-Cresol	C7H8O	0.406268%	0.247485%
M-Cresol	C7H8O	0.412861%	
P-Cresol	C7H8O	0.121193%	
Guaiacol	C7H8O2	0.478313%	3.697653%
Methylcyclohexane	C7H14		0.087826%
2-Pentanone,-3-Ethyl-	C7H14O		0.319367%
Benzofuran	C8H6O	0.142478%	
Styrene	C8H8	0.087806%	
Vanillin	C8H8O3	5.052388%	
O-Xylene	C8H10	0.806088%	2.625854%
P-Xylene	C8H10	3.520454%	
Ethylbenzene	C8H10	0.174449%	
2,3-Xylenol	C8H10O		4.321594%
2,5-Xylenol	C8H10O	0.366302%	
4-Methyl-2-Methoxyphenol	C8H10O2	0.389861%	
Syringol	C8H10O3	10.737977%	1.048450%
C8H10O3-N1	C8H10O3		1.691445%
2,4,6-Trimethylpyridine	C8H11N	1.633478%	0.637586%
1-Ethyl-2-Methylcyclopentene	C8H14		0.039848%
Cis-1,2-Dimethylcyclohexane	C8H16		0.031746%
Trans-1,2-Dimethylcyclohexane	C8H16		0.088559%
Indene	C9H8	0.452281%	
2-Methylbenzofuran	C9H8O	0.235734%	0.868838%
Indane	C9H10	0.250841%	
5-Indanol	C9H10O	0.069247%	
2-Methoxy-4-Vinylphenol	C9H10O2	0.265995%	0.026170%
Acetovanillone	C9H10O3	0.116465%	
Syringaldehyde	C9H10O4	0.652240%	
N-Propylbenzene	C9H12		0.600361%
1,3,5-Trimethylbenzene	C9H12	0.404755%	
1-Methyl-4-Ethylbenzene	C9H12		0.055641%

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Name	Formula	CFP-A	CFP-B
1,2,3-Trimethylbenzene	C9H12		0.041551%
Naphthalene	C10H8	1.002643%	2.255695%
1-Naphthalenol	C10H8O	0.715995%	2.191306%
1-Methylindene	C10H10	0.447027%	
Coniferyl-Aldehyde	C10H10O3	0.072708%	2.258845%
4-Allyl-2-Methoxyphenol	C10H12O2	0.188017%	
Isoeugenol	C10H12O2	0.666943%	2.408934%
Propyl-P-Hydroxybenzoate	C10H12O3	0.039121%	
Decahydronaphthalene	C10H18		1.156483%
1-Methylnaphthalene	C11H10	1.610353%	0.358757%
Naphthalen-1-Ylmethanol	C11H10O	0.044974%	
1-Methyl-Decalin	C11H20		0.397920%
Dibenzothiophene	C12H8S	1.695093%	0.164986%
2,6-Dimethylnaphthalene	C12H12	0.442461%	
2-Ethylnaphthalene	C12H12	0.101160%	
Fluorene	C13H10	0.067737%	
Biphenyl-4-Carboxaldehyde	C13H10O	5.118564%	
5-Hexyl-O-Cresol	C13H20O	3.982309%	0.457474%
Phenanthrene	C14H10	0.143389%	0.000000%
1-Methylfluorene	C14H12	0.167868%	
9-Fluorenemethanol	C14H12O	5.125758%	
Tetradecahydrophenanthrene	C14H24		0.953042%
1-Methylanthracene	C15H12	0.167872%	
4-Methylphenanthrene	C15H12	0.118659%	0.000001%
1,2,4A,5,6,7,8,8A-Octahydro-3-Propyl-4-Ethylnaphthalene	C15H26		1.009980%
1,4-Dimethylphenanthrene	C16H14	0.167873%	0.815630%
Chrysene	C18H12	0.167873%	
4,5-Dimethylpyrene	C18H14	0.167873%	
Retene	C18H18	0.046883%	
4'-(Trans-4-N-Butylcyclohexyl)Acetophenone	C18H26O		7.320436%
High-MW-Lignin-A	C20H26O8	5.080411%	
High-MW-Lignin-B	C21H26O8	5.080411%	11.516975%
13H-Dibenzo[A,G]Fluorene	C21H14	0.167873%	
1,1,3-Triphenyl-1-Propene	C21H18	0.167873%	
(E)-1-(4-Methoxyphenyl)-2-(4-Hexoxyphenyl)Ethene	C21H26O2	5.124714%	7.401786%
Eicosahydrodibenzo(a,i)fluorene	C21H34		9.310290%
1,3-Benzene-Carboxylic-Acid,bis(4-Methoxyphenyl)-Ester	C22H18O6	5.580417%	
1,4-Cyclohexane-Dicarboxylic-Acid,trans-,di(P-Methoxyphenyl)-Ester	C22H24O6	5.580416%	17.355042%
4-Methoxy-4'-Heptoxy-trans-Stilbene	C22H28O2	4.330403%	1.547247%

Structures of the High-MW-Lignin-A and High-MW-Lignin-B were entered in Aspen Plus to estimate their properties. Those structures are available in Table F-3 of Jones et al. 2013 [1].

Reference for Appendix C

1. S. Jones, et al. 2013. *Process Design and Economics for the Conversion of Lignocellulosic Biomass to Hydrocarbon Fuels: Fast Pyrolysis and Hydrotreating Bio-oil Pathway*. PNNL-23053; NREL/TP-5100-61178. www.pnnl.gov/main/publications/external/technical_reports/PNNL-23053.pdf.

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Appendix D. Surrogate Compounds in Aspen Plus Hydroprocessing Models

Straight-Run Diesel Surrogate Composition

Surrogate compounds to represent straight-run diesel (SRD) were based on literature information [Mueller et al., 2016]. Nitrogen and sulfur compounds were then added in proportions to match their mass fractions in SRD streams of some of our refinery optimization models (Table D-1).

Table D-1. Surrogate Model Compounds for SRD

Compound Name	Formula	Mass %
Indole	C8H7N	0.38%
2,4,6-trimethylpyridine	C8H11N	0.38%
1,2,3,4-tetrahydronaphthalene	C10H12	11.07%
n-butylcyclohexane	C10H20	13.66%
1-methylnaphthalene	C11H10	10.06%
2-butylbenzothiophene	C12H14S	3.11%
2,3-dimethyl-4-ethylquinoline	C13H15N	0.38%
4,6-dimethyldibenzothiophene	C14H12S	3.11%
phenanthrene,-tetradecahydro-	C14H24	5.91%
1,3,7-trimethylcarbazole	C15H15N	0.38%
1,3,5-triisopropylbenzene	C15H24	15.23%
1,3,5-triisopropylcyclohexane	C15H30	11.81%
n-octadecane	C18H38	14.03%
16-methylheptadecane	C18H38	9.41%
n-eicosane	C20H42	1.11%
Elemental Analysis, wt %	**Value**	
% C	86.72	
% H	12.15	
% N	0.14	
% S	0.99	
H:C Molar Ratio	**1.67**	

Reference: Mueller, C. J.; Cannella W. J.; Bays, J. T.; Bruno, T. J.; DeFabio, K.; Dettman, H. D.; Gieleciak, R. M.; Huber, M. L.; Chol-Bum, K.; McConnell, S. S.; Pitz, W. J.; Ratcliff, M. A. Diesel Surrogate Fuels for Engine Testing and Chemical-Kinetic Modeling: Compositions and Properties. Energy Fuels 2016, 30, 1445–1461. dx.doi.org/10.1021/acs.energyfuels.5b02879.

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Light Cycle Oil (LCO) and Vacuum Gas Oil (VGO) Surrogate Compositions with Accurate Compositional and Volatility Characteristics

Unlike SRD, we did not find suitable LCO and VGO surrogates in the literature and needed to create surrogate compound slates for use in our models. This work was performed by Dr. Suphat Watanasiri.

LCO is a middle distillate produced in the FCC unit. It generally has high sulfur content (up to 4 wt %), nitrogen content (up to 600 ppm by weight), and aromatic content (up to 90 wt %, the majority of which are diaromatics) [1]. Vacuum gas oil (VGO) is a standard feedstock for industrial FCC plants and is the top product of the vacuum distillation [2]. VGO usually contains hydrocarbon molecules ranging from C17 to C33 with initial boiling point of ~320 °C and final boiling point of ~560°C [3].

Since both LCO and VGO contain thousands of complex molecules and isomers, it is difficult to develop detailed understanding of kinetic reactions involved. Instead, using a smaller number of realistic surrogate compounds can enable rational selection of reactions and products for the hydroprocessing reactors. In this work, the surrogate compounds for a LCO and a VGO were selected from a set of palette compounds and the compositions (mole fractions) of the selected surrogate compounds were determined by multi-property regression analysis of the thermodynamic, compositional, and volatility properties of the LCO or VGO of interest. The resultant surrogate compounds and their compositions constitute the surrogate liquid hydrocarbon feedstock which provides the best match to the properties of the actual feedstock of interest and is used in our process model.

The experimental properties of the LCO fuel that can be used in the multi-property regression analysis include specific gravity, S wt %, N wt %, basic nitrogen wt %, mono-aromatics wt %, diaromatics wt %, tri-aromatics wt %, paraffins wt %, and D86 distillation temperatures at the following vol %: 0, 10, 30, 50, 70, 90, and 100. For the VGO fuel, the properties include density at 15 °C, C wt %, H wt %, S wt %, N wt %, paraffins wt %, mono-aromatics wt %, diaromatics wt %, tri-aromatics wt %, poly-aromatics wt %, and TBP distillation temperatures at the following wt %: 0, 5, 10, 20, 30, 40, 50, 60, 70, 80, 90, 95 and 100.

The Aspen Plus (version 10.0) data-fit feature [4] was used to perform the regression. The data regression algorithm [5,6,7] minimizes the following objective function:

$$Q = \sum_{i=1}^{N} \left(\frac{q^{exp} - q^{calc}}{\sigma} \right)_i^2$$

where q^{exp} is the experiment property, q^{calc} is the calculated property and σ is the standard deviation of the data. The summation is over all the N data points used.

Liquid density was calculated using a linear mass-fraction-based mixing rule of specific volume ν_L, neglecting excess volume contribution [8].

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$$\nu_{\mathrm{L}} = \sum_i w_i \nu_{\mathrm{L},i} \;\Rightarrow\; \rho_{\mathrm{L}} = \left(\sum_i \frac{w_i}{\rho_{\mathrm{L},i}} \right)^{-1}$$

where ν_{L} is specific volume of the mixture, ρ_L is liquid density of the mixture, w_i, $\nu_{\mathrm{L},i}$, and $\rho_{\mathrm{L},i}$ are mass fraction, specific volume, and liquid density of component i, respectively.

Elemental analysis (C wt %, H wt %, S wt %, and N wt %) were calculated from the composition and molecular formula of the compounds and the atomic weight of the elements in each compound. Hydrocarbon analysis (paraffins wt %, naphthene wt %, mono-aromatics wt %, diaromatics wt %, tri-aromatics wt %, and poly-aromatics wt %) were calculated from the composition (i.e., weight fraction) of the compounds involved and by categorizing the compounds into the appropriate compound classes. D86 and TBP distillation temperatures were calculated using the methods described in Watanasiri et al. [9].

LCO Results

The optimized surrogate LCO fuel used to represent the LCO in this work (from data in Laredo et al. [10]) are summarized in Table D-2.

Table D-2. Optimized Composition of Surrogate Compounds Used To Represent Properties of the LCO Used in This Work

Compound Name	Formula	Mass Flow kg/h	Mole Fraction	Mass Fraction
n-decane	C10H22	2.47	0.02865	0.02124
n-octadecane	C18H38	12.45	0.08073	0.10708
n-docosane	C22H46	4.35	0.02311	0.03742
2,2,4-trimethylnonane	C12H26	2.7	0.02616	0.02322
1,3,5-triisopropylbenzene	C15H24	11.96	0.09658	0.10287
n-tetradecylbenzene	C20H34	4	0.02405	0.0344
Acenaphthene	C12H10	27.69	0.29631	0.23817
1-octylnaphthalene	C18H24	15.49	0.10634	0.13323
Phenanthrene	C14H10	5.2	0.04815	0.04473
1-propylphenanthrene	C17H16	6.57	0.04921	0.05651
1-butylphenanthrene	C18H18	4.36	0.0307	0.0375
3-methylquinoline	C10H9N	0.083	0.00096	0.00071
1,3,7-trimethylcarbazole	C15H15N	1.56	0.0123	0.01342
2,3-dimethylbenzothiophene	C10H10S	17.38	0.17676	0.14949

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The predicted properties of the surrogate LCO are compared with experimental data of the actual LCO [10] in Table D-3. The results in Table D-3 show that the calculated properties are generally in good agreement with experimental data.

Table D-3. Comparison Between Experimental Properties of LCO [10] and Model Prediction Using the Proposed Surrogate Fuel in Table D-2

Property	Experimental	Prediction
sp. gravity (20/4 °C)	0.9573	0.9651
Elemental Analysis, wt %		
Carbon (C)	N/A	87.56
Hydrogen (H)	N/A	9.39
Atomic H/C	N/A	1.28
Sulfur (S)	3.52	2.95
Nitrogen (N), wt %	0.0972	0.0968
Nitrogen (N), ppm	972	967.8
basic nitrogen, wt %	0.0070	0.0070
basic nitrogen, ppm	70	69.8
Hydrocarbon Analysis, wt %		
Aromatics	74.7	81.10
Mono-aromatics	15.21	13.73
Di-aromatics	38.91	37.14
Tri-aromatics	15.94	13.87
Benzo and dibenzo thiophenes	N/A	14.95
Other aromatics	N/A	1.41
Paraffins	20.23	18.90
Napthenes	N/A	0.00
Total	**98.5**	**100.0**
D86 Distillation, °C		
Initial boiling point	216	214.8
10 vol %	259	258.6
30	284	283.0
50	302	302.4
70	327	326.2
90	355	355.1
Final boiling point	380	380.2

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Figure D-1 shows excellent agreement between the experimental D86 distillation temperatures and model predictions.

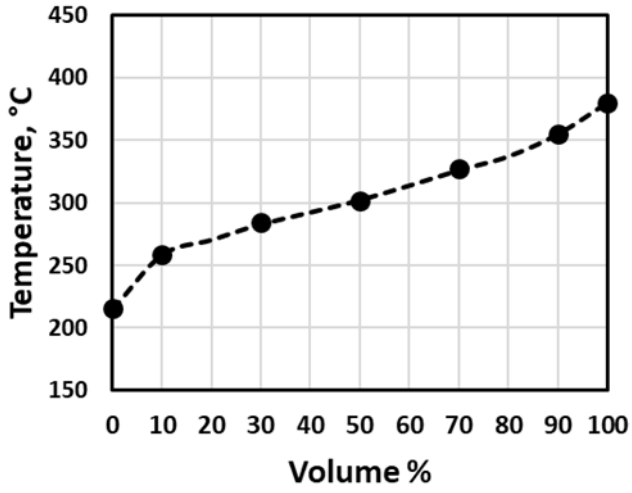

Figure D-1. Comparison between experimental D86 distillation temperatures (●) of an LCO fuel [10] and model predictions (- - -) using a surrogate fuel.

VGO Results

The optimized surrogate fuel used to represent the VGO in this work (from data in Ma et al. [11]) are summarized in Table D-4, and the predicted properties of the surrogate VGO are compared with experimental data of the actual VGO [11] in Table D-5.

Table D-4. Optimized Composition of Surrogate Compounds Used To Represent Properties of the VGO Used in This Work

Compound Name	Formula	Mass Flow kg/h	Mole Fraction	Mass Fraction
n-dodecane	C12H26	0.309	0.02567	0.01416
n-tritriacontane	C33H68	8.721	0.26545	0.39955
n-hexadecylbenzene	C22H38	1.419	0.06637	0.06501
1-dodecylnaphthalene	C22H32	0.861	0.04109	0.03945
1-butylphenanthrene	C18H18	3.24	0.19565	0.14844
hexamethylchrysene	C24H24	3.596	0.16181	0.16475
2-methyl-5,6-benzoquinoline	C14H11N	0.445	0.03259	0.02039
2,3-dimethylbenzothiophene	C10H10S	0.993	0.08660	0.04549
pentyldibenzothiophene	C17H18S	2.243	0.12477	0.10276

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Table D-5. Comparison Between Experimental Properties of VGO [11] and Model Predictions Using the Proposed Surrogate Fuel in Table D-4

Property	Experimental	Prediction
Density at 15 °C	0.9263	0.9451
Elemental Analysis, wt %		
Carbon (C)	85.27	86.64
Hydrogen (H)	12	11.01
Atomic H/C	1.68	1.51
Sulfur (S)	2.45	2.19
Nitrogen (N), wt %	0.17	0.148
Hydrocarbon Analysis, wt %		
Paraffins	45	41.4
Napthenes	N/A	0
Aromatics	44	41.8
Mono-aromatics	10	6.5
Diaromatics	5	3.9
Tri-aromatics	16	14.8
Poly-aromatics	13	16.5
Sulfur compounds		14.8
Nitrogen compounds		2.0
Polar compounds	11	16.9
Total	**100**	**100**
TBP Distillation, °C		
Initial boiling point	205	210.1
5 wt %	341	339.6
10	378	381.3
20	410	409.4
30	434	434.4
40	450	454.0
50	473	471.2
60	494	487.5
70	510	500.8
80	530	520.6
90		555.0
95		576.2
Final boiling point		597.3

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The results show that the calculated properties are in good agreement with experimental data in general. Figure D-2 shows excellent agreement between experimental TBP distillation temperatures and model predictions.

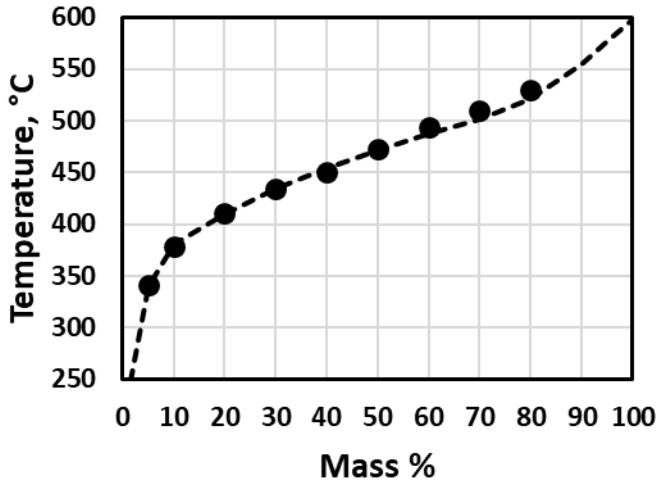

Figure D-2. Comparison between experimental TBP distillation temperatures (●) of a VGO fuel [11] and model predictions (- - -) using a surrogate VGO fuel

References for Appendix D (LCO and VGO)

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Surrogate Compounds in Hydroprocessing Models

A list of compounds with nonzero flows in the hydroprocessing models are listed in Table D-6. Subsequent stream flow summaries in Appendix L refer to these compounds by their IDs.

Table D-6. List of Surrogate Compounds in the Hydroprocessing Models

ID	Name	Formula
HYDROGEN	HYDROGEN	H2
AMMONIA	AMMONIA	H3N
H_2S	HYDROGEN SULFIDE	H2S
WATER	WATER	H2O
CO	CARBON MONOXIDE	CO
CO_2	CARBON DIOXIDE	CO2
METHANE	METHANE	CH4
FORMALDH	FORMALDEHYDE	CH2O
FORMACID	FORMIC-ACID	CH2O2
GLYOXAL	GLYOXAL	C2H2O2
ETHYLENE	ETHYLENE	C2H4
ACALDEHY	ACETALDEHYDE	C2H4O
HYDACTLD	GLYCOL-ALDEHYDE	C2H4O2
ACETACID	ACETIC-ACID	C2H4O2
ETHANE	ETHANE	C2H6
ETHANOL	ETHANOL	C2H6O
PROPYLEN	PROPYLENE	C3H6
ACETONE	ACETONE	C3H6O
ACETOL	ACETOL	C3H6O2
PROPACID	PROPIONIC-ACID	C3H6O2
PROPANE	PROPANE	C3H8
PROPANOL	1-PROPANOL	C3H8O
FURAN	FURAN	C4H4O
2BUTENAL	TRANS-CROTONALDEHYDE	C4H6O
2BTE3ONE	2-BUTEN-3-ONE	C4H6O
DIACESUL	DIACETYL-SULFIDE	C4H6O2S
BUTENE	1-BUTENE	C4H8
THF	TETRAHYDROFURAN	C4H8O
2BUTNONE	METHYL-ETHYL-KETONE	C4H8O
1BUTANAL	N-BUTYRALDEHYDE	C4H8O
NBUTANE	N-BUTANE	C4H10
P-IC4	ISOBUTANE	C4H10
FURFURAL	FURFURAL	C5H4O2
2MEFURAN	2-METHYLFURAN	C5H6O
3MEFURAN	3-METHYLFURAN	C5H6O
CYPNTENO	CYCLOPENTENONE	C5H6O
FURFUALC	FURFURYL-ALCOHOL	C5H6O2
13CPDION	1,3-CYCLOPENTANEDIONE	C5H6O2

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ID	Name	Formula
CYPNTANO	CYCLOPENTANONE	C5H8O
2MBTENAL	2-METHYL-2-BUTENAL	C5H8O
2PNTENAL	2-PENTENAL	C5H8O
3PNTE2ON	TRANS-3-PENTEN-2-ONE	C5H8O
2PENTENE	CIS-2-PENTENE	C5H10
2PENTNON	METHYL-N-PROPYL-KETONE	C5H10O
3PENTNON	DIETHYL-KETONE	C5H10O
THFA	TETRAHYDROFURFURYL-ALCOHOL	C5H10O2
PENTANE	N-PENTANE	C5H12
BENZENE	BENZENE	C6H6
PHENOL	PHENOL	C6H6O
12BZDIOL	1,2-BENZENEDIOL	C6H6O2
5METFURF	5-METHYLFURFURAL	C6H6O2
HMF	5-HYDROXYMETHYLFURFURAL	C6H6O3
CYHXDIEN	1,3-CYCLOHEXADIENE	C6H8
DIMEFURA	2,5-DIMETHYLOXOLE	C6H8O
METCY5ON	2-CYCLOPENTEN-1-ONE,-2-METHYL-	C6H8O
2ETFURAN	2-ETHYLFURAN	C6H8O
HYMET5ON	2-HYDROXY-3-METHYL-2-CYCLOPENTEN-1-ONE	C6H8O2
CYHXENE	CYCLOHEXENE	C6H10
CYHEXNON	CYCLOHEXANONE	C6H10O
2MECY5ON	2-METHYLCYCLOPENTANONE	C6H10O
LEVOGLUC	LEVOGLUCOSAN	C6H10O5
CYHXNOL	CYCLOHEXANOL	C6H12O
1HEXANAL	1-HEXANAL	C6H12O
2HEXNONE	2-HEXANONE	C6H12O
3HEXNONE	3-HEXANONE	C6H12O
CYHEXANE	CYCLOHEXANE	C6H12
METPIPR	4-METHYLPIPERIDINE	C6H13N
NHEXANE	N-HEXANE	C6H14
SORBTOL	SORBITOL	C6H14O6
TOLUENE	TOLUENE	C7H8
OCRESOL	O-CRESOL	C7H8O
MCRESOL	M-CRESOL	C7H8O
PCRESOL	P-CRESOL	C7H8O
GUAIACOL	GUAIACOL	C7H8O2
MECYHEXN	METHYLCYCLOHEXANE	C7H14
2M3HXENE	3-METHYL-2-HEXENE	C7H14
METHHEXN	2-METHYLHEXANE	C7H16
3ETPNTNO	2-PENTANONE,-3-ETHYL-	C7H14O
BZOFURAN	BENZOFURAN	C8H6O
D-INDOLE	INDOLE	C8H7N
STYRENE	STYRENE	C8H8

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ID	Name	Formula
VANILLIN	VANILLIN	C8H8O3
OXYLENE	O-XYLENE	C8H10
PXYLENE	P-XYLENE	C8H10
ETYBENZN	ETHYLBENZENE	C8H10
23XYLNOL	2,3-XYLENOL	C8H10O
25XYLNOL	2,5-XYLENOL	C8H10O
CREOSOL	4-METHYL-2-METHOXYPHENOL	C8H10O2
SYRINGOL	SYRINGOL	C8H10O3
VANALC	VANILLYL-ALCOHOL	C8H10O3
TRIMEPYR	2,4,6-TRIMETHYLPYRIDINE	C8H11N
MEETCY5E	1-ETHYL-2-METHYLCYCLOPENTENE	C8H14
DIMECYHX	CIS-1,2-DIMETHYLCYCLOHEXANE	C8H16
TRDICYHN	TRANS-1,2-DIMETHYLCYCLOHEXANE	C8H16
ETCYHEXN	ETHYLCYCLOHEXANE	C8H16
4MEHEPTN	4-METHYLHEPTANE	C8H18
INDENE	INDENE	C9H8
MEBZFURA	2-METHYLBENZOFURAN	C9H8O
INDANE	INDANE	C9H10
5INDANOL	5-INDANOL	C9H10O
MEVINPHE	2-METHOXY-4-VINYLPHENOL	C9H10O2
ETMOXETN	ACETOVANILLONE	C9H10O3
SYRNGALD	SYRINGALDEHYDE	C9H10O4
PROPBENZ	N-PROPYLBENZENE	C9H12
ETMEBENZ	1-METHYL-4-ETHYLBENZENE	C9H12
TRIMEBEN	1,2,3-TRIMETHYLBENZENE	C9H12
135TMBZ	1,3,5-TRIMETHYLBENZENE	C9H12
HXHYINDN	BICYCLO[4.3.0]NONANE	C9H16
PROPCYHX	N-PROPYLCYCLOHEXANE	C9H18
TRIMCYHX	1-TRANS-3,5-TRIMETHYLCYCLOHEXANE	C9H18
NAPTHALE	NAPHTHALENE	C10H8
1NAPTHOL	1-NAPHTHALENOL	C10H8O
MEINDENE	1-METHYLINDENE	C10H10
CONIFALD	CONIFERYL-ALDEHYDE	C10H10O3
TETRALIN	1,2,3,4-TETRAHYDRONAPHTHALENE	C10H12
EUGENOL	4-ALLYL-2-METHOXYPHENOL	C10H12O2
IEUGENOL	ISOEUGENOL	C10H12O2
PRHYBNZO	PROPYL-P-HYDROXYBENZOATE	C10H12O3
D-DECALN	TRANS-DECALIN	C10H18
DECALIN	DECAHYDRONAPHTHALENE	C10H18
D-C4CYC6	N-BUTYLCYCLOHEXANE	C10H20
DIMEOCTN	2,5-DIMETHYLOCTANE	C10H22
1MENAPHT	1-METHYLNAPHTHALENE	C11H10
1NAPMNOL	NAPHTHALEN-1-YLMETHANOL	C11H10O

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ID	Name	Formula
METDEC	1-METHYL-DECALIN	C11H20
DIBZTHIO	DIBENZOTHIOPHENE	C12H8S
D-C4BZ-S	2-BUTYLBENZOTHIOPHENE	C12H14S
BIPHENYL	DIPHENYL	C12H10
DIMENAPH	2,6-DIMETHYLNAPHTHALENE	C12H12
2ETYNAPH	2-ETHYLNAPHTHALENE	C12H12
D-C12H18	BENZENE,-(4-METHYLPENTYL)-	C12H18
D-C12H1B	BENZENE,-(2,3-DIMETHYLBUTYL)-	C12H18
CNYLXNE	1-(1-CYCLOHEXENYL)-1-CYCLOHEXENE	C12H18
2ETDECLN	3-ETHYLBICYCLO[4.4.0]DECANE	C12H22
FLUORENE	FLUORENE	C13H10
BIPHCALD	BIPHENYL-4-CARBOXALDEHYDE	C13H10O
D-CQUINL	2,3-DIMETHYL-4-ETHYLQUINOLINE	C13H15N
HXLCRESO	5-HEXYL-O-CRESOL	C13H20O
PHFLOREN	FLUORENE,-DODECAHYDRO-	C13H22
HEPCYHEX	HEPTYLCYCLOHEXANE	C13H26
PHENANTH	PHENANTHRENE	C14H10
MEFLOREN	1-METHYLFLUORENE	C14H12
FLORMEOH	9-FLUORENEMETHANOL	C14H12O
D-2C2BZS	4,6-DIMETHYLDIBENZOTHIOPHENE	C14H12S
D-C14H14	M,M*-BITOLYL	C14H14
ETPHCYHX	1-ETHYL-4-PHENYLCYCLOHEXANE	C14H20
D-HYPHEN	PHENANTHRENE,-TETRADECAHYDRO-	C14H24
THYDPHEN	CIS-ANTI-TRANS-TETRADECAHYDROPHENANTHRENE	C14H24
DIMDODEC	3,5-DIMETHYLDODECANE	C14H30
MEANTCEN	1-METHYLANTHRACENE	C15H12
MEPHENAN	4-METHYLPHENANTHRENE	C15H12
D-3CZOLE	1,3,7-TRIMETHYLCARBAZOLE	C15H15N
D-C15H16	BIPHENYL,-3,4*,5-TRIMETHYL-	C15H16
D-3C3BEN	1,3,5-TRIISOPROPYLBENZENE	C15H24
OCHYENAP	1,2,4A,5,6,7,8,8A-OCTAHYDRO-3-PROPYL-4-ETHYLNAPHTHALENE	C15H26
D-DIMTRD	6,8-DIMETHYLTRIDECANE	C15H32
DIMPHENA	1,4-DIMETHYLPHENANTHRENE	C16H14
CETANE	N-HEXADECANE	C16H34
CHRYSENE	CHRYSENE	C18H12
DIMEPYRE	4,5-DIMETHYLPYRENE	C18H14
RETENE	RETENE	C18H18
CYHXPHON	4'-(TRANS-4-N-BUTYLCYCLOHEXYL)ACETOPHENONE	C18H26O
CYHXHXBZ	1-CYCLOHEXYL-4-HEXYLBENZENE	C18H28
D-NC18	N-OCTADECANE	C18H38
D-C1C17	16-METHYLHEPTADECANE	C18H38
HMWLIGNA	HIGH-MW-LIGNIN-A	C20H26O8
D-NC20	N-EICOSANE	C20H42

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ID	Name	Formula
HMWLIGNB	HIGH-MW-LIGNIN-B	C21H26O8
DBZFLORE	13H-DIBENZO[A,G]FLUORENE	C21H14
TRIPHNPR	1,1,3-TRIPHENYL-1-PROPENE	C21H18
MOXHXPHE	(E)-1-(4-METHOXYPHENYL)-2-(4-HEXOXYPHENYL)ETHENE	C21H26O2
DIBZFLOR	PERHYDRO-1H-DIBENZO[A.I]FLUORENE	C21H34
BZACMOXP	1,3-BENZENE-CARBOXYLIC-ACID,BIS(4-METHOXYPHENYL)-ESTER	C22H18O6
HXACMOXP	1,4-CYCLOHEXANE-DICARBOXYLIC-ACID,TRANS-,DI(P-METHOXYPHENYL)-ESTER	C22H24O6
MOXHEPOX	4-METHOXY-4'-HEPTOXY-TRANS-STILBENE	C22H28O2
HXDECBEN	N-HEXADECYLBENZENE	C22H38
L-3MQUIN	QUINOLINE,-3-METHYL-	C10H9N
L-NC10	N-DECANE	C10H22
L-ACENPH	ACENAPHTHENE	C12H10
L-TMC9	2,2,4-TRIMETHYLNONANE	C12H26
L-C3PHEN	1-PROPYLPHENANTHRENE	C17H16
L-C4PHEN	1-BUTYLPHENANTHRENE	C18H18
L-C8NPH	1-OCTYLNAPHTHALENE	C18H24
L-NC14BZ	N-TETRADECYLBENZENE	C20H34
L-NC22	N-DOCOSANE	C22H46
V-23DMBT	2,3-DIMETHYLBENZOTHIOPHENE	C10H10S
V-NC12	N-DODECANE	C12H26
V-2MBZQN	3-METHYL-1-AZAPHENANTHRENE	C14H11N
V-C5DBT	PENTYLDIBENZOTHIOPHENE	C17H18S
V-C12NPH	DODECANE,-1-(1-NAPHTHYL)-	C22H32
V-HMCHRY	HEXAMETHYCHRYSENE	C24H24
V-NC33	N-TRITRIACONTANE	C33H68
P-IC4CY6	ISOBUTYLCYCLOHEXANE	C10H20
P-C12H20	PERHYDROACENAPHTHYLENE	C12H20
P-C14H20	1-ISOBUTYLTETRALIN	C14H20
P-3C3CY6	1,3,5-TRIISOPROPYLCYCLOHEXANE	C15H30
P-C17H20	4-(1,1-DIMETHYLPROPYL)-1,1*-BIPHENYL	C17H20
P-C17H24	1,2,3,4,41,9,10,10A-OCTAHYDRO-2,6,9-TRIMETHYLPHENANTHRENE	C17H24
P-NC17	N-HEPTADECANE	C17H36
P-C18H32	6-METHYL-13-(1-METHYLETHYL)TRICYCLO[8.4.0.02,6]TETRADECANE	C18H32
P-C18H34	2-OCTYLBICYCLO[4.4.0]DECANE	C18H34
P-C20H40	TETRADECYLCYCLOHEXANE	C20H40

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Appendix E. TEA Summary for CFP-A Oil Production

For equipment cost basis and other assumptions details please refer to the 2015 design report [1].

[1] 2015 Design Report, NREL/TP-5100-62455, www.nrel.gov/docs/fy15osti/62455.pdf

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Table E-1. CFP-A Equipment Cost Summary

Equip ID	Required	Spares	Equipment Name	Scaling Stream Flow (per unit)	Scaling Stream Flow (total)	New Stream Flow (total)	Units	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req+Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2020$	Scaled Uninstalled Cost in 2020$
C-101			Hopper Feeder						INCLUDED								
C-102			Screener Feeder Conveyor						INCLUDED								
C-103			Radial Stacker Conveyor						INCLUDED								
C-104			Dryer Feed Screw Conveyor						INCLUDED								
C-105			Biomass Feed Screw Conveyor						INCLUDED								
K-101	5		Flue Gas Blower	204	1,020	855	hp	0.84	$59,300	2013	$296,500	0.78	$258,339	1.94	$500,123	$525,601	$271,499
M-101			Hydraulic Truck Dump with Scale						INCLUDED								
M-102			Hammermill						INCLUDED								
M-103			Front End Loaders						INCLUDED								
M-104	2		Cross Flow Pellet Dryer	102,066	204,131	204,131	lb/h	1.00	$50,000	2011	$100,000	0.80	$100,000	2.00	$200,000	$203,586	$101,793
S-101			Magnetic Head Pulley						INCLUDED								
S-102			Vibratory Conveyor						INCLUDED								
T-101			Dump Hopper						INCLUDED								
T-102			Hammermill surge bin						INCLUDED								
T-103			Dryer Feed Bin						INCLUDED								
T-104			Dried Biomass Hopper						INCLUDED								
T-105			Lock Hopper						INCLUDED								
T-106			Feed Hopper						INCLUDED								
A100	1		Pinch Exchanger CAPEX							2008	$0		$0	2.64	$0	$0	$0
Area A100: Feed Handling & Drying											**$396,500**		**$358,339**	**1.95**	**$700,123**	**$729,187**	**$373,292**
C-212	2		Fines purge cooler						INCLUDED								
K-211	2		Fluidizing gas recycle compressor	3,648	7,296	7,059	hp	0.97	$4,700,000	2014	$9,400,000	0.90	$9,124,381	1.60	$14,599,009	$15,108,365	$9,442,728
K-212	2		Combustor air compressor	10,000	20,000	7,105	hp	0.36	$2,600,000	2014	$5,200,000	0.78	$2,319,567	1.60	$3,711,306	$3,840,793	$2,400,496
R-211	2		Fluidized Bed Reactor	1,222,474	2,444,948	5,295,003	ACFH	2.17	$3,449,500	2011	$6,899,000	0.50	$10,152,768	3.23	$32,793,440	$33,381,337	$10,334,779
R-212	2		Char Combustor	719,240	1,438,480	7,439,237	ACFH	5.17	$3,483,784	2011	$6,967,568	0.50	$15,845,046	3.06	$48,485,840	$49,355,059	$16,129,104
R-212C	2		Sand Cooler	3,671,901	7,343,802	78,466,071	Btu/h	10.68	$623,441	2011	$1,246,882	0.80	$8,295,339	3.00	$24,886,016	$25,332,154	$8,444,051
S-211	2		Primary Cyclone						INCLUDED								
S-212	2		Secondary Cyclone						INCLUDED								
S-213	2		Char Combustor Primary Cyclone						INCLUDED								
S-214	2		Char Combustor Secondary Cyclone						INCLUDED								
T-211	2		Bed Media Bin						INCLUDED								
T-212	2		Depleted bed media and ash storage bin						INCLUDED								
T-217	2		Char combustor bed media feed bin						INCLUDED								
A211	1		Pinch Exchanger CAPEX							2008	$1,637,964		$1,637,964	2.64	$4,316,635	$4,472,676	$1,697,174
Area A211: Ex Situ Fast Pyrolysis											**$31,351,414**		**$47,375,063**	**2.72**	**$128,792,246**	**$131,490,384**	**$48,448,333**

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Table E-1. CFP-A Equipment Cost Summary (continued)

Equip ID	Required	Spares	Equipment Name	Scaling Stream Flow (per unit)	Scaling Stream Flow (total)	New Stream Flow (total)	Units	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req+Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2020$	Scaled Uninstalled Cost in 2020$
C-262	1		Fines purge cooler						INCLUDED								
H-271	1		Secondary catalyst cooler (info only)			44,766,306	Btu/h										
K-262	1		Regenerator air compressor	10,000	10,000	4,491	hp	0.45	$2,600,000	2014	$2,600,000	0.78	$1,392,439	1.60	$2,227,903	$2,305,634	$1,441,021
R-261	1		Catalytic Fluidized Bed Reactor	2,534,057	2,534,057	6,755,997	lb/h	2.67	$3,818,400	2011	$3,818,400	0.50	$6,234,734	3.17	$19,764,108	$20,118,424	$6,346,506
R-262	1		Catalyst Regenerator	1,084,318	1,084,318	4,449,810	ACFH	4.10	$4,207,313	2011	$4,207,313	0.50	$8,523,093	3.06	$26,080,665	$26,548,220	$8,675,889
R-262C	1		Catalyst Cooler	272,000,000	272,000,000	116,155,148	Btu/h	0.43	$3,968,900	2011	$3,968,900	0.80	$2,009,305	3.00	$6,027,914	$6,135,978	$2,045,326
S-261	1		Primary Cyclone						INCLUDED								
S-262	1		Secondary Cyclone						INCLUDED								
S-263	1		Char Combustor Primary Cyclone						INCLUDED								
S-264	1		Char Combustor Secondary Cyclone						INCLUDED								
T-261	1		Catalyst steam stripper	2,756	2,756	2,756	lb/h	1.00	$347,005	2011	$347,005	0.30	$346,987	3.00	$1,040,962	$1,059,624	$353,208
T-262	1		Depleted catalyst and ash storage bin						INCLUDED								
T-267	1		Catalyst feed bin						INCLUDED								
A261	1		Pinch Exchanger CAPEX							2008	$686,194		$686,194	2.64	$1,808,372	$1,873,743	$710,999
R-261	2	1	VPU Fixed Bed Reactor	161,200	322,400	0	lb/h	0.00	$3,841,800	2013	$0	0.70	$0	1.21	$0	$0	$0
A261	1		Pinch Exchanger CAPEX							2008	$0		$0	2.64	$0	$0	$0
Area A261: Ex Situ Vapor Phase Upgrading											**$15,627,812**		**$19,192,752**	**2.97**	**$56,949,924**	**$58,041,622**	**$19,572,949**
D-301	1		Heavy Fraction Condenser Absorber		12.0	16.5	ft	1.90	$1,085,800	2013	$1,085,800	0.60	$1,596,809	1.40	$2,233,591	$2,347,377	$1,678,155
D-302	1		Light Fraction Condenser Absorber		13.0	16.6	ft	1.64	$1,085,300	2013	$1,085,300	0.60	$1,459,666	1.33	$1,935,237	$2,033,824	$1,534,026
H-313	1		D-301 Vapor ACHE		61,000,000	23,253,279	Btu/h	0.38	$578,900	2013	$578,900	0.60	$324,561	1.32	$427,609	$449,392	$341,095
H-314	1		D-302 Feed Cooler		12,067,840	41,342,788	Btu/h	3.43	$172,000	2013	$172,000	0.70	$407,256	2.03	$825,877	$867,950	$428,003
H-315	1		D-302 Feed/overhead exchanger		4,850,345	3,466,062	Btu/h	0.71	$758,400	2013	$758,400	0.70	$599,436	1.52	$911,563	$958,001	$629,973
H-316	1		D-302 Liquid feed chiller		10,138,754	22,957,983	Btu/h	2.26	$132,400	2013	$132,400	0.70	$234,613	2.33	$547,017	$574,884	$246,565
H-320	1		Heavy fraction ACHE		280,000	235,635	Btu/h	0.84	$33,300	2013	$33,300	0.60	$30,026	2.42	$72,765	$76,472	$31,555
H-320A	1		Heavy fraction cooler		845,500	195,013	Btu/h	0.23	$45,300	2013	$45,300	0.70	$16,224	3.23	$52,361	$55,029	$17,051
H-326	1	1	D-302 Quench circulation chiller		13,087,000	12,272,990	Btu/h	0.94	$78,300	2013	$156,600	0.70	$149,716	4.63	$693,896	$729,245	$157,343
P-318	1	1	D-301 Bottoms pump		56.0	37	gpm liq	0.66	$5,800	2013	$11,600	0.80	$8,282	4.38	$36,269	$38,117	$8,704
P-323	1	1	Circulation filter charge pump		41	32	hp	0.79	$25,400	2013	$50,800	0.80	$41,893	4.61	$193,302	$203,149	$44,027
P-329	1	1	Aqueous filter charge pump		182.0	149	gpm liq	0.82	$7,500	2013	$15,000	0.80	$12,786	4.20	$53,701	$56,437	$13,437
P-331	1	1	Organic product pump		124.0	79	gpm liq	0.64	$6,500	2013	$13,000	0.80	$9,057	4.63	$41,939	$44,075	$9,518
P-332	1	1	Quench recycle pump		0.53	0.06	hp	0.11	$7,800	2013	$15,600	0.80	$2,721	4.88	$13,289	$13,966	$2,859
S-319	1	1	D-301 Bottoms filter		52	34	gpm liq	0.66	$14,600	2013	$29,200	0.80	$21,004	1.84	$38,700	$40,672	$22,075
S-322	1		D-302 Bottoms decanter		1,013,050	832,689	lb/h	0.82	$294,700	2013	$294,700	0.70	$256,907	1.82	$467,960	$491,800	$269,995
S-324	2	1	Quench circulation filter		2,340.0	1,789	gpm liq	0.76	$83,100	2013	$249,300	0.80	$201,110	1.71	$343,896	$361,415	$211,356
S-329	1	1	Aqueous phase filter		182.0	149	gpm liq	0.82	$26,500	2013	$53,000	0.80	$45,177	1.77	$79,955	$84,028	$47,478
T-328	1		Organic product surge tank		1,000	633	ACFH	0.63	$48,000	2013	$48,000	0.70	$34,854	3.24	$112,985	$118,741	$36,629
T-332	1		D-301 Circulation surge tank		1,000	1,366	ACFH	1.37	$48,000	2013	$48,000	0.70	$59,718	3.24	$193,586	$203,448	$62,760
A310	1		Pinch exchanger CAPEX							2008	$1,153,458		$1,153,458	2.64	$3,039,784	$3,149,668	$1,195,154
Area A310: Product Condensation and Separation											**$6,029,658**		**$6,665,274**	**1.85**	**$12,315,283**	**$12,897,690**	**$6,987,759**

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Table E-1. CFP-A Equipment Cost Summary (continued)

Equip ID	Required	Spares	Equipment Name	Scaling Stream Flow (per unit)	Scaling Stream Flow (total)	New Stream Flow (total)	Units	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req+Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2020$	Scaled Uninstalled Cost in 2020$
H-568A	1		WGS Effluent ACHE		5,740,000	6	Btu/h	0.00	$76,300	2013	$76,300	0.60	$19	1.63	$31	$32	$20
H-568C	1		WGS Effluent cooler		3,159,000	11	Btu/h	0.00	$32,600	2013	$32,600	0.70	$5	3.60	$18	$19	$5
K-511	1		Pyrolysis offgas compressor to reformer		3,648	0	hp	0.00	$4,700,000	2014	$4,700,000	0.60	$1,012	1.60	$1,619	$1,675	$1,047
K-545	1		Reformer combustor air blower						INCLUDED								
K-580	1		PSA Compressor multistage						INCLUDED								
M-575	1		PSA						INCLUDED								
P-576	1		Compressor condensate pump		50.0	0	gpm liq	0.00	$5,800	2013	$5,800	0.80	$0	4.38	$0	$0	$0
R-530	1		Steam reformer package (PNNL)		49	0.0	MMSCFD	0.00	$66,875,000	2007	$34,830,729	0.65	$9,114	1.92	$17,498	$19,856	$10,342
R-565	1		WGS reactor						INCLUDED								
R-566	1		LTS reactor		354,424	0	lb/h	0.00	$1,243,000	2013	$1,243,000	0.60	$312	1.42	$442	$465	$328
S-575	1		Compressor suction KO drum						INCLUDED								
S-586	1		PSA Compressor discharge KO drum						INCLUDED								
T-576	1		Compressor condensate tank		1,000	0	ACFH	0.00	$48,000	2013	$48,000	0.70	$1	3.24	$3	$3	$1
A500			Pinch exchanger CAPEX							2008	$4,113,627		$4,113,627	2.64	$10,840,917	$11,232,802	$4,262,330
Area A500: Off-Gas Utilization											**$45,050,057**		**$4,124,090**	**2.63**	**$10,860,528**	**$11,254,853**	**$4,274,073**
H-601	1		Steam turbine condenser		1.00	10.00	Btu/h	10.00	$1,300,000	2010	$1,300,000	1.00	$13,000,000	1.40	$18,200,000	$19,700,145	$14,071,532
H-603	1		Blowdown cooler		381,671	640,880	Btu/h	1.68	$16,780	2007	$16,780	0.65	$23,502	4.32	$101,542	$115,225	$26,669
M-601	1		BFW electrodeionization (EDI)+polishing		300,000	537,581	Btu/h	1.79	$1,325,000	2010	$1,325,000	0.60	$1,880,224	2.00	$3,760,448	$4,070,405	$2,035,202
M-602A	1		Extraction steam turbine / generator Stage 1		-40,418	-26,212	hp	0.65	$7,700,000	2010	$7,700,000	0.70	$5,686,370	1.80	$10,235,465	$11,079,129	$6,155,072
M-602B	1		Extraction steam turbine / generator Stage 2		-40,418	-72,315	hp	1.79	$7,700,000	2010	$7,700,000	0.70	$11,570,305	1.80	$20,826,549	$22,543,189	$12,523,994
M-603	1		Startup boiler		204,131	204,131	lb/h	1.00	$275,500	2013	$275,500	0.60	$275,500	1.69	$466,519	$490,285	$289,535
P-601	1	1	Makeup pump		80,411	10,748	lb/h	0.13	$6,528	2007	$13,056	0.30	$7,139	4.72	$33,681	$38,220	$8,101
P-602	1	1	Condensate pump		247,010	537,581	lb/h	2.18	$9,810	2007	$19,620	0.30	$24,775	4.61	$114,153	$129,536	$28,114
P-603	1	1	EDI pump		247,010	537,581	lb/h	2.18	$9,810	2007	$19,620	0.30	$24,775	4.61	$114,153	$129,536	$28,114
P-604	1	1	Boiler feed water pump		494,622	537,581	lb/h	1.09	$304,578	2007	$609,156	0.30	$624,568	1.35	$844,848	$958,695	$708,731
P-605	1	1	Condensate collection pump		247,010	537,581	lb/h	2.18	$9,810	2007	$19,620	0.30	$24,775	4.61	$114,153	$129,536	$28,114
S-601	1		Blowdown flash drum		9,892	10,752	lb/h	1.09	$47,205	2007	$47,205	0.65	$49,831	3.41	$170,168	$193,099	$56,546
T-601	1		Condensate collection tank		500,400	537,581	lb/h	1.07	$28,505	2007	$28,505	0.65	$29,864	6.83	$203,985	$231,473	$33,889
T-602	1		Condensate surge tank		500,400	537,581	lb/h	1.07	$27,704	2007	$27,704	0.65	$29,025	6.51	$189,003	$214,472	$32,936
T-603	1		Deaerator		494,619	537,581	lb/h	1.09	$53,299	2007	$53,299	0.65	$56,264	5.07	$285,126	$323,548	$63,846
T-603A	1		Deaerator packed column		494,619	537,581	lb/h	1.09	$18,405	2007	$18,405	0.65	$19,429	5.18	$100,602	$114,158	$22,047
T-605	1		Steam drum		494,622	537,585	lb/h	1.09	$104,100	2007	$104,100	0.65	$109,891	2.28	$250,291	$284,018	$124,700
T-606	1		Steam turbine condensate tank		500,400	537,581	lb/h	1.07	28,505	2007	$28,505	0.65	$29,864	6.83	$203,985	$231,473	$33,889
A600			Pinch exchanger CAPEX							2008	$6,119,429		$6,119,429	2.64	$16,126,938	$16,709,907	$6,340,639
Area A600: Steam System & Power Generation											**$25,425,504**		**$39,585,531**	**1.83**	**$72,341,608**	**$77,686,048**	**$42,611,669**

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Table E-1. CFP-A Equipment Cost Summary (continued)

Equip ID	Required	Spares	Equipment Name	Scaling Stream Flow (per unit)	Scaling Stream Flow (total)	New Stream Flow (total)	Units	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req+Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2020$	Scaled Uninstalled Cost in 2020$
K-701	2	1	Plant air compressor		262,454	204,131	lb/h	0.78	$87,922	2007	$263,766	0.30	$244,611	1.57	$383,656	$435,356	$277,573
M-701	1		Cooling tower system		7,506,000	3,618,114	lb/h	0.48	$260,852	2010	$260,852	0.78	$147,636	2.47	$364,660	$394,718	$159,805
M-702	1		Hydraulic truck dump with scale		367,437	204,131	lb/h	0.56	$80,000	1998	$80,000	0.60	$56,225	2.47	$138,875	$212,573	$86,062
M-703	1		Flue gas scrubber		489,600	1,592,641	lb/h	3.25	$436,250	2010	$436,250	0.65	$939,107	2.47	$2,319,595	$2,510,789	$1,016,514
M-704	1		Flue gas stack		939,119	1,592,641	lb/h	1.70	$169,187	2007	$169,187	0.65	$238,493	1.30	$309,839	$351,591	$270,631
M-708	1		Chilled water system		28,200,000	35,230,974	Btu/h	1.25	$637,500	2011	$637,500	0.60	$728,594	1.80	$1,311,469	$1,334,980	$741,656
M-710	1		Product loading rack		1	1			$25,000	2011	$25,000	1.00	$25,000	2.47	$61,750	$62,857	$25,448
P-701	1	1	Cooling water pump		7,001,377	3,618,114	lb/h	0.52	$239,375	2007	$478,750	0.30	$392,734	2.14	$840,512	$953,775	$445,657
P-702	1	1	Firewater pump		262,454	204,131	lb/h	0.78	$23,043	2007	$46,086	0.30	$42,739	3.70	$158,025	$179,320	$48,498
P-703	1	1	Diesel pump		262,454	204,131	lb/h	0.78	$3,842	2007	$7,684	0.30	$7,126	5.39	$38,393	$43,567	$8,086
P-704	1	1	Ammonia pump		262,454	204,131	lb/h	0.78	$3,842	2007	$7,684	0.30	$7,126	5.21	$37,095	$42,094	$8,086
P-705	1	1	Caustic pump		1	1			$4,906	2007	$9,812	0.30	$9,812	4.30	$42,200	$47,887	$11,134
P-707	1	1	BFW chemical pump		262,454	204,131	lb/h	0.78	$3,842	2007	$7,684	0.30	$7,126	5.21	$37,095	$42,094	$8,086
P-708	1	1	Flue gas scrubber circulation pump		489,600	1,592,641	lb/h	3.25	$12,510	2007	$25,020	0.30	$35,643	4.12	$146,730	$166,503	$40,446
P-790	1	1	Gasoline product pump		108.0	0	gpm liq	0.00	$7,400	2013	$14,800	0.80	$0	3.96	$0	$0	$0
P-792	1	1	Diesel product pump		108.0	0	gpm liq	0.00	$7,400	2013	$14,800	0.80	$0	3.96	$0	$0	$0
S-701	1	1	Instrument air dryer		262,454	204,131	lb/h	0.78	$8,349	2002	$16,698	0.60	$14,361	2.47	$35,471	$53,458	$21,643
T-701	1		Plant air receiver		262,454	204,131	lb/h	0.78	$21,005	2007	$21,005	0.65	$17,839	5.44	$97,074	$110,155	$20,243
T-702	1		Firewater storage tank		262,454	204,131	lb/h	0.78	$229,900	2007	$229,900	0.65	$195,252	1.46	$285,447	$323,912	$221,563
T-703	1		Purchased diesel storage tank		262,454	204,131	lb/h	0.78	$104,674	2007	$104,674	0.65	$88,899	1.35	$119,665	$135,791	$100,878
T-704	1		Ammonia storage tank		262,454	204,131	lb/h	0.78	$15,704	2007	$15,704	0.65	$13,337	5.39	$71,850	$81,532	$15,135
T-705	1		Caustic storage tank		262,454	204,131	lb/h	0.78	$16,005	2007	$16,005	0.65	$13,593	3.01	$40,936	$46,452	$15,425
T-708	1		BFW chemical storage tank		262,454	204,131	lb/h	0.78	$22,004	2007	$22,004	0.65	$18,688	6.70	$125,186	$142,055	$21,206
T-790	1	1	Gasoline product storage tanks		50.0	0	gpm liq	0.00	$442,700	2013	$885,400	0.70	$885,400	1.75	$1,553,400	$1,632,535	$930,505
T-792	1	1	Diesel product storage tank		50.0	0	gpm liq	0.00	$442,700	2013	$885,400	0.70	$885,400	1.75	$1,553,400	$1,632,535	$930,505
Area A700: Cooling Water & Other Utilities											**$4,681,665**		**$5,014,740**	**2.01**	**$10,072,324**	**$10,936,527**	**$5,424,784**

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Table E-1. CFP-A Equipment Cost Summary (continued)

Equip ID	Required	Spares	Equipment Name	Scaling Stream Flow (per unit)	Scaling Stream Flow (total)	New Stream Flow (total)	Units	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Original Equip Cost (Req+Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2020$	Scaled Uninstalled Cost in 2020$
H-840	1		RTO Feed/Effluent exchanger						INCLUDED								
K-831	1		RTO blower		900	7,880	hp	8.76	$360,420	2009	$360,420	0.60	$1,324,864	1.60	$2,119,783	$2,421,565	$1,513,478
M-805	1		Clarifier		220	220			$96,221	2007	$96,221	0.65	$96,221	2.41	$232,300	$263,603	$109,187
M-806	1		Belt press		1	1			$135,000	2010	$135,000	0.65	$135,000	2.47	$333,450	$360,935	$146,127
M-807	1		Sand filter		265	265			$120,400	2010	$120,400	0.65	$120,400	2.47	$297,388	$321,900	$130,324
M-833	1		Aqueous regenerative thermal oxidizer		18,500	152,853	SCFM	8.26	$1,850,000	2002	$1,850,000	0.60	$6,568,005	2.40	$15,763,212	$23,756,388	$9,898,495
M-861	1		Reverse osmosis unit		197,000	30,122	lb/h	0.15	$350,000	2012	$350,000	0.70	$94,008	1.15	$108,109	$110,254	$95,873
P-801	1	1	Clarifier overflow pump		50	18	gpm liq	0.36	$4,500	2013	$9,000	0.80	$3,967	4.38	$17,365	$18,250	$4,169
P-809	1	1	Slurry pump		1	1			$3,242	2007	$3,242	0.80	$3,242	3.10	$10,050	$11,405	$3,679
P-810	1	1	Boiler charge pump		200	0	gpm liq	0.00	$6,500	2013	$13,000	0.80	$0	4.14	$1	$1	$0
P-812	1	1	Condensate Pump		140	0	gpm liq	0.00	$5,100	2013	$10,200	0.80	$0	4.59	$0	$0	$0
P-814	1	1	Boiler charge pump		145	0	gpm liq	0.00	$5,100	2013	$10,200	0.80	$1	4.59	$4	$4	$1
P-815	1	1	Boiler makeup pump		5	0	gpm liq	0.00	$14,700	2013	$29,400	0.80	$26	1.84	$47	$50	$27
P-822	1	1	Concentrated waste water pump		7	0	gpm liq	0.00	$4,900	2013	$9,800	0.80	$5	4.31	$21	$22	$5
P-825	1	1	RTO Feed pump		185	158	gpm liq	0.85	$6,400	2013	$12,800	0.80	$11,291	3.84	$43,400	$45,611	$11,866
P-860	1	1	RO Feed pump		185	72	gpm liq	0.39	$6,400	2013	$12,800	0.80	$6,004	3.84	$23,077	$24,252	$6,310
P-866	1	1	Clean water pump		140	54	gpm liq	0.38	$5,200	2013	$5,200	0.80	$2,422	4.35	$10,526	$11,062	$2,545
T-801	1		Clarifier overflow tank		1	1		1.00	$85,800	2013	$85,800	0.65	$85,800	2.27	$195,000	$204,934	$90,171
T-809	1		Slurry tank		1	1		1.00	$85,800	2013	$85,800	0.65	$85,800	2.27	$195,000	$204,934	$90,171
T-810	1		Boiler Feed Tank		200	0	gpm liq	0.00	$24,600	2013	$24,600	0.70	$1	4.49	$6	$6	$1
T-812	1		Steam drum		100,800	#VALUE!	ft3/h vap	0.00	$24,400	2013	$24,400	0.70	$0	4.46	$0	$0	$0
T-814	1		Boiler feed tank		145	0	gpm liq	0.00	$24,400	2013	$24,400	0.70	$6	4.46	$28	$29	$6
T-822	1		Steam drum		156,000	1	ft3/h vap	0.00	$19,400	2013	$19,400	0.70	$3	5.23	$17	$17	$3
T-860	1		RO Feed tank		185	72	gpm liq	0.39	$76,000	2013	$76,000	0.70	$39,185	2.24	$87,651	$92,116	$41,181
T-866	1		Clean water tank		140	54	gpm liq	0.38	$194,200	2013	$194,200	0.70	$194,200	2.23	$434,000	$456,109	$204,093
A800	1		Pinch exchanger CAPEX							2008	$950,884		$950,884	2.64	$2,505,929	$2,596,515	$985,258
Area A800: Water Management											**$4,513,167**		**$9,721,335**	**2.30**	**$22,376,362**	**$30,899,962**	**$13,332,971**
Plant Total											**$140,284,676**		**$132,037,124**	**2.38**	**$314,408,399**	**$333,936,273**	**$141,025,830**

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Table E-2. CFP-A Operating Cost Summary

Raw Material	Scaling Flow		Quoted Price		Year of Price Quote	2020 Cost	2020 Cost ($/h)	2020 Cost (MM$/yr)	2020 Cost (¢/GGE)
Variable Operating Costs									
Feedstock, Catalyst & Chemicals									
Feedstock - Wood Chips	102	US ton/h	$ 60.33	/US ton	2016	$ 65.27	6,662.17	52.525	166.72
Sand Makeup	126	lb/h	$ 0.21	/lb	2011	$ 0.24	30.02	0.237	0.75
VPU Zeolite Catalyst	229	lb/h	$ 1.89	/lb	2016	$ 2.05	467.88	3.689	11.71
CFP Zeolite Catalyst	-	lb/h	$ 1.89	/lb	2016	$ 2.05	-	-	-
Natural Gas to SMR	0	lb/h	$ 0.12	/lb	2011	$ 0.14	0.01	0.000	0.00
50 wt% Caustic	1,565	lb/h	$ 0.08	/lb	2010	$ 0.09	145.14	1.144	3.63
Boiler Chemicals	10,752	lb/h blowdn	$ 75.00	/MM lb	2014	$ 80.52	0.87	0.007	0.02
Cooling Tower Chemicals	6,489	ton cool	$ 33.84	/ton cool/yr	2014	$ 36.33	29.90	0.236	0.75
Regeneration Gas: FP	0	lb/h	$ 0.12	/lb	2011	$ 0.14	0.00	0.000	0.00
Regeneration Gas: CFP	-	lb/h	$ 0.12	/lb	2011	$ 0.14	-	-	-
Regeneration Gas: VPU	0	lb/h	$ 0.12	/lb	2011	$ 0.14	0.00	0.000	0.00
Net Water Makeup	93,097	lb/h	$ 19.96	lb/h	2001	$ 36.41	16.95	0.134	0.42
Diesel Fuel	71	lb/h	$ 0.40	/lb	2011	$ 0.46	32.75	0.258	0.82
Bio-Oil Transportation (800 miles)	37,387	lb/h	$ 40.80	/ton/800 miles	2007	$ 58.00	1,084.20	8.548	27.13
Subtotal							**8,469.89**	**66.777**	**211.96**
Waste Streams									
FP Purge	2,101	lb/h	$ 0.016	/lb	1998	$ 0.032	66.67	0.526	1.67
CFP Purge	-	lb/h	$ 0.016	/lb	1998	$ 0.032	-	-	-
VPU Purge	262	lb/h	$ 0.016	/lb	1998	$ 0.032	8.31	0.065	0.21
Quench Filter Solids	35	lb/h	$ 0.016	/lb	1998	$ 0.032	1.13	0.009	0.03
WWT Cost	7,531	lb/h	$ 0.003	/lb	2011	$ 0.003	22.79	0.180	0.57
Subtotal							**98.89**	**0.780**	**2.47**
Coproduct Credits									
Electricity	41,049	kW	$ 5.852	/kWh	2020	$ 5.852	(2,402)	(18.94)	(60.12)
Subtotal							**(2,402.16)**	**(18.94)**	**(60.12)**
Total Variable Operating Costs							**6,166.62**	**48.618**	**154.32**
Fixed Operating Costs									
	2007 Salary	**# Positions**	**2020 Cost**		**2020 Hourly Wage**				
Plant Manager	$ 147,000	1	$ 195,724		$ 70.67				
Plant Engineer	$ 70,000	1	$ 93,202		$ 33.65				
Maintenance Supr	$ 57,000	1	$ 75,893		$ 27.40				
Lab Manager	$ 56,000	1	$ 74,562		$ 26.92				
Shift Supervisor	$ 48,000	5	$ 319,550		$ 23.08				
Lab Technician	$ 40,000	2	$ 106,517		$ 19.23				
Maintenance Tech	$ 40,000	16	$ 852,133		$ 19.23				
Shift Operators	$ 48,000	20	$ 1,278,199		$ 23.08				
Yard Employees	$ 28,000	12	$ 447,370		$ 13.46				
Clerks & Secretaries	$ 36,000	3	$ 143,797		$ 17.31				
Total Salaries		**62**	**$ 3,586,947**					**3.587**	**11.39**
					Avg Salary (w/ Benefits)				
Overhead and Benefits	% Supervison =	90.0%	$3,228,253		$52.85	per Hour		3.228	10.25
Maintenance	% of FCI =	3.0%	$17,832,992		$109,923	per Year		17.833	56.61
Insurance & Taxes	% of FCI =	0.7%	$4,161,031					4.161	13.21
Total Fixed Operating Costs								**28.809**	**91.45**
Total Operating Costs								**77.427**	**245.77**

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Table E-3. Discounted Cash Flow Rate of Return (DCFROR) for CFP-A

Year		−2	−1	0	1	2	3	4	5	6	7	8
Fixed Capital Investment		$20,631,858	$142,663,936	$76,087,432								
Working Capital				$29,721,653								
Loan Payment					$53,152,833	$53,152,833	$53,152,833	$53,152,833	$53,152,833	$53,152,833	$53,152,833	$53,152,833
Loan Interest Payment		$2,282,623	$19,402,295	$28,532,787	$28,532,787	$26,563,183	$24,436,011	$22,138,666	$19,657,532	$16,977,908	$14,083,914	$10,958,401
Loan Principal		$28,532,787	$242,528,691	$356,659,839	$332,039,793	$305,450,143	$276,733,321	$245,719,153	$212,223,852	$176,048,926	$136,980,007	$94,785,574
Fuel Product Sales					$118,191,608	$157,588,811	$157,588,811	$157,588,811	$157,588,811	$157,588,811	$157,588,811	$157,588,811
Byproduct Credit					$14,203,987	$18,938,650	$18,938,650	$18,938,650	$18,938,650	$18,938,650	$18,938,650	$18,938,650
Total Annual Sales					$132,395,595	$176,527,460	$176,527,460	$176,527,460	$176,527,460	$176,527,460	$176,527,460	$176,527,460
Annual Manufacturing Cost												
Raw Materials					$45,958,974	$52,524,541	$52,524,541	$52,524,541	$52,524,541	$52,524,541	$52,524,541	$52,524,541
Catalyst Costs		$0	$0	$0	$358,793	$2	$2	$2	$2	$2	$2	$2
Other Variable Costs					$13,152,747	$15,031,711	$15,031,711	$15,031,711	$15,031,711	$15,031,711	$15,031,711	$15,031,711
Fixed Operating Costs					$28,809,223	$28,809,223	$28,809,223	$28,809,223	$28,809,223	$28,809,223	$28,809,223	$28,809,223
Total Product Cost					$88,279,737	$96,365,477	$96,365,477	$96,365,477	$96,365,477	$96,365,477	$96,365,477	$96,365,477
Annual Depreciation												
General Plant % Depreciation					14.3%	24.5%	17.5%	12.5%	8.9%	8.9%	8.9%	4.5%
General Plant Depreciation					$59,278,079	$101,589,933	$72,552,386	$51,811,281	$37,043,614	$37,002,132	$37,043,614	$18,501,066
General Plant Remaining					$355,544,025	$253,954,092	$181,401,706	$129,590,425	$92,546,811	$55,544,680	$18,501,066	$0
Steam Plant % Depreciation					3.75%	7.22%	6.68%	6.18%	5.71%	5.29%	4.89%	4.52%
Steam Plant Depreciation					$6,735,411	$12,966,115	$11,992,624	$11,094,569	$10,261,174	$9,492,439	$8,779,384	$8,122,008
Steam Plant Remaining					$172,875,550	$159,909,435	$147,916,811	$136,822,242	$126,561,068	$117,068,628	$108,289,245	$100,167,237
Total Depreciation					$66,013,490	$114,556,049	$84,545,010	$62,905,850	$47,304,788	$46,494,571	$45,822,998	$26,623,074
Net Revenue					-$50,430,419	-$60,957,249	-$28,819,038	-$4,882,532	$13,199,663	$16,689,504	$20,255,071	$42,580,509
Losses Forward						-$50,430,419	-$111,387,667	-$140,206,706	-$145,089,238	-$131,889,575	-$115,200,071	-$94,945,000
Taxable Income					-$50,430,419	-$111,387,667	-$140,206,706	-$145,089,238	-$131,889,575	-$115,200,071	-$94,945,000	-$52,364,491
Income Tax					$0	$0	$0	$0	$0	$0	$0	$0
Annual Cash Income					-$9,036,975	$27,009,150	$27,009,150	$27,009,150	$27,009,150	$27,009,150	$27,009,150	$27,009,150
Discount Factor		1.210	1.100	1.000	0.909	0.826	0.751	0.683	0.621	0.564	0.513	0.467
Annual Present Value	$338,545,677				-$8,215,432	$22,321,611	$20,292,374	$18,447,613	$16,770,557	$15,245,961	$13,859,964	$12,599,968
Total Capital Investment + Interest		$27,726,522	$178,272,854	$134,341,873								
Net Present Worth				$0								

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Table E-3. DCFROR for CFP-A (continued)

Year	9	10	11	12	13	14	15	16	17	18	19
Fixed Capital Investment											
Working Capital											
Loan Payment	$53,152,833	$53,152,833	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$7,582,846	$3,937,247	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$49,215,587	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Product Sales	$157,588,811	$157,588,811	$157,588,811	$157,588,811	$157,588,811	$157,588,811	$157,588,811	$157,588,811	$157,588,811	$157,588,811	$157,588,811
Byproduct Credit	$18,938,650	$18,938,650	$18,938,650	$18,938,650	$18,938,650	$18,938,650	$18,938,650	$18,938,650	$18,938,650	$18,938,650	$18,938,650
Total Annual Sales	$176,527,460	$176,527,460	$176,527,460	$176,527,460	$176,527,460	$176,527,460	$176,527,460	$176,527,460	$176,527,460	$176,527,460	$176,527,460
Annual Manufacturing Cost											
Raw Materials	$52,524,541	$52,524,541	$52,524,541	$52,524,541	$52,524,541	$52,524,541	$52,524,541	$52,524,541	$52,524,541	$52,524,541	$52,524,541
Catalyst Costs	$2	$2	$2	$2	$2	$2	$2	$2	$2	$2	$2
Other Variable Costs	$15,031,711	$15,031,711	$15,031,711	$15,031,711	$15,031,711	$15,031,711	$15,031,711	$15,031,711	$15,031,711	$15,031,711	$15,031,711
Fixed Operating Costs	$28,809,223	$28,809,223	$28,809,223	$28,809,223	$28,809,223	$28,809,223	$28,809,223	$28,809,223	$28,809,223	$28,809,223	$28,809,223
Total Product Cost	$96,365,477	$96,365,477	$96,365,477	$96,365,477	$96,365,477	$96,365,477	$96,365,477	$96,365,477	$96,365,477	$96,365,477	$96,365,477
Annual Depreciation											
General Plant % Depreciation											
General Plant Depreciation											
General Plant Remaining											
Steam Plant % Depreciation	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%
Steam Plant Depreciation	$8,014,241	$8,012,445	$8,014,241	$8,012,445	$8,014,241	$8,012,445	$8,014,241	$8,012,445	$8,014,241	$8,012,445	$8,014,241
Steam Plant Remaining	$92,152,996	$84,140,551	$76,126,310	$68,113,865	$60,099,624	$52,087,179	$44,072,938	$36,060,493	$28,046,252	$20,033,807	$12,019,566
Total Depreciation	$8,014,241	$8,012,445	$8,014,241	$8,012,445	$8,014,241	$8,012,445	$8,014,241	$8,012,445	$8,014,241	$8,012,445	$8,014,241
Net Revenue	$64,564,896	$68,212,291	$72,147,742	$72,149,538	$72,147,742	$72,149,538	$72,147,742	$72,149,538	$72,147,742	$72,149,538	$72,147,742
Losses Forward	-$52,364,491	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$12,200,405	$68,212,291	$72,147,742	$72,149,538	$72,147,742	$72,149,538	$72,147,742	$72,149,538	$72,147,742	$72,149,538	$72,147,742
Income Tax	$2,562,085	$14,324,581	$15,151,026	$15,151,403	$15,151,026	$15,151,403	$15,151,026	$15,151,403	$15,151,026	$15,151,403	$15,151,026
Annual Cash Income	$24,447,065	$12,684,568	$65,010,957	$65,010,580	$65,010,957	$65,010,580	$65,010,957	$65,010,580	$65,010,957	$65,010,580	$65,010,957
Discount Factor	0.424	0.386	0.350	0.319	0.290	0.263	0.239	0.218	0.198	0.180	0.164
Annual Present Value	$10,367,942	$4,890,450	$22,785,944	$20,714,374	$18,831,359	$17,119,318	$15,563,106	$14,148,196	$12,862,071	$11,692,724	$10,629,811

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Table E-3. DCFROR for CFP-A (continued)

Year	20	21	22	23	24	25	26	27	28	29	30
Fixed Capital Investment											
Working Capital											-$31,331,653
Loan Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Product Sales	$157,588,811	$157,588,811	$157,588,811	$157,588,811	$157,588,811	$157,588,811	$157,588,811	$157,588,811	$157,588,811	$157,588,811	$157,588,811
Byproduct Credit	$18,938,650	$18,938,650	$18,938,650	$18,938,650	$18,938,650	$18,938,650	$18,938,650	$18,938,650	$18,938,650	$18,938,650	$18,938,650
Total Annual Sales	$176,527,460	$176,527,460	$176,527,460	$176,527,460	$176,527,460	$176,527,460	$176,527,460	$176,527,460	$176,527,460	$176,527,460	$176,527,460
Annual Manufacturing Cost											
Raw Materials	$52,524,541	$52,524,541	$52,524,541	$52,524,541	$52,524,541	$52,524,541	$52,524,541	$52,524,541	$52,524,541	$52,524,541	$52,524,541
Catalyst Costs	$2	$2	$2	$2	$2	$2	$2	$2	$2	$2	$2
Other Variable Costs	$15,031,711	$15,031,711	$15,031,711	$15,031,711	$15,031,711	$15,031,711	$15,031,711	$15,031,711	$15,031,711	$15,031,711	$15,031,711
Fixed Operating Costs	$28,809,223	$28,809,223	$28,809,223	$28,809,223	$28,809,223	$28,809,223	$28,809,223	$28,809,223	$28,809,223	$28,809,223	$28,809,223
Total Product Cost	$96,365,477	$96,365,477	$96,365,477	$96,365,477	$96,365,477	$96,365,477	$96,365,477	$96,365,477	$96,365,477	$96,365,477	$96,365,477
Annual Depreciation											
General Plant % Depreciation											
General Plant Depreciation											
General Plant Remaining											
Steam Plant % Depreciation	4.46%	2.23%									
Steam Plant Depreciation	$8,012,445	$4,007,121									
Steam Plant Remaining	$4,007,121	$0									
Total Depreciation	$8,012,445	$4,007,121									
Net Revenue	$72,149,538	$76,154,863	$80,161,983	$80,161,983	$80,161,983	$80,161,983	$80,161,983	$80,161,983	$80,161,983	$80,161,983	$80,161,983
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$72,149,538	$76,154,863	$80,161,983	$80,161,983	$80,161,983	$80,161,983	$80,161,983	$80,161,983	$80,161,983	$80,161,983	$80,161,983
Income Tax	$15,151,403	$15,992,521	$16,834,016	$16,834,016	$16,834,016	$16,834,016	$16,834,016	$16,834,016	$16,834,016	$16,834,016	$16,834,016
Annual Cash Income	$65,010,580	$64,169,462	$63,327,967	$63,327,967	$63,327,967	$63,327,967	$63,327,967	$63,327,967	$63,327,967	$63,327,967	$63,327,967
Discount Factor	0.149	0.135	0.123	0.112	0.102	0.092	0.084	0.076	0.069	0.063	0.057
Annual Present Value	$9,663,408	$8,671,256	$7,779,586	$7,072,351	$6,429,410	$5,844,918	$5,313,562	$4,830,511	$4,391,373	$3,992,158	$3,629,234
Total Capital Investment + Interest											-$1,795,572

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Appendix F. TEA Summary for CFP-B Oil Production

For equipment cost basis and other details please refer to previous references [1,2].

[1] 2015 Design Report, NREL/TP-5100-62455, www.nrel.gov/docs/fy15osti/62455.pdf

[2] Topics in Catalysis 59 (1): 2–18. doi.org/10.1007/s11244-015-0500-z.

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Table F-1. CFP-B Equipment Cost Summary

Equip ID	Required	Spares	Equipment Name	Design Flow (per unit)	Design Flow (total)	Actual Flow (total)	Units	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Orig Equip Cost (Req+Spare) in Base Year	Scaling Exp.	Scaled Cost in Base Year	Install Factor	Installed Cost in Base Year	Installed Cost in 2020$	Scaled Uninstalled Cost in 2020$
C-101			Hopper Feeder						INCLUDED								
C-102			Screener Feeder Conveyor						INCLUDED								
C-103			Radial Stacker Conveyor						INCLUDED								
C-104			Dryer Feed Screw Conveyor						INCLUDED								
C-105			Biomass Feed Screw Conveyor						INCLUDED								
K-101	3		Flue Gas Blower	204	612	496	HP	0.81	$59,300	2013	$177,900	0.78	$151,121	1.94	$292,557	$307,461	$158,819
M-101			Hydraulic Truck Dump with Scale						INCLUDED								
M-102			Hammermill						INCLUDED								
M-103			Front End Loaders						INCLUDED								
M-104	2		Cross Flow Pellet Dryer	102,066	204,131	204,131	LB/HR	1.00	$50,000	2011	$100,000	0.80	$100,000	2.00	$200,000	$203,586	$101,793
S-101			Magnetic Head Pulley						INCLUDED								
S-102			Vibratory Conveyor						INCLUDED								
T-101			Dump Hopper						INCLUDED								
T-102			Hammermill surge bin						INCLUDED								
T-103			Dryer Feed Bin						INCLUDED								
T-104			Dried Biomass Hopper						INCLUDED								
T-105			Lock Hopper						INCLUDED								
T-106			Feed Hopper						INCLUDED								
A100	1		Pinch Exchanger CAPEX							2008	$0		$0	2.64	$0	$0	$0
Area A100: Feed Handling & Drying ↑											**$277,900**		**$251,121**	**1.96**	**$492,558**	**$511,047**	**$260,612**
C-212	2		Fines purge cooler						INCLUDED								
K-211	2		Fluidizing gas recycle compressor	3,648	7,296	3,964	HP	0.54	$4,700,000	2014	$9,400,000	0.90	$5,428,810	1.60	$8,686,095	$8,989,151	$5,618,220
K-212	2		Combustor air compressor	10,000	20,000	12,512	HP	0.63	$2,600,000	2014	$5,200,000	0.78	$3,606,713	1.60	$5,770,741	$5,972,081	$3,732,551
R-211	2		Fluidized Bed Reactor	1,222,474	2,444,948	2,513,102	CUFT/HR	1.03	$3,449,500	2011	$6,899,000	0.50	$6,994,495	4.01	$28,047,926	$28,550,749	$7,119,888
R-212	2		Char Combustor	719,240	1,438,480	1,316,739	CUFT/HR	0.92	$3,483,784	2011	$6,967,568	0.50	$6,666,213	3.97	$26,459,662	$26,934,012	$6,785,720
R-212C	2		Sand Cooler	3,671,901	7,343,802	0	BTU/HR	0.00	$623,441	2011	$0	0.80	$0	3.00	$0	$0	$0
R-212C	2		Sand Cooler (alternate)	52,000,000	104,000,000	2,216,640	BTU/HR	0.02	$1,285,000	2011	$2,570,000	0.80	$118,267	6.00	$709,600	$722,321	$120,387
S-211	2		Primary Cyclone						INCLUDED								
S-212	2		Secondary Cyclone						INCLUDED								
S-213	2		Char Combustor Primary Cyclone						INCLUDED								
S-214	2		Char Combustor Secondary Cyclone						INCLUDED								
T-211	2		Bed Media Bin						INCLUDED								
T-212	2		Depleted bed media and ash storage bin						INCLUDED								
T-217	2		Char combustor bed media feed bin						INCLUDED								
A211	1		Pinch Exchanger CAPEX							2008	$1,559,517		$1,559,517	2.64	$4,109,899	$4,258,467	$1,615,892
Area A211: Fast Pyrolysis ↑											**$32,596,085**		**$24,374,014**	**3.03**	**$73,783,924**	**$75,426,781**	**$24,992,656**

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Table F-1. CFP-B Equipment Cost Summary (continued)

Equip ID	Required	Spares	Equipment Name	Design Flow (per unit)	Design Flow (total)	Actual Flow (total)	Units	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Orig Equip Cost (Req+Spare) in Base Year	Scaling Exp.	Scaled Cost in Base Year	Install Factor	Installed Cost in Base Year	Installed Cost in 2020$	Scaled Uninstalled Cost in 2020$
M-251	1		Hot Gas Filter package		5,670,000	2,556,952	CUFT/HR	0.45	$16,100,000	2011	$16,100,000	0.90	$7,862,333	1.80	$14,152,199	$14,405,909	$8,003,283
Area A251: Hot Gas Filter ↑											**$16,100,000**		**$7,862,333**	**1.80**	**$14,152,199**	**$14,405,909**	**$8,003,283**
R-261X1	2	2	#1 Fixed Bed VPU Reactor	156,670	313,340	324,159	LB/HR	1.03	$2,733,673	2013	$10,934,692	0.70	$11,197,633	1.59	$17,804,236	$18,711,238	$11,768,074
R-261X2	0	0	#2 Fixed Bed VPU Reactor	161,200	0			0.00	$3,841,800	2013	$0	0.70	$0	1.21	$0	$0	$0
R-261X3	0	0	#3 Fixed Bed VPU Reactor	161,200	0			0.00	$3,841,800	2013	$0	0.70	$0	1.21	$0	$0	$0
CONT-REG	1	0	Regeneration and other fixed bed contingency					1.00	$1,000,000	2016	$1,000,000	0.70	$1,000,000	1.60	$1,600,000	$1,760,975	$1,100,609
K-280X	2		Fixed Bed Regenerator Air Compressor	10,000	20,000	1,988	HP	0.10	$2,600,000	2014	$5,200,000	0.78	$859,075	1.60	$1,374,521	$1,422,477	$889,048
A261X	1		Pinch Exchanger CAPEX							2008	$1,162,751		$1,162,751	2.64	$3,064,276	$3,175,046	$1,204,784
Area A261: Ex Situ Vapor Phase Upgrading ↑											**$18,297,443**		**$14,219,459**	**1.68**	**$23,843,033**	**$25,069,737**	**$14,962,515**
D-301	1		Heavy Fraction Condenser Absorber		12.0	12	ft	1.03	$1,085,800	2013	$1,085,800	0.60	$1,102,900	1.40	$1,542,719	$1,621,310	$1,159,085
D-302	1		Light Fraction Condenser Absorber		13.0	10	ft	0.60	$1,085,300	2013	$1,085,300	0.60	$801,607	1.33	$1,062,778	$1,116,919	$842,443
H-313	1		D-301 Vapor ACHE		61,000,000	65,810,684	BTU/HR	1.08	$578,900	2013	$578,900	0.60	$605,876	1.32	$798,240	$838,905	$636,741
H-314	1		D-302 Feed cooler (water-cooled)		12,067,840	13,001,471	BTU/HR	1.08	$172,000	2013	$172,000	0.70	$181,210	2.03	$367,477	$386,198	$190,442
H-315	1		D-302 Feed/overhead exchanger		4,850,345	0	BTU/HR	0.00	$758,400	2013	$758,400	0.70	$0	1.52	$0	$0	$0
H-300IX-6	1		Indirect Cond (other tags are switched)		4,850,345	#N/A	#N/A	0.00	$758,400	2013	$0	0.70	$0	1.52	$0	$0	$0
H-316	1		D-302 Liquid feed chiller		10,138,754	6,662,346	BTU/HR	0.66	$132,400	2013	$132,400	0.70	$98,682	2.33	$230,084	$241,805	$103,709
H-320	1		Heavy fraction ACHE		280,000	348,285	BTU/HR	1.24	$33,300	2013	$33,300	0.60	$37,959	2.42	$91,990	$96,676	$39,892
H-320A	1		Heavy fraction cooler		845,500	287,609	BTU/HR	0.34	$45,300	2013	$45,300	0.70	$21,295	3.23	$68,727	$72,229	$22,380
H-326	1	1	D-302 Quench circulation chiller		13,087,000	2,820,652	BTU/HR	0.22	$78,300	2013	$156,600	0.70	$53,487	4.63	$247,898	$260,527	$56,212
P-318	1	1	D-301 Bottoms pump		56.0	57	gpm	1.02	$5,800	2013	$11,600	0.80	$11,770	4.38	$51,547	$54,173	$12,370
P-323	1	1	Circulation filter charge pump		41	12	HP	0.30	$25,400	2013	$50,800	0.80	$19,373	4.61	$89,391	$93,945	$20,360
P-329	1	1	Aqueous filter charge pump		182.0	136	gpm	0.75	$7,500	2013	$15,000	0.80	$11,867	4.20	$49,842	$52,381	$12,472
P-331	1	1	Organic product pump		124.0	97	gpm	0.78	$6,500	2013	$13,000	0.80	$10,696	4.63	$49,531	$52,055	$11,241
P-316IX4	1	1	Indirect Cond Pump		124.0	0	gpm	0.00	$6,500	2013	$0	0.80	$0	4.63	$0	$0	$0
P-332	1	1	Quench recycle pump		0.53	0.73	HP	1.37	$7,800	2013	$15,600	0.80	$20,067	4.88	$98,020	$103,013	$21,089
S-319	1	1	D-301 Bottoms filter		52	54	gpm	1.04	$14,600	2013	$29,200	0.80	$30,186	1.84	$55,616	$58,449	$31,723
S-316IX3	1	1	Indirect Cond Filter		52	0	gpm	0.00	$14,600	2013	$0	0.80	$0	1.84	$0	$0	$0
S-322	1		D-302 Bottoms decanter		1,013,050	290,772	LB/HR	0.29	$294,700	2013	$294,700	0.70	$123,006	1.82	$224,057	$235,471	$129,272
S-324	2	1	Quench circulation filter		2,340.0	616	gpm	0.26	$83,100	2013	$249,300	0.80	$85,762	1.71	$146,652	$154,123	$90,131
S-329	1	1	Aqueous phase filter		182.0	136	gpm	0.75	$26,500	2013	$53,000	0.80	$41,931	1.77	$74,210	$77,990	$44,067
T-328	1		Organic product surge tank		1,000	779	CUFT/HR	0.78	$48,000	2013	$48,000	0.70	$40,313	3.24	$130,681	$137,338	$42,367
T-332	1		D-301 Circulation surge tank		1,000	890	CUFT/HR	0.89	$48,000	2013	$48,000	0.70	$44,241	3.24	$143,415	$150,721	$46,495
LOSEP	1		Light Oxygenates Separation & Purification		1	1	UNIT	1.00	$19,771,797	2016	$19,771,797	0.70	$19,771,797	1.32	$26,006,271	$28,622,741	$21,761,022
A310	1		Pinch exchanger CAPEX							2008	$2,877,275		$2,877,275	2.64	$7,582,674	$7,856,779	$2,981,285
Area A310: Product Condensation and Separation ↑											**$27,525,272**		**$25,991,300**	**1.50**	**$39,111,821**	**$42,283,747**	**$28,254,798**

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Table F-1. CFP-B Equipment Cost Summary (continued)

Equip ID	Required	Spares	Equipment Name	Design Flow (per unit)	Design Flow (total)	Actual Flow (total)	Units	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Orig Equip Cost (Req+Spare) in Base Year	Scaling Exp.	Scaled Cost in Base Year	Install Factor	Installed Cost in Base Year	Installed Cost in 2020$	Scaled Uninstalled Cost in 2020$
H-373	1		Post-WGS ACHE		5,160,000	13,334,900	BTU/HR	2.58	$88,300	2013	$88,300	0.60	$156,086	1.71	$267,627	$281,261	$164,038
H-374	1		Post-WGS cooler		2,929,000	5,061,157	BTU/HR	1.73	$45,600	2013	$45,600	0.70	$66,871	4.05	$270,709	$284,500	$70,277
K-390	1		PSA compressor: multistage		3,648	3,962	HP	1.09	$4,700,000	2014	$4,700,000	0.90	$5,062,907	1.60	$8,100,651	$8,383,281	$5,239,551
M-385	1		PSA		5,218	8,951	LB/HR	1.72	$975,000	2013	$975,000	0.60	$1,347,822	1.80	$2,426,080	$2,549,672	$1,416,484
P-377	1	1	Compressor condensate pump		56.0	66	gpm	1.17	$5,800	2013	$11,600	0.80	$13,185	4.38	$57,741	$60,683	$13,857
R-365	1		High-temperature WGS		354,424	109,331	LB/HR	0.31	$1,243,000	2013	$1,243,000	0.60	$613,766	1.42	$869,444	$913,736	$645,033
S-385	1		PSA Compressor suction KO		288,000	571,599	CUFT/HR	1.98	$44,900	2013	$44,900	0.70	$72,550	2.00	$145,099	$152,491	$76,246
S-386	1		PSA Compr disch KO		103,650	209,410	CUFT/HR	2.02	$52,500	2013	$52,500	0.70	$85,893	2.56	$220,050	$231,260	$90,269
T-377	1		Compressor condensate tank		1,000	527	CUFT/HR	0.53	$48,000	2013	$48,000	0.70	$30,660	3.24	$99,390	$104,453	$32,222
A360	1		Pinch exchanger CAPEX							2008	$567,648		$567,648	2.64	$1,495,960	$1,550,037	$588,168
Area A360: Hydrogen Recycle and WGS ↑											**$7,776,548**		**$8,017,388**	**1.74**	**$13,952,752**	**$14,511,375**	**$8,336,144**
H-568A	1		WGS Effluent ACHE		5,740,000	4,933,613	BTU/HR	0.86	$76,300	2013	$76,300	0.60	$69,675	1.63	$113,324	$119,098	$73,224
H-568C	1		WGS Effluent cooler		3,159,000	1,925,355	BTU/HR	0.61	$32,600	2013	$32,600	0.70	$23,051	3.60	$82,871	$87,092	$24,225
K-511	1		Pyrolysis offgas compressor to reformer		3,648	1,695	HP	0.46	$4,700,000	2014	$4,700,000	0.60	$2,967,036	1.60	$4,747,258	$4,912,889	$3,070,555
K-545	1		Reformer combustor air blower						INCLUDED								
K-580	1		PSA Compressor multistage						INCLUDED								
M-575	1		PSA						INCLUDED								
P-576	1		Compressor condensate pump		50.0	21	gpm	0.41	$5,800	2013	$5,800	0.80	$2,863	4.38	$12,540	$13,178	$3,009
R-530	1		Steam reformer package (PNNL)		49	14.8	MMSCFD	0.30	$66,875,000	2007	$34,830,729	0.65	$15,973,284	1.92	$30,668,705	$34,801,451	$18,125,755
R-565	1		WGS reactor						INCLUDED								
R-566	1		LTS reactor		354,424	59,037	LB/HR	0.17	$1,243,000	2013	$1,243,000	0.60	$424,063	1.42	$600,716	$631,318	$445,666
S-575	1		Compressor suction KO drum						INCLUDED								
S-586	1		PSA Compressor discharge KO drum						INCLUDED								
T-576	1		Compressor condensate tank		1,000	166	CUFT/HR	0.17	$48,000	2013	$48,000	0.70	$13,653	3.24	$44,258	$46,513	$14,348
A500			Pinch exchanger CAPEX							2008	$5,267,691		$5,267,691	2.64	$13,882,297	$14,384,124	$5,458,111
Area A500: Hydrogen Plant ↑											**$46,204,120**		**$24,741,316**	**2.03**	**$50,151,969**	**$54,995,663**	**$27,214,895**

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Table F-1. CFP-B Equipment Cost Summary (continued)

Equip ID	Required	Spares	Equipment Name	Design Flow (per unit)	Design Flow (total)	Actual Flow (total)	Units	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Orig Equip Cost (Req+Spare) in Base Year	Scaling Exp.	Scaled Cost in Base Year	Install Factor	Installed Cost in Base Year	Installed Cost in 2020$	Scaled Uninstalled Cost in 2020$
H-601	1		Steam turbine condenser		1.00	6.46	BTU/HR	6.46	$1,300,000	2010	$1,300,000	1.00	$8,403,113	1.40	$11,764,358	$12,734,041	$9,095,744
H-616DUM	1		Steam turbine condenser		1	#N/A	#N/A	0.00	$1,300,000	2010	$1,300,000	1.00	$0	1.40	$0	$0	$0
H-603	1		Blowdown cooler		381,671	444,036	BTU/HR	1.16	$16,780	2007	$16,780	0.65	$18,515	4.32	$79,995	$90,775	$21,010
M-601	1		BFW EDI+polishing		300,000	372,465	LB/HR	1.24	$1,325,000	2010	$1,325,000	0.60	$1,508,671	2.00	$3,017,341	$3,266,047	$1,633,024
M-602A	1		Steam turbine generator Stage 1		-40,418	-18,161	hp	0.45	$7,700,000	2010	$7,700,000	0.70	$4,398,293	1.80	$7,916,927	$8,569,484	$4,760,824
M-602B	1		Steam turbine generator Stage 2		-40,418	-50,104	hp	1.24	$7,700,000	2010	$7,700,000	0.70	$8,949,398	1.80	$16,108,917	$17,436,703	$9,687,057
M-603	1		Startup boiler		204,131	203,962	LB/HR	1.00	$275,500	2013	$275,500	0.60	$275,363	1.69	$466,286	$490,040	$289,391
P-601	1	1	Makeup pump		80,411	7,447	LB/HR	0.09	$6,528	2007	$13,056	0.30	$6,394	4.72	$30,170	$34,236	$7,256
P-602	1	1	Condensate pump		247,010	372,465	LB/HR	1.51	$9,810	2007	$19,620	0.30	$22,193	4.61	$102,254	$116,033	$25,183
P-603	1	1	EDI pump		247,010	372,465	LB/HR	1.51	$9,810	2007	$19,620	0.30	$22,193	4.61	$102,254	$116,033	$25,183
P-604	1	1	Boiler feed water pump		494,622	372,465	LB/HR	0.75	$304,578	2007	$609,156	0.30	$559,464	1.35	$756,782	$858,762	$634,854
P-605	1	1	Condensate collection pump		247,010	372,465	LB/HR	1.51	$9,810	2007	$19,620	0.30	$22,193	4.61	$102,254	$116,033	$25,183
S-601	1		Blowdown flash drum		9,892	7,449	LB/HR	0.75	$47,205	2007	$47,205	0.65	$39,257	3.41	$134,059	$152,124	$44,547
T-601	1		Condensate collection tank		500,400	372,465	LB/HR	0.74	$28,505	2007	$28,505	0.65	$23,527	6.83	$160,700	$182,355	$26,698
T-602	1		Condensate surge tank		500,400	372,465	LB/HR	0.74	$27,704	2007	$27,704	0.65	$22,866	6.51	$148,897	$168,962	$25,947
T-603	1		Deaerator		494,619	372,465	LB/HR	0.75	$53,299	2007	$53,299	0.65	$44,325	5.07	$224,623	$254,892	$50,298
T-603A	1		Deaerator packed column		494,619	372,465	LB/HR	0.75	$18,405	2007	$18,405	0.65	$15,306	5.18	$79,254	$89,934	$17,369
T-605	1		Steam drum		494,622	372,468	LB/HR	0.75	$104,100	2007	$104,100	0.65	$86,573	2.28	$197,179	$223,750	$98,239
T-606	1		Steam turbine condensate tank		500,400	372,465	LB/HR	0.74	28,505	2007	$28,505	0.65	$23,527	6.83	$160,700	$182,355	$26,698
A600			Pinch exchanger CAPEX							2008	$5,026,208		$5,026,208	2.64	$13,245,900	$13,724,723	$5,207,899
Area A600: Steam System & Power Generation ↑											**$25,632,283**		**$29,467,377**	**1.86**	**$54,798,849**	**$58,807,281**	**$31,702,404**

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Table F-1. CFP-B Equipment Cost Summary (continued)

Equip ID	Required	Spares	Equipment Name	Design Flow (per unit)	Design Flow (total)	Actual Flow (total)	Units	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Orig Equip Cost (Req+Spare) in Base Year	Scaling Exp.	Scaled Cost in Base Year	Install Factor	Installed Cost in Base Year	Installed Cost in 2020$	Scaled Uninstalled Cost in 2020$
K-701	2	1	Plant air compressor		262,454	204,131	LB/HR	0.78	$87,922	2007	$263,766	0.30	$244,611	1.57	$383,656	$435,356	$277,573
M-701	1		Cooling tower system		7,506,000	2,389,092	LB/HR	0.32	$260,852	2010	$260,852	0.78	$106,806	2.47	$263,812	$285,557	$115,610
M-702	1		Hydraulic truck dump with scale		367,437	204,131	LB/HR	0.56	$80,000	1998	$80,000	0.60	$56,225	2.47	$138,875	$212,573	$86,062
M-703	1		Flue gas scrubber		489,600	937,884	LB/HR	1.92	$436,250	2010	$436,250	0.65	$665,634	2.47	$1,644,116	$1,779,633	$720,499
M-704	1		Flue gas stack		939,119	937,884	LB/HR	1.00	$169,187	2007	$169,187	0.65	$169,042	1.30	$219,612	$249,206	$191,822
M-708	1		Chilled water system		28,200,000	9,482,998	BTU/HR	0.34	$637,500	2011	$637,500	0.60	$331,511	1.80	$596,720	$607,418	$337,454
M-710	1		Product loading rack		1	1			$25,000	2011	$25,000	1.00	$25,000	2.47	$61,750	$62,857	$25,448
P-701	1	1	Cooling water pump		7,001,377	2,389,092	LB/HR	0.34	$239,375	2007	$478,750	0.30	$346,756	2.14	$742,112	$842,114	$393,483
P-702	1	1	Firewater pump		262,454	204,131	LB/HR	0.78	$23,043	2007	$46,086	0.30	$42,739	3.70	$158,025	$179,320	$48,498
P-703	1	1	Diesel pump		262,454	204,131	LB/HR	0.78	$3,842	2007	$7,684	0.30	$7,126	5.39	$38,393	$43,567	$8,086
P-704	1	1	Ammonia pump		262,454	204,131	LB/HR	0.78	$3,842	2007	$7,684	0.30	$7,126	5.21	$37,095	$42,094	$8,086
P-705	1	1	Caustic pump		1	1			$4,906	2007	$9,812	0.30	$9,812	4.30	$42,200	$47,887	$11,134
P-707	1	1	BFW chemical pump		262,454	204,131	LB/HR	0.78	$3,842	2007	$7,684	0.30	$7,126	5.21	$37,095	$42,094	$8,086
P-708	1	1	Flue gas scrubber circulation pump		489,600	937,884	LB/HR	1.92	$12,510	2007	$25,020	0.30	$30,407	4.12	$125,178	$142,047	$34,505
P-790	1	1	Gasoline product pump		108.0	0	gpm	0.00	$7,400	2013	$14,800	0.80	$0	3.96	$0	$0	$0
P-792	1	1	Diesel product pump		108.0	0	gpm	0.00	$7,400	2013	$14,800	0.80	$0	3.96	$0	$0	$0
S-701	1	1	Instrument air dryer		262,454	204,131	LB/HR	0.78	$8,349	2002	$16,698	0.60	$14,361	2.47	$35,471	$53,458	$21,643
T-701	1		Plant air receiver		262,454	204,131	LB/HR	0.78	$21,005	2007	$21,005	0.65	$17,839	5.44	$97,074	$110,155	$20,243
T-702	1		Firewater storage tank		262,454	204,131	LB/HR	0.78	$229,900	2007	$229,900	0.65	$195,252	1.46	$285,447	$323,912	$221,563
T-703	1		Purchased diesel storage tank		262,454	204,131	LB/HR	0.78	$104,674	2007	$104,674	0.65	$88,899	1.35	$119,665	$135,791	$100,878
T-704	1		Ammonia storage tank		262,454	204,131	LB/HR	0.78	$15,704	2007	$15,704	0.65	$13,337	5.39	$71,850	$81,532	$15,135
T-705	1		Caustic storage tank		262,454	204,131	LB/HR	0.78	$16,005	2007	$16,005	0.65	$13,593	3.01	$40,936	$46,452	$15,425
T-708	1		BFW chemical storage tank		262,454	204,131	LB/HR	0.78	$22,004	2007	$22,004	0.65	$18,688	6.70	$125,186	$142,055	$21,206
T-790	1	1	Gasoline product storage tanks		50	0	gpm	0.00	$442,700	2013	$885,400	0.70	$885,400	1.75	$1,553,400	$1,632,535	$930,505
T-792	1	1	Diesel product storage tank		50	0	gpm	0.00	$442,700	2013	$885,400	0.70	$885,400	1.75	$1,553,400	$1,632,535	$930,505
Area A700: Cooling Water & Other Utilities ↑											**$4,681,665**		**$4,182,690**	**2.00**	**$8,371,068**	**$9,130,145**	**$4,543,450**

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Table F-1. CFP-B Equipment Cost Summary (continued)

Equip ID	Required	Spares	Equipment Name	Design Flow (per unit)	Design Flow (total)	Actual Flow (total)	Units	Size Ratio	Original Equip Cost (per unit)	Base Year	Total Orig Equip Cost (Req+Spare) in Base Year	Scaling Exp.	Scaled Cost in Base Year	Install Factor	Installed Cost in Base Year	Installed Cost in 2020$	Scaled Uninstalled Cost in 2020$
H-840	1		RTO Feed/Effluent exchanger						INCLUDED								
K-831	1		RTO blower		900	1,890	HP	2.10	$360,420	2009	$360,420	0.60	$562,485	1.60	$899,976	$1,028,100	$642,563
M-805	1		Clarifier		220	220			$96,221	2007	$96,221	0.65	$96,221	2.41	$232,300	$263,603	$109,187
M-806	1		Belt press		1	1			$135,000	2010	$135,000	0.65	$135,000	2.47	$333,450	$360,935	$146,127
M-807	1		Sand filter		265	265			$120,400	2010	$120,400	0.65	$120,400	2.47	$297,388	$321,900	$130,324
M-833	1		Aqueous regenerative thermal oxidizer		18,500	33,400	SCFM	1.81	$1,850,000	2002	$1,850,000	0.60	$2,637,019	2.40	$6,328,846	$9,538,064	$3,974,193
M-861	1		Reverse osmosis unit		197,000	68,022	LB/HR	0.35	$350,000	2012	$350,000	0.70	$166,263	1.15	$191,203	$194,997	$169,563
P-801	1	1	Clarifier overflow pump		50	43	gpm	0.86	$4,500	2013	$9,000	0.80	$7,970	4.38	$34,891	$36,668	$8,376
P-809	1	1	Slurry pump		1	1			$3,242	2007	$6,484	0.80	$6,484	3.10	$20,100	$22,809	$7,358
P-810	1	1	Boiler charge pump		200	136	gpm	0.68	$6,500	2013	$13,000	0.80	$9,537	4.14	$39,469	$41,480	$10,023
P-812	1	1	Condensate Pump		140	39	gpm	0.28	$5,100	2013	$10,200	0.80	$3,656	4.59	$16,774	$17,629	$3,842
P-814	1	1	Boiler charge pump		145	89	gpm	0.61	$5,100	2013	$10,200	0.80	$6,912	4.59	$31,715	$33,331	$7,264
P-815	1	1	Boiler makeup pump		5	43	gpm	8.66	$14,700	2013	$29,400	0.80	$165,263	1.84	$304,668	$320,188	$173,681
P-822	1	1	Concentrated waste water pump		7	18	gpm	2.57	$4,900	2013	$9,800	0.80	$20,852	4.31	$89,790	$94,364	$21,914
P-860	1	1	RO Feed pump		185	163	gpm	0.88	$6,400	2013	$12,800	0.80	$11,552	3.84	$44,405	$46,667	$12,141
P-866	1	1	Clean water pump		140	121	gpm	0.87	$5,200	2013	$10,400	0.80	$9,275	4.35	$40,311	$42,365	$9,748
T-801	1		Clarifier overflow tank		1	1		1.00	$85,800	2013	$85,800	0.65	$85,800	2.27	$195,000	$204,934	$90,171
T-809	1		Slurry tank		1	1		1.00	$85,800	2013	$85,800	0.65	$85,800	2.27	$195,000	$204,934	$90,171
T-810	1		Boiler Feed Tank		200	136	gpm	0.68	$24,600	2013	$24,600	0.70	$18,760	4.49	$84,266	$88,558	$19,715
T-812	1		Steam drum		100,800	139,501	CUFT/HR	1.38	$24,400	2013	$24,400	0.70	$30,632	4.46	$136,713	$143,678	$32,192
T-814	1		Boiler feed tank		145	89	gpm	0.61	$24,400	2013	$24,400	0.70	$17,359	4.46	$77,477	$81,423	$18,244
T-822	1		Steam drum		156,000	88,790	CUFT/HR	0.57	$19,400	2013	$19,400	0.70	$13,076	5.23	$68,345	$71,826	$13,742
T-860	1		RO Feed tank		185	163	gpm	0.88	$76,000	2013	$76,000	0.70	$69,478	2.24	$155,411	$163,328	$73,017
T-866	1		Clean water tank		140	121	gpm	0.87	$194,200	2013	$194,200	0.70	$194,200	2.23	$434,000	$456,109	$204,093
A800	1		Pinch exchanger CAPEX							2008	$846,318		$846,318	2.64	$2,230,359	$2,310,984	$876,912
Area A800: Water Management ↑											**$4,404,243**		**$5,320,312**	**2.35**	**$12,481,855**	**$16,088,874**	**$6,844,561**
Plant Total											**$187,385,459**		**$144,427,311**	**2.02**	**$291,140,028**	**$311,230,559**	**$155,115,318**

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Table F-2. CFP-B Operating Cost Summary

Raw Material	Scaling Flow		Quoted Price		Year of Price Quote	2020 Cost	2020 Cost ($/h)	2020 Cost (MM$/yr)	2020 Cost (¢/GGE)
Variable Operating Costs									
Feedstock, Catalyst & Chemicals									
Feedstock - Wood Chips	102	US ton/h	$60.33	/US ton	2016	$ 65.27	6,662.17	52.525	126.01
Sand Makeup	159	lb/h	$ 0.21	/lb	2011	$ 0.24	37.72	0.297	0.71
Natural Gas to SMR	0	MMBtu/h	$ 3.500	/MMBtu	2016	$ 3.79	0.01	0.000	0.00
50 wt% Caustic	231	lb/h	$ 0.08	/lb	2010	$ 0.09	21.39	0.169	0.40
Solvent 1	111	lb/h	$ 1.21	/lb	2016	$ 1.31	144.94	1.143	2.74
Solvent 2	3	lb/h	$ 0.40	/lb	2016	$ 0.43	1.22	0.010	0.02
Boiler Chemicals	7,449	lb/h blowdn	$ 75.00	/MM lb	2014	$ 80.52	0.60	0.005	0.01
Cooling Tower Chemicals	4,264	ton cool	$ 33.84	/ton cool/yr	2014	$ 36.33	19.65	0.155	0.37
Regeneration Gas: FP	0	lb/h	$ 0.12	/lb	2011	$ 0.14	0.00	0.000	0.00
Net Water Makeup	52,226	lb/h	$ 0.20	/ton	2004	$ 0.33	8.73	0.069	0.17
Purchased Hydrogen	-	lb/h	$ 0.680	lb/h	2016	$ 0.74	-	-	-
Bio-Oil Transportation (800 miles)	44,130	lb/h	$ 40.80	/ton/800 miles	2007	$ 58.00	1,279.75	10.090	24.21
Diesel Fuel at Biorefinery	71	lb/h	$ 0.40	/lb	2011	$ 0.46	32.75	0.258	0.62
Subtotal							**8,208.92**	**64.719**	**155.27**
Waste Streams									-
FP Purge	3,717	lb/h	$ 0.016	/lb	1998	$ 0.032	117.97	0.930	2.23
Quench Filter Solids	25	lb/h	0.016	/lb	1998	$ 0.032	0.80	0.006	0.02
WWT Cost	17,215	lb/h	$ 0.003	/lb	2011	$ 0.003	52.10	0.411	0.99
Subtotal							**170.87**	**1.347**	**3.23**
Coproduct Credits									
Acetone	5,336	lb/h	$ 0.442	/lb	2020	$ 0.442	(2,360)	(18.61)	(44.64)
2-Butanone (MEK)	926	lb/h	$ 0.771	/lb	2020	$ 0.771	(714)	(5.63)	(13.51)
Hydrogen	3	lb/h	$ 0.680	/lb	2020	$ 0.680	(2)	(0.01)	(0.03)
Electricity	18,190	kW	$ 0.059	/kWh	2020	$ 0.059	(1,064)	(8.39)	(20.13)
Subtotal							**(4,140.80)**	**(32.65)**	**(78.32)**
Total Variable Operating Costs							**4,238.99**	**33.420**	**80.18**
Fixed Operating Costs								2020 Cost (MM$/yr)	2020 Cost (¢/GGE)
	2007 Salary	**# Positions**	**2020 Cost**		**2020 Hourly Wage**				
Plant Manager	$ 147,000	1	$ 195,724		$ 70.67				
Plant Engineer	$ 70,000	1	$ 93,202		$ 33.65				
Maintenance Supr	$ 57,000	1	$ 75,893		$ 27.40				
Lab Manager	$ 56,000	1	$ 74,562		$ 26.92				
Shift Supervisor	$ 48,000	5	$ 319,550		$ 23.08				
Lab Technician	$ 40,000	2	$ 106,517		$ 19.23				
Maintenance Tech	$ 40,000	16	$ 852,133		$ 19.23				
Shift Operators	$ 48,000	20	$ 1,278,199		$ 23.08				
Yard Employees	$ 28,000	12	$ 447,370		$ 13.46				
Clerks & Secretaries	$ 36,000	3	$ 143,797		$ 17.31				
Total Salaries		**62**	**$ 3,586,947**					**3.587**	**8.61**
					Avg Salary (w/ Benefits)				
Overhead and Benefits	Labor & Supervison =	90.0%	$3,228,253		$52.85	per Hour		3.228	7.74
Maintenance	% of FCI =	3.0%	$16,468,279		$109,923	per Year		16.468	39.51
Insurance & Taxes	% of FCI =	0.7%	$3,842,598					3.843	9.22
Total Fixed Operating Costs								**27.126**	**65.08**
Total Operating Costs								**60.546**	**145.25**

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Table F-3. Discounted Cash Flow Rate of Return (DCFROR) for CFP-B

Year	Cumulative	-2	-1	0	1	2	3	4	5	6	7	8
Fixed Capital Investment		$17,566,164	$131,746,233	$70,264,658								
Land		$1,848,000										
Working Capital				$27,447,132								
Loan Payment					$49,085,184	$49,085,184	$49,085,184	$49,085,184	$49,085,184	$49,085,184	$49,085,184	$49,085,184
Loan Interest Payment		$2,107,940	$17,917,488	$26,349,247	$26,349,247	$24,530,372	$22,565,987	$20,444,451	$18,153,192	$15,678,633	$13,006,109	$10,119,782
Loan Principal		$26,349,247	$223,968,597	$329,365,583	$306,629,646	$282,074,833	$255,555,635	$226,914,901	$195,982,909	$162,576,357	$126,497,281	$87,531,879
Annual Income												
Fuel Sales					$104,847,459	$139,796,611	$139,796,611	$139,796,611	$139,796,611	$139,796,611	$139,796,611	$139,796,611
Byproduct Credit					$24,484,566	$32,646,089	$32,646,089	$32,646,089	$32,646,089	$32,646,089	$32,646,089	$32,646,089
Total Annual Sales					$129,332,025	$172,442,700	$172,442,700	$172,442,700	$172,442,700	$172,442,700	$172,442,700	$172,442,700
Annual Manufacturing Cost												
Raw Materials (Feedstock)					$45,958,976	$52,524,544	$52,524,544	$52,524,544	$52,524,544	$52,524,544	$52,524,544	$52,524,544
Catalyst Costs		$0	$0	$0	$22,079,314	$168,390	$6,724,520	$168,390	$6,724,520	$168,390	$6,724,520	$168,390
Other Variable Costs					$11,849,012	$13,541,728	$13,541,728	$13,541,728	$13,541,728	$13,541,728	$13,541,728	$13,541,728
Fixed Operating Costs					$27,126,078	$27,126,078	$27,126,078	$27,126,078	$27,126,078	$27,126,078	$27,126,078	$27,126,078
Total Product Cost					$107,013,379	$93,360,740	$99,916,869	$93,360,740	$99,916,869	$93,360,740	$99,916,869	$93,360,740
Annual Depreciation												
General Plant Writedown					14.29%	24.49%	17.49%	12.49%	8.93%	8.92%	8.93%	4.46%
Depreciation Charge					$70,040,343	$120,034,149	$85,724,674	$61,217,906	$43,769,087	$43,720,074	$43,769,087	$21,860,037
Remaining Value					$420,095,015	$300,060,866	$214,336,192	$153,118,286	$109,349,198	$65,629,124	$21,860,037	$0
Steam Plant Writedown					3.75%	7.22%	6.68%	6.18%	5.71%	5.29%	4.89%	4.52%
Depreciation Charge					$2,205,273	$4,245,298	$3,926,562	$3,632,526	$3,359,660	$3,107,965	$2,874,500	$2,659,265
Remaining Value					$56,602,008	$52,356,711	$48,430,149	$44,797,623	$41,437,963	$38,329,998	$35,455,498	$32,796,233
Annual Tax												
Net Revenue					($76,276,217)	($69,727,858)	($39,691,392)	($6,212,923)	$7,243,891	$16,575,289	$12,876,135	$44,442,876
Losses Forward						($76,276,217)	($146,004,075)	($185,695,467)	($191,908,390)	($184,664,499)	($168,089,210)	($155,213,075)
Taxable Income					($76,276,217)	($146,004,075)	($185,695,467)	($191,908,390)	($184,664,499)	($168,089,210)	($155,213,075)	($110,770,200)
Income Tax					$0	$0	$0	$0	$0	$0	$0	$0
Annual Cash Flow												
Annual Cash Income					($26,766,539)	$29,996,776	$23,440,646	$29,996,776	$23,440,646	$29,996,776	$23,440,646	$29,996,776
Discount Factor		1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830	0.6209	0.5645	0.5132	0.4665
Annual Present Value					($24,333,217)	$24,790,724	$17,611,304	$20,488,201	$14,554,797	$16,932,398	$12,028,758	$13,993,717
Total Capital Investment + Interest		$26,041,746	$164,630,093	$124,061,036								
Cumulative Present Value	**$313,054,014**											
Cumulative Investments + Interest	**$313,054,014**											
Net Present Worth	**$0**											

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Table F-3. DCFROR for CFP-B (continued)

Year	9	10	11	12	13	14	15	16	17	18	19
Fixed Capital Investment											
Land											
Working Capital											
Loan Payment	$49,085,184	$49,085,184	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$7,002,550	$3,635,940	($0)	($0)	($0)	($0)	($0)	($0)	($0)	($0)	($0)
Loan Principal	$45,449,245	($0)	($0)	($0)	($0)	($0)	($0)	($0)	($0)	($0)	($0)
Annual Income											
Fuel Sales	$139,796,611	$139,796,611	$139,796,611	$139,796,611	$139,796,611	$139,796,611	$139,796,611	$139,796,611	$139,796,611	$139,796,611	$139,796,611
Byproduct Credit	$32,646,089	$32,646,089	$32,646,089	$32,646,089	$32,646,089	$32,646,089	$32,646,089	$32,646,089	$32,646,089	$32,646,089	$32,646,089
Total Annual Sales	$172,442,700	$172,442,700	$172,442,700	$172,442,700	$172,442,700	$172,442,700	$172,442,700	$172,442,700	$172,442,700	$172,442,700	$172,442,700
Annual Manufacturing Cost											
Raw Materials (Feedstock)	$52,524,544	$52,524,544	$52,524,544	$52,524,544	$52,524,544	$52,524,544	$52,524,544	$52,524,544	$52,524,544	$52,524,544	$52,524,544
Catalyst Costs	$6,724,520	$168,390	$6,724,520	$168,390	$6,724,520	$168,390	$6,724,520	$168,390	$6,724,520	$168,390	$6,724,520
Other Variable Costs	$13,541,728	$13,541,728	$13,541,728	$13,541,728	$13,541,728	$13,541,728	$13,541,728	$13,541,728	$13,541,728	$13,541,728	$13,541,728
Fixed Operating Costs	$27,126,078	$27,126,078	$27,126,078	$27,126,078	$27,126,078	$27,126,078	$27,126,078	$27,126,078	$27,126,078	$27,126,078	$27,126,078
Total Product Cost	$99,916,869	$93,360,740	$99,916,869	$93,360,740	$99,916,869	$93,360,740	$99,916,869	$93,360,740	$99,916,869	$93,360,740	$99,916,869
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%
Depreciation Charge	$2,623,981	$2,623,393	$2,623,981	$2,623,393	$2,623,981	$2,623,393	$2,623,981	$2,623,393	$2,623,981	$2,623,393	$2,623,981
Remaining Value	$30,172,252	$27,548,859	$24,924,878	$22,301,485	$19,677,504	$17,054,112	$14,430,131	$11,806,738	$9,182,757	$6,559,364	$3,935,383
Annual Tax											
Net Revenue	$62,899,299	$72,822,628	$69,901,850	$76,458,567	$69,901,850	$76,458,567	$69,901,850	$76,458,567	$69,901,850	$76,458,567	$69,901,850
Losses Forward	($110,770,200)	($47,870,901)	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	($47,870,901)	$24,951,727	$69,901,850	$76,458,567	$69,901,850	$76,458,567	$69,901,850	$76,458,567	$69,901,850	$76,458,567	$69,901,850
Income Tax	$0	$5,239,863	$14,679,388	$16,056,299	$14,679,388	$16,056,299	$14,679,388	$16,056,299	$14,679,388	$16,056,299	$14,679,388
Annual Cash Flow											
Annual Cash Income	$23,440,646	$24,756,913	$57,846,442	$63,025,661	$57,846,442	$63,025,661	$57,846,442	$63,025,661	$57,846,442	$63,025,661	$57,846,442
Discount Factor	0.4241	0.3855	0.3505	0.3186	0.2897	0.2633	0.2394	0.2176	0.1978	0.1799	0.1635
Annual Present Value	$9,941,122	$9,544,862	$20,274,825	$20,081,918	$16,756,054	$16,596,626	$13,847,978	$13,716,220	$11,444,610	$11,335,719	$9,458,356
Total Capital Investment + Interest											

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Table F-3. DCFROR for CFP-B (continued)

Year	20	21	22	23	24	25	26	27	28	29	30
Fixed Capital Investment											
Land											($1,848,000)
Working Capital											($27,447,132)
Loan Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	($0)	($0)	($0)	($0)	($0)	($0)	($0)	($0)	($0)	($0)	($0)
Loan Principal	($0)	($0)	($0)	($0)	($0)	($0)	($0)	($0)	($0)	($0)	($0)
Annual Income											
Fuel Sales	$139,796,611	$139,796,611	$139,796,611	$139,796,611	$139,796,611	$139,796,611	$139,796,611	$139,796,611	$139,796,611	$139,796,611	$139,796,611
Byproduct Credit	$32,646,089	$32,646,089	$32,646,089	$32,646,089	$32,646,089	$32,646,089	$32,646,089	$32,646,089	$32,646,089	$32,646,089	$32,646,089
Total Annual Sales	$172,442,700	$172,442,700	$172,442,700	$172,442,700	$172,442,700	$172,442,700	$172,442,700	$172,442,700	$172,442,700	$172,442,700	$172,442,700
Annual Manufacturing Cost											
Raw Materials (Feedstock)	$52,524,544	$52,524,544	$52,524,544	$52,524,544	$52,524,544	$52,524,544	$52,524,544	$52,524,544	$52,524,544	$52,524,544	$52,524,544
Catalyst Costs	$168,390	$6,724,520	$168,390	$6,724,520	$168,390	$6,724,520	$168,390	$6,724,520	$168,390	$6,724,520	$168,390
Other Variable Costs	$13,541,728	$13,541,728	$13,541,728	$13,541,728	$13,541,728	$13,541,728	$13,541,728	$13,541,728	$13,541,728	$13,541,728	$13,541,728
Fixed Operating Costs	$27,126,078	$27,126,078	$27,126,078	$27,126,078	$27,126,078	$27,126,078	$27,126,078	$27,126,078	$27,126,078	$27,126,078	$27,126,078
Total Product Cost	$93,360,740	$99,916,869	$93,360,740	$99,916,869	$93,360,740	$99,916,869	$93,360,740	$99,916,869	$93,360,740	$99,916,869	$93,360,740
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	2.23%									
Depreciation Charge	$2,623,393	$1,311,990									
Remaining Value	$1,311,990	($0)									
Annual Tax											
Net Revenue	$76,458,567	$71,213,840	$79,081,960	$72,525,831	$79,081,960	$72,525,831	$79,081,960	$72,525,831	$79,081,960	$72,525,831	$79,081,960
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$76,458,567	$71,213,840	$79,081,960	$72,525,831	$79,081,960	$72,525,831	$79,081,960	$72,525,831	$79,081,960	$72,525,831	$79,081,960
Income Tax	$16,056,299	$14,954,906	$16,607,212	$15,230,424	$16,607,212	$15,230,424	$16,607,212	$15,230,424	$16,607,212	$15,230,424	$16,607,212
Annual Cash Flow											
Annual Cash Income	$63,025,661	$57,570,924	$62,474,749	$57,295,406	$62,474,749	$57,295,406	$62,474,749	$57,295,406	$62,474,749	$57,295,406	$62,474,749
Discount Factor	0.1486	0.1351	0.1228	0.1117	0.1015	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
Annual Present Value	$9,368,363	$7,779,592	$7,674,771	$6,398,645	$6,342,786	$5,288,137	$5,241,972	$4,370,361	$4,332,208	$3,611,869	$3,580,337
Total Capital Investment + Interest											($1,678,862)

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Appendix G. TEA Summary for 100% SRD Hydroprocessing (HT Configuration)

Case included:

- 12201011: HT of 100% SRD in H-TC configuration with no liquid recycle, 85 mol % H_2 makeup purity, process model with 75k BPD capacity.

Please see Section 3.4.1 for interpretation of Case IDs.

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Table G-1. Equipment Costs and Fixed Capital Investment for 75k BPD SRD HT, 85% H_2 Makeup Purity (12201011)

Equip ID	Equipment Description	Process Variable Value	Size Ratio	Scaled Cost (Base Year)	Scaled Cost (2020$)	Pressure (psig)	Pres. Adj. Scaled Cost (2020)	Lang/TCI Factor (Low)	Lang/TCI Factor (High)	Average Lang/TCI Factor	FCI (2020$) for Single Units of Each	FCI (2020$)
C450	Amine Scrubber	189	0.79	$ 91,983	$ 68,808		68,808	13.21999	22.35638	17.78818	$ 1,223,971	$ 1,223,971
C470	Product Recovery Column 1	921094	0.97	$ 5,079,989	$ 3,800,112		3,800,112	2.996629	3.676469	3.336549	$ 12,679,260	$ 12,679,260
C480	Product Recovery Column 2	691825	1.02	$ 3,736,173	$ 2,794,864		2,794,864	3.357408	4.221599	3.789504	$ 10,591,147	$ 10,591,147
MX405	CFP and H2 Static Mixer	0	0.00	$ -	$ -		-	0	0	0	$ -	$ -
MX406	Rx Inlet Static Mixer	0	0.00	$ -	$ -		-	0	0	0	$ -	$ -
P400A/B	Cat Pyrolysis Oil 2 Pumps	0	0.00	$ -	$ -	1,085	-	0	0	0	$ -	$ -
P405A/B	Liquid Feed 2 Pumps	75000	1.00	$ 1,538,999	$ 1,151,256	1,085	1,030,121	3614196	5168928	4391562	$ 4,391,562	$ 8,783,124
P455A/B	Liquid Recycle 2 Pumps	0	0.00	$ -	$ -	1,035	-	0	0	0	$ -	$ -
P471A/B	Prod Recov Reflux 2 Pumps - 1	921094	0.97	$ 60,703	$ 45,409	25	45,510	258953.1	426162.4	342557.8	$ 342,558	$ 685,116
P472A/B	Prod Recov Btms 2 Pumps - 1	921094	0.97	$ 64,632	$ 48,348	30	46,631	264332.6	434539.3	349436	$ 349,436	$ 698,872
P481A/B	Prod Recov Reflux 2 Pumps - 2	691825	1.02	$ 55,912	$ 41,825	20	42,805	245884	405771.9	325828	$ 325,828	$ 651,656
P482A/B	Prod Recov Btms 2 Pumps - 2	691825	1.02	$ 42,978	$ 32,150	15	31,268	188574.8	315626	252100.4	$ 252,100	$ 504,201
PIPE412	H2 MU Comp Jumpover Line	8923	0.73	$ 5,148	$ 3,851		3,851	6.5	7	6.75	$ 25,992	$ 25,992
PIPE442	H2 Rec Comp Jumpover Line	23242	0.31	$ 5,101	$ 3,816		3,816	6.5	7	6.75	$ 25,757	$ 25,757
PIPE500	CFP Pipe from Storage	0	0.00	$ -	$ -		-	0	0	0	$ -	$ -
PSA460	Pressure Swing Adsorption	25660831	0.31	$ 3,376,892	$ 2,526,102		2,526,102	3.6	4.4	4	$ 10,104,407	$ 10,104,407
QAH413	Rx Effluent Air Cooler	28702736	0.23	$ 576,682	$ 431,390		431,390	4.75	5.75	5.25	$ 2,264,796	$ 2,264,796
QCH413	Rx Effluent Water Cooler	27243313	0.69	$ 795,549	$ 595,114	985	1,336,356	7296633	9442973	8369803	$ 8,369,803	$ 8,369,803
QCH444	Water Cooler H2 Quench 1	2034741	0.43	$ 93,503	$ 69,946	1,085	131,287	1031884	1607804	1319844	$ 1,319,844	$ 1,319,844
QCH445	Water Cooler H2 Quench 2	2034741	0.43	$ 93,503	$ 69,946	1,085	131,287	1031884	1607804	1319844	$ 1,319,844	$ 1,319,844
QFH411-L	Reactor Preheat Furnace - Liquid Feed	34501233	0.81	$ 1,245,243	$ 931,510	1,085	931,510	4.47	7.21	5.84	$ 4,165,054	$ 4,165,054
QFH411-G	Reactor Preheat Furnace - Gas Feed	0	0.00	$ -	$ -	1,085	-	0	0	0	$ -	$ -
QH412	Process Rx Eff Cooler	61096489	0.45	$ 1,692,806	$ 1,266,313	995	2,939,992	14183576	17233681	15708629	$ 15,708,629	$ 15,708,629
QH471	Prod Recov Condenser 1	921094	0.97	$ 4,237,785	$ 3,170,097	25	3,170,097	15113818	18253594	16683706	$ 16,683,706	$ 16,683,706
QH472	Prod Recov Reboiler 1	921094	0.97	$ 3,854,777	$ 2,883,586	30	2,883,586	13953829	16980824	15467327	$ 15,467,327	$ 15,467,327
QH481	Prod Recov Condenser 2	691825	1.02	$ 2,217,010	$ 1,658,446	20	1,658,446	8753430	11134280	9943855	$ 9,943,855	$ 9,943,855
QH482	Prod Recov Reboiler 2	691825	1.02	$ 1,688,931	$ 1,263,413	15	1,263,413	6959417	9047096	8003257	$ 8,003,257	$ 8,003,257
QXH410	Feed Preheater to Furnace	302948178	0.83	$ 1,062,370	$ 794,711	1,085	1,935,366	9970371	12526515	11248443	$ 11,248,443	$ 11,248,443
RX405	CFP Cat Stabilization Rx	0	0.00	$ -	$ -	1,085	-	0	0	0	$ -	$ -
RX410-HT	Reactor System - Hydrotreater	1063442	0.83	$ 9,562,559	$ 7,153,322	1,030	6,527,552	3.5	5.3	4.4	$ 28,721,230	$ 28,721,230
RX410-HC	Reactor System - Hydrocracker	1063442	0.83	$ 9,562,559	$ 7,153,322	1,030	6,527,552	3.5	5.3	4.4	$ 28,721,230	$ -
TK400	Cat Pyrolysis Oil Storage	0	0.00	$ -	$ -		-	0	0	0	$ -	$ -
V451	High Pressure Flash Drum	1310556	1.03	$ 832,679	$ 622,890	980	1,211,695	6.456222	10.52633	8.491276	$ 10,288,835	$ 10,288,835
V452	Low Pressure Flash Drum	943623	0.95	$ 212,960	$ 159,306	1,085	159,306	13.40284	25.70327	19.55306	$ 2,135,147	$ 2,135,147
V471	Prod Recov Ovhd Drum 1	793095	0.84	$ 51,925	$ 38,843	1,085	38,843	22.27659	47.82711	35.05185	$ 865,282	$ 865,282
V481	Prod Recov Ovhd Drum 2	743479	1.10	$ 70,023	$ 52,381	1,085	52,381	20.00306	41.93073	30.9669	$ 1,047,787	$ 1,047,787
WK412	H2 Make-up Compressor	8818	0.66	$ 8,215,337	$ 6,145,526		6,145,526	3	3.8	3.4	$ 20,894,788	$ 20,894,788
WK442	H2 Recycle Compressor	1529	0.38	$ 2,285,203	$ 1,709,458		1,709,458	3	3.8	3.4	$ 5,812,157	$ 5,812,157
QH470	Dist Col 1 Liquid Feed Preheat Furnace	40691885	0.96	$ 1,355,035	$ 1,013,641	36	1,013,641	4.337327	6.945616	5.641472	$ 4,396,494	$ 4,396,494
Total												**$ 214,629,778**

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Table G-2. Operating Costs for 75k BPD SRD HT (12201011)

	Quoted Price	Unit	Cost Year	2020$ price	Scaling Flow	Unit	2020$ Cost ($/h)	2020$ Cost (MM$/y)	2020$ Cost (cents/GGE)
Feedstocks									
Crude Oil Basis	$ 85.00	/bbl	2020	$ 85.00					
Straight Run Diesel (SRD)	$ 105.88	/bbl	2020	$ 105.88	75,000	bbl/day	330,871	2,753.51	233.0
Light Cycle Oil (LCO)	$ 102.95	/bbl	2020	$ 102.95	0	bbl/day	0	0.00	0.0
Vacuuum Gas Oil (VGO)	$ 100.78	/bbl	2020	$ 100.78	0	bbl/day	0	0.00	0.0
Catalytic Fast Pyrolysis Oil (21.6%)	$ 215.62	/bbl	2020	$ 215.62	0	bbl/day	1	0.01	0.0
Bio-Oil Transportation 800 miles	$ -	/ton-mile	2007	$ -	1	lb/h	0	0.00	0.0
Subtotal							**330,872**	**2,753.52**	**233.0**
Process Inputs									
Hydrogen (pure H2 basis)									
Total Hydrogen Feed					15,289	lb/h			
Hydrogen Recovered via PSA					3,714	lb/h			
Net Hydrogen Makeup	$ 0.680	/lb	2020	$ 0.680	11,576	lb/h	7,876	65.54	5.5
Electricity for Hydroprocessing	$ 0.082	/kWh	2020	$ 0.082	10,573	kWh	864	7.19	0.6
Natural Gas	$ 3.231	/MMBtu	2020	$ 3.231	0	MMBtu/h	0	0.00	0.0
Fuel Gas	$ 2.746	/MMBtu	2020	$ 2.746	381	MMBtu/h	1,045	8.70	0.7
Amine Consumption	$ 0.00122600	/lbmol inlet gas	2020	$ 0.00122600	37,796	lbmol/h	46	0.39	0.033
Water Makeup	$ 0.200	/ton	2004	$ 0.334	124	ton/h	41	0.34	0.0
BFW to Steam	$ 0.001	/lb	2016	$ 0.001		lb/h			
Cooling Water Circulation	$ 0.00768	/ton	2020	$ 0.008	1,032	ton/h	8	0.07	0.0
Boiler Chemicals	$ 75.00	/MM lb bldwn	2014	$ 80.516		MMlb/h			
Cooling Tower Chemicals	$ 33.84	/ton-cool/y	2014	$ 36.329		ton-cool/y			
Diesel Fuel	$ 2.57	/gallon	2020	$ 2.571	10	gallon/h	26	0.21	0.0
Subtotal							**9,906**	**82.44**	**7.0**
Waste Streams									
Wastewater Treatment	$ 0.003	/lb	2011	$ 0.003	248,631	lb/h	752	6.26	0.5
Acid Gas Conversion and Recovery	$ 0.477	/lbmol acid gas	2020	$ 0.477	189	lbmol/h	90	0.75	0.1
Subtotal							**843**	**7.01**	**0.6**
Coproducts									
Fuel Gas	$ (2.746)	/MMBtu	2020	$ (2.746)	479	MMBtu/h	-1,315	-10.95	-0.9
HP Steam (900 psig)	$ (0.0045)	/lb	2023	$ (0.004)	0	lb/h	0	0.00	0.0
MP Steam (350 psig)	$ (0.0045)	/lb	2023	$ (0.004)	0	lb/h	0	0.00	0.0
LP Steam (150 psig)	$ (0.0045)	/lb	2023	$ (0.004)	171,299	lb/h	-626	-5.21	-0.4
VLP Steam (15 psig)	$ (0.0044)	/lb	2023	$ (0.004)	153,672	lb/h	-546	-4.55	-0.4
Subtotal							**-2,488**	**-20.70**	**-1.8**
Catalysts and Royalties									
Royalties	$ 0.20	/Bbl Feed	2005	$ 0.309	3,125	Bbl/h	964	8.03	0.7
Subtotal							**964**	**8.03**	**0.7**
Total Variable Operating Costs							**340,097**	**2,830.29**	**239.5**
Fixed Operating Costs									
Total Salaries								2.55	0.2
Overhead and Benefits	% of Labor & Supervison =		0.9					2.29	0.2
Maintenance	% of FCI =		0.03					6.4	0.5
Insurance & Taxes	% of FCI =		0.007					1.5	0.1
Total Fixed Operating Costs								**12.78**	**1.1**
Total Operating Costs								**2,843.07**	**240.6**

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Table G-3. Discounted Cash Flow Rate of Return (DCFROR) for 75k BPD SRD HT (12201011)

Year		-2	-1	0	1	2	3	4	5	6	7	8
Fixed Capital Investment		$6,868,153	$51,511,147	$27,472,612								
Land		$0										
Working Capital				$10,731,489								
Loan Payment					$19,191,700	$19,191,700	$19,191,700	$19,191,700	$19,191,700	$19,191,700	$19,191,700	$19,191,700
Loan Interest Payment		$824,178	$7,005,516	$10,302,229	$10,302,229	$9,591,072	$8,823,021	$7,993,527	$7,097,673	$6,130,151	$5,085,227	$3,956,710
Loan Principal		$10,302,229	$87,568,950	$128,777,867	$119,888,397	$110,287,769	$99,919,091	$88,720,918	$76,626,892	$63,565,344	$49,458,871	$34,223,881
Fuel Sales					$2,157,611,929	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905
Co-Product Sales					$15,528,608	$20,704,811	$20,704,811	$20,704,811	$20,704,811	$20,704,811	$20,704,811	$20,704,811
Total Annual Sales					$2,173,140,537	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716
Annual Manufacturing Cost												
Feedstocks					$2,065,137,778	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037
Process Inputs					$61,829,447	$82,439,263	$82,439,263	$82,439,263	$82,439,263	$82,439,263	$82,439,263	$82,439,263
Waste Streams					$5,260,411	$7,013,881	$7,013,881	$7,013,881	$7,013,881	$7,013,881	$7,013,881	$7,013,881
Catalysts					$9,797,798	$0	$9,797,798	$0	$9,797,798	$0	$9,797,798	$0
Royalties					$6,019,065	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420
Fixed Operating Costs					$12,781,510	$12,781,510	$12,781,510	$12,781,510	$12,781,510	$12,781,510	$12,781,510	$12,781,510
Total Product Cost					$2,160,826,010	$2,863,777,112	$2,873,574,910	$2,863,777,112	$2,873,574,910	$2,863,777,112	$2,873,574,910	$2,863,777,112
Annual Depreciation												
General Plant Writedown					14%	24.49%	17.49%	12.49%	8.93%	8.92%	8.93%	4.46%
Depreciation Charge					$30,670,595	$52,562,833	$37,538,748	$26,807,259	$19,166,439	$19,144,976	$19,166,439	$9,572,488
Remaining Value					$183,959,183	$131,396,350	$93,857,602	$67,050,343	$47,883,904	$28,738,927	$9,572,488	$0
Steam Plant Writedown					3.75%	7.22%	6.68%	6.18%	5.71%	5.29%	4.89%	4.52%
Depreciation Charge					$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value					$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue					($28,658,297)	($28,410,300)	($22,415,964)	($1,057,182)	($2,318,307)	$8,468,477	-$305,861	$20,214,406
Losses Forward						($28,658,297)	($57,068,598)	($79,484,561)	($80,541,744)	($82,860,051)	($74,391,574)	($74,697,435)
Taxable Income					($28,658,297)	($57,068,598)	($79,484,561)	($80,541,744)	($82,860,051)	($74,391,574)	($74,697,435)	($54,483,028)
Income Tax					$0	$0	$0	$0	$0	$0	$0	$0
Annual Cash Income					-$6,877,172	$14,551,905	$4,754,106	$14,551,905	$4,754,106	$14,551,905	$4,754,106	$14,551,905
Discount Factor		1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830	0.6209	0.5645	0.5132	0.4665
Annual Present Value	$121,567,374				-$6,251,975	$12,026,367	$3,571,830	$9,939,147	$2,951,926	$8,214,171	$2,439,608	$6,788,571
Total Capital Investment + Interest		$9,307,721	$64,368,329	$48,506,330								
Net Present Worth				$0								

Year	9	10	11	12	13	14	15	16	17	18	19
Fixed Capital Investment											
Land											
Working Capital											
Loan Payment	$19,191,700	$19,191,700	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$2,737,911	$1,421,607	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$17,770,092	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905
Co-Product Sales	$20,704,811	$20,704,811	$20,704,811	$20,704,811	$20,704,811	$20,704,811	$20,704,811	$20,704,811	$20,704,811	$20,704,811	$20,704,811
Total Annual Sales	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716
Annual Manufacturing Cost											
Feedstocks	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037
Process Inputs	$82,439,263	$82,439,263	$82,439,263	$82,439,263	$82,439,263	$82,439,263	$82,439,263	$82,439,263	$82,439,263	$82,439,263	$82,439,263
Waste Streams	$7,013,881	$7,013,881	$7,013,881	$7,013,881	$7,013,881	$7,013,881	$7,013,881	$7,013,881	$7,013,881	$7,013,881	$7,013,881
Catalysts	$9,797,798	$0	$9,797,798	$0	$9,797,798	$0	$9,797,798	$0	$9,797,798	$0	$9,797,798
Royalties	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420
Fixed Operating Costs	$12,781,510	$12,781,510	$12,781,510	$12,781,510	$12,781,510	$12,781,510	$12,781,510	$12,781,510	$12,781,510	$12,781,510	$12,781,510
Total Product Cost	$2,873,574,910	$2,863,777,112	$2,873,574,910	$2,863,777,112	$2,873,574,910	$2,863,777,112	$2,873,574,910	$2,863,777,112	$2,873,574,910	$2,863,777,112	$2,873,574,910
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%
Depreciation Charge	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue	$21,207,895	$32,321,997	$23,945,806	$33,743,604	$23,945,806	$33,743,604	$23,945,806	$33,743,604	$23,945,806	$33,743,604	$23,945,806
Losses Forward	($54,483,028)	-$33,275,133	-$953,136	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	-$33,275,133	-$953,136	$22,992,670	$33,743,604	$23,945,806	$33,743,604	$23,945,806	$33,743,604	$23,945,806	$33,743,604	$23,945,806
Income Tax	$0	$0	$4,828,461	$7,086,157	$5,028,619	$7,086,157	$5,028,619	$7,086,157	$5,028,619	$7,086,157	$5,028,619
Annual Cash Income	$4,754,106	$14,551,905	$19,117,345	$26,657,447	$18,917,187	$26,657,447	$18,917,187	$26,657,447	$18,917,187	$26,657,447	$18,917,187
Discount Factor	0.4241	0.3855	0.3505	0.3186	0.2897	0.2633	0.2394	0.2176	0.1978	0.1799	0.1635
Annual Present Value	$2,016,205	$5,610,389	$6,700,513	$8,493,884	$5,479,635	$7,019,739	$4,528,624	$5,801,437	$3,742,665	$4,794,576	$3,093,111
Total Capital Investment + Interest											
Net Present Worth											

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Year	20	21	22	23	24	25	26	27	28	29	30
Fixed Capital Investment											
Land											$0
Working Capital											($10,731,489)
Loan Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905	$2,876,815,905
Co-Product Sales	$20,704,811	$20,704,811	$20,704,811	$20,704,811	$20,704,811	$20,704,811	$20,704,811	$20,704,811	$20,704,811	$20,704,811	$20,704,811
Total Annual Sales	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716	$2,897,520,716
Annual Manufacturing Cost											
Feedstocks	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037	$2,753,517,037
Process Inputs	$82,439,263	$82,439,263	$82,439,263	$82,439,263	$82,439,263	$82,439,263	$82,439,263	$82,439,263	$82,439,263	$82,439,263	$82,439,263
Waste Streams	$7,013,881	$7,013,881	$7,013,881	$7,013,881	$7,013,881	$7,013,881	$7,013,881	$7,013,881	$7,013,881	$7,013,881	$7,013,881
Catalysts	$0	$9,797,798	$0	$9,797,798	$0	$9,797,798	$0	$9,797,798	$0	$9,797,798	$0
Royalties	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420
Fixed Operating Costs	$12,781,510	$12,781,510	$12,781,510	$12,781,510	$12,781,510	$12,781,510	$12,781,510	$12,781,510	$12,781,510	$12,781,510	$12,781,510
Total Product Cost	$2,863,777,112	$2,873,574,910	$2,863,777,112	$2,873,574,910	$2,863,777,112	$2,873,574,910	$2,863,777,112	$2,873,574,910	$2,863,777,112	$2,873,574,910	$2,863,777,112
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	2.23%									
Depreciation Charge	$0	$0									
Remaining Value	$0	$0									
Net Revenue	$33,743,604	$23,945,806	$33,743,604	$23,945,806	$33,743,604	$23,945,806	$33,743,604	$23,945,806	$33,743,604	$23,945,806	$33,743,604
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$33,743,604	$23,945,806	$33,743,604	$23,945,806	$33,743,604	$23,945,806	$33,743,604	$23,945,806	$33,743,604	$23,945,806	$33,743,604
Income Tax	$7,086,157	$5,028,619	$7,086,157	$5,028,619	$7,086,157	$5,028,619	$7,086,157	$5,028,619	$7,086,157	$5,028,619	$7,086,157
Annual Cash Income	$26,657,447	$18,917,187	$26,657,447	$18,917,187	$26,657,447	$18,917,187	$26,657,447	$18,917,187	$26,657,447	$18,917,187	$26,657,447
Discount Factor	0.1486	0.1351	0.1228	0.1117	0.1015	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
Annual Present Value	$3,962,460	$2,556,290	$3,274,760	$2,112,637	$2,706,413	$1,745,981	$2,236,705	$1,442,959	$1,848,517	$1,192,528	$1,527,700
Total Capital Investment + Interest											($615,006)
Net Present Worth											

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Appendix H. TEA Summary for 50% LCO + 50% VGO Hydroprocessing (HT+HC Configuration)

Case included:

- 42201011: HT+HC of 50 vol % LCO + 50 vol % VGO in H-TC configuration with no liquid recycle, 85 mol % H_2 makeup purity, process model with 75k BPD capacity.

Please see Section 3.4.1 for interpretation of Case IDs.

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Table H-1. Equipment Costs and Fixed Capital Investment for 75k BPD 50% LCO + 50% VGO HT+HC, 85% H_2 Makeup Purity (42201011)

Equip ID	Equipment Description	Process Variable Value	Size Ratio	Scaled Cost (Base Year)	Scaled Cost (2020$)	Pressure (psig)	Pres. Adj. Scaled Cost (2020)	Lang/TCI Factor (Low)	Lang/TCI Factor (High)	Average Lang/TCI Factor	FCI (2020$) for Single Units of Each	FCI (2020$)
C450	Amine Scrubber	726	3.02	$ 271,433	$ 203,047		203,047	8.858234	13.73793	11.29808	$ 2,294,037	$ 2,294,037
C470	Product Recovery Column 1	1036504	1.09	$ 5,459,029	$ 4,083,655		4,083,655	2.917894	3.559321	3.238608	$ 13,225,356	$ 13,225,356
C480	Product Recovery Column 2	891015	1.31	$ 4,325,880	$ 3,235,997		3,235,997	3.180201	3.95217	3.566185	$ 11,540,167	$ 11,540,167
MX405	CFP and H2 Static Mixer	0	0.00	$ -	$ -		-	0	0	0	$ -	$ -
MX406	Rx Inlet Static Mixer	0	0.00	$ -	$ -		-	0	0	0	$ -	$ -
P400A/B	Cat Pyrolysis Oil 2 Pumps	0	0.00	$ -	$ -	2,085	-	0	0	0	$ -	$ -
P405A/B	Liquid Feed 2 Pumps	75000	1.00	$ 1,538,999	$ 1,151,256	2,085	2,881,828	8620664	11771299	10195982	$ 10,195,982	$ 20,391,964
P455A/B	Liquid Recycle 2 Pumps	0	0.00	$ -	$ -	2,035	-	0	0	0	$ -	$ -
P471A/B	Prod Recov Reflux 2 Pumps - 1	1036504	1.09	$ 62,871	$ 47,031	25	47,135	266747.1	438296.3	352521.7	$ 352,522	$ 705,043
P472A/B	Prod Recov Btms 2 Pumps - 1	1036504	1.09	$ 67,111	$ 50,203	30	48,422	272885	447838.7	360361.9	$ 360,362	$ 720,724
P481A/B	Prod Recov Reflux 2 Pumps - 2	891015	1.31	$ 60,267	$ 45,083	20	46,137	261963.5	430851.2	346407.3	$ 346,407	$ 692,815
P482A/B	Prod Recov Btms 2 Pumps - 2	891015	1.31	$ 45,441	$ 33,992	15	33,061	197670.3	330020.6	263845.4	$ 263,845	$ 527,691
PIPE412	H2 MU Comp Jumpover Line	34606	2.84	$ 13,294	$ 9,945		9,945	6.5	7	6.75	$ 67,127	$ 67,127
PIPE442	H2 Rec Comp Jumpover Line	115179	1.52	$ 15,640	$ 11,700		11,700	6.5	7	6.75	$ 78,973	$ 78,973
PIPE500	CFP Pipe from Storage	0	0.00	$ -	$ -		-	0	0	0	$ -	$ -
PSA460	Pressure Swing Adsorption	137098297	1.64	$ 7,740,323	$ 5,790,189		5,790,189	3.6	4.4	4	$ 23,160,754	$ 23,160,754
QAH413	Rx Effluent Air Cooler	67674311	0.53	$ 1,235,671	$ 924,350		924,350	4.75	5.75	5.25	$ 4,852,839	$ 4,852,839
QCH413	Rx Effluent Water Cooler	57651926	1.47	$ 1,505,274	$ 1,126,028	1,985	4,291,259	19509176	22998110	21253643	$ 21,253,643	$ 21,253,643
QCH444	Water Cooler H2 Quench 1	6204044	1.31	$ 237,751	$ 177,851	2,085	887,532	5167605	6910263	6038934	$ 6,038,934	$ 6,038,934
QCH445	Water Cooler H2 Quench 2	6204044	1.31	$ 237,751	$ 177,851	2,085	887,532	5167605	6910263	6038934	$ 6,038,934	$ 6,038,934
QFH411-L	Reactor Preheat Furnace - Liquid Feed	46801417	1.10	$ 1,455,644	$ 1,088,902	2,085	1,088,902	4.23	6.73	5.48	$ 7,328,475	$ 7,328,475
QFH411-G	Reactor Preheat Furnace - Gas Feed	0	0.00	$ -	$ -	2,085	-	0	0	0	$ -	$ -
QH412	Process Rx Eff Cooler	255856537	1.87	$ 6,944,341	$ 5,194,750	1,995	19,175,341	68916375	72071402	70493889	$ 70,493,889	$ 70,493,889
QH471	Prod Recov Condenser 1	1036504	1.09	$ 4,670,162	$ 3,493,539	25	3,493,539	16403742	19658106	18030924	$ 18,030,924	$ 18,030,924
QH472	Prod Recov Reboiler 1	1036504	1.09	$ 4,245,569	$ 3,175,920	30	3,175,920	15137217	18279170	16708193	$ 16,708,193	$ 16,708,193
QH481	Prod Recov Condenser 2	891015	1.31	$ 2,696,563	$ 2,017,178	20	2,017,178	10324507	12928549	11626528	$ 11,626,528	$ 11,626,528
QH482	Prod Recov Reboiler 2	891015	1.31	$ 2,055,400	$ 1,537,553	15	1,537,553	8212356	10509483	9360919	$ 9,360,919	$ 9,360,919
QXH410	Feed Preheater to Furnace	452036050	1.23	$ 1,513,643	$ 1,132,288	2,085	4,484,989	20249058	23786136	22017597	$ 22,017,597	$ 22,017,597
RX405	CFP Cat Stabilization Rx	0	0.00	$ -	$ -	2,085	-	0	0	0	$ -	$ -
RX410-HT	Reactor System - Hydrotreater	1768564	1.39	$ 15,184,507	$ 11,358,849	2,030	15,961,079	3.5	5.3	4.4	$ 70,228,748	$ 70,228,748
RX410-HC	Reactor System - Hydrocracker	1768564	1.39	$ 15,184,507	$ 11,358,849	2,030	15,961,079	3.5	5.3	4.4	$ 70,228,748	$ 70,228,748
TK400	Cat Pyrolysis Oil Storage	0	0.00	$ -	$ -		-	0	0	0	$ -	$ -
V451	High Pressure Flash Drum	2003666	1.57	$ 1,248,861	$ 934,217	1,980	4,949,847	3.889976	5.666953	4.778465	$ 23,652,671	$ 23,652,671
V452	Low Pressure Flash Drum	1102254	1.11	$ 238,702	$ 178,563	2,085	178,563	12.86339	24.44456	18.65397	$ 4,364,884	$ 4,364,884
V471	Prod Recov Ovhd Drum 1	409796	0.43	$ 41,615	$ 31,131	2,085	31,131	24.12417	52.71903	38.4216	$ 1,641,176	$ 1,641,176
V481	Prod Recov Ovhd Drum 2	794200	1.17	$ 71,607	$ 53,566	2,085	53,566	19.84265	41.52012	30.68138	$ 2,224,069	$ 2,224,069
WK412	H2 Make-up Compressor	51849	3.89	$ 33,666,338	$ 25,184,280		25,184,280	3	3.8	3.4	$ 85,626,551	$ 85,626,551
WK442	H2 Recycle Compressor	3892	0.98	$ 3,035,062	$ 2,270,394		2,270,394	3	3.8	3.4	$ 7,719,338	$ 7,719,338
QH470	Dist Col 1 Liquid Feed Preheat Furnace	19383848	0.46	$ 926,941	$ 693,403	36	693,403	4.972626	8.208545	6.590585	$ 3,448,032	$ 3,448,032
Total												**$ 536,289,740**

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Table H-2. Operating Costs for 75k BPD 50% LCO + 50% VGO HT+HC (42201011)

	Quoted Price	Unit	Cost Year	2020$ price	Scaling Flow	Unit	2020$ Cost ($/h)	2020$ Cost (MM$/y)	2020$ Cost (cents/GGE)
Feedstocks									
Crude Oil Basis	$ 85.00	/bbl	2020	$ 85.00					
Straight Run Diesel (SRD)	$ 105.88	/bbl	2020	$ 105.88	0	bbl/day	0	0.00	0.0
Light Cycle Oil (LCO)	$ 102.95	/bbl	2020	$ 102.95	37,500	bbl/day	160,863	1,338.70	101.2
Vacuuum Gas Oil (VGO)	$ 100.78	/bbl	2020	$ 100.78	37,500	bbl/day	157,474	1,310.50	99.0
Catalytic Fast Pyrolysis Oil (21.6%)	$ 215.62	/bbl	2020	$ 215.62	0	bbl/day	1	0.01	0.0
Bio-Oil Transportation 800 miles	$ -	/ton-mile	2007	$ -	1	lb/h	0	0.00	0.0
Subtotal							**318,338**	**2,649.21**	**200.2**
Process Inputs									
Hydrogen (pure H2 basis)									
Total Hydrogen Feed					59,297	lb/h			
Hydrogen Recovered via PSA					20,079	lb/h			
Net Hydrogen Makeup	$ 0.680	/lb	2020	$ 0.680	39,218	lb/h	26,684	222.06	16.8
Electricity for Hydroprocessing	$ 0.082	/kWh	2020	$ 0.082	51,899	kWh	4,240	35.29	2.7
Natural Gas	$ 3.231	/MMBtu	2020	$ 3.231	0	MMBtu/h	0	0.00	0.0
Fuel Gas	$ 2.746	/MMBtu	2020	$ 2.746	296	MMBtu/h	813	6.76	0.5
Amine Consumption	$ 0.00122600	/lbmol inlet gas	2020	$ 0.00122600	117,908	lbmol/h	145	1.20	0.091
Water Makeup	$ 0.200	/ton	2004	$ 0.334	118	ton/h	39	0.33	0.0
BFW to Steam	$ 0.001	/lb	2016	$ 0.001		lb/h			
Cooling Water Circulation	$ 0.00768	/ton	2020	$ 0.008	1,971	ton/h	15	0.13	0.0
Boiler Chemicals	$ 75.00	/MM lb bldwn	2014	$ 80.516		MMlb/h			
Cooling Tower Chemicals	$ 33.84	/ton-cool/y	2014	$ 36.329		ton-cool/y			
Diesel Fuel	$ 2.57	/gallon	2020	$ 2.571	10	gallon/h	26	0.21	0.0
Subtotal							**31,961**	**265.98**	**20.1**
Waste Streams									
Wastewater Treatment	$ 0.003	/lb	2011	$ 0.003	235,852	lb/h	714	5.94	0.4
Acid Gas Conversion and Recovery	$ 0.477	/lbmol acid gas	2020	$ 0.477	726	lbmol/h	347	2.88	0.2
Subtotal							**1,060**	**8.82**	**0.7**
Coproducts									
Fuel Gas	$ (2.746)	/MMBtu	2020	$ (2.746)	938	MMBtu/h	-2,577	-21.44	-1.6
HP Steam (900 psig)	$ (0.0045)	/lb	2023	$ (0.004)	0	lb/h	0	0.00	0.0
MP Steam (350 psig)	$ (0.0045)	/lb	2023	$ (0.004)	45,446	lb/h	-167	-1.39	-0.1
LP Steam (150 psig)	$ (0.0045)	/lb	2023	$ (0.004)	156,352	lb/h	-571	-4.76	-0.4
VLP Steam (15 psig)	$ (0.0044)	/lb	2023	$ (0.004)	260,834	lb/h	-927	-7.72	-0.6
Subtotal							**-4,243**	**-35.31**	**-2.7**
Catalysts and Royalties									
Royalties	$ 0.20	/Bbl Feed	2005	$ 0.309	3,125	Bbl/h	964	8.03	0.6
Subtotal							**964**	**8.03**	**0.6**
Total Variable Operating Costs							**348,081**	**2,896.73**	**218.9**
Fixed Operating Costs									
Total Salaries								2.55	0.2
Overhead and Benefits	% of Labor & Supervison =		0.9					2.29	0.2
Maintenance	% of FCI =		0.03					16.1	1.2
Insurance & Taxes	% of FCI =		0.007					3.8	0.3
Total Fixed Operating Costs								**24.68**	**1.9**
Total Operating Costs								**2,921.41**	**220.8**

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Table H-3. DCFROR for 75k BPD 50% LCO + 50% VGO HT+HC (42201011)

Year		-2	-1	0	1	2	3	4	5	6	7	8
Fixed Capital Investment		$17,161,272	$128,709,538	$68,645,087								
Land		$0										
Working Capital				$26,814,487								
Loan Payment					$47,953,791	$47,953,791	$47,953,791	$47,953,791	$47,953,791	$47,953,791	$47,953,791	$47,953,791
Loan Interest Payment		$2,059,353	$17,504,497	$25,741,908	$25,741,908	$23,964,957	$22,045,850	$19,973,215	$17,734,769	$15,317,247	$12,706,323	$9,886,526
Loan Principal		$25,741,908	$218,806,214	$321,773,844	$299,561,960	$275,573,126	$249,665,184	$221,684,607	$191,465,584	$158,829,040	$123,581,572	$85,514,306
Fuel Sales					$2,256,848,844	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792
Co-Product Sales					$26,481,473	$35,308,631	$35,308,631	$35,308,631	$35,308,631	$35,308,631	$35,308,631	$35,308,631
Total Annual Sales					$2,283,330,317	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423
Annual Manufacturing Cost												
Feedstocks					$1,986,907,675	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234
Process Inputs					$199,485,292	$265,980,389	$265,980,389	$265,980,389	$265,980,389	$265,980,389	$265,980,389	$265,980,389
Waste Streams					$6,617,873	$8,823,831	$8,823,831	$8,823,831	$8,823,831	$8,823,831	$8,823,831	$8,823,831
Catalysts					$36,613,677	$0	$18,306,838	$18,306,838	$18,306,838	$0	$36,613,677	$0
Royalties					$6,019,065	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420
Fixed Operating Costs					$24,682,929	$24,682,929	$24,682,929	$24,682,929	$24,682,929	$24,682,929	$24,682,929	$24,682,929
Total Product Cost					$2,260,326,512	$2,956,722,803	$2,975,029,642	$2,975,029,642	$2,975,029,642	$2,956,722,803	$2,993,336,480	$2,956,722,803
Annual Depreciation												
General Plant Writedown					14%	24.49%	17.49%	12.49%	8.93%	8.92%	8.93%	4.46%
Depreciation Charge					$76,635,804	$131,337,357	$93,797,076	$66,982,589	$47,890,674	$47,837,045	$47,890,674	$23,918,522
Remaining Value					$459,653,936	$328,316,579	$234,519,503	$167,536,915	$119,646,241	$71,809,196	$23,918,522	$0
Steam Plant Writedown					3.75%	7.22%	6.68%	6.18%	5.71%	5.29%	4.89%	4.52%
Depreciation Charge					$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value					$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue					($79,373,906)	($67,584,695)	($46,432,145)	($17,545,022)	$3,785,339	$24,563,328	-$9,493,054	$53,912,571
Losses Forward						($79,373,906)	($146,958,601)	($193,390,745)	($210,935,768)	($207,150,429)	($182,587,101)	($192,080,156)
Taxable Income					($79,373,906)	($146,958,601)	($193,390,745)	($210,935,768)	($207,150,429)	($182,587,101)	($192,080,156)	($138,167,585)
Income Tax					$0	$0	$0	$0	$0	$0	$0	$0
Annual Cash Income					-$24,949,986	$39,763,828	$21,456,989	$21,456,989	$21,456,989	$39,763,828	$3,150,151	$39,763,828
Discount Factor		1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830	0.6209	0.5645	0.5132	0.4665
Annual Present Value	$303,757,175				-$22,681,805	$32,862,668	$16,120,954	$14,655,413	$13,323,102	$22,445,644	$1,616,526	$18,550,119
Total Capital Investment + Interest		$23,256,955	$160,835,438	$121,201,481								
Net Present Worth				$0								

Year	9	10	11	12	13	14	15	16	17	18	19
Fixed Capital Investment											
Land											
Working Capital											
Loan Payment	$47,953,791	$47,953,791	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$6,841,144	$3,552,133	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$44,401,659	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792
Co-Product Sales	$35,308,631	$35,308,631	$35,308,631	$35,308,631	$35,308,631	$35,308,631	$35,308,631	$35,308,631	$35,308,631	$35,308,631	$35,308,631
Total Annual Sales	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423
Annual Manufacturing Cost											
Feedstocks	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234
Process Inputs	$265,980,389	$265,980,389	$265,980,389	$265,980,389	$265,980,389	$265,980,389	$265,980,389	$265,980,389	$265,980,389	$265,980,389	$265,980,389
Waste Streams	$8,823,831	$8,823,831	$8,823,831	$8,823,831	$8,823,831	$8,823,831	$8,823,831	$8,823,831	$8,823,831	$8,823,831	$8,823,831
Catalysts	$18,306,838	$18,306,838	$18,306,838	$0	$36,613,677	$0	$18,306,838	$18,306,838	$18,306,838	$0	$36,613,677
Royalties	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420
Fixed Operating Costs	$24,682,929	$24,682,929	$24,682,929	$24,682,929	$24,682,929	$24,682,929	$24,682,929	$24,682,929	$24,682,929	$24,682,929	$24,682,929
Total Product Cost	$2,975,029,642	$2,975,029,642	$2,975,029,642	$2,956,722,803	$2,993,336,480	$2,956,722,803	$2,975,029,642	$2,975,029,642	$2,975,029,642	$2,956,722,803	$2,993,336,480
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%
Depreciation Charge	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue	$62,569,636	$65,858,648	$69,410,781	$87,717,619	$51,103,943	$87,717,619	$69,410,781	$69,410,781	$69,410,781	$87,717,619	$51,103,943
Losses Forward	($138,167,585)	-$75,597,948	-$9,739,300	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	-$75,597,948	-$9,739,300	$59,671,481	$87,717,619	$51,103,943	$87,717,619	$69,410,781	$69,410,781	$69,410,781	$87,717,619	$51,103,943
Income Tax	$0	$0	$12,531,011	$18,420,700	$10,731,828	$18,420,700	$14,576,264	$14,576,264	$14,576,264	$18,420,700	$10,731,828
Annual Cash Income	$21,456,989	$21,456,989	$56,879,770	$69,296,919	$40,372,115	$69,296,919	$54,834,517	$54,834,517	$54,834,517	$69,296,919	$40,372,115
Discount Factor	0.4241	0.3855	0.3505	0.3186	0.2897	0.2633	0.2394	0.2176	0.1978	0.1799	0.1635
Annual Present Value	$9,099,858	$8,272,598	$19,936,012	$22,080,134	$11,694,364	$18,248,045	$13,126,947	$11,933,589	$10,848,717	$12,463,660	$6,601,163
Total Capital Investment + Interest											
Net Present Worth											

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Year	20	21	22	23	24	25	26	27	28	29	30
Fixed Capital Investment											
Land											$0
Working Capital											($26,814,487)
Loan Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792	$3,009,131,792
Co-Product Sales	$35,308,631	$35,308,631	$35,308,631	$35,308,631	$35,308,631	$35,308,631	$35,308,631	$35,308,631	$35,308,631	$35,308,631	$35,308,631
Total Annual Sales	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423	$3,044,440,423
Annual Manufacturing Cost											
Feedstocks	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234	$2,649,210,234
Process Inputs	$265,980,389	$265,980,389	$265,980,389	$265,980,389	$265,980,389	$265,980,389	$265,980,389	$265,980,389	$265,980,389	$265,980,389	$265,980,389
Waste Streams	$8,823,831	$8,823,831	$8,823,831	$8,823,831	$8,823,831	$8,823,831	$8,823,831	$8,823,831	$8,823,831	$8,823,831	$8,823,831
Catalysts	$0	$18,306,838	$18,306,838	$18,306,838	$0	$36,613,677	$0	$18,306,838	$18,306,838	$18,306,838	$0
Royalties	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420
Fixed Operating Costs	$24,682,929	$24,682,929	$24,682,929	$24,682,929	$24,682,929	$24,682,929	$24,682,929	$24,682,929	$24,682,929	$24,682,929	$24,682,929
Total Product Cost	$2,956,722,803	$2,975,029,642	$2,975,029,642	$2,975,029,642	$2,956,722,803	$2,993,336,480	$2,956,722,803	$2,975,029,642	$2,975,029,642	$2,975,029,642	$2,956,722,803
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	2.23%									
Depreciation Charge	$0	$0									
Remaining Value	$0	$0									
Net Revenue	$87,717,619	$69,410,781	$69,410,781	$69,410,781	$87,717,619	$51,103,943	$87,717,619	$69,410,781	$69,410,781	$69,410,781	$87,717,619
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$87,717,619	$69,410,781	$69,410,781	$69,410,781	$87,717,619	$51,103,943	$87,717,619	$69,410,781	$69,410,781	$69,410,781	$87,717,619
Income Tax	$18,420,700	$14,576,264	$14,576,264	$14,576,264	$18,420,700	$10,731,828	$18,420,700	$14,576,264	$14,576,264	$14,576,264	$18,420,700
Annual Cash Income	$69,296,919	$54,834,517	$54,834,517	$54,834,517	$69,296,919	$40,372,115	$69,296,919	$54,834,517	$54,834,517	$54,834,517	$69,296,919
Discount Factor	0.1486	0.1351	0.1228	0.1117	0.1015	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
Annual Present Value	$10,300,545	$7,409,820	$6,736,200	$6,123,818	$7,035,411	$3,726,185	$5,814,389	$4,182,650	$3,802,409	$3,456,736	$3,971,306
Total Capital Investment + Interest											($1,536,699)
Net Present Worth											

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Appendix I. TEA Summary for CFP-A Hydroprocessing and Co-Hydroprocessing

Cases included:

- 12199011: HT of 100% CFP-A oil in H-TC configuration with no liquid recycle, 99 mol % H_2 makeup purity, process model with 75k BPD capacity.
- 42199011: HT+HC of 100% CFP-A oil in H-TC configuration with no liquid recycle, 99 mol % H_2 makeup purity, process model with 75k BPD capacity.
- 12199513: HT of 100% CFP-A oil in H-TC configuration with 50% liquid recycle, 99 mol % H_2 makeup purity, process model with 10k BPD capacity.
- 12120211: HT of 20 vol % CFP-A oil with 80 vol % SRD in H-TC configuration with 20% liquid recycle, 99% H_2 makeup purity, process model with 75k BPD capacity.
- 12120811: HT of 20 vol% CFP-A oil with 80 vol % SRD in L-TC configuration, 99 mol % H_2 makeup purity, process model with 75k BPD capacity.

Please see Section 3.4.1 for interpretation of Case IDs.

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Table I-1. Equipment Costs and Fixed Capital Investment for 75k BPD CFP-A (21.6% O) HT Configuration Without Liquid Recycle (12199011)

Equip ID	Equipment Description	Process Variable Value	Size Ratio	Scaled Cost (Base Year)	Scaled Cost (2020$)	Pressure (psig)	Pres. Adj. Scaled Cost (2020)	Lang/TCI Factor (Low)	Lang/TCI Factor (High)	Average Lang/TCI Factor	FCI (2020$) for Single Units of Each	FCI (2020$)
C450	Amine Scrubber	903	3.75	$ 323,250	$ 241,809		241,809	8.303726	12.69922	10.50147	$ 2,539,347	$ 2,539,347
C470	Product Recovery Column 1	918328	0.97	$ 5,070,685	$ 3,793,152		3,793,152	2.998662	3.679503	3.339083	$ 12,665,648	$ 12,665,648
C480	Product Recovery Column 2	654981	0.96	$ 3,619,603	$ 2,707,663		2,707,663	3.397016	4.282247	3.839632	$ 10,396,428	$ 10,396,428
MX405	CFP and H2 Static Mixer	75000	7.50	$ 15,195	$ 11,367		11,367	90	100	95	$ 1,079,850	$ 1,079,850
MX406	Rx Inlet Static Mixer	75000	10.00	$ 59,583	$ 44,572		44,572	90	100	95	$ 4,234,312	$ 4,234,312
P400A/B	Cat Pyrolysis Oil 2 Pumps	75000	7.50	$ 905,506	$ 677,368	1,085	599,839	2288551	3353650	2821101	$ 2,821,101	$ 5,642,201
P405A/B	Liquid Feed 2 Pumps	0	0.00	$ -	$ -	1,085	-	0	0	0	$ -	$ -
P455A/B	Liquid Recycle 2 Pumps	0	0.00	$ -	$ -	1,035	-	0	0	0	$ -	$ -
P471A/B	Prod Recov Reflux 2 Pumps - 1	918328	0.97	$ 60,648	$ 45,368	25	45,469	258757.6	425857.7	342307.6	$ 342,308	$ 684,615
P472A/B	Prod Recov Btms 2 Pumps - 1	918328	0.97	$ 64,570	$ 48,302	30	46,586	264118.2	434205.7	349162	$ 349,162	$ 698,324
P481A/B	Prod Recov Reflux 2 Pumps - 2	654981	0.96	$ 55,012	$ 41,152	20	42,116	242538.1	400542.5	321540.3	$ 321,540	$ 643,081
P482A/B	Prod Recov Btms 2 Pumps - 2	654981	0.96	$ 42,463	$ 31,765	15	30,894	186663.4	312596.3	249629.8	$ 249,630	$ 499,260
PIPE412	H2 MU Comp Jumpover Line	61910	5.08	$ 19,975	$ 14,942		14,942	6.5	7	6.75	$ 100,861	$ 100,861
PIPE442	H2 Rec Comp Jumpover Line	439846	5.80	$ 39,957	$ 29,890		29,890	6.5	7	6.75	$ 201,755	$ 201,755
PIPE500	CFP Pipe from Storage	75000	7.50	$ 198,322	$ 148,356		148,356	6.5	7	6.75	$ 1,001,404	$ 1,001,404
PSA460	Pressure Swing Adsorption	179553208	2.15	$ 8,846,140	$ 6,617,401		6,617,401	3.6	4.4	4	$ 26,469,604	$ 26,469,604
QAH413	Rx Effluent Air Cooler	703424905	5.55	$ 9,892,915	$ 7,400,447		7,400,447	4.75	5.75	5.25	$ 38,852,346	$ 38,852,346
QCH413	Rx Effluent Water Cooler	145191044	3.70	$ 3,302,580	$ 2,470,512	985	5,764,526	25020368	28806639	26913504	$ 26,913,504	$ 26,913,504
QCH444	Water Cooler H2 Quench 1	28782676	6.08	$ 859,039	$ 642,608	1,085	1,556,753	8298722	10609469	9454095	$ 9,454,095	$ 9,454,095
QCH445	Water Cooler H2 Quench 2	28782676	6.08	$ 859,039	$ 642,608	1,085	1,556,753	8298722	10609469	9454095	$ 9,454,095	$ 9,454,095
QFH411-L	Reactor Preheat Furnace - Liquid Feed	99130795	2.33	$ 2,137,674	$ 1,599,098	1,085	1,599,098	3.68	5.68	4.68	$ 5,886,005	$ 5,886,005
QFH411-G	Reactor Preheat Furnace - Gas Feed	0	0.00	$ -	$ -	1,085	-	0	0	0	$ -	$ -
QH412	Process Rx Eff Cooler	760910837	5.55	$ 20,330,695	$ 15,208,482	995	36,038,552	1.17E+08	1.17E+08	1.17E+08	$ 116,973,083	$ 116,973,083
QH471	Prod Recov Condenser 1	918328	0.97	$ 4,227,309	$ 3,162,261	25	3,162,261	15082316	18219154	16650735	$ 16,650,735	$ 16,650,735
QH472	Prod Recov Reboiler 1	918328	0.97	$ 3,845,305	$ 2,876,501	30	2,876,501	13924921	16948981	15436951	$ 15,436,951	$ 15,436,951
QH481	Prod Recov Condenser 2	654981	0.96	$ 2,125,074	$ 1,589,673	20	1,589,673	8446416	10780225	9613320	$ 9,613,320	$ 9,613,320
QH482	Prod Recov Reboiler 2	654981	0.96	$ 1,618,699	$ 1,210,876	15	1,210,876	6714644	8758606	7736625	$ 7,736,625	$ 7,736,625
QXH410	Feed Preheater to Furnace	995909704	2.71	$ 3,044,267	$ 2,277,280	1,085	5,625,759	24511659	28276015	26393837	$ 26,393,837	$ 26,393,837
RX405	CFP Cat Stabilization Rx	75000	7.50	$ 28,245	$ 21,128	1,085	20,683	35.5	45.5	40.5	$ 837,681	$ 837,681
RX410-HT	Reactor System - Hydrotreater	3681426	2.89	$ 29,570,167	$ 22,120,117	1,030	21,494,347	3.5	5.3	4.4	$ 94,575,128	$ 94,575,128
RX410-HC	Reactor System - Hydrocracker	3681426	2.89	$ 29,570,167	$ 22,120,117	1,030	21,494,347	3.5	5.3	4.4	$ 94,575,128	$ -
TK400	Cat Pyrolysis Oil Storage	75000	7.50	$ 14,423,781	$ 10,789,784		10,789,784	6	7.2	6.6	$ 71,212,576	$ 71,212,576
V451	High Pressure Flash Drum	4103789	3.22	$ 2,476,283	$ 1,852,396	980	2,441,201	5.017215	7.734361	6.375788	$ 15,564,580	$ 15,564,580
V452	Low Pressure Flash Drum	934300	0.94	$ 211,412	$ 158,148	1,085	158,148	13.43807	25.78587	19.61197	$ 2,125,206	$ 2,125,206
V471	Prod Recov Ovhd Drum 1	792588	0.84	$ 51,914	$ 38,834	1,085	38,834	22.27831	47.83162	35.05496	$ 865,163	$ 865,163
V481	Prod Recov Ovhd Drum 2	845909	1.25	$ 73,154	$ 54,723	1,085	54,723	19.69054	41.13142	30.41098	$ 1,077,534	$ 1,077,534
WK412	H2 Make-up Compressor	61993	4.65	$ 38,813,506	$ 29,034,645		29,034,645	3	3.8	3.4	$ 98,717,793	$ 98,717,793
WK442	H2 Recycle Compressor	28662	7.21	$ 5,566,706	$ 4,164,203		4,164,203	3	3.8	3.4	$ 14,158,291	$ 14,158,291
QH470	Dist Col 1 Liquid Feed Preheat Furnace	66103302	1.56	$ 1,737,148	$ 1,299,483	36	1,299,483	3.966278	6.226454	5.096366	$ 5,154,110	$ 5,154,110
Total												**$ 654,509,349**

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Table I-2. Operating Costs for 75k BPD CFP-A (21.6% O) HT Configuration (12199011)

	Quoted Price	Unit	Cost Year	2020$ price	Scaling Flow	Unit	2020$ Cost ($/h)	2020$ Cost (MM$/y)	2020$ Cost (cents/GGE)
Feedstocks									
Crude Oil Basis	$ 85.00	/bbl	2020	$ 85.00					
Straight Run Diesel (SRD)	$ 105.88	/bbl	2020	$ 105.88	0	bbl/day	0	0.00	0.0
Light Cycle Oil (LCO)	$ 102.95	/bbl	2020	$ 102.95	0	bbl/day	0	0.00	0.0
Vacuuum Gas Oil (VGO)	$ 100.78	/bbl	2020	$ 100.78	0	bbl/day	0	0.00	0.0
Catalytic Fast Pyrolysis Oil (21.6%)	$ 215.62	/bbl	2020	$ 215.62	75,000	bbl/day	673,801	5,607.37	473.7
Bio-Oil Transportation 800 miles	$ -	/ton-mile	2007	$ -	1,260,728	lb/h	0	0.00	0.0
Subtotal							**673,801**	**5,607.37**	**473.7**
Process Inputs									
Hydrogen (pure H2 basis)									
Total Hydrogen Feed					123,555	lb/h			
Hydrogen Recovered via PSA					25,964	lb/h			
Net Hydrogen Makeup	$ 0.680	/lb	2020	$ 0.680	97,591	lb/h	66,400	552.58	46.7
Electricity for Hydroprocessing	$ 0.082	/kWh	2020	$ 0.082	81,398	kWh	6,650	55.34	4.7
Natural Gas	$ 3.231	/MMBtu	2020	$ 3.231	0	MMBtu/h	0	0.00	0.0
Fuel Gas	$ 2.746	/MMBtu	2020	$ 2.746	455	MMBtu/h	1,248	10.39	0.9
Amine Consumption	$ 0.00122600	/lbmol inlet gas	2020	$ 0.00122600	445,812	lbmol/h	547	4.55	0.384
Water Makeup	$ 0.200	/ton	2004	$ 0.334	211	ton/h	70	0.59	0.0
BFW to Steam	$ 0.001	/lb	2016	$ 0.001		lb/h			
Cooling Water Circulation	$ 0.00768	/ton	2020	$ 0.008	5,083	ton/h	39	0.32	0.0
Boiler Chemicals	$ 75.00	/MM lb bldwn	2014	$ 80.516		MMlb/h			
Cooling Tower Chemicals	$ 33.84	/ton-cool/y	2014	$ 36.329		ton-cool/y			
Diesel Fuel	$ 2.57	/gallon	2020	$ 2.571	10	gallon/h	26	0.21	0.0
Subtotal							**74,981**	**623.99**	**52.7**
Waste Streams									
Wastewater Treatment	$ 0.003	/lb	2011	$ 0.003	723,582	lb/h	2,190	18.22	1.5
Acid Gas Conversion and Recovery	$ 0.477	/lbmol acid gas	2020	$ 0.477	903	lbmol/h	431	3.58	0.3
Subtotal							**2,620**	**21.81**	**1.8**
Coproducts									
Fuel Gas	$ (2.746)	/MMBtu	2020	$ (2.746)	2,545	MMBtu/h	-6,989	-58.17	-4.9
HP Steam (900 psig)	$ (0.0045)	/lb	2023	$ (0.004)	0	lb/h	0	0.00	0.0
MP Steam (350 psig)	$ (0.0045)	/lb	2023	$ (0.004)	0	lb/h	0	0.00	0.0
LP Steam (150 psig)	$ (0.0045)	/lb	2023	$ (0.004)	408,169	lb/h	-1,492	-12.42	-1.0
VLP Steam (15 psig)	$ (0.0044)	/lb	2023	$ (0.004)	641,162	lb/h	-2,280	-18.97	-1.6
Subtotal							**-10,761**	**-89.55**	**-7.6**
Catalysts and Royalties									
Royalties	$ 0.20	/Bbl Feed	2005	$ 0.309	3,125	Bbl/h	964	8.03	0.7
Subtotal							**964**	**8.03**	**0.7**
Total Variable Operating Costs							**741,606**	**6,171.64**	**521.4**
Fixed Operating Costs									
Total Salaries								2.55	0.2
Overhead and Benefits	% of Labor & Supervison =		0.9					2.29	0.2
Maintenance	% of FCI =		0.03					19.6	1.7
Insurance & Taxes	% of FCI =		0.007					4.6	0.4
Total Fixed Operating Costs								**29.06**	**2.5**
Total Operating Costs								**6,200.70**	**523.9**

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Table I-3. DCFROR for 75k BPD CFP-A (21.6% O) HT Configuration (12199011)

Year		-2	-1	0	1	2	3	4	5	6	7	8
Fixed Capital Investment		$20,944,299	$157,082,244	$83,777,197								
Land		$0										
Working Capital				$32,725,467								
Loan Payment					$58,524,716	$58,524,716	$58,524,716	$58,524,716	$58,524,716	$58,524,716	$58,524,716	$58,524,716
Loan Interest Payment		$2,513,316	$21,363,185	$31,416,449	$31,416,449	$29,247,787	$26,905,633	$24,376,106	$21,644,218	$18,693,778	$15,507,303	$12,065,910
Loan Principal		$31,416,449	$267,039,814	$392,705,609	$365,597,342	$336,320,413	$304,701,330	$270,552,720	$233,672,222	$193,841,283	$150,823,870	$104,365,063
Fuel Sales					$4,764,711,893	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191
Co-Product Sales					$67,164,876	$89,553,168	$89,553,168	$89,553,168	$89,553,168	$89,553,168	$89,553,168	$89,553,168
Total Annual Sales					$4,831,876,769	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358
Annual Manufacturing Cost												
Feedstocks					$4,205,531,004	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672
Process Inputs					$467,991,054	$623,988,072	$623,988,072	$623,988,072	$623,988,072	$623,988,072	$623,988,072	$623,988,072
Waste Streams					$16,355,016	$21,806,688	$21,806,688	$21,806,688	$21,806,688	$21,806,688	$21,806,688	$21,806,688
Catalysts					$121,704,941	$0	$121,704,941	$0	$121,704,941	$0	$121,704,941	$0
Royalties					$6,019,065	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420
Fixed Operating Costs					$29,057,054	$29,057,054	$29,057,054	$29,057,054	$29,057,054	$29,057,054	$29,057,054	$29,057,054
Total Product Cost					$4,846,658,135	$6,290,251,907	$6,411,956,848	$6,290,251,907	$6,411,956,848	$6,290,251,907	$6,411,956,848	$6,290,251,907
Annual Depreciation												
General Plant Writedown					14%	24.49%	17.49%	12.49%	8.93%	8.92%	8.93%	4.46%
Depreciation Charge					$93,529,386	$160,289,339	$114,473,685	$81,748,218	$58,447,685	$58,382,234	$58,447,685	$29,191,117
Remaining Value					$560,979,963	$400,690,623	$286,216,938	$204,468,720	$146,021,036	$87,638,802	$29,191,117	$0
Steam Plant Writedown					3.75%	7.22%	6.68%	6.18%	5.71%	5.29%	4.89%	4.52%
Depreciation Charge					$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value					$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue					($139,727,201)	($37,286,676)	($110,833,808)	$46,126,127	($49,546,392)	$75,174,439	-$43,409,478	$110,993,425
Losses Forward						($139,727,201)	($177,013,877)	($287,847,685)	($241,721,558)	($291,267,950)	($216,093,511)	($259,502,989)
Taxable Income					($139,727,201)	($177,013,877)	($287,847,685)	($241,721,558)	($291,267,950)	($216,093,511)	($259,502,989)	($148,509,564)
Income Tax					$0	$0	$0	$0	$0	$0	$0	$0
Annual Cash Income					-$73,306,083	$93,725,735	-$27,979,206	$93,725,735	-$27,979,206	$93,725,735	-$27,979,206	$93,725,735
Discount Factor		1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830	0.6209	0.5645	0.5132	0.4665
Annual Present Value	$370,717,349				-$66,641,893	$77,459,285	-$21,021,192	$64,015,938	-$17,372,886	$52,905,734	-$14,357,757	$43,723,747
Total Capital Investment + Interest		$28,383,714	$196,289,972	$147,919,113								
Net Present Worth				$0								

Year	9	10	11	12	13	14	15	16	17	18	19
Fixed Capital Investment											
Land											
Working Capital											
Loan Payment	$58,524,716	$58,524,716	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$8,349,205	$4,335,164	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$54,189,552	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191
Co-Product Sales	$89,553,168	$89,553,168	$89,553,168	$89,553,168	$89,553,168	$89,553,168	$89,553,168	$89,553,168	$89,553,168	$89,553,168	$89,553,168
Total Annual Sales	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358
Annual Manufacturing Cost											
Feedstocks	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672
Process Inputs	$623,988,072	$623,988,072	$623,988,072	$623,988,072	$623,988,072	$623,988,072	$623,988,072	$623,988,072	$623,988,072	$623,988,072	$623,988,072
Waste Streams	$21,806,688	$21,806,688	$21,806,688	$21,806,688	$21,806,688	$21,806,688	$21,806,688	$21,806,688	$21,806,688	$21,806,688	$21,806,688
Catalysts	$121,704,941	$0	$121,704,941	$0	$121,704,941	$0	$121,704,941	$0	$121,704,941	$0	$121,704,941
Royalties	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420
Fixed Operating Costs	$29,057,054	$29,057,054	$29,057,054	$29,057,054	$29,057,054	$29,057,054	$29,057,054	$29,057,054	$29,057,054	$29,057,054	$29,057,054
Total Product Cost	$6,411,956,848	$6,290,251,907	$6,411,956,848	$6,290,251,907	$6,411,956,848	$6,290,251,907	$6,411,956,848	$6,290,251,907	$6,411,956,848	$6,290,251,907	$6,411,956,848
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%
Depreciation Charge	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue	$22,196,305	$147,915,287	$30,545,510	$152,250,451	$30,545,510	$152,250,451	$30,545,510	$152,250,451	$30,545,510	$152,250,451	$30,545,510
Losses Forward	($148,509,564)	-$126,313,259	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	-$126,313,259	$21,602,028	$30,545,510	$152,250,451	$30,545,510	$152,250,451	$30,545,510	$152,250,451	$30,545,510	$152,250,451	$30,545,510
Income Tax	$0	$4,536,426	$6,414,557	$31,972,595	$6,414,557	$31,972,595	$6,414,557	$31,972,595	$6,414,557	$31,972,595	$6,414,557
Annual Cash Income	-$27,979,206	$89,189,309	$24,130,953	$120,277,856	$24,130,953	$120,277,856	$24,130,953	$120,277,856	$24,130,953	$120,277,856	$24,130,953
Discount Factor	0.4241	0.3855	0.3505	0.3186	0.2897	0.2633	0.2394	0.2176	0.1978	0.1799	0.1635
Annual Present Value	-$11,865,915	$34,386,340	$8,457,752	$38,324,232	$6,989,877	$31,672,919	$5,776,758	$26,175,966	$4,774,180	$21,633,030	$3,945,604
Total Capital Investment + Interest											
Net Present Worth											

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Year	20	21	22	23	24	25	26	27	28	29	30
Fixed Capital Investment											
Land											$0
Working Capital											($32,725,467)
Loan Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191	$6,352,949,191
Co-Product Sales	$89,553,168	$89,553,168	$89,553,168	$89,553,168	$89,553,168	$89,553,168	$89,553,168	$89,553,168	$89,553,168	$89,553,168	$89,553,168
Total Annual Sales	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358	$6,442,502,358
Annual Manufacturing Cost											
Feedstocks	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672	$5,607,374,672
Process Inputs	$623,988,072	$623,988,072	$623,988,072	$623,988,072	$623,988,072	$623,988,072	$623,988,072	$623,988,072	$623,988,072	$623,988,072	$623,988,072
Waste Streams	$21,806,688	$21,806,688	$21,806,688	$21,806,688	$21,806,688	$21,806,688	$21,806,688	$21,806,688	$21,806,688	$21,806,688	$21,806,688
Catalysts	$0	$121,704,941	$0	$121,704,941	$0	$121,704,941	$0	$121,704,941	$0	$121,704,941	$0
Royalties	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420
Fixed Operating Costs	$29,057,054	$29,057,054	$29,057,054	$29,057,054	$29,057,054	$29,057,054	$29,057,054	$29,057,054	$29,057,054	$29,057,054	$29,057,054
Total Product Cost	$6,290,251,907	$6,411,956,848	$6,290,251,907	$6,411,956,848	$6,290,251,907	$6,411,956,848	$6,290,251,907	$6,411,956,848	$6,290,251,907	$6,411,956,848	$6,290,251,907
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	2.23%									
Depreciation Charge	$0	$0									
Remaining Value	$0	$0									
Net Revenue	$152,250,451	$30,545,510	$152,250,451	$30,545,510	$152,250,451	$30,545,510	$152,250,451	$30,545,510	$152,250,451	$30,545,510	$152,250,451
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$152,250,451	$30,545,510	$152,250,451	$30,545,510	$152,250,451	$30,545,510	$152,250,451	$30,545,510	$152,250,451	$30,545,510	$152,250,451
Income Tax	$31,972,595	$6,414,557	$31,972,595	$6,414,557	$31,972,595	$6,414,557	$31,972,595	$6,414,557	$31,972,595	$6,414,557	$31,972,595
Annual Cash Income	$120,277,856	$24,130,953	$120,277,856	$24,130,953	$120,277,856	$24,130,953	$120,277,856	$24,130,953	$120,277,856	$24,130,953	$120,277,856
Discount Factor	0.1486	0.1351	0.1228	0.1117	0.1015	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
Annual Present Value	$17,878,537	$3,260,829	$14,775,650	$2,694,900	$12,211,281	$2,227,190	$10,091,968	$1,840,653	$8,340,469	$1,521,201	$6,892,950
Total Capital Investment + Interest											($1,875,449)
Net Present Worth											

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Table I-4. Equipment Costs and Fixed Capital Investment for 75k BPD CFP-A (21.6% O) HT+HC Configuration, Without Liquid Recycle (42199011)

Equip ID	Equipment Description	Process Variable Value	Size Ratio	Scaled Cost (Base Year)	Scaled Cost (2020$)	Pressure (psig)	Pres. Adj. Scaled Cost (2020)	Lang/TCI Factor (Low)	Lang/TCI Factor (High)	Average Lang/TCI Factor	FCI (2020$) for Single Units of Each	FCI (2020$)
C450	Amine Scrubber	903	3.75	$ 323,250	$ 241,809		241,809	8.303727	12.69922	10.50147	$ 2,539,346	$ 2,539,346
C470	Product Recovery Column 1	917314	0.97	$ 5,067,271	$ 3,790,599		3,790,599	2.999409	3.680618	3.340014	$ 12,660,651	$ 12,660,651
C480	Product Recovery Column 2	668409	0.98	$ 3,662,400	$ 2,739,678		2,739,678	3.382274	4.259656	3.820965	$ 10,468,212	$ 10,468,212
MX405	CFP and H2 Static Mixer	75000	7.50	$ 15,195	$ 11,367		11,367	90	100	95	$ 1,079,850	$ 1,079,850
MX406	Rx Inlet Static Mixer	75000	10.00	$ 59,583	$ 44,572		44,572	90	100	95	$ 4,234,312	$ 4,234,312
P400A/B	Cat Pyrolysis Oil 2 Pumps	75000	7.50	$ 905,506	$ 677,368	2,085	1,847,193	5920034	8247073	7083553	$ 7,083,553	$ 14,167,107
P405A/B	Liquid Feed 2 Pumps	0	0.00	$ -	$ -	2,085	-	0	0	0	$ -	$ -
P455A/B	Liquid Recycle 2 Pumps	0	0.00	$ -	$ -	2,035	-	0	0	0	$ -	$ -
P471A/B	Prod Recov Reflux 2 Pumps - 1	917314	0.97	$ 60,628	$ 45,353	25	45,454	258685.8	425745.8	342215.8	$ 342,216	$ 684,432
P472A/B	Prod Recov Btms 2 Pumps - 1	917314	0.97	$ 64,547	$ 48,285	30	46,570	264039.5	434083.3	349061.4	$ 349,061	$ 698,123
P481A/B	Prod Recov Reflux 2 Pumps - 2	668409	0.98	$ 55,344	$ 41,400	20	42,370	243773.5	402473.8	323123.6	$ 323,124	$ 646,247
P482A/B	Prod Recov Btms 2 Pumps - 2	668409	0.98	$ 42,653	$ 31,907	15	31,032	187369.9	313716.4	250543.1	$ 250,543	$ 501,086
PIPE412	H2 MU Comp Jumpover Line	60237	4.95	$ 19,596	$ 14,659		14,659	6.5	7	6.75	$ 98,945	$ 98,945
PIPE442	H2 Rec Comp Jumpover Line	433932	5.73	$ 39,580	$ 29,608		29,608	6.5	7	6.75	$ 199,853	$ 199,853
PIPE500	CFP Pipe from Storage	75000	7.50	$ 198,322	$ 148,356		148,356	6.5	7	6.75	$ 1,001,404	$ 1,001,404
PSA460	Pressure Swing Adsorption	156177070	1.87	$ 8,255,982	$ 6,175,930		6,175,930	3.6	4.4	4	$ 24,703,721	$ 24,703,721
QAH413	Rx Effluent Air Cooler	704334705	5.56	$ 9,904,283	$ 7,408,951		7,408,951	4.75	5.75	5.25	$ 38,896,991	$ 38,896,991
QCH413	Rx Effluent Water Cooler	141224642	3.60	$ 3,225,670	$ 2,412,979	1,985	8,977,294	36347254	40390568	38368911	$ 38,368,911	$ 38,368,911
QCH444	Water Cooler H2 Quench 1	16382234	3.46	$ 535,952	$ 400,922	2,085	1,728,328	9063354	11490497	10276926	$ 10,276,926	$ 10,276,926
QCH445	Water Cooler H2 Quench 2	16382234	3.46	$ 535,952	$ 400,922	2,085	1,728,328	9063354	11490497	10276926	$ 10,276,926	$ 10,276,926
QFH411-L	Reactor Preheat Furnace - Liquid Feed	96355830	2.27	$ 2,106,823	$ 1,576,020	2,085	1,576,020	3.70	5.72	4.71	$ 9,014,351	$ 9,014,351
QFH411-G	Reactor Preheat Furnace - Gas Feed	0	0.00	$ -	$ -	2,085	-	0	0	0	$ -	$ -
QH412	Process Rx Eff Cooler	770236118	5.62	$ 20,576,246	$ 15,392,168	1,995	56,436,596	1.71E+08	1.64E+08	1.68E+08	$ 167,724,789	$ 167,724,789
QH471	Prod Recov Condenser 1	917314	0.97	$ 4,223,468	$ 3,159,387	25	3,159,387	15070761	18206521	16638641	$ 16,638,641	$ 16,638,641
QH472	Prod Recov Reboiler 1	917314	0.97	$ 3,841,832	$ 2,873,903	30	2,873,903	13914318	16937299	15425809	$ 15,425,809	$ 15,425,809
QH481	Prod Recov Condenser 2	668409	0.98	$ 2,158,713	$ 1,614,837	20	1,614,837	8558988	10910185	9734587	$ 9,734,587	$ 9,734,587
QH482	Prod Recov Reboiler 2	668409	0.98	$ 1,644,395	$ 1,230,098	15	1,230,098	6804392	8864497	7834445	$ 7,834,445	$ 7,834,445
QXH410	Feed Preheater to Furnace	980080043	2.67	$ 3,001,424	$ 2,245,231	2,085	8,679,878	35329459	39365511	37347485	$ 37,347,485	$ 37,347,485
RX405	CFP Cat Stabilization Rx	75000	7.50	$ 28,245	$ 21,128	2,085	24,784	35.5	45.5	40.5	$ 1,003,747	$ 1,003,747
RX410-HT	Reactor System - Hydrotreater	3595783	2.82	$ 28,944,129	$ 21,651,807	2,030	26,254,037	3.5	5.3	4.4	$ 115,517,761	$ 115,517,761
RX410-HC	Reactor System - Hydrocracker	3595783	2.82	$ 28,944,129	$ 21,651,807	2,030	26,254,037	3.5	5.3	4.4	$ 115,517,761	$ 115,517,761
TK400	Cat Pyrolysis Oil Storage	75000	7.50	$ 14,423,781	$ 10,789,784		10,789,784	6	7.2	6.6	$ 71,212,576	$ 71,212,576
V451	High Pressure Flash Drum	4021883	3.15	$ 2,429,072	$ 1,817,080	1,980	5,832,711	3.666795	5.272141	4.469468	$ 26,069,114	$ 26,069,114
V452	Low Pressure Flash Drum	948771	0.96	$ 213,812	$ 159,943	2,085	159,943	13.38358	25.65812	19.52085	$ 4,103,846	$ 4,103,846
V471	Prod Recov Ovhd Drum 1	1254756	1.33	$ 60,556	$ 45,299	2,085	45,299	21.0769	44.69814	32.88752	$ 2,024,798	$ 2,024,798
V481	Prod Recov Ovhd Drum 2	857932	1.26	$ 73,505	$ 54,986	2,085	54,986	19.65666	41.04495	30.35081	$ 2,256,890	$ 2,256,890
WK412	H2 Make-up Compressor	91907	6.89	$ 53,105,695	$ 39,725,992		39,725,992	3	3.8	3.4	$ 135,068,371	$ 135,068,371
WK442	H2 Recycle Compressor	14661	3.69	$ 4,540,934	$ 3,396,870		3,396,870	3	3.8	3.4	$ 11,549,357	$ 11,549,357
QH470	Dist Col 1 Liquid Feed Preheat Furnace	63138990	1.49	$ 1,696,817	$ 1,269,313	36	1,269,313	3.999962	6.291144	5.145553	$ 5,077,203	$ 5,077,203
Total												**$ 924,624,570**

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Table I-5. Operating Costs for 75k BPD CFP-A (21.6% O) HT+HC Configuration, Without Liquid Recycle (42199011)

	Quoted Price	Unit	Cost Year	2020$ price	Scaling Flow	Unit	2020$ Cost ($/h)	2020$ Cost (MM$/y)	2020$ Cost (cents/GGE)
Feedstocks									
Crude Oil Basis	$ 85.00	/bbl	2020	$ 85.00					
Straight Run Diesel (SRD)	$ 105.88	/bbl	2020	$ 105.88	0	bbl/day	0	0.00	0.0
Light Cycle Oil (LCO)	$ 102.95	/bbl	2020	$ 102.95	0	bbl/day	0	0.00	0.0
Vacuuum Gas Oil (VGO)	$ 100.78	/bbl	2020	$ 100.78	0	bbl/day	0	0.00	0.0
Catalytic Fast Pyrolysis Oil (21.6%)	$ 215.62	/bbl	2020	$ 215.62	75,000	bbl/day	673,801	5,607.37	473.4
Bio-Oil Transportation 800 miles	$ -	/ton-mile	2007	$ -	1,260,728	lb/h	0	0.00	0.0
Subtotal							**673,801**	**5,607.37**	**473.4**
Process Inputs									
Hydrogen (pure H2 basis)									
Total Hydrogen Feed					120,217	lb/h			
Hydrogen Recovered via PSA					22,613	lb/h			
Net Hydrogen Makeup	$ 0.680	/lb	2020	$ 0.680	97,604	lb/h	66,409	552.65	46.7
Electricity for Hydroprocessing	$ 0.082	/kWh	2020	$ 0.082	96,694	kWh	7,900	65.74	5.5
Natural Gas	$ 3.231	/MMBtu	2020	$ 3.231	0	MMBtu/h	0	0.00	0.0
Fuel Gas	$ 2.746	/MMBtu	2020	$ 2.746	562	MMBtu/h	1,544	12.85	1.1
Amine Consumption	$ 0.00122600	/lbmol inlet gas	2020	$ 0.00122600	440,152	lbmol/h	540	4.49	0.379
Water Makeup	$ 0.200	/ton	2004	$ 0.334	213	ton/h	71	0.59	0.0
BFW to Steam	$ 0.001	/lb	2016	$ 0.001		lb/h			
Cooling Water Circulation	$ 0.00768	/ton	2020	$ 0.008	4,422	ton/h	34	0.28	0.0
Boiler Chemicals	$ 75.00	/MM lb bldwn	2014	$ 80.516		MMlb/h			
Cooling Tower Chemicals	$ 33.84	/ton-cool/y	2014	$ 36.329		ton-cool/y			
Diesel Fuel	$ 2.57	/gallon	2020	$ 2.571	10	gallon/h	26	0.21	0.0
Subtotal							**76,523**	**636.82**	**53.8**
Waste Streams									
Wastewater Treatment	$ 0.003	/lb	2011	$ 0.003	727,370	lb/h	2,201	18.32	1.5
Acid Gas Conversion and Recovery	$ 0.477	/lbmol acid gas	2020	$ 0.477	903	lbmol/h	431	3.58	0.3
Subtotal							**2,632**	**21.90**	**1.8**
Coproducts									
Fuel Gas	$ (2.746)	/MMBtu	2020	$ (2.746)	2,535	MMBtu/h	-6,961	-57.93	-4.9
HP Steam (900 psig)	$ (0.0045)	/lb	2023	$ (0.004)	0	lb/h	0	0.00	0.0
MP Steam (350 psig)	$ (0.0045)	/lb	2023	$ (0.004)	0	lb/h	0	0.00	0.0
LP Steam (150 psig)	$ (0.0045)	/lb	2023	$ (0.004)	428,052	lb/h	-1,565	-13.02	-1.1
VLP Steam (15 psig)	$ (0.0044)	/lb	2023	$ (0.004)	687,740	lb/h	-2,445	-20.35	-1.7
Subtotal							**-10,971**	**-91.30**	**-7.7**
Catalysts and Royalties									
Royalties	$ 0.20	/Bbl Feed	2005	$ 0.309	3,125	Bbl/h	964	8.03	0.7
Subtotal							**964**	**8.03**	**0.7**
Total Variable Operating Costs							**742,949**	**6,182.82**	**521.9**
Fixed Operating Costs									
Total Salaries								2.55	0.2
Overhead and Benefits	% of Labor & Supervison =		0.9					2.29	0.2
Maintenance	% of FCI =		0.03					27.7	2.3
Insurance & Taxes	% of FCI =		0.007					6.5	0.5
Total Fixed Operating Costs								**39.05**	**3.3**
Total Operating Costs								**6,221.88**	**525.2**

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Table I-6. DCFROR for 75k BPD CFP-A (21.6% O) HT+HC Configuration, Without Liquid Recycle (42199011)

Year		-2	-1	0	1	2	3	4	5	6	7	8
Fixed Capital Investment		$29,587,986	$221,909,897	$118,351,945								
Land		$0										
Working Capital				$46,231,229								
Loan Payment					$82,677,796	$82,677,796	$82,677,796	$82,677,796	$82,677,796	$82,677,796	$82,677,796	$82,677,796
Loan Interest Payment		$3,550,558	$30,179,746	$44,381,979	$44,381,979	$41,318,314	$38,009,555	$34,436,096	$30,576,760	$26,408,677	$21,907,148	$17,045,496
Loan Principal		$44,381,979	$377,246,825	$554,774,742	$516,478,925	$475,119,443	$430,451,202	$382,209,502	$330,108,467	$273,839,348	$213,068,699	$147,436,399
Fuel Sales					$4,799,842,383	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844
Co-Product Sales					$68,476,604	$91,302,138	$91,302,138	$91,302,138	$91,302,138	$91,302,138	$91,302,138	$91,302,138
Total Annual Sales					$4,868,318,987	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983
Annual Manufacturing Cost												
Feedstocks					$4,205,530,926	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568
Process Inputs					$477,617,814	$636,823,752	$636,823,752	$636,823,752	$636,823,752	$636,823,752	$636,823,752	$636,823,752
Waste Streams					$16,426,565	$21,902,087	$21,902,087	$21,902,087	$21,902,087	$21,902,087	$21,902,087	$21,902,087
Catalysts					$121,406,054	$0	$60,703,027	$60,703,027	$60,703,027	$0	$121,406,054	$0
Royalties					$6,019,065	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420
Fixed Operating Costs					$39,051,318	$39,051,318	$39,051,318	$39,051,318	$39,051,318	$39,051,318	$39,051,318	$39,051,318
Total Product Cost					$4,866,051,743	$6,313,177,145	$6,373,880,172	$6,373,880,172	$6,373,880,172	$6,313,177,145	$6,434,583,200	$6,313,177,145
Annual Depreciation												
General Plant Writedown					14%	24.49%	17.49%	12.49%	8.93%	8.92%	8.93%	4.46%
Depreciation Charge					$132,128,851	$226,440,557	$161,716,837	$115,485,609	$82,568,974	$82,476,512	$82,568,974	$41,238,256
Remaining Value					$792,495,719	$566,055,162	$404,338,324	$288,852,716	$206,283,742	$123,807,230	$41,238,256	$0
Steam Plant Writedown					3.75%	7.22%	6.68%	6.18%	5.71%	5.29%	4.89%	4.52%
Depreciation Charge					$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value					$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue					($174,243,586)	($89,844,034)	($82,514,583)	($32,709,895)	$4,066,076	$69,029,648	-$47,967,339	$119,631,086
Losses Forward						($174,243,586)	($264,087,620)	($346,602,203)	($379,312,098)	($375,246,022)	($306,216,373)	($354,183,712)
Taxable Income					($174,243,586)	($264,087,620)	($346,602,203)	($379,312,098)	($375,246,022)	($306,216,373)	($354,183,712)	($234,552,627)
Income Tax					$0	$0	$0	$0	$0	$0	$0	$0
Annual Cash Income					-$80,410,552	$95,237,041	$34,534,014	$34,534,014	$34,534,014	$95,237,041	-$26,169,013	$95,237,041
Discount Factor		1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830	0.6209	0.5645	0.5132	0.4665
Annual Present Value	$523,711,954				-$73,100,502	$78,708,299	$25,945,916	$23,587,196	$21,442,906	$53,758,827	-$13,428,842	$44,428,783
Total Capital Investment + Interest		$40,097,639	$277,298,607	$208,965,153								
Net Present Worth				$0								

Year	9	10	11	12	13	14	15	16	17	18	19
Fixed Capital Investment											
Land											
Working Capital											
Loan Payment	$82,677,796	$82,677,796	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$11,794,912	$6,124,281	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$76,553,515	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844
Co-Product Sales	$91,302,138	$91,302,138	$91,302,138	$91,302,138	$91,302,138	$91,302,138	$91,302,138	$91,302,138	$91,302,138	$91,302,138	$91,302,138
Total Annual Sales	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983
Annual Manufacturing Cost											
Feedstocks	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568
Process Inputs	$636,823,752	$636,823,752	$636,823,752	$636,823,752	$636,823,752	$636,823,752	$636,823,752	$636,823,752	$636,823,752	$636,823,752	$636,823,752
Waste Streams	$21,902,087	$21,902,087	$21,902,087	$21,902,087	$21,902,087	$21,902,087	$21,902,087	$21,902,087	$21,902,087	$21,902,087	$21,902,087
Catalysts	$60,703,027	$60,703,027	$60,703,027	$0	$121,406,054	$0	$60,703,027	$60,703,027	$60,703,027	$0	$121,406,054
Royalties	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420
Fixed Operating Costs	$39,051,318	$39,051,318	$39,051,318	$39,051,318	$39,051,318	$39,051,318	$39,051,318	$39,051,318	$39,051,318	$39,051,318	$39,051,318
Total Product Cost	$6,373,880,172	$6,373,880,172	$6,373,880,172	$6,313,177,145	$6,434,583,200	$6,313,177,145	$6,373,880,172	$6,373,880,172	$6,373,880,172	$6,313,177,145	$6,434,583,200
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%
Depreciation Charge	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue	$105,416,898	$111,087,529	$117,211,810	$177,914,837	$56,508,783	$177,914,837	$117,211,810	$117,211,810	$117,211,810	$177,914,837	$56,508,783
Losses Forward	($234,552,627)	-$129,135,728	-$18,048,199	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	-$129,135,728	-$18,048,199	$99,163,611	$177,914,837	$56,508,783	$177,914,837	$117,211,810	$117,211,810	$117,211,810	$177,914,837	$56,508,783
Income Tax	$0	$0	$20,824,358	$37,362,116	$11,866,844	$37,362,116	$24,614,480	$24,614,480	$24,614,480	$37,362,116	$11,866,844
Annual Cash Income	$34,534,014	$34,534,014	$96,387,452	$140,552,722	$44,641,939	$140,552,722	$92,597,330	$92,597,330	$92,597,330	$140,552,722	$44,641,939
Discount Factor	0.4241	0.3855	0.3505	0.3186	0.2897	0.2633	0.2394	0.2176	0.1978	0.1799	0.1635
Annual Present Value	$14,645,793	$13,314,357	$33,783,214	$44,784,429	$12,931,179	$37,011,924	$22,167,065	$20,151,877	$18,319,888	$25,279,642	$7,299,314
Total Capital Investment + Interest											
Net Present Worth											

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Year	20	21	22	23	24	25	26	27	28	29	30
Fixed Capital Investment											
Land											$0
Working Capital											($46,231,229)
Loan Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844	$6,399,789,844
Co-Product Sales	$91,302,138	$91,302,138	$91,302,138	$91,302,138	$91,302,138	$91,302,138	$91,302,138	$91,302,138	$91,302,138	$91,302,138	$91,302,138
Total Annual Sales	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983	$6,491,091,983
Annual Manufacturing Cost											
Feedstocks	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568	$5,607,374,568
Process Inputs	$636,823,752	$636,823,752	$636,823,752	$636,823,752	$636,823,752	$636,823,752	$636,823,752	$636,823,752	$636,823,752	$636,823,752	$636,823,752
Waste Streams	$21,902,087	$21,902,087	$21,902,087	$21,902,087	$21,902,087	$21,902,087	$21,902,087	$21,902,087	$21,902,087	$21,902,087	$21,902,087
Catalysts	$0	$60,703,027	$60,703,027	$60,703,027	$0	$121,406,054	$0	$60,703,027	$60,703,027	$60,703,027	$0
Royalties	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420
Fixed Operating Costs	$39,051,318	$39,051,318	$39,051,318	$39,051,318	$39,051,318	$39,051,318	$39,051,318	$39,051,318	$39,051,318	$39,051,318	$39,051,318
Total Product Cost	$6,313,177,145	$6,373,880,172	$6,373,880,172	$6,373,880,172	$6,313,177,145	$6,434,583,200	$6,313,177,145	$6,373,880,172	$6,373,880,172	$6,373,880,172	$6,313,177,145
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	2.23%									
Depreciation Charge	$0	$0									
Remaining Value	$0	$0									
Net Revenue	$177,914,837	$117,211,810	$117,211,810	$117,211,810	$177,914,837	$56,508,783	$177,914,837	$117,211,810	$117,211,810	$117,211,810	$177,914,837
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$177,914,837	$117,211,810	$117,211,810	$117,211,810	$177,914,837	$56,508,783	$177,914,837	$117,211,810	$117,211,810	$117,211,810	$177,914,837
Income Tax	$37,362,116	$24,614,480	$24,614,480	$24,614,480	$37,362,116	$11,866,844	$37,362,116	$24,614,480	$24,614,480	$24,614,480	$37,362,116
Annual Cash Income	$140,552,722	$92,597,330	$92,597,330	$92,597,330	$140,552,722	$44,641,939	$140,552,722	$92,597,330	$92,597,330	$92,597,330	$140,552,722
Discount Factor	0.1486	0.1351	0.1228	0.1117	0.1015	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
Annual Present Value	$20,892,266	$12,512,730	$11,375,209	$10,341,099	$14,269,699	$4,120,272	$11,793,140	$7,063,110	$6,421,009	$5,837,281	$8,054,873
Total Capital Investment + Interest											($2,649,445)
Net Present Worth											

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Table I-7. Equipment Costs and FCI for 10k BPD (Including Recycle) CFP-A (21.6% O) HT Configuration, With 50% Liquid Recycle (12199513)

Equip ID	Equipment Description	Process Variable Value	Size Ratio	Scaled Cost (Base Year)	Scaled Cost (2020$)	Pressure (psig)	Pres. Adj. Scaled Cost (2020)	Lang/TCI Factor (Low)	Lang/TCI Factor (High)	Average Lang/TCI Factor	FCI (2020$) for Single Units of Each	FCI (2020$)
C450	Amine Scrubber	31	0.13	$ 21,479	$ 16,068		16,068	22.64411	43.01904	32.83158	$ 527,526	$ 527,526
C470	Product Recovery Column 1	61151	0.06	$ 972,335	$ 727,360		727,360	5.524721	7.736634	6.630678	$ 4,822,889	$ 4,822,889
C480	Product Recovery Column 2	43635	0.06	$ 753,867	$ 563,934		563,934	6.070202	8.675341	7.372771	$ 4,157,757	$ 4,157,757
MX405	CFP and H2 Static Mixer	5000	0.50	$ 4,555	$ 3,407		3,407	90	100	95	$ 323,683	$ 323,683
MX406	Rx Inlet Static Mixer	5000	0.67	$ 12,787	$ 9,565		9,565	90	100	95	$ 908,678	$ 908,678
P400A/B	Cat Pyrolysis Oil 2 Pumps	5000	0.50	$ 550,162	$ 411,551	1,085	362,332	1494741	2240655	1867698	$ 1,867,698	$ 3,735,396
P405A/B	Liquid Feed 2 Pumps	0	0.00	$ -	$ -	1,085	-	0	0	0	$ -	$ -
P455A/B	Liquid Recycle 2 Pumps	5000	0.20	$ 388,779	$ 290,829	1,035	240,462	1057075	1614088	1335581	$ 1,335,581	$ 2,671,163
P471A/B	Prod Recov Reflux 2 Pumps - 1	61151	0.06	$ 27,103	$ 20,275	25	20,320	131015.7	223581.1	177298.4	$ 177,298	$ 354,597
P472A/B	Prod Recov Btms 2 Pumps - 1	61151	0.06	$ 27,215	$ 20,359	30	19,624	127212.8	217432.3	172322.5	$ 172,323	$ 344,645
P481A/B	Prod Recov Reflux 2 Pumps - 2	43635	0.06	$ 24,644	$ 18,435	20	18,873	123086.4	210749.1	166917.7	$ 166,918	$ 333,835
P482A/B	Prod Recov Btms 2 Pumps - 2	43635	0.06	$ 23,387	$ 17,495	15	17,011	112741.8	193941.7	153341.7	$ 153,342	$ 306,683
PIPE412	H2 MU Comp Jumpover Line	4155	0.34	$ 3,015	$ 2,255		2,255	6.5	7	6.75	$ 15,222	$ 15,222
PIPE442	H2 Rec Comp Jumpover Line	21872	0.29	$ 4,889	$ 3,657		3,657	6.5	7	6.75	$ 24,686	$ 24,686
PIPE500	CFP Pipe from Storage	5000	0.50	$ 29,793	$ 22,286		22,286	6.5	7	6.75	$ 150,434	$ 150,434
PSA460	Pressure Swing Adsorption	10978516	0.13	$ 2,218,183	$ 1,659,323		1,659,323	3.6	4.4	4	$ 6,637,292	$ 6,637,292
QAH413	Rx Effluent Air Cooler	29089550	0.23	$ 583,582	$ 436,551		436,551	4.75	5.75	5.25	$ 2,291,894	$ 2,291,894
QCH413	Rx Effluent Water Cooler	10372460	0.26	$ 349,866	$ 261,719	985	549,147	3447686	4790851	4119268	$ 4,119,268	$ 4,119,268
QCH444	Water Cooler H2 Quench 1	1501900	0.32	$ 72,517	$ 54,247	1,085	92,210	766083.2	1227874	996978.8	$ 996,979	$ 996,979
QCH445	Water Cooler H2 Quench 2	1501900	0.32	$ 72,517	$ 54,247	1,085	92,210	766083.2	1227874	996978.8	$ 996,979	$ 996,979
QFH411-L	Reactor Preheat Furnace - Liquid Feed	0	0.00	$ -	$ -	1,085	-	0.00	0.00	0.00	$ -	$ -
QFH411-G	Reactor Preheat Furnace - Gas Feed	5667262	0.48	$ 484,147	$ 362,169	1,085	362,169	13.2	15.8	14.5	$ 5,251,451	$ 5,251,451
QH412	Process Rx Eff Cooler	57059402	0.42	$ 1,582,507	$ 1,183,803	995	2,744,114	13382680	16350495	14866587	$ 14,866,587	$ 14,866,587
QH471	Prod Recov Condenser 1	61151	0.06	$ 454,703	$ 340,143	25	340,143	2302295	3324207	2813251	$ 2,813,251	$ 2,813,251
QH472	Prod Recov Reboiler 1	61151	0.06	$ 419,255	$ 313,626	30	313,626	2150032	3124583	2637308	$ 2,637,308	$ 2,637,308
QH481	Prod Recov Condenser 2	43635	0.06	$ 261,195	$ 195,388	20	195,388	1442769	2177635	1810202	$ 1,810,202	$ 1,810,202
QH482	Prod Recov Reboiler 2	43635	0.06	$ 197,773	$ 147,945	15	147,945	1141210	1761236	1451223	$ 1,451,223	$ 1,451,223
QXH410	Feed Preheater to Furnace	55565727	0.15	$ 236,981	$ 177,275	1,085	398,450	2630783	3750724	3190753	$ 3,190,753	$ 3,190,753
RX405	CFP Cat Stabilization Rx	5000	0.50	$ 15,994	$ 11,964	1,085	11,519	35.5	45.5	40.5	$ 466,534	$ 466,534
RX410-HT	Reactor System - Hydrotreater	262441	0.21	$ 2,679,973	$ 2,004,768	1,030	1,378,998	3.5	5.3	4.4	$ 6,067,590	$ 6,067,590
RX410-HC	Reactor System - Hydrocracker	262441	0.21	$ 2,679,973	$ 2,004,768	1,030	1,378,998	3.5	5.3	4.4	$ 6,067,590	$ -
TK400	Cat Pyrolysis Oil Storage	5000	0.50	$ 1,371,429	$ 1,025,905		1,025,905	6	7.2	6.6	$ 6,770,972	$ 6,770,972
V451	High Pressure Flash Drum	289484	0.23	$ 196,920	$ 147,307	980	736,112	7.72503	13.10734	10.41619	$ 7,667,481	$ 7,667,481
V452	Low Pressure Flash Drum	123755	0.13	$ 47,905	$ 35,836	1,085	35,836	22.9322	49.55304	36.24262	$ 821,795	$ 821,795
V471	Prod Recov Ovhd Drum 1	52595	0.06	$ 20,911	$ 15,643	1,085	15,643	30.90624	71.36265	51.13444	$ 483,462	$ 483,462
V481	Prod Recov Ovhd Drum 2	56356	0.08	$ 29,212	$ 21,852	1,085	21,852	27.40191	61.60147	44.50169	$ 598,795	$ 598,795
WK412	H2 Make-up Compressor	4160	0.31	$ 4,517,337	$ 3,379,217		3,379,217	3	3.8	3.4	$ 11,489,338	$ 11,489,338
WK442	H2 Recycle Compressor	1430	0.36	$ 2,238,915	$ 1,674,832		1,674,832	3	3.8	3.4	$ 5,694,428	$ 5,694,428
QH470	Dist Col 1 Liquid Feed Preheat Furnace	4352466	0.10	$ 431,435	$ 322,737	36	322,737	6.548698	11.49228	9.020492	$ 2,113,507	$ 2,113,507
Total												**$ 107,914,285**

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Table I-8. Operating Costs for 10k BPD (Including Recycle) CFP-A (21.6% O) HT Configuration, With 50% Liquid Recycle (12199513)

	Quoted Price	Unit	Cost Year	2020$ price	Scaling Flow	Unit	2020$ Cost ($/h)	2020$ Cost (MM$/y)	2020$ Cost (cents/GGE)
Feedstocks									
Crude Oil Basis	$ 85.00	/bbl	2020	$ 85.00					
Straight Run Diesel (SRD)	$ 105.88	/bbl	2020	$ 105.88	0	bbl/day	0	0.00	0.0
Light Cycle Oil (LCO)	$ 102.95	/bbl	2020	$ 102.95	0	bbl/day	0	0.00	0.0
Vacuuum Gas Oil (VGO)	$ 100.78	/bbl	2020	$ 100.78	0	bbl/day	0	0.00	0.0
Catalytic Fast Pyrolysis Oil (21.6%)	$ 215.62	/bbl	2020	$ 215.62	5,000	bbl/day	44,920	373.82	473.1
Bio-Oil Transportation 800 miles	$ -	/ton-mile	2007	$ -	84,049	lb/h	0	0.00	0.0
Subtotal							**44,920**	**373.82**	**473.1**
Process Inputs									
Hydrogen (pure H2 basis)									
Total Hydrogen Feed					8,292	lb/h			
Hydrogen Recovered via PSA					1,560	lb/h			
Net Hydrogen Makeup	$ 0.680	/lb	2020	$ 0.680	6,732	lb/h	4,580	38.12	48.2
Electricity for Hydroprocessing	$ 0.082	/kWh	2020	$ 0.082	5,187	kWh	424	3.53	4.5
Natural Gas	$ 3.231	/MMBtu	2020	$ 3.231	0	MMBtu/h	0	0.00	0.0
Fuel Gas	$ 2.746	/MMBtu	2020	$ 2.746	29	MMBtu/h	80	0.67	0.8
Amine Consumption	$ 0.00122600	/lbmol inlet gas	2020	$ 0.00122600	22,577	lbmol/h	28	0.23	0.292
Water Makeup	$ 0.200	/ton	2004	$ 0.334	14	ton/h	5	0.04	0.0
BFW to Steam	$ 0.001	/lb	2016	$ 0.001		lb/h			
Cooling Water Circulation	$ 0.00768	/ton	2020	$ 0.008	336	ton/h	3	0.02	0.0
Boiler Chemicals	$ 75.00	/MM lb bldwn	2014	$ 80.516		MMlb/h			
Cooling Tower Chemicals	$ 33.84	/ton-cool/y	2014	$ 36.329		ton-cool/y			
Diesel Fuel	$ 2.57	/gallon	2020	$ 2.571	10	gallon/h	26	0.21	0.3
Subtotal							**5,145**	**42.81**	**54.2**
Waste Streams									
Wastewater Treatment	$ 0.003	/lb	2011	$ 0.003	48,154	lb/h	146	1.21	1.5
Acid Gas Conversion and Recovery	$ 0.477	/lbmol acid gas	2020	$ 0.477	31	lbmol/h	15	0.12	0.2
Subtotal							**161**	**1.34**	**1.7**
Coproducts									
Fuel Gas	$ (2.746)	/MMBtu	2020	$ (2.746)	178	MMBtu/h	-489	-4.07	-5.2
HP Steam (900 psig)	$ (0.0045)	/lb	2023	$ (0.004)	0	lb/h	0	0.00	0.0
MP Steam (350 psig)	$ (0.0045)	/lb	2023	$ (0.004)	5,826	lb/h	-21	-0.18	-0.2
LP Steam (150 psig)	$ (0.0045)	/lb	2023	$ (0.004)	31,573	lb/h	-115	-0.96	-1.2
VLP Steam (15 psig)	$ (0.0044)	/lb	2023	$ (0.004)	40,599	lb/h	-144	-1.20	-1.5
Subtotal							**-770**	**-6.41**	**-8.1**
Catalysts and Royalties									
Royalties	$ 0.20	/Bbl Feed	2005	$ 0.309	208	Bbl/h	64	0.54	0.7
Subtotal							**64**	**0.54**	**0.7**
Total Variable Operating Costs							**49,519**	**412.10**	**521.6**
Fixed Operating Costs									
Total Salaries								2.55	3.2
Overhead and Benefits	% of Labor & Supervison =		0.9					2.29	2.9
Maintenance	% of FCI =		0.03					3.2	4.1
Insurance & Taxes	% of FCI =		0.007					0.8	1.0
Total Fixed Operating Costs								**8.83**	**11.2**
Total Operating Costs								**420.93**	**532.8**

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Table I-9. DCFROR for 10k BPD (Including Recycle) CFP-A (21.6% O) HT Configuration, With 50% Liquid Recycle (12199513)

Year		-2	-1	0	1	2	3	4	5	6	7	8
Fixed Capital Investment		$3,453,257	$25,899,428	$13,813,028								
Land		$0										
Working Capital				$5,395,714								
Loan Payment					$9,649,446	$9,649,446	$9,649,446	$9,649,446	$9,649,446	$9,649,446	$9,649,446	$9,649,446
Loan Interest Payment		$414,391	$3,522,322	$5,179,886	$5,179,886	$4,822,321	$4,436,151	$4,019,087	$3,568,658	$3,082,195	$2,556,815	$1,989,405
Loan Principal		$5,179,886	$44,029,028	$64,748,571	$60,279,010	$55,451,885	$50,238,589	$44,608,230	$38,527,442	$31,960,190	$24,867,559	$17,207,518
Fuel Sales					$329,877,446	$439,836,595	$439,836,595	$439,836,595	$439,836,595	$439,836,595	$439,836,595	$439,836,595
Co-Product Sales					$4,807,994	$6,410,658	$6,410,658	$6,410,658	$6,410,658	$6,410,658	$6,410,658	$6,410,658
Total Annual Sales					$334,685,440	$446,247,253	$446,247,253	$446,247,253	$446,247,253	$446,247,253	$446,247,253	$446,247,253
Annual Manufacturing Cost												
Feedstocks					$280,368,734	$373,824,978	$373,824,978	$373,824,978	$373,824,978	$373,824,978	$373,824,978	$373,824,978
Process Inputs					$32,110,354	$42,813,806	$42,813,806	$42,813,806	$42,813,806	$42,813,806	$42,813,806	$42,813,806
Waste Streams					$1,002,179	$1,336,239	$1,336,239	$1,336,239	$1,336,239	$1,336,239	$1,336,239	$1,336,239
Catalysts					$8,394,470	$0	$8,394,470	$0	$8,394,470	$0	$8,394,470	$0
Royalties					$401,271	$535,028	$535,028	$535,028	$535,028	$535,028	$535,028	$535,028
Fixed Operating Costs					$8,833,037	$8,833,037	$8,833,037	$8,833,037	$8,833,037	$8,833,037	$8,833,037	$8,833,037
Total Product Cost					$331,110,045	$427,343,088	$435,737,558	$427,343,088	$435,737,558	$427,343,088	$435,737,558	$427,343,088
Annual Depreciation												
General Plant Writedown					14%	24.49%	17.49%	12.49%	8.93%	8.92%	8.93%	4.46%
Depreciation Charge					$15,420,951	$26,428,208	$18,874,208	$13,478,494	$9,636,746	$9,625,954	$9,636,746	$4,812,977
Remaining Value					$92,493,333	$66,065,125	$47,190,917	$33,712,423	$24,075,677	$14,449,723	$4,812,977	$0
Steam Plant Writedown					3.75%	7.22%	6.68%	6.18%	5.71%	5.29%	4.89%	4.52%
Depreciation Charge					$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value					$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue					($17,025,443)	($12,346,365)	($12,800,664)	$1,406,583	($2,695,709)	$6,196,015	-$1,683,866	$12,101,783
Losses Forward						($17,025,443)	($29,371,807)	($42,172,472)	($40,765,888)	($43,461,598)	($37,265,583)	($38,949,449)
Taxable Income					($17,025,443)	($29,371,807)	($42,172,472)	($40,765,888)	($43,461,598)	($37,265,583)	($38,949,449)	($26,847,666)
Income Tax					$0	$0	$0	$0	$0	$0	$0	$0
Annual Cash Income					-$6,074,052	$9,254,718	$860,248	$9,254,718	$860,248	$9,254,718	$860,248	$9,254,718
Discount Factor		1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830	0.6209	0.5645	0.5132	0.4665
Annual Present Value	$61,123,187				-$5,521,866	$7,648,527	$646,317	$6,321,097	$534,147	$5,224,047	$441,443	$4,317,394
Total Capital Investment + Interest		$4,679,854	$32,363,926	$24,388,628								
Net Present Worth				$0								

Year	9	10	11	12	13	14	15	16	17	18	19
Fixed Capital Investment											
Land											
Working Capital											
Loan Payment	$9,649,446	$9,649,446	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$1,376,601	$714,774	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$8,934,673	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$439,836,595	$439,836,595	$439,836,595	$439,836,595	$439,836,595	$439,836,595	$439,836,595	$439,836,595	$439,836,595	$439,836,595	$439,836,595
Co-Product Sales	$6,410,658	$6,410,658	$6,410,658	$6,410,658	$6,410,658	$6,410,658	$6,410,658	$6,410,658	$6,410,658	$6,410,658	$6,410,658
Total Annual Sales	$446,247,253	$446,247,253	$446,247,253	$446,247,253	$446,247,253	$446,247,253	$446,247,253	$446,247,253	$446,247,253	$446,247,253	$446,247,253
Annual Manufacturing Cost											
Feedstocks	$373,824,978	$373,824,978	$373,824,978	$373,824,978	$373,824,978	$373,824,978	$373,824,978	$373,824,978	$373,824,978	$373,824,978	$373,824,978
Process Inputs	$42,813,806	$42,813,806	$42,813,806	$42,813,806	$42,813,806	$42,813,806	$42,813,806	$42,813,806	$42,813,806	$42,813,806	$42,813,806
Waste Streams	$1,336,239	$1,336,239	$1,336,239	$1,336,239	$1,336,239	$1,336,239	$1,336,239	$1,336,239	$1,336,239	$1,336,239	$1,336,239
Catalysts	$8,394,470	$0	$8,394,470	$0	$8,394,470	$0	$8,394,470	$0	$8,394,470	$0	$8,394,470
Royalties	$535,028	$535,028	$535,028	$535,028	$535,028	$535,028	$535,028	$535,028	$535,028	$535,028	$535,028
Fixed Operating Costs	$8,833,037	$8,833,037	$8,833,037	$8,833,037	$8,833,037	$8,833,037	$8,833,037	$8,833,037	$8,833,037	$8,833,037	$8,833,037
Total Product Cost	$435,737,558	$427,343,088	$435,737,558	$427,343,088	$435,737,558	$427,343,088	$435,737,558	$427,343,088	$435,737,558	$427,343,088	$435,737,558
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%
Depreciation Charge	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue	$9,133,093	$18,189,391	$10,509,695	$18,904,165	$10,509,695	$18,904,165	$10,509,695	$18,904,165	$10,509,695	$18,904,165	$10,509,695
Losses Forward	($26,847,666)	-$17,714,573	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	-$17,714,573	$474,818	$10,509,695	$18,904,165	$10,509,695	$18,904,165	$10,509,695	$18,904,165	$10,509,695	$18,904,165	$10,509,695
Income Tax	$0	$99,712	$2,207,036	$3,969,875	$2,207,036	$3,969,875	$2,207,036	$3,969,875	$2,207,036	$3,969,875	$2,207,036
Annual Cash Income	$860,248	$9,155,006	$8,302,659	$14,934,290	$8,302,659	$14,934,290	$8,302,659	$14,934,290	$8,302,659	$14,934,290	$8,302,659
Discount Factor	0.4241	0.3855	0.3505	0.3186	0.2897	0.2633	0.2394	0.2176	0.1978	0.1799	0.1635
Annual Present Value	$364,829	$3,529,651	$2,910,031	$4,758,525	$2,404,985	$3,932,665	$1,987,591	$3,250,137	$1,642,637	$2,686,063	$1,357,551
Total Capital Investment + Interest											
Net Present Worth											

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Year	20	21	22	23	24	25	26	27	28	29	30
Fixed Capital Investment											
Land											$0
Working Capital											($5,395,714)
Loan Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$439,836,595	$439,836,595	$439,836,595	$439,836,595	$439,836,595	$439,836,595	$439,836,595	$439,836,595	$439,836,595	$439,836,595	$439,836,595
Co-Product Sales	$6,410,658	$6,410,658	$6,410,658	$6,410,658	$6,410,658	$6,410,658	$6,410,658	$6,410,658	$6,410,658	$6,410,658	$6,410,658
Total Annual Sales	$446,247,253	$446,247,253	$446,247,253	$446,247,253	$446,247,253	$446,247,253	$446,247,253	$446,247,253	$446,247,253	$446,247,253	$446,247,253
Annual Manufacturing Cost											
Feedstocks	$373,824,978	$373,824,978	$373,824,978	$373,824,978	$373,824,978	$373,824,978	$373,824,978	$373,824,978	$373,824,978	$373,824,978	$373,824,978
Process Inputs	$42,813,806	$42,813,806	$42,813,806	$42,813,806	$42,813,806	$42,813,806	$42,813,806	$42,813,806	$42,813,806	$42,813,806	$42,813,806
Waste Streams	$1,336,239	$1,336,239	$1,336,239	$1,336,239	$1,336,239	$1,336,239	$1,336,239	$1,336,239	$1,336,239	$1,336,239	$1,336,239
Catalysts	$0	$8,394,470	$0	$8,394,470	$0	$8,394,470	$0	$8,394,470	$0	$8,394,470	$0
Royalties	$535,028	$535,028	$535,028	$535,028	$535,028	$535,028	$535,028	$535,028	$535,028	$535,028	$535,028
Fixed Operating Costs	$8,833,037	$8,833,037	$8,833,037	$8,833,037	$8,833,037	$8,833,037	$8,833,037	$8,833,037	$8,833,037	$8,833,037	$8,833,037
Total Product Cost	$427,343,088	$435,737,558	$427,343,088	$435,737,558	$427,343,088	$435,737,558	$427,343,088	$435,737,558	$427,343,088	$435,737,558	$427,343,088
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	2.23%									
Depreciation Charge	$0	$0									
Remaining Value	$0	$0									
Net Revenue	$18,904,165	$10,509,695	$18,904,165	$10,509,695	$18,904,165	$10,509,695	$18,904,165	$10,509,695	$18,904,165	$10,509,695	$18,904,165
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$18,904,165	$10,509,695	$18,904,165	$10,509,695	$18,904,165	$10,509,695	$18,904,165	$10,509,695	$18,904,165	$10,509,695	$18,904,165
Income Tax	$3,969,875	$2,207,036	$3,969,875	$2,207,036	$3,969,875	$2,207,036	$3,969,875	$2,207,036	$3,969,875	$2,207,036	$3,969,875
Annual Cash Income	$14,934,290	$8,302,659	$14,934,290	$8,302,659	$14,934,290	$8,302,659	$14,934,290	$8,302,659	$14,934,290	$8,302,659	$14,934,290
Discount Factor	0.1486	0.1351	0.1228	0.1117	0.1015	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
Annual Present Value	$2,219,887	$1,121,943	$1,834,617	$927,226	$1,516,213	$766,302	$1,253,068	$633,308	$1,035,594	$523,395	$855,863
Total Capital Investment + Interest											($309,221)
Net Present Worth											

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Table I-10. Capital Costs and FCI for 75k BPD (Including Recycle) 20% CFP-A With SRD, HT Configuration, 20% Liquid Recycle (12120211)

Equip ID	Equipment Description	Process Variable Value	Size Ratio	Scaled Cost (Base Year)	Scaled Cost (2020$)	Pressure (psig)	Pres. Adj. Scaled Cost (2020)	Lang/TCI Factor (Low)	Lang/TCI Factor (High)	Average Lang/TCI Factor	FCI (2020$) for Single Units of Each	FCI (2020$)
C450	Amine Scrubber	169	0.70	$ 83,875	$ 62,743		62,743	13.6791	23.30418	18.49164	$ 1,160,227	$ 1,160,227
C470	Product Recovery Column 1	731323	0.77	$ 4,413,512	$ 3,301,551		3,301,551	3.15669	3.916663	3.536677	$ 11,676,517	$ 11,676,517
C480	Product Recovery Column 2	545906	0.80	$ 3,257,167	$ 2,436,540		2,436,540	3.532249	4.490463	4.011356	$ 9,773,832	$ 9,773,832
MX405	CFP and H2 Static Mixer	12000	1.20	$ 6,724	$ 5,030		5,030	90	100	95	$ 477,834	$ 477,834
MX406	Rx Inlet Static Mixer	12000	1.60	$ 21,029	$ 15,731		15,731	90	100	95	$ 1,494,461	$ 1,494,461
P400A/B	Cat Pyrolysis Oil 2 Pumps	12000	1.20	$ 646,323	$ 483,485	1,085	426,333	1714981	2552051	2133516	$ 2,133,516	$ 4,267,032
P405A/B	Liquid Feed 2 Pumps	48000	0.64	$ 1,186,111	$ 887,276	1,085	789,341	2886091	4177359	3531725	$ 3,531,725	$ 7,063,450
P455A/B	Liquid Recycle 2 Pumps	15000	0.60	$ 738,161	$ 552,185	1,035	460,394	1830062	2713899	2271981	$ 2,271,981	$ 4,543,962
P471A/B	Prod Recov Reflux 2 Pumps - 1	731323	0.77	$ 56,679	$ 42,399	25	42,493	244372.3	403409.6	323890.9	$ 323,891	$ 647,782
P472A/B	Prod Recov Btms 2 Pumps - 1	731323	0.77	$ 60,047	$ 44,919	30	43,320	248385	409678.4	329031.7	$ 329,032	$ 658,063
P481A/B	Prod Recov Reflux 2 Pumps - 2	545906	0.80	$ 52,120	$ 38,989	20	39,903	231725.5	383616.4	307670.9	$ 307,671	$ 615,342
P482A/B	Prod Recov Btms 2 Pumps - 2	545906	0.80	$ 40,794	$ 30,516	15	29,678	180439.5	302719.7	241579.6	$ 241,580	$ 483,159
PIPE412	H2 MU Comp Jumpover Line	13090	1.07	$ 6,731	$ 5,035		5,035	6.5	7	6.75	$ 33,990	$ 33,990
PIPE442	H2 Rec Comp Jumpover Line	66460	0.88	$ 10,643	$ 7,962		7,962	6.5	7	6.75	$ 53,741	$ 53,741
PIPE500	CFP Pipe from Storage	12000	1.20	$ 54,987	$ 41,133		41,133	6.5	7	6.75	$ 277,647	$ 277,647
PSA460	Pressure Swing Adsorption	21423308	0.26	$ 3,088,282	$ 2,310,206		2,310,206	3.6	4.4	4	$ 9,240,823	$ 9,240,823
QAH413	Rx Effluent Air Cooler	41883676	0.33	$ 806,786	$ 603,520		603,520	4.75	5.75	5.25	$ 3,168,481	$ 3,168,481
QCH413	Rx Effluent Water Cooler	42736707	1.09	$ 1,166,846	$ 872,865	985	1,992,179	10216540	12806120	11511330	$ 11,511,330	$ 11,511,330
QCH444	Water Cooler H2 Quench 1	4593588	0.97	$ 184,868	$ 138,291	1,085	301,413	2079232	3031308	2555270	$ 2,555,270	$ 2,555,270
QCH445	Water Cooler H2 Quench 2	4593588	0.97	$ 184,868	$ 138,291	1,085	301,413	2079232	3031308	2555270	$ 2,555,270	$ 2,555,270
QFH411-L	Reactor Preheat Furnace - Liquid Feed	39414119	0.93	$ 1,333,080	$ 997,218	1,085	997,218	4.36	7.00	5.68	$ 4,350,770	$ 4,350,770
QFH411-G	Reactor Preheat Furnace - Gas Feed	0	0.00	$ -	$ -	1,085	-	0	0	0	$ -	$ -
QH412	Process Rx Eff Cooler	181290820	1.32	$ 4,944,984	$ 3,699,121	995	8,715,452	35451482	39488551	37470017	$ 37,470,017	$ 37,470,017
QH471	Prod Recov Condenser 1	731323	0.77	$ 3,504,927	$ 2,621,879	25	2,621,879	12878356	15791792	14335074	$ 14,335,074	$ 14,335,074
QH472	Prod Recov Reboiler 1	731323	0.77	$ 3,191,833	$ 2,387,668	30	2,387,668	11901507	14703613	13302560	$ 13,302,560	$ 13,302,560
QH481	Prod Recov Condenser 2	545906	0.80	$ 1,845,654	$ 1,380,651	20	1,380,651	7499991	9680862	8590426	$ 8,590,426	$ 8,590,426
QH482	Prod Recov Reboiler 2	545906	0.80	$ 1,405,297	$ 1,051,240	15	1,051,240	5960251	7863002	6911627	$ 6,911,627	$ 6,911,627
QXH410	Feed Preheater to Furnace	336061248	0.92	$ 1,164,467	$ 871,085	1,085	2,125,475	10789852	13454854	12122353	$ 12,122,353	$ 12,122,353
RX405	CFP Cat Stabilization Rx	12000	1.20	$ 19,222	$ 14,379	1,085	13,934	35.5	45.5	40.5	$ 564,334	$ 564,334
RX410-HT	Reactor System - Hydrotreater	1328698	1.04	$ 11,708,345	$ 8,758,488	1,030	8,132,718	3.5	5.3	4.4	$ 35,783,961	$ 35,783,961
RX410-HC	Reactor System - Hydrocracker	1328698	1.04	$ 11,708,345	$ 8,758,488	1,030	8,132,718	3.5	5.3	4.4	$ 35,783,961	$ -
TK400	Cat Pyrolysis Oil Storage	12000	1.20	$ 2,934,535	$ 2,195,194		2,195,194	6	7.2	6.6	$ 14,488,283	$ 14,488,283
V451	High Pressure Flash Drum	1557712	1.22	$ 982,015	$ 734,601	980	1,323,406	6.25447	10.1257	8.190084	$ 10,838,807	$ 10,838,807
V452	Low Pressure Flash Drum	927189	0.94	$ 210,229	$ 157,263	1,085	157,263	13.46524	25.84961	19.65743	$ 2,117,588	$ 2,117,588
V471	Prod Recov Ovhd Drum 1	623966	0.66	$ 47,914	$ 35,842	1,085	35,842	22.93076	49.54925	36.24001	$ 821,886	$ 821,886
V481	Prod Recov Ovhd Drum 2	603541	0.89	$ 65,246	$ 48,807	1,085	48,807	20.51849	43.25504	31.88676	$ 1,001,452	$ 1,001,452
WK412	H2 Make-up Compressor	13107	0.98	$ 11,263,878	$ 8,426,003		8,426,003	3	3.8	3.4	$ 28,648,410	$ 28,648,410
WK442	H2 Recycle Compressor	4347	1.09	$ 3,138,723	$ 2,347,938		2,347,938	3	3.8	3.4	$ 7,982,989	$ 7,982,989
QH470	Dist Col 1 Liquid Feed Preheat Furnace	33843314	0.80	$ 1,233,027	$ 922,373	36	922,373	4.487188	7.240041	5.863614	$ 4,138,859	$ 4,138,859
Total												**$ 275,727,608**

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Table I-11. Operating Costs for 75k BPD (Including Recycle) 20% CFP-A With SRD, HT Configuration, 20% Liquid Recycle (12120211)

	Quoted Price	Unit	Cost Year	2020$ price	Scaling Flow	Unit	2020$ Cost ($/h)	2020$ Cost (MM$/y)	2020$ Cost (cents/GGE)
Feedstocks									
Crude Oil Basis	$ 85.00	/bbl	2020	$ 85.00					
Straight Run Diesel (SRD)	$ 105.88	/bbl	2020	$ 105.88	48,000	bbl/day	211,758	1,762.25	186.6
Light Cycle Oil (LCO)	$ 102.95	/bbl	2020	$ 102.95	0	bbl/day	0	0.00	0.0
Vacuuum Gas Oil (VGO)	$ 100.78	/bbl	2020	$ 100.78	0	bbl/day	0	0.00	0.0
Catalytic Fast Pyrolysis Oil (21.6%)	$ 215.62	/bbl	2020	$ 215.62	12,000	bbl/day	107,808	897.18	95.0
Bio-Oil Transportation 800 miles	$ -	/ton-mile	2007	$ -	201,717	lb/h	0	0.00	0.0
Subtotal							**319,566**	**2,659.43**	**281.7**
Process Inputs									
Hydrogen (pure H2 basis)									
Total Hydrogen Feed					26,124	lb/h			
Hydrogen Recovered via PSA					3,037	lb/h			
Net Hydrogen Makeup	$ 0.680	/lb	2020	$ 0.680	23,087	lb/h	15,708	130.73	13.8
Electricity for Hydroprocessing	$ 0.082	/kWh	2020	$ 0.082	16,862	kWh	1,378	11.46	1.2
Natural Gas	$ 3.231	/MMBtu	2020	$ 3.231	0	MMBtu/h	0	0.00	0.0
Fuel Gas	$ 2.746	/MMBtu	2020	$ 2.746	314	MMBtu/h	862	7.17	0.8
Amine Consumption	$ 0.00122600	/lbmol inlet gas	2020	$ 0.00122600	77,968	lbmol/h	96	0.80	0.084
Water Makeup	$ 0.200	/ton	2004	$ 0.334	115	ton/h	38	0.32	0.0
BFW to Steam	$ 0.001	/lb	2016	$ 0.001		lb/h			
Cooling Water Circulation	$ 0.00768	/ton	2020	$ 0.008	1,446	ton/h	11	0.09	0.0
Boiler Chemicals	$ 75.00	/MM lb bldwn	2014	$ 80.516		MMlb/h			
Cooling Tower Chemicals	$ 33.84	/ton-cool/y	2014	$ 36.329		ton-cool/y			
Diesel Fuel	$ 2.57	/gallon	2020	$ 2.571	10	gallon/h	26	0.21	0.0
Subtotal							**18,119**	**150.78**	**16.0**
Waste Streams									
Wastewater Treatment	$ 0.003	/lb	2011	$ 0.003	281,722	lb/h	853	7.10	0.8
Acid Gas Conversion and Recovery	$ 0.477	/lbmol acid gas	2020	$ 0.477	169	lbmol/h	81	0.67	0.1
Subtotal							**933**	**7.77**	**0.8**
Coproducts									
Fuel Gas	$ (2.746)	/MMBtu	2020	$ (2.746)	735	MMBtu/h	-2,019	-16.80	-1.8
HP Steam (900 psig)	$ (0.0045)	/lb	2023	$ (0.004)	0	lb/h	0	0.00	0.0
MP Steam (350 psig)	$ (0.0045)	/lb	2023	$ (0.004)	0	lb/h	0	0.00	0.0
LP Steam (150 psig)	$ (0.0045)	/lb	2023	$ (0.004)	164,237	lb/h	-600	-5.00	-0.5
VLP Steam (15 psig)	$ (0.0044)	/lb	2023	$ (0.004)	232,308	lb/h	-826	-6.87	-0.7
Subtotal							**-3,445**	**-28.67**	**-3.0**
Catalysts and Royalties									
Royalties	$ 0.20	/Bbl Feed	2005	$ 0.309	2,500	Bbl/h	771	6.42	0.7
Subtotal							**771**	**6.42**	**0.7**
Total Variable Operating Costs							**335,944**	**2,795.73**	**296.1**
Fixed Operating Costs									
Total Salaries								2.55	0.3
Overhead and Benefits	% of Labor & Supervison =		0.9					2.29	0.2
Maintenance	% of FCI =		0.03					8.3	0.9
Insurance & Taxes	% of FCI =		0.007					1.9	0.2
Total Fixed Operating Costs								**15.04**	**1.6**
Total Operating Costs								**2,810.77**	**297.7**

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Table I-12. DCFROR for 75k BPD (Including Recycle) 20% CFP-A With SRD, HT Configuration, 20% Liquid Recycle (12120211)

Year		-2	-1	0	1	2	3	4	5	6	7	8
Fixed Capital Investment		$8,823,283	$66,174,626	$35,293,134								
Land		$0										
Working Capital				$13,786,380								
Loan Payment					$24,654,927	$24,654,927	$24,654,927	$24,654,927	$24,654,927	$24,654,927	$24,654,927	$24,654,927
Loan Interest Payment		$1,058,794	$8,999,749	$13,234,925	$13,234,925	$12,321,325	$11,334,637	$10,269,014	$9,118,141	$7,875,198	$6,532,820	$5,083,051
Loan Principal		$13,234,925	$112,496,864	$165,436,565	$154,016,564	$141,682,962	$128,362,672	$113,976,759	$98,439,973	$81,660,244	$63,538,137	$43,966,262
Fuel Sales					$2,144,625,305	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407
Co-Product Sales					$21,502,049	$28,669,399	$28,669,399	$28,669,399	$28,669,399	$28,669,399	$28,669,399	$28,669,399
Total Annual Sales					$2,166,127,354	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806
Annual Manufacturing Cost												
Feedstocks					$1,994,572,111	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482
Process Inputs					$113,087,294	$150,783,059	$150,783,059	$150,783,059	$150,783,059	$150,783,059	$150,783,059	$150,783,059
Waste Streams					$5,824,401	$7,765,868	$7,765,868	$7,765,868	$7,765,868	$7,765,868	$7,765,868	$7,765,868
Catalysts					$22,613,288	$0	$22,613,288	$0	$22,613,288	$0	$22,613,288	$0
Royalties					$4,815,252	$6,420,336	$6,420,336	$6,420,336	$6,420,336	$6,420,336	$6,420,336	$6,420,336
Fixed Operating Costs					$15,042,130	$15,042,130	$15,042,130	$15,042,130	$15,042,130	$15,042,130	$15,042,130	$15,042,130
Total Product Cost					$2,155,954,478	$2,839,440,876	$2,862,054,164	$2,839,440,876	$2,862,054,164	$2,839,440,876	$2,862,054,164	$2,839,440,876
Annual Depreciation												
General Plant Writedown					14%	24.49%	17.49%	12.49%	8.93%	8.92%	8.93%	4.46%
Depreciation Charge					$39,401,475	$67,525,691	$48,224,759	$34,438,378	$24,622,475	$24,594,903	$24,622,475	$12,297,451
Remaining Value					$236,326,133	$168,800,442	$120,575,683	$86,137,305	$61,514,829	$36,919,927	$12,297,451	$0
Steam Plant Writedown					3.75%	7.22%	6.68%	6.18%	5.71%	5.29%	4.89%	4.52%
Depreciation Charge					$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value					$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue					($42,463,524)	($31,118,086)	($33,443,754)	$4,021,538	($7,624,974)	$16,258,830	-$5,039,653	$31,348,428
Losses Forward						($42,463,524)	($73,581,610)	($107,025,364)	($103,003,826)	($110,628,800)	($94,369,971)	($99,409,624)
Taxable Income					($42,463,524)	($73,581,610)	($107,025,364)	($103,003,826)	($110,628,800)	($94,369,971)	($99,409,624)	($68,061,196)
Income Tax					$0	$0	$0	$0	$0	$0	$0	$0
Annual Cash Income					-$14,482,050	$24,074,003	$1,460,715	$24,074,003	$1,460,715	$24,074,003	$1,460,715	$24,074,003
Discount Factor		1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830	0.6209	0.5645	0.5132	0.4665
Annual Present Value	$156,173,488				-$13,165,500	$19,895,871	$1,097,457	$16,442,868	$906,989	$13,589,147	$749,578	$11,230,700
Total Capital Investment + Interest		$11,957,314	$82,691,813	$62,314,439								
Net Present Worth				$0								

Year	9	10	11	12	13	14	15	16	17	18	19
Fixed Capital Investment											
Land											
Working Capital											
Loan Payment	$24,654,927	$24,654,927	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$3,517,301	$1,826,291	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$22,828,636	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407
Co-Product Sales	$28,669,399	$28,669,399	$28,669,399	$28,669,399	$28,669,399	$28,669,399	$28,669,399	$28,669,399	$28,669,399	$28,669,399	$28,669,399
Total Annual Sales	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806
Annual Manufacturing Cost											
Feedstocks	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482
Process Inputs	$150,783,059	$150,783,059	$150,783,059	$150,783,059	$150,783,059	$150,783,059	$150,783,059	$150,783,059	$150,783,059	$150,783,059	$150,783,059
Waste Streams	$7,765,868	$7,765,868	$7,765,868	$7,765,868	$7,765,868	$7,765,868	$7,765,868	$7,765,868	$7,765,868	$7,765,868	$7,765,868
Catalysts	$22,613,288	$0	$22,613,288	$0	$22,613,288	$0	$22,613,288	$0	$22,613,288	$0	$22,613,288
Royalties	$6,420,336	$6,420,336	$6,420,336	$6,420,336	$6,420,336	$6,420,336	$6,420,336	$6,420,336	$6,420,336	$6,420,336	$6,420,336
Fixed Operating Costs	$15,042,130	$15,042,130	$15,042,130	$15,042,130	$15,042,130	$15,042,130	$15,042,130	$15,042,130	$15,042,130	$15,042,130	$15,042,130
Total Product Cost	$2,862,054,164	$2,839,440,876	$2,862,054,164	$2,839,440,876	$2,862,054,164	$2,839,440,876	$2,862,054,164	$2,839,440,876	$2,862,054,164	$2,839,440,876	$2,862,054,164
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%
Depreciation Charge	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue	$22,598,341	$46,902,639	$26,115,642	$48,728,930	$26,115,642	$48,728,930	$26,115,642	$48,728,930	$26,115,642	$48,728,930	$26,115,642
Losses Forward	($68,061,196)	-$45,462,856	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	-$45,462,856	$1,439,783	$26,115,642	$48,728,930	$26,115,642	$48,728,930	$26,115,642	$48,728,930	$26,115,642	$48,728,930	$26,115,642
Income Tax	$0	$302,355	$5,484,285	$10,233,075	$5,484,285	$10,233,075	$5,484,285	$10,233,075	$5,484,285	$10,233,075	$5,484,285
Annual Cash Income	$1,460,715	$23,771,649	$20,631,357	$38,495,855	$20,631,357	$38,495,855	$20,631,357	$38,495,855	$20,631,357	$38,495,855	$20,631,357
Discount Factor	0.4241	0.3855	0.3505	0.3186	0.2897	0.2633	0.2394	0.2176	0.1978	0.1799	0.1635
Annual Present Value	$619,486	$9,165,000	$7,231,165	$12,265,966	$5,976,169	$10,137,162	$4,938,983	$8,377,820	$4,081,804	$6,923,818	$3,373,392
Total Capital Investment + Interest											
Net Present Worth											

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Year	20	21	22	23	24	25	26	27	28	29	30
Fixed Capital Investment											
Land											$0
Working Capital											($13,786,380)
Loan Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407	$2,859,500,407
Co-Product Sales	$28,669,399	$28,669,399	$28,669,399	$28,669,399	$28,669,399	$28,669,399	$28,669,399	$28,669,399	$28,669,399	$28,669,399	$28,669,399
Total Annual Sales	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806	$2,888,169,806
Annual Manufacturing Cost											
Feedstocks	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482	$2,659,429,482
Process Inputs	$150,783,059	$150,783,059	$150,783,059	$150,783,059	$150,783,059	$150,783,059	$150,783,059	$150,783,059	$150,783,059	$150,783,059	$150,783,059
Waste Streams	$7,765,868	$7,765,868	$7,765,868	$7,765,868	$7,765,868	$7,765,868	$7,765,868	$7,765,868	$7,765,868	$7,765,868	$7,765,868
Catalysts	$0	$22,613,288	$0	$22,613,288	$0	$22,613,288	$0	$22,613,288	$0	$22,613,288	$0
Royalties	$6,420,336	$6,420,336	$6,420,336	$6,420,336	$6,420,336	$6,420,336	$6,420,336	$6,420,336	$6,420,336	$6,420,336	$6,420,336
Fixed Operating Costs	$15,042,130	$15,042,130	$15,042,130	$15,042,130	$15,042,130	$15,042,130	$15,042,130	$15,042,130	$15,042,130	$15,042,130	$15,042,130
Total Product Cost	$2,839,440,876	$2,862,054,164	$2,839,440,876	$2,862,054,164	$2,839,440,876	$2,862,054,164	$2,839,440,876	$2,862,054,164	$2,839,440,876	$2,862,054,164	$2,839,440,876
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	2.23%									
Depreciation Charge	$0	$0									
Remaining Value	$0	$0									
Net Revenue	$48,728,930	$26,115,642	$48,728,930	$26,115,642	$48,728,930	$26,115,642	$48,728,930	$26,115,642	$48,728,930	$26,115,642	$48,728,930
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$48,728,930	$26,115,642	$48,728,930	$26,115,642	$48,728,930	$26,115,642	$48,728,930	$26,115,642	$48,728,930	$26,115,642	$48,728,930
Income Tax	$10,233,075	$5,484,285	$10,233,075	$5,484,285	$10,233,075	$5,484,285	$10,233,075	$5,484,285	$10,233,075	$5,484,285	$10,233,075
Annual Cash Income	$38,495,855	$20,631,357	$38,495,855	$20,631,357	$38,495,855	$20,631,357	$38,495,855	$20,631,357	$38,495,855	$20,631,357	$38,495,855
Discount Factor	0.1486	0.1351	0.1228	0.1117	0.1015	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
Annual Present Value	$5,722,164	$2,787,927	$4,729,061	$2,304,072	$3,908,315	$1,904,192	$3,230,012	$1,573,712	$2,669,432	$1,300,589	$2,206,142
Total Capital Investment + Interest											($790,078)
Net Present Worth											

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Table I-13. Capital Costs and FCI for 75k BPD (Including Recycle) 20% CFP-A With SRD, With L-TC or Liquid Temperature Control (12120811)

Equip ID	Equipment Description	Process Variable Value	Size Ratio	Scaled Cost (Base Year)	Scaled Cost (2020$)	Pressure (psig)	Pres. Adj. Scaled Cost (2020)	Lang/TCI Factor (Low)	Lang/TCI Factor (High)	Average Lang/TCI Factor	FCI (2020$) for Single Units of Each	FCI (2020$)
C450	Amine Scrubber	86	0.36	$ 48,794	$ 36,500		36,500	16.71503	29.73753	23.22628	$ 847,771	$ 847,771
C470	Product Recovery Column 1	452138	0.48	$ 3,292,126	$ 2,462,692		2,462,692	3.518324	4.468942	3.993633	$ 9,835,089	$ 9,835,089
C480	Product Recovery Column 2	337819	0.50	$ 2,466,692	$ 1,845,222		1,845,222	3.914891	5.088824	4.501857	$ 8,306,925	$ 8,306,925
MX405	CFP and H2 Static Mixer	7430	0.74	$ 5,432	$ 4,064		4,064	90	100	95	$ 386,048	$ 386,048
MX406	Rx Inlet Static Mixer	7430	0.99	$ 16,014	$ 11,979		11,979	90	100	95	$ 1,138,036	$ 1,138,036
P400A/B	Cat Pyrolysis Oil 2 Pumps	7430	0.74	$ 591,749	$ 442,661	1,085	389,986	1590589	2376455	1983522	$ 1,983,522	$ 3,967,044
P405A/B	Liquid Feed 2 Pumps	29718	0.40	$ 896,625	$ 670,725	1,085	593,870	2269292	3326925	2798109	$ 2,798,109	$ 5,596,217
P455A/B	Liquid Recycle 2 Pumps	37852	1.52	$ 1,266,942	$ 947,742	1,035	800,270	2919821	4223566	3571694	$ 3,571,694	$ 7,143,388
P471A/B	Prod Recov Reflux 2 Pumps - 1	452138	0.48	$ 49,128	$ 36,751	25	36,833	216566.2	359814.6	288190.4	$ 288,190	$ 576,381
P472A/B	Prod Recov Btms 2 Pumps - 1	452138	0.48	$ 51,511	$ 38,533	30	37,157	218173.3	362342	290257.7	$ 290,258	$ 580,515
P481A/B	Prod Recov Reflux 2 Pumps - 2	337819	0.50	$ 45,208	$ 33,818	20	34,614	205488.5	342365.5	273927	$ 273,927	$ 547,854
P482A/B	Prod Recov Btms 2 Pumps - 2	337819	0.50	$ 36,702	$ 27,455	15	26,701	165017.7	278167.2	221592.4	$ 221,592	$ 443,185
PIPE412	H2 MU Comp Jumpover Line	7827	0.64	$ 4,697	$ 3,513		3,513	6.5	7	6.75	$ 23,715	$ 23,715
PIPE442	H2 Rec Comp Jumpover Line	32668	0.43	$ 6,474	$ 4,843		4,843	6.5	7	6.75	$ 32,689	$ 32,689
PIPE500	CFP Pipe from Storage	7430	0.74	$ 39,310	$ 29,406		29,406	6.5	7	6.75	$ 198,492	$ 198,492
PSA460	Pressure Swing Adsorption	6714790	0.08	$ 1,739,039	$ 1,300,897		1,300,897	3.6	4.4	4	$ 5,203,588	$ 5,203,588
QAH413	Rx Effluent Air Cooler	35543861	0.28	$ 697,310	$ 521,626		521,626	4.75	5.75	5.25	$ 2,738,537	$ 2,738,537
QCH413	Rx Effluent Water Cooler	27538652	0.70	$ 802,879	$ 600,598	985	1,349,305	7356188	9512705	8434447	$ 8,434,447	$ 8,434,447
QCH444	Water Cooler H2 Quench 1	0	0.00	$ -	$ -	1,085	-	0	0	0	$ -	$ -
QCH445	Water Cooler H2 Quench 2	0	0.00	$ -	$ -	1,085	-	0	0	0	$ -	$ -
QFH411-L	Reactor Preheat Furnace - Liquid Feed	31257675	0.74	$ 1,183,860	$ 885,593	1,085	885,593	4.55	7.37	5.96	$ 4,032,463	$ 4,032,463
QFH411-G	Reactor Preheat Furnace - Gas Feed	0	0.00	$ -	$ -	1,085	-	0	0	0	$ -	$ -
QH412	Process Rx Eff Cooler	114073602	0.83	$ 3,132,358	$ 2,343,176	995	5,496,455	24035861	27778773	25907317	$ 25,907,317	$ 25,907,317
QH471	Prod Recov Condenser 1	452138	0.48	$ 2,359,419	$ 1,764,976	25	1,764,976	9225093	11675935	10450514	$ 10,450,514	$ 10,450,514
QH472	Prod Recov Reboiler 1	452138	0.48	$ 2,153,826	$ 1,611,180	30	1,611,180	8542649	10891332	9716990	$ 9,716,990	$ 9,716,990
QH481	Prod Recov Condenser 2	337819	0.50	$ 1,273,043	$ 952,306	20	952,306	5483765	7291832	6387798	$ 6,387,798	$ 6,387,798
QH482	Prod Recov Reboiler 2	337819	0.50	$ 968,283	$ 724,329	15	724,329	4354077	5917812	5135944	$ 5,135,944	$ 5,135,944
QXH410	Feed Preheater to Furnace	291636993	0.79	$ 1,027,205	$ 768,406	1,085	1,869,887	9685240	12201835	10943537	$ 10,943,537	$ 10,943,537
RX405	CFP Cat Stabilization Rx	7430	0.74	$ 17,381	$ 13,002	1,085	12,557	35.5	45.5	40.5	$ 508,556	$ 508,556
RX410-HT	Reactor System - Hydrotreater	1122253	0.88	$ 10,042,138	$ 7,512,073	1,030	6,886,303	3.5	5.3	4.4	$ 30,299,734	$ 30,299,734
RX410-HC	Reactor System - Hydrocracker	1122253	0.88	$ 10,042,138	$ 7,512,073	1,030	6,886,303	3.5	5.3	4.4	$ 30,299,734	$ -
TK400	Cat Pyrolysis Oil Storage	7430	0.74	$ 1,934,735	$ 1,447,288		1,447,288	6	7.2	6.6	$ 9,552,102	$ 9,552,102
V451	High Pressure Flash Drum	1261862	0.99	$ 803,114	$ 600,774	980	1,189,579	6.499179	10.61199	8.555587	$ 10,177,543	$ 10,177,543
V452	Low Pressure Flash Drum	926182	0.94	$ 210,062	$ 157,138	1,085	157,138	13.46911	25.85869	19.6639	$ 2,116,507	$ 2,116,507
V471	Prod Recov Ovhd Drum 1	379860	0.40	$ 40,571	$ 30,349	1,085	30,349	24.34601	53.31215	38.82908	$ 738,878	$ 738,878
V481	Prod Recov Ovhd Drum 2	373585	0.55	$ 55,456	$ 41,485	1,085	41,485	21.7551	46.46224	34.10867	$ 902,500	$ 902,500
WK412	H2 Make-up Compressor	7838	0.59	$ 7,479,709	$ 5,595,236		5,595,236	3	3.8	3.4	$ 19,023,801	$ 19,023,801
WK442	H2 Recycle Compressor	2142	0.54	$ 2,531,606	$ 1,893,781		1,893,781	3	3.8	3.4	$ 6,438,856	$ 6,438,856
QH470	Dist Col 1 Liquid Feed Preheat Furnace	20259271	0.48	$ 948,143	$ 709,264	36	709,264	4.932303	8.127266	6.529785	$ 3,498,303	$ 3,498,303
Total												**$ 211,831,264**

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Table I-14. Operating Costs for 75k BPD (Including Recycle) 20% CFP-A With SRD, With L-TC or Liquid Temperature Control (12120811)

	Quoted Price	Unit	Cost Year	2020$ price	Scaling Flow	Unit	2020$ Cost ($/h)	2020$ Cost (MM$/y)	2020$ Cost (cents/GGE)
Feedstocks									
Crude Oil Basis	$ 85.00	/bbl	2020	$ 85.00					
Straight Run Diesel (SRD)	$ 105.88	/bbl	2020	$ 105.88	29,718	bbl/day	131,107	1,091.07	186.4
Light Cycle Oil (LCO)	$ 102.95	/bbl	2020	$ 102.95	0	bbl/day	0	0.00	0.0
Vacuuum Gas Oil (VGO)	$ 100.78	/bbl	2020	$ 100.78	0	bbl/day	0	0.00	0.0
Catalytic Fast Pyrolysis Oil (21.6%)	$ 215.62	/bbl	2020	$ 215.62	7,430	bbl/day	66,748	555.48	94.9
Bio-Oil Transportation 800 miles	$ -	/ton-mile	2007	$ -	124,890	lb/h	0	0.00	0.0
Subtotal							**197,854**	**1,646.54**	**281.3**
Process Inputs									
Hydrogen (pure H2 basis)									
Total Hydrogen Feed					15,621	lb/h			
Hydrogen Recovered via PSA					935	lb/h			
Net Hydrogen Makeup	$ 0.680	/lb	2020	$ 0.680	14,687	lb/h	9,993	83.16	14.2
Electricity for Hydroprocessing	$ 0.082	/kWh	2020	$ 0.082	10,371	kWh	847	7.05	1.2
Natural Gas	$ 3.231	/MMBtu	2020	$ 3.231	0	MMBtu/h	0	0.00	0.0
Fuel Gas	$ 2.746	/MMBtu	2020	$ 2.746	201	MMBtu/h	551	4.58	0.8
Amine Consumption	$ 0.00122600	/lbmol inlet gas	2020	$ 0.00122600	41,675	lbmol/h	51	0.43	0.073
Water Makeup	$ 0.200	/ton	2004	$ 0.334	70	ton/h	23	0.19	0.0
BFW to Steam	$ 0.001	/lb	2016	$ 0.001		lb/h			
Cooling Water Circulation	$ 0.00768	/ton	2020	$ 0.008	791	ton/h	6	0.05	0.0
Boiler Chemicals	$ 75.00	/MM lb bldwn	2014	$ 80.516		MMlb/h			
Cooling Tower Chemicals	$ 33.84	/ton-cool/y	2014	$ 36.329		ton-cool/y			
Diesel Fuel	$ 2.57	/gallon	2020	$ 2.571	10	gallon/h	26	0.21	0.0
Subtotal							**11,497**	**95.68**	**16.3**
Waste Streams									
Wastewater Treatment	$ 0.003	/lb	2011	$ 0.003	172,796	lb/h	523	4.35	0.7
Acid Gas Conversion and Recovery	$ 0.477	/lbmol acid gas	2020	$ 0.477	86	lbmol/h	41	0.34	0.1
Subtotal							**564**	**4.69**	**0.8**
Coproducts									
Fuel Gas	$ (2.746)	/MMBtu	2020	$ (2.746)	461	MMBtu/h	-1,267	-10.55	-1.8
HP Steam (900 psig)	$ (0.0045)	/lb	2023	$ (0.004)	0	lb/h	0	0.00	0.0
MP Steam (350 psig)	$ (0.0045)	/lb	2023	$ (0.004)	0	lb/h	0	0.00	0.0
LP Steam (150 psig)	$ (0.0045)	/lb	2023	$ (0.004)	86,965	lb/h	-318	-2.65	-0.5
VLP Steam (15 psig)	$ (0.0044)	/lb	2023	$ (0.004)	158,215	lb/h	-563	-4.68	-0.8
Subtotal							**-2,148**	**-17.87**	**-3.1**
Catalysts and Royalties									
Royalties	$ 0.20	/Bbl Feed	2005	$ 0.309	1,548	Bbl/h	478	3.98	0.7
Subtotal							**478**	**3.98**	**0.7**
Total Variable Operating Costs							**208,245**	**1,733.02**	**296.1**
Fixed Operating Costs									
Total Salaries								2.55	0.4
Overhead and Benefits	% of Labor & Supervison =		0.9					2.29	0.4
Maintenance	% of FCI =		0.03					6.4	1.1
Insurance & Taxes	% of FCI =		0.007					1.5	0.3
Total Fixed Operating Costs								**12.68**	**2.2**
Total Operating Costs								**1,745.70**	**298.2**

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Table I-15. DCFROR for 75k BPD (Including Recycle) 20% CFP-A With SRD, With L-TC or Liquid Temperature Control (12120811)

Year		-2	-1	0	1	2	3	4	5	6	7	8
Fixed Capital Investment		$6,778,600	$50,839,503	$27,114,402								
Land		$0										
Working Capital				$10,591,563								
Loan Payment					$18,941,463	$18,941,463	$18,941,463	$18,941,463	$18,941,463	$18,941,463	$18,941,463	$18,941,463
Loan Interest Payment		$813,432	$6,914,172	$10,167,901	$10,167,901	$9,466,016	$8,707,980	$7,889,301	$7,005,128	$6,050,222	$5,018,922	$3,905,119
Loan Principal		$10,167,901	$86,427,156	$127,098,758	$118,325,196	$108,849,749	$98,616,266	$87,564,104	$75,627,769	$62,736,528	$48,813,987	$33,777,643
Fuel Sales					$1,336,140,132	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176
Co-Product Sales					$13,404,589	$17,872,785	$17,872,785	$17,872,785	$17,872,785	$17,872,785	$17,872,785	$17,872,785
Total Annual Sales					$1,349,544,721	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961
Annual Manufacturing Cost												
Feedstocks					$1,234,908,581	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774
Process Inputs					$71,757,033	$95,676,044	$95,676,044	$95,676,044	$95,676,044	$95,676,044	$95,676,044	$95,676,044
Waste Streams					$3,520,561	$4,694,081	$4,694,081	$4,694,081	$4,694,081	$4,694,081	$4,694,081	$4,694,081
Catalysts					$14,334,823	$0	$14,334,823	$0	$14,334,823	$0	$14,334,823	$0
Royalties					$2,981,289	$3,975,052	$3,975,052	$3,975,052	$3,975,052	$3,975,052	$3,975,052	$3,975,052
Fixed Operating Costs					$12,677,965	$12,677,965	$12,677,965	$12,677,965	$12,677,965	$12,677,965	$12,677,965	$12,677,965
Total Product Cost					$1,340,180,252	$1,763,567,917	$1,777,902,740	$1,763,567,917	$1,777,902,740	$1,763,567,917	$1,777,902,740	$1,763,567,917
Annual Depreciation												
General Plant Writedown					14%	24.49%	17.49%	12.49%	8.93%	8.92%	8.93%	4.46%
Depreciation Charge					$30,270,688	$51,877,477	$37,049,288	$26,457,725	$18,916,532	$18,895,349	$18,916,532	$9,447,674
Remaining Value					$181,560,576	$129,683,100	$92,633,812	$66,176,087	$47,259,555	$28,364,206	$9,447,674	$0
Steam Plant Writedown					3.75%	7.22%	6.68%	6.18%	5.71%	5.29%	4.89%	4.52%
Depreciation Charge					$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value					$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue					($31,074,120)	($25,518,448)	($24,267,047)	$1,478,018	($4,431,439)	$10,879,474	-$2,445,233	$22,472,251
Losses Forward						($31,074,120)	($56,592,568)	($80,859,615)	($79,381,597)	($83,813,036)	($72,933,562)	($75,378,795)
Taxable Income					($31,074,120)	($56,592,568)	($80,859,615)	($79,381,597)	($83,813,036)	($72,933,562)	($75,378,795)	($52,906,544)
Income Tax					$0	$0	$0	$0	$0	$0	$0	$0
Annual Cash Income					-$9,576,994	$16,883,581	$2,548,758	$16,883,581	$2,548,758	$16,883,581	$2,548,758	$16,883,581
Discount Factor		1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830	0.6209	0.5645	0.5132	0.4665
Annual Present Value	$119,982,281				-$8,706,358	$13,953,373	$1,914,920	$11,531,713	$1,582,578	$9,530,341	$1,307,916	$7,876,315
Total Capital Investment + Interest		$9,186,359	$63,529,043	$47,873,866								
Net Present Worth				$0								

Year	9	10	11	12	13	14	15	16	17	18	19
Fixed Capital Investment											
Land											
Working Capital											
Loan Payment	$18,941,463	$18,941,463	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$2,702,211	$1,403,071	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$17,538,392	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176
Co-Product Sales	$17,872,785	$17,872,785	$17,872,785	$17,872,785	$17,872,785	$17,872,785	$17,872,785	$17,872,785	$17,872,785	$17,872,785	$17,872,785
Total Annual Sales	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961
Annual Manufacturing Cost											
Feedstocks	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774
Process Inputs	$95,676,044	$95,676,044	$95,676,044	$95,676,044	$95,676,044	$95,676,044	$95,676,044	$95,676,044	$95,676,044	$95,676,044	$95,676,044
Waste Streams	$4,694,081	$4,694,081	$4,694,081	$4,694,081	$4,694,081	$4,694,081	$4,694,081	$4,694,081	$4,694,081	$4,694,081	$4,694,081
Catalysts	$14,334,823	$0	$14,334,823	$0	$14,334,823	$0	$14,334,823	$0	$14,334,823	$0	$14,334,823
Royalties	$3,975,052	$3,975,052	$3,975,052	$3,975,052	$3,975,052	$3,975,052	$3,975,052	$3,975,052	$3,975,052	$3,975,052	$3,975,052
Fixed Operating Costs	$12,677,965	$12,677,965	$12,677,965	$12,677,965	$12,677,965	$12,677,965	$12,677,965	$12,677,965	$12,677,965	$12,677,965	$12,677,965
Total Product Cost	$1,777,902,740	$1,763,567,917	$1,777,902,740	$1,763,567,917	$1,777,902,740	$1,763,567,917	$1,777,902,740	$1,763,567,917	$1,777,902,740	$1,763,567,917	$1,777,902,740
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%
Depreciation Charge	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue	$18,788,010	$34,421,973	$21,490,221	$35,825,044	$21,490,221	$35,825,044	$21,490,221	$35,825,044	$21,490,221	$35,825,044	$21,490,221
Losses Forward	($52,906,544)	-$34,118,535	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	-$34,118,535	$303,438	$21,490,221	$35,825,044	$21,490,221	$35,825,044	$21,490,221	$35,825,044	$21,490,221	$35,825,044	$21,490,221
Income Tax	$0	$63,722	$4,512,946	$7,523,259	$4,512,946	$7,523,259	$4,512,946	$7,523,259	$4,512,946	$7,523,259	$4,512,946
Annual Cash Income	$2,548,758	$16,819,859	$16,977,275	$28,301,785	$16,977,275	$28,301,785	$16,977,275	$28,301,785	$16,977,275	$28,301,785	$16,977,275
Discount Factor	0.4241	0.3855	0.3505	0.3186	0.2897	0.2633	0.2394	0.2176	0.1978	0.1799	0.1635
Annual Present Value	$1,080,922	$6,484,784	$5,950,431	$9,017,821	$4,917,712	$7,452,744	$4,064,225	$6,159,293	$3,358,863	$5,090,325	$2,775,920
Total Capital Investment + Interest											
Net Present Worth											

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Year	20	21	22	23	24	25	26	27	28	29	30
Fixed Capital Investment											
Land											$0
Working Capital											($10,591,563)
Loan Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176	$1,781,520,176
Co-Product Sales	$17,872,785	$17,872,785	$17,872,785	$17,872,785	$17,872,785	$17,872,785	$17,872,785	$17,872,785	$17,872,785	$17,872,785	$17,872,785
Total Annual Sales	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961	$1,799,392,961
Annual Manufacturing Cost											
Feedstocks	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774	$1,646,544,774
Process Inputs	$95,676,044	$95,676,044	$95,676,044	$95,676,044	$95,676,044	$95,676,044	$95,676,044	$95,676,044	$95,676,044	$95,676,044	$95,676,044
Waste Streams	$4,694,081	$4,694,081	$4,694,081	$4,694,081	$4,694,081	$4,694,081	$4,694,081	$4,694,081	$4,694,081	$4,694,081	$4,694,081
Catalysts	$0	$14,334,823	$0	$14,334,823	$0	$14,334,823	$0	$14,334,823	$0	$14,334,823	$0
Royalties	$3,975,052	$3,975,052	$3,975,052	$3,975,052	$3,975,052	$3,975,052	$3,975,052	$3,975,052	$3,975,052	$3,975,052	$3,975,052
Fixed Operating Costs	$12,677,965	$12,677,965	$12,677,965	$12,677,965	$12,677,965	$12,677,965	$12,677,965	$12,677,965	$12,677,965	$12,677,965	$12,677,965
Total Product Cost	$1,763,567,917	$1,777,902,740	$1,763,567,917	$1,777,902,740	$1,763,567,917	$1,777,902,740	$1,763,567,917	$1,777,902,740	$1,763,567,917	$1,777,902,740	$1,763,567,917
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	2.23%									
Depreciation Charge	$0	$0									
Remaining Value	$0	$0									
Net Revenue	$35,825,044	$21,490,221	$35,825,044	$21,490,221	$35,825,044	$21,490,221	$35,825,044	$21,490,221	$35,825,044	$21,490,221	$35,825,044
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$35,825,044	$21,490,221	$35,825,044	$21,490,221	$35,825,044	$21,490,221	$35,825,044	$21,490,221	$35,825,044	$21,490,221	$35,825,044
Income Tax	$7,523,259	$4,512,946	$7,523,259	$4,512,946	$7,523,259	$4,512,946	$7,523,259	$4,512,946	$7,523,259	$4,512,946	$7,523,259
Annual Cash Income	$28,301,785	$16,977,275	$28,301,785	$16,977,275	$28,301,785	$16,977,275	$28,301,785	$16,977,275	$28,301,785	$16,977,275	$28,301,785
Discount Factor	0.1486	0.1351	0.1228	0.1117	0.1015	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
Annual Present Value	$4,206,880	$2,294,149	$3,476,760	$1,895,991	$2,873,356	$1,566,935	$2,374,674	$1,294,987	$1,962,541	$1,070,237	$1,621,934
Total Capital Investment + Interest											($606,987)
Net Present Worth											

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Appendix J. TEA Summary for Select CFP-B Hydroprocessing Cases

Cases included:

- 11199011: HT of 100% CFP-B oil in H-TC configuration with no liquid recycle, 99 mol % H_2 makeup purity, process model with 75k BPD capacity.

- 11220811: HT of 20 vol% CFP-B oil with 80 vol % SRD in L-TC configuration, 85 mol % H_2 makeup purity, process model with 75k BPD capacity.

Please see Section 3.4.1 for interpretation of Case IDs.

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Table J-1. Equipment Costs and Fixed Capital Investment for 75k BPD CFP-B (16.0% O) HT Configuration Without Liquid Recycle (11199011)

Equip ID	Equipment Description	Process Variable Value	Size Ratio	Scaled Cost (Base Year)	Scaled Cost (2020$)	Pressure (psig)	Pres. Adj. Scaled Cost (2020)	Lang/TCI Factor (Low)	Lang/TCI Factor (High)	Average Lang/TCI Factor	FCI (2020$) for Single Units of Each	FCI (2020$)
C450	Amine Scrubber	866	3.60	$ 312,688	$ 233,908		233,908	8.406418	12.89048	10.64845	$ 2,490,757	$ 2,490,757
C470	Product Recovery Column 1	909836	0.96	$ 5,042,049	$ 3,771,731		3,771,731	3.004952	3.688892	3.346922	$ 12,623,690	$ 12,623,690
C480	Product Recovery Column 2	645636	0.95	$ 3,589,602	$ 2,685,220		2,685,220	3.407493	4.298316	3.852905	$ 10,345,898	$ 10,345,898
MX405	CFP and H2 Static Mixer	75000	7.50	$ 15,195	$ 11,367		11,367	90	100	95	$ 1,079,850	$ 1,079,850
MX406	Rx Inlet Static Mixer	75000	10.00	$ 59,583	$ 44,572		44,572	90	100	95	$ 4,234,312	$ 4,234,312
P400A/B	Cat Pyrolysis Oil 2 Pumps	75000	7.50	$ 905,506	$ 677,368	1,085	599,839	2288551	3353650	2821101	$ 2,821,101	$ 5,642,201
P405A/B	Liquid Feed 2 Pumps	0	0.00	$ -	$ -	1,085	-	0	0	0	$ -	$ -
P455A/B	Liquid Recycle 2 Pumps	0	0.00	$ -	$ -	1,035	-	0	0	0	$ -	$ -
P471A/B	Prod Recov Reflux 2 Pumps - 1	909836	0.96	$ 60,481	$ 45,243	25	45,344	258154.4	424917.8	341536.1	$ 341,536	$ 683,072
P472A/B	Prod Recov Btms 2 Pumps - 1	909836	0.96	$ 64,379	$ 48,159	30	46,448	263457.2	433176.8	348317	$ 348,317	$ 696,634
P481A/B	Prod Recov Reflux 2 Pumps - 2	645636	0.95	$ 54,778	$ 40,977	20	41,937	241667.1	399180.5	320423.8	$ 320,424	$ 640,848
P482A/B	Prod Recov Btms 2 Pumps - 2	645636	0.95	$ 42,329	$ 31,664	15	30,796	186164.7	311805.6	248985.1	$ 248,985	$ 497,970
PIPE412	H2 MU Comp Jumpover Line	42478	3.49	$ 15,345	$ 11,479		11,479	6.5	7	6.75	$ 77,483	$ 77,483
PIPE442	H2 Rec Comp Jumpover Line	272590	3.60	$ 28,585	$ 21,383		21,383	6.5	7	6.75	$ 144,335	$ 144,335
PIPE500	CFP Pipe from Storage	75000	7.50	$ 198,322	$ 148,356		148,356	6.5	7	6.75	$ 1,001,404	$ 1,001,404
PSA460	Pressure Swing Adsorption	114321159	1.37	$ 7,074,583	$ 5,292,179		5,292,179	3.6	4.4	4	$ 21,168,715	$ 21,168,715
QAH413	Rx Effluent Air Cooler	418693602	3.30	$ 6,239,161	$ 4,667,237		4,667,237	4.75	5.75	5.25	$ 24,502,995	$ 24,502,995
QCH413	Rx Effluent Water Cooler	105661209	2.69	$ 2,520,206	$ 1,885,253	985	4,382,619	19858733	23370760	21614747	$ 21,614,747	$ 21,614,747
QCH444	Water Cooler H2 Quench 1	18051567	3.81	$ 581,303	$ 434,847	1,085	1,039,595	5904547	7796459	6850503	$ 6,850,503	$ 6,850,503
QCH445	Water Cooler H2 Quench 2	18051567	3.81	$ 581,303	$ 434,847	1,085	1,039,595	5904547	7796459	6850503	$ 6,850,503	$ 6,850,503
QFH411-L	Reactor Preheat Furnace - Liquid Feed	72630620	1.71	$ 1,822,955	$ 1,363,671	1,085	1,363,671	3.90	6.10	5.00	$ 5,315,629	$ 5,315,629
QFH411-G	Reactor Preheat Furnace - Gas Feed	0	0.00	$ -	$ -	1,085	-	0	0	0	$ -	$ -
QH412	Process Rx Eff Cooler	524597441	3.83	$ 14,091,925	$ 10,541,538	995	24,959,277	86066983	88128831	87097907	$ 87,097,907	$ 87,097,907
QH471	Prod Recov Condenser 1	909836	0.96	$ 4,195,112	$ 3,138,175	25	3,138,175	14985418	18113179	16549298	$ 16,549,298	$ 16,549,298
QH472	Prod Recov Reboiler 1	909836	0.96	$ 3,816,194	$ 2,854,724	30	2,854,724	13836001	16850991	15343496	$ 15,343,496	$ 15,343,496
QH481	Prod Recov Condenser 2	645636	0.95	$ 2,101,573	$ 1,572,092	20	1,572,092	8367601	10689139	9528370	$ 9,528,370	$ 9,528,370
QH482	Prod Recov Reboiler 2	645636	0.95	$ 1,600,746	$ 1,197,447	15	1,197,447	6651812	8684393	7668102	$ 7,668,102	$ 7,668,102
QXH410	Feed Preheater to Furnace	719176662	1.96	$ 2,282,517	$ 1,707,449	1,085	4,207,343	19187070	22654163	20920617	$ 20,920,617	$ 20,920,617
RX405	CFP Cat Stabilization Rx	75000	7.50	$ 28,245	$ 21,128	1,085	20,683	35.5	45.5	40.5	$ 837,681	$ 837,681
RX410-HT	Reactor System - Hydrotreater	2838375	2.23	$ 23,343,952	$ 17,462,565	1,030	16,836,795	3.5	5.3	4.4	$ 74,081,896	$ 74,081,896
RX410-HC	Reactor System - Hydrocracker	2838375	2.23	$ 23,343,952	$ 17,462,565	1,030	16,836,795	3.5	5.3	4.4	$ 74,081,896	$ -
TK400	Cat Pyrolysis Oil Storage	75000	7.50	$ 14,423,781	$ 10,789,784		10,789,784	6	7.2	6.6	$ 71,212,576	$ 71,212,576
V451	High Pressure Flash Drum	2838375	2.23	$ 1,741,495	$ 1,302,734	980	1,891,539	5.499795	8.65309	7.076443	$ 13,385,369	$ 13,385,369
V452	Low Pressure Flash Drum	937953	0.95	$ 212,019	$ 158,602	1,085	158,602	13.42421	25.75337	19.58879	$ 2,129,108	$ 2,129,108
V471	Prod Recov Ovhd Drum 1	799230	0.84	$ 52,059	$ 38,943	1,085	38,943	22.25589	47.77278	35.01433	$ 866,714	$ 866,714
V481	Prod Recov Ovhd Drum 2	641458	0.94	$ 66,607	$ 49,826	1,085	49,826	20.36653	42.86382	31.61517	$ 1,014,775	$ 1,014,775
WK412	H2 Make-up Compressor	42534	3.19	$ 28,755,840	$ 21,510,956		21,510,956	3	3.8	3.4	$ 73,137,249	$ 73,137,249
WK442	H2 Recycle Compressor	17675	4.44	$ 4,806,356	$ 3,595,420		3,595,420	3	3.8	3.4	$ 12,224,427	$ 12,224,427
QH470	Dist Col 1 Liquid Feed Preheat Furnace	57348348	1.35	$ 1,615,271	$ 1,208,312	36	1,208,312	4.071516	6.428965	5.250241	$ 4,919,660	$ 4,919,660
Total												**$ 537,378,791**

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Table J-2. Operating Costs for 75k BPD CFP-B (16.0% O) HT Configuration Without Liquid Recycle (11199011)

	Quoted Price	Unit	Cost Year	2020$ price	Scaling Flow	Unit	2020$ Cost ($/h)	2020$ Cost (MM$/y)	2020$ Cost (cents/GGE)
Feedstocks									
Crude Oil Basis	$ 85.00	/bbl	2020	$ 85.00					
Straight Run Diesel (SRD)	$ 105.88	/bbl	2020	$ 105.88	0	bbl/day	0	0.00	0.0
Light Cycle Oil (LCO)	$ 102.95	/bbl	2020	$ 102.95	0	bbl/day	0	0.00	0.0
Vacuuum Gas Oil (VGO)	$ 100.78	/bbl	2020	$ 100.78	0	bbl/day	0	0.00	0.0
Catalytic Fast Pyrolysis Oil (16.0%)	$ 148.30	/bbl	2020	$ 148.30	75,000	bbl/day	463,437	3,856.73	329.9
Bio-Oil Transportation 800 miles	$ -	/ton-mile	2007	$ -	1,153,385	lb/h	0	0.00	0.0
Subtotal							**463,438**	**3,856.73**	**329.9**
Process Inputs									
Hydrogen (pure H2 basis)									
Total Hydrogen Feed					84,774	lb/h			
Hydrogen Recovered via PSA					16,447	lb/h			
Net Hydrogen Makeup	$ 0.680	/lb	2020	$ 0.680	68,327	lb/h	46,489	386.88	33.1
Electricity for Hydroprocessing	$ 0.082	/kWh	2020	$ 0.082	54,557	kWh	4,457	37.09	3.2
Natural Gas	$ 3.231	/MMBtu	2020	$ 3.231	0	MMBtu/h	0	0.00	0.0
Fuel Gas	$ 2.746	/MMBtu	2020	$ 2.746	396	MMBtu/h	1,088	9.05	0.8
Amine Consumption	$ 0.00122600	/lbmol inlet gas	2020	$ 0.00122600	260,970	lbmol/h	320	2.66	0.228
Water Makeup	$ 0.200	/ton	2004	$ 0.334	0	ton/h	0	0.00	0.0
BFW to Steam	$ 0.001	/lb	2016	$ 0.001		lb/h			
Cooling Water Circulation	$ 0.00768	/ton	2020	$ 0.008	3,666	ton/h	28	0.23	0.0
Boiler Chemicals	$ 75.00	/MM lb bldwn	2014	$ 80.516		MMlb/h			
Cooling Tower Chemicals	$ 33.84	/ton-cool/y	2014	$ 36.329		ton-cool/y			
Diesel Fuel	$ 2.57	/gallon	2020	$ 2.571	10	gallon/h	26	0.21	0.0
Subtotal							**52,408**	**436.14**	**37.3**
Waste Streams									
Wastewater Treatment	$ 0.003	/lb	2011	$ 0.003	182,342	lb/h	552	4.59	0.4
Acid Gas Conversion and Recovery	$ 0.477	/lbmol acid gas	2020	$ 0.477	866	lbmol/h	413	3.44	0.3
Subtotal							**965**	**8.03**	**0.7**
Coproducts									
Fuel Gas	$ (2.746)	/MMBtu	2020	$ (2.746)	2,389	MMBtu/h	-6,560	-54.59	-4.7
HP Steam (900 psig)	$ (0.0045)	/lb	2023	$ (0.004)	0	lb/h	0	0.00	0.0
MP Steam (350 psig)	$ (0.0045)	/lb	2023	$ (0.004)	33,787	lb/h	-124	-1.03	-0.1
LP Steam (150 psig)	$ (0.0045)	/lb	2023	$ (0.004)	254,807	lb/h	-931	-7.75	-0.7
VLP Steam (15 psig)	$ (0.0044)	/lb	2023	$ (0.004)	481,347	lb/h	-1,711	-14.24	-1.2
Subtotal							**-9,327**	**-77.62**	**-6.6**
Catalysts and Royalties									
Royalties	$ 0.20	/Bbl Feed	2005	$ 0.309	3,125	Bbl/h	964	8.03	0.7
Subtotal							**964**	**8.03**	**0.7**
Total Variable Operating Costs							**508,448**	**4,231.30**	**361.9**
Fixed Operating Costs									
Total Salaries								2.55	0.2
Overhead and Benefits		% of Labor & Supervison =	0.9					2.29	0.2
Maintenance		% of FCI =	0.03					16.1	1.4
Insurance & Taxes		% of FCI =	0.007					3.8	0.3
Total Fixed Operating Costs								**24.72**	**2.1**
Total Operating Costs								**4,256.03**	**364.1**

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Table J-3. DCFROR for 75k BPD CFP-B (16.0% O) HT Configuration Without Liquid Recycle (11199011)

Year		-2	-1	0	1	2	3	4	5	6	7	8
Fixed Capital Investment		$17,196,121	$128,970,910	$68,784,485								
Land		$0										
Working Capital				$26,868,940								
Loan Payment					$48,051,172	$48,051,172	$48,051,172	$48,051,172	$48,051,172	$48,051,172	$48,051,172	$48,051,172
Loan Interest Payment		$2,063,535	$17,540,044	$25,794,182	$25,794,182	$24,013,623	$22,090,619	$20,013,775	$17,770,783	$15,348,352	$12,732,126	$9,906,602
Loan Principal		$25,794,182	$219,250,547	$322,427,274	$300,170,285	$276,132,736	$250,172,182	$222,134,785	$191,854,396	$159,151,576	$123,832,530	$85,687,961
Fuel Sales					$3,276,886,118	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491
Co-Product Sales					$58,214,713	$77,619,617	$77,619,617	$77,619,617	$77,619,617	$77,619,617	$77,619,617	$77,619,617
Total Annual Sales					$3,335,100,831	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108
Annual Manufacturing Cost												
Feedstocks					$2,892,546,319	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425
Process Inputs					$327,103,544	$436,138,059	$436,138,059	$436,138,059	$436,138,059	$436,138,059	$436,138,059	$436,138,059
Waste Streams					$6,022,703	$8,030,271	$8,030,271	$8,030,271	$8,030,271	$8,030,271	$8,030,271	$8,030,271
Catalysts					$77,955,077	$0	$77,955,077	$0	$77,955,077	$0	$77,955,077	$0
Royalties					$6,019,065	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420
Fixed Operating Costs					$24,723,224	$24,723,224	$24,723,224	$24,723,224	$24,723,224	$24,723,224	$24,723,224	$24,723,224
Total Product Cost					$3,334,369,933	$4,333,645,399	$4,411,600,476	$4,333,645,399	$4,411,600,476	$4,333,645,399	$4,411,600,476	$4,333,645,399
Annual Depreciation												
General Plant Writedown					14%	24.49%	17.49%	12.49%	8.93%	8.92%	8.93%	4.46%
Depreciation Charge					$76,791,429	$131,604,066	$93,987,551	$67,118,611	$47,987,926	$47,934,188	$47,987,926	$23,967,094
Remaining Value					$460,587,362	$328,983,296	$234,995,745	$167,877,134	$119,889,208	$71,955,020	$23,967,094	$0
Steam Plant Writedown					3.75%	7.22%	6.68%	6.18%	5.71%	5.29%	4.89%	4.52%
Depreciation Charge					$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value					$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue					($101,854,713)	($42,461,979)	($80,877,538)	$26,023,324	($30,558,077)	$49,873,169	-$25,519,420	$79,282,013
Losses Forward						($101,854,713)	($144,316,692)	($225,194,230)	($199,170,906)	($229,728,984)	($179,855,814)	($205,375,235)
Taxable Income					($101,854,713)	($144,316,692)	($225,194,230)	($199,170,906)	($229,728,984)	($179,855,814)	($205,375,235)	($126,093,222)
Income Tax					$0	$0	$0	$0	$0	$0	$0	$0
Annual Cash Income					-$47,320,273	$65,104,537	-$12,850,540	$65,104,537	-$12,850,540	$65,104,537	-$12,850,540	$65,104,537
Discount Factor		1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830	0.6209	0.5645	0.5132	0.4665
Annual Present Value	$304,374,019				-$43,018,430	$53,805,403	-$9,654,801	$44,467,275	-$7,979,174	$36,749,814	-$6,594,359	$30,371,747
Total Capital Investment + Interest		$23,304,184	$161,162,049	$121,447,607								
Net Present Worth				$0								

Year	9	10	11	12	13	14	15	16	17	18	19
Fixed Capital Investment											
Land											
Working Capital											
Loan Payment	$48,051,172	$48,051,172	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$6,855,037	$3,559,346	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$44,491,826	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491
Co-Product Sales	$77,619,617	$77,619,617	$77,619,617	$77,619,617	$77,619,617	$77,619,617	$77,619,617	$77,619,617	$77,619,617	$77,619,617	$77,619,617
Total Annual Sales	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108
Annual Manufacturing Cost											
Feedstocks	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425
Process Inputs	$436,138,059	$436,138,059	$436,138,059	$436,138,059	$436,138,059	$436,138,059	$436,138,059	$436,138,059	$436,138,059	$436,138,059	$436,138,059
Waste Streams	$8,030,271	$8,030,271	$8,030,271	$8,030,271	$8,030,271	$8,030,271	$8,030,271	$8,030,271	$8,030,271	$8,030,271	$8,030,271
Catalysts	$77,955,077	$0	$77,955,077	$0	$77,955,077	$0	$77,955,077	$0	$77,955,077	$0	$77,955,077
Royalties	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420
Fixed Operating Costs	$24,723,224	$24,723,224	$24,723,224	$24,723,224	$24,723,224	$24,723,224	$24,723,224	$24,723,224	$24,723,224	$24,723,224	$24,723,224
Total Product Cost	$4,411,600,476	$4,333,645,399	$4,411,600,476	$4,333,645,399	$4,411,600,476	$4,333,645,399	$4,411,600,476	$4,333,645,399	$4,411,600,476	$4,333,645,399	$4,411,600,476
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%
Depreciation Charge	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue	$28,345,595	$109,596,363	$35,200,632	$113,155,709	$35,200,632	$113,155,709	$35,200,632	$113,155,709	$35,200,632	$113,155,709	$35,200,632
Losses Forward	($126,093,222)	-$97,747,627	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	-$97,747,627	$11,848,736	$35,200,632	$113,155,709	$35,200,632	$113,155,709	$35,200,632	$113,155,709	$35,200,632	$113,155,709	$35,200,632
Income Tax	$0	$2,488,235	$7,392,133	$23,762,699	$7,392,133	$23,762,699	$7,392,133	$23,762,699	$7,392,133	$23,762,699	$7,392,133
Annual Cash Income	-$12,850,540	$62,616,303	$27,808,499	$89,393,010	$27,808,499	$89,393,010	$27,808,499	$89,393,010	$27,808,499	$89,393,010	$27,808,499
Discount Factor	0.4241	0.3855	0.3505	0.3186	0.2897	0.2633	0.2394	0.2176	0.1978	0.1799	0.1635
Annual Present Value	-$5,449,883	$24,141,295	$9,746,709	$28,483,368	$8,055,132	$23,539,974	$6,657,134	$19,454,524	$5,501,763	$16,078,119	$4,546,912
Total Capital Investment + Interest											
Net Present Worth											

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Year	20	21	22	23	24	25	26	27	28	29	30
Fixed Capital Investment											
Land											$0
Working Capital											($26,868,940)
Loan Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491	$4,369,181,491
Co-Product Sales	$77,619,617	$77,619,617	$77,619,617	$77,619,617	$77,619,617	$77,619,617	$77,619,617	$77,619,617	$77,619,617	$77,619,617	$77,619,617
Total Annual Sales	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108	$4,446,801,108
Annual Manufacturing Cost											
Feedstocks	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425	$3,856,728,425
Process Inputs	$436,138,059	$436,138,059	$436,138,059	$436,138,059	$436,138,059	$436,138,059	$436,138,059	$436,138,059	$436,138,059	$436,138,059	$436,138,059
Waste Streams	$8,030,271	$8,030,271	$8,030,271	$8,030,271	$8,030,271	$8,030,271	$8,030,271	$8,030,271	$8,030,271	$8,030,271	$8,030,271
Catalysts	$0	$77,955,077	$0	$77,955,077	$0	$77,955,077	$0	$77,955,077	$0	$77,955,077	$0
Royalties	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420	$8,025,420
Fixed Operating Costs	$24,723,224	$24,723,224	$24,723,224	$24,723,224	$24,723,224	$24,723,224	$24,723,224	$24,723,224	$24,723,224	$24,723,224	$24,723,224
Total Product Cost	$4,333,645,399	$4,411,600,476	$4,333,645,399	$4,411,600,476	$4,333,645,399	$4,411,600,476	$4,333,645,399	$4,411,600,476	$4,333,645,399	$4,411,600,476	$4,333,645,399
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	2.23%									
Depreciation Charge	$0	$0									
Remaining Value	$0	$0									
Net Revenue	$113,155,709	$35,200,632	$113,155,709	$35,200,632	$113,155,709	$35,200,632	$113,155,709	$35,200,632	$113,155,709	$35,200,632	$113,155,709
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$113,155,709	$35,200,632	$113,155,709	$35,200,632	$113,155,709	$35,200,632	$113,155,709	$35,200,632	$113,155,709	$35,200,632	$113,155,709
Income Tax	$23,762,699	$7,392,133	$23,762,699	$7,392,133	$23,762,699	$7,392,133	$23,762,699	$7,392,133	$23,762,699	$7,392,133	$23,762,699
Annual Cash Income	$89,393,010	$27,808,499	$89,393,010	$27,808,499	$89,393,010	$27,808,499	$89,393,010	$27,808,499	$89,393,010	$27,808,499	$89,393,010
Discount Factor	0.1486	0.1351	0.1228	0.1117	0.1015	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
Annual Present Value	$13,287,701	$3,757,778	$10,981,571	$3,105,602	$9,075,679	$2,566,613	$7,500,561	$2,121,168	$6,198,811	$1,753,031	$5,122,984
Total Capital Investment + Interest											($1,539,820)
Net Present Worth											

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Table J-4. Capital Costs and FCI for 75k BPD (Including Recycle) 20% CFP-B With SRD, With L-TC or Liquid Temperature Control (11220811)

Equip ID	Equipment Description	Process Variable Value	Size Ratio	Scaled Cost (Base Year)	Scaled Cost (2020$)	Pressure (psig)	Pres. Adj. Scaled Cost (2020)	Lang/TCI Factor (Low)	Lang/TCI Factor (High)	Average Lang/TCI Factor	FCI (2020$) for Single Units of Each	FCI (2020$)
C450	Amine Scrubber	104	0.43	$ 56,744	$ 42,447		42,447	15.80713	27.78476	21.79595	$ 925,179	$ 925,179
C470	Product Recovery Column 1	538384	0.57	$ 3,661,821	$ 2,739,244		2,739,244	3.382472	4.259959	3.821216	$ 10,467,243	$ 10,467,243
C480	Product Recovery Column 2	386231	0.57	$ 2,665,649	$ 1,994,053		1,994,053	3.804127	4.914256	4.359192	$ 8,692,458	$ 8,692,458
MX405	CFP and H2 Static Mixer	8763	0.88	$ 5,846	$ 4,373		4,373	90	100	95	$ 415,473	$ 415,473
MX406	Rx Inlet Static Mixer	8763	1.17	$ 17,589	$ 13,158		13,158	90	100	95	$ 1,249,986	$ 1,249,986
P400A/B	Cat Pyrolysis Oil 2 Pumps	8763	0.88	$ 610,002	$ 456,315	1,085	402,136	1632361	2435501	2033931	$ 2,033,931	$ 4,067,862
P405A/B	Liquid Feed 2 Pumps	35053	0.47	$ 987,319	$ 738,569	1,085	654,912	2464880	3597794	3031337	$ 3,031,337	$ 6,062,673
P455A/B	Liquid Recycle 2 Pumps	31183	1.25	$ 1,131,454	$ 846,390	1,035	712,373	2646421	3848184	3247303	$ 3,247,303	$ 6,494,605
P471A/B	Prod Recov Reflux 2 Pumps - 1	538384	0.57	$ 51,746	$ 38,709	25	38,795	226275.4	375069	300672.2	$ 300,672	$ 601,344
P472A/B	Prod Recov Btms 2 Pumps - 1	538384	0.57	$ 54,460	$ 40,739	30	39,286	228691.4	378859.3	303775.3	$ 303,775	$ 607,551
P481A/B	Prod Recov Reflux 2 Pumps - 2	386231	0.57	$ 47,039	$ 35,188	20	36,015	212495.8	353408.8	282952.3	$ 282,952	$ 565,905
P482A/B	Prod Recov Btms 2 Pumps - 2	386231	0.57	$ 37,801	$ 28,277	15	27,500	169183.4	284810.8	226997.1	$ 226,997	$ 453,994
PIPE412	H2 MU Comp Jumpover Line	13979	1.15	$ 7,048	$ 5,272		5,272	6.5	7	6.75	$ 35,589	$ 35,589
PIPE442	H2 Rec Comp Jumpover Line	30488	0.40	$ 6,168	$ 4,614		4,614	6.5	7	6.75	$ 31,147	$ 31,147
PIPE500	CFP Pipe from Storage	8763	0.88	$ 44,126	$ 33,009		33,009	6.5	7	6.75	$ 222,811	$ 222,811
PSA460	Pressure Swing Adsorption	61076462	0.73	$ 5,187,233	$ 3,880,337		3,880,337	3.6	4.4	4	$ 15,521,347	$ 15,521,347
QAH413	Rx Effluent Air Cooler	50907671	0.40	$ 959,507	$ 717,764		717,764	4.75	5.75	5.25	$ 3,768,263	$ 3,768,263
QCH413	Rx Effluent Water Cooler	26600302	0.68	$ 779,547	$ 583,144	985	1,308,092	7166319	9290200	8228260	$ 8,228,260	$ 8,228,260
QCH444	Water Cooler H2 Quench 1	0	0.00	$ -	$ -	1,085	-	0	0	0	$ -	$ -
QCH445	Water Cooler H2 Quench 2	0	0.00	$ -	$ -	1,085	-	0	0	0	$ -	$ -
QFH411-L	Reactor Preheat Furnace - Liquid Feed	33956152	0.80	$ 1,235,131	$ 923,946	1,085	923,946	4.48	7.23	5.86	$ 4,143,376	$ 4,143,376
QFH411-G	Reactor Preheat Furnace - Gas Feed	0	0.00	$ -	$ -	1,085	-	0	0	0	$ -	$ -
QH412	Process Rx Eff Cooler	98447686	0.72	$ 2,709,025	$ 2,026,500	995	4,744,669	21233011	24829907	23031459	$ 23,031,459	$ 23,031,459
QH471	Prod Recov Condenser 1	538384	0.57	$ 2,723,993	$ 2,037,697	25	2,037,697	10412971	13028772	11720872	$ 11,720,872	$ 11,720,872
QH472	Prod Recov Reboiler 1	538384	0.57	$ 2,484,461	$ 1,858,514	30	1,858,514	9635558	12145169	10890363	$ 10,890,363	$ 10,890,363
QH481	Prod Recov Condenser 2	386231	0.57	$ 1,412,068	$ 1,056,304	20	1,056,304	5984450	7891891	6938171	$ 6,938,171	$ 6,938,171
QH482	Prod Recov Reboiler 2	386231	0.57	$ 1,074,342	$ 803,667	15	803,667	4752798	6406241	5579519	$ 5,579,519	$ 5,579,519
QXH410	Feed Preheater to Furnace	307833238	0.84	$ 1,077,510	$ 806,037	1,085	1,963,557	10092662	12665497	11379080	$ 11,379,080	$ 11,379,080
RX405	CFP Cat Stabilization Rx	8763	0.88	$ 17,994	$ 13,461	1,085	13,016	35.5	45.5	40.5	$ 527,134	$ 527,134
RX410-HT	Reactor System - Hydrotreater	1141436	0.90	$ 10,198,067	$ 7,628,717	1,030	7,002,947	3.5	5.3	4.4	$ 30,812,965	$ 30,812,965
RX410-HC	Reactor System - Hydrocracker	1141436	0.90	$ 10,198,067	$ 7,628,717	1,030	7,002,947	3.5	5.3	4.4	$ 30,812,965	$ -
TK400	Cat Pyrolysis Oil Storage	8763	0.88	$ 2,233,186	$ 1,670,546		1,670,546	6	7.2	6.6	$ 11,025,607	$ 11,025,607
V451	High Pressure Flash Drum	1247707	0.98	$ 794,511	$ 594,338	980	1,183,143	6.511884	10.63736	8.57462	$ 10,144,998	$ 10,144,998
V452	Low Pressure Flash Drum	950998	0.96	$ 214,181	$ 160,219	1,085	160,219	13.37528	25.63869	19.50699	$ 2,142,974	$ 2,142,974
V471	Prod Recov Ovhd Drum 1	488037	0.52	$ 44,126	$ 33,008	1,085	33,008	23.62083	51.37776	37.4993	$ 779,686	$ 779,686
V481	Prod Recov Ovhd Drum 2	423204	0.62	$ 57,851	$ 43,275	1,085	43,275	21.4266	45.6062	33.5164	$ 927,245	$ 927,245
WK412	H2 Make-up Compressor	13814	1.04	$ 11,744,848	$ 8,785,795		8,785,795	3	3.8	3.4	$ 29,871,703	$ 29,871,703
WK442	H2 Recycle Compressor	2011	0.51	$ 2,483,439	$ 1,857,750		1,857,750	3	3.8	3.4	$ 6,316,348	$ 6,316,348
QH470	Dist Col 1 Liquid Feed Preheat Furnace	34178297	0.80	$ 1,239,261	$ 927,036	36	927,036	4.479049	7.223994	5.851522	$ 4,152,239	$ 4,152,239
Total												**$ 238,795,428**

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Table J-5. Operating Costs for 75k BPD (Including Recycle) 20% CFP-B With SRD, With L-TC or Liquid Temperature Control (11220811)

	Quoted Price	Unit	Cost Year	2020$ price	Scaling Flow	Unit	2020$ Cost ($/h)	2020$ Cost (MM$/y)	2020$ Cost (cents/GGE)
Feedstocks									
Crude Oil Basis	$ 85.00	/bbl	2020	$ 85.00					
Straight Run Diesel (SRD)	$ 105.88	/bbl	2020	$ 105.88	35,053	bbl/day	154,642	1,286.93	187.3
Light Cycle Oil (LCO)	$ 102.95	/bbl	2020	$ 102.95	0	bbl/day	0	0.00	0.0
Vacuuum Gas Oil (VGO)	$ 100.78	/bbl	2020	$ 100.78	0	bbl/day	0	0.00	0.0
Catalytic Fast Pyrolysis Oil (16.0%)	$ 148.30	/bbl	2020	$ 148.30	8,763	bbl/day	54,150	450.64	65.6
Bio-Oil Transportation 800 miles	$ -	/ton-mile	2007	$ -	134,767	lb/h	0	0.00	0.0
Subtotal							**208,793**	**1,737.57**	**252.8**
Process Inputs									
Hydrogen (pure H2 basis)									
Total Hydrogen Feed					23,952	lb/h			
Hydrogen Recovered via PSA					8,716	lb/h			
Net Hydrogen Makeup	$ 0.680	/lb	2020	$ 0.680	15,236	lb/h	10,367	86.27	12.6
Electricity for Hydroprocessing	$ 0.082	/kWh	2020	$ 0.082	15,543	kWh	1,270	10.57	1.5
Natural Gas	$ 3.231	/MMBtu	2020	$ 3.231	0	MMBtu/h	0	0.00	0.0
Fuel Gas	$ 2.746	/MMBtu	2020	$ 2.746	240	MMBtu/h	659	5.49	0.8
Amine Consumption	$ 0.00122600	/lbmol inlet gas	2020	$ 0.00122600	37,844	lbmol/h	46	0.39	0.056
Water Makeup	$ 0.200	/ton	2004	$ 0.334	53	ton/h	18	0.15	0.0
BFW to Steam	$ 0.001	/lb	2016	$ 0.001		lb/h			
Cooling Water Circulation	$ 0.00768	/ton	2020	$ 0.008	810	ton/h	6	0.05	0.0
Boiler Chemicals	$ 75.00	/MM lb bldwn	2014	$ 80.516		MMlb/h			
Cooling Tower Chemicals	$ 33.84	/ton-cool/y	2014	$ 36.329		ton-cool/y			
Diesel Fuel	$ 2.57	/gallon	2020	$ 2.571	10	gallon/h	26	0.21	0.0
Subtotal							**12,392**	**103.13**	**15.0**
Waste Streams									
Wastewater Treatment	$ 0.003	/lb	2011	$ 0.003	130,799	lb/h	396	3.29	0.5
Acid Gas Conversion and Recovery	$ 0.477	/lbmol acid gas	2020	$ 0.477	104	lbmol/h	50	0.41	0.1
Subtotal							**445**	**3.71**	**0.5**
Coproducts									
Fuel Gas	$ (2.746)	/MMBtu	2020	$ (2.746)	620	MMBtu/h	-1,704	-14.18	-2.1
HP Steam (900 psig)	$ (0.0045)	/lb	2023	$ (0.004)	0	lb/h	0	0.00	0.0
MP Steam (350 psig)	$ (0.0045)	/lb	2023	$ (0.004)	0	lb/h	0	0.00	0.0
LP Steam (150 psig)	$ (0.0045)	/lb	2023	$ (0.004)	62,432	lb/h	-228	-1.90	-0.3
VLP Steam (15 psig)	$ (0.0044)	/lb	2023	$ (0.004)	184,643	lb/h	-657	-5.46	-0.8
Subtotal							**-2,588**	**-21.54**	**-3.1**
Catalysts and Royalties									
Royalties	$ 0.20	/Bbl Feed	2005	$ 0.309	1,826	Bbl/h	563	4.69	0.7
Subtotal							**563**	**4.69**	**0.7**
Total Variable Operating Costs							**219,605**	**1,827.55**	**265.9**
Fixed Operating Costs									
Total Salaries								2.55	0.4
Overhead and Benefits	% of Labor & Supervison =		0.9					2.29	0.3
Maintenance	% of FCI =		0.03					7.2	1.0
Insurance & Taxes	% of FCI =		0.007					1.7	0.2
Total Fixed Operating Costs								**13.68**	**2.0**
Total Operating Costs								**1,841.23**	**267.9**

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Table J-6. DCFROR for 75k BPD (Including Recycle) 20% CFP-B With SRD, With L-TC or Liquid Temperature Control (11220811)

Year		-2	-1	0	1	2	3	4	5	6	7	8
Fixed Capital Investment		$7,641,454	$57,310,903	$30,565,815								
Land		$0										
Working Capital				$11,939,771								
Loan Payment					$21,352,536	$21,352,536	$21,352,536	$21,352,536	$21,352,536	$21,352,536	$21,352,536	$21,352,536
Loan Interest Payment		$916,974	$7,794,283	$11,462,181	$11,462,181	$10,670,952	$9,816,425	$8,893,536	$7,896,816	$6,820,359	$5,657,785	$4,402,205
Loan Principal		$11,462,181	$97,428,534	$143,277,257	$133,386,901	$122,705,317	$111,169,206	$98,710,206	$85,254,486	$70,722,309	$55,027,557	$38,077,225
Fuel Sales					$1,409,946,434	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578
Co-Product Sales					$16,154,810	$21,539,747	$21,539,747	$21,539,747	$21,539,747	$21,539,747	$21,539,747	$21,539,747
Total Annual Sales					$1,426,101,244	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325
Annual Manufacturing Cost												
Feedstocks					$1,303,179,238	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317
Process Inputs					$77,344,706	$103,126,275	$103,126,275	$103,126,275	$103,126,275	$103,126,275	$103,126,275	$103,126,275
Waste Streams					$2,780,173	$3,706,897	$3,706,897	$3,706,897	$3,706,897	$3,706,897	$3,706,897	$3,706,897
Catalysts					$13,067,769	$0	$13,067,769	$0	$13,067,769	$0	$13,067,769	$0
Royalties					$3,516,479	$4,688,639	$4,688,639	$4,688,639	$4,688,639	$4,688,639	$4,688,639	$4,688,639
Fixed Operating Costs					$13,675,639	$13,675,639	$13,675,639	$13,675,639	$13,675,639	$13,675,639	$13,675,639	$13,675,639
Total Product Cost					$1,413,564,004	$1,862,769,767	$1,875,837,536	$1,862,769,767	$1,875,837,536	$1,862,769,767	$1,875,837,536	$1,862,769,767
Annual Depreciation												
General Plant Writedown					14%	24.49%	17.49%	12.49%	8.93%	8.92%	8.93%	4.46%
Depreciation Charge					$34,123,867	$58,481,000	$41,765,320	$29,825,549	$21,324,432	$21,300,552	$21,324,432	$10,650,276
Remaining Value					$204,671,561	$146,190,561	$104,425,241	$74,599,692	$53,275,260	$31,974,708	$10,650,276	$0
Steam Plant Writedown					3.75%	7.22%	6.68%	6.18%	5.71%	5.29%	4.89%	4.52%
Depreciation Charge					$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value					$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue					($33,048,807)	($30,453,394)	($25,950,956)	($20,527)	($3,590,459)	$10,577,647	-$1,351,427	$23,646,078
Losses Forward						($33,048,807)	($63,502,201)	($89,453,158)	($89,473,685)	($93,064,144)	($82,486,496)	($83,837,924)
Taxable Income					($33,048,807)	($63,502,201)	($89,453,158)	($89,473,685)	($93,064,144)	($82,486,496)	($83,837,924)	($60,191,846)
Income Tax					$0	$0	$0	$0	$0	$0	$0	$0
Annual Cash Income					-$8,815,296	$17,346,022	$4,278,253	$17,346,022	$4,278,253	$17,346,022	$4,278,253	$17,346,022
Discount Factor		1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830	0.6209	0.5645	0.5132	0.4665
Annual Present Value	$135,254,918				-$8,013,906	$14,335,555	$3,214,315	$11,847,566	$2,656,459	$9,791,377	$2,195,420	$8,092,047
Total Capital Investment + Interest		$10,355,698	$71,615,704	$53,967,767								
Net Present Worth				$0								

Year	9	10	11	12	13	14	15	16	17	18	19
Fixed Capital Investment											
Land											
Working Capital											
Loan Payment	$21,352,536	$21,352,536	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$3,046,178	$1,581,669	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$19,770,867	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578
Co-Product Sales	$21,539,747	$21,539,747	$21,539,747	$21,539,747	$21,539,747	$21,539,747	$21,539,747	$21,539,747	$21,539,747	$21,539,747	$21,539,747
Total Annual Sales	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325
Annual Manufacturing Cost											
Feedstocks	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317
Process Inputs	$103,126,275	$103,126,275	$103,126,275	$103,126,275	$103,126,275	$103,126,275	$103,126,275	$103,126,275	$103,126,275	$103,126,275	$103,126,275
Waste Streams	$3,706,897	$3,706,897	$3,706,897	$3,706,897	$3,706,897	$3,706,897	$3,706,897	$3,706,897	$3,706,897	$3,706,897	$3,706,897
Catalysts	$13,067,769	$0	$13,067,769	$0	$13,067,769	$0	$13,067,769	$0	$13,067,769	$0	$13,067,769
Royalties	$4,688,639	$4,688,639	$4,688,639	$4,688,639	$4,688,639	$4,688,639	$4,688,639	$4,688,639	$4,688,639	$4,688,639	$4,688,639
Fixed Operating Costs	$13,675,639	$13,675,639	$13,675,639	$13,675,639	$13,675,639	$13,675,639	$13,675,639	$13,675,639	$13,675,639	$13,675,639	$13,675,639
Total Product Cost	$1,875,837,536	$1,862,769,767	$1,875,837,536	$1,862,769,767	$1,875,837,536	$1,862,769,767	$1,875,837,536	$1,862,769,767	$1,875,837,536	$1,862,769,767	$1,875,837,536
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%
Depreciation Charge	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue	$22,584,611	$37,116,889	$25,630,789	$38,698,558	$25,630,789	$38,698,558	$25,630,789	$38,698,558	$25,630,789	$38,698,558	$25,630,789
Losses Forward	($60,191,846)	-$37,607,235	-$490,346	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	-$37,607,235	-$490,346	$25,140,443	$38,698,558	$25,630,789	$38,698,558	$25,630,789	$38,698,558	$25,630,789	$38,698,558	$25,630,789
Income Tax	$0	$0	$5,279,493	$8,126,697	$5,382,466	$8,126,697	$5,382,466	$8,126,697	$5,382,466	$8,126,697	$5,382,466
Annual Cash Income	$4,278,253	$17,346,022	$20,351,296	$30,571,861	$20,248,324	$30,571,861	$20,248,324	$30,571,861	$20,248,324	$30,571,861	$20,248,324
Discount Factor	0.4241	0.3855	0.3505	0.3186	0.2897	0.2633	0.2394	0.2176	0.1978	0.1799	0.1635
Annual Present Value	$1,814,397	$6,687,642	$7,133,005	$9,741,137	$5,865,218	$8,050,526	$4,847,288	$6,653,328	$4,006,023	$5,498,618	$3,310,763
Total Capital Investment + Interest											
Net Present Worth											

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Year	20	21	22	23	24	25	26	27	28	29	30
Fixed Capital Investment											
Land											$0
Working Capital											($11,939,771)
Loan Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578	$1,879,928,578
Co-Product Sales	$21,539,747	$21,539,747	$21,539,747	$21,539,747	$21,539,747	$21,539,747	$21,539,747	$21,539,747	$21,539,747	$21,539,747	$21,539,747
Total Annual Sales	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325	$1,901,468,325
Annual Manufacturing Cost											
Feedstocks	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317	$1,737,572,317
Process Inputs	$103,126,275	$103,126,275	$103,126,275	$103,126,275	$103,126,275	$103,126,275	$103,126,275	$103,126,275	$103,126,275	$103,126,275	$103,126,275
Waste Streams	$3,706,897	$3,706,897	$3,706,897	$3,706,897	$3,706,897	$3,706,897	$3,706,897	$3,706,897	$3,706,897	$3,706,897	$3,706,897
Catalysts	$0	$13,067,769	$0	$13,067,769	$0	$13,067,769	$0	$13,067,769	$0	$13,067,769	$0
Royalties	$4,688,639	$4,688,639	$4,688,639	$4,688,639	$4,688,639	$4,688,639	$4,688,639	$4,688,639	$4,688,639	$4,688,639	$4,688,639
Fixed Operating Costs	$13,675,639	$13,675,639	$13,675,639	$13,675,639	$13,675,639	$13,675,639	$13,675,639	$13,675,639	$13,675,639	$13,675,639	$13,675,639
Total Product Cost	$1,862,769,767	$1,875,837,536	$1,862,769,767	$1,875,837,536	$1,862,769,767	$1,875,837,536	$1,862,769,767	$1,875,837,536	$1,862,769,767	$1,875,837,536	$1,862,769,767
Annual Depreciation											
General Plant Writedown											
Depreciation Charge											
Remaining Value											
Steam Plant Writedown	4.46%	2.23%									
Depreciation Charge	$0	$0									
Remaining Value	$0	$0									
Net Revenue	$38,698,558	$25,630,789	$38,698,558	$25,630,789	$38,698,558	$25,630,789	$38,698,558	$25,630,789	$38,698,558	$25,630,789	$38,698,558
Losses Forward	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Taxable Income	$38,698,558	$25,630,789	$38,698,558	$25,630,789	$38,698,558	$25,630,789	$38,698,558	$25,630,789	$38,698,558	$25,630,789	$38,698,558
Income Tax	$8,126,697	$5,382,466	$8,126,697	$5,382,466	$8,126,697	$5,382,466	$8,126,697	$5,382,466	$8,126,697	$5,382,466	$8,126,697
Annual Cash Income	$30,571,861	$20,248,324	$30,571,861	$20,248,324	$30,571,861	$20,248,324	$30,571,861	$20,248,324	$30,571,861	$20,248,324	$30,571,861
Discount Factor	0.1486	0.1351	0.1228	0.1117	0.1015	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
Annual Present Value	$4,544,312	$2,736,168	$3,755,630	$2,261,295	$3,103,826	$1,868,839	$2,565,146	$1,544,495	$2,119,955	$1,276,442	$1,752,029
Total Capital Investment + Interest											($684,251)
Net Present Worth											

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Appendix K. Summary for Hydroprocessing Cases

Please refer to Section 3.4.1 for interpreting Case IDs.

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Table K-1. Hydroprocessing Cases 11101011 to 11140011

Case ID	11101011	11110011	11120011	11120211	11120811	11130011	11140011
Total Feed, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
SRD Feed, BPD	75,000	67,500	60,000	48,000	34,643	52,500	45,000
LCO Feed, BPD	0	0	0	0	0	0	0
VGO Feed, BPD	0	0	0	0	0	0	0
CFP Oil Feed, BPD	0	7,500	15,000	12,000	8,661	22,500	30,000
Liquid Product Recycle, BPD	0	0	0	15,000	31,696	0	0
CFP Oil in Fresh Feed, vol %	0%	10%	20%	20%	20%	30%	40%
Liquid Product Recycle, vol %	0%	0%	0%	20%	42%	0%	0%
CFP Oil Oxygen (dry basis), wt %	16.0%	16.0%	16.0%	16.0%	16.0%	16.0%	16.0%
Total Liquid Fuel Product, lb/h	902,001	898,501	896,447	717,264	518,038	895,012	894,158
Total Liquid Fuel Product (all cuts), BGE/day	81,316	81,037	80,858	64,692	46,730	80,732	80,679
Total Liquid Fuel Product (all cuts), BJE/day	75,941	75,680	75,513	60,416	43,641	75,396	75,346
Total Liquid Fuel Product (all cuts), BDE/day	73,491	73,239	73,078	58,467	42,234	72,964	72,916
Total Liquid Product Molar H:C Ratio	1.79	1.80	1.80	1.80	1.80	1.79	1.80
Modeled Gasoline Product, wt %	23%	25%	26%	26%	26%	28%	29%
Modeled Jet Product, wt %	47%	48%	48%	48%	48%	48%	48%
Modeled Diesel/Heavies, wt %	30%	28%	26%	26%	26%	24%	23%
Hydroprocessing Reactor							
Inlet Pressure, psia	1,045	1,045	1,045	1,045	1,045	1,045	1,045
Inlet Temperature, F	644	644	644	644	644	644	644
Outlet Temperature, F	684	684	684	684	684	684	684
Carbon-Efficiency for CFP Oil, %	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%
Carbon-Efficiency for Petroleum Feed, %	98.8%	98.7%	98.7%	98.6%	98.5%	98.6%	98.6%
WHSV proration basis from SRD using chem. H2							
Base SRD Case H2 Consumption, scf/bbl	615.0	615.0	615.0	615.0	615.0	615.0	615.0
Base SRD Case WHSV Assumption, 1/h	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrogen							
Makeup H2 Purity, mol %	99%	99%	99%	99%	99%	99%	99%
H2 Purity at Reactor Inlet, mol %	91%	80%	80%	80%	80%	80%	80%
Chemical H2 per Total Feed, scf/bbl	613	980	1,309	1,056	776	1,612	1,946
Chemical H2 per Fresh Feed, scf/bbl	613	980	1,309	1,320	1,344	1,612	1,946
Reactor H2 Feed per Total Feed, scf/bbl	2,838	5,125	7,676	3,780	2,926	9,982	12,410
Reactor H2 Feed per Fresh Feed, scf/bbl	2,838	5,125	7,676	4,725	5,068	9,982	12,410
Makeup H2 per Total Feed, scf/bbl	659	1,046	1,494	1,153	824	1,897	2,358
Makeup H2 per Fresh Feed, scf/bbl	659	1,046	1,494	1,441	1,427	1,897	2,358
Net H2 after PSA, per Total Feed, scf/bbl	659	1,026	1,374	1,106	819	1,693	2,047
Hydrotreating							
WHSV, 1/h	2.01	1.26	0.94	0.93	0.92	0.76	0.63
Catalyst Load, lb	550,208	901,544	1,234,557	995,793	732,021	1,556,837	1,923,771
Catalyst Replacement Period, years	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrocracking							
WHSV, 1/h	N/A	N/A	N/A	N/A	N/A	N/A	N/A
Catalyst Load, lb	N/A	N/A	N/A	N/A	N/A	N/A	N/A

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Case ID	11101011	11110011	11120011	11120211	11120811	11130011	11140011
Catalyst Replacement Period, years	N/A	N/A	N/A	N/A	N/A	N/A	N/A
TEA							
Capital							
Fixed Capital Investment (FCI), million $	196.3	243.6	285.4	246.6	209.3	320.3	355.3
Working Capital, million $	9.8	12.2	14.3	12.3	10.5	16.0	17.8
Operating Costs, $/GGE	2.40	2.53	2.66	2.66	2.67	2.79	2.91
Petroleum Feed Cost, $/GGE	2.33	2.10	1.87	1.87	1.87	1.64	1.41
CFP Oil Feed Cost, $/GGE	0.00	0.33	0.66	0.65	0.65	0.98	1.31
Net Hydrogen Makeup Cost, $/GGE	0.05	0.08	0.11	0.11	0.11	0.14	0.16
Fixed Operating Costs, $/GGE	0.01	0.01	0.01	0.01	0.02	0.01	0.02
Coproduct/Utility Credits, $/GGE	−0.02	−0.02	−0.03	−0.03	−0.03	−0.03	−0.04
Other Operating Costs, $/GGE	0.03	0.04	0.05	0.05	0.05	0.06	0.07
Capital Charges, Income Tax etc.	0.02	0.03	0.03	0.04	0.04	0.04	0.04
MFSP, with Full Capital Costs, $/GGE	2.42	2.56	2.69	2.70	2.71	2.82	2.95
MFSP, with Full Capital Costs, $/GJE	2.59	2.74	2.89	2.89	2.90	3.02	3.16
MFSP, with Full Capital Costs, $/GDE	2.68	2.83	2.98	2.98	3.00	3.12	3.27
Other TEA Metrics							
Total Operating Cost without Feed, cents/GGE	7.6	10.6	13.6	13.6	14.3	16.3	19.3
Hydrogen Cost, cents/GGE	5.2	8.2	11.0	11.1	11.3	13.6	16.4
Total Catalyst Cost per Feed Barrel, $/bbl	0.18	0.29	0.40	0.41	0.41	0.51	0.63
Delta Capital for Repurposed Unit, million $	0.4	37.9	75.2	49.2	34.0	108.7	144.1
Repurposed Case ID	11201011	11201011	11201011	11201011	11201011	11201011	11201011
MFSP with Repurposed Equipment, $/GGE	2.40	2.54	2.67	2.67	2.68	2.80	2.93
Repurposed Petroleum Unit Size, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
Repurposed Unit FCI, $/annual bbl fresh feed	0.0	1.5	2.9	2.4	2.3	4.2	5.5
New Unit FCI, $/annual bbl fresh feed	7.5	9.4	11.0	11.9	13.9	12.3	13.7
Hydrotreating reactor FCI, million $	27.4	33.4	38.2	31.6	30.0	42.6	47.3
Hydrocracking reactor FCI, million $	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Reactor(s) Pressure Correction On/Off	On	On	On	On	On	On	On
Makeup compressor FCI, million $	14.3	20.7	27.5	22.4	17.1	33.3	39.6
Recycle compressor FCI, million $	6.0	7.4	8.4	6.8	6.3	9.1	9.7
Fuel Lifecycle CI, g CO2e/MJ	90.6	82.9	75.3	75.1	75.3	67.7	60.1
CI for Petroleum Diesel Case, g CO2e/MJ	90.8	90.8	90.8	90.8	90.8	90.8	90.8
CI Reduction, %	0.2%	8.7%	17.1%	17.3%	17.0%	25.5%	33.8%

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Table K-2. Hydroprocessing Cases 11150011 to 11201011

Case ID	11150011	11160011	11170011	11180011	11190011	11199011	11201011
Total Feed, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
SRD Feed, BPD	37,500	30,000	22,500	15,000	7,500	0	75,000
LCO Feed, BPD	0	0	0	0	0	0	0
VGO Feed, BPD	0	0	0	0	0	0	0
CFP Oil Feed, BPD	37,500	45,000	52,500	60,000	67,500	75,000	0
Liquid Product Recycle, BPD	0	0	0	0	0	0	0
CFP Oil in Fresh Feed, vol %	50%	60%	70%	80%	90%	100%	0%
Liquid Product Recycle, vol %	0%	0%	0%	0%	0%	0%	0%
CFP Oil Oxygen (dry basis), wt %	16.0%	16.0%	16.0%	16.0%	16.0%	16.0%	16.0%
Total Liquid Fuel Product, lb/h	893,402	892,884	891,984	890,627	889,076	888,545	899,973
Total Liquid Fuel Product (all cuts), BGE/day	80,630	80,641	80,578	80,426	80,298	80,274	81,148
Total Liquid Fuel Product (all cuts), BJE/day	75,300	75,311	75,252	75,109	74,990	74,968	75,784
Total Liquid Fuel Product (all cuts), BDE/day	72,872	72,882	72,824	72,687	72,572	72,550	73,340
Total Liquid Product Molar H:C Ratio	1.80	1.80	1.80	1.80	1.80	1.80	1.79
Modeled Gasoline Product, wt %	31%	31%	31%	30%	27%	27%	23%
Modeled Jet Product, wt %	47%	47%	46%	45%	45%	45%	47%
Modeled Diesel/Heavies, wt %	22%	22%	23%	25%	28%	28%	30%
Hydroprocessing Reactor							
Inlet Pressure, psia	1,045	1,045	1,045	1,045	1,045	1,045	1,045
Inlet Temperature, F	644	644	644	644	644	644	644
Outlet Temperature, F	684	684	683	684	683	684	684
Carbon-Efficiency for CFP Oil, %	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%
Carbon-Efficiency for Petroleum Feed, %	98.5%	98.5%	98.5%	98.4%	98.4%	99.5%	98.6%
WHSV proration basis from SRD using chem. H2							
Base SRD Case H2 Consumption, scf/bbl	615.0	615.0	615.0	615.0	615.0	615.0	615.0
Base SRD Case WHSV Assumption, 1/h	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrogen							
Makeup H2 Purity, mol %	99%	99%	99%	99%	99%	99%	85%
H2 Purity at Reactor Inlet, mol %	80%	80%	80%	80%	80%	80%	80%
Chemical H2 per Total Feed, scf/bbl	2,276	2,627	2,953	3,248	3,568	3,891	612
Chemical H2 per Fresh Feed, scf/bbl	2,276	2,627	2,953	3,248	3,568	3,891	612
Reactor H2 Feed per Total Feed, scf/bbl	14,786	17,330	19,726	21,836	24,230	26,442	2,267
Reactor H2 Feed per Fresh Feed, scf/bbl	14,786	17,330	19,726	21,836	24,230	26,442	2,267
Makeup H2 per Total Feed, scf/bbl	2,815	3,299	3,757	4,185	4,653	5,100	921
Makeup H2 per Fresh Feed, scf/bbl	2,815	3,299	3,757	4,185	4,653	5,100	921
Net H2 after PSA, per Total Feed, scf/bbl	2,396	2,768	3,114	3,428	3,769	4,111	696
Hydrotreating							
WHSV, 1/h	0.54	0.47	0.42	0.38	0.34	0.32	2.01
Catalyst Load, lb	2,301,310	2,716,745	3,120,337	3,505,025	3,930,716	4,374,350	549,019
Catalyst Replacement Period, years	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrocracking							
WHSV, 1/h	N/A	N/A	N/A	N/A	N/A	N/A	N/A
Catalyst Load, lb	N/A	N/A	N/A	N/A	N/A	N/A	N/A

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Case ID	11150011	11160011	11170011	11180011	11190011	11199011	11201011
Catalyst Replacement Period, years	N/A	N/A	N/A	N/A	N/A	N/A	N/A
TEA							
Capital							
Fixed Capital Investment (FCI), million $	388.4	422.0	453.1	481.5	511.2	537.4	214.6
Working Capital, million $	19.4	21.1	22.7	24.1	25.6	26.9	10.7
Operating Costs, $/GGE	3.04	3.16	3.29	3.42	3.55	3.67	2.41
Petroleum Feed Cost, $/GGE	1.17	0.94	0.70	0.47	0.24	0.00	2.33
CFP Oil Feed Cost, $/GGE	1.64	1.97	2.30	2.63	2.97	3.30	0.00
Net Hydrogen Makeup Cost, $/GGE	0.19	0.22	0.25	0.28	0.30	0.33	0.06
Fixed Operating Costs, $/GGE	0.02	0.02	0.02	0.02	0.02	0.02	0.01
Coproduct/Utility Credits, $/GGE	−0.04	−0.05	−0.05	−0.06	−0.06	−0.07	−0.02
Other Operating Costs, $/GGE	0.07	0.08	0.09	0.10	0.11	0.12	0.03
Capital Charges, Income Tax etc.	0.05	0.05	0.05	0.06	0.06	0.07	0.02
MFSP, with Full Capital Costs, $/GGE	3.08	3.21	3.34	3.47	3.61	3.74	2.43
MFSP, with Full Capital Costs, $/GJE	3.30	3.44	3.58	3.72	3.86	4.00	2.61
MFSP, with Full Capital Costs, $/GDE	3.41	3.56	3.70	3.84	3.99	4.14	2.69
Other TEA Metrics							
Total Operating Cost without Feed, cents/GGE	22.3	25.5	28.5	31.2	34.3	37.3	8.0
Hydrogen Cost, cents/GGE	19.2	22.2	25.0	27.5	30.3	33.1	5.5
Total Catalyst Cost per Feed Barrel, $/bbl	0.75	0.88	1.02	1.14	1.28	1.42	0.18
Delta Capital for Repurposed Unit, million $	178.1	212.4	244.2	273.4	303.7	332.9	0.0
Repurposed Case ID	11201011	11201011	11201011	11201011	11201011	11201011	11201011
MFSP with Repurposed Equipment, $/GGE	3.06	3.19	3.32	3.45	3.58	3.71	2.41
Repurposed Petroleum Unit Size, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
Repurposed Unit FCI, $/annual bbl fresh feed	6.8	8.2	9.4	10.5	11.7	12.8	0.0
New Unit FCI, $/annual bbl fresh feed	14.9	16.2	17.4	18.5	19.7	20.7	8.3
Hydrotreating reactor FCI, million $	51.8	56.6	61.2	65.3	69.8	74.1	28.7
Hydrocracking reactor FCI, million $	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Reactor(s) Pressure Correction On/Off	On	On	On	On	On	On	On
Makeup compressor FCI, million $	45.6	51.7	57.3	62.5	68.0	73.1	20.9
Recycle compressor FCI, million $	10.2	10.8	11.2	11.5	11.9	12.2	5.8
Fuel Lifecycle CI, g CO2e/MJ	52.5	45.1	37.7	30.2	22.8	15.2	90.8
CI for Petroleum Diesel Case, g CO2e/MJ	90.8	90.8	90.8	90.8	90.8	90.8	90.8
CI Reduction, %	42.2%	50.3%	58.4%	66.7%	74.9%	83.2%	0.0%

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Table K-3. Hydroprocessing Cases 11220011 to 12120811

Case ID	11220011	11220811	12101011	12110011	12120011	12120211	12120811
Total Feed, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
SRD Feed, BPD	60,000	35,053	75,000	67,500	60,000	48,000	29,718
LCO Feed, BPD	0	0	0	0	0	0	0
VGO Feed, BPD	0	0	0	0	0	0	0
CFP Oil Feed, BPD	15,000	8,763	0	7,500	15,000	12,000	7,430
Liquid Product Recycle, BPD	0	31,183	0	0	0	15,000	37,852
CFP Oil in Fresh Feed, vol %	20%	20%	0%	10%	20%	20%	20%
Liquid Product Recycle, vol %	0%	42%	0%	0%	0%	20%	50%
CFP Oil Oxygen (dry basis), wt %	16.0%	16.0%	21.6%	21.6%	21.6%	21.6%	21.6%
Total Liquid Fuel Product, lb/h	895,086	523,024	902,018	900,082	899,086	719,338	445,758
Total Liquid Fuel Product (all cuts), BGE/day	80,723	47,192	81,319	81,136	81,035	64,833	40,190
Total Liquid Fuel Product (all cuts), BJE/day	75,387	44,072	75,944	75,773	75,679	60,548	37,534
Total Liquid Fuel Product (all cuts), BDE/day	72,955	42,651	73,494	73,329	73,238	58,595	36,323
Total Liquid Product Molar H:C Ratio	1.79	1.80	1.79	1.79	1.79	1.79	1.80
Modeled Gasoline Product, wt %	24%	26%	23%	24%	24%	24%	24%
Modeled Jet Product, wt %	47%	48%	47%	47%	47%	47%	47%
Modeled Diesel/Heavies, wt %	29%	26%	30%	29%	29%	29%	29%
Hydroprocessing Reactor							
Inlet Pressure, psia	1,045	1,045	1,045	1,045	1,045	1,045	1,045
Inlet Temperature, F	644	644	644	644	644	644	644
Outlet Temperature, F	683	684	684	684	684	684	684
Carbon-Efficiency for CFP Oil, %	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%
Carbon-Efficiency for Petroleum Feed, %	98.2%	98.2%	98.8%	98.7%	98.6%	98.6%	98.5%
WHSV proration basis from SRD using chem. H2							
Base SRD Case H2 Consumption, scf/bbl	615.0	615.0	615.0	615.0	615.0	615.0	615.0
Base SRD Case WHSV Assumption, 1/h	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrogen							
Makeup H2 Purity, mol %	85%	85%	99%	99%	99%	99%	99%
H2 Purity at Reactor Inlet, mol %	80%	80%	91%	80%	80%	80%	80%
Chemical H2 per Total Feed, scf/bbl	1,280	777	614	1,134	1,630	1,316	834
Chemical H2 per Fresh Feed, scf/bbl	1,280	1,331	614	1,134	1,630	1,645	1,684
Reactor H2 Feed per Total Feed, scf/bbl	7,200	2,960	2,832	6,769	11,066	6,448	3,170
Reactor H2 Feed per Fresh Feed, scf/bbl	7,200	5,067	2,832	6,769	11,066	8,060	6,399
Makeup H2 per Total Feed, scf/bbl	2,931	1,441	660	1,299	2,004	1,572	940
Makeup H2 per Fresh Feed, scf/bbl	2,931	2,467	660	1,299	2,004	1,965	1,897
Net H2 after PSA, per Total Feed, scf/bbl	1,568	917	660	1,194	1,721	1,389	884
Hydrotreating							
WHSV, 1/h	0.96	0.92	2.00	1.09	0.75	0.75	0.73
Catalyst Load, lb	1,207,687	733,281	551,022	1,055,257	1,571,320	1,268,916	804,380
Catalyst Replacement Period, years	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrocracking							
WHSV, 1/h	N/A	N/A	N/A	N/A	N/A	N/A	N/A
Catalyst Load, lb	N/A	N/A	N/A	N/A	N/A	N/A	N/A

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Case ID	11220011	11220811	12101011	12110011	12120011	12120211	12120811
Catalyst Replacement Period, years	N/A	N/A	N/A	N/A	N/A	N/A	N/A
TEA							
Capital							
Fixed Capital Investment (FCI), million $	330.0	238.8	196.4	262.5	317.5	275.7	211.8
Working Capital, million $	16.5	11.9	9.8	13.1	15.9	13.8	10.6
Operating Costs, $/GGE	2.69	2.69	2.40	2.69	2.99	2.99	2.99
Petroleum Feed Cost, $/GGE	1.87	1.87	2.33	2.10	1.87	1.87	1.86
CFP Oil Feed Cost, $/GGE	0.66	0.66	0.00	0.47	0.95	0.95	0.95
Net Hydrogen Makeup Cost, $/GGE	0.13	0.13	0.05	0.10	0.14	0.14	0.14
Fixed Operating Costs, $/GGE	0.01	0.02	0.01	0.01	0.01	0.02	0.02
Coproduct/Utility Credits, $/GGE	−0.03	−0.03	−0.02	−0.02	−0.03	−0.03	−0.03
Other Operating Costs, $/GGE	0.06	0.05	0.03	0.05	0.06	0.06	0.06
Capital Charges, Income Tax etc.	0.04	0.05	0.02	0.03	0.04	0.04	0.05
MFSP, with Full Capital Costs, $/GGE	2.72	2.74	2.42	2.73	3.02	3.03	3.04
MFSP, with Full Capital Costs, $/GJE	2.92	2.93	2.59	2.92	3.24	3.24	3.26
MFSP, with Full Capital Costs, $/GDE	3.01	3.03	2.68	3.02	3.35	3.35	3.37
Other TEA Metrics							
Total Operating Cost without Feed, cents/GGE	15.6	16.0	7.6	12.3	17.1	17.2	18.1
Hydrogen Cost, cents/GGE	12.6	12.6	5.2	9.5	13.7	13.8	14.2
Total Catalyst Cost per Feed Barrel, $/bbl	0.39	0.41	0.18	0.34	0.51	0.52	0.53
Delta Capital for Repurposed Unit, million $	116.5	50.0	0.4	52.5	104.2	73.0	35.8
Repurposed Case ID	11201011	11201011	12201011	12201011	12201011	12201011	12201011
MFSP with Repurposed Equipment, $/GGE	2.70	2.70	2.40	2.70	3.00	3.00	3.00
Repurposed Petroleum Unit Size, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
Repurposed Unit FCI, $/annual bbl fresh feed	4.5	3.3	0.0	2.0	4.0	3.5	2.8
New Unit FCI, $/annual bbl fresh feed	12.7	15.7	7.6	10.1	12.2	13.3	16.4
Hydrotreating reactor FCI, million $	38.9	30.8	27.4	35.8	43.0	35.8	30.3
Hydrocracking reactor FCI, million $	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Reactor(s) Pressure Correction On/Off	On	On	On	On	On	On	On
Makeup compressor FCI, million $	52.6	29.9	14.4	24.6	34.8	28.6	19.0
Recycle compressor FCI, million $	8.3	6.3	6.0	8.1	9.4	8.0	6.4
Fuel Lifecycle CI, g CO2e/MJ	76.5	76.1	90.6	83.3	76.1	76.0	76.1
CI for Petroleum Diesel Case, g CO2e/MJ	90.8	90.8	90.8	90.8	90.8	90.8	90.8
CI Reduction, %	15.7%	16.2%	0.2%	8.2%	16.1%	16.3%	16.1%

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Table K-4. Hydroprocessing Cases 12130011 to 12190011

Case ID	12130011	12140011	12150011	12160011	12170011	12180011	12190011
Total Feed, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
SRD Feed, BPD	52,500	45,000	37,500	30,000	22,500	15,000	7,500
LCO Feed, BPD	0	0	0	0	0	0	0
VGO Feed, BPD	0	0	0	0	0	0	0
CFP Oil Feed, BPD	22,500	30,000	37,500	45,000	52,500	60,000	67,500
Liquid Product Recycle, BPD	0	0	0	0	0	0	0
CFP Oil in Fresh Feed, vol %	30%	40%	50%	60%	70%	80%	90%
Liquid Product Recycle, vol %	0%	0%	0%	0%	0%	0%	0%
CFP Oil Oxygen (dry basis), wt %	21.6%	21.6%	21.6%	21.6%	21.6%	21.6%	21.6%
Total Liquid Fuel Product, lb/h	898,545	898,463	898,537	898,838	899,254	899,796	900,357
Total Liquid Fuel Product (all cuts), BGE/day	80,986	80,984	80,999	81,035	81,082	81,141	81,202
Total Liquid Fuel Product (all cuts), BJE/day	75,632	75,631	75,645	75,678	75,722	75,777	75,834
Total Liquid Fuel Product (all cuts), BDE/day	73,193	73,192	73,205	73,237	73,280	73,333	73,388
Total Liquid Product Molar H:C Ratio	1.79	1.79	1.79	1.79	1.79	1.80	1.80
Modeled Gasoline Product, wt %	24%	25%	25%	26%	26%	27%	27%
Modeled Jet Product, wt %	47%	47%	47%	47%	47%	47%	46%
Modeled Diesel/Heavies, wt %	28%	28%	27%	27%	27%	27%	27%
Hydroprocessing Reactor							
Inlet Pressure, psia	1,045	1,045	1,045	1,045	1,045	1,045	1,045
Inlet Temperature, F	644	644	644	644	644	644	644
Outlet Temperature, F	684	684	684	684	685	684	684
Carbon-Efficiency for CFP Oil, %	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%
Carbon-Efficiency for Petroleum Feed, %	98.6%	98.5%	98.5%	98.5%	98.4%	98.4%	98.3%
WHSV proration basis from SRD using chem. H2							
Base SRD Case H2 Consumption, scf/bbl	615.0	615.0	615.0	615.0	615.0	615.0	615.0
Base SRD Case WHSV Assumption, 1/h	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrogen							
Makeup H2 Purity, mol %	99%	99%	99%	99%	99%	99%	99%
H2 Purity at Reactor Inlet, mol %	80%	80%	80%	80%	80%	80%	80%
Chemical H2 per Total Feed, scf/bbl	2,107	2,606	3,104	3,599	4,091	4,581	5,069
Chemical H2 per Fresh Feed, scf/bbl	2,107	2,606	3,104	3,599	4,091	4,581	5,069
Reactor H2 Feed per Total Feed, scf/bbl	15,085	19,132	23,095	27,039	30,876	34,860	38,711
Reactor H2 Feed per Fresh Feed, scf/bbl	15,085	19,132	23,095	27,039	30,876	34,860	38,711
Makeup H2 per Total Feed, scf/bbl	2,666	3,366	4,061	4,747	5,426	6,103	6,772
Makeup H2 per Fresh Feed, scf/bbl	2,666	3,366	4,061	4,747	5,426	6,103	6,772
Net H2 after PSA, per Total Feed, scf/bbl	2,226	2,756	3,283	3,807	4,328	4,846	5,360
Hydrotreating							
WHSV, 1/h	0.58	0.47	0.40	0.34	0.30	0.27	0.24
Catalyst Load, lb	2,101,138	2,685,706	3,301,923	3,948,083	4,624,401	5,330,232	6,065,400
Catalyst Replacement Period, years	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrocracking							
WHSV, 1/h	N/A	N/A	N/A	N/A	N/A	N/A	N/A
Catalyst Load, lb	N/A	N/A	N/A	N/A	N/A	N/A	N/A

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Case ID	12130011	12140011	12150011	12160011	12170011	12180011	12190011
Catalyst Replacement Period, years	N/A	N/A	N/A	N/A	N/A	N/A	N/A
TEA							
Capital							
Fixed Capital Investment (FCI), million $	365.1	413.9	458.0	500.3	541.0	580.7	618.9
Working Capital, million $	18.3	20.7	22.9	25.0	27.1	29.0	30.9
Operating Costs, $/GGE	3.28	3.57	3.86	4.14	4.43	4.72	5.00
Petroleum Feed Cost, $/GGE	1.63	1.40	1.17	0.93	0.70	0.47	0.23
CFP Oil Feed Cost, $/GGE	1.43	1.90	2.38	2.85	3.32	3.80	4.27
Net Hydrogen Makeup Cost, $/GGE	0.18	0.22	0.26	0.30	0.34	0.39	0.43
Fixed Operating Costs, $/GGE	0.02	0.02	0.02	0.02	0.02	0.02	0.02
Coproduct/Utility Credits, $/GGE	−0.04	−0.04	−0.05	−0.05	−0.06	−0.06	−0.07
Other Operating Costs, $/GGE	0.07	0.09	0.10	0.12	0.13	0.15	0.17
Capital Charges, Income Tax etc.	0.04	0.05	0.05	0.06	0.07	0.07	0.08
MFSP, with Full Capital Costs, $/GGE	3.32	3.62	3.91	4.20	4.50	4.79	5.08
MFSP, with Full Capital Costs, $/GJE	3.56	3.87	4.19	4.50	4.82	5.13	5.44
MFSP, with Full Capital Costs, $/GDE	3.67	4.00	4.33	4.65	4.98	5.30	5.62
Other TEA Metrics							
Total Operating Cost without Feed, cents/GGE	21.7	26.5	31.3	36.1	40.9	45.6	50.4
Hydrogen Cost, cents/GGE	17.8	22.0	26.2	30.4	34.5	38.6	42.7
Total Catalyst Cost per Feed Barrel, $/bbl	0.68	0.87	1.07	1.29	1.51	1.73	1.97
Delta Capital for Repurposed Unit, million $	152.3	201.8	246.5	289.6	331.1	371.7	411.1
Repurposed Case ID	12201011	12201011	12201011	12201011	12201011	12201011	12201011
MFSP with Repurposed Equipment, $/GGE	3.30	3.59	3.89	4.18	4.47	4.77	5.06
Repurposed Petroleum Unit Size, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
Repurposed Unit FCI, $/annual bbl fresh feed	5.9	7.8	9.5	11.1	12.7	14.3	15.8
New Unit FCI, $/annual bbl fresh feed	14.0	15.9	17.6	19.2	20.8	22.3	23.8
Hydrotreating reactor FCI, million $	49.8	56.5	63.0	69.4	75.7	82.1	88.2
Hydrocracking reactor FCI, million $	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Reactor(s) Pressure Correction On/Off	On	On	On	On	On	On	On
Makeup compressor FCI, million $	43.6	52.5	61.0	69.1	76.8	84.4	91.7
Recycle compressor FCI, million $	10.3	11.1	11.7	12.3	12.8	13.3	13.7
Fuel Lifecycle CI, g CO2e/MJ	69.0	61.9	54.8	47.7	40.6	33.5	26.4
CI for Petroleum Diesel Case, g CO2e/MJ	90.8	90.8	90.8	90.8	90.8	90.8	90.8
CI Reduction, %	24.0%	31.8%	39.7%	47.5%	55.3%	63.1%	70.9%

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Table K-5. Hydroprocessing Cases 12199011 to 12199513.

Changes in "Base" reflect the different cases used for a base case for calculations related to repurposing existing equipment.

Case ID	12199011	Base: 30k BPD 12199313	Base: 10k BPD 12199313	12199313	Base: 30k BPD 12199513	Base: 10k BPD 12199513	12199513
Total Feed, BPD	75,000	10,000	10,000	10,000	10,000	10,000	10,000
SRD Feed, BPD	0	0	0	0	0	0	0
LCO Feed, BPD	0	0	0	0	0	0	0
VGO Feed, BPD	0	0	0	0	0	0	0
CFP Oil Feed, BPD	75,000	7,000	7,000	7,000	5,000	5,000	5,000
Liquid Product Recycle, BPD	0	3,000	3,000	3,000	5,000	5,000	5,000
CFP Oil in Fresh Feed, vol %	100%	100%	100%	100%	100%	100%	100%
Liquid Product Recycle, vol %	0%	30%	30%	30%	50%	50%	50%
CFP Oil Oxygen (dry basis), wt %	21.6%	21.6%	21.6%	21.6%	21.6%	21.6%	21.6%
Total Liquid Fuel Product, lb/h	901,040	84,175	84,175	84,175	60,157	60,157	60,157
Total Liquid Fuel Product (all cuts), BGE/day	81,274	7,592	7,592	7,592	5,425	5,425	5,425
Total Liquid Fuel Product (all cuts), BJE/day	75,901	7,090	7,090	7,090	5,067	5,067	5,067
Total Liquid Fuel Product (all cuts), BDE/day	73,453	6,861	6,861	6,861	4,903	4,903	4,903
Total Liquid Product Molar H:C Ratio	1.80	1.80	1.80	1.80	1.80	1.80	1.80
Modeled Gasoline Product, wt %	27%	27%	27%	27%	27%	27%	27%
Modeled Jet Product, wt %	45%	45%	45%	45%	45%	45%	45%
Modeled Diesel/Heavies, wt %	27%	27%	27%	27%	27%	27%	27%
Hydroprocessing Reactor							
Inlet Pressure, psia	1,045	1,045	1,045	1,045	1,045	1,045	1,045
Inlet Temperature, F	644	644	644	644	644	644	644
Outlet Temperature, F	683	684	684	684	684	684	684
Carbon-Efficiency for CFP Oil, %	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%
Carbon-Efficiency for Petroleum Feed, %	98.2%	97.5%	97.5%	97.5%	234.0%	234.0%	234.0%
WHSV proration basis from SRD using chem. H2							
Base SRD Case H2 Consumption, scf/bbl	615.0	615.0	615.0	615.0	615.0	615.0	615.0
Base SRD Case WHSV Assumption, 1/h	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrogen							
Makeup H2 Purity, mol %	99%	99%	99%	99%	99%	99%	99%
H2 Purity at Reactor Inlet, mol %	80%	80%	80%	80%	80%	80%	80%
Chemical H2 per Total Feed, scf/bbl	5,553	3,969	3,969	3,969	2,873	2,873	2,873
Chemical H2 per Fresh Feed, scf/bbl	5,553	5,670	5,670	5,670	5,745	5,745	5,745
Reactor H2 Feed per Total Feed, scf/bbl	42,665	26,767	26,767	26,767	15,913	15,913	15,913
Reactor H2 Feed per Fresh Feed, scf/bbl	42,665	38,238	38,238	38,238	31,827	31,827	31,827
Makeup H2 per Total Feed, scf/bbl	7,433	5,276	5,276	5,276	3,742	3,742	3,742
Makeup H2 per Fresh Feed, scf/bbl	7,433	7,537	7,537	7,537	7,483	7,483	7,483
Net H2 after PSA, per Total Feed, scf/bbl	5,871	4,201	4,201	4,201	3,038	3,038	3,038
Hydrotreating							
WHSV, 1/h	0.22	0.22	0.22	0.22	0.21	0.21	0.21
Catalyst Load, lb	6,829,318	650,884	650,884	650,884	471,045	471,045	471,045

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Case ID	12199011	Base: 30k BPD 12199313	Base: 10k BPD 12199313	12199313	Base: 30k BPD 12199513	Base: 10k BPD 12199513	12199513
Catalyst Replacement Period, years	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrocracking							
WHSV, 1/h	N/A	N/A	N/A	N/A	N/A	N/A	N/A
Catalyst Load, lb	N/A	N/A	N/A	N/A	N/A	N/A	N/A
Catalyst Replacement Period, years	N/A	N/A	N/A	N/A	N/A	N/A	N/A
TEA							
Capital							
Fixed Capital Investment (FCI), million $	654.5	133.0	133.0	133.0	107.9	107.9	107.9
Working Capital, million $	32.7	6.7	6.7	6.7	5.4	5.4	5.4
Operating Costs, $/GGE	5.29	5.36	5.36	5.36	5.38	5.38	5.38
Petroleum Feed Cost, $/GGE	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CFP Oil Feed Cost, $/GGE	4.74	4.73	4.73	4.73	4.73	4.73	4.73
Net Hydrogen Makeup Cost, $/GGE	0.47	0.48	0.48	0.48	0.48	0.48	0.48
Fixed Operating Costs, $/GGE	0.02	0.09	0.09	0.09	0.11	0.11	0.11
Coproduct/Utility Credits, $/GGE	−0.08	−0.08	−0.08	−0.08	−0.08	−0.08	−0.08
Other Operating Costs, $/GGE	0.18	0.19	0.19	0.19	0.18	0.18	0.18
Capital Charges, Income Tax etc.	0.08	0.17	0.17	0.17	0.19	0.19	0.19
MFSP, with Full Capital Costs, $/GGE	5.37	5.52	5.52	5.52	5.57	5.57	5.57
MFSP, with Full Capital Costs, $/GJE	5.75	5.91	5.91	5.91	5.96	5.96	5.96
MFSP, with Full Capital Costs, $/GDE	5.94	6.11	6.11	6.11	6.16	6.16	6.16
Other TEA Metrics							
Total Operating Cost without Feed, cents/GGE	55.0	62.2	62.2	62.2	64.7	64.7	64.7
Hydrogen Cost, cents/GGE	46.7	47.7	47.7	47.7	48.2	48.2	48.2
Total Catalyst Cost per Feed Barrel, $/bbl	2.22	2.27	2.27	2.27	2.30	2.30	2.30
Delta Capital for Repurposed Unit, million $	449.6	49.8	77.4	29.0	33.0	55.8	20.3
Repurposed Case ID	12201011	12201012	12201013	12201011	12201012	12201013	12201011
MFSP with Repurposed Equipment, $/GGE	5.34	5.42	5.46	5.40	5.44	5.48	5.42
Repurposed Petroleum Unit Size, BPD	75,000	30,000	10,000	75,000	30,000	10,000	75,000
Repurposed Unit FCI, $/annual bbl fresh feed	17.3	20.5	31.9	12.0	19.0	32.2	11.7
New Unit FCI, $/annual bbl fresh feed	25.2	54.8	54.8	54.8	62.2	62.2	62.2
Hydrotreating reactor FCI, million $	94.6	8.8	8.8	8.8	6.1	6.1	6.1
Hydrocracking reactor FCI, million $	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Reactor(s) Pressure Correction On/Off	On	On	On	On	On	On	On
Makeup compressor FCI, million $	98.7	15.1	15.1	15.1	11.5	11.5	11.5
Recycle compressor FCI, million $	14.2	6.7	6.7	6.7	5.7	5.7	5.7
Fuel Lifecycle CI, g CO2e/MJ	19.4	19.3	19.3	19.3	19.1	19.1	19.1
CI for Petroleum Diesel Case, g CO2e/MJ	90.8	90.8	90.8	90.8	90.8	90.8	90.8
CI Reduction, %	78.6%	78.7%	78.7%	78.7%	79.0%	79.0%	79.0%

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Table K-6. Hydroprocessing Cases 12199811 to 12220011

Case ID	SCALEx0.13 12199811	12199811	12201011	12210811	12210821	12210851	12220011
Total Feed, BPD	10,000	75,000	75,000	75,000	75,000	75,000	75,000
SRD Feed, BPD	0	0	75,000	40,803	40,802	40,604	60,000
LCO Feed, BPD	0	0	0	0	0	0	0
VGO Feed, BPD	0	0	0	0	0	0	0
CFP Oil Feed, BPD	2,165	16,235	0	4,534	4,534	4,512	15,000
Liquid Product Recycle, BPD	7,835	58,765	0	29,664	29,665	29,884	0
CFP Oil in Fresh Feed, vol %	100%	100%	0%	10%	10%	10%	20%
Liquid Product Recycle, vol %	78%	78%	0%	40%	40%	40%	0%
CFP Oil Oxygen (dry basis), wt %	21.6%	21.6%	21.6%	21.6%	21.6%	21.6%	21.6%
Total Liquid Fuel Product, lb/h	26,096	195,722	899,990	543,297	543,323	540,672	897,075
Total Liquid Fuel Product (all cuts), BGE/day	2,352	17,637	81,151	49,009	49,010	48,767	80,889
Total Liquid Fuel Product (all cuts), BJE/day	2,196	16,472	75,787	45,770	45,770	45,543	75,542
Total Liquid Fuel Product (all cuts), BDE/day	2,125	15,940	73,343	44,293	44,294	44,074	73,106
Total Liquid Product Molar H:C Ratio	1.79	1.79	1.79	1.80	1.80	1.80	1.79
Modeled Gasoline Product, wt %	27%	27%	23%	24%	24%	24%	24%
Modeled Jet Product, wt %	45%	45%	47%	47%	47%	47%	47%
Modeled Diesel/Heavies, wt %	28%	28%	30%	29%	29%	29%	29%
Hydroprocessing Reactor							
Inlet Pressure, psia	1,045	1,045	1,045	1,045	1,045	1,045	1,045
Inlet Temperature, F	644	644	644	644	644	644	644
Outlet Temperature, F	684	684	684	684	684	684	684
Carbon-Efficiency for CFP Oil, %	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%
Carbon-Efficiency for Petroleum Feed, %	97.8%	97.8%	98.6%	98.3%	98.3%	98.3%	98.1%
WHSV proration basis from SRD using chem. H2							
Base SRD Case H2 Consumption, scf/bbl	615.0	615.0	615.0	615.0	615.0	615.0	615.0
Base SRD Case WHSV Assumption, 1/h	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrogen							
Makeup H2 Purity, mol %	99%	99%	85%	85%	85%	85%	85%
H2 Purity at Reactor Inlet, mol %	80%	80%	80%	80%	75%	70%	80%
Chemical H2 per Total Feed, scf/bbl	1,256	1,256	612	703	703	699	1,622
Chemical H2 per Fresh Feed, scf/bbl	5,804	5,804	612	1,163	1,164	1,161	1,622
Reactor H2 Feed per Total Feed, scf/bbl	4,800	4,800	2,256	2,681	2,682	2,663	10,266
Reactor H2 Feed per Fresh Feed, scf/bbl	22,173	22,173	2,256	4,435	4,437	4,427	10,266
Makeup H2 per Total Feed, scf/bbl	1,524	1,524	920	1,272	985	858	4,005
Makeup H2 per Fresh Feed, scf/bbl	7,040	7,040	920	2,104	1,630	1,427	4,005
Net H2 after PSA, per Total Feed, scf/bbl	1,329	1,329	696	826	781	756	2,020
Hydrotreating							
WHSV, 1/h	0.21	0.21	2.01	1.06	1.06	1.06	0.76
Catalyst Load, lb	206,022	1,545,168	549,791	654,482	654,734	650,228	1,563,345
Catalyst Replacement Period, years	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrocracking							
WHSV, 1/h	N/A	N/A	N/A	N/A	N/A	N/A	N/A

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Case ID	SCALEx0.13 12199811	12199811	12201011	12210811	12210821	12210851	12220011
Catalyst Load, lb	N/A	N/A	N/A	N/A	N/A	N/A	N/A
Catalyst Replacement Period, years	N/A	N/A	N/A	N/A	N/A	N/A	N/A
TEA							
Capital							
Fixed Capital Investment (FCI), million $	71.5	236.0	214.6	225.9	218.0	214.4	372.6
Working Capital, million $	3.6	11.8	10.7	11.3	10.9	10.7	18.6
Operating Costs, $/GGE	5.48	5.31	2.41	2.72	2.71	2.71	3.02
Petroleum Feed Cost, $/GGE	0.00	0.00	2.33	2.10	2.10	2.10	1.87
CFP Oil Feed Cost, $/GGE	4.73	4.73	0.00	0.47	0.47	0.47	0.95
Net Hydrogen Makeup Cost, $/GGE	0.49	0.49	0.06	0.11	0.10	0.10	0.16
Fixed Operating Costs, $/GGE	0.22	0.05	0.01	0.02	0.02	0.02	0.02
Coproduct/Utility Credits, $/GGE	−0.08	−0.08	−0.02	−0.03	−0.03	−0.02	−0.04
Other Operating Costs, $/GGE	0.18	0.18	0.03	0.05	0.05	0.05	0.07
Capital Charges, Income Tax etc.	0.29	0.13	0.02	0.04	0.04	0.04	0.04
MFSP, with Full Capital Costs, $/GGE	5.77	5.44	2.43	2.76	2.75	2.75	3.07
MFSP, with Full Capital Costs, $/GJE	6.18	5.82	2.61	2.96	2.95	2.94	3.28
MFSP, with Full Capital Costs, $/GDE	6.38	6.02	2.69	3.06	3.05	3.04	3.39
Other TEA Metrics							
Total Operating Cost without Feed, cents/GGE	75.7	58.5	8.0	14.5	13.8	13.5	20.2
Hydrogen Cost, cents/GGE	48.7	48.7	5.5	10.9	10.3	10.0	16.1
Total Catalyst Cost per Feed Barrel, $/bbl	2.32	2.32	0.18	0.35	0.35	0.35	0.51
Delta Capital for Repurposed Unit, million $	24.5	78.7	0.0	35.9	27.7	27.8	159.1
Repurposed Case ID	12201011	12201011	12201011	12201011	12201011	12201011	12201011
MFSP with Repurposed Equipment, $/GGE	5.59	5.36	2.41	2.73	2.72	2.72	3.04
Repurposed Petroleum Unit Size, BPD	10,000	75,000	75,000	75,000	75,000	75,000	75,000
Repurposed Unit FCI, $/annual bbl fresh feed	32.6	14.0	0.0	2.3	1.8	1.8	6.1
New Unit FCI, $/annual bbl fresh feed	95.3	41.9	8.3	14.4	13.9	13.7	14.3
Hydrotreating reactor FCI, million $	5.8	33.7	28.7	30.1	30.9	31.9	44.7
Hydrocracking reactor FCI, million $	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Reactor(s) Pressure Correction On/Off	Off	On	On	On	On	On	On
Makeup compressor FCI, million $	5.6	28.0	20.9	27.0	22.1	19.8	67.4
Recycle compressor FCI, million $	4.0	7.3	5.8	6.1	6.2	6.3	9.2
Fuel Lifecycle CI, g CO2e/MJ	19.1	18.9	90.8	84.1	83.8	83.6	77.7
CI for Petroleum Diesel Case, g CO2e/MJ	90.8	90.8	90.8	90.8	90.8	90.8	90.8
CI Reduction, %	78.9%	79.2%	0.0%	7.4%	7.7%	7.9%	14.4%

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Table K-7. Hydroprocessing Cases 12220021 to 41120811

Case ID	12220021	12320211	41101011	41110011	41120011	41120211	41120811
Total Feed, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
SRD Feed, BPD	60,000	48,000	0	0	0	0	0
LCO Feed, BPD	0	0	37,500	33,750	30,000	24,000	12,680
VGO Feed, BPD	0	0	37,500	33,750	30,000	24,000	12,680
CFP Oil Feed, BPD	15,000	12,000	0	7,500	15,000	12,000	6,340
Liquid Product Recycle, BPD	0	15,000	0	0	0	15,000	43,300
CFP Oil in Fresh Feed, vol %	20%	20%	0%	10%	20%	20%	20%
Liquid Product Recycle, vol %	0%	20%	0%	0%	0%	20%	58%
CFP Oil Oxygen (dry basis), wt %	21.6%	21.6%	16.0%	16.0%	16.0%	16.0%	16.0%
Total Liquid Fuel Product, lb/h	896,945	719,359	1,009,195	995,517	983,473	787,456	416,600
Total Liquid Fuel Product (all cuts), BGE/day	80,878	64,834	91,046	89,818	88,715	71,059	37,594
Total Liquid Fuel Product (all cuts), BJE/day	75,531	60,548	85,028	83,881	82,851	66,362	35,109
Total Liquid Fuel Product (all cuts), BDE/day	73,095	58,595	82,285	81,175	80,178	64,221	33,977
Total Liquid Product Molar H:C Ratio	1.79	1.79	1.79	1.79	1.79	1.80	1.80
Modeled Gasoline Product, wt %	24%	24%	12%	14%	17%	18%	18%
Modeled Jet Product, wt %	47%	47%	33%	33%	34%	34%	34%
Modeled Diesel/Heavies, wt %	29%	29%	56%	52%	49%	49%	48%
Hydroprocessing Reactor							
Inlet Pressure, psia	1,045	1,045	2,045	2,045	2,045	2,045	2,045
Inlet Temperature, F	644	644	644	644	644	644	644
Outlet Temperature, F	685	684	684	684	684	684	684
Carbon-Efficiency for CFP Oil, %	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%
Carbon-Efficiency for Petroleum Feed, %	98.3%	98.6%	98.4%	98.3%	98.2%	98.2%	97.9%
WHSV proration basis from SRD using chem. H2							
Base SRD Case H2 Consumption, scf/bbl	615.0	615.0	615.0	615.0	615.0	615.0	615.0
Base SRD Case WHSV Assumption, 1/h	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrogen							
Makeup H2 Purity, mol %	85%	99%	99%	99%	99%	99%	99%
H2 Purity at Reactor Inlet, mol %	75%	80%	92%	86%	82%	84%	91%
Chemical H2 per Total Feed, scf/bbl	1,626	1,316	2,035	2,246	2,418	1,964	1,048
Chemical H2 per Fresh Feed, scf/bbl	1,626	1,645	2,035	2,246	2,418	2,454	2,481
Reactor H2 Feed per Total Feed, scf/bbl	9,340	11,014	14,812	15,019	15,187	10,764	4,627
Reactor H2 Feed per Fresh Feed, scf/bbl	9,340	13,767	14,812	15,019	15,187	13,455	10,947
Makeup H2 per Total Feed, scf/bbl	2,840	1,619	2,134	2,345	2,517	2,060	1,146
Makeup H2 per Fresh Feed, scf/bbl	2,840	2,024	2,134	2,345	2,517	2,576	2,711
Net H2 after PSA, per Total Feed, scf/bbl	1,847	1,396	2,133	2,344	2,516	2,060	1,146
Hydrotreating							
WHSV, 1/h	0.76	0.75	1.21	1.10	1.02	1.00	0.99
Catalyst Load, lb	1,567,847	1,268,572	1,027,527	1,146,962	1,248,561	1,014,088	541,484
Catalyst Replacement Period, years	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrocracking							
WHSV, 1/h	N/A	N/A	1.21	1.10	1.02	1.00	0.99
Catalyst Load, lb	N/A	N/A	1,027,527	1,146,962	1,248,561	1,014,088	541,484

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Case ID	12220021	12320211	41101011	41110011	41120011	41120211	41120811
Catalyst Replacement Period, years	N/A	N/A	3.0	3.0	3.0	3.0	3.0
TEA							
Capital							
Fixed Capital Investment (FCI), million $	349.6	289.2	462.2	508.7	537.5	483.6	366.1
Working Capital, million $	17.5	14.5	23.1	25.4	26.9	24.2	18.3
Operating Costs, $/GGE	3.01	2.99	2.19	2.33	2.46	2.46	2.48
Petroleum Feed Cost, $/GGE	1.87	1.87	2.00	1.82	1.64	1.64	1.64
CFP Oil Feed Cost, $/GGE	0.95	0.95	0.00	0.29	0.60	0.60	0.60
Net Hydrogen Makeup Cost, $/GGE	0.15	0.14	0.15	0.17	0.18	0.19	0.20
Fixed Operating Costs, $/GGE	0.02	0.02	0.02	0.02	0.02	0.02	0.03
Coproduct/Utility Credits, $/GGE	−0.03	−0.03	−0.02	−0.03	−0.03	−0.03	−0.03
Other Operating Costs, $/GGE	0.07	0.06	0.06	0.06	0.07	0.07	0.07
Capital Charges, Income Tax etc.	0.04	0.04	0.05	0.05	0.06	0.06	0.09
MFSP, with Full Capital Costs, $/GGE	3.05	3.04	2.24	2.38	2.52	2.53	2.57
MFSP, with Full Capital Costs, $/GJE	3.26	3.25	2.40	2.55	2.70	2.71	2.76
MFSP, with Full Capital Costs, $/GDE	3.37	3.36	2.48	2.63	2.79	2.80	2.85
Other TEA Metrics							
Total Operating Cost without Feed, cents/GGE	18.4	17.8	19.2	20.9	22.4	23.0	25.3
Hydrogen Cost, cents/GGE	14.8	13.9	15.1	16.9	18.3	18.7	19.7
Total Catalyst Cost per Feed Barrel, $/bbl	0.51	0.52	0.56	0.62	0.68	0.69	0.70
Delta Capital for Repurposed Unit, million $	136.1	85.7	0.8	22.2	57.5	43.6	39.0
Repurposed Case ID	12201011	12201011	41201011	41201011	41201011	41201011	41201011
MFSP with Repurposed Equipment, $/GGE	3.02	3.01	2.19	2.33	2.47	2.47	2.50
Repurposed Petroleum Unit Size, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
Repurposed Unit FCI, $/annual bbl fresh feed	5.2	4.1	0.0	0.9	2.2	2.1	3.6
New Unit FCI, $/annual bbl fresh feed	13.4	13.9	17.8	19.6	20.7	23.2	33.3
Hydrotreating reactor FCI, million $	46.3	42.5	63.2	68.5	72.9	64.3	54.8
Hydrocracking reactor FCI, million $	0.0	0.0	63.2	68.5	72.9	64.3	54.8
Reactor(s) Pressure Correction On/Off	On	On	On	On	On	On	On
Makeup compressor FCI, million $	51.3	29.3	51.1	55.1	58.3	49.7	31.2
Recycle compressor FCI, million $	9.1	9.4	8.1	8.3	8.4	7.5	5.7
Fuel Lifecycle CI, g CO2e/MJ	76.8	76.4	95.2	87.9	80.3	80.5	81.0
CI for Petroleum Diesel Case, g CO2e/MJ	90.8	90.8	90.8	90.8	90.8	90.8	90.8
CI Reduction, %	15.4%	15.8%	-4.9%	3.2%	11.5%	11.4%	10.8%

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Table K-8. Hydroprocessing Cases 41130011 to 41190011

Case ID	41130011	41140011	41150011	41160011	41170011	41180011	41190011
Total Feed, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
SRD Feed, BPD	0	0	0	0	0	0	0
LCO Feed, BPD	26,250	22,500	18,750	15,000	11,250	7,500	3,750
VGO Feed, BPD	26,250	22,500	18,750	15,000	11,250	7,500	3,750
CFP Oil Feed, BPD	22,500	30,000	37,500	45,000	52,500	60,000	67,500
Liquid Product Recycle, BPD	0	0	0	0	0	0	0
CFP Oil in Fresh Feed, vol %	30%	40%	50%	60%	70%	80%	90%
Liquid Product Recycle, vol %	0%	0%	0%	0%	0%	0%	0%
CFP Oil Oxygen (dry basis), wt %	16.0%	16.0%	16.0%	16.0%	16.0%	16.0%	16.0%
Total Liquid Fuel Product, lb/h	972,348	960,691	949,496	938,388	927,315	916,212	904,378
Total Liquid Fuel Product (all cuts), BGE/day	87,746	86,668	85,673	84,703	83,742	82,747	81,668
Total Liquid Fuel Product (all cuts), BJE/day	81,946	80,939	80,010	79,104	78,206	77,277	76,270
Total Liquid Fuel Product (all cuts), BDE/day	79,303	78,328	77,430	76,552	75,684	74,785	73,810
Total Liquid Product Molar H:C Ratio	1.80	1.79	1.79	1.80	1.80	1.80	1.80
Modeled Gasoline Product, wt %	20%	23%	26%	29%	32%	35%	38%
Modeled Jet Product, wt %	35%	35%	36%	37%	38%	38%	40%
Modeled Diesel/Heavies, wt %	45%	42%	39%	35%	31%	27%	22%
Hydroprocessing Reactor							
Inlet Pressure, psia	2,045	2,045	2,045	2,045	2,045	2,045	2,045
Inlet Temperature, F	644	644	644	644	644	644	644
Outlet Temperature, F	683	684	684	684	684	683	683
Carbon-Efficiency for CFP Oil, %	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%
Carbon-Efficiency for Petroleum Feed, %	98.2%	98.1%	98.1%	98.1%	98.1%	98.0%	98.0%
WHSV proration basis from SRD using chem. H2							
Base SRD Case H2 Consumption, scf/bbl	615.0	615.0	615.0	615.0	615.0	615.0	615.0
Base SRD Case WHSV Assumption, 1/h	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrogen							
Makeup H2 Purity, mol %	99%	99%	99%	99%	99%	99%	99%
H2 Purity at Reactor Inlet, mol %	80%	80%	80%	80%	80%	80%	80%
Chemical H2 per Total Feed, scf/bbl	2,594	2,754	2,937	3,129	3,324	3,499	3,666
Chemical H2 per Fresh Feed, scf/bbl	2,594	2,754	2,937	3,129	3,324	3,499	3,666
Reactor H2 Feed per Total Feed, scf/bbl	16,185	17,397	18,827	20,319	21,882	23,299	24,667
Reactor H2 Feed per Fresh Feed, scf/bbl	16,185	17,397	18,827	20,319	21,882	23,299	24,667
Makeup H2 per Total Feed, scf/bbl	2,726	2,981	3,262	3,554	3,847	4,117	4,391
Makeup H2 per Fresh Feed, scf/bbl	2,726	2,981	3,262	3,554	3,847	4,117	4,391
Net H2 after PSA, per Total Feed, scf/bbl	2,699	2,875	3,075	3,284	3,496	3,687	3,871
Hydrotreating							
WHSV, 1/h	0.95	0.89	0.84	0.79	0.74	0.70	0.67
Catalyst Load, lb	1,354,648	1,453,912	1,567,497	1,687,837	1,812,165	1,927,472	2,040,427
Catalyst Replacement Period, years	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrocracking							
WHSV, 1/h	0.95	0.89	0.84	0.79	0.74	0.70	0.67
Catalyst Load, lb	1,354,648	1,453,912	1,567,497	1,687,837	1,812,165	1,927,472	2,040,427

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Case ID	41130011	41140011	41150011	41160011	41170011	41180011	41190011
Catalyst Replacement Period, years	3.0	3.0	3.0	3.0	3.0	3.0	3.0
TEA							
Capital							
Fixed Capital Investment (FCI), million $	569.8	598.7	627.6	655.7	683.3	708.0	730.8
Working Capital, million $	28.5	29.9	31.4	32.8	34.2	35.4	36.5
Operating Costs, $/GGE	2.60	2.74	2.88	3.03	3.18	3.34	3.50
Petroleum Feed Cost, $/GGE	1.45	1.26	1.06	0.86	0.65	0.44	0.22
CFP Oil Feed Cost, $/GGE	0.91	1.22	1.55	1.88	2.21	2.56	2.92
Net Hydrogen Makeup Cost, $/GGE	0.20	0.21	0.23	0.25	0.27	0.29	0.31
Fixed Operating Costs, $/GGE	0.02	0.02	0.02	0.02	0.02	0.03	0.03
Coproduct/Utility Credits, $/GGE	−0.03	−0.04	−0.04	−0.05	−0.05	−0.05	−0.06
Other Operating Costs, $/GGE	0.07	0.08	0.08	0.09	0.09	0.10	0.11
Capital Charges, Income Tax etc.	0.06	0.07	0.07	0.07	0.08	0.08	0.09
MFSP, with Full Capital Costs, $/GGE	2.66	2.80	2.95	3.10	3.26	3.42	3.58
MFSP, with Full Capital Costs, $/GJE	2.85	3.00	3.16	3.32	3.49	3.66	3.84
MFSP, with Full Capital Costs, $/GDE	2.94	3.10	3.27	3.43	3.61	3.78	3.96
Other TEA Metrics							
Total Operating Cost without Feed, cents/GGE	24.0	25.7	27.6	29.6	31.7	33.6	35.6
Hydrogen Cost, cents/GGE	19.9	21.4	23.2	25.1	27.0	28.8	30.6
Total Catalyst Cost per Feed Barrel, $/bbl	0.73	0.79	0.85	0.92	0.98	1.05	1.11
Delta Capital for Repurposed Unit, million $	83.4	105.9	129.9	153.6	177.4	200.0	225.5
Repurposed Case ID	41201011	41201011	41201011	41201011	41201011	41201011	41201011
MFSP with Repurposed Equipment, $/GGE	2.61	2.75	2.90	3.05	3.20	3.36	3.53
Repurposed Petroleum Unit Size, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
Repurposed Unit FCI, $/annual bbl fresh feed	3.2	4.1	5.0	5.9	6.8	7.7	8.7
New Unit FCI, $/annual bbl fresh feed	21.9	23.0	24.1	25.2	26.3	27.2	28.1
Hydrotreating reactor FCI, million $	76.4	78.8	81.6	84.3	87.2	89.9	92.3
Hydrocracking reactor FCI, million $	76.4	78.8	81.6	84.3	87.2	89.9	92.3
Reactor(s) Pressure Correction On/Off	On	On	On	On	On	On	On
Makeup compressor FCI, million $	62.1	66.7	71.6	76.7	81.7	86.2	90.8
Recycle compressor FCI, million $	8.6	8.8	9.0	9.2	9.5	9.6	9.8
Fuel Lifecycle CI, g CO2e/MJ	72.7	64.9	56.9	48.9	41.0	32.7	24.2
CI for Petroleum Diesel Case, g CO2e/MJ	90.8	90.8	90.8	90.8	90.8	90.8	90.8
CI Reduction, %	19.9%	28.5%	37.3%	46.1%	54.9%	63.9%	73.4%

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Table K-9. Hydroprocessing Cases 41199011 to 41220211

Case ID	41199011	41201011	41201011	41220011	RxP 1795psi 41220211	RxP 1545psi 41220211	41220211
Total Feed, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
SRD Feed, BPD	0	0	0	0	0	0	0
LCO Feed, BPD	0	37,500	37,500	30,000	24,000	24,000	24,000
VGO Feed, BPD	0	37,500	37,500	30,000	24,000	24,000	24,000
CFP Oil Feed, BPD	75,000	0	0	15,000	12,000	12,000	12,000
Liquid Product Recycle, BPD	0	0	0	0	15,000	15,000	15,000
CFP Oil in Fresh Feed, vol %	100%	0%	0%	20%	20%	20%	20%
Liquid Product Recycle, vol %	0%	0%	0%	0%	20%	20%	20%
CFP Oil Oxygen (dry basis), wt %	16.0%	16.0%	16.0%	16.0%	16.0%	16.0%	16.0%
Total Liquid Fuel Product, lb/h	889,922	1,007,029	1,007,029	981,408	785,119	784,982	785,208
Total Liquid Fuel Product (all cuts), BGE/day	80,395	90,860	90,860	88,595	70,856	70,845	70,863
Total Liquid Fuel Product (all cuts), BJE/day	75,081	84,854	84,854	82,738	66,172	66,162	66,178
Total Liquid Fuel Product (all cuts), BDE/day	72,659	82,117	82,117	80,070	64,038	64,028	64,044
Total Liquid Product Molar H:C Ratio	1.80	1.79	1.79	1.80	1.80	1.80	1.79
Modeled Gasoline Product, wt %	27%	12%	12%	17%	17%	17%	17%
Modeled Jet Product, wt %	45%	33%	33%	35%	34%	34%	34%
Modeled Diesel/Heavies, wt %	28%	56%	56%	49%	49%	49%	49%
Hydroprocessing Reactor							
Inlet Pressure, psia	2,045	2,045	2,045	2,045	1,795	1,545	2,045
Inlet Temperature, F	644	644	644	644	644	644	644
Outlet Temperature, F	684	684	684	684	684	684	684
Carbon-Efficiency for CFP Oil, %	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%
Carbon-Efficiency for Petroleum Feed, %	97.9%	97.8%	97.8%	97.6%	97.6%	97.7%	97.6%
WHSV proration basis from SRD using chem. H2							
Base SRD Case H2 Consumption, scf/bbl	615.0	615.0	615.0	615.0	615.0	615.0	615.0
Base SRD Case WHSV Assumption, 1/h	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrogen							
Makeup H2 Purity, mol %	99%	85%	85%	85%	85%	85%	85%
H2 Purity at Reactor Inlet, mol %	80%	80%	80%	80%	80%	80%	80%
Chemical H2 per Total Feed, scf/bbl	3,872	2,034	2,034	2,430	1,946	1,948	1,945
Chemical H2 per Fresh Feed, scf/bbl	3,872	2,034	2,034	2,430	2,433	2,435	2,431
Reactor H2 Feed per Total Feed, scf/bbl	26,393	11,182	11,182	14,013	9,244	9,221	9,276
Reactor H2 Feed per Fresh Feed, scf/bbl	26,393	11,182	11,182	14,013	11,555	11,526	11,595
Makeup H2 per Total Feed, scf/bbl	4,810	3,569	3,569	4,913	3,702	3,822	3,610
Makeup H2 per Fresh Feed, scf/bbl	4,810	3,569	3,569	4,913	4,628	4,777	4,513
Net H2 after PSA, per Total Feed, scf/bbl	4,100	2,360	2,360	2,903	2,291	2,297	2,290
Hydrotreating							
WHSV, 1/h	0.64	1.21	1.21	1.01	1.01	1.01	1.01
Catalyst Load, lb	2,176,642	1,027,236	1,027,236	1,254,778	1,005,040	1,005,900	1,004,348
Catalyst Replacement Period, years	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrocracking							

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Case ID	41199011	41201011	41201011	41220011	RxP 1795psi 41220211	RxP 1545psi 41220211	41220211
WHSV, 1/h	0.64	1.21	1.21	1.01	1.01	1.01	1.01
Catalyst Load, lb	2,176,642	1,027,236	1,027,236	1,254,778	1,005,040	1,005,900	1,004,348
Catalyst Replacement Period, years	3.0	3.0	3.0	3.0	3.0	3.0	3.0
TEA							
Capital							
Fixed Capital Investment (FCI), million $	758.6	536.3	536.3	630.3	514.1	483.6	552.7
Working Capital, million $	37.9	26.8	26.8	31.5	25.7	24.2	27.6
Operating Costs, $/GGE	3.67	2.22	2.22	2.51	2.50	2.50	2.50
Petroleum Feed Cost, $/GGE	0.00	2.00	2.00	1.64	1.64	1.64	1.64
CFP Oil Feed Cost, $/GGE	3.29	0.00	0.00	0.60	0.60	0.60	0.60
Net Hydrogen Makeup Cost, $/GGE	0.33	0.17	0.17	0.21	0.21	0.21	0.21
Fixed Operating Costs, $/GGE	0.03	0.02	0.02	0.02	0.02	0.02	0.02
Coproduct/Utility Credits, $/GGE	−0.07	−0.03	−0.03	−0.04	−0.04	−0.04	−0.04
Other Operating Costs, $/GGE	0.11	0.07	0.07	0.08	0.08	0.08	0.08
Capital Charges, Income Tax etc.	0.09	0.06	0.06	0.07	0.07	0.06	0.07
MFSP, with Full Capital Costs, $/GGE	3.76	2.27	2.27	2.57	2.57	2.56	2.58
MFSP, with Full Capital Costs, $/GJE	4.03	2.44	2.44	2.76	2.75	2.75	2.76
MFSP, with Full Capital Costs, $/GDE	4.16	2.52	2.52	2.85	2.84	2.84	2.85
Other TEA Metrics							
Total Operating Cost without Feed, cents/GGE	38.0	21.7	21.7	26.6	26.1	25.9	26.3
Hydrogen Cost, cents/GGE	33.0	16.8	16.8	21.2	20.9	20.9	20.9
Total Catalyst Cost per Feed Barrel, $/bbl	1.18	0.56	0.56	0.68	0.68	0.68	0.68
Delta Capital for Repurposed Unit, million $	256.8	0.0	0.0	99.1	39.7	37.1	46.6
Repurposed Case ID	41201011	41201011	41201011	41201011	41201011	41201011	41201011
MFSP with Repurposed Equipment, $/GGE	3.71	2.22	2.22	2.52	2.51	2.51	2.51
Repurposed Petroleum Unit Size, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
Repurposed Unit FCI, $/annual bbl fresh feed	9.9	0.0	0.0	3.8	1.9	1.8	2.2
New Unit FCI, $/annual bbl fresh feed	29.2	20.6	20.6	24.2	24.7	23.2	26.6
Hydrotreating reactor FCI, million $	95.7	70.2	70.2	75.9	61.2	55.5	66.9
Hydrocracking reactor FCI, million $	95.7	70.2	70.2	75.9	61.2	55.5	66.9
Reactor(s) Pressure Correction On/Off	On	On	On	On	On	On	On
Makeup compressor FCI, million $	97.6	85.7	85.7	110.5	83.4	79.9	86.4
Recycle compressor FCI, million $	10.0	7.7	7.7	8.3	7.6	7.9	7.3
Fuel Lifecycle CI, g CO2e/MJ	15.5	96.3	96.3	82.4	81.9	81.9	82.0
CI for Petroleum Diesel Case, g CO2e/MJ	90.8	90.8	90.8	90.8	90.8	90.8	90.8
CI Reduction, %	82.9%	−6.1%	−6.1%	9.3%	9.8%	9.8%	9.7%

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Table K-10. Hydroprocessing Cases 41220811 to 42140011

Case ID	41220811	42110011	42120011	42120211	42120811	42130011	42140011
Total Feed, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
SRD Feed, BPD	0	0	0	0	0	0	0
LCO Feed, BPD	12,875	33,750	30,000	24,000	11,455	26,250	22,500
VGO Feed, BPD	12,875	33,750	30,000	24,000	11,455	26,250	22,500
CFP Oil Feed, BPD	6,438	7,500	15,000	12,000	5,728	22,500	30,000
Liquid Product Recycle, BPD	42,812	0	0	15,000	46,362	0	0
CFP Oil in Fresh Feed, vol %	20%	10%	20%	20%	20%	30%	40%
Liquid Product Recycle, vol %	57%	0%	0%	20%	62%	0%	0%
CFP Oil Oxygen (dry basis), wt %	16.0%	21.6%	21.6%	21.6%	21.6%	21.6%	21.6%
Total Liquid Fuel Product, lb/h	421,774	996,974	985,754	788,895	376,860	974,899	964,288
Total Liquid Fuel Product (all cuts), BGE/day	38,069	89,959	88,951	71,180	33,992	87,970	87,010
Total Liquid Fuel Product (all cuts), BJE/day	35,553	84,013	83,071	66,474	31,745	82,155	81,259
Total Liquid Fuel Product (all cuts), BDE/day	34,406	81,303	80,392	64,330	30,721	79,505	78,638
Total Liquid Product Molar H:C Ratio	1.80	1.80	1.80	1.80	1.80	1.80	1.80
Modeled Gasoline Product, wt %	18%	13%	14%	15%	15%	16%	17%
Modeled Jet Product, wt %	34%	34%	35%	34%	34%	36%	37%
Modeled Diesel/Heavies, wt %	49%	53%	51%	51%	51%	48%	45%
Hydroprocessing Reactor							
Inlet Pressure, psia	2,045	2,045	2,045	2,045	2,045	2,045	2,045
Inlet Temperature, F	644	644	644	644	644	644	644
Outlet Temperature, F	684	684	684	685	684	684	683
Carbon-Efficiency for CFP Oil, %	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%
Carbon-Efficiency for Petroleum Feed, %	97.5%	98.3%	98.2%	98.1%	97.9%	98.2%	98.1%
WHSV proration basis from SRD using chem. H2							
Base SRD Case H2 Consumption, scf/bbl	615.0	615.0	615.0	615.0	615.0	615.0	615.0
Base SRD Case WHSV Assumption, 1/h	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrogen							
Makeup H2 Purity, mol %	85%	99%	99%	99%	99%	99%	99%
H2 Purity at Reactor Inlet, mol %	80%	81%	80%	80%	88%	80%	80%
Chemical H2 per Total Feed, scf/bbl	1,057	2,427	2,789	2,241	1,074	3,121	3,474
Chemical H2 per Fresh Feed, scf/bbl	2,463	2,427	2,789	2,801	2,813	3,121	3,474
Reactor H2 Feed per Total Feed, scf/bbl	4,128	15,348	18,134	12,395	4,584	21,109	24,200
Reactor H2 Feed per Fresh Feed, scf/bbl	9,619	15,348	18,134	15,493	12,004	21,109	24,200
Makeup H2 per Total Feed, scf/bbl	1,575	2,520	3,052	2,408	1,165	3,551	4,090
Makeup H2 per Fresh Feed, scf/bbl	3,669	2,520	3,052	3,010	3,052	3,551	4,090
Net H2 after PSA, per Total Feed, scf/bbl	1,224	2,519	2,907	2,342	1,165	3,264	3,646
Hydrotreating							
WHSV, 1/h	1.00	1.01	0.88	0.88	0.87	0.79	0.71
Catalyst Load, lb	546,036	1,252,167	1,469,480	1,180,871	566,021	1,678,967	1,907,043
Catalyst Replacement Period, years	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrocracking							
WHSV, 1/h	1.00	1.01	0.88	0.88	0.87	0.79	0.71
Catalyst Load, lb	546,036	1,252,167	1,469,480	1,180,871	566,021	1,678,967	1,907,043

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Case ID	41220811	42110011	42120011	42120211	42120811	42130011	42140011
Catalyst Replacement Period, years	3.0	3.0	3.0	3.0	3.0	3.0	3.0
TEA							
Capital							
Fixed Capital Investment (FCI), million $	401.3	530.7	585.7	521.6	367.5	631.2	679.4
Working Capital, million $	20.1	26.5	29.3	26.1	18.4	31.6	34.0
Operating Costs, $/GGE	2.51	2.47	2.76	2.76	2.79	3.05	3.35
Petroleum Feed Cost, $/GGE	1.64	1.82	1.64	1.64	1.63	1.45	1.25
CFP Oil Feed Cost, $/GGE	0.60	0.43	0.87	0.87	0.87	1.31	1.77
Net Hydrogen Makeup Cost, $/GGE	0.21	0.18	0.21	0.21	0.22	0.24	0.27
Fixed Operating Costs, $/GGE	0.04	0.02	0.02	0.02	0.04	0.02	0.02
Coproduct/Utility Credits, $/GGE	−0.04	−0.03	−0.03	−0.03	−0.04	−0.04	−0.04
Other Operating Costs, $/GGE	0.08	0.07	0.08	0.08	0.08	0.09	0.10
Capital Charges, Income Tax etc.	0.10	0.06	0.06	0.07	0.10	0.07	0.07
MFSP, with Full Capital Costs, $/GGE	2.61	2.53	2.82	2.83	2.89	3.12	3.43
MFSP, with Full Capital Costs, $/GJE	2.79	2.71	3.02	3.03	3.09	3.34	3.67
MFSP, with Full Capital Costs, $/GDE	2.88	2.80	3.12	3.13	3.19	3.45	3.79
Other TEA Metrics							
Total Operating Cost without Feed, cents/GGE	27.1	22.5	26.0	26.2	28.6	29.3	32.9
Hydrogen Cost, cents/GGE	20.8	18.1	21.1	21.3	22.2	24.0	27.1
Total Catalyst Cost per Feed Barrel, $/bbl	0.69	0.68	0.80	0.80	0.80	0.91	1.03
Delta Capital for Repurposed Unit, million $	39.6	46.6	86.6	53.6	38.3	123.0	160.0
Repurposed Case ID	41201011	42201011	42201011	42201011	42201011	42201011	42201011
MFSP with Repurposed Equipment, $/GGE	2.52	2.48	2.77	2.77	2.80	3.07	3.37
Repurposed Petroleum Unit Size, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
Repurposed Unit FCI, $/annual bbl fresh feed	3.5	1.8	3.3	2.6	3.9	4.7	6.2
New Unit FCI, $/annual bbl fresh feed	36.0	20.4	22.5	25.1	37.0	24.3	26.1
Hydrotreating reactor FCI, million $	57.8	72.7	78.1	69.0	55.5	82.9	87.8
Hydrocracking reactor FCI, million $	57.8	72.7	78.1	69.0	55.5	82.9	87.8
Reactor(s) Pressure Correction On/Off	On	On	On	On	On	On	On
Makeup compressor FCI, million $	44.7	58.3	68.0	56.3	31.6	76.7	85.8
Recycle compressor FCI, million $	5.7	8.5	8.9	8.0	5.8	9.4	9.8
Fuel Lifecycle CI, g CO2e/MJ	81.9	88.2	81.2	81.2	81.7	74.0	66.7
CI for Petroleum Diesel Case, g CO2e/MJ	90.8	90.8	90.8	90.8	90.8	90.8	90.8
CI Reduction, %	9.8%	2.8%	10.5%	10.5%	10.0%	18.4%	26.5%

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Table K-11. Hydroprocessing Cases 42150011 to 42201011

Case ID	42150011	42160011	42170011	42180011	42190011	42199011	42201011
Total Feed, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
SRD Feed, BPD	0	0	0	0	0	0	0
LCO Feed, BPD	18,750	15,000	11,250	7,500	3,750	0	37,500
VGO Feed, BPD	18,750	15,000	11,250	7,500	3,750	0	37,500
CFP Oil Feed, BPD	37,500	45,000	52,500	60,000	67,500	75,000	0
Liquid Product Recycle, BPD	0	0	0	0	0	0	0
CFP Oil in Fresh Feed, vol %	50%	60%	70%	80%	90%	100%	0%
Liquid Product Recycle, vol %	0%	0%	0%	0%	0%	0%	0%
CFP Oil Oxygen (dry basis), wt %	21.6%	21.6%	21.6%	21.6%	21.6%	21.6%	21.6%
Total Liquid Fuel Product, lb/h	953,792	943,309	932,832	922,494	912,197	901,758	1,007,030
Total Liquid Fuel Product (all cuts), BGE/day	86,059	85,109	84,158	83,220	82,286	81,339	90,860
Total Liquid Fuel Product (all cuts), BJE/day	80,370	79,483	78,595	77,719	76,847	75,962	84,854
Total Liquid Fuel Product (all cuts), BDE/day	77,778	76,919	76,060	75,213	74,368	73,512	82,117
Total Liquid Product Molar H:C Ratio	1.80	1.80	1.80	1.80	1.80	1.80	1.79
Modeled Gasoline Product, wt %	19%	20%	22%	23%	24%	26%	12%
Modeled Jet Product, wt %	39%	40%	42%	43%	45%	47%	33%
Modeled Diesel/Heavies, wt %	43%	40%	37%	34%	31%	28%	56%
Hydroprocessing Reactor							
Inlet Pressure, psia	2,045	2,045	2,045	2,045	2,045	2,045	2,045
Inlet Temperature, F	644	644	644	644	644	644	644
Outlet Temperature, F	684	684	684	684	684	684	684
Carbon-Efficiency for CFP Oil, %	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%
Carbon-Efficiency for Petroleum Feed, %	98.1%	98.1%	98.1%	98.0%	98.0%	98.2%	97.8%
WHSV proration basis from SRD using chem. H2							
Base SRD Case H2 Consumption, scf/bbl	615.0	615.0	615.0	615.0	615.0	615.0	615.0
Base SRD Case WHSV Assumption, 1/h	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrogen							
Makeup H2 Purity, mol %	99%	99%	99%	99%	99%	99%	85%
H2 Purity at Reactor Inlet, mol %	80%	80%	80%	80%	80%	80%	80%
Chemical H2 per Total Feed, scf/bbl	3,823	4,171	4,516	4,859	5,200	5,539	2,035
Chemical H2 per Fresh Feed, scf/bbl	3,823	4,171	4,516	4,859	5,200	5,539	2,035
Reactor H2 Feed per Total Feed, scf/bbl	27,193	30,290	33,277	36,279	39,194	42,113	11,175
Reactor H2 Feed per Fresh Feed, scf/bbl	27,193	30,290	33,277	36,279	39,194	42,113	11,175
Makeup H2 per Total Feed, scf/bbl	4,620	5,154	5,680	6,203	6,721	7,233	3,567
Makeup H2 per Fresh Feed, scf/bbl	4,620	5,154	5,680	6,203	6,721	7,233	3,567
Net H2 after PSA, per Total Feed, scf/bbl	4,022	4,398	4,770	5,140	5,507	5,872	2,359
Hydrotreating							
WHSV, 1/h	0.64	0.59	0.54	0.51	0.47	0.44	1.21
Catalyst Load, lb	2,140,943	2,381,404	2,628,129	2,881,206	3,140,262	3,406,273	1,027,265
Catalyst Replacement Period, years	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrocracking							
WHSV, 1/h	0.64	0.59	0.54	0.51	0.47	0.44	1.21
Catalyst Load, lb	2,140,943	2,381,404	2,628,129	2,881,206	3,140,262	3,406,273	1,027,265

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Case ID	42150011	42160011	42170011	42180011	42190011	42199011	42201011
Catalyst Replacement Period, years	3.0	3.0	3.0	3.0	3.0	3.0	3.0
TEA							
Capital							
Fixed Capital Investment (FCI), million $	742.5	771.4	813.1	853.5	891.9	924.6	536.3
Working Capital, million $	37.1	38.6	40.7	42.7	44.6	46.2	26.8
Operating Costs, $/GGE	3.66	3.97	4.29	4.62	4.95	5.29	2.22
Petroleum Feed Cost, $/GGE	1.06	0.85	0.65	0.44	0.22	0.00	2.00
CFP Oil Feed Cost, $/GGE	2.24	2.71	3.20	3.70	4.21	4.73	0.00
Net Hydrogen Makeup Cost, $/GGE	0.30	0.33	0.37	0.40	0.43	0.47	0.17
Fixed Operating Costs, $/GGE	0.03	0.03	0.03	0.03	0.03	0.03	0.02
Coproduct/Utility Credits, $/GGE	−0.05	−0.05	−0.06	−0.07	−0.07	−0.08	−0.03
Other Operating Costs, $/GGE	0.11	0.12	0.14	0.15	0.16	0.18	0.07
Capital Charges, Income Tax etc.	0.08	0.09	0.09	0.10	0.10	0.11	0.06
MFSP, with Full Capital Costs, $/GGE	3.74	4.06	4.39	4.72	5.06	5.40	2.27
MFSP, with Full Capital Costs, $/GJE	4.01	4.35	4.70	5.05	5.41	5.79	2.44
MFSP, with Full Capital Costs, $/GDE	4.14	4.49	4.85	5.22	5.59	5.98	2.52
Other TEA Metrics							
Total Operating Cost without Feed, cents/GGE	36.6	40.3	44.2	48.0	52.0	55.9	21.7
Hydrogen Cost, cents/GGE	30.2	33.4	36.6	39.9	43.2	46.7	16.8
Total Catalyst Cost per Feed Barrel, $/bbl	1.16	1.29	1.43	1.56	1.70	1.85	0.56
Delta Capital for Repurposed Unit, million $	222.1	251.0	293.4	334.9	375.6	416.8	0.0
Repurposed Case ID	42201011	42201011	42201011	42201011	42201011	42201011	42201011
MFSP with Repurposed Equipment, $/GGE	3.69	4.00	4.33	4.66	5.00	5.35	2.22
Repurposed Petroleum Unit Size, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
Repurposed Unit FCI, $/annual bbl fresh feed	8.5	9.7	11.3	12.9	14.4	16.0	0.0
New Unit FCI, $/annual bbl fresh feed	28.5	29.7	31.3	32.8	34.3	35.6	20.6
Hydrotreating reactor FCI, million $	92.5	97.3	101.9	106.5	111.0	115.5	70.2
Hydrocracking reactor FCI, million $	92.5	97.3	101.9	106.5	111.0	115.5	70.2
Reactor(s) Pressure Correction On/Off	On	On	On	On	On	On	On
Makeup compressor FCI, million $	94.5	103.1	111.4	119.5	127.4	135.1	85.6
Recycle compressor FCI, million $	10.1	10.5	10.8	11.0	11.3	11.5	7.7
Fuel Lifecycle CI, g CO2e/MJ	59.2	51.6	43.8	36.0	28.1	19.6	96.3
CI for Petroleum Diesel Case, g CO2e/MJ	90.8	90.8	90.8	90.8	90.8	90.8	90.8
CI Reduction, %	34.7%	43.2%	51.7%	60.3%	69.1%	78.4%	−6.1%

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Table K-12. Hydroprocessing Cases 42220011 to 42220811, and Process Sensitivity Cases With 12120211

Case ID	42220011	42220811	Rx delta T Reactor T 12120211	Reactor T 12120211	Prod.H:C 1.85 12120211	FRecy 12120211	CFP CEff 86% 12120211
Total Feed, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
SRD Feed, BPD	0	0	48,000	48,000	48,000	48,000	48,000
LCO Feed, BPD	30,000	11,653	0	0	0	0	0
VGO Feed, BPD	30,000	11,653	0	0	0	0	0
CFP Oil Feed, BPD	15,000	5,826	12,000	12,000	12,000	12,000	12,000
Liquid Product Recycle, BPD	0	45,868	15,000	15,000	15,000	15,000	15,000
CFP Oil in Fresh Feed, vol %	20%	20%	20%	20%	20%	20%	20%
Liquid Product Recycle, vol %	0%	61%	20%	20%	20%	20%	20%
CFP Oil Oxygen (dry basis), wt %	21.6%	21.6%	21.6%	21.6%	21.6%	21.6%	21.6%
Total Liquid Fuel Product, lb/h	983,558	382,508	719,314	719,219	722,908	715,835	712,309
Total Liquid Fuel Product (all cuts), BGE/day	88,770	34,514	64,828	64,812	65,448	64,538	64,203
Total Liquid Fuel Product (all cuts), BJE/day	82,902	32,233	60,543	60,527	61,122	60,272	59,959
Total Liquid Fuel Product (all cuts), BDE/day	80,228	31,193	58,590	58,575	59,151	58,328	58,025
Total Liquid Product Molar H:C Ratio	1.80	1.80	1.79	1.79	1.85	1.79	1.79
Modeled Gasoline Product, wt %	14%	15%	24%	24%	26%	24%	24%
Modeled Jet Product, wt %	35%	35%	47%	47%	46%	47%	48%
Modeled Diesel/Heavies, wt %	51%	51%	29%	29%	28%	29%	28%
Hydroprocessing Reactor							
Inlet Pressure, psia	2,045	2,045	1,045	1,045	1,045	1,045	1,045
Inlet Temperature, F	644	644	702	712	644	644	644
Outlet Temperature, F	684	684	752	753	684	684	684
Carbon-Efficiency for CFP Oil, %	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%	86.0%
Carbon-Efficiency for Petroleum Feed, %	97.5%	97.4%	98.6%	98.6%	98.6%	98.3%	98.5%
WHSV proration basis from SRD using chem. H2							
Base SRD Case H2 Consumption, scf/bbl	615.0	615.0	615.0	615.0	615.0	615.0	615.0
Base SRD Case WHSV Assumption, 1/h	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrogen							
Makeup H2 Purity, mol %	85%	85%	99%	99%	99%	99%	99%
H2 Purity at Reactor Inlet, mol %	80%	80%	80%	80%	80%	80%	80%
Chemical H2 per Total Feed, scf/bbl	2,769	1,087	1,314	1,310	1,501	1,337	1,356
Chemical H2 per Fresh Feed, scf/bbl	2,769	2,798	1,643	1,637	1,877	1,671	1,695
Reactor H2 Feed per Total Feed, scf/bbl	16,772	4,252	5,711	5,939	8,051	8,626	6,609
Reactor H2 Feed per Fresh Feed, scf/bbl	16,772	10,946	7,138	7,423	10,064	10,782	8,261
Makeup H2 per Total Feed, scf/bbl	5,987	1,711	1,556	1,556	1,768	1,628	1,680
Makeup H2 per Fresh Feed, scf/bbl	5,987	4,406	1,945	1,945	2,210	2,035	2,099
Net H2 after PSA, per Total Feed, scf/bbl	3,350	1,268	1,385	1,381	1,577	1,416	1,439
Hydrotreating							
WHSV, 1/h	0.89	0.88	0.75	0.75	0.66	0.74	0.73
Catalyst Load, lb	1,459,027	572,771	1,267,050	1,262,537	1,447,371	1,288,997	1,307,077
Catalyst Replacement Period, years	2.0	2.0	2.0	2.0	2.0	2.0	2.0

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Case ID	42220011	42220811	Rx delta T Reactor T 12120211	Reactor T 12120211	Prod.H:C 1.85 12120211	FRecy 12120211	CFP CEff 86% 12120211
Hydrocracking							
WHSV, 1/h	0.89	0.88	N/A	N/A	N/A	N/A	N/A
Catalyst Load, lb	1,459,027	572,771	N/A	N/A	N/A	N/A	N/A
Catalyst Replacement Period, years	3.0	3.0	N/A	N/A	N/A	N/A	N/A
TEA							
Capital							
Fixed Capital Investment (FCI), million $	682.2	406.4	275.1	275.9	289.3	285.6	280.1
Working Capital, million $	34.1	20.3	13.8	13.8	14.5	14.3	14.0
Operating Costs, $/GGE	2.82	2.82	2.99	2.99	2.98	3.01	3.02
Petroleum Feed Cost, $/GGE	1.64	1.64	1.87	1.87	1.85	1.87	1.88
CFP Oil Feed Cost, $/GGE	0.87	0.87	0.95	0.95	0.94	0.95	0.96
Net Hydrogen Makeup Cost, $/GGE	0.24	0.24	0.14	0.14	0.16	0.14	0.14
Fixed Operating Costs, $/GGE	0.02	0.04	0.02	0.02	0.02	0.02	0.02
Coproduct/Utility Credits, $/GGE	−0.04	−0.04	−0.03	−0.03	−0.03	−0.03	−0.03
Other Operating Costs, $/GGE	0.10	0.09	0.06	0.06	0.06	0.06	0.06
Capital Charges, Income Tax etc.	0.07	0.11	0.04	0.04	0.04	0.04	0.04
MFSP, with Full Capital Costs, $/GGE	2.89	2.93	3.03	3.03	3.02	3.05	3.06
MFSP, with Full Capital Costs, $/GJE	3.09	3.13	3.24	3.24	3.24	3.26	3.28
MFSP, with Full Capital Costs, $/GDE	3.20	3.24	3.35	3.35	3.35	3.37	3.39
Other TEA Metrics							
Total Operating Cost without Feed, cents/GGE	30.9	31.2	17.0	17.0	19.2	17.6	17.5
Hydrogen Cost, cents/GGE	24.4	23.7	13.8	13.8	15.6	14.2	14.5
Total Catalyst Cost per Feed Barrel, $/bbl	0.79	0.80	0.52	0.51	0.59	0.52	0.53
Delta Capital for Repurposed Unit, million $	150.9	38.4	72.7	73.4	86.5	82.3	76.9
Repurposed Case ID	42201011	42201011	12201011	12201011	12201011	12201011	12201011
MFSP with Repurposed Equipment, $/GGE	2.83	2.83	3.00	3.00	3.00	3.02	3.03
Repurposed Petroleum Unit Size, BPD	75,000	75,000	75,000	75,000	75,000	75,000	75,000
Repurposed Unit FCI, $/annual bbl fresh feed	5.8	3.8	3.5	3.5	4.2	4.0	3.7
New Unit FCI, $/annual bbl fresh feed	26.2	40.2	13.2	13.3	13.9	13.7	13.5
Hydrotreating reactor FCI, million $	81.0	58.2	34.7	35.0	38.1	39.1	36.3
Hydrocracking reactor FCI, million $	81.0	58.2	0.0	0.0	0.0	0.0	0.0
Reactor(s) Pressure Correction On/Off	On	On	On	On	On	On	On
Makeup compressor FCI, million $	129.3	47.7	28.4	28.4	31.5	29.5	30.2
Recycle compressor FCI, million $	8.7	5.8	7.7	7.8	8.5	8.7	8.0
Fuel Lifecycle CI, g CO2e/MJ	83.6	82.9	75.9	75.9	76.9	76.2	76.2
CI for Petroleum Diesel Case, g CO2e/MJ	90.8	90.8	90.8	90.8	90.8	90.8	90.8
CI Reduction, %	7.9%	8.6%	16.4%	16.4%	15.3%	16.0%	16.1%

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Table K-13. Hydroprocessing Cases, CAPEX, and Scale Changes With Case 12120211

Case ID	CAPEXx1.3 12120211	SCALEx0.67 CAPEXx1.3 12120211	SCALEx0.67 12120211	SCALEx0.33 CAPEXx1.3 12120211	SCALEx0.33 12120211	SCALEx0.13 CAPEXx1.3 12120211	SCALEx0.13 12120211
Total Feed, BPD	75,000	50,000	50,000	25,000	25,000	10,000	10,000
SRD Feed, BPD	48,000	32,000	32,000	16,000	16,000	6,400	6,400
LCO Feed, BPD	0	0	0	0	0	0	0
VGO Feed, BPD	0	0	0	0	0	0	0
CFP Oil Feed, BPD	12,000	8,000	8,000	4,000	4,000	1,600	1,600
Liquid Product Recycle, BPD	15,000	10,000	10,000	5,000	5,000	2,000	2,000
CFP Oil in Fresh Feed, vol %	20%	20%	20%	20%	20%	20%	20%
Liquid Product Recycle, vol %	20%	20%	20%	20%	20%	20%	20%
CFP Oil Oxygen (dry basis), wt %	21.6%	21.6%	21.6%	21.6%	21.6%	21.6%	21.6%
Total Liquid Fuel Product, lb/h	719,338	479,559	479,559	239,779	239,779	95,912	95,912
Total Liquid Fuel Product (all cuts), BGE/day	64,833	43,222	43,222	21,611	21,611	8,644	8,644
Total Liquid Fuel Product (all cuts), BJE/day	60,548	40,365	40,365	20,183	20,183	8,073	8,073
Total Liquid Fuel Product (all cuts), BDE/day	58,595	39,063	39,063	19,532	19,532	7,813	7,813
Total Liquid Product Molar H:C Ratio	1.79	1.79	1.79	1.79	1.79	1.79	1.79
Modeled Gasoline Product, wt %	24%	24%	24%	24%	24%	24%	24%
Modeled Jet Product, wt %	47%	47%	47%	47%	47%	47%	47%
Modeled Diesel/Heavies, wt %	29%	29%	29%	29%	29%	29%	29%
Hydroprocessing Reactor							
Inlet Pressure, psia	1,045	1,045	1,045	1,045	1,045	1,045	1,045
Inlet Temperature, F	644	644	644	644	644	644	644
Outlet Temperature, F	684	684	684	684	684	684	684
Carbon-Efficiency for CFP Oil, %	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%
Carbon-Efficiency for Petroleum Feed, %	98.6%	98.6%	98.6%	98.6%	98.6%	98.6%	98.6%
WHSV proration basis from SRD using chem. H2							
Base SRD Case H2 Consumption, scf/bbl	615.0	615.0	615.0	615.0	615.0	615.0	615.0
Base SRD Case WHSV Assumption, 1/h	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrogen							
Makeup H2 Purity, mol %	99%	99%	99%	99%	99%	99%	99%
H2 Purity at Reactor Inlet, mol %	80%	80%	80%	80%	80%	80%	80%
Chemical H2 per Total Feed, scf/bbl	1,316	1,316	1,316	1,316	1,316	1,316	1,316
Chemical H2 per Fresh Feed, scf/bbl	1,645	1,645	1,645	1,645	1,645	1,645	1,645
Reactor H2 Feed per Total Feed, scf/bbl	6,448	6,448	6,448	6,448	6,448	6,448	6,448
Reactor H2 Feed per Fresh Feed, scf/bbl	8,060	8,060	8,060	8,060	8,060	8,060	8,060
Makeup H2 per Total Feed, scf/bbl	1,572	1,572	1,572	1,572	1,572	1,572	1,572
Makeup H2 per Fresh Feed, scf/bbl	1,965	1,965	1,965	1,965	1,965	1,965	1,965
Net H2 after PSA, per Total Feed, scf/bbl	1,389	1,389	1,389	1,389	1,389	1,389	1,389
Hydrotreating							
WHSV, 1/h	0.75	0.75	0.75	0.75	0.75	0.75	0.75
Catalyst Load, lb	1,268,916	845,944	845,944	422,972	422,972	169,189	169,189
Catalyst Replacement Period, years	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrocracking							

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Case ID	CAPEXx1.3 12120211	SCALEx0.67 CAPEXx1.3 12120211	SCALEx0.67 12120211	SCALEx0.33 CAPEXx1.3 12120211	SCALEx0.33 12120211	SCALEx0.13 CAPEXx1.3 12120211	SCALEx0.13 12120211
WHSV, 1/h	N/A	N/A	N/A	N/A	N/A	N/A	N/A
Catalyst Load, lb	N/A	N/A	N/A	N/A	N/A	N/A	N/A
Catalyst Replacement Period, years	N/A	N/A	N/A	N/A	N/A	N/A	N/A
TEA							
Capital							
Fixed Capital Investment (FCI), million $	358.4	275.1	211.6	178.2	137.1	107.7	82.9
Working Capital, million $	17.9	13.8	10.6	8.9	6.9	5.4	4.1
Operating Costs, $/GGE	2.99	3.00	2.99	3.01	3.00	3.04	3.04
Petroleum Feed Cost, $/GGE	1.87	1.87	1.87	1.87	1.87	1.87	1.87
CFP Oil Feed Cost, $/GGE	0.95	0.95	0.95	0.95	0.95	0.95	0.95
Net Hydrogen Makeup Cost, $/GGE	0.14	0.14	0.14	0.14	0.14	0.14	0.14
Fixed Operating Costs, $/GGE	0.02	0.02	0.02	0.04	0.03	0.07	0.06
Coproduct/Utility Credits, $/GGE	−0.03	−0.03	−0.03	−0.03	−0.03	−0.03	−0.03
Other Operating Costs, $/GGE	0.06	0.06	0.06	0.06	0.06	0.06	0.06
Capital Charges, Income Tax etc.	0.05	0.06	0.05	0.08	0.06	0.12	0.09
MFSP, with Full Capital Costs, $/GGE	3.04	3.06	3.04	3.09	3.06	3.16	3.13
MFSP, with Full Capital Costs, $/GJE	3.26	3.27	3.25	3.30	3.28	3.38	3.35
MFSP, with Full Capital Costs, $/GDE	3.37	3.38	3.36	3.41	3.39	3.50	3.46
Other TEA Metrics							
Total Operating Cost without Feed, cents/GGE	17.5	18.0	17.6	19.3	18.8	22.7	22.0
Hydrogen Cost, cents/GGE	13.8	13.8	13.8	13.8	13.8	13.8	13.8
Total Catalyst Cost per Feed Barrel, $/bbl	0.52	0.52	0.52	0.52	0.52	0.52	0.52
Delta Capital for Repurposed Unit, million $	94.9	71.1	54.7	44.2	34.0	24.0	18.5
Repurposed Case ID	12201011	12201011	12201011	12201011	12201011	12201011	12201011
MFSP with Repurposed Equipment, $/GGE	3.01	3.01	3.01	3.03	3.02	3.07	3.06
Repurposed Petroleum Unit Size, BPD	75,000	50,000	50,000	25,000	25,000	10,000	10,000
Repurposed Unit FCI, $/annual bbl fresh feed	4.6	5.1	3.9	6.4	4.9	8.7	6.7
New Unit FCI, $/annual bbl fresh feed	17.2	19.8	15.3	25.7	19.8	38.8	29.9
Hydrotreating reactor FCI, million $	35.8	23.9	23.9	11.4	11.4	6.2	6.2
Hydrocracking reactor FCI, million $	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Reactor(s) Pressure Correction On/Off	On	On	On	On	On	Off	Off
Makeup compressor FCI, million $	28.6	20.7	20.7	11.9	11.9	5.8	5.8
Recycle compressor FCI, million $	8.0	7.1	7.1	5.7	5.7	4.3	4.3
Fuel Lifecycle CI, g CO2e/MJ	76.0	76.0	76.0	76.0	76.0	76.0	76.0
CI for Petroleum Diesel Case, g CO2e/MJ	90.8	90.8	90.8	90.8	90.8	90.8	90.8
CI Reduction, %	16.3%	16.3%	16.3%	16.3%	16.3%	16.2%	16.2%

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Table K-14. Hydroprocessing Cases, On-Stream Factor, CAPEX, and Scale Changes With Case 12120211

Case ID	SCALEx0.04 CAPEXx1.3 12120211	SCALEx0.04 12120211	Onstream 98% SCALEx0.67 12120211	Onstream 98% SCALEx0.33 12120211	Onstream 98% SCALEx0.13 12120211	Onstream 98% SCALEx0.04 12120211	Onstream 98% 12120211
Total Feed, BPD	3,000	3,000	50,000	25,000	10,000	3,000	75,000
SRD Feed, BPD	1,920	1,920	32,000	16,000	6,400	1,920	48,000
LCO Feed, BPD	0	0	0	0	0	0	0
VGO Feed, BPD	0	0	0	0	0	0	0
CFP Oil Feed, BPD	480	480	8,000	4,000	1,600	480	12,000
Liquid Product Recycle, BPD	600	600	10,000	5,000	2,000	600	15,000
CFP Oil in Fresh Feed, vol %	20%	20%	20%	20%	20%	20%	20%
Liquid Product Recycle, vol %	20%	20%	20%	20%	20%	20%	20%
CFP Oil Oxygen (dry basis), wt %	21.6%	21.6%	21.6%	21.6%	21.6%	21.6%	21.6%
Total Liquid Fuel Product, lb/h	28,774	28,774	479,559	239,779	95,912	28,774	719,338
Total Liquid Fuel Product (all cuts), BGE/day	2,593	2,593	43,222	21,611	8,644	2,593	64,833
Total Liquid Fuel Product (all cuts), BJE/day	2,422	2,422	40,365	20,183	8,073	2,422	60,548
Total Liquid Fuel Product (all cuts), BDE/day	2,344	2,344	39,063	19,532	7,813	2,344	58,595
Total Liquid Product Molar H:C Ratio	1.79	1.79	1.79	1.79	1.79	1.79	1.79
Modeled Gasoline Product, wt %	24%	24%	24%	24%	24%	24%	24%
Modeled Jet Product, wt %	47%	47%	47%	47%	47%	47%	47%
Modeled Diesel/Heavies, wt %	29%	29%	29%	29%	29%	29%	29%
Hydroprocessing Reactor							
Inlet Pressure, psia	1,045	1,045	1,045	1,045	1,045	1,045	1,045
Inlet Temperature, F	644	644	644	644	644	644	644
Outlet Temperature, F	684	684	684	684	684	684	684
Carbon-Efficiency for CFP Oil, %	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%	90.0%
Carbon-Efficiency for Petroleum Feed, %	98.6%	98.6%	98.6%	98.6%	98.6%	98.6%	98.6%
WHSV proration basis from SRD using chem. H2							
Base SRD Case H2 Consumption, scf/bbl	615.0	615.0	615.0	615.0	615.0	615.0	615.0
Base SRD Case WHSV Assumption, 1/h	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrogen							
Makeup H2 Purity, mol %	99%	99%	99%	99%	99%	99%	99%
H2 Purity at Reactor Inlet, mol %	80%	80%	80%	80%	80%	80%	80%
Chemical H2 per Total Feed, scf/bbl	1,316	1,316	1,316	1,316	1,316	1,316	1,316
Chemical H2 per Fresh Feed, scf/bbl	1,645	1,645	1,645	1,645	1,645	1,645	1,645
Reactor H2 Feed per Total Feed, scf/bbl	6,448	6,448	6,448	6,448	6,448	6,448	6,448
Reactor H2 Feed per Fresh Feed, scf/bbl	8,060	8,060	8,060	8,060	8,060	8,060	8,060
Makeup H2 per Total Feed, scf/bbl	1,572	1,572	1,572	1,572	1,572	1,572	1,572
Makeup H2 per Fresh Feed, scf/bbl	1,965	1,965	1,965	1,965	1,965	1,965	1,965
Net H2 after PSA, per Total Feed, scf/bbl	1,389	1,389	1,389	1,389	1,389	1,389	1,389
Hydrotreating							
WHSV, 1/h	0.75	0.75	0.75	0.75	0.75	0.75	0.75
Catalyst Load, lb	50,757	50,757	845,944	422,972	169,189	50,757	1,268,916

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Case ID	SCALEx0.04 CAPEXx1.3 12120211	SCALEx0.04 12120211	Onstream 98% SCALEx0.67 12120211	Onstream 98% SCALEx0.33 12120211	Onstream 98% SCALEx0.13 12120211	Onstream 98% SCALEx0.04 12120211	Onstream 98% 12120211
Catalyst Replacement Period, years	2.0	2.0	2.0	2.0	2.0	2.0	2.0
Hydrocracking							
WHSV, 1/h	N/A	N/A	N/A	N/A	N/A	N/A	N/A
Catalyst Load, lb	N/A	N/A	N/A	N/A	N/A	N/A	N/A
Catalyst Replacement Period, years	N/A	N/A	N/A	N/A	N/A	N/A	N/A
TEA							
Capital							
Fixed Capital Investment (FCI), million $	59.4	45.7	211.6	137.1	82.9	45.7	275.7
Working Capital, million $	3.0	2.3	10.6	6.9	4.1	2.3	13.8
Operating Costs, $/GGE	3.16	3.15	2.99	3.00	3.03	3.15	2.99
Petroleum Feed Cost, $/GGE	1.87	1.87	1.87	1.87	1.87	1.87	1.87
CFP Oil Feed Cost, $/GGE	0.95	0.95	0.95	0.95	0.95	0.95	0.95
Net Hydrogen Makeup Cost, $/GGE	0.14	0.14	0.14	0.14	0.14	0.14	0.14
Fixed Operating Costs, $/GGE	0.19	0.17	0.02	0.03	0.06	0.17	0.02
Coproduct/Utility Credits, $/GGE	−0.03	−0.03	−0.03	−0.03	−0.03	−0.03	−0.03
Other Operating Costs, $/GGE	0.06	0.06	0.06	0.06	0.06	0.06	0.06
Capital Charges, Income Tax etc.	0.21	0.16	0.04	0.06	0.09	0.16	0.04
MFSP, with Full Capital Costs, $/GGE	3.38	3.32	3.04	3.06	3.12	3.30	3.03
MFSP, with Full Capital Costs, $/GJE	3.62	3.55	3.25	3.28	3.34	3.54	3.24
MFSP, with Full Capital Costs, $/GDE	3.74	3.67	3.36	3.39	3.45	3.66	3.35
Other TEA Metrics							
Total Operating Cost without Feed, cents/GGE	34.8	33.4	17.5	18.7	21.8	32.9	17.1
Hydrogen Cost, cents/GGE	13.8	13.8	13.8	13.8	13.8	13.8	13.8
Total Catalyst Cost per Feed Barrel, $/bbl	0.52	0.52	0.52	0.52	0.52	0.52	0.52
Delta Capital for Repurposed Unit, million $	12.5	9.6	54.7	34.0	18.5	9.6	73.0
Repurposed Case ID	12201011	12201011	12201011	12201011	12201011	12201011	12201011
MFSP with Repurposed Equipment, $/GGE	3.21	3.19	3.00	3.02	3.06	3.18	3.00
Repurposed Petroleum Unit Size, BPD	3,000	3,000	50,000	25,000	10,000	3,000	75,000
Repurposed Unit FCI, $/annual bbl fresh feed	15.0	11.5	3.9	4.9	6.7	11.5	3.5
New Unit FCI, $/annual bbl fresh feed	71.4	54.9	15.3	19.8	29.9	54.9	13.3
Hydrotreating reactor FCI, million $	2.1	2.1	23.9	11.4	6.2	2.1	35.8
Hydrocracking reactor FCI, million $	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Reactor(s) Pressure Correction On/Off	Off	Off	On	On	Off	Off	On
Makeup compressor FCI, million $	2.2	2.2	20.7	11.9	5.8	2.2	28.6
Recycle compressor FCI, million $	3.0	3.0	7.1	5.7	4.3	3.0	8.0
Fuel Lifecycle CI, g CO2e/MJ	76.2	76.2	76.0	76.0	76.0	76.2	76.0
CI for Petroleum Diesel Case, g CO2e/MJ	90.8	90.8	90.8	90.8	90.8	90.8	90.8
CI Reduction, %	16.1%	16.1%	16.3%	16.3%	16.2%	16.1%	16.3%

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Table K-15. Hydroprocessing Cases, On-Stream Factor and Scale Changes With Case 12120211

Case ID	Onstream 90% SCALEx0.67 12120211	Onstream 90% SCALEx0.33 12120211	Onstream 90% SCALEx0.13 12120211	Onstream 90% SCALEx0.04 12120211	Onstream 90% 12120211
Total Feed, BPD	50,000	25,000	10,000	3,000	75,000
SRD Feed, BPD	32,000	16,000	6,400	1,920	48,000
LCO Feed, BPD	0	0	0	0	0
VGO Feed, BPD	0	0	0	0	0
CFP Oil Feed, BPD	8,000	4,000	1,600	480	12,000
Liquid Product Recycle, BPD	10,000	5,000	2,000	600	15,000
CFP Oil in Fresh Feed, vol %	20%	20%	20%	20%	20%
Liquid Product Recycle, vol %	20%	20%	20%	20%	20%
CFP Oil Oxygen (dry basis), wt %	21.6%	21.6%	21.6%	21.6%	21.6%
Total Liquid Fuel Product, lb/h	479,559	239,779	95,912	28,774	719,338
Total Liquid Fuel Product (all cuts), BGE/day	43,222	21,611	8,644	2,593	64,833
Total Liquid Fuel Product (all cuts), BJE/day	40,365	20,183	8,073	2,422	60,548
Total Liquid Fuel Product (all cuts), BDE/day	39,063	19,532	7,813	2,344	58,595
Total Liquid Product Molar H:C Ratio	1.79	1.79	1.79	1.79	1.79
Modeled Gasoline Product, wt %	24%	24%	24%	24%	24%
Modeled Jet Product, wt %	47%	47%	47%	47%	47%
Modeled Diesel/Heavies, wt %	29%	29%	29%	29%	29%
Hydroprocessing Reactor					
Inlet Pressure, psia	1,045	1,045	1,045	1,045	1,045
Inlet Temperature, F	644	644	644	644	644
Outlet Temperature, F	684	684	684	684	684
Carbon-Efficiency for CFP Oil, %	90.0%	90.0%	90.0%	90.0%	90.0%
Carbon-Efficiency for Petroleum Feed, %	98.6%	98.6%	98.6%	98.6%	98.6%
WHSV proration basis from SRD using chem. H2					
Base SRD Case H2 Consumption, scf/bbl	615.0	615.0	615.0	615.0	615.0
Base SRD Case WHSV Assumption, 1/h	2.0	2.0	2.0	2.0	2.0
Hydrogen					
Makeup H2 Purity, mol %	99%	99%	99%	99%	99%
H2 Purity at Reactor Inlet, mol %	80%	80%	80%	80%	80%
Chemical H2 per Total Feed, scf/bbl	1,316	1,316	1,316	1,316	1,316
Chemical H2 per Fresh Feed, scf/bbl	1,645	1,645	1,645	1,645	1,645
Reactor H2 Feed per Total Feed, scf/bbl	6,448	6,448	6,448	6,448	6,448
Reactor H2 Feed per Fresh Feed, scf/bbl	8,060	8,060	8,060	8,060	8,060
Makeup H2 per Total Feed, scf/bbl	1,572	1,572	1,572	1,572	1,572
Makeup H2 per Fresh Feed, scf/bbl	1,965	1,965	1,965	1,965	1,965
Net H2 after PSA, per Total Feed, scf/bbl	1,389	1,389	1,389	1,389	1,389
Hydrotreating					
WHSV, 1/h	0.75	0.75	0.75	0.75	0.75
Catalyst Load, lb	845,944	422,972	169,189	50,757	1,268,916
Catalyst Replacement Period, years	2.0	2.0	2.0	2.0	2.0
Hydrocracking					

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Case ID	Onstream 90% SCALEx0.67 12120211	Onstream 90% SCALEx0.33 12120211	Onstream 90% SCALEx0.13 12120211	Onstream 90% SCALEx0.04 12120211	Onstream 90% 12120211
WHSV, 1/h	N/A	N/A	N/A	N/A	N/A
Catalyst Load, lb	N/A	N/A	N/A	N/A	N/A
Catalyst Replacement Period, years	N/A	N/A	N/A	N/A	N/A
TEA					
Capital					
Fixed Capital Investment (FCI), million $	211.6	137.1	82.9	45.7	275.7
Working Capital, million $	10.6	6.9	4.1	2.3	13.8
Operating Costs, $/GGE	2.99	3.01	3.04	3.16	2.99
Petroleum Feed Cost, $/GGE	1.87	1.87	1.87	1.87	1.87
CFP Oil Feed Cost, $/GGE	0.95	0.95	0.95	0.95	0.95
Net Hydrogen Makeup Cost, $/GGE	0.14	0.14	0.14	0.14	0.14
Fixed Operating Costs, $/GGE	0.02	0.03	0.07	0.18	0.02
Coproduct/Utility Credits, $/GGE	−0.03	−0.03	−0.03	−0.03	−0.03
Other Operating Costs, $/GGE	0.06	0.06	0.06	0.06	0.06
Capital Charges, Income Tax etc.	0.05	0.06	0.10	0.17	0.04
MFSP, with Full Capital Costs, $/GGE	3.04	3.07	3.14	3.33	3.03
MFSP, with Full Capital Costs, $/GJE	3.26	3.29	3.36	3.57	3.25
MFSP, with Full Capital Costs, $/GDE	3.37	3.40	3.47	3.69	3.36
Other TEA Metrics					
Total Operating Cost without Feed, cents/GGE	17.7	18.9	22.3	34.4	17.3
Hydrogen Cost, cents/GGE	13.8	13.8	13.8	13.8	13.8
Total Catalyst Cost per Feed Barrel, $/bbl	0.52	0.52	0.52	0.52	0.52
Delta Capital for Repurposed Unit, million $	54.7	34.0	18.5	9.6	73.0
Repurposed Case ID	12201011	12201011	12201011	12201011	12201011
MFSP with Repurposed Equipment, $/GGE	3.01	3.02	3.06	3.20	3.00
Repurposed Petroleum Unit Size, BPD	50,000	25,000	10,000	3,000	75,000
Repurposed Unit FCI, $/annual bbl fresh feed	3.9	4.9	6.7	11.5	3.5
New Unit FCI, $/annual bbl fresh feed	15.3	19.8	29.9	54.9	13.3
Hydrotreating reactor FCI, million $	23.9	11.4	6.2	2.1	35.8
Hydrocracking reactor FCI, million $	0.0	0.0	0.0	0.0	0.0
Reactor(s) Pressure Correction On/Off	On	On	Off	Off	On
Makeup compressor FCI, million $	20.7	11.9	5.8	2.2	28.6
Recycle compressor FCI, million $	7.1	5.7	4.3	3.0	8.0
Fuel Lifecycle CI, g CO2e/MJ	76.0	76.0	76.0	76.2	76.0
CI for Petroleum Diesel Case, g CO2e/MJ	90.8	90.8	90.8	90.8	90.8
CI Reduction, %	16.3%	16.3%	16.2%	16.1%	16.3%

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Appendix L. Stream Summaries for Four Cases

Simplified Process Flow Diagram and Stream Summaries for Four Cases

A simplified flow diagram for the hydroprocessing models is shown in Figure L-1. Subsequent stream tables refer to the stream numbers, heat duties, and power requirements tags shown in the figure. The compounds associated with component IDs listed can be found in Appendix D.

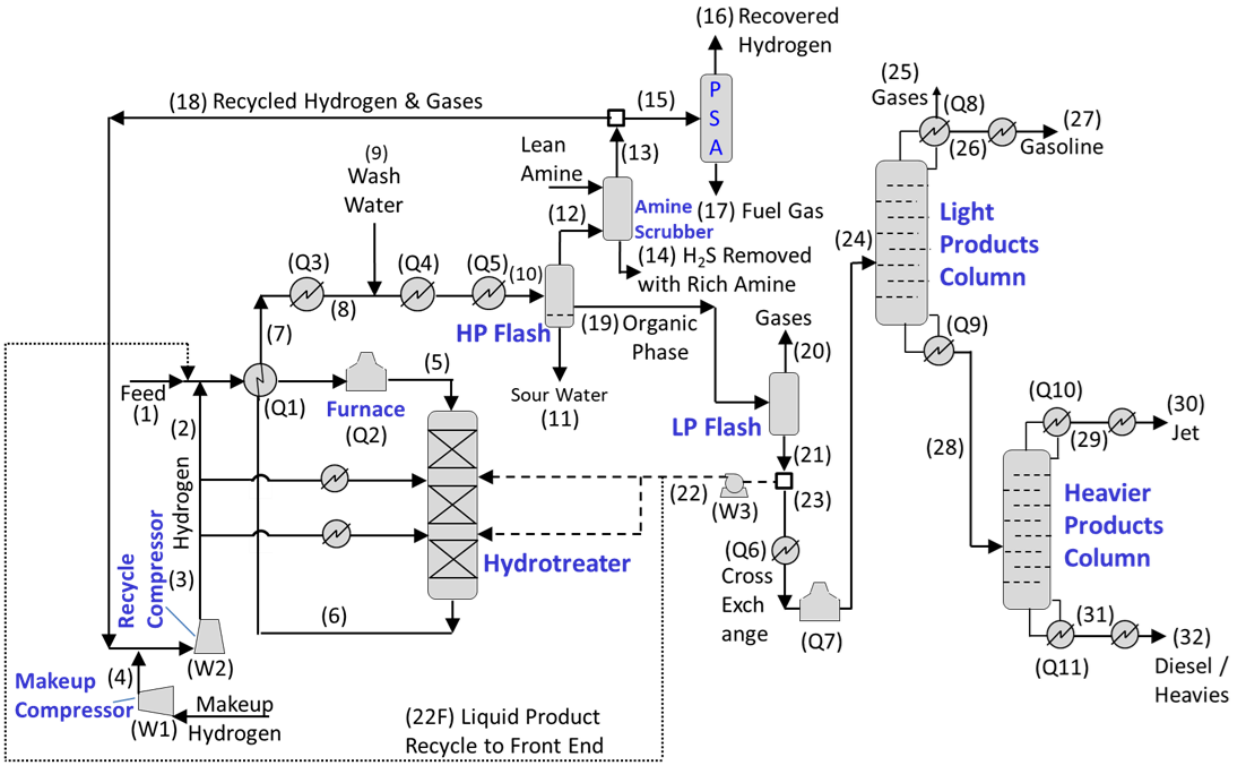

Figure L-1. Flow diagram with stream (number), heat duty (Q followed by number), and power (W followed by number) tags for model results presented in following tables in this appendix.

Four bookend cases are included to cover the boundaries of feed compositions. All flows reported in the following tables of this appendix are from 75k BPD process models.

The cases included are for hydroprocessing with inter-bed hydrogen quench and no liquid product recycle. Please refer to Table 32 for interpreting case IDs below:

- 100% petroleum SRD HT (Case 11201011) in Tables L-1 and L-2
- 100% petroleum with 50/50 LCO/VGO HT+HC (Case 41201011) in Tables L-3 and L-4
- 100% CFP-B Oil HT (Case 11199011) in Tables L-5 and L-6
- 100% CFP-A Oil HT (Case 12199011) in Tables L-7 and L-8

Notes: (i) A negligible 0.0001 vol % of the other category of feed (either CFP oil or petroleum feed) were introduced in the 100% cases to avoid simulation warnings. (ii) For the *100% CFP cases* Stream 14 shows NH_3 and CO_2. The NH_3 in Stream 14 should be titrated out as a

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carbonate salt in Stream 11. Stream 14 should have zero NH_3 and a corresponding reduction of the titrated CO_2. Potential acids due to the presence of CO_2 are not accounted for in Stream 11.

Stream, Heat Duty, Power Requirements Summary for 100% SRD (11201011)

Table L-1. Stream Flows for 100% SRD HT (Case 11201011)

Stream →	(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)
Temp, °F	92.4	142.9	142.9	140.0	644.0	683.5	361.7	276.7
Pres, psia	1100	1100	1100	995.00002	1045	1015	1010	1005
Total Flow	920080.58	48967.35	144021.62	49942.638	969047.02	1064101.3	1064101.3	1064101.3
HYDROGEN	2.709E-05	12814.265	37689.014	15301.39	12814.071	27522.75	27522.75	27522.75
AMMONIA	0	1.015E-19	2.985E-19	0	4.873E-15	1567.164	1567.164	1567.164
H2S	0	7.677E-18	2.258E-17	0	3.686E-13	9591.2913	9591.2913	9591.2913
WATER	0.0095204	147.07411	432.57091	0	147.08447	432.78654	432.78654	432.78654
CO	0.0001235	0	0	0	0.0001235	0	0	0
CO_2	0.001016	0	0	0	0.001016	0.001016	0.001016	0.001016
METHANE	8.798E-05	16539.649	48646.027	10744.535	16539.142	49987.228	49987.228	49987.228
ETHYLENE	0.0001172	0.0039017	0.0114755	0	0.0040189	0.0184346	0.0184346	0.0184346
ACALDEHY	0.0073147	0	0	0	0.0073147	0	0	0
ACETACID	5.897E-10	0	0	0	5.897E-10	0	0	0
ETHANE	0.0001066	11462.782	33714.064	13425.965	11462.511	35807.063	35807.063	35807.063
PROPYLEN	0.0004741	8.776E-06	2.581E-05	0	0.0004829	6.748E-05	6.748E-05	6.748E-05
ACETONE	0.0126477	0	0	0	0.0126477	0	0	0
PROPANE	0.0001774	5212.3879	15330.553	7875.5627	5212.4305	20998.033	20998.033	20998.033
FURAN	0.0020856	0	0	0	0.0020856	0	0	0
2BUTENAL	0.0031142	0	0	0	0.0031142	0	0	0
BUTENE	0.0003908	3.511E-06	1.033E-05	0	0.0003943	5.349E-05	5.349E-05	5.349E-05
THF	0.0018007	0	0	0	0.0018007	0	0	0
2BUTNONE	0.0070162	0	0	0	0.0070162	0	0	0
1BUTANAL	0.002299	0	0	0	0.002299	0	0	0
NBUTANE	0	827.67479	2434.3376	865.06161	827.67413	8502.5891	8502.5891	8502.5891
P-IC4	0	1848.397	5436.4616	1730.1232	1848.407	17247.52	17247.52	17247.52
2MEFURAN	0.007166	0	0	0	0.007166	0	0	0
3MEFURAN	0.0011353	0	0	0	0.0011353	0	0	0
CYPNTENO	0.0018906	0	0	0	0.0018906	0	0	0
CYPNTANO	0.0043652	0	0	0	0.0043652	0	0	0
2PNTENAL	0.0059755	0	0	0	0.0059755	0	0	0
3PNTE2ON	0.0018075	0	0	0	0.0018075	0	0	0
2PENTENE	0	2.842E-05	8.36E-05	0	2.843E-05	0.001184	0.001184	0.001184
2PENTNON	0.0025236	0	0	0	0.0025236	0	0	0
3PENTNON	0.0015262	0	0	0	0.0015262	0	0	0
PENTANE	0.0001287	0.0005803	0.0017067	0	0.000709	0.0234721	0.0234721	0.0234721
BENZENE	0.0034481	0.0003119	0.0009173	0	0.00376	0.0366895	0.0366895	0.0366895
PHENOL	0.0586712	1.162E-25	3.417E-25	0	0.0586712	0	0	0
12BZDIOL	0.0088288	0	0	0	0.0088288	0	0	0
CYHXDIEN	0.0001115	0	0	0	0.0001115	0	0	0
DIMEFURA	0.0029189	1.086E-06	3.195E-06	0	0.00292	0.0002161	0.0002161	0.0002161
METCY5ON	0.0013683	0	0	0	0.0013683	0	0	0
2ETFURAN	0.002633	0	0	0	0.002633	0	0	0
CYHXENE	0	8.492E-07	2.498E-06	0	8.493E-07	0.0001168	0.0001168	0.0001168
2MECY5ON	0.0018052	0	0	0	0.0018052	0	0	0
1HEXANAL	0.0017517	0	0	0	0.0017517	0	0	0
2HEXNONE	0.0008445	0	0	0	0.0008445	0	0	0
3HEXNONE	0.0007792	0	0	0	0.0007792	0	0	0
CYHEXANE	0.0088221	0.0003786	0.0011136	0	0.0092008	0.0501178	0.0501178	0.0501178
NHEXANE	0	0.0004377	0.0012875	0	0.0004377	0.0438469	0.0438469	0.0438469

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Stream →	(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)
TOLUENE	0.0025755	1.308E-06	3.848E-06	0	0.0025768	0.0003539	0.0003539	0.0003539
OCRESOL	0.002855	0	0	0	0.002855	0	0	0
GUAIACOL	0.0426492	0	0	0	0.0426492	0	0	0
MECYHEXN	0.0010128	7.294E-06	2.145E-05	0	0.00102	0.0018111	0.0018111	0.0018111
2M3HXENE	0	0.0001505	0.0004427	0	0.0001505	0.0334303	0.0334303	0.0334303
METHHEXN	0	0.0001038	0.0003052	0	0.0001038	0.019318	0.019318	0.019318
3ETPNTNO	0.0036838	0	0	0	0.0036838	0	0	0
D-INDOLE	3459.3821	0.001599	0.0047028	0	3459.3837	46.637108	46.637108	46.637108
OXYLENE	0.0302823	1.657E-05	4.873E-05	0	0.0302988	0.0163698	0.0163698	0.0163698
ETYBENZN	0	3.9970189	11.755938	0	3.9970456	3088.6905	3088.6905	3088.6905
23XYLNOL	0.0498463	0	0	0	0.0498463	0	0	0
SYRINGOL	0.0120934	0	0	0	0.0120934	0	0	0
VANALC	0.0195095	0	0	0	0.0195095	0	0	0
TRIMEPYR	3459.3895	0.0169706	0.0499134	0	3459.4065	46.638305	46.638305	46.638305
MEETCY5E	0.0004595	4.374E-06	1.286E-05	0	0.0004639	0.0039607	0.0039607	0.0039607
DIMECYHX	0.0003661	4.604E-05	0.0001354	0	0.0004122	0.0258966	0.0258966	0.0258966
TRDICYHN	0.0010212	5.973E-05	0.0001757	0	0.001081	0.028264	0.028264	0.028264
ETCYHEXN	0	2.182E-05	6.417E-05	0	2.182E-05	0.0135134	0.0135134	0.0135134
4MEHEPTN	0	6.9945654	20.572251	0	6.9946168	3221.0187	3221.0187	3221.0187
MEBZFURA	0.0100215	0	0	0	0.0100215	0	0	0
MEVINPHE	0.0003019	0	0	0	0.0003019	0	0	0
PROPBENZ	0.0069226	4.845E-06	1.425E-05	0	0.0069275	0.008728	0.008728	0.008728
ETMEBENZ	0.0006418	3.089E-07	9.084E-07	0	0.0006421	0.0006427	0.0006427	0.0006427
TRIMEBEN	0.0004792	1.516E-07	4.46E-07	0	0.0004794	0.0004796	0.0004796	0.0004796
PROPCYHX	0	1.46E-05	4.293E-05	0	1.46E-05	0.0195232	0.0195232	0.0195232
NAPTHALE	0.0260192	2.757E-08	8.109E-08	0	0.0260192	0.0002417	0.0002417	0.0002417
1NAPTHOL	0.0252745	0	0	0	0.0252745	0	0	0
CONIFALD	0.0260545	0	0	0	0.0260545	0	0	0
TETRALIN	101878.34	8.6041787	25.306408	0	101886.94	56887.059	56887.059	56887.059
IEUGENOL	0.0277852	0	0	0	0.0277852	0	0	0
D-DECALN	0	11.084496	32.601459	0	11.084533	44009.061	44009.061	44009.061
DECALIN	0.0133396	4.413E-06	1.298E-05	0	0.0133441	0.0330709	0.0330709	0.0330709
D-C4CYC6	125649.95	42.575387	125.22173	0	125692.53	122119.26	122119.26	122119.26
DIMEOCTN	0	5.571E-06	1.639E-05	0	5.571E-06	0.0104944	0.0104944	0.0104944
1MENAPHT	92539.496	1.9291631	5.674009	0	92541.425	51661.442	51661.442	51661.442
METDEC	0.0045895	3.0297504	8.9110306	0	3.0343546	40908.029	40908.029	40908.029
DIBZTHIO	0.001903	0	0	0	0.001903	0	0	0
D-C4BZ-S	28627.867	0.0005465	0.0016075	0	28627.868	385.94104	385.94104	385.94104
BIPHENYL	0	1.199E-07	3.526E-07	0	1.199E-07	0.0059369	0.0059369	0.0059369
DIMENAPH	0	1.778E-08	5.229E-08	0	1.778E-08	0.0012855	0.0012855	0.0012855
D-C12H18	0	1.5708776	4.6202281	0	1.5708939	23963.534	23963.534	23963.534
D-C12H1B	0	0.2178582	0.6407595	0	0.2178605	2974.5921	2974.5921	2974.5921
2ETDECLN	0	5.318E-08	1.564E-07	0	5.319E-08	0.0007345	0.0007345	0.0007345
D-CQUINL	3459.3821	1.466E-05	4.312E-05	0	3459.3822	46.636985	46.636985	46.636985
HXLCRESO	0.0052768	0	0	0	0.0052768	0	0	0
HEPCYHEX	0	6.719E-07	1.976E-06	0	6.719E-07	0.0330912	0.0330912	0.0330912
D-2C2BZS	28627.867	3.665E-05	0.0001078	0	28627.867	385.941	385.941	385.941
D-C14H14	0	0.0728216	0.2141812	0	0.0728226	24121.396	24121.396	24121.396
ETPHCYHX	0	7.306E-08	2.149E-07	0	7.306E-08	0.0073303	0.0073303	0.0073303
D-HYPHEN	54335.115	0.3318159	0.9759291	0	54335.446	52755.678	52755.678	52755.678
THYDPHEN	0.0109926	2.889E-07	8.498E-07	0	0.0109929	0.0286174	0.0286174	0.0286174
DIMDODEC	0	8.305E-06	2.443E-05	0	8.305E-06	0.2712905	0.2712905	0.2712905
D-3CZOLE	3459.3821	5.69E-06	1.674E-05	0	3459.3821	46.636984	46.636984	46.636984
D-C15H16	0	0.0049768	0.0146376	0	0.0049769	3184.4922	3184.4922	3184.4922
D-3C3BEN	140082.72	1.6904566	4.971931	0	140084.41	78200.809	78200.809	78200.809
OCHYENAP	0.0116495	1.143E-08	3.362E-08	0	0.0116495	0.0085097	0.0085097	0.0085097

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Stream →	(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)
D-DIMTRD	0	7.39E-07	2.173E-06	0	7.39E-07	0.0376732	0.0376732	0.0376732
DIMPHENA	0.009408	2.263E-10	6.657E-10	0	0.009408	0.0025483	0.0025483	0.0025483
CETANE	0	1.111E-07	3.266E-07	0	1.111E-07	0.0443249	0.0443249	0.0443249
CYHXPHON	0.0844325	0	0	0	0.0844325	0	0	0
CYHXHXBZ	0	2.752E-09	8.095E-09	0	2.752E-09	0.0175158	0.0175158	0.0175158
D-NC18	129045.9	0.0853107	0.2509138	0	129045.98	125292.69	125292.69	125292.69
D-C1C17	86596.589	0.0500844	0.147307	0	86596.639	84077.968	84077.968	84077.968
D-NC20	10187.834	0.0019435	0.0057162	0	10187.836	9891.5142	9891.5142	9891.5142
HMWLIGNB	0.1328401	0	0	0	0.1328401	0	0	0
TRIPHNPR	0	3.769E-12	1.108E-11	0	3.769E-12	0.0001731	0.0001731	0.0001731
MOXHXPHE	0.0853743	0	0	0	0.0853743	0	0	0
DIBZFLOR	0.107389	4.567E-11	1.343E-10	0	0.107389	0.0108875	0.0108875	0.0108875
HXACMOXP	0.2001781	0	0	0	0.2001781	0	0	0
MOXHEPOX	0.0178464	0	0	0	0.0178464	0	0	0
P-3C3CY6	108670.23	32.854737	96.63158	0	108703.08	165128.34	165128.34	165128.34

Table L-1 (continued). Stream Flows for 100% SRD HT (Case 11201011)

Stream →	(9)	(10)	(11)	(12)	(13)	(14)	(15)	(16)
Temp, °F	110.0	110.0	109.4	110.0	110.0	110.0	110.0	110.0
Pres, psia	1050	995	993	993	993	60	993	973
Total Flow	247091.98	1311193.3	248608.65	119002.26	112547.2	6455.0539	18465.985	3735.1455
HYDROGEN	0	27522.75	0.6785054	26782.483	26782.483	0	4394.2888	3735.1455
AMMONIA	0	1567.164	1567.164	0	0	0	0	0
H2S	0	9591.2913	3136.2384	6455.0529	0	6455.0529	0	0
WATER	247091.98	247524.77	243899.83	517.48667	517.48667	0	84.905715	0
CO	0	0	0	0	0	0	0	0
CO2	0	0.001016	0	0.001016	0	0.001016	0	0
METHANE	0	49987.228	2.8912986	45340.756	45340.756	0	7439.2047	0
ETHYLENE	0	0.0184346	1.494E-06	0.013728	0.013728	0	0.0022524	0
ACALDEHY	0	0	0	0	0	0	0	0
ACETACID	0	0	0	0	0	0	0	0
ETHANE	0	35807.063	1.7581652	24271.834	24271.834	0	3982.3584	0
PROPYLEN	0	6.748E-05	1.091E-09	3.088E-05	3.088E-05	0	5.066E-06	0
ACETONE	0	0	0	0	0	0	0	0
PROPANE	0	20998.033	0.0861138	8918.3882	8918.3882	0	1463.2688	0
FURAN	0	0	0	0	0	0	0	0
2BUTENAL	0	0	0	0	0	0	0	0
BUTENE	0	5.349E-05	4.585E-11	1.235E-05	1.235E-05	0	2.027E-06	0
THF	0	0	0	0	0	0	0	0
2BUTNONE	0	0	0	0	0	0	0	0
1BUTANAL	0	0	0	0	0	0	0	0
NBUTANE	0	8502.5891	0.0019038	1877.3226	1877.3226	0	308.01841	0
P-IC4	0	17247.52	0.0018503	4433.8782	4433.8782	0	727.48077	0
2MEFURAN	0	0	0	0	0	0	0	0
3MEFURAN	0	0	0	0	0	0	0	0
CYPNTENO	0	0	0	0	0	0	0	0
CYPNTANO	0	0	0	0	0	0	0	0
2PNTENAL	0	0	0	0	0	0	0	0
3PNTE2ON	0	0	0	0	0	0	0	0
2PENTENE	0	0.001184	5.106E-11	0.0001	0.0001	0	1.641E-05	0
2PENTNON	0	0	0	0	0	0	0	0
3PENTNON	0	0	0	0	0	0	0	0
PENTANE	0	0.0234721	1.313E-10	0.0020417	0.0020417	0	0.000335	0
BENZENE	0	0.0366895	6.678E-07	0.0010974	0.0010974	0	0.00018	0

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Stream →	(9)	(10)	(11)	(12)	(13)	(14)	(15)	(16)
PHENOL	0	0	0	0	0	0	0	0
12BZDIOL	0	0	0	0	0	0	0	0
CYHXDIEN	0	0	0	0	0	0	0	0
DIMEFURA	0	0.0002161	1.34E-10	3.823E-06	3.823E-06	0	6.272E-07	0
METCY5ON	0	0	0	0	0	0	0	0
2ETFURAN	0	0	0	0	0	0	0	0
CYHXENE	0	0.0001168	1.332E-10	2.988E-06	2.988E-06	0	4.903E-07	0
2MECY5ON	0	0	0	0	0	0	0	0
1HEXANAL	0	0	0	0	0	0	0	0
2HEXNONE	0	0	0	0	0	0	0	0
3HEXNONE	0	0	0	0	0	0	0	0
CYHEXANE	0	0.0501178	1.115E-08	0.0013322	0.0013322	0	0.0002186	0
NHEXANE	0	0.0438469	3.986E-12	0.0015402	0.0015402	0	0.0002527	0
TOLUENE	0	0.0003539	9.538E-11	4.603E-06	4.603E-06	0	7.553E-07	0
OCRESOL	0	0	0	0	0	0	0	0
GUAIACOL	0	0	0	0	0	0	0	0
MECYHEXN	0	0.0018111	2.922E-12	2.567E-05	2.567E-05	0	4.211E-06	0
2M3HXENE	0	0.0334303	3.091E-13	0.0005296	0.0005296	0	8.689E-05	0
METHHEXN	0	0.019318	2.383E-14	0.0003651	0.0003651	0	5.99E-05	0
3ETPNTNO	0	0	0	0	0	0	0	0
D-INDOLE	0	46.637108	5.06E-05	0.005626	0.005626	0	0.0009231	0
OXYLENE	0	0.0163698	3.108E-10	5.83E-05	5.83E-05	0	9.565E-06	0
ETYBENZN	0	3088.6905	1.641E-05	14.063663	14.063663	0	2.3074708	0
23XYLNOL	0	0	0	0	0	0	0	0
SYRINGOL	0	0	0	0	0	0	0	0
VANALC	0	0	0	0	0	0	0	0
TRIMEPYR	0	46.638305	1.528E-08	0.059712	0.059712	0	0.0097971	0
MEETCY5E	0	0.0039607	4.424E-12	1.539E-05	1.539E-05	0	2.525E-06	0
DIMECYHX	0	0.0258966	2.383E-14	0.000162	0.000162	0	2.658E-05	0
TRDICYHN	0	0.028264	3.744E-14	0.0002102	0.0002102	0	3.448E-05	0
ETCYHEXN	0	0.0135134	5.782E-14	7.677E-05	7.677E-05	0	1.26E-05	0
4MEHEPTN	0	3221.0187	6.822E-11	24.610642	24.610642	0	4.0379477	0
MEBZFURA	0	0	0	0	0	0	0	0
MEVINPHE	0	0	0	0	0	0	0	0
PROPBENZ	0	0.008728	5.157E-13	1.705E-05	1.705E-05	0	2.797E-06	0
ETMEBENZ	0	0.0006427	7.006E-14	1.087E-06	1.087E-06	0	1.783E-07	0
TRIMEBEN	0	0.0004796	4.946E-13	5.335E-07	5.335E-07	0	8.754E-08	0
PROPCYHX	0	0.0195232	1.085E-15	5.136E-05	5.136E-05	0	8.427E-06	0
NAPTHALE	0	0.0002417	1.362E-11	9.701E-08	9.701E-08	0	1.592E-08	0
1NAPTHOL	0	0	0	0	0	0	0	0
CONIFALD	0	0	0	0	0	0	0	0
TETRALIN	0	56887.059	0.0001515	30.274201	30.274201	0	4.9671864	0
IEUGENOL	0	0	0	0	0	0	0	0
D-DECALN	0	44009.061	5.244E-07	39.001191	39.001191	0	6.3990517	0
DECALIN	0	0.0330709	1.335E-20	1.553E-05	1.553E-05	0	2.548E-06	0
D-C4CYC6	0	122119.26	3.15E-11	149.80315	149.80315	0	24.578688	0
DIMEOCTN	0	0.0104944	1.742E-20	1.96E-05	1.96E-05	0	3.216E-06	0
1MENAPHT	0	51661.442	4.604E-05	6.7878458	6.7878458	0	1.1137039	0
METDEC	0	40908.029	9.02E-10	10.660283	10.660283	0	1.7490673	0
DIBZTHIO	0	0	0	0	0	0	0	0
D-C4BZ-S	0	385.94104	4.279E-11	0.001923	0.001923	0	0.0003155	0
BIPHENYL	0	0.0059369	9.36E-13	4.218E-07	4.218E-07	0	6.921E-08	0
DIMENAPH	0	0.0012855	1.225E-14	6.256E-08	6.256E-08	0	1.026E-08	0
D-C12H18	0	23963.534	2.303E-12	5.5271929	5.5271929	0	0.9068645	0
D-C12H1B	0	2974.5921	3.323E-13	0.7665426	0.7665426	0	0.1257691	0
2ETDECLN	0	0.0007345	5.066E-20	1.871E-07	1.871E-07	0	3.07E-08	0

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Stream →	(9)	(10)	(11)	(12)	(13)	(14)	(15)	(16)
D-CQUINL	0	46.636985	2.638E-13	5.159E-05	5.159E-05	0	8.464E-06	0
HXLCRESO	0	0	0	0	0	0	0	0
HEPCYHEX	0	0.0330912	2.542E-31	2.364E-06	2.364E-06	0	3.879E-07	0
D-2C2BZS	0	385.941	4.709E-11	0.000129	0.000129	0	2.116E-05	0
D-C14H14	0	24121.396	2.222E-09	0.2562256	0.2562256	0	0.0420398	0
ETPHCYHX	0	0.0073303	9.652E-21	2.571E-07	2.571E-07	0	4.218E-08	0
D-HYPHEN	0	52755.678	1.294E-11	1.167507	1.167507	0	0.1915567	0
THYDPHEN	0	0.0286174	1.358E-21	1.017E-06	1.017E-06	0	1.668E-07	0
DIMDODEC	0	0.2712905	2.983E-31	2.922E-05	2.922E-05	0	4.795E-06	0
D-3CZOLE	0	46.636984	6.374E-13	2.002E-05	2.002E-05	0	3.285E-06	0
D-C15H16	0	3184.4922	6.723E-15	0.017511	0.017511	0	0.0028731	0
D-3C3BEN	0	78200.809	3.724E-18	5.9479407	5.9479407	0	0.9758979	0
OCHYENAP	0	0.0085097	4.672E-28	4.022E-08	4.022E-08	0	6.6E-09	0
D-DIMTRD	0	0.0376732	8.275E-36	2.6E-06	2.6E-06	0	4.266E-07	0
DIMPHENA	0	0.0025483	2.209E-17	7.964E-10	7.964E-10	0	1.307E-10	0
CETANE	0	0.0443249	1.268E-38	3.908E-07	3.908E-07	0	6.411E-08	0
CYHXPHON	0	0	0	0	0	0	0	0
CYHXHXBZ	0	0.0175158	2.071E-32	9.684E-09	9.684E-09	0	1.589E-09	0
D-NC18	0	125292.69	1.008E-39	0.3001688	0.3001688	0	0.0492497	0
D-C1C17	0	84077.968	2.529E-37	0.1762237	0.1762237	0	0.0289136	0
D-NC20	0	9891.5142	1.985E-46	0.0068383	0.0068383	0	0.001122	0
HMWLIGNB	0	0	0	0	0	0	0	0
TRIPHNPR	0	0.0001731	2.053E-32	1.326E-11	1.326E-11	0	2.176E-12	0
MOXHXPHE	0	0	0	0	0	0	0	0
DIBZFLOR	0	0.0108875	2.617E-27	1.607E-10	1.607E-10	0	2.636E-11	0
HXACMOXP	0	0	0	0	0	0	0	0
MOXHEPOX	0	0	0	0	0	0	0	0
P-3C3CY6	0	165128.34	3.197E-28	115.60061	115.60061	0	18.966966	0

Table L-1 (continued). Stream Flows for 100% SRD HT (Case 11201011)

Stream →	(17)	(18)	(19)	(20)	(21)	(22)	(23)	(24)
Temp, °F	110.0	110.0	110.0	110.0	110.0	110.0	110.0	450.0
Pres, psia	22	993	993	53	53	1050	53	45.5
Total Flow	14730.839	94078.981	943582.36	22500.492	921081.87	1.802E-05	921081.87	921081.87
HYDROGEN	659.14332	22387.624	739.5888	724.78557	14.803227	0	14.803227	14.803227
AMMONIA	0	2.985E-19	0	0	0	0	0	0
H2S	0	2.258E-17	0	0	0	0	0	0
WATER	84.905715	432.57091	3107.4522	415.77579	2691.6764	1.802E-05	2691.6764	2691.6764
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
METHANE	7439.2047	37901.492	4643.5807	4278.9146	364.66603	0	364.66603	364.66603
ETHYLENE	0.0022524	0.0114755	0.0047051	0.0035742	0.0011309	0	0.0011309	0.0011309
ACALDEHY	0	0	0	0	0	0	0	0
ACETACID	0	0	0	0	0	0	0	0
ETHANE	3982.3584	20288.099	11533.471	7906.5159	3626.9551	0	3626.9551	3626.9551
PROPYLEN	5.066E-06	2.581E-05	3.66E-05	1.619E-05	2.04E-05	0	2.04E-05	2.04E-05
ACETONE	0	0	0	0	0	0	0	0
PROPANE	1463.2688	7454.99	12079.559	4893.6241	7185.9347	0	7185.9347	7185.9347
FURAN	0	0	0	0	0	0	0	0
2BUTENAL	0	0	0	0	0	0	0	0
BUTENE	2.027E-06	1.033E-05	4.114E-05	8.423E-06	3.272E-05	0	3.272E-05	3.272E-05
THF	0	0	0	0	0	0	0	0

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Stream →	(17)	(18)	(19)	(20)	(21)	(22)	(23)	(24)
2BUTNONE	0	0	0	0	0	0	0	0
1BUTANAL	0	0	0	0	0	0	0	0
NBUTANE	308.01841	1569.276	6625.2645	1174.768	5450.4965	0	5450.4965	5450.4965
P-IC4	727.48077	3706.3384	12813.64	2933.763	9879.8765	0	9879.8765	9879.8765
2MEFURAN	0	0	0	0	0	0	0	0
3MEFURAN	0	0	0	0	0	0	0	0
CYPNTENO	0	0	0	0	0	0	0	0
CYPNTANO	0	0	0	0	0	0	0	0
2PNTENAL	0	0	0	0	0	0	0	0
3PNTE2ON	0	0	0	0	0	0	0	0
2PENTENE	1.641E-05	8.36E-05	0.0010839	7.091E-05	0.001013	0	0.001013	0.001013
2PENTNON	0	0	0	0	0	0	0	0
3PENTNON	0	0	0	0	0	0	0	0
PENTANE	0.000335	0.0017067	0.0214304	0.0014302	0.0200002	0	0.0200002	0.0200002
BENZENE	0.00018	0.0009173	0.0355915	0.0007765	0.034815	0	0.034815	0.034815
PHENOL	0	3.417E-25	0	0	0	0	0	0
12BZDIOL	0	0	0	0	0	0	0	0
CYHXDIEN	0	0	0	0	0	0	0	0
DIMEFURA	6.272E-07	3.195E-06	0.0002122	2.542E-06	0.0002097	0	0.0002097	0.0002097
METCY5ON	0	0	0	0	0	0	0	0
2ETFURAN	0	0	0	0	0	0	0	0
CYHXENE	4.903E-07	2.498E-06	0.0001138	2.048E-06	0.0001117	0	0.0001117	0.0001117
2MECY5ON	0	0	0	0	0	0	0	0
1HEXANAL	0	0	0	0	0	0	0	0
2HEXNONE	0	0	0	0	0	0	0	0
3HEXNONE	0	0	0	0	0	0	0	0
CYHEXANE	0.0002186	0.0011136	0.0487856	0.0009097	0.0478759	0	0.0478759	0.0478759
NHEXANE	0.0002527	0.0012875	0.0423067	0.0010051	0.0413016	0	0.0413016	0.0413016
TOLUENE	7.553E-07	3.848E-06	0.0003493	2.477E-06	0.0003468	0	0.0003468	0.0003468
OCRESOL	0	0	0	0	0	0	0	0
GUAIACOL	0	0	0	0	0	0	0	0
MECYHEXN	4.211E-06	2.145E-05	0.0017855	1.61E-05	0.0017694	0	0.0017694	0.0017694
2M3HXENE	8.689E-05	0.0004427	0.0329007	0.0003231	0.0325776	0	0.0325776	0.0325776
METHHEXN	5.99E-05	0.0003052	0.0189529	0.0002225	0.0187304	0	0.0187304	0.0187304
3ETPNTNO	0	0	0	0	0	0	0	0
D-INDOLE	0.0009231	0.0047028	46.631432	0.0024899	46.628942	0	46.628942	46.628942
OXYLENE	9.565E-06	4.873E-05	0.0163115	3.332E-05	0.0162781	0	0.0162781	0.0162781
ETYBENZN	2.3074708	11.755938	3074.6268	8.1278752	3066.499	0	3066.499	3066.499
23XYLNOL	0	0	0	0	0	0	0	0
SYRINGOL	0	0	0	0	0	0	0	0
VANALC	0	0	0	0	0	0	0	0
TRIMEPYR	0.0097971	0.0499134	46.578593	0.0304432	46.548149	0	46.548149	46.548149
MEETCY5E	2.525E-06	1.286E-05	0.0039454	8.682E-06	0.0039367	0	0.0039367	0.0039367
DIMECYHX	2.658E-05	0.0001354	0.0257346	9.036E-05	0.0256442	0	0.0256442	0.0256442
TRDICYHN	3.448E-05	0.0001757	0.0280539	0.0001192	0.0279347	0	0.0279347	0.0279347
ETCYHEXN	1.26E-05	6.417E-05	0.0134366	4.295E-05	0.0133937	0	0.0133937	0.0133937
4MEHEPTN	4.0379477	20.572251	3196.4081	13.583366	3182.8247	0	3182.8247	3182.8247
MEBZFURA	0	0	0	0	0	0	0	0
MEVINPHE	0	0	0	0	0	0	0	0
PROPBENZ	2.797E-06	1.425E-05	0.008711	8.896E-06	0.0087021	0	0.0087021	0.0087021
ETMEBENZ	1.783E-07	9.084E-07	0.0006416	5.627E-07	0.000641	0	0.000641	0.000641
TRIMEBEN	8.754E-08	4.46E-07	0.0004791	2.71E-07	0.0004788	0	0.0004788	0.0004788
PROPCYHX	8.427E-06	4.293E-05	0.0194718	2.615E-05	0.0194457	0	0.0194457	0.0194457
NAPTHALE	1.592E-08	8.109E-08	0.0002416	4.653E-08	0.0002415	0	0.0002415	0.0002415
1NAPTHOL	0	0	0	0	0	0	0	0
CONIFALD	0	0	0	0	0	0	0	0

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Stream →	(17)	(18)	(19)	(20)	(21)	(22)	(23)	(24)
TETRALIN	4.9671864	25.306408	56856.785	12.20902	56844.576	0	56844.576	56844.576
IEUGENOL	0	0	0	0	0	0	0	0
D-DECALN	6.3990517	32.601459	43970.06	18.829549	43951.23	0	43951.23	43951.23
DECALIN	2.548E-06	1.298E-05	0.0330554	6.288E-06	0.0330491	0	0.0330491	0.0330491
D-C4CYC6	24.578688	125.22173	121969.45	68.862311	121900.59	0	121900.59	121900.59
DIMEOCTN	3.216E-06	1.639E-05	0.0104748	9.006E-06	0.0104658	0	0.0104658	0.0104658
1MENAPHT	1.1137039	5.674009	51654.654	2.9725644	51651.681	0	51651.681	51651.681
METDEC	1.7490673	8.9110306	40897.369	4.4388069	40892.93	0	40892.93	40892.93
DIBZTHIO	0	0	0	0	0	0	0	0
D-C4BZ-S	0.0003155	0.0016075	385.93912	0.0005798	385.93854	0	385.93854	385.93854
BIPHENYL	6.921E-08	3.526E-07	0.0059365	1.676E-07	0.0059363	0	0.0059363	0.0059363
DIMENAPH	1.026E-08	5.229E-08	0.0012854	2.36E-08	0.0012854	0	0.0012854	0.0012854
D-C12H18	0.9068645	4.6202281	23958.006	2.191445	23955.815	0	23955.815	23955.815
D-C12H1B	0.1257691	0.6407595	2973.8255	0.3071101	2973.5184	0	2973.5184	2973.5184
2ETDECLN	3.07E-08	1.564E-07	0.0007343	7.464E-08	0.0007342	0	0.0007342	0.0007342
D-CQUINL	8.464E-06	4.312E-05	46.636933	1.341E-05	46.63692	0	46.63692	46.63692
HXLCRESO	0	0	0	0	0	0	0	0
HEPCYHEX	3.879E-07	1.976E-06	0.0330888	6.85E-07	0.0330881	0	0.0330881	0.0330881
D-2C2BZS	2.116E-05	0.0001078	385.94087	3.096E-05	385.94084	0	385.94084	385.94084
D-C14H14	0.0420398	0.2141812	24121.14	0.0822434	24121.058	0	24121.058	24121.058
ETPHCYHX	4.218E-08	2.149E-07	0.0073301	8.431E-08	0.00733	0	0.00733	0.00733
D-HYPHEN	0.1915567	0.9759291	52754.511	0.3822616	52754.129	0	52754.129	52754.129
THYDPHEN	1.668E-07	8.498E-07	0.0286164	3.26E-07	0.028616	0	0.028616	0.028616
DIMDODEC	4.795E-06	2.443E-05	0.2712612	8.646E-06	0.2712526	0	0.2712526	0.2712526
D-3CZOLE	3.285E-06	1.674E-05	46.636964	4.834E-06	46.636959	0	46.636959	46.636959
D-C15H16	0.0028731	0.0146376	3184.4747	0.0049272	3184.4698	0	3184.4698	3184.4698
D-3C3BEN	0.9758979	4.971931	78194.862	1.9365383	78192.925	0	78192.925	78192.925
OCHYENAP	6.6E-09	3.362E-08	0.0085097	9.869E-09	0.0085097	0	0.0085097	0.0085097
D-DIMTRD	4.266E-07	2.173E-06	0.0376706	6.973E-07	0.0376699	0	0.0376699	0.0376699
DIMPHENA	0	6.657E-10	0.0025483	1.866E-10	0.0025483	0	0.0025483	0.0025483
CETANE	6.411E-08	3.266E-07	0.0443245	8.4E-08	0.0443244	0	0.0443244	0.0443244
CYHXPHON	0	0	0	0	0	0	0	0
CYHXHXBZ	0	8.095E-09	0.0175158	1.836E-09	0.0175158	0	0.0175158	0.0175158
D-NC18	0.0492497	0.2509138	125292.39	0.0510707	125292.34	0	125292.34	125292.34
D-C1C17	0.0289136	0.147307	84077.792	0.030922	84077.761	0	84077.761	84077.761
D-NC20	0.001122	0.0057162	9891.5073	0.0009178	9891.5064	0	9891.5064	9891.5064
HMWLIGNB	0	0	0	0	0	0	0	0
TRIPHNPR	0	1.108E-11	0.0001731	2.198E-12	0.0001731	0	0.0001731	0.0001731
MOXHXPHE	0	0	0	0	0	0	0	0
DIBZFLOR	0	1.343E-10	0.0108875	2.446E-11	0.0108875	0	0.0108875	0.0108875
HXACMOXP	0	0	0	0	0	0	0	0
MOXHEPOX	0	0	0	0	0	0	0	0
P-3C3CY6	18.966966	96.63158	165012.74	38.291425	164974.45	0	164974.45	164974.45

Table L-1 (continued). Stream Flows for 100% SRD HT (Case 11201011)

Stream →	(25)	(26)	(27)	(28)	(29)	(30)	(31)	(32)
Temp, °F	200.0	200.0	60.0	559.9	434.5	60.0	593.3	60.0
Pres, psia	35	35	15	45	15	15	20	15
Total Flow	23543.154	205655.38	208089.82	691883.33	423784.52	423784.52	268098.82	268098.82
HYDROGEN	14.693608	0.1096197	0.1109643	5.785E-18	0	0	0	0
AMMONIA	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0
WATER	1387.1354	1304.5411	1367.7111	2.006E-16	0	0	0	0
CO	0	0	0	0	0	0	0	0

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Stream →	(25)	(26)	(27)	(28)	(29)	(30)	(31)	(32)
CO2	0	0	0	0	0	0	0	0
METHANE	356.65957	8.0064645	8.1560799	4.357E-19	0	0	0	0
ETHYLENE	0.0010599	7.102E-05	7.271E-05	9.794E-29	0	0	0	0
ACALDEHY	0	0	0	0	0	0	0	0
ACETACID	0	0	0	0	0	0	0	0
ETHANE	3322.9093	304.04576	311.89098	0	0	0	0	0
PROPYLEN	1.684E-05	3.567E-06	3.68E-06	0	0	0	0	0
ACETONE	0	0	0	0	0	0	0	0
PROPANE	5813.7516	1372.1831	1416.7422	8.257E-18	0	0	0	0
FURAN	0	0	0	0	0	0	0	0
2BUTENAL	0	0	0	0	0	0	0	0
BUTENE	2.149E-05	1.123E-05	1.167E-05	2.262E-25	0	0	0	0
THF	0	0	0	0	0	0	0	0
2BUTNONE	0	0	0	0	0	0	0	0
1BUTANAL	0	0	0	0	0	0	0	0
NBUTANE	3415.6521	2034.8444	2117.3971	8.407E-17	0	0	0	0
P-IC4	6795.9254	3083.9511	3200.4264	2.032E-17	0	0	0	0
2MEFURAN	0	0	0	0	0	0	0	0
3MEFURAN	0	0	0	0	0	0	0	0
CYPNTENO	0	0	0	0	0	0	0	0
CYPNTANO	0	0	0	0	0	0	0	0
2PNTENAL	0	0	0	0	0	0	0	0
3PNTE2ON	0	0	0	0	0	0	0	0
2PENTENE	0.0004079	0.0006051	0.0006349	1.946E-20	0	0	0	0
2PENTNON	0	0	0	0	0	0	0	0
3PENTNON	0	0	0	0	0	0	0	0
PENTANE	0.0082602	0.01174	0.0123079	3.057E-19	0	0	0	0
BENZENE	0.006562	0.0282531	0.0297234	4.455E-15	0	0	0	0
PHENOL	0	0	0	0	0	0	0	0
12BZDIOL	0	0	0	0	0	0	0	0
CYHXDIEN	0	0	0	0	0	0	0	0
DIMEFURA	2.783E-05	0.0001819	0.0001914	2.235E-16	0	0	0	0
METCY5ON	0	0	0	0	0	0	0	0
2ETFURAN	0	0	0	0	0	0	0	0
CYHXENE	1.876E-05	9.296E-05	9.774E-05	5.919E-17	0	0	0	0
2MECY5ON	0	0	0	0	0	0	0	0
1HEXANAL	0	0	0	0	0	0	0	0
2HEXNONE	0	0	0	0	0	0	0	0
3HEXNONE	0	0	0	0	0	0	0	0
CYHEXANE	0.008329	0.0395469	0.0415668	2.349E-14	0	0	0	0
NHEXANE	0.0095839	0.0317177	0.0334371	4.424E-16	0	0	0	0
TOLUENE	2.996E-05	0.0003169	0.0003311	4.346E-14	0	0	0	0
OCRESOL	0	0	0	0	0	0	0	0
GUAIACOL	0	0	0	0	0	0	0	0
MECYHEXN	0.0001889	0.0015804	0.0016545	1.007E-13	0	0	0	0
2M3HXENE	0.0041761	0.0284016	0.0298715	8.866E-14	0	0	0	0
METHHEXN	0.0027746	0.0159558	0.0167968	1.864E-14	0	0	0	0
3ETPNTNO	0	0	0	0	0	0	0	0
D-INDOLE	0.0005405	0.3510058	0.351543	46.277395	45.663972	45.663972	0.6134234	0.6134234
OXYLENE	0.0005451	0.015733	0.0161497	4.24E-09	4.24E-09	4.24E-09	4.2E-14	4.2E-14
ETYBENZN	127.60022	2938.8986	3029.1842	0.0001716	0.0001716	0.0001716	1.092E-09	1.092E-09
23XYLNOL	0	0	0	0	0	0	0	0
SYRINGOL	0	0	0	0	0	0	0	0
VANALC	0	0	0	0	0	0	0	0
TRIMEPYR	0.6552498	45.883257	46.479774	0.0096429	0.0096421	0.0096421	8.003E-07	8.003E-07
MEETCY5E	0.0001454	0.0037913	0.0038991	4.763E-10	0	0	0	0

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Stream →	(25)	(26)	(27)	(28)	(29)	(30)	(31)	(32)
DIMECYHX	0.0013615	0.0242827	0.0251129	2.271E-09	2.271E-09	2.271E-09	2.253E-14	2.253E-14
TRDICYHN	0.001736	0.0261987	0.0271746	5.206E-10	0	0	0	0
ETCYHEXN	0.0006584	0.0127353	0.0131558	1.413E-09	1.413E-09	1.413E-09	1.404E-14	1.404E-14
4MEHEPTN	230.24178	2952.5829	3078.2986	1.484E-06	1.484E-06	1.484E-06	2.775E-12	2.775E-12
MEBZFURA	0	0	0	0	0	0	0	0
MEVINPHE	0	0	0	0	0	0	0	0
PROPBENZ	0.0001766	0.0085253	0.0086773	1.715E-07	1.715E-07	1.715E-07	7.516E-12	7.516E-12
ETMEBENZ	1.161E-05	0.0006294	0.0006396	1.989E-08	1.989E-08	1.989E-08	9.676E-13	9.676E-13
TRIMEBEN	5.853E-06	0.0004726	0.000478	3.604E-07	3.604E-07	3.604E-07	4.27E-11	4.27E-11
PROPCYHX	0.000487	0.0189577	0.0193499	9.982E-07	9.982E-07	9.982E-07	7.672E-11	7.672E-11
NAPTHALE	1.84E-07	4.123E-05	4.14E-05	0.0002001	0.0001997	0.0001997	3.844E-07	3.844E-07
1NAPTHOL	0	0	0	0	0	0	0	0
CONIFALD	0	0	0	0	0	0	0	0
TETRALIN	111.5539	21515.853	21624.23	35217.169	35172.371	35172.371	44.798009	44.798009
IEUGENOL	0	0	0	0	0	0	0	0
D-DECALN	394.68477	41358.974	41729.102	2197.5712	2196.4724	2196.4724	1.0987837	1.0987837
DECALIN	0.0001985	0.0296784	0.0298699	0.0031721	0.0031693	0.0031693	2.796E-06	2.796E-06
D-C4CYC6	1507.3936	116854.13	118217.99	3539.0623	3536.932	3536.932	2.130314	2.130314
DIMEOCTN	0.0002356	0.0102302	0.0104302	7.308E-08	7.307E-08	7.307E-08	3.58E-12	3.58E-12
1MENAPHT	0.5896328	345.56242	346.14769	51305.529	50309.987	50309.987	995.54169	995.54169
METDEC	8.1003033	2418.7707	2426.7343	38466.059	38246.089	38246.089	219.97036	219.97036
DIBZTHIO	0	0	0	0	0	0	0	0
D-C4BZ-S	1.113E-07	0.0011444	0.0011445	385.93739	52.710416	52.710416	333.22698	333.22698
BIPHENYL	1.049E-08	1.055E-05	1.056E-05	0.0059258	0.0056413	0.0056413	0.0002845	0.0002845
DIMENAPH	6.025E-10	8.417E-07	8.417E-07	0.0012845	0.0011579	0.0011579	0.0001267	0.0001267
D-C12H18	8.1191381	2449.3467	2457.341	21498.349	21406.301	21406.301	92.047551	92.047551
D-C12H1B	1.2604308	349.81832	351.05681	2622.4397	2611.9513	2611.9513	10.488307	10.488307
2ETDECLN	4.259E-08	1.383E-05	1.387E-05	0.0007204	0.0007077	0.0007077	1.263E-05	1.263E-05
D-CQUINL	9.513E-11	3.695E-06	3.695E-06	46.636916	0.0012581	0.0012581	46.635658	46.635658
HXLCRESO	0	0	0	0	0	0	0	0
HEPCYHEX	2.054E-07	0.0001508	0.000151	0.0329371	0.0302294	0.0302294	0.0027077	0.0027077
D-2C2BZS	3.131E-12	4.149E-07	4.149E-07	385.94084	8.594E-08	8.594E-08	385.94084	385.94084
D-C14H14	0.0001153	0.5866704	0.5867856	24120.471	12639.972	12639.972	11480.499	11480.499
ETPHCYHX	1.18E-09	2.05E-06	2.05E-06	0.007328	0.0056575	0.0056575	0.0016704	0.0016704
D-HYPHEN	0.0009096	2.5958604	2.596769	52751.532	29595.919	29595.919	23155.612	23155.612
THYDPHEN	2.042E-09	3.728E-06	3.73E-06	0.0286123	0.0183556	0.0183556	0.0102567	0.0102567
DIMDODEC	1.12E-05	0.0054255	0.0054366	0.2658159	0.2573238	0.2573238	0.008492	0.008492
D-3CZOLE	8.912E-13	9.321E-08	9.321E-08	46.636959	4.958E-08	4.958E-08	46.636959	46.636959
D-C15H16	6.641E-07	0.0063157	0.0063164	3184.4635	135.04535	135.04535	3049.4181	3049.4181
D-3C3BEN	0.5312058	400.48139	401.00999	77791.913	73949.904	73949.904	3842.0089	3842.0089
OCHYENAP	1.28E-12	1.269E-08	1.269E-08	0.0085097	0.0001043	0.0001043	0.0084054	0.0084054
D-DIMTRD	2.626E-07	0.0001858	0.0001861	0.0374838	0.033689	0.033689	0.0037948	0.0037948
DIMPHENA	4.41E-17	5.606E-12	5.606E-12	0.0025483	4.649E-12	4.649E-12	0.0025483	0.0025483
CETANE	4.763E-10	1.975E-06	1.975E-06	0.0443224	0.0150002	0.0150002	0.0293223	0.0293223
CYHXPHON	0	0	0	0	0	0	0	0
CYHXHXBZ	2.497E-15	1.586E-10	1.586E-10	0.0175158	3.06E-10	3.06E-10	0.0175158	0.0175158
D-NC18	1.738E-05	0.2139867	0.2140041	125292.12	626.45814	626.45814	124665.67	124665.67
D-C1C17	1.793E-05	0.2361609	0.2361789	84077.525	1814.8301	1814.8301	82262.695	82262.695
D-NC20	2.33E-08	0.0007991	0.0007991	9891.5056	0.0116251	0.0116251	9891.494	9891.494
HMWLIGNB	0	0	0	0	0	0	0	0
TRIPHNPR	3.851E-21	1.794E-15	1.794E-15	0.0001731	2.862E-20	2.862E-20	0.0001731	0.0001731
MOXHXPHE	0	0	0	0	0	0	0	0
DIBZFLOR	0	0	0	0.0108875	7.018E-18	7.018E-18	0.0108875	0.0108875
HXACMOXP	0	0	0	0	0	0	0	0
MOXHEPOX	0	0	0	0	0	0	0	0
P-3C3CY6	45.648553	5913.087	5955.4905	159015.71	151443.51	151443.51	7572.1972	7572.1972

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Table L-2. Heat Duties and Power for 100% SRD HT (Case 11201011)

Tag	Description	Value	Unit
Duty			
Q1	Feed-effluent exchanger	303.2	MMBtu/h
Q2	Feed preheat furnace	34.5	MMBtu/h
Q3	Effluent cooler (makes steam)	60.8	MMBtu/h
Q4	Air cooler (cools to 140°F)	28.6	MMBtu/h
Q5	Water cooler (cools to 110°F)	27.3	MMBtu/h
Q6	Distillation cross exchange preheat	138.6	MMBtu/h
Q7	Distillation feed preheat furnace	40.7	MMBtu/h
Q8	Light products column condenser	184.1	MMBtu/h
Q9	Light products column reboiler	192.7	MMBtu/h
Q10	Heavier products column condenser	82.9	MMBtu/h
Q11	Heavier products column reboiler	55.6	MMBtu/h
Power			
W1	Makeup hydrogen compressor	6,580	kW
W2	Recycle compressor	1,146	kW
W3	Liquid recycle pump	-	kW

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Stream, Heat Duty, and Power Requirements Summary for 50/50 LCO/VGO (41201011)

Table L-3. Stream Flows for 50/50 LCO/VGO HT+HC (Case 41201011)

Stream →	(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)
Temp, °F	94.3	129.6	129.6	140.0	644.0	683.7	437.3	276.7
Pres, psia	2100	2100	2100	1995.0001	2045	2015	2010	2005
Total Flow	1035090.5	249523.02	733891.23	193620.6	1284612.9	1768981.2	1768981.2	1768981.2
HYDROGEN	2.709E-05	63196.361	185871.65	59321.343	63196.355	152055.7	152055.7	152055.7
AMMONIA	0	6.508E-28	1.914E-27	0	2.689E-26	1499.8573	1499.8573	1499.8573
H2S	0	4.923E-26	1.448E-25	0	2.034E-24	27759.932	27759.932	27759.932
WATER	0.0095204	587.30608	1727.3708	0	587.31508	1727.5851	1727.5851	1727.5851
CO	0.0001235	0	0	0	0.0001235	0	0	0
CO2	0.001016	0	0	0	0.001016	0.001016	0.001016	0.001016
METHANE	8.798E-05	75146.124	221018.01	41655.056	75146.149	221982.15	221982.15	221982.15
ETHYLENE	0.0001172	0.0036508	0.0107378	0	0.0037681	0.0142885	0.0142885	0.0142885
ACALDEHY	0.0073147	0	0	0	0.0073147	0	0	0
ETHANE	0.0001066	65832.513	193625.04	52050.584	65832.614	194983.78	194983.78	194983.78
PROPYLEN	0.0004741	2.029E-05	5.968E-05	0	0.0004944	9.636E-05	9.636E-05	9.636E-05
ACETONE	0.0126477	0	0	0	0.0126477	0	0	0
PROPANE	0.0001774	30337.402	89227.654	30532.452	30336.649	97863.188	97863.188	97863.188
FURAN	0.0020856	0	0	0	0.0020856	0	0	0
2BUTENAL	0.0031142	0	0	0	0.0031142	0	0	0
BUTENE	0.0003908	1.135E-05	3.339E-05	0	0.0004021	7.597E-05	7.597E-05	7.597E-05
THF	0.0018007	0	0	0	0.0018007	0	0	0
2BUTNONE	0.0070162	0	0	0	0.0070162	0	0	0
1BUTANAL	0.002299	0	0	0	0.002299	0	0	0
NBUTANE	0	4536.0834	13341.422	3353.7226	4536.0918	22154.796	22154.796	22154.796
P-IC4	0	9427.8053	27728.839	6707.4451	9427.8165	44978.623	44978.623	44978.623
2MEFURAN	0.007166	0	0	0	0.007166	0	0	0
3MEFURAN	0.0011353	0	0	0	0.0011353	0	0	0
CYPNTENO	0.0018906	0	0	0	0.0018906	0	0	0
CYPNTANO	0.0043652	0	0	0	0.0043652	0	0	0
2PNTENAL	0.0059755	0	0	0	0.0059755	0	0	0
3PNTE2ON	0.0018075	0	0	0	0.0018075	0	0	0
2PENTENE	0	0.0001634	0.0004807	0	0.0001634	0.002065	0.002065	0.002065
2PENTNON	0.0025236	0	0	0	0.0025236	0	0	0
3PENTNON	0.0015262	0	0	0	0.0015262	0	0	0
PENTANE	0.0001287	0.0022735	0.0066869	0	0.0024023	0.0279545	0.0279545	0.0279545
BENZENE	0.0034481	0.0013802	0.0040594	0	0.0048283	0.03954	0.03954	0.03954
PHENOL	0.0586712	7.451E-34	2.191E-33	0	0.0586712	0	0	0
12BZDIOL	0.0088288	0	0	0	0.0088288	0	0	0
CYHXDIEN	0.0001115	0	0	0	0.0001115	0	0	0
DIMEFURA	0.0029189	7.123E-06	2.095E-05	0	0.002926	0.0003113	0.0003113	0.0003113
METCY5ON	0.0013683	0	0	0	0.0013683	0	0	0
2ETFURAN	0.002633	0	0	0	0.002633	0	0	0
CYHXENE	0	3.906E-06	1.149E-05	0	3.906E-06	0.0001258	0.0001258	0.0001258
2MECY5ON	0.0018052	0	0	0	0.0018052	0	0	0
1HEXANAL	0.0017517	0	0	0	0.0017517	0	0	0
2HEXNONE	0.0008445	0	0	0	0.0008445	0	0	0
3HEXNONE	0.0007792	0	0	0	0.0007792	0	0	0
CYHEXANE	0.0088221	0.0016192	0.0047622	0	0.0104413	0.0504263	0.0504263	0.0504263
NHEXANE	0	0.0019651	0.0057796	0	0.0019651	0.0471075	0.0471075	0.0471075
SORBTOL	0	5.519E-47	1.623E-46	0	2.282E-45	0	0	0
TOLUENE	0.0025755	7.911E-06	2.327E-05	0	0.0025834	0.0003789	0.0003789	0.0003789
OCRESOL	0.002855	0	0	0	0.002855	0	0	0

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Stream →	(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)
GUAIACOL	0.0426492	0	0	0	0.0426492	0	0	0
MECYHEXN	0.0010128	2.861E-05	8.415E-05	0	0.0010414	0.0014759	0.0014759	0.0014759
2M3HXENE	0	0.0007543	0.0022186	0	0.0007543	0.0347232	0.0347232	0.0347232
METHHEXN	0	0.0005019	0.0014761	0	0.0005019	0.0197695	0.0197695	0.0197695
3ETPNTNO	0.0036838	0	0	0	0.0036838	0	0	0
D-INDOLE	0	3.214E-29	9.453E-29	0	7.237E-27	1.625E-28	1.625E-28	1.625E-28
OXYLENE	0.0302823	0.000116	0.0003412	0	0.0303983	0.0201305	0.0201305	0.0201305
ETYBENZN	0	3.21E-05	9.443E-05	0	3.21E-05	0.004461	0.004461	0.004461
23XYLNOL	0.0498463	0	0	0	0.0498463	0	0	0
SYRINGOL	0.0120934	0	0	0	0.0120934	0	0	0
VANALC	0.0195095	0	0	0	0.0195095	0	0	0
TRIMEPYR	0.007354	2.776E-27	8.164E-27	0	0.007354	1.406E-26	1.406E-26	1.406E-26
MEETCY5E	0.0004595	2.498E-05	7.346E-05	0	0.0004845	0.0039675	0.0039675	0.0039675
DIMECYHX	0.0003661	0.0002818	0.000829	0	0.000648	0.0286425	0.0286425	0.0286425
TRDICYHN	0.0010212	0.0002723	0.0008009	0	0.0012935	0.0238323	0.0238323	0.0238323
ETCYHEXN	0	0.0001239	0.0003644	0	0.0001239	0.013781	0.013781	0.013781
4MEHEPTN	0	8.54E-05	0.0002512	0	8.54E-05	0.0071833	0.0071833	0.0071833
MEBZFURA	0.0100215	0	0	0	0.0100215	0	0	0
MEVINPHE	0.0003019	0	0	0	0.0003019	0	0	0
PROPBENZ	0.0069226	3.027E-05	8.903E-05	0	0.0069529	0.008752	0.008752	0.008752
ETMEBENZ	0.0006418	1.965E-06	5.778E-06	0	0.0006437	0.0006475	0.0006475	0.0006475
TRIMEBEN	0.0004792	9.971E-07	2.933E-06	0	0.0004802	0.0004821	0.0004821	0.0004821
PROPCYHX	0	9.177E-05	0.0002699	0	9.177E-05	0.0198036	0.0198036	0.0198036
NAPTHALE	0.0260192	3.066E-07	9.016E-07	0	0.0260195	0.0003756	0.0003756	0.0003756
1NAPTHOL	0.0252745	0	0	0	0.0252745	0	0	0
CONIFALD	0.0260545	0	0	0	0.0260545	0	0	0
TETRALIN	0	1.487E-05	4.374E-05	0	1.487E-05	0.0117471	0.0117471	0.0117471
IEUGENOL	0.0277852	0	0	0	0.0277852	0	0	0
DECALIN	0.0133396	6.1010836	17.944363	0	6.1144281	5858.2282	5858.2282	5858.2282
D-C4CYC6	0	23.119083	67.997302	0	23.11911	9637.8477	9637.8477	9637.8477
DIMEOCTN	0	3.644E-05	0.0001072	0	3.644E-05	0.0102303	0.0102303	0.0102303
1MENAPHT	0.0041378	1.08E-24	3.176E-24	0	0.0041378	1.49E-22	1.49E-22	1.49E-22
METDEC	0.0045895	2.725E-05	8.014E-05	0	0.0046167	0.0463862	0.0463862	0.0463862
DIBZTHIO	0.001903	0	0	0	0.001903	0	0	0
BIPHENYL	0	1.1E-06	3.236E-06	0	1.1E-06	0.0062458	0.0062458	0.0062458
DIMENAPH	0	1.557E-07	4.58E-07	0	1.557E-07	0.0012284	0.0012284	0.0012284
2ETDECLN	0	32.654452	96.042505	0	32.654467	55426.011	55426.011	55426.011
HXLCRESO	0.0052768	0	0	0	0.0052768	0	0	0
HEPCYHEX	0	7.113E-06	2.092E-05	0	7.113E-06	0.033586	0.033586	0.033586
PHENANTH	23515.222	0.0462881	0.1361415	0	23515.269	3970.5495	3970.5495	3970.5495
ETPHCYHX	0	8.536E-07	2.511E-06	0	8.536E-07	0.0084858	0.0084858	0.0084858
D-HYPHEN	0	1.3080949	3.847338	0	1.308094	20098.762	20098.762	20098.762
THYDPHEN	0.0109926	2.965E-06	8.72E-06	0	0.0109956	0.0287486	0.0287486	0.0287486
DIMDODEC	0	8.328E-05	0.000245	0	8.328E-05	0.2697062	0.2697062	0.2697062
D-3CZOLE	7054.5667	0.0002876	0.000846	0	7054.567	156.18178	156.18178	156.18178
D-C15H16	0	0.1224143	0.3600419	0	0.1224141	6461.4627	6461.4627	6461.4627
D-3C3BEN	54085.012	1.9213805	5.6511191	0	54086.933	9133.4163	9133.4163	9133.4163
OCHYENAP	0.0116495	4.052E-07	1.192E-06	0	0.0116499	0.0228428	0.0228428	0.0228428
D-DIMTRD	0	8.18E-06	2.406E-05	0	8.18E-06	0.0381978	0.0381978	0.0381978
DIMPHENA	0.009408	5.238E-09	1.541E-08	0	0.009408	0.0037786	0.0037786	0.0037786
CETANE	0	2.1334893	6.2749685	0	2.1334879	61802.116	61802.116	61802.116
CYHXPHON	0.0844325	0	0	0	0.0844325	0	0	0
CYHXHXBZ	0	4.878E-08	1.435E-07	0	4.878E-08	0.0181707	0.0181707	0.0181707
D-NC18	56300.869	0.6199238	1.8233053	0	56301.489	54090.189	54090.189	54090.189
HMWLIGNB	0.1328401	0	0	0	0.1328401	0	0	0
TRIPHNPR	0	8.241E-11	2.424E-10	0	8.241E-11	0.0001853	0.0001853	0.0001853

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Stream →	(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)
MOXHXPHE	0.0853743	0	0	0	0.0853743	0	0	0
DIBZFLOR	0.107389	1.469E-09	4.321E-09	0	0.107389	0.0150028	0.0150028	0.0150028
HXACMOXP	0.2001781	0	0	0	0.2001781	0	0	0
MOXHEPOX	0.0178464	0	0	0	0.0178464	0	0	0
HXDECBEN	33112.147	0.0028558	0.0083994	0	33112.149	11167.903	11167.903	11167.903
L-3MQUIN	375.33913	0.0014873	0.0043745	0	375.34061	8.3098085	8.3098085	8.3098085
L-NC10	11169.731	20.615118	60.632699	0	11190.346	10790.612	10790.612	10790.612
L-ACENPH	125218.56	1.9138415	5.6289455	0	125220.47	21144.235	21144.235	21144.235
L-TMC9	12209.827	18.026843	53.020125	0	12227.854	11782.324	11782.324	11782.324
L-C3PHEN	29710.579	0.0021824	0.0064189	0	29710.581	5016.6008	5016.6008	5016.6008
L-C4PHEN	95321.511	0.0020881	0.0061415	0	95321.513	16094.936	16094.936	16094.936
L-C8NPH	70048.23	0.0262192	0.0771153	0	70048.256	11827.585	11827.585	11827.585
L-NC14BZ	18088.633	0.0031284	0.0092012	0	18088.636	3054.2482	3054.2482	3054.2482
L-NC22	19671.388	0.0335643	0.0987184	0	19671.449	18898.477	18898.477	18898.477
V-23DMBT	101766.61	0.2833248	0.8333082	0	101766.89	2253.0475	2253.0475	2253.0475
V-NC12	7210.4674	11.326938	33.314522	0	7221.7944	25778.386	25778.386	25778.386
V-2MBZQN	10384.006	0.0008598	0.0025289	0	10384.007	229.89263	229.89263	229.89263
V-C5DBT	52340.06	5.104E-05	0.0001501	0	52340.06	1158.7618	1158.7618	1158.7618
V-C12NPH	20091.302	0.000699	0.0020559	0	20091.303	6776.2952	6776.2952	6776.2952
V-HMCHRY	83912.106	3.036E-06	8.931E-06	0	83912.106	28301.457	28301.457	28301.457
V-NC33	203503.19	0.0006916	0.0020341	0	203503.19	68636.544	68636.544	68636.544
P-IC4CY6	0	218.31123	642.09185	0	218.31149	86904.776	86904.776	86904.776
P-C12H20	0	39.567685	116.37554	0	39.567694	105755.06	105755.06	105755.06
P-C14H20	0	0.9585405	2.8192369	0	0.9585401	9884.0752	9884.0752	9884.0752
P-3C3CY6	0	75.36962	221.67535	0	75.369714	44318.034	44318.034	44318.034
P-C17H20	0	0.1762572	0.5184036	0	0.1762569	45074.513	45074.513	45074.513
P-C17H24	0	0.2614644	0.7690128	0	0.261464	24388.264	24388.264	24388.264
P-NC17	0	1.283395	3.7746911	0	1.2833938	65627.466	65627.466	65627.466
P-C18H32	0	2.4659638	7.2528347	0	2.4659619	80033.02	80033.02	80033.02
P-C18H34	0	0.6889533	2.0263332	0	0.6889525	57795.757	57795.757	57795.757
P-C20H40	0	0.0594716	0.1749164	0	0.054448	14639.305	14639.305	14639.305

Table L-3 (continued). Stream Flows for 50/50 LCO/VGO HT+HC (Case 41201011)

Stream →	(9)	(10)	(11)	(12)	(13)	(14)	(15)	(16)
Temp, °F	110.0	110.0	109.3	110.0	110.0	110.0	110.0	110.0
Pres, psia	2050	1995	1993	1993	1993	20	1993	1973
Total Flow	235097.33	2004078.5	235846.66	665985.72	641227.33	24758.39	100955.48	20100.167
HYDROGEN	0	152055.7	1.2728033	150197.55	150197.55	0	23647.255	20100.167
AMMONIA	0	1499.8573	1499.8573	0	0	0	0	0
H2S	0	27759.932	3001.5428	24758.389	0	24758.389	0	0
WATER	235097.33	236824.91	231336.65	2050.1497	2050.1497	0	322.77765	0
CO	0	0	0	0	0	0	0	0
CO2	0	0.001016	0	0.001016	0	0.001016	0	0
METHANE	0	221982.15	4.0474665	212878.7	212878.7	0	33515.838	0
ETHYLENE	0	0.0142885	3.743E-07	0.0127442	0.0127442	0	0.0020065	0
ACALDEHY	0	0	0	0	0	0	0	0
ETHANE	0	194983.78	3.1337962	168028.95	168028.95	0	26454.648	0
PROPYLEN	0	9.636E-05	5.849E-10	7.084E-05	7.084E-05	0	1.115E-05	0
ACETONE	0	0	0	0	0	0	0	0
PROPANE	0	97863.188	0.153508	69664.86	69664.86	0	10968.106	0
FURAN	0	0	0	0	0	0	0	0
2BUTENAL	0	0	0	0	0	0	0	0
BUTENE	0	7.597E-05	3.115E-11	3.963E-05	3.963E-05	0	6.239E-06	0
THF	0	0	0	0	0	0	0	0

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Stream →	(9)	(10)	(11)	(12)	(13)	(14)	(15)	(16)
2BUTNONE	0	0	0	0	0	0	0	0
1BUTANAL	0	0	0	0	0	0	0	0
NBUTANE	0	22154.796	0.0022496	11853.985	11853.985	0	1866.3035	0
P-IC4	0	44978.623	0.0021923	24949.433	24949.433	0	3928.0641	0
2MEFURAN	0	0	0	0	0	0	0	0
3MEFURAN	0	0	0	0	0	0	0	0
CYPNTENO	0	0	0	0	0	0	0	0
CYPNTANO	0	0	0	0	0	0	0	0
2PNTENAL	0	0	0	0	0	0	0	0
3PNTE2ON	0	0	0	0	0	0	0	0
2PENTENE	0	0.002065	5.436E-11	0.0005705	0.0005705	0	8.983E-05	0
2PENTNON	0	0	0	0	0	0	0	0
3PENTNON	0	0	0	0	0	0	0	0
PENTANE	0	0.0279545	9.421E-11	0.0079364	0.0079364	0	0.0012495	0
BENZENE	0	0.03954	5.353E-07	0.0048179	0.0048179	0	0.0007585	0
PHENOL	0	0	0	0	0	0	0	0
12BZDIOL	0	0	0	0	0	0	0	0
CYHXDIEN	0	0	0	0	0	0	0	0
DIMEFURA	0	0.0003113	1.458E-10	2.487E-05	2.487E-05	0	3.915E-06	0
METCY5ON	0	0	0	0	0	0	0	0
2ETFURAN	0	0	0	0	0	0	0	0
CYHXENE	0	0.0001258	1.055E-10	1.363E-05	1.363E-05	0	2.147E-06	0
2MECY5ON	0	0	0	0	0	0	0	0
1HEXANAL	0	0	0	0	0	0	0	0
2HEXNONE	0	0	0	0	0	0	0	0
3HEXNONE	0	0	0	0	0	0	0	0
CYHEXANE	0	0.0504263	8.182E-09	0.0056521	0.0056521	0	0.0008899	0
NHEXANE	0	0.0471075	2.877E-12	0.0068596	0.0068596	0	0.00108	0
SORBTOL	0	0	0	0	0	0	0	0
TOLUENE	0	0.0003789	7.656E-11	2.762E-05	2.762E-05	0	4.348E-06	0
OCRESOL	0	0	0	0	0	0	0	0
GUAIACOL	0	0	0	0	0	0	0	0
MECYHEXN	0	0.0014759	1.766E-12	9.987E-05	9.987E-05	0	1.572E-05	0
2M3HXENE	0	0.0347232	2.294E-13	0.0026331	0.0026331	0	0.0004146	0
METHHEXN	0	0.0197695	1.7E-14	0.0017519	0.0017519	0	0.0002758	0
3ETPNTNO	0	0	0	0	0	0	0	0
D-INDOLE	0	1.625E-28	0	0	0	0	0	0
OXYLENE	0	0.0201305	3.061E-10	0.000405	0.000405	0	6.376E-05	0
ETYBENZN	0	0.004461	1.87E-11	0.0001121	0.0001121	0	1.764E-05	0
23XYLNOL	0	0	0	0	0	0	0	0
SYRINGOL	0	0	0	0	0	0	0	0
VANALC	0	0	0	0	0	0	0	0
TRIMEPYR	0	1.406E-26	0	0	0	0	0	0
MEETCY5E	0	0.0039675	3.485E-12	8.719E-05	8.719E-05	0	1.373E-05	0
DIMECYHX	0	0.0286425	1.957E-14	0.0009839	0.0009839	0	0.0001549	0
TRDICYHN	0	0.0238323	2.331E-14	0.0009506	0.0009506	0	0.0001497	0
ETCYHEXN	0	0.013781	4.434E-14	0.0004325	0.0004325	0	6.809E-05	0
4MEHEPTN	0	0.0071833	1.096E-16	0.0002981	0.0002981	0	4.693E-05	0
MEBZFURA	0	0	0	0	0	0	0	0
MEVINPHE	0	0	0	0	0	0	0	0
PROPBENZ	0	0.008752	4.085E-13	0.0001057	0.0001057	0	1.664E-05	0
ETMEBENZ	0	0.0006475	5.617E-14	6.858E-06	6.858E-06	0	1.08E-06	0
TRIMEBEN	0	0.0004821	4.046E-13	3.481E-06	3.481E-06	0	5.48E-07	0
PROPCYHX	0	0.0198036	8.295E-16	0.0003203	0.0003203	0	5.044E-05	0
NAPTHALE	0	0.0003756	1.8E-11	1.07E-06	1.07E-06	0	1.685E-07	0
1NAPTHOL	0	0	0	0	0	0	0	0

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Stream →	(9)	(10)	(11)	(12)	(13)	(14)	(15)	(16)
CONIFALD	0	0	0	0	0	0	0	0
TETRALIN	0	0.0117471	2.582E-11	5.192E-05	5.192E-05	0	8.174E-06	0
IEUGENOL	0	0	0	0	0	0	0	0
DECALIN	0	5858.2282	1.743E-15	21.297452	21.297452	0	3.3530924	0
D-C4CYC6	0	9637.8477	1.858E-12	80.703274	80.703274	0	12.706005	0
DIMEOCTN	0	0.0102303	1.221E-20	0.0001272	0.0001272	0	2.003E-05	0
1MENAPHT	0	1.49E-22	0	0	0	0	0	0
METDEC	0	0.0463862	8.094E-16	9.512E-05	9.512E-05	0	1.498E-05	0
DIBZTHIO	0	0	0	0	0	0	0	0
BIPHENYL	0	0.0062458	8.38E-13	3.841E-06	3.841E-06	0	6.047E-07	0
DIMENAPH	0	0.0012284	9.881E-15	5.435E-07	5.435E-07	0	8.558E-08	0
2ETDECLN	0	55426.011	2.933E-12	113.98904	113.98904	0	17.946551	0
HXLCRESO	0	0	0	0	0	0	0	0
HEPCYHEX	0	0.033586	1.725E-31	2.483E-05	2.483E-05	0	3.909E-06	0
PHENANTH	0	3970.5495	1.804E-08	0.1615812	0.1615812	0	0.0254395	0
ETPHCYHX	0	0.0084858	8.709E-21	2.98E-06	2.98E-06	0	4.691E-07	0
D-HYPHEN	0	20098.762	3.985E-12	4.5662557	4.5662557	0	0.7189159	0
THYDPHEN	0	0.0287486	1.044E-21	1.035E-05	1.035E-05	0	1.629E-06	0
DIMDODEC	0	0.2697062	1.947E-31	0.0002907	0.0002907	0	4.577E-05	0
D-3CZOLE	0	156.18178	1.893E-12	0.0010041	0.0010041	0	0.0001581	0
D-C15H16	0	6461.4627	1.102E-14	0.4273199	0.4273199	0	0.0672777	0
D-3C3BEN	0	9133.4163	3.162E-19	6.7070929	6.7070929	0	1.0559715	0
OCHYENAP	0	0.0228428	9.244E-28	1.415E-06	1.415E-06	0	2.227E-07	0
D-DIMTRD	0	0.0381978	5.304E-36	2.855E-05	2.855E-05	0	4.495E-06	0
DIMPHENA	0	0.0037786	2.915E-17	1.829E-08	1.829E-08	0	2.879E-09	0
CETANE	0	61802.116	1.12E-32	7.4475192	7.4475192	0	1.172545	0
CYHXPHON	0	0	0	0	0	0	0	0
CYHXHXBZ	0	0.0181707	1.553E-32	1.703E-07	1.703E-07	0	2.681E-08	0
D-NC18	0	54090.189	2.567E-40	2.164012	2.164012	0	0.3407042	0
HMWLIGNB	0	0	0	0	0	0	0	0
TRIPHNPR	0	0.0001853	1.687E-32	2.877E-10	2.877E-10	0	4.529E-11	0
MOXHXPHE	0	0	0	0	0	0	0	0
DIBZFLOR	0	0.0150028	2.971E-27	5.129E-09	5.129E-09	0	8.075E-10	0
HXACMOXP	0	0	0	0	0	0	0	0
MOXHEPOX	0	0	0	0	0	0	0	0
HXDECBEN	0	11167.903	1.873E-39	0.009969	0.009969	0	0.0015695	0
L-3MQUIN	0	8.3098085	3.019E-07	0.0051919	0.0051919	0	0.0008174	0
L-NC10	0	10790.612	2.283E-15	71.962537	71.962537	0	11.329855	0
L-ACENPH	0	21144.235	1.788E-08	6.6807764	6.6807764	0	1.0518282	0
L-TMC9	0	11782.324	6.719E-20	62.927481	62.927481	0	9.9073665	0
L-C3PHEN	0	5016.6008	3.969E-15	0.0076184	0.0076184	0	0.0011994	0
L-C4PHEN	0	16094.936	1.268E-17	0.0072891	0.0072891	0	0.0011476	0
L-C8NPH	0	11827.585	1.1E-23	0.0915253	0.0915253	0	0.0144098	0
L-NC14BZ	0	3054.2482	1.577E-34	0.0109206	0.0109206	0	0.0017193	0
L-NC22	0	18898.477	2.843E-46	0.1046996	0.1046996	0	0.016484	0
V-23DMBT	0	2253.0475	1.053E-06	0.989021	0.989021	0	0.1557125	0
V-NC12	0	25778.386	5.346E-20	39.539695	39.539695	0	6.2251697	0
V-2MBZQN	0	229.89263	4.988E-09	0.0030015	0.0030015	0	0.0004726	0
V-C5DBT	0	1158.7618	1.301E-17	0.0001782	0.0001782	0	2.805E-05	0
V-C12NPH	0	6776.2952	1.09E-36	0.0024401	0.0024401	0	0.0003842	0
V-HMCHRY	0	28301.457	4.074E-36	1.06E-05	1.06E-05	0	1.669E-06	0
V-NC33	0	68636.544	1.033E-45	0.0021605	0.0021605	0	0.0003402	0
P-IC4CY6	0	86904.776	1.766E-10	762.07312	762.07312	0	119.98157	0
P-C12H20	0	105755.06	8.038E-11	138.12154	138.12154	0	21.745996	0
P-C14H20	0	9884.0752	1.584E-15	3.3460425	3.3460425	0	0.5268043	0
P-3C3CY6	0	44318.034	5.241E-29	263.0977	263.0977	0	41.422368	0

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Stream →	(9)	(10)	(11)	(12)	(13)	(14)	(15)	(16)
P-C17H20	0	45074.513	5.238E-19	0.6152735	0.6152735	0	0.0968693	0
P-C17H24	0	24388.264	1.149E-19	0.9127119	0.9127119	0	0.1436983	0
P-NC17	0	65627.466	2.265E-36	4.4800368	4.4800368	0	0.7053415	0
P-C18H32	0	80033.02	5.297E-30	8.6081112	8.6081112	0	1.3552698	0
P-C18H34	0	57795.757	3.672E-32	2.4049778	2.4049778	0	0.3786421	0
P-C20H40	0	14639.305	2.202E-46	0.184279	0.184279	0	0.0290131	0

Table L-3 (continued). Stream Flows for 50/50 LCO/VGO HT+HC (Case 41201011)

Stream →	(17)	(18)	(19)	(20)	(21)	(22)	(23)	(24)
Temp, °F	110.0	110.0	110.0	110.0	110.0	110.0	110.0	450.0
Pres, psia	22	1993	1993	53	53	2050	53	45.5
Total Flow	80855.315	540270.63	1102246.1	65741.543	1036504.6	1.802E-05	1036504.6	1036504.6
HYDROGEN	3547.0883	126550.31	1856.875	1842.4919	14.38307	0	14.38307	14.38307
AMMONIA	0	1.914E-27	0	0	0	0	0	0
H2S	0	1.448E-25	0	0	0	0	0	0
WATER	322.77765	1727.3708	3438.1118	1049.6909	2388.421	1.802E-05	2388.421	2388.421
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
METHANE	33515.838	179362.96	9099.4017	8796.7432	302.6585	0	302.6585	302.6585
ETHYLENE	0.0020065	0.0107378	0.0015439	0.0013789	0.000165	0	0.000165	0.000165
ACALDEHY	0	0	0	0	0	0	0	0
ETHANE	26454.648	141574.45	26951.702	22976.65	3975.0524	0	3975.0524	3975.0524
PROPYLEN	1.115E-05	5.968E-05	2.552E-05	1.735E-05	8.175E-06	0	8.175E-06	8.175E-06
ACETONE	0	0	0	0	0	0	0	0
PROPANE	10968.106	58695.202	28198.174	18189.358	10008.816	0	10008.816	10008.816
FURAN	0	0	0	0	0	0	0	0
2BUTENAL	0	0	0	0	0	0	0	0
BUTENE	6.239E-06	3.339E-05	3.634E-05	1.487E-05	2.147E-05	0	2.147E-05	2.147E-05
THF	0	0	0	0	0	0	0	0
2BUTNONE	0	0	0	0	0	0	0	0
1BUTANAL	0	0	0	0	0	0	0	0
NBUTANE	1866.3035	9987.6992	10300.808	3778.9454	6521.863	0	6521.863	6521.863
P-IC4	3928.0641	21021.394	20029.188	8857.0337	11172.154	0	11172.154	11172.154
2MEFURAN	0	0	0	0	0	0	0	0
3MEFURAN	0	0	0	0	0	0	0	0
CYPNTENO	0	0	0	0	0	0	0	0
CYPNTANO	0	0	0	0	0	0	0	0
2PNTENAL	0	0	0	0	0	0	0	0
3PNTE2ON	0	0	0	0	0	0	0	0
2PENTENE	8.983E-05	0.0004807	0.0014944	0.0002386	0.0012559	0	0.0012559	0.0012559
2PENTNON	0	0	0	0	0	0	0	0
3PENTNON	0	0	0	0	0	0	0	0
PENTANE	0.0012495	0.0066869	0.0200181	0.0032381	0.01678	0	0.01678	0.01678
BENZENE	0.0007585	0.0040594	0.0347215	0.0020417	0.0326799	0	0.0326799	0.0326799
PHENOL	0	2.191E-33	0	0	0	0	0	0
12BZDIOL	0	0	0	0	0	0	0	0
CYHXDIEN	0	0	0	0	0	0	0	0
DIMEFURA	3.915E-06	2.095E-05	0.0002864	9.397E-06	0.000277	0	0.000277	0.000277
METCY5ON	0	0	0	0	0	0	0	0
2ETFURAN	0	0	0	0	0	0	0	0
CYHXENE	2.147E-06	1.149E-05	0.0001121	5.45E-06	0.0001067	0	0.0001067	0.0001067
2MECY5ON	0	0	0	0	0	0	0	0

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Stream →	(17)	(18)	(19)	(20)	(21)	(22)	(23)	(24)
1HEXANAL	0	0	0	0	0	0	0	0
2HEXNONE	0	0	0	0	0	0	0	0
3HEXNONE	0	0	0	0	0	0	0	0
CYHEXANE	0.0008899	0.0047622	0.0447742	0.0022412	0.042533	0	0.042533	0.042533
NHEXANE	0.00108	0.0057796	0.040248	0.0024871	0.0377609	0	0.0377609	0.0377609
SORBTOL	0	1.623E-46	0	0	0	0	0	0
TOLUENE	4.348E-06	2.327E-05	0.0003512	6.914E-06	0.0003443	0	0.0003443	0.0003443
OCRESOL	0	0	0	0	0	0	0	0
GUAIACOL	0	0	0	0	0	0	0	0
MECYHEXN	1.572E-05	8.415E-05	0.001376	3.375E-05	0.0013422	0	0.0013422	0.0013422
2M3HXENE	0.0004146	0.0022186	0.03209	0.0008433	0.0312467	0	0.0312467	0.0312467
METHHEXN	0.0002758	0.0014761	0.0180176	0.0005599	0.0174578	0	0.0174578	0.0174578
3ETPNTNO	0	0	0	0	0	0	0	0
D-INDOLE	0	9.453E-29	0	0	0	0	0	0
OXYLENE	6.376E-05	0.0003412	0.0197255	0.0001139	0.0196116	0	0.0196116	0.0196116
ETYBENZN	1.764E-05	9.443E-05	0.004349	3.225E-05	0.0043167	0	0.0043167	0.0043167
23XYLNOL	0	0	0	0	0	0	0	0
SYRINGOL	0	0	0	0	0	0	0	0
VANALC	0	0	0	0	0	0	0	0
TRIMEPYR	0	8.164E-27	0	0	0	0	0	0
MEETCY5E	1.373E-05	7.346E-05	0.0038803	2.393E-05	0.0038564	0	0.0038564	0.0038564
DIMECYHX	0.0001549	0.000829	0.0276587	0.0002647	0.027394	0	0.027394	0.027394
TRDICYHN	0.0001497	0.0008009	0.0228817	0.0002642	0.0226175	0	0.0226175	0.0226175
ETCYHEXN	6.809E-05	0.0003644	0.0133485	0.000117	0.0132315	0	0.0132315	0.0132315
4MEHEPTN	4.693E-05	0.0002512	0.0068852	7.864E-05	0.0068066	0	0.0068066	0.0068066
MEBZFURA	0	0	0	0	0	0	0	0
MEVINPHE	0	0	0	0	0	0	0	0
PROPBENZ	1.664E-05	8.903E-05	0.0086463	2.485E-05	0.0086214	0	0.0086214	0.0086214
ETMEBENZ	1.08E-06	5.778E-06	0.0006407	1.588E-06	0.0006391	0	0.0006391	0.0006391
TRIMEBEN	5.48E-07	2.933E-06	0.0004787	7.748E-07	0.0004779	0	0.0004779	0.0004779
PROPCYHX	5.044E-05	0.0002699	0.0194832	7.194E-05	0.0194113	0	0.0194113	0.0194113
NAPTHALE	1.685E-07	9.016E-07	0.0003746	2.119E-07	0.0003744	0	0.0003744	0.0003744
1NAPTHOL	0	0	0	0	0	0	0	0
CONIFALD	0	0	0	0	0	0	0	0
TETRALIN	8.174E-06	4.374E-05	0.0116951	7.307E-06	0.0116878	0	0.0116878	0.0116878
IEUGENOL	0	0	0	0	0	0	0	0
DECALIN	3.3530924	17.944363	5836.9307	3.0409149	5833.8898	0	5833.8898	5833.8898
D-C4CYC6	12.706005	67.997302	9557.1444	14.813902	9542.3305	0	9542.3305	9542.3305
DIMEOCTN	2.003E-05	0.0001072	0.0101031	2.338E-05	0.0100797	0	0.0100797	0.0100797
1MENAPHT	0	3.176E-24	0	0	0	0	0	0
METDEC	1.498E-05	8.014E-05	0.0462911	1.429E-05	0.0462768	0	0.0462768	0.0462768
DIBZTHIO	0	0	0	0	0	0	0	0
BIPHENYL	6.047E-07	3.236E-06	0.006242	5.228E-07	0.0062414	0	0.0062414	0.0062414
DIMENAPH	8.558E-08	4.58E-07	0.0012279	6.676E-08	0.0012278	0	0.0012278	0.0012278
2ETDECLN	17.946551	96.042505	55312.021	15.750809	55296.271	0	55296.271	55296.271
HXLCRESO	0	0	0	0	0	0	0	0
HEPCYHEX	3.909E-06	2.092E-05	0.0335612	1.833E-06	0.0335593	0	0.0335593	0.0335593
PHENANTH	0.0254395	0.1361415	3970.3879	0.012539	3970.3754	0	3970.3754	3970.3754
ETPHCYHX	4.691E-07	2.511E-06	0.0084828	2.785E-07	0.0084825	0	0.0084825	0.0084825
D-HYPHEN	0.7189159	3.847338	20094.196	0.4241121	20093.772	0	20093.772	20093.772
THYDPHEN	1.629E-06	8.72E-06	0.0287382	9.259E-07	0.0287373	0	0.0287373	0.0287373
DIMDODEC	4.577E-05	0.000245	0.2694155	2.241E-05	0.2693931	0	0.2693931	0.2693931
D-3CZOLE	0.0001581	0.000846	156.18078	5.046E-05	156.18073	0	156.18073	156.18073
D-C15H16	0.0672777	0.3600419	6461.0353	0.0293445	6461.006	0	6461.006	6461.006
D-3C3BEN	1.0559715	5.6511191	9126.7092	0.6208599	9126.0883	0	9126.0883	9126.0883
OCHYENAP	2.227E-07	1.192E-06	0.0228413	7.442E-08	0.0228413	0	0.0228413	0.0228413

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Stream →	(17)	(18)	(19)	(20)	(21)	(22)	(23)	(24)
D-DIMTRD	4.495E-06	2.406E-05	0.0381692	1.814E-06	0.0381674	0	0.0381674	0.0381674
DIMPHENA	2.879E-09	1.541E-08	0.0037785	8.654E-10	0.0037785	0	0.0037785	0.0037785
CETANE	1.172545	6.2749685	61794.668	0.3044882	61794.364	0	61794.364	61794.364
CYHXPHON	0	0	0	0	0	0	0	0
CYHXHXBZ	2.681E-08	1.435E-07	0.0181705	5.366E-09	0.0181705	0	0.0181705	0.0181705
D-NC18	0.3407042	1.8233053	54088.025	0.0558044	54087.969	0	54087.969	54087.969
HMWLIGNB	0	0	0	0	0	0	0	0
TRIPHNPR	0	2.424E-10	0.0001853	6.906E-12	0.0001853	0	0.0001853	0.0001853
MOXHXPHE	0	0	0	0	0	0	0	0
DIBZFLOR	0	4.321E-09	0.0150028	1.034E-10	0.0150028	0	0.0150028	0.0150028
HXACMOXP	0	0	0	0	0	0	0	0
MOXHEPOX	0	0	0	0	0	0	0	0
HXDECBEN	0.0015695	0.0083994	11167.893	0.000132	11167.893	0	11167.893	11167.893
L-3MQUIN	0.0008174	0.0043745	8.3046163	0.0007688	8.3038475	0	8.3038475	8.3038475
L-NC10	11.329855	60.632699	10718.649	11.687081	10706.962	0	10706.962	10706.962
L-ACENPH	1.0518282	5.6289455	21137.554	0.7452773	21136.809	0	21136.809	21136.809
L-TMC9	9.9073665	53.020125	11719.397	8.2587027	11711.138	0	11711.138	11711.138
L-C3PHEN	0.0011994	0.0064189	5016.5932	0.0002536	5016.5929	0	5016.5929	5016.5929
L-C4PHEN	0.0011476	0.0061415	16094.929	0.0001747	16094.929	0	16094.929	16094.929
L-C8NPH	0.0144098	0.0771153	11827.494	0.003078	11827.491	0	11827.491	11827.491
L-NC14BZ	0.0017193	0.0092012	3054.2373	0.0002256	3054.2371	0	3054.2371	3054.2371
L-NC22	0.016484	0.0987184	18898.373	0.0010705	18898.371	0	18898.371	18898.371
V-23DMBT	0.1557125	0.8333082	2252.0584	0.128632	2251.9298	0	2251.9298	2251.9298
V-NC12	6.2251697	33.314522	25738.847	4.0390374	25734.808	0	25734.808	25734.808
V-2MBZQN	0.0004726	0.0025289	229.88963	0.0001922	229.88944	0	229.88944	229.88944
V-C5DBT	2.805E-05	0.0001501	1158.7616	3.618E-06	1158.7616	0	1158.7616	1158.7616
V-C12NPH	0.0003842	0.0020559	6776.2927	3.005E-05	6776.2927	0	6776.2927	6776.2927
V-HMCHRY	1.669E-06	8.931E-06	28301.457	3.56E-08	28301.457	0	28301.457	28301.457
V-NC33	0.0003402	0.0020341	68636.541	1.686E-06	68636.541	0	68636.541	68636.541
P-IC4CY6	119.98157	642.09185	86142.703	145.99248	85996.71	0	85996.71	85996.71
P-C12H20	21.745996	116.37554	105616.94	18.203927	105598.74	0	105598.74	105598.74
P-C14H20	0.5268043	2.8192369	9880.7292	0.302308	9880.4269	0	9880.4269	9880.4269
P-3C3CY6	41.422368	221.67535	44054.936	25.521512	44029.415	0	44029.415	44029.415
P-C17H20	0.0968693	0.5184036	45073.897	0.0264013	45073.871	0	45073.871	45073.871
P-C17H24	0.1436983	0.7690128	24387.352	0.0467711	24387.305	0	24387.305	24387.305
P-NC17	0.7053415	3.7746911	65622.985	0.1453431	65622.84	0	65622.84	65622.84
P-C18H32	1.3552698	7.2528347	80024.412	0.3751515	80024.037	0	80024.037	80024.037
P-C18H34	0.3786421	2.0263332	57793.352	0.0819022	57793.27	0	57793.27	57793.27
P-C20H40	0.0290131	0.1749164	14639.121	0.0024992	14639.118	0	14639.118	14639.118

Table L-3 (continued). Stream Flows for 50/50 LCO/VGO HT+HC (Case 41201011)

Stream →	(25)	(26)	(27)	(28)	(29)	(30)	(31)	(32)
Temp, °F	200.0	200.0	60.0	608.9	447.3	60.0	633.3	60.0
Pres, psia	35	35	15	45	15	15	20	15
Total Flow	33426.564	112060.66	116011.68	891017.33	329253.58	329253.58	561763.76	561763.76
HYDROGEN	14.336271	0.0467982	0.0484301	8.336E-22	0	0	0	0
AMMONIA	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0
WATER	1790.3938	598.02714	684.35766	9.855E-15	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
METHANE	299.63114	3.0273615	3.1836678	2.841E-27	0	0	0	0
ETHYLENE	0.0001605	4.503E-06	4.807E-06	1.127E-28	0	0	0	0
ACALDEHY	0	0	0	0	0	0	0	0

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Stream →	(25)	(26)	(27)	(28)	(29)	(30)	(31)	(32)
ETHANE	3828.6829	146.3695	157.14833	2.39E-19	0	0	0	0
PROPYLEN	7.517E-06	6.573E-07	7.169E-07	4.977E-27	0	0	0	0
ACETONE	0	0	0	0	0	0	0	0
PROPANE	9115.6221	893.19407	976.4347	1.478E-17	0	0	0	0
FURAN	0	0	0	0	0	0	0	0
2BUTENAL	0	0	0	0	0	0	0	0
BUTENE	1.769E-05	3.781E-06	4.208E-06	1.527E-23	0	0	0	0
THF	0	0	0	0	0	0	0	0
2BUTNONE	0	0	0	0	0	0	0	0
1BUTANAL	0	0	0	0	0	0	0	0
NBUTANE	5245.603	1276.26	1425.1707	1.228E-14	0	0	0	0
P-IC4	9395.9462	1776.2079	1968.6566	3.34E-15	0	0	0	0
2MEFURAN	0	0	0	0	0	0	0	0
3MEFURAN	0	0	0	0	0	0	0	0
CYPNTENO	0	0	0	0	0	0	0	0
CYPNTANO	0	0	0	0	0	0	0	0
2PNTENAL	0	0	0	0	0	0	0	0
3PNTE2ON	0	0	0	0	0	0	0	0
2PENTENE	0.0007861	0.0004698	0.0005362	2.086E-18	0	0	0	0
2PENTNON	0	0	0	0	0	0	0	0
3PENTNON	0	0	0	0	0	0	0	0
PENTANE	0.0106128	0.0061672	0.0070205	2.422E-17	0	0	0	0
BENZENE	0.0125475	0.0201323	0.0231764	1.718E-13	0	0	0	0
PHENOL	0	0	0	0	0	0	0	0
12BZDIOL	0	0	0	0	0	0	0	0
CYHXDIEN	0	0	0	0	0	0	0	0
DIMEFURA	8.011E-05	0.0001969	0.0002265	1.175E-14	0	0	0	0
METCY5ON	0	0	0	0	0	0	0	0
2ETFURAN	0	0	0	0	0	0	0	0
CYHXENE	3.714E-05	6.952E-05	7.981E-05	2.358E-15	0	0	0	0
2MECY5ON	0	0	0	0	0	0	0	0
1HEXANAL	0	0	0	0	0	0	0	0
2HEXNONE	0	0	0	0	0	0	0	0
3HEXNONE	0	0	0	0	0	0	0	0
CYHEXANE	0.0151032	0.0274298	0.0314502	9.325E-13	0	0	0	0
NHEXANE	0.016054	0.0217069	0.025016	2.97E-14	0	0	0	0
SORBTOL	0	0	0	0	0	0	0	0
TOLUENE	6.984E-05	0.0002745	0.0003095	1.324E-12	0	0	0	0
OCRESOL	0	0	0	0	0	0	0	0
GUAIACOL	0	0	0	0	0	0	0	0
MECYHEXN	0.0003164	0.0010259	0.0011604	3.031E-12	0	0	0	0
2M3HXENE	0.0083274	0.0229193	0.0261741	4.821E-12	0	0	0	0
METHHEXN	0.0051768	0.012281	0.0140547	1.124E-12	0	0	0	0
3ETPNTNO	0	0	0	0	0	0	0	0
D-INDOLE	0	0	0	0	0	0	0	0
OXYLENE	0.0016983	0.0179132	0.0192395	1.048E-07	1.048E-07	1.048E-07	1.512E-12	1.512E-12
ETYBENZN	0.0004517	0.003865	0.0041943	5.771E-09	5.771E-09	5.771E-09	5.493E-14	5.493E-14
23XYLNOL	0	0	0	0	0	0	0	0
SYRINGOL	0	0	0	0	0	0	0	0
VANALC	0	0	0	0	0	0	0	0
TRIMEPYR	0	0	0	0	0	0	0	0
MEETCY5E	0.0003588	0.0034976	0.003771	1.124E-08	1.124E-08	1.124E-08	1.353E-13	1.353E-13
DIMECYHX	0.0033925	0.0240014	0.0261957	8.079E-08	8.079E-08	8.079E-08	1.213E-12	1.213E-12
TRDICYHN	0.0032276	0.0193899	0.0213304	1.526E-08	1.526E-08	1.526E-08	1.406E-13	1.406E-13
ETCYHEXN	0.0015431	0.0116884	0.0127258	4.31E-08	4.31E-08	4.31E-08	6.426E-13	6.426E-13
4MEHEPTN	0.0010824	0.0057241	0.0063637	1.637E-10	0	0	0	0

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Stream →	(25)	(26)	(27)	(28)	(29)	(30)	(31)	(32)
MEBZFURA	0	0	0	0	0	0	0	0
MEVINPHE	0	0	0	0	0	0	0	0
PROPBENZ	0.0004533	0.0081647	0.0085607	3.352E-06	3.352E-06	3.352E-06	1.88E-10	1.88E-10
ETMEBENZ	3.03E-05	0.0006084	0.0006354	3.705E-07	3.705E-07	3.705E-07	2.256E-11	2.256E-11
TRIMEBEN	1.574E-05	0.0004572	0.0004718	4.947E-06	4.947E-06	4.947E-06	7.057E-10	7.057E-10
PROPCYHX	0.0011885	0.0181979	0.0191826	2.487E-05	2.487E-05	2.487E-05	2.51E-09	2.51E-09
NAPTHALE	3.73E-07	2.82E-05	2.856E-05	0.0003458	0.0003451	0.0003451	7.364E-07	7.364E-07
1NAPTHOL	0	0	0	0	0	0	0	0
CONIFALD	0	0	0	0	0	0	0	0
TETRALIN	2.417E-05	0.001615	0.0016385	0.0100487	0.0100347	0.0100347	1.396E-05	1.396E-05
IEUGENOL	0	0	0	0	0	0	0	0
DECALIN	31.694357	1930.594	1961.3409	3871.6014	3868.5376	3868.5376	3.0638124	3.0638124
D-C4CYC6	199.25011	6147.5181	6330.4717	3195.5623	3193.3568	3193.3568	2.2055656	2.2055656
DIMEOCTN	0.0005223	0.0095544	0.0100105	3.015E-06	3.015E-06	3.015E-06	1.614E-10	1.614E-10
1MENAPHT	0	0	0	0	0	0	0	0
METDEC	9.055E-06	0.000998	0.001007	0.0452697	0.0450278	0.0450278	0.0002419	0.0002419
DIBZTHIO	0	0	0	0	0	0	0	0
BIPHENYL	1.716E-08	5.756E-06	5.773E-06	0.0062357	0.0059844	0.0059844	0.0002512	0.0002512
DIMENAPH	8.473E-10	3.977E-07	3.977E-07	0.0012274	0.0011308	0.0011308	9.658E-05	9.658E-05
2ETDECLN	3.2396048	401.7336	404.92452	54891.297	54001.468	54001.468	889.8291	889.8291
HXLCRESO	0	0	0	0	0	0	0	0
HEPCYHEX	8.014E-08	2.676E-05	2.684E-05	0.0335325	0.0317455	0.0317455	0.001787	0.001787
PHENANTH	1.329E-07	0.0006849	0.000685	3970.3747	65.151555	65.151555	3905.2231	3905.2231
ETPHCYHX	1.295E-09	8.34E-07	8.34E-07	0.0084817	0.0070335	0.0070335	0.0014482	0.0014482
D-HYPHEN	0.0004127	0.4139624	0.4143747	20093.357	13045.712	13045.712	7047.6451	7047.6451
THYDPHEN	1.922E-09	1.326E-06	1.328E-06	0.028736	0.0205895	0.0205895	0.0081465	0.0081465
DIMDODEC	3.67E-06	0.0008317	0.0008354	0.2685577	0.2630737	0.2630737	0.005484	0.005484
D-3CZOLE	4.385E-12	1.385E-07	1.385E-07	156.18073	1.408E-05	1.408E-05	156.18071	156.18071
D-C15H16	1.406E-06	0.0046557	0.0046571	6461.0013	985.06039	985.06039	5475.9409	5475.9409
D-3C3BEN	0.0374517	11.553459	11.590759	9114.4974	8799.9914	8799.9914	314.50601	314.50601
OCHYENAP	2.154E-12	8.345E-09	8.345E-09	0.0228413	0.0009548	0.0009548	0.0218864	0.0218864
D-DIMTRD	7.439E-08	2.572E-05	2.579E-05	0.0381416	0.0357597	0.0357597	0.0023819	0.0023819
DIMPHENA	8.786E-17	3.364E-12	3.364E-12	0.0037785	6.381E-10	6.381E-10	0.0037785	0.0037785
CETANE	0.0001607	0.3205476	0.3207083	61794.043	30084.203	30084.203	31709.84	31709.84
CYHXPHON	0	0	0	0	0	0	0	0
CYHXHXBZ	1.201E-15	3.008E-11	3.008E-11	0.0181705	3.828E-09	3.828E-09	0.0181705	0.0181705
D-NC18	1.137E-06	0.0072831	0.0072842	54087.962	270.43974	270.43974	53817.522	53817.522
HMWLIGNB	0	0	0	0	0	0	0	0
TRIPHNPR	2.679E-21	4.437E-16	4.437E-16	0.0001853	2.905E-18	2.905E-18	0.0001853	0.0001853
MOXHXPHE	0	0	0	0	0	0	0	0
DIBZFLOR	5.585E-20	2.614E-14	2.614E-14	0.0150028	2.468E-15	2.468E-15	0.0150028	0.0150028
HXACMOXP	0	0	0	0	0	0	0	0
MOXHEPOX	0	0	0	0	0	0	0	0
HXDECBEN	2.841E-13	6.735E-08	6.735E-08	11167.893	8.114E-09	8.114E-09	11167.893	11167.893
L-3MQUIN	6.389E-05	0.0190889	0.0191525	8.2846948	8.146411	8.146411	0.1382837	0.1382837
L-NC10	315.95888	10250.75	10546.652	140.253	140.23102	140.23102	0.0219738	0.0219738
L-ACENPH	0.0021686	1.5014942	1.5036596	21135.305	16889.362	16889.362	4245.9433	4245.9433
L-TMC9	141.50032	6447.5652	6582.4777	5122.0723	5118.7059	5118.7059	3.3664375	3.3664375
L-C3PHEN	1.229E-12	1.503E-07	1.503E-07	5016.5929	2.565E-07	2.565E-07	5016.5929	5016.5929
L-C4PHEN	3.222E-14	1.323E-08	1.323E-08	16094.929	1.463E-10	1.463E-10	16094.929	16094.929
L-C8NPH	1.185E-09	3.161E-05	3.161E-05	11827.491	0.0175408	0.0175408	11827.473	11827.473
L-NC14BZ	1.554E-11	9.277E-07	9.277E-07	3054.2371	2.384E-05	2.384E-05	3054.237	3054.237
L-NC22	5.816E-11	3.129E-06	3.129E-06	18898.371	5.816E-06	5.816E-06	18898.371	18898.371
V-23DMBT	0.0019496	0.8927032	0.8946483	2251.0351	2103.6322	2103.6322	147.40296	147.40296
V-NC12	3.3765188	486.41013	489.75254	25245.021	25129.086	25129.086	115.93531	115.93531
V-2MBZQN	3.482E-10	4.808E-06	4.808E-06	229.88943	0.0828617	0.0828617	229.80657	229.80657

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Stream →	(25)	(26)	(27)	(28)	(29)	(30)	(31)	(32)
V-C5DBT	5.069E-17	5.762E-11	5.762E-11	1158.7616	1.764E-14	1.764E-14	1158.7616	1158.7616
V-C12NPH	1.055E-15	6.966E-10	6.966E-10	6776.2927	1.127E-13	1.127E-13	6776.2927	6776.2927
V-HMCHRY	6.246E-27	3.686E-18	3.686E-18	28301.457	2.113E-34	2.113E-34	28301.457	28301.457
V-NC33	5.891E-20	7.681E-13	7.681E-13	68636.541	2.917E-23	2.917E-23	68636.541	68636.541
P-IC4CY6	3034.2657	80991.601	83762.964	1970.8431	1970.418	1970.418	0.4251011	0.4251011
P-C12H20	2.0900036	376.98696	379.05895	105219.66	102955.02	102955.02	2264.6371	2264.6371
P-C14H20	0.0010219	0.7006501	0.7016704	9879.7252	7812.5379	7812.5379	2067.1872	2067.1872
P-3C3CY6	4.846727	318.61871	323.22029	43705.949	41868.791	41868.791	1837.1586	1837.1586
P-C17H20	4.386E-08	0.0006529	0.000653	45073.87	3.8774511	3.8774511	45069.993	45069.993
P-C17H24	2.435E-07	0.0015711	0.0015714	24387.303	22.507366	22.507366	24364.796	24364.796
P-NC17	1.456E-05	0.0528085	0.0528231	65622.787	9630.4094	9630.4094	55992.378	55992.378
P-C18H32	1.105E-05	0.0307331	0.0307441	80024.006	1215.8183	1215.8183	78808.188	78808.188
P-C18H34	7.542E-07	0.0047391	0.0047399	57793.266	70.586033	70.586033	57722.68	57722.68
P-C20H40	9.693E-10	2.315E-05	2.315E-05	14639.118	0.0008936	0.0008936	14639.118	14639.118

Table L-4. Heat Duties and Power for 50/50 LCO/VGO HT+HC (Case 41201011)

Tag	Description	Value	Unit
Duty			
Q1	Feed-effluent exchanger	452.1	MMBtu/h
Q2	Feed preheat furnace	46.8	MMBtu/h
Q3	Effluent cooler (makes steam)	255.8	MMBtu/h
Q4	Air cooler (cools to 140°F)	67.7	MMBtu/h
Q5	Water cooler (cools to 110°F)	57.7	MMBtu/h
Q6	Distillation cross exchange preheat	217.2	MMBtu/h
Q7	Distillation feed preheat furnace	19.4	MMBtu/h
Q8	Light products column condenser	86.1	MMBtu/h
Q9	Light products column reboiler	154.8	MMBtu/h
Q10	Heavier products column condenser	95.5	MMBtu/h
Q11	Heavier products column reboiler	69.3	MMBtu/h
Power			
W1	Makeup hydrogen compressor	38,679	kW
W2	Recycle compressor	2,904	kW
W3	Liquid recycle pump	-	kW

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Table L-5. Stream Flows for 100% CFP-B Oil HT (Case 11199011)

Stream →	(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)
Temp, °F	116.4	135.0	135.0	140.0	644.0	683.9	448.0	276.7
Pres, psia	1100	1100	1100	995.00002	1045	1015	1010	1005
Total Flow	1152435	573220.36	1685942.2	91589.098	1725653.4	2838375.3	2838375.3	2838375.3
HYDROGEN	2.0504679	149433.19	439509.39	84774.439	149435.18	374842.19	374842.19	374842.19
AMMONIA	0	1.024E-14	3.012E-14	0	3.785E-14	1033.4505	1033.4505	1033.4505
H2S	0	7.747E-13	2.279E-12	0	2.863E-12	351.99548	351.99548	351.99548
WATER	9455.4757	2378.4896	6995.5578	0	11833.961	192755.49	192755.49	192755.49
CO	14.999983	2.299E-11	6.762E-11	0	14.999983	0	0	0
CO2	730.68154	0	0	0	730.68154	35898.238	35898.238	35898.238
METHANE	2.79E+01	1.79E+05	5.27E+05	6.81E+03	1.79E+05	5.53E+05	5.53E+05	5.53E+05
ETHYLENE	77.984554	6354.3659	18689.312	0	6432.367	20217.45	20217.45	20217.45
ACALDEHY	7221.827	1.734E-09	5.1E-09	0	7221.8269	0	0	0
HYDACTLD	0.0003988	4.384E-18	1.29E-17	0	0.0003988	0	0	0
ACETACID	0.0005895	0	0	0	0.0005895	0	0	0
ETHANE	79.213975	182990.09	538206.16	0	183068.93	587885.25	587885.25	587885.25
PROPYLEN	425.4552	26.96632	79.312707	0	452.42153	91.146403	91.146403	91.146403
ACETONE	12599.423	1.41E-09	4.147E-09	0	12599.423	0	0	0
ACETOL	0.0001924	2.005E-18	5.898E-18	0	0.0001924	0	0	0
PROPACID	8.503E-05	0	0	0	8.503E-05	0	0	0
PROPANE	160.66143	21343.156	62773.989	0	21504.998	72935.221	72935.221	72935.221
PROPANOL	0	2.73E-18	8.031E-18	0	1.191E-17	0	0	0
FURAN	2066.7711	0	0	0	2066.7711	0	0	0
2BUTENAL	3111.9633	1.068E-10	3.142E-10	0	3111.9633	0	0	0
BUTENE	376.79638	20.160901	59.296769	0	396.95729	78.316942	78.316942	78.316942
THF	1795.6591	3.708E-10	1.091E-09	0	1795.6591	0	0	0
2BUTNONE	7004.4404	4.346E-10	1.278E-09	0	7004.4404	0	0	0
1BUTANAL	2294.3455	1.643E-10	4.833E-10	0	2294.3455	0	0	0
NBUTANE	0	13851.206	40738.84	0	13848.779	54395.225	54395.225	54395.225
P-IC4	0	0.0334995	0.0985278	0	0.037538	0.1144026	0.1144026	0.1144026
2MEFURAN	7144.5512	6.287E-10	1.849E-09	0	7144.5512	0	0	0
3MEFURAN	1132.1445	1.019E-10	2.996E-10	0	1132.1445	0	0	0
CYPNTENO	1889.9617	4.646E-11	1.367E-10	0	1889.9617	0	0	0
FURFUALC	0.0006193	2.936E-18	8.634E-18	0	0.0006193	0	0	0
CYPNTANO	4364.0107	8.447E-11	2.484E-10	0	4364.0107	0	0	0
2PNTENAL	5973.715	1.18E-10	3.472E-10	0	5973.715	0	0	0
3PNTE2ON	1807.0692	3.073E-11	9.038E-11	0	1807.0692	0	0	0
2PENTENE	0	364.5842	1072.3065	0	364.58479	2061.1357	2061.1357	2061.1357
2PENTNON	2521.6268	1.046E-10	3.078E-10	0	2521.6268	0	0	0
3PENTNON	1524.9925	5.41E-11	1.591E-10	0	1524.9925	0	0	0
THFA	0	2.317E-06	6.813E-06	0	2.317E-06	0.0006516	0.0006516	0.0006516
PENTANE	127.27705	8367.0937	24609.099	0	8494.3404	46460.024	46460.024	46460.024
BENZENE	3442.0375	210.93373	620.39331	0	3652.9717	2299.4484	2299.4484	2299.4484
PHENOL	58669.168	3.272E-10	9.622E-10	0	58669.168	0	0	0
12BZDIOL	8828.7598	9.702E-12	2.853E-11	0	8828.7598	0	0	0
HMF	0.000722	7.524E-08	2.213E-07	0	0.000722	0.0007222	0.0007222	0.0007222
CYHXDIEN	111.27228	7.246E-12	2.131E-11	0	111.27228	0	0	0
DIMEFURA	2915.804	18.936033	55.694216	0	2934.7401	312.06011	312.06011	312.06011
METCY5ON	1368.226	1.173E-11	3.451E-11	0	1368.226	0	0	0
2ETFURAN	2630.1155	1.234E-10	3.628E-10	0	2630.1155	0	0	0
HYMET5ON	0.0003509	4.989E-07	1.467E-06	0	0.0003514	0.0003523	0.0003523	0.0003523
CYHXENE	0	12.242715	36.007986	0	12.242727	150.07964	150.07964	150.07964
CYHEXNON	5.151E-05	3.288E-07	9.671E-07	0	5.184E-05	2.984E-05	2.984E-05	2.984E-05

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Stream →	(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)
2MECY5ON	1804.8846	2.648E-11	7.788E-11	0	1804.8846	0	0	0
LEVOGLUC	1.76E-01	3.29E-19	9.68E-19	0.00E+00	1.76E-01	0.00E+00	0.00E+00	0.00E+00
1HEXANAL	1751.1627	3.087E-11	9.08E-11	0	1751.1627	0	0	0
2HEXNONE	844.20065	1.461E-11	4.296E-11	0	844.20065	0	0	0
3HEXNONE	778.92188	1.517E-11	4.463E-11	0	778.92188	0	0	0
CYHEXANE	8805.4529	2192.6101	6448.8532	0	10998.065	25958.562	25958.562	25958.562
NHEXANE	0	1547.9507	4552.7961	0	1547.952	14714.841	14714.841	14714.841
SORBTOL	0	8.686E-34	2.555E-33	0	3.21E-33	0	0	0
TOLUENE	2573.9985	17.513558	51.510465	0	2591.5121	374.60779	374.60779	374.60779
OCRESOL	2854.9532	1.027E-11	3.021E-11	0	2854.9532	0	0	0
GUAIACOL	42648.873	7.467E-11	2.196E-10	0	42648.873	0	0	0
MECYHEXN	1011.6659	79.315944	233.28219	0	1090.982	1554.2781	1554.2781	1554.2781
2M3HXENE	0	2242.0239	6594.1879	0	2242.0281	39426.781	39426.781	39426.781
METHHEXN	0	125.39736	368.81577	0	125.39766	1894.8572	1894.8572	1894.8572
3ETPNTNO	3682.8458	5.33E-11	1.568E-10	0	3682.8458	0	0	0
D-INDOLE	0.0034594	3.63E-08	1.068E-07	0	0.0034594	7.618E-05	7.618E-05	7.618E-05
OXYLENE	30277.395	388.64342	1143.0689	0	30666.038	27519.796	27519.796	27519.796
ETYBENZN	0	173.73834	510.99511	0	173.73878	9673.9005	9673.9005	9673.9005
23XYLNOL	49845.995	7.265E-11	2.137E-10	0	49845.995	0	0	0
SYRINGOL	12093.404	2.565E-12	7.545E-12	0	12093.404	0	0	0
VANALC	19509.451	1.137E-13	3.344E-13	0	19509.451	0	0	0
TRIMEPYR	7353.5779	3.984E-07	1.172E-06	0	7353.5779	7.621E-05	7.621E-05	7.621E-05
MEETCY5E	459.41243	64.171032	188.73833	0	523.58362	4142.5007	4142.5007	4142.5007
DIMECYHX	365.91085	0.0157765	0.0464015	0	365.92662	0.6342353	0.6342353	0.6342353
TRDICYHN	1020.5333	763.16279	2244.5964	0	1783.6981	26144.449	26144.449	26144.449
ETCYHEXN	0	206.21681	606.52002	0	206.21732	9115.3544	9115.3544	9115.3544
4MEHEPTN	0	230.39631	677.6362	0	230.39686	7609.4039	7609.4039	7609.4039
MEBZFURA	10021.329	2.735E-11	8.043E-11	0	10021.329	0	0	0
MEVINPHE	301.85451	2.326E-13	6.84E-13	0	301.85451	0	0	0
PROPBENZ	6921.9097	222.85467	655.45493	0	7144.7649	27915.566	27915.566	27915.566
ETMEBENZ	641.71769	4.5266388	13.313643	0	646.24434	655.03135	655.03135	655.03135
TRIMEBEN	479.17869	2.1991775	6.4681693	0	481.37787	485.64686	485.64686	485.64686
NAPTHALE	26018.909	44.577106	131.10913	0	26063.486	27642.879	27642.879	27642.879
1NAPTHOL	25274.476	4.396E-12	1.293E-11	0	25274.476	0	0	0
CONIFALD	26054.436	1.853E-13	5.449E-13	0	26054.436	0	0	0
TETRALIN	0.1018772	60.994312	179.39504	0	61.096348	28144.102	28144.102	28144.102
IEUGENOL	27785.146	5.069E-12	1.491E-11	0	27785.146	0	0	0
D-DECALN	0	0.0001462	0.0004301	0	0.0001462	0.0399649	0.0399649	0.0399649
DECALIN	13339.051	26.387755	77.611044	0	13365.439	12930.74	12930.74	12930.74
D-C4CYC6	0.125637	0.0006472	0.0019034	0	0.1262842	0.1238469	0.1238469	0.1238469
DIMEOCTN	0	83.594888	245.86732	0	83.595088	10390.065	10390.065	10390.065
1MENAPHT	4137.878	0.9013199	2.6509409	0	4138.7794	1688.3908	1688.3908	1688.3908
METDEC	4589.4258	17.078494	50.230865	0	4606.5043	15633.439	15633.439	15633.439
DIBZTHIO	1902.9898	1.14E-13	3.352E-13	0	1902.9898	0	0	0
D-C4BZ-S	0.0286279	1.267E-08	3.726E-08	0	0.0286279	0.0006304	0.0006304	0.0006304
BIPHENYL	0	1.7784076	5.2306105	0	1.778411	6195.6093	6195.6093	6195.6093
DIMENAPH	0	0.0424352	0.1248094	0	0.0424353	214.66773	214.66773	214.66773
D-C12H18	0	2.345E-05	6.896E-05	0	2.345E-05	0.0238685	0.0238685	0.0238685
D-C12H1B	0	3.253E-06	9.569E-06	0	3.253E-06	0.0029637	0.0029637	0.0029637
CNYLXNE	0	1.093E-08	3.216E-08	0	1.093E-08	1.052E-05	1.052E-05	1.052E-05
2ETDECLN	0	1.4111616	4.1504754	0	1.4111644	1298.4117	1298.4117	1298.4117
D-CQUINL	0.0034594	3.382E-10	9.948E-10	0	0.0034594	7.617E-05	7.617E-05	7.617E-05
HXLCRESO	5276.7663	4.383E-14	1.289E-13	0	5276.7663	0	0	0
HEPCYHEX	0	1.5733395	4.6274691	0	1.5733424	4757.9005	4757.9005	4757.9005
D-2C2BZS	0.0286279	8.283E-10	2.436E-09	0	0.0286279	0.0006304	0.0006304	0.0006304
D-C14H14	0	1.034E-06	3.043E-06	0	1.034E-06	0.0239644	0.0239644	0.0239644

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Stream →	(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)
ETPHCYHX	0	6.9668294	20.490675	0	6.9668408	46737.674	46737.674	46737.674
D-HYPHEN	0.0543351	4.864E-06	1.431E-05	0	0.05434	0.0527601	0.0527601	0.0527601
THYDPHEN	10992.6	6.6736762	19.628459	0	10999.273	43747.327	43747.327	43747.327
DIMDODEC	0	108.75864	319.87834	0	108.75884	216132.25	216132.25	216132.25
D-3CZOLE	0.0034594	1.298E-10	3.817E-10	0	0.0034594	7.617E-05	7.617E-05	7.617E-05
D-C15H16	0	7.26E-08	2.135E-07	0	7.26E-08	0.0031636	0.0031636	0.0031636
D-3C3BEN	0.1400819	2.828E-05	8.319E-05	0	0.1401102	0.0842298	0.0842298	0.0842298
OCHYENAP	11649.522	0.2483706	0.7305016	0	11649.771	11951.974	11951.974	11951.974
D-DIMTRD	0	12.039837	35.411286	0	12.039837	36513.454	36513.454	36513.454
DIMPHENA	9407.9708	0.0047208	0.0138846	0	9407.9756	3812.6195	3812.6195	3812.6195
CETANE	0	1.8189827	5.3499491	0	1.8189858	43098.954	43098.954	43098.954
CYHXPHON	84432.429	2.225E-13	6.545E-13	0	84432.429	0	0	0
CYHXHXBZ	0	0.1913734	0.562863	0	0.1913737	77716.004	77716.004	77716.004
D-NC18	0.1290459	1.497E-06	4.404E-06	0	0.1290474	0.125276	0.125276	0.125276
D-C1C17	0.0865966	8.621E-07	2.536E-06	0	0.0865975	0.0840664	0.0840664	0.0840664
D-NC20	0.0101878	3.569E-08	1.05E-07	0	0.0101879	0.00989	0.00989	0.00989
HMWLIGNB	132839.97	1.42E-30	4.178E-30	0	132839.97	0	0	0
TRIPHNPR	0	6.052E-05	0.000178	0	6.052E-05	183.20618	183.20618	183.20618
MOXHXPHE	85374.244	4.514E-14	1.328E-13	0	85374.244	0	0	0
DIBZFLOR	107388.93	0.0008952	0.0026329	0	107388.94	14621.375	14621.375	14621.375
HXACMOXP	200177.85	9.604E-15	2.825E-14	0	200177.85	0	0	0
MOXHEPOX	17846.363	6.45E-15	1.897E-14	0	17846.363	0	0	0
HXDECBEN	0	0.0157232	0.0462446	0	0.0157232	93011.672	93011.672	93011.672
P-3C3CY6	0.1086282	0.0005397	0.0015873	0	0.1091679	0.1603942	0.1603942	0.1603942

Table L-5 (continued). Stream Flows for 100% CFP-B Oil HT (Case 11199011)

Stream →	(9)	(10)	(11)	(12)	(13)	(14)	(15)	(16)
Temp, °F	110.0	110.0	109.9	110.0	110.0	110.0	110.0	110.0
Pres, psia	1050	995	993	993	993	20	993	973
Total Flow	0	2838375.3	182341.95	1718080.1	1681324.3	36755.798	86966.158	16447.031
HYDROGEN	0	374842.19	0.5034302	374084.56	374084.56	0	19349.448	16447.031
AMMONIA	0	1033.4505	175.89053	857.55995	0	857.55995	0	0
H2S	0	351.99548	351.99548	0	0	0	0	0
WATER	0	192755.49	181809.38	7377.2045	7377.2045	0	381.58441	0
CO	0	0	0	0	0	0	0	0
CO2	0	35898.238	0	35898.238	0	35898.238	0	0
METHANE	0.00E+00	5.53E+05	1.85E+00	5.49E+05	5.49E+05	0.00E+00	2.84E+04	0.00E+00
ETHYLENE	0	20217.45	0.1128376	19708.696	19708.696	0	1019.4283	0
ACALDEHY	0	0	0	0	0	0	0	0
HYDACTLD	0	0	0	0	0	0	0	0
ACETACID	0	0	0	0	0	0	0	0
ETHANE	0	587885.25	2.1602411	567563.71	567563.71	0	29357.118	0
PROPYLEN	0	91.146403	0.0001546	83.638935	83.638935	0	4.326207	0
ACETONE	0	0	0	0	0	0	0	0
ACETOL	0	0	0	0	0	0	0	0
PROPACID	0	0	0	0	0	0	0	0
PROPANE	0	72935.221	0.0334084	66195.351	66195.351	0	3423.9411	0
PROPANOL	0	0	0	0	0	0	0	0
FURAN	0	0	0	0	0	0	0	0
2BUTENAL	0	0	0	0	0	0	0	0
BUTENE	0	78.316942	1.208E-05	62.531232	62.531232	0	3.2344153	0
THF	0	0	0	0	0	0	0	0
2BUTNONE	0	0	0	0	0	0	0	0
1BUTANAL	0	0	0	0	0	0	0	0

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Stream →	(9)	(10)	(11)	(12)	(13)	(14)	(15)	(16)
NBUTANE	0	54395.225	0.0022683	42967.371	42967.371	0	2222.4786	0
P-IC4	0	0.1144026	2.039E-09	0.0939788	0.0939788	0	0.004861	0
2MEFURAN	0	0	0	0	0	0	0	0
3MEFURAN	0	0	0	0	0	0	0	0
CYPNTENO	0	0	0	0	0	0	0	0
FURFUALC	0	0	0	0	0	0	0	0
CYPNTANO	0	0	0	0	0	0	0	0
2PNTENAL	0	0	0	0	0	0	0	0
3PNTE2ON	0	0	0	0	0	0	0	0
2PENTENE	0	2061.1357	2.988E-05	1130.7995	1130.7995	0	58.490375	0
2PENTNON	0	0	0	0	0	0	0	0
3PENTNON	0	0	0	0	0	0	0	0
THFA	0	0.0006516	1.493E-07	7.185E-06	7.185E-06	0	3.716E-07	0
PENTANE	0	46460.024	8.627E-05	25951.551	25951.551	0	1342.3387	0
BENZENE	0	2299.4484	0.0205921	654.23619	654.23619	0	33.840235	0
PHENOL	0	0	0	0	0	0	0	0
12BZDIOL	0	0	0	0	0	0	0	0
HMF	0	0.0007222	2.289E-06	2.334E-07	2.334E-07	0	1.207E-08	0
CYHXDIEN	0	0	0	0	0	0	0	0
DIMEFURA	0	312.06011	0.000106	58.732444	58.732444	0	3.0379238	0
METCY5ON	0	0	0	0	0	0	0	0
2ETFURAN	0	0	0	0	0	0	0	0
HYMET5ON	0	0.0003523	4.589E-07	1.547E-06	1.547E-06	0	8.004E-08	0
CYHXENE	0	150.07964	8.727E-05	37.972258	37.972258	0	1.9641074	0
CYHEXNON	0	2.984E-05	1.744E-12	1.02E-06	1.02E-06	0	5.275E-08	0
2MECY5ON	0	0	0	0	0	0	0	0
LEVOGLUC	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
1HEXANAL	0	0	0	0	0	0	0	0
2HEXNONE	0	0	0	0	0	0	0	0
3HEXNONE	0	0	0	0	0	0	0	0
CYHEXANE	0	25958.562	0.0029339	6800.644	6800.644	0	351.76193	0
NHEXANE	0	14714.841	6.379E-07	4801.1536	4801.1536	0	248.3387	0
SORBTOL	0	0	0	0	0	0	0	0
TOLUENE	0	374.60779	5.767E-05	54.320424	54.320424	0	2.8097129	0
OCRESOL	0	0	0	0	0	0	0	0
GUAIACOL	0	0	0	0	0	0	0	0
MECYHEXN	0	1554.2781	1.438E-06	246.00828	246.00828	0	12.724728	0
2M3HXENE	0	39426.781	2.08E-07	6953.9148	6953.9148	0	359.68984	0
METHHEXN	0	1894.8572	1.301E-09	388.93545	388.93545	0	20.117607	0
3ETPNTNO	0	0	0	0	0	0	0	0
D-INDOLE	0	7.618E-05	5.117E-11	1.126E-07	1.126E-07	0	5.823E-09	0
OXYLENE	0	27519.796	0.0003284	1205.4294	1205.4294	0	62.350592	0
ETYBENZN	0	9673.9005	3.215E-05	538.8719	538.8719	0	27.87304	0
23XYLNOL	0	0	0	0	0	0	0	0
SYRINGOL	0	0	0	0	0	0	0	0
VANALC	0	0	0	0	0	0	0	0
TRIMEPYR	0	7.621E-05	1.607E-14	1.235E-06	1.235E-06	0	6.39E-08	0
MEETCY5E	0	4142.5007	2.922E-06	199.0348	199.0348	0	10.295035	0
DIMECYHX	0	0.6342353	3.673E-13	0.0489357	0.0489357	0	0.0025312	0
TRDICYHN	0	26144.449	2.154E-08	2367.0468	2367.0468	0	122.43502	0
ETCYHEXN	0	9115.3544	2.459E-08	639.60802	639.60802	0	33.083595	0
4MEHEPTN	0	7609.4039	1.01E-10	714.60367	714.60367	0	36.96273	0
MEBZFURA	0	0	0	0	0	0	0	0
MEVINPHE	0	0	0	0	0	0	0	0
PROPBENZ	0	27915.566	1.064E-06	691.21331	691.21331	0	35.752869	0
ETMEBENZ	0	655.03135	4.604E-08	14.039972	14.039972	0	0.7262147	0

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Stream →	(9)	(10)	(11)	(12)	(13)	(14)	(15)	(16)
TRIMEBEN	0	485.64686	3.214E-07	6.8210436	6.8210436	0	0.3528171	0
NAPTHALE	0	27642.879	0.0009844	138.26197	138.26197	0	7.1515722	0
1NAPTHOL	0	0	0	0	0	0	0	0
CONIFALD	0	0	0	0	0	0	0	0
TETRALIN	0	28144.102	4.768E-05	189.18179	189.18179	0	9.7853897	0
IEUGENOL	0	0	0	0	0	0	0	0
D-DECALN	0	0.0399649	3.091E-13	0.0004535	0.0004535	0	2.346E-05	0
DECALIN	0	12930.74	3.536E-15	81.84527	81.84527	0	4.23343	0
D-C4CYC6	0	0.1238469	2.134E-17	0.0020073	0.0020073	0	0.0001038	0
DIMEOCTN	0	10390.065	1.165E-14	259.28078	259.28078	0	13.411246	0
1MENAPHT	0	1688.3908	9.495E-07	2.7955708	2.7955708	0	0.1446003	0
METDEC	0	15633.439	2.256E-10	52.971317	52.971317	0	2.7399307	0
DIBZTHIO	0	0	0	0	0	0	0	0
D-C4BZ-S	0	0.0006304	4.338E-17	3.929E-08	3.929E-08	0	2.032E-09	0
BIPHENYL	0	6195.6093	6.152E-07	5.5159815	5.5159815	0	0.285313	0
DIMENAPH	0	214.66773	1.292E-09	0.1316188	0.1316188	0	0.006808	0
D-C12H18	0	0.0238685	1.522E-18	7.272E-05	7.272E-05	0	3.762E-06	0
D-C12H1B	0	0.0029637	2.198E-19	1.009E-05	1.009E-05	0	5.22E-07	0
CNYLXNE	0	1.052E-05	2.291E-18	3.391E-08	3.391E-08	0	1.754E-09	0
2ETDECLN	0	1298.4117	5.952E-14	4.3769149	4.3769149	0	0.226395	0
D-CQUINL	0	7.617E-05	2.644E-19	1.049E-09	1.049E-09	0	5.426E-11	0
HXLCRESO	0	0	0	0	0	0	0	0
HEPCYHEX	0	4757.9005	2.593E-26	4.879941	4.879941	0	0.252414	0
D-2C2BZS	0	0.0006304	4.608E-17	2.569E-09	2.569E-09	0	1.329E-10	0
D-C14H14	0	0.0239644	1.385E-15	3.209E-06	3.209E-06	0	1.66E-07	0
ETPHCYHX	0	46737.674	4.039E-14	21.608628	21.608628	0	1.1177019	0
D-HYPHEN	0	0.0527601	8.324E-18	1.509E-05	1.509E-05	0	7.804E-07	0
THYDPHEN	0	43747.327	1.375E-15	20.699371	20.699371	0	1.0706708	0
DIMDODEC	0	216132.25	1.704E-25	337.33062	337.33062	0	17.448358	0
D-3CZOLE	0	7.617E-05	6.294E-19	4.025E-10	4.025E-10	0	2.082E-11	0
D-C15H16	0	0.0031636	4.275E-21	2.252E-07	2.252E-07	0	1.165E-08	0
D-3C3BEN	0	0.0842298	2.732E-24	8.773E-05	8.773E-05	0	4.538E-06	0
OCHYENAP	0	11951.974	4.392E-22	0.7703674	0.7703674	0	0.0398471	0
D-DIMTRD	0	36513.454	5.851E-30	37.343337	37.343337	0	1.9315765	0
DIMPHENA	0	3812.6195	1.993E-11	0.0146422	0.0146422	0	0.0007574	0
CETANE	0	43098.954	8.923E-33	5.6418501	5.6418501	0	0.2918236	0
CYHXPHON	0	0	0	0	0	0	0	0
CYHXHXBZ	0	77716.004	6.157E-26	0.5935744	0.5935744	0	0.0307025	0
D-NC18	0	0.125276	7.514E-46	4.644E-06	4.644E-06	0	2.402E-07	0
D-C1C17	0	0.0840664	1.851E-43	2.674E-06	2.674E-06	0	1.383E-07	0
D-NC20	0	0.00989	1.362E-52	1.107E-07	1.107E-07	0	5.726E-09	0
HMWLIGNB	0	0	0	0	0	0	0	0
TRIPHNPR	0	183.20618	1.401E-26	0.0001877	0.0001877	0	9.709E-06	0
MOXHXPHE	0	0	0	0	0	0	0	0
DIBZFLOR	0	14621.375	2.173E-21	0.0027766	0.0027766	0	0.0001436	0
HXACMOXP	0	0	0	0	0	0	0	0
MOXHEPOX	0	0	0	0	0	0	0	0
HXDECBEN	0	93011.672	1.682E-38	0.048768	0.048768	0	0.0025225	0
P-3C3CY6	0	0.1603942	2.301E-34	0.001674	0.001674	0	8.658E-05	0

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Table L-5 (continued). Stream Flows for 100% CFP-B Oil HT (Case 11199011)

Stream →	(17)	(18)	(19)	(20)	(21)	(22)	(23)	(24)
Temp, °F	110.0	110.0	110.0	110.0	110.0	110.0	110.0	450.0
Pres, psia	22	993.00001	993	53	53	1050	53	45.5
Total Flow	70519.126	1594353.1	937953.24	28117.427	909835.81	1.802E-05	909835.81	909835.81
HYDROGEN	2902.4172	354734.96	757.12887	744.02975	13.099119	0	13.099119	13.099119
AMMONIA	0	3.012E-14	0	0	0	0	0	0
H2S	0	2.279E-12	0	0	0	0	0	0
WATER	381.58441	6995.5578	3568.9082	491.17082	3077.7374	1.802E-05	3077.7374	3077.7374
CO	0	6.762E-11	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
METHANE	2.84E+04	5.20E+05	4.33E+03	4.02E+03	3.13E+02	0.00E+00	3.13E+02	3.13E+02
ETHYLENE	1019.4283	18689.312	508.64098	396.3115	112.32948	0	112.32948	112.32948
ACALDEHY	0	5.1E-09	0	0	0	0	0	0
HYDACTLD	0	1.29E-17	0	0	0	0	0	0
ACETACID	0	0	0	0	0	0	0	0
ETHANE	29357.118	538206.16	20319.383	14395.922	5923.4618	0	5923.4618	5923.4618
PROPYLEN	4.326207	79.312707	7.5073138	3.5065088	4.000805	0	4.000805	4.000805
ACETONE	0	4.147E-09	0	0	0	0	0	0
ACETOL	0	5.898E-18	0	0	0	0	0	0
PROPACID	0	0	0	0	0	0	0	0
PROPANE	3423.9411	62773.989	6739.8373	2904.6673	3835.17	0	3835.17	3835.17
PROPANOL	0	8.031E-18	0	0	0	0	0	0
FURAN	0	0	0	0	0	0	0	0
2BUTENAL	0	3.142E-10	0	0	0	0	0	0
BUTENE	3.2344153	59.296769	15.785699	3.4820779	12.303621	0	12.303621	12.303621
THF	0	1.091E-09	0	0	0	0	0	0
2BUTNONE	0	1.278E-09	0	0	0	0	0	0
1BUTANAL	0	4.833E-10	0	0	0	0	0	0
NBUTANE	2222.4786	40738.84	11427.852	2197.5583	9230.2934	0	9230.2934	9230.2934
P-IC4	0.004861	0.0985278	0.0204238	0.0050572	0.0153665	0	0.0153665	0.0153665
2MEFURAN	0	1.849E-09	0	0	0	0	0	0
3MEFURAN	0	2.996E-10	0	0	0	0	0	0
CYPNTENO	0	1.367E-10	0	0	0	0	0	0
FURFUALC	0	8.634E-18	0	0	0	0	0	0
CYPNTANO	0	2.484E-10	0	0	0	0	0	0
2PNTENAL	0	3.472E-10	0	0	0	0	0	0
3PNTE2ON	0	9.038E-11	0	0	0	0	0	0
2PENTENE	58.490375	1072.3065	930.3362	66.094754	864.24144	0	864.24144	864.24144
2PENTNON	0	3.078E-10	0	0	0	0	0	0
3PENTNON	0	1.591E-10	0	0	0	0	0	0
THFA	3.716E-07	6.813E-06	0.0006443	3.202E-07	0.0006439	0	0.0006439	0.0006439
PENTANE	1342.3387	24609.099	20508.472	1496.9075	19011.565	0	19011.565	19011.565
BENZENE	33.840235	620.39331	1645.1916	38.124092	1607.0675	0	1607.0675	1607.0675
PHENOL	0	9.622E-10	0	0	0	0	0	0
12BZDIOL	0	2.853E-11	0	0	0	0	0	0
HMF	1.207E-08	2.213E-07	0.0007196	7.785E-09	0.0007196	0	0.0007196	0.0007196
CYHXDIEN	0	2.131E-11	0	0	0	0	0	0
DIMEFURA	3.0379238	55.694216	253.32756	3.2169168	250.11064	0	250.11064	250.11064
METCY5ON	0	3.451E-11	0	0	0	0	0	0
2ETFURAN	0	3.628E-10	0	0	0	0	0	0
HYMET5ON	8.004E-08	1.467E-06	0.0003503	6.444E-08	0.0003503	0	0.0003503	0.0003503
CYHXENE	1.9641074	36.007986	112.10729	2.1458715	109.96142	0	109.96142	109.96142
CYHEXNON	5.275E-08	9.671E-07	2.882E-05	4.716E-08	2.877E-05	0	2.877E-05	2.877E-05
2MECY5ON	0	7.788E-11	0	0	0	0	0	0

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Stream →	(17)	(18)	(19)	(20)	(21)	(22)	(23)	(24)
LEVOGLUC	0.00E+00	9.68E-19	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
1HEXANAL	0	9.08E-11	0	0	0	0	0	0
2HEXNONE	0	4.296E-11	0	0	0	0	0	0
3HEXNONE	0	4.463E-11	0	0	0	0	0	0
CYHEXANE	351.76193	6448.8532	19157.915	382.18686	18775.728	0	18775.728	18775.728
NHEXANE	248.3387	4552.7961	9913.6871	258.05046	9655.6366	0	9655.6366	9655.6366
SORBTOL	0	2.555E-33	0	0	0	0	0	0
TOLUENE	2.8097129	51.510465	320.28731	2.3983426	317.88897	0	317.88897	317.88897
OCRESOL	0	3.021E-11	0	0	0	0	0	0
GUAIACOL	0	2.196E-10	0	0	0	0	0	0
MECYHEXN	12.724728	233.28219	1308.2698	12.694201	1295.5756	0	1295.5756	1295.5756
2M3HXENE	359.68984	6594.1879	32472.866	348.89429	32123.972	0	32123.972	32123.972
METHHEXN	20.117607	368.81577	1505.9217	19.498568	1486.4231	0	1486.4231	1486.4231
3ETPNTNO	0	1.568E-10	0	0	0	0	0	0
D-INDOLE	5.823E-09	1.068E-07	7.607E-05	4.027E-09	7.606E-05	0	7.606E-05	7.606E-05
OXYLENE	62.350592	1143.0689	26314.366	56.400443	26257.966	0	26257.966	26257.966
ETYBENZN	27.87304	510.99511	9135.0285	25.518525	9109.51	0	9109.51	9109.51
23XYLNOL	0	2.137E-10	0	0	0	0	0	0
SYRINGOL	0	7.545E-12	0	0	0	0	0	0
VANALC	0	3.344E-13	0	0	0	0	0	0
TRIMEPYR	6.39E-08	1.172E-06	7.497E-05	5.135E-08	7.492E-05	0	7.492E-05	7.492E-05
MEETCY5E	10.295035	188.73833	3943.4659	9.1943299	3934.2716	0	3934.2716	3934.2716
DIMECYHX	0.0025312	0.0464015	0.5852995	0.0022378	0.5830617	0	0.5830617	0.5830617
TRDICYHN	122.43502	2244.5964	23777.403	110.12378	23667.279	0	23667.279	23667.279
ETCYHEXN	33.083595	606.52002	8475.7464	29.331839	8446.4145	0	8446.4145	8446.4145
4MEHEPTN	36.96273	677.6362	6894.8002	32.350162	6862.45	0	6862.45	6862.45
MEBZFURA	0	8.043E-11	0	0	0	0	0	0
MEVINPHE	0	6.84E-13	0	0	0	0	0	0
PROPBENZ	35.752869	655.45493	27224.353	29.450017	27194.903	0	27194.903	27194.903
ETMEBENZ	0.7262147	13.313643	640.99137	0.5932875	640.39809	0	640.39809	640.39809
TRIMEBEN	0.3528171	6.4681693	478.82582	0.2823467	478.54347	0	478.54347	478.54347
NAPTHALE	7.1515722	131.10913	27504.616	5.383819	27499.233	0	27499.233	27499.233
1NAPTHOL	0	1.293E-11	0	0	0	0	0	0
CONIFALD	0	5.449E-13	0	0	0	0	0	0
TETRALIN	9.7853897	179.39504	27954.92	6.1664236	27948.753	0	27948.753	27948.753
IEUGENOL	0	1.491E-11	0	0	0	0	0	0
D-DECALN	2.346E-05	0.0004301	0.0395114	1.782E-05	0.0394936	0	0.0394936	0.0394936
DECALIN	4.23343	77.611044	12848.895	2.6879778	12846.207	0	12846.207	12846.207
D-C4CYC6	0.0001038	0.0019034	0.1218397	7.515E-05	0.1217645	0	0.1217645	0.1217645
DIMEOCTN	13.411246	245.86732	10130.784	9.7137913	10121.07	0	10121.07	10121.07
1MENAPHT	0.1446003	2.6509409	1685.5952	0.0984558	1685.4967	0	1685.4967	1685.4967
METDEC	2.7399307	50.230865	15580.468	1.7859933	15578.682	0	15578.682	15578.682
DIBZTHIO	0	3.352E-13	0	0	0	0	0	0
D-C4BZ-S	2.032E-09	3.726E-08	0.0006303	9.445E-10	0.0006303	0	0.0006303	0.0006303
BIPHENYL	0.285313	5.2306105	6190.0933	0.1766942	6189.9166	0	6189.9166	6189.9166
DIMENAPH	0.006808	0.1248094	214.53612	0.0039962	214.53212	0	214.53212	214.53212
D-C12H18	3.762E-06	6.896E-05	0.0237958	2.333E-06	0.0237935	0	0.0237935	0.0237935
D-C12H1B	5.22E-07	9.569E-06	0.0029536	3.273E-07	0.0029533	0	0.0029533	0.0029533
CNYLXNE	1.754E-09	3.216E-08	1.049E-05	1.162E-09	1.048E-05	0	1.048E-05	1.048E-05
2ETDECLN	0.226395	4.1504754	1294.0348	0.1413085	1293.8934	0	1293.8934	1293.8934
D-CQUINL	0	9.948E-10	7.617E-05	2.161E-11	7.617E-05	0	7.617E-05	7.617E-05
HXLCRESO	0	1.289E-13	0	0	0	0	0	0
HEPCYHEX	0.252414	4.6274691	4753.0206	0.1136704	4752.9069	0	4752.9069	4752.9069
D-2C2BZS	0	2.436E-09	0.0006304	4.866E-11	0.0006304	0	0.0006304	0.0006304
D-C14H14	1.66E-07	3.043E-06	0.0239612	8.236E-08	0.0239611	0	0.0239611	0.0239611
ETPHCYHX	1.1177019	20.490675	46716.065	0.5690385	46715.496	0	46715.496	46715.496

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Stream →	(17)	(18)	(19)	(20)	(21)	(22)	(23)	(24)
D-HYPHEN	7.804E-07	1.431E-05	0.0527451	3.96E-07	0.0527447	0	0.0527447	0.0527447
THYDPHEN	1.0706708	19.628459	43726.628	0.5328555	43726.095	0	43726.095	43726.095
DIMDODEC	17.448358	319.87834	215794.92	8.0352664	215786.88	0	215786.88	215786.88
D-3CZOLE	0	3.817E-10	7.617E-05	7.673E-12	7.617E-05	0	7.617E-05	7.617E-05
D-C15H16	1.165E-08	2.135E-07	0.0031634	5.051E-09	0.0031634	0	0.0031634	0.0031634
D-3C3BEN	4.538E-06	8.319E-05	0.0841421	2.299E-06	0.0841398	0	0.0841398	0.0841398
OCHYENAP	0.0398471	0.7305016	11951.203	0.0150407	11951.188	0	11951.188	11951.188
D-DIMTRD	1.9315765	35.411286	36476.111	0.8044795	36475.306	0	36475.306	36475.306
DIMPHENA	0.0007574	0.0138846	3812.6049	0.0002705	3812.6046	0	3812.6046	3812.6046
CETANE	0.2918236	5.3499491	43093.312	0.0965987	43093.216	0	43093.216	43093.216
CYHXPHON	0	6.545E-13	0	0	0	0	0	0
CYHXHXBZ	0.0307025	0.562863	77715.41	0.008874	77715.402	0	77715.402	77715.402
D-NC18	2.402E-07	4.404E-06	0.1252714	6.255E-08	0.1252713	0	0.1252713	0.1252713
D-C1C17	1.383E-07	2.536E-06	0.0840637	3.714E-08	0.0840637	0	0.0840637	0.0840637
D-NC20	5.726E-09	1.05E-07	0.0098899	1.169E-09	0.0098899	0	0.0098899	0.0098899
HMWLIGNB	0	4.178E-30	0	0	0	0	0	0
TRIPHNPR	9.709E-06	0.000178	183.20599	2.435E-06	183.20599	0	183.20599	183.20599
MOXHXPHE	0	1.328E-13	0	0	0	0	0	0
DIBZFLOR	0.0001436	0.0026329	14621.372	3.289E-05	14621.372	0	14621.372	14621.372
HXACMOXP	0	2.825E-14	0	0	0	0	0	0
MOXHEPOX	0	1.897E-14	0	0	0	0	0	0
HXDECBEN	0.0025225	0.0462446	93011.623	0.0004641	93011.622	0	93011.622	93011.622
P-3C3CY6	8.658E-05	0.0015873	0.1587202	4.495E-05	0.1586753	0	0.1586753	0.1586753

Table L-5 (continued). Stream Flows for 100% CFP-B Oil HT (Case 11199011)

Stream →	(25)	(26)	(27)	(28)	(29)	(30)	(31)	(32)
Temp, °F	200.0	200.0	60.0	574.7	450.4	60.0	642.0	60.0
Pres, psia	35	35	15	45	15	15	20	15
Total Flow	39951.889	224247.49	242908.14	645636.44	398421.48	398421.48	247214.96	247214.96
HYDROGEN	13.002955	0.0961632	0.1091059	0	0	0	0	0
AMMONIA	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0
WATER	1570.6804	1507.0571	2038.3279	1.614E-16	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
METHANE	3.06E+02	6.90E+00	8.26E+00	4.98E-29	0.00E+00	0.00E+00	0.00E+00	0.00E+00
ETHYLENE	105.28048	7.0489962	8.8064258	5.861E-20	0	0	0	0
ACALDEHY	0	0	0	0	0	0	0	0
HYDACTLD	0	0	0	0	0	0	0	0
ACETACID	0	0	0	0	0	0	0	0
ETHANE	5429.7345	493.72724	626.21017	4.292E-18	0	0	0	0
PROPYLEN	3.3022125	0.6985925	0.9210653	1.769E-21	0	0	0	0
ACETONE	0	0	0	0	0	0	0	0
ACETOL	0	0	0	0	0	0	0	0
PROPACID	0	0	0	0	0	0	0	0
PROPANE	3110.154	725.01607	954.06824	1E-18	0	0	0	0
PROPANOL	0	0	0	0	0	0	0	0
FURAN	0	0	0	0	0	0	0	0
2BUTENAL	0	0	0	0	0	0	0	0
BUTENE	8.0913249	4.2122959	5.6982937	7.721E-20	0	0	0	0
THF	0	0	0	0	0	0	0	0
2BUTNONE	0	0	0	0	0	0	0	0
1BUTANAL	0	0	0	0	0	0	0	0
NBUTANE	5808.4097	3421.8837	4616.2993	1.415E-16	0	0	0	0

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Stream →	(25)	(26)	(27)	(28)	(29)	(30)	(31)	(32)
P-IC4	0.0105927	0.0047738	0.006448	2.671E-23	0	0	0	0
2MEFURAN	0	0	0	0	0	0	0	0
3MEFURAN	0	0	0	0	0	0	0	0
CYPNTENO	0	0	0	0	0	0	0	0
FURFUALC	0	0	0	0	0	0	0	0
CYPNTANO	0	0	0	0	0	0	0	0
2PNTENAL	0	0	0	0	0	0	0	0
3PNTE2ON	0	0	0	0	0	0	0	0
2PENTENE	349.5621	514.67934	674.81063	1.332E-14	0	0	0	0
2PENTNON	0	0	0	0	0	0	0	0
3PENTNON	0	0	0	0	0	0	0	0
THFA	7.772E-06	0.0006362	0.0006439	1.8E-10	0	0	0	0
PENTANE	7915.34	11096.225	14570.317	2.418E-13	0	0	0	0
BENZENE	305.39156	1301.6759	1529.3954	1.17E-10	0	0	0	0
PHENOL	0	0	0	0	0	0	0	0
12BZDIOL	0	0	0	0	0	0	0	0
HMF	2.401E-10	5.159E-07	5.159E-07	0.0007191	0.0007007	0.0007007	1.847E-05	1.847E-05
CYHXDIEN	0	0	0	0	0	0	0	0
DIMEFURA	33.275019	216.83562	245.05777	1.431E-10	0	0	0	0
METCY5ON	0	0	0	0	0	0	0	0
2ETFURAN	0	0	0	0	0	0	0	0
HYMET5ON	1.941E-06	0.0003483	0.0003503	1.173E-08	1.173E-08	1.173E-08	3.272E-13	3.272E-13
CYHXENE	18.512075	91.449348	105.97218	3.328E-11	0	0	0	0
CYHEXNON	7.899E-07	2.798E-05	2.875E-05	3.094E-11	0	0	0	0
2MECY5ON	0	0	0	0	0	0	0	0
LEVOGLUC	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
1HEXANAL	0	0	0	0	0	0	0	0
2HEXNONE	0	0	0	0	0	0	0	0
3HEXNONE	0	0	0	0	0	0	0	0
CYHEXANE	3280.7617	15494.967	18029.812	5.44E-09	5.44E-09	5.44E-09	6.196E-16	6.196E-16
NHEXANE	2262.6907	7392.946	8981.7065	7.341E-11	0	0	0	0
SORBTOL	0	0	0	0	0	0	0	0
TOLUENE	27.550325	290.33864	315.30538	1.941E-08	1.941E-08	1.941E-08	1.306E-14	1.306E-14
OCRESOL	0	0	0	0	0	0	0	0
GUAIACOL	0	0	0	0	0	0	0	0
MECYHEXN	139.62179	1155.9538	1277.8745	4.079E-08	4.079E-08	4.079E-08	2.34E-14	2.34E-14
2M3HXENE	4180.9958	27942.976	31509.547	5.421E-08	5.421E-08	5.421E-08	1.072E-14	1.072E-14
METHHEXN	224.00899	1262.4141	1446.7868	9.727E-10	9.727E-10	9.727E-10	1.415E-16	1.415E-16
3ETPNTNO	0	0	0	0	0	0	0	0
D-INDOLE	7.23E-10	4.702E-07	4.702E-07	7.559E-05	7.463E-05	7.463E-05	9.568E-07	9.568E-07
OXYLENE	884.68687	25373.276	26233.025	0.0028181	0.0028181	0.0028181	2.258E-08	2.258E-08
ETYBENZN	381.99809	8727.5117	9095.2234	0.000223	0.000223	0.000223	1.119E-09	1.119E-09
23XYLNOL	0	0	0	0	0	0	0	0
SYRINGOL	0	0	0	0	0	0	0	0
VANALC	0	0	0	0	0	0	0	0
TRIMEPYR	1.066E-06	7.385E-05	7.491E-05	6.015E-09	6.015E-09	6.015E-09	3.859E-13	3.859E-13
MEETCY5E	146.70142	3787.57	3929.6125	0.0002076	0.0002076	0.0002076	1.331E-09	1.331E-09
DIMECYHX	0.0315283	0.5515333	0.5812763	2.702E-08	2.702E-08	2.702E-08	1.952E-13	1.952E-13
TRDICYHN	1497.2801	22169.998	23565.39	0.0002376	0.0002376	0.0002376	1.019E-09	1.019E-09
ETCYHEXN	422.10449	8024.3096	8425.1422	0.0004484	0.0004484	0.0004484	3.293E-09	3.293E-09
4MEHEPTN	508.04855	6354.4015	6825.732	1.873E-06	1.873E-06	1.873E-06	2.126E-12	2.126E-12
MEBZFURA	0	0	0	0	0	0	0	0
MEVINPHE	0	0	0	0	0	0	0	0
PROPBENZ	559.36884	26635.313	27186.268	0.2209657	0.2209584	0.2209584	7.305E-06	7.305E-06
ETMEBENZ	11.740638	628.64945	640.24077	0.0079927	0.0079924	0.0079924	2.933E-07	2.933E-07
TRIMEBEN	5.9060271	472.50056	478.36068	0.1368908	0.1368779	0.1368779	1.292E-05	1.292E-05

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Stream →	(25)	(26)	(27)	(28)	(29)	(30)	(31)	(32)
NAPTHALE	29.312396	6545.6692	6574.9125	20924.251	20887.089	20887.089	37.162452	37.162452
1NAPTHOL	0	0	0	0	0	0	0	0
CONIFALD	0	0	0	0	0	0	0	0
TETRALIN	78.392868	15009.437	15087.613	12860.923	12846.792	12846.792	14.131934	14.131934
IEUGENOL	0	0	0	0	0	0	0	0
D-DECALN	0.0003707	0.0382757	0.038644	0.0008472	0.0008469	0.0008469	3.383E-07	3.383E-07
DECALIN	84.167989	12119.832	12203.708	642.2067	641.88559	641.88559	0.3211036	0.3211036
D-C4CYC6	0.0015734	0.118497	0.1200549	0.0016942	0.0016934	0.0016934	7.134E-07	7.134E-07
DIMEOCTN	236.10424	9884.9288	10117.155	0.0370815	0.0370805	0.0370805	1E-06	1E-06
1MENAPHT	0.0166729	9.694478	9.711139	1675.7856	1646.917	1646.917	28.86858	28.86858
METDEC	3.4877189	1018.4699	1021.9519	14556.724	14495.114	14495.114	61.610141	61.610141
DIBZTHIO	0	0	0	0	0	0	0	0
D-C4BZ-S	9.943E-14	1.005E-09	1.005E-09	0.0006303	0.0001939	0.0001939	0.0004365	0.0004365
BIPHENYL	0.0085495	8.5130359	8.5215825	6181.395	5927.5414	5927.5414	253.85366	253.85366
DIMENAPH	7.551E-05	0.1041104	0.1041859	214.42793	196.79565	196.79565	17.632284	17.632284
D-C12H18	9.91E-06	0.0029021	0.002912	0.0208815	0.0208234	0.0208234	5.808E-05	5.808E-05
D-C12H1B	1.581E-06	0.000426	0.0004276	0.0025257	0.0025191	0.0025191	6.634E-06	6.634E-06
CNYLXNE	2.388E-09	7.265E-07	7.288E-07	9.756E-06	9.719E-06	9.719E-06	3.677E-08	3.677E-08
2ETDECLN	0.0793685	25.017382	25.096619	1268.7967	1252.4385	1252.4385	16.358178	16.358178
D-CQUINL	6.793E-17	2.585E-12	2.585E-12	7.617E-05	9.16E-08	9.16E-08	7.608E-05	7.608E-05
HXLCRESO	0	0	0	0	0	0	0	0
HEPCYHEX	0.0291892	20.011865	20.04104	4732.8659	4541.9472	4541.9472	190.91869	190.91869
D-2C2BZS	1.847E-18	2.41E-13	2.41E-13	0.0006304	2.091E-11	2.091E-11	0.0006304	0.0006304
D-C14H14	6.994E-11	3.495E-07	3.495E-07	0.0239608	0.0145615	0.0145615	0.0093992	0.0093992
ETPHCYHX	0.0059077	9.9230541	9.9289607	46705.567	38776.187	38776.187	7929.3802	7929.3802
D-HYPHEN	6.582E-10	1.83E-06	1.83E-06	0.0527428	0.0337097	0.0337097	0.0190331	0.0190331
THYDPHEN	0.0025534	4.4873601	4.4899131	43721.605	31511.419	31511.419	12210.186	12210.186
DIMDODEC	9.5704175	4316.4659	4326.0287	211460.84	208552.82	208552.82	2908.0252	2908.0252
D-3CZOLE	5.011E-19	5.146E-14	5.146E-14	7.617E-05	9.539E-12	9.539E-12	7.617E-05	7.617E-05
D-C15H16	3.961E-13	3.649E-09	3.649E-09	0.0031634	0.0005515	0.0005515	0.0026118	0.0026118
D-3C3BEN	5.253E-07	0.0003774	0.000378	0.0837618	0.0813945	0.0813945	0.0023673	0.0023673
OCHYENAP	1.164E-06	0.0109189	0.0109201	11951.177	706.45681	706.45681	11244.721	11244.721
D-DIMTRD	0.2642483	172.17449	172.4386	36302.867	34682.029	34682.029	1620.8381	1620.8381
DIMPHENA	2.374E-11	2.961E-06	2.961E-06	3812.6046	0.000746	0.000746	3812.6038	3812.6038
CETANE	0.0003671	1.3911936	1.3915606	43091.824	21755.315	21755.315	21336.509	21336.509
CYHXPHON	0	0	0	0	0	0	0	0
CYHXHXBZ	5.363E-09	0.0003185	0.0003185	77715.401	0.0164033	0.0164033	77715.385	77715.385
D-NC18	1.247E-11	1.372E-07	1.372E-07	0.1252712	0.0006264	0.0006264	0.1246448	0.1246448
D-C1C17	1.196E-11	1.419E-07	1.419E-07	0.0840636	0.0023776	0.0023776	0.0816859	0.0816859
D-NC20	1.52E-14	4.553E-10	4.553E-10	0.0098899	6.057E-09	6.057E-09	0.0098899	0.0098899
HMWLIGNB	0	0	0	0	0	0	0	0
TRIPHNPR	1.43E-15	6.276E-10	6.276E-10	183.20599	1.701E-12	1.701E-12	183.20599	183.20599
MOXHXPHE	0	0	0	0	0	0	0	0
DIBZFLOR	2.305E-14	3.137E-08	3.137E-08	14621.372	6.505E-10	6.505E-10	14621.372	14621.372
HXACMOXP	0	0	0	0	0	0	0	0
MOXHEPOX	0	0	0	0	0	0	0	0
HXDECBEN	4.867E-12	2.325E-06	2.325E-06	93011.622	1.669E-08	1.669E-08	93011.622	93011.622
P-3C3CY6	6.367E-05	0.0076165	0.0076798	0.1509951	0.147281	0.147281	0.0037141	0.0037141

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Table L-6. Heat Duties and Power for 100% CFP-B Oil HT (Case 11199011)

Tag	Description	Value	Unit
Duty			
Q1	Feed-effluent exchanger	719.2	MMBtu/h
Q2	Feed preheat furnace	72.6	MMBtu/h
Q3	Effluent cooler (makes steam)	524.6	MMBtu/h
Q4	Air cooler (cools to 140°F)	418.7	MMBtu/h
Q5	Water cooler (cools to 110°F)	105.7	MMBtu/h
Q6	Distillation cross exchange preheat	143.1	MMBtu/h
Q7	Distillation feed preheat furnace	57.3	MMBtu/h
Q8	Light products column condenser	164.3	MMBtu/h
Q9	Light products column reboiler	155.8	MMBtu/h
Q10	Heavier products column condenser	73.4	MMBtu/h
Q11	Heavier products column reboiler	50.8	MMBtu/h
Power			
W1	Makeup hydrogen compressor	31,718	kW
W2	Recycle compressor	13,180	kW
W3	Liquid recycle pump	-	kW

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Stream, Heat Duty, and Power Requirements Summary for 100% CFP-A (Case 12199011)

Table L-7. Stream Flows for 100% CFP-A Oil HT (Case 12199011)

Stream →	(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)
Temp, °F	93.0	135.5	135.5	140.0	644.0	683.5	449.5	276.7
Pres, psia	1100	1100	1100	995.00002	1045	1015	1010	1005
Total Flow	1260609	823078	2420819	133488	2083685	3681426	3681426	3681426
HYDROGEN	0	241113.92	709158.6	123555.44	241113.91	616859.56	616859.56	616859.56
AMMONIA	0	4.243E-63	1.248E-62	0	6.548E-62	4982.364	4982.364	4982.364
H2S	0	3.21E-61	9.441E-61	0	4.954E-60	9387.9559	9387.9559	9387.9559
WATER	12118.572	3791.8418	11152.476	0	15910.411	302708.87	302708.87	302708.87
CO	91.007124	0	0	0	91.007124	0	0	0
CO2	498.84563	0	0	0	498.84563	38967.419	38967.419	38967.419
METHANE	33.686691	409825.05	1205367.8	9932.0999	409858.81	1263651.3	1263651.3	1263651.3
FORMALDH	23.975112	0	0	0	23.975112	0	0	0
FORMACID	66992.251	0	0	0	66992.251	0	0	0
GLYOXAL	3119.6061	0	0	0	3119.6061	0	0	0
ETHYLENE	186.33587	1260.6047	3707.6609	0	1446.9609	3961.6703	3961.6703	3961.6703
ACALDEHY	5881.3442	0	0	0	5881.3442	0	0	0
HYDACTLD	1961.4693	0	0	0	1961.4693	0	0	0
ACETACID	1909.5095	0	0	0	1909.5095	0	0	0
ETHANE	18.996647	74367.291	218727.33	0	74386.116	235135.25	235135.25	235135.25
PROPYLEN	591.56099	36.417823	107.11124	0	627.97882	118.98772	118.98772	118.98772
ACETONE	15012.641	0	0	0	15012.641	0	0	0
ACETOL	536.89364	0	0	0	536.89364	0	0	0
PROPANE	96.910941	35271.292	103739.1	0	35368.383	116038.32	116038.32	116038.32
FURAN	4225.2659	0	0	0	4225.2659	0	0	0
2BUTENAL	382.75145	0	0	0	382.75145	0	0	0
2BTE3ONE	10979.951	0	0	0	10979.951	0	0	0
DIACESUL	19055.933	0	0	0	19055.933	0	0	0
BUTENE	523.75782	28.759694	84.587336	0	552.51753	103.18167	103.18167	103.18167
NBUTANE	0	34235.29	100692.03	0	34232.891	123768.47	123768.47	123768.47
P-IC4	0	0.0180837	0.0531872	0	0.017724	0.0646773	0.0646773	0.0646773
FURFURAL	461.22797	0	0	0	461.22797	0	0	0
2MEFURAN	7446.3995	0	0	0	7446.3995	0	0	0
CYPNTENO	5792.3432	0	0	0	5792.3432	0	0	0
13CPDION	1860.3622	0	0	0	1860.3622	0	0	0
CYPNTANO	215.12117	0	0	0	215.12117	0	0	0
PENTANE	0	7763.7793	22834.645	0	7763.809	36141.246	36141.246	36141.246
BENZENE	1252.0322	3089.0464	9085.4305	0	4341.0841	25943.337	25943.337	25943.337
PHENOL	330.12309	8.123E-66	2.389E-65	0	330.12309	0	0	0
12BZDIOL	37331.831	0	0	0	37331.831	0	0	0
5METFURF	330.70457	0	0	0	330.70457	0	0	0
METCY5ON	2017.0019	0	0	0	2017.0019	0	0	0
CYHXENE	7367.0144	1168.5471	3436.9033	0	8535.5646	10803.921	10803.921	10803.921
LEVOGLUC	39097.031	6.573E-75	1.933E-74	0	39097.031	0	0	0
CYHEXANE	0	4541.8096	13358.263	0	4541.8221	40647.893	40647.893	40647.893
METPIPR	12464.256	0	0	0	12464.256	0	0	0
NHEXANE	0	102.60813	301.78861	0	102.60491	742.65612	742.65612	742.65612
TOLUENE	1943.1485	67.460256	198.41252	0	2010.6089	1056.7589	1056.7589	1056.7589
OCRESOL	5047.2486	0	0	0	5047.2486	0	0	0
MCRESOL	5128.7226	0	0	0	5128.7226	0	0	0
PCRESOL	1507.5329	0	0	0	1507.5329	0	0	0
GUAIACOL	6372.1077	0	0	0	6372.1077	0	0	0
MECYHEXN	0	342.93534	1008.6333	0	342.936	4844.0469	4844.0469	4844.0469

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Stream →	(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)
2M3HXENE	0	1834.545	5395.7206	0	1834.5494	23111.19	23111.19	23111.19
METHHEXN	0	307.75632	905.16565	0	307.75709	3350.8328	3350.8328	3350.8328
3ETPNTNO	0	0	0	0	6.34E-65	0	0	0
BZOFURAN	1769.0156	0	0	0	1769.0156	0	0	0
D-INDOLE	0.0034594	4.17E-08	1.227E-07	0	0.0034594	6.779E-05	6.779E-05	6.779E-05
STYRENE	1082.4064	0	0	0	1082.4064	0	0	0
VANILLIN	108332.58	1.307E-05	3.843E-05	0	108332.58	0.2621471	0.2621471	0.2621471
OXYLENE	9924.3046	45.374509	133.45444	0	9969.679	2288.2396	2288.2396	2288.2396
PXYLENE	43040.483	0	0	0	43040.483	0	0	0
ETYBENZN	2131.1924	61.523359	180.95106	0	2192.7158	2428.3289	2428.3289	2428.3289
23XYLNOL	0	0	0	0	8.579E-64	0	0	0
24XYLNOL	0	0	0	0	0	0	0	0
25XYLNOL	4549.8574	0	0	0	4549.8574	0	0	0
CREOSOL	4842.7075	0	0	0	4842.7075	0	0	0
SYRINGOL	116398.8	0	0	0	116398.8	0	0	0
VANALC	0	0	0	0	3.358E-64	0	0	0
TRIMEPYR	20222.388	4.966E-07	1.461E-06	0	20222.388	6.783E-05	6.783E-05	6.783E-05
MEETCY5E	0	68.22323	200.65656	0	68.223364	3109.677	3109.677	3109.677
DIMECYHX	0	691.49939	2033.8217	0	691.50032	19269.851	19269.851	19269.851
TRDICYHN	0	896.73395	2637.4528	0	896.7359	21367.268	21367.268	21367.268
ETCYHEXN	0	323.19993	950.58803	0	323.20062	9936.0243	9936.0243	9936.0243
4MEHEPTN	0	962.80666	2831.7843	0	962.80877	21894.192	21894.192	21894.192
INDENE	5613.1332	2.6621031	7.8297151	0	5615.7953	470.90738	470.90738	470.90738
MEBZFURA	2926.917	0	0	0	2926.917	0	0	0
INDANE	3112.333	42.068207	123.73002	0	3154.4012	5760.7087	5760.7087	5760.7087
5INDANOL	860.15155	0	0	0	860.15155	0	0	0
MEVINPHE	3304.0664	0	0	0	3304.0664	0	0	0
ETMOXETN	1454.6429	0	0	0	1454.6429	0	0	0
SYRNGALD	3535.2661	0	0	0	3535.2661	0	0	0
PROPBENZ	0	22.207188	65.315259	0	22.207227	1961.376	1961.376	1961.376
135TMBZ	4997.9784	0	0	0	4997.9784	0	0	0
HXHYINDN	0	41.015625	120.63419	0	41.015698	3770.6993	3770.6993	3770.6993
PROPCYHX	0	102.73035	302.14809	0	102.73055	6475.2586	6475.2586	6475.2586
TRIMCYHX	0	109.90984	323.26422	0	109.91005	5572.7189	5572.7189	5572.7189
NAPTHALE	12500.319	1.6322129	4.8006261	0	12501.952	753.4804	753.4804	753.4804
1NAPTHOL	91180.394	0	0	0	91180.394	0	0	0
MEINDENE	5547.7915	18.28067	53.766677	0	5566.0722	5601.5582	5601.5582	5601.5582
CONIFALD	903.08018	0	0	0	903.08018	0	0	0
TETRALIN	0.1018784	68.321661	200.94606	0	68.42364	23435.926	23435.926	23435.926
EUGENOL	2335.3915	0	0	0	2335.3915	0	0	0
IEUGENOL	8284.3463	0	0	0	8284.3463	0	0	0
PRHYBNZO	485.94304	0	0	0	485.94304	0	0	0
D-DECALN	0	0.0002114	0.0006217	0	0.0002114	0.0411428	0.0411428	0.0411428
DECALIN	0	78.820054	231.82369	0	78.820181	26466.791	26466.791	26466.791
D-C4CYC6	0.1256491	0.0009518	0.0027995	0	0.1266009	0.1247402	0.1247402	0.1247402
DIMEOCTN	0	41.091046	120.85602	0	41.091123	3422.142	3422.142	3422.142
1MENAPHT	20001.811	3.835E-05	0.0001128	0	20001.811	0.0548604	0.0548604	0.0548604
1NAPMNOL	558.92025	0	0	0	558.92025	0	0	0
METDEC	0	151.76015	446.35338	0	151.76036	99866.716	99866.716	99866.716
DIBZTHIO	21038.164	0	0	0	21038.164	0	0	0
D-C4BZ-S	0.0286279	1.407E-08	4.139E-08	0	0.0286279	0.000561	0.000561	0.000561
BIPHENYL	0	11.500474	33.824924	0	11.500486	30611.094	30611.094	30611.094
DIMENAPH	5495.8287	2.6246105	7.7194428	0	5498.4533	10156.003	10156.003	10156.003
2ETYNAPH	1256.5073	0	0	0	1256.5073	0	0	0
D-C12H18	0	3.321E-05	9.767E-05	0	3.321E-05	0.0239703	0.0239703	0.0239703
D-C12H1B	0	4.619E-06	1.359E-05	0	4.619E-06	0.0029768	0.0029768	0.0029768

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Stream →	(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)
2ETDECLN	0	13.609234	40.02716	0	13.609252	8847.4918	8847.4918	8847.4918
FLUORENE	841.41371	0	0	0	841.41371	0	0	0
BIPHCALD	63645.638	0	0	0	63645.638	0	0	0
D-CQUINL	0.0034594	3.645E-10	1.072E-09	0	0.0034594	6.779E-05	6.779E-05	6.779E-05
HXLCRESO	13735.182	0	0	0	13735.182	0	0	0
PHFLOREN	0	0.288728	0.8492	0	0.2887283	903.48892	903.48892	903.48892
HEPCYHEX	0	55.483814	163.18769	0	55.483881	111629.2	111629.2	111629.2
PHENANTH	1781.9086	0.020469	0.0602029	0	1781.9291	1231.8938	1231.8938	1231.8938
MEFLOREN	2085.0667	0.0562564	0.1654599	0	2085.123	2085.2322	2085.2322	2085.2322
FLORMEOH	63653.411	0	0	0	63653.411	0	0	0
D-2C2BZS	0.0286279	8.566E-10	2.519E-09	0	0.0286279	0.000561	0.000561	0.000561
D-C14H14	0	1.322E-06	3.889E-06	0	1.322E-06	0.024039	0.024039	0.024039
ETPHCYHX	0	3.972057	11.682521	0	3.9720609	19572.897	19572.897	19572.897
D-HYPHEN	0.0543352	6.437E-06	1.893E-05	0	0.0543416	0.0527623	0.0527623	0.0527623
THYDPHEN	0	5.1300534	15.088392	0	5.1300585	24438.791	24438.791	24438.791
DIMDODEC	0	101.29281	297.92004	0	101.29295	131719.27	131719.27	131719.27
MEANTCEN	2085.1364	0.0117149	0.0344555	0	2085.1481	2085.1709	2085.1709	2085.1709
MEPHENAN	1472.1301	0.0067907	0.0199726	0	1472.1369	1017.7116	1017.7116	1017.7116
D-3CZOLE	0.0034594	1.36E-10	4E-10	0	0.0034594	6.779E-05	6.779E-05	6.779E-05
D-C15H16	0	9.429E-08	2.773E-07	0	9.429E-08	0.0031734	0.0031734	0.0031734
D-3C3BEN	0.1400828	3.997E-05	0.0001176	0	0.1401228	0.0830109	0.0830109	0.0830109
OCHYENAP	0	3.7593837	11.057011	0	3.7593823	132862.19	132862.19	132862.19
D-DIMTRD	0	0.1201876	0.3534928	0	0.1201875	234.95259	234.95259	234.95259
DIMPHENA	2085.1387	0.0016641	0.0048943	0	2085.1403	1149.0632	1149.0632	1149.0632
CETANE	0	1.647943	4.8468913	0	1.6479445	26017.688	26017.688	26017.688
CHRYSENE	2085.1434	0.0001116	0.0003281	0	2085.1435	2085.1437	2085.1437	2085.1437
DIMEPYRE	2085.1407	7.056E-05	0.0002075	0	2085.1408	2085.1409	2085.1409	2085.1409
RETENE	582.32898	0	0	0	582.32898	0	0	0
CYHXHXBZ	0	0.0864324	0.2542129	0	0.0864324	26382.994	26382.994	26382.994
D-NC18	0.129046	2.292E-06	6.741E-06	0	0.1290483	0.1252726	0.1252726	0.1252726
D-C1C17	0.0865967	1.291E-06	3.797E-06	0	0.086598	0.0840638	0.0840638	0.0840638
HMWLIGNA	47577.13	0	0	0	47577.13	0	0	0
D-NC20	0.0101878	5.601E-08	1.647E-07	0	0.0101879	0.0098896	0.0098896	0.0098896
HMWLIGNB	47577.124	0	0	0	47577.124	0	0	0
DBZFLORE	2085.1341	2.103E-06	6.186E-06	0	2085.1341	481.58487	481.58487	481.58487
TRIPHNPR	2085.145	0.0003411	0.0010033	0	2085.1453	829.59128	829.59128	829.59128
MOXHXPHE	63653.761	0	0	0	63653.761	0	0	0
DIBZFLOR	0	0.0004922	0.0014476	0	0.0004922	6907.0202	6907.0202	6907.0202
BZACMOXP	69314.009	0	0	0	69314.009	0	0	0
HXACMOXP	69313.988	0	0	0	69313.988	0	0	0
MOXHEPOX	14971.825	0	0	0	14971.825	0	0	0
HXDECBEN	0	0.002729	0.0080264	0	0.002729	11913.685	11913.685	11913.685
P-3C3CY6	0.1086578	0.0008915	0.0026222	0	0.1095494	0.1627271	0.1627271	0.1627271

Table L-7 (continued). Stream Flows for 100% CFP-A Oil HT (Case 12199011)

Stream →	(9)	(10)	(11)	(12)	(13)	(14)	(15)	(16)
Temp, °F	110.0	110.0	109.4	110.0	110.0	110.0	110.0	110.0
Pres, psia	1050	995	993	993	993	20	993	973
Total Flow	422363	4103789	723582	2445907	2406648	39259	119311	25964
HYDROGEN	0	616859.56	1.9674115	616149.25	616149.25	0	30546.064	25964.154
AMMONIA	0	4982.364	4691.1187	291.2453	0	291.2453	0	0
H2S	0	9387.9559	9387.9559	0	0	0	0	0
WATER	422363.15	725072.03	709487.36	11734.216	11734.216	0	581.7326	0
CO	0	0	0	0	0	0	0	0

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Stream →	(9)	(10)	(11)	(12)	(13)	(14)	(15)	(16)
CO2	0	38967.419	0	38967.419	0	38967.419	0	0
METHANE	0	1263651.3	10.167926	1257791.4	1257791.4	0	62355.957	0
FORMALDH	0	0	0	0	0	0	0	0
FORMACID	0	0	0	0	0	0	0	0
GLYOXAL	0	0	0	0	0	0	0	0
ETHYLENE	0	3961.6703	0.0541133	3901.0008	3901.0008	0	193.39506	0
ACALDEHY	0	0	0	0	0	0	0	0
HYDACTLD	0	0	0	0	0	0	0	0
ACETACID	0	0	0	0	0	0	0	0
ETHANE	0	235135.25	2.1281927	230136.86	230136.86	0	11409.209	0
PROPYLEN	0	118.98772	0.0005109	112.69833	112.69833	0	5.587105	0
ACETONE	0	0	0	0	0	0	0	0
ACETOL	0	0	0	0	0	0	0	0
PROPANE	0	116038.32	0.1353833	109149.85	109149.85	0	5411.1862	0
FURAN	0	0	0	0	0	0	0	0
2BUTENAL	0	0	0	0	0	0	0	0
2BTE3ONE	0	0	0	0	0	0	0	0
DIACESUL	0	0	0	0	0	0	0	0
BUTENE	0	103.18167	4.26E-05	88.999523	88.999523	0	4.4122185	0
NBUTANE	0	123768.47	0.0138782	105950.58	105950.58	0	5252.58	0
P-IC4	0	0.0646773	3.069E-09	0.0569407	0.0569407	0	0.0028229	0
FURFURAL	0	0	0	0	0	0	0	0
2MEFURAN	0	0	0	0	0	0	0	0
CYPNTENO	0	0	0	0	0	0	0	0
13CPDION	0	0	0	0	0	0	0	0
CYPNTANO	0	0	0	0	0	0	0	0
PENTANE	0	36141.246	0.0002007	24025.662	24025.662	0	1191.0903	0
BENZENE	0	25943.337	0.7558078	9559.3273	9559.3273	0	473.91086	0
PHENOL	0	0	0	0	0	0	0	0
12BZDIOL	0	0	0	0	0	0	0	0
5METFURF	0	0	0	0	0	0	0	0
METCY5ON	0	0	0	0	0	0	0	0
CYHXENE	0	10803.921	0.0209928	3616.1701	3616.1701	0	179.27436	0
LEVOGLUC	0	0	0	0	0	0	0	0
CYHEXANE	0	40647.893	0.0153381	14055.02	14055.02	0	696.78823	0
METPIPR	0	0	0	0	0	0	0	0
NHEXANE	0	742.65612	1.077E-07	317.53917	317.53917	0	15.742244	0
TOLUENE	0	1056.7589	0.0005692	208.76173	208.76173	0	10.34952	0
OCRESOL	0	0	0	0	0	0	0	0
MCRESOL	0	0	0	0	0	0	0	0
PCRESOL	0	0	0	0	0	0	0	0
GUAIACOL	0	0	0	0	0	0	0	0
MECYHEXN	0	4844.0469	1.583E-05	1061.2436	1061.2436	0	52.611953	0
2M3HXENE	0	23111.19	4.348E-07	5677.1589	5677.1589	0	281.44944	0
METHHEXN	0	3350.8328	8.157E-09	952.37859	952.37859	0	47.214887	0
3ETPNTNO	0	0	0	0	0	0	0	0
BZOFURAN	0	0	0	0	0	0	0	0
D-INDOLE	0	6.779E-05	1.547E-10	1.291E-07	1.291E-07	0	6.398E-09	0
STYRENE	0	0	0	0	0	0	0	0
VANILLIN	0	0.2621471	1.341E-05	4.044E-05	4.044E-05	0	2.005E-06	0
OXYLENE	0	2288.2396	9.85E-05	140.41602	140.41602	0	6.9612303	0
PXYLENE	0	0	0	0	0	0	0	0
ETYBENZN	0	2428.3289	2.922E-05	190.38946	190.38946	0	9.438701	0
23XYLNOL	0	0	0	0	0	0	0	0
24XYLNOL	0	0	0	0	0	0	0	0
25XYLNOL	0	0	0	0	0	0	0	0

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Stream →	(9)	(10)	(11)	(12)	(13)	(14)	(15)	(16)
CREOSOL	0	0	0	0	0	0	0	0
SYRINGOL	0	0	0	0	0	0	0	0
VANALC	0	0	0	0	0	0	0	0
TRIMEPYR	0	6.783E-05	5.208E-14	1.537E-06	1.537E-06	0	7.619E-08	0
MEETCY5E	0	3109.677	7.993E-06	211.12279	211.12279	0	10.466571	0
DIMECYHX	0	19269.851	4.149E-08	2139.9069	2139.9069	0	106.0875	0
TRDICYHN	0	21367.268	6.512E-08	2775.0216	2775.0216	0	137.57379	0
ETCYHEXN	0	9936.0243	9.929E-08	1000.1705	1000.1705	0	49.584207	0
4MEHEPTN	0	21894.192	1.09E-09	2979.4893	2979.4893	0	147.71043	0
INDENE	0	470.90738	2.218E-05	8.2381148	8.2381148	0	0.4084108	0
MEBZFURA	0	0	0	0	0	0	0	0
INDANE	0	5760.7087	0.0006377	130.18379	130.18379	0	6.4539599	0
5INDANOL	0	0	0	0	0	0	0	0
MEVINPHE	0	0	0	0	0	0	0	0
ETMOXETN	0	0	0	0	0	0	0	0
SYRNGALD	0	0	0	0	0	0	0	0
PROPBENZ	0	1961.376	2.749E-07	68.722108	68.722108	0	3.4069504	0
135TMBZ	0	0	0	0	0	0	0	0
HXHYINDN	0	3770.6993	1.443E-06	126.92648	126.92648	0	6.2924758	0
PROPCYHX	0	6475.2586	8.894E-10	317.90811	317.90811	0	15.760535	0
TRIMCYHX	0	5572.7189	1.427E-08	340.12566	340.12566	0	16.861986	0
NAPTHALE	0	753.4804	9.415E-05	5.051029	5.051029	0	0.2504086	0
1NAPTHOL	0	0	0	0	0	0	0	0
MEINDENE	0	5601.5582	5.227E-06	56.571162	56.571162	0	2.8045581	0
CONIFALD	0	0	0	0	0	0	0	0
TETRALIN	0	23435.926	0.0001422	211.42748	211.42748	0	10.481677	0
EUGENOL	0	0	0	0	0	0	0	0
IEUGENOL	0	0	0	0	0	0	0	0
PRHYBNZO	0	0	0	0	0	0	0	0
D-DECALN	0	0.0411428	1.168E-12	0.0006542	0.0006542	0	3.243E-05	0
DECALIN	0	26466.791	2.813E-14	243.91566	243.91566	0	12.092303	0
D-C4CYC6	0	0.1247402	8.25E-17	0.0029455	0.0029455	0	0.000146	0
DIMEOCTN	0	3422.142	1.506E-14	127.15987	127.15987	0	6.3040463	0
1MENAPHT	0	0.0548604	1.086E-10	0.0001187	0.0001187	0	5.883E-06	0
1NAPMNOL	0	0	0	0	0	0	0	0
METDEC	0	99866.716	5.318E-09	469.63536	469.63536	0	23.282527	0
DIBZTHIO	0	0	0	0	0	0	0	0
D-C4BZ-S	0	0.000561	1.318E-16	4.355E-08	4.355E-08	0	2.159E-09	0
BIPHENYL	0	30611.094	1.059E-05	35.589255	35.589255	0	1.7643642	0
DIMENAPH	0	10156.003	2.138E-07	8.1220947	8.1220947	0	0.402659	0
2ETYNAPH	0	0	0	0	0	0	0	0
D-C12H18	0	0.0239703	5.747E-18	0.0001028	0.0001028	0	5.095E-06	0
D-C12H1B	0	0.0029768	8.313E-19	1.43E-05	1.43E-05	0	7.087E-07	0
2ETDECLN	0	8847.4918	1.53E-12	42.114996	42.114996	0	2.0878827	0
FLUORENE	0	0	0	0	0	0	0	0
BIPHCALD	0	0	0	0	0	0	0	0
D-CQUINL	0	6.779E-05	7.901E-19	1.128E-09	1.128E-09	0	5.592E-11	0
HXLCRESO	0	0	0	0	0	0	0	0
PHFLOREN	0	903.48892	5.235E-15	0.8934948	0.8934948	0	0.0442957	0
HEPCYHEX	0	111629.2	2.522E-24	171.69965	171.69965	0	8.5121395	0
PHENANTH	0	1231.8938	1.295E-08	0.0633431	0.0633431	0	0.0031403	0
MEFLOREN	0	2085.2322	2.772E-06	0.1740905	0.1740905	0	0.0086307	0
FLORMEOH	0	0	0	0	0	0	0	0
D-2C2BZS	0	0.000561	1.331E-16	2.651E-09	2.651E-09	0	1.314E-10	0
D-C14H14	0	0.024039	4.811E-15	4.092E-06	4.092E-06	0	2.029E-07	0
ETPHCYHX	0	19572.897	6.254E-14	12.29189	12.29189	0	0.6093797	0

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Stream →	(9)	(10)	(11)	(12)	(13)	(14)	(15)	(16)
D-HYPHEN	0	0.0527623	2.99E-17	1.992E-05	1.992E-05	0	9.876E-07	0
THYDPHEN	0	24438.791	2.878E-15	15.875414	15.875414	0	0.7870356	0
DIMDODEC	0	131719.27	4.368E-25	313.45969	313.45969	0	15.54	0
MEANTCEN	0	2085.1709	2.371E-07	0.0362527	0.0362527	0	0.0017973	0
MEPHENAN	0	1017.7116	5.14E-13	0.0210144	0.0210144	0	0.0010418	0
D-3CZOLE	0	6.779E-05	1.842E-18	4.209E-10	4.209E-10	0	2.087E-11	0
D-C15H16	0	0.0031734	1.53E-20	2.918E-07	2.918E-07	0	1.447E-08	0
D-3C3BEN	0	0.0830109	1.051E-23	0.0001237	0.0001237	0	6.133E-06	0
OCHYENAP	0	132862.19	1.86E-20	11.633768	11.633768	0	0.5767528	0
D-DIMTRD	0	234.95259	1.624E-31	0.3719318	0.3719318	0	0.0184388	0
DIMPHENA	0	1149.0632	1.967E-11	0.0051496	0.0051496	0	0.0002553	0
CETANE	0	26017.688	2.298E-32	5.0997092	5.0997092	0	0.2528219	0
CHRYSENE	0	2085.1437	6.159E-16	0.0003452	0.0003452	0	1.711E-05	0
DIMEPYRE	0	2085.1409	8.752E-14	0.0002184	0.0002184	0	1.083E-05	0
RETENE	0	0	0	0	0	0	0	0
CYHXHXBZ	0	26382.994	7.982E-26	0.267473	0.267473	0	0.0132602	0
D-NC18	0	0.1252726	3.349E-45	7.092E-06	7.092E-06	0	3.516E-07	0
D-C1C17	0	0.0840638	8.044E-43	3.995E-06	3.995E-06	0	1.981E-07	0
HMWLIGNA	0	0	0	0	0	0	0	0
D-NC20	0	0.0098896	5.318E-52	1.733E-07	1.733E-07	0	8.594E-09	0
HMWLIGNB	0	0	0	0	0	0	0	0
DBZFLORE	0	481.58487	3.532E-16	6.508E-06	6.508E-06	0	3.226E-07	0
TRIPHNPR	0	829.59128	2.295E-25	0.0010556	0.0010556	0	5.233E-05	0
MOXHXPHE	0	0	0	0	0	0	0	0
DIBZFLOR	0	6907.0202	3.495E-21	0.0015231	0.0015231	0	7.551E-05	0
BZACMOXP	0	0	0	0	0	0	0	0
HXACMOXP	0	0	0	0	0	0	0	0
MOXHEPOX	0	0	0	0	0	0	0	0
HXDECBEN	0	11913.685	8.766E-39	0.0084451	0.0084451	0	0.0004187	0
P-3C3CY6	0	0.1627271	1.036E-33	0.002759	0.002759	0	0.0001368	0

Table L-7 (continued). Stream Flows for 100% CFP-A Oil HT (Case 12199011)

Stream →	(17)	(18)	(19)	(20)	(21)	(22)	(23)	(24)
Temp, °F	110.0	110.0	110.0	110.0	110.0	110.0	110.0	450.0
Pres, psia	22	993	993	53	53	1050	53	45.5
Total Flow	93347	2287331	934300	15972	918328	0	918328	918328
HYDROGEN	4581.9096	585603.16	708.34456	691.40497	16.939593	0	16.939593	16.939593
AMMONIA	0	1.248E-62	0	0	0	0	0	0
H2S	0	9.441E-61	0	0	0	0	0	0
WATER	581.7326	11152.476	3850.4487	348.36467	3502.084	1.802E-05	3502.084	3502.084
CO	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0
METHANE	62355.957	1195435.7	5849.7444	5264.2915	585.45294	0	585.45294	585.45294
FORMALDH	0	0	0	0	0	0	0	0
FORMACID	0	0	0	0	0	0	0	0
GLYOXAL	0	0	0	0	0	0	0	0
ETHYLENE	193.39506	3707.6609	60.615408	42.98896	17.626449	0	17.626449	17.626449
ACALDEHY	0	0	0	0	0	0	0	0
HYDACTLD	0	0	0	0	0	0	0	0
ACETACID	0	0	0	0	0	0	0	0
ETHANE	11409.209	218727.33	4996.2612	3126.1241	1870.137	0	1870.137	1870.137
PROPYLEN	5.587105	107.11124	6.2888717	2.3410154	3.9478563	0	3.9478563	3.9478563
ACETONE	0	0	0	0	0	0	0	0

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Stream →	(17)	(18)	(19)	(20)	(21)	(22)	(23)	(24)
ACETOL	0	0	0	0	0	0	0	0
PROPANE	5411.1862	103739.1	6888.3299	2339.719	4548.611	0	4548.611	4548.611
FURAN	0	0	0	0	0	0	0	0
2BUTENAL	0	0	0	0	0	0	0	0
2BTE3ONE	0	0	0	0	0	0	0	0
DIACESUL	0	0	0	0	0	0	0	0
BUTENE	4.4122185	84.587336	14.182106	2.264218	11.917888	0	11.917888	11.917888
NBUTANE	5252.58	100692.03	17817.872	2457.5049	15360.367	0	15360.367	15360.367
P-IC4	0.0028229	0.0531872	0.0077366	0.0014111	0.0063255	0	0.0063255	0.0063255
FURFURAL	0	0	0	0	0	0	0	0
2MEFURAN	0	0	0	0	0	0	0	0
CYPNTENO	0	0	0	0	0	0	0	0
13CPDION	0	0	0	0	0	0	0	0
CYPNTANO	0	0	0	0	0	0	0	0
PENTANE	1191.0903	22834.645	12115.584	608.78565	11506.798	0	11506.798	11506.798
BENZENE	473.91086	9085.4305	16383.254	242.67836	16140.576	0	16140.576	16140.576
PHENOL	0	2.389E-65	0	0	0	0	0	0
12BZDIOL	0	0	0	0	0	0	0	0
5METFURF	0	0	0	0	0	0	0	0
METCY5ON	0	0	0	0	0	0	0	0
CYHXENE	179.27436	3436.9033	7187.7296	88.747334	7098.9823	0	7098.9823	7098.9823
LEVOGLUC	0	1.933E-74	0	0	0	0	0	0
CYHEXANE	696.78823	13358.263	26592.858	344.21368	26248.644	0	26248.644	26248.644
METPIPR	0	0	0	0	0	0	0	0
NHEXANE	15.742244	301.78861	425.11695	7.4432408	417.67371	0	417.67371	417.67371
TOLUENE	10.34952	198.41252	847.99658	4.0331633	843.96341	0	843.96341	843.96341
OCRESOL	0	0	0	0	0	0	0	0
MCRESOL	0	0	0	0	0	0	0	0
PCRESOL	0	0	0	0	0	0	0	0
GUAIACOL	0	0	0	0	0	0	0	0
MECYHEXN	52.611953	1008.6333	3782.8033	23.859475	3758.9438	0	3758.9438	3758.9438
2M3HXENE	281.44944	5395.7206	17434.031	124.26182	17309.769	0	17309.769	17309.769
METHHEXN	47.214887	905.16565	2398.4542	20.83117	2377.623	0	2377.623	2377.623
3ETPNTNO	0	0	0	0	0	0	0	0
BZOFURAN	0	0	0	0	0	0	0	0
D-INDOLE	6.398E-09	1.227E-07	6.766E-05	2.07E-09	6.766E-05	0	6.766E-05	6.766E-05
STYRENE	0	0	0	0	0	0	0	0
VANILLIN	2.005E-06	3.843E-05	0.2620933	5.452E-07	0.2620927	0	0.2620927	0.2620927
OXYLENE	6.9612303	133.45444	2147.8234	2.8814471	2144.942	0	2144.942	2144.942
PXYLENE	0	0	0	0	0	0	0	0
ETYBENZN	9.438701	180.95106	2237.9394	3.9499114	2233.9895	0	2233.9895	2233.9895
23XYLNOL	0	0	0	0	0	0	0	0
24XYLNOL	0	0	0	0	0	0	0	0
25XYLNOL	0	0	0	0	0	0	0	0
CREOSOL	0	0	0	0	0	0	0	0
SYRINGOL	0	0	0	0	0	0	0	0
VANALC	0	0	0	0	0	0	0	0
TRIMEPYR	7.619E-08	1.461E-06	6.629E-05	2.825E-08	6.627E-05	0	6.627E-05	6.627E-05
MEETCY5E	10.466571	200.65656	2898.5542	4.2792922	2894.2749	0	2894.2749	2894.2749
DIMECYHX	106.0875	2033.8217	17129.944	42.919111	17087.025	0	17087.025	17087.025
TRDICYHN	137.57379	2637.4528	18592.247	56.559809	18535.687	0	18535.687	18535.687
ETCYHEXN	49.584207	950.58803	8935.8538	20.117341	8915.7365	0	8915.7365	8915.7365
4MEHEPTN	147.71043	2831.7843	18914.703	59.166047	18855.537	0	18855.537	18855.537
INDENE	0.4084108	7.8297151	462.66925	0.1529123	462.51633	0	462.51633	462.51633
MEBZFURA	0	0	0	0	0	0	0	0
INDANE	6.4539599	123.73002	5630.5242	2.4816207	5628.0426	0	5628.0426	5628.0426

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Stream →	(17)	(18)	(19)	(20)	(21)	(22)	(23)	(24)
5INDANOL	0	0	0	0	0	0	0	0
MEVINPHE	0	0	0	0	0	0	0	0
ETMOXETN	0	0	0	0	0	0	0	0
SYRNGALD	0	0	0	0	0	0	0	0
PROPBENZ	3.4069504	65.315259	1892.6539	1.2919707	1891.3619	0	1891.3619	1891.3619
135TMBZ	0	0	0	0	0	0	0	0
HXHYINDN	6.2924758	120.63419	3643.7729	2.3961228	3641.3767	0	3641.3767	3641.3767
PROPCYHX	15.760535	302.14809	6157.3505	5.8376862	6151.5128	0	6151.5128	6151.5128
TRIMCYHX	16.861986	323.26422	5232.5932	6.4959611	5226.0972	0	5226.0972	5226.0972
NAPTHALE	0.2504086	4.8006261	748.42928	0.0875438	748.34173	0	748.34173	748.34173
1NAPTHOL	0	0	0	0	0	0	0	0
MEINDENE	2.8045581	53.766677	5544.987	0.9768653	5544.0101	0	5544.0101	5544.0101
CONIFALD	0	0	0	0	0	0	0	0
TETRALIN	10.481677	200.94606	23224.498	3.0991275	23221.399	0	23221.399	23221.399
EUGENOL	0	0	0	0	0	0	0	0
IEUGENOL	0	0	0	0	0	0	0	0
PRHYBNZO	0	0	0	0	0	0	0	0
D-DECALN	3.243E-05	0.0006217	0.0404886	1.141E-05	0.0404772	0	0.0404772	0.0404772
DECALIN	12.092303	231.82369	26222.875	3.5947029	26219.281	0	26219.281	26219.281
D-C4CYC6	0.000146	0.0027995	0.1217947	4.903E-05	0.1217457	0	0.1217457	0.1217457
DIMEOCTN	6.3040463	120.85602	3294.9821	2.115862	3292.8662	0	3292.8662	3292.8662
1MENAPHT	5.883E-06	0.0001128	0.0547417	1.895E-06	0.0547398	0	0.0547398	0.0547398
1NAPMNOL	0	0	0	0	0	0	0	0
METDEC	23.282527	446.35338	99397.081	7.1075795	99389.973	0	99389.973	99389.973
DIBZTHIO	0	0	0	0	0	0	0	0
D-C4BZ-S	2.159E-09	4.139E-08	0.0005609	4.833E-10	0.0005609	0	0.0005609	0.0005609
BIPHENYL	1.7643642	33.824924	30575.505	0.5149077	30574.99	0	30574.99	30574.99
DIMENAPH	0.402659	7.7194428	10147.881	0.1117983	10147.769	0	10147.769	10147.769
2ETYNAPH	0	0	0	0	0	0	0	0
D-C12H18	5.095E-06	9.767E-05	0.0238675	1.484E-06	0.023866	0	0.023866	0.023866
D-C12H1B	7.087E-07	1.359E-05	0.0029625	2.085E-07	0.0029623	0	0.0029623	0.0029623
2ETDECLN	2.0878827	40.02716	8805.3768	0.6116941	8804.7651	0	8804.7651	8804.7651
FLUORENE	0	0	0	0	0	0	0	0
BIPHCALD	0	0	0	0	0	0	0	0
D-CQUINL	0	1.072E-09	6.779E-05	1.085E-11	6.779E-05	0	6.779E-05	6.779E-05
HXLCRESO	0	0	0	0	0	0	0	0
PHFLOREN	0.0442957	0.8492	902.59542	0.0111915	902.58423	0	902.58423	902.58423
HEPCYHEX	8.5121395	163.18769	111457.5	1.8355982	111455.66	0	111455.66	111455.66
PHENANTH	0.0031403	0.0602029	1231.8304	0.0006999	1231.8297	0	1231.8297	1231.8297
MEFLOREN	0.0086307	0.1654599	2085.0581	0.0020703	2085.056	0	2085.056	2085.056
FLORMEOH	0	0	0	0	0	0	0	0
D-2C2BZS	0	2.519E-09	0.000561	2.364E-11	0.000561	0	0.000561	0.000561
D-C14H14	2.029E-07	3.889E-06	0.0240349	4.823E-08	0.0240349	0	0.0240349	0.0240349
ETPHCYHX	0.6093797	11.682521	19560.605	0.1479407	19560.457	0	19560.457	19560.457
D-HYPHEN	9.876E-07	1.893E-05	0.0527424	2.393E-07	0.0527422	0	0.0527422	0.0527422
THYDPHEN	0.7870356	15.088392	24422.915	0.1869878	24422.728	0	24422.728	24422.728
DIMDODEC	15.54	297.92004	131405.81	3.4191211	131402.39	0	131402.39	131402.39
MEANTCEN	0.0017973	0.0344555	2085.1346	0.0003745	2085.1342	0	2085.1342	2085.1342
MEPHENAN	0.0010418	0.0199726	1017.6906	0.000201	1017.6904	0	1017.6904	1017.6904
D-3CZOLE	0	4E-10	6.779E-05	3.774E-12	6.779E-05	0	6.779E-05	6.779E-05
D-C15H16	1.447E-08	2.773E-07	0.0031731	3.031E-09	0.0031731	0	0.0031731	0.0031731
D-3C3BEN	6.133E-06	0.0001176	0.0828872	1.479E-06	0.0828857	0	0.0828857	0.0828857
OCHYENAP	0.5767528	11.057011	132850.56	0.1059812	132850.45	0	132850.45	132850.45
D-DIMTRD	0.0184388	0.3534928	234.58066	0.0036923	234.57696	0	234.57696	234.57696
DIMPHENA	0.0002553	0.0048943	1149.0581	4.485E-05	1149.058	0	1149.058	1149.058
CETANE	0.2528219	4.8468913	26012.588	0.0409401	26012.547	0	26012.547	26012.547

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Stream →	(17)	(18)	(19)	(20)	(21)	(22)	(23)	(24)
CHRYSENE	1.711E-05	0.0003281	2085.1434	2.144E-06	2085.1434	0	2085.1434	2085.1434
DIMEPYRE	1.083E-05	0.0002075	2085.1407	1.357E-06	2085.1407	0	2085.1407	2085.1407
RETENE	0	0	0	0	0	0	0	0
CYHXHXBZ	0.0132602	0.2542129	26382.727	0.0019025	26382.725	0	26382.725	26382.725
D-NC18	3.516E-07	6.741E-06	0.1252656	4.55E-08	0.1252655	0	0.1252655	0.1252655
D-C1C17	1.981E-07	3.797E-06	0.0840598	2.64E-08	0.0840598	0	0.0840598	0.0840598
HMWLIGNA	0	0	0	0	0	0	0	0
D-NC20	8.594E-09	1.647E-07	0.0098894	8.857E-10	0.0098894	0	0.0098894	0.0098894
HMWLIGNB	0	0	0	0	0	0	0	0
DBZFLORE	3.226E-07	6.186E-06	481.58487	3.345E-08	481.58487	0	481.58487	481.58487
TRIPHNPR	5.233E-05	0.0010033	829.59023	6.599E-06	829.59022	0	829.59022	829.59022
MOXHXPHE	0	0	0	0	0	0	0	0
DIBZFLOR	7.551E-05	0.0014476	6907.0186	8.773E-06	6907.0186	0	6907.0186	6907.0186
BZACMOXP	0	0	0	0	0	0	0	0
HXACMOXP	0	0	0	0	0	0	0	0
MOXHEPOX	0	0	0	0	0	0	0	0
HXDECBEN	0.0004187	0.0080264	11913.676	3.941E-05	11913.676	0	11913.676	11913.676
P-3C3CY6	0.0001368	0.0026222	0.1599681	3.355E-05	0.1599346	0	0.1599346	0.1599346

Table L-7 (continued). Stream Flows for 100% CFP-A Oil HT (Case 12199011)

Stream →	(25)	(26)	(27)	(28)	(29)	(30)	(31)	(32)
Temp, °F	200.0	200.0	60.0	565.2	444.7	60.0	598.9	60.0
Pres, psia	35	35	15	45	15	15	20	15
Total Flow	35965	227382	246059	654981	407889	407889	247092	247092
HYDROGEN	16.790904	0.1486898	0.1692992	5.745E-19	0	0	0	0
AMMONIA	0	0	0	0	0	0	0	0
H2S	0	0	0	0	0	0	0	0
WATER	1664.8099	1837.2741	2577.1948	3.143E-16	0	0	0	0
CO	0	0	0	0	0	0	0	0
CO_2	0	0	0	0	0	0	0	0
METHANE	570.1858	15.26714	18.405647	1.19E-23	0	0	0	0
FORMALDH	0	0	0	0	0	0	0	0
FORMACID	0	0	0	0	0	0	0	0
GLYOXAL	0	0	0	0	0	0	0	0
ETHYLENE	16.346255	1.2801939	1.613926	0	0	0	0	0
ACALDEHY	0	0	0	0	0	0	0	0
HYDACTLD	0	0	0	0	0	0	0	0
ACETACID	0	0	0	0	0	0	0	0
ETHANE	1690.5422	179.59488	231.01839	0	0	0	0	0
PROPYLEN	3.1602128	0.7876435	1.0610009	0	0	0	0	0
ACETONE	0	0	0	0	0	0	0	0
ACETOL	0	0	0	0	0	0	0	0
PROPANE	3571.3929	977.21807	1309.8371	1.671E-24	0	0	0	0
FURAN	0	0	0	0	0	0	0	0
2BUTENAL	0	0	0	0	0	0	0	0
2BTE3ONE	0	0	0	0	0	0	0	0
DIACESUL	0	0	0	0	0	0	0	0
BUTENE	7.3916777	4.5262105	6.2124387	1.096E-19	0	0	0	0
NBUTANE	9065.6238	6294.7433	8597.5266	3.518E-16	0	0	0	0
P-IC4	0.0041296	0.0021959	0.0029992	0	0	0	0	0
FURFURAL	0	0	0	0	0	0	0	0
2MEFURAN	0	0	0	0	0	0	0	0
CYPNTENO	0	0	0	0	0	0	0	0
13CPDION	0	0	0	0	0	0	0	0

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Stream →	(25)	(26)	(27)	(28)	(29)	(30)	(31)	(32)
CYPNTANO	0	0	0	0	0	0	0	0
PENTANE	4376.2232	7130.5749	9294.6664	2.146E-13	0	0	0	0
BENZENE	2702.8309	13437.745	15634.302	1.827E-09	1.827E-09	1.827E-09	8.47E-17	8.47E-17
PHENOL	0	0	0	0	0	0	0	0
12BZDIOL	0	0	0	0	0	0	0	0
5METFURF	0	0	0	0	0	0	0	0
METCY5ON	0	0	0	0	0	0	0	0
CYHXENE	1053.8592	6045.1231	6927.8015	3.282E-09	3.282E-09	3.282E-09	2.541E-16	2.541E-16
LEVOGLUC	0	0	0	0	0	0	0	0
CYHEXANE	4041.7683	22206.876	25544.933	1.153E-08	1.153E-08	1.153E-08	8.788E-16	8.788E-16
METPIPR	0	0	0	0	0	0	0	0
NHEXANE	86.515814	331.15789	396.18641	4.506E-12	0	0	0	0
TOLUENE	63.891638	780.07177	839.67268	7.615E-08	7.615E-08	7.615E-08	3.894E-14	3.894E-14
OCRESOL	0	0	0	0	0	0	0	0
MCRESOL	0	0	0	0	0	0	0	0
PCRESOL	0	0	0	0	0	0	0	0
GUAIACOL	0	0	0	0	0	0	0	0
MECYHEXN	352.75248	3406.1913	3725.0095	1.721E-07	1.721E-07	1.721E-07	7.029E-14	7.029E-14
2M3HXENE	1956.2199	15353.549	17080.86	4.052E-08	4.052E-08	4.052E-08	5.271E-15	5.271E-15
METHHEXN	311.26943	2066.3536	2332.5759	2.139E-09	2.139E-09	2.139E-09	1.97E-16	1.97E-16
3ETPNTNO	0	0	0	0	0	0	0	0
BZOFURAN	0	0	0	0	0	0	0	0
D-INDOLE	8.756E-10	6.439E-07	6.439E-07	6.702E-05	6.566E-05	6.566E-05	1.354E-06	1.354E-06
STYRENE	0	0	0	0	0	0	0	0
VANILLIN	9.178E-09	4.661E-05	4.662E-05	0.2620461	0.2139661	0.2139661	0.04808	0.04808
OXYLENE	62.834042	2082.1076	2143.7128	0.0003156	0.0003156	0.0003156	2.23E-09	2.23E-09
PXYLENE	0	0	0	0	0	0	0	0
ETYBENZN	81.33783	2152.6516	2231.8341	7.592E-05	7.592E-05	7.592E-05	3.211E-10	3.211E-10
23XYLNOL	0	0	0	0	0	0	0	0
24XYLNOL	0	0	0	0	0	0	0	0
25XYLNOL	0	0	0	0	0	0	0	0
CREOSOL	0	0	0	0	0	0	0	0
SYRINGOL	0	0	0	0	0	0	0	0
VANALC	0	0	0	0	0	0	0	0
TRIMEPYR	8.182E-07	6.544E-05	6.625E-05	6.33E-09	6.329E-09	6.329E-09	4.042E-13	4.042E-13
MEETCY5E	93.552015	2800.7227	2892.1555	0.000208	0.000208	0.000208	1.134E-09	1.134E-09
DIMECYHX	795.58275	16291.441	17052.203	0.0010808	0.0010808	0.0010808	6.078E-09	6.078E-09
TRDICYHN	1010.5308	17525.156	18482.279	0.0002561	0.0002561	0.0002561	8.312E-10	8.312E-10
ETCYHEXN	384.17617	8531.5597	8901.026	0.0006466	0.0006466	0.0006466	3.77E-09	3.77E-09
4MEHEPTN	1198.5315	17657.006	18785.828	6.778E-06	6.778E-06	6.778E-06	5.396E-12	5.396E-12
INDENE	4.3907712	457.52335	461.89417	0.6022112	0.6021174	0.6021174	9.387E-05	9.387E-05
MEBZFURA	0	0	0	0	0	0	0	0
INDANE	66.108193	5560.1704	5625.8787	1.7640128	1.7638381	1.7638381	0.0001747	0.0001747
5INDANOL	0	0	0	0	0	0	0	0
MEVINPHE	0	0	0	0	0	0	0	0
ETMOXETN	0	0	0	0	0	0	0	0
SYRNGALD	0	0	0	0	0	0	0	0
PROPBENZ	33.654929	1857.6877	1890.9818	0.0193219	0.0193213	0.0193213	5.929E-07	5.929E-07
135TMBZ	0	0	0	0	0	0	0	0
HXHYINDN	59.94662	3581.2514	3640.5905	0.1787734	0.1787637	0.1787637	9.737E-06	9.737E-06
PROPCYHX	135.21258	6016.1131	6149.0283	0.1871464	0.1871373	0.1871373	9.126E-06	9.126E-06
TRIMCYHX	156.59721	5069.4993	5222.7553	0.0006934	0.0006934	0.0006934	4.603E-09	4.603E-09
NAPTHALE	0.830556	211.57715	212.40645	535.93403	534.70235	534.70235	1.2316746	1.2316746
1NAPTHOL	0	0	0	0	0	0	0	0
MEINDENE	27.810973	4804.292	4832.033	711.90713	711.36657	711.36657	0.5405602	0.5405602
CONIFALD	0	0	0	0	0	0	0	0

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Stream →	(25)	(26)	(27)	(28)	(29)	(30)	(31)	(32)
TETRALIN	60.58572	13289.615	13350.091	9871.1985	9857.8862	9857.8862	13.312332	13.312332
EUGENOL	0	0	0	0	0	0	0	0
IEUGENOL	0	0	0	0	0	0	0	0
PRHYBNZO	0	0	0	0	0	0	0	0
D-DECALN	0.0003287	0.0392369	0.0395642	0.0009116	0.0009112	0.0009112	3.904E-07	3.904E-07
DECALIN	146.28254	24762.143	24908.02	1310.8551	1310.1997	1310.1997	0.6554263	0.6554263
D-C4CYC6	0.0013439	0.1185386	0.1198723	0.0018631	0.0018624	0.0018624	7.594E-07	7.594E-07
DIMEOCTN	65.230303	3227.6217	3291.9668	0.0142454	0.0142451	0.0142451	3.108E-07	3.108E-07
1MENAPHT	6.907E-07	0.0004578	0.0004585	0.0542813	0.0528432	0.0528432	0.0014381	0.0014381
1NAPMNOL	0	0	0	0	0	0	0	0
METDEC	23.750743	8025.5428	8049.2661	91340.68	90854.533	90854.533	486.14696	486.14696
DIBZTHIO	0	0	0	0	0	0	0	0
D-C4BZ-S	1.276E-13	1.46E-09	1.46E-09	0.0005609	8.896E-07	8.896E-07	0.00056	0.00056
BIPHENYL	0.0553163	62.710811	62.766115	30512.224	28402.03	28402.03	2110.1941	2110.1941
DIMENAPH	0.0046778	7.3512696	7.3559468	10140.413	8657.0106	8657.0106	1483.4026	1483.4026
2ETYNAPH	0	0	0	0	0	0	0	0
D-C12H18	9.959E-06	0.0033952	0.0034051	0.0204609	0.0203929	0.0203929	6.795E-05	6.795E-05
D-C12H1B	1.57E-06	0.0004925	0.0004941	0.0024682	0.0024605	0.0024605	7.661E-06	7.661E-06
2ETDECLN	0.572503	209.95858	210.53037	8594.234	8450.6973	8450.6973	143.53664	143.53664
FLUORENE	0	0	0	0	0	0	0	0
BIPHCALD	0	0	0	0	0	0	0	0
D-CQUINL	9.33E-17	4.005E-12	4.005E-12	6.779E-05	7.13E-12	7.13E-12	6.779E-05	6.779E-05
HXLCRESO	0	0	0	0	0	0	0	0
PHFLOREN	0.0003868	0.5306223	0.5310091	902.05322	707.95048	707.95048	194.10274	194.10274
HEPCYHEX	0.5973515	489.95219	490.54927	110965.11	105142.16	105142.16	5822.953	5822.953
PHENANTH	1.946E-08	0.0003605	0.0003606	1231.8294	0.0064501	0.0064501	1231.8229	1231.8229
MEFLOREN	6.581E-07	0.0071153	0.007116	2085.0489	10.425552	10.425552	2074.6233	2074.6233
FLORMEOH	0	0	0	0	0	0	0	0
D-2C2BZS	2.88E-18	4.196E-13	4.196E-13	0.000561	6.592E-16	6.592E-16	0.000561	0.000561
D-C14H14	9.613E-11	5.464E-07	5.464E-07	0.0240343	0.003908	0.003908	0.0201264	0.0201264
ETPHCYHX	0.0028752	5.5918098	5.5946846	19554.863	13557.817	13557.817	5997.0459	5997.0459
D-HYPHEN	8.303E-10	2.65E-06	2.65E-06	0.0527395	0.0142171	0.0142171	0.0385224	0.0385224
THYDPHEN	0.0016253	3.317367	3.3189921	24419.409	11553.656	11553.656	12865.753	12865.753
DIMDODEC	4.8487792	2628.0531	2632.8982	128769.49	126777.05	126777.05	1992.436	1992.436
MEANTCEN	4.084E-09	0.000168	0.000168	2085.1341	0.0008831	0.0008831	2085.1332	2085.1332
MEPHENAN	5.022E-10	2.095E-05	2.095E-05	1017.6904	3.2E-06	3.2E-06	1017.6904	1017.6904
D-3CZOLE	7.134E-19	8.209E-14	8.209E-14	6.779E-05	3.497E-16	3.497E-16	6.779E-05	6.779E-05
D-C15H16	5.221E-13	5.514E-09	5.514E-09	0.0031731	7.225E-07	7.225E-07	0.0031724	0.0031724
D-3C3BEN	5.161E-07	0.0004363	0.0004368	0.0824489	0.0792433	0.0792433	0.0032056	0.0032056
OCHYENAP	1.472E-05	0.1612841	0.1612989	132850.29	4.2206809	4.2206809	132846.07	132846.07
D-DIMTRD	0.0013537	1.0673847	1.0687377	233.50822	220.57957	220.57957	12.928659	12.928659
DIMPHENA	1.21E-11	1.69E-06	1.69E-06	1149.058	9.058E-09	9.058E-09	1149.058	1149.058
CETANE	0.0001881	0.8603652	0.8605533	26011.687	1133.2322	1133.2322	24878.454	24878.454
CHRYSENE	1.071E-16	2.853E-10	2.853E-10	2085.1434	3.153E-19	3.153E-19	2085.1434	2085.1434
DIMEPYRE	9.052E-17	3.179E-10	3.179E-10	2085.1407	9.296E-19	9.296E-19	2085.1407	2085.1407
RETENE	0	0	0	0	0	0	0	0
CYHXHXBZ	2.161E-09	0.00015	0.00015	26382.725	5.911E-07	5.911E-07	26382.725	26382.725
D-NC18	9.88E-12	1.329E-07	1.329E-07	0.1252654	3.678E-07	3.678E-07	0.125265	0.125265
D-C1C17	9.946E-12	1.433E-07	1.433E-07	0.0840597	1.704E-06	1.704E-06	0.084058	0.084058
HMWLIGNA	0	0	0	0	0	0	0	0
D-NC20	1.11E-14	4.119E-10	4.119E-10	0.0098894	2.373E-12	2.373E-12	0.0098894	0.0098894
HMWLIGNB	0	0	0	0	0	0	0	0
DBZFLORE	4.437E-21	1.059E-13	1.059E-13	481.58487	5.371E-27	5.371E-27	481.58487	481.58487
TRIPHNPR	9.381E-15	4.737E-09	4.737E-09	829.59022	2.846E-16	2.846E-16	829.59022	829.59022
MOXHXPHE	0	0	0	0	0	0	0	0
DIBZFLOR	1.77E-14	2.735E-08	2.735E-08	6907.0186	7.612E-15	7.612E-15	6907.0186	6907.0186

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Stream →	(25)	(26)	(27)	(28)	(29)	(30)	(31)	(32)
BZACMOXP	0	0	0	0	0	0	0	0
HXACMOXP	0	0	0	0	0	0	0	0
MOXHEPOX	0	0	0	0	0	0	0	0
HXDECBEN	6.668E-13	3.8E-07	3.8E-07	11913.676	1.729E-13	1.729E-13	11913.676	11913.676
P-3C3CY6	4.79E-05	0.0069767	0.0070242	0.15291	0.149193	0.149193	0.003717	0.003717

Table L-8. Heat Duties and Power for 100% CFP-A Oil HT (Case 12199011)

Tag	Description	Value	Unit
Duty			
Q1	Feed-effluent exchanger	995.9	MMBtu/h
Q2	Feed preheat furnace	99.1	MMBtu/h
Q3	Effluent cooler (makes steam)	760.9	MMBtu/h
Q4	Air cooler (cools to 140°F)	703.4	MMBtu/h
Q5	Water cooler (cools to 110°F)	145.2	MMBtu/h
Q6	Distillation cross exchange preheat	134.5	MMBtu/h
Q7	Distillation feed preheat furnace	66.1	MMBtu/h
Q8	Light products column condenser	163.8	MMBtu/h
Q9	Light products column reboiler	151.4	MMBtu/h
Q10	Heavier products column condenser	97.8	MMBtu/h
Q11	Heavier products column reboiler	69.8	MMBtu/h
Power			
W1	Makeup hydrogen compressor	46,228	kW
W2	Recycle compressor	21,373	kW
W3	Liquid recycle pump	-	kW

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Appendix M. Reviewer Comments on Previous Draft(s) and Responses

This appendix presents a summary of reviewer comments on draft versions of the report, issued for peer review. Comments from the reviewers are summarized below and followed by a response from the authors and sections/tables in the report where the information is included in the final publication. Some of the comments are paraphrased along with added context for clarity. Comments were combined when there were commonalities among observations from multiple reviewers. Minor/editorial comments were addressed but left out of this appendix.

General

1. **Comment:** The introduction does not provide a sense of the scale of production of liquid fuels via pyrolysis oil.

   Response: A paragraph has been added to the introduction (Section 1.1) to provide the potential for hypothetical full-scale implementation with resources in the United States.

2. **Comment:** How many (2000 TPD) CFP plants to satisfy one hydrotreater?

   Response: We added CFP production information in barrels per day with our results (under the "Products" sub-section in Table 26 and Table 27) so one can quickly estimate this metric for any hydrotreater capacity. Approximately four 2,000 dry metric tonnes per day CFP plants will generate 10,000 barrels per day of CFP liquid intermediates for hydroprocessing.

3. **Comment:** In a previously reviewed version, the base price of CFP oil was set at $150/bbl (2020$). What was the basis? How was the CI for CFP oil calculated?

   Response: The final version includes a detailed analysis of CFP oil production, with related costs and CI calculations for two different cases, CFP-A and CFP-B.

4. **Comment:** The term "sustainable" has regulatory implications that would require a 50% reduction in GHG emissions. CFP-derived fuels may or may not meet that definition. Alternatively, the terms "renewable" or "synthetic" could also be used.

   Response: We removed "sustainable" or sustainable aviation fuel when referring to jet fuel product cuts. The U.S. Congressional Research Service reference where this current 50% GHG emission reduction is stated is included below [1].

5. **Comment:** Where is the assumed location for hydroprocessing?

   Response: U.S. Gulf Coast basis was used for hydroprocessing capital cost estimates, as stated in Appendix B. Most of the other costs are based on U.S. Gulf Coast.

6. **Comment:** Discuss the impact of policy incentives.

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Response: We did not include policy incentives in our analysis because they vary with location and time. The interested reader can roughly estimate those impacts by applying the incentives applicable for their situation, on top of our cost estimates. We provide GHG emission estimates for each case (Appendix K), which can help enable this exercise.

7. **Comment:** Does the analysis include additional startup capital costs? Is a 6 month start up included for both the new build cases and revamp/integration cases?

   Response: 5% of FCI is included as working capital. Additional startup capital is not included. We include a 6-month startup time with assumptions of: 50% of production capacity output, 75% of variable operating costs, and 100% of fixed operating costs. The same assumption is maintained for our DCFROR calculations for all cases.

8. **Comment:** In a previous draft it was mentioned that refinery units often include 15-20% design contingency, and this may help alleviate some of the new constraints that may appear when CFP oil is introduced. A reviewer asked for clarification about contingencies applied in this report.

   Response: In subsequent drafts and in the final report, we removed any assumptions of design contingencies helping alleviate constraints and reduce costs; assumptions of design contingencies without site-specific information will be incorrect.

   For CFP, the cost contingency factor is included in Table 19 and Table 20. For hydroprocessing, FCI estimation methods provided by KBR include cost contingency within the Lang Factors (Section 1.2.2); the cost estimates factor usual refinery assumptions.

9. **Comment:** Which cost indices were used to convert to 2020 dollars?

   Response: The information is now included in Section 1.2.2.

10. **Comment:** Please add columns to allow comparison of CFP oil properties with typical hydroprocessing feeds.

    Response: Table 29 lists H:C ratios and densities for the various liquid feedstocks along with the total liquid product for hydroprocessing. Further details of CFP oil properties are provided in Table 3. Appendix D discusses properties of petroleum feedstocks used in our models and lists surrogate compounds used for the fuel products.

11. **Comment:** For 99 mol % and 85 mol % purity hydrogen, where are the hydrogen purities coming from? If this is based off of typical refinery purities, please state.

    Response: The model assumptions for the hydrogen streams are now included in Table 30. The 99 mol % H2 specification was maintained from the previous 2015 CFP design report [2], with an assumption of nominal impurities after a steam reformer. The 85 mol %

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composition was based on an approximate composition for some refinery streams based on discussion and guidance from KBR.

12. **Comment:** Hydrogen purity (from PSA, or in recycle or makeup gas) is often in mol %.

    Response: We had used vol % in the previous draft because vol % and mol % are often used interchangeably for the gas phase. However, in this case mol % is the correct unit because all our model specifications for hydrogen purity are in moles. We have made this change for all hydrogen purity specifications in the text.

13. **Comment:** What is the system boundary for GHG emission calculations and are upstream impacts from natural gas production/processing/transportation, etc., included? What is the basis for emissions calculations? What is the location assumed for intensity of the electricity grid, etc.? For the coprocessing and revamp scenarios are associated utilities for crude production, transportation, desalting and fractionation included? In a real refinery, there are many other associated downstream processes that produce streams that are blended into product fuels, are those reflected here?

    Response: The system boundary is the entire supply chain ending in fuel combustion and the emission factors include impacts from production/processing/transportation of inputs. As mentioned in the text, GREET 2023 is the basis for emissions calculations. Average U.S. grid electricity is assumed. The emissions basis for the SRD refinery intermediate fed into the hydroprocessing units is discussed in Section 3.4.2.2, and the derived number used (Table 38) should approximately reflect all operations to produce the SRD stream. As a simplification for this exercise, LCO/VGO inputs were assigned the same value as SRD, on an energy basis. Note that GHG emissions for the hydroprocessing step for stand-alone and coprocessing scenarios (including the use of utilities) are modeled and included as part of this analysis. Product blending is not accounted for, with an assumption that the calculated emissions for the produced fuel will be based on the calculations presented, and specific blends will likely have calculated emissions based on proportions of their constituents.

14. **Comment:** Is CI based on all liquid fuel products?

    Response: Yes, we calculate CIs on an energy basis for the entire pool of liquid fuel product. We did not distinguish emissions from the different cuts. This may be possible in the future if we integrate experimental fuel specifications and yields of specific cuts in the model and distinguish emissions associated with hydrogen consumption for each cut. This can be done once a predictive reactor model is developed.

15. **Comment:** Is there a tool available for calculations that determine monetary incentives from federal and state policies for the yielded fuel products?

    Response: Although we had initially considered including some sample results discussing the impacts of incentives, we did not include them in the final version. All the economics presented are without incentives. As the reviewer pointed out in the full comment (not

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included here), some incentives are transient, and one needs to be careful when making decisions based on those.

16. **Comment:** In Table 23, does "Maintenance" mean "Maintenance and Materials"?

    Response: We assume a flat 3% of fixed capital investment (FCI) for estimating maintenance costs. This includes the cost of associated materials, since there is no separate line item for materials.

17. **Comment:** In Table 23, "Insurance and Taxes" included are 0.7% of FCI. This seems low.

    Response: We maintained this assumption for consistency with our other analysis. Note that a 21% income tax is accounted separately in our DCFROR calculations. The 0.7% of FCI included here accounts for other taxes, such as property and other local taxes.

18. **Comment:** Key sustainability details would be helpful to understand what data was used to calculate the CI.

    Response: Table 26, Table 27, and Table 38 are provided along with emission factor values.

19. **Comment:** Suggest caveating the mention of jet fuel production since this pathway is not currently ASTM approved.

    Response: A sentence has been added to the first paragraph in the Introduction (Section 1) to state that this pathway is not ASTM approved.

Process Model

20. **Comment:** How sensitive are the numbers to small changes the model compounds used?

    Response: With the modeling assumptions explained in the text, the biggest impacts on economic and sustainability assessments are from the overall oxygen content of the CFP oil and overall H:C ratio of the product fuel. The surrogate compound slates chosen for (1) CFP-A and CFP-B are shown in Appendix C; (2) SRD, LCO, and VGO in Appendix D; and (3) final fuel products in Appendix D and Appendix L. As noted in the text, the overall H:C ratio of the total final liquid fuel product was ~1.8, and no attempts were made during the modeling exercise for this report to match fuel properties of individual cuts (if measured); separate work on fuel property predictions is ongoing and may be integrated with the current models in the future. Sensitivity to deliberately tweaking individual model compounds and their proportions was not explored as part of the modeling for this report. However, model compounds and their proportions were changed for previous annual state of technology assessments [3]; those exercises informed that our overall model outputs were generally representative if the oxygen content of CFP oils and H:C ratio of final liquid hydrocarbon fuels were matched with the experimental analysis. Although heating value measurements are not available for the CFP oils presented in this report, the process model reports LHVs of 12,408 Btu/lb (28.86 MJ/kg) for CFP-A (21.6 wt % O) and 13,908 Btu/lb (32.35 MJ/kg) for

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CFP-B (16.0 wt % O). These values are generally consistent with measured LHVs reported by Paasikallio et al. [4]; upon deriving a linear fit for their narrow-range data, 17.5 to 19.5 wt % O CFP oils, which is near our region of interest (*LHV in MJ/kg* = −0.6944 × *CFP oil wt % O* + 43.974), the LHVs for 16 wt % O and 21.6 wt % O CFP oils were 32.86 MJ/kg and 28.97 MJ/kg, respectively. Similarly, for the total hydrocarbon liquid fuel product from Case 11120211, with a H:C ratio of 1.8, the model reports an LHV of 42.62 MJ/kg compared to 42.45 MJ/kg according to a correlation (*LHV in MJ/kg* = 4.596 × *H:C ratio of liquid fuel* + 34.197) from Yue et al. [5]. Thus, these additional verifications indicate that our approach of matching overall oxygen contents for CFP oils and overall H:C in the final liquid product is reasonable.

21. **Comment:** The properties of the CFP oils used must be presented, e.g., CHNO, density, etc. Comparison of CFP oil properties with SRD/LCO/VGO and FP oil are needed.

    Response: Table 3 shows FP and CFP oil properties and SRD, LCO, VGO properties are included in the Appendix D. Finished fuel H:C molar ratio was approximately 1.8 in all modeled cases for the entire liquid fuel product slate. Density assumptions are included in Table 29.

22. **Comment:** Recommend providing a typical range of carbon loss for FCC coprocessing of CFP oils, with references.

    Response: This has been included in Section 1.1.2 of the final report.

23. **Comment:** Specify “biomass feedstock” when used, instead of just “feedstock”. Add a row for chlorine with the biomass specification.

    Response: We made this change to distinguish biomass feedstock for CFP vs. liquid feedstocks for hydroprocessing. We did not add chlorine because we focus on woody feedstocks where chlorine is not a concern because it is not a major constituent. We appreciate that chlorine needs to be included when processing non-woody feedstocks.

24. **Comment:** Model representations when working with surrogate model compound slates are never perfect. It is often better to work with pseudo-compound mixtures for oil cuts. Has this been attempted for pyrolysis or other bio-mixtures? What thermo package was used in the models?

    Response: Pseudo-components are most often used to represent hydrocarbon systems. Introduction of oxygen, typical of bio-derived intermediates like pyrolysis liquids, introduces various physical property behavior based on the functional groups associated with the oxygen in the molecules. As a result, the relatively simple approach with pseudo-components fails to capture properties in such cases. This was the primary reason for our choice of using surrogate model compounds. The primary property method used in the process models is stated in Section 1.2.4.

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Catalytic Fast Pyrolysis

25. **Comment:** One can change the carbon yield by changing the biomass-to-catalyst-mass ratio in a ZSM-5 system, at the expense of oxygen content.

    Response: This is now mentioned in Section 2.2.2.

26. **Comment:** What is meant by higher yield CFP scenario is mentioned in the text?

    Response: We clarified that this is the potential future case using a Pt/TiO_2 catalyst in the presence of hydrogen, with improved CFP yields and 16.0 wt % oxygen content in the CFP oil (Section 2.2.1).

27. **Comment:** Pt/TiO_2 catalyst process involves the addition of hydrogen, and the resulting oil has already been lightly hydrotreated, this should be stated.

    Response: A sentence has been added to Section 1.1.1.

28. **Comment:** There is a need for the pyrolysis community to define a specification for CFP oil quality that is suitable for coprocessing, as well as storage stability. Long-term storage stability needs to be addressed. Recommend expanding this section in the body and discuss potential next steps.

    Response: A paragraph on stability and corrosivity has been added to the discussion in Section 2.2.2. The lack of larger volumetric production and limited availability are constraints toward maturity of this process, and the next steps should evolve rapidly once this constraint is overcome, because many of these challenges can be addressed through materials, process, and engineering adaptations. Similarly, requirements and specifications for CFP oils will evolve once sufficient quantities of different CFP oils are available for hydroprocessing tests at different scales, and that can help identify undesirable properties and associated critical limits.

29. **Comment:** At a high level, include pilot-scale testing requirements needed to further de-risk CFP oil coprocessing. This can address details on the pilot plant physical scale, test campaign durations, etc.

    Response: We did include details of bench-scale results so far (Appendix A). Pilot facilities are necessary for the next stage of testing toward maturation. However, the supply of CFP oils remains a constraint for large-scale testing, although there are upcoming facilities pursuing biomass CFP scale-up and production of larger volumes of CFP oils [6]. Progress in hydroprocessing at the pilot scale can occur with increased CFP oil availability, and these scale-up efforts will likely be pursued by industrial entities.

30. **Comment:** Add a property table for CFP oil with expected ranges. List chlorine content and any other known contaminants of concern for refinery integration.

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Response: We added Table 3 showing some experimental results. CHONS values are included, but chlorine was not measured because its proportion in woody biomass is known to be very small.

31. **Comment:** The stabilization (pretreatment) of raw pyrolysis oils is the big challenge. Can this be addressed?

    Response: We limited the scope of this report to CFP oils. We appreciate that hydroprocessing approaches for raw pyrolysis oils (non-catalytic pyrolysis) are significantly more challenging and we did not attempt to address that in this report.

32. **Comment:** The CFP costs shown in Table 18 are lower than publicly available information for FP and CFP projects. Show the impacts of higher investment costs.

    Response: We understand that project cost is one of the primary areas of TEA uncertainty.

    As a clarification, the costs presented in Table 18 show the total purchased equipment cost (TPEC) and the total installed cost (TIC). The TIC is not the cost typically reported publicly. Our factored approach for estimating investment costs adds 78% (for CFP-A) and 76% (for CFP-B) to the TIC for estimating the fixed capital investment or FCI (see Table 20). Part of the reviewer's concerns about the low investment cost estimates are addressed through this factored increase from TIC to FCI.

    Information in the public-domain or in press releases may be incomplete and we typically refrain from publishing direct comparisons with our TEA results. However, since the reviewer pointed out specific public information for some projects, we made high level comparisons to understand how our mature plant estimates compare with reported commercialization costs for pioneer plants. The companies and publicly available sources (not peer reviewed or otherwise vetted independently) are deliberately not named to avoid misrepresenting any of the entities. Costs were reported in different years and currencies; the approximate figures below reflect conversion to 2020 U.S. dollars through simple factors based on approximate conversion rates and indices. Average values were used where ranges were reported. Scaling exponents were applied directly to the publicly reported costs to gauge the approximate impacts of scaling. Recent higher inflation after 2020 will result in higher costs once the values are converted using indices to estimate 2024-2025 costs. Note that plant cost increases between 2011 and 2020 were relatively not significant based on the available values of indices.

    Based on our worksheets (Table 18 and Table 20) and cost breakdown in Appendix E (for CFP-A), the sum of TIC for the FP and condensation sections (not including the *ex situ* vapor upgrading reactor) is approximately $144.5 million, with an estimated FCI of $257 million using the estimation factors shown in Table 20.

    In one case, a 120 tonnes biomass/day FP plant was projected to cost approximately $30 million based on a 2019 source. Hypothetically, the approximate investment for a 2,000 tonnes biomass/day unit will be $480 million when scaling by numbers (approximately 16x

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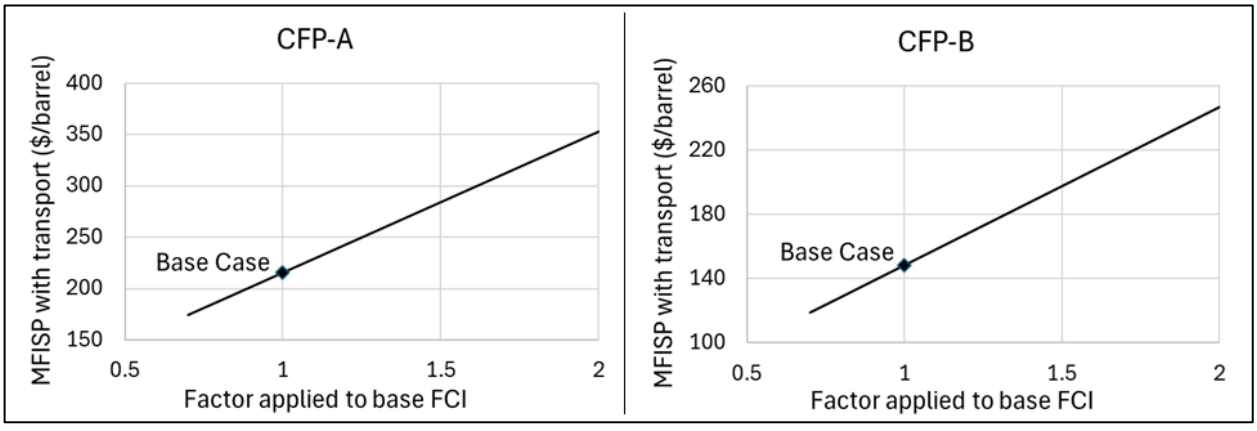
going up from 120 to 2,000 tonnes biomass/day) and $240 million if using a scaling exponent of 0.75 for the 16x capacity.

In a second case a 400 tonnes biomass/day FP plant was projected to incur approximately $80 million CAPEX (assumed to mean investment cost) based on a 2011 source. Scaling this cost to 2,000 tonnes/day by numbers (5x going up from 400 to 2,000 tonnes biomass/day) provides in investment cost estimate of $400 million. If a 0.75 scaling exponent is used to scale up to 1,000 tonnes biomass/day and multiplied 2x to estimate the cost of a 2,000 tonnes/day unit, the resulting investment cost estimate is ~$320 million.

A third public source from 2012 reported an investment of approximately $220 million for a 500 tonnes biomass/day plant producing fuels via *in situ* CFP. Estimating by numbers (4x going up from 500 to 2,000 tonnes biomass/day), the investment necessary for a 2,000 tonnes biomass/day plant is $880 million. If a 0.75 scaling exponent is used to scale up to 1,000 tonnes biomass/day and 2x that cost to estimate the cost of a 2,000 tonnes biomass/day unit, the resulting investment cost estimate is approximately $740 million. Scaling 4x with a 0.75 scaling exponent gives an estimated investment cost of $622 million. The total capital investment or TCI for the modeled CFP-A plant in this report is $624 million (Table 20) which includes an additional *ex situ* vapor upgrading reactor but does not include biomass feedstock processing and a hydrotreating section for producing finished fuels (one or both of these may have been included in the publicly reported investment estimate above).

These comparisons illustrate the uncertainties in plant cost estimates. Some of those publicly available investment costs align better with our cost estimates than others. The chosen FP configuration in this report is scalable and will likely benefit from cost reduction associated with larger-scales in a mature plant scenario, although some of those scales (e.g., single-train 1,000 tonnes biomass/day) have not been achieved yet in installed FP units. As discussed in the report, mature plant cases can provide additional cost reductions over pioneer plants. In addition, all the biomass feedstock preprocessing costs are included as an operating cost in this report (based on INL modeled estimates); some of the publicly reported investment costs may include part of the capital costs incurred for feedstock preprocessing facilities.

The charts below illustrate MFISP impacts of varying capital costs for the modeled CFP-A and CFP-B cases (using the financial assumptions discussed in Section 2.3.12):



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33. **Comment:** Provide more discussion about the CFP process with its pros and cons and commercialization efforts with potential challenges.

    Response: We added Section 2 on CFP in the final report and provide the background of our CFP assumptions. Some entities are working to successfully commercialize CFP processes, and we included references. We did not discuss those efforts in this report but have other ongoing work that will compile current commercialization efforts. We hope to publish those findings separately.

Hydroprocessing

34. **Comment:** What was the longest run in the small-scale testing (hydrotreating) and was it long enough to draw conclusions about catalyst fouling and the production fuels that meet current specifications? What was the proportion of CFP oil during co-hydroprocessing?

    Response: A detailed discussion of experimental data and associated fuel specification tests are now included in Section 3.1 and a summary of experiments is provided in Appendix A. It is noted that long duration experiments are necessary, but the short-term experimental results are encouraging. Durations of the tests are also reported, if available.

35. **Comment:** Regarding the discussion of phenols in wastewater, is the toxicity of phenolics in wastewater the primary concern?

    Response: We have added toxicity to the list of potential problems as part of the discussion in Section 3.2.11.

36. **Comment:** It is mentioned that incremental CO_2 may impact the apparent capacity of the sulfur recovery unit, but the impact is expected to be minimal and offset by reduced sulfur load to the refinery with CFP oil processing. Is this assumption valid given the relatively low loading of sulfur in SRD?

    Response: Augmentation of amine systems based on changes in acid gas loading is included in the cost analysis. It is not assumed that no changes will be necessary to the sulfur recovery unit in all circumstances.

37. **Comment:** In a previously reviewed version: need to include hydroprocessing catalyst information.

    Response: Available experimental information and catalysts used are presented in detail in Section 3.1 and Appendix A.

38. **Comment:** In a previously reviewed version: need to include sensitivity to CFP oil oxygen content and H:C ratio. These will have a large impact on hydrogen consumption.

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Response: Sensitivity cases are included to show two different CFP oil oxygen contents (CFP-A and CFP-B) and changes in H:C ratio in the final product.

39. **Comment:** Explore the effect of using green hydrogen.

    Response: This is not included explicitly in the report. However, the granularity in the results presented allows the interested reader to prorate the cost and GHG emission impacts using different hydrogen costs and associated emission factors.

40. **Comment:** 2-stage hydroprocessing configurations offer more flexibility over 1-stage configurations. Which is more common in existing refineries?

    Response: While we did not look into the current inventory of refinery setups, we followed up with a discussion of 1-stage and 2-stage systems in Section 3.2.1, highlighting that both these configurations are prevalent in current refineries and the reason we chose only 1-stage configurations for the analysis in this report. Further discussion 1-stage and 2-stage systems is available in literature including an EPA background information document [7].

41. **Comment:** How do these CI calculations in this report compare against baseline values for ultra-low-sulfur diesel and other petroleum products under GREET and the CARB LCFS?

    Response: The lifecycle GHG emissions for the 100% SRD hydrotreating case is matched with the value for diesel in GREET (90.8 gCO2e/MJ). A discussion of related emission factors is included in Section 3.4.2.2. We did not look into CARB LCFS for this report.

42. **Comment:** Operations to maintain hydroprocessor performance with CFP oils, as they do with petroleum feedstocks, will require modifications or replacement of reactor internals.

    Response: In the absence of further detailed general guidance from available experimental results, we captured the impacts of CFP oil processing on reactor costs through WHSV adjustments discussed in Section 3.2.7, and adjustment of reactor capital costs based on gas + liquid throughput.

43. **Comment:** Include temperature ranges in degrees F in Table 28.

    Response: A row showing temperatures in degrees F has been added.

44. **Comment:** What is the impact of shorter cycle length for the hydroprocessing catalyst when CFP oil is introduced?

    Response: Catalyst cost contributions are included for all the cases in Appendix K. Catalyst costs can be estimated by prorating costs for shorter cycles (e.g., reducing the cycle length from 2 years to 1 year will approximately double the catalyst cost contribution). However, a significant impact of more frequent catalyst replacement will be a lower on-stream factor (discussed in Section 3.4.4.11) with lost revenue for the refinery.

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45. **Comment:** Steam categories, by pressure, are defined as VLP, LP, MP, HP. Should they be named otherwise: LP, MP, HP, HHP?

    Response: We kept our original nomenclature but clearly state the pressure assumed for each category (Table 30).

46. **Comment:** Expand results beyond MFSP and CI to include capital costs of each of the configuration, how product yields change, process constraints identified in the model such as exotherm and hydrogen consumption, penalties suffered (temperature, catalyst life, product yield shifts) on fossil hydroprocessing objectives due to the inhibitory effect of coprocessing biomass-derived liquid intermediates.

    Response: We included additional metrics and many sensitivity studies in the final report. Yield and quality impacts need to be addressed when more experimental data is available for the development of kinetic models; they are not addressed in this report.

47. **Comment:** The choice of gasoline equivalent as the basis for Minimum Fuel Selling Price (MFSP) reporting may be debatable. Industrial biofuel production is focused on making diesel and jet fuel, not gasoline. Naphtha (gasoline boiling range hydrocarbon) produced from bio-hydroprocessing has very low octane value, and is often considered an undesirable byproduct. It would be most helpful if the product yields (gases, naphtha, jet, diesel) are reported individually for each case studied. More detail discussion on how the varying yields are summarized by a single MFSP-gasoline-equivalent number would be very helpful.

    Response: We chose to report in gasoline equivalent because it is a relatable metric in the U.S. As explained in the final version, the use of GGE does not imply that gasoline is the primary product. We added other energy metrics for the total fuel product, including jet and diesel equivalents. We mention that the gasoline, jet, and diesel cuts reported from the model (shown in Appendix K) are nominal and specifications of the cuts are not tied fuel specifications. The primary metric of the model was to achieve a H:C ratio of ~1.8 for the total liquid fuel product in order to ensure consistency across all the cases and provide similar bases for comparison.

48. **Comment:** Details of the process and kinetic models are of interest.

    Response: As clarified in the text, we do not have enough experimental information at this point to predict kinetic behavior. Details of other process model inputs and assumptions are explained in Section 3.2; compounds used in the process model and stream summaries from bookend cases are included in appendices of the final report.

49. **Comment:** Use straight-run distillate instead of straight-run diesel for the SRD acronym.

    Response: We used straight-run diesel for SRD because the material received and used in experiments was named as such. We have used that term in previous publications [8] and are

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maintaining the same in this report for consistency, understanding the reviewer's note that diesel is in distillate range.

50. **Comment:** The process simulation can help understand the limits with more extensive parameter space for co-hydroprocessing.

    Response: We included many sensitivity cases in the final report where the impacts of single parameter variations are shown. Since the reactor model is not predictive, the single parameter impacts are not convoluted. The interested reader may be able to understand approximate impacts for specific cases of their interest through linear interpolations of the results. However, for some parameters, one needs to be careful with specific quantification using interpolation, especially when the impact is known to be nonlinear.

51. **Comment:** Comment related to the statement Section 3.2.12: "Any specially designed catalyst technologies for optimization of CFP oil upgrading, if necessary, are assumed to negligibly impact catalyst unit cost". A reviewer commented that it had been stated earlier in the text that the report was constrained to existing catalyst systems, and not specially designed ones.

    Response: The reviewer is correct that we assume the use of commercial petroleum hydroprocessing catalysts. We retained the statement above, in Section 3.2.12, to reinforce that the catalyst unit cost assumption was not varied and needs to be adjusted if other catalysts are used. Although we do not include a specific sensitivity with catalyst unit cost, that impact can be quantified by prorating using the cost breakdowns for each case provided in Appendix K.

52. **Comment:** Comment related to the statement Section 3.2.12: "The fuel properties of products from CFP oil are assumed to meet quality specifications of fossil derived equivalents after adequate hydroprocessing." This needs to be checked. One needs to meet specifications beyond distillation points, such as cold flow properties, smoke point, viscosity, etc.

    Response: We included an extensive discussion of experimental results, including multiple jet fuel properties measured as part of experimental efforts. Based on the experimental data so far, we feel that with adequate hydroprocessing, one can meet required fuel specifications. Once again, as stated in the report, for this process modeling effort we did not try to match fuel properties in the modeled product streams; the primary model specification with respect to quality was to achieve a H:C molar ratio of ~1.8 in the total liquid fuel product.

53. **Comment:** Additional discussion was requested about the CFP oil feed system including changes in CFP oil properties before and after the stabilization reactor.

    Response: The feed system is discussed in Section 3.2.13 of the final version. The composition of the stream after stabilization was not quantified during our limited experiments, although we intend to analyze this intermediate stream if resources are

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available in the future. As mentioned in Section 3.2.13, the design of the feed system may need to be adapted based on the quality of CFP oil. Thus, an analysis of the product after stabilization will be important.

54. **Comment:** Include the yields and heat content of the products in the results. That would be a good indicator of how the processing options change based on differing amounts of CFP oil and indirectly indicate hydrogen usage. Regulatory incentives for renewable fuels are not geared toward gasoline and instead favor sustainable aviation fuel (SAF) and renewable diesel. It will be useful to understand what products would really be maximized (e.g., SAF) and which would be minimized (low-octane gasoline).

    Response: As stated in the report, the included reactor model is not predictive and does not predict yields and properties of product fuel cuts. The amounts of distillation cuts reported in Appendix K are based on the same boiling cuts specifications for all the cases, as stated in Section 3.2.3. Mass flow of the total liquid product, along with weight fractions of each cut are reported along with the LHV of the liquid fuel product. LHVs are reported in barrels gasoline equivalent (BGE), with a 42-gal/bbl conversion used to convert GGE to BGE. Experimental distributions of cuts from bench-scale experiments are discussed in Section 3.1, indicating that with sufficient hydrotreatment one can achieve 40-50 wt % yields of jet fuel from CFP oils. In this context it is pointed out that the jet fuel cuts after hydroprocessing 100% CFP-A and CFP-B oils, if certified for use as jet fuel based on fuel properties, can qualify as SAF because they achieve greater than 50% GHG reduction compared to corresponding petroleum-derived jet fuel, per criteria in current legislation [1]. However, the CFP oils hydroprocessing pathway is not currently certified for acceptance as jet fuel, and more R&D and toward increased production along with associated testing is necessary before this conversion pathway can successfully integrate with jet fuel supplies.

55. **Comment:** In the previous draft version case for 20% CFP oil in the feedstock, the cost impact between 25k BPD and 75k BPD units was 4 cents per GGE and the impact from using an existing hydroprocessing unit was between 1-3 cents per GGE. One would have expected the MFSP to be much larger for new units, compared to the repurposed case.

    Response: Figure 28 in Section 3.4.4.12 shows similar results in the final report. As mentioned in the report, the impact of capital on the MFSP decreases when large volumes are processed, noting that the quantifications presented are based on our TEA model assumptions. One can see that the impact of capital costs, and hence the difference in the MFSP between new and repurposed units, increase significantly at the smaller scales of 2.4k BPD and 8k BPD shown in Figure 28.

56. **Comment:** Plots presented in the earlier draft switched between wt % and vol % for the proportion of CFP oil included during coprocessing. One should stick with vol % of CFP oil in the feed.

    Response: The plots in the current version only use vol % of CFP oil.

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57. **Comment:** In the previous draft version, explain the non-linearity in CI and hydrogen consumption when 10% increments of CFP oil were introduced in the cofeed with petroleum feedstocks.

    Response: In the current version, the modeled CI based on 10 vol % increments of CFP oil follows a linear trend (Figure 14 and Figure 15). The reason for the non-linearity in the previous version was because the plot in the previous version was presented using wt % CFP oil on the x-axis (instead of vol %). Since the analysis assumption includes constant total feed volume in these cases, and petroleum and CFP oil feeds differ significantly in their densities, the trend with total weight of the feed is not linear (when constant volume of feed is maintained). Hence the use of wt % on the x-axis results showed a nonlinear trend for CI (and hydrogen consumption) in the previous draft version.

58. **Comment:** In the previous draft, CI presented for the same case in two different figures with 20 % CFP oil did not match.

    Response: The reason for the difference is the same as indicated in the previous comment. One of the figures reported the value for 20 wt % CFP oil, while the other reported for 20 vol % CFP oil. The values at the 20 wt % and 20 vol % differ because of differences in the densities of CFP oil and petroleum feedstocks. All figures in the final version use vol % CFP oil for consistency. Most of the values presented in the figures are also tabulated in Appendix K for the interested reader.

59. **Comment:** The use of the term "biogenic CI" (in the previous draft) is confusing. The CI of feedstocks like SRD and CFP oil should be fixed.

    Response: The term "biogenic CI" has been removed from the text in the final publication. The CI values used for SRD, LCO, and VGO are shown in Table 38, as are the calculated CIs for CFP oils from each pathway CFP-A (Table 26) and CFP-B (Table 27).

60. **Comment:** It may be useful to have a total heteroatoms vs. CFP oil content plot. This can be a surrogate for hydrogen consumption.

    Response: We agree that CFP oil heteroatoms content is an indicator of hydrogen consumption. There is an additional key component in the calculation, the H:C molar ratio (in the feed and products) that will impact the hydrogen consumption and require the inclusion of multiple parameters to complete this plot. We presented two CFP oil cases (CFP-A and CFP-B) with different oxygen contents in the final report but did not include the mentioned plot. It can be derived for any specific choices of feed and their proportions. Once the petroleum feed (with its H:C ratio and constituent heteroatoms) and CFP oil feed (with its H:C ratio and constituent oxygen) are determined, the hydrogen consumption for achieving the product's H:C ratio (along with removal of heteroatoms) will be dictated by reaction stoichiometry.

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61. **Comment:** Include % GHG reduction to allow lookup of various credits associated with the % GHG reduction. E.g., if 50% GHG reduction is necessary to qualify as in SAF then one can extend the MFSP results to include those credits after accounting for credits based on the % GHG Reduction.

    Response: We included the calculated GHGs for all the cases presented (in Appendix K). We also included the GHG reduction based on a diesel benchmark, which can easily be modified with a jet fuel benchmark. This will allow the reader to estimate the impact of credits for a specific situation. We, however, did not present the MFSPs after subtracting any specific credits in this report.

62. **Comment:** Color coding for different plant sections in the process flow diagram (in the previous draft) was confusing. Based on the color coding it seemed that repurposing an existing asset for dedicated CFP oil will require the addition of a new distillation train, but that coprocessing would not need the new distillation system. Unless the distillation unit would be shared between multiple different hydroprocessing units it seems like this cost should be the same for the two scenarios.

    Response: The color coding has been removed from the diagram (now Figure B-1). The reviewer is correct that the cost when repurposing distillation units should be similar irrespective of whether the unit is repurposed for stand-alone or coprocessing operations. Some of the costs incurred for modifications will be based on capacity changes, and any additional changes in product cuts. As mentioned in the main text, the FCI for each repurposed equipment was estimated by subtracting the cost of new equipment from the cost of the original unit being repurposed.

63. **Comment:** Is the reported MFSP an overall weighted average of the different products? If so, it should be described. It would be helpful to first discuss the overall yield structure of gases/gasoline/jet/diesel for the different feedstocks before jumping into the economic summary.

    Response: This is an important point, but since we don't currently have enough information to develop predictive reactor models that can include interactions among various liquid feed mixtures in a hydroprocessing environment, we chose to present an overall MFSP for the entire liquid fuel product slate on an energy basis. The total liquid product slate has a H:C molar ratio of ~1.8 across all the cases (except for specific sensitivity cases varying the H:C parameter), because this helps make direct comparisons among the cases. The overall H:C ratio of ~1.8 was chosen to represent hydroprocessing to a point where one can consider distillation into different potentially viable fuel cuts and is convenient for interpreting process model results without being convoluted by incomplete or premature reactor model predictions of fuel properties and associated H:C ratios. Although we do report the different cuts for each case (in Appendix K), these cuts are purely distillation (boiling point) based, and the model compounds do not represent any detailed consideration of required quality specifications within individual cuts; hence we de-emphasize that aspect for this report and

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focus on the overall MFSP on an energy basis. We have other ongoing work [9] to tie that aspect into future process models when more data becomes available.

64. **Comment:** Comment regarding the trend in the plot shown in Figure 30. One would expect a graph like this to have a local minimum at some optimum WHSV below which the yield is too poor and above which the capital costs and catalyst replacement costs are too high. This graph would imply one should go to a much higher WHSV.

    Response: As mentioned earlier, the reactor behavior is not predictive. The goal of this sensitivity case is to analyze cost impacts if more or less catalyst is necessary to achieve the same hydroprocessing performance (in this case, a fuel product H:C ratio of ~1.8). The reviewer's observation can be illustrated if we develop a kinetic model that can be used to predict reactor behavior, including classification of product species to represent different fuel cuts.

65. **Comment:** Include the MFSP for a 0% CFP (100% fossil) baseline for comparison purposes. Expand the results to include potential incentives with the MFSPs presented. One insightful way to present that data would be to show what policy incentive would be required for the various CFP pathways to be on economic parity with the fossil baseline.

    Response: In the final draft we provided information for all the cases that include both MFSPs and CI (in Appendix K). The interested reader may apply potential incentives based on the information provided. We did not present those calculations using specific incentives. The delta from the fossil baseline provided will allow the reader to easily calculate the cost differential, hence the incentives required for cost parity.

66. **Comment:** Include the CI for 0% CFP oil cases.

    Response: This has been included in the relevant plots and all values are included in Appendix K.

67. **Comment:** Different CFP oils do not behave the same during hydrotreatment, even if they might have the same oxygen content. Is there enough data to discuss any differences in this report?

    Response: We currently have limited data, and additional experimentation is necessary before we can predict reactor behavior. However, we included relevant experimental data available to us and ones we identified in the literature (Appendix A).

68. **Comment:** In the previous draft there was a discussion that appeared to favor stand-alone hydroprocessing over coprocessing in potential initial implementations. The reviewer pointed out that this was contrary to the evolution of hydroprocessed esters and fatty acids (HEFA) derived fuels technology. HEFA developed primarily via coprocessing first.

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Response: We have revised the language in the final version of this report to reflect what the reviewer mentioned.

69. **Comment:** For bioliquid coprocessing, one should assume the cost of refinery modification increases with the level of bioliquid inserted. For HEFA coprocessing, typically 0-5% blend requires just a catalyst change, 5-15% blend requires additional changes, and for >20% it often seems that a 100% HEFA product unit may be more reasonable. Can one make that analogy here?

    Response: We thank the reviewer for providing these insights. A constraints-focused analysis is in our future plans. In this final version we already have the information necessary for such analysis, although not directly presented in that manner. Most of the tables and plots show gas flows with total liquid flow capacity maintained. The results show how some of the parameters start running into process constraints by exceeding what may be design allowances of the systems. Often ~15% allowances are included in equipment designs but will vary for specific situations; we did not include potential cost savings from design allowances because we did not want to make a general design allowance assumption applicable to all units. We present options for mitigating some of the constraints, e.g., through liquid recycle or increasing hydrogen makeup purity. We feel we have presented information (along with details in Appendix K) along the lines of the reviewer's request but can further specifically address this in subsequent publications since we already have the model framework to do so. Guidelines in this regard will depend on multiple parameters, including the quality (primarily oxygen content) of the CFP oil feed.

70. **Comment:** For HEFA hydroprocessing unit conversions to 100% biofeed, one typically sees a 60-80% reduction in unit throughput capacity compared to petroleum feed processing due to the generation of additional heat.

    Response: Thank you for this information. One can consider different parameters as the constraint that limits the capacity of the unit. For illustration, using the total catalyst load (included in Appendix K) as the parameter to approximate the necessary capacity reduction we can come up with approximate reductions for different feed combinations:

Petroleum Feed	CFP Oil Feed	Catalyst load for 100% Petroleum for 75k BPD (lb)	Catalyst load for 100% CFP Oil for 75k BPD (lb)	Estimated Capacity Reduction
SRD	CFP-A	549,791	6,829,318	92%
SRD	CFP-B	549,019	4,374,350	87%
LCO/VGO	CFP-A	2,054,530	6,812,546	70%
LCO/VGO	CFP-B	2,054,471	4,353,285	53%

    Note that these reduction estimates do not include any other mitigation strategies like liquid recycle. In many cases, additional mitigation strategies will need to be deployed for practical implementations.

71. **Comment:** Coprocessing can mean different things to different people. Some people assume coprocessing means blending the liquids in the same storage tank, while generally

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blending liquids in line to the first reactor or even providing separate feeds to the first reactor is much more likely.

Response: In our case we assume providing separate feeds to the first reactor, as discussed in Section 3.2.13.

72. **Comment:** Include typical gas and liquid recycle ratios from literature in Table 28.

Response: These values differ with reactors, feedstocks, and desired product cuts. We did not include these values in Table 28, but do include relevant information in Appendix K so the interested reader has these values for our modeled cases. Based on values presented by Gary et al. [10] one can derive a hydrogen recycle to consumption ratio range of 2.5 to 10. In our modeled single-stage configurations, the total liquid product is recycled to help maintain the total liquid feed rate to the reactor and provide reactor cooling. No additional conversion of recycled material is assumed in these cases. In refinery operations recycle of heavier cuts is often included for additional cracking of heavier cuts. Gary et al. [10] show an example of a hydrocracker yield structure with 35 vol % of the initial feed (coker gas oil + FCC decant oil) distilled and yielded as gas oil, which can potentially be recycled to the hydrocracker or a second stage hydrocracker. Recycle in such cases will be dictated by the yield structure and the yield targets for product cuts that are attainable within the reactor system configuration.

73. **Comment:** In Figure B-1, recommend including equipment labels in parenthesis to make it easier to follow. E.g., stabilization reactor (RX405).

Response: We did not include the equipment type, e.g., “stabilization reactor”, on the diagram. Only the equipment number, e.g., RX405, is included in the diagram to avoid it becoming too crowded. The table immediately following the diagram provides the corresponding information about the type of equipment, and this is now mentioned in the Figure B-1 caption.

74. **Comment:** Is two years of catalyst lifetime supported by experimental data for extended time on stream testing at pilot scale? If not, this should be stated. Likewise, what is the assumed lifetime of the pre-treat catalyst section? Recommend a sensitivity case for catalyst lifetime and replacement costs/LCA impacts, beyond just liquid hourly space velocity.

Response: The two-year catalyst lifetime is an assumption for the long term and not proven yet through long-term testing. This has been noted along with a sensitivity analysis presented in Section 3.4.4.14. The pretreatment catalyst cost is assumed to be lumped with the main reactor catalyst (Section 3.2.13) and thus also assumed to have the same lifetime. Depending on experimental results and how elaborate the pretreatment configuration is, one may consider parallel and redundant pretreatment train(s) to allow achieving the same catalyst changeout or maintenance cycles.

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75. **Comment:** Recommend including necessary hydrotreater pilot plant physical parameters and test campaign durations in the appendix. This would include minimum reactor tube diameters and L/D ratios, liquid flow rate, gas flux, etc. Extended test campaigns (e.g., 1000-2000 hours) should also be included to ensure modeling results and conclusions are based on sound pilot plant de-risking data.

    Response: We included data from bench-scale experiments available in the literature along with relevant references. We hope more information will be publicly available in the future, especially from pilot-scale testing.

76. **Comment:** There may be a need to stabilize bulk oil supplies before processing CFP oil material in other units. In addition to stability, there may be other concerns like metals contamination, molecular weight, etc.

    Response: Content has been added to the conclusion section suggesting the potential need for pre-stabilization if there are variations in CFP oil supplies beyond acceptable limits.

77. **Comment:** Language in a previous draft seemed to advocate repurposing existing hydrotreating units for rapid scale-up of stand-alone upgrading without doing the appropriate work to identify and mitigate risks at smaller scales. Is this what the authors intended?

    Response: This was not the intent, and the need for experiments enabling incremental scale-up has been added to the conclusion of the final version. Additional feedback, beyond this review process, suggested that the capital investment barrier for stand-alone units can be significant in the near term, and especially difficult to justify without significant stable supplies of CFP oil. Thus, some of the near-term industrial scale-up efforts may pursue incremental introduction of CFP oils in coprocessing setups when CFP oils become available in larger volumes.

78. **Comment:** Feed filters should be included, e.g., in Figure 13.

    Response: We did not add filters to Figure 13 and other relevant figures with flow diagrams but understand that those are important to include in the feed system. We have added "(feed filters not shown)" to the caption of Figure 13 to note the need for feed filters in the process.

79. **Comment:** In the case of the use of existing equipment, metaling up may be required. Was that captured in costing? One needs to be careful of metallurgy choices. One of the big suitability criteria for CFP oil coprocessing is potential acidity and corrosion management.

    Response: The cost of metaling up parts of the process was not explicitly included in the cost estimates when existing equipment was reused for coprocessing in the conceptual design cases. As mentioned in the report, additional retrofit costs for each equipment were estimated by subtracting 'the cost of equipment with the new capacities necessary for coprocessing' from 'the cost estimate of the corresponding existing equipment'. The base

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case for existing equipment costs are stated in Appendix K (under "Repurposed Case ID" in the tables). A zero cost was assumed when this difference was negative for any specific equipment. A separate CFP oil feed system is proposed to address some of the potential corrosion issues, to allow the improvement of chemical properties of the CFP oil before it enters the regular petroleum hydroprocessing equipment. Additional detailed assessments of metallurgy are necessary for specific implementations.

80. **Comment:** The 90% onstream factor for hydroprocessing (in the previous draft) was very low relative to conventional refinery operations. Onstream factors should be at least 95% and even then, that is not even 3rd quartile.

    Response: We changed the hydroprocessing onstream factor to 95% in the final report per reviewer feedback.

81. **Comment:** It is good that relatively recent, real-world cost estimates were used for the hydroprocessing section. Try to include some estimates after the recent super-inflation as a sensitivity case.

    Response: As mentioned in Appendix B, KBR cost estimates were based on the first quarter of 2024. The basis for this report is 2020 U.S. dollars. Known indices, cited in this report, were used to convert to 2020 dollars. The reader may use the same indices to estimate costs for a different cost year.

82. **Comment:** In the previous draft we assumed that beyond a 20% increase in operational burden, the equipment would have to be replaced or reconfigured to handle CFP oil coprocessing or repurposing. A reviewer asked what constraints, such as exotherms etc. dictated this decision. Note that each unit may already be bottlenecked by some particular design choice or history of modifications.

    Response: In the final version we did not include the threshold criteria of 20% mentioned in the comment. We simply subtracted the cost of equipment with the new capacities necessary for coprocessing from the cost estimate of the corresponding existing equipment to estimate additional costs of retrofits. This was done for simplicity and consistency of our methods, and also understanding that some of the potential 20% design leeway in a unit may already be used up by the refinery during different modifications in the unit's lifetime. For simplicity, the scaling variable for each unit (see Table B-2 column "Scaling Unit") was used as the resizing criterion/basis for each unit, understanding that other constraints beyond our chosen scaling variable may become the basis for resizing a specific unit.

83. **Comment:** With the choice of surrogate model compounds, it will be good to understand how well they represent the actual chemistry.

    Response: Details of surrogate compounds for each feed stream are included in Appendix C (for CFP oils) and Appendix D (for SRD, LCO, VGO). Appendix L provides surrogate compounds in the product streams. As noted in the report, we did not focus on matching the

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properties of product streams with fuel specifications, instead we used a consistent H:C ratio of ~1.8 in the overall liquid product after hydroprocessing for consistency across all cases presented, especially to capture hydrogen consumption and exotherms. Note that we did not have sufficient data for developing predictive reaction/reactor models for this report; development of such reaction/reactor models in the future can further enhance the value of similar process modeling exercises. While we defer the capture of reactor chemistry to the future development of predictive reactor models, our response to Comment 20 shows that we captured the energy and mass flow impacts adequately in our models.

84. **Comment:** Why are biomass feedstock and CFP presented here. The main focus is on hydroprocessing, and one might just state that CFP oil is an input for hydroprocessing and has a purchase price.

    Response: We wanted to present a picture for the entire supply chain including biomass feedstock and CFP. Although we leveraged previous work for presenting those sections, we felt it would be important since the biggest potential for improving maturity, increasing supply volumes, and reducing costs for this pathway currently lie in those front-end steps. Other reviewers also recommended that we present a detailed basis for our CFP assumptions.

85. **Comment:** It should be mentioned that hydrogen consumption is higher for CFP oils relative to HDS/HDN in part because (a) oxygen levels are higher than typical S/N levels and (b) because of the chemistry (ring saturation is needed to completely remove phenolic oxygen).

    Response: Content has been added in the final version to address the reviewer's comment.

86. **Comment:** The hydroprocessing base capacity of 75,000 barrels per day is for relatively large units and may not represent many smaller refineries or units. This large capacity may also imply large amounts of CFP oil are needed to coprocess significant percentages in the total feed.

    Response: We acknowledge that the 75K BPD capacity is large. We maintained that for consistency in our process models. However, we have added separate sensitivity cases (Section 3.4.4.12) to explore smaller capacity units as well.

87. **Comment:** Clarify the volume of catalyst assumed. For the coprocessing case this will likely limit the quantity of CFP oil one can include because of heat management and reduce the amount of fresh feed one can incorporate for stand-alone operation (assuming once can use product recycle for thermal management).

    Response: The assumptions regarding WHSV are included in Section (3.2.7) of the final version. Cases showing required curtailment of fresh feed for heat management are also included. A constraints-mitigation-based approach was not the format in which results are

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presented in this report, although we have all the information necessary for such an analysis; this may be the subject of future publications if resources permit.

88. **Comment:** Recycle of liquid with the front-end feed should be included as an option, in addition to inter-bed recycle.

    Response: This has been included in the final report (Section 3.4.4.2).

89. **Comment:** Why was the configuration shown in Section 3.2.4 (for hydrotreating followed by hydrocracking) adopted?

    Response: As noted in the final version, this was adopted primarily because we had some bench-scale experimental data reflecting this 1-stage configuration with two catalysts for hydrotreating and hydrocracking in series. The hydrocracker is intended to crack and convert heavy molecules to fuel blendstocks. This is consistent with refinery operations where fuel-range molecules are recovered through fractionation and heavier cuts are further processed in heavy feed conversion units like fluid catalytic cracking and hydrocracking.

90. **Comment:** In Section 3.2.5 of the final version, it is stated that in many petroleum refining configurations processing heavier feedstocks such as LCO/VGO, the recycled liquid is a heavier cut after distillation that requires further cracking. This was not included in this report. It may be important to understand those recycle effects but understood that currently available data does not allow accurate modeling.

    Response: We will look into this in the future if relevant experimental data becomes available.

91. **Comment:** It was stated in the draft version that when liquid recycle is used for temperature control (L-TC) the hydrogen concentration at the reactor outlet can be specified. This may not be realistic, and it should be determined by process operations.

    Response: This has been clarified in Section 3.2.5 of the final version, stating that this was only a modeling assumption, and that actual operations may need modifications based on reactor and process needs. Note that the model also captures equipment capacity changes (and associated costs) when an assumption of hydrogen concentration for L-TC cases is applied. Our criteria for maintaining a hydrogen flow of approximately four times the chemical hydrogen demand for the L-TC cases is mentioned in Section 3.4.4.2.

92. **Comment:** As with the quality of CFP oil, the petroleum streams used as co-feed will also likely impact stability; one should use caution if extrapolating the initial bench-scale experimental results too far.

    Response: This is a valuable comment, and we are noting it in this Appendix for future reference. Current data sets used as the basis for our models are limited to SRD and

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LCO/VGO blends. This aspect will be further explored as new experimental data becomes available with different biomass derived intermediates and base fossil feedstocks.

93. **Comment:** There are a lot of nuances during catalyst stability testing. Encourage a breakout section or a section in the Appendix to explain more details of experiments, such as time onstream for the tests, conditions run, and % CFP oil tested.

    Response: Details are provided in Appendix A of the final report.

94. **Comment:** While hydrodeoxygenation may inhibit HDS and HDN, the reverse may also be true.

    Response: Per the reviewer's comment, a note has been added that there may be inhibitory impacts and competition among HDS, HDN, and hydrodeoxygenation, and those effects need to be understood through experimental studies.

95. **Comment:** When mentioning the removal of CO_2 with H_2S using an amine system, point out that CO_2 and H_2S do not react the same with all amines.

    Response: Content has been added to the final version per this comment.

96. **Comment:** With respect to controlling CO in the gas stream, some vendors promote conversion of CO to methane to avoid permanent catalyst deactivation.

    Response: The reviewer's comment is noted here.

97. **Comment:** CO_2 may not be totally consumed/removed in the amine system, based on current coprocessing experience. This depends on the quantity of CO_2 produced and the amine capacity.

    Response: A statement has been added in the final version of this report, citing that this may happen depending on the biogenic feedstock, such as for HEFA coprocessing. Inert CO_2 remaining in the fuel gas system (downstream of amine treatment) is considered low risk but remains an important consideration for refinery operations as it will reduce the heating value of the fuel gas supply.

98. **Comment:** Put more emphasis on the CFP oil feeding system.

    Response: We included a CFP oil feed system with a relatively simple design that includes catalyst within enhanced piping (Section 3.2.13). It is also noted that more extensive designs may be necessary depending on the CFP oil and its chemical behavior. On the other extreme, separate and elaborate catalytic stabilization reactors are necessary for non-catalytic pyrolysis oils.

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99. **Comment:** Running CFP oil reduces capacity. Is this opportunity cost accounted for in the analysis?

    Response: The opportunity cost does not appear in the MFSP calculations. However, the downside of reduced capacity is mentioned in discussion content, noting that refineries may have slim margins and reducing processing capacity can impact the bottom-lines of these units. There will be optimization opportunities that can be assessed for specific situations; those cannot be effectively analyzed in a generic setup such as the one presented in this report. Crude and CFP oil prices, associated incentives, processing costs, and other key inputs for the specific location will need to be considered for such optimization analysis.

100. **Comment:** Are different crude costs or margin environments considered?

    Response: We did not consider different crude costs and margin environments. The MFSP is calculated to provide estimated price points for the different feed options, with our underlying assumptions. The modeling framework can be adapted for such analysis, but that was not considered for this report. Only the impact of changing unit costs of hydroprocessing feeds on the MFSP is discussed in this report (Section 3.4.4.15).

101. **Comment:** The cost basis for petroleum feeds, CFP oil, and hydrogen should be included.

    Response: Those are now included in the final version. It is noted that these major unit costs may be prorated by the reader to estimate results with their own assumptions.

102. **Comment:** Are the cost of feed system modifications for CFP oil included in the MFSP?

    Response: The costs are included based on the assumptions described in Section 3.2.13.

103. **Comment:** Pertaining to the text in the report about small refining margins and how they can be affected by the introduction of CFP oil, it will be helpful to see the details of additional hydrogen consumption when CFP oil is introduced at different levels. It would also be helpful to include more details about how the yield vectors change as a function of percent CFP oil cofed.

    Response: Appendix K of the final version includes both hydrogen consumption and cost information for all the cases we studied. As discussed in the report, we do not have sufficient experimental data at this point to understand yield vectors. Key assumptions for yields in our models are included in the report (a) 90% carbon efficiency for CFP oil to hydrocarbons, (b) ~1.8 H:C ratio in the total hydrocarbon fuel product, (c) complete deoxygenation of CFP oil.

104. **Comment:** The MFSP charts at 75k BPD show that capital is an insignificant factor vs. feedstock, which is typical for commodity processes, and especially true for fuel production. Does this imply that the operating configuration matters less than the coprocessed amount, the source of the CFP oil, and other factors? That is what the figures imply.

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Response: That was also our conclusion from this analysis, but sensitivity analysis for lower processing capacities (Section 3.4.4.12) show a larger impact of capital costs on the MFSP, as is expected.

105. **Comment:** One might investigate CI minimization strategies like purge minimization (i.e. allowing H_2 purity to drop w/associated catalyst lifetime impact), purge gas reprocessing, or additional recycle gas purification.

Response: Some of these impacts are presented in sensitivity cases in the final version. A PSA is included within our system analysis boundary to capture and recycle hydrogen; this is accounted in our CI calculations.

106. **Comment:** Since a constant capacity of 75k BPD is maintained in most of the cases even as the WHSV was changed, was reactor volume adjusted to maintain constant processing capacity?

Response: That is correct, the reactor volume was adjusted. In this report we estimated capital costs by scaling using the total flow into the reactor. For cost estimation purposes, this scaling may be done using other process metrics, such as changes in the catalyst volume. In the future we can report results based on constraints of existing equipment, such as using the criteria for maintaining the same reactor volume after curtailing capacity when CFP oil is introduced.

107. **Comment:** It would be helpful to clarify the impact that WHSV has on the unit. Is the assumption a lower throughput at lower WHSV? Are there any yield implications or H_2 uptake impacts assumed for a change in WHSV.

Response: As explained in the final version of the report, we assumed chemical hydrogen consumption as the basis for adjusting the WHSV. This approach was adopted because we currently do not have the necessary experimental data to develop predictive reactor models.

108. **Comment:** One would expect the temperature control problem to be more pronounced when running high CFP oil content feeds. Considering catalyst deactivation due to hot spots/H_2 starvation effects in such situations may be out of scope for this report.

Response: That is correct, we did not consider detailed behavior that may be seen inside the reactor. The process model explores macro aspects. However, one can notice that our model assumptions automatically increase the hydrogen flow based on reaction needs and those larger flows will likely alleviate some of these concerns. These aspects need to be evaluated in detail for any specific installation.

109. **Comment:** The results in the report (earlier draft version) will be more pronounced by running higher proportions of CFP oil in the feed or CFP oils with high hydrogen consumption.

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Response: That is correct, and this is illustrated by the results in the final report.

110. **Comment:** Provide more explanation related to hydrogen consumption, e.g., high purge rate leads to higher H_2 losses for the low H_2 supply purity case.

Response: This is included in the final report (Section 3.4.4.5).

111. **Comment:** Suggest a deeper explanation of the effects of process changes throughout the report, rather than just reporting cost changes. This will lead to more interesting process operational choices to model.

Response: We did not have sufficient time for those write-ups when we prepared the initial draft. Explanations are available in the final version.

112. **Comment:** Discuss the stability challenges of raw fast pyrolysis liquid which leads to the need for producing stable pyrolysis derived intermediates.

Response: We included a review article on the hydrotreatment of fast pyrolysis liquid intermediates in Section 1.1. We did not discuss those challenges in this report, but the interested reader may refer to the review article and other related references.

113. **Comment:** When discussing FCC vs. hydroprocessing mention the distillate/gasoline ratio. Heavier fuels from hydroprocessing may be desirable for supplying heavy-duty vehicles because of the electrification of light-duty vehicles.

Response: We mentioned that FCC products are typically dominated by the gasoline-range fuel (Section 1.1.2).

114. **Comment:** For hydroprocessing catalyst, another adjustment (handle) could be reaction temperature and therefore consequently reduced lifetime.

Response: We discuss catalyst lifetime, and the process impacts of temperature separately as sensitivity cases. Those impacts can be approached concurrently if the relation is understood for specific feedstocks and catalysts based on experimental data.

115. **Comment:** The stabilization reactor is the key for ensuring the two years lifetime of the main hydrotreater. It will handle all coke precursors and inorganic contaminants in the biofeed. This unit operation has not been well defined for CFP oils. It will be important to provide the assumption of this unit. What are the operating cost inputs for the stabilization reactor? This may need to be regenerated or refilled frequently.

Response: Our assumptions regarding the CFP oil feed system are included in Section 3.2.13 of the final report. A minimal design is proposed in that section, but the design may need to be more elaborate based on long-term testing. If necessary, a redundant train will

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need to be added so it can be isolated for regeneration without interrupting the entire hydroprocessing operation. Ultimately the design should be determined after experimental evaluation with specific feedstocks and catalysts in use.

116. **Comment:** For co-hydroprocessing, changes in gas yields and gas composition are important because recirculation of gas is common in hydrotreaters. Changes in gas make would also impact downstream gas stripper design.

    Response: Although a predictive model was not used for yields, the significant differences in gas yields from petroleum feedstocks vs. CFP oils are generally captured in the process models. The modeling framework factors in changes in gas yields when estimating equipment and operating costs for each case.

117. **Comment:** What impact did coprocessing have on conversion? From a refining perspective, maintaining product quality upon co-feeding CFP oils will be an important criteria.

    Response: Quality aspects for diesel and jet fuels are discussed in Section 3.1. Impacts of using CFP oil as hydroprocessing feed are stated, such as the necessity to saturate aromatics to improve the cetane number of the fuel. Impacts of CFP oil coprocessing on product quality and cuts will need to be factored into refinery economics after understanding the chemistry. Maintaining identical product slates will be difficult when high proportions of CFP oils are introduced, but those impacts can be modeled before implementation if experimental data is available. Decision-making will ultimately be guided by overall economics based on policy and demand tradeoffs for specific locations (and many other criteria guiding the overall company business).

118. **Comment:** Regarding hydroprocessing design types (1-stage vs. 2-stage included in the report), one can make the distinction by reactor types instead: fixed bed (trickle or in some cases vapor) or slurry reactors. Almost all industrial reactors are packed with multiple layers of catalyst, with staging included for functions such as treating or cracking, heat management, hydrogen addition etc.

    Response: We included the 1-stage and 2-stage distinction in Section 3.2.1, highlighting primarily that there is no intermediate processing, such as separation of certain products, as the reactants pass through different beds and reactor functions (treating and cracking); intermediate injection of hydrogen and liquid are included within our interpretation of a 1-stage configuration. All hydroprocessing reactors assessed in this project are assumed to be distillate-range units. Based on operating conditions, all reactors are fixed, trickle bed designs. There are no naphtha hydrotreaters (vapor phase) or resid units (slurry reactors) considered in this analysis.

119. **Comment:** In a previous draft, it was stated that single-stage systems may have limited performance. An absolute statement regarding limited performance cannot be made because this depends on the feed stream.

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Response: We removed the "limited performance" from our description of the single-stage system but maintained that they have less operational flexibility compared to two-stage systems.

120. **Comment:** The yield structures in Table A-2 (Appendix A) suggest hydrocracking severity was very low (low conversion). Many refineries use hydrocracking to reduce residue and maximize more valuable streams. Implicit in Section 3.2.1, the two-stage hydrocracking option is portrayed to offer flexibility; however, this flexibility will be limited by refinery-wide optimization (same for design/operational choices for each hydrotreater). Returning to the low severity experimental hydrocracking conditions with CFP oil (Table A-2), was it done deliberately to avoid cracking CFP oils too far? This local choice of conditions may not represent overall optimization strategies adopted by refiners.

    Response: This is a valuable comment. We hope that additional experimental data will allow the development of models that predict reaction behavior when CFP oils are introduced. These predictive reactor models can then be used by refinery optimization models that can determine the optimal choices for refinery operations when CFP oil is introduced. The proportions of CFP oils will greatly impact the optimal choices.

121. **Comment:** Regarding the use of the total condensed liquid product (prior to fractionation) as a potential means for cooling the reactor, it may be more favorable if the post-fractionation heavy product is recirculated instead of the total liquid product.

    Response: As mentioned in the text, the recycle of heavy products is a common industrial implementation. However, we made the simplifying assumption of total liquid product recycle, assuming it is nearly inert after the first-pass of hydroprocessing. This allows easy interpretation of the results for assessing the impact of CFP oil processing without superimposing additional hydrocracking reactions that are typical with the recycle of the fractionated heavy products.

122. **Comment:** If there are many hydrocarbons going with the gaseous feed stream to the amine scrubber per the model outputs there may be accumulation of hydrocarbons with increased surface tension and foaming leading to operational problems.

    Response: Stream(s) 12 in Appendix L show the process model compositions of the amine scrubber feed for four bookend cases (100% CFP oil and 100% petroleum feed). These compositions (with the caveat that these are modeled compositions) may be assessed from a high level for potential operational problems such as foaming. Molecular weight ranges of hydrocarbon species after HP Flash will likely be similar with and without CFP oil feed, although their proportions will usually change (may increase if there are higher amounts of smaller fragmented molecules formed during the removal of oxygen bonds in CFP oil) relative to petroleum feed hydroprocessing. The other big difference will be the presence of large proportions of CO_2 after CFP oil hydroprocessing. As mentioned in the text, CO formation has been reported to be low in the CFP oil bench-scale hydroprocessing

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experiments; the potential presence of CO may be another difference between gases formed from CFP oils and petroleum feed hydroprocessing.

123. **Comment:** Makeup hydrogen is added to cover chemical hydrogen consumption and purge losses. For coprocessing, if there's much CO/CO_2 and light hydrocarbons then the purge can become significant, requiring more makeup hydrogen to manage purity

    Response: The increased purge required during CFP oil hydroprocessing has been captured in the process models which include design specifications to maintain specified purities of hydrogen.

124. **Comment:** The process models assume an H:C ratio of ~1.8 in the total liquid fuel products. Hydrocrackers are usually operated with a measure of conversion of the heavy feed, and not H:C ratio of the products. The H:C choices for hydrotreater and hydrocracker products should have been different, since they operate at different severities and have different chemical functions.

    Response: In practice these assumptions will need to change, especially when specific situations and associated strategies have been identified. The combinations of cases with the different hydroprocessing feed options considered in this report require different and complicated assumptions after careful analysis, especially in tandem with any experimental data available. Since we attempted to generalize and present high-level impacts in this report it was important for us to have the simplified consistent assumptions included for this analysis. We do present some sensitivity cases to show the impacts of changing some of our assumptions. Our simplified assumption reflects hydrogen and other resources required for the ultimate conversion of all feedstocks (including CFP oil and petroleum) to usable products; in practice this may be achieved after multi-pass recycle of the heavier fractionated molecules in the petroleum feed to a hydrocracker. Repeating from the report, the recycle of unconverted hydrocarbons after fractionation of the hydrocracker products was not included since we did not have predictive reactor models that could be adapted to reflect reaction behavior after the introduction of CFP oil feed. We believe our simplified assumptions allow easy interpretation of the impacts of the introduction of CFP oils with existing petroleum feed, delineated from the complexities of individual hydroprocessing operations.

125. **Comment:** Process safety would require options for rapid depressurization of the hydrocracker and significantly more thermometry to avoid runaway exothermic reactions. Does scaling up for higher pressure ratings account for these requirements?

    Response: The cost prediction correlations with ranges included in Appendix B capture a broad range of operating conditions, configurations, and operations. Some of these requirements should be captured by those methods and correlations.

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126. **Comment:** Hydroprocessing unit operators prefer to change temperature rather than space velocity when making operational adjustments. Adjustments in space velocities result in de-rating the unit, and there isn't usually spare "space velocity".

Response: We considered space velocity adjustments since de-rating is an important option when oxygenated feeds are introduced, resulting in higher hydrogen demand and associated higher exotherm. We appreciate the reviewer's comment and understand that de-rating decisions will be taken at a higher level, but unit operators will likely continue to use temperature adjustments (within allowable ranges) as the primary means to control reactor performance after baseline operational parameters have been determined.

127. **Comment:** The simplifying assumption for calculating costs by splitting the costs evenly between the hydrotreating and hydrocracking reactors (mentioned in Section 3.2.7) will be different in practice. These are two different reactors with different catalysts and hydrogen consumptions.

Response: We agree and have noted this comment here. This is a simplifying assumption and can be revisited when we have more data to provide additional granularity for the reactor system. Our preliminary bench-scale experiments for hydrotreating + hydrocracking (that we used as a potential basis) were configured with two sequential catalyst beds with no intermediate operations between the catalyst beds.

128. **Comment:** Regarding the minimal implementation of the modified feed system in Section 3.2.13, miscibility and thermal stability evaluations will be important for scale-up. Depending on the service, fouling may already be a concern with an existing feed, and it will likely be exacerbated by addition of CFP oil.

Response: We agree that the system shown in Section 3.2.13 is minimal. More elaborate systems, including stabilization reactors may be necessary for more reactive feedstocks, especially if attempting to hydroprocess non-catalytic fast pyrolysis liquids. As the reviewer pointed out, interactions with existing feed will also need to be assessed.

129. **Comment:** Most of the metrics are presented on an energy basis. Refiners use volume basis (not energy) for road fuels, aviation fuel, etc. It would be useful to have volumetric information presented side-by-side.

Response: We used energy basis to avoid potential errors that often get introduced with liquid density calculations. To provide an idea of the metrics corresponding to average volume bases for different fuels we provided the MFSP metrics in dollars per GGE, GJE, and GDE in the results tables (Appendix K). We agree with the reviewer's comment that refiners make choices based on market conditions, but we did not pick specific market drivers for determining price differentials among different fuels; we instead used an energy basis to determine the MFSP of the entire fuel pool.

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130. **Comment:** As with the MFSP sensitivity with the on-stream factor, the MFSP changes with WHSV does not capture the economic impact on the refinery. Changes in WHSV are associated with changes in throughput and capacity curtailment which affects the bottom line of the refinery.

    Response: We added text to point out that the MFSP metric does not fully capture the economic impact for the refiner in these sensitivity cases.

131. **Comment:** Regarding the stated basis for hydrogen consumption and WHSV corrections: chemical hydrogen consumption in the 100% SRD hydrotreater model, 615 scf/barrel of fresh SRD feed with a corresponding WHSV of 2 h^{-1} was used as the basis. The 615 scf/barrel is relatively low and a WHSV of 2 h^{-1} is on the high side.

    Response: We understand that these values can change with the petroleum feedstock chosen, as shown in Table 28. We picked the relatively mild hydrotreating case with SRD as the basis, although we could have picked a higher hydrogen consumption with lower WHSV based on another petroleum feedstock, as shown in Table 28. We chose the SRD case because it was already modeled within this same framework and we could ensure consistency when we analyzed other feedstock combinations.

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