PNNL-36528

Hydrothermal Processing of Wastewater Solids (HYPOWERs) Project Preliminary Techno-Economic Analysis

September 2024

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U.S. DEPARTMENT OF ENERGY

Prepared for the U.S. Department of Energy
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Executive Summary

A techno-economic analysis was conducted for a conceptual full-scale n^{th} plant for hydrothermal processing (HTP) of Central Contra Costa Sanitary District's waste sludge. The HTP plant includes hydrothermal liquefaction (HTL) and catalytic hydrothermal gasification (CHG) of the HTL aqueous phase. Figure ES.1 gives a summary of the resulting process economics for the HTP plant. The minimum biocrude selling price (MBSP) at which the net present value of the project is zero is $2.77/gasoline gallon equivalent (gge) or $3.00/gallon, including the avoided current cost sludge disposal (incineration) and renewable fuel credits from federal Renewable Identification Number (RIN) credits and California's Low Carbon Fuel Standard (LCFS) credits. Estimated renewable fuel credits are based on an assumed 60% reduction in carbon intensity relative to petroleum, a D3 RIN price of $1.48 and an LCFS credit price of $85.8/tonne CO_2-e avoided. RIN and LCFS prices are based on two-year market averages, 80% discounted to reflect after-trading net revenue. Inclusion of renewable fuel credits and avoided cost of sludge incineration (-$146/dry ton) reduces the MBSP by $4.06/gge and $1.70/gge, respectively.

Sensitivity analysis around key assumptions for the HTP plant reveals that the MBSP is very sensitive to renewable fuel credit prices, carbon intensity reduction, avoided cost of current sludge treatment, and economic assumptions including internal rate of return, Lang factor, and labor cost. Evaluation of two alternative configurations, the first where co-product methane gas is burned on-site for heat, and the second where the aqueous phase from hydrothermal liquefaction is recycled directly to the headworks, results in MBSPs of $3.82/gge and $1.37/gge, respectively. Preliminary analysis of pioneer plant costs suggests an increase in capital investment of 75% and decrease in average on-stream factor of 9% relative to the n^{th} plant analysis, resulting in a MBSP of $7.21/gge. Preliminary comparative analysis with anaerobic digestion (AD) suggests that HTP results in revenue approximately 4 times higher than AD where biogas is used for combined heat and power, and approximately 40% higher than AD where biogas is sold as transportation fuel.

Biocrude is assumed to be converted to transportation fuel blendstocks at a separate (stand-alone) upgrading plant. Figure ES.2 gives a summary of the process economics for the upgrading plant. The upgrading plant is envisioned to be in a centralized location that can collect biocrude from multiple sources in a region and therefore takes advantage of economies of scale. Co-processing at a petroleum refinery is also an opportunity that is currently being explored. The minimum fuel selling price (MFSP) of the combined upgraded fuel blendstocks is $3.83/gge ($4.10/gal diesel; $3.78/gal naphtha).

Biocrude from Sludge Hydrothermal Processing - Nth Plant Analysis

Primary/Secondary Wastewater Treatment Sludge (20 wt% solids; 12% ash)

Sludge Credit/Cost: -$146 $/U.S. ton sludge (dry, ash free basis)

		w/out RINs & LCFS
Minimum Selling Price, $/gge	**$2.77**	$6.83
Minimum Selling Price, $/gal	**$3.00**	$7.38

Renewable Fuel Credit Prices Assumed (60% GHG reduction)

D3 RIN Price, $/RIN:	1.48
LCFS Credit, $/tonne CO2e:	85.8

Biocrude Production	1.37	million gge/yr
	159	billion Btu/yr
	102	gge/US ton AFDW sludge
	11,839	million Btu/US ton AFDW sludge
Internal Rate of Return (After-Tax)	10%	
Equity Percent of Total Investment	40%	
Cost Year	2017	

CAPITAL COSTS

Sludge Dewatering	$900,000	4%
HTL Oil Production	$14,000,000	55%
CHG	$9,600,000	38%
Balance of Plant	$1,000,000	4%
Total Installed Capital Cost	$25,500,000	100%
Building, site development, piping	$4,800,000	
Indirect Costs	$18,200,000	
Fixed Capital Investment (FCI)	$48,500,000	
Working Capital	$2,400,000	
Land (plant located at WWTP)	$0	
Total Capital Investment (TCI)	$50,900,000	
Installed Capital / Annual gge Biocrude	$18.6	
TCI per Annual gge Biocrude	$37.1	
Loan Rate	8.0%	
Term (years)	10	
Capital Charge Factor (computed)	0.128	

MANUFACTURING COSTS

Plant Hours per year	7,920	
Feed rate, dry sludge	46	ton/day
Feed rate, dry ash-free sludge	41	ton/day
Feed rate, slurry	36	gal/min

	$/gge biocrude	$/year	$/gal
Avoided sludge disposal cost	-1.70	-$2,300,000	-1.84
Natural Gas	0.22	$300,000	0.23
Catalysts and Chemicals	0.52	$700,000	0.57
Waste Disposal	0.25	$300,000	0.27
Electricity and other utilities	0.47	$600,000	0.51
CHG Gas Co-Product	-0.18	-$200,000	-0.19
Renewable Fuel Credits	-4.06	-$5,600,000	-4.38
Fixed Costs	2.52	$3,400,000	2.72
Capital Depreciation	1.17	$1,600,000	1.26
Average Income Tax	0.35	$500,000	0.38
Average Return on Investment	3.21	$4,400,000	3.47
	2.77		**3.00**

RINs and LCFS Credits Breakdown

	$/gge biocrude	$/year	$/MMBtu
Biocrude RINs	-2.19	-$3,003,011	-18.87
CHG Gas RINs	-0.87	-$1,194,085	-19.77
Biocrude LCFS	-0.71	-$976,220	-6.13
CHG Gas LCFS	-0.28	-$388,174	-6.43
Total Credits	-4.06	-$5,600,000	-51.20

PERFORMANCE

Net Electricity Purchased (KW)	534
Net Electricity Purchased (KWh/gge product)	3.1
(Energy in Biocrude) / (Electricity+Natural Gas)	3.2
Overall Carbon Yield to Biocrude	
On sludge + natural gas	68%
On sludge	72%

Figure ES.1. Process economics summary of the HTP plant.

Liquid Fuels from Biocrude Upgrading - Nth Plant Analysis

Biocrude Feedstock Cost:	$2.86	$/gge biocrude (includes transport cost)
Minimum Fuel Selling Price (MFSP)	**$3.83**	**$/gge**
Diesel Fuel Selling Price	$4.10	$/gal
Naphtha Fuel Selling Price	$3.78	$/gal

Naphtha	Diesel	Total	
663	1985	2,600	BPSD
9.1	29.5	39	million gge/yr
1.1	3.4	4.5	trillion Btu/yr, LHV basis
0.22	0.73	0.95	gge fuel/gge biocrude

Internal Rate of Return (After-Tax)	10%
Equity Percent of Total Investment	40%
Cost Year	2017

CAPITAL COSTS

Hydrotreating	$32,900,000	44%
Hydrocracking	$6,500,000	9%
Hydrogen Plant	$26,900,000	36%
Steam cycle	$1,600,000	2%
Balance of Plant	$6,400,000	9%
Total Installed Capital Cost	$74,300,000	100%
Building, site development, add'l piping	$12,200,000	
Indirect Costs	$51,900,000	
Working Capital	$6,900,000	
Land (included in feedstock cost)	$2,600,000	
Total Capital Investment (TCI)	$148,000,000	
Installed Capital per Annual GGE Fuel	$1.9	
TCI per Annual GGE Fuel	$3.8	
Loan Rate	8.0%	
Term (years)	10	
Capital Charge Factor (computed)	0.159	

MANUFACTURING COSTS

Plant Hours per year	7920	
Biocrude feed rate	37	mmgal/y
	$/gge fuel blendstock	$/year
Biocrude	3.00	$115,700,000
Natural Gas	0.05	$1,800,000
Catalysts & Chemicals	0.01	$500,000
Waste Disposal	0.002	$100,000
Electricity and other utilities	0.02	$1,000,000
Fixed Costs	0.25	$9,600,000
Capital Depreciation	0.002	$4,600,000
Average Income Tax	0.04	$1,400,000
Average Return on Investment	0.45	$17,500,000
	3.83	

PERFORMANCE

Net Electricity Purchased (KW)	1,637
Electricity Produced Onsite (KW)	1,812
Electricity Used (KW)	3,449
Net Electricity Purchased (KWh/gge product)	0.3
Overall Carbon Yield (Naphtha + Diesel)	
On biocrude + natural gas	83%
On biocrude	89%

Figure ES.2. Process economics summary of the biocrude upgrading plant.

Contents

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Tables

1.0 Introduction

A techno-economic analysis (TEA) was conducted for a conceptual commercial-scale plant for hydrothermal processing (HTP) of waste sludge from the Central Contra Costa Sanitary District (CCCSD) waste water treatment plant (WWTP). The HTP process includes hydrothermal liquefaction (HTL) and catalytic hydrothermal gasification (CHG) of the HTL aqueous phase. Data from bench scale testing is used as the basis of Aspen process modeling and the resulting mass and heat balances for the modeled HTP and upgrading plants. Experimental data is presented in Section 2.0 and includes compositional analysis of the sludge mixture, and process testing conditions and results from sludge HTL, CHG of the HTL aqueous phase, and hydrotreating of the sludge-derived biocrude. The HTP and biocrude upgrading operations are modeled as separate plants that are not geographically co-located (see Figure 1). This assumption is consistent with the previously published design case (Snowden-Swan et al. 2017) and takes advantage of economies of scale for the upgrading plant by collecting biocrude at a centralized location. Co-processing at a petroleum refinery is also an opportunity that is being pursued.

1.1 Conversion Plant Model Overview

Figure 1 shows the overall block flow diagram for the conversion of wastewater treatment plant (WWTP) sludge to fuel blendstocks via HTP and biocrude upgrading. The modeled scale for the HTP plant is 46 dry ton sludge per day, corresponding to the CCCSD's average daily sludge production rate. The HTP plant is assumed to be co-located with the WWTP to avoid the cost of transporting sludge. While collection of sludge from nearby WWTP facilities is possible and could conceivably benefit economics by increasing the HTP plant scale, this scenario is not considered here. Future analysis is needed to better understand the cost tradeoffs between HTL scale and transporting sludge between WWTP facilities. The HTL biocrude is assumed to be shipped to a larger scale upgrading plant where it is combined with biocrude from multiple WWTP/HTL plants for catalytic hydroprocessing and product fractionation into naphtha and diesel fuel blendstocks. The modeled upgrading plant scale is 2,700 barrel per day of biocrude feed, consistent with the design case established in earlier work (Snowden-Swan et al 2017). The plant produces 2,650 barrel/day of diesel and naphtha range fuel blendstock. The aqueous phase from HTL is treated with CHG to produce co-product methane for transportation fuel. This stream contains high levels of nitrogen, primarily in the form of ammonium ion, which may need removal prior to recycling back to the WWTP. Alternatives for sidestream nitrogen removal are currently being investigated.

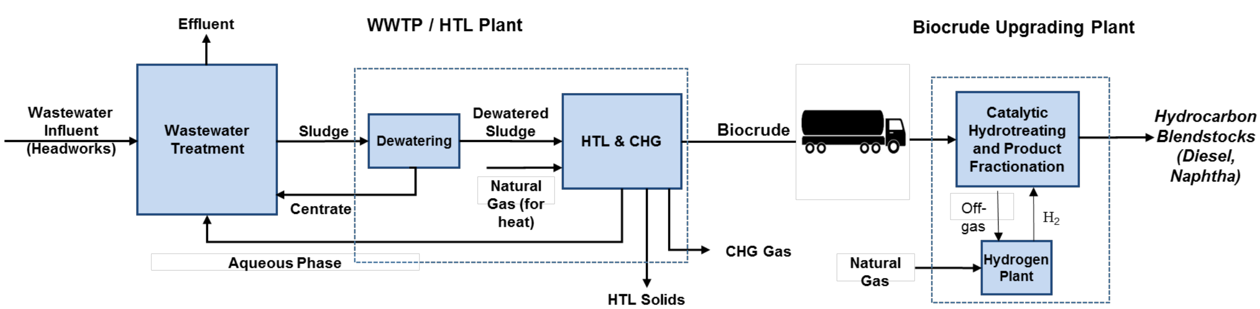


Figure 1. Sludge HTP and biocrude upgrading.

1.2 Techno-Economic Analysis Methods

The approach to developing conversion process techno-economics is similar to that employed in previous analyses conducted for the Bioenergy Technologies Office (BETO) (Dutta et al. 2015; Jones et al. 2013, 2014; Tan et al. 2015). Process flow diagrams and models are based on experimental results from completed and ongoing research, as well as information from commercial vendors for mature and similar technologies. To assure consistency across all biomass conversion pathways, BETO has developed a set of economic assumptions that are used for all TEAs and are documented in BETO's Multi-Year Program Plan (DOE 2016). An important aspect of these assumptions is that they reflect an "n^{th} plant" design, as described below. For this work, a preliminary pioneer plant analysis was also performed for the HTP plant using literature methods. In addition, renewable fuel credits are estimated for the biocrude/blendstocks and CHG gas and are included in the economics.

1.2.1 Nth Plant Analysis

A standard reference basis used in previous design reports, known as the "n^{th}" plant design, is used for all BETO TEAs, thus allowing a consistent basis for comparison of different technologies within the context of a well-defined hypothetical plant. These assumptions do not account for additional costs that would normally be incurred for a first-of-a-kind plant, including special financing, equipment redundancies, large contingencies and longer startup times necessary for the first few plants. For n^{th} plant designs, it is assumed that the costs reflect a future time when the technology is mature and several plants have already been built and are operating. The specific assumptions are shown in Table 1.

Table 1. Nth-plant economic assumptions.

Assumption Description	Assumed Value
Internal rate of return (IRR)	10%
Plant financing debt/equity	60% / 40% of total capital investment (TCI)
Plant life	30 years
Income tax rate	21%
Interest rate for debt financing	8.0% annually
Term for debt financing	10 years
Working capital cost	5.0% of fixed capital investment (excluding land)
Depreciation schedule	7-years MACRS[a] schedule
Construction period	3 years (8% 1st yr, 60% 2nd yr, 32% 3rd yr)
Plant salvage value	No value
Start-up time	6 months
Revenue and costs during start-up	Revenue = 50% of normal Variable costs = 75% of normal Fixed costs = 100% of normal
On-stream factor	90.4% (7,920 operating hours per year)

(a) Modified accelerated cost recovery system

1.2.2 General Cost Estimation Basis

All costs in this report are on a 2017 constant dollar basis. The annual Chemical Engineering Plant Cost Indices (CE Index) from Chemical Engineering magazine are used to convert capital costs to 2017 dollars. Bureau of Labor Statistics indices for average hourly earnings of production workers (Series ID: CEU3232500008) and producer price for chemical manufacturing (Series ID: PCU325---325---) are used to convert labor rates and chemical prices, respectively, to 2017 dollars. The indices can be found in Appendix D.

Capital costs are estimated from a variety of resources. The heat and material balances generated by the simulation software (Aspen Plus [AspenTech 2013]) are used to size the major pieces of equipment. These are used to either scale vendor quotes or input to Aspen Capital Cost Estimator (ACCE) software to determine the installed capital cost. In addition, costs from the literature are used when necessary.

The original cost reflects the year of the cost quote or estimate, and the scale of the equipment. All capital costs are adjusted to an annualized 2017 basis using the Chemical Engineering (CE) magazine's published indices:

$$\text{Cost in 2017 \$} = \text{equipment cost in quote year} \times \left(\frac{\text{2017 index} = 567.5}{\text{quote cost year index}}\right)$$

The scale is adjusted to match the appropriate scaling term (heat exchanger area for example) by using the following expression:

$$\text{Scaled equipment cost} = \text{cost at original scale} \times \left(\frac{\text{scale up capacity}}{\text{original capacity}}\right)^{n}$$

where 'n' is the scale factor, typically, 0.6 to 0.7.

Once the equipment is scaled and adjusted to the common cost year, factors are applied to calculate the total capital investment. Individual installation factors calculated by ACCE are multiplied to equipment costs, unless installed costs are already available from vendors. The total direct cost is the sum of all the installed equipment costs, plus the costs for buildings, additional piping and site development. Indirect costs are estimated as 60% of the total installed costs. Factors for the calculation of these additional direct and indirect costs are listed in Table 2. The sum of the direct and indirect costs is the fixed capital investment (FCI). The total capital investment is the fixed capital plus working capital and land costs.

Table 2. Direct and indirect cost factors for capital investment estimation.

Direct Costs	
Item	% of Total Installed Cost (TIC)
Buildings	4.0%
Site development	10.0%
Additional piping	4.5%
Total Direct Costs (TDC)	**18.5%**
Indirect Costs	
Item	% of TDC
Prorated expenses	10%
Home office & construction fees	20%
Field expenses	10%
Project contingency	10%
Startup and permits	10%
Total Indirect Costs	**60%**
Working Capital	5% of FCI

Operating costs are estimated using the results from the Aspen Plus heat and material balances and applying the assumptions shown in Section 2.0. For the cooling tower, it is assumed that water is available at 90 °F with a 20 °F allowable temperature rise.

1.2.3 Pioneer Plant Analysis

The nth plant analysis described in Section 1.2.1 is the standard assumption used for BETO's biofuel pathway TEAs. However, additional costs such as longer start-up times and equipment redundancies can be expected for projects involving first-of-a-kind plants. Empirical equations developed by Merrow et al (1981), and applied to nth plant costs in Anex et al (2010), Huang et al (2017) and Tao et al (2017), can be used as a first estimate of the cost impacts of first-of-a-kind plants. The equations were developed from a statistical analysis of cost estimation error in pioneer plants (Merrow et al 1981). The first equation describes the cost growth for the actual capital relative to the estimated capital investment for a new facility:

$$\begin{aligned} Cost\ Growth &= 1.1219 - 0.00297\ x\ Pctnew - 0.02125\ x\ Impurities - 0.01137\ x\ Complexity \\ &+ 0.00111\ x\ Inclusiveness - 0.06361\ x\ Project\ Definition \end{aligned}$$

The independent variables, *Pctnew*, *Impurities*, *Complexity*, *Inclusiveness*, and *Project Definition* in the above equation are defined in Table 3, along with the estimated values for the HTP plant. The resulting Cost Growth factor is then used to predict the pioneer plant capital cost relative to the nth plant capital cost as:

$$\text{Pioneer Plant Cost} = \frac{\text{Nth Plant Cost}}{Cost\ Growth}$$

Table 3. Cost growth equation variable definition and values assigned for HTP plant.

Variable	Definition	Range	Value for HTP Plant
Pctnew	Percentage of capital cost of commercially undemonstrated equipment	0-100	41
Impurities	Impurities buildup and corrosion issues	0-5	4
Complexity	Number of continuously linked process steps	≥0	6
Inclusiveness	Percentage of 3 factors included in cost estimate: pre-startup personnel costs, pere-startup inventory cost, and land purchase	0-100	33
Project Definition	Levels of site-specific information and engineering included in estimate (basic plant layout, process flow conditions, definition of major equipment)	2-8 (2 is max definition)	5
Cost Growth (Result)	Ratio of estimated to actual costs	≥0	0.57

In addition to the cost growth factor, Merrow et al (1981) developed an equation to predict the shortfall in production after the second six months after startup that is often experienced with first-of-a-kind plants:

$$Plant\ Performance = 85.77 - 9.69\ x\ Newsteps + 0.33\ x\ Baleqs - 4.12\ x\ Waste - 17.91\ x\ Solids$$

Plant Performance is defined as the actual average production in months 7 to 12 after start-up as a percentage of plant design capacity. The independent variables, *Newsteps*, *Baleqs*, *Waste* and *Solids* in the above equation are defined in Table 4 and values assigned to each of the variables for the HTP plant are given. For this analysis, the calculated *Plant Performance* factor is assumed for the first year of operation, with a 20% production growth each year until the plant reaches an on-stream factor of 90% (the Nth plant assumption), as shown in Figure 2. The average of the values over the lifetime of the plant is then used as the overall on-stream factor used in the DCFROR calculation.

Table 4. Plant performance equation variable definition and values assigned for HTP plant.

Variable	Definition	Range	Value for HTP Plant
Newsteps	Number of process units that incorporate technology unproven in commercial use	≥0	3
Baleqs	Percent of heat and mass balance equations based on actual data from prior plants	0-100	0%
Waste	Assessment of difficulties with waste handling encountered during development	0-5	2
Solids	Designates that a plant processes primarily solid feedstocks of products	0 or 1	1
Plant Performance (Result)	Actual average production in months 7 to 12 after start-up as a percent of plant design capacity production	0-86%	31%

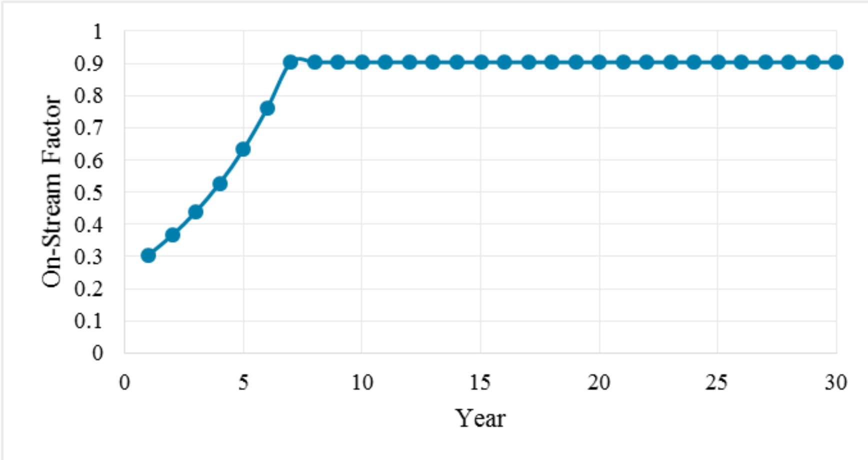


Figure 2. Annual plant performance for pioneer plant analysis.

1.2.4 Calculation of Renewable Fuel Credits

This analysis considers the potential impact of renewable fuel credits on the economics of the sludge-to-fuel blendstock pathway economics. Both renewable identification number (RIN) credits under the EPA's Renewable Fuel Standard (RFS) and credits under California's Low Carbon Fuel Standard (LCFS) are included for the CHG gas and combined fuel blendstock from the biocrude. Credits are determined based on the trading prices of RFS and LCFS credits and the greenhouse gas (GHG) intensity of the product fuels. As the GHG assessment has not yet been finalized, it is assumed that the fuels produced from the pathway result in a 60% GHG reduction from petroleum, which would qualify them for D3 RINs (aka, cellulosic biofuel). Preliminary analysis suggests that this assumption is reasonable. The TEA should be adjusted after the GHG analysis is finalized. The estimated renewable fuel credits are reflected in the production price of the biocrude and carried forward in the effective feedstock price to the centralized upgrading plant.

1.2.4.1 Renewable Fuel Standard (RFS)

The RFS program was put into effect as a requirement of the Energy Independence and Security Act of 2007 and is described in 40 CFR 80, Subpart M. Credits under the RFS are awarded via renewable identification numbers (RINs). From 40 CFR 80.1415, the number of gallon-RINs for a renewable fuel is calculated as

$V^{RIN} = EV * V_S$ [§ 80.1426 paragraph (f)(2)]

Where EV = equivalence value for a gallon of the renewable fuel

V_S = the standardized volume of the renewable fuel @ 60F

The equivalence value (EV) of the renewable fuel is calculated as

EV = (R/0.972) x (EC/77000) [§ 80.1415 paragraph (c)(1)]

Where R = renewable content of the fuel

EC = energy content of the renewable fuel, in Btu/gal (LHV)

The equivalence values for the naphtha and diesel fuel blendstocks from biocrude upgrading and for the CHG gas were calculated using the above equation. The naphtha/diesel blend has an estimated EV of 1.6 RIN/gal, using a combined blendstock lower heating value (LHV) of 121,934 Btu/gal from the Aspen model. Because the EV must be calculated based on energy content in Btu/gal, an adjusted value using these units was needed for the CHG gas. The CHG gas LHV is 670 Btu/scf (Aspen model results), which is 0.68 the heating value of natural gas (LHV of 983 Btu/scf from GREET 2018). The RFS states that 77,000 Btu of CNG shall represent 1 gallon of renewable fuel with an EV of 1. Therefore, compressed CHG gas is estimated to have a heating value proportional to that of CNG at 53,000 Btu/gal, with a resulting EV of 0.7 RIN/gal (13.4 RIN/MMBtu). It is assumed that the fuel products have at least a 60% reduction in carbon intensity compared to petroleum, which will qualify them for D3 RINs (cellulosic biofuel). A two-year average D3 RIN price, discounted 20% to account for the actual after-trading net revenue, was used for the baseline case and was varied in the sensitivity analysis by +/- 50%. Sensitivity analysis varying the GHG reduction from 50% to 70% is also included, using a two-year average D5 RIN price (also 20% discounted).

1.2.4.2 Low Carbon Fuel Standard (LCFS)

The state of California's LCFS program is described in the California Code of Regulations (CCR), Title 17 LCFS credits are calculated, in metric tonnes of CO_2-eq avoided relative to petroleum fuel use, by the following equation:

Credits = $(CI_{standard} - CI_{renewable}) \times E_{displaced} \times C$ [§ 95485 paragraph (a)(3)]

where $CI_{standard}$ is the carbon intensity of either gasoline or diesel in g CO_2-e/MJ (according to Tables 1 and 2 of the regulation

$CI_{renewable}$ is the carbon intensity of the alternative fuel in g CO_2-e/MJ

$E_{displaced}$ is the total amount of gasoline or diesel fuel energy displaced in MJ and includes

C is the conversion factor from grams to tonnes (1×10^{-6})

The credits can be calculated in tonnes CO_2-e avoided per gallon of fuel by using units of MJ/gal for $E_{displaced}$ (LHV of the fuel). As in the RIN calculation, a two-year average LCFS credit price ($/tonne CO2-e), discounted 20% to reflect the after-trading net revenue, was used for the baseline case and varied in the sensitivity analysis. The CI of the petroleum baseline was estimated by taking weighted average (based on the diesel and naphtha yields of the upgrading plant) of the 2016 CI values for petroleum diesel and gasoline from the compliance tables of the regulation, resulting in a CI of 94.7 g CO2-e/MJ. The CI of the fuel blendstocks and CHG gas from the process is 60% reduced from the petroleum baseline CI, or 56.84 g CO2-e/MJ. The CI was varied from 50% to 70% GHG reduction in the sensitivity analysis.

2.0 Process Design and Cost Estimation

Key experimental results used in the process model include compositional analysis and HTL processing of the CCCSD sludge, CHG of HTL wastewater from previous work, and sludge biocrude hydrotreating from previous work. The experimental data and the primary process model assumptions are presented here.

2.1 Sludge Feedstock Composition

The modeled feedstock is a mixture of sludges from the primary (solids settling) and secondary (biological) treatment stages of the CCCSD plant. Table 5 shows ultimate analysis for the CCCSD sludge, as received. CCCSD adds lime (CaO) to their process in order to facilitate better sludge incineration, which results in a 17.4% ash content. However, implementation of the HTP process would eliminate the need for this lime. The adjusted sludge composition, without lime addition, assumed in the model is 11.6%.

Table 5. Ultimate analysis of CCCSD wastewater sludge, as received.

Component	CCCSD wt% dry basis	Model wt% dry basis	Model wt% dry, ash free basis
C	43.3	45.1	51.1
H	6.3	6.6	7.4
O	30.2	31.5	35.6
N	4.5	4.7	5.3
S	0.4	0.5	0.5
Ash	16.7[(a)]	11.6	
P	2.5		

(a) Central San currently treats their wastewater with lime to aid their incineration process. Ash content of their mixed sludge without lime addition is estimated at 11.3%.

2.2 Sludge Dewatering

Prior to processing, the sludge is dewatered from 3% to 20% solids. Dewatering to this level serves to minimize capital and operating costs, while still ensuring pumpability. Figure 3 shows a simple diagram of the dewatering step, where the mixed raw primary and secondary sludge feedstock is dewatered in a centrifuge. Polymer is added at a rate of 20 lb/ton dry sludge, commensurate with CCCSD's current dewatering operations. The dewatered sludge is then routed to the HTL process.

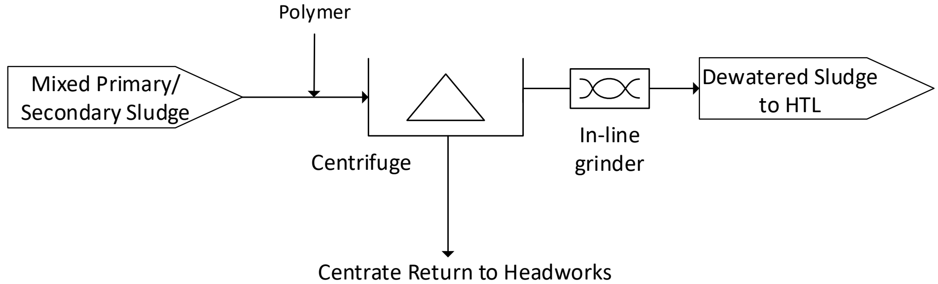


Figure 3. Process flow for slurry feed preparation.

The capital cost for the centrifuge and in-line grinder is based on vendor budgetary estimates. The installed equipment cost for the centrifuge and grinder is $0.95 million (2017 $) for a 15,350 gal/hr feed capacity.

2.3 Sludge Hydrothermal Liquefaction (HTL)

The main process steps in the HTL section of the plant are shown in Figure 4. The 20% solids slurry feed is pumped to 2900-3000 psia and then preheated to 550°F (288°C). Heat recovered from the HTL reactor effluent via a hot oil system is used to bring the slurry up to reaction temperature (656°F; 347°C), similar to the pilot scale system configuration. The hot oil system was selected over direct exchange of the high pressure feed/product as it is likely more economical due to the large surface area needed for sludge heating. Further work is needed to optimize the economics of the heating and cooling steps, including exploring staged heating at lower pressures. Off-gas from the HTL reactor and purchased natural gas are combusted for heat in the hot oil system. An electrically heated hot oil system is also plausible. The heated, pressurized slurry is fed to the HTL reactor where the contents are converted to an organic biocrude phase, an aqueous phase, solids and a small amount of gas. The HTL reactor effluent is fed to a hot filter where solids, consisting of 60-70% water, ash, char, and low levels of organics, are separated. The biocrude tends to adhere to the solid particles and therefore the amount lost to the solids depends upon the ash content in the feed. The solids are assumed to be disposed of in a landfill. An attractive alternative to landfilling is to recover value from the HTL solids, either in the form of separate nutrients (e.g., phosphorus) or a combined fertilizer product. Further work is needed to validate the technical and economical feasibility of this option. After solids separation, the remaining biocrude-aqueous-gas mixture is cooled to 140°F (60°C) and separated in a three-phase separator. The biocrude product is stored and then shipped to a centralized plant where it is upgraded into fuel blendstocks. The aqueous stream, which contains effluent water, any remaining soluble organics, ammonia, and metal salts, is routed to the CHG section of the plant.

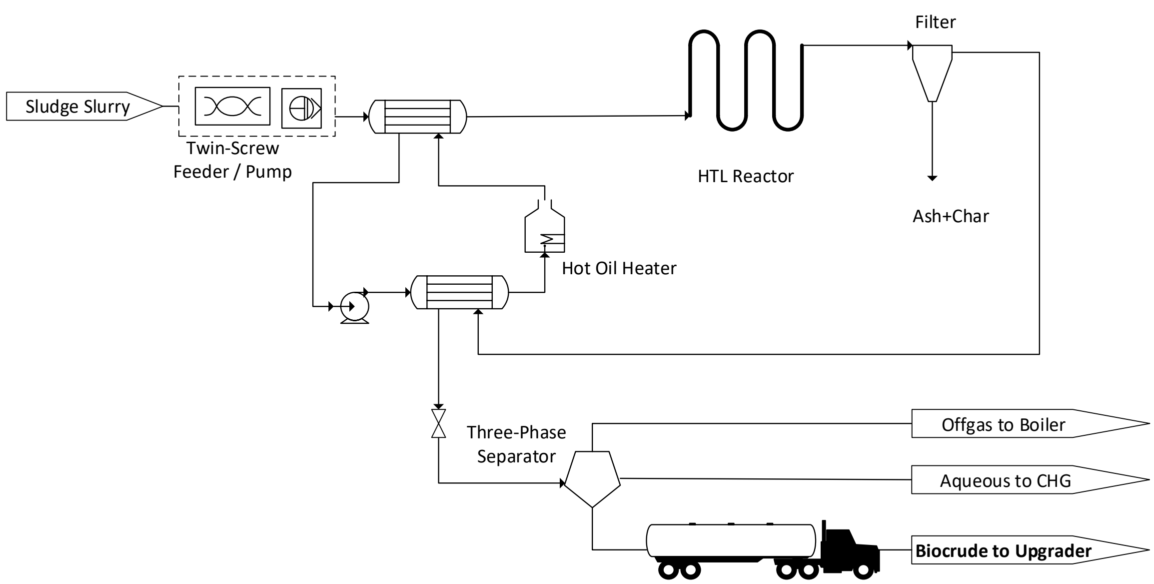


Figure 4. Process flow for sludge HTL.

Table 6 lists the experimental testing conditions and product results for CCCSD sludge and the parameters used in the process model. The experimental data were collected from PNNL's bench scale system similar to previous studies (Elliott et al. 2013; Marrone 2016).

Table 6. Sludge HTL experimental results and model assumptions.

Operating Conditions and Results	Experimental Results (WW09)	CCCSD Model
Temperature, °F (°C)	654 (346)	656 (347)
Pressure, psia (MPa)	2975 (19.8)	2979 (20.5)
Feed solids, wt%		
Ash included	17.4%	20.0%
Ash-free basis	14.5%	17.7%
LHSV, vol./h per vol. reactor	3.6 Hybrid CSTR-PFR[d]	3.6 PFR
Equivalent residence time, min.	17	17
Product yields[a] (dry, ash free sludge), wt%		
Oil (biocrude)	37.3%	37.5%
Aqueous	34.3%	33.3%
Gas	23.1%	21.9%
Solids	5.4%	7.4%
Carbon yields, wt%		
Oil (biocrude)	52.1%	57.0%
Aqueous	29.3%	27.9%
Gas	12.2%	12.5%
Solids	6.4%	2.6%
HTL dry biocrude analysis, wt%		
C	77.6%	77.7%
H	9.9%	10.8%
O	6.8%	6.2%
N	5.2%	4.7%
S	0.4%	0.6%
P	0.0	Not modeled[b]
Ash	0.07%	0.0%
HTL dry biocrude molar H:C Ratio	1.53	1.65
HTL biocrude dry HHV, Btu/lb (MJ/kg)	16,359 (38.0)[c]	16,290 (39.4)
HTL biocrude moisture, wt%	4.0 wt%	4.0 wt%
HTL biocrude wet density @25°C (g/ml)	0.99	0.99
Aqueous phase COD (mg/L)	75,200	92,000
Aqueous phase NH3 (mg/L)	3,100	4,000
Aqueous phase TC (mg/L)	26,800	31,800
Aqueous phase TOC (mg/L	25,100	28,000

(a) Recovered after separations.
(b) Phosphorus partitioning is not directly modeled in Aspen because of the small quantity, most of which reports to the solid phase.
(c) Calculated using Boie's equation (Boie 1953).
(d) The experimental system includes a continuous flow stirred-tank reactor (CSTR) followed by a plug-flow reactor (PFR). The CSTR helps prevent overheating of the feed.

Costing of equipment in the HTL area is from scaling of vendor equipment quotes (Merrick), Aspen Capital Cost Estimator v. 8.8, and scaling of equipment from Knorr et al (2013). Scaling of equipment follows general engineering guidelines (Peters et al. 2003) and does not consider the cost implications of numbering up with additional modular units. Additional work is needed to evaluate the impact of modular scale up. Installation factors from Knorr et al. (2013) were applied to vendor-quoted equipment. The capital costs for each of the major process components of the HTL section are given in Table 7. Individual equipment costs, scaling factors and installation factors can be found in Appendix B.

Table 7. Sludge HTP capital costs.

Item	Purchased, million USD (2017)	Installed, million USD (2017)	Source
HTL Reactor System	3.96	8.73	Vendor quotes, Knorr et al. 2013
Phase separation	1.68	3.19	ACCE, Knorr et al. 2013, vendor quote
Hot oil system for reactor and trim heater	1.90	2.06	ACCE; vendor quote
Total	**7.5**	**14.0**	

2.4 Catalytic Hydrothermal Gasification (CHG)

Figure 5 shows the main process steps in the CHG section of the plant. As shown, the aqueous phase from HTL is treated with CHG to recover energy from the dissolved organics and reduce the chemical oxygen demand (COD) of the water being returned to CCCSD's headworks. The HTL aqueous stream is heated with the hot CHG reactor effluent to 580F in a direct feed/product exchanger, and then to 665F with a trim heater. Table 8 lists the CHG reactor conditions and product results from the experimental data and the process model. The product gas (67% methane) is sold as coproduct renewable natural gas for transportation fuel. The treated CHG water is recycled back to the headworks of CCCSD. This sidestream contains elevated nitrogen levels that may or may not be problematic for CCCSD's biological treatment step. The nitrogen levels in this sidestream are not unique to HTL and must be managed regardless of the sludge treatment technology that is used (e.g., anaerobic digestion). The necessity for nitrogen removal/recovery from the CHG water is currently under investigation.

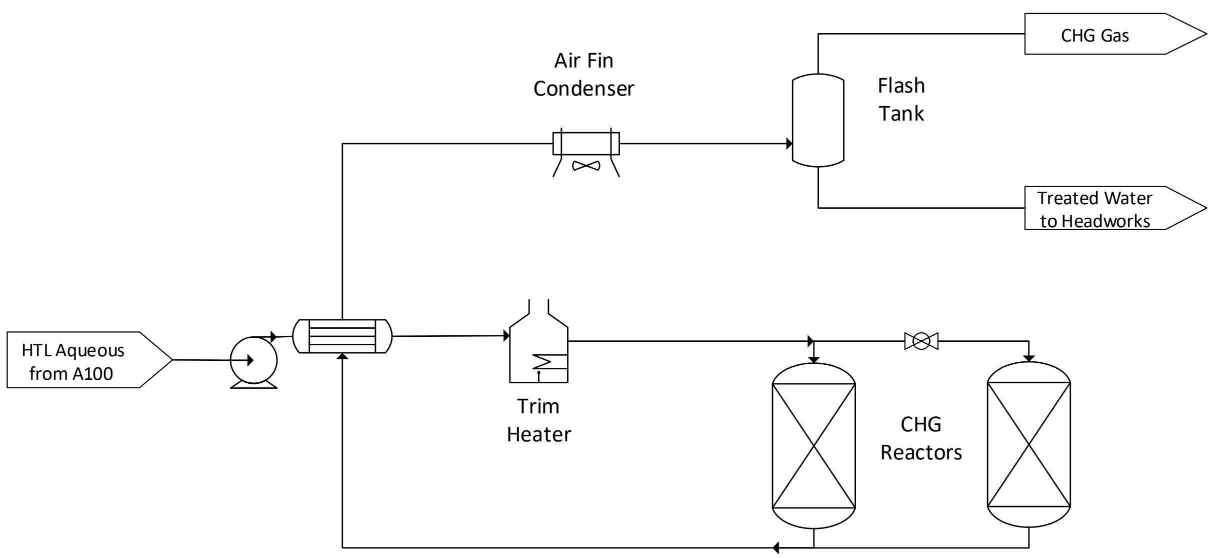


Figure 5. Process flow for CHG.

Table 8. HTL aqueous phase CHG experimental results and model assumptions.

Component	Experimental (WERF 02)	Model
Guard Bed	Raney nickel	None – assume catalyst regeneration
Temperature, °F (°C)	638 (337)[(a)]	662 (350)
Pressure, psia	2890	3079
Catalyst	7.8 wt% Ru/C	7.8% Ru/C
LHSV, vol./hour per vol. catalyst	2.0	2.0
WHSV, wt./hr per wt. catalyst	3.6	3.6
% COD conversion	99.9%	94.3%
% Carbon to gas [(b)]	61%	85%
Gas analysis, volume %		
CO_2	28.6%	25.7%
H_2	3.1%	1.5%
CH_4	66.9%	67.0%
C_{2+}	--	1.0%
N_2+O_2	1.5%	--
Water	--	4.7%
COD of CHG treated water (mg/L)	55	Low

(a) Average of reactor inlet and outlet temperatures
(b) Organic carbon to gas. Note that the remaining converted carbon is dissolved bicarbonate.

Experimental testing was conducted with a raney nickel guard bed upstream of the main Ru/C bed to remove sulfur components. However, raney nickel is not effective for removal of sulfate, the primary sulfur component in the HTL aqueous phase. Therefore, regeneration of the Ru/C catalyst using dilute peroxide solution and washing is currently being explored in the laboratory and is assumed for the analysis. To accommodate the regeneration scenario, an extra catalyst bed is used in the capital equipment estimation. It is assumed that each reactor is taken offline every three days and regenerated while the other is online. Initial regeneration testing in the laboratory was carried out using the methods of Dreher et al (2014), with 5 bed volumes of 3% hydrogen peroxide solution for oxidation of the Ru and 5 bed volumes of wash water to complete the regeneration treatment.

Costing of purchased equipment in the CHG area is from Aspen Capital Cost Estimator v. 8.8, vendor quotes (Merrick) and Knorr et al (2013). The capital costs for each of the major process components of the CHG section are listed in Table 9. Individual equipment costs as well as scaling assumptions and installation factors can be found in Appendix B. Note that the energy and cost of purifying and compressing the CHG gas into a compressed natural gas (CNG) for transportation fuel is not included in the analysis.

Table 9. CHG capital costs.

Item	Purchased, million USD (2017)	Installed, million USD (2017)	Source
CHG System	4.56	9.24	ACCE, Knorr et al. 2013; vendor quote
Phase separation	0.18	0.35	ACCE
Total	**4.7**	**9.6**	

2.5 Biocrude Upgrading

The HTL biocrude is transported to a centralized upgrading facility and is supplied to the plant at 26 psia and 110°F. The heart of the upgrading process is hydrotreating, shown in Figure 6. The biocrude feed is pumped to 1540 psia, mixed with compressed hydrogen, and preheated to the hydrotreater reactor temperature of 752°F (400°C). Hydrogen is produced onsite via steam reforming of the upgrading offgas and purchased natural gas. During the hydrotreating process, biocrude oxygen is converted to CO_2 and water, nitrogen is converted to ammonia, and sulfur is converted to hydrogen sulfide. The reactor effluent is cooled to condense the produced water and hydrocarbons, the latter of which is then fractionated into lights, naphtha, diesel and heavy oil. Heavy oil is sent to the hydrocracker to produce additional naphtha and diesel blendstocks. Hydrotreator reactor conditions and product results from the experimental data and the model are listed in Table 10. Hydrocracking and other sections of the upgrading plant are presented in detail in Appendix A.

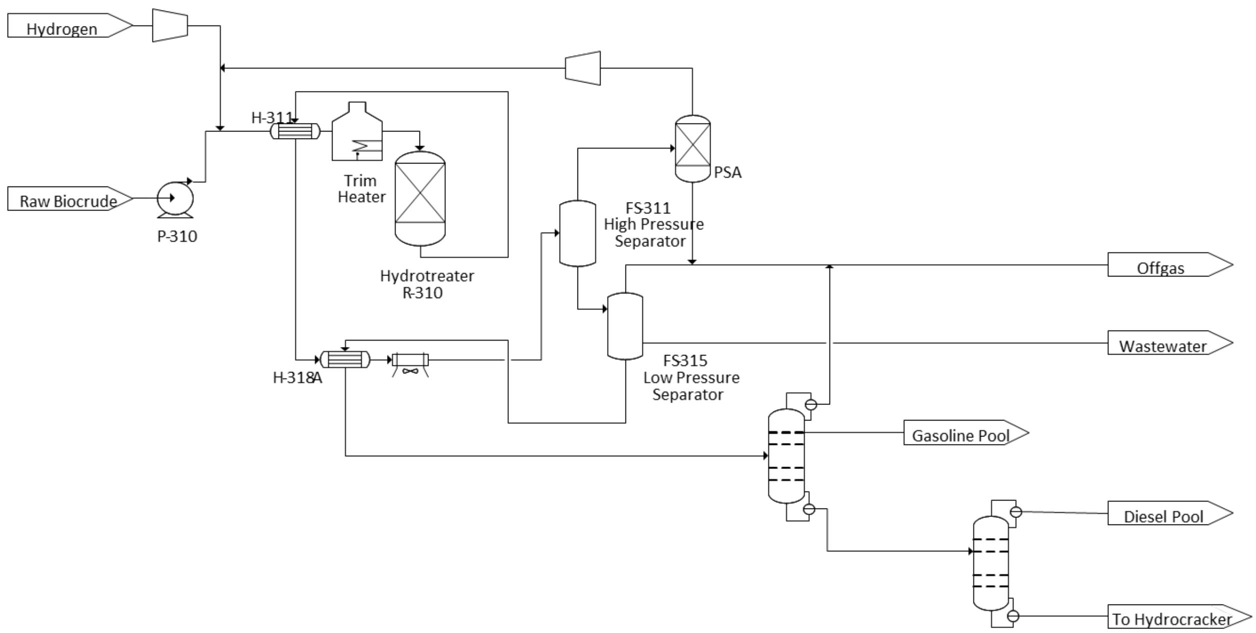


Figure 6. HTL biocrude hydrotreating process diagram.

Previous bench scale hydrotreating testing of biocrude (from Detroit sludge) conducted at PNNL is used as the basis of the modeled biocrude upgrading process. Hydrotreating of the HTL biocrude is in development, specifically with respect to improving catalyst lifetime, reactor space velocity, and the hydrotreated yield. For the purposes of this study, however, it is assumed that the upgrading process has been improved for these parameters.

Table 10. Sludge biocrude hydrotreating experimental results and model assumptions.

Component	Experimental (HT-62005-60)	Model
Temperature, °F (°C)	752 (400)	752 (400)
Pressure, psia	1540	1515
Catalyst[a]	CoMo/alumina	CoMo/alumina
Sulfided?	Yes	Purchased presulfided
LHSV, vol./hour per vol. catalyst	0.11	0.50
WHSV, wt./hr per wt. catalyst	0.18	0.81
HTL biocrude feed rate, ml/h	5.6	Commercial scale
Catalyst life	302 hours (demonstrated)	2 years
Chemical H_2 consumption, wt/wt HTL biocrude (wet)	0.046	0.044
Product yields, lb/lb dry biocrude (vol/vol wet biocrude)		
Hydrotreated oil[b]	0.82 (0.99)	0.84 (0.97)
Aqueous phase	0.14 (0.13)	0.13 (0.19)
Gas	0.08	0.07
Product oil, wt%		
C	85.6%	85.3%
H	14.6%	14.1%
O	1.0%	0.6%
N	<0.05%	0.04%
S	7-10 ppm	0.0%
Aqueous carbon, wt%	0.10%	0.2%
Gas analysis, volume%		
CO_2, CO	0%	0%
CH_4	51%	33%
C_2+	49%	38%
NH_3	Not measured	26%
NH4HS	Not measured	3%
TAN, feed (product)	59 (<0.01)	Not calculated
Viscosity@40 °C, cSt, feed (product)	400 (2.7)	Not calculated
Density@40 °C, g/ml, feed (product)	0.98 (0.79)	0.98 (0.79)

(a) Fixed bed reactor includes 40% (vol) guard bed of CoMo whole extrudate
(b) Yield after phase separation

Figure 7 shows the boiling point curve from simulated distillation (ASTM Method D2887) of the hydrotreated product from sludge-derived biocrude. The curve was used to match the modeled product boiling point curve and distribution as closely to the testing results as possible, as shown.

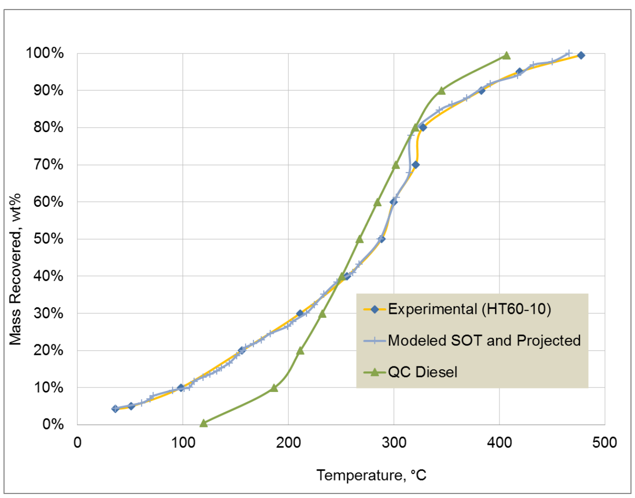


Figure 7. Boiling point distribution (ASTM D2887) for hydrotreated product from sludge biocrude.

Capital costs for the sludge biocrude upgrading plant are scaled from previous work (Zhu et al. 2014; Snowden-Swan et al. 2017). The purchased capital and installed capital cost for the main processing sections of the plant are given in Table 11. Additional cost details can be found in Appendix B.

Table 11. Capital costs for main sections of biocrude upgrading plant.

Item or Area	Purchased, million USD (2017)	Installed, million USD (2017)	Source
Hydrotreater system (2732 BPSD feed)	20.65	32.88	IHS 2014a
Hydrocracker system (1020 BPSD feed)	4.32	6.52	IHS 2014a
Hydrogen Plant (9.6 mmscf/day)	14.03	26.93	HIS 2014b

3.0 Process Economics

The process economic analysis involves first determining the total capital investment (TCI), the variable operating costs, and the fixed operating costs for each of the HTP and biocrude upgrading plants. Discounted cash flow rate of return analysis is then used to determine the minimum biocrude selling price (MBSP) and minimum fuel selling price (MFSP) of blendstocks from the biocrude upgrading, using standard methodology used for all BETO TEA studies. Sensitivity analysis around key technical and economic assumptions is also presented.

3.1 HTP Plant

Table 12 lists the feed and biocrude product flow rates for the sludge HTP plant. The overall mass balance from the process model is shown in Figure 8.

Table 12. Annual feed and product rates for the sludge HTP plant.

Stream	Million gallons/year	Million lb/year
Dry sludge (12% ash) feed	N/A	30.4
Dry, ash-free sludge feed	N/A	26.9
Total slurry feed (20% solids)	17.1	152.0
HTL biocrude product	1.27	10.1
CHG gas product	90.1 (scf/year)	5.9
HTL solids (wet)	N/A	15.5

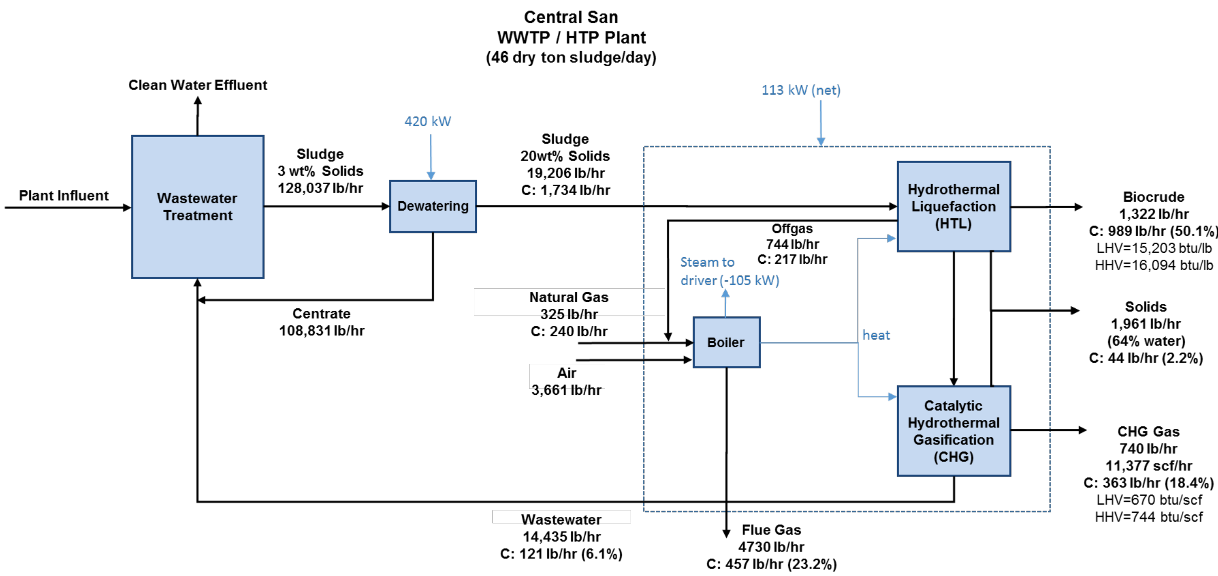


Figure 8. Mass and carbon flows for HTP plant.

3.1.1 Total Capital Investment

Table 13 summarizes the costs presented in Section 2.0 for the HTP plant, including the balance of plant items such as the steam system, tank farm, flare and cooling water system. The total capital investment for the project is \$50.9 million.

Table 13. Total capital investment for sludge HTP plant.

Cost Item	Million US Dollars (2017$)
Sludge dewatering	0.9
HTL biocrude production	14.0
CHG water treatment	9.6
Balance of plant	1.0
Total Installed Cost (TIC)	**25.5**
Buildings (4% of TIC)	1.1
Site development (10% of TIC)	2.5
Additional piping (4.5% of TIC)	1.1
Total Direct Costs (TDC)	**30.3**
Indirect Costs	
Prorated expenses (10% TDC)	3.0
Home office & construction fees (20% TDC)	6.1
Field expenses (10% TDC)	3.0
Project contingency (10% TDC)	3.0
Startup and permits (10% TDC)	3.0
Total Indirect	**18.2**
Fixed Capital Investment (FCI)	**48.5**
Working Capital (5% of FCI)	2.4
Land[a]	0.004
Total Capital Investment (TCI)	**50.9**

(a) Scaled on Dutta et al. 2011

3.1.2 Operating Costs

Variable operating costs and supporting assumptions for the sludge HTL plant are given in Table 14. The largest contributors to annual operating cost are electricity (66% of which is for dewatering), polymer, solids disposal, and natural gas. Extra aeration power or other operating costs that would be necessary for treating the nitrogen in the CHG water recycle are not included in the analysis.

Table 14. HTP plant variable operating costs.

	Quote Year Cost		Year of Price Quote	MM$/yr (2017)	Source
Raw Materials					
Sludge feedstock cost	-146	$/dry ton	2015	-2.33	CCCSD
Polymer for dewatering sludge	1.24	$/lb	2017	0.38	CCCSD
Natural Gas	4.90	$/1000scf	2016	0.30	CCCSD
Boiler Chemicals	2.266	$/lb	2007	0.00029	Humbird 2011
poCooling Tower Chemicals	1.36	$/lb	2007	0.00056	Humbird 2011
70% H2O2 Solution for Cat Regen	0.48	$/lb	2014	0.09	PEP
CHG Catalyst	58	$/lb	2014	0.25	Estimated
Subtotal				**-1.32**	
Waste Disposal					
HTL Solid Waste	42.38	$/ton	2015	0.35	BACWA
Subtotal				**0.35**	
Utilities					
Water makup	0.22	$/metric ton	2001	0.01	Humbird 2011
Electricity	15	¢/kwh	2017	0.63	CCCSD
Subtotal				**0.64**	
Total Variable Operating Costs				**-0.33**	
Total w/Out Feedstock Credit				**2.0**	

Fixed costs for the HTL + CHG plant are shown in Table 15. Salaries are taken from Dutta et al. (2011) and converted to a 2017 dollar basis using US Bureau of Labor Statistics labor cost indices. The factors for benefits and maintenance, insurance, and taxes are the standard assumptions used for BETO design cases.

Table 15. HTL plant fixed operating costs.

Position Title	Number	Total Cost (2017), million USD/year
Conversion Plant (unburdened)		
Plant Manager	1	0.17
Plant Engineer	1	0.08
Maintenance Super	1	0.06
Lab Manager	1	0.06
Shift Supervisor	3	0.16
Lab Technician	1	0.05
Maintenance Tech	1	0.05
Shift Operators	4	0.22
Yard Employees	1	0.03
Clerks & Secretaries	1	0.04
Subtotal		**0.92**
Overhead & maintenance	90% of labor & supervision	0.82
Maintenance capital	3% of TIC	1.37
Insurance and taxes	0.7% of FCI	0.34
Total Other Fixed Costs		**3.45**

3.1.3 Co-Product, RINs and LCFS Credits

Table 16 gives the results of the revenue for the HTP plant from the sale of CHG gas and from RINs and LCFS credits associated with the biocrude and CHG gas. The CHG gas is included in the analysis as a co-product of the process, assuming it can be sold at a commodity price equal to natural gas (per MMBTU). The renewable fuel credits are reflected in the production price of the biocrude and carried forward in the feedstock price to the centralized upgrading plant (including $0.9/gge for transportation). Annual revenue from credits for the biocrude and CHG gas are $4.0 million and $1.6 million, respectively. The overall CHG gas selling price, including commodity price and renewable credit value, is $34/mmBtu.

Table 16. Annual revenues from CHG gas and renewable fuel credits (with a 60% GHG reduction).

Co-product and Renewable Credits				MM$/yr (2017)
CHG Gas - commodity price[(a)]	4.01	$/MMBTU	2017	-0.24
Biocrude RIN Credit[(b)]	1.48	$/D3 RIN	2018	-3.00
Biocrude LCFS Credit[(c)]	85.8	$/tonne CO2-e	2018	-0.98
CHG RIN Credit[(b)]	1.48	$/D3 RIN	2018	-1.19
CHG LCFS Credit[(c)]	85.8	$/tonne CO2-e	2018	-0.39
Total Credits				**-5.42**

(a) 2017 average citygate price, EIA 2018.
(b) Two-year average $1.85/D3 RIN (EPA 2018), 80% discounted.
(c) Two-year average $107.2/tonne CO_2-e avoided (Neste 2018), 80% discounted.

3.1.4 Minimum Biocrude Selling Price

The minimum biocrude selling price (MBSP) is determined using a discounted cash flow rate of return (DCFROR) analysis, which takes into account the time value of money. The MBSP is the plant gate selling price of the biocrude that makes the net present value of the project equal to zero with a 10% discounted cash flow rate of return over a 30 year plant life and 40% equity with the remainder debt financed at 8% interest for a 10 year term (see Table 1, Section 1.2). The resulting MBSP for the 46 dry ton/day sludge HTL plant is $2.77/gge biocrude, or $3.00/gallon. For more details on the DCFROR calculation, see Appendix C. The modeled biocrude has a lower heating value (LHV) of 125,480 Btu/gal. A LHV of 116,090 Btu/gal (ANL 2018) is used to convert the heat value of the fuel products to a gasoline gallon equivalent basis. Table 17 shows the contribution of costs making up the MBSP.

Table 17. Minimum biocrude selling price cost breakdown.

	$/gge biocrude	$/year	$/gal
Avoided Sludge Disposal Cost	-1.70	-$2,300,000	-1.84
Natural Gas	0.22	$300,000	0.23
Catalysts and Chemicals	0.52	$700,000	0.57
Waste Disposal	0.25	$300,000	0.27
Electricity and Other Utilities	0.47	$600,000	0.51
CHG Gas Co-Product (commodity price)	-0.18	-$200,000	-0.19
Renewable Fuel Credits[1]	-4.06	-$5,600,000	-4.38
Fixed Costs	2.52	$2,400,000	2.72
Capital Depreciation	1.17	$1,600,000	1.26
Average Income Tax	0.35	$500,000	0.38
Average Return on Investment	3.21	$4,400,000	3.47
Total MBSP	**2.77**		**3.00**
Total MBSP w/out Renewable Credits	**6.83**		**7.38**

Figure 9 shows the contribution of each of the areas of the HTP plant and credits to the MBSP. Renewable fuel credits offset the production price by $2.90/gge from the sale of biocrude and by an additional $1.16/gge from the sale of CHG gas, assuming the pathway meets a 60% reduction in GHGs from the petroleum baseline. Inclusion of the avoided cost of CCCSD's current incineration process for treatment of their sludge reduces the MBSP by $1.70/gge.

Sensitivity analysis around key financial and technical assumptions and their impact on the MBSP is presented in Figure 10. Each bar on the tornado chart represents a single-point sensitivity in the parameter that is varied. As shown, MBSP fluctuates significantly with the uncertainty or variability associated with several assumptions. An increase of $2.55/gge is seen for the case where GHG reductions for the fuels only meet 50%. This is because the credits would only qualify for D5 RINs, which are a much lower value than D3 RINs (average price of $0.33/D5 RIN vs. $1.48/D3 RIN, discounted 80%). Also, the LCFS credit is reduced by 10%. A 70% reduction in GHG reductions results in only a slightly decreased MBSP. The reason for the assimetry in this sensitivity is because a 70% GHG reduction compared to the base case (60% reduction) gives a slight increase in LCFS credits (which are measured corresponding to the exact carbon intensity), but no further benefit is realized with regard to RINs credits (60% reduction gives the maximum credit). Related to this is the sensitivity in RINs and LCFS price, which can be volatile over time. A ±50% change in RIN and LCFS price gives a ±$2.03/gge price variation. Varying IRR by 50% impacts the MBSP by about ±$1.58/gge, relative to the n^{th} plant assumption of 10% IRR. Dry, ash-free percentage yields for biocrude have ranged from the low 30s to a high of 44 in experimental testing for WWTP sludge (Snowden-Swan et al. 2017), which impacts the MBSP by +$1.93/gge to -$0.96/gge. The project Lang factor, defined as the fixed capital investment (FCI) divided by the purchased equipment cost, is 3.8 for the base case. When the factor is varied from 3 to 5, the MBSP varies by -$1.02/gge to $1.56/gge. The inclusion of the avoided cost of incinerating sludge, CCCSD's current treatment method, assumes that the plant would be willing to continue to pay this much to an HTP owner/operator to take its sludge. If the sludge was given away ($0/dry ton credit), the MBSP would increase by $1.70/gge. The capital for HTL and CHG equipment was varied from -10 to +25% of the base case, resulting in a -$0.55 to $1.37/gge impact on the MBSP. Plant economics are sensitive to other assumptions included in Figure 10, but to a less significant degree than those at the top of the tornado chart.

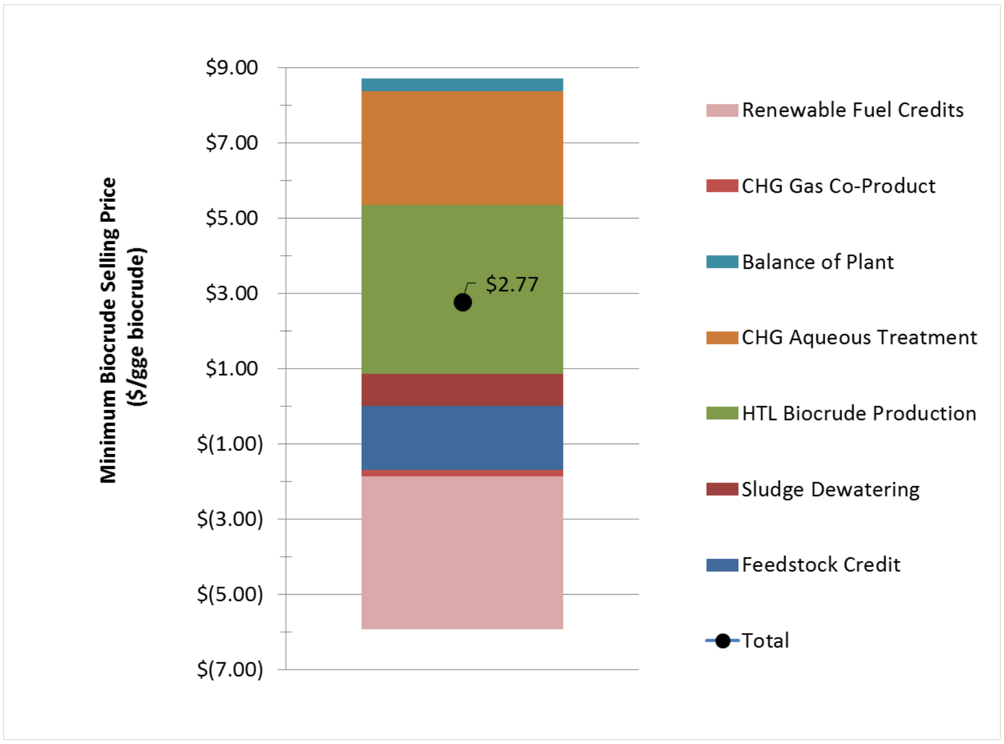


Figure 9. Sludge HTL biocrude production cost for goal case (renewable credits - 60% GHG reduction).

Economics for other possible HTP plant configurations are listed for comparison to the base case in Table 18. These include: 1) burning the CHG gas on-site for heat and power, and; 2) an HTL-only case, where the HTL aqueous phase is sent directly back to the CCCSD's headworks. The case where CHG gas is used on-site for heat results in a higher MBSP of $3.82/gge due to a reduction in RINs and LCFS credits from the CHG gas. The HTL-only case results in a reduced MBSP of $1.37/gge due primarily to a 40% reduction in capital cost (including removal of the steam system), while still receiving most of the revenue from the renewable fuel credits associated with the biocrude.

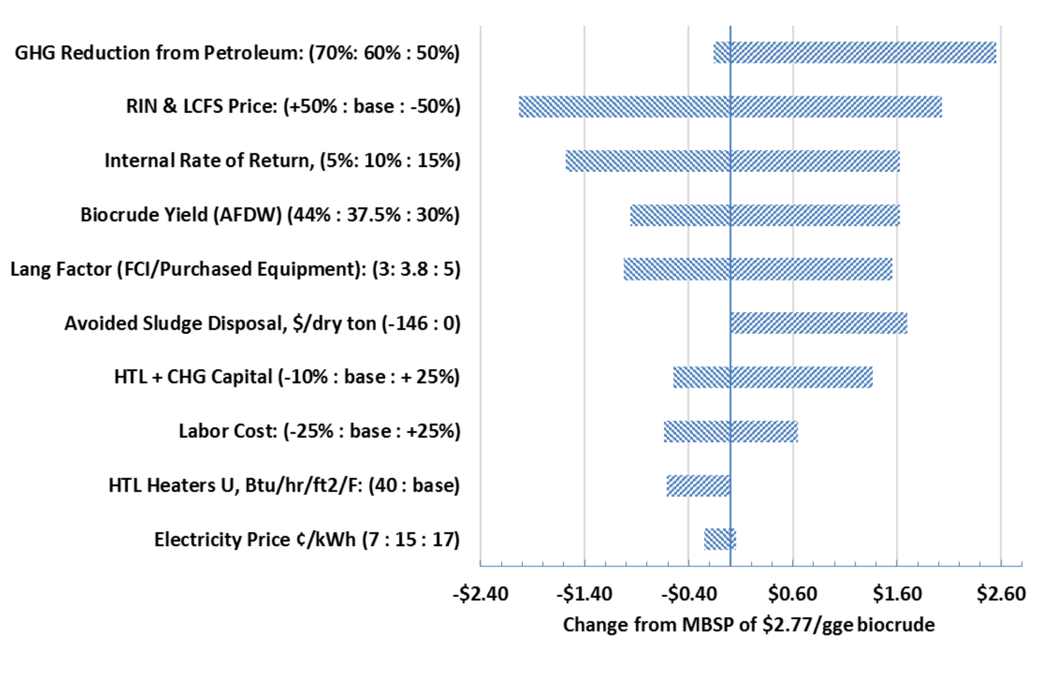


Figure 10. Sensitivity of MBSP to process and economic assumptions.

Table 18. Comparison of economics for alternative HTP plant configurations.

	HTP Base Case (CHG gas sold)	HTP (CHG Gas burned on-site)	HTL Only[(b)]
Sludge feed rate, dry ton/day:	46	46	46
Biocrude Yield, MM GGE/yr	1.4	1.4	1.4
CHG Gas Yield, MM scf/yr	90	0	0
Capital Costs, $million (2017 USD)			
Sludge Dewatering	0.9	0.9	0.9
HTL	14.0	14.0	14.0
CHG	9.6	9.6	0
Balance of Plant	1.0	1.0	0.5
Total Installed Capital	**25.5**	**25.5**	**15.4**
Total Capital Investment	**50.9**	**50.9**	**30.8**
Operating Costs, $million/yr (2017 USD)			
Catalyst and Chemicals	0.7	0.7	0.3
Natural Gas	0.3	0	0.1
Waste Disposal	0.3	0.3	0.3
Electricity	0.6	0.6	0.4
Fixed Costs	3.4	3.4	2.8
Total O&M Cost	**5.3**	**5.0**	**4.3**
Avoided Cost of Incineration	2.3	2.3	2.3
Revenue from Biocrude	3.8	5.2	1.9
Revenue from CHG gas	0.2	0	0
Revenue from RINs and LCFS[(a)]	5.6	4.0	4.0
MBSP	$2.77/gge	$3.82/gge	$1.37/gge

(a) Assuming 60% GHG reduction for all scenarios.
(b) Steam cycle removed; does not include extra cost for processing BOD at WWTP.

3.1.5 Preliminary Pioneer Plant Analysis

Results of the preliminary pioneer plant analysis are presented in Table 19. The results suggest that the capital investment for a pioneer plant is 75% greater than the estimated n^{th} plant cost. The average on-stream factor (over the 30-year life of the plant) is estimated to be 9% less than the n^{th} plant performance. This analysis is preliminary and is intended to provide a rough approximate of the impact of building a first-of-a-kind plant on HTP process economics. Additional industry data is needed, preferably from similar projects (e.g., biorefineries), to conduct a more detailed analysis and comparison with n^{th} plant assumptions.

Table 19. Pioneer plant vs. nth plant economics for the HTP process.

Parameter	Pioneer Plant	N^{th} Plant
Cost Growth (n^{th} plant / pioneer plant)	0.57	n/a
Plant Performance in months 7-12, %	31%	90%
Average on-stream factor, %	82%	90%
TCI, $MM	89.7	50.9
MBSP, $/gge	7.21	2.77

3.1.6 Comparison of HTP with Anaerobic Digestion

Anaerobic digestion (AD) is a conventional sludge treatment technology that is in broad use today and is attracting attention because of improved economics associated with selling the biogas as renewable transportation fuel. As such, comparison of HTP with AD as a sludge treatment alternative is of interest. A preliminary cost estimation was made for AD for comparison to HTP. Costs for AD were derived from a detailed project feasibility study and proposal conducted for the city of Ann Arbor, Michigan (State of Michigan 2007). Equipment costs from the study were scaled from a capacity of 28 dry ton/day of sludge feed to the CCCSD scale using a 0.6 exponent and escalated to 2017 using the CE Indices. Operating costs were scaled proportionally and harmonized with the HTP case to provide as consistent a basis for comparison as possible. For the purposes of this comparison, the biocrude selling price from HTP was assumed to be equal to the the average 2017 first purchase price for crude oil at $1.14/gal (EIA 2018), or $1.0/gge.

Table 20 compares the economics of the HTP plant and AD plant, with and without CHP. The case with CHP was the original presented in the study, whereas the case without CHP was derived from removing the capital cost of the CHP equipment, subtracting the power credits, and including the revenue from the sale of biogas for transportation fuel. The installed equipment costs are relatively similar between the HTP and AD w/CHP plants, however, because of the revenue from biocrude and renewable fuel credits, the sludge treatment profit (negative cost) for HTP is about 4 times that of AD. When the biogas from AD is sold as renewable natural gas (consistent with the HTP case), AD profit per ton of sludge compares more favorably with HTP, but is still about 30% lower than HTP.

Table 20. Preliminary cost comparison of HTP and AD.

	HTP		AD w/ CHP		AD w/out CHP	
Biocrude, gge/yr	1,371,001					
Methane Gas, MMBtu/yr	60,385					129,577
Capital Costs, $MM						
Purchased Equip.	$	13.2	$	12.0	$	10.1
Installed Equip.	$	25.5	$	18.0	$	15.1
FCI	$	48.5	$	43.5	$	36.5
Operating Costs, $/yr						
Materials and Utilities	$	1,650,000	$	(1,270,000)	$	710,000
Solids Disposal	$	350,000	$	1,190,000	$	1,190,000
Labor	$	3,450,000	$	1,200,000	$	1,200,000
Total Operating	**$**	**5,450,000**	**$**	**1,120,000**	**$**	**3,100,000**
Revenues, $/yr						
Revenue Biocrude (@ $1/gge)	$	1,370,000	$	-	$	-
Revenue Methane	$	240,000	$	-	$	520,000
Revenue RINs	$	4,200,000	$	-	$	2,250,000
Revenue LCFS	$	1,360,000	$	-	$	830,000
Avoided Cost of Incineration	$	2,330,000	$	2,330,000	$	2,330,000
Total Revenue	**$**	**9,500,000**	**$**	**2,330,000**	**$**	**5,930,000**
Treament Cost, $/ton sludge	(41)		(10)		(29)	
Treatment Cost w/o avoided incineration cost, $/ton sludge	(16)		20		(2)	

3.2 Biocrude Upgrading Plant

Table 21 lists the feed and product flow rates for the centralized sludge biocrude upgrading facility. The plant processes 2,697 BPSD of biocrude feed and produces 2,649 BPSD of diesel and naphtha blendstocks. Mass and carbon flows for the upgrading plant are shown in Figure 11.

Table 21. Annual feed and product flows for centralized biocrude upgrading plant.

Stream	Million gal/year	Million lbs/year
Biocrude feed [a]	37.4	297.2
Hydrotreated oil	36.6	242.6
Diesel blendstock	27.5 (29.5 million gge/year)[a]	183.3
Naphtha (gasoline blendstock)	9.2 (9.1 million gge/year) [b]	56.3

(a) Wet, including 3.7% water.
(b) Based on a gasoline LHV of 116,090 Btu/gal (ANL 2018).

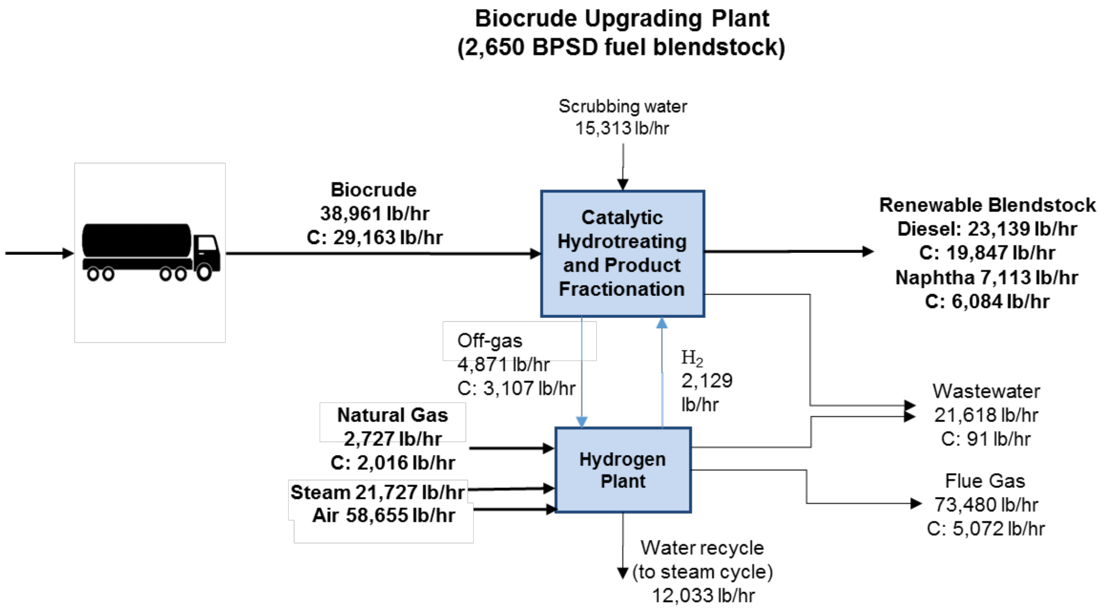


Figure 11. Mass and carbon flows for upgrading plant.

3.2.1 Total Capital Investment

Table 22 summarizes the costs for the biocrude upgrading plant including the steam cycle and balance of plant items such as the tank farm, flare and cooling water system. The hydrotreating section of the plant has the highest installed capital cost, with the hydrogen plant next highest.

Table 22. Total capital investment for biocrude upgrading plant.

Cost Item	Million US Dollars (2017$)
Hydrotreating	32.9
Hydrocracking	6.5
Hydrogen plant	26.9
Steam cycle	1.6
Balance of plant	6.4
Total Installed Cost (TIC)	**74.3**
Buildings (4% of TIC)	3.0
Site development (10% of TIC)	7.4
Additional piping (4.5% of TIC)	17.7
Total Direct Costs (TDC)	**86.5**
Indirect Costs	
Prorated expenses (10% TDC)	8.7
Home office & construction fees (20% TDC)	17.3
Field expenses (10% TDC)	8.7
Project contingency (10% TDC)	8.7
Startup and permits (10% TDC)	8.7
Total Indirect	**51.9**
Fixed Capital Investment (FCI)	**138.5**
Working Capital (5% of FCI)	6.9
Land (6% of TPEC)[(a)]	2.6
Total Capital Investment (TCI)	**148.0**

(a) Total purchased equipment cost

3.2.2 Operating Costs

Table 23 and Table 24 list the plant variable and fixed operating costs and supporting assumptions for the biocrude upgrading facility. Biocrude cost is the large majority (95%) of the operating cost and therefore efforts aimed at reducing its production cost are critical to reducing the final fuel MFSP. The biocrude price includes $0.092/gge for transportation to the upgrading facility (Snowden-Swan et al. 2017).

Table 23. Biocrude upgrading plant variable operating costs.

	Quote Year Cost		Year of Price Quote	MM$/yr (2017)	Source
Raw Materials					
Biocrude Feedstock[(a)]	2.86	$/gge	2017	115.65	This analysis
Natural Gas	3.51	$/1000scf	2016	1.78	2016 EIA
Boiler Chemicals	2.27	$/lb	2007	0.01	Humbird 2011
Cooling Tower Chemicals	1.36	$/lb	2007	0.006	Humbird 2011
Hydrogen Plant Catalyst	2.05	¢/1000scf H2	2014	0.07	PEP 2014
Hydrotreating Catalyst	16.6	$/lb	2014	0.42	PEP 2014
Hydrocracking Catalyst	16.6	$/lb	2014	0.03	PEP 2014
Subtotal				**117.96**	
Waste Disposal				**MM$/yr (2017)**	
Wastewater POTW fee	0.53	$/tonne	2001	0.07	Dutta 2011
Subtotal				**0.07**	
Utilities				**MM$/yr (2017)**	
Water Makeup	0.22000	$/metric ton	2001	0.05	Humbird 2011
Electricity	6.76000	¢/kwh	2016	0.91	EIA 2016

Subtotal	**0.96**
Total Variable Operating Costs	**118.99**
(a) Includes transportation cost of $0.092/gge biocrude.	

Table 24. Biocrude upgrading plant fixed operating costs.

Position Title	Number	Total Cost (2017), million USD/year
Conversion Plant (unburdened)		
Plant Manager	1	0.17
Plant Engineer	1	0.08
Maintenance Super	1	0.06
Lab Manager	1	0.06
Shift Supervisor	3	0.16
Lab Technician	3	0.15
Maintenance Tech	6	0.27
Shift Operators	25	1.36
Yard Employees	4	0.13
Clerks & Secretaries	1	0.04
Subtotal		**2.47**
Overhead & Maintenance	90% of labor & supervision	2.22
Maintenance Capital	3% of TIC	3.98
Insurance and Taxes	0.7% of FCI	0.97
Total Other Fixed Costs		**9.64**

3.2.3 Minimum Fuel Selling Price

The minimum fuel selling price (MFSP) for the upgrading plant is $3.83/gge fuel blendstock. Table 25 shows the breakdown of costs contributing to the fuel blendstock MFSP. Note that the cost of the biocrude feed is given per gge of upgraded fuel blendstock, considering that 1.054 gge of biocrude is needed to make 1 gge of fuel. Figure 12 shows the contribution of each section of the upgrading plant to the overall MFSP. The biocrude price contributes 56% of the fuel production price.

Table 25. Upgraded fuel blendstock MFSP cost breakdown.

	$/gge blendstock	$/year
Biocrude Feed (including transport cost of $0.092/gge)	3.00	$115,700,000
Natural Gas	0.05	$1,800,000
Catalysts & Chemicals	0.01	$500,000
Waste Disposal	0.002	$100,000
Electricity and Other Utilities	0.02	$1,000,000
Fixed Costs	0.25	$9,600,000
Capital Depreciation	0.002	$4,600,000
Average Income Tax	0.04	$1,400,000
Average Return on Investment	0.45	$17,500,000
Total	**3.83**	
Total w/out Renewable Credits	**8.08**	

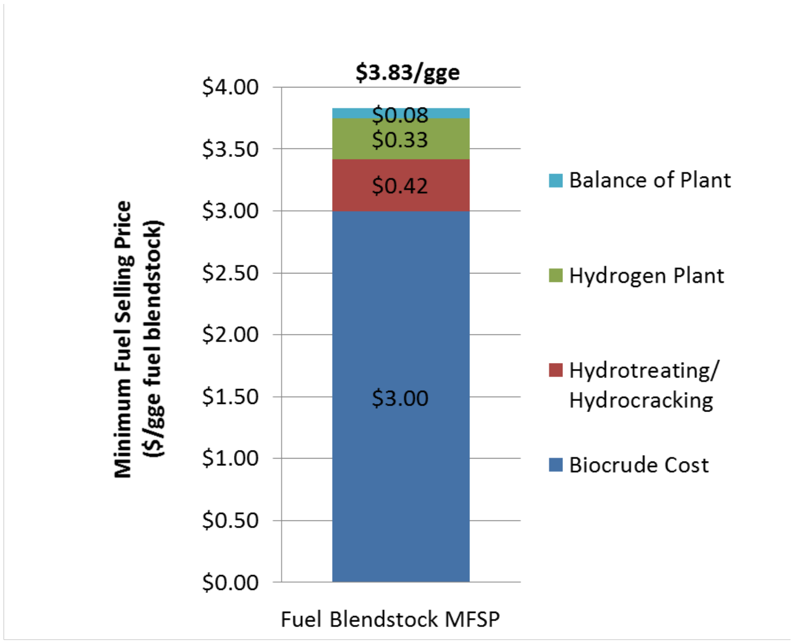


Figure 12. Contribution of biocrude cost and upgrading process areas to blendstock production cost.

4.0 References

40 CFR 80, Subpart M, "Renewable Fuel Standard." U.S. Code of Federal Regulations.

Anex RP, A Aden, FK Kazi, J Fortman, RM Swanson, MM Wright, JA Satrio, RC Brown, DE Daugaard, A Platon, G Kothandaraman, DD Hsu, and A Dutta. 2010. "Techno-economic comparison of biomass-to-transportation fuels via pyrolysis, gasification, and biochemical pathways." *Fuel* 89(2010):S29–S35. Doi:10.1016/j.fuel.2010.07.015

ANL. 2018. *Greenhouse Gases, Regulated Emissions, and Energy use in Transportation (GREET) Model.* Argonne National Laboratory, Argonne, IL. http://greet.es.anl.gov.

Boie W. 1953. "Fuel Technology Calculations." *Energietechnik* 3:309-316.

California Code of Regulations, Title 17, "Public Health." State of California, Sacramento, CA.

Dreher M, M Steib, M Nachtegall, J Wambach, and F Vogel. 2014. "On-Stream Regeneration of a Sulfur-Poisoned Ruthenium–Carbon Catalyst Under Hydrothermal Gasification Conditions." *ChemCatChem* 6:626-633. Doi:10.1002/cctc.201300791

EIA. 2018. *Domestic Crude Oil First Purchase Prices by Area.* U.S. Energy Information Administration, Washington, D.C. https://www.eia.gov/dnav/pet/pet_pri_dfp1_k_a.htm.

EPA. 2018. *RIN Trades and Price Information.* U.S. Environmental Protection Agency, Washington, D.C. https://www.epa.gov/fuels-registration-reporting-and-compliance-help/rin-trades-and-price-information.

Huang K, ZJ Brentzel, KJ Barnett, JA Dumesic, GW Huber, and CT Maravelias. 2017. "Conversion of Furfural to 1,5-Pentanediol: Process Synthesis and Analysis," *ACS Sustainable Chem. Eng.* 5:4699-4706. Doi:10.1021/acssuschemeng.7b00059

Merrow EW, KE Phillips, and CW Myers. 1981. *Understanding Cost Growth and Performance Shortfalls in Pioneer Process Plants*. Report DE-AC01-79PE70078. The Rand Corporation, Santa Monica, CA.

State of Michigan Biomass Energy Program Grant PLA-06-48. 2007. "Feasibility Study: Biodigester for Combined Heat and Power at Ann Arbor Wastewater Treatment Plant." https://www.a2gov.org/departments/finance-admin-services/purchasing/Documents/A2%20CHP%20Feasibility%20Study%20Report%20FINAL.pdf

Neste. 2018. *California Low Carbon Fuel Standard Credit Price.* Neste, Finland. https://www.neste.com/corporate-info/investors/market-data/lcfs-credit-price

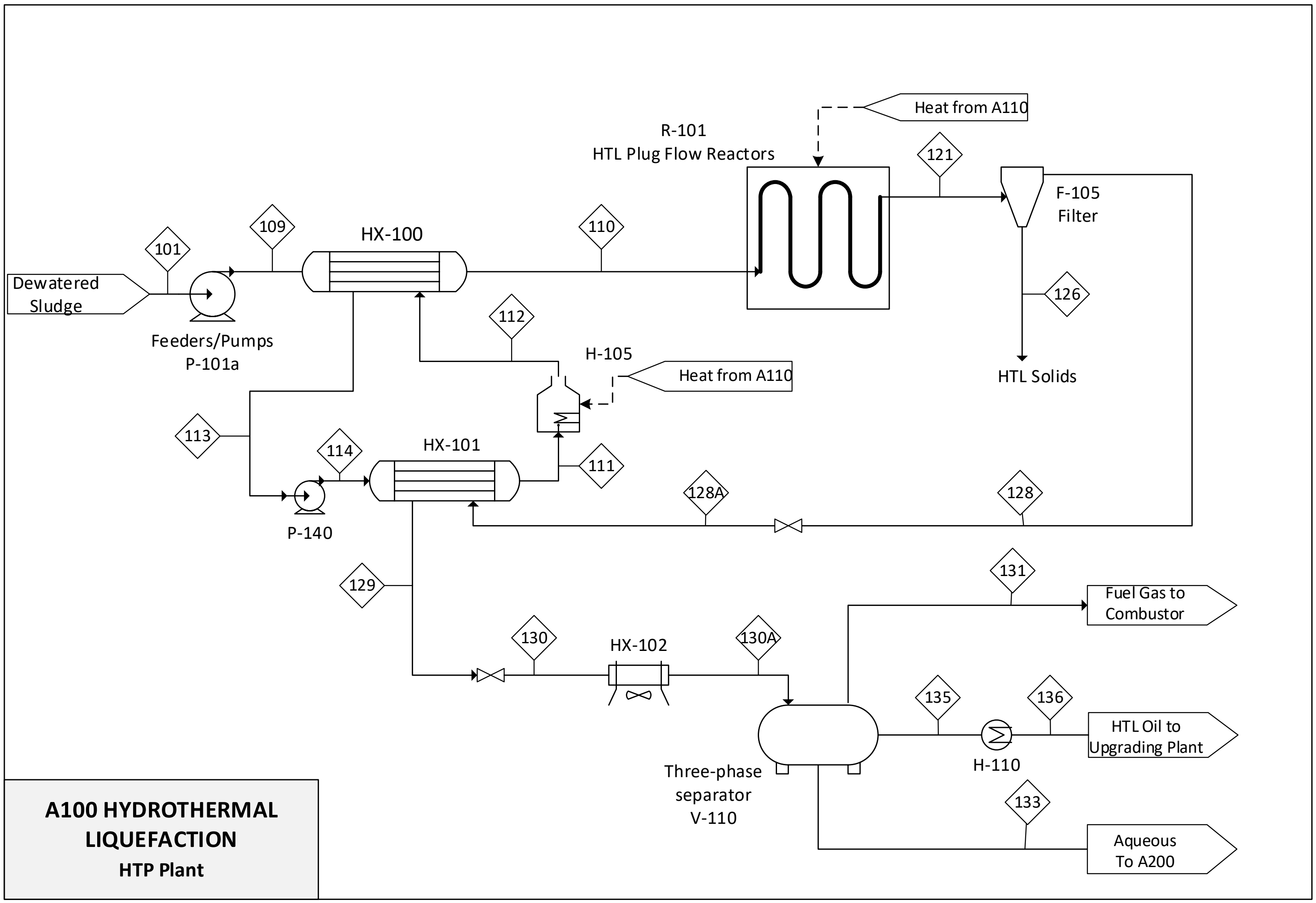
Peters, M, K Timmerhaus, and R West. 2003. *Plant Design and Economics for Chemical* Engineers. Mc-Graw-Hill, New York, NY.

Snowden-Swan LJ, Y Zhu, MD Bearden, TE Seiple, SB Jones, AJ Schmidt, JM Billing, RT Hallen, TR Hart, J Liu, KO Albrecht, SP Fox, GD Maupin, and DC Elliott. 2017. *Conceptual Biorefinery Design and Research Targeted for 2022: Hydrothermal Liquefaction Processing of Wet Waste to Fuels, PNNL-27186,* Pacific Northwest National Laboratory, Richland, WA.

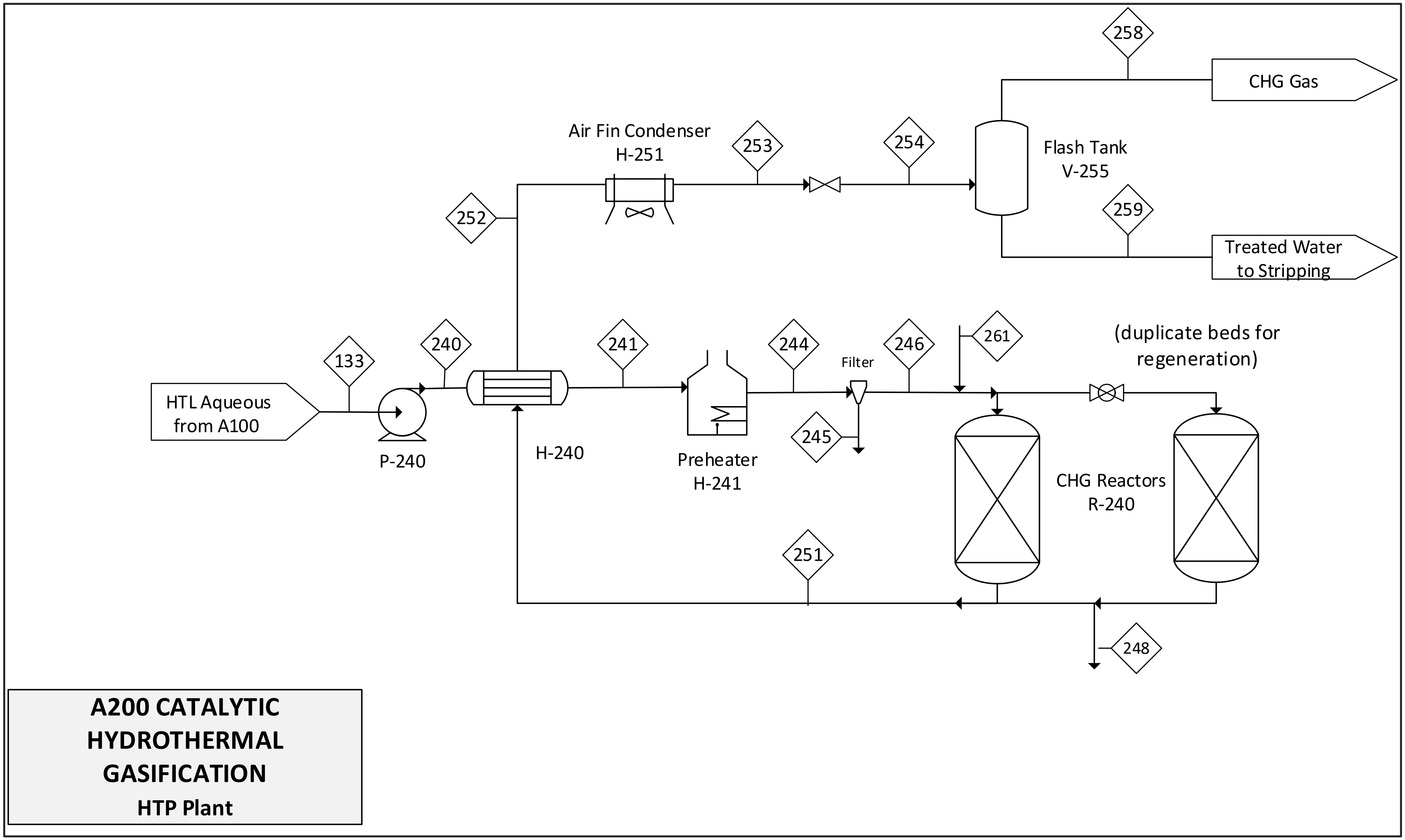
Tao L, J Markham, Z Haz, and M Biddy. 2017. "Techno-economic analysis for upgrading the biomass-derived ethanol-to-jet blendstocks." *Green Chem* 19:1082-1101.

Appendix A

PFDs and Stream Tables

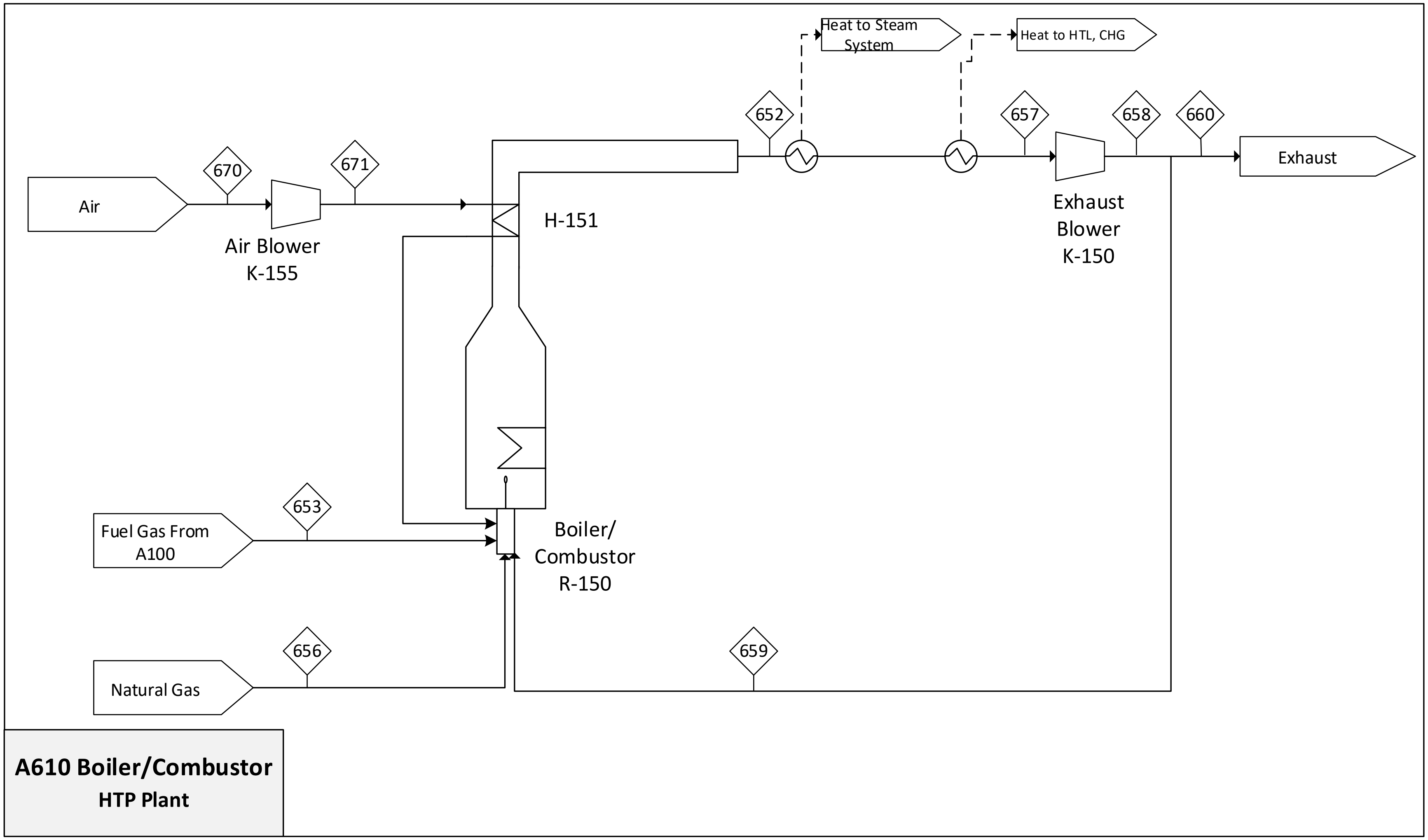


Stream	101	109	110	111	112	113	114	121	126	128	128A	129	130	130A	131	133	135	136
Temperature F	60.0	64.9	656.0	630.0	690.0	134.5	134.8	656.0	656.0	656.0	652.8	227.7	204.0	140.0	140.0	140.0	140.0	140.0
Pressure psia	14.7	3055.0	3010.0	170.0	165.0	155.0	180.0	3000.0	3000.0	3000.0	2900.0	2855.0	30.0	28.0	28.0	28.0	28.0	28.0
Vapor Frac	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.1	0.0	0.2	0.2	0.0	0.1	0.0	1.0	0.0	0.0	0.0
Total Mass Flow lb/hr	19205.5	19205.5	19205.5	40453.0	40453.0	40453.0	40453.0	19205.5	1953.9	17251.6	17251.6	17251.6	17251.6	17251.6	743.9	15185.9	1321.9	1321.9
VLSTDMX gal/hr	1844.6	1844.6	1844.6	4580.7	4580.7	4580.7	4580.7	2259.7	151.3	2108.5	2108.5	2108.5	2108.5	2108.5	112.3	1851.4	144.8	144.8
Mass Flow lb/hr																		
H2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.3	0.0	0.3	0.3	0.3	0.3	0.3	0.3	0.0	0.0	0.0
CO2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	932.3	0.0	932.3	932.3	932.3	932.3	932.3	717.0	215.4	0.0	0.0
H2O	15364.4	15364.4	15364.4	0.0	0.0	0.0	0.0	15363.7	1259.8	14103.9	14103.9	14103.9	14103.9	14103.9	0.0	14055.3	48.6	48.6
NH3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	61.1	0.0	61.1	61.1	61.1	61.1	61.1	0.0	61.1	0.0	0.0
CH4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	3.4	0.0	3.4	3.4	3.4	3.4	3.4	3.4	0.0	0.0	0.0
C2H6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	3.5	0.0	3.5	3.5	3.5	3.5	3.5	3.5	0.0	0.0	0.0
C3H8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	5.0	0.0	5.0	5.0	5.0	5.0	5.0	5.0	0.0	0.0	0.0
N-C4H10	0.0	0.0	0.0	0.0	0.0	0.0	0.0	10.0	0.0	10.0	10.0	10.0	10.0	10.0	10.0	0.0	0.0	0.0
N-PENTAN	0.0	0.0	0.0	0.0	0.0	0.0	0.0	4.7	0.0	4.7	4.7	4.7	4.7	4.7	4.7	0.0	0.0	0.0
METHANOL	0.0	0.0	0.0	0.0	0.0	0.0	0.0	78.1	0.0	78.1	78.1	78.1	78.1	78.1	0.0	78.1	0.0	0.0
ETHANOL	0.0	0.0	0.0	0.0	0.0	0.0	0.0	78.1	0.0	78.1	78.1	78.1	78.1	78.1	0.0	78.1	0.0	0.0
ACETONE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	67.9	0.0	67.9	67.9	67.9	67.9	67.9	0.0	67.9	0.0	0.0
ACEACID	0.0	0.0	0.0	0.0	0.0	0.0	0.0	203.7	0.0	203.7	203.7	203.7	203.7	203.7	0.0	203.7	0.0	0.0
PROACID	0.0	0.0	0.0	0.0	0.0	0.0	0.0	169.8	0.0	169.8	169.8	169.8	169.8	169.8	0.0	169.8	0.0	0.0
ETHAMIN	0.0	0.0	0.0	0.0	0.0	0.0	0.0	84.9	0.0	84.9	84.9	84.9	84.9	84.9	0.0	84.9	0.0	0.0
2-PYRRLD	0.0	0.0	0.0	0.0	0.0	0.0	0.0	34.0	0.0	34.0	34.0	34.0	34.0	34.0	0.0	34.0	0.0	0.0
2-PIPERD	0.0	0.0	0.0	0.0	0.0	0.0	0.0	34.0	0.0	34.0	34.0	34.0	34.0	34.0	0.0	34.0	0.0	0.0
7-LACTAM	0.0	0.0	0.0	0.0	0.0	0.0	0.0	47.5	0.0	47.5	47.5	47.5	47.5	47.5	0.0	47.5	0.0	0.0
C5H9NS	0.0	0.0	0.0	0.0	0.0	0.0	0.0	43.5	0.0	43.5	43.5	43.5	43.5	43.5	0.0	16.5	27.0	27.0
TOLUENE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	50.9	0.0	50.9	50.9	50.9	50.9	50.9	0.0	0.0	50.9	50.9
RYRO3ETM	0.0	0.0	0.0	0.0	0.0	0.0	0.0	34.0	0.0	34.0	34.0	34.0	34.0	34.0	0.0	0.0	34.0	34.0
PHENO4M	0.0	0.0	0.0	0.0	0.0	0.0	0.0	34.0	0.0	34.0	34.0	34.0	34.0	34.0	0.0	0.0	34.0	34.0
AMIPHENO	0.0	0.0	0.0	0.0	0.0	0.0	0.0	13.6	0.0	13.6	13.6	13.6	13.6	13.6	0.0	0.0	13.6	13.6
INDOLE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	71.3	0.0	71.3	71.3	71.3	71.3	71.3	0.0	0.0	71.3	71.3
2-PYTENE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	47.5	0.0	47.5	47.5	47.5	47.5	47.5	0.0	0.0	47.5	47.5
C15OLEF	0.0	0.0	0.0	0.0	0.0	0.0	0.0	54.3	0.0	54.3	54.3	54.3	54.3	54.3	0.0	0.0	54.3	54.3
MC12AMID	0.0	0.0	0.0	0.0	0.0	0.0	0.0	105.3	0.0	105.3	105.3	105.3	105.3	105.3	0.0	0.0	105.3	105.3
C16AMIDE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	88.3	0.0	88.3	88.3	88.3	88.3	88.3	0.0	0.0	88.3	88.3
C18AMIDE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	159.6	0.0	159.6	159.6	159.6	159.6	159.6	0.0	0.0	159.6	159.6
C16:0FA	0.0	0.0	0.0	0.0	0.0	0.0	0.0	151.4	0.0	151.4	151.4	151.4	151.4	151.4	0.0	5.4	146.0	146.0
C18:1FA	0.0	0.0	0.0	0.0	0.0	0.0	0.0	180.0	0.0	180.0	180.0	180.0	180.0	180.0	0.0	0.0	180.0	180.0
C13H18	0.0	0.0	0.0	0.0	0.0	0.0	0.0	67.9	0.0	67.9	67.9	67.9	67.9	67.9	0.0	0.0	67.9	67.9
HEVOIL1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	153.0	0.0	153.0	153.0	153.0	153.0	153.0	0.0	27.2	125.8	125.8
HEVOIL2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	67.9	0.0	67.9	67.9	67.9	67.9	67.9	0.0	0.0	67.9	67.9
DOWTH-01	0.0	0.0	0.0	40453.0	40453.0	40453.0	40453.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SLUDGE	3841.1	3841.1	3841.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ASH	0.0	0.0	0.0	0.0	0.0	0.0	0.0	446.4	441.9	4.5	4.5	4.5	4.5	4.5	0.0	4.5	0.0	0.0
SOLID	0.0	0.0	0.0	0.0	0.0	0.0	0.0	254.7	252.1	2.5	2.5	2.5	2.5	2.5	0.0	2.5	0.0	0.0
Enthalpy MMBtu/hr	-112.0	-111.8	-100.2	9.3	10.9	-0.7	-0.7	-99.2	-8.7	-90.5	-90.5	-100.5	-100.5	-101.8	-2.8	-98.0	-1.2	-1.2

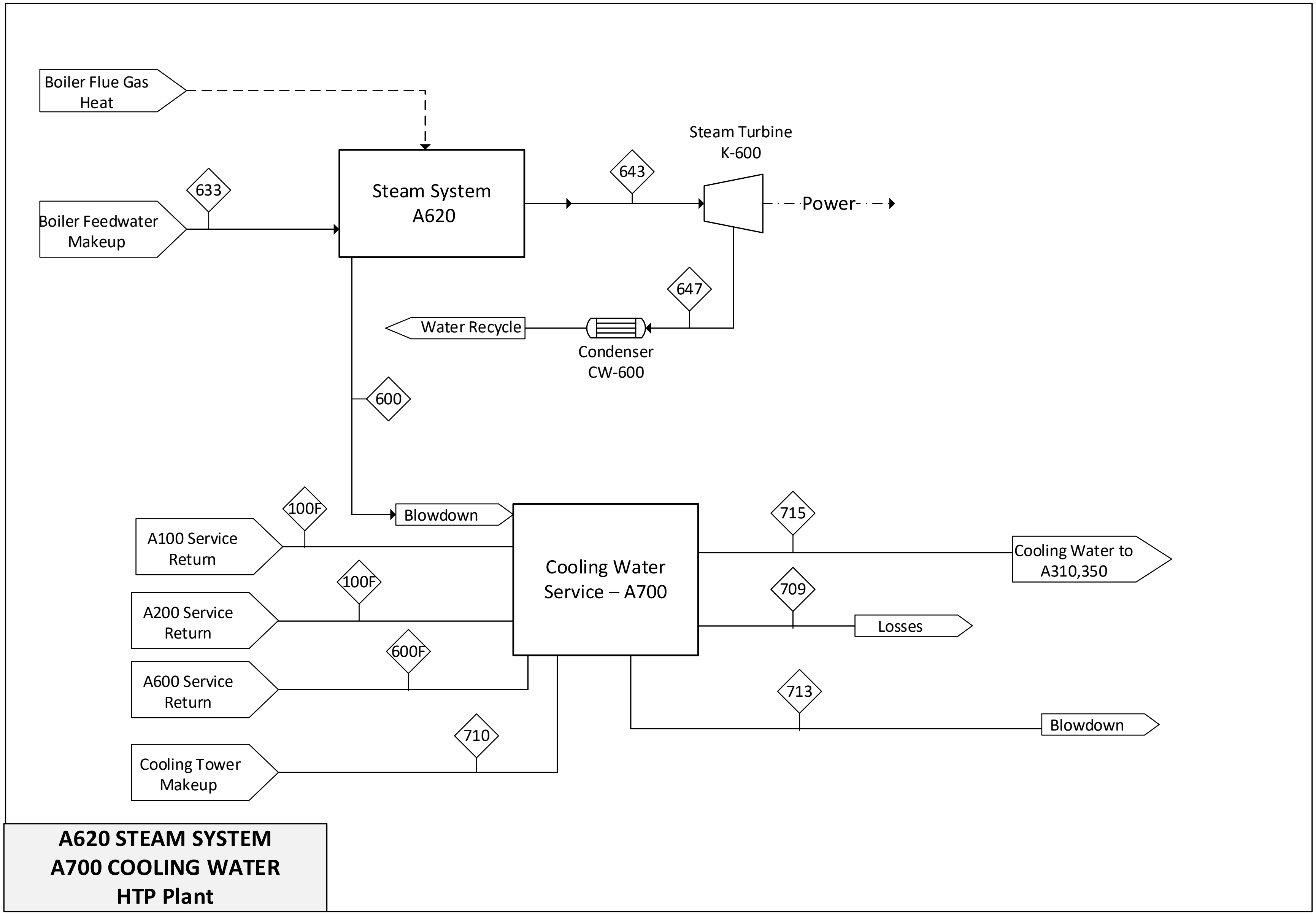


HTL Aqueous from A100
133
240
P-240
H-240
252
241
Air Fin Condenser
H-251
253
254
Flash Tank
V-255
258
CHG Gas
259
Treated Water to Stripping
244
Preheater
H-241
Filter
245
246
261
(duplicate beds for regeneration)
CHG Reactors
R-240
251
248
A200 CATALYTIC HYDROTHERMAL GASIFICATION
HTP Plant

Stream	133	240	241	244	24	246	248	251	252	253	254	258	259	261
Temperature F	140	144.5	580	665		664.9	689.	637.4	330.1	140	138.6	132	132	77
Pressure psia	28	2950	2945	2944		2939	2934	2924	2923	2921	50	50	50	3084.
Vapor Frac	0	0	0	0		0	0.01	0.238	0.044	0.038	0.052	1	0	1
Total Mass Flow	15185.	15185.	15185.	15185.	7.0	15178.	4.6	15174.	15174.	15174.	15174.	739.	14434.	0.292
VLSTDMX gal/hr	1851.4	1851.4	1851.4	1851.4		1851.4	0.3	1973.9	1973.9	1973.9	1973.9	198.	1640.1	0.929
Mass Flow lb/hr														
H2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1.0	1.0	1.0	1.0	1.0	0.0	0.292
CO2	215.4	215.4	215.4	215.4	0.0	215.4	0.0	799.3	799.3	799.3	799.3	358.	9.3	0.0
H2O	14055.	14055.	14055.	14055.	0.0	14055.	0.0	13899.	13899.	13899.	13899.	27.0	13647.	0.0
NH3	61.1	61.1	61.1	61.1	0.0	61.1	0.0	121.6	121.6	121.6	121.6	0.0	0.0	0.0
CH4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	341.0	341.0	341.0	341.0	340.	0.4	0.0
C2H6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	3.8	3.8	3.8	3.8	3.8	0.0	0.0
C3H8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	8.6	8.6	8.6	8.6	8.6	0.0	0.0
SULFUR	0.0	0.0	0.0	0.0	0.0	0.0	4.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NH4+	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	128.8	0.0
H3O+	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	50.6	0.0
HCO3-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	597.9	0.0
METHANOL	78.1	78.1	78.1	78.1	0.0	78.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHANOL	78.1	78.1	78.1	78.1	0.0	78.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACETONE	67.9	67.9	67.9	67.9	0.0	67.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ACEACID	203.7	203.7	203.7	203.7	0.0	203.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
PROACID	169.8	169.8	169.8	169.8	0.0	169.8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ETHAMIN	84.9	84.9	84.9	84.9	0.0	84.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
2-PYRRLD	34.0	34.0	34.0	34.0	0.0	34.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
2-PIPERD	34.0	34.0	34.0	34.0	0.0	34.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
7-LACTAM	47.5	47.5	47.5	47.5	0.0	47.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
C5H9NS	16.5	16.5	16.5	16.5	0.0	16.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
C16:0FA	5.4	5.4	5.4	5.4	0.0	5.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HEVOIL1	27.2	27.2	27.2	27.2	0.0	27.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ASH	4.5	4.5	4.5	4.5	4.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SOLID	2.5	2.5	2.5	2.5	2.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Enthalpy	-98.0	-97.8	-90.6	-87.9	0.0	-87.9	0.0	-87.9	-95.1	-97.9	-97.9	-2.2	-96.0	<


Heat to Steam System
Heat to HTL, CHG
652
657
658
660
670
671
Air
Air Blower
K-155
H-151
Exhaust
Exhaust Blower
K-150
653
Fuel Gas From A100
Boiler/ Combustor
R-150
656
Natural Gas
659
A610 Boiler/Combustor
HTP Plant

Stream	652	653	656	657	658	659	660	670	671
Temperature F	1794	140	60	312	347	347	347	90	132
Pressure psia	16	28	450	15	17	17	17	15	18
Vapor Frac	1	1	1	1	1	1	1	1	1
Mass Flow lb/hr	15765	744	325	15765	15765	11036	4730	3661	3661
Volume Flow cuft/hr	835368	3910	235	309548	284213	198949	85264	50141	44812
Mass Flow lb/hr									
H2	0.0	0.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N2	7071.5	0.0	1.8	7071.5	7071.5	4950.1	2121.5	2119.7	2119.7
O2	242.2	0.0	0.0	242.2	242.2	169.5	72.6	1447.6	1447.6
AR	120.5	0.0	0.0	120.5	120.5	84.4	36.2	36.2	36.2
CO2	5593.4	717.0	3.6	5593.4	5593.4	3915.4	1678.0	1.4	1.4
H2O	2737.8	0.0	0.0	2737.8	2737.8	1916.5	821.3	55.9	55.9
H2S	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SO2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NH3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NO2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CH4	0.0	3.4	319.6	0.0	0.0	0.0	0.0	0.0	0.0
C2H6	0.0	3.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0
C3H8	0.0	5.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N-C4H10	0.0	10.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N-PENTAN	0.0	4.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Enthalpy MMBtu/hr	-28.9	-2.8	-0.7	-36.3	-36.1	-25.3	-10.8	-0.3	-0.3


Boiler Flue Gas Heat
Boiler Feedwater Makeup
633
Steam System A620
643
Steam Turbine K-600
Power
647
Water Recycle
Condenser CW-600
600
Blowdown
A100 Service Return
100F
A200 Service Return
100F
A600 Service Return
600F
Cooling Tower Makeup
710
Cooling Water Service – A700
715
Cooling Water to A310,350
709
Losses
713
Blowdown
A620 STEAM SYSTEM
A700 COOLING WATER
HTP Plant

Stream	600	633	643	644	645	CW100F	CW200F	CW600F	709	710	713	715
Temperature F	111.7	232.3	700	179.9	179.3	110	110	110	89.9	60	89.2	89.3
Pressure psia	14.7	21.7	659.2	7.5	7.4	59.7	59.64	59.7	14.7	14.7	14.7	74.7
Vapor Frac	0	0	1	0.993	0	0	0	0	0	0	0	0
Mass Flow lb/hr	34.136	1706.771	1672.648	1672.648	1672.648	3430.171	15342.65	83332.41	100145.9	2364.015	438.792	102105.2
VLSTDMX gal/hr	4.098	204.91	200.814	200.814	200.814	411.817	1841.998	10004.67	12023.25	283.817	52.68	12258.49
Mass Flow lb/hr												
H2O	34.136	1706.771	1672.648	1672.648	1672.648	3430.171	15342.65	83332.41	100145.9	2364.015	438.792	102105.2
Enthalpy MMBtu/hr	-0.232	-11.377	-9.23	-9.592	-11.238	-23.515	-105.179	-571.27	-688.859	-16.342	-3.019	-702.389

Appendix B

Equipment Lists for HTP and Upgrading Plants

HTP Plant:

Equipment Number	Number Required	Number Spares	Equipment Name	Scaling Stream	Original Equipment Stream Flow	New Flows	stream flow units	Size Ratio	Original Equip Cost (per unit)	Base Year	COST BASIS: installed (i) or bare (b)	Total Original Equip Cost in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2017$	Scaled Uninstalled Cost in 2017$	Ref
A 10	**Feed Prep - Sludge Dewatering and Grinding**																		
C-10	1	0	Bowl Centrifuge - GEA	sludge	26,417	15,345	gal/hr	0.6	$700,000	2016	b	$700,000	0.67	486,438	1.80	875,588	917,290	$509,606	2
G-10	1	0	Grinder (assumed after dewatering)	S101	4,935	2,161	gal/hr	0.4	$24,980	2016	b	$24,980	0.6	15,219	1.80	27,394	28,698	$15,944	2
	A10 Total																**945,988**	**525,549**	
A100	**HTL Oil Production**																		
P-101	1	0	Twin Screw Feeder and Feed Pump	S101	192	36	gpm	0.19	$364,700	2011	b	$364,700	0.8	95,600	2.30	219,881	213,048	$92,630	1
H-100	1	0	Sludge Heater	Area	1,441.0	18,866.3	ft2	13.09	$464,812	2Q 2018	b	$464,812	0.6	2,175,155	2.20	4,785,341	4,515,599	$2,052,545	2
H-101	1	0	HTL Product Cooler	Area	3,154.0	7,334.1	ft2	2.33	$642,172	2Q 2018	b	$642,172	0.6	1,065,468	2.20	2,344,030	2,211,901	$1,005,410	2
R-101	1	0	HTL Reactor	S101	1,375	19,201	lb/hr	13.96	$61,568	2Q 2018	b	$61,568	1	859,772	2.20	1,891,498	1,784,877	$811,308	2
	1	0	Hot Oil Tank	Vol	730.0	730.0	gal	1.00	$22,300	1Q 2014	b	$22,300	0.6	22,300	4.20	93,660	92,745	$22,082	3
H-105	1	0	Hot Oil Heater (fired heater)	Duty	2.6	2.6	mmBtu/hr	1.00	$191,800	1Q 2014	b	$191,800	0.65	191,800	1.40	268,520	265,896	$189,926	3
	1	0	Hot Oil Pump	Flow	79.0	76.4	gpm	0.97	$6,300	1Q 2014	b	$6,300	0.7	6,154	3.60	22,155	21,938	$6,094	3
	1	0	Dowtherm	Flow	6.3	76.4	gpm	12.13	$136,080	1Q 2017	b	$136,080	1	1,650,240	1.00	1,650,240	1,678,937	$1,678,937	2
F-105	1	0	Solids filter	S101	2,420	36	pgm	0.01	$1,017,998	2011	b	$1,017,998	0.6	81,533	1.70	138,605	134,298	$78,999	1
	1	0	Solids Precipitation Vessel	Flow	1,047.0	1,047.0	gallon	1.00	$407,100	1Q 2014	b	$407,100	0.6	407,100	1.70	692,070	685,307	$403,122	3
	1	0	Solids Blowdown Precipitation Vessel	Flow	838.0	838.0	gallon	1.00	$379,100	1Q 2014	b	$379,100	0.6	379,100	1.60	606,560	600,633	$375,396	3
	1	0	Solids Pump	Flow	25.00	30.00	gpm	1.20	$18,700	1Q 2014	b	$18,700	0.6	20,862	2.10	43,810	43,382	$20,658	3
H-102	1	0	Reactor product air cooler	Duty	1.25	1.25	mmBtu/hr	1.00	$41,900	1Q 2014	b	$41,900	0.6	41,900	2.17	90,923	90,035	$41,491	3
V-110	1	0	3-phase separator (horizontal vessel)	S140	8,376	8,400	gal	1.00	$98,900	1Q 2014	b	$98,900	0.6	99,070	2.70	267,489	264,875	$98,102	3
	1	0	Iron Sponge - (H2S removal)	S101	1,375	19,201	lb/hr	13.96	$141,425	2Q2018	b	$141,425	0.6	687,928	2.00	1,375,856	1,298,301	$649,151	2
H-110 (CW-103	1	0	Biocrude cooler	Duty	0.068	0.068	mmBtu/hr	1.00	$11,900	1Q 2014	b	$11,900	0.6	11,900	6.00	71,400	70,702	$11,784	3
	A100 Total																**13,972,475**	**7,537,632**	
A200	**CHG HTL Water Treatment System**																		
P-240a	1	0	Booster Pump (plunger type)	S133	15,186	15,186	lb/hr	1.00	$16,500	1Q 2014	b	$16,500	0.8	16,500	2.40	39,600	39,213	$16,339	3
P-240	1	0	Feed Pump (plunger type)	S133	15,186	15,186	lb/hr	1.00	$79,100	1Q 2014	b	$79,100	0.8	79,100	1.74	137,634	136,289	$78,327	3
H-240	1	0	Feed/Product Exchanger	Area	7,720	1,463	ft2	0.19	$5,013,647	2011	b	$5,013,647	0.6	1,848,119	2.20	4,065,862	3,939,520	$1,790,691	1
H-241	1	0	Fired Heater	Duty	2.7	2.7	mmBtu/hr	1.00	$333,000	1Q 2014	b	$333,000	0.65	333,000	1.30	432,900	428,670	$329,746	3
R-240	1	1	CHG Reactor	S133	1,117	15,186	lb/hr	13.60	$260,100	2Q 2018	b	$520,200	0.6	2,490,029	2.00	4,980,059	4,699,341	$2,349,670	2
H-251	1	0	Product Air Fin Cooler	S133	15,186	15,186	lb/hr	1.00	$132,500	1Q 2014	b	$132,500	0.6	132,500	1.19	157,675	156,134	$131,205	3
V-255	1	0	CHG Product Separator	S133	15,186	15,186	lb/hr	1.00	$47,900	1Q 2014	b	$47,900	0.6	47,900	4.15	198,785	196,843	$47,432	3
	A200 Total																**$9,596,009**	**$4,743,410**	
A600	**Steam system**																		
P-604	1	0	BFW Pump	S637	320	202	gal/hr	0.63	$52,100	1Q 2014	b	$52,100	0.85	35,238	1.32	46,514	46,059	$34,893	3
H-605A	1	0	BFW Preheater		546,113	477,046	mmBtu/hr	0.87	$16,600	1Q 2014	b	$16,600	0.70	15,101	6.35	95,891	94,954	$14,953	3
R-650	1	0	Boiler - Packaged	S633	10,000	1,680	lb/hr	0.17	$152,500	1Q 2014	b	$152,500	0.85	33,480	3.22	107,805	106,752	$33,153	3
			steam drum, aerator included w/superheat																
H-651	1	0	Boiler air preheater	Duty	77,656	77,656	Btu/hr	1.00	$10,600	1Q 2014	b	$10,600	0.40	10,600	7.42	78,652	77,883	$10,496	3
K-600	1	0	Steam engine - condensing	Duty	104	104	kW	1.00	$58,100	1Q 2014	b	$58,100	0.85	58,100	1.75	101,675	100,681	$57,532	3
										Subtotal							**$426,330**	**$151,028**	
A700	**OSBL - including cooling water system**																		
	1	0	Cooling Tower System - packaged	circ rate	150	201	gpm	1.34	$5,900	1Q 2014	b	$5,900	0.6	7,041	10.10	71,110	70,415	$6,972	3
	1	0	Cooling Water Pump	circ rate	201	201	gpm	1.00	$6,800	1Q 2014	b	$6,800	0.6	6,808	3.97	27,027	26,763	$6,741	3
	0	0	Plant Air Compressor	plant	2,000	42	tonne/day	0.02	$87,922	2007	b	$0	0.3	0	1.57	0	0	$0	4
	1	0	Plant Air Compressor	flow	1,000	750	scf/min	0.75	$169,700	1Q 2014	b	$169,700	0.6	142,797	1.80	257,034	254,523	$141,402	3
	1	0	Firewater Pump	plant	2,000	42	tonne/day	0.02	$23,043	2007	b	$23,043	0.3	7,222	3.70	26,720	28,861	$7,800	4
	1	0	Instrument Air Dryer	plant scale	2,000	42	tonne/day	0.02	$8,349	2002	b	$8,349	0.6	820	2.47	2,025	2,906	$1,176	4
	1	0	Plant Air Receiver	plant scale	2,000	42	tonne/day	0.02	$21,005	2007	b	$21,005	0.65	1,700	5.44	9,250	9,991	$1,837	4
	1	0	Firewater Storage Tank	plant scale	2,000	42	tonne/day	0.02	$229,900	2007	b	$229,900	0.65	18,610	1.46	27,170	29,347	$20,101	4
	1	0	Biocrude Storage - 3 day (S316)	S135	10,427	11,531	gal	1.11	$52,300	1Q 2014	b	$52,300	0.65	55,836	2.15	120,046	118,873	$55,290	3
	A700 Total									**Subtotal**							**$541,677**	**$241,318**	
			References																
		1	Knorr 2013											**Total Equipment Cost**			**$25,482,480**	**$13,198,938**	
		2	Vendor Budget Estimate																
		3	Aspen Capital Cost Estimator, 8.8																
		4	Dutta et al. NREL//TP-5100-51400																
		5	Aspen Process Economic Analyzer, V.8.8																

Upgrading Plant:

Equipment Number	Number Required	Number Spares	Equipment Name	Scaling Stream	Original Equipment Stream Flow	New Flows	stream flow units	Size Ratio	Original Equip Cost (per unit)	Base Year	COST BASIS: installed (i) or bare (b)	Equip Cost (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2017$	Scaled Uninstalled Cost in 2017$	Ref
A300	**HTL Oil Upgrading and Product Separation**																		
A310	**HTL Oil Hydrotreating & Separations**																		
	1	0	Desalter	S301	2,697	2,697	bpd fd	1.00	$763,177	2007	i	$763,177	0.75	763,177	2.47	763,177	824,330	$333,737	7
	1	0	Hydrotreater Fe Bed	S301	15,472	2,697	bpd fd	0.17	$16,302,021	2014	i	$16,302,021	0.75	4,398,334	1.51	4,398,334	4,332,676	$2,869,322	1
R-301a	1	0	Hydrotreater Reactor, vessels,	S301	15,472	2,697	bpd fd	0.17	$77,087,500	2014	i	$77,087,500	0.75	20,798,436	1.51	20,798,436	20,487,958	$13,568,184	1
K-310	1	0	Hydrogen Compressor	S304-H2	3,786.0	4,965.0	lb/hr H2	1.31	$1,385,600	1Q 2011	b	$1,385,600	0.8	1,721,190	1.1	1,893,309	1,879,400	$1,708,545	2
PSA-310	1	0	PSA for Hydrogen Recycle	S425-H2	10	9.6	mmscfd H2	0.96	$1,750,000	2004	b	$1,750,000	0.8	1,696,444	2.47	4,190,216	5,353,326	$2,167,339	3
	A310 Total									**Subtotal**					**A310**		**$32,877,690**	**$20,647,126**	
A350	**Hydrocracking and Separations**																		
R-350	1	0	Hydrocracker Unit + auxiliaries	S338	15,472	1,021	bpd fd	0.07	$50,858,783	2014	i	$50,858,783	0.75	6,619,706	1.51	6,619,706	6,520,887	$4,318,468	1
	A350 Total									**Subtotal**					**A350 HCK**		**$6,520,887**	**$4,318,468**	
											0.66								
A400	**Hydrogen Plant - OSBL**																		
	1	0	Stm Reformer system w/ associated O	S425-H2	24.5	9.6	mmscfd H2	0.39	$50,198,438	2014	i	$50,198,438	0.65	27,337,868	1.92	27,337,868	26,929,769	$14,025,921	4
	A400 Total									**Subtotal**					**A400 H2 Plnat**		**$26,929,769**	**$14,025,921**	
A600	**Power Generation - OSBL**			Costs accounted for in Hydrogen Plant except for turbine															
E-601	1	0	Steam turbine + generator	work	1,652	1,812	kW	1.10	$1,268,300	1Q 2014	b	$1,268,300	0.85	1,371,980	1.19	1,632,657	1,616,703	$1,358,574	8
	A600 Total									**Subtotal**							**$1,616,703**	**$1,358,574**	
															A600 STM				
A700	**OSBL - including cooling water system**																		
	1	0	Wastewater treatment (ammonia	S320+SCRB-	34,000	21,280	lb/hr	0.63	1,217,800	1Q 2014	b	$1,217,800	0.65	898,041	2.4	2,155,299	2,134,238	$889,266	8
	1	0	Cooling Tower System	S715	7,506,000	1,038,497	lb/hr	0.14	260,852	2010	b	$260,852	0.78	55,766	2.47	137,743	141,919	$57,457	6
	1	0	Cooling Water Pump	S715	7,001,377	1,038,497	lb/hr	0.15	239,375	2007	b	$239,375	0.3	135,035	2.14	288,974	312,130	$145,855	6
	1	0	Makeup water pump	S618+S710	80,411	34,095	lb/hr	0.42	6,528	2007	b	$6,528	0.3	5,047	4.72	23,820	25,728	$5,451	6
	1	0	Plant Air Compressor	S101(dry) X	262,454	61,113	lb/hr	0.23	$87,922	2007	b	$87,922	0.3	56,784	1.57	89,150	96,294	$61,334	6
	1	0	Firewater Pump	S101(dry) X	262,454	61,113	lb/hr	0.23	$23,043	2007	b	$23,043	0.3	14,882	3.7	55,064	59,476	$16,075	6
	1	0	Instrument Air Dryer	S101(dry) X 10	262,454	61,113	lb/hr	0.23	$8,349	2002	b	$8,349	0.6	3,482	2.47	8,602	12,339	$4,996	6
	1	0	Plant Air Receiver	S101(dry) X 10	262,454	61,113	lb/hr	0.23	$21,005	2007	b	$21,005	0.65	8,146	5.44	44,312	47,863	$8,798	6
	1	0	Firewater Storage Tank	S101(dry) X 10	262,454	61,113	lb/hr	0.23	$229,900	2007	b	$229,900	0.65	89,154	1.46	130,165	140,596	$96,298	6
	1	0	Feed Storage - 3 day	S-301	1,000,000	339,870	gallons	0.34	2,371,900	2010	b	$2,371,900	0.7	1,114,334	1.4	1,560,067	1,544,823	$1,103,445	5
	1	0	Product Storage - 3 day	S392	1,000,000	83,592	gallons	0.08	2,371,900	2010	b	$2,371,900	0.7	417,457	1.4	584,440	602,160	$430,114	5
	1	0	Product Storage - 3 day	S393	1,000,000	250,128	gallons	0.25	2,371,900	2010	b	$2,371,900	0.7	899,104	1.4	1,258,746	1,296,910	$926,365	5
	A700 Total									**Subtotal**					**A700 OSBL**		**$6,414,476**	**$3,745,453**	
			References																
		1	IHS PEP Yearbook 2014a - "Diesel from High Pressure Hydrocracking" (P=2000 psia)												**Total Equipment Cost**		**$74,359,525**	**$44,095,543**	
		2	Aspen Capital Cost Estimator, 7.3.2																
		3	Ron Pasadan Quote from 2004																
		4	IHS PEP Yearbook 2014b - "Hydrogen from Natural Gas by Steam Reforming"																
		5	Aspen Capital Cost Estimator, 7.3.1																
		6	Dutta et al. NREL//TP-5100-51400, 2011																
		7	Kaiser and Gary, Study updates refinery investment cost curves, Oil&Gas Journal, April 2007																
		8	Aspen Capital Cost Estimator, 8.8																

Appendix C

Discounted Cash Flow Worksheets

Table C.1. HTP plant DCFROR sheet.

Year		-2	-1	0	1	2	3	4
Fixed Capital Investment		$ 1,552,597	$ 11,644,474	$ 6,210,386				
Land		$ 35,752						
Working Capital				$ 2,425,932				
Loan Payment					$ 4,338,425	$ 4,338,425	$ 4,338,425	$ 4,338,425
Loan Interest Payment		$ 186,312	$ 1,583,648	$ 2,328,895	$ 2,328,895	$ 2,168,132	$ 1,994,509	$ 1,806,996
Loan Principal		$ 2,328,895	$ 19,795,606	$ 29,111,185	$ 27,101,655	$ 24,931,362	$ 22,587,446	$ 20,056,017
Fuel Sales					$ 2,849,717	$ 3,799,622	$ 3,799,622	$ 3,799,622
By-Product and Renewable Credits)					$ 4,352,817	$ 5,803,756	$ 5,803,756	$ 5,803,756
Total Annual Sales					$ 7,202,534	$ 9,603,378	$ 9,603,378	$ 9,603,378
Annual Manufacturing Cost								
Sludge production					$ (1,746,089)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)
CHG catalyst					$ 249,455	$ 249,455	$ 249,455	$ 249,455
Baghouse Bags					$ -			
Other Variable Costs					$ 1,530,708	$ 1,749,380	$ 1,749,380	$ 1,749,380
Fixed Operating Costs					$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275
Total Product Cost					$ 3,483,349	$ 3,119,992	$ 3,119,992	$ 3,119,992
Annual Depreciation								
General Plant Writedown					14.29%	24.49%	17.49%	12.49%
Depreciation Charge					$ 6,817,318	$ 11,683,422	$ 8,343,939	$ 5,958,593
Steam Plant Writedown					3.75%	7.22%	6.68%	6.18%
Depreciation Charge					$ 30,440	$ 58,599	$ 54,199	$ 50,141
Net Revenue	net profit before taxes				$ (5,457,467)	$ (7,426,767)	$ (3,909,260)	$ (1,332,343)
Losses Forward						$ (5,457,467)	$ (12,884,234)	$ (16,793,495)
Taxable Income					$ (5,457,467)	$ (12,884,234)	$ (16,793,495)	$ (18,125,838)
Income Tax					$ -	$ -	$ -	$ -
Annual Cash Income					$ (619,240)	$ 2,144,962	$ 2,144,962	$ 2,144,962
Discount Factor		1.2100	1.1000	1.0000	0.9091	0.8264	0.7513	0.6830
Annual Present Value	$ 27,522,411				$ (562,946)	$ 1,772,696	$ 1,611,541	$ 1,465,038
Total Capital Investment + Interest		$ 2,147,339	$ 14,550,935	$ 10,965,213				
Net Present Worth				$ -				
Annual Cash Flows								
Payback Period (PBP)= FCI/(Avg Annual Cash Flow)								
PBP, years (including depreciation)	12.0							
Annual Cash Flow (including loan payment)	This is using Peters and Timmerhaus eqn 6.1 (net profit after taxes + deprecia				$ 1,390,290	$ 4,315,254	$ 4,488,878	$ 4,676,391
Payback Period, years (not including depreciation)	9.7							
Disposal Cost/Revenue, $/wet ton sludge	$ 66							

5	6	7	8	9	10	11	12	13	14	15	16	17
$ 4,338,425	$ 4,338,425	$ 4,338,425	$ 4,338,425	$ 4,338,425	$ 4,338,425	$ –	$ –	$ –	$ –	$ –	$ –	$ –
$ 1,604,481	$ 1,385,766	$ 1,149,553	$ 894,443	$ 618,925	$ 321,365	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)
$ 17,322,073	$ 14,369,414	$ 11,180,542	$ 7,736,560	$ 4,017,060	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)
$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622
$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756
$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378
$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)
$ 249,455	$ 249,455	$ 249,455	$ 249,455	$ 249,455	$ 249,455	$ 249,455	$ 249,455	$ 249,455	$ 249,455	$ 249,455	$ 249,455	$ 249,455
	$ –					$ –					$ –	
$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380
$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275
$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992
8.93%	8.92%	8.93%	4.46%									
$ 4,260,227	$ 4,255,456	$ 4,260,227	$ 2,127,728									
5.71%	5.29%	4.89%	4.52%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%	4.46%
$ 46,374	$ 42,900	$ 39,677	$ 36,706	$ 36,219	$ 36,211	$ 36,219	$ 36,211	$ 36,219	$ 36,211	$ 36,219	$ 36,211	$ 36,219
$ 572,304	$ 799,264	$ 1,033,929	$ 3,424,509	$ 5,828,242	$ 6,125,810	$ 6,447,167	$ 6,447,175	$ 6,447,167	$ 6,447,175	$ 6,447,167	$ 6,447,175	$ 6,447,167
$ (18,125,838)	$ (17,553,534)	$ (16,754,269)	$ (15,720,340)	$ (12,295,832)	$ (6,467,590)	$ (341,779)	$ –	$ –	$ –	$ –	$ –	$ –
$ (17,553,534)	$ (16,754,269)	$ (15,720,340)	$ (12,295,832)	$ (6,467,590)	$ (341,779)	$ 6,105,388	$ 6,447,175	$ 6,447,167	$ 6,447,175	$ 6,447,167	$ 6,447,175	$ 6,447,167
$ –	$ –	$ –	$ –	$ –	$ –	$ 1,282,131	$ 1,353,907	$ 1,353,905	$ 1,353,907	$ 1,353,905	$ 1,353,907	$ 1,353,905
$ 2,144,962	$ 2,144,962	$ 2,144,962	$ 2,144,962	$ 2,144,962	$ 2,144,962	$ 5,201,255	$ 5,129,480	$ 5,129,482	$ 5,129,480	$ 5,129,482	$ 5,129,480	$ 5,129,482
0.6209	0.5645	0.5132	0.4665	0.4241	0.3855	0.3505	0.3186	0.2897	0.2633	0.2394	0.2176	0.1978
$ 1,331,852	$ 1,210,775	$ 1,100,704	$ 1,000,640	$ 909,673	$ 826,976	$ 1,823,008	$ 1,634,410	$ 1,485,828	$ 1,350,752	$ 1,227,957	$ 1,116,324	$ 1,014,841
$ 4,878,905	$ 5,097,621	$ 5,333,834	$ 5,588,943	$ 5,864,462	$ 6,162,022	$ 5,201,255	$ 5,129,480	$ 5,129,482	$ 5,129,480	$ 5,129,482	$ 5,129,480	$ 5,129,482

18	19	20	21	22	23	24	25	26	27	28	29	30
												$ (35,752)
			$ –	$ –	$ –	$ –	$ –	$ –	$ –	$ –	$ –	$ (2,425,932)
$ –	$ –	$ –	$ –	$ –	$ –	$ –	$ –	$ –	$ –	$ –	$ –	$ –
$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)
$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)	$ (0)
$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622	$ 3,799,622
$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756	$ 5,803,756
$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378	$ 9,603,378
$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)	$ (2,328,119)
$ 249,455	$ 249,455	$ 249,455	$ 249,455	$ 249,455	$ 249,455	$ 249,455	$ 249,455	$ 249,455	$ 249,455	$ 249,455	$ 249,455	$ 249,455
$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380	$ 1,749,380
$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275	$ 3,449,275
$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992	$ 3,119,992
4.46%	4.46%	4.46%	2.23%									
$ 36,211	$ 36,219	$ 36,211	$ 18,110									
$ 6,447,175	$ 6,447,167	$ 6,447,175	$ 6,465,277	$ 6,483,387	$ 6,483,387	$ 6,483,387	$ 6,483,387	$ 6,483,387	$ 6,483,387	$ 6,483,387	$ 6,483,387	$ 6,483,387
$ –	$ –	$ –	$ –	$ –	$ –	$ –	$ –	$ –	$ –	$ –	$ –	$ –
$ 6,447,175	$ 6,447,167	$ 6,447,175	$ 6,465,277	$ 6,483,387	$ 6,483,387	$ 6,483,387	$ 6,483,387	$ 6,483,387	$ 6,483,387	$ 6,483,387	$ 6,483,387	$ 6,483,387
$ 1,353,907	$ 1,353,905	$ 1,353,907	$ 1,357,708	$ 1,361,511	$ 1,361,511	$ 1,361,511	$ 1,361,511	$ 1,361,511	$ 1,361,511	$ 1,361,511	$ 1,361,511	$ 1,361,511
$ 5,129,480	$ 5,129,482	$ 5,129,480	$ 5,125,678	$ 5,121,875	$ 5,121,875	$ 5,121,875	$ 5,121,875	$ 5,121,875	$ 5,121,875	$ 5,121,875	$ 5,121,875	$ 5,121,875
0.1799	0.1635	0.1486	0.1351	0.1228	0.1117	0.1015	0.0923	0.0839	0.0763	0.0693	0.0630	0.0573
$ 922,582	$ 838,711	$ 762,464	$ 692,636	$ 629,202	$ 572,002	$ 520,001	$ 472,729	$ 429,753	$ 390,685	$ 355,168	$ 322,880	$ 293,527
												$ (141,076)
$ 5,129,480	$ 5,129,482	$ 5,129,480	$ 5,125,678	$ 5,121,875	$ 5,121,875	$ 5,121,875	$ 5,121,875	$ 5,121,875	$ 5,121,875	$ 5,121,875	$ 5,121,875	$ 5,121,875

Appendix D

Compound Selection

D.1 HTL Liquid Products Composition

HTL organic products are a complex mixture of hundreds of compounds. The number and type of compounds used in the Aspen model to represent HTL oil and the associated aqueous phase must reasonably match key properties, such as CHONS, density, heating value, GC/MS data, expected HTL oil distillation range, and aqueous solubility. The compounds chosen for the Aspen model are shown in Table D.1. Note that this list does not imply that these compounds occur in the given percentages in actual HTL oil, rather each compound represents a group of compounds that taken together exhibit the bulk properties. Carbon dioxide and ammonia in the aqueous phase actually form their ionic species in various amounts and types, including NH_4^+, NH_2COO^-, HCO_3^-, CO_3^{2-}. For simplification purposes, ion formation is not simulated in the HTL model but their pure original compounds are considered.

Table D.1. Compounds used to model HTL liquid products.

HTL Oil	Heat & Mat'l Balance Names	Wt%	C	H	O	N	S	CAS
N-methylthiopyrrolidone	C5H9NS	4.167%	5	9		1	1	10441-57-3
Toluene	TOLUENE	2.986%	7	8				108-88-3
1H-Pyrrole, 3-ethyl-2,4,5-trimethyl-	RYRO3ETM	5.284%	9	15		1		520-69-4
Phenol, 4-methyl-	PHENO4M	0.459%	7	8	1			106-44-5
3-methyl-4-aminophenol	AMIPHENO	0.919%	7	9	1	1		2835-99-6
Indole	INDOLE	5.513%	8	7		1		120-72-9
3,7,11,15-tetramethyl-2-Hexadecene	2-PYTENE	6.892%	20	40				3452-07-1
1-Pentadecene	C15OLEF	6.892%	15	30				13360-61-7
N-Methyldodecanamide	MC12AMID	5.973%	13	27	1	1		27563-67-3
Palmitamide(Hexadecanamide)	C16AMIDE	5.513%	16	33	1	1		629-54-9
9-Octadecenamide, (Z)-	C18AMIDE	11.486%	18	35	1	1		301-02-0
Palmitic acid (n-Hexadecanoic acid)	C16:0FA	7.811%	16	32	2			57-10-3
Oleic-acid	C18:1FA	11.257%	18	34	2			112-80-1
Naphthalene, 1,2,3,4-tetrahydro-1,1,6-trimethyl-	C13H18	9.189%	13	18				91-20-3
Triphenylformazan	HEVOIL1	9.915%	19	16		4		531-52-2
Dibenzyl-sebacate	HEVOIL2	5.743%	24	30	4			140-24-9
		100.000%						
HTL Aqueous Phase		**wt%**	**C**	**H**	**O**	**N**	**S**	**CAS**
Methanol	METHANOL	2.347%	1	4	1			67-56-1
Ethanol	ETHANOL	0.391%	2	6	1			64-17-5
Acetone	ACETONE	0.196%	3	6	1			67-64-1
Acetic acid	ACEACID	2.347%	2	4	2			64-19-7
propanoic acid	PROACID	1.173%	3	6	2			79-09-4
CO2	CO2	68.178%	1		2			124-38-9
NH3	NH3	15.646%		3		1		7664-41-7
Ethyl-amine	ETHAMIN	1.956%	2	7		1		75-04-7
Triphenylformazan	HEVOIL1	2.738%	19	16		4		531-52-2
N-METHYLTHIOPYRROLIDONE		5.028%	5	9		1	1	10441-57-3
		100.000%						

D.2 Hydrotreated Oil Model Compounds

Similar to HTL oil, hydrotreated oil contains numerous compounds and a limited number are used for modeling purposes. Table D.2 shows the mixture of compounds used to represent the hydrotreated oil in the models. Note, that this list does not imply that these compounds occur in the given percentages in actual hydrotreated oil, rather each compound represents a group of compounds that taken together exhibit the bulk properties.

Table D.2. Compounds used to model hydrotreated product.

Compound	Heat & Mat'l Balance Names	C	H	O	N	Wt%	CAS
Pentane	N-PENTAN	5	12			4.47%	109-66-0
Pentane, 2-methyl-	2MPENTA	6	14			1.45%	107-83-5
Hexane	HEXANE	6	14			1.12%	110-54-3
Cyclopentane, methyl-	CC5-METH	6	12			0.78%	96-37-7
Hexane, 2-methyl-	2MHEXAN	7	16			1.45%	591-76-4
Cyclohexane, methyl-	CC6-METH	7	14			0.56%	108-87-2
Piperidine	PIPERDIN	5	11		1	0.22%	110-89-4
Toluene	TOLUENE	7	8			1.68%	108-88-3
Heptane, 3-methyl-	3MHEPTA	8	18			1.12%	589-81-1
Octane	OCTANE	8	18			0.78%	111-65-9
Cyclohexane, ethyl-	ETHCYC6	8	16			0.84%	1678-91-7
Ethylbenzene	ETHBENZ	8	10			0.84%	100-41-4
o-Xylene	O-XYLENE	8	10			1.34%	95-47-6
Nonane	C9H20	9	20			1.90%	111-84-2
Cyclohexane, propyl-	CC6-PRO	9	18			1.45%	1678-92-8
Benzene, propyl-	C3BENZ	9	12			0.89%	103-65-1
Nonane, 4-methyl-	4MNONAN	10	22			0.89%	17301-94-9
Decane	C10H22	10	22			1.12%	124-18-5
Benzene, butyl-	C4BENZ	10	14			1.68%	104-51-8
Benzene, 1-butenyl-	C10H12	10	12			1.90%	824-90-8
Dodecane	C12H26	12	26			1.34%	112-40-3
Naphthalene, 1,2,3,4-tetrahydro-	1234NA	10	12			2.24%	119-64-2
Cyclohexane, hexyl-	CC6-HEX	12	24			2.24%	4292-75-5
Naphthalene, 1,2,3,4-tetrahydro-1,4-dimethyl-	14DMNAPH	12	16			2.80%	4175-54-6
Benzene, heptyl-	C7BENZ	13	20			3.36%	1078-71-3
Benzene, octyl-	C8BENZ	14	22			2.57%	2189-60-8
Pentadecane	C15H32	15	32			2.24%	629-62-9
Hexadecane	C16H34	16	34			6.71%	544-76-3
Heptadecane	C17H36	17	36			11.19%	629-78-7
2,6,10,14-Tetramethyl-hexadecane	TETMC16	20	42			6.71%	638-36-8
Octadecane	C18H38	18	38			10.07%	593-45-3
Eicosane	C20H42	20	42			6.71%	112-95-8
Palmitic acid	C16:0FA	16	32	2		1.57%	57-10-3
Heneicosane	C21H44	22	46			1.68%	629-94-7
Tetracosane	C24H50	24	50			3.91%	646-31-1
Di-2-ethylhexyl-adipate	C22H42O4	22	42	4		2.24%	103-23-1
Octacosane	C28H58	28	58			2.80%	630-02-4
Triacontane	C30H62	30	62			0.89%	638-68-6
Dotriacontane	C32H66	32	66			2.24%	544-85-3

Appendix E

Cost Indices

Table E.1. Labor indices.

Source: Bureau of Labor Statistics
Series ID: CEU3232500008 Chemicals
Average Hourly Earnings of Production Workers
Current indices @ http://data.bls.gov/cgi-bin/srgate

YEAR	INDEX	YEAR	INDEX
2005	19.67	2012	21.45
2006	19.60	2013	21.40
2007	19.55	2014	21.49
2008	19.50	2015	21.76
2009	20.30	2016	22.71
2010	21.07	2017	24.28
2011	21.46		

Table E.2. Capital cost indices.

Chemical Engineering Magazine, CEI annual index

YEAR	INDEX	YEAR	INDEX
1990	357.6	2005	468.2
1991	361.3	2006	499.6
1992	358.2	2007	525.4
1993	359.2	2008	575.4
1994	368.1	2009	521.9
1995	381.1	2010	550.8
1996	381.7	2011	585.7
1997	386.5	2012	584.6
1998	389.5	2013	567.3
1999	390.6	2014	576.1
2000	394.1	2015	556.8
2001	394.3	2016	541.7
2002	395.6	2017	567.5
2003	402.0	2Q 2018	601.4
2004	444.2		

Table E.3. Inorganic chemical indices.

Source: Bureau of Labor Statistics
Series ID: PCU325---325---
Producer Price Index Industry Data for Chemical mfg
Current indices @ http://data.bls.gov/cgi-bin/srgate

YEAR	INDEX	YEAR	INDEX
1997	147.1	2008	228.2
1998	148.7	2009	224.7
1999	149.7	2010	233.7
2000	156.7	2011	252.1
2001	158.4	2012	260.3
2002	157.3	2013	263.9
2003	164.6	2014	269.2
2004	172.8	2015	264.8
2005	187.3	2016	267.1
2006	196.8	2017	277.6
2007	203.3		

Appendix F

Life Cycle Inventory for HTP and Upgrading

Table F.1. HTL plant parameters for GHG and water analysis.

HTL PLANT	CHG Gas Export Case
Sludge Properties	
Moisture content, %	80%
Ash content (dry basis), %	11.6%
Biocrude Properties	
Moisture content, %	3.70%
Density, lb/gal	8.25
LHV, Btu/gal	125480
Inputs	
Sludge, lb/hr (dry basis)	3841
Natural gas, lb/hr (for heat)	325
Electricity, kW (process)	533
Dewatering polymer, lb/hr	38.4
Cooling water makeup, lb/hr	2334
Boiler Feedwater makeup, lb/hr	1680
CHG catalyst regeneration:	
H2O2, lb/hr (70% solution)	23
Makeup water, lb/hr	513
Rinse water, lb/hr	536
Outputs	
Biocrude, lb/hr	1322
CHG Gas, lb/hr	740
Solids, lb/hr (64% moisture)	1961
Wastewater, lb/hr	14435
Wastewater, lb/hr (cooling water BD)	433
CHG Catalyst Regeneration	
Wastewater, lb/hr, including rinse	1072

Table F.2. Upgrading plant parameters for GHG and water analysis.

UPGRADING PLANT	
Fuel Product Properties	
Diesel density, lb/gal	6.66
Diesel LHV, Btu/gal	124,394
Naphtha density, lb/gal	6.13
Naphtha LHV, Btu/gal	114650
Inputs	
Biocrude, lb/hr	38961
Natural gas, lb/hr	2727
Electricity, kW	1653
Cooling tower chemical, lb/hr	0.4
Boiler chemical, lb/hr	0.3
Hydrotreating catalyst, lb/hr	3.1
Hydrocracking catalyst, lb/hr	0.3
Hydrogen plant catalyst, lb/hr	0.4
Cooling water makeup, lb/hr	23,537
Boiler feedwater makeup, lb/hr	10,558
Scrubbing water makeup, lb/hr	15,832
Outputs	
Diesel, lb/hr	23139
Naphtha, lb/hr	7113
Wastewater, lb/hr	21618

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