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Conceptual Biorefinery Design and Research Targeted for 2022: Hydrothermal Liquefaction Processing of Wet Waste to Fuels

December 2017

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UNITED STATES DEPARTMENT OF ENERGY
under Contract DE-AC05-76RL01830

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Summary

The Department of Energy Bioenergy Technologies Office (BETO) invests in research and development of new pathways for commercially viable conversion of biomass into drop-in ready transportation fuels, fuel blendstocks and products. The primary emphasis has been on terrestrial and algae feedstocks, but more recently BETO has begun to explore the potential of wet wastes for biofuel production, with focus on wastewater residuals, manure, food waste, and fats, oils and grease. A recent resource analysis estimates that 77 million dry tons per year of these wastes are generated annually, 65% of which are underutilized for any beneficial purpose.[1] Approximately 14 million dry tons of the total resource is wastewater residuals (sludge and biosolids) generated at the nation's wastewater treatment plants (WWTPs).[2] Conversion of this resource into transportation fuels could significantly contribute to the creation of a new domestic bioenergy and bioproduct industry, while providing an economically and environmentally sustainable alternative for current waste disposal practices.

Hydrothermal liquefaction (HTL) is a process that uses hot, pressurized water in the condensed phase to convert biomass to a thermally stable oil product, also known as "biocrude", which can then be thermo-catalytically upgraded to hydrocarbon fuel blendstocks. HTL is conceptually simple, has a high carbon efficiency, and can be applied to a wide range of wet feedstocks at similar processing conditions. The purpose of this report is to document the conceptual design, economics and supporting data for a sludge-to-fuel pathway via HTL and biocrude upgrading. The configuration includes a HTL plant that is co-located with a WWTP and a larger scale biocrude upgrading plant for production of hydrocarbon fuel blendstocks. Experimental data from bench scale testing of a 1:1 mixture of primary:secondary sludges are used to establish the economic and technical assumptions for the analysis. The design represents a goal case for the pathway, targeting performance that is anticipated to be achievable by 2022 with further research and development. The year 2022 is BETO's target year for verification of hydrocarbon biofuel pathways.[3] As this analysis represents a goal case, assumed values of several design parameters represent improvements in the technology relative to what has currently been demonstrated in the laboratory. While HTL is fairly well developed and may therefore be ready for commercialization prior to 2022, there are specific advancements addressed in this analysis that are necessary to enhance performance compared to what has been demonstrated to date. In addition, an important aspect to the pathway is the upgrading of biocrude to fuel blendstock, an area that has received much less attention and requires significant research to validate the goal case performance parameters.

Summary economics for the sludge HTL plant and the biocrude upgrading plant are presented in Figures ES.1 and ES.2, respectively. The estimated plant gate minimum fuel selling price for fuel blendstock from sludge HTL and upgrading is $3.46/gasoline gallon equivalent (gge) (Figure ES.2). This price is within the tolerance (+$0.49/gge) of BETO's $3/gge programmatic cost target[3] and illustrates that fuel blendstocks generated from HTL of sludge and centralized biocrude upgrading have the potential to

[1] DOE. 2017. *Biofuels and Bioproducts from Wet and Gaseous Waste Streams: Challenges and Opportunities*. Bionergy Technologies Office, Energy Efficiency and Renewable Energy, U.S. Department of Energy, Washington, D.C.

[2] Seiple T, AM Coleman, and RL Skaggs. 2017. "Municipal wastewater sludge as a sustainable bioresource in the United States." *J. Environ. Manage.* 197:(2017) 673-680. http://dx.doi.org/10.1016/j.jenvman.2017.04.032

[3] DOE. 2016. *Biomass Multi-Year Program Plan*. Bionergy Technologies Office, Energy Efficiency and Renewable Energy, U.S. Department of Energy, Washington, D.C.

be competitive with fossil fuels. This analysis illustrates the feasibility of HTL for point-of-generation conversion of waste feedstock at a scale 1/20th that of the standard lignocellulosic biorefinery scale typically used in BETO design cases. The relevance of this work reaches beyond wastewater treatment sludge to lay the groundwork for application to other distributed wet wastes and blends that together represent a significant resource of underutilized biomass.

Biocrude from Sludge Hydrothermal Liquefaction

Primary/Secondary Wastewater Treatment Sludge (25 wt% solids; 15% ash)

Minimum Selling Price	**$2.35**	**$/gge biocrude**
Minimum Selling Price	$2.53	$/gal biocrude
Biocrude	4.1	million gge/yr
	0.47	trillion Btu/yr
	132	gge/US ton AFDW sludge
	15	million Btu/US ton AFDW sludge
Internal Rate of Return (After-Tax)	10%	
Equity Percent of Total Investment	40%	
Cost Year	2014	

CAPITAL COSTS

Sludge Dewatering	$1,400,000	8%
HTL Oil Production	$13,100,000	72%
HTL Water Recycle Treatment	$3,100,000	17%
Balance of Plant	$600,000	3%
Total Installed Capital Cost	$18,200,000	100%
Building, site development, add'l piping	$3,300,000	
Indirect Costs	$12,900,000	
Working Capital	$1,700,000	
Land (plant located at WWTP)	$100,000	
Total Capital Investment (TCI)	$36,200,000	
Installed Capital per Annual GGE Biocrude	$4.5	
TCI per Annual GGE Biocrude	$8.9	
Loan Rate	8.0%	
Term (years)	10	
Capital Charge Factor (computed)	0.141	

MANUFACTURING COSTS

Plant Hours per year	7920
Feed rate, dry sludge	110 ton/day
Feed rate, dry ash-free sludge	93.5 ton/day
Feed rate, slurry	69 gal/min

	$/gge biocrude	$/year
Avoided sludge disposal cost	0.00	$0
Natural Gas	0.14	$600,000
Chemicals	0.18	$800,000
Electricity	0.11	$500,000
Fixed Costs	0.68	$2,800,000
Capital Depreciation	0.27	$1,100,000
Average Income Tax	0.16	$700,000
Average Return on Investment	0.81	$3,300,000
	2.35	

PERFORMANCE

Net Electricity Purchased (KW)	818
Net Electricity Purchased (KWh/gge product)	1.6
(Energy in Biocrude) / (Electricity+Natural Gas Input)	4.1
Overall Carbon Yield to Biocrude	
On sludge + natural gas	66%
On sludge	72%

Figure ES.1. Summary economics for sludge HTL plant.

Liquid Fuels from Sludge HTL Biocrude Upgrading

Biocrude Feedstock Cost:	$2.45	$/gge biocrude (includes $0.10/gge transport cost)

Minimum Fuel Selling Price (MFSP)	**$3.46**	**$/gge**
Diesel Fuel Selling Price	$3.71	$/gal
Naphtha Fuel Selling Price	$3.42	$/gal

Naphtha	Diesel	Total	
666	1991	2,700	BPSD
9.1	29.6	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	2014

CAPITAL COSTS

Hydrotreating	$33,600,000	45%
Hydrocracking	$6,600,000	9%
Hydrogen Plant	$27,200,000	36%
Steam cycle	$1,600,000	2%
Balance of Plant	$6,500,000	9%
Total Installed Capital Cost	$75,500,000	100%
Building, site development, add'l piping	$12,400,000	
Indirect Costs	$52,800,000	
Working Capital	$7,000,000	
Land (included in feedstock cost)	$2,700,000	
Total Capital Investment (TCI)	$150,400,000	
Installed Capital per Annual GGE Fuel	$2.0	
TCI per Annual GGE Fuel	$3.9	

Loan Rate	8.0%
Term (years)	10
Capital Charge Factor (computed)	0.168

MANUFACTURING COSTS

Plant Hours per year	7920	
Biocrude feed rate	38 mmgal/y	

	$/gge fuel blendstock	$/year
Biocrude	2.58	$99,900,000
Natural Gas	0.07	$2,800,000
Catalysts & Chemicals	0.01	$500,000
Waste Disposal	0.002	$100,000
Electricity and other utilities	0.02	$1,000,000
Fixed Costs	0.24	$9,200,000
Capital Depreciation	0.002	$4,700,000
Average Income Tax	0.07	$2,800,000
Average Return on Investment	0.46	$17,700,000
	3.46	

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. Summary economics for biocrude upgrading plant.

Acknowledgments

Abbreviations

ACCE	Aspen Capital Cost Estimator
AD	anaerobic digestion
AFDW	ash-free dry weight
BETO	Bioenergy Technologies Office
BOD	biochemical oxygen demand
BPSD	barrels per stream day
Btu	British thermal unit
COD	chemical oxygen demand
CSTR	continuous stirred tank reactor
CWNS	Clean Watershed Needs Survey
DOE	U.S. Department of Energy
EPA	Environmental Protection Agency
FCI	fixed capital investment
FOG	fats, oils, and grease
gge	gasoline gallon equivalent
HHV	Higher Heating Value
HTL	hydrothermal liquefaction
IRR	internal rate of return
LHSV	liquid hourly space velocity
LHV	lower heating value
MFSP	minimum fuel selling price
MGD	million gallons per day
PFR	plug flow reactor
PNNL	Pacific Northwest National Laboratory
POTW	publicly owned treatment works
PSA	pressure swing adsorption
TAN	total acid number
TCI	total capital investment
TCLP	toxicity characteristic leaching procedure
TDC	total direct cost
TEA	techno-economic analysis
U	overall heat transfer coefficient
USD	United States dollar
WAS	waste activated sludge
WE&RF	Water Environment & Reuse Foundation
WHSV	weight hourly space velocity
WWTP	waste water treatment plant

Contents

Figures

Tables

1.0   Introduction

For several decades, the Department of Energy Bioenergy Technologies Office (BETO) has supported research and development for the advancement of bioenergy pathways for terrestrial and algal feedstocks. More recently, BETO has begun to explore the potential of wet wastes for biofuel production, and estimates that 77 million dry tons per year of wastewater residuals, manure, food waste, and fats, oils and grease (FOG) are generated annually, 65% of which are underutilized for any beneficial purpose, such as for fertilizer, biodiesel or compost (DOE 2017). Conversion of this resource into transportation fuels could significantly contribute to the nation's renewable energy goals and provide an economically and environmentally sustainable alternative for current waste disposal practices.

Approximately 14 million dry tons of wastewater residuals are generated annually at the nation's wastewater treatment plants (WWTPs) (Seiple et al. 2017). Costs associated with management and disposal of these residuals accounts for about 45-65% of the total WWTP's operating expenses (Nowak 2006; Gray 2010; City of Detroit 2014). Shipping is a significant portion of total sludge management cost, with some municipalities having to transport their sludge great distances for treatment and disposal (Peccia and Westerhoff 2015; Hsieh 2013). While cost estimates vary widely, it is evident that reducing sludge management and disposal costs is a strong industry incentive that introduces the opportunity for new technology.

Hydrothermal liquefaction (HTL) is a process that uses hot, pressurized water in the condensed phase to convert wet biomass to an oil product. As such, it is particularly well suited for processing wet waste feedstocks and eliminates the need for drying that is required for other biomass conversion technologies. The oil product, also known as "biocrude", is analogous to petroleum crude in that it contains a mixture of hydrocarbons with carbon numbers in the gasoline/jet/diesel range. However, it contains higher oxygen and nitrogen than petroleum and therefore must be hydrotreated to improve compability with petroleum fuels.

HTL research over the past several decades has included a range of feedstocks from lignocellulosics to algae to wet wastes. The summary here will focus on liquefaction of sludge feedstock into an oil product and will not cover the breadth of work done on other feedstocks nor on carbonization of sludge. Much of the literature on sewage sludge describes testing in lab-scale batch systems. Early work appears to have originated with Appell et al. (1970), where several organic waste feedstocks were tested in a 500-ml autoclave system, where the effect of temperature, pressure, and the presence of carbon monoxide (a reducing agent) and sodium carbonate catalyst on oil yield and composition were examined. A series of studies in Japan tested various sewage sludges also in the presence of sodium carbonate catalyst (Yokoyama et al. 1987; Suzuki et al. 1988, 1990; Inoue et al. 1997). More recently, Vardon et al. (2011) examined properties of biocrude from anaerobically digested sludge, manure and algae. Huang et al. (2013) and Leng et al. (2015) characterized biocrude from HTL of sludge using ethanol solvent and studied its emulsification with petroleum diesel, respectively. Malins et al. (2015) examined the effect of several different catalysts as well as other reaction parameters on biocrude properties. Several recent studies focused on the migration and ultimate fate of metals during HTL of sludge (Huang et al. 2011; Huang and Yuan 2016; Leng et al. 2016; Shao et al. 2015; Zhai et al. 2014).

While batch system investigations provide useful information, continuous system testing is necessary to facilitate engineering scale up and economic analysis of commercial systems. Several studies of

continuous HTL focused on sewage sludge can be found in the literature. Appell et al. (1971) conducted initial trials of a continuous 100-500 g/hr unit developed based on their initial batch testing (Appell et al. 1970) with the use of reducing agents. Solvent extraction was used to separate the oil product from the reaction mixture. While the focus of this effort was conversion of wastes, including sewage sludge, they reported results only for cellulose. A study by PNNL for the U.S. Environmental Protection Agency (EPA) conducted larger scale continuous thermochemical liquefaction as a possible alternative sludge disposal technology (Molton et al. 1986). In this work, a pilot-scale (30 L/hr) reactor system processing primary sludge at 20% solids and 5% sodium carbonate concentration was demonstrated and solvent extraction was used to recover the oil. Itoh et al. (1992) tested a mixture of primary and secondary sludge with no added catalyst in a 500 kg/day (21 L/hr) continuous system and demonstrated continuous separation of the oil using high pressure distillation. Villadsen et al. (2012) tested a range of waste feedstocks including sludge in a 30 L/hr continuous pilot system called CatLiq® (Toor et al. 2012), with the focus being to develop and apply analysis methodology for identification of specific compounds in HTL biocrudes. Their process utilizes both homogeneous and heterogeneous catalysts in the reactor, and a centrifuge for biocrude separation.

As part of a collaborative project between BETO and the Water Environment & Reuse Foundation (WE&RF), PNNL recently conducted continuous HTL of primary, secondary and anaerobically digested sludges in a 1.5 L/hr bench system (Marrone 2016). Representatives from WWTPs, the EPA, academia, and engineering consulting firms participated in the joint project. This work was the basis of a preliminary techno-economic analysis (TEA) for sludge HTL and subsequent biocrude upgrading (Snowden-Swan et al. 2016) and helped to guide follow-on experiments with more targeted run conditions, the results of which provide the basis for the present analysis. Complementing earlier continuous system studies, PNNL work described in Marrone (2016) and herein provides extensive characterization of sludge feedstock and products from HTL, introduces potential strategies for reducing ammonia levels in the HTL water stream recycled to the WWTP, and provides testing and product characterization data from hydrotreating of sludge-derived biocrude.

The purpose of this report is to document the conceptual design, economics and supporting data for a sludge HTL plant co-located with a WWTP and a centralized sludge biocrude upgrading plant for production of hydrocarbon fuel blendstocks. The design and TEA represents a goal case for the pathway, targeting performance that is anticipated to be achievable by 2022 (BETO's target year for verification of biofuel hydrocarbon pathways) with further research and development. As such, values of key design parameters, such as biocrude yield, sludge feed solids, ash content, and sludge heat transfer rate, represent improvements in the technology relative to what has currently been demonstrated in the laboratory. This work builds upon and complements the algae HTL design case (Jones et al. 2014) and the preliminary sludge HTL TEA (Snowden-Swan et al. 2016) to provide goal case economics for an example of HTL for onsite, point-of-generation conversion of distributed wet waste. In addition, the resource and fuel potential analysis illustrates potential regional biocrude collection scenarios that could enable increased fuel production by taking advantage of the low fuel production cost of the largest scale WWTPs in the nation.

1.1 Wastewater Treatment Background

Municipal wastewater treatment includes various combinations of process steps, depending on the constituents needing removal and the levels of removal required (Tchobanoglous et al. 2014). The

complete set of steps for taking sewage to clean water for discharge to the environment is commonly referred to as a "treatment train." Figure 1 shows a typical treatment train using conventional primary settling and secondary biological treatment. Preliminary treatment consists of screening of large debris and settling of grit, both of which are typically landfilled. Primary treatment involves physical settling of solids from the wastewater while secondary treatment includes biological and/or chemical means to remove/reduce biodegradable organics, as normally measured by biochemical oxygen demand (BOD) and total suspended solids (TSS). Primary sludge results from sedimentation of organic solids (e.g., toilet paper and excrement) from the wastewater. Approximately 50% of the suspended solids and 33% of the BOD are removed in the primary step (Amuda et al. 2008). Removal of FOG/scum may be carried out in separate tanks or at the primary sedimentation tanks. The scum is either sent to the plant's anaerobic digester (if present) or landfilled. Secondary treatment typically uses aerobic microbes to convert the remaining organic material into microbial biomass which is then settled and removed in the secondary clarifiers. The removed solids are commonly referred to as waste activated sludge (WAS). Typical solids concentrations of primary sludge and secondary sludge are 3% and 0.8%, respectively (Tchobanoglous et al. 2014). Secondary sludge is often concentrated to 3-4% solids using dissolved air floatation thickeners. Tertiary treatment with filtration/screening is sometimes used to remove residual suspended solids, particularly for any water reuse applications. In addition, some plants include nutrient removal in their secondary or tertiary treatment steps, especially if they are near their permitted nitrogen and/or phosphorus effluent limits (Tchobanoglous et al. 2014). Nutrient removal may be carried out either biologically or through chemical methods. More restrictive limits on effluent nutrient levels generally exist in areas near water bodies that are vulnerable to eutrophication and degraded water quality. The final step in the liquid treatment train is disinfection, after which the treated water is discharged to the environment. Various water streams resulting from thickening and dewatering of sludges and digested biosolids are generally routed back to the plant influent or "headworks" (Tchobanoglous et al. 2014). Given increasingly strict regulatory limits, the trend is toward more thorough water cleaning to remove a greater number of constitutents and at higher removal levels (Tchobanoglous et al. 2014). The industry as a whole is also moving beyond the conventional end-of-pipe treatment role to becoming sustainable resource recovery facilities of the future, driving toward options that maximize energy and nutrient recovery while producing clean water.
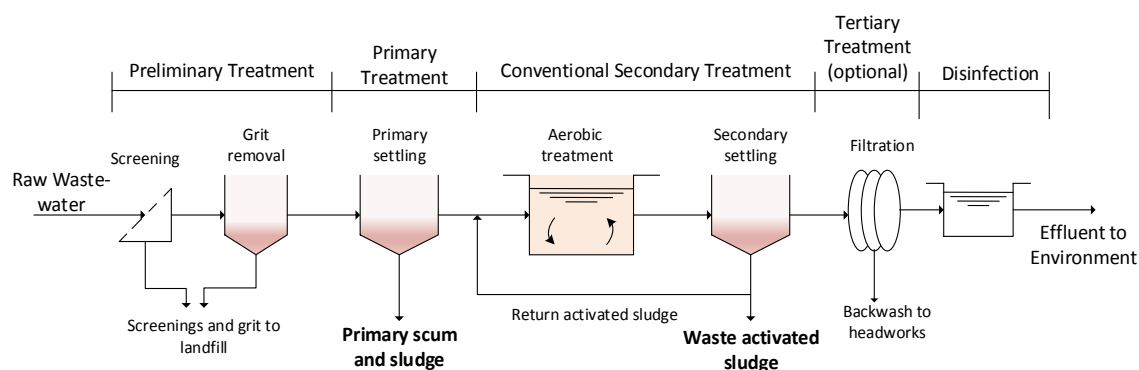
Figure 1. Typical steps for conventional biological treatment at a WWTP (adapted from Tchobanoglous et al. 2014).

The most common methods that WWTPs use to manage their sludge include stabilization and land application, landfill disposal, or incineration. Stabilization is commonly achieved through anaerobic digestion (AD), but can also be accomplished by alkaline addition, drying, or composting. The AD

process produces biogas that can be used for onsite heat or power, and "biosolids", which can be land applied as a fertilizer alternative. Land application of biosolids is regulated according to CFR 40 Part 503. The type of land to which biosolids may be applied depends on the pathogen and pollutant (metal) concentrations in the biosolids. Land application of biosolids provides benefit to soils by adding nutrients and decreasing evaporative water loss, but in some areas faces the challenge of public concern over health risks (SCAP 2013; Peccia and Westerhoff 2015). HTL could provide a more sustainable alternative than current sludge management methods, with additional benefits of relatively small footprint, high operational certainty, high percent solids reduction, and destruction of bioactive components.

1.2   Overall Sludge to Fuel Process Summary

A simple block diagram of the overall HTL plant and biocrude upgrading process scheme is shown in Figure 2. The HTL facility is assumed to be co-located with the WWTP to avoid the cost of transporting sludge. It is possible that collection of sludge within a reasonable radius within densely populated regions may be feasible and should be further investigated to enable larger economies of scale for the HTL facility. The HTL process produces an oil phase (biocrude), a solids stream containing mostly ash and some char, and an aqueous stream containing 1-3% carbon. The flow of the aqueous stream is less than 1% of the headworks flow, however it is highly concentrated in nutrients (relative to the influent) and therefore may need treatment prior to recycling back to the headworks of the WWTP. The biocrude is transported by tanker truck to a larger scale centralized upgrading plant for conversion to fuel naphtha and diesel blendstocks. The use of natural gas is assumed at both the HTL and upgrading facilities for process heat and hydrogen, respectively.
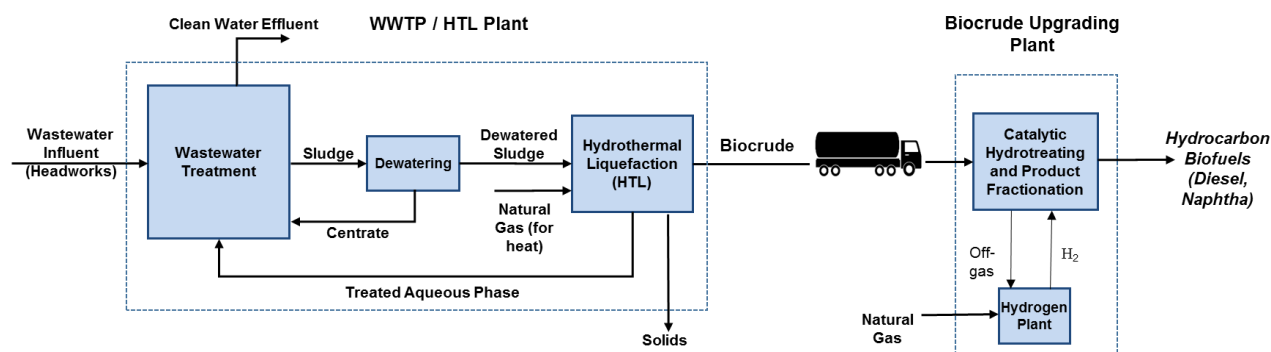
Figure 2. Simple block diagram for the WWTP/HTL plant and centralized biocrude upgrading plant.

1.3   Techno-Economic Analysis Approach

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 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 "nth plant" design, as described below.

1.3.1 Definition of Nth Plant

A standard reference basis common to the conceptual design reports, known as the "nth" plant design, is used. 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 nth 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. These assumptions are consistent across BETO design cases, thus allowing a standard basis for comparison of different conversion technologies within the context of a well-defined hypothetical plant. While WWTPs may use other economic assumptions or methods of estimating project feasibility when evaluating alternative sludge treatment technologies, this analysis should provide the key information needed to apply to a specific plant's circumstances. It is also worth noting that tax incentives and other credits that may be applicable (e.g., credits under the Renewable Fuel Standard or cellulosic biofuels bonus depreciation) but are excluded from the analysis to represent plant economics independent of any government subsidies.

Table 1. Nth-plant 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	35%
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% (7,884 operating hours per year)

(a) Modified accelerated cost recovery system

1.3.2 General Cost Estimation Basis

All costs in this report are on a 2014 constant dollar basis. This is the current reference year that BETO uses to facilitate comparison of various conversion technologies (DOE 2016). Indices used to convert capital and operating costs to 2014 dollars 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 input to Aspen Capital Cost Estimator (ACCE) software to determine the installed capital cost.

In addition, select data from commercial vendors, either as budgetary estimates or from their published literature are used when available.

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 2014 basis using the Chemical Engineering (CE) magazine's published indices:

$$\text{Cost in 2014 \$} = \text{equipment cost in quote year} \times \left(\frac{2014 \text{ index} = 576.1}{\text{quote cost year index}}\right)$$

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. Cost factors for direct and indirect costs.

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 by using the results from the Aspen Plus heat and material balances and applying the assumptions shown in Section 3.0. For the cooling tower, it is assumed that water is available at 90 °F with a 20 °F allowable temperature rise.

2.0 Process Design and Cost Estimation

This section describes the bases for the design and costing of the sludge HTL plant and the centralized biocrude upgrading plant. Detailed heat and material balances are presented in Appendix A.

2.1 Basis for Sludge Feedstock, Plant Scale, and Sludge Composition

Sludges from primary and secondary treatment, as well as biosolids from AD have all been shown to be viable feedstocks for the HTL process (Marrone 2016). A common process configuration includes both primary and secondary treatment, and therefore, mixed primary/secondary sludge was selected as the feedstock for the baseline model. While there will be variation from plant to plant in the relative proportions of primary and secondary sludge generated, a 1:1 ratio was chosen as representative of a typical plant (Turovskiy and Mathai 2006; ASCE 2000). HTL can also be applied to biosolids from AD, however direct conversion of sludge obviously results in a greater yield of liquid fuel from the plant's influent wastewater carbon.

Existing municipal wastewater treatment plants range from < 0.1 million gallon per day (MGD) to 950 MGD (EPA 2016). The largest 3% of these plants (> 13 MGD) produce approximately 60% of the total sludge generated in the U.S. (Seiple et al. 2017). As the size of WWTPs is highly variable, the scale of the HTL and upgrading plant were chosen through an iterative process whereby economics were evaluated and then sensitivity and regional analyses were performed to support a reasonable base case scale that is in context of the current WWTP size range and geographical distribution. Through this process, a base case scale of 110 dry ton/day sludge (including ash) was selected as the approximate minimum size that is economically feasible (due to economies of scale). This corresponds to a WWTP that processes about 110 MGD incoming wastewater and serves approximately 1.3 million people (EPA 2016). While this plant scale represents the minimal scale economically feasible for an individual HTL plant, the regional analysis presented in Section 4.0 explores how collecting biocrude in the most densely populated areas of the country can make smaller plants viable as part of a larger supply network to a centralized upgrading facility. It is important to note that the economics presented in this analysis reflect the cost of conventional on-site construction ("stick-built") that is typical for chemical plants and refineries. Alternatively, modularization of HTL systems is possible, which could enable smaller scale applications through reduced investment costs[1]. As an example, one company estimated that, through reduced field costs, modularization resulted in a 15% lower project cost than stick-built construction

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[1] Weber R, J Holladay, C Jenks, E Panisko, L Snowden-Swan, M Gaalswyck, M Ramirez-Corredores, B Baynes, B Rittman, M Wright, C Hamstra, L Angenent, SJ Hu, L Harmon, and D Boysen. "Modularized Production of Fuels and other Value-Added Products from Distributed, Waste or Stranded Feedstocks." Submitted to *WIREs.*

(Jameson 2007). Further work is needed to investigate and develop methods for estimating the impact of modularized manufacturing of HTL systems on the plant project economics and MFSP.

The upgrading plant is assumed to receive biocrude from 10 HTL plants within a 100-mile radius and produces 2,700 barrels per stream day (BPSD) of fuel blendstocks. Further analysis by the resource assessment team is needed to determine the optimum radius for individual regions in the country. As illustrated by the regional analysis presented in Section 4.2, this scale and associated biocrude intermediate availability is reasonable in several of the most densely populated regions of the country. For perspective, the upgrader scale is 54% of other BETO design cases of ~5,000 BPSD fuel production and only 5% of the average scale for combined gasoline and diesel production at U.S. refineries of about 51,000 BPSD (EIA 2016).

Sludge treatment and disposal currently represents a significant expense to WWTPs. Therefore, the solids reduction resulting from the HTL process would conceivably result in significant savings in avoided disposal costs to the plant. However, the base case model does not include this savings (i.e., a zero feedstock cost is used) in order to account for potential market adjustments that may result from implementation of this technology in the future. Sensitivity analysis around this assumption is presented in Section 4.0.

Wastewater treatment sludge is composed of water, three basic types of organic macromolecules: proteins, lipids/fats, and carbohydrates, as well as some inorganic material (ash and grit). Sludge composition will vary from plant to plant depending on the treatment processes used, types of industrial dischargers in the region (i.e., sewage chemistry), weather, and other factors. Compositional data on wastewater sludge is sparse in the literature and those that are available are highly variable and often do not indicate the source of the sludge (e.g., primary, secondary, mixed, or biosolids). While lipid, carbohydrate and protein content will vary to some degree from plant to plant, ash content is perhaps the single most important compositional parameter, as it most directly impacts the overall sludge to biocrude yield for a given plant size. In other words, the ash in the sludge increases equipment size and therefore capital expense, but does not contribute any fuel product. Table 3 lists ranges of sludge ash content found in the literature indicating a wide range of 15-40% for mixed primary/secondary.

Table 3. Municipal sludge ash content (% of total solids)

Primary	Secondary	Mixed Sludge	Source[a]
36	14	25	This work
8	16	12[b]	Marrone 2016
35	23-33	28	Manara and Zabaniotou 2012
20-30	NR	15-25	Pennsylvania DEP 2001
10-30	25-35	18-33[b]	New Mexico Environment Department 2007
15-40	15-40	15-40[b]	Tchobanoglous et al. 2014
20-40	15-40	18-40[b]	ASCE 2000
20-25	20-25	20-25[b]	von Sperling and Gonçalves 2007
	20-30		WEF 2013

(a) All sources except this work and Marrone reported as % volatile matter (VM). Ash is calculated as 100-%VM.
(b) Calculated average mix of 1:1 primary, secondary mixture.

Ultimate analysis for a 1:1 mixture of primary/secondary sludge provided by the City of Detroit and Great Lakes Water Authority for HTL testing is given in Table 4. This composition is used for the modeled sludge feedstock with the exception of ash content, which was decreased from 26% to 15% for the goal case. A lower ash content was chosen based on two reasons: 1) process knowledge from Detroit indicates their ash content is likely on the high end of the potential range due to aging infrastructure and addition of $FeCl_3$ for phosphorus removal in its primary sedimentation step; and 2) the goal case assumes that future renovations in aging collection systems and outdated processes will lead to reduced sludge ash content over time. Ideally, enhanced de-gritting and de-ashing technology should also be pursued for pretreatment of the sludge. The cost of additional processing steps needed for producing a lower ash feed should be considered in future analyses. Sensitivity analysis around ash content is presented in Section 4.0.

Table 4. Sludge composition and ash content.

50/50 Primary/Secondary Sludge Mixture Characteristics	Experimental Data	Model	Experimental & Model
Component	Wt% dry basis	Wt% dry basis	Wt% dry, ash free basis
C	41.06	46.83	52.11
H	5.67	6.46	7.19
O	26.06	29.72	33.07
N	4.98	5.68	6.32
S	1.03	1.170	1.30
Ash	26.08	15.0	
P	1.86	1.86	
HHV Btu/lb[a]			9,936

(a) Calculated by the Boie Equation: HHV (Btu/lb) = (151.2 C + 499.77 H +45.0 S -47.7 O + 27 N) *100 - 189.0.

2.2 Feedstock Preparation

Dewatering of sludge feedstock is necessary to minimize capital and operating costs for the HTL process. This section describes the equipment and associated capital necessary to prepare the sludge slurry feedstock.

Figure 3 shows a simple diagram of the proposed feed preparation steps. Sludges are generated from the WWTP's primary and secondary treatment steps at 3% and 0.8% solids content, respectively. Primary sludge and secondary sludge may be thickened separately before they are mixed as part of normal WWTP operations. For this analysis, it is assumed that the mixed sludge is intercepted and mechanically dewatered to the target solids content of 25%. This target is considered the maximum that is attainable with conventional dewatering technology while still enabling effective pumping through the HTL system. Secondary sludge is typically more difficult to dewater than primary sludge because of interstitial water contained in cell mass, therefore mixing the sludge prior to dewatering is likely to produce the target solids content.

Dewatering of sludges and biosolids is currently carried out to varying degrees at WWTPs to minimize costs for treatment, transport, and disposition. Common units for dewatering to a high solids content are the centrifuge, belt filter press, and screw press. Although dewatering equipment are likely to already be present at the WWTP, capital for investing in a new centrifuge is included in the analysis to

account for situations in which the existing dewatering unit is toward the end of its life and/or is not a large enough capacity to handle raw sludges (e.g., in the case of replacing AD). Chemical conditioning of the secondary sludge is assumed to be required to maximize dewatering efficiency. Water soluble polymers, or organic polyelectrolytes, are the most commonly used conditioning agents for mechanical dewatering and act as flocculants that enhance solid/liquid separation in the sludge (Tchobanoglous et al. 2014). Cationic polyacrylamide is the backbone of most polymer used in the industry (Tchobanoglous et al. 2014). Polymers provide added benefit over inorganic conditioners by not adding ash to the sludge.
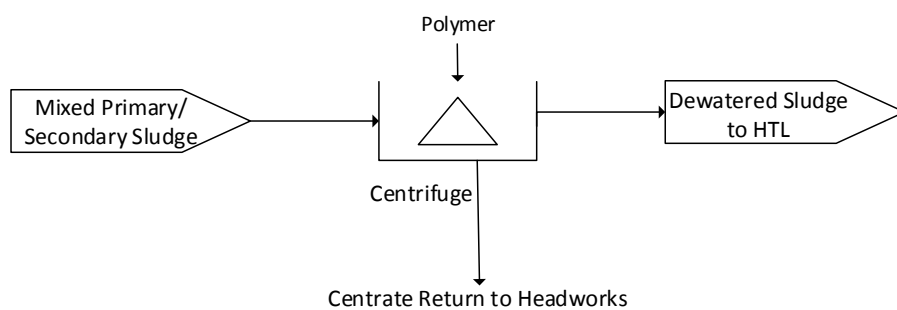
Figure 3. Process flow for slurry feed preparation.

The capital cost for the centrifuge is based on a vendor budgetary estimate. The installed equipment cost for the centrifuge is $1.4 million (2014 $) for 30,000 gal/hr feed capacity. Individual plants have differing configurations and therefore slight modifications to capital investment and operating costs are to be expected. Also, certain configurations of biological treatment, such as the use of membrane bioreactors in lieu of the conventional activated sludge process, or the use of nutrient removal, make sludge that is difficult to dewater (Tchobanoglous et al. 2014). Centrifugation of a mixture of primary sludge/WAS of 2-4% solids content, with added polymer at 5-16 lb/dry ton dry solids, is expected to achieve 25-35% solids content in the product (Stubbart et al. 2006).

2.3 Hydrothermal Liquefaction

The main process steps in the HTL section of the plant are shown in Figure 4. The base case plant processes 100 dry tons per day of sludge. The modeled process is largely based on previous conceptual designs for wood HTL and algae HTL (Knorr et al. 2013; Jones et al. 2014). The 25% solids ground slurry feed is pumped to 2900-3000 psia and then preheated to 550°F (288°C) in two double-pipe heat exchangers in series using heat from the reactor liquid product (biocrude/aqueous mixture). High-pressure pumping of biomass slurries has only been demonstrated at bench and smaller pilot scale, however, it is generally thought to be more easily accomplished at large scale (Elliott 2011). Berglin et al. (2012) indicated that pumping a 15% solids wood slurry at HTL pressures is viable with commercial off-the-shelf equipment, such as a twin-screw feeder/piston pump type of system. Sludge is more easily pumpable than wood slurry and therefore should be possible at scale and at the higher feed solids assumed in this analysis. A fired heater with hot oil system is used to bring the pressurized feed up to the reactor temperature of 656°F (347°C) and provide reactor heat. Off-gas from the reactor along with purchased natural gas are burned in the fired heater to provide this heat. Scrubbing of the off-gas may be required, depending on local permit requirements. Approximately 20% of the heat is provided by the HTL off-gas, which is composed of 92% CO_2 and 8% C_1-C_5 gasses).

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 and a small amount of gases. Water at subcritical conditions has a much lower dielectric constant and higher ion product than water at normal conditions and therefore provides a reaction medium with improved solvent and catalytic properties (Elliott 2011; Elliott 2015). The HTL chemistry is complicated, however the general reaction pathways can be put into three basic categories: 1) depolymerization of the biomass components, 2) decomposition of biomass monomers by cleavage, dehydration, decarboxylization and deamination, and 3) recombination of reactive fragments (Toor et al. 2012). The biocrude from sludge is similar to biocrude from algae HTL and comprises a mixture of fatty acids, amides, ketones, hydrocarbons, phenols, alcohols and other components.

The HTL reactor effluent is fed to a hot filter where solids are removed. The solids stream consists of 60-70% water, ash, char, and low levels of organics from the aqueous phase and biocrude phase. 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 mass balance from testing indicates that the solid phase from testing of Detroit sludge (20% feed solids and 26% ash) contained 7-9% of the total produced biocrude. It is estimated that about half of this biocrude can be recovered through further work on improved separations methods. The solids are assumed to be disposed of in a landfill. Solids from experimental HTL testing were analyzed for hazardous metals content with the toxicity characteristic leaching procedure (TCLP) and determined to be under the federal regulatory limits. While these results indicate that HTL solids are non-hazardous, the metals content will depend on the incoming sludge composition and therefore TCLP testing would need to be performed at each plant and probably on a periodic basis to verify that the solids can be landfilled.. Metals tested include arsenic, barium, cadmium, chromium, copper, lead, mercury, nickel, selenium, silver, and zinc (see Appendix E). 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 explore the technical and economical feasibility of this option.

After solids separation, the remaining biocrude-aqueous-gas mixture is cooled to 140°F (60°C), reduced to 30 psia, and separated in a three-phase separator. The biocrude product is stored and then shipped to a centralized upgrading facility where biocrude is assumed to be collected from multiple HTL plants in the area. The aqueous stream, which contains effluent water, any remaining soluble organics, ammonia, and metal salts, is routed to the HTL aqueous phase treatment section of the plant (see section 2.4).
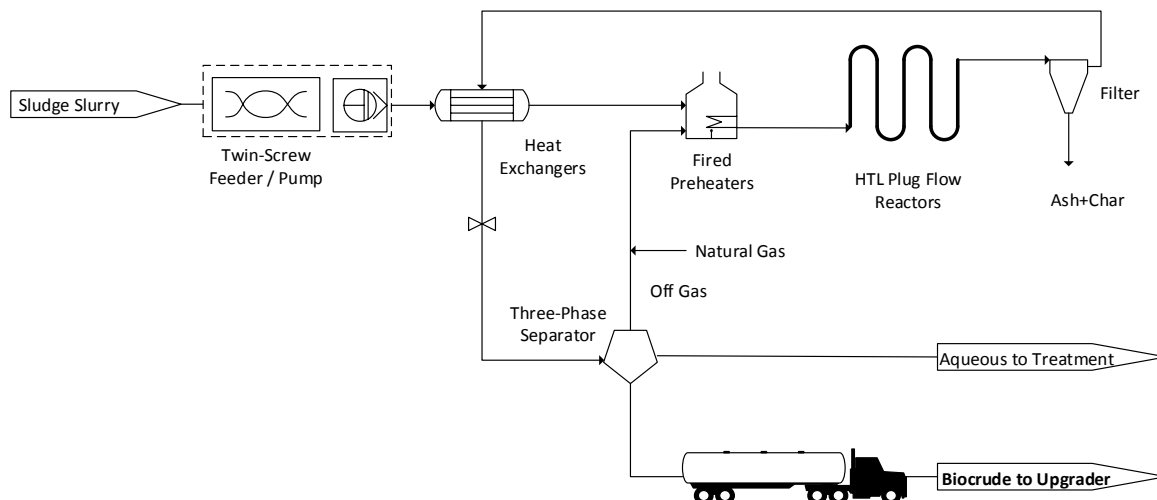
Figure 4. Process flow for sludge HTL.

2.3.1 Sludge HTL Design Basis

Table 5 shows the experimental testing conditions and product results for the primary/secondary sludge and the modeled parameters used for the goal case design. The experimental data were collected from PNNL's bench scale system similar to previous studies (Elliott et al. 2013; Marrone 2016). Biocrude yield and feed solids content are important cost drivers that can be improved with further research. As such, these parameters have been improved for the goal case relative to the experimental results, with values the experimental team expects can be realized with further research. While there are potentially thousands of individual compounds present in the biocrude, a simplified list of chemical compounds is used in the process model. This analysis assumes the same set of compounds modeled in the previous analysis for algae HTL and biocrude upgrading (Jones et al. 2014), as listed in Appendix C. This is a reasonable assumption, as biocrude from wastewater sludge has been shown to be similar in composition to biocrude from algae (Jarvis et al. 2017).

Table 5. Sludge HTL experimental results and model assumptions.

Operating Conditions and Results	Experimental Results (WW06 SS-2)	Aspen Model Assumption (Goal Case)
Temperature, °F (°C)	656 (347)	656 (347)
Pressure, psia (MPa)	2979 (20.5)	2979 (20.5)
Feed solids, wt%		
Ash included	20%	25%
Ash-free basis	15%	21%
LHSV, vol./h per vol. reactor	3.6 Hybrid CSTR-PFR[d]	6 PFR
Equivalent residence time, minutes	17	10
Product yields[a] (dry, ash free sludge), wt%		
Oil	44%	48%
Aqueous	31%	25%
Gas	16%	16%
Solids	9%	11%
HTL dry oil analysis, wt%		
C	78.5%	78.3%
H	10.7%	10.8%
O	4.7%	4.8%
N	4.8%	4.9%
S	1.2%	1.2%
P	0.0	Not modeled[b]
Ash	0.06%	0.0%
HTL dry oil H:C Ratio	1.6	1.6
HTL oil dry HHV, Btu/lb (MJ/kg)	17,000 (39.5)[c]	17,100 (39.7)
HTL oil moisture, wt%	4.4 wt%	4.0 wt%
HTL oil wet density	0.98	0.98
Aqueous phase COD (mg/L)	61,300	61,100
Aqueous phase density (g/ml)	1.0	0.98 Aspen est.

(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.

2.3.2 HTL Capital Costs

The HTL section capital costs are primarily based on estimates developed by Knorr et al. (2013) for a wood HTL system under the National Advanced Biofuels Consortium (NABC), and scaled to the rate of the sludge HTL plant. Major equipment costs adapted from the NABC work include the feed/product exchangers, trim preheater, hot oil system, solids filter, and oil/water separator. Due to the corrosive nature of the HTL environment, stainless steel 316-L is selected for process equipment The experimental HTL system is constructed of stainless steel 316-L (with the exception of the CSTR, which is Inconel), for which no measurable corrosion has been detected during testing of sludges. In addition, materials of contruction testing, including systematic ex-situ coupon testing, in-situ coupon testing, and destructive evaluation of system components indicated no measurable corrosion of stainless steel 316-L coupons during 1000 hours exposure time. Feed/product heat exchange is modeled using two double-pipe type exchangers in series. Attempts to obtain budgetary estimates for these exchangers were unsuccessful and therefore capital costs are scaled on estimates provided by Knorr et al. (2013). The feed/product

exchanger areas are adjusted based on a more optimistic heat transfer rate that is thought to be attainable for the goal case. The heat transfer coefficient is dependent on the physical properties of the sludge (e.g., viscosity, heat capacity, density) as well as flow properties (e.g., pipe diameter, line velocity). Testing of the PNNL bench scale tube-in-tube heat exchanger has shown an overall heat transfer coefficient (U) of 50 Btu/hr/ft^2/°F for heating a 15% wood slurry from 77°F to 549°F (25-287°C) with hot oil. The general trend in viscosity among feedstock slurries at similar solids content has been observed in PNNL testing as algae<sludge<wood and therefore a higher heat rate is expected for sludge compare to wood. In addition, sludge viscosity should decrease with shear rate and consequently scaled up systems should provide for higher heat rates. With higher line velocities and higher Reynolds Numbers, an overall U of 100 Btu/hr/ft^2/°F for the combined feed/product exchangers is considered achievable in the goal case timeframe. Further work is needed to characterize the rheology of the sludge at relevant line velocities and temperature ranges, investigate configurations/strategies for improved sludge heating, evaluate the effect of fouling on the heat transfer coefficient and ultimately demonstrate the achievable heat exchange rate at larger than bench scale.

The HTL reactor is a jacketed serpentine pipe with heating oil in the annular space, similar to a double-pipe heat exchanger. To minimize pressure drop, a configuration of two reactors in parallel is assumed, similar to the design in Knorr et al. (2013). The developed parameters for each reactor are listed in Table 6. ACCE was used to estimate the equipment cost of a fully jacketed pipe with elbow fittings for the reactor. Based on the PNNL experimental runs to date, an oil jacket is recommended to minimize heat losses over the length of the reactor. The reactor design in Knorr et al. (2013) also assumed a serpentine pipe type reactor, but without an oil jacket. Therefore, the capital estimate presented in this analysis is slightly more expensive than the Knorr design.

Table 6. HTL reactor specifications.

Spec	Value	Spec	Value
Feed Rate, lb/hr (gal/hr)	18,300 (2,100)	Overall Length, ft	530
LHSV, hr-1	6	Line velocity, ft/min	53
Reactor Volume, gal	350	Surface Area, ft^2	550
Inside Pipe Diameter, inch	4	U, Btu/hr/ft^2/°F	154 (Knorr et al. 2013)
Pipe Thickness, inch	0.67	Q, mmBtu/hr	2.4
Pipe Material	SS316	Pipe section length, ft	25
Pipe Schedule	XXH	Number of pipe sections	21

Note: Two of the specified reactors are used in parallel for the process.

The capital costs for each of the major process components of the HTL section are given in Table 7. Individual equipment costs as well as scaling assumptions and installation factors can be found in Appendix B.

Table 7. Sludge HTL capital costs.

Item	Purchased, million USD (2014)	Installed, million USD (2014)	Source
HTL Reactor System: Pumps, heat integration, HTL reactor, knockout drums	$5.67	$11.34	Knorr et al. 2013; ACCE (reactor)
Phase separation	$0.47	$1.12	ACCE (product coolers); Knorr et al. 2013 (separators)
Hot oil system for reactor and trim heater	$0.48	$0.64	Knorr et al. 2013
Total	**$6.6**	**$13.1**	

2.4 HTL Aqueous Phase Treatment

2.4.1 Aqueous Treatment Design Basis

After separation from the biocrude product, the HTL aqueous phase is routed back to the WWTP's treatment train. Table 8 lists the measured HTL aqueous phase constituent concentrations from the experimental run with Detroit sludge. While the organic and ammonia levels in the HTL water are high relative to a typical WWTP influent, the overall volume is small. Table 9 provides a comparison of the component loadings from the HTL water versus the average WWTP headworks (calculated for 100 MGD at concentrations from Tchobanoglous et al. (2014). As shown in Table 9, the HTL water contributes 9% and 18% of the BOD and ammonia load to the WWTP, respectively. The organic content, which is essentially all dissolved, would likely be processed without issue by the activated sludge process (note that the extra aerating power required is included in the analysis). The high ammonia, on the other hand, could cause problems at some plants, such as toxicity impacts and/or pass through to the WWTP's discharge to the environment. Significantly high phosphate levels may also pose a problem. Some plants may be able to recycle the HTL aqueous stream directly back to primary or secondary treatment without pre-treatment. Others, particularly those that are already near their permitted effluent nitrogen and/or phosphorus limit and/or do not already have an enhanced nutrient removal step in the main treatment train, might have to treat this stream to reduce loads on the plant. In any case, it is anticipated that effluent limits will become increasingly stringent and nutrient recovery will become more important for WWTPs in the future. Therefore, as a conservative measure, treatment of the HTL water is included in the design.

Table 8. HTL aqueous phase constituent analysis from experimental run

Constituent	mg/L[a]
BOD[d]	22,900
COD	61,300
COD$_s$	Not measured
TOC (total organic carbon)	20,550
Total Carbon	24,700
TKN[b] (total Keldahl nitrogen)	7,670
TSS (total suspended solids)	Not measured
Nitrogen (total)	7,700
Free ammonia-Nitrogen	4,900
Calcium	<1
Phosphorus (total)	11
Potassium	280
Sodium	140
Sulfur	461
Silicon	294
Chloride	280
Nitrate	36
Phosphate	<10
Sulfate	196
Alkalinity as CaCO3, total[b]	15,700
Dissolved Organics	
Acetic Acid	5,450
Acetone	1,090
Ethanol	240
Methanol	6,870
P-Cresol	1,170
Propanoic acid	2,040
Phenolics, total recoverable[b]	69
pH	7.8

(a) Detroit sludge; WW-06 SS-2 (20% solids feed; LHSV=4 hr^{-1})
(b) Detroit sludge; WW-06 (composite of LHSV 2 hr^{-1} and 4 hr^{-1})

Table 9. Comparison of HTL effluent water and WWTP headworks component flows.

	Untreated WWTP Headworks Wastewater	HTL Aqueous Phase Effluent	Contribution of HTL Effluent water to Headworks Loadings(%)
Flow Rate for 100 MGD WWTP, gal/min	69,444	67[a]	0.2%
Constituent	lb/day[b]	lb/day	% Contribution
BOD[d]	166,907	16,512	9%
COD	423,945	44,200	9%
COD_s	147,713	Not measured	Not measured
TOC (total organic carbon)	136,864	14,817	10%
Total Carbon	Not listed	17,810	Not listed
TKN[d] (total Keldahl nitrogen)	29,209	5,530	16%
TSS (total suspended solids)	162,735	Not measured	Not measured
Nitrogen (total)	29,209	5,552	16%
Free ammonia-Nitrogen	16,691	3,533	17%
Calcium	Not listed	<0.7	Not listed
Phosphorus (total)	4,673	8	0.2%
Potassium	13,353	202	1.5%
Sodium	Not listed	101	Not listed
Sulfur	Not listed	332	Not listed
Silicon	Not listed	212	Not listed
Chloride	49,238	202	0.4%
Nitrate	0	26	100%
Phosphate	653	<7	1%
Sulfate	30,043	141	0.5%
Alkalinity as CaCO3, total[d]	Not listed	11,320	Not listed

(a) Using Table 8 data and assuming 100 dry ton/day sludge is generated
(b) Based on concentrations given in Tchobanoglous et al. 2014

Possible treatment options for ammonia removal from the HTL aqueous stream include physical methods, such as ammonia stripping or membranes, as well as biological methods. An air stripping system was selected for the design as this option is well established and reliable for lowering the ammonia content of wastewater (EPA 2000). As shown in Figure 5, the HTL aqueous stream is initially treated with lime (CaO) to raise the pH to 11 and shift the NH_3/NH_4^+ equilibrium to the gas phase. Lime treatment causes $CaCO_3$ and $CaPO_4$ to precipitate and is therefore effective for removing phosphate, if present. After clarification of precipitated $CaCO_3$ and $CaPO_4$, the effluent water containing dissolved ammonia gas is contacted with an air stream in a packed tower where the ammonia gas is stripped from the wastewater. Preliminary modeling indicates that air stripping will also remove volatile oxygenates (e.g., ethanol, acetone, acetic acid, butanol) from the wastewater at levels that are likely too high to recover a marketable ammonia product. Therefore, the ammonia/air stream is treated with a thermal oxidation unit (THROX) where ammonia and organics are catalytically combusted to N_2, CO_2 and H_2O and released to the atmosphere.

A membrane system and biological based methods were also examined as possible options for ammonia removal from the HTL aqueous stream. It is possible a membrane system could be used to recover saleable ammonia byproduct, but the efficacy of a membrane is highly uncertain as fouling from precipitated salts (following pH adjustment) is a significant risk. Biological treatment systems for the anaerobic oxidation of ammonium (anammox) in WWTP sidestreams have been installed at several facilities in the U.S. and Europe WWTP (Pugh 2015). Preliminary discussions with vendors indicate that a hybrid system consisting of an anaerobic blanket and anammox (Lu et al. 2016) is a plausible option. However, this option does not allow for nutrient product recovery as ammonia is oxidized to nitrogen gas in the biological process. Recovery of a nitrogen product pure enough to be saleable will likely require removal of the organics prior to the ammonia recovery step. For example, catalytic hydrothermal gasification (CHG), which can be used to convert organics in wastewater to methane (Jones et al. 2014), could be coupled with air stripping or a membrane system to recover a pure ammonia byproduct.
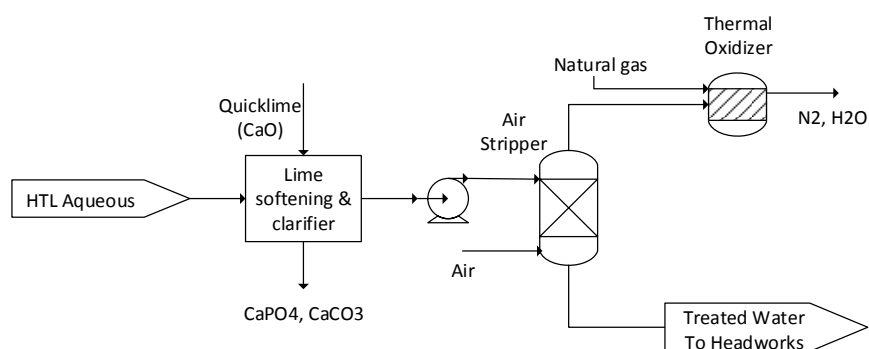
Figure 5. Process flow for HTL aqueous phase treatment.

Table 10 summarizes some advantages and disadvantages of each treatment method considered for this initial technology screening. Note that this list is not intended to be comprehensive, as there are likely other possible treatment options that could not be included in this initial screening. The treatment scheme using ammonia stripping is preliminary and as such, further experimental testing and analysis is needed to test its feasibility and to determine the most appropriate and economical treatment option for the HTL aqueous phase. Additional work is also needed to determine the overall impact of recycling the HTL water on the WWTP. For example, there may be components other than ammonia that could build up in the HTL water and at the least, require a purge.

Table 10. Possible HTL water ammonia removal/recovery options.

Option	Pros	Cons
No Treatment (direct recycle to WWTP)	Little/no cost	May cause WWTP effluent nitrogen level to increase.
NH3 Stripping	Conventional technology; Potential recovery of NH_3 byproduct	Organics in NH_3 stream may be too high for saleable product; destruction of NH_3 is expensive and GHG-intensive; Does not remove other nitrogen.
Membrane	Off-the-shelf; Potential recovery of NH_3 byproduct; Simple system; small footprint	Potential membrane fouling with organics and/or suspended solids; Organics in NH_3 stream may be too high for saleable product; Does not remove other nitrogen.
Biological (Anammox or Nitrification-Denitrification)	Fairly well-established for WWTP AD centrate treatment	Additional biological operations and associated potential for upsets; May require methanol (for nitrification-denitrification; No option for product recovery (produces N_2 gas).

2.4.2   Aqueous Treatment Capital Costs

A separate, preliminary model for the aqueous phase treatment section was developed with Aspen's electrolyte package to enable more accurate prediction of lime requirements for pH adjustment and get initial estimates for capital and operating costs. Capital costs for the lime softening, air stripping, and thermal oxidation equipment were estimated with ACCE. Table 11 shows the purchased and installed cost for the complete treatment process.

Table 11. HTL aqueous phase ammonia removal system capital costs

Equipment	Purchased, million USD (2014)	Installed, million USD (2014)	Source
Lime softening and clarification	$0.43	$1.04	ACCE, v. 8.8
Air stripping	$0.63	$1.51	ACCE v. 8.8
Thermal oxidation	$0.22	$0.53	ACCE, v. 8.8
Total	**$1.3**	**$3.1**	

2.5   Centralized Biocrude Upgrading

The HTL biocrude requires catalytic processing to remove oxygen, nitrogen, and sulfur before it can be used as fuel blendstock. Biocrude produced from multiple plants within a geographical area is collected at a central facility where it is hydroprocessed and finished into fuel products. The upgrading plant is assumed to process 2,700 BPSD of biocrude, which is equivalent to the output of ten HTL facilities, each processing 110 dry ton/day sludge. The upgrading plant flow diagram is shown in Figure 6 and based on that modeled in Jones et al. (2014).   Prior to upgrading, a desalting process similar to that used in a petroleum refinery is used to remove inorganic components from the biocrude to prevent deactivation of the hydrotreating catalyst. The biocrude is then pumped to ~1500 psia, mixed with compressed hydrogen, and preheated to the hydrotreater reactor temperature of ~750°F (400°C). Hydrogen is produced onsite via steam reforming of the process off-gas and additional purchased natural

gas (see Section 2.5.3). A guard bed directly upstream of the hydrotreater is used to remove soluble iron that may be present in the biocrude. In contrast to algae HTL biocrude (Jarvis et al. 2016), no solubilized iron complexes were observed in the sludge HTL biocrude and therefore a filter may be sufficient for mineral removal in lieu of a guard bed in the future. However, there may be other mineral compounds present that are unlikely to be captured by a filter. Further investigation is needed in this area to verify effective biocrude cleanup strategies. 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 resulting hydrocarbon product consists of a mixture of paraffins, olefins, naphthenes and aromatics that lie within the gasoline, jet and diesel boiling ranges. The hydrotreater reactor effluent is cooled to condense the produced water and hydrocarbons and the organic phase is fractionated into four boiling point cuts: C4 minus, naphtha range, diesel range and heavy oil range material. The heavy cut is cracked in a conventional hydrocracker to produce additional naphtha and diesel range products. As product cut analysis is not yet available, simulated distillation data are used to estimate the volumes of naphtha, diesel and heavy oil boiling range material. The wastewater from the process is high in ammonia and therefore will likely require treatment prior to sending to a WWTP. A stripping system similar to that described for HTL water treatment is assumed for the upgrading plant.
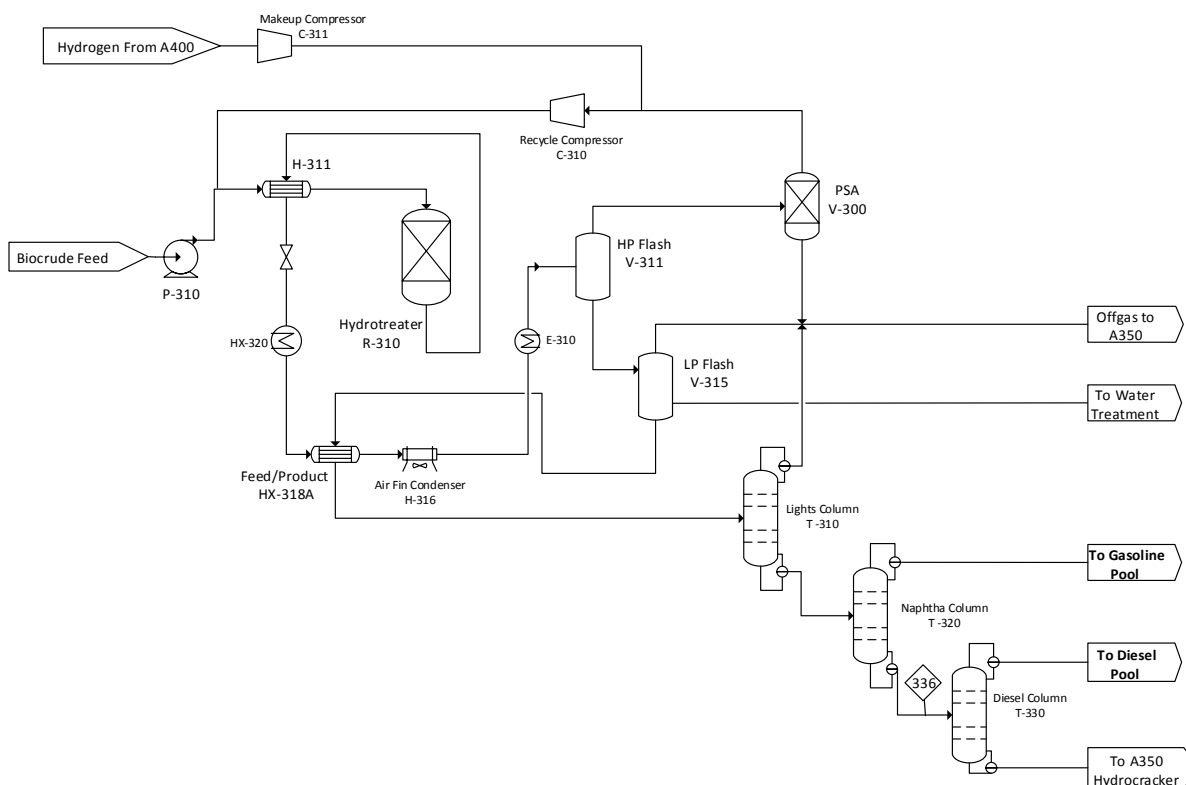
Figure 6. Process flow for HTL biocrude hydrotreating.

2.5.1   Biocrude Hydroprocessing Design Basis

The design basis for hydrotreating is presented in Table 12. The experimental data was collected from bench scale flow reactors at PNNL. The modeled goal case assumes improved hydrotreated product yield and reactor space velocity relative to that demonstrated in the testing. Based on the experience of

the experimental team, these values are expected to be attainable within the time frame of the goal case with further research and development.

Table 12. Sludge biocrude hydrotreating experimental results and model assumptions.

Component	Experimental (HT-62005-60)	Model (goal case)
Temperature, °F (°C)	752 (400)	752 (400)
Pressure, psia	1540	1515
Catalyst	CoMo/alumina-F	CoMo/alumina
Sulfided?	Yes	Purchased presulfided
LHSV, vol./hour per vol. catalyst	0.18	0.50
WHSV, wt./hr per wt. catalyst	0.29	0.81
HTL biocrude feed rate, ml/h	5.6	Commercial scale
Total continuous run time, hours	302 (total run)	Not applicable
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[a]	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%	0.17%
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) Yield after phase separation		

The hydrocarbon compounds used to model the hydrotreated product are similar to those used for previous algae HTL modeling (Jones et al. 2014) and are listed in Appendix C. Figure 7 shows the boiling point curves from simulated distillation (ASTM Method D2887) for the hydrotreated product from experimental testing and the modeled hydrotreated product. Actual distillation was also performed on the hydrotreated product and matched well with the D2887 results.
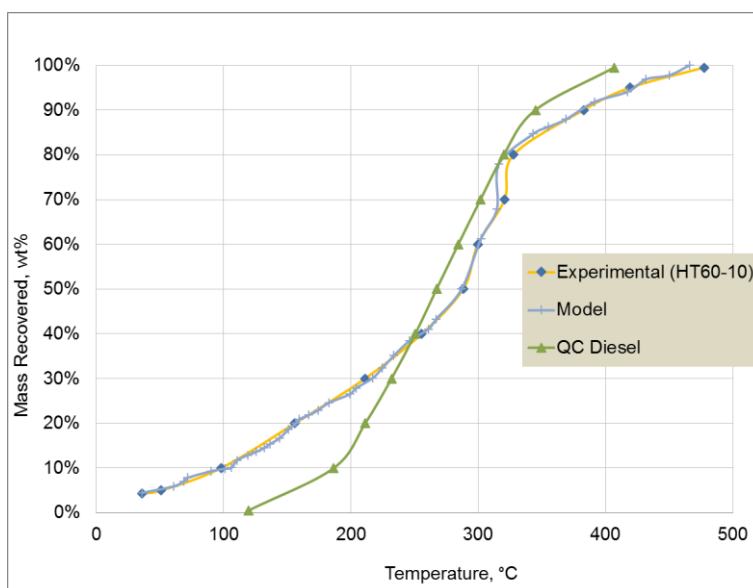
Figure 7. Boiling point distribution (ASTM D2887) for hydrotreated product from sludge biocrude.

The heavy oil fraction from hydrotreating is assumed to be hydrocracked into additional gasoline and diesel range fuel. No experimental data are yet available for hydrocracking of hydrotreated HTL biocrude, so it is assumed that the heavy cut is processed similar to petroleum operations. The hydrocracking assumptions for the model are given in Table 13.

Table 13. Hydrocracking model assumptions.

Process	Basis	Assumptions
Hydrocracking heavier than diesel portion of hydrotreated HTL biocrude	No experimental data, assumed to be similar to conventional hydrocrackers LHSV=1 Temperature: 734 °F (390 °C) Pressure: 1035 psia	H_2 chemical consumption: 0.004 wt/wt heavy oil Product breakdown: Gas (excluding excess H_2); 3 wt% Liquid fuels: 96 wt% Aqueous: 1 wt%

2.5.2 Biocrude Hydroprocessing Capital Costs

Table 14 shows the capital costs for the hydrotreater and hydrocracker systems, which are based on a review of conventional naphtha, diesel and kerosene hydrotreaters as published in the IHS 2014 Yearbook and from Petroleum Refining Technology and Economics (Gary 2007). Additional cost details can be found in Appendix B.

Table 14. Hydrotreater and hydrocracker capital costs.

Item or Area	Purchased, million USD (2014)	Installed, million USD (2014)	Source
Hydrotreater system (2732 BPSD feed)	21.1	33.6	IHS 2014a
Hydrocracker system (1020 BPSD feed)	4.4	6.6	IHS 2014a

The costs spanned a broad range from simple naphtha hydrodesulfurization units, to multi-stage hydrocrackers. In addition to the reactor(s), each system at least includes recycle compressors, multi-stage flash systems and distillation. The hydrotreater system also includes the cost of a desalter and guard bed. Costs for a generic hydrocracking system (2000 psia) were chosen as the basis for hydrotreating and hydrocracking. While these costs are generally applicable to HTL oil hydrotreating as they employ similar temperatures and pressures, conventional refining space velocities tend to be higher. Reactor cost sensitivity is considered in Section 4.0.

2.5.3   Hydrogen Generation Design and Cost Basis

The hydrogen plant is a conventional natural gas based steam reformer. Most of the off-gas is used to fire the reformer. However, a portion of the off-gas is compressed and mixed with makeup natural gas which is then sent to a hydrodesulfurizer (HDS) unit. Figure 8 (Jones et al. 2014) shows the simplified flow scheme for hydrogen generation by steam reforming of natural gas (IHS 2014b; Meyers 2004; H_2A 2013) combined with the off-gas streams from hydrotreating and hydrocracking. Hydrogen for the HDS unit is supplied by the off-gas stream. The gas exiting the HDS unit is then mixed with superheated steam and sent through an adiabatic pre-reformer to convert C_2+ compounds to methane prior to entering the main steam reformer to produce syngas. This reduces the rate of coking in the main reformer. The syngas hydrogen content is increased by high temperature shift. After condensing out the water, the hydrogen is purified by pressure swing adsorption (PSA). Off-gas from the PSA is recycled to the reformer burners.
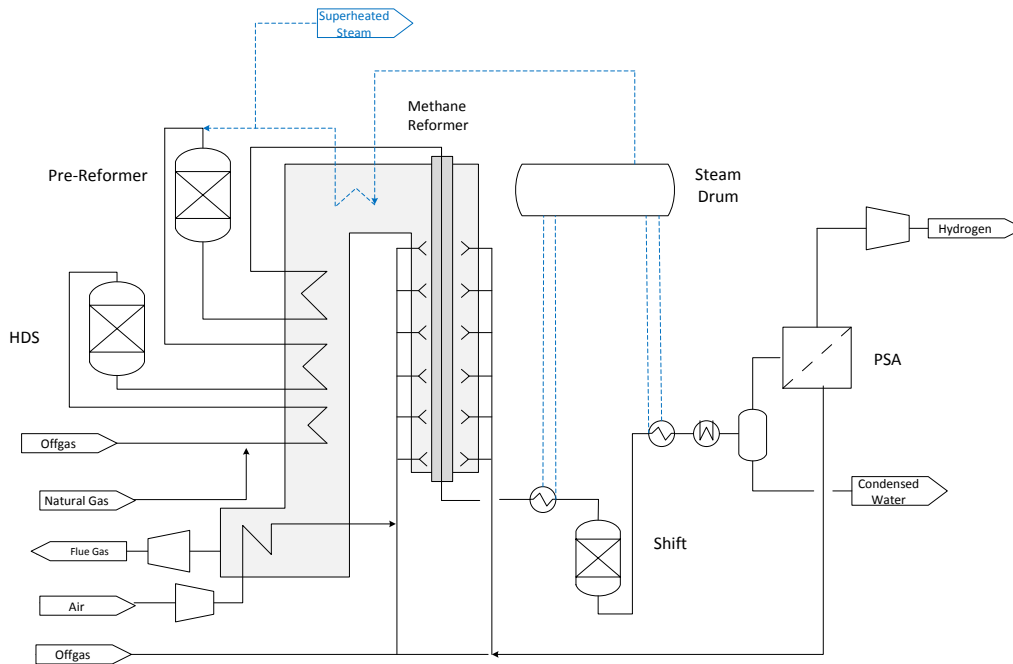
Figure 8. Process flow for hydrogen production (from Jones et al. 2014).

Saturated and superheated steams are generated by recuperating heat from the reformer exhaust and cooling the product from the water gas shift reactor. The generated steam is used in the reformer and also to provide process heat, including the distillation column reboilers.

The design assumptions are shown in Table 15. Gibbs minimization reactors are used to model the pre-reforming, methane reforming and burner reactions. The reactor methane conversion of 80 mole % matches that reported by IHS (2014b).

Table 15. Hydrogen plant model assumptions.

Equipment	Assumptions	
Pre-reformer	Outlet temperature	925 °F (496 °C)
	Outlet pressure	429 psia
	Steam/carbon ratio	3.5
Methane Reformer	Steam pressure	670 psia
	Outlet temperature	1562 °F (850 °C)
	Outlet pressure	399 psia
Burners	Bridge wall temperature	1800 °F (982 °C)
	Pressure	Slightly positive
Shift Reactor	Outlet temperature	568 °F (300 °C)
	Outlet pressure	388 psia
	Approach to equilibrium	98%
PSA	Hydrogen delivery pressure	376 psia
	Hydrogen recovery	90%

Capital costs for hydrogen generation are taken from the IHS PEP 2014 Yearbook and scaled to the necessary hydrogen production rate using the IHS scale factor. The equipment includes a sulfur guard

bed, pre-reformer, primary reformer with nickel catalyst, high temperature shift reactor, pressure swing adsorption unit, waste heat recovery producing high pressure steam and all associated outside battery limit equipment. Conventional steam reformer hydrogen plants range in scale from 1 to 100 million scf of hydrogen per day. The hydrogen plant scale needed for the centralized biocrude upgrading plant is at the low end: 9 million scf per day. The purchased capital and installed capital cost for this plant is $14.2 million and $27.2 million (in 2014 dollars), respectively. Additional cost details can be found in Appendix B.

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 HTL and biocrude upgrading plants. Discounted cash flow rate of return analysis is then used to determine the fuel production cost using standard methodology used for all BETO design cases. The summary economics and performance for the HTL plant and the upgrading plant are presented in the following sections. Sensitivity analysis around key technical and economic assumptions is presented in Section 4.0.

3.1   Sludge HTL Plant

Table 16 lists the feed and biocrude product flow rates for the sludge HTL plant goal case.

Table 16. Annual feed and biocrude product rates for the sludge HTL plant.

Stream	Million gallons/year	Million lb/year
Dry sludge (15% ash) feed	N/A	73
Dry, ash-free sludge feed	N/A	62
Total slurry feed (25% solids)	33	290
HTL biocrude product	3.8	30
HTL aqueous phase	25	9.4
HTL solids (dry)	N/A	2000

3.1.1   Total Capital Investment

Table 17 summarizes the costs presented in Section 2.0 for the sludge HTL plant goal case including the balance of plant items such as the tank farm, flare and cooling water system. The HTL section of the plant contributes 72% the total installed capital.

Table 17. Total capital investment for sludge HTL plant.

Cost Item	Million US Dollars (2014$)
Sludge dewatering	1.43
HTL biocrude production	13.1
HTL aqueous phase treatment	3.09
Balance of plant	0.62
Total Installed Cost (TIC)	**18.2**
Buildings (4% of TIC)	0.73
Site development (10% of TIC)	1.82
Additional piping (4.5% of TIC)	0.73
Total Direct Costs (TDC)	**21.5**
Indirect Costs	
Prorated expenses (10% TDC)	2.15
Home office & construction fees (20% TDC)	4.30
Field expenses (10% TDC)	2.15
Project contingency (10% TDC)	2.15
Startup and permits (10% TDC)	2.15
Total Indirect	**12.9**
Fixed Capital Investment (FCI)	**34.4**
Working Capital (5% of FCI)	1.72
Land[a]	0.009
Total Capital Investment (TCI)	**36.2**
(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 18. The largest contributors to annual operating cost are quicklime for HTL aqueous phase ammonia stripping, natural gas for the HTL process, extra aeration power needed for processing the recycled HTL water in the WWTP's biological treatment step, and polymer consumed for sludge dewatering.

Table 18. HTL plant variable operating costs.

Variable	Value	Source	Total Cost (2014), million USD/year
Sludge Dewatering			
Polymer, $/lb (2013$)	1.73	City of Detroit 2014	0.33[a]
HTL Processing			
Natural Gas, $/1000 scf (2014$)	5.62	EIA 2014a	0.39
Electricity, ¢/kWh (2014$)	7.09	EIA 2014b	0.10
HTL Aqueous Phase Treatment			
Quicklime, $/ton (2014$)	107	USGS 2016	0.42
Natural gas (for THROX unit), $/1000 scf (2014$)	5.62	EIA 2014a	0.18
Extra Aeration at WWTP			
Electricity, kWh/lb COD removed (assuming100% COD removal)	0.40	Wan et al. 2016	0.36
Total			**1.78**
(a) For a polymer dose of 10.5 lb/dry ton secondary sludge.			

Fixed costs for the HTL plant are shown in Table 19. Salaries are taken from Dutta et al. (2011) and converted to a 2014 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 19. HTL plant fixed operating costs.

Position Title	Number	Total Cost (2014), million USD/year
Conversion Plant (unburdened)		
Plant Manager	1	0.15
Plant Engineer	1	0.07
Maintenance Super	1	0.06
Lab Manager	1	0.06
Shift Supervisor	3	0.14
Lab Technician	1	0.04
Maintenance Tech	1	0.04
Shift Operators	4	0.19
Yard Employees	1	0.03
Clerks & Secretaries	1	0.04
Subtotal		**0.81**
Overhead & maintenance	90% of labor & supervision	0.73
Maintenance capital	3% of TIC	0.98
Insurance and taxes	0.7% of FCI	0.24
Total Other Fixed Costs		**2.76**

3.1.3   Minimum Biocrude Selling Price

The minimum fuel selling price (MFSP) of the biocrude is determined using a discounted cash flow rate of return analysis. The MFSP is the plant gate selling price of the fuel product 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 1.3). The resulting MFSP for the 100 dry ton/day sludge HTL plant is $2.35/gge biocrude, or $2.53/gallon. The modeled biocrude has a lower heating value (LHV) of 124,990 Btu/gal. A LHV of 116,090 Btu/gal (ANL 2016) is used to convert the heat value of the fuel products to a gasoline gallon equivalent basis. Table 20 shows the breakdown of costs contributing to the MFSP of the biocrude. This MFSP corresponds to a revenue from biocrude sales of $9.6 million/year. The payback period of the project, calculated as the fixed capital investment divided by the average annual cash flow (Peters et al. 2004), is 13 years at the assumed 10% internal rate of return (IRR). This payback period does not consider any avoided sludge disposal costs that would be incurred due to reduction of solids from HTL. Figure 9 shows the contribution of each section of the HTL plant to the overall MFSP. The HTL section of the plant constitutes 72% of the production cost. Sensitivity analysis around key financial and technical assumptions and their impact on the MFSP is presented in Section 4.0.

Table 20. Biocrude MFSP cost breakdown.

	$/gge biocrude	$/year (2014 USD)
Natural Gas	0.14	$600,000
Quicklime, Polymer	0.18	$800,000
Electricity	0.11	$500,000
Fixed Costs	0.68	$2,800,000
Capital Depreciation	0.27	$1,100,000
Average Income Tax	0.16	$700,000
Average Return on Investment	0.81	$3,300,000
Total	**2.35**	
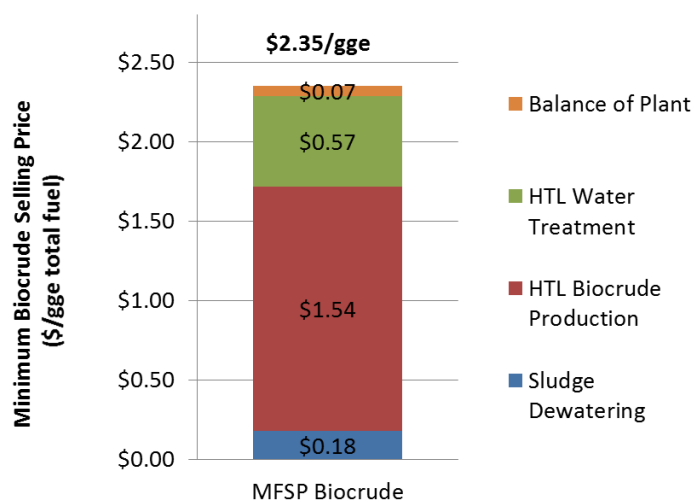
Figure 9. Sludge HTL biocrude production cost for goal case.

3.2 Biocrude Upgrading Plant

Table 21 lists the feed and product flow rates for the centralized sludge biocrude upgrading facility. The plant processes 2,730 BPSD of biocrude feed (from 10 HTL plants) and produces 2,660 BPSD of diesel and gasoline blendstocks.

Table 21. Annual feed and product flows for centralized biocrude upgrading plant (processing biocrude from 10 HTL plants).

Stream	Million gal/year	Million lbs/year
Biocrude feed	37.9	296.2
Hydrotreated oil	36.6	242.6
Diesel blendstock	27.6 (29.6 million gge/year)[a]	183.8
Naphtha (gasoline blendstock)	9.2 (9.1 million gge/year) [a]	56.5
(a) Based on a gasoline LHV of 116,090 Btu/gal (ANL 2016).		

3.2.1 Total Capital Investment

Table 22 summarizes the costs presented in Section 2.0 for the sludge HTL 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 single highest installed capital cost.

Table 22. Total capital investment for biocrude upgrading plant.

Cost Item	Million US Dollars (2014$)
Hydrotreating	33.6
Hydrocracking	6.6
Hydrogen plant	27.2
Steam cycle	1.6
Balance of plant	6.5
Total Installed Cost (TIC)	**75.6**
Buildings (4% of TIC)	3.0
Site development (10% of TIC)	7.6
Additional piping (4.5% of TIC)	1.8
Total Direct Costs (TDC)	**87.9**
Indirect Costs	
Prorated expenses (10% TDC)	8.8
Home office & construction fees (20% TDC)	17.6
Field expenses (10% TDC)	8.8
Project contingency (10% TDC)	8.8
Startup and permits (10% TDC)	8.8
Total Indirect	**52.8**
Fixed Capital Investment (FCI)	**140.7**
Working Capital (5% of FCI)	7.0
Land (6% of TPEC)[a]	2.7
Total Capital Investment (TCI)	**150.4**
(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.

Table 23. Biocrude upgrading plant variable operating costs.

Variable	Value	Source	Total Cost (2014), million USD/year
Biocrude cost, $/gge (2014$)	2.42[a]	This analysis	99.9
Hydrotreating catalyst, $/lb (2014$) (2 year life)	16.6	IHS 2014c	0.40
Hydrocracking catalyst, $/lb (2014$) (5 year life)	16.6	IHS 2014c	0.03
Hydrogen plant catalyst, $/1000scf H_2 (2014$) (5 year life)	0.0205	IHS 2014b	0.06
Natural gas, $/1000scf (2014$)	5.62	EIA 2014a	2.79
Cooling tower chemical, $/lb (2007$)	1.36	Humbird et al. 2011	0.01
Boiler chemical, $/lb (2007$)	2.27	Humbird et al. 2011	0.01
Electricity, ¢/kWh (2014$)	7.09	EIA 2014b	0.92
Water makeup, $/ton (2001$)	0.20	Humbird et al. 2011	0.05
Wastewater fee, $/ton (2001$)	0.48	Peters et al. 2004	0.07
Total			**104.2**

(a) Includes 10 cents/gge cost for biocrude transportation to the refinery (Sheppard 2011).

Table 24. Biocrude upgrading plant fixed operating costs.

Position Title	Number	Total Cost (2014), million USD/year
Conversion Plant (unburdened)		
Plant Manager	1	0.15
Plant Engineer	1	0.07
Maintenance Super	1	0.06
Lab Manager	1	0.06
Shift Supervisor	3	0.14
Lab Technician	3	0.12
Maintenance Tech	6	0.24
Shift Operators	25	1.20
Yard Employees	4	0.11
Clerks & Secretaries	1	0.04
Subtotal		**2.19**
Overhead & maintenance	90% of labor & supervision	1.97
Maintenance capital	3% of TIC	4.05
Insurance and taxes	0.7% of FCI	0.99
Total Other Fixed Costs		**9.18**

3.2.3 Minimum Fuel Selling Price

The MFSP for the goal case centralized upgrading plant is $3.46/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, considering that 1.05 gge of biocrude is needed to make 1 gge of upgraded fuel. Figure 10 shows the contribution of each section of the upgrading plant to the overall MFSP. Again, the overriding impact of biocrude price on the fuel production price is evident. Sensitivity analysis around key financial and technical assumptions and their impact on the MFSP is presented in Section 4.0.

Table 25. Upgraded fuel blendstock MFSP cost breakdown.

	$/gge final fuel	$/year
Biocrude Feed	2.58	$99,900,000
Natural Gas	0.07	$2,800,000
Catalysts & Chemicals	0.01	$500,000
Waste Disposal	0.002	$100,000
Electricity and other utilities	0.02	$1,000,000
Fixed Costs	0.24	$9,200,000
Capital Depreciation	0.002	$4,700,000
Average Income Tax	0.07	$2,800,000
Average Return on Investment	0.46	$17,700,000
Total	**3.46**	
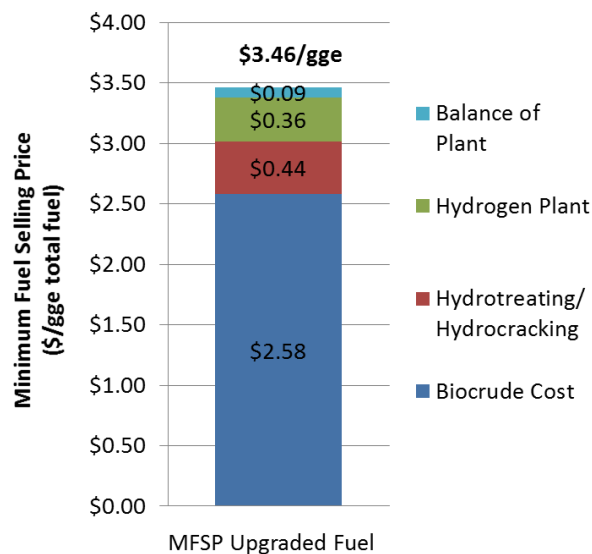
Figure 10. Upgraded fuel production cost for the goal case.

4.0 Economic and Technical Sensitivities

4.1 Sensitivity Analysis

Figure 11 shows the results of sensitivity analysis around the key technical and economic modeling assumptions. A wide range of plant scale of 50 to 950 dry ton/day was selected to include the largest plants in the country. Collectively this size range covers 52% of the total WWTP capacity represented in the 2012 CWNS data (EPA 2016). Smaller sized plants are not economically feasible under conventional equipment scaling assumptions for stick-built plants (i.e., the sixth-tenths rule), however biocrude collection scenarios centered around large WWTPs in densely populated regions could include smaller plants while still meeting the goal fuel MFSP, as explored in Section 4.2. As sludge management and disposal currently constitutes a major expense to the wastewater industry, the impact of avoided sludge disposal cost (feedstock cost) is an important aspect of reflecting the current reality for WWTPs. In some cases, the HTL plant would be owned by the WWTP, in others by a private company. In the latter case, the avoided sludge disposal cost may be in the form of a fee that WWTPs would pay to the HTL owner/operator that would allow sharing of the cost savings between the two parties. A range of $200/dry ton credit (avoided cost) to a $25/dry ton sludge cost was assessed for the sensitivity analysis. Biocrude yield and feed solids content were varied based on minimum values observed in experimental testing and maximum values thought to be possible with extensive research and development. The biocrude yield will vary depending on the composition of the sludge, generally following the trend of lipids>proteins>carbohydrates (Biller and Ross 2011). Hence, blending of sludge with high-lipid waste feedstocks such as FOG from WWTP skimming operations and animal rendering wastes could result in significantly higher yields. Note that there may be competition for these feedstocks in certain areas and this aspect needs to be considered on an individual plant basis when determining project feasibility. Sludge ash content will also vary considerably from plant to plant. As a comparative example, ash content of sludges provided by MetroVancouver in the LIFT study ranged from 7 to 12 wt% (Marrone 2016), while the mixed sludge provided by the City of Detroit was 26 wt%. We believe the Detroit sludge to be on the upper end and MetroVancouver to be on the lower end of the spectrum and accordingly, a range of 5-30% ash content was investigated. Note that the ash sensitivity only considers changes in yield that are proportional to organic content and does not include possible influences of ash content on HTL chemistry (e.g., catalytic) or oil separations (e.g., biocrude adhering to solids). There is uncertainty regarding the amount of extra heat exchange capacity that will be needed due to fouling of the sludge preheaters. The overall heat exchanger coefficient (U) was varied by -50% to +25% of the base case value to illustrate the potential impact of this factor on cost.

As shown in Figure 12, variability in the plant scale has the greatest impact on MFSP. Biocrude yield, sludge ash content and the overall heat transfer coefficient for the feed/product heat exchangers are key technical parameters that significantly affect production cost as well. Consideration of avoided sludge disposal cost, uncertainty in HTL capital cost, and the assumed IRR used in the discounted cash flow calculations are key economic assumptions impacting the MFSP. A 10% IRR is assumed for all conversion pathway TEAs in BETO's portfolio, however WWTP/HTL plant owners may want a higher project IRR.
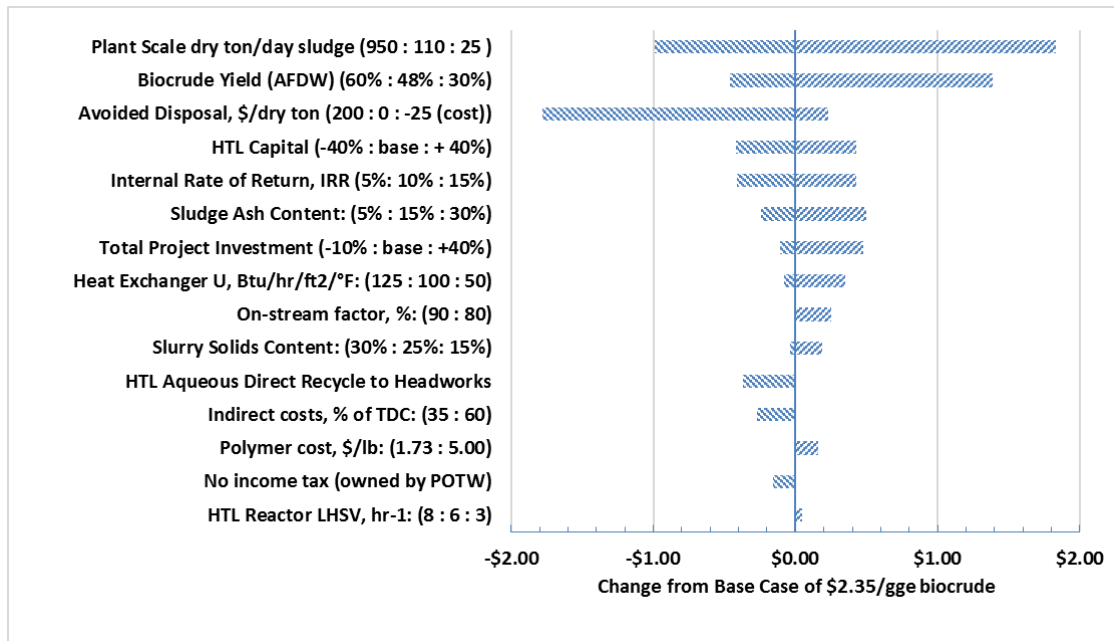
Figure 11. Sensitivity analysis for sludge HTL plant.

Figure 11 shows the sensitivity of final upgraded fuel MFSP to several economic and technical parameters for the biocrude upgrading plant model. Biocrude price was varied widely according to the approximate range seen from the variable HTL plant scale in Figure 11. Upgrader plant scale was varied from 1,000 to 5,000 BPSD final fuel. For plant scales above the base case, it is likely that a biocrude supply draw radius larger than the base case of 100 miles would be necessary. A sensitivity case where the biocrude is upgraded at an existing petroleum refinery (TCI=0) is also considered. As shown, this option could potentially reduce the MFSP by $0.50/gge. However, more work is needed to characterize the processing needs of biocrude and the oxygen, nitrogen and sulfur limits that could be tolerated in a petroleum refinery. The hydrotreated oil yield range shown represents the minimum observed in the laboratory and the maximum thought to be achievable with the highest quality biocrude. It may be possible to eliminate the hydrocracker through the use of a specially designed hydrotreating catalyst that also has cracking capabilities. In this case, the heavy residual would be recycled to the reactor with a small purge. This case is also included as a sensitivity. The sensitivity analysis shows that biocrude price has the greatest impact on upgrading production cost. Plant scale, upgrading at an existing refinery, and hydrotreated oil yield significantly affect cost as well. It is important to note that inherent to the success of a centralized biocrude upgrader scenario are the contractual arrangements and logistical coordination that would be required between owners/operators of the HTL/WWTP plants and the regional upgrading plant.
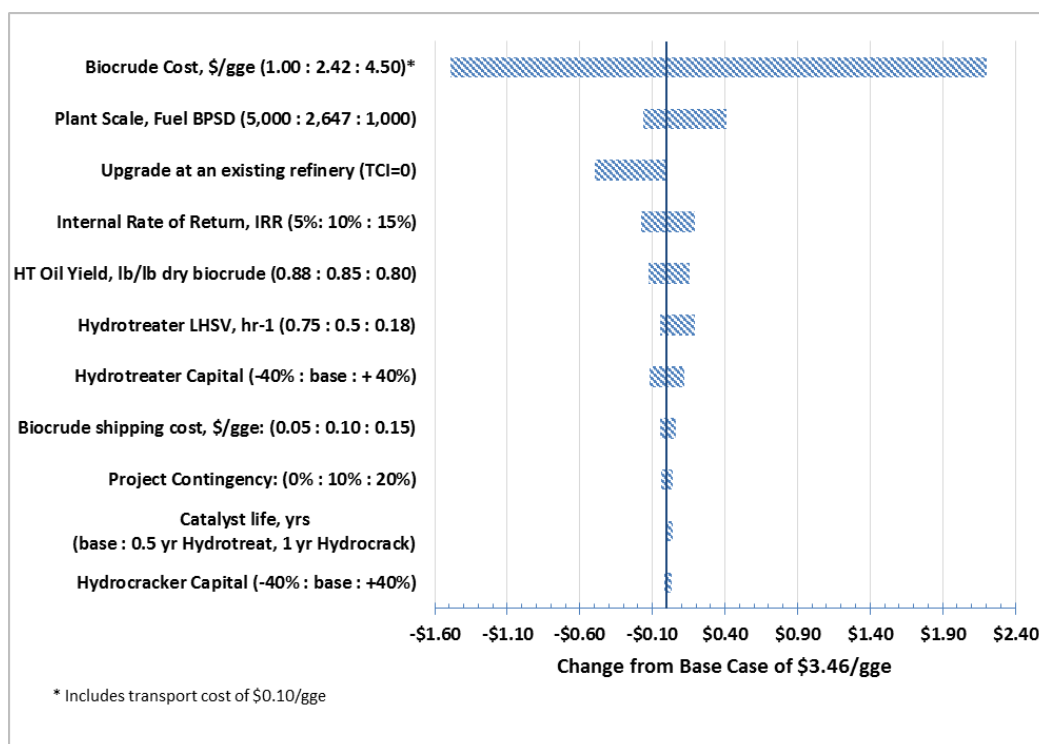
Figure 12. Sensitivity analysis for biocrude upgrading plant.

4.2   Regional Resource and Fuel Potential Analysis

As shown in the sensitivity analysis, plant scale is a key cost driver for the sludge biocrude production cost. The base case of 110 dry ton/day plant is approximately the minimum HTL plant scale that is feasible within BETO's $3/gge target when only considering a single WWTP size feeding an upgrader of 2,700 BPSD fuel production capacity. However, WWTPs larger than 110 dry ton/day can produce biocrude cheaper than the base case price of $2.35/gge. In this way, blending biocrude from smaller WWTPs with biocrude from the largest WWTPs could maximize fuel production while still meeting a feasible average biocrude price for the upgrader.

Using the sludge production dataset published by Seiple et al. (2017) and the HTL and biocrude upgrading plant cost production models, a siting analysis was performed to identify specific regions of the country that could support upgraded fuel production at or below $3.50/gge. This analysis is intended to provide an initial estimate of the general regions of interest rather than identify specific siting locations.

Siting was constrained to the 15,014 existing WWTPs ranging from 0.001 to 812 MGD influent flow. This was done in part to simplify the siting model, but also because WWTPs are highly spatially distributed in the U.S. and large scale upgrader facilities are likely to be proximal to biocrude sources due to transport costs and logistics. It is assumed that each WWTP has its own HTL plant and could send its biocrude to a central upgrader site. An initial pool of 2,621 siting candidates was developed by identifying all WWTPs capable of producing $3.50/gge fuel when aggregating biocrude within a travel radius of 100 miles, without considering resource competition from neighboring facilities. Although only candidate WWTPs could represent possible upgrader locations, resources from all WWTPs were considered during siting. Siting begins at the candidate upgrading plant location with the highest total

fuel blendstock production potential. Resource competition is simulated by ensuring that each WWTP can only participate in a single service territory. Within a given service territory, biocrude is incrementally aggregated by prioritizing neighbors by total biocrude rather than transport distance. Aggregation continues until the fuel price limit can no longer be achieved. Unused neighbors remain available for subsequent siting iterations, allowing service territories to overlap but ensuring resource pools are mutually exclusive. After each siting, the total fuel production potential is re-calculated for remaining candidates using any remaining neighbors. As siting progresses, some viable candidates become non-viable as their neighbors are scavenged by higher priority candidates. Siting continues until all candidates are either sited or no longer able to achieve the fuel price limit.

The results of the upgrading plant siting analysis are presented in Figure 13. A total of seven centralized upgrading sites are feasible in the U.S. given a fuel price limit of $3.50/gge. In total, these sites utilize 34% (13,000 dry ton/day) of the total daily sludge feedstock in the U.S. to generate approximately 1.32 million gallons/day of fuel blendstock. A total of 295 WWTPs (2%) were utilized during siting. They ranged in size from 0.05 to 812 MGD wastewater influent flow, with an average of 40 MGD. Only two facilities had a flow rate less than 1 MGD. On average, candidate sites utilized 13% (range of 3-31%) of their available neighbors, but consumed 72% (range of 29-94%) of available biocrude produced within the service area. Regional biocrude aggregation increased fuel blendstock production output by 225%, over the 0.41 million gallons/day of fuel that could be produced at 9 WWTPs capable of independently producing fuel at or below $3.50/gge.
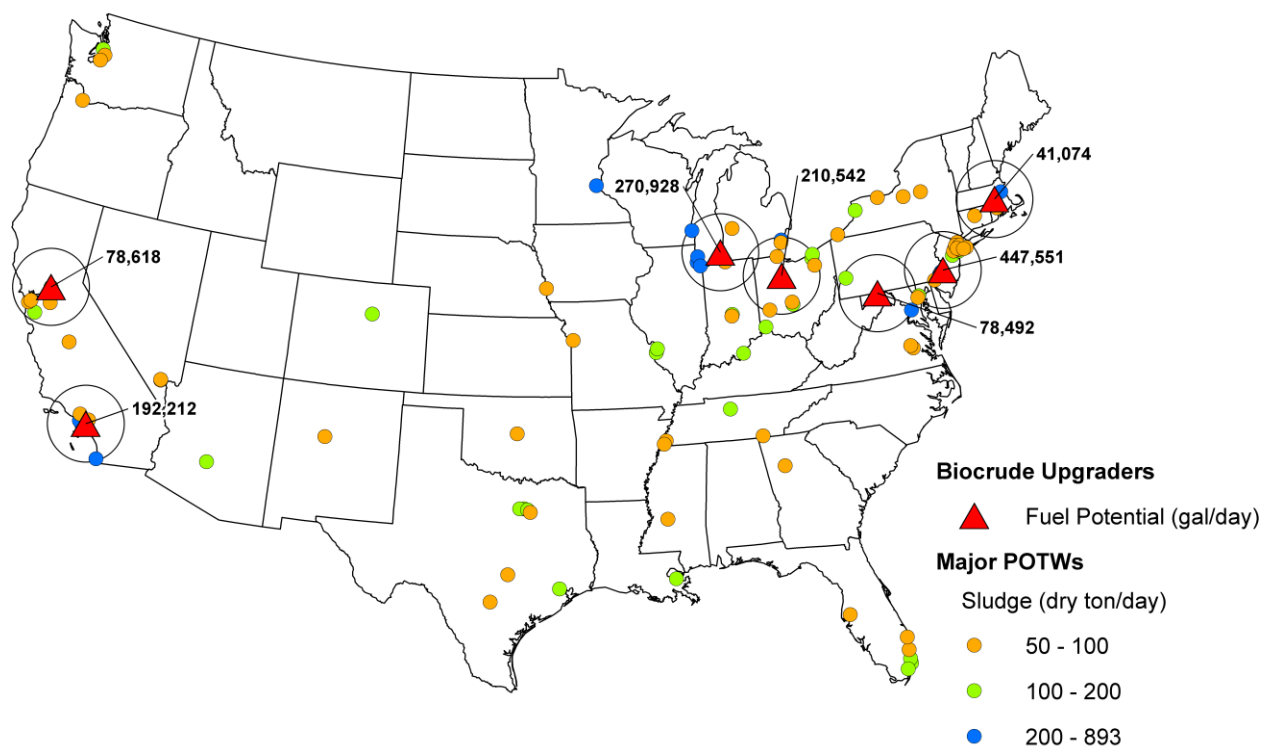
Figure 13. Regional fuel production at $3.50/gge from sludge HTL and biocrude upgrading.

5.0   Conclusions and Recommendations

This analysis shows that fuel blendstocks generated from HTL of sludge and centralized biocrude upgrading have the potential to be competitive with fossil fuels. The estimated plant gate MFSP of $3.46/gge final fuel blendstock for the goal case conceptual design for 2022 is within the tolerance of BETO's $3/gge programmatic goal for biofuel production cost. This analysis illustrates the feasibility of HTL for point-of-generation conversion of waste feedstock at a scale 1/20th that of the standard lignocellulosic biorefinery scale typically used in BETO design cases. The relevance of this work reaches beyond wastewater treatment sludge to lay the groundwork for application to other distributed wet wastes and blends that together represent a significant resource of underutilized biomass.

Future research in several key areas is needed to support the goal case design, including:

- Increased biocrude yields through

    – Improved continuous solid/liquid separations and liquid/liquid separations. Research is needed to investigate and test improved methods for enhanced phase separation such as coalescers, continuous solids separation, solids washing, feed de-ashing, and other approaches.

    – Incorporation of FOG and other fat-rich waste resources, such as skimmed FOG from the WWTP, collected restaurant grease trap waste, and regional rendering. Further work is necessary to determine and test appropriate sludge/FOG blends.

    – Increased feed solids concentration. Concentrating the solids increases the amount of carbon reporting to the biocrude phase during HTL and improves economics through decreased capital costs and increased energy efficiency. Further work is needed to assess high pressure pumping and HTL of sludge feed at 25% solids at various ash contents.

    – Decreased ash/grit content in the sludge solids through WWTP operations such as improved degritting and/or deashing technology, and reduced inorganic flocculating agent usage.

- Strategies for more effective and scalable heat transfer at HTL operating pressure:

    – Improved understanding of fundamental heat and mass transfer limitations for sludge heating and associated improvements to the design for application at scale. This includes heat transfer studies to better characterize the preheater equipment needs and rheology studies to determine viscosity profile as a function of temperature and flow rate.

    – Investigation of alternative heat exchanger designs for reduced material costs such as static mixer heat exchangers, dual-shell configurations (Fassbender 1992), and staged heating with shell/tube units for the latter portion of preheating.

- Cost-effective treatment technologies for the HTL aqueous phase. Baseline testing is needed to determine the effectiveness of various options (e.g., ammonia stripping, biological methods and membranes) for nitrogen removal. Further investigation of conversion (e.g., catalytic gasification) and treatment technologies through collaborative experimental and modeling work with universities and industry is needed to understand the impacts of full integration with the WWTP.

- Address scale challenges for distributed waste feedstocks through:

    – Integration of additional waste feedstocks on a regional basis. Resource analysis will help to identify available waste feedstock blends including manure, FOG, pulp sludge, food waste, etc…

Testing of different blends to establish overall yields and impact on quality is needed. Dry waste may also be beneficial as it is cheap to transport relative to wet materials and could reduce dewatering energy.

- Investigation of the feasibility of transporting sludge to enable larger scale HTL plants.

- Investigation of modular HTL systems for smaller scale applications.

- Improved performance of biocrude upgrading through (Jones et al. 2014):

  - Improved catalyst performance. Hydrotreating catalyst maintenance and stability are unknown, as are regeneration protocols and lifetimes. Longer-term testing with biocrude and detailed characterization of catalyst performance and deactivation modes are needed. Pretreatment steps, such as desalting, need to be demonstrated.

  - Development of hydrodeoxygenation and hydrodenitrogenation reaction kinetics to assist reactor designs and better inform the choice of co-processing in a petroleum refinery. A possible method to be more generally applicable might be to establish rate laws for sludge components such as, lipids, carbohydrates and proteins.

  - Both standard and comprehensive characterization of the major distillation fractions, gasoline range, diesel range, and gas oil range for the HTL oil and the hydrotreated oil. The jet fuel range should also be characterized, and an understanding of how to produce a jet cut without degrading naphtha and diesel properties would also be useful. Testing for key final fuel qualities, such as flash, octane, cetane, and cold flow properties is desirable.

  - Demonstration of hydrocracking yields of the gas oil fraction. Combined hydrotreating-hydrocracking is a possible way to reduce cost and could be achieved through catalyst research and development.

  - Investigation of compatibility of biocrude with petroleum refinery insertion points and other coprocessing strategies.

6.0   References

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Appendix A - Heat and Material Balances

A.2
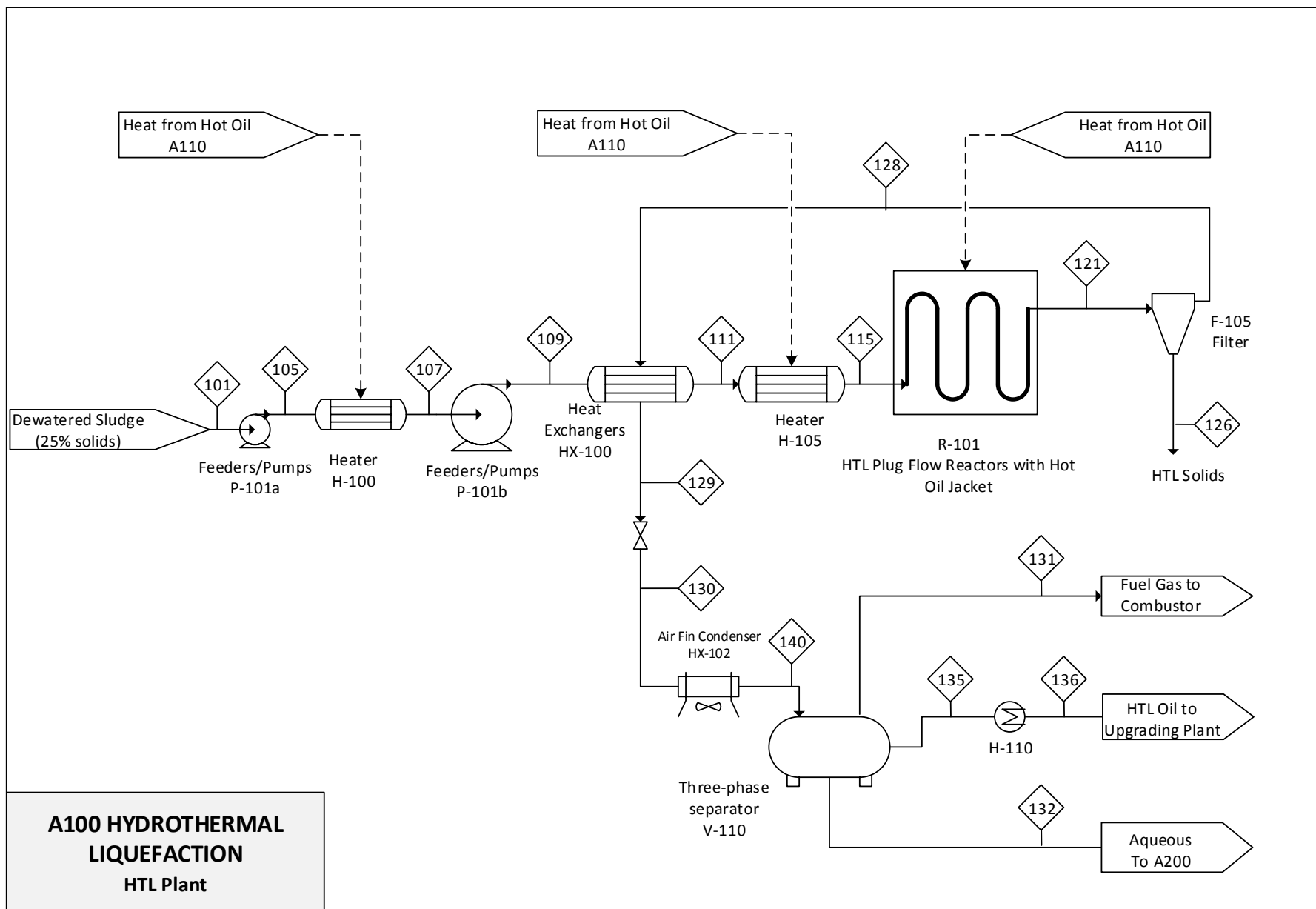
	101	105	107	109	111	115	121	126	128	129	130	131	132	135	136	140
Total Flow lb/hr	36667	36667	36667	36667	36667	36667	36667	5522	31144	31144	31144	1226	26022	3896	3896	31144
Temperature F	60	60	80	85	555	656	656	656	656	351	233	140	140	140	110	140
Pressure psia	15	85	84	3074	2984	2979	2969	2969	2969	1498	30	28	28	28	26	28
Vapor Frac	0	0	0	0	0	0	0	0	1	0	0	1	0	0	0	0
Mass Flow lb/hr																
H2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	2045	0	2045	2045	2045	1126	919	0	0	2045
H2O	27500	27500	27500	27500	27500	27500	27506	3301	24205	24205	24205	0	24049	156	156	24205
NH3	0	0	0	0	0	0	130	0	130	130	130	0	130	0	0	130
CH4	0	0	0	0	0	0	37	0	37	37	37	37	0	0	0	37
C2H6	0	0	0	0	0	0	12	0	12	12	12	12	0	0	0	12
C3H8	0	0	0	0	0	0	4	0	4	4	4	4	0	0	0	4
N-C4H10	0	0	0	0	0	0	34	0	34	34	34	34	0	0	0	34
N-PENTAN	0	0	0	0	0	0	12	0	12	12	12	12	0	0	0	12
METHANOL	0	0	0	0	0	0	218	0	218	218	218	0	218	0	0	218
ETHANOL	0	0	0	0	0	0	9	0	9	9	9	0	9	0	0	9
ACETONE	0	0	0	0	0	0	38	0	38	38	38	0	38	0	0	38
ACEACID	0	0	0	0	0	0	179	0	179	179	179	0	179	0	0	179
PROACID	0	0	0	0	0	0	65	0	65	65	65	0	65	0	0	65
ETHAMIN	0	0	0	0	0	0	96	0	96	96	96	0	96	0	0	96
2-PYRRLD	0	0	0	0	0	0	47	0	47	47	47	0	47	0	0	47
2-PIPERD	0	0	0	0	0	0	47	0	47	47	47	0	47	0	0	47
7-LACTAM	0	0	0	0	0	0	78	0	78	78	78	0	78	0	0	78
C5H9NS	0	0	0	0	0	0	245	0	245	245	245	0	89	156	156	245
TOLUENE	0	0	0	0	0	0	112	0	112	112	112	0	0	112	112	112
RYRO3ETM	0	0	0	0	0	0	198	0	198	198	198	0	0	198	198	198
PHENO4M	0	0	0	0	0	0	17	0	17	17	17	0	0	17	17	17
AMIPHENO	0	0	0	0	0	0	34	0	34	34	34	0	0	34	34	34
INDOLE	0	0	0	0	0	0	206	0	206	206	206	0	0	206	206	206
2-PYTENE	0	0	0	0	0	0	258	0	258	258	258	0	0	258	258	258
C15OLEF	0	0	0	0	0	0	258	0	258	258	258	0	0	258	258	258
MC12AMID	0	0	0	0	0	0	223	0	223	223	223	0	0	223	223	223
C16AMIDE	0	0	0	0	0	0	206	0	206	206	206	0	0	206	206	206
C18AMIDE	0	0	0	0	0	0	430	0	430	430	430	0	0	430	430	430
C16:0FA	0	0	0	0	0	0	304	0	304	304	304	0	12	292	292	304
C18:1FA	0	0	0	0	0	0	421	0	421	421	421	0	0	421	421	421
C13H18	0	0	0	0	0	0	344	0	344	344	344	0	0	344	344	344
HEVOIL1	0	0	0	0	0	0	418	0	418	418	418	0	47	371	371	418
HEVOIL2	0	0	0	0	0	0	215	0	215	215	215	0	0	215	215	215
SLUDGE	9167	9167	9167	9167	9167	9167	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	1377	1377	0	0	0	0	0	0	0	0
SOLID	0	0	0	0	0	0	845	845	0	0	0	0	0	0	0	0
Enthalpy mmBtu/hr	-202.74	-202.73	-202.11	-201.70	-186.30	-181.06	-178.06	-23.28	-154.74	-170.14	-170.14	-4.45	-168.18	-3.35	-3.53	-175.76

A.3

Air

170 171

Air Blower
K-155

H-151

157

Exhaust
Blower
K-150

160

Exhaust

172

Fuel Gas From
A100

131

Combustor
R-150

Natural Gas

156

Circ. Heat
Transfer Oil

(Details not shown)

Heat from Hot Oil
To A100 Feed Prep

Heat from Hot Oil
To A100 HTL

**A110 Combustor & Heat
Transfer Circulation Oil**
HTL Plant

A.4
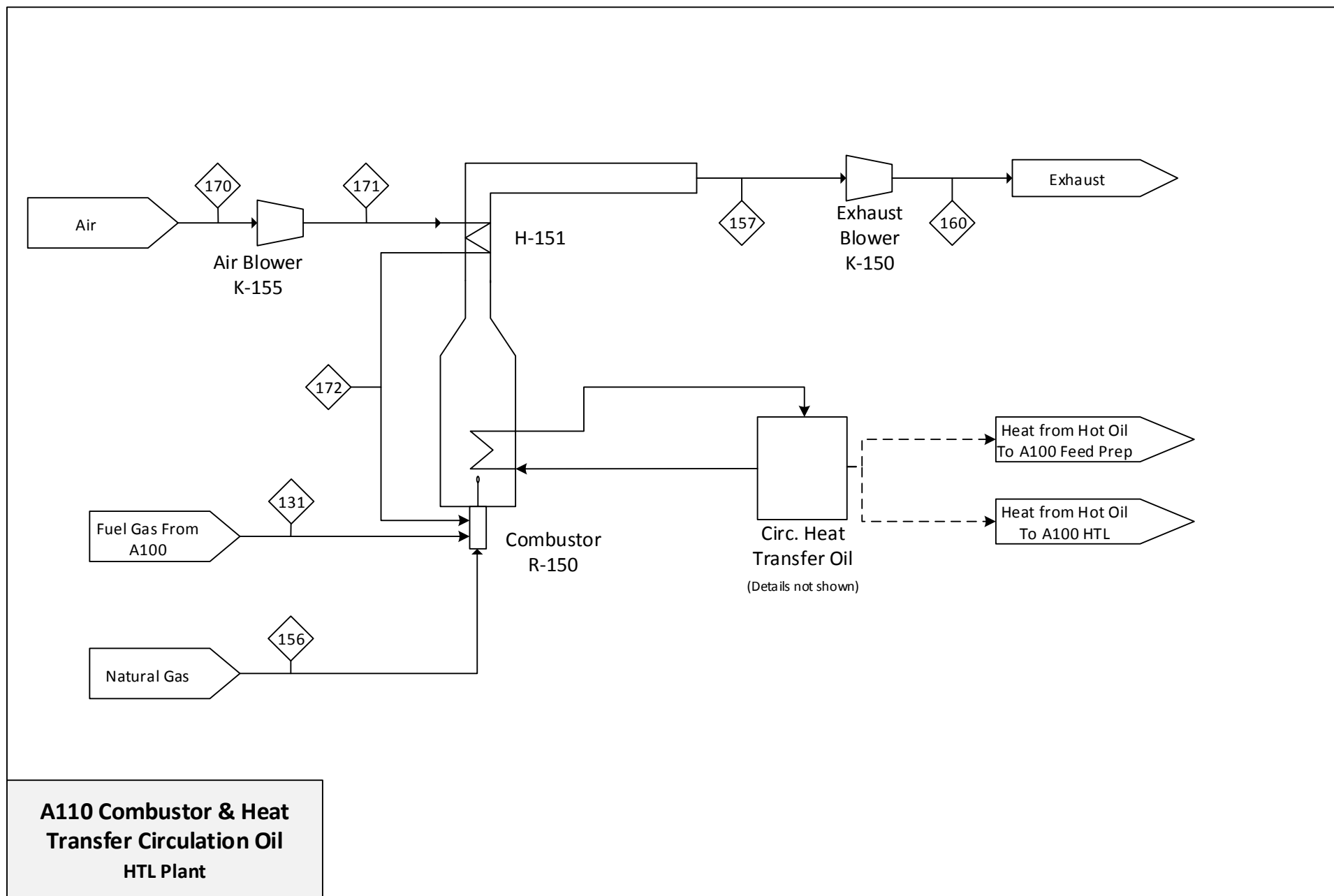
	131	156	157	160	170	171	172
Total Flow lb/hr	1226	374	11342	11342	9742	9742	9742
Temperature F	140	60	491	533	90	132	350
Pressure psia	28	450	15	17	15	18	16
Vapor Frac	1	1	1	1	1	1	1
Mass Flow lb/hr							
N2	0	2	7216	7216	7214	7214	7214
O2	0	0	368	368	2210	2210	2210
AR	0	0	123	123	123	123	123
CO2	1126	4	2433	2433	5	5	5
H2O	0	0	1202	1202	190	190	190
NH3	0	0	0	0	0	0	0
CH4	37	368	0	0	0	0	0
C2H6	12	0	0	0	0	0	0
C3H8	4	0	0	0	0	0	0
N-C4H10	34	0	0	0	0	0	0
N-PENTAN	12	0	0	0	0	0	0
HEXANE	0	0	0	0	0	0	0
METHANOL	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0
ACETONE	0	0	0	0	0	0	0
ACEACID	0	0	0	0	0	0	0
PROACID	0	0	0	0	0	0	0
ETHAMIN	0	0	0	0	0	0	0
2-PYRRLD	0	0	0	0	0	0	0
2-PIPERD	0	0	0	0	0	0	0
7-LACTAM	0	0	0	0	0	0	0
C5H9NS	0	0	0	0	0	0	0
TOLUENE	0	0	0	0	0	0	0
RYRO3ETM	0	0	0	0	0	0	0
PHENO4M	0	0	0	0	0	0	0
AMIPHENO	0	0	0	0	0	0	0
INDOLE	0	0	0	0	0	0	0
2-PYTENE	0	0	0	0	0	0	0
C15OLEF	0	0	0	0	0	0	0
MC12AMID	0	0	0	0	0	0	0
C16AMIDE	0	0	0	0	0	0	0
C18AMIDE	0	0	0	0	0	0	0
C16:0FA	0	0	0	0	0	0	0
C18:1FA	0	0	0	0	0	0	0
C13H18	0	0	0	0	0	0	0
HEVOIL1	0	0	0	0	0	0	0
HEVOIL2	0	0	0	0	0	0	0
Enthalpy, mmBtu/hr	-4.45	-0.78	-15.04	-14.91	-1.09	-0.99	-0.46

THROX Unit
TX-230

Natural Gas

203

235

Exhaust

222

218

232

H-230

230

202

225

Air

Hydrated Lime

Clarifier
CL-215

132

K-230

237

HTL Aqueous
From A100

NH3 Stripper
T-220

242

220

Lime Sludge

Water recycle to
headworks

A200 HTL Aqueous Treatment
HTL Plant
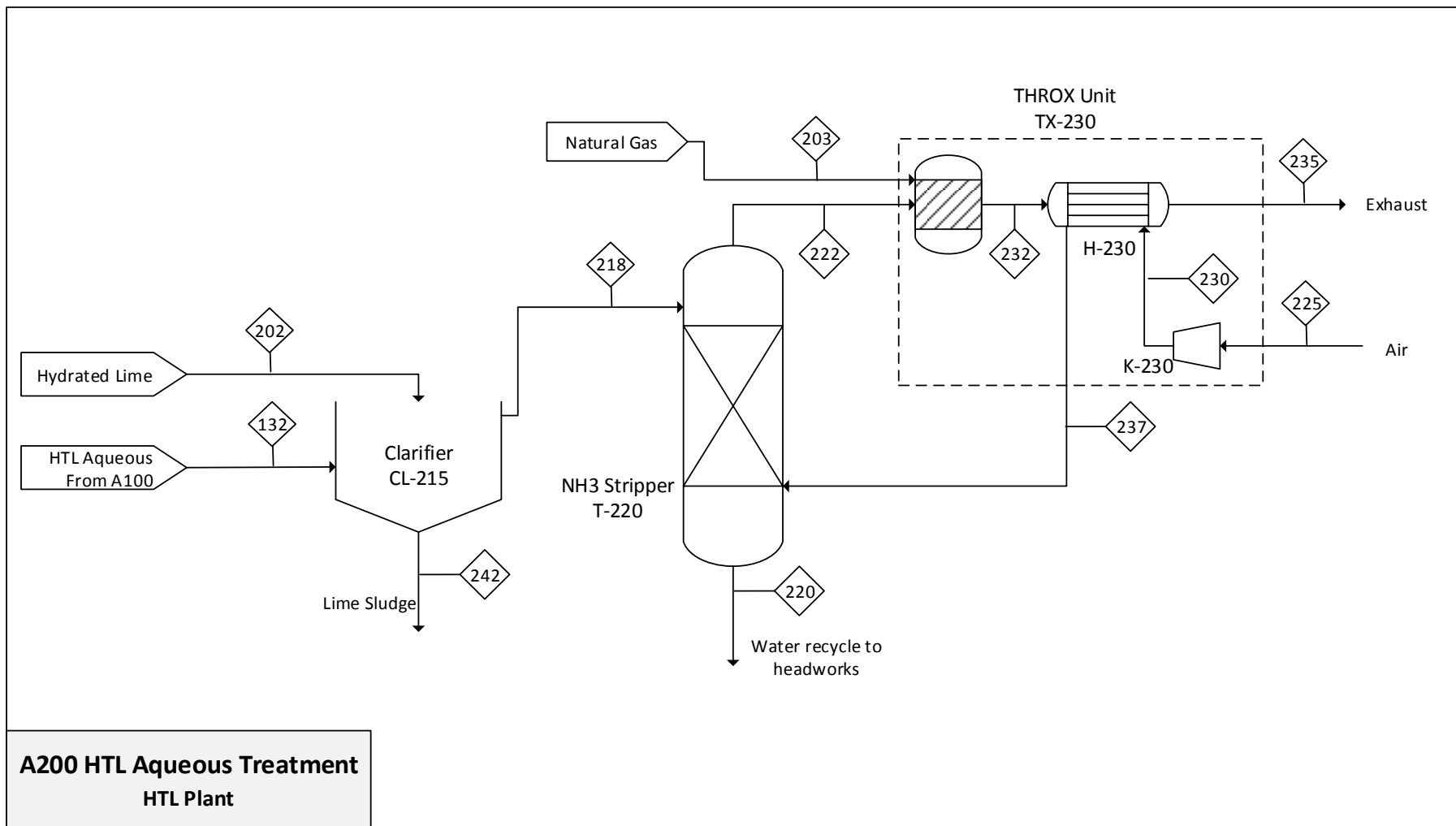
	132	202	203	218	220	222	225	230	232	235	237	242
Mass Flow lb/hr	25901	1314	175	25125	21139	58986	55000	55000	59161	59161	55000	2091
Temperature F	140	70	70	164	106	116	59	66	600	401	301	165
Pressure psia	28	15	25	60	15	15	15	15	15	15	15	60
Vapor Frac	0	0	1	0	0	1	1	1	1	1	1	0
Solid Frac	0	1	0	0	0	0	0	0	0	0	0	1
Volume Flow cuft/hr	2883	9	2476	415	341	885766	721495	706958	1658570	1346800	1023720	16
Mass Flow lb/hr												
H2O	23919	0	0	24097	20533	3911	347	347	4815	4815	347	268
N2	0	0	0	0	28	41249	41277	41277	41355	41355	41277	0
O2	0	0	0	0	22	12628	12650	12650	11285	11285	12650	0
AR	0	0	0	0	1	699	701	701	699	699	701	0
CO2	454	0	0	0	1	24	25	25	963	963	25	0
NH3	6	0	0	129	0	129	0	0	0	0	0	1
METHANE	0	0	175	0	0	0	0	0	0	0	0	0
METHANOL	218	0	0	216	10	206	0	0	0	0	0	2
ETHANOL	9	0	0	9	0	9	0	0	0	0	0	0
ETHAMIN	96	0	0	43	0	43	0	0	43	43	0	53
ACETONE	38	0	0	38	0	38	0	0	0	0	0	0
ACEACID	179	0	0	177	130	47	0	0	0	0	0	2
PROACID	65	0	0	64	62	2	0	0	0	0	0	1
2-PYRRLD	47	0	0	46	46	0	0	0	0	0	0	1
2-PIPERD	47	0	0	46	46	0	0	0	0	0	0	1
7-LACTAM	78	0	0	77	77	0	0	0	0	0	0	1
C5H9NS	89	0	0	88	88	0	0	0	0	0	0	1
C16:0FA	12	0	0	12	12	0	0	0	0	0	0	0
HEVOIL1	47	0	0	46	46	0	0	0	0	0	0	1
CA++	0	0	0	7	7	0	0	0	0	0	0	0
K+	7	0	0	7	7	0	0	0	0	0	0	0
NA+	4	0	0	4	4	0	0	0	0	0	0	0
NH4+	131	0	0	0	0	0	0	0	0	0	0	0
CL-	7	0	0	7	7	0	0	0	0	0	0	0
PO4---	0	0	0	0	0	0	0	0	0	0	0	0
SO4--	5	0	0	5	5	0	0	0	0	0	0	0
HCO3-	442	0	0	0	0	0	0	0	0	0	0	0
CO3--	2	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	6	6	0	0	0	0	0	0	0
CA3PO4*2	0	0	0	0	0	0	0	0	0	0	0	0
CACO3	0	0	0	0	0	0	0	0	0	0	0	1759
CA(OH)2	0	1314	0	0	0	0	0	0	0	0	0	0
Enthalpy mmBtu/hr	-168.01	-7.52	-0.35	-164.58	-140.62	-23.07	-2.35	-2.25	-23.42	-26.56	0.89	-10.96

A.7

A310 HYDROTREATING
Upgrading Plant
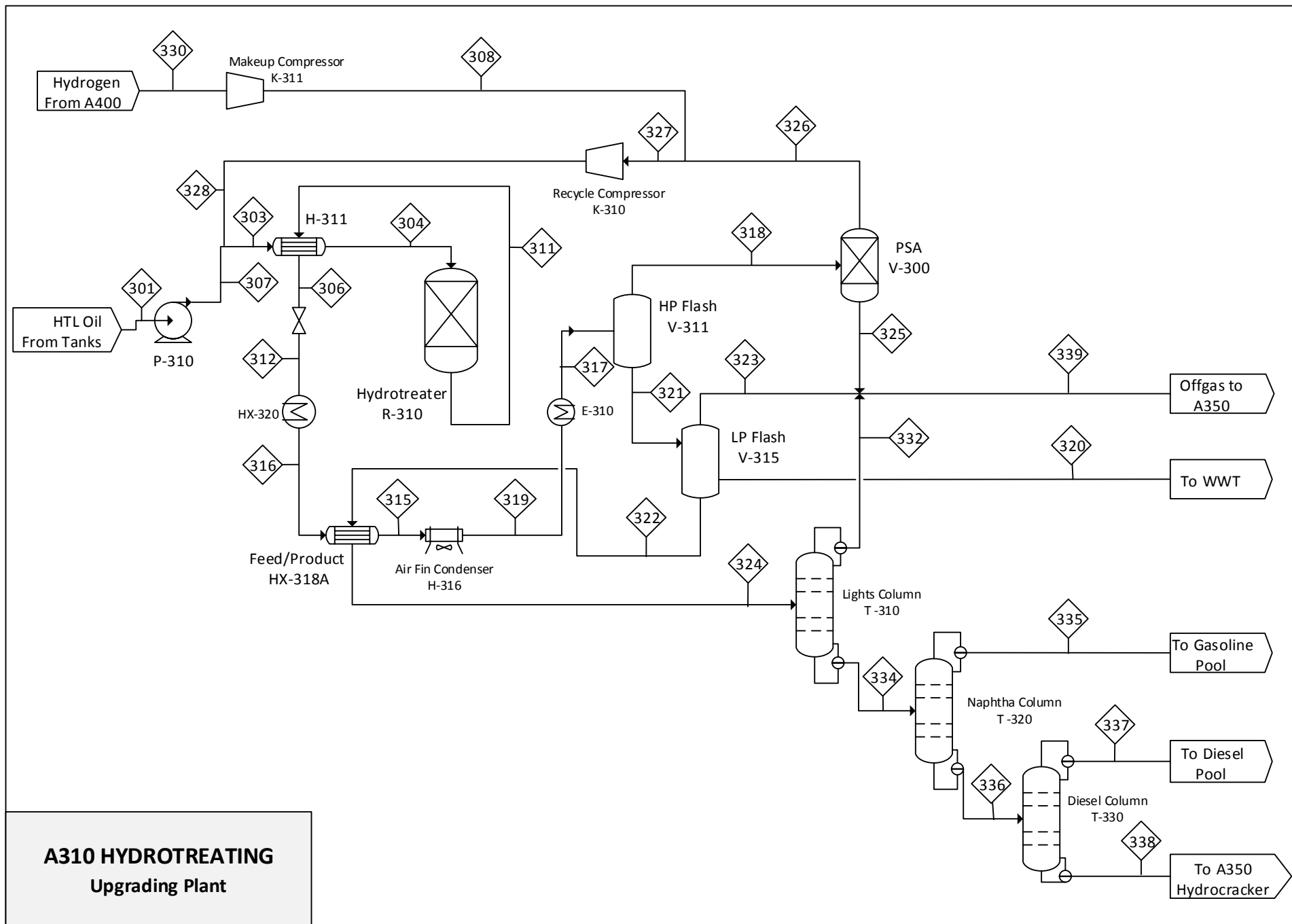
	301	304	306	307	308	311	312	315	316	317	320	321	322	324	325	326	327	328	330	331	332	334	335	336	337	338	339	341
Total Flow lb/hr	38961	43887	43887	38961	1988	43887	43887	43887	43887	43887	5171	36821	31476	31476	4128	2938	4926	4926	1988	36821	841	30635	6423	24212	12262	11950	5143	7066
Temperature F	80	335	591	92	354	750	581	246	300	110	78	110	78	220	110	110	209	244	140	78	108	409	205	560	487	656	108	110
Pressure psia	26	1525	1515	1530	710	1515	725	724	725	717	55	717	55	54	20	707	707	1530	336	55	48	50	25	25	19	19	20	717
Vapor Frac	0	1	1	0	1	1	1	1	1	1	0	0	0	0	0	1	1	1	1	0	1	0	0	0	0	0	1	1
Mass Flow lb/hr																												
H2	0	4926	3284	0	1988	3284	3284	3284	3284	3284	0	19	1	1	326	2938	4926	4926	1988	19	1	0	0	0	0	0	346	3265
H2O	1560	1560	3525	1560	0	3525	3525	3525	3525	3525	3375	3383	6	6	142	0	0	0	0	3383	6	0	0	0	0	0	150	142
NH3	0	0	1969	0	0	1969	1969	1969	1969	1969	1259	1290	16	16	679	0	0	0	0	1290	16	0	0	0	0	0	711	679
CH4	0	0	850	0	0	850	850	850	850	850	0	31	6	6	819	0	0	0	0	31	6	0	0	0	0	0	850	819
C2H6	0	0	1028	0	0	1028	1028	1028	1028	1028	0	133	75	75	895	0	0	0	0	133	75	0	0	0	0	0	1027	895
C3H8	0	0	557	0	0	557	557	557	557	557	0	170	139	139	388	0	0	0	0	170	136	2	2	0	0	0	555	388
N-C4H10	0	0	188	0	0	188	188	188	188	188	1	104	99	99	84	0	0	0	0	104	57	42	42	0	0	0	145	84
N-PENTAN	0	0	1521	0	0	1521	1521	1521	1521	1521	6	1167	1147	1147	354	0	0	0	0	1167	287	860	860	0	0	0	655	354
HEXANE	0	0	354	0	0	354	354	354	354	354	0	317	315	315	37	0	0	0	0	317	39	276	276	0	0	0	78	37
C5H9NS	1559	1559	0	1559	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
TOLUENE	1117	1117	531	1117	0	531	531	531	531	531	0	516	516	516	15	0	0	0	0	516	23	492	492	0	0	0	39	15
RYRO3ETM	1976	1976	0	1976	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PHENO4M	172	172	0	172	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMIPHENO	344	344	0	344	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
INDOLE	2062	2062	0	2062	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
2-PYTENE	2578	2578	0	2578	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C15OLEF	2578	2578	0	2578	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MC12AMID	2234	2234	0	2234	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C16AMIDE	2062	2062	0	2062	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C18AMIDE	4296	4296	0	4296	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C16:0FA	2921	2921	496	2921	0	496	496	496	496	496	0	496	495	495	0	0	0	0	0	496	0	495	0	495	25	470	0	0
C18:1FA	4210	4210	0	4210	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C13H18	3437	3437	0	3437	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HEVOIL1	3708	3708	0	3708	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HEVOIL2	2148	2148	0	2148	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
2MPENTA	0	0	460	0	0	460	460	460	460	460	1	397	394	394	63	0	0	0	0	397	61	334	334	0	0	0	126	63
CC5-METH	0	0	248	0	0	248	248	248	248	248	1	225	224	224	22	0	0	0	0	225	24	200	200	0	0	0	47	22
2MHEXAN	0	0	460	0	0	460	460	460	460	460	1	433	431	431	28	0	0	0	0	433	33	398	398	0	0	0	61	28
CC6-METH	0	0	177	0	0	177	177	177	177	177	0	170	169	169	7	0	0	0	0	170	9	160	160	0	0	0	17	7
PIPERDIN	0	0	71	0	0	71	71	71	71	71	16	70	54	54	0	0	0	0	0	70	2	52	52	0	0	0	2	0
3MHEPTA	0	0	354	0	0	354	354	354	354	354	0	345	345	345	8	0	0	0	0	345	12	333	333	0	0	0	21	8
OCTANE	0	0	248	0	0	248	248	248	248	248	0	243	243	243	4	0	0	0	0	243	7	237	237	0	0	0	11	4
ETHCYC6	0	0	265	0	0	265	265	265	265	265	0	261	261	261	4	0	0	0	0	261	6	255	255	0	0	0	10	4
ETHBENZ	0	0	265	0	0	265	265	265	265	265	0	262	262	262	3	0	0	0	0	262	6	256	256	0	0	0	9	3
O-XYLENE	0	0	425	0	0	425	425	425	425	425	0	421	420	420	4	0	0	0	0	421	7	413	413	0	0	0	11	4
C9H20	0	0	602	0	0	602	602	602	602	602	0	597	597	597	4	0	0	0	0	597	7	590	590	0	0	0	12	4
CC6-PRO	0	0	460	0	0	460	460	460	460	460	0	457	457	457	3	0	0	0	0	457	5	452	451	0	0	0	8	3
C3BENZ	0	0	283	0	0	283	283	283	283	283	0	282	281	281	2	0	0	0	0	282	3	278	278	0	0	0	5	2
4MNONAN	0	0	283	0	0	283	283	283	283	283	0	282	282	282	1	0	0	0	0	282	2	280	279	1	1	0	4	1
C10H22	0	0	354	0	0	354	354	354	354	354	0	353	353	353	1	0	0	0	0	353	2	351	337	14	14	0	3	1
C4BENZ	0	0	531	0	0	531	531	531	531	531	0	530	530	530	1	0	0	0	0	530	2	527	175	352	352	0	4	1
C10H12	0	0	602	0	0	602	602	602	602	602	0	601	600	600	1	0	0	0	0	601	2	598	2	596	596	0	3	1
C12H26	0	0	425	0	0	425	425	425	425	425	0	425	425	425	0	0	0	0	0	425	0	424	0	424	424	0	1	0
1234NA	0	0	708	0	0	708	708	708	708	708	0	707	707	707	1	0	0	0	0	707	2	705	0	705	705	0	2	1
CC6-HEX	0	0	708	0	0	708	708	708	708	708	0	708	708	708	0	0	0	0	0	708	1	707	0	707	707	0	1	0
14DMNAPH	0	0	885	0	0	885	885	885	885	885	0	885	884	884	0	0	0	0	0	885	1	884	0	884	884	0	1	0
C7BENZ	0	0	1062	0	0	1062	1062	1062	1062	1062	0	1062	1061	1061	0	0	0	0	0	1062	0	1061	0	1061	1061	0	1	0
C8BENZ	0	0	814	0	0	814	814	814	814	814	0	814	814	814	0	0	0	0	0	814	0	814	0	814	814	0	0	0
TETMC16	0	0	2124	0	0	2124	2124	2124	2124	2124	0	2124	2124	2124	0	0	0	0	0	2124	0	2124	0	2124	675	1448	0	0
C22H42O4	0	0	708	0	0	708	708	708	708	708	0	708	708	708	0	0	0	0	0	708	0	708	0	708	0	708	0	0
C15H32	0	0	708	0	0	708	708	708	708	708	0	708	708	708	0	0	0	0	0	708	0	708	0	708	707	1	0	0
C16H34	0	0	2124	0	0	2124	2124	2124	2124	2124	0	2124	2124	2124	0	0	0	0	0	2124	0	2124	0	2124	2053	71	0	0
C17H36	0	0	3540	0	0	3540	3540	3540	3540	3540	0	3539	3539	3539	0	0	0	0	0	3539	0	3539	0	3539	2247	1292	0	0
C18H38	0	0	3186	0	0	3186	3186	3186	3186	3186	0	3186	3186	3186	0	0	0	0	0	3186	0	3186	0	3186	930	2255	0	0
C20H42	0	0	2124	0	0	2124	2124	2124	2124	2124	0	2124	2124	2124	0	0	0	0	0	2124	0	2124	0	2124	65	2059	0	0
C22H46	0	0	531	0	0	531	531	531	531	531	0	531	531	531	0	0	0	0	0	531	0	531	0	531	1	530	0	0
C24H50	0	0	1239	0	0	1239	1239	1239	1239	1239	0	1239	1239	1239	0	0	0	0	0	1239	0	1239	0	1239	0	1239	0	0
C28H58	0	0	885	0	0	885	885	885	885	885	0	885	885	885	0	0	0	0	0	885	0	885	0	885	0	885	0	0
C30H62	0	0	283	0	0	283	283	283	283	283	0	283	283	283	0	0	0	0	0	283	0	283	0	283	0	283	0	0
C32H66	0	0	708	0	0	708	708	708	708	708	0	708	708	708	0	0	0	0	0	708	0	708	0	708	0	708	0	0
NH4HS	0	0	689	0	0	689	689	689	689	689	459	459	1	1	229	0	0	0	0	459	0	1	0	1	0	1	229	229
2*NH4CO3	0	0	49	0	0	49	49	49	49	49	49	49	0	0	0	0	0	0	0	49	0	0	0	0	0	0	0	0
Enthalpy, mmBtu/hr	-35.88	-24.19	-31.70	-35.40	1.93	-24.48	-31.73	-46.31	-44.00	-52.51	-26.52	-49.47	-23.81	-21.50	-5.46	0.36	2.29	2.97	0.44	-49.47	-0.75	-17.05	-4.12	-11.23	-4.72	-6.28	-6.40	-5.13

A.9

Off-Gas
From A310

339

366

To Ammonia
Scrubber

377

378

Make-up Compressor
K-355

380

376

368

385

Hydrogen
From A400

351

352

353

355

362

365

H-351

Trim Heater
H-352

Make-up Compressor
K-350

Hydrocracker
R-350

Heavy Oil
From A310

350

361

370

P-350

374

372

V-375

T-350 Reboiler
HX-381

Air Fin Condenser
H-371

381

V-380

383

386

To Gasoline
Pool

Naphtha Splitter
T-350

389

To Diesel
Pool

A350 HYDROCRACKING

Upgrading Plant

A.10

	339	350	351	352	353	355	361	362	365	366	368	370	372	374	376	377	378	380	381	383	385	386	389
Total Flow lb/hr	5143	11950	125	125	12676	12676	11950	12676	12676	5557	182	12676	12676	12676	752	602	150	602	11924	11742	82	716	10944
Temperature F	108	653	140	297	584	700	661	741	753	135	137	655	140	300	140	140	140	148	140	137	140	140	520
Pressure psia	20	19	336	1040	1035	1030	1035	1010	1010	20	30	1009	1006	1008	1006	1006	1006	1040	1006	30	20	20	24
Vapor Frac	1	0	1	1	1	1	0	1	1	1	1	1	1	1	1	1	1	1	0	0	1	0	0
Mass Flow lb/hr																							
H2	346	0	125	125	420	420	0	420	377	428	8	377	377	377	368	295	74	295	8	0	0	0	0
CO2	0	0	0	0	262	262	0	262	430	168	91	430	430	430	327	262	65	262	103	12	11	0	0
H2O	150	0	0	0	12	12	0	12	78	216	51	78	78	78	15	12	3	12	63	12	11	1	0
NH3	711	0	0	0	0	0	0	0	0	711	0	0	0	0	0	0	0	0	0	0	0	0	0
CH4	850	0	0	0	0	0	0	0	0	850	0	0	0	0	0	0	0	0	0	0	0	0	0
C2H6	1027	0	0	0	0	0	0	0	0	1027	0	0	0	0	0	0	0	0	0	0	0	0	0
C3H8	555	0	0	0	0	0	0	0	0	555	0	0	0	0	0	0	0	0	0	0	0	0	0
N-C4H10	145	0	0	0	26	26	0	26	137	215	23	137	137	137	33	26	7	26	104	82	40	41	1
N-PENTAN	655	0	0	0	0	0	0	0	0	655	0	0	0	0	0	0	0	0	0	0	0	0	0
HEXANE	78	0	0	0	1	1	0	1	15	80	1	15	15	15	1	1	0	1	15	14	2	12	0
TOLUENE	39	0	0	0	0	0	0	0	0	39	0	0	0	0	0	0	0	0	0	0	0	0	0
C16:0FA	0	470	0	0	470	470	470	470	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
2MPENTA	126	0	0	0	0	0	0	0	0	126	0	0	0	0	0	0	0	0	0	0	0	0	0
CC5-METH	47	0	0	0	0	0	0	0	0	47	0	0	0	0	0	0	0	0	0	0	0	0	0
2MHEXAN	61	0	0	0	0	0	0	0	0	61	0	0	0	0	0	0	0	0	0	0	0	0	0
HEPTANE	0	0	0	0	1	1	0	1	75	5	1	75	75	75	1	1	0	1	74	72	4	63	6
CC6-METH	17	0	0	0	0	0	0	0	0	17	0	0	0	0	0	0	0	0	0	0	0	0	0
PIPERDIN	2	0	0	0	0	0	0	0	0	2	0	0	0	0	0	0	0	0	0	0	0	0	0
3MHEPTA	21	0	0	0	3	3	0	3	440	35	4	440	440	440	4	3	1	3	436	432	10	344	78
OCTANE	11	0	0	0	1	1	0	1	166	15	1	166	166	166	1	1	0	1	165	163	3	123	38
ETHCYC6	10	0	0	0	0	0	0	0	0	10	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHBENZ	9	0	0	0	0	0	0	0	0	9	0	0	0	0	0	0	0	0	0	0	0	0	0
O-XYLENE	11	0	0	0	0	0	0	0	0	11	0	0	0	0	0	0	0	0	0	0	0	0	0
C9H20	12	0	0	0	1	1	0	1	253	14	1	253	253	253	1	1	0	1	252	252	1	114	137
CC6-PRO	8	0	0	0	0	0	0	0	0	8	0	0	0	0	0	0	0	0	0	0	0	0	0
C3BENZ	5	0	0	0	0	0	0	0	0	5	0	0	0	0	0	0	0	0	0	0	0	0	0
4MNONAN	4	0	0	0	0	0	0	0	0	4	0	0	0	0	0	0	0	0	0	0	0	0	0
C10H22	3	0	0	0	0	0	0	0	391	4	0	391	391	391	0	0	0	0	390	390	0	19	370
C4BENZ	4	0	0	0	0	0	0	0	0	4	0	0	0	0	0	0	0	0	0	0	0	0	0
C11H24	0	0	0	0	0	0	0	0	501	0	0	501	501	501	0	0	0	0	501	501	0	0	501
C10H12	3	0	0	0	0	0	0	0	0	3	0	0	0	0	0	0	0	0	0	0	0	0	0
C12H26	1	0	0	0	0	0	0	0	563	1	0	563	563	563	0	0	0	0	563	563	0	0	563
1234NA	2	0	0	0	0	0	0	0	0	2	0	0	0	0	0	0	0	0	0	0	0	0	0
CC6-HEX	1	0	0	0	0	0	0	0	0	1	0	0	0	0	0	0	0	0	0	0	0	0	0
14DMNAPH	1	0	0	0	0	0	0	0	0	1	0	0	0	0	0	0	0	0	0	0	0	0	0
C13H28	0	0	0	0	0	0	0	0	520	0	0	520	520	520	0	0	0	0	520	520	0	0	520
C7BENZ	1	0	0	0	0	0	0	0	0	1	0	0	0	0	0	0	0	0	0	0	0	0	0
TETMC16	0	1448	0	0	1448	1448	1448	1448	1448	0	0	1448	1448	1448	0	0	0	0	1448	1448	0	0	1448
C22H42O4	0	708	0	0	708	708	708	708	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C14H30	0	0	0	0	0	0	0	0	384	0	0	384	384	384	0	0	0	0	384	384	0	0	384
C15H32	0	1	0	0	1	1	1	1	337	0	0	337	337	337	0	0	0	0	337	337	0	0	337
C16H34	0	71	0	0	71	71	71	71	695	0	0	695	695	695	0	0	0	0	695	695	0	0	695
C17H36	0	1292	0	0	1292	1292	1292	1292	1423	0	0	1423	1423	1423	0	0	0	0	1423	1423	0	0	1423
C18H38	0	2255	0	0	2255	2255	2255	2255	2324	0	0	2324	2324	2324	0	0	0	0	2324	2324	0	0	2324
C19H40	0	0	0	0	0	0	0	0	35	0	0	35	35	35	0	0	0	0	35	35	0	0	35
C20H42	0	2059	0	0	2059	2059	2059	2059	2074	0	0	2074	2074	2074	0	0	0	0	2074	2074	0	0	2074
C21H44	0	0	0	0	0	0	0	0	9	0	0	9	9	9	0	0	0	0	9	9	0	0	9
C22H46	0	530	0	0	530	530	530	530	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C24H50	0	1239	0	0	1239	1239	1239	1239	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C28H58	0	885	0	0	885	885	885	885	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C30H62	0	283	0	0	283	283	283	283	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C32H66	0	708	0	0	708	708	708	708	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4HS	229	1	0	0	1	1	1	1	1	229	0	1	1	1	0	0	0	0	0	0	0	0	0
Enthalpy, mmBtu/hr	-6.40	-6.36	0.03	0.10	-7.20	-5.84	-6.27	-5.30	-5.30	-7.49	-0.67	-6.67	-11.85	-10.49	-1.29	-1.03	-0.26	-1.02	-10.56	-9.93	-0.16	-0.66	-6.38

A.11

A310 Naphtha — 335

A350 Naphtha — 386 — 390 — Air Cooler H-360 — 392 — Naphtha Storage

A310 Diesel — 337

A350 Diesel — 389 — 391 — Heat Recovery HX-385 — 393 — Air Cooler H-365 — 394 — Diesel Storage

A300 PRODUCT COOLING
Upgrading Plant

A.12
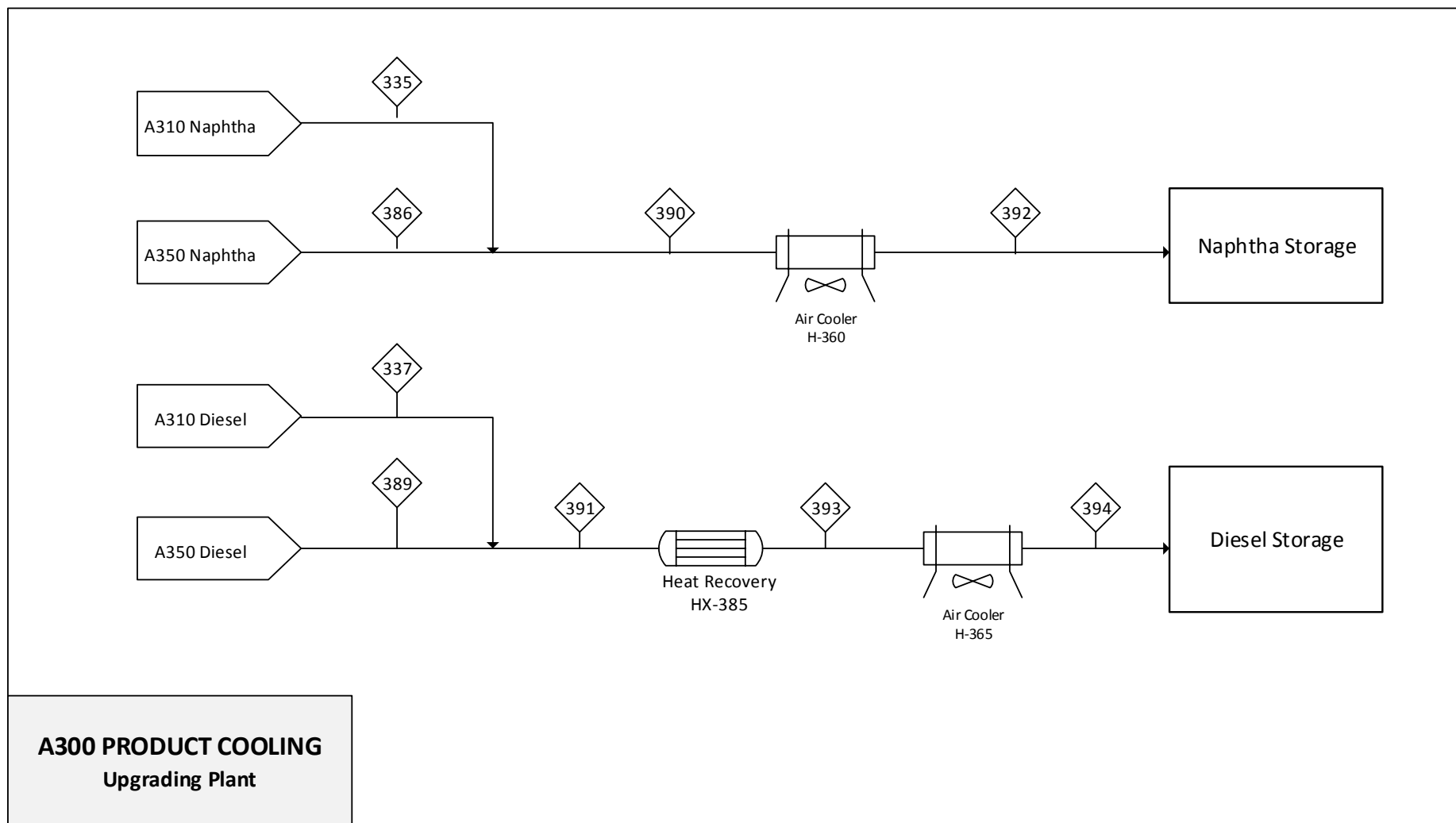
	335	337	386	389	390	391	392	393	394
Total Flow lb/hr	6423	12262	716	10944	7140	23206	7140	23206	23206
Temperature F	205	487	140	520	192	496	140	140	300
Pressure psia	25	19	20	24	20	19	19	17	18
Vapor Frac	0	0	0	0	0	0	0	0	0
Mass Flow lb/hr									
H2O	0	0	1	0	1	0	1	0	0
C3H8	2	0	0	0	2	0	2	0	0
N-C4H10	42	0	41	1	83	1	83	1	1
N-PENTAN	860	0	0	0	860	0	860	0	0
HEXANE	276	0	12	0	288	0	288	0	0
TOLUENE	492	0	0	0	492	0	492	0	0
C16:0FA	0	25	0	0	0	25	0	25	25
2MPENTA	334	0	0	0	334	0	334	0	0
CC5-METH	200	0	0	0	200	0	200	0	0
2MHEXAN	398	0	0	0	398	0	398	0	0
HEPTANE	0	0	63	6	63	6	63	6	6
CC6-METH	160	0	0	0	160	0	160	0	0
PIPERDIN	52	0	0	0	52	0	52	0	0
3MHEPTA	333	0	344	78	677	78	677	78	78
OCTANE	237	0	123	38	360	38	360	38	38
ETHCYC6	255	0	0	0	255	0	255	0	0
ETHBENZ	256	0	0	0	256	0	256	0	0
O-XYLENE	413	0	0	0	413	0	413	0	0
C9H20	590	0	114	137	703	137	703	137	137
CC6-PRO	451	0	0	0	451	0	451	0	0
C3BENZ	278	0	0	0	278	0	278	0	0
4MNONAN	279	1	0	0	279	1	279	1	1
C10H22	337	14	19	370	356	384	356	384	384
C4BENZ	175	352	0	0	175	352	175	352	352
C11H24	0	0	0	501	0	501	0	501	501
C10H12	2	596	0	0	2	596	2	596	596
C12H26	0	424	0	563	0	987	0	987	987
1234NA	0	705	0	0	0	705	0	705	705
CC6-HEX	0	707	0	0	0	707	0	707	707
14DMNAPH	0	884	0	0	0	884	0	884	884
C13H28	0	0	0	520	0	520	0	520	520
C7BENZ	0	1061	0	0	0	1061	0	1061	1061
C8BENZ	0	814	0	0	0	814	0	814	814
TETMC16	0	675	0	1448	0	2124	0	2124	2124
C14H30	0	0	0	384	0	384	0	384	384
C15H32	0	707	0	337	0	1044	0	1044	1044
C16H34	0	2053	0	695	0	2748	0	2748	2748
C17H36	0	2247	0	1423	0	3670	0	3670	3670
C18H38	0	930	0	2324	0	3255	0	3255	3255
C19H40	0	0	0	35	0	35	0	35	35
C20H42	0	65	0	2074	0	2139	0	2139	2139
C21H44	0	0	0	9	0	9	0	9	9
C22H46	0	1	0	0	0	1	0	1	1
Enthalpy, mmBtu/hr	-4.12	-4.72	-0.66	-6.38	-4.78	-11.10	-5.01	-16.36	-14.27

To A400

367

Water

366

Offgas From
A350

To WWT

NH$_3$ Scrubber

A300 AMMONIA SCRUBBER
Upgrading Plant
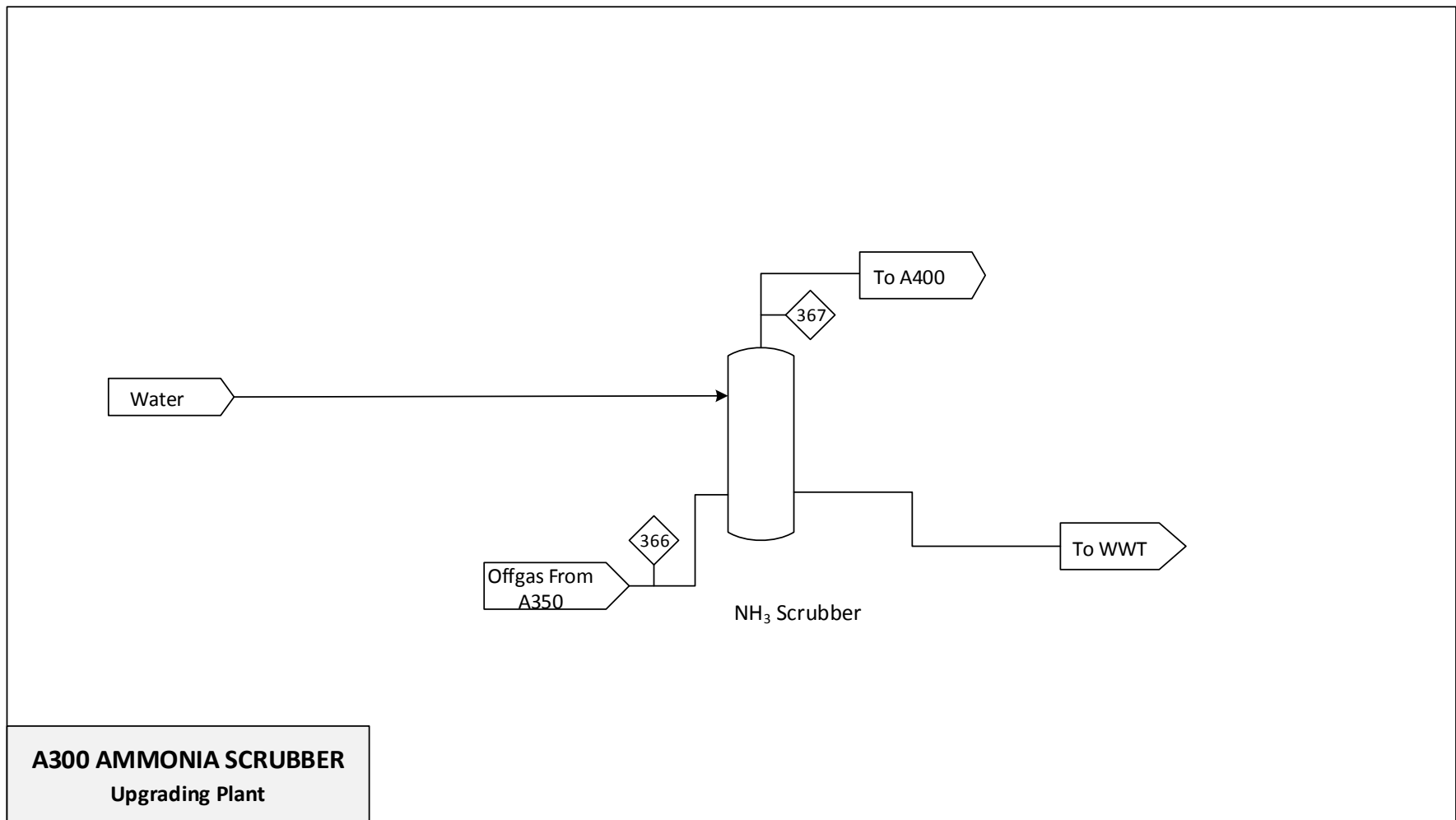
	366	WATER	367	To WWT
Total Flow lb/hr	5557	15313	4905	15965
Temperature F	135	90	114	116
Pressure psia	20	30	20	20
Vapor Frac	1	0	1	0
Mass Flow lb/hr				
H2	428	0	428	0
CO2	168	0	167	0
H2O	216	15313	472	15057
NH3	711	0	36	674
CH4	850	0	850	0
C2H6	1027	0	1027	0
C3H8	555	0	555	0
N-C4H10	215	0	215	0
N-PENTAN	655	0	654	1
HEXANE	80	0	80	0
TOLUENE	39	0	38	0
C16:0FA	0	0	0	0
2MPENTA	126	0	125	0
CC5-METH	47	0	47	0
2MHEXAN	61	0	61	0
HEPTANE	5	0	5	0
CC6-METH	17	0	17	0
PIPERDIN	2	0	1	1
3MHEPTA	35	0	35	0
OCTANE	15	0	15	0
ETHCYC6	10	0	10	0
ETHBENZ	9	0	9	0
O-XYLENE	11	0	11	0
C9H20	14	0	14	0
CC6-PRO	8	0	8	0
C3BENZ	5	0	5	0
4MNONAN	4	0	4	0
C10H22	4	0	4	0
C4BENZ	4	0	4	0
C11H24	0	0	0	0
C10H12	3	0	3	0
C12H26	1	0	1	0
1234NA	2	0	2	0
CC6-HEX	1	0	1	0
14DMNAPH	1	0	1	0
C7BENZ	1	0	1	0
NH4HS	229	0	0	229
Enthalpy, mmBtu/hr	-7.49	-104.21	-7.88	-103.82

A.15

Pre-Reformer

Sulfur Removal

Steam Reformer

Steam Drum

PSA

Hydrogen To A310

Hydrogen To A350

A350 Offgas from NH₃ Scrubber

K-400

K-462

K-461

H-415

H-414

H-420

H-425

H-430

Water Gas Shift

Flue Gas

Air

Natural Gas

Condensate to BFW makeup

Condensate to WWT

415 439 330

413 351

411 425

410 421

367 434 417

402 419 423 420 426

468 416

460 428

461 427 436

401

A400 HYDROGEN PLANT
Upgrading Plant

	330	351	367	401	402	410	411	413	415	417	419	420	421	423	425	426	427	428	434	436	439	460	461	468
Total Flow lb/hr	1988	125	4905	1203	1475	27575	27575	27575	21563	27575	27575	27575	15632	27575	2113	11943	13519	58212	4537	368	27575	58212	58212	72934
Temperature F	140	140	114	60	60	542	900	957	700	300	1562	140	140	572	140	140	140	400	291	114	1050	90	175	322
Pressure psia	336	336	20	450	450	450	450	410	659	348	359	346	346	359	336	346	20	21	450	20	409	15	21	16
Vapor Frac	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	0	1	1	1	0	1	1	1	1
Mass Flow lb/hr																								
H2	1988	125	428	0	0	428	428	434	0	2347	2066	2347	2347	2066	2113	0	235	0	428	0	434	0	0	0
CO	0	0	0	0	0	0	0	41	0	593	4505	593	593	4505	0	0	593	0	0	0	41	0	0	0
N2	0	0	0	7	8	8	8	35	0	36	36	36	36	36	0	0	36	43105	0	0	35	43105	43105	43147
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	13207	0	0	0	13207	13207	2199
AR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	735	0	0	0	735	735	735
CO2	0	0	167	13	16	184	184	1638	0	11427	5279	11427	11418	5279	0	9	11418	29	167	0	1638	29	29	18448
H2O	0	0	472	0	0	21731	21731	20514	21563	12145	14662	12145	211	14662	0	11934	211	1136	168	303	20514	1136	1136	8403
H2S	0	0	0	0	0	0	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	0	0	0	0	0	0
NH3	0	0	36	0	0	36	36	4	0	2	2	2	2	2	0	0	2	0	36	0	4	0	0	2
CH4	0	0	850	1183	1450	2300	2300	4909	0	1024	1024	1024	1024	1024	0	0	1024	0	849	0	4909	0	0	0
C2H6	0	0	1027	0	0	1027	1027	0	0	0	0	0	0	0	0	0	0	0	1027	1	0	0	0	0
C3H8	0	0	555	0	0	554	554	0	0	0	0	0	0	0	0	0	0	0	554	1	0	0	0	0
N-C4H10	0	0	215	0	0	214	214	0	0	0	0	0	0	0	0	0	0	0	214	1	0	0	0	0
N-PENTAN	0	0	654	0	0	646	646	0	0	0	0	0	0	0	0	0	0	0	646	8	0	0	0	0
HEXANE	0	0	80	0	0	78	78	0	0	0	0	0	0	0	0	0	0	0	78	2	0	0	0	0
TOLUENE	0	0	38	0	0	34	34	0	0	0	0	0	0	0	0	0	0	0	34	4	0	0	0	0
C16:0FA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
2MPENTA	0	0	125	0	0	123	123	0	0	0	0	0	0	0	0	0	0	0	123	3	0	0	0	0
CC5-METH	0	0	47	0	0	46	46	0	0	0	0	0	0	0	0	0	0	0	46	2	0	0	0	0
2MHEXAN	0	0	61	0	0	58	58	0	0	0	0	0	0	0	0	0	0	0	58	3	0	0	0	0
HEPTANE	0	0	5	0	0	5	5	0	0	0	0	0	0	0	0	0	0	0	5	0	0	0	0	0
CC6-METH	0	0	17	0	0	15	15	0	0	0	0	0	0	0	0	0	0	0	15	1	0	0	0	0
3MHEPTA	0	0	35	0	0	30	30	0	0	0	0	0	0	0	0	0	0	0	30	5	0	0	0	0
OCTANE	0	0	15	0	0	13	13	0	0	0	0	0	0	0	0	0	0	0	13	3	0	0	0	0
ETHCYC6	0	0	10	0	0	8	8	0	0	0	0	0	0	0	0	0	0	0	8	2	0	0	0	0
ETHBENZ	0	0	9	0	0	7	7	0	0	0	0	0	0	0	0	0	0	0	7	2	0	0	0	0
O-XYLENE	0	0	11	0	0	8	8	0	0	0	0	0	0	0	0	0	0	0	8	3	0	0	0	0
C9H20	0	0	14	0	0	9	9	0	0	0	0	0	0	0	0	0	0	0	9	5	0	0	0	0
CC6-PRO	0	0	8	0	0	5	5	0	0	0	0	0	0	0	0	0	0	0	5	3	0	0	0	0
C3BENZ	0	0	5	0	0	3	3	0	0	0	0	0	0	0	0	0	0	0	3	2	0	0	0	0
4MNONAN	0	0	4	0	0	2	2	0	0	0	0	0	0	0	0	0	0	0	2	2	0	0	0	0
C10H22	0	0	4	0	0	2	2	0	0	0	0	0	0	0	0	0	0	0	2	2	0	0	0	0
C4BENZ	0	0	4	0	0	1	1	0	0	0	0	0	0	0	0	0	0	0	1	2	0	0	0	0
C10H12	0	0	3	0	0	1	1	0	0	0	0	0	0	0	0	0	0	0	1	2	0	0	0	0
C12H26	0	0	1	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1	0	0	0	0
1234NA	0	0	2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2	0	0	0	0
CC6-HEX	0	0	1	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1	0	0	0	0
14DMNAPH	0	0	1	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1	0	0	0	0
C13H28	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C7BENZ	0	0	1	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1	0	0	0	0
TETMC16	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C22H42O4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C15H32	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C16H34	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C17H36	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C18H38	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C19H40	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C20H42	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C21H44	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C22H46	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C24H50	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C28H58	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C30H62	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
C32H66	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4HS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Enthalpy, mmBtu/hr	0.44	0.03	-7.88	-2.44	-2.99	-127.43	-121.36	-121.36	-118.93	-118.61	-87.74	-128.91	-47.49	-106.37	0.47	-81.42	-47.92	-2.04	-5.50	-2.11	-119.71	-6.49	-5.27	-114.72

A.17

A300, A400 Process Heat

415 — Steam to A400

Steam Turbine E-601

Steam System A600

645

648 — Condenser HX-600 — 647

618 633 — Boiler Feedwater Makeup

426 — A400 Condensate

600

Power

Blowdown

Cooling Water Service – A700

715 — Cooling Water to A310,350

709 — Losses

713 — Blowdown

730 — A310, 350 Service Return

760 — A600 Service Return

710 — Cooling Tower Makeup

**A600 STEAM SYSTEM
A700 COOLING WATER**
Upgrading Plant

A.18

	415	426	600	618	633	645	648	647	709	710	713	715	730	760
Total Flow lb/hr	21563	11943	851	10479	42546	20124	20124	20124	1016260	23485	4453	1036140	100370	935773
Temperature F	700	140	112	60	232	700	148	148	90	60	89	89	110	110
Pressure psia	659	346	15	15	22	659	4	4	15	15	15	75	60	60
Vapor Frac	1	0	0	0	0	1	0	1	0	0	0	0	0	0
Mass Flow lb/hr														
H2		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
N2		0	0	0	0	0	0	0	0	0	0	0	0	0
CO2		9	0	0	9	0	0	0	0	0	0	0	0	0
H2O	21563	11934	851	10479	42538	20124	20124	20124	1016260	23485	4453	1036140	100370	935773
NH3		0	0	0	0	0	0	0	0	0	0	0	0	0
CH4		0	0	0	0	0	0	0	0	0	0	0	0	0
C2H6		0	0	0	0	0	0	0	0	0	0	0	0	0
Enthalpy, mmBtu/hr	-118.93	-81.42	-5.78	-71.66	-283.58	-111.05	-135.85	-117.56	-6990.40	-162.35	-30.63	-7127.60	-688.07	-6415.00

A.19

Heat Exchanger Duty Summary			Power Summary	
Process Flow Diagram	Equip ID	Duty, mmBTU/hr	Equip ID	HP
A100 - HTL				
	H-100	0.61	P101a	3.34
	Hx-100	15.41	P101b	160.87
	H-105	5.24		
	R-101	3		
	HX-102	5.26		
	H-110	0.18		
A110 - Combustor				
	H-151	0.52	K-150	50.92
A200 - Aqueous Treatment			K-155	39.09
	H-230	3.14	K-230	36.01
A310 - Hydrotreating				
	H-311	7.25	P-310	106.57
	HX-318A	2.31	K-311	585.59
	H-316	4.84	K-310	1497.52
	E-310	1.36		
	T-310 Condenser	0.64		
	T-310 Reboiler	4.34		
	T-320 Condenser	4.55		
	T-320 Reboiler	6.26		
	T-336 Condenser	2.64		
	T-336 Reboiler	2.87		
A350 – Hydrocracking				
	H-351	1.36	P-350	35.53
	H-352	0.53	K-350	53.91
	H-371	1.36	K-355	3.51
	T-350 Condenser	0.24		
	T-350 Reboiler	2.98		
A300 - Product Cooling				
	H-360	0.23		
	HX-385	3.18		
	H-365	2.08		
A400 - Hydrogen Plant				
	H-414	6.07	K-400	841.39
	H-415	1.64	K-461	477.53
	H-420	18.63	K-462	277.69
	H-425	12.24		
	H-430	10.3		
A600 - Steam System				
	HX-600	18.29	E-601	2556.50

Appendix B- Equipment Cost Details

HTL 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 (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2014$	Scaled Uninstalled Cost in 2014$	Ref
A 10			**Feed Prep - Sludge Dewatering and Grinding**																
C-10	1	0	Bowl Centrifuge - GEA	raw sludge rate	26,417	29,296	gal/hr	1.11	$700,000	2016	b	$700,000	0.6	744,817	1.8	1,340,670	1,425,808	$792,115	2
			A10 Total														**1,425,808**	**792,115**	
A100			**HTL Oil Production**																
P-101	1	0	Twin Screw Feeder and Feed Pump	S101	575	69	gal/min	0.12	$1,094,100	2011	b	$1,094,100	0.8	200,098	2.3	460,226	452,683	$196,819	1
HX-100	1	0	Feed/Product Exchangers	Area	35,540	1,284	ft2	0.04	$31,860,000	2012	b	$31,860,000	0.7	3,117,016	2.2	6,857,436	6,757,730	$3,071,695	1
H-105	1	0	Feed Trim Heater	duty	32.1	5.2	mmBtu/hr	0.16	$998,850	2012	b	$998,850	0.7	279,350	2.2	614,571	605,635	$275,289	1
R-101	1	0	HTL Reactor	S101	33,333	36,667	lb/h	1.10	$770,992	1Q 2014	b	$770,992	1	848,109	1.2	975,326	980,431	$852,549	3
K-110	1	0	Reactor K/O Drum	S101	304,939	36,667	lb/h	0.12	$5,700,243	2011	b	$5,700,243	0.7	1,294,010	2.0	2,588,021	2,545,601	$1,272,801	1
	1	0	Hot Oil System	duty	64	8.6	mmBtu/hr	0.14	$1,265,600	2011	b	$1,265,600	0.6	380,997	1.4	545,968	537,020	$374,752	1
	1	0	Dowtherm	capacity of system	23,200	3,137	gal	0.14	$782,847	2011	b	$782,847	1	105,857	1.0	105,857	104,122	$104,122	1
HX-102	1	0	Reactor Product Air Cooler	duty	5.12	5.26	mmBtu/hr	1.03	$63,200	1Q 2014	b	$63,200	0.7	64,405	2.0	127,521	128,189	$64,742	3
F-105	1	0	Solids Filter	S101	2,420	69	gal/min	0.03	$1,017,998	2011	b	$1,017,998	0.6	120,198	1.7	204,337	200,988	$118,228	1
V-110	1	0	Oil/Water Separator	S101	2,420	69	gal/min	0.03	$2,653,959	2011	b	$2,653,959	0.7	219,485	2.0	438,969	431,774	$215,887	1
H-110	1	0	Biocrude Cooler	duty	0.170	0.181	mmBtu/hr	1.06	$63,200	1Q 2014	b	$63,200	0.7	66,008	5.4	356,442	358,308	$66,353	3
			A100 Total														**13,102,481**	**6,613,237**	
A200			**HTL Water Treatment**																
CL-215	1	0	Lime Softening	S101	34,000	36,667	lb/hr	1.08	$410,100	1Q2014	b	$410,100	0.65	430,730	2.4	1,033,752	1,039,163	$432,985	5
T-220	1	0	Ammonia Stripping	S101	34,000	36,667	lb/hr	1.08	$596,800	1Q2014	b	$596,800	0.65	626,822	2.4	1,504,373	1,512,248	$630,103	5
TX-230	1	0	THROX Unit	S101	34,000	36,667	lb/hr	1.08	$210,900	1Q2014	b	$210,900	0.65	221,509	2.4	531,622	534,405	$222,669	5
			A200						$542,000								**$3,085,816**	**$1,285,757**	
A700			**OSBL - including cooling water system**																
	1	0	Cooling Tower System (packaged)	circ rate	150	18	gal/min	0.12	$5,900	1Q 2014	b	$5,900	0.6	1,659	10.1	16,756	16,844	$1,668	3
	1	1	Cooling Water Pump	circ rate	16	18	gal/min	1.13	$4,400	1Q 2014	b	$8,800	0.6	9,477	4	37,907	38,105	$9,526	3
	1	0	Plant Air Compressor	S101 - dry	2,205	110	ton/day	0.05	$87,922	2007	b	$87,922	0.3	35,770	1.57	56,158	61,577	$39,221	4
	1	0	Firewater Pump	S101 - dry	2,205	110	ton/day	0.05	$23,043	2007	b	$23,043	0.3	9,375	3.7	34,686	38,033	$10,279	4
	1	0	Instrument Air Dryer	S101 - dry	2,205	110	ton/day	0.05	$8,349	2002	b	$8,349	0.6	1,382	2.47	3,413	4,971	$2,012	4
	1	0	Plant Air Receiver	S101 - dry	2,205	110	ton/day	0.05	$21,005	2007	b	$21,005	0.65	2,993	5.44	16,280	17,851	$3,281	4
	1	0	Firewater Storage Tank	S101 - dry	2,205	110	ton/day	0.05	$229,900	2007	b	$229,900	0.65	32,755	1.46	47,822	52,437	$35,916	4
	1		Biocrude Storage - 3 day	S136	31,719	34,419	gal	1.09	$280,900		b	$280,900	0.65	296,216	1.3	385,081	387,097	$297,767	3
			A700 Total							Subtotal							**$616,915**	**$399,671**	
			References												Total Equipment Cost		**$18,231,020**	**$9,090,780**	
			1 Knorr et al. 2013																
			2 Vendor Budget Estimate																
			3 Aspen Capital Cost Estimator, 8.8																
			4 Dutta et al. 2011																
			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)	Total Original 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 2014$	Scaled Uninstalled Cost in 2014$	Ref
A300			**HTL Oil Upgrading and Product Separation**																
A310			**HTL Oil Hydrotreating & Separations**																
	1	0	Desalter	S301	2,732	2,732	bpd feed	1.00	$768,541	2007	i	$768,541	0.75	768,541	2.5	768,541	842,704	$341,176	7
	1	0	Hydrotreater Fe Bed	S301	15,472	2,732	bpd feed	0.18	$16,302,021	2014	i	$16,302,021	0.75	4,440,161	1.5	4,440,161	4,440,161	$2,940,504	1
R-310	1	0	Hydrotreater Reactor, vessels, columns	S301	15,472	2,732	bpd feed	0.18	$77,087,500	2014	i	$77,087,500	0.75	20,996,224	1.5	20,996,224	20,996,224	$13,904,784	1
K-311	1	0	Hydrogen Compressor	S304-H2	3,786.0	4,926.0	lb/hr H2	1.30	$1,385,600	1Q 2011	b	$1,385,600	0.8	1,710,365	1.1	1,881,402	1,895,882	$1,723,529	2
V-300	1	0	PSA for Hydrogen Recycle	S425-H2	10	9.5	mmscfd H2	0.95	$1,750,000	2004	b	$1,750,000	0.8	1,686,237	2.5	4,165,005	5,401,754	$2,186,945	3
			A310 Total							Subtotal					HDO/N/S			$33,576,724	$21,096,937	
A350			**Hydrocracking and Separations**																
R-350	1	0	Hydrocracker Unit + auxiliaries	S338	15,472	1,021	bpd feed	0.07	$50,858,783	2014	i	$50,858,783	0.75	6,619,706	1.5	6,619,706	6,619,706	$4,383,911	1
			A350 Total							Subtotal					A350 HCK			$6,619,706	$4,383,911	
A400			**Hydrogen Plant - OSBL**																
	1	0	Stm Reformer system w/ associated OSBL	S425-H2	24.5	9.5	mmscfd H2	0.39	$50,198,438	2014	i	$50,198,438	0.65	27,204,148	1.9	27,204,148	27,204,148	$14,168,827	4
A400 Total										Subtotal				A400 H2 Plant			$27,204,148	$14,168,827	
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	1.2	1,632,657	1,641,203	$1,379,162	8
A600 Total										Subtotal				A600 STM Gen			$1,641,203	$1,379,162	
A700			**OSBL - including cooling water system**																
	1	0	Wastewater treatment (NH3 stripping)	S320+SCRB-	34,000	21,135	lb/hr	0.62	$1,217,800	1Q 2014	b	$1,217,800	0.65	894,059	2.4	2,145,741	2,156,974	$898,739	8
	1	0	Cooling Tower System	S715	7,506,000	1,036,143	lb/hr	0.14	$260,852	2010	b	$260,852	0.78	55,668	2.5	137,499	143,815	$58,225	6
	1	0	Cooling Water Pump	S715	7,001,377	1,036,143	lb/hr	0.15	$239,375	2007	b	$239,375	0.3	134,943	2.1	288,778	316,644	$147,964	6
	1	0	Makeup water pump	S618+S710	80,411	33,964	lb/hr	0.42	$6,528	2007	b	$6,528	0.3	5,041	4.7	23,792	26,088	$5,527	6
	1	0	Plant Air Compressor	S101(dry) X 10	262,454	61,113	lb/hr	0.23	$87,922	2007	b	$87,922	0.3	56,784	1.6	89,150	97,753	$62,263	6
	1	0	Firewater Pump	S101(dry) X 10	262,454	61,113	lb/hr	0.23	$23,043	2007	b	$23,043	0.3	14,882	3.7	55,064	60,377	$16,318	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	2.5	8,602	12,526	$5,071	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	5.4	44,312	48,588	$8,932	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	1.5	130,165	142,726	$97,758	6
	1	0	Feed Storage - 3 day	S301	1,000,000	344,186	gallons	0.34	2,371,900	2010	b	$2,371,900	0.7	1,124,221	1.4	1,573,909	1,582,148	$1,130,106	5
	1	0	Product Storage - 3 day	S392	1,000,000	83,952	gallons	0.08	2,371,900	2010	b	$2,371,900	0.7	418,715	1.4	586,200	613,126	$437,947	5
	1	0	Product Storage - 3 day	S393	1,000,000	250,848	gallons	0.25	2,371,900	2010	b	$2,371,900	0.7	900,915	1.4	1,261,281	1,319,216	$942,297	5
			A700 Total							Subtotal					A700 OSBL			$6,519,982	$3,811,148	
														Total Equipment Cost			$75,561,763	$44,839,985	

References
1. IHS PEP Yearbook 2014a - "Diesel from High Pressure Hydrocracking" (P=2000 psia)
2. Aspen Capital Cost Estimator, 7.3.2
3. Vendor Budget Estimate
4. IHS PEP Yearbook 2014b - "Hydrogen from Natural Gas by Steam Reforming"
5. Aspen Capital Cost Estimator, 7.3.1
6. Dutta et al. 2011
7. Kaiser and Gary 2007
8. Aspen Capital Cost Estimator, 8.8

Appendix C - Compound Selection

C.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 C.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 C.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%						

C.2 Hydrotreated Oil Model Compounds

Similar to HTL oil, hydrotreated oil contains numerous compounds and a limited number are used for modeling purposes. Table C.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 C.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 D- Indices

Table D.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	2010	21.07
2006	19.60	2011	21.46
2007	19.55	2012	21.45
2008	19.50	2013	21.40
2009	20.30	2014	21.49

Table D.2. Capital cost indices.

Chemical Engineering Magazine, CEI annual index

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

Table D.3. Inorganic chemical indices.

Source: SRI International Chemical Economics Handbook, Economic Environment of the Chemical Industry, September 2006
Current indices @https://www.sriconsulting.com/CEH/Private/EECI/EECI.pdf

YEAR	INDEX	YEAR	INDEX
1990	123.6	2003	164.6
1991	125.6	2004	172.8
1992	125.9	2005	187.3
1993	128.2	2006	196.8
1994	132.1	2007	203.3
1995	139.5	2008	228.2
1996	142.1	2009	224.7
1997	147.1	2010	233.7
1998	148.7	2011	252.1
1999	149.7	2012	260.3
2000	156.7	2013	263.9
2001	158.4	2014	269.2
2002	157.3	2015	264.8

Appendix E – TCLP Testing Results

Table E.1. TCLP analysis for HTL solids from Detroit sludge (methods 6010B for metals and 7470A for mercury).

Metal	WW-06 BC Comp, mg/L	Federal Limits, mg/L
As	n.d.	5
Ba	0.11	100
Cd	n.d.	1
Cr	n.d.	5
Pb	0.12	5
Hg	n.d.	0.2
Ni	0.48	n/a
Se	n.d.	1
Ag	n.d.	5
Zn	16.7	n/a

HTL Solids TCLP Analysis

Appendix F - Process Efficiency Metrics

In addition to setting technical and economic targets, BETO determines process efficiency metrics that can be used to estimate environmental sustainability of biofuels/products pathways, including product yield, carbon efficiency, energy inputs, and water consumption. These metrics are used in supply chain sustainability analyses for the pathways using the GREET (ANL 2016a) and WATER (ANL 2016b) models. Table F.1 lists metric values for the HTL plant (sludge to biocrude) and the overall sludge to upgraded fuel conversion process. Solids reduction efficiency is also listed for the sludge HTL process as it is of particular interest to WWTPs for any sludge treatment strategy under consideration.

Table F.1. Process efficiency metrics for sludge HTL and biocrude upgrading goal case.

Process Efficiency Metric	Sludge to Biocrude	Overall Sludge to Fuel Blendstocks
Product Yield (% w/w of feedstock)[a]	48%	39%
Carbon Efficiency of Conversion Process		
% C in fuel / feedstock	72%	64%
% C in fuel / feedstock + natural gas	66%	53%
Net Electricity Import (kWh/gge product)	1.6	2.0
Purchased Natural Gas Import (million Btu/gge product)[b]	0.02	0.04
Product Energy / Input Energy (LHV)[c]	4.2	2.8
Water Consumption (gal/gge product)	0.05	1.2
Solids Reduction, %		
Dry basis	76%	n/a
Dry, ash-free basis	89%	

(a) On a dry, ash-free basis

(b) Calculated using natural gas lower heating value of 21,152 Btu/lb from process model.

(c) Includes only direct inputs of electricity and natural gas (for HTL and THROX unit). Biocrude transport energy not included.

References

ANL. 2016a. *Greenhouse Gases, Regulated Emissions, and Energy Use in Transportation (GREET) Model*. Argonne National Laboratory, Argonne, IL. http://greet.es.anl.gov.

ANL. 2016b. *Water Analysis Tool for Energy Resources (WATER)*. Argonne National Laboratory, Argonne, IL. http://Water.es.anl.gov