A Techno-Economic and Environmental Assessment of Hydroprocessed Renewable Distillate Fuels

by

Matthew Noah Pearlson

B.S., University of Massachusetts (2007)

Submitted to the Engineering Systems Division
in partial fulfillment of the requirements for the degree of

Master of Science in Technology and Policy

at the

MASSACHUSETTS INSTITUTE OF TECHNOLOGY

June 2011

Author . ...
Engineering Systems Division
May 6, 2011

Certified ....
James I. Hileman
Principal Research Engineer, Department of Aeronautics and Astronautics
Thesis Supervisor

Accepted by .. ..
Dava J. Newman
Director, Technology and Policy Program

A Techno-Economic and Environmental Assessment of Hydroprocessed Renewable Distillate Fuels

by

Matthew Noah Pearlson

Submitted to the Engineering Systems Division
on May 6, 2011, in partial fulfillment of the
requirements for the degree of
Master of Science in Technology and Policy

Abstract

This thesis presents a model to quantify the economic costs and environmental impacts of producing fuels from hydroprocessed renewable oils (HRO) process. Aspen Plus was used to model bio-refinery operations and supporting utilities. Material and energy balances for electricity, carbon dioxide, and water requirements as well as economic costs were obtained from these models. A discounted-cash-flow-rate-of-return (DCFROR) economic model was used to evaluate minimum product values for diesel and jet fuels under various economic conditions. The baseline gate cost for distillate fuel production were found to range between \$3.80 and \$4.38 per gallon depending on the size of the facility. The additional cost for maximizing jet fuel production ranged between \$0.25 and \$0.30 per gallon. While the cost of feedstock is the most significant portion of fuel cost, facility size, financing, and capacity utilization were found to be sensitive parameters of the gate cost. The total water use of the system was found to be 0.9 pounds of water per pound of vegetable oil processed. Lifecycle greenhouse gas emissions (GHGs) for the processing step were found to range between 10.1 and 13.0 gCO_2e per MJ of distillate fuel using an energy allocation method consistent with methods in the literature. Finally, the policy landscape for producing jet and diesel fuels from renewable oils was reviewed from the perspective of a fuel producer. It was found that the potential of HRO fuels penetrating the market is dependent on the availability of feedstocks and access to capital.

Thesis Supervisor: James I. Hileman
Title: Principal Research Engineer, Department of Aeronautics and Astronautics

Acknowledgments

I would like to appreciate the following individuals and organizations for their assistance and support. First, Dr. James Hileman, Dr. Christoph Wollersheim, Mr. Russell Stratton, and Mr. Michael Hagerty for their assistance and feedback in preparing this work. The Federal Aviation Administration and the Air Force Research Lab for their financial support of this project. Mr. Michael Peters of McKinsey and Co., and Mr. Brian Morrissey of Citizen's Energy for their wealth of economic and project financing knowledge and acumen. Mr. Ramin Abhari of Syntroleum for sharing his HRO processing expertise. I'd also like to appreciate AspenTech for providing AspenPlus software, Dr. Randy Field of the MIT Energy Initiative for his generosity and kindness assisting with Aspen, and Wilfried Mofor and Enrique Citron of the MIT chemical engineering undergraduate program who provided modeled some of the auxiliary processes. Finally, to Leslie, who changed my life forever with a question and a dare, and has been supportive ever since.

Contents

1 Introduction **15**

2 Transportation Fuels Background **17**

2.1 Motivation 17

2.1.1 Petroleum: a primer on the scale of the problem 17

2.1.2 Alternative Fuel Process Options and Scope of Work 19

2.2 Transportation Fuels 19

2.3 Renewable Diesel and Jet Fuels from the Hydroprocessed Renewable Oil Process 21

2.3.1 Process Chemistry 22

2.3.2 Renewable Oils 24

2.4 Naphtha and Gasoline 25

2.4.1 Background 26

2.4.2 On-site Naphtha Upgrade 26

2.4.3 On-site Naphtha Use for Utilities 27

2.4.4 Naphtha Sold for Upgrading or Blending Elsewhere 27

2.5 FAME: a first generation alternative middle distillate fuel 28

3 Hydroprocessing Plant Design **31**

3.1 The Hydroprocessed Renewable Oil Process 31

3.1.1 Feedstock and Feed Storage 33

3.1.2 Hydro-Deoxygenation 33

3.1.3 Selective Isomerization and Catalytic Cracking 34

3.1.4 Heat Integration for Steam Generation and Cooling Water . . 34
3.1.5 Fuel Gas Cleanup and Recycle 35
3.1.6 Hydrogen Gas Production 35
3.1.7 Product Separation 36
3.1.8 Product Storage and Blending 36
3.2 Offsites, Special Costs, and Other aspects not modeled explicitly . . . 36
3.3 Product Profiles 37
3.4 Hydrogen Production and Purchase 37
3.5 Process Utilities 38

4 Economic Modeling 41

4.1 Introduction 41
4.2 Capital and Operating Expenses 41
4.2.1 Capital Expenses 42
4.2.2 Operating Expenses 43
4.3 Gross Income: Estimation of Refinery Sales 47
4.3.1 Product Profiles and Material Balance 47
4.3.2 Products and Historic Gate Prices 48
4.3.3 Return On Investment Analysis 50
4.4 Discounted Cash Flow Rate of Return 51
4.4.1 Plant Sizes 54
4.4.2 Hydrogen Source 55
4.4.3 Equity Structure 55
4.4.4 Input and Output Price Sensitivity 55
4.4.5 Production Capacity: Start-up Penalty and Feedstock Shortages 56

5 Environmental Model 59

5.1 Introduction 59
5.2 Water Usage Model 59
5.2.1 Cooling Water and Steam Systems 59
5.2.2 Water Integration Designs 60

5.2.3 Water Production 62

5.3 Green House Gas Model 62

6 Results **63**

6.1 Introduction 63

6.2 Economic Costs of Production 63

6.2.1 Capital Expenses and Economies of Scale 64

6.2.2 Direct and Variable Operating Costs 66

6.3 Cash Flow Modeling Using ROI 69

6.4 Discounted Cash Flow Rate of Return 70

6.4.1 Baseline Results 71

6.4.2 Product Slate and Co-Product Price 72

6.4.3 Brownfield and On-Site Hydrogen Production 73

6.4.4 Financing 73

6.4.5 Production Level and Ramp-Up 74

6.4.6 Vegetable Oil Cost Sensitivity 74

6.5 Environmental Results 77

6.5.1 Water Usage 77

6.5.2 Lifecycle Greenhouse Gas Emissions 78

7 Policy and Market Review, Discussion, and Implications **81**

7.1 Introduction and Motivation 81

7.2 New HRO Plants versus Old Biodiesel Infrastructure 82

7.3 Trade-offs of Environmental and Economic Costs 83

7.4 Feedstocks Supplies 85

7.5 Large versus Small Suppliers 86

7.6 Financing 87

7.7 Viability 88

8 Concluding Remarks and Future Work **91**

8.1 Summary 91

8.2 Future Work 93

A Aspen Models **95**

B Excel Model **97**

List of Figures

2-1 Distribution of finished petroleum products from a barrel of crude oil 18
2-2 Carbon number and boiling point of gasoline, jet, and diesel fuel cuts 20
2-3 Renewable oil deoxygenation reaction pathways 22
2-4 Example of an isomerization reaction for dodecane (n-$C_{12}H_{26}$) 23
2-5 Example of a cracking reaction for dodecane ($C_{12}H_{26}$) 24
2-6 Biodiesel Reaction Pathway 28

3-1 Simplified Hydroprocessed Renewable Oil (HRO) system design . . . 32

4-1 Historic refinery product gate prices 49

5-1 Cooling water design reported in Huo et al. 61
5-2 Alternate water integration design. 61

A-1 Process flowsheet in Aspen Plus 96

B-1 Screenshot of the DCFROR spreadsheet model 98

List of Tables

2.1 Properties and Specifications of Fossil and Renewable Diesel and Jet-A Fuels 21
2.2 Current renewable oil and fat resources in the United States 24
2.3 Renewable oil chain length profiles for selected renewable oils 25
2.4 Typical properties of biodiesel (FAME) and petroleum diesel 29

3.1 Mass-based product yields by product profile 38
3.2 Process utility requirements 39

4.1 List of capital equipment and other project costs 42
4.2 List of fixed operating expenses 44
4.3 Variable operating expenses 46
4.4 Historic annual average refinery product gate prices 48
4.5 Cash flows needed for ROI calculation 52
4.6 DCFROR Assumptions 53
4.7 Production ramp-up schedule 57

6.1 Capital Expenses for each plant size 65
6.2 Direct annual operating expenses 67
6.3 Variable Operating Expenses 68
6.4 Summary of cost per gallon contributions 69
6.5 Cash flows results for ROI calculation 70
6.6 DCFROR gate cost and sensitivity results 72
6.7 Feed vegetable oil cost for fixed fuel gate cost 75

6.8 Overall system water production and makeup demand. 78
6.9 Process greenhouse gas emissions. 80

Chapter 1

Introduction

"When you don't know where you are going, all roads will take you there." - Yiddish Proverb

The technical means of producing alternative fuels from renewable oils, and the resulting carbon intensity has been documented in previous work [24, 85, 33, 59, 47]. However, an accurate cost of production for distillate fuels is not available in the literature. The questions investigated in this work are, what are the costs associated with a hydroprocessed renewable oil (HRO) facility? How much does it cost to produce renewable distillate fuels from vegetable oils? What additional cost penalty is incurred for producing more jet fuel in addition to, or instead of, diesel? What are the carbon, water, and cost intensities for this production method? What is the policy landscape for producing jet and diesel fuels from renewable oils?

The purpose of this techno-economic and environmental analysis is to examine the economic and environmental costs of producing liquid transportation fuels from renewable oils and fats. This work models the hydroprocessing of renewable oils (HRO) in Aspen Plus for jet and diesel fuel production, and determines the economic and environmental costs under various process designs and economic scenarios. The process was modeled based on published literature and interviews with industry professionals familiar with hydroprocessing technology. This model was refined using Aspen Plus modeling software to determine material and energy balances. The process economics were calculated using the cost curve method from petroleum handbooks based on the results of the modeling effort. Finally, a Discounted Cash Flow Rate of Return (DCFROR) model combined the process material and energy flows with the estimated cost of production to calculate the gate cost of fuel and feedstocks under various plant size and cost conditions.

The remainder of this thesis is broken out as follows. Chapter 2 describes the current state of petroleum fuels and non-petroleum based alternatives that are in production. Chapter 3 describes the design of the hydroprocessing plant and the details of the model. Chapter 4 describes the economic model and the assumptions used for evaluating the economic costs of production. Chapter 5 describes the environmental model, with specific focus on carbon dioxide emissions, and water requirements for the process. Chapter 6 presents the results of the models. Chapter 7 discusses the results in light of current policy, and suggests future policy mechanisms for meeting alternative and renewable domestic fuel production targets. Chapter 8 summarizes the thesis work, major findings, and makes suggestions for future work. Appendix A contains information on the Aspen Plus model. Appendix B contains information on the techno-economic models used for the economic and environmental analysis.

Chapter 2

Transportation Fuels Background

"Energy and persistence alter all things." - Benjamin Franklin

2.1 Motivation

Alternative fuel use is motivated by trends in petroleum prices, concerns about the environment, and the desire for a distributed, domestic, and renewable fuel production infrastructure. The introduction of alternative fuels is not only dependent on technical feasibility and environmental impact, but also on economic viability. In this effort, a fuel production method that looks promising for reducing life cycle GHG emissions was examined for the economic cost of production. A techno-economic model for the production of hydroprocessed renewable oil (HRO) diesel and jet fuels was created from well-established petroleum and chemical engineering methods and cost estimation techniques.

2.1.1 Petroleum: a primer on the scale of the problem

Petroleum is the single largest source of energy in the transportation sector [91]. It is used in single stroke lawn mower engines, diesel powered luxury cruise liners, and everything in between. In 2010 the United States consumed approximately 20 million barrels of oil products per day, or roughly 23% of the total world wide demand [91].

Approximately 43% is used in motor gasoline for passenger car and light duty truck engines, 22% for diesel, and 9% for jet fuel. Between 2005 and 2010, total jet and diesel fuel demand ranged between five and six million barrels per day (BPD) of consumption. The remainder of petroleum products are natural gas and liquid propane gases (LPG), heavy fuel oil, which is burned in large cargo ships, and other products used for the specialty chemical and polymer industries. This distribution is depicted in Figure 2-1.

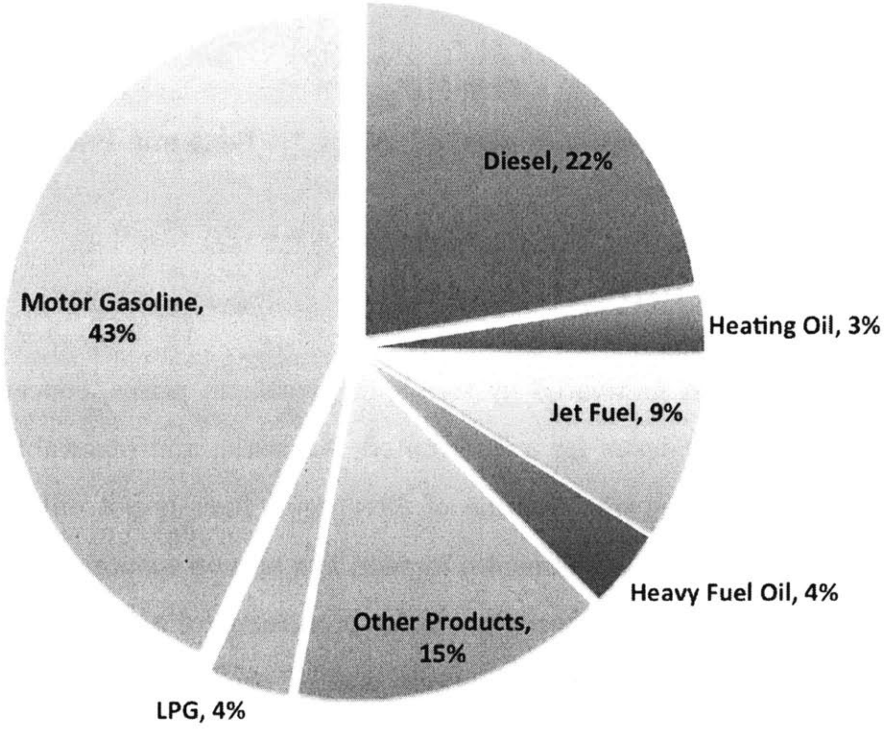


Figure 2-1: The distribution of finished fuel products from a barrel of crude oil in 2009. Source: [91]

There were 148 refineries operating in 2010 in the United States with a total capacity of 17.5 million BPD [92]. This means that the average refining capacity in the US is on the order of 100,000 BPD. By contrast, the entire biodiesel capacity in the United States in 2010 was approximately 163,000 BPD with 120-170 consutrctured facilities and average capacity of 1,200 BPD [55].

2.1.2 Alternative Fuel Process Options and Scope of Work

Because of the scale of the petroleum industry, creating sufficient alternative fuels to meet petroleum demand is not trivial. No single alternative fuel technology or production pathway is sufficient to satisfy the fuel demand for various technical and economic reasons. For example, even a large bio-fuel facility of 6,500 BPD is not capable of supplying a medium sized airport, such as Logan International Airport in Boston, Massachusetts, which consumes 25,000 BPD [45].

The scope of this work deals exclusively with hydroprocessing renewable oils and compares it to FAME production of biodiesel and petroleum diesel and jet fuels. Biodiesel is reviewed in detail in Section 2.5 for completeness. However, there are a number of other technologies for converting biomass into transportation fuels including aqueous phase reforming, gasification and Fischer-Tropsch synthesis, fatty-acid trans-esterification (FAME), hydrothermal catalysis, methanol to olefins, and pyrolysis to name a few. There are excellent reviews of each technology, as well as economic and policy considerations in the literature. A non-exhaustive list is given here: General biofuels review: [53, 32, 69]. Ethanol and food versus fuel debate: [70, 74]. Algae and synthetic feedstocks: [6, 95, 97, 98]. Biodiesel: [46, 18, 52, 9, 1]. Fischer-Tropsch (coal, natural gas, and biomass to liquids): [27, 38, 73, 84]. Pyrolysis: [13, 8, 35].

2.2 Transportation Fuels

Transportation fuels are usually liquids to facilitate handling. Motor gasoline, jet, and diesel fuels are mixtures of different chemicals of various shapes and sizes. These sizes affect the range at which they boil. In general, the higher the boiling point the longer the length of the carbon chains. The carbon number and boiling point for motor gasoline, jet, and diesel fuels are shown in Figure 2-2. It can be seen that motor gasoline is the lightest liquid transportation fuel, followed by jet, and then diesel. Gasoline is in the C_4 to C_{12} range. Jet is the next heaviest from C_9 to C_{16}. Diesel fuel is in the range of C_9 to C_{24}.

Although turbine engines are fuel omnivores, jet fuel formulation is very important

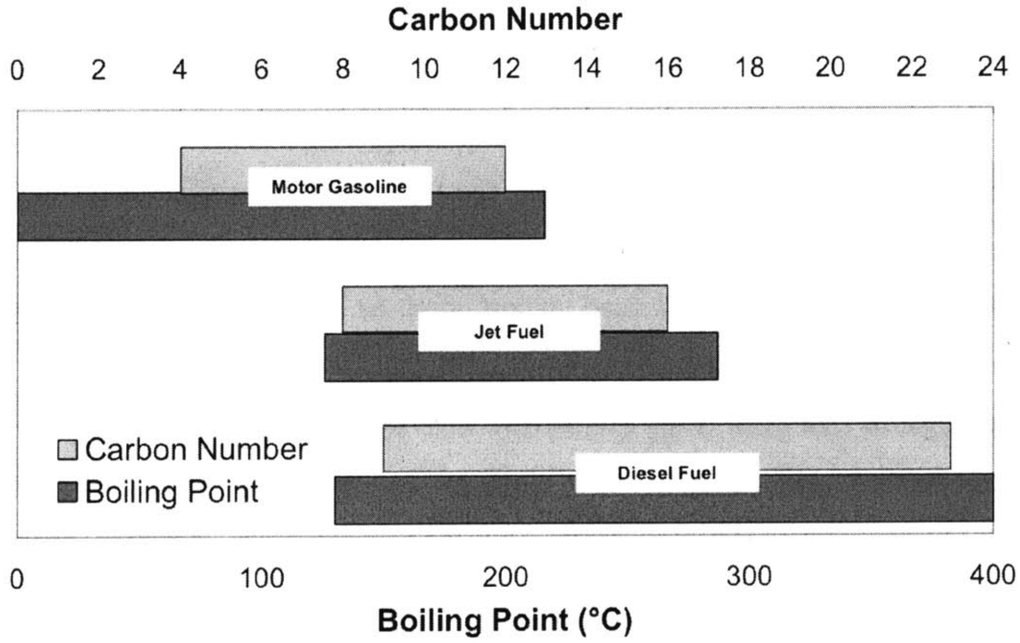


Figure 2-2: The carbon number and boiling point for motor gasoline, jet, and diesel fuel. Source: [24]

for safety and performance reasons. For example, if the fuel was blended with too much gasoline range molecules it would volatilize and evaporate at cruise altitude because of the reduced pressure. Similarly, fuel will gel at the low temperatures of cruise altitude if it includes too many molecules from the diesel range. In both cases, the fuel will not get delivered to the turbine engine causing performance issues, and possibly a catastrophic safety situation.

Diesel fuel includes a wider range of boiling points and includes all of the jet fuel range[1] This means that in operational conditions diesel engines can burn jet fuel without modification, but not the other way around. As a result the United States military procures jet fuel as the single strategic battle fuel to simplify logistics for both aircraft, helicopters, and non-aviation equipment such as tanks and humvees [42].

In addition to molecular weight and boiling point, there are many other specifications that a finished fuel must meet to ensure safe operation. These are described

---

[1] *N.B.* It can be assumed that diesel cuts will always include a portion jet fuel. The amount of jet in a diesel cut depends on the feedstocks and refining process.

Property		Diesel		Jet	
		Fossil	Renewable[1]	Fossil	Renewable[2]
Oxygen content	%	0	0	0	0
Specific gravity	-	0.84	0.78	0.75 to 0.84	0.73 to 0.77
Cloud point	^{o}C	-5	<10	>40	>40
Cetane	-	40-52	70-90	-	-
Sulphur	ppm	<10	<2	<3000	<15
Specific energy	MJ/kg	43	44	>42.8	44.1 (typical)
Aromatics	vol-%	<12	0	<25	<0.5

Table 2.1: Properties and Specifications of fossil and renewable diesel and Jet-A fuels. [1]Properties of renewable diesel from UOP Green Diesel. [2]ASTM D7566 Annex 1 used for hydroprocessed renewable oil specification. Sources: [88, 4, 2, 3]

in petroleum manufacturing handbooks [63] and in ASTM fuel specifications [4, 2].

2.3 Renewable Diesel and Jet Fuels from the Hydroprocessed Renewable Oil Process

Hydroprocessed renewable oils (HROs) are a "drop-in" quality biofuel. This means the fuels are chemical equivalents and are compatible with existing production, storage, distribution, and combustion infrastructure. The performance properties are equivalent to conventional petroleum fuels, but with the added benefit of potentially lower greenhouse gas emissions if renewable feedstocks are used [85, 20, 59, 5, 88, 33, 71, 47].

HRO fuels produced from vegetable oils and animal fats have high cetane values, low aromatic content, and are naturally low in sulfur compounds. A comparison of fossil and renewable fuels is presented in Table 2.1.

Several companies are already producing HRO diesel fuel at commercial scales. For example, Neste Oil has three facilities for the European market with a version of the process known as NExBTL [71, 58, 57, 56]. In the United States, Syntroleum has partnered with Tyson Foods and licensed their Bio-Synfining hydroprocessing technology to Dynamic Fuels for a plant in Geismar, Louisiana [16, 68]. In addition, Honeywell-UOP has licensed their Ecofining technology to Diamond Green Diesel, a Valero and Darling International joint venture, which received a conditional $230 million loan guarantee from the United States. Department of Energy in 2010 [90].

2.3.1 Process Chemistry

HRO fuel is produced in two steps. The first step uses hydrogen gas and catalyst to saturate double bonds, cleave the propane backbone, and remove oxygen from a feed of oils and fats. The second processing step, known as isomerization and cracking, rearranges and reduces the molecular chain lengths to improve cold weather performance.

The first step of the process involves a set of chemical reactions, as shown in Figure 2-3. The first step reacts unsaturated bonds in the triglyceride with hydrogen over a catalyst. Next, the propane backbone is cleaved from the molecule, leaving three long fatty acid chains. Finally, the oxygen in the fatty acid molecules are removed. This last step occurs via two pathways. One reaction removes oxygen in the form of H_2O and is called hydrodeoxygenation. The other pathway removes oxygen in the form of CO_2 and is known as decarboxylation. The hydrodeoxygenation reaction requires nine more moles of hydrogen gas than decarboxylation.

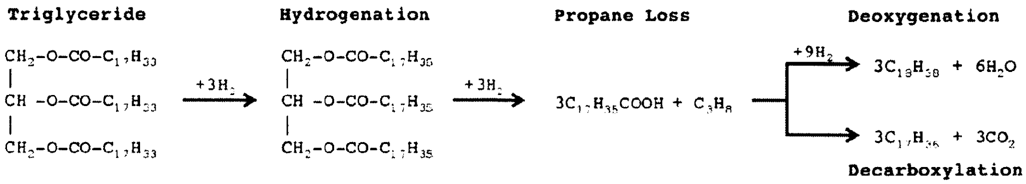


Figure 2-3: Renewable oil deoxygenation reaction pathways.

The selectivity of the process catalyst maybe tuned to favor one reaction route over another [68, 39]. Honeywell-UOP has opted for the decarboxylation route in order to reduce capital and operating expenses resulting from extra hydrogen production and circulation in the process [87, 39]. Whereas, Syntroleum prefers the longer carbon chains that result from the deoxygenation reaction [68]. This work assumes a decarboxylation mechanism in the process model to minimize hydrogen requirements.

At this point, the renewable oil has been converted from an unsaturated triglyceride to a fully saturated hydrocarbon. The resulting product could be blended in small quantities with a fossil diesel stream, but it would not meet the cloud point specification of a finished fuel without further processing.

The second step takes the straight carbon chains, and rearranges them into branched structures as seen in Figure 2-4. Branching a molecule reduces the freeze point relative to the straight chain configuration. For example, unbranched dodecane (n-$C_{12}H_{26}$) has a freezing point of -23 °C. It is isomerized into two different configurations. The first molecule, 2,2-dimethyldecane, has a 10 carbon backbone chain and two branches, with a freeze point 22 °C below the unbranched molecule. The second isomer has an eight carbon backbone and four branches, with a freeze point 53 °C below the normal molecule [37].

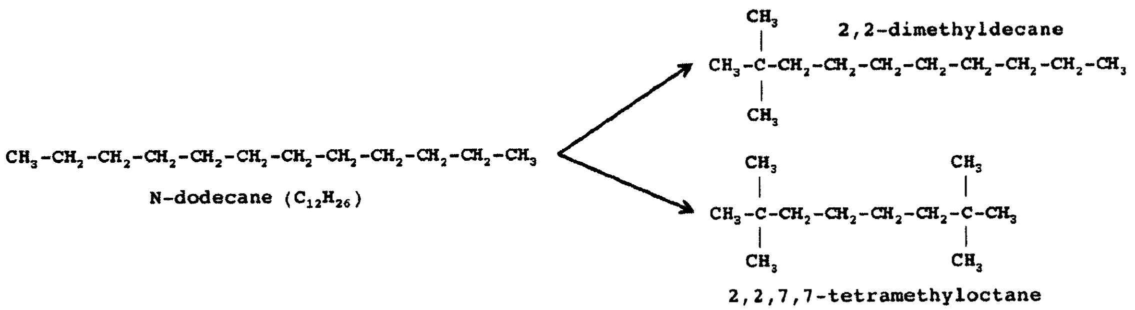


Figure 2-4: Example of an isomerization and cracking reactions for dodecane (n-$C_{12}H_{26}$)

Sometimes the rearrangement mechanism does not recombine, and results in molecular cracking [81, 10]. Cracking means the chain length is reduced and two molecules are produced. For example, n-dodecane may crack from a straight 12 carbon chain into a four carbon chain and an eight carbon chain. This is shown in Figure 2-5. The chemistry of isomerization and cracking reactions are discussed in detail in the literature [81, 10, 31, 33, 47, 72, 71, 59] and in Section 3.1.3 on page 34.

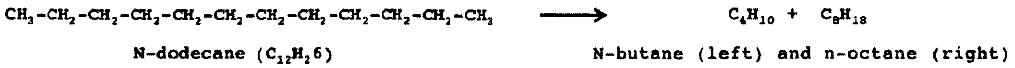


Figure 2-5: Example of a cracking reaction for dodecane ($C_{12}H_{26}$)

2.3.2 Renewable Oils

There are several sources for renewable oil including pure vegetable oils, recycled products, animal fats, and pyrolysis oils. Table 2.2 presents the availability of selected sources in the United States [87]. The table shows that 6.6% of the United States diesel demand could be met if all the currently available fats and greases in the US were used for HRO diesel production.

Source	Supply [BPD]
Vegetable oils[1]	194,000
Recycled products[2]	51,700
Animal fats[3]	71,000
Pyrolysis oil[4]	1,500
Total	318,200
Potential HRD production[5]	270,470
Percent of diesel demand[6]	6.6%
2010 Biodiesel production	42,000

Table 2.2: Current renewable oil and fat resources in the United States, shown in BPD. Notes: [1]Soybeans, corn, canola, palm. [2]Yellow, brown, and trap greases. [3]Tallow, lard, fish oil. [4]Wood slash waste, and biomass wastes. [5]Assuming 85%wt conversion efficiency. [6]Assuming 4.1x10^6 bbl/year of diesel consumption in the US. Source: [87]

Renewable oils contain three long carbon chains connected with an acid group to a glycerine backbone. The carbon chains are of various lengths and saturation (double bonds). The products of a HRO facility are primarily determined by the chain-length of the feedstock oil used. The carbon chain lengths for various vegetable and algae oils are reported in Table 2.3.

In general, most natural vegetable oils, animal fats, and algae oils are in the diesel range, namely C_{16} to C_{22}. These molecules can be processed into smaller chains for jet and motor gasoline range fuels, but this extra step creates by-products amd reduces the overall volumetric middle distillate yield. For example, imagine snapping the ends off of a spaghetti noodle to make it fit length-wise into a container. The spaghetti ends are the by-products of the catalytic cracking reaction. If the process starts with smaller chain length oils, then there is less by-product created and the initial product volume is preserved. As shown in Table 2.3, shorter chain oils exist that are suitable for jet fuel production.

C[1]	Soybean	Palm	Palm Kernel	Canola[2]	Jatropha[3]	Cyanophyta[5]	Tallow
8	-	-	2	-	-	-	-
10	-	-	7	-	-	27-55	-
12	-	-	47	-	-	-	-
14	-	-	14	-	-	7	4
16	11	44	9	4	12	15-24	28
18	87.6	56	21	62	87	6-17	67
20	1.4	-	-	34	1	-	1

Table 2.3: Renewable oil chain length profiles for selected renewable oils. Differences between single and double bonds are not noted. Notes: [1]Carbon chain length. [2]Brassica campestris. [3]Jatropha curcas. [4]Salicornia bigevolii. [5]Trichodesmium erythraeum. Sources: [85, 30, 52].

Approximately half of the carbon chains in palm kernel oil and the cyanophyta organism are suitable for jet fuel production. However these oils are not currently as plentiful or inexpensive as large commodity vegetable oils such as soybean and palm oils [19]. Work on algal derived oil is being investigated by many researchers as a future source of renewable, inexpensive oil that would be suitable for jet and diesel production [82, 51, 77]. The cost of oil and fat feedstock is discussed in the economic model and results Chapters 4 and 6, respectively.

2.4 Naphtha and Gasoline

In the process of producing jet fuel from diesel, naphtha range co-products are produced. Determining options for how to upgrade these low-value co-products is a determining factor in the economics of HRO diesel and jet fuel production. There are three basic options to explore for naphtha use: (1) upgrading to high-octane gasoline on-site; (2) use as process fuel or hydrogen production feed; (3) sold for upgrading or blending into gasoline elsewhere.

2.4.1 Background

Naphtha consists of C_4 to C_{10} hydrocarbons. Naphtha is used as a feedstock for plastics, synthetic fibers, pesticides, insecticides, and solvents manufacture; it is also upgraded and blended to formulate gasoline. Commercial processes to upgrade light naphtha to improve octane ratings may include isomerization, alkylation, and polymerization to increase carbon number and branching; as well as ring formation and dehydrogenation to create double bonds for use as feedstock for polymerization processes. The choice among these or other alternatives depends the location and unit size, and cannot be generalized [17].

2.4.2 On-site Naphtha Upgrade

Naphtha can be upgraded to high-octane gasoline by branching and by creating double bonds or rings. For example, the unleaded research octane number (RON) of normal pentane is increased from 61.7 to 92.3 when it is converted to iso-pentane [63]. Yet, the additional capital and operating expenses for these processes may not be justified at the small scales of HRO bio-refineries. For example, the typical size of a reforming unit is between 3,000 and 20,000 barrels per day (BPD), whereas a 6,500 BPD HRO facility would only produce 470 BPD of naphtha. The capital cost of a 470 BPD naphtha reformer would be approximately $7.6 million [2]. That equates to $11,000 per BPD of naphtha reforming capacity, compared to less than $2,000/BPD for a typical reactor at a large petroleum refinery [17]. In other words, it would cost more than five times as much to upgrade the naphtha to gasoline on-site at a small facility as it would to do so at a larger refinery.

---

[2] Based on a Lang exponent of 0.6 and a $18 million, 3,000 BPD naphtha reforming unit [17].

2.4.3 On-site Naphtha Use for Utilities

A stand-alone greenfield refinery requires process fuel to run utility units, and a hydrocarbon source for conversion to hydrogen gas. Naphtha streams from a stand alone HRO facility could be used as a process fuel or as a feed for the production of hydrogen, instead of natural gas. However, it has a lower carbon to hydrogen ratio, and therefore modifications to reactor and utility equipment would be required to accommodate the larger volumes of naphtha needed to produce the same amount of hydrogen [63]. Additionally, modification of boiler equipment is necessary to burn naphtha, as it is more corrosive than natural gas [63]. The cost of modifying equipment is not available in the literature.

2.4.4 Naphtha Sold for Upgrading or Blending Elsewhere

The naphtha could also be sold as-is to be upgraded or blended at other facilities. Selling the naphtha requires no additional capital or operating costs to the refinery, since it is a by-product of the HRO process and storage is included in the economics model of the plant. Although the actual cost of upgrading naphtha at an off-site refinery depends on the volume and distance transported, a few estimates present a range of possible costs. The estimated additional cost of upgrading might be similar to the cost of other refinery operations, and be on the order of $0.05 to 0.55 per gallon as a rough estimate for processing costs [92]. A second estimate can be derived from historical data from the Organization for Economic Co-operation and Development (OECD) showing a gasoline premium of approximately $10 per barrel, or $0.24 per gallon over naphtha [61]. However, this price is not just the cost of upgrading, but also includes the value-add of higher octane for use as a transportation fuel in automobile engines. A third estimate would build up the cost with assumptions of $0.08 per gallon gal to transport to the refinery, $0.04 per gallon for capital recovery, and $0.05 per gallon for blending with the gasoline pool, results in a total cost of $0.32 per gallon.

Naphtha might also be blended without octane upgrading, as a non-oxygenated biofuel additive for gasoline. According to personal communications with Syntroleum, the firm that processed Solazyme algal oil in 2009, naphtha co-product from HRO processing was blended at 5-10% volumes without degrading the octane rating of the gasoline pool [68]. Since ethanol and other gasoline additives are blended in 10% volumes, this operation is attractive because of its capital efficiency and it was assumed for this analysis.

2.5 FAME: a first generation alternative middle distillate fuel

Biodiesel is the common name for fatty acid methyl esters (FAME). Biodiesel is made by the base catalyzed transesterification of triglycerides with methanol. The reaction is shown in Figure 2-6. Notice the oxygen is left on the FAME molecules after reaction.

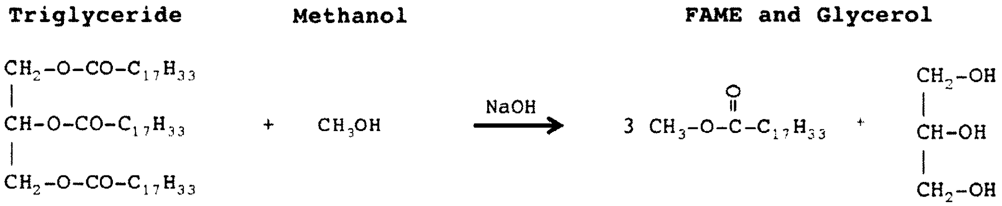


Figure 2-6: The biodiesel reaction.

There are some advantages to biodiesel relative to conventional diesel fuel, such as renewable energy production and supply diversification, increased engine lubricity, and the potential for greenhouse gas emission reductions[3] [33]. There is infrastructure and awareness for FAME processing in the United States because of the relatively simple, low temperature and pressure process that can be scaled from home garage units to industrial facilities that can process several thousand barrels continuously.

According to the National Biodiesel Board, the trade group for the FAME industry, there are between 120 and 170 FAME production facilities in the United States, with combined production capacity of three billion gallons per year [55]. However,

---

[3] Lifecycle greenhouse gas depend on the feedstock and processing [85].

little of the production capacity is currently being realized. Estimates for total 2010 production are around 10% or 300 million gallons per year [7]. This is approximately 1/5 of the production possible with currently available feedstocks as was shown in Table 2.2. Underutilization is primarily due to the expiration of the Blender's Credit tax subsidy and the relatively high cost of feedstock vegetable oil [66]. These shortcomings are discussed in more detail in Chapter 7.

FAME has different chemical properties then petroleum hydrocarbons, which results in infrastructure incompatibilities, decreased energy content and cold weather performance issues [52, 46, 26, 18]. The most significant chemical difference between biodiesel and petroleum diesel are the oxygen atoms in the FAME molecules. Oxygen makes the molecules polar and hydrophylic, (i.e, water loving). This means that biodiesel will mix with water it comes in contact with and become contaminated. Water contamination is problematic since petroleum pipelines have water pools that normally do not mix with petroleum fuel plugs[4]. Another big issue is jet fuel contamination with FAME from biodiesel blends that sticks to the pipeline walls [24]. As a result, biodiesel must necessarily be transported by tanker truck, which is both more expensive and may result in higher lifecycle greenhouse gas emissions. Oxygen molecules also add extra mass, which decreases energy content and causes freeze point elevation compared to petroleum diesel. The FAME properties are shown in Table 2.4 and compared to petroleum diesel.

Property		Biodiesel	Petroleum Diesel
Oxygen	%	11	0
Density	g/ml	0.883	0.78
Sulfur content	ppm	<10	<15
Heating Value (lower)	MJ/kg	38	44
Cloud Point	^{o}C	-5	-5 to -30
Distillation Range	10-90%	340-355	265-320
Cetane	-	50	42-45

Table 2.4: Typical properties of biodiesel (FAME) and petroleum diesel. Source: [88]

---

[4] This is like salad dressing: oil and water don't mix.

Chapter 3

Hydroprocessing Plant Design

> *"To whom does design address itself: to the greatest number, to the specialist of an enlightened matter, to a privileged social class? Design addresses itself to the need."* -Charles Eames

The hydroprocessed renewable oil (HRO) plant design and modeling efforts are described in this chapter. The plant was modeled in Aspen Plus with process information from the literature [33, 88, 17, 63]. See Appendix A for information about the Aspen model. The model was used to examine two production profiles and two hydrogen gas sources. The design is based on hydrotreating and isomerization technology available from the literature and other standard petrochemical support processes such as storage tanks, hydrogen gas production, cooling water tower, etc.

3.1 The Hydroprocessed Renewable Oil Process

The purpose of the HRO process is to convert vegetable oils and animal fats into liquid transportation fuels that are chemically equivalent to transportation fuels from fossil resources. The process was developed based on the work from [33] and engineering judgement. Because of the similarity with petroleum refining, the additional plant costs, known as balance of plant expenses, were taken from petroleum industry handbooks, such as [17].

In the hydroprocessed renewable oil (HRO) process, vegetable oils and fats are reacted with hydrogen gas and converted to diesel, jet, and motor gasoline fuels, as well as lighter paraffin molecules. This is achieved through the catalytic hydrodeoxygenation and subsequent selective cracking and isomerization of triglycerides as described in Section 2.3.1 on page 2.3.1. This model assumes the use of soybean oil as the triglyceride source.

An overview of the process is presented in Figure 3-1. Vegetable oil is taken from feed storage and fed into a hydrotreator with hydrogen gas. The effluent is cooled by steam generation, and sent to an isomerization unit. The isomerized product is then cooled with cooling water before being sent to a separation tower where gasses, including mixed paraffin gases, carbon dioxide, and excess hydrogen, are separated from the liquid products. The paraffin gases and hydrogen are separated from carbon dioxide and recycled to the hydrotreator. Liquid products are separated into liquified natural gases (LNG), naphtha, jet, and diesel streams and then sent to product storage tank farms. Wastewater is separated from the product stream and sent to treatment units.

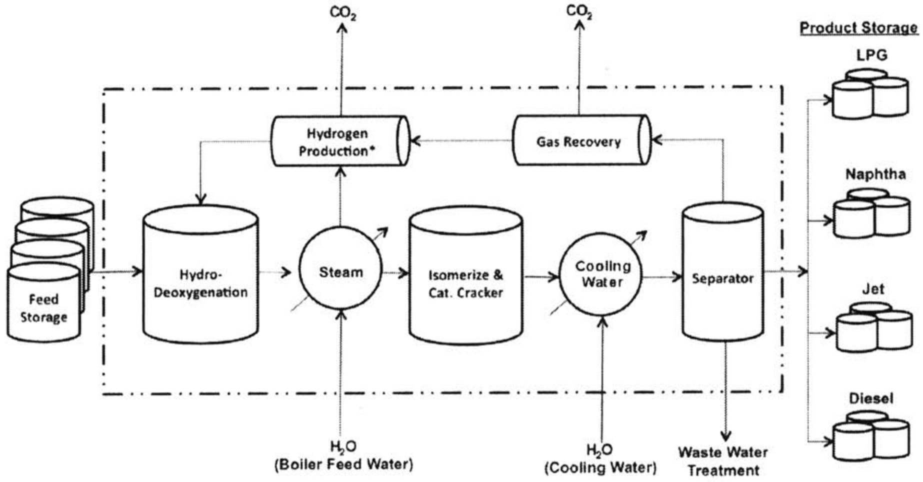


Figure 3-1: Simplified Hydroprocessed Renewable Oil (HRO) system design.

The plant was modeled as eight unit processes, which are described in the following sections: (1) Vegetable Oil Feedstock Storage, (2) Hydro-Deoxygenation, (3) Selective Isomerization and Catalytic Cracking, (4) Heat Integration for Steam Generation and

Cooling Water, (5) Fuel Gas Cleanup and Recycle, (6) Hydrogen Gas Production, (7) Product Separation, and (8) Product Storage and Blending.

3.1.1 Feedstock and Feed Storage

The plant will have 13 days of liquid storage at a cost of $50/bbl in 2005 United States Gulf Coast (USGC) dollars [17]. The capital expense is escalated to 2010 prices in the economic analysis section. It is assumed that refined, deodorized, and bleached oils are purchased from oil suppliers. This is a common way to purchase oils and fats, and therefore, feed pretreatment is not included in this analysis. Although soybean oil is used for this analysis, mixed feed storage and processing is not a problem since the process can accommodate a mixture of feedstock oils and fats [87]. However, depending on the acidity and nature of the feeds, metallurgical considerations might need to be considered.

3.1.2 Hydro-Deoxygenation

Hydrotreatment of the feed removes oxygen, saturates double bonds, and cleaves the propane backbone of triglycerides by reaction with hydrogen in the presence of catalyst. The vegetable oil and hydrogen feed ratios were taken from the literature [5, 33, 87, 88]. The hydrogen to vegetable oil ratio is 2.7% for the maximum distillate profile, and 4.0% for the maximum jet production scenario. The products of the deoxygenation reaction are water, carbon dioxide, and propane, and a range of straight chain alkanes. Water and carbon dioxide are produced when hydrogen reacts with the oxygen atoms and either decarboxylates or hydrodeoxygenates the triglycerides. Propane is produced when the glycerin backbone of the triglyceride is removed [33]. The main products from this reaction are a range of straight chain alkanes covering the diesel and jet fuel carbon lengths from C_9 through C_{20}, with the exact distribution being feedstock dependent. When soybean oil is used, the resulting fuel stream can be blended with fossil based diesel fuel in appropriate quantities, but the neat (unblended) fuel stream would not meet cloud point requirements. Additional isomerization processing is necessary to create a finished fuel that meets ASTM specifications.

3.1.3 Selective Isomerization and Catalytic Cracking

The cloud point of the deoxygenated product is improved by isomerization and chain length reduction. The products are liquid middle distillates in the jet and diesel range, with naphtha, and liquified natural gases (LNG) as co-products. Significant effort was spent attempting to model the product profile and distribution using models from the literature [34, 41, 102, 103, 10, 79, 88, 94, 96, 60]. For example, an exhaustive list of isomers from C_4 to C_{20} was created using SMILES notation, imported into the component list of Aspen Plus, and then run in an equilibrium reactor model using published reactor conditions to determine product distributions based on thermodynamic equilibrium. However, empirical results are required since equilibrium reactor models in Aspen Plus do not accurately model catalytic processes.

Additional attempts to model with existing literature and petroleum handbooks were also unsuccessful. Although, the exact process conditions and product distributions are not available for analysis, lumped product yields were obtained from the literature [5, 33, 87, 59]. Detailed product yields have been reported for diesel products from soybean oil in the literature [33], and jet fuel yields from jatropha oil are inferred from various reports [5, 33, 87, 89, 40]. Since soybean and jatropha oils are similar in chain length compositions (see Table 2.3), these reports provide enough information to calculate product yields without having to model the isomerization process explicitly.

3.1.4 Heat Integration for Steam Generation and Cooling Water

The hydrotreating and isomerization processes are exothermic and must be controlled by removing heat between stages. The modeling of heat integration is important because of water use and associated greenhouse gas emissions associated with the electricity used to run pumps and fans for cooling. A previously reported model used cooling water to remove heat from the reactor effluent stream [33]. This is a simple, but water intensive design. An alternative method generates steam from the process streams. A notional system was modeled in Aspen Plus to compare the steam and cooling water requirements. This water model is described in more detail in Chapter 5. Additional cooling water demands for other process requirements were calculated from process equipment handbooks. The cooling water tower was sized based on requirements from the literature and model results. A 15% water contingency was included in the sizing and cost estimates.

3.1.5 Fuel Gas Cleanup and Recycle

After the product stream leaves the hydrotreating and isomerization reactors, the gas products are separated from liquid products and purified. Hydrogen is separated by pressure swing absorption (PSA) and recycled back to the deoxygenation reactor. Other light gases such as methane, ethane, and propane are used as process fuel in the model. Amine scrubbing and acid gas processing are utilized for gas cleanup. Pure streams of carbon dioxide are recovered from the PSA unit and available for possible sequestration, though this is not included in the scope of the analysis. Utility estimates were taken from an industrial handbook [17] and modeled as black-box separators in Aspen Plus.

3.1.6 Hydrogen Gas Production

A hydrogen plant is needed to produce the gas volumes required in this analysis, therefore bulk hydrogen delivery was not considered. Steam methane reformation (SMR) was modeled for hydrogen production in this system. Capital and operating costs, and utility demands were taken from the literature for modeling purposes [78, 28, 21, 17]. Detailed models were created in Aspen Plus, and the utility demands for electricity, cooling water, and boiler feed water (BFW) agree with literature values [28, 76, 29, 49].

3.1.7 Product Separation

The liquid product streams are separated by boiling point in an atmospheric distillation column. The capital, operating, and utility costs associated with product separation were included in the hydrotreator and isomerization units [17]. Once separated, the products are sent to the product storage tank farm.

3.1.8 Product Storage and Blending

After separation, diesel, jet, naphtha, and liquid natural gases (LNG) products were sent to product storage. Twenty-five days of product storage are assumed for the product tank farm to cover shutdown and other production anomalies [17]. In some instances, off-specification products may be "blended off" in sufficiently small quantities to avoid large economic losses for waste and disposal. Having sufficient storage is required for these types of blending operations.

3.2 Offsites, Special Costs, and Other aspects not modeled explicitly

Other facility systems were not modeled explicitly, but instead they were included in offsites and special costs for inclusion in the economic evaluation. Offsites (or outside-battery limits, OSBL) include: electric power distribution, fuel gas facilities, water supply and treatment, plant air systems, fire protection, flare, drain and waste containment, as well as plant communication, roads and walks, railroads, fences, buildings, vehicles, product and additive blending facilities, and product loading facilities. Special costs include: land, spare parts, inspection, project management, chemicals, miscellaneous supplies, and office and laboratory furniture. Sulfur recovery was not included because of insufficient information on the sulfur requirements of the catalyst.

3.3 Product Profiles

Soybean oil was used as a surrogate for all oils in the process model. Since this feedstock is predominately a C_{18} oil, the products of the reaction will mostly be diesel fuel. However, a producer could choose to produce more jet fuel by cracking the diesel down to the jet range. Two product profiles were modeled to compare this choice: (1) maximum distillate production, and (2) maximum jet production.

The maximum distillate profile meets diesel specifications, and minimizes LNG and naphtha co-products. The option to separate the jet fuel fraction from the distillate product stream would be available, and is considered herein. UOP reports that the jet fraction is approximately 15% by volume [88]. The maximum jet profile produces jet fuel by catalytically cracking diesel range molecules. In theory, jet range fuel and naphtha could be created by converting $C_{18} \rightarrow C_{10} + C_8$, with no additional by-products. However, in reality the selectivity of the cracking reaction is difficult to control and the products range in size from C_3 through C_{15}. Since the higher molecular weight products have more economic value, not all of the diesel fuel is cracked in the maximum jet scenario even though it is technically possible. The product quantities are calculated from reported literature values [5, 33, 87, 14], and will be used in subsequent calculations. Actual product distributions and hydrogen consumption depend on the feedstock being used in the process, and product profiles and hydrogen consumption will change over the lifetime of the process and vary from start-of run to end-of-run. However, only soybean oil and a consistent product profile were assumed in this work. Table 3.1 summarizes the product profiles for both scenarios.

3.4 Hydrogen Production and Purchase

Two additional scenarios were considered to understand hydrogen production: (1) an on-site SMR hydrogen production scenario, and (2) an over-the-fence hydrogen purchase scenario. In the first scenario, an on-site steam-methane-reformation (SMR)

Product Profiles [wt%]	Maximum Distillate	Maximum Jet
Vegetable Oil	100.0	100.0
Hydrogen	2.7	4.0
Total In	102.7	104.0
Water	8.7	8.7
Carbon Dioxide	5.5	5.4
Propane	4.2	4.2
LPG	1.6	6.0
Naphtha	1.8	7.0
Jet	12.8	49.4
Diesel	68.1	23.3
Total Out	102.7	104.0

Table 3.1: Mass-based product yields by product profile. The product yields for each product profile are both based on 100 pounds of soybean vegetable oil feed. Quantities are based on material balances provided in the literature for a decarboxylation reaction, such as the UOP process. Sources: [5, 33, 87, 14]

facility is included in the material and energy balances of the model. The second scenario assumes an industrial gas supplier, such as PraxAir, Linde Gas, Topsoe, or Air Products, could erect, operate, and supply an over the fence, (a.k.a., sale of gas), hydrogen supply [64]. These scenarios are identical in the assumptions and processing equipment except for the capital and operating expenses associated with a SMR hydrogen production facility.

3.5 Process Utilities

The utility requirements for the process units were obtained from petroleum handbooks, literature reports on the process, and the modified Aspen Simulation [29, 63, 89, 59, 87, 33, 17]. These include boiler feed water, steam, cooling water, electric power, and natural gas. The utility requirements were normalized on a per pound of vegetable oil basis. Table 3.2 summarizes the utility requirements for each process unit and the total utility demand of the system.

Boiler feed water (BFW) is used to generate steam. BFW undergoes a polishing process that removes dissolved mineral content to prevent scaling in the boiler

Process Unit per lb feed	BFW [lb/hr]	Cooling Water [lb/hr]	Steam [lb/hr]	Power [kW]	Natural Gas [lb/hr]
Hydrotreator	0.25[1]	-	(0.25)	0.01	0.02
Isomerization	-	2.55	-	0.00	0.03
Gas Processing Unit	-	5.26	-	0.01	0.00
Subtotal	0.25	5.33	(0.25)	0.02	0.06
Hydrogen SMR	0.5	0.25	-	0.00	0.06
Addition Max. Jet	-	-	-	-	0.06

Table 3.2: Process utilities requirements per pound of vegetable oil. The process utility requirements are the same for both product profiles. Notes: [1]Results of Aspen Plus simulation.

systems. BFW fed to a process unit comes out as process steam, and sometimes the condensate is recovered, other times the water vapor is lost in gas venting downstream. The SMR unit uses BFW to generate high pressure steam to convert natural gas into hydrogen. Process steam is used for heating process streams to increase the rate of reaction, or for heating the reboiler section of a distillation unit. Steam is normally generated at a steam island, such as an SMR, and distributed around a facility. However, the HRO process generates steam by heat exchange with the high product temperature stream leaving the hydrotreator. No attempt was made to integrate the steam from the hydrotreator for use at the SMR because the engineering required would be beyond the scope of this work.

Cooling water is used at a refinery to reduce stream temperatures. This is important in controlling reactions that produce heat, and preventing temperatures and pressure from exceeding safety limits. Cooling streams will also sometimes be used to condense products from the gas to the liquid phase for easier handing and storage of products. As an example, the fuel products leaving the isomerization unit are in the gas phase and cooling water is used to condense the naphtha, jet, and diesel range molecules to the liquid products. This also allows the non-condensable gases, such as hydrogen, carbon dioxide, and propane to be separated easily from the liquid fuel products. The gas processing unit uses cooling water to facilitate the separation of methane, ethane, and propane, from water and other impurities to produce a dry gas suitable for use as a fuel.

Electricity is used to power pumps and compressors as well as other electrical controls around the refinery. Natural gas is used as process fuel in various units around the refinery. Process fuel can be burned to produce heat in direct fired heaters, such as in the boiler of an SMR unit.

Chapter 4

Economic Modeling

"There are three rules to a successful business. Rule #1, if it don't make money the other two rules don't matter." - T. Boone Pickens

4.1 Introduction

This chapter introduces the economic model used to estimate the price of distillate fuel under various economic conditions. It begins by describing the capital (CapEx) and operational (OpEx) expenses, and gross income. It then provides a brief discussion for evaluating the economics of a project with return on investment (ROI), and discounted cash flow rate of return (DCFROR) models.

The gate cost of diesel and jet fuel were found for three plant sizes, two product profiles, two hydrogen sources, two project finance structures, and two production ramp-up schedules. Additionally, separate scenarios to determine the maximum cost of vegetable oil for profitable operation for fixed diesel and jet off-take prices were conducted.

4.2 Capital and Operating Expenses

The hydrotreating facility is analyzed as an Nth plant, not a pioneer plant, with the assumption that it will be built from traditional and well established petrochemical

Capital Expenses
Inside Battery Limits
Hydrotreator
Isomerizer
Hydrogen island
Saturated gas plant
Outside Battery Limits
Storage, feed
Storage, liquid products
Storage, gas products
Cooling water tower
Offsites, greenfield
Special costs
Contingency
Escalation
Location factor

Table 4.1: List of capital equipment and other project costs.

plant designs and equipment. Pioneer plants have economic penalties in capital and operating expenses due to unforeseen complications with scale-up. Nth plants have the "kinks worked out," resulting in smoother engineering, procurement, construction and optimized operation. These cost reductions are known in the economic literature as learning by doing effects [62]. The capital and operating expenses of the process are described in 4.2.1 and 4.2.2. These expenses were estimated by following the cost-curve method from Gary, Handwerk, and Kaiser [17].

4.2.1 Capital Expenses

The installed capital cost for each unit process was obtained from cost-curves provided in [17]. This data was used to estimate capital costs as a function of unit size. Table 4.1 summarizes the process equipment and other capital expenses included in the total project investment (TPI). The results are presented in Section 6.2.1.

The inside battery limits (ISBL) include engineered equipment expenses such as the cost of purchasing and installing process units and supporting processes. These costs are estimated from cost curves in [17]. Outside batter limits (OSBL) include storage and basic process utilities. These costs are estimated from heuristics in [17]. Additional capital costs include: offsites, which are supplementary infrastructure items like roads and fences; and special costs, which include land, project management, and office and lab furniture. These were both estimated as a percentage of the engineered equipment costs. A detailed list of what was included in offsites and special costs can be found in Section 3.2 on page 36.

Estimates from this method are assumed to be accurate to +/-25% of the actual project cost [17]. In addition, a contingency of 15% is added to the subtotal of the ISBL, OSBL, offsites, and special costs. This is standard practice in cost-curve estimations to account for the "lack of complete definition of facilities required" [17]. The escalation and location factor are used to adjust for inflation and regional differences in capital expenditures. The Chemical Engineering Progress Cost Index (CEPCI) was used for cost escalation. All prices in [17] are based on 2005 with construction in the United States Gulf Coast area and were then scaled to 2010 prices using the CEPCI. No location factor was specified in the model for this analysis.

4.2.2 Operating Expenses

Operating expenses occur annually, and are characterized as either fixed or variable. Fixed operating expenses are constant expenses and independent of production levels. Variable expenses are not constant and are proportional to the level of production. Each type of expense is explained in the following section.

Fixed Operating Expenses

Fixed operating expenses recur every year regardless of the amount of finished fuel the refinery produces. These expenses include insurance, taxes, maintenance, and plant staff salaries. For example, insurance premiums and staff salaries need to be paid regardless of the production level of the plant. Fixed operating expenses were estimated from capital expenses based on heuristics in the literature and personal

Fixed Operating Expenses	
Catalyst	$/lb feed
Insurance	0.5% Total Plant Investment
Local Taxes	1.0% Total Plant Investment
Maintenance	5.5% Total Plant Investment
Miscellaneous Supplies	0.2% Total Plant Investment
Plant staff and operators	12 staff @ $72k/yr
Contingency	10% of above subtotal

Table 4.2: Fixed operating expenses based on the total project investment (TPI) and heuristics. Source: [17].

interviews with industry experts familiar with the process [17, 64, 68]. A 10% contingency is included on top of these estimates in a manner similar to the capital expenses to account for uncertainty. Table 4.2 summarizes the fixed operating expenses and the model results are presented in Section 6.2.2.

Catalysts are used to lower the energy requirements or increase the rates of specific chemical processes. Catalysts have a finite lifetime and need to be replaced or regenerated periodically. Catalyst expenses were estimated from [17] for standard hydrotreators. According to [17] insurance and local taxes account for 0.5% and 1% of the TPI respectively. Maintenance can vary between three and eight percent, in this case a value of 5.5% of the TPI was used, and includes costs for materials and labor. Miscellaneous supplies include chemicals used for controlling corrosion, office supplies, and drinking water. These costs are small, and [17] uses an average value of 0.2% for miscellaneous supplies. Plant staff and operators are the personnel required to run the plant. The facility is not as complex as a traditional refinery, and so a reduced number of staff was assumed. Interviews with industry professionals suggested having 12 staff, instead of 32, with an average annual salary of $72,000 was used for the staff [68]. Maintenance personnel were not included in this figure since they were included within the maintenance line item.

Variable Operating Expenses

Variable operating expenses depend on the amount of production at the refinery. For example, if the facility cuts its production rate in half, then the variable operating expenses would be exactly one half as expensive because producing less products requires less material and energy. These expenses include utility costs, such as electricity, natural gas, boiler feed water, and reaction feedstock such as vegetable oil and are described in Section 3.5.

Electric power is used to power pumps and compressors as well as other electrical equipment in the refinery. Natural gas is used for process fuel for heating, as well as for feed gas for certain reactions such as SMR hydrogen production. Water can be procured from surface or ground sources and then subsequently polished on-site, or it can be purchased from a local utility. It was assumed that make-up water is purchased from a utility in this analysis. Soybean oil was assumed to be the primary feedstock for producing liquid transportation fuels. Hydrogen gas may be produced on-site, or purchased from a supplier. Both hydrogen source scenarios are considered separately in the subsequent analysis.

Since utility and feedstock prices change with supply and demand of the commodity market, they add uncertainty into the economic analysis. The price for industrial electric power, process fuel, and hydrogen are dependent on the price of the commodities from which they are made, specifically natural gas. Vegetable oil is also an important variable for production. Vegetable oils, such as soy and canola, are food commodities and are also used as a feedstock in biodiesel production. Waste oils, fats, and greases are also commodities sold for use as lubricants, and caloric additives for animal feed [66]. These commodities have recently seen historically high prices due to inclement weather, increased costs of energy, and political unrest in many parts of the world [66, 67]. Finally, in a more water-constrained world, makeup water could have added costs associated with it compared to today [36, 83, 100, 22].

In general purchase price agreements (PPAs) could be established with utility suppliers to lock in prices and alleviate uncertainty in operational expenses. The

Varaiable Expense	Unit	5yr Avg	20yr Avg	Max	Min
Electric power	(\$/kWh)	0.06	0.05	0.07	0.04
Natural gas	(\$/10^3·lb)	373.18	233.15	461.93	139.65
Makeup water	(\$/10^3·lb)	0.04	0.04	0.04	0.03
Soybean oil	(\$/gal)	2.62	1.58	3.98	0.92
Hydrogen	(\$/lb)	0.66	0.54	0.68	0.41

Table 4.3: Variable operating expenses. Sources: [19, 92, 17, 75, 35, 101, 8, 13]

duration of these PPAs could range from one to 20 years in length. However, in reality, these PPAs would likely be short term and market based due to the high variability in the commodities markets, and high substitutability of the products [67]. For example, a vegetable oil broker would not sell their product at \$0.20 per pound when the market is clearing at \$0.43 per pound. Similarly, if the cost of producing electricity increases because of increased natural gas prices, an electric utility would not want to sell power at a loss because of a PPA. As a result, the utility and feedstock agreements would likely be volume contracts for a fixed period of time, and the price would be determined by a market-based formula [67].

Scenarios were developed to explore how variable operating expenses influence product values and the economic feasibility of a hydroprocessed renewable oil project. Table 4.3 lists the variable expenses used in the baseline scenario, as well as historic ranges. The results from the scenario analyses are presented in Section 6.2.2.

Commodity prices in Table 4.3 were taken from government reporting agencies, commodity market prices, industry sources, and recent techno-economic reports in the literature. Natural gas and electricity prices were estimated from national monthly historic industrial prices given by [92]. Power was assumed to cost \$0.06 per kilowatt-hour, with historic maximum and minimum cost of \$0.07 and \$0.04per kilowatt-hour. The five year average natural gas cost was \$373.18 per thousand pound, with a twenty year range of 461.93 and 139.65 per thousand pound, respectively. The price of makeup water was taken from [17] as \$0.04 per thousand gallon. There was no historical information available for the cost of water, so it was scaled using the consumer price index [75]. Soybean oil cost was calculated as the average annual price in the United States. Soybean oil is traded on the commodity market, and over the last 20 years the price has fluctuated between $0.92 and $3.98 per gallon with a five year historic average of $2.62 per gallon [19]. When hydrogen is purchased in 2010, the cost was assumed to be $0.66 per pound; this agrees with published literature values [35, 8, 13]. Price ranges for industrial hydrogen gas were not available, but were scaled following the consumer price index [75].

4.3 Gross Income: Estimation of Refinery Sales

The gross income from refinery sales depends on the quantity of products for sale, and the price at which the products are sold. These are the prices at the refinery gate, (a.k.a., "gate prices"), and include the cost of production and refiner profit, but do not include the costs for distribution, transportation, retail markup, or taxes. Price supports for renewable fuel production, including the $1 per gallon Blender's Credit, and renewable identification numbers (RINs) are also not included in the gate costs. This section will explain the basis for the prices used for the analysis.

4.3.1 Product Profiles and Material Balance

The hydroprocessing plant converts vegetable oil and hydrogen into a variety of products. Table 3.1 in Section 3.3 presents the material balances for each profile. Some of these products do not have an inherent value, such as produced water and carbon dioxide. The water can be treated and used to balance the water demands of the plant, or treated and disposed. The carbon dioxide is vented to the atmosphere as there are no economical uses for it at this time. Carbon dioxide, and water account for approximately 14% by weight of the products.

The valuable fuel products include propane, LNG, naphtha, jet, and diesel fuel. Propane is created by breaking the three carbon backbone of the triglyceride. It accounts for 4% by weight of products, and is the same in both product profiles. The distribution of the other products available for sale depends on the product slate and processing conditions. The maximum distillate profile will have 80.9% distillate (diesel

Product Values [$/gal]	5yr Avg	20yr Avg	Max	Min
Propane	1.77	1.04	2.22	0.58
LNG	1.77	1.04	2.22	0.58
Naphtha	2.03	1.09	2.59	0.53
Jet	2.12	1.05	3.02	0.45
Diesel	2.13	1.05	2.99	0.44

Table 4.4: Historic annual average refinery product gate prices. Gasoline prices were used as a surrogate for naphtha, and propane as a surrogate for LNG. Source: [92]

and jet fuel combined) fuels for sale, 1.6% LNG, and 1.8% naphtha co-products. The maximum jet profile cracks diesel down into the jet range, and will increase LNG and naphtha to 6.0% and 7.0%. Although it is possible to crack the diesel to extinction and create even more jet fuel, this would increase hydrogen consumption, result in more naphtha and LNG co-products, and an overall reduction in the higher molecular products [68][1].

4.3.2 Products and Historic Gate Prices

Historic gate prices were used to calculate refinery sales. Refinery products were relatively constant in price between 1990 and 2000, but then increased by as much as 300% before peaking in 2008. Figure 4-1 shows annual, national average product gate prices from the EIA [92]. The five year average, 20 year average, maximum and minimum prices are reported in Table 4.4. The five year average price was used as the basis for the gross income.

The data were taken from the US Energy Information Agency [92]. Data was scaled using the consumer price index from the Bureau of Labor Statistics when historic information was not available [75]. The price of propane was used as a surrogate for the price of LPG and gasoline was used as a surrogate for naphtha because costs were not available from public services.

Propane, butane, and pentane are products used for heating applications and sold

---

[1] The exact product slate for a "no-diesel" scenario was not calculated because experimental data was not available.

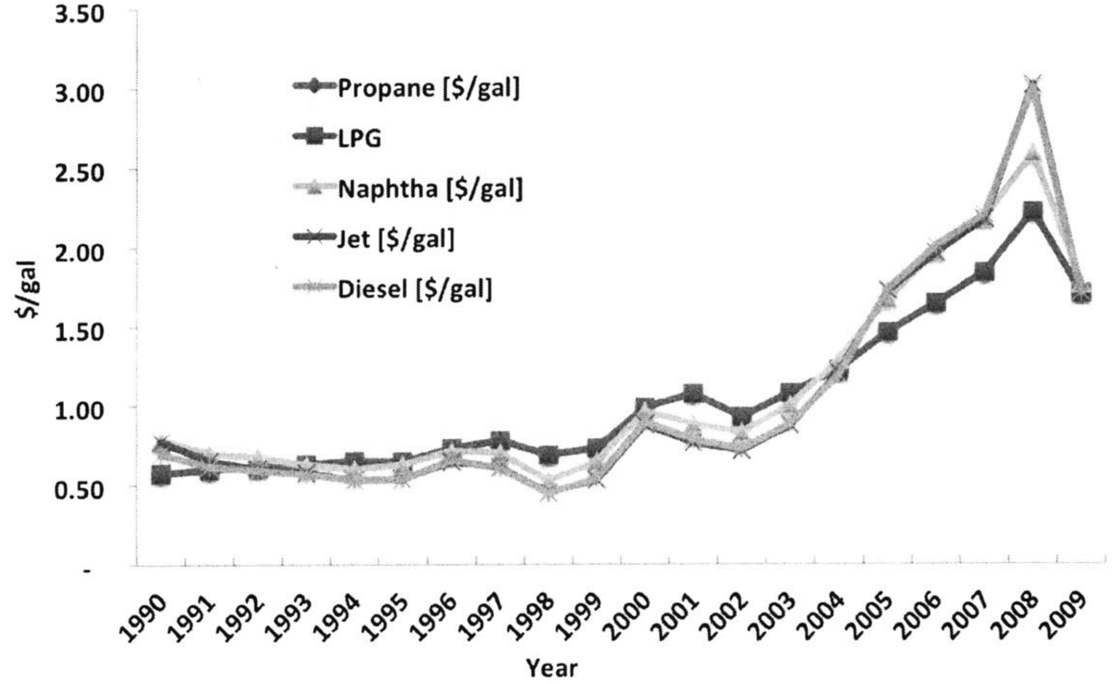


Figure 4-1: Historic annual average refinery product gate prices. Gasoline and propane prices were used as surrogates for naphtha and LNG, respectively. Values reported in \$/gal. Source: [92]

as liquid natural gases (LNG). The five year average for propane is \$1.77 per gallon, with a 20 year high of \$2.22 and low of \$0.58 per gallon. LNG prices were not available on the EIA website, and so propane prices were used as a substitute.

The naphtha product stream consists entirely of straight chain paraffins from the cracking reaction. As such, it has a low octane rating, approximately 50 [87]. Since the EIA does not report prices for naphtha, gasoline for these purposes even though it would likely have a lower price because of the octane penalty[2]. This assumption is almost certainly less influential then the price variability of gasoline. Historically, motor gasoline has annual average cost \$1.09 per gallon over the last twenty years with a historic annual average high of \$2.59 and low of \$0.53 per gallon. The jet and diesel prices were determined in the model, and the five year average prices, \$2.13 per gallon for diesel and \$2.12 per gallon for jet, will be used for comparison. There is a \$0.01 difference over the five year period sampled, but that premium disappears in the 20 year average.

4.3.3 Return On Investment Analysis

Given the input costs and product prices, a return on investment (ROI) analysis was performed to screen economic and operational factors and determine if an investment is profitable. An ROI does not take into account the time-value of money, but does provide a more realistic prediction of cash flows during the later years of the investment as opposed to a method that discounts future earnings. ROI is calculated from the quantities in Table 4.5, using Equation 4.1.

$$ROI = \frac{\text{Net Annual Cash Flow} \times 100}{\text{Total Project Investment} + \text{Working Capital}} \tag{4.1}$$

The capital investment and operating expenses were estimated and discussed in Section 4.2. The total project investment (TPI) was estimated using the cost curve method for a complete hydroprocessing renewable oil facility on page 42. Working capital was assumed to be 5% of the TPI. Working capital includes 30 days worth

---

[2] Although ethanol is also a renewable gasoline additive, it has a higher octane and lower energy content than the naphtha stream and is therefore not a proper substitute for economic analysis.

of product and feed inventory, cash for wages and materials, and spare parts [17]. The operating expenses were estimated from heuristics described on page 43. The total sales were calculated using the five year national average industrial prices for utilities, feedstocks, and products described as listed in Tables 4.4 and 4.3. No start-up penalty, (e.g., 100% capacity utilization), during the first year of operation was assumed. This is not a realistic assumption because the market may have feedstock limitations similar to the biodiesel industry, or the producer may elect for a staggered production ramp-up to verify the system is performing properly [18]. Ramp-up effects are discussed in Section 4.4.5 on page 57. Depreciation was calculated with a 10-year straight-line schedule, and income tax was taken to be 40% of taxable income. These assumptions agree with generally accepted accounting principles (GAAP) [17].

The difference of total sales, operating costs, and income tax, is the net annual cash flow. These terms are described in various economic texts and in [17]. The ROI is then calculated as the quotient of net annual cash flow, and the sum of the total project investment and working capital. In general, a refinery project should have a ROI greater than 15% to be economically attractive. The results of the ROI study are reported in Section 6.3.

In addition to ROI analysis, a more complete economic model was developed to take into account the time value of money and various economic and operating scenarios. These scenarios are described in Section 4.4 and results are reported in Chapter 6.

4.4 Discounted Cash Flow Rate of Return

For this work, a discounted cash flow rate of return (DCFROR) analysis was the preferred method for process evaluation. A DCFROR determines the "true rate of return" by taking into account the time value of money [17]. A DCFROR requires the same cash flows as an ROI, but it includes the net present value (NPV) of those cash flows over the entire economic lifetime of the project. The baseline case for analysis used the assumptions summarized in Table 4.6.

Total project investment
Working capital
Direct operating costs
Variable pperating costs
Total sales
Less depreciation allowance
Taxable income
Income tax
Income after tax
Net annual cash flow
ROI (%/yr)

Table 4.5: Cash flows for ROI calculation.

Equity financing was assumed to be 20% based on current incentives and loan guarantee programs from US Federal government agencies such as the Department of Agriculture and Department of Energy [93, 90, 16]. The loan terms for capital expenses is 10 years with an interest rate of 5.5%. A 15% internal rate of return, with 20% equity, and 2% annual inflation rate was assumed. These rates were taken from the literature [17]. The construction period is estimated to be three years (36 months), which includes engineering, procurement, and construction phases. Construction costs were distributed as 8%, 60%, and 32% of the TPI for each year, respectively. The depreciation is calculated from the IRS Modified Accelerated Cost Recovery System (MACRS). Depreciation is based on a variable declining balance method over 10 years based on IRS Assets Class 13.30 for petroleum refining equipment [12]. This allows for the largest deductions over the shortest recovery period. State tax is included in fixed operating costs as a percentage of the total capital investment even though the plant location is not specified. The cost basis was scaled form 2005 to 2010 using the Chemical Engineering Progress Cost Index (CEPCI) [44]. The plant operates for 8,400 hours per year, which is equivalent to a 96% on-stream capacity. The net present value was calculated in the last year of the construction period (year 0) [17].

A cash flow spreadsheet was created with these assumptions and the input and

Baseline DCFROR Assumptions	**Value (Range) and Units**
Facility Size	(2000-6500) BPD
Total Plant Investment	(calculated) \$
Working Capital (% of TPI)	5%
Equity	20%
Loan Interest	5.5%
Loan Term	10 yrs
Annual Loan Payment	(calculated) \$
Depreciation Period	10 yrs
Construction Period	3 yrs
% Spent in Year -3	8%
% Spent in Year -2	60%
% Spent in Year -1	32%
Internal Rate of Return	15%
Income Tax Rate	40%
Operating Hours per Year	8,400
Cost Year for Analysis	2010
Inflation	2%

Table 4.6: Assumptions in the DCFROR analysis.

product prices described in Sections 4.2.2 and 4.3. With all other costs and prices held constant, the gate cost of diesel was found that sets the net present value (NPV) of the system to zero. This cost is the minimum price needed in order to satisfy the constraints on the system, (i.e., internal rate of return investment, loan payment, etc.).

The base case was then expanded to investigate the effects of the following on the gate cost of fuel: plant size, hydrogen source, equity structure, cost of inputs (commodity market volatility), and production capacity utilization. These sensitivity studies are discussed in Sections 4.4.1 through 4.4.5. Additional case studies and opportunities for future analysis are discussed in Section 8.2 on page 93.

4.4.1 Plant Sizes

The size of a bio-refinery is bound by economies of scale and feedstock availability limitations. Because it determines capital costs, the size of the plant effects the gate price of fuel. Larger plants have economies of scale because capital costs are not linearly proportional to the size of the plant [63, 17]. For example, doubling the size of a plant will only increase capital costs by 1.5 times [17]. Therefore more fuel could be produced for less, and a refiner would want to build the largest plant it can afford.

On the other hand, biomass materials are not as easy to produce and aggregate as crude oil. Biomass must be grown, harvested, processed, and transported. Furthermore, increasing transportation distance of feedstocks increases cost and environmental footprint[69, 82]. As a result, local feedstock availability limits facility size of a renewable oil refinery by typically one or two orders of magnitude smaller than petroleum refineries. Three HRO facilities considered herein; 2,000, 4,000, and 6,500 BPD, which account for the limited local feedstock availabilities and economies of scale [87, 88].

4.4.2 Hydrogen Source

The source of hydrogen was also investigated to determine the effect of economies of scope. Economies of scope are defined as "proportionate savings gained by producing two or more distinct goods, when the cost of doing so is less than that of producing each separately" [48]. In this case, purchasing hydrogen from an industrial supplier, is compared to the cost of on-site production. Economies of scope would occur because of the complimentary nature of the renewable oil process and hydrogen production. For example, the renewable oil process generates steam and natural gas as by-products, which could be used by the hydrogen facility to offset the cost of production. The cost of on-site production was determined by including a steam-methane-reformation (SMR) unit for hydrogen production in the capital and operating costs of the plant. The cost curve and utilities of the system were taken from [17].

4.4.3 Equity Structure

The equity structure of the refinery is how the project is financed. For example, a 100% equity structure means that the project has no financing, and the entire cost of the capital project was paid with cash. Large petroleum companies, or other heavy industry companies have sufficient capital to finance projects without leverage. On the other hand, a 20% equity structure means that the refinery has leveraged 80% of the project with loans from a lender, and must pay the lender the principle plus accrued interest. For a capital project costing many millions of dollars, the interest payment may not trivial. The additional cost of financing on diesel fuel prices was determined by comparing a 100% equity structure with a 20% equity structured project at 5.5% interest.

4.4.4 Input and Output Price Sensitivity

The minimum price for the distillate products depends on the cost of inputs, (e.g., vegetable oil feedstock and natural gas), as well as the market price of co-products.

For example, if the cost of inputs goes up, and the co-product prices stay the same, then the distillate gate cost will have to increase in order to offset input cost increases. Similarly, if the price of co-products increases, the the gate cost could decrease while maintaining the rate of return. The DCFROR model was used historic average input costs and co-product market prices. The computed gate-price was then compared with historic petroleum prices.

4.4.5 Production Capacity: Start-up Penalty and Feedstock Shortages

In addition to input prices, the plant capacity also affects the gate price. The refinery production levels were varied to investigate changes in distillate gate price values. The baseline case assumed 100% overnight utilization. This means that the plant has no start-up penalty. In reality plants would normally start operating at a minimum production level, and then ramp-up to verify systems are functioning properly. It was assumed that the plant can be "turned-down" to 50% of its nameplate capacity without operational consequences and below this, the plant would have to be shut down completely [88].

As part of the sensitivity analysis, two ramp-up scenarios were investigated. The first scenario is a notional start-up period starting with 50% capacity utilization the first year, followed by 75% in the second year, and full capacity in the beginning of the third year. The second case assumes a persistent market-wide shortage of feedstocks, which would force the refinery to operate at 50% capacity over the entire 20-year plant lifetime. The availability of affordable feedstock is of paramount concern because issues with feedstock availability have limited biodiesel production in the past [11, 26, 23, 52, 46, 1, 54, 18]. USDA and DOE are trying to address this with federal programs to incentivize oil crop production [93]. However, current feedstock availability would likely limit production if the plant were to be built overnight. The ramp-up scenarios are summarized in Table 4.4.5.

Scenario	Initial Capacity [% of nameplate capacity]	Ramp-Up Rate [%/yr]
Baseline	100 %	0 %
Start-Up	50 %	25 %
Feedstock Shortage	50 %	0 %

Table 4.7: Production ramp-up schedule.

Chapter 5

Environmental Model

> *"Thank God men cannot fly, and lay waste the sky as well as the earth."* - Henry David Thoreau

5.1 Introduction

This section presents the environmental impacts of hydroprocessed renewable oil (HRO) plants in terms of water use and greenhouse gas emissions (GHG). The water intensity is based on the refining process and the GHG calculations are on a life cycle basis.

5.2 Water Usage Model

Water requirements of a plant were considered because it determines regional feasibility due to variation in fresh water availability. Water usage in industrial processes and two water integration designs for the plant are presented. The results of the model are found in Section 6.5.1 on page 77.

5.2.1 Cooling Water and Steam Systems

Water is used for cooling applications in alternative fuel production. Cooling water systems are either open or closed. Open systems, also known as once-through systems, withdraw water from a source and exchange heat with the system before being returned at an elevated temperature. In a closed-loop system water is circulated around the plant, and heat is removed in a cooling tower or cooling pond [17, 63].

Steam is used to heat process streams or to drive thermal separation processes. For example, steam may be used to pre-heat a feed stream before entering a reactor. Similarly, steam might be used to boil the bottoms section of a distillation column for product purification. High pressure steam is typically generated by boiling water at a central process unit and it is then transported around the plant. To prevent build-up of scale and other fouling materials in the boiler system, sodium bicarbonate and other minerals are removed from the boiler feed water in a process known as polishing [17, 63].

These systems must initially be charged with water; however, only a portion of this is consumed and needs to be made-up. Water consumption is water that is lost due to blow-down, evaporative or windage losses, and water export [17, 63]. Blowdown is a maintenance procedure to clear water systems of particulate matter. Evaporative and windage losses take place in a cooling pond, or in the cooling tower where water is lost by natural convective forces. Water export occurs whenever process water is not recovered, such as steam leaving the system boundaries or water vapor being vented to the atmosphere. All of these losses consume approximately 5% of total circulated water in cooling systems, and 100% of the boiler feed water that is not recovered as condensate.

5.2.2 Water Integration Designs

Two water designs were considered in this work. The first is a simple heat integration design published by Huo et al [33]. It uses a combination of medium pressure steam and product recycle to heat the incoming feed stream, and it uses cooling water to reduce product stream temperatures. This scheme is shown in Figure 5-1. The overall water requirement is 23 pounds per hour of cooling water and 0.92 pound per hour of boiler feed water, both are on a per pound of feed oil basis. Huo et al. report that the cooling water value is higher than the UOP report of 13.5 pound per hour because optimized heat integration was not in the scope of their analysis.

The second water system design aims to reduce the amount of cooling water required by producing steam from the hot hydrotreator effluent streams. This heat integration approach reduces the size of the cooling tower and electricity requirement for pumping. Water consumption may increase with this design if the steam condensate is not recovered. This integrated water design in shown in Figure 5-2.

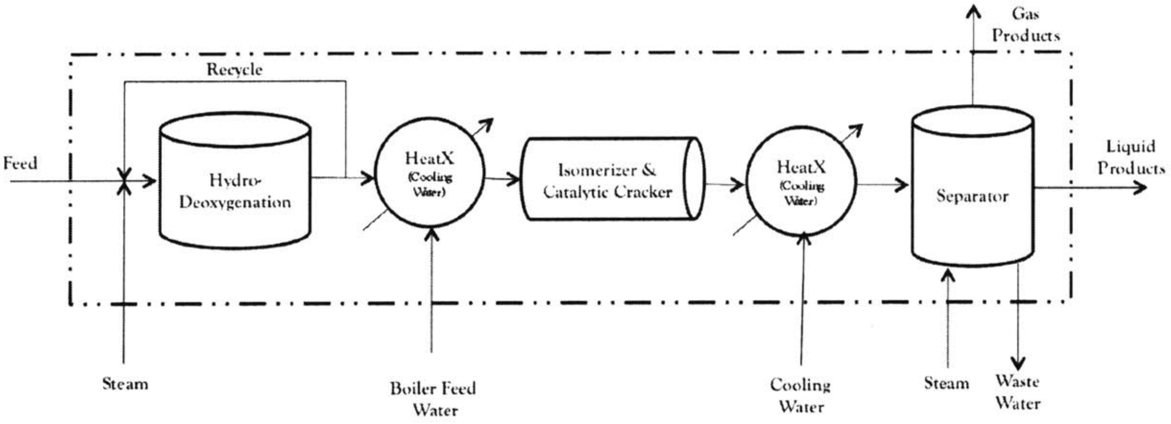


Figure 5-1: Cooling water design reported in Huo et al. Source: [33]

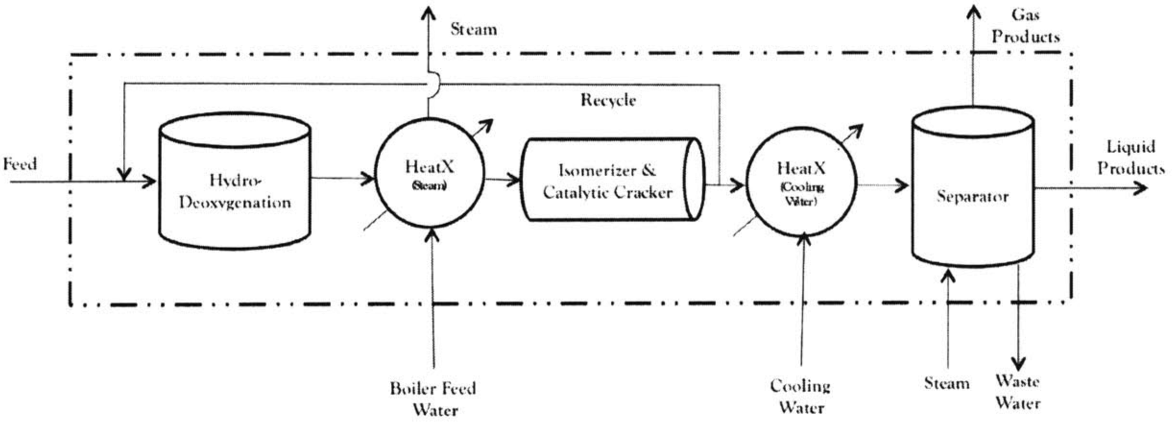


Figure 5-2: Alternate integrated water usage design.

5.2.3 Water Production

As discussed in Section 3.3, water is a by-product of the deoxygenation reaction. The amount of water production is determined by catalyst activity and selectivity [14, 39]. Although catalyst performance degrades over time, due to process-related fouling such as metal sintering or coking, the selectivity and activity of the catalyst was assumed to be constant over the entire lifetime of plant operation. This assumption simplified the analysis and was required because empirical data was not available.

5.3 Green House Gas Model

The GHG emissions of the process were evaluated on a lifecycle basis using data from Stratton et al. [85] and the energy and material balances from the Aspen Plus model. Details on lifecycle GHG modeling are available in Stratton et al [85] and are not given here. Lifecycle GHG emissions were obtained from GREET version 1.8D and [25, 85] for electricity and natural gas utilities. Natural gas GHG emissions were 65.96 gCO_2e per MJ for gas sourced from North America supplies and used in stationary applications. This value includes extraction, upgrading, and transport, as well as combustion emissions. Natural gas is combusted in large industrial boilers and direct fired heaters, as well as used as a feedstock for steam-methane-reformation to produce hydrogen gas. The carbon dioxide produced from the decarboxylation step is not included in the lifecycle analysis because it comes from a renewable source [85].

The GHG emissions from the electric utility were assumed to be U.S. grid average at a value of 200.39 gCO_2e/MJ. Grid average electricity consists of the following primary energy sources: 1% residual oil, 20.2% natural gas, 46.7% coal, 0.3% biomass, 21% nuclear, and 10.7% other renewables, (e.g., hydro, geothermal, wind, solar, etc.). The lifecycle GHGs of these sources are taken into account as a weighted average for the electricity GHG value used. Electricity is used to run motors for pumps, fans, and compressors, as well as for control equipment, such as, valves and sensors. The electric demand for the process was taken from [17] and the Aspen Plus model. The results of this modeling effort are reported in section 6.5.2.

Chapter 6

Results

"Once the toothpaste is out of the tube, it's hard to get it back in." - Harold Robbins Haldeman

6.1 Introduction

The results of the economic and environmental analyses are presented in this section. The cost contributions for capital and operating expenses are presented in sections 6.2.1 and 6.2.2 for different plant size, hydrogen source, and production profile. The baseline gate cost for distillate fuels using a discounted cash flow rate of return (DCFROR) model is presented in Section 6.4 for three plant sizes. The additional cost for maximum jet fuel production, on-site hydrogen gas production, financing, and ramp-up are given in Section 6.4.2. An additional cost sensitivity analysis is performed for vegetable oil feedstocks for several gate-prices in Section 6.4.6. Environmental results are presented for greenhouse gases and water consumption in Section 6.5.

6.2 Economic Costs of Production

The economic costs of distillate fuel production are presented in this section. The contribution of capital and operating expenses for 2,000, 4,000, and 6,500 BPD plant sizes are reported. Next, the cash-flow model results for the base case and several sensitivity analyses determined, with the results described in Chapter 6.4.

6.2.1 Capital Expenses and Economies of Scale

The capital expenses of the project are a function of equipment size and refinery scope. Capital expenses for inside-battery-limit (ISBL) equipment and balance of plant costs for the outside-battery-limits (OSBL) and offsites are reported in Table 6.1. Values are given on a cent per gallon of total liquid products basis, (e.g., naphtha, jet, and diesel). The estimate assumes a ten year straight-line depreciation for the capital equipment. The baseline scenario of maximum diesel fuel production was described in Section 4. The additional costs of hydrogen production, and maximum jet production are given as additional costs per gallon. Costs of capital are additive. For example, the capital expenditure would be 19 cents per gallon for a 4,000 BPD plant producing maximum jet fuel and on-site hydrogen: 15 cents per gallon for the baseline case price, plus three cents per gallon for hydrogen production, and one cent per gallon for maximum jet fuel production.

TPI is based on maximum distillate production with off-site hydrogen production. The OSBL and offsites are given for a greenfield facility with no existing infrastructure to leverage. Gary et al. [17] estimate the total cost of offsites at a greenfield to be 50% of the cost of storage, and ISBL equipment. Similarly, the credit for brownfield infrastructure, such as existing roads, offices and laboratories, railways, and distribution terminals might account for as much as 25% less than a greenfield site. A credit for the off-sites at a brownfield site is provided at the bottom of the table. The added cost for on-site hydrogen production and maximum jet fuel production profile are also shown.

Capital costs decrease with increasing plant size because of economies of scale for the engineered equipment, (e.g., hydrotreator, isomerizer, and saturated gas plant) costs. The total plant investment is 13 cents per gallon for the 6,500 BPD facility, and increases to 15 and 21 cents per gallon for the 4,000 and 2,000 BPD facilities, respectively. The economy of scale for 2,000 to 4,000 BPD is a 27% reduction in

Capital Expenses [cents/gal]		2,000	4,000	6,500
Hydrotreator		3	2	1
Isomerizer		5	3	3
Saturated gas plant		1	1	1
Subtotal A	(ISBL)	*9*	*6*	*5*
Storage	feed, 13 days	0	0	0
Storage	liquid product, 25 days	1	1	1
Storage	gas product, 25 days	0	0	0
Cooling water system		0	0	0
Subtotal B	(OSBL)	*10*	*7*	*6*
Offsites, greenfield	50% of Subtotal B	5	4	3
Subtotal C	(Balance of Plant)	*15*	*11*	*9*
Special costs	4% of Subtotal C	1	0	0
Contingency	15% of Subtotal C	2	1	0
Escalation	2010 CEPCI	3	2	2
Total plant investment	(TPI)	**21**	**15**	**13**
On-site hydrogen production	additional cost	5	3	3
Maximum Jet Production	additional cost	1	1	0
Brownfield offsites credit	25% of Subtotal B	(3)	(2)	(2)

Table 6.1: Capital Expenses, in cents per gallon, for each plant size [BPD]. Capital cost credits are shown in parenthesis.

capital expense for a 200% increase in capacity, and 39% capital reduction for a 325% increase in capacity for 6,500 BPD compared to 2,000 BPD.

The additional capital cost for hydrogen production, between three and five cents per gallon, is a result of additional capital cost for on-site hydrogen equipment raising the capital expense subtotals, and subsequently increasing the special costs, contingency, and escalation costs. Maximizing jet fuel production adds approximately one cent per gallon because of increased SMR facility size, and a 3% reduction in overall liquid fuel yield.

6.2.2 Direct and Variable Operating Costs

Annual direct and variable operating costs were calculated for three plant sizes with both products and hydrogen production scenarios, with the results shown in Tables 6.2 and 6.3. The additional costs of hydrogen production, and maximum jet production are shown in a manner similar to Table 6.1.

The direct annual operating expenses, shown in Table 6.2, account for 13, 16, and 22 cents per gallon of the gate cost for 6,500, 4,000, 2,000 BPD of capacity, respectively. Even though expenses increase with plant size, the contribution to the cost per gallon decreases because direct operating expenses are a function of total project investment, which are influenced by economies of scale. The additional direct operating expenses for on-site hydrogen production increase costs between two and four cents per gallon, because of increases to insurance, local taxes, maintenance, and miscellaneous supplies, which are based on the total plant investment. The additional cost of maximum jet fuel production, which is a result of 3% reduction in total liquid fuel production, is also shown.

The variable operating expenses for the maximum distillate and maximum jet fuel production profiles are reported in Table 6.3. The makeup water, electric power, and vegetable oil requirements are the same for both production profiles and are independent of plant size because the variable expenses are scaled linearly by production capacity in the economic model. Makeup water, though a non-zero annual expense for the process, is not significant in the cost per gallon calculation. Electric power is

Direct Operating Expenses		2,000	4,000	6,500
Catalyst		2	2	1
Insurance	0.5% of TPI	1	1	1
Local taxes	1% of TPI	2	2	1
Maintenance	5.5% of TPI	12	9	7
Miscellaneous supplies	0.15% of TPI	0	0	0
Plant staff		3	1	1
Subtotal		*20*	*14*	*11*
Contingency	10% of Subtotal	2	1	1
Total		**22**	**16**	**13**
On-site hydrogen production	additional cost	4	3	2
Maximum jet production	additional cost	1	1	0

Table 6.2: Direct annual operating expenses in cents per gallon, for each plant size [BPD].

two cents per gallon, and natural gas is 16 cents per gallon for the base case. Hydrogen gas is 13 cents per gallon for the maximum distillate fuel profile, and 21 cents per gallon for the maximum get scenario. The maximum jet profile consumed more hydrogen in the isomerization unit due to cracking reactions. The vegetable oil is the largest portion of the gate price, and is between 264 and 274 cents per gallon, while the total utility costs account for between 31 and 38 cents per gallon.

The variable expenses for maximum jet production profile are 17 cents per gallon higher than the maximum distillate profile because of hydrogen requirements, nearly seven cents per gallon more, and lower liquid product yields that contribute to eight cents per gallon higher vegetable oil expenses. On-site hydrogen production will increase the costs of utilities, (e.g., makeup water, electric power, and natural gas), because of additional demands from the SMR equipment. The overall cost increase is two cents per gallon for the maximum distillate case, and six cents per gallon for the maximum jet case. The maximum jet profile is higher again because of the additional hydrogen demand required by the process ,and the 3% reduction in liquid product

Variable Operating Expenses	Max. Distillate	Max. Jet
Makeup water	0	0
Electric power	2	2
Natural gas	16	16
Hydrogen purchase	13	21
Subtotal	*31*	*38*
Vegetable oil	264	272
Total	**294**	**311**
On-site hydrogen, additional cost	2	6

Table 6.3: Variable operating expenses reported in cents per gallon.

yield.

Table 6.4 summarizes the totals for capital, direct, and variable expenses. The three plant sizes are shown, as well as the additional cost for hydrogen production, and the additional cost for producing maximizing jet fuel production. Upgrading vegetable oil to liquid products costs between 74 and 55 cents per gallon. Vegetable oil is shown as its own line item, as it is the single largest contributor to the product price and contributes between 87% and 89% of the finished fuel costs.

On-site hydrogen production adds an additional cost of 11 cents per gallon for 2,000 BPD, eight cents per gallon for 4,000 BPD, and seven cents per gallon for 6,500 BPD. The variable expenses are two cents per gallon for on-site hydrogen production regardless of plant size. The cost reductions are primarily due to economies of scale, which reduce capital expenses from five cents per gallon to two cents per gallon, and subsequently reduce direct operating expense from four cents per gallon to two cents per gallon. Furthermore, maximizing jet fuel production adds nine cents per gallon, the variable operating expenses account for seven cents of that due primarily to increased hydrogen demand for this process.

These costs do not account for financing, taxes, or revenues from co-product sales. In reality, the costs could be reduced by co-product sales, and they could increase

Summary of Expenses	2,000	4,000	6,500
Capital expense	21	15	13
Direct operating expense	22	16	11
Variable operating expense	31	31	31
Subtotal	*74*	*62*	*55*
Vegetable oil	264	264	264
Total	**338**	**326**	**319**
On-site hydrogen production	11	8	7
Maximum jet production	9	9	7

Table 6.4: Summary of cost per gallon contributions.

because of interest and tax payments. The entire cash flow, including co-product sales, financing, and depreciation are accounted for in Section 6.3.

6.3 Cash Flow Modeling Using ROI

Cash flow models determine project profitability. A simple return on investment (ROI) analysis was used to compare plant size and production profile profitability using the historic costs of utilities, feedstocks, and total sales of finished fuel products. The ROI model was discussed in Section 4.3.3

The return on investment (ROI) analysis was used to determine the profitability of different plant sizes, and to compare maximum jet to maximum distillate fuel production. Table 6.5 shows the results of the ROI analysis. Using the five year national industrial average for utility and finished fuel product costs, described in Chapter 4, the facility is not economically viable for any size or production profile. This is because the variable expenses, and in particular the vegetable oil feedstocks, cost more than the finished products. Additionally, maximizing jet fuel production is even less profitable because it requires additional operating costs and reduces the

ROI Results	2,000	4,000	6,500
Direct operating costs	(7.22)	(10.12)	(13.36)
Variable operating costs	(85.73)	(171.45)	(278.61)
Total sales	64.35	128.71	209.15
Less depreciation allowance	(4.11)	(6.02)	(8.15)
Taxable income	0	0	0
Income tax	-	-	-
Net annual cash flow	(32.70)	(58.88)	(90.97)
Maximum Jet, (additional loss)	(1.33)	(2.66)	(4.32)
ROI (%/yr)	-	-	-

Table 6.5: Cash flows for ROI calculation. Values reported are in millions of dollars.

amount of liquid fuel available for sale, which are higher priced products than the propane and LNG products. Finally, because the project is not economical, increasing the size of the facility causes higher losses.

Since the project is not profitable with the given economic conditions, the return on investment is not reported. Input costs would have to go down to reduce variable operating expenses, or product prices would need to increase and boost total sales in order to off-set costs for the project to be economically sensible.

The distillate fuel gate prices and the vegetable oil costs needed to achieve given ROIs is determined with a more comprehensive cash flow model next. The results are reported using a DCFROR analysis that takes into account the time value of money, as well as capacity utilization and variable capital depreciation.

6.4 Discounted Cash Flow Rate of Return

A discounted cash flow rate of return (DCFROR) analysis was used to determine the gate cost of distillate fuel, and input price for vegetable oil that would be needed to satisfy the constraints on the model, such as internal rate of return, taxes, and loan payments. A description of the method used for the analysis was provided in Section 4.4 on page 51.

6.4.1 Baseline Results

The baseline scenario assumes an internal rate of return of 15%, with a 20 % equity structure, 5.5% loan interest and 10 year term. Capital expenses during construction are spread over three years and distributed as 8%, 60%, and 32% of the total project investment respectively. Depreciation is scheduled over 10 years, and uses the variable declining balance method. The plant operates at 100% of its name plate capacity, which means there is no production penalty during start up. The gate cost of fuel was found for the maximum distillate product profile produced for the three plant sizes: 2,000, 4,000, and 6,500 BPD. These assumptions were reported in Table 4.6 on page 53.

The input utility and feedstock costs for the baseline case are based on five year historical prices as reported in Table 4.3 on page 46. The costs were assumed to be \$0.06 per kolowatt-hour of electricity, \$373.18 per thousand pound of natural gas, \$0.04 per thousand pound of water, \$2.62 per gallon of soybean vegetable oil, and \$0.66 per pound of hydrogen gas. The product values were also based on historical data: \$1.77 per gallon of propane and LNG and \$2.03 per gallon of naphtha.

The gate cost for the distillate streams was found by setting the net present value of the cash flow spreadsheet to zero. The results of the baseline scenario are reported in Table 6.6 with the gate cost range being \$4.38, \$3.98, and \$3.80 per gallon for 2,000, 4,000, and 6,500 BPD facilities, respectively. Moreover, these gate prices are higher than the annual average maximum cost of \$3.87 per gallon for No. 2 diesel fuel, and \$3.98 per gallon for jet fuel reported by the EIA [92].

This means the HRO products would require a price-support, between \$2.25 per gallon and \$1.67 per gallon, in order to be economically competitive for the base case. These values are equivalent to March 2011 subsidies available from the \$1 per gallon Blender's Credit, and \$1.77 per gallon for renewable identification number (RIN)

DCFROR Results	2,000	4,000	6,500
Baseline	\$4.38	\$3.98	\$3.80
Sensitivities			
Product Slate			
Maximum jet production	\$0.30	\$0.26	\$0.25
Naphtha discount, 15%	\$0.00	\$0.00	\$0.00
Location Brownfield Offsites Credit	\$(0.25)	\$(0.18)	\$(0.15)
On-site Hydrogen Production	\$0.34	\$0.25	\$0.20
Financing Financing: 100% Equity	\$(0.38)	\$(0.28)	\$(0.23)
Financing: 25% IRR	\$0.14	\$0.10	\$0.08
Production Level			
Notional Ramp-up: 50%/75%/100%	\$0.52	\$0.37	\$0.31
Feedstock Shortage: 50%	\$2.13	\$1.53	\$1.27

Table 6.6: DCFROR gate cost and sensitivity results given in \$/gal. Credits shown in parenthesis.

trading [66]. However, with crude oil trading over \$110 per barrel and April 2011 price for diesel fuel around \$3.33 per gallon [92], only the Blender's credit would be needed to make the baseline competitive, and no subsidy would be needed for the 6,500 BPD plant with 15% IRR.

In addition to the baseline price for diesel and jet fuels, the effect of the following parameters on gate cost was determined: hydrogen source, brownfield offsites, financing equity structure, ramp-up, and input and product price sensitivities. These sensitivities were described previously in Section 4.4. The results are presented as baseline prices with additional "add-on" costs that account for changes from the baseline in Table 6.6.

6.4.2 Product Slate and Co-Product Price

The difference in gate cost for the maximum jet production profile and lower naphtha co-product price were investigated. Producing the maximum jet fuel profile requires an additional cost of $0.30 per gallon for the smaller plant and $0.25 for the larger plants thus yielding a finished jet fuel price between $4.05 and $4.68 per gallon. The additional cost comes from the reduced revenues of higher value liquid products. The cost is higher than the expense reported in Table 6.4 by the discount rate over the 20 year economic lifetime of the plant. These values are roughly twice the five year averages for jet and diesel fuels, which were used for the ROI analysis in Section 6.3. A parametric study was performed on the naphtha price with a 15% discount (approximately 25 cents per gallon) being applied to account for transport, upgrading, and blending at another facility. However, it has a negligible effect on the results. The five year average prices are $2.12 per gallon for jet and $2.13 for diesel. These values were reported in Table 4.4.

6.4.3 Brownfield and On-Site Hydrogen Production

The credit for building the facility at a brownfield site with existing infrastructure was found to range between 15 and 25 cents per gallon as shown in Table 6.6. The savings is more pronounced for smaller plants and can be the difference between producing economically competitive fuels for a small scale producer. The additional cost of on-site hydrogen production was investigated and found to add $0.34 cents per gallon for 2,000 BPD, $0.25 cents per gallon for 4,000 BPD, and $0.20 per gallon for 6,500 BPD capacity. This cost comes from additional capital costs, and operating expenses as described in Section 6.2.2. However, the gate cost of liquid fuels produced with on-site hydrogen could be reduced by $0.04 and $0.08 per gallon if the facility were 100% financed (not shown in the table).

6.4.4 Financing

The equity structure was changed from 20% to 100% to determine the effect of capital financing on the gate price. Financing the project costs approximately $0.23 per gallon to $0.38 per gallon depending on the size of the facility. This additional cost comes from interest on the capital project over the 10 year lifetime of the loan. In other words, the gate cost of diesel would be reduced by at least $0.23 per gallon if the project does not need to leverage the refinery with debit financing. Additionally, changing the internal rate of return from 15% to 25% would result in an additional cost of $0.14 per gallon for 2,000 BPD, $0.10 per gallon for 4,000 BPD, and $0.08 for 6,500 BPD of capacity on top of the baseline distillate fuel cost. The additional revenues on the distillate fuel are required to cover the discounting rate.

6.4.5 Production Level and Ramp-Up

It is unrealistic to assume the plant can achieve 100% utilization of capacity in the first year after construction is completed. This is because of industrial start-up practices, which turn on the plant slowly for system verification, and because of possible feedstock shortages. Two ramp-up scenarios were considered. The first ramp up scenario assumes a notional schedule of 50% capacity in the first year, 75% utilization in the second year, and full capacity is achieved in the beginning of the third year and maintained thereafter. This ramp-up schedule requires an additional cost of $0.52, $0.37, and $0.31 per gallon for 2,000, 4,000, and 6,500 facility sizes, respectively, over the baseline cost.

The second scenario assumes the plant operating at 50% of nameplate capacity for the entire lifetime because of chronic industrial feedstock limitations. This means, for instance, that a 4,000 BPD facility will process 2,000 BPD of feedstock. As a result of lower fuel production volumes, the price increases above the baseline cost by an additional $2.13 per gallon for 2,000 BPD, $1.53 per gallon for 4,000 BPD, and $1.27 per gallon for 6,500 BPD of nameplate capacity. This sensitivity shows that the cost of capital can have profound consequences on the gate cost if a refiner decides to build a plant that she can "grow into".

6.4.6 Vegetable Oil Cost Sensitivity

The dominant cost in Table 6.6 is the vegetable oil feedstock. This section examines the maximum vegetable oil cost that could be supported with comparison to historic prices. It is assumed in this analysis that refineries are price-takers because the market for commodity fuels is competitive. If the market price for distillate fuel is $3.00 per gallon, the firm with the lowest cost of production will be the most profitable. Since vegetable oil costs are the largest contributor to the gate price of the finished fuel, the feedstock cost is critical to profitability. This sensitivity study examines the maximum price that a refiner can purchase feedstocks for a given fuel price with all other costs constant.

Five distillate fuel prices were examined: $1.50 per gallon, $3.00 per gallon, $3.50 per gallon, $4.00 per gallon, and $6.00 per gallon; to correspond to an optimistic price, several notional near term domestic prices, and a price typical of other countries. The vegetable oil prices for each fuel price were then calculated with the DCFROR cash-flow spreadsheet model.The maximum cost that each plant size could pay for feedstock and maintain a 15% IRR are reported in Table 6.7.

The maximum price a refinery can afford to pay for vegetable oil decreases with plant size. Larger plants can afford to pay higher feedstock prices than smaller plants. This means that smaller plants are more sensitive to changes in feedstock price and need access to lower cost vegetable oils to be able to produce fuels at competitive prices. If a plant can purchase feedstocks below these costs, then the firm would have lower production costs and higher profits.

Distillate Fuel Gate Price	Vegetable Oil Price		
	2,000	4,000	6500
$1.50/gal	$(0.08)	$0.36	$0.45
$3.00/gal	$1.35	$1.77	$1.88
$3.50/gal	$1.80	$2.18	$2.33
$4.00/gal	$2.25	$2.63	$2.78
$6.00/gal	$4.13	$4.50	$4.65

Table 6.7: Maximum feed vegetable oil cost that a refinery could pay for fixed distillate fuel gate cost. Values in $/gal of feedstock. Negative values shown in parenthesis.

To be profitable with $1.50 per gallon distillate fuel, the 6,500 BPD refinery could pay up to $0.45 per gallon for vegetable oil for 6500 BPD, but only $0.36 per gallon with a 4,000 BPD capacity. These prices are two to three times lower than the 20 year historic minimum of $0.90 per gallon, which makes both scenarios very unrealistic. Additionally, the 2,000 BPD plant could afford to produce fuel for $1.50 per gallon if it were paid $0.08 per gallon of vegetable oil. This is an unrealistic scenario for vegetable oil, or waste greases, and animal fats because restaurants and rendering facilities are not paying to have these products removed [67].

Furthermore, 2,000 BPD of waste grease is a sizable amount and a producer would have difficulty getting these sources locally. For example, according to a 1998 NREL report on urban waste grease, 33 BPD of grease and fat available in all of the metro-Boston area [99]. Personal communications with Baker Commodities in Billerica, Massachusetts and Kraft Gelatin in Woburn, Massachusetts have indicated additional grease and fat inventories of 452 BPD and 200 BPD, respectively in the Boston metro area [67, 65]. Furthermore, 430 BPD might be available in all of Massachusetts if one gallon of waste grease were generated per capita and the population is 6,500,000 [66]. These values indicate 700-1000 BPD could be available from all sources in Massachusetts.

For the $3.00 per gallon, $3.50 per gallon, and $4.00 per gallon scenarios, the maximum feedstock costs would be between $1.35 per gallon and $2.78 per gallon. These prices are not unprecedented for waste greases, and might be achieved with additional production of next generation oils such as jatropha or algal oils, or with government subsidies for production [66, 67]. However, only the 4,000 and 6,500 BPD facilities can afford to pay for feedstock and profitably produce $4.00 per gallon fuels given the five year average soybean oil price of $2.63 per gallon. Whereas, production at current soybean oil prices of around $4.00 per gallon only make sense at the $6.00 per gallon of fuel level. This could explain why most of the fats and oils are being exported out of the country and to the international biofuels market [67].

As discussed in Section 6.4.2, maximum jet fuel production results in lower revenues. This means that maximum distillate producers can afford to pay more while maintaining a given IRR. This puts the jet producers at a strategic disadvantage. Jet producers could afford to pay $1.20, $1.73, $2.25, and $4.20 per gallon of vegetable oil less for the \$3.00 per gallon, \$3.50 per gallon, \$4.00 per gallon, and \$6.00 per gallon gate prices, respectively.

Reducing production costs would increase the maximum feedstock price the refinery could pay while maintaining a fixed IRR. For example, 100% equity financing was found to increase the maximum vegetable oil cost between \$2.18 per gallon and \$3.53 per gallon depending on the size of the facility; these values are based on a \$3.00 per gallon distillate fuel gate price. This additional purchasing power improves economic competitiveness of fully financed plants and makes them better prepared for fluctuations in feedstock prices.

6.5 Environmental Results

The results of the water usage, and life-cycle GHG emissions are discussed in this section. The models used for this analysis are described in chapter 5.

6.5.1 Water Usage

The section reports the results of using steam to remove heat from the product streams and compares it to previous reports in the literature that used cooling water. The water models were described in Section 5.2.

The Huo et al. model, which used medium pressure steam to pre-heat reactor streams, and cooling water to reduce product stream temperatures requires 2.3 pounds of water per pound of feed vegetable oil [33]. Huo et al. reports that the UOP process, which follows a similar design, uses 1.4 pounds of water per pound of vegetable feed oil. The alternative water integration design, which uses reduces makeup water by generating steam instead of using cooling water, was found to require 0.8 pounds of makeup water for cooling and steam production per pound of feed vegetable oil processed. This is a savings of as much as 54% compared to previously reported water demand.

SMR production requirements, which were not reported by the other authors, were found to be 0.2 pounds of water per pound of vegetable oil. From the chemistry of the system, it was found that nine pounds of water are produced per 100 pounds of vegetable oil. The produced water could be treated and polished for reuse in the boiler or cooling water systems, rather than discharged to the sewer and reduce approximately 10% of the total makeup demand. The total water demand for the process is 0.9 gallons per gallon of vegetable oil and includes These results are reported in Table 6.8.

System water usage	lb water
Huo et al.	2.3
UOP [33]	1.4
Makeup water demand	0.8
SMR Demand	0.2
Water Co-Product Produced	(0.1)
Total Water Required	0.9

Table 6.8: Overall system water production and makeup demand in gallons of water per gallon of vegetable oil feedstock.

6.5.2 Lifecycle Greenhouse Gas Emissions

The greenhouse gas emissions (GHG) of the HRO process are evaluated on a lifecycle basis and compared to results from Stratton et al. [85]. In this analysis hydrogen production contributes the same emissions to the fuel product regardless of where the SMR is sited. Therefore the GHG emissions are reported for finished fuels and include contributions from hydrogen production. Data from GREET [86] and [85, 25] were used to calculate the life-cycle GHG emissions of the process. United States domestic natural gas used in a stationary application has a life-cycle GHG value of 66.0 gCO_2e per MJ. United States grid average electricity has a value of 200.4 gCO_2e per MJ. Energy allocation was used for the diesel, jet, naphtha, LNG, and propane products.

It is assumed that the fuel demand is partially satisfied by the propane and LNG co-products of the process, and the remainder by purchased natural gas. Naphtha would be sold as a liquid product and not used as an on-site fuel because it is not compatible with the gas-fuel burners, and therefore is not included in the fuel balance. The process requires a total demand of 0.12 pounds of process fuel per pound of vegetable oil feed for maximum distillate production, and 0.15 pounds of fuel more for the maximum jet fuel profile to produce additional hydrogen. Propane is produced in equal quantities for both the maximum distillate, and maximum jet fuel profiles because it comes from the glycerol backbone of the triglycerides. LNG co-products are tripled in maximum jet profile because of increased cracking reactions.

The maximum distillate profile produces 0.06 pounds of fuel per pound of vegetable oil and the maximum jet fuel profile produces 0.10 pounds of process fuel per pound of vegetable oil. The processes require a total demand of 0.12 pounds of fuel per pound of vegetable oil feed for maximum distillate production, and 0.15 pounds of fuel for the maximum jet fuel profile in order to accommodate the production of additional hydrogen. This leaves a net demand of 0.06 and 0.05 pounds of fuel per pound of feed that must be made up with natural gas. These values were used with data from Stratton et al. [85] to calculate the life-cycle GHG emissions of 8.9 gCO_2 per MJ for the maximum distillate fuel profile, and 11.8 gCO_2 per MJ for the maximum jet fuel profile.

Similarly, the electricity demand of 0.03 kilowatt per pound of vegetable oil for the process was found based on the process design. No change in electricity consumption was found for the different product profiles, since there was no additional electric demand for additional utilities or process equipment. The SMR equipment was found to have a non-zero electric demand, but is smaller than the significant figures reported here. The GHG emissions from electricity of the processes were found to be 1.2 gCO_2e per MJ for the maximum distillate and the maximum jet profiles.

The combined GHG emissions for the processes are 10.1 and 13.0 gCO_2e per MJ for the maximum distillate and the maximum jet profile respectively. The maximum jet profile has higher GHG emissions because it uses 1.7 times more natural gas than the maximum distillate profile. These results agree with the range of 7-13 gCO_2e per MJ reported in [85, 86]. Table 6.9 summarizes the lifecycle GHG emissions for producing both production profiles.

Lifecycle GHG Emissions	Maximum Distillate	Maximum Jet
Electricity	1.2	1.2
Natural Gas	8.9	11.8
Total	10.1	13.0

Table 6.9: Process greenhouse gas emissions in gCO_2e/MJ.

Chapter 7

Policy and Market Review, Discussion, and Implications

> *"Giving money and power to government is like giving whiskey and car keys to teenage boys."* - P.J. O'Rourke

7.1 Introduction and Motivation

Biomass based fuels were utilized well before petroleum became the dominant source of liquid transportation fuels. Early Americans used whale blubber to light lamps. In 1900 Rudolf Diesel ran his compression ignition engine on peanut oil at the World's Fair in Paris. Farmers used biomass-derived syngas, called "wood-gas", to run farm equipment during the 1940's. For additional historical context see the work by [15, 43, 50, 80].

Today biomass based fuels are being produced as substitutes to petroleum for environmental, national security, and economic reasons. Yet, current policies have not addressed feedstock price volatility and supply limitations, nor have they addressed project finance difficulties for smaller-scale independent producers. A review of stakeholders and efforts in the market are discussed in this chapter with respect to the potential of hydroprocessed renewable oil technologies in contributing to the creation of a renewable, distributed, and diversified jet fuel production infrastructure. As will be explained, the potential for HRO jet fuels to penetrate the market is limited by feedstock availability and access to capital.

7.2 New HRO Plants versus Old Biodiesel Infrastructure

The revised Renewable Fuels Standard (RFS2) is the main renewable fuels legislation in the United States. The RFS sets mandates for fuel production volumes, lifecycle greenhouse gas (GHG) emissions standards, and price supports. While, the RFS2 states that fuels must be produced, it is does not specify which technology must be used. Therefore, if there are limited feedstocks available, why would new HRO fuel facilities be construction when the biodiesel industry is utilizing only 10% of the 2.5 billion gallons of capacity already available?

The production of HRO fuels is motivated by several technical advantages. For example, HRO is more similar to petroleum based fuels than biodiesel. In particular, HRO fuels have better cold weather properties, and have higher energy content than biodiesel making them more suited for a wider geographical area. Furthermore, the chemical similarity of HRO allow these fuels to be transported in the existing fuel infrastructure, and used in transportation equipment without the need for modification or retrofit [33, 25]. Finally, the HRO process creates fuels suitable for use in aircraft, heavy and light duty, as well as passenger equipment, whereas biodiesel may only be used in a small percentage blend in diesel engines and home heating oil. While these advantages are strong motivators, the existing biodiesel infrastructure has several economic advantages over new HRO plants.

Although most biodiesel facilities have been deactivated, reactivating existing biodiesel plants have several economic advantages. For example, purchasing distressed assets from holding companies or banks that have already written off the capital costs as a loss, or have been sold in bankruptcy may have lower costs than constructing new facilities. Additionally, the operations and logistics of the supply chain are more mature for biodiesel than HRO, which could also have positive effects on operating expenses, and thus drive down the gate cost of fuel. Finally, purchasing an existing facility would allow a producer to come to market faster since it has already been constructed, thus mitigating most permitting and licensing risks.

It maybe possible to achieve the best of worlds by repurposing assets from deactivated biodiesel[1] and turning them into HRO product facilities. This brownfield approach could mitigate permitting and construction risks, reduce capital costs and operating costs, while producing technically superior products, such as aviation grade jet fuels. The applicability of this approach in practice however, may be limited to just a few brownfield facilities across the country that are suitable for reactivation and that are also located near suitable sources of feedstocks.

7.3 Trade-offs of Environmental and Economic Costs

Policy decisions can influence economic and environmental trade-offs for HRO processed including, location, size, and product slate. For example, location may impact the cost of fuel production because capital and operating costs change in different parts of the country [17]. For example, co-locating a facility in the Gulf Coast of the United States may reduce construction costs and GHG emissions by taking advantage of the existing hydrogen supply pipeline, fuel storage, and distribution infrastructure that was established for the refinery industry, as well as the fats from the regional poultry and hog industry. While co-location has potential economic and environmental advantages, the RFS2 states that biomass derived feedstocks that are co-processed at an existing petroleum facility do not qualify as renewable fuels, and therefore do not generate RINs or other price supports. Therefore, refineries might be built as "greenfields" farther away from existing utility and transportation infrastructure.

The size of a facility may also be influenced by policies like accelerated depreciation, favorable loan terms, and other tax credits. A refinery is most profitable when

---

[1] These may include other deactivated or distressed assets such as oleo-chemical or petroleum refineries

it is utilizing 100% of its capacity, and while the minimum refinery size is limited by economies of scale, the maximum size is usually determined by the availability of feedstock. However, with the right economic incentives, a refiner may choose to build a facility that can accommodate more than the available feedstock supply. In the short term this approach could inspire "if you built it they will come" effects and promote new feedstock providers, but requires sufficient price supports to off-set the extra cost of capital until such feedstocks are available.

In the case of the ramp-up sensitivity analysis discussed in Section 6.4.5, fuels would be between \$0.31 and \$0.51 per gallon more expensive that the baseline case if a refinery was underutilized, as seen in Table 6.6 on page 72. This equates to \$19M and \$31M per year, or roughly 1/2 the total fixed capital expense of the project, required in subsidies, tax credits, or depreciation to off-set the period of under utilization. Without these price supports the refinery would have to pass these costs on to the fuel purchaser, and risk being less economically competitive in the market. If the operator is never able to reach the name full capacity of the plant, even higher subsidies are required to offset the additional cost of fuel production. For example, if the plant operates at 50% of its capacity over the entire economic lifetime, fuels would be between \$1.27 and \$2.13 per gallon more expensive than the baseline case as seen in Table 6.6. The total economic cost for this scenario is between \$65M and \$126M per year, which is equivalent to producing a second greenfield plant when feedstocks become available, if ever.

The product slate may also be determined by policy decisions. It was found in Section 6.4.2 that the maximum jet fuel profile costs between \$0.25 and \$0.30 per gallon more than the maximum distillate profile, due to additional natural gas demand for hydrogen production and reduced revenues from high value liquid products. Since the historic annual average price difference for diesel and jet fuel has never been greater than \$0.08 per gallon, a policy instrument would be required to incentivize jet fuel production over diesel production. Such a policy might take the form of additional mandates for renewable LNG or non-ethanol gasoline additives in order to derive more revenues from cracking co-products and in effect subsidize the cost of maximum jet fuel production. Another policy mechanism might subsidize the cost of jet fuel production outright in order to ensure renewable jet fuels are available on the market. However, these price-supports would be required indefinitely since the process is inherently more expensive than HRO diesel production.

On the other hand, policies that encourage the production and availability of shorter chain renewable oils could make HRO jet fuel more economically and environmentally competitive over the long term. For example, the environmental impact of maximum jet production is approximately three gCO_2e per MJ, and \$0.26 per gallon higher than the maximum distillate profile when soybean oil is used as a feedstock at a 4,000 BPD facility. This GHG and cost penalty might be avoided if shorter chain oils were available at economically attractive prices. Such feedstocks would require less natural gas to process since the oil chain lengths are already in the right range for jet fuels. Policies, which encourage shorter chain oil feedstocks, might include supporting algal, synthetic biological fermentation, and other fuel-specific terrestrial oil crops with research and development, scale-up, and demonstration financial assistance.

7.4 Feedstocks Supplies

The prices for commodity oils and fats have tripled during the last ten years [19, 66]. Historically, used grease and fat feedstocks have traded between \$0.70 and \$1.60 per gallon depending on the quality and market conditions, and recently these same feedstocks are trading between \$3.50 to \$4.50 per gallon, which is only a few cents lower than virgin quality vegetable oils [67]. This increase has been caused by increased demands in the international biofuels market, as well as rising food prices. These high prices limit the economic competitiveness of HRO and biodiesel fuels at current crude oil prices. Until the gate price of distillate fuel rises above \$4.00 per gallon, as reported in the vegetable oil sensitivity analysis on page 74, renewable fuel producers in the United States may not be able to purchase feedstocks.

In 2008 when the biodiesel companies in the United States reduced production, the feedstock brokers and renderers had to find new off-take agreements with credit-worthy counter-parties [67]. The oils and fats found their way onto the international market and are now exported to Europe, South America, and Asia, because the international market, particularly in Europe, has greater economic flexibility to purchase feedstock. A policy that limits the exporting of waste greases and oils for international biofuels production, might enhance the competitiveness of domestic biofuels production without the need for fuel prices to rise. This is because it would protect domestic fuel producers against high international market prices and overcome the reluctance of oil and fat brokers to engage in long-term price and volume contracts. The inherent increase in lifecycle GHGs incurred by transporting these renewable feedstocks across the ocean may also be solved by this type of policy.

7.5 Large versus Small Suppliers

Assuming that all 318,000 BPD of fats and greases in the United States were available for conversion to HRO fuels (see table 2.2 on page 24), then roughly 80 plants, 4,000 BPD in size would be required. At approximately $25,000 per BPD of capacity, the estimated total capital layout would be nearly $8,000,000,000 (eight billion dollars). This equates to 4.1 billion gallons per year of distillate fuel[2], or 10% of the total jet fuel demand of the United States in 2010 [92].

As discussed previously, renewable oils co-processed in petroleum facilities do no qualify as renewable fuels. Though it is not clear if an explicit intention of the Energy Security and Independence Act of 2007 was to limit the oil majors and spur the growth of a distributed infrastructure of smaller plants, it is a curious consequence of the legislature. This begs the question, of who will build and operate these plants?

Two HRO projects in the United States, both located in Louisiana, have created vertically integrated joint ventures. Tyson Foods and Darling International supply fats and grease from their operations to Syntroleum and Valero refineries, respectively. While this is efficient because it allows the fuel producers to build larger facilities,

---

[2] assuming an 84% conversion

which take advantage of economies of scale discussed in chapter 6.2.1, it signals that only larger firms have access to feedstocks through strategic joint ventures.

If the two HRO plants in Louisiana are successful, then these producers might repeat it across the country. With 50 - 100 rendering and packing facilities across the country holding inventories of fats and greases, then 318,000 BPD of capacity could be installed assuming firms built on average 80, 4,000 BPD facilities. This estimation assumes sufficient site locations exists, and environmental permits are obtainable; both of which could cause cost over-runs for additional site development, and legal fees for a prolonged permitting appeal or court battle. Additionally, 80 new refineries would increase the number of refineries by 50% while only increasing the refining capacity of the United States by 2% [3].

In addition to the rendered sources, companies investing in new forms of feedstocks are likely to site those projects near the refining capacity in order to de-risk their investments. This could mean that new feedstock production projects, either through the United States Department of Agriculture and Department of Energy Farm-To-Fly initiative to produce oil-seed crops, or advanced biological pathways of photosynthetic or fermentation organisms, will likely be co-located near HRO facilities.

Smaller regional fuel producers may be able to take advantage of smaller markets that do not meet the return on investment larger firms require. However, the opportunities for these smaller firms to partner with feedstock suppliers and create markets on their own, are arguably more economically risky and less economically competitive than larger producers. As a result, the small regional firms may not have access to the capital they require to enter the market.

7.6 Financing

Financing is also a major challenge for small regional firms to secure because of the outstanding underutilized biodiesel capacity, and the lack of in-take and off-take agreements. First of all, HRO facilities cost between $65-$100 million for 2,000-6,500

---

[3] The EIA reported total operable atmospheric crude tower capacity at 17,500,000 BPD[92].

BPD of capacity. Because of the high capital costs, these projects cannot be funded by bootstrapping, or going into "start-up mode" with a venture capital (VC) firm like some other clean-tech or high-tech firms who need $3-4 million. The only path to finance this type of project are with equity investments from a multi-million dollar company[4] with a large balance sheet and high risk tolerance, and debit financing with a large private equity firm or project finance bank.

In either case, both the equity and debit partners would want to reduce their investment risk by making sure the business has in-take and off-take agreements in place. This creates a chicken and egg problem. Feedstock suppliers will not provide long term contracts without proof that the business has a credible financial plan, which often requires long-term price-fixed off agreements. Creating meaningful off-take agreements requires having a product to sell, and a price that you can sell it at. Without knowing the up-stream costs, there is no realistic way to do this and mitigate risk. Bringing both players together at the same time in order to facilitate negotiations seems to be one of the only methods for mitigating this problem.

In addition to contracts for cash-flow, the financing institutions would also prefer to see an experienced operator engaged in the project. These are complex, chemical unit operations that necessitate sophisticated understanding of process engineering and production. Additionally, experience storing, blending, and scheduling feedstock and product delivery logistics are also required skills for an operator. Producing HRO fuels is not a scaled up garage biodiesel enterprise that can be learned from watching online videos or attending a seminar at conference. It requires a seasoned oleo-chemical veteran who understands the process, the business, and the market. Such experienced operators are few and far between, especially outside the typical US refining geography of the Gulf Coast.

---

[4] Or very high net worth individual or network of individuals.

7.7 Viability

In conclusion, while the biomass based fuels are being produced as petroleum substitutes, there are several economic and environmental trade offs that have not been completely addressed by policy. While the idle biodiesel infrastructure in the US has several economic advantages over new HRO facilities, the technical advantages of HRO fuels make these fuels more attractive. However, the limited availability of inexpensive feedstocks and easy access to project finance limit the viability of HRO market penetration. Such limitations could be addressed with policies that limit the exportation of waste greases to the international biofuels market, as well as support of new feedstock technologies and industries, specifically for producing shorter chain oils to make HRO jet fuel more economically and environmentally competitive.

Chapter 8

Concluding Remarks and Future Work

"If I had more time, I would have written less." - Mark Twain

8.1 Summary

Aggressive targets for renewable fuel programs and mandates have created market signals, and comprehensive price supports, including incentives and credits, spurring and supporting projects in research and development, education, infrastructure and production. Key stakeholders including producers, manufacturers, and consumers have been involved in the process to address and attempt to overcome the limited successes of first generation biofuels, such as chronic under capacity production, limited fungibility, and distribution issues. Hydroprocessed renewable oils are one such fuel production pathway that address the need for drop-in quality, synthetic fuel.

As part of continuing research on alternative jet fuels, this work modeled the cost of production under various economic scenarios, process designs, and product profiles. The baseline gate cost for distillate fuels, and the additional cost for maximum jet fuel production, on-site hydrogen gas production, financing, and ramp-up were found and presented in Chapter 6. It was found that the baseline cost for HRO fuel production ranges between \$3.80 and \$4.39 per gallon depending on the size of the facility.

Producing a maximum jet fuel profile requires approximately \$0.25 per gallon more to cover increased operating expenses and reduced revenues from liquid products.

The location, hydrogen source, and ramp-up schedule also strongly influence the gate price of finished fuels. Co-locating on existing industrial sites, or deactivated refineries can reduce gate prices by up to \$0.25 per gallon. Similarly, the cost of production can be reduced by up to \$0.38 per gallon is the plant is paid in cash and does not require financing. The ramp-up scenarios and feedstock simulations demonstrated that it in the worst case, it is better to build to size then to over-build and risk cost penalties in the range of \$0.31 to \$2.31 per gallon.

Additional cost sensitivities revealed that vegetable oil feedstock prices must return to historic average prices to compete with petroleum fuels at current prices. However, the international fuel market, especially in European can afford to pay nearly twice as much for feedstocks because of higher fuel costs abroad. New policy mechanisms, such as export tariffs or other economic measures could support domestic fuel production and off-set imported petroleum products.

The lifecycle greenhouse gas emissions and the water requirements for fuel production were also found and reported in 6.5. The water usage for an improved heat integration system were between 0.5 and 1.4 pounds of water per pound of vegetable oil feedstock processed lower than the values reported in the literature. This water usage improvement comes from generating steam instead of using cooling to control the exothermic reaction in the hydrotreator.

Finally, the policy landscape for producing jet and diesel fuels from renewable oils was reviewed from the perspective of a fuel producer. It was found, that the potential of HRO fuels penetrating the market is dependent on the availability of feedstocks and access to capital. While there are comprehensive mandates and price supports for the production of advanced renewable fuels, it is difficult for small regional producers to enter the market because of limited feedstock supplies and access to capital issues. Whereas larger, legacy firms such as Valero and Syntroleum have formed strategic joint ventures to secure feedstock, reduce project risk, and secure loan guarantees from the Federal government.

8.2 Future Work

This model is a comprehensive tool with much potential for conducting scenario analysis. Changes in inflation and depreciation rates, as well as utility and process input costs are obvious choices for sensitivity studies. More complicated risk analysis with predictive simulations and forecasting could also be used with software programs such as Crystal Ball from Oracle. For example, a Monte Carlo simulation for stochastic modeling could be used to look into the sensitivity of product prices under uncertain prices in the future. In addition, historical price data and location cost factors could be used to evaluate how the economic viability might change in different regions across the United States.

Additional process equipment could be added to reform the naphtha product into a finished gasoline fuel, or cracked into products for the petro-chemical industry such as olefins for polymer and specialty chemical production. The additional capital costs and utility demands would have to be instrumented in the process model, and the product profiles would also have to be updated. Other improvements to the model might include having small scale Fischer-Tropsch (F-T) reactors and upgrading units on site to take advantage of natural gas products or additional biomass materials that might be available at an integrated oil-seed or micro-biological, (e.g., algal or yeast), oil processing growth facility.

Different oil feedstock profiles could be added to the model. In addition to soy bean oil, other vegetable oils such as canola, jatropha, and camelina could be modeled. Other grease and fat profiles could also be used such as algal strains, beef tallow, choice white hog grease, yellow grease and used vegetable oils. Each feedstocks has a different carbon chain distribution and level of saturation. As a result, the hydrogen consumption and product yields would vary with a resulting impact on the economics of the project. For example, oils from certain algal strains are more saturated and have a lower carbon number and could be used to produce more jet fuel without cracking the diesel range molecules. This could reduce hydrogen consumption and naphtha production while maximizing jet fuel yields at a potentially lower cost.

There are several additional issue with the model that I would have addressed if I had more time. These include high net cash on hand at the end of every year, and the independence of diesel and jet fuel price with the price of co-products, utilities, and inputs. A more detailed financial model could use those cash flows to service high debt costs, and therefore reduce the cost of production over the lifetime of the plant. This was out of scope for this project because it would require detailed assumptions for the corporate structure, and special purpose vehicles, all of which are highly dependent on negotiations between equity finance and project developers.

While it is informative to see what the price of distillate fuel would be holding all other things constant, the independence of diesel and jet fuel prices with the other costs and prices of utilities and co-products is unrealistic. In other words, if the cost of diesel shoots up from \$2.12/gal to \$5.76, the costs of gasoline and LNG should also increase. The same is true for the cost of utilities and inputs. The current model does not allow for this, and as a result, the price of distillate fuel is higher than it otherwise would be in order to compensate for lower revenue streams from the other products. Coupling these prices together with forecasts, or correlations from regressed historical data, or formula based price indexes are potential ways of solving this issue instead of just escalating the prices with a constant inflation multiplier.

Appendix A

Aspen Models

The HRO process is modeled in Aspen Plus. The process flowsheet, showing an example of material and energy streams is shown in figure A-1.

The soy oil feed (OILFEED) stream is pressurized (OILFEED2) and heated with the reactor effluent stream followed by medium pressure steam. It is then mixed with hydrogen and reacted. The effluent (RXEFF) is cooled to 100 oF (RXEFF2) to reduce the vapor fraction and then (RXEFF3) sent to a preflash drum to separate the MPG and from the HRD and waste water mixture. The vapor product (MIXEDGAS) is separated into MPG (MPG) and hydrogen (H2REC) in a PSA unit. The hydrogen is recycled and pressurized for further reaction. The liquid products are separated in an 8-stage column with no reboiler or condenser. Live low-pressure (MPSTM) steam is feed to tray nine to facilitate separation of the HRD (GDSL) from remaining MPG (LIGHTS) and water. Storage tanks, cooling water, and steam boiler units are added in the economic analyzer because such units are not available in Aspen PlusTM. The quoted price for an SMR hydrogen island (60MM USD) is added to the capital costs after the economic analyzer simulation was completed because a packaged unit was not available, and it was too complicated to build up from scratch.

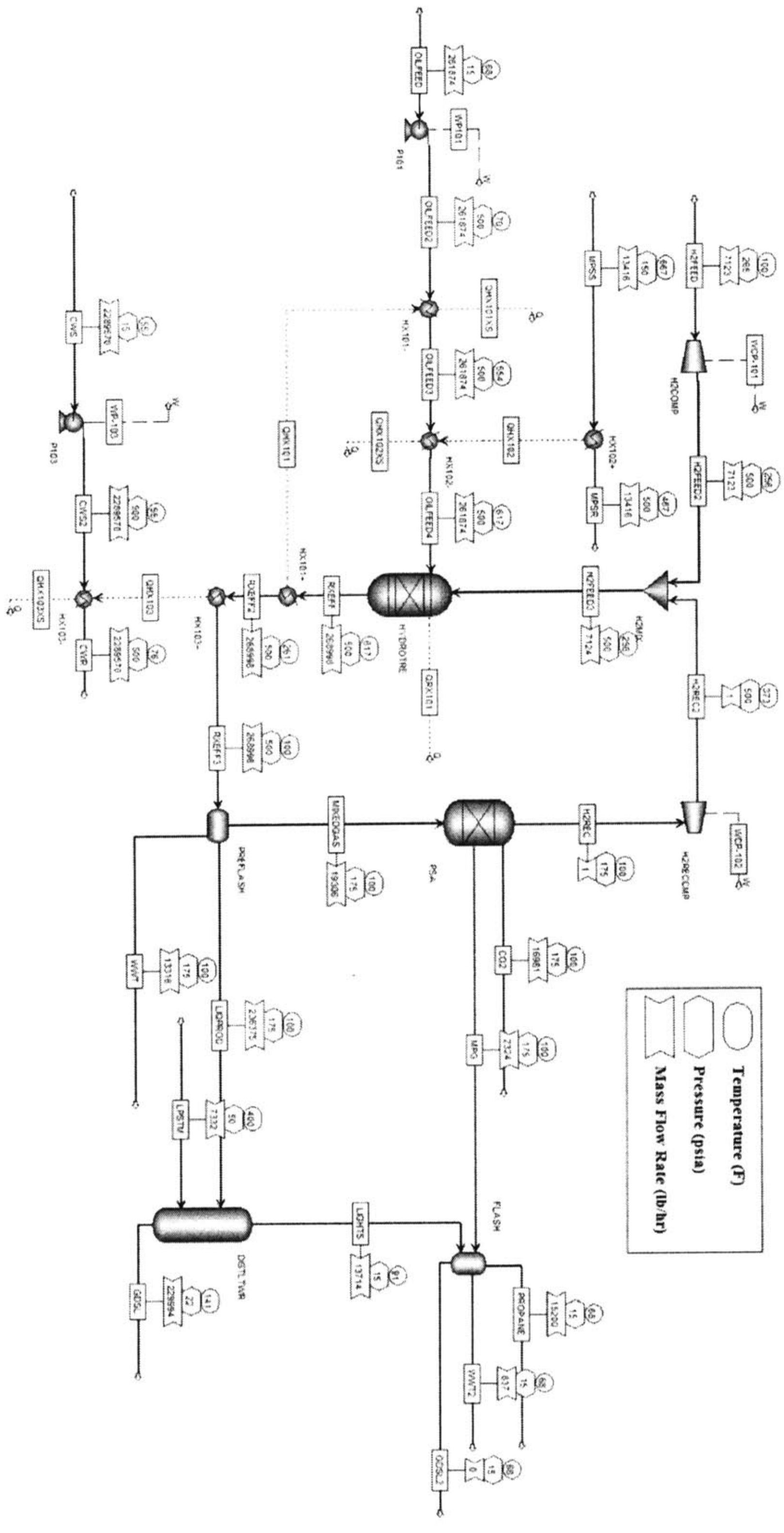


Figure A-1: The process flowsheet from Aspen Plus simulation.

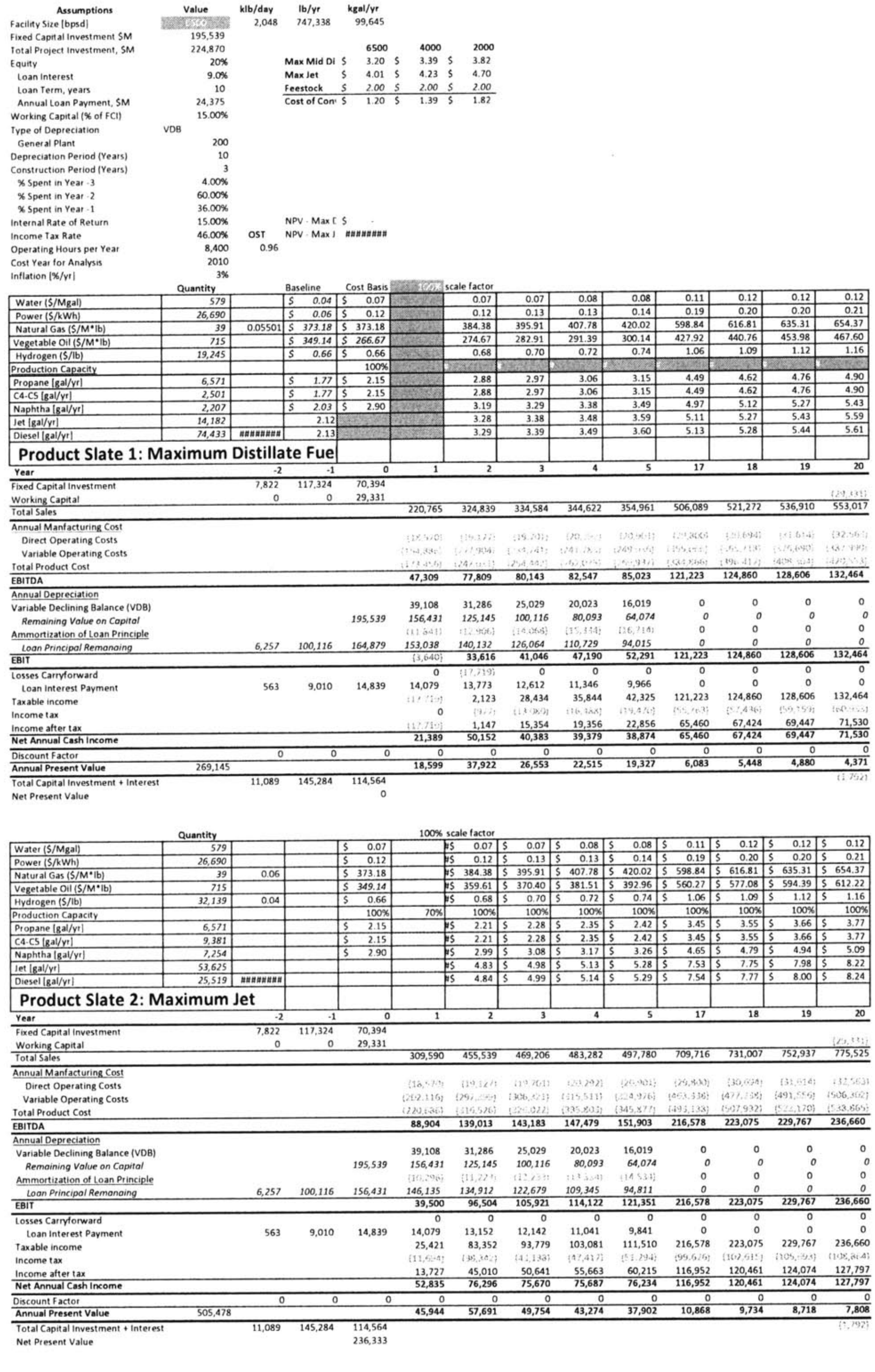
Appendix B

Excel Model

A screen shot of the DCFROR Excel model used for the economic analysis is shown in figure B-1. The semantics of the model are encoded with colors: blue cells are numbers that should be changed, such as plant size or utilization; yellow fields numbers that are changed for sensitivity analysis such as financing or interest rate; white fields are calculated. Note that years 6-16 are hidden to allow for the last years of the model to be shown in the figure. This obfuscates that the financing and depreciation end in years 9 and 10 respectively (not shown).

The general organization of the model is as follows. Assumptions for plant size, financing, and contruction are in the upper right hand corner. Quantities and prices for inputs, utilities, and products are below the assumptions and are scaled by the inflation factor for each year. The cash flow for each fiscal year is shown below the costs and quantities. The same arrangement is used for the maximum jet fuel production model with the appropriate changes to the quantities of hydrogen, natural gas, and product distributions.

There were several other linked spreadsheets that populated this model including digitized cost curves with polynomial functions, material and energy balances that scaled with plant size, and historic product and input costs that are not shown in this screenshot. For access to the model, or for more information, please contact the author at, pearlson@alum.mit.edu.



Assumptions	Value	klb/day	lb/yr	kgal/yr
Facility Size [bpsd]	6500	2,048	747,338	99,645
Fixed Capital Investment $M	195,539			
Total Project Investment, $M	224,870			
Equity	20%			
Loan Interest	9.0%			
Loan Term, years	10			
Annual Loan Payment, $M	24,375			
Working Capital (% of FCI)	15.00%			
Type of Depreciation	VDB			
General Plant	200			
Depreciation Period (Years)	10			
Construction Period (Years)	3			
% Spent in Year -3	4.00%			
% Spent in Year -2	60.00%			
% Spent in Year -1	36.00%			
Internal Rate of Return	15.00%			
Income Tax Rate	46.00%	OST		
Operating Hours per Year	8,400	0.96		
Cost Year for Analysis	2010			
Inflation [%/yr]	3%			

	6500	4000	2000
Max Mid Di	$ 3.20	$ 3.39	$ 3.82
Max Jet	$ 4.01	$ 4.23	$ 4.70
Feestock	$ 2.00	$ 2.00	$ 2.00
Cost of Con	$ 1.20	$ 1.39	$ 1.82

NPV - Max D $ -

NPV - Max J ########

	Quantity		Baseline	Cost Basis	100%	scale factor							
Water ($/Mgal)	579		$ 0.04	$ 0.07		0.07	0.07	0.08	0.08	0.11	0.12	0.12	0.12
Power ($/kWh)	26,690		$ 0.06	$ 0.12		0.12	0.13	0.13	0.14	0.19	0.20	0.20	0.21
Natural Gas ($/M*lb)	39	0.05501	$ 373.18	$ 373.18		384.38	395.91	407.78	420.02	598.84	616.81	635.31	654.37
Vegetable Oil ($/M*lb)	715		$ 349.14	$ 266.67		274.67	282.91	291.39	300.14	427.92	440.76	453.98	467.60
Hydrogen ($/lb)	19,245		$ 0.66	$ 0.66		0.68	0.70	0.72	0.74	1.06	1.09	1.12	1.16
Production Capacity				100%									
Propane [gal/yr]	6,571		$ 1.77	$ 2.15		2.88	2.97	3.06	3.15	4.49	4.62	4.76	4.90
C4-C5 [gal/yr]	2,501		$ 1.77	$ 2.15		2.88	2.97	3.06	3.15	4.49	4.62	4.76	4.90
Naphtha [gal/yr]	2,207		$ 2.03	$ 2.90		3.19	3.29	3.38	3.49	4.97	5.12	5.27	5.43
Jet [gal/yr]	14,182		2.12			3.28	3.38	3.48	3.59	5.11	5.27	5.43	5.59
Diesel [gal/yr]	74,433	########	2.13			3.29	3.39	3.49	3.60	5.13	5.28	5.44	5.61
Product Slate 1: Maximum Distillate Fue													
Year		-2	-1	0	1	2	3	4	5	17	18	19	20
Fixed Capital Investment		7,822	117,324	70,394									
Working Capital		0	0	29,331									(29,331)
Total Sales					220,765	324,839	334,584	344,622	354,961	506,089	521,272	536,910	553,017
Annual Manfacturing Cost													
Direct Operating Costs					(18,570)	(19,127)	(19,701)	(20,292)	(20,901)	(29,800)	(30,694)	(31,614)	(32,563)
Variable Operating Costs					(154,886)	(227,904)	(234,741)	(241,783)	(249,036)	(355,066)	(365,718)	(376,690)	(387,990)
Total Product Cost					(173,456)	(247,031)	(254,442)	(262,075)	(269,937)	(384,866)	(396,412)	(408,304)	(420,553)
EBITDA					**47,309**	**77,809**	**80,143**	**82,547**	**85,023**	**121,223**	**124,860**	**128,606**	**132,464**
Annual Depreciation													
Variable Declining Balance (VDB)					39,108	31,286	25,029	20,023	16,019	0	0	0	0
Remaining Value on Capital				*195,539*	*156,431*	*125,145*	*100,116*	*80,093*	*64,074*	*0*	*0*	*0*	*0*
Ammortization of Loan Principle					(11,641)	(12,906)	(14,068)	(15,334)	(16,714)	0	0	0	0
Loan Principal Remanaing		*6,257*	*100,116*	*164,879*	*153,038*	*140,132*	*126,064*	*110,729*	*94,015*	*0*	*0*	*0*	*0*
EBIT					(3,640)	**33,616**	**41,046**	**47,190**	**52,291**	**121,223**	**124,860**	**128,606**	**132,464**
Losses Carryforward					0	(17,719)	0	0	0	0	0	0	0
Loan Interest Payment		563	9,010	14,839	14,079	13,773	12,612	11,346	9,966	0	0	0	0
Taxable income					(17,719)	2,123	28,434	35,844	42,325	121,223	124,860	128,606	132,464
Income tax					0	(977)	(13,080)	(16,488)	(19,470)	(55,763)	(57,436)	(59,159)	(60,933)
Income after tax					(17,719)	1,147	15,354	19,356	22,856	65,460	67,424	69,447	71,530
Net Annual Cash Income					**21,389**	**50,152**	**40,383**	**39,379**	**38,874**	**65,460**	**67,424**	**69,447**	**71,530**
Discount Factor		0	0	0	0	0	0	0	0	0	0	0	0
Annual Present Value	269,145				**18,599**	**37,922**	**26,553**	**22,515**	**19,327**	**6,083**	**5,448**	**4,880**	**4,371**
Total Capital Investment + Interest		11,089	145,284	114,564									(1,752)
Net Present Value				0									

	Quantity				100% scale factor								
Water ($/Mgal)	579			$ 0.07		#$ 0.07	$ 0.07	$ 0.08	$ 0.08	$ 0.11	$ 0.12	$ 0.12	$ 0.12
Power ($/kWh)	26,690			$ 0.12		#$ 0.12	$ 0.13	$ 0.13	$ 0.14	$ 0.19	$ 0.20	$ 0.20	$ 0.21
Natural Gas ($/M*lb)	39	0.06		$ 373.18		#$ 384.38	$ 395.91	$ 407.78	$ 420.02	$ 598.84	$ 616.81	$ 635.31	$ 654.37
Vegetable Oil ($/M*lb)	715			$ 349.14		#$ 359.61	$ 370.40	$ 381.51	$ 392.96	$ 560.27	$ 577.08	$ 594.39	$ 612.22
Hydrogen ($/lb)	32,139	0.04		$ 0.66		#$ 0.68	$ 0.70	$ 0.72	$ 0.74	$ 1.06	$ 1.09	$ 1.12	$ 1.16
Production Capacity				100%	70%	100%	100%	100%	100%	100%	100%	100%	100%
Propane [gal/yr]	6,571			$ 2.15		#$ 2.21	$ 2.28	$ 2.35	$ 2.42	$ 3.45	$ 3.55	$ 3.66	$ 3.77
C4-C5 [gal/yr]	9,381			$ 2.15		#$ 2.21	$ 2.28	$ 2.35	$ 2.42	$ 3.45	$ 3.55	$ 3.66	$ 3.77
Naphtha [gal/yr]	7,254			$ 2.90		#$ 2.99	$ 3.08	$ 3.17	$ 3.26	$ 4.65	$ 4.79	$ 4.94	$ 5.09
Jet [gal/yr]	53,625					#$ 4.83	$ 4.98	$ 5.13	$ 5.28	$ 7.53	$ 7.75	$ 7.98	$ 8.22
Diesel [gal/yr]	25,519	########				#$ 4.84	$ 4.99	$ 5.14	$ 5.29	$ 7.54	$ 7.77	$ 8.00	$ 8.24
Product Slate 2: Maximum Jet													
Year		-2	-1	0	1	2	3	4	5	17	18	19	20
Fixed Capital Investment		7,822	117,324	70,394									
Working Capital		0	0	29,331									(29,331)
Total Sales					309,590	455,539	469,206	483,282	497,780	709,716	731,007	752,937	775,525
Annual Manfacturing Cost													
Direct Operating Costs					(18,570)	(19,127)	(19,701)	(20,292)	(20,901)	(29,800)	(30,694)	(31,614)	(32,563)
Variable Operating Costs					(202,116)	(297,399)	(306,321)	(315,511)	(324,976)	(463,336)	(477,238)	(491,556)	(506,302)
Total Product Cost					(220,686)	(316,526)	(326,022)	(335,803)	(345,877)	(493,138)	(507,932)	(523,170)	(538,865)
EBITDA					**88,904**	**139,013**	**143,183**	**147,479**	**151,903**	**216,578**	**223,075**	**229,767**	**236,660**
Annual Depreciation													
Variable Declining Balance (VDB)					39,108	31,286	25,029	20,023	16,019	0	0	0	0
Remaining Value on Capital				*195,539*	*156,431*	*125,145*	*100,116*	*80,093*	*64,074*	*0*	*0*	*0*	*0*
Ammortization of Loan Principle					(10,296)	(11,223)	(12,233)	(13,334)	(14,534)	0	0	0	0
Loan Principal Remanaing		*6,257*	*100,116*	*156,431*	*146,135*	*134,912*	*122,679*	*109,345*	*94,811*	*0*	*0*	*0*	*0*
EBIT					**39,500**	**96,504**	**105,921**	**114,122**	**121,351**	**216,578**	**223,075**	**229,767**	**236,660**
Losses Carryforward					0	0	0	0	0	0	0	0	0
Loan Interest Payment		563	9,010	14,839	14,079	13,152	12,142	11,041	9,841	0	0	0	0
Taxable income					25,421	83,352	93,779	103,081	111,510	216,578	223,075	229,767	236,660
Income tax					(11,694)	(38,342)	(43,138)	(47,417)	(51,294)	(99,626)	(102,615)	(105,693)	(108,864)
Income after tax					13,727	45,010	50,641	55,663	60,215	116,952	120,461	124,074	127,797
Net Annual Cash Income					**52,835**	**76,296**	**75,670**	**75,687**	**76,234**	**116,952**	**120,461**	**124,074**	**127,797**
Discount Factor		0	0	0	0	0	0	0	0	0	0	0	0
Annual Present Value	505,478				**45,944**	**57,691**	**49,754**	**43,274**	**37,902**	**10,868**	**9,734**	**8,718**	**7,808**
Total Capital Investment + Interest		11,089	145,284	114,564									(1,792)
Net Present Value				236,333									

Figure B-1: Screenshot of the DCFROR spreadsheet model

Bibliography

[1] International Energy Agency. *Renewables information 2010: with 2009 data.* The Stationery Office, 2010.

[2] ASTM. ASTM D1655 - Standard Specification for Aviation Turbine Fuels, 2010.

[3] ASTM. ASTM D7566 - Standard Specification for Aviation Turbine Fuel Containing Synthesized Hydrocarbons, 2010.

[4] ASTM. ASTM D975 - Standard Specification for Diesel Fuel Oil, 2010.

[5] Robert E Bailis and Jennifer E Baka. Greenhouse Gas Emissions and Land Use Change from Jatropha Curcas-Based Jet Fuel in Brazil. *Environmental science & technology*, 44(22):8684–8691, October 2010.

[6] Goran Berndes. Bioenergy and water—the implications of large-scale bioenergy production for water use and supply. *Global Environmental Change*, 12(4):253–271, December 2002.

[7] Philip Brasher. Biodiesel industry primed to flourish, 2010.

[8] Robert C Brown and Jennifer Holmgren. Fast Pyrolysis and Bio-Oil Upgrading. In *ACS Symposium Series*, 2009.

[9] R Cascone. Biobutanol–A Replacement for Bioethanol? *Chemical Engineering Progress*, 104(8):S4–S9, 2008.

[10] H Deldari. Suitable catalysts for hydroisomerization of long-chain normal paraffins. *Applied Catalysis A: General*, 293:1–10, September 2005.

[11] Ayhan Demirbas. *The Biodiesel Economy and Biodiesel Policy*, chapter 10, pages 195–208. Number 1. Springer, 1 edition, 2008.

[12] Department Of Treasury Internal Revenue Service. Publication 946: How to Depreciate Property. Technical report, Department of Treasury, Washington, D.C., 2009.

[13] Matthew J. DeWitt, Edwin Corporan, John Graham, and Donald Minus. Effects of Aromatic Type and Concentration in FischerTropsch Fuel on Emissions

Production and Material Compatibility. *Energy & Fuels*, 22(4):2411–2418, July 2008.

[14] Bjorn Donnis, Rasmus Gottschalck Egeberg, Peder Blom, and Kim Gron Knudsen. Hydroprocessing of Bio-Oils and Oxygenates to Hydrocarbons. Understanding the Reaction Routes. *Topics in Catalysis*, 52(3):229–240, January 2009.

[15] HM Draper III. *New oilseed crops for fuels and chemicals: ecological and agricultural considerations.* Phd thesis/dissertation, Washington University, 1982.

[16] Dynamic Fuels LLC. Dynamic Fuels Website, 2010.

[17] G. Gary, J., Handwerk. *Petroleum Refining Technology and Economics.* CRC Press, Boca Raton, 5 edition, 2007.

[18] RG Ginder and ND Paulson. The Growth and Direction of the Biodiesel Industry in the US. 2007.

[19] Global Financial Data. Soya Oil Commodity Prices, 2010.

[20] Marcos Sebastião De Paula Gomes and Maria Silvia Muylaert de Araújo. Biofuels production and the environmental indicators. *Renewable and Sustainable Energy Reviews*, 13(8):2201–2204, 2009.

[21] N. Hallale, I. Moore, and D. Vauk. *Hydrogen: Under New Management*, chapter 26, pages 819–840. Springer, 2006.

[22] MJ Hammer and M HAMMER. Water and Wastewater Technology. *New Jersey*, 1996.

[23] D.R. Hardy, E.J. Beal, G.W. Mushrush, J.M. Hughes, J.H. Wynne, and J.V. Sakran. Biodiesel fuels: the use of soy oil as a blending stock for middle distillate petroleum fuels, 2000.

[24] James I Hileman, Hsin Min Wong, Ian A Waitz, David S Ortiz, James T Bartis, Malcolm A Weiss, and Pearl E Donohoo. Near-Term Feasibility of Alternative Jet Fuels. Technical report, MIT and RAND Corporation, Santa Monica, 2009.

[25] JI Hileman, RW Stratton, and P Donohoo. Energy Content and Alternative Jet Fuel Viability. *Journal of Propulsion and Power*, 26(6):1184–1195, 2010.

[26] Jason Hill, Erik Nelson, David Tilman, Stephen Polasky, and Douglas Tiffany. Environmental, economic, and energetic costs and benefits of biodiesel and ethanol biofuels. *Proceedings of the National Academy of Sciences of the United States of America*, 103(30):11206–10, July 2006.

[27] Z Hoffman. *Simulation and Economic Evaluation of Coal Gasification with SETS Reforming.* Master thesis, Louisiana State University, 2005.

[28] Chang Hsu, Paul Robinson, M. Crews, and B. Shumake. *Practical Advances in Petroleum Processing.* Springer New York, New York, NY, 2006.

[29] C.S. Hsu and P.R. Robinson. *Practical advances in petroleum processing.* Springer New York, New York, 2006.

[30] Qiang Hu, Milton Sommerfeld, Eric Jarvis, Maria Ghirardi, Matthew Posewitz, Michael Seibert, and Al Darzins. Microalgal triacylglycerols as feedstocks for biofuel production: perspectives and advances. *The Plant journal : for cell and molecular biology*, 54(4):621–39, May 2008.

[31] George W Huber and Avelino Corma. Synergies between bio- and oil refineries for the production of fuels from biomass. *Angewandte Chemie (International ed. in English)*, 46(38):7184–201, 2007.

[32] George W. Huber, Sara Iborra, and Avelino Corma. Synthesis of Transportation Fuels from Biomass: Chemistry, Catalysts, and Engineering. *Chemical Reviews*, 106(9):4044–4098, September 2006.

[33] Hong Huo, Michael Wang, Cary Bloyd, and Vicky Putsche. Life-Cycle Assessment of Energy Use and Greenhouse Gas Emissions of Soybean-Derived Biodiesel and Renewable Fuels. *Environmental Science & Technology*, 43(3):750–756, February 2009.

[34] K.G. Joback and R.C. Reid. Estimation of Pure-Component Properties From Group-Contributions. *Chemical Engineering Communications*, 57(1):233–243, 1987.

[35] S.B. Jones, C Valkenburg, C.W. Walton, D.C. Elliott, J.E. Holladay, Pacific Northwest National Laboratory (US), and United States. Dept. Energy. *Production of gasoline and diesel from biomass via fast pyrolysis, hydrotreating and hydrocracking: a design case.* Number February. PNNL, Richland, Washintong, rev. 1. edition, 2009.

[36] Jyllian Kemsley. Treating Sewage For Drinking Water. *Chemical & Engineering News*, 86(4):71–73, 2008.

[37] Márton Krár, Tamás Kasza, Sándor Kovács, Dénes Kalló, and Jenő Hancsók. Bio gas oils with improved low temperature properties. *Fuel Processing Technology*, 92(5):886–892, May 2011.

[38] T.G. Kreutz, E.D. Larson, G. Liu, and R.H. Williams. Fischer-Tropsch fuels from coal and biomass. In *Princeton Environmental Institute, Princeton University, prepared for 25th Annual International Pittsburgh Coal Conference, Pittsburgh, Pa*, number August, 2008.

[39] David Kubička and Luděk Kaluža. Deoxygenation of vegetable oils over sulfided Ni, Mo and NiMo catalysts. *Applied Catalysis A: General*, 372(2):199–208, 2010.

[40] Rohit Kumar, Bharat S. Rana, Rashmi Tiwari, Deepak Verma, Rakesh Kumar, Rakesh K. Joshi, Madhukar O. Garg, and Anil K. Sinha. Hydroprocessing of jatropha oil and its mixtures with gas oil. *Green Chemistry*, 12(12):2232, November 2010.

[41] John R. Lacher. The chemical thermodynamics of organic compounds (Stull, Daniel R.; Westrum, Edgar F.; Sinke, Gerard C.). *Journal of Chemical Education*, 47(4):A300, April 1970.

[42] Delanie Lamprecht. FischerTropsch Fuel for Use by the U.S. Military as Battlefield-Use Fuel of the Future. *Energy & Fuels*, 21(3):1448–1453, May 2007.

[43] E S Lipinsky. Chemicals from biomass: petrochemical substitution options. *Science (New York, N.Y.)*, 212(4502):1465–71, June 1981.

[44] Dorothy Lozowski. Economic Indicators. *Chemical Engineering*, 117(13):2, 2010.

[45] Massachusetts Port Authority. Environmental Data Report, 2009.

[46] Jeremy Mattson. *Structure of the canola and biodiesel industries.* Center of Excellence for Agbiotechnology Oilseed Development Dept. of Agribusiness and Applied Economics North Dakota State University, Fargo N.D., 2007.

[47] M.J. McCall, TL Marker, JA Petri, TN Kalnes, DE Mackowiak, DC Elliott, S. Czernik, and D.R. Shonnard. Opportunities for Biorenewables in Petroleum Refineries. In *The 2006 Spring National Meeting*, pages 1–9, 2006.

[48] E McKean. *The New Oxford American Dictionary.* 2005.

[49] I. Mochida and K.H. Choi. *Current Progress in Catalysts and Catalysis for Hydrotreating*, pages 257–296. Springer, 2006.

[50] RP Morgan and EB Shultz. Fuels and chemicals from novel seed oils. *Chem. Eng. News;(United States)*, 59(36), 1981.

[51] Michael Morweiser, Olaf Kruse, Ben Hankamer, and Clemens Posten. Developments and perspectives of photobioreactors for biofuel production. *Applied microbiology and biotechnology*, 87(4):1291–301, July 2010.

[52] Bryan R. Moser. Biodiesel production, properties, and feedstocks. *In Vitro Cellular & Developmental Biology - Plant*, 45(3):229–266, March 2009.

[53] National Academies. *Liquid Transportation Fuels from Coal and Biomass: Technological Status, Costs, and Environmental Impacts.* The National Academies Press, Washington, D.C., 2009.

[54] National Biodiesel Board. Biodiesel Myths Busted. Technical report, Jefferson City, 2009.

[55] National Biodiesel Board. U.S. Biodiesel Production Capacity, 2009.

[56] Neste Oil Corporation. Press Release Nov 30, 2006: Neste Oil to build a second biodiesel plant at Porvoo, 2006.

[57] Neste Oil Corporation. Press Release Nov 30, 2007: Neste Oil to build a NExBTL Renewable Diesel plant in Singapore, 2007.

[58] Neste Oil Corporation. Press Release Jun 12 2008: Neste Oil to build a NExBTL renewable diesel plant in Rotterdam, 2008.

[59] NREL. Biodiesel and Other Renewable Diesel Fuels (Fact Sheet), 2006.

[60] P. O'Connor. *Kinetics and Mechanisms of Fluid Catalytic Cracking*, pages 169–175. Springer, 2006.

[61] OECD. OECD: Periodical: Oil Gas Coal & Electricity - Quarterly Statistics, http://www.oecd.org. Technical report, OECD, Paris, 2010.

[62] Richard P. Palluzi. *Pilot Plant Design, Construction, and Operation.* Mcgraw-Hill (Tx), 1992.

[63] Surinder Parkash. *Petroleum fuels manufacturing handbook : including specialty products and sustainable manufacturing techniques.* McGraw-Hill, New York, 1st edition, 2010.

[64] Matthew Pearlson. Pearlson Communication Mar 31, 2011 - Air Products and Chemicals, Inc., Jennifer Huebschman, 2011.

[65] Matthew Pearlson. Personal Communication Feb 22, 2011: Kraft Geletin, Harry Vassilakos, Sales Manager, 2011.

[66] Matthew Pearlson. Personal Communication Mar 18, 2011: The Jacobsen, Ryan Standard, Biofuels Specialist, 2011.

[67] Matthew Pearlson. Personal Communication Mar 22, 2011 - Baker Commodities, Victoria Armstrong, finished product sales specialist, 2011.

[68] Matthew Pearlson. Personal Communication Mar 8, 2011: Syntroleum, Ramin Abhari, techinical lead and patent specialist, 2011.

[69] J. Pickett, D. Anderson, D. Bowles, T. Bridgwater, P. Jarvis, N. Mortimer, M. Poliakoff, and J. Woods. Sustainable Biofuels: Prospects and Challenges. *The Royal Society, London, UK*, (January), 2008.

[70] D Pimentel. Biofuel Food Disasters and Cellulosic Ethanol Problems. *Bulletin of Science, Technology & Society*, 29(3):205–212, April 2009.

[71] L. Rantanen, R. Linnaila, P. Aakko, and T. Harju. NExBTL- Biodiesel Fuel of the Second Generation, SAE Technical Paper 2005-01-3771. Technical report, San Antonio, TX, 2005.

[72] Robert Rapier. *Renewable Diesel*, chapter 7, pages 153–171. Springer Science+Business Media, 2008.

[73] Michael Reed. Increasing Security and Reducing Carbon Emissions of the U.S. Transportation Sector: A Transformational Role for Coal with Biomass. Technical report, NETL, 2007.

[74] Arnold W. Reitze. Should the Clean Air Act Be Used to Turn Petroleum Addicts into Alcoholics? *Environmental Forum, p. 50, July/August 2007*, 2007.

[75] Economic News Release. U . S . Bureau of Labor Statistics Economic News Release area , seasonally adjusted. *Statistics*, pages 1–2, 2009.

[76] P. Robinson and G. Dolbear. *Hydrotreating and Hydrocracking: Fundamentals*, pages 177–218. Springer, 2006.

[77] Liliana Rodolfi, Graziella Chini Zittelli, Niccolò Bassi, Giulia Padovani, Natascia Biondi, Gimena Bonini, and Mario R Tredici. Microalgae for oil: strain selection, induction of lipid synthesis and outdoor mass cultivation in a low-cost photobioreactor. *Biotechnology and bioengineering*, 102(1):100–12, January 2009.

[78] M Rosen. Thermodynamic comparison of hydrogen production processes. *International Journal of Hydrogen Energy*, 21(5):349–365, May 1996.

[79] Ilenia Rossetti, Chiara Gambaro, and Vincenzo Calemma. Hydrocracking of long chain linear paraffins. *Chemical Engineering Journal*, 154(1-3):295–301, November 2009.

[80] E.G. Shay. Diesel fuel from vegetable oils: status and opportunities. *Biomass and Bioenergy*, 4(4):227–242, 1993.

[81] S.T. Sie. Acid-catalyzed cracking of paraffinic hydrocarbons. 2. Evidence for the protonated cyclopropane mechanism from catalytic cracking experiments. *Industrial & engineering chemistry research*, 32(3):397–402, 1993.

[82] Anoop Singh and Stig Irving Olsen. A critical review of biochemical conversion, sustainability and life cycle assessment of algal biofuels. *Applied Energy*, January 2011.

[83] M Specter. The last drop: Confronting the possibility of a global catastrophe. *The New Yorker Magazine*, 23:60–71, 2006.

[84] Seethamraju Srinivas, S.M. Mahajani, and R.K. Malik. Reactive Distillation for Fischer- Tropsch Synthesis: Simulation-Based Design Methodology Using Aspen Plus. *Industrial & Engineering Chemistry Research*, (iii), 2010.

[85] RW Stratton. *Life Cycle Assessment of Greenhouse Gas Emissions and Non-CO2 Combustion Effects from Alternative Jet Fuels, Thesis available: http://hdl.handle.net/1721.1/59694*. PhD thesis, 2010.

[86] Systems Assessment Section Center for Transporation Resaerch - Argonne National Laboratory. GREET, The Greenhouse Gases, Regulated Emissions, and Energy Use in Transportation (GREET) Model, version 1.8D, 2010.

[87] UOP. Opportunities for Biorenewables in oil refineries (report available http://www.osti.gov/bridge/servlets/purl/861458-Wv5uum/861458.pdf). Technical report, Des Plaines, Ill, 2005.

[88] UOP. Renewable Diesel Process Techincal Paper (http://www.uop.com/wp-content/uploads/2011/01/UOP-Hydrorefining-Green-Diesel-Tech-Paper.pdf). Technical report, 2008.

[89] UOP. Controlling production of transportation fuels from renewable feedstocks (patent: WO 2009/151692 A2), 2009.

[90] US DOE. Diamond Green Diesel: Diversifying Our Transportation Fuel Supply — DOE Blog (press release, available: http://blog.energy.gov), 2011.

[91] US Energy Information Administration. How much renewable energy do we use (available, http://tonto.eia.doe.gov/energy_in_brief/), 2010.

[92] US Energy Information Administration. Official Energy Statistics from the U.S. Government, 2010.

[93] USDA. USDA Rural Development-RD Home (http://www.rurdev.usda.gov/Home.html), 2011.

[94] Qiang WANG, Peisheng MA, Chang WANG, and Shuqian XIA. Position Group Contribution Method for Predicting the Normal Boiling Point of Organic Compounds. *Chinese Journal of Chemical Engineering*, 17(2):254–258, April 2009.

[95] J C Weissman, R P Goebel, and J R Benemann. Photobioreactor design: Mixing, carbon utilization, and oxygen accumulation. *Biotechnology and bioengineering*, 31(4):336–44, March 1988.

[96] Jens Weitkamp, Peter A. Jacobs, and Johan A. Martens. Isomerization and hydrocracking of C9 through C16 n-alkanes on Pt/HZSM-5 zeolite. *Applied Catalysis*, 8(1):123–141, October 1983.

[97] R. H. Wijffels and M. J. Barbosa. An Outlook on Microalgal Biofuels. *Science*, 329(5993):796–799, August 2010.

[98] R.H. Wijffels, M.J. Barbosa, and M.H.M. Eppink. Microalgae for the production of bulk chemicals and biofuels. *Biofuels, Bioproducts and Biorefining*, 4(3):287–295, 2010.

[99] G. Wiltsee. EPA Report: Urban waste grease resource assessment (available: http://www.epa.gov/region9/waste/biodiesel/docs/NRELwaste-grease-assessment.pdf). Technical report, NREL, 1998.

[100] World Health Organization. Desalination for Safe Water Supply: Guidance for the Health and Environmental Aspects Applicable to Desalination. Technical report, 2007.

[101] Mark M. Wright, Daren E. Daugaard, Justinus A. Satrio, and Robert C. Brown. Techno-economic analysis of biomass fast pyrolysis to transportation fuels. *Fuel*, 89:S2–S10, November 2010.

[102] S Zhang. Hydroisomerization of normal hexadecane with platinum-promoted tungstate-modified zirconia catalysts. *Applied Catalysis A: General*, 193(1-2):155–171, February 2000.

[103] S Zhang. Anion-modified zirconia: effect of metal promotion and hydrogen reduction on hydroisomerization of n-hexadecane and Fischer–Tropsch waxes. *Fuel Processing Technology*, 69(1):59–71, January 2001.